Process Design of Oil and Gas Separators and Scrubbers

API RECOMMENDED PRACTICE 12J
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1 Scope

This recommended practice covers minimum requirements for the process design of pressure vessel based gas-liquid and gas-liquid-liquid separators operating on static facilities and facilities subject to motion such as floating production, storage, and offloading (FPSO) and tension leg platforms. Separators covered by this document may be vertical (e.g. scrubbers, knockout drums) or horizontal (e.g. free water knockout) and are limited to those containing conventional, longstanding, separation internals such as inlet devices, mist eliminators, agglomerators, and distribution baffles. Other internals such as solids jetting systems are addressed by company requirements and/or technology supplier designs.

As instrumentation is a key part of separator design, guidelines are given for the pressure, temperature, and level sensors necessary for separator process control and troubleshooting.

While there are well established mechanical design codes for pressure vessels, the process and mechanical designs are related and as such should not be performed as separate exercises. General steps in establishing the parameters to consider when designing the mechanical aspects of the separator are also included.

Separators outside the scope of this document include spherical separators, flare knockout drums, electrostatic separators, inline separators, mono-cyclone separators, filter separators, de-oiling equipment, and de-sanding equipment.

2 Normative References

No normative documents are referred to in this document.

3 Terms, Definitions, Abbreviations and Symbols

3.1 Terms and Definitions

3.1.1
Company
The organization that specifies and purchases a separator for their own use or an intermediary that specifies and purchases the separator for an end user. An intermediary can be a contractor, a packager, a manufacturer, or another similar entity.

3.1.2
Technology Supplier
The organization that designs the internal components for a separator and may also provide the complete process design for the separator.

3.2 Abbreviations and Symbols

A_f	Liquid film flow cross sectional area (m^2)
$A_{g,h}$	Horizontal vessel gas space area (m^2)
$A_{g,v}$	Vertical vessel cross-sectional area (m^2)
A_N	Inlet nozzle area (m^2)
BOV	Bottom of vessel (horizontal vessel) (m)
BTL	Bottom tangent line (vertical vessel) (m)
C_D	Drag coefficient (--)
CFD	Computational Fluid Dynamics

D	Vessel ID (m)
D_g	Gas outlet nozzle ID (m)
D_l	Liquid outlet nozzle ID (m)
D_{in}	Inlet nozzle ID (m)
D_h	Hydraulic diameter (m)
D_P	Inlet pipe ID (m)
ΔP	Total pressure drop across mist eliminator (Pa)
ΔP_d	Drainage pressure drop of mist eliminator (Pa)
d_p	Particle diameter (m)
D_w	Water outlet nozzle ID (m)
FEA	Finite element analysis
F_B	Buoyancy force (N)
F_D	Drag force (N)
F_G	Gravity force (N)
FIV	Flow induced vibration
FPSO	Floating Production, Storage, and Offloading
Fr	Froude number (--)
g	Gravitational acceleration (m/s^2)
g'	Modified gravitational acceleration (m/s^2)
GVF	Gas volume fraction (--)
h	Droplet vertical fall distance (m)
Δh_d	Drain pipe liquid height above vessel liquid level (m)
H_S	Nozzle submergence (m)
H	Gas outlet nozzle-mist eliminator spacings (with subscripts 1 to 4) in a vertical vessel (m)
ID	Inside diameter (m)
K	K factor (m/s) – as defined in Equation 1 or H.3
LSH	High liquid level trip (m)
LAH	High liquid level alarm (m)
NFA	Net free area
NLL	Normal liquid level (m)
LAL	Low liquid level alarm (m)
LSL	Low liquid level trip (m)
LISH	High interface level trip (m)
LIAH	High interface level alarm (m)
NIL	Normal interface level (m)
LIAL	Low interface liquid level alarm (m)
LISL	Low interface level trip (m)
L_B	Baffle spacing (m)
LG	Level gauge
$L_{g, eff}$	Gas space effective separation length (m)
L_H	Droplet horizontal travel distance (m)
L_{IVD}	IVD Length (m)
$L_{l, eff}$	Liquid space effective separation length (2-phase separator) (m)
$L_{oiw, eff}$	Water space effective separation length (m)
L_N	Spacing from closest edge of gas outlet nozzle to the outlet of the mist eliminator (m)
L_P	Length of inlet pipe (m)
LT	Level transmitter
L_{TT}	Horizontal vessel tangent to tangent length (m)
$L_{wio, eff}$	Oil space effective separation length (m)
M	Mass flow rate (kg/s)
M_g	Gas mass flow rate (kg/s)
M_o	Oil mass flow rate (kg/s)
M_w	Water mass flow rate (kg/s)
MMscf	Standard ft^3 x 10^6 = Millions standard ft^3
MSm^3	Standard m^3 x 10^6 = Millions Standard m^3

MSm^3/d	Standard m^3 x 10^6 / day = Millions Standard m^3 per day
N_μ	Viscosity number, (--)
P	Pressure (N/m^2)
P_w	Wetted perimeter (m)
Q	Volumetric flow rate (m^3/s)
Q_g	Gas volumetric flow rate (m^3/s)
Q_o	Oil volumetric flow rate (m^3/s)
Q_w	Water volumetric flow rate (m^3/s)
Re	Reynolds number (--)
Re_f	Liquid Film Reynolds number (--)
RVP	Reid vapor pressure (bar)
T	Temperature (deg. C)
TT	Tangent to tangent
TVP	True vapor pressure (bara)
USG	U.S. gallon
VB	Vortex breaker
V_B	Bulk liquid axial velocity
V_g	Gas velocity (m/s)
$V_{g,h}$	Horizontal vessel gas space velocity (m/s)
$V_{g,v}$	Vertical vessel gas space velocity (m/s)
V_m	Mixture velocity (m/s)
V_l	Liquid nozzle velocity (m/s)
V_T	Particle terminal velocity (m/s)
α	Roll or pitch angle (deg.)
θ	Gas outlet nozzle-mist eliminator angles (with subscripts 1 to 3) in a horizontal vessel (deg.)
μ_c	Dynamic viscosity of continuous phase (kg/m/s)
μ_l	Dynamic viscosity of liquid phase (kg/m/s)
ρ_c	Density of continuous phase (kg/m^3)
ρ_d	Density of dispersed droplet phase (kg/m^3)
ρ_g	Gas density (kg/m^3)
ρ_H	Density of heavy phase (kg/m^3)
ρ_l	Liquid density (kg/m^3)
ρ_L	Density of light phase (kg/m^3)
ρ_m	Mixture density (kg/m^3)
σ	Interfacial tension (N/m) – gas-liquid surface tension (σ_{gl}), liquid-liquid (σ_{ow})

4 Design Parameters

4.1 Separation Requirements

Vessels should be designed to achieve a separation performance sufficient to meet the requirements of the downstream processing equipment or end use. Hence, different separators do not necessarily require the same level of performance, depending on the application and the equipment downstream.

Some equipment are designed for bulk separation, in which coarse separation is acceptable, provided the downstream equipment can handle the liquid carryover or gas carryunder. Production separators are a good example of a bulk separator. The gas leaving a production separator is frequently mixed with other gases and normally routed to a compressor suction scrubber. In applications where condensation may occur, targeting 0.1-1.0% of incoming liquid carried over into the gas phase may be acceptable, provided this will not overload the compressor suction scrubber. In other cases, even a total 99.9% separation efficiency of liquid from gas may not be sufficient depending upon the absolute liquid volumes. Similarly, ~5000 ppmw oil in water and/or ~15% (by volume) water in oil may be acceptable if downstream processing can handle the entrained liquid.

Other applications may require better separation as the consequences of carryover are more severe. In some applications, carryover can result in lost or deferred production, a significant financial loss, or can result in damage or degradation of equipment. Scrubbers upstream of a gas glycol dehydration unit are good examples of separators that require a high level of liquid removal efficiency. Poor separation of free liquid hydrocarbons can result in excessive foaming in the absorption column. Foaming can result in solvent losses and poor performance.

Appropriate and sufficient separation performance specifications linked to the application of the separator should be established by company and technology supplier. Typical specifications include:

- Absolute entrainment, e.g. liters/MSm^3 liquid carryover in a gas scrubber, vol% water in oil, ppmv oil in water
- Droplet cut off criteria, e.g. requirements from manufacturers of downstream equipment such as compressors)
- Removal efficiency, e.g. %liquid from gas
- Liquid control volumes, e.g. downstream pump requirements

Over specification of the design should be avoided as it can result in costly designs, rework, and conflicting requirements. For example, degassing of the liquid in a scrubber is rarely critical and vessel design is typically governed by the diameter required for gas-liquid separation and liquid control volume.

The same separator performance requirements should not be used for all services. For example, applying a scrubber specification to a production separator can result in impractical separation efficiencies and/or oversized separators with overdesigned internals.

Commonly quoted high separation performance claims such as 0.1 USG/MMscf (13.4 liters/MSm^3) liquid carryover in gas [1] or > 99% removal of droplets larger than 10μ diameter should not be used by default. These are often impractical and unnecessary, and in many cases difficult to prove. Liquid carryover testing [2] show that the widely quoted 0.1 USG/MMscf specification was only met in a very small percentage of the test conditions.

Specifications should be consistent with the available internals technology. For example, a too rigorous liquid carryover specification may require internals that are very sensitive to fouling or unnecessarily large vessels.

4.2 Testing

Test data may be used to justify or validate designs. However, care should be taken to ensure that the fluids and the conditions employed in the testing are representative of the intended application. This can be accomplished by either using the actual fluids being processed or representative model fluids. See Section 4.4.

4.3 Process Design Cases

The process design of a separator should cover the anticipated operating envelope which can vary depending on the uncertainty in the system. Several process design cases should be developed that cover the entire expected operating envelope of the separator with a minimum of:

- design flowrate
- normal operating flowrate
- minimum flowrate

For example, a compressor suction scrubber is expected to process 3 MSm^3/d of gas. The selected compressor has a maximum throughput of 3.5 MSm^3/d and the flow at the surge control line is 1.5 MSm^3/d. Therefore, the scrubber should be designed as follows:

– design flowrate (3.5 MSm^3/d)
– normal operating flowrate (3 MSm^3/d)
– minimum flowrate (1.5 MSm^3/d)

Other systems can have different ranges of minimum and design flow. Expected ranges in liquid density, watercut, gas volume fraction (GVF), and fluid properties (density, viscosity, surface tension) shall also be evaluated.

No margin is required for sizing criteria values provided in this document. All margins should be accounted for in flowrates and fluid properties of the cases described above.

4.4 Fluid Properties

Fluid densities, viscosities, and interfacial tensions for all phases shall be provided for the flow sizing cases of Section 4.3. Data related to the foaming and emulsion tendency of the oil-water phases should also be supplied.

Note that interfacial tension and hydrocarbon liquid viscosity are difficult to measure and are typically poorly predicted at the desired pressure and temperature. Interfacial tension is a major factor in the estimation of drop/bubble sizes. Liquid viscosity is a determining factor in oil-water separation and degassing. A sensitivity analysis of separator design with these two parameters should be conducted.

4.5 Separator Orientation Selection

The design features of horizontal and vertical separators are summarized in Annex A.

4.6 Mist Eliminator Selection

Performance characteristics for knitted mesh, vane pack, and axial flow cyclone mist eliminators are provided in Annex B.

4.7 Motion Effects

For separators that are installed on offshore floating production facilities, specific design considerations and requirements apply. These specifics along with some background information are presented in Annex C.

4.8 Configurations

Typical configurations of separators and scrubbers are shown in Annex D. Sizing examples using the guidelines in this document are contained in Annex E.

4.9 Process Sizing Data Requirements (Datasheet)

Input process data necessary for designing a separator or scrubber is presented in Annex F in the form of a typical datasheet. Parameters and requirements discussed in the previous Sections 4.1 to 4.8 as well as in Section 5 are listed in the datasheet.

4.10 Inlet Piping

Although inlet piping is not part of the separator design, it does play a key role in the separator's performance. Good flow distribution in a separator's inlet nozzle is vital to the performance of a separator. Bends and/or restrictions in the piping upstream of the separator materially impact the flow within the separator. Recommendations of inlet pipe configurations and lengths given in Annex G should be followed.

5 Separator Sizing Specifications

5.1 General Specifications

5.1.1 Separation Components

These specifications cover two-phase and three-phase horizontal separators/scrubbers and two-phase vertical separators/scrubbers. See Annex A for design aspects of vessel orientation.

Design specifications for separators and scrubbers are categorized into areas corresponding to the main separation components as shown in Figure 1 for horizontal vessels and Figure 2 for vertical vessels. The components are:

a) Nozzles
   1) Inlet nozzle/inlet devices
   2) Gas outlet nozzle
   3) Liquid outlet nozzles
   4) Solids (Sand) outlet nozzles
b) Liquid gravity settling section,
c) Gas gravity settling section, and
d) Gas demisting.

As noted in Section 4.10 and shown in Figure 1 and Figure 2, inlet piping is not part of the separator design although it does play a key role in the separator's performance [3].

The separator shall be designed for safe control and operation and to meet (as applicable):

- liquid in gas specifications
- gas in liquid specifications
- hydrocarbon liquid in water specifications
- water in hydrocarbon liquid specifications

The above can be accomplished by using the information and proven industry practices within this document to design the main areas of a separator as follows.

e) Inlet nozzle/inlet devices shall be designed to:
   1) Mitigate droplet shatter and foam generation
   2) Promote plug flow to the gravity section

   Outlet nozzles shall be designed to:
   3) Remove the desired phase while minimizing entrainment of unwanted phases
   4) Minimize effect on the upstream gravity section or mist eliminator (e.g. short-circuiting, mal-distribution)
   5) Mitigate erosion

f) The liquid gravity section shall be designed to:
   1) Allow for control and safe operation of the vessel (i.e. level control)
   2) Meet gas in liquid specifications
   3) Meet hydrocarbon liquid in water specifications
   4) Meet water in hydrocarbon liquid specifications
   5) Accommodate liquid slugs
   6) Meet the liquid requirements of downstream equipment (e.g. pump)

g) The gas gravity section shall be designed to:
   1) Reduce liquid load to within the operating range of the mist eliminator (if applicable)
   2) Meet the liquid in gas specifications (if applicable, e.g. no mist eliminator)

h) The gas demisting section (if applicable) shall be designed to meet the liquid in gas specifications

Many of the sizing criteria are commonly specified using a K factor (adapted from the original Souders-Brown B capacity parameter [4]). The maximum gas velocity, V_g (m/s), through the gas flow area or mist eliminator is determined from the K factor (m/s) in Equation 1:

$$V_g = K\sqrt{\frac{\rho_l - \rho_g}{\rho_g}} \tag{1}$$

where

ρ_l = liquid density (kg/m^3), and
ρ_g = gas density (kg/m^3).

The physical properties, drop size, and liquid loading have a significant impact on the allowable K factor and may appreciably reduce its value and hence the maximum gas velocity. See Annex H for discussion on the K factor and drop settling models.

Typical configurations/sketches of separators are given in Annex D. Sizing calculation examples using the herein described guidelines are provided in Annex E. An example datasheet for process design data necessary to size and specify internals is given in Annex F.

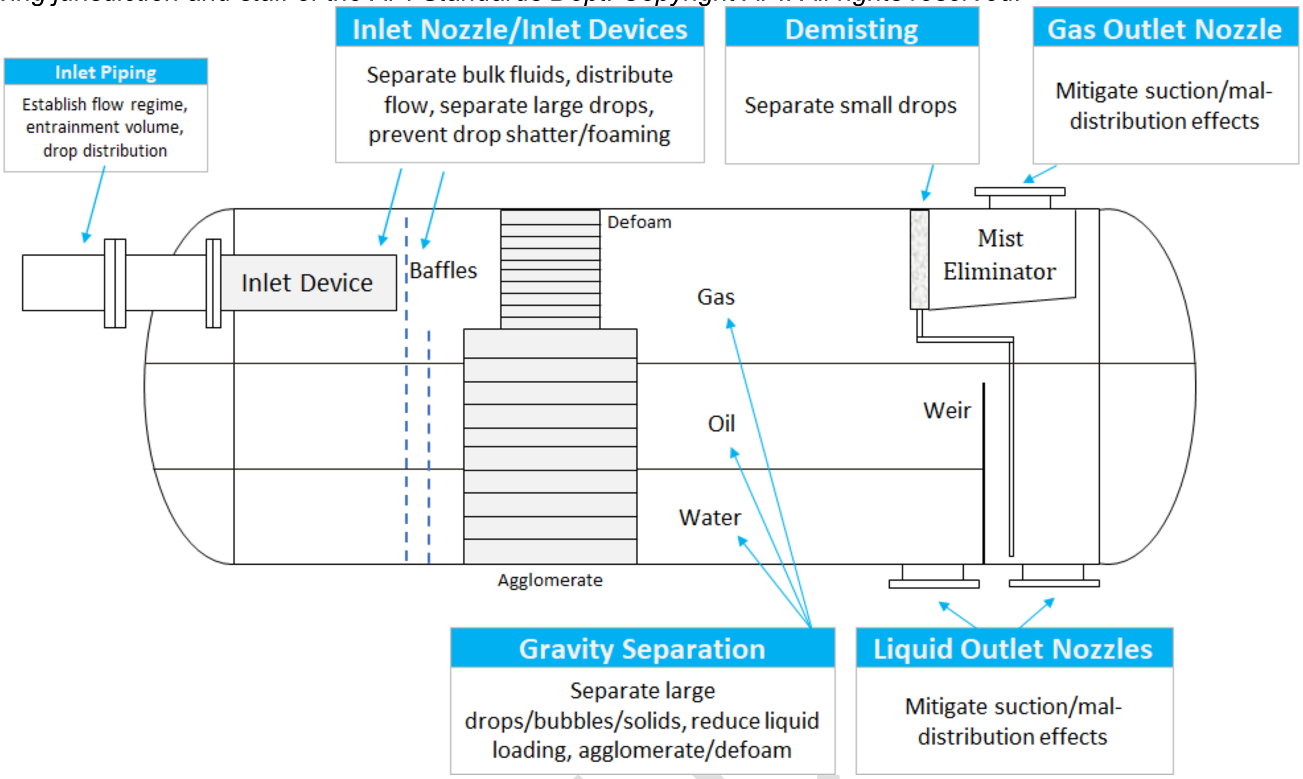


Figure 1 – Key separation areas of a horizontal separator

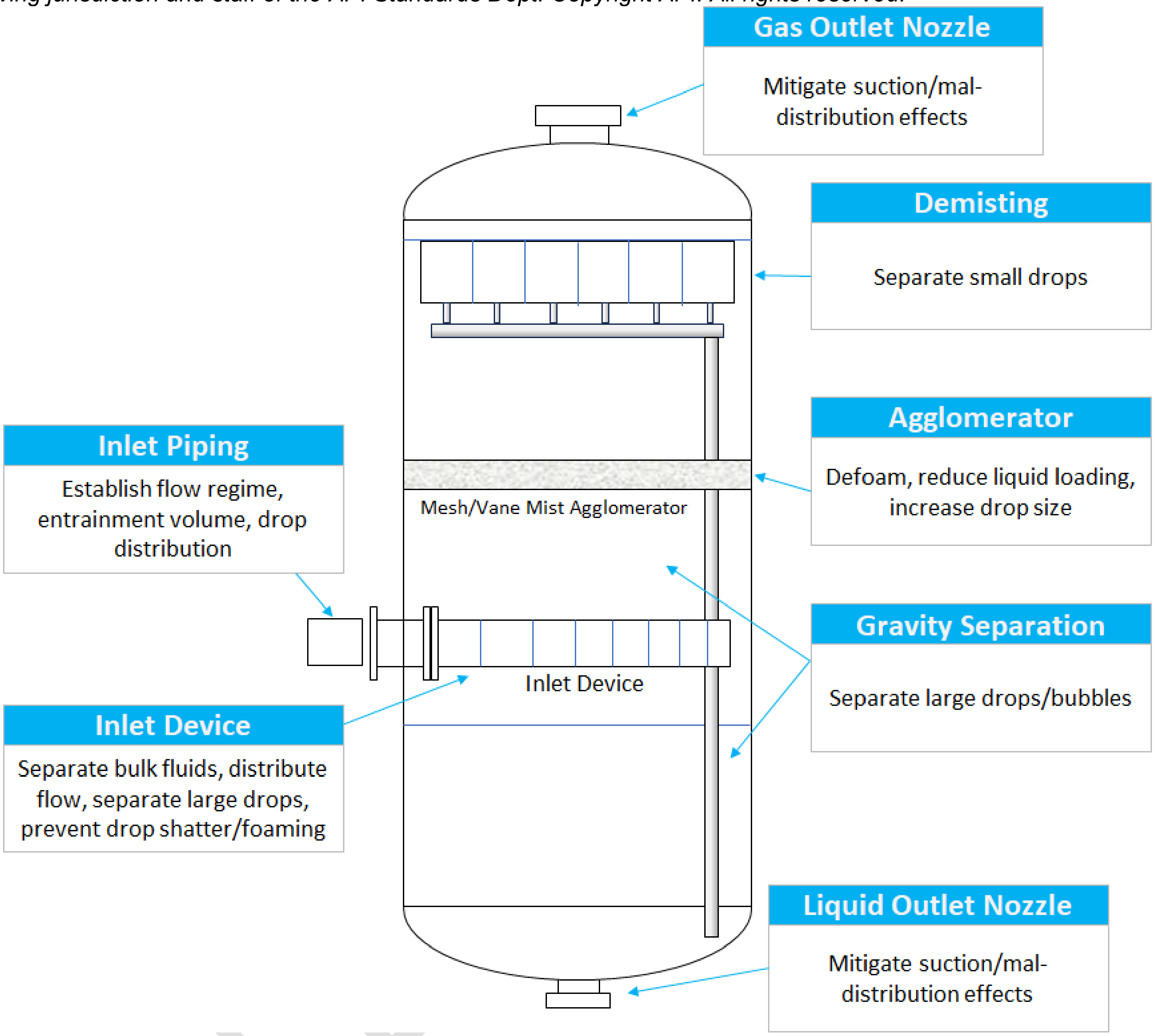


Figure 2 – Key separation areas of a vertical separator

5.1.2 Nozzles

5.1.2.1 Inlet nozzle/inlet devices

The inlet nozzle internal diameter should be sized in conjunction with the selection of the inlet device. The inlet device, depending on the design, can serve many purposes that include separation of bulk liquid from gas, separation of large drops and bubbles, flow distribution into the vessel, and prevention of droplet shatter and foaming. Guidelines for sizing and selecting inlet nozzles and inlet devices are given below.

a) The inlet nozzle internal diameter (ID), D_{in}, shall be at least the size of the inlet pipe ID, D_P. If the inlet nozzle ID is larger than the inlet piping ID, the piping ID should be expanded to the nozzle ID at least

10 pipe diameters upstream of the nozzle (inlet pipe length/diameter $L_P/D_P \geq 10$). The inlet piping configurations of Figure G.*1*, Figure G.2, and Figure G.3 in Annex G still apply.

b) Inlet devices are commonly sized based on a momentum flux, $\rho_m V_m^2$, which is determined assuming homogenous flow. The mixture velocity, V_m (m/s), and density, ρ_m (kg/m^3), are calculated as in Equation 2 and Equation 3:

$$V_m = \frac{Q_o + Q_w + Q_g}{A_N} \tag{2}$$

$$\rho_m = \frac{M_o + M_w + M_g}{Q_o + Q_w + Q_g} \tag{3}$$

where

Q	=	volumetric flow rate at operating P, T (m^3/s)
M	=	Mass flow rate (kg/s)
A_N	=	Inlet nozzle area (m^2)

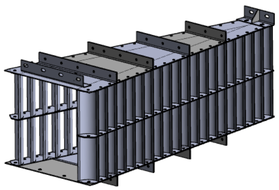
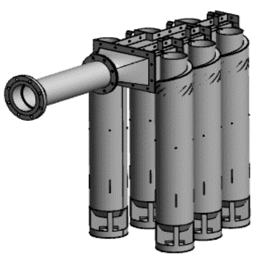
and the subscripts g, o, w refer to gas, oil, and water phases respectively.

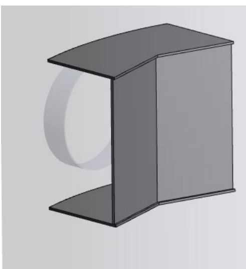
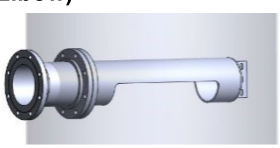
– For typical inlet devices, Table 1 lists the maximum inlet momentum flux and velocity limits that should be followed.

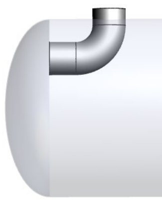
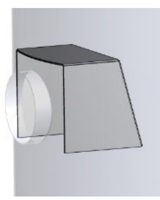
– Some devices, such as inlet cyclones, operate within certain liquid levels to prevent carryunder of the bulk gas phase into the liquid phase and to prevent liquid carryover out the top of the cyclones into the gas phase. The liquid levels in Section 5.1.3.2 shall be checked and adjusted if required to account for these requirements.

c) For horizontal vessels, distribution baffles should be used in conjunction with the inlet device to promote plug flow to the gravity sections. See Section 5.2.2.1.

Table 1 – Maximum allowable inlet momentum fluxes of different inlet devices

Multiphase Inlet device[1,6,8]	**Maximum nozzle Momentum Flux**[5,7] $\rho_m V_m^2$	**Maximum nozzle Velocity**[2,3,5,7] V_m	**Recommendations**[4]		**Comments**
	Pa (kg/m-s^2)	**m/s**	**Vertical Vessel**	**Horizontal Vessel**	
Inlet Vane Diffuser (IVD) 	≤ 6,000 – 10,000 Typical Design Point ≤ 8000 Lower is better for separation	≤ 10 – 20	Recommended Typically spans the vessel ID	Recommended Typically L_{IVD}/D_{in} = 3 - 6	- Generally, the recommended choice for most services, including GVF > 70%. - The design is technology supplier specific, and details shall therefore be confirmed by the technology supplier. - Liquid splashing onto the liquid surface can generate foam. Guides or splashing reduction plates can be evaluated. - Designs for higher momentum (heavy duty) are available based on specific technology supplier evaluations, including FIV. Under slugging service, the momentum flux can be much higher than noted. The higher momentum flux shall be used for mechanical design.
Inlet Cyclone 	Maximum is by technology supplier / service specific Do not exceed erosion limits	Typical ≤ 20	Acceptable *Inlet cyclones are generally not used due to liquid volume storage loss in the bottom of the separator.*	Recommended depending on the specific application	- The design is highly technology supplier specific; design limits can vary and shall be confirmed by the technology supplier in each case. - There is a risk for gas carry-under to the liquid outlets of the device and liquid carryover with the gas. For this reason, the critical submergence and margins shall be well documented and known to the operators. Similarly, capacity / performance for turndown should be documented. Maldistribution between cyclone cans shall be evaluated by technology supplier. - If solids are present in the fluids, the technology supplier should verify the design for internal erosion risk. - Can be used in foaming service.

Multiphase Inlet device[1,6,8]	**Maximum nozzle Momentum Flux**[5,7] $\rho_m V_m^2$	**Maximum nozzle Velocity**[2,3,5,7] V_m	**Recommendations**[4]		**Comments**
	Pa (kg/m-s²)	**m/s**	**Vertical Vessel**	**Horizontal Vessel**	
V-Baffle (shown), **Tangential baffle** 	≤ 1500	≤ 5 – 18	Acceptable *Maldistribution of gas flow and droplet shatter* should *be assessed by CFD.*	Not Recommended	- Inlet impact baffles are high-shear devices which can shatter and mix the inlet flow generating smaller bubbles and droplets potentially making separation more difficult. - Can result in the creation of foam and emulsion. - Vertical Vessels: Can result in maldistribution of gas flow in the vessel diameter and into the demisting device due to swirling flow in the vessel invalidating gravity separation and demisting calculations. - When the flow is directed towards (around) the separator walls, erosion concerns shall be investigated / mitigated.
Half Pipe (with or without Elbow)  *Includes sparger, slotted pipe designs and trapezoidal cut-outs in a pipe*	≤ 1500	≤ 5 – 18	Not Recommended	Acceptable in bulk separation service where stringent separation is not required. Typically GVF ≤ 70%.	- Vertical Vessel: Directing the inlet flow down can result in maldistribution of gas flow in the vessel and invalidate general gravity separation and demisting calculations (e.g., K factor). Velocity estimates should be based on CFD simulations. Downward flow can result in gas impacting liquid level resulting in entrainment from the liquid surface. - Horizontal Vessel: Downward flow can result in gas impacting liquid level resulting in entrainment from the liquid surface. - The pipe opening can be facing up or down; both orientations can be problematical for the vessel. A slotted pipe design can be acceptable for single phase gas inlets or submerged liquid inlets. - For a gas sparger design the holes/slots should be directed to the sides or angled down between 0 – 45 deg. from horizontal.

Multiphase Inlet device[1,6,8]	**Maximum nozzle Momentum Flux[5,7]** $\rho_m V_m^2$	**Maximum nozzle Velocity[2,3,5,7]** V_m	**Recommendations[4]**		**Comments**
	Pa (kg/m-s^2)	**m/s**	**Vertical Vessel**	**Horizontal Vessel**	
Elbow, **Baffle Box (hood)**  	≤ 950 – 1500 Typical ≤ 750	≤ 5– 10	Not Recommended	Acceptable Inlet on shell: where stringent separation is not required. Elbow should direct the flow back to upstream head. Inlet in head: where stringent separation is not required. Full diameter distribution baffle should be present. Typically GVF ≤ 15% for elbow inlets.	- Vertical Vessel: Directing the inlet flow down can result in maldistribution of gas flow in the vessel and invalidate general gravity separation and demisting calculations (e.g., K factor). Velocity estimates should be based on CFD simulations. Downward flow can result in gas impacting liquid level resulting in entrainment from the liquid surface. - Horizontal Vessel: Downward flow can result in liquid entrainment from the liquid surface. - If flow from elbow is directed towards the vessel head or wall a wear plate should be installed where the flow impacts the vessel.
None 	≤ 1000 Typical ≤ 750	≤ 5– 10	Not Recommended	Not Recommended Inlet on shell: gas is directed down at liquid level and falling liquid can result in foaming. Acceptable Inlet in head: Full diameter distribution baffle should be present.	- Vertical Vessel: Can result in inlet flow hitting the opposite wall (erosion concern) and severe maldistribution of gas flow. - Horizontal Vessel: Causes extended turbulent inlet section and reduced separation length.

Multiphase Inlet device[1,6,8]	**Maximum nozzle Momentum Flux**[5,7] $\rho_m V_m^2$	**Maximum nozzle Velocity**[2,3,5,7] V_m	**Recommendations**[4]		**Comments**
	Pa (kg/m-s^2)	**m/s**	**Vertical Vessel**	**Horizontal Vessel**	

Notes:

1. These represent groups of typical inlet device designs. Design details can vary between suppliers, and the sketches are for illustration purposes only. Other types of inlet distributors may also be acceptable, but their suitability shall then be documented and accepted by the company.
2. Always check that erosion velocity is not exceeded, especially when solids are present. In gas dominated, low pressure (typically ≤ 15 barg) and vacuum applications, the maximum velocity can exceed that noted, and is likely limited by erosion limits. Erosion limits in some services may limit fluid velocity below that indicated by the momentum flux limit. Erosion limits are per company guidelines / specifications.
3. In gas dominated, low pressure (typically ≤ 15 barg) and vacuum applications, the maximum velocity may exceed that noted and might be limited by maximum allowed inlet momentum flux.
4. " Acceptable” = requires further verification by company subject matter expert / technology supplier for the specific service.
5. Recommended momentum and velocities are applicable to typical oil and gas operation services. The inlet sizing and distributor selection and sizing should always be verified by company subject matter expert / technology supplier for the specific service.
6. Any inlet which directs the flow towards the liquid level such as an elbow, baffle box, or downward facing impact baffle can cause liquid short-circuiting which can impact oil-water separation. The impact can drive gas into the liquid impacting degassing and potentially causing foaming as well.
7. The maximum momentum fluxes and velocities represent typical values for a system with a low liquid load and moderate physical properties. High liquid loads, high liquid viscosities, and low interfacial tensions can detrimentally impact the performance. The final design of the system, by the technology supplier, should account for these effects for all fluid phases.
8. Fouling and plugging should be considered in the inlet device design.

5.1.2.2 Gas Outlet Nozzle

Gas outlet nozzle sizing is dependent on the demisting technology installed in the vessel and more specifically on the pressure drop, ΔP, of the final demisting device. The sizing is such that the nozzle should not negatively impact gas and liquid distribution into and through the demisting device. The flow distribution is also dependent on the distance the demisting device is from the nozzle in the vessel.

Gas outlet nozzles should be sized per Table 2 where V_g is the outlet nozzle gas velocity and ρ_g is the gas density. The outlet piping diameter should match the nozzle diameter for 3D – 5D downstream of the nozzle before being reduced.

Table 2 – Gas outlet nozzle sizing criteria

Demisting Technology	Gas Outlet Nozzle Momentum Flux Max.(1,2) $\rho_g V_g^2$, Pa (kg/m/s^2)	Gas Outlet Nozzle Velocity Max.(2,3) m/s
None	≤ 4500	20
Mesh or Vanes	≤ 4500	20
Cyclones	≤ 9000 < 5400 (Typically)	20

Notes:

1. To ensure that the nozzle does not impact the flow distribution through the mist eliminator.
2. Erosion limits in some services may limit fluid velocity below that indicated by the momentum limit. Erosion limits are per company guidelines / specifications.
3. In gas dominated, low pressure (typically ≤ 15 barg) and vacuum applications, the maximum velocity may exceed that noted.

5.1.2.3 Liquid Outlet Nozzles

Liquid outlet nozzles should be designed to not negatively affect the performance of the gravity section. In most applications, liquid outlet nozzles should be fitted with vortex breakers (VB). Examples of acceptable designs can be found in IOGP S-619 [7]. Nozzles may be elevated due to solids accumulation. The liquid outlet nozzles should be sized as follows:

a) 2-Phase Gas-Liquid Separators:
   - Hydrocarbon or Mixed Phase Nozzle: 2 m/s velocity maximum

b) 3-Phase Gas-Oil-Water Separators:
   - Hydrocarbon Nozzle: 2 m/s velocity maximum
   - Water Nozzle: 1 m/s velocity maximum

c) Feed to Pump Suction (any service): 1 m/s velocity maximum

In critical separation services and/or where a vortex breaker is not provided, to mitigate risk of carryunder of the phase above, the liquid outlet nozzle should be designed for self-venting flow (see Annex I).

Critical separation services include operations where gas carryunder or where contamination of liquid flows cannot be tolerated.

Velocity and self-venting criteria are normally acceptable for design if the nozzle ID is less than 24 in. However, if the nozzle ID is greater than 24 in., an additional submergence criterion (see Annex I) should be considered to provide a minimum value for low liquid level trip (LSL) [or low interface level trip (LISL)] above the nozzle.

For separators with solids removal systems (e.g., sand jetting), liquid outlets in the same compartment/section should be designed to reduce solids carry-over (e.g. sand dams or elevated nozzles).

5.1.2.4 Solids (Sand) Outlet Nozzles

Solids outlet nozzles are periodically used to remove solids accumulated at the bottom of a vessel.

The ID of the nozzles should be a minimum of DN50 to avoid blockage and the vessel should be fitted with internals to limit the lines filling with solids when not in use.

The exit velocity should be limited to avoid erosion; a typical value is 3 m/s.

5.1.3 Liquid Gravity Settling Section

5.1.3.1 General

The liquid gravity section allows large liquid drops, gas bubbles, and solids to separate out to their respective phases, for example, water drops settle out of the oil phase, oil drops rise out of the water phase, and gas bubbles rise out of the liquid phase. In addition, the liquid section must have sufficient volume to allow proper operation of the vessel.

The liquid section for two or three-phase separators shall meet minimum requirements for:

a) level control/reaction time,
b) degassing and defoaming, and
c) solids settling (if applicable).

For horizontal three-phase separators (see Section 5.2.3), there are additional separation requirements for:

d) velocity limits
e) water droplets settling out of oil, and
f) oil droplets settling out of water.

Liquid levels shall be adjusted for proper operation of the inlet device if present. Some devices, such as inlet cyclones, may need to be liquid sealed [e.g., between low liquid level alarm (LAL) and high liquid level alarm (LAH) for two-phase separators] while other inlet devices should not be submerged.

If the distance of the inlet device above the liquid level is excessive, then risk of foaming due to splashing is higher and should be addressed. Also, in systems with large slug volumes the distance between normal and high level can be large and a large fall height may result. If foaming due to falling liquid is a concern, properly designed mitigation schemes such as downcomer baffles or packing may be used.

The largest of the above requirements should dictate the sizing of the liquid section.

5.1.3.2 Level Control/Reaction Times

Level control volume refers to the volume in a vessel between the settings and alarms on the control system. The vessel liquid volumes shall provide sufficient level control time which allows time for the operators and control systems to respond to disturbances.

Level settings are normally dictated by project specifications, process requirements, and/or safety criteria which may override the recommendations given below. The recommendations below should not be interpreted as limiting the specific project requirements to minimum or maximum values. In the absence of such requirements, Table 3 may be used, provided that the separation performance is met.

- The basis of the flow rate, used to determine the control times, should be clearly stated and is normally the design flow rate.
- Table 3 lists minimum liquid levels for two-phase and three-phase separators where no other overriding criteria are available. Company guidelines should be consulted to determine if some operating levels are not required. See also Figure 3, Figure 4, and Figure 6 through Figure 8.
- See Annex C for vessels subject to motion.

In addition, the level instrumentation for safety shut downs (low or high level) shall be separate from the operating level instruments, as required by the applicable safety standard or practice.

Other level control time/volume requirements for safety systems, downstream equipment protection, etc. should also be considered in the level settings.

Table 3 – Minimum level control spacings and control times

	Level	Minimum Spacing Between Levels (mm)	Minimum Time between Levels (sec)[1]
High Liquid Level Trip[4,5]	LSH		
		150	30
High Liquid Level Alarm[3]	LAH		
		150	30
Normal Liquid Level[3]	NLL		
		150	30
Low Liquid Level Alarm	LAL		
		150	30
Low Liquid Level Trip	LSL[2]		
		See values below for LISL	
Bottom Tan Line (vertical) or Bottom of Vessel[2] (horizontal)	BTL or BOV		
High Interface Level Trip	LISH		
		150	30
High Interface Level Alarm	LIAH		
		150	30
Normal Interface Level	NIL		
		150	30
Low Interface Level Alarm	LIAL		
		150	30
Low Interface Level Trip	LISL[2]		
	If vertical vessel :	Greater of 1)150 mm from BTL, 2)150 mm above elevated outlet, or 3)height of VB[6]	
	If horizontal vessel :	Greater of 1)150 mm from BOV, 2)150 mm from top of elevated outlet, or 3)height of VB[6]	
Bottom Tan Line (vertical) or Bottom of Vessel[2] (horizontal)	BTL or BOV		

Notes:

1. The listed control times include any vessel head space for horizontal vessels.
2. For horizontal vessels, depending upon the outlet design, the minimum LSL and LISL should be the greater of 1) 150 mm above BOV, 2) 150 mm above an elevated outlet, or 3) the height of the vortex breaker. For vertical vessels, depending upon the outlet design, the minimum LSL should be the greater of 1) 150 mm above BTL, 2) 150 mm above an elevated outlet, or 3) the height of the vortex breaker.
3. Where liquid slugs are expected, the separator should be designed to accommodate the volume of the slug within the normal to high liquid level (NLL to LAH). Overall system dynamic behavior including level control response, initial steady state condition, and downstream surge capacity should be considered in the determination of the slug volume to be accommodated.

4. Specifications of LSH spacing from the bottom of the mist eliminator should be followed as shown in Figure 3, Figure 4, and Figure 6 through Figure 8.
5. Specifications of LSH spacing from the bottom of a non-submerged inlet device should be followed as shown in Figure 3, Figure 4, and Figure 6 through Figure 8.
6. The height of a vortex breaker is typically one-half of the nozzle ID.

5.1.3.3 Degassing and foaming

Many common applications require a specific limit on gas entrainment in liquid, and therefore the design needs to incorporate a degassing criterion. Examples include:

- Pumped systems
- Feeds to electrostatic dehydrators
- Systems with RVP/TVP requirements

If degassing is required, a commonly used degassing criteria for critical applications may be used: the rise of a 200 μm gas bubble from the bottom of the vessel to NLL (two-phase) or from NIL to NLL (three-phase). In many applications, this may be relaxed to a 600 μm bubble. See Annex H for bubble rising models. For a two-phase separator containing 2 liquid phases, the highest liquid viscosity should be used in the bubble rise model.

- For horizontal vessels, the effective gas bubble disengagement lengths should be used as shown in Figure 3 and Figure 4.

- For vertical vessels, the calculated bubble rise velocity should exceed the downward liquid bulk velocity.

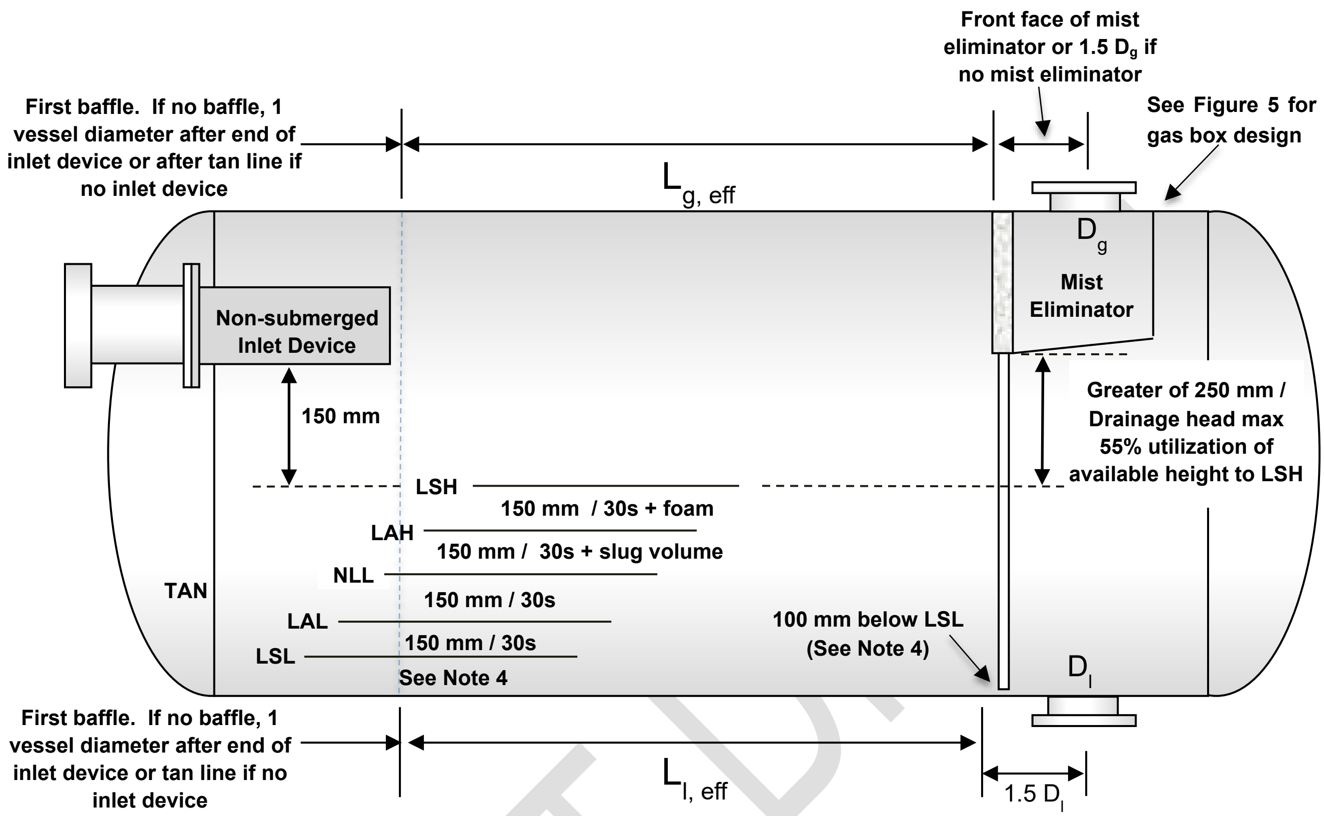
In viscous services, it is not always possible to degas to this level. In such scenarios the requirements downstream of the system shall be assessed to determine the acceptability of residual gas.

If the system is foamy, the spacing from LAH to LSH should be increased by at least 250 mm to accommodate a foam layer.

The selection of inlet devices and internals can have a significant impact on volume of the dispersed gas in the liquid. Where degassing is required, these impacts should be considered in the design.

5.1.3.4 Solids Removal

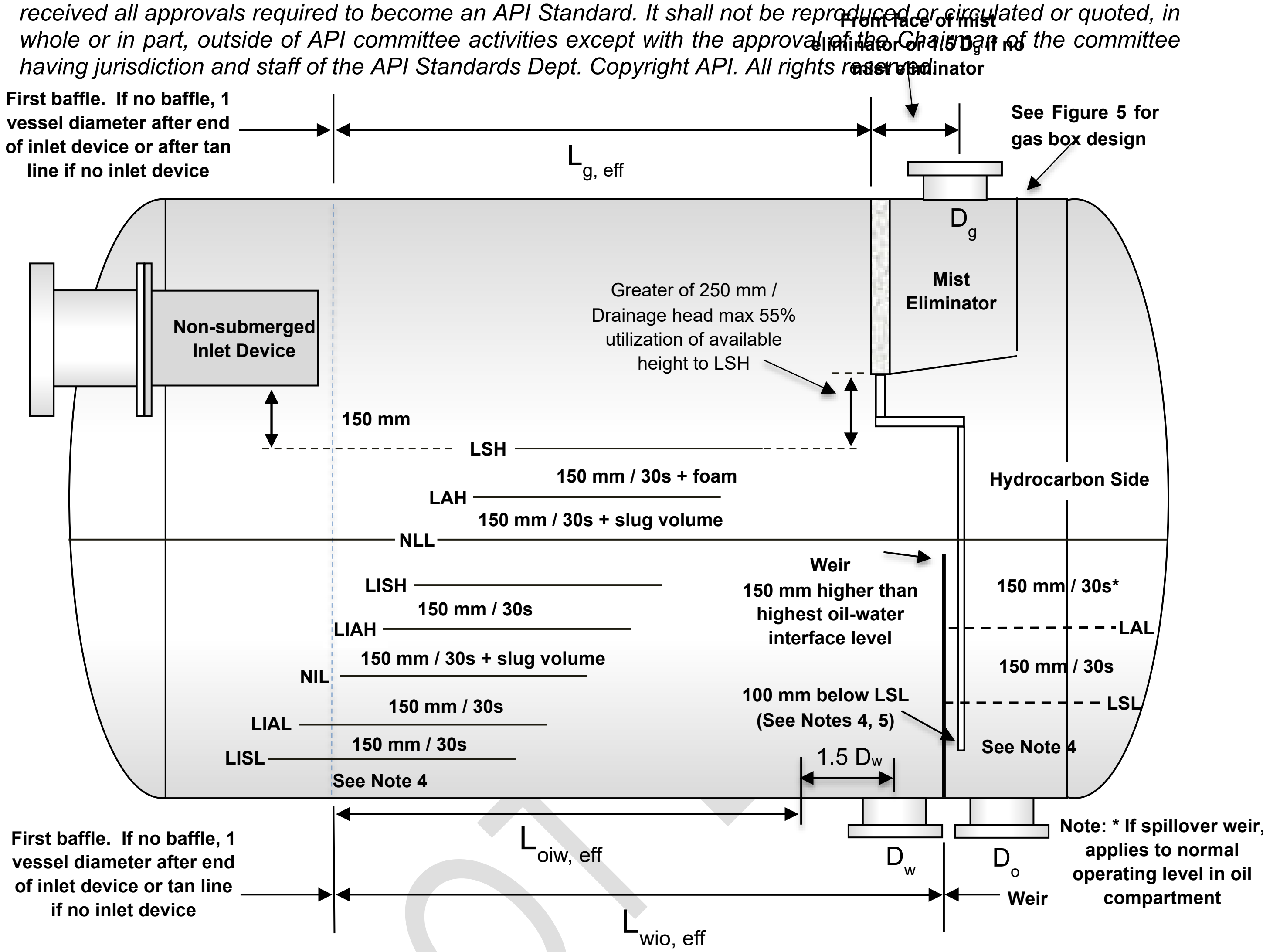
It can be advantageous to remove solids in horizontal production separators. A vessel can be sized in order to provide space for the collected solids and appropriate solids handling infrastructure (e.g. jetting systems, sand dams). Solids are typically not accumulated in vertical separators. However, if the liquid outlet is not placed at the bottom of the separator, or if stagnant sections can be expected, solids accumulation/removal should be evaluated. The design of solids handling systems are per company/technology supplier specifications.



Notes:

1. Control times are based on vessel length including heads
2. Separation lengths are determined from a number of parameters, in particular the inlet device configuration, baffle configuration and nozzle positioning.
   - bubble rise models are based on effective length $L_{l, eff}$
   - droplet in gas settling model is based on effective length $L_{g, eff}$
3. Additional spacing, due to mechanical requirements, may be necessary and should be accounted for in the design
4. The height of LSL from BOV is the greater of 1) 150 mm from BOV, or 2) 150 mm from top of elevated outlet, or 3) height of VB. Solids that may settle at the bottom should also be taken into account in the setting of LSL and the bottom of the drainage pipe.

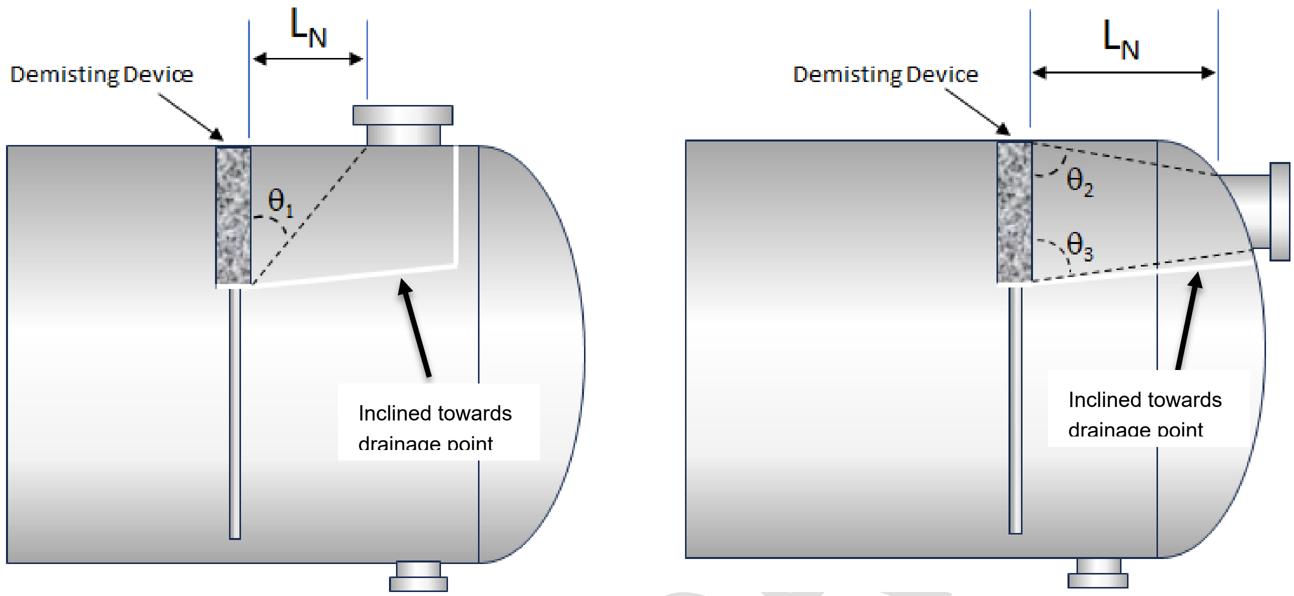
Figure 3 – Overview of specifications for a horizontal two-phase separator



Notes:

1. Control times are based on vessel length including head(s)
2. Separation lengths are determined from several parameters, in particular the inlet device configuration, baffle configuration, and nozzle positioning.
   - water droplet settling in oil and bubble rise degassing model are based on effective length $L_{wio,\ eff}$
   - oil droplet rising in water model is based on effective length $L_{oiw,\ eff}$
   - droplet in gas settling model is based on length $L_{g,\ eff}$
3. Additional spacings, due to mechanical requirements, may be necessary and should be accounted for in the design.
4. The height of LSL from BOV is the greater of 1) 150 mm from BOV, or 2) 150 mm from top of elevated outlet, or 3) height of VB. Solids that may settle at the bottom should also be taken into account in the setting of LSL and the bottom of the drainage pipe.
5. The mist eliminator drains should normally extend to the oil side of the weir. Exceptions should be made when meeting the water in hydrocarbon requirement is challenging and locating the drain upstream of the weir should be considered. In this case, the downcomer should end in the hydrocarbon phase.
6. For a spill-over weir, LAH and LSH can be below or above the weir depending upon the size of the oil compartment.

Figure 4 – Overview of specifications for a horizontal three-phase separator



Where θ_1, θ_2, $\theta_3 \geq$ 45-deg and distance $L_N \geq$ 250 mm

Figure 5 – Gas outlet nozzle spacings from the mist eliminator for a horizontal separator

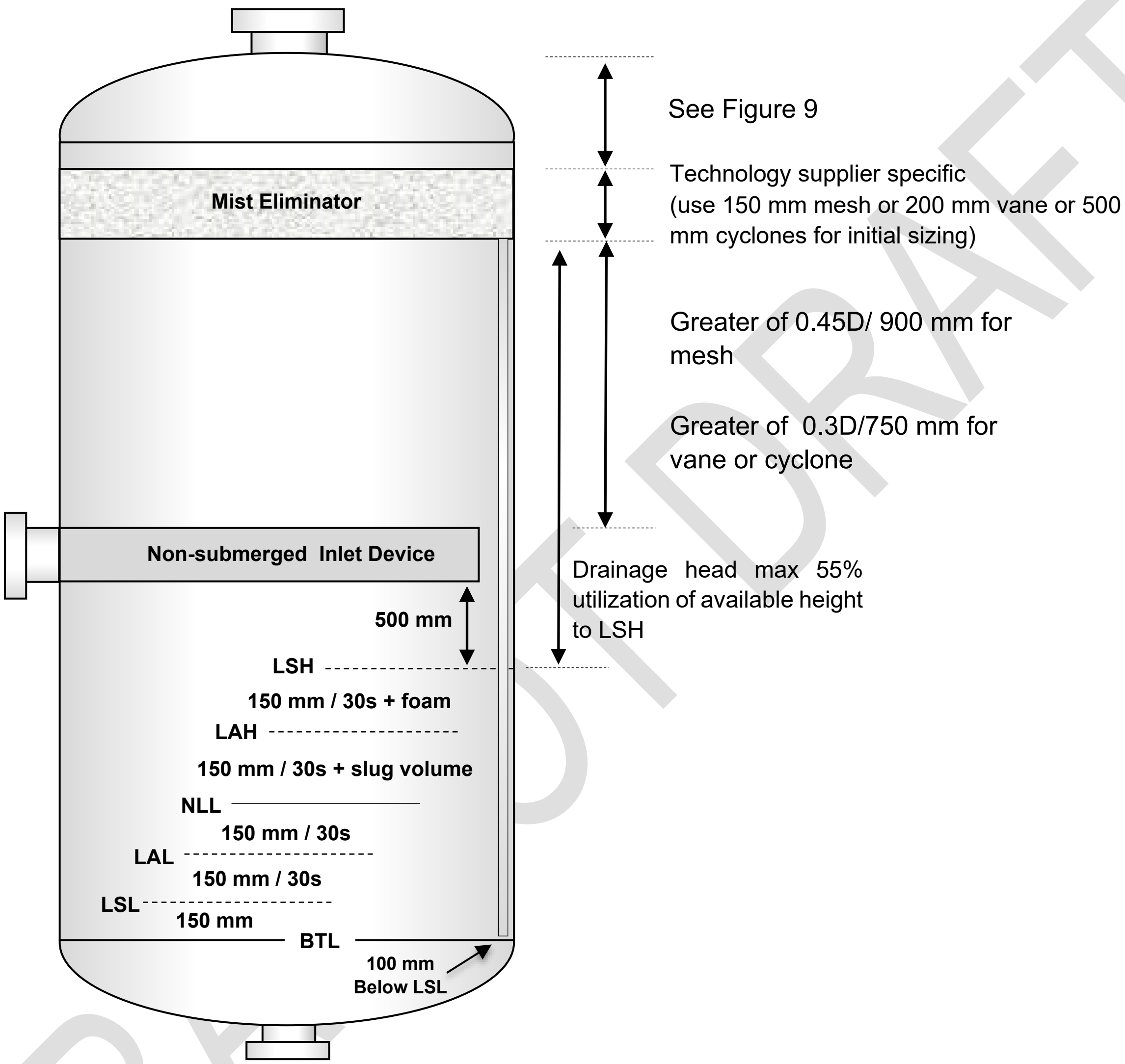
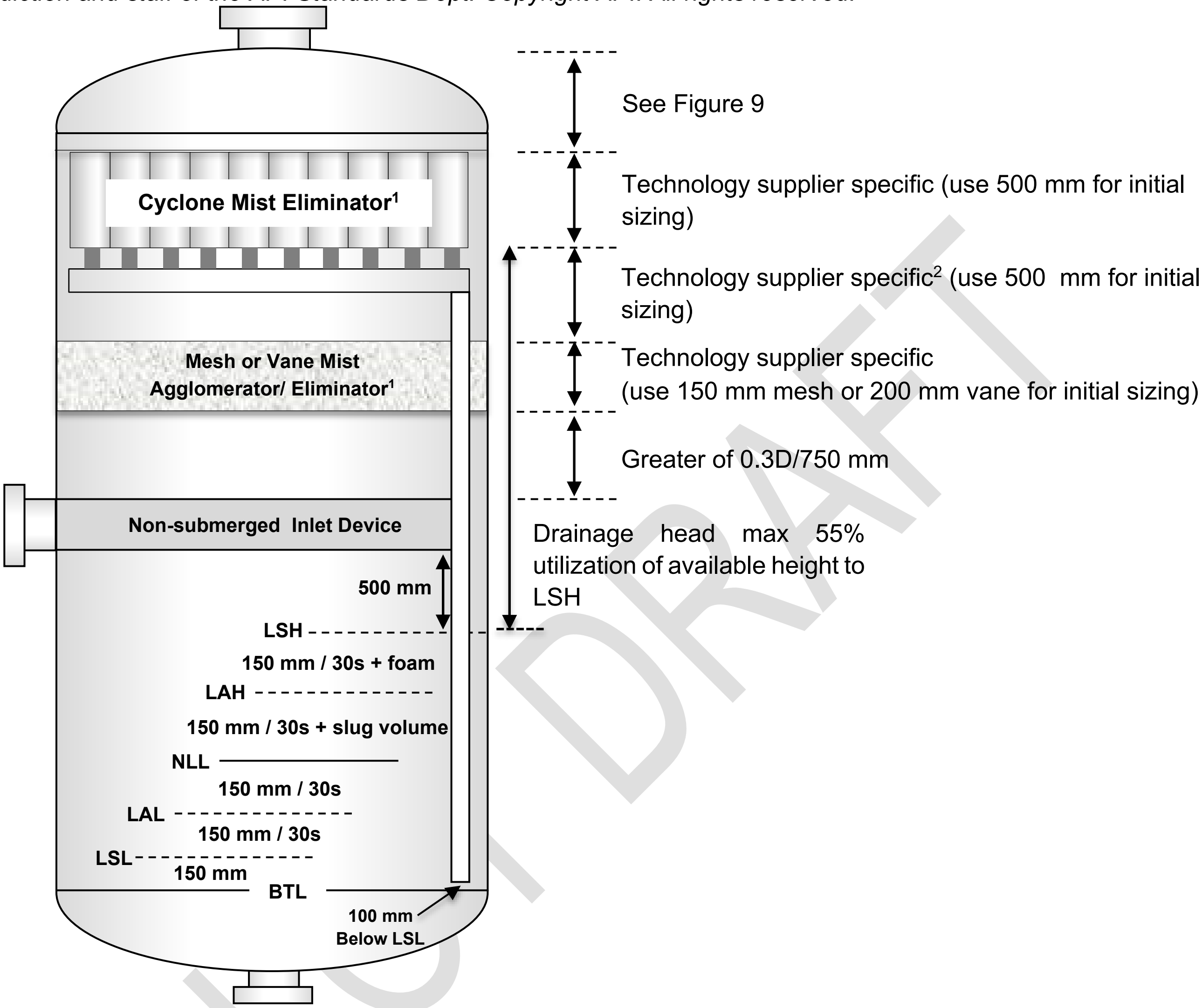


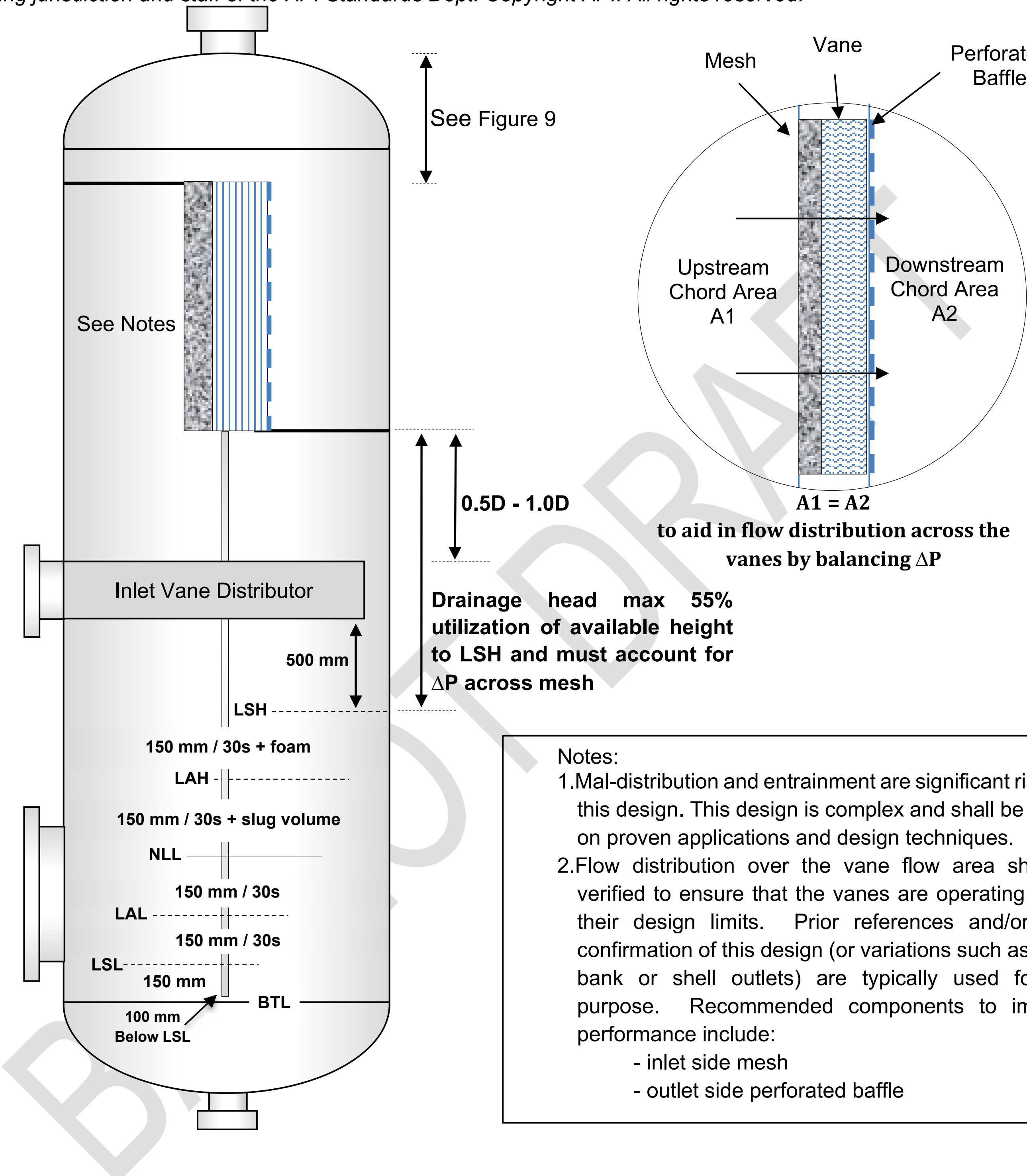
Figure 6 – Overview of specifications for a vertical two-phase separator with a single mist eliminator



Note:

1. Gas demisting capacity shall be determined by the cyclone mist eliminator. The mesh or vane agglomerator/eliminator should normally be full diameter.
2. When the first device is operated in a flooded condition, the minimum distance between the mist eliminators should be 500 mm.

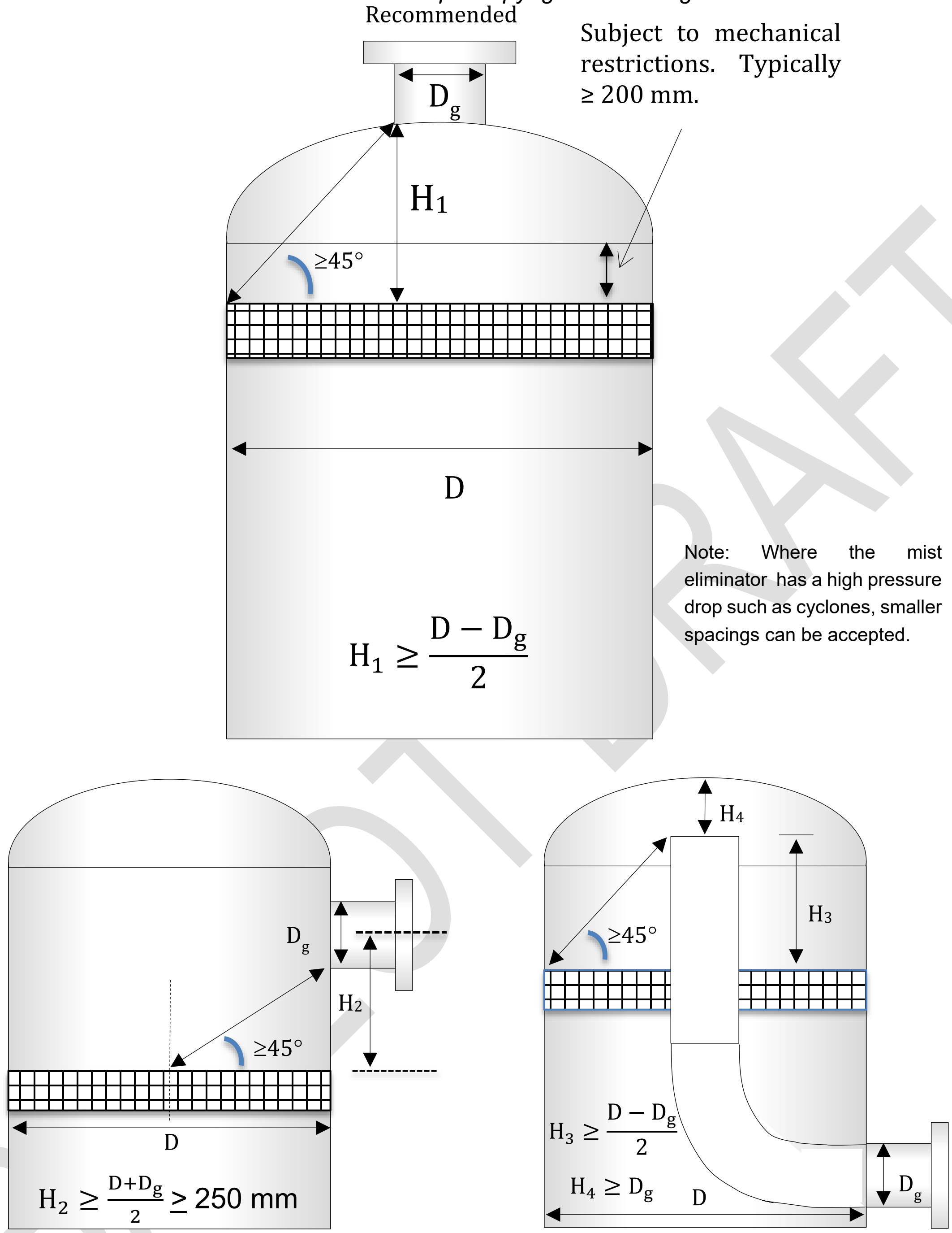
Figure 7 – Overview of specifications for a vertical two-phase separator with dual mist eliminators



Notes:

1. Mal-distribution and entrainment are significant risks for this design. This design is complex and shall be based on proven applications and design techniques.
2. Flow distribution over the vane flow area shall be verified to ensure that the vanes are operating within their design limits. Prior references and/or CFD confirmation of this design (or variations such as multi-bank or shell outlets) are typically used for this purpose. Recommended components to improve performance include:
   - inlet side mesh
   - outlet side perforated baffle

Figure 8 – Overview of specifications for vertical two-phase separator with horizontal flowing vane mist eliminator



Internal pipe or baffles can be used on the nozzle to centralize the gas outlet for distribution of flow on the mist eliminator

Subtract vertical pipe area from available demisting flow area – if horizontal pipe is too close to mist eliminator, it can restrict flow as well

Note: If a partial diameter mist eliminator is used, use the diameter of the mist eliminator instead of vessel diameter D.

Figure 9 – Gas outlet nozzle spacings from the mist eliminator for a vertical separator

5.1.4 Gas Gravity Settling Section

The gas gravity section allows large liquid droplets to fall to the gas-liquid interface.

The required gas space area shall be the larger of the two areas sized by the following criteria:

a) The gas space is sized by a K factor that depends on the selected mist eliminator. Recommended K factors are listed in Table 5 (horizontal vessels) and Table 6 (vertical vessels).
b) The gas velocity is limited to mitigate the risk of entrainment of liquid from the gas-liquid interface (see Annex J).

For flows with light and heavy liquids, the physical properties of the light liquid should be used.

5.1.5 Gas Demisting Section

5.1.5.1 General

Liquid drops that do not settle out in the gas gravity section are removed by a mist eliminator. Figure B.1 provides some typical droplet sizes formed by various processes and droplet removal ranges of mist eliminators.

5.1.5.2 Gas Demisting Technology

The selection of an appropriate mist eliminator depends on a number of parameters such as gas capacity, liquid loading, fluid properties, and drop size. Table 4 summarizes relative performance characteristics of commonly used mist eliminators.

Table 4 – Summary of relative performance characteristics for mist eliminators

Parameter	Gravity/ No Mist Eliminator	Knitted Mesh	Vane Packs (no pockets)	Vane Packs (with pockets)	Axial Flow Cyclones
Separation Mechanism	Gravitational deposition	Inertial/direct interception	Inertial interception	Inertial interception	Inertial interception
Gas Handling Capacity	Low	Moderate	High	High	Very High
Turndown gas capacity (Max/Min)	NA	4	3	3	3
Droplet Removal Range	Large droplet sizes	See Figure B.1	See Figure B.1	See Figure B.1	See Figure B.1
Separation Efficiency above Max Gas	Progressive decline	Very Abrupt decline	Abrupt decline	Abrupt decline	Progressive decline
Liquid Load Capacity	Very high	Moderate	Moderate	High	High
Solids Handling Capacity	Very high	Low	Moderate	Low	Moderate-high
Fouling/plugging Tendency (e.g. waxes, asphaltenes, hydrates)	Very low	High	Moderate	Moderate-High	Low-Moderate
Derating for High Liquid Viscosity	N/A	High	Moderate	High	Low
Derating for Low Interfacial Tension	N/A	Low	High	High	Moderate
Pressure Drop[1,2]	Very Low	Low	Low	Low	Moderate

Notes:
1. For low pressure drop mist eliminators, flow distribution can be an issue.
2. For moderate ΔP mist eliminators, retrograde condensation can be an issue and should be investigated.

5.1.5.3 Gas Demisting Sizing

If a mist eliminator is required, the allowable K factor for the mist eliminator should be followed as listed in Table *5* (horizontal vessels) and Table *6* (vertical vessels). If the inlet piping and nozzle are sized, selected, and designed appropriately (see Section 5.1.2.1 and Annex G), sufficient gravity separation will normally avoid flooding the mist eliminator. Evaluation of the entrainment, droplet size and liquid loading should be performed by the technology supplier.

The bottom of the mist eliminator drainage pipes shall be sealed in the liquid phase above any expected solids level. In lieu of other requirements, a seal of 100 mm below LSL should be used.

Drainage pipes should be designed to be self-venting (Refer to Annex I for Froude number sizing criteria).

- In special cases, downcomer drainage may require external drainage to a dedicated vessel.

The drainage space of the mist eliminators operates at a lower pressure than the vessel space. The lower pressure results in a higher liquid level in the mist eliminator drainage line. The maximum calculated liquid level in the drainage pipe, at the maximum design gas rate through the mist eliminator, should be less than 55% of the height from the LSH to the bottom of the mist eliminator. This requirement is critically important with mist eliminators with moderate pressure drop, such as cyclones.

- When calculating the height of liquid inside the drainpipe above the vessel liquid level, Δh_d, the light liquid density shall be used in the calculations as in Equation 4:

$$\Delta \mathrm{h_d} = \frac{\Delta P_d}{(\rho_L - \rho_g) \cdot g} \tag{4}$$

where

Δh_d	=	drain pipe liquid height above external liquid level (m)
ΔP_d	=	Mist eliminator drainage pressure drop (Pa)
ρ_L	=	Light liquid density (kg/m^3)
ρ_g	=	Gas density (kg/m^3)
g	=	Gravitational acceleration, (m/s^2).

The minimum spacing between LSH and the bottom of the mist eliminator is then $\Delta h_d/0.55$.

Table 5 – Typical K factors of demisting internals and vessel gas flow area for horizontal vessels

Demisting Internals	Vessel Gas Space K Factor, m/s	Demisting Internals K Factor, m/s	Comments
No demisting Device	0.07	NA	Does not remove fine mists
Vertical Mesh pad (Horizontal Flow)	None specified	0.11 standard mesh	150 mm thickness (typical) Actual mesh K factor maximum limits in the range 0.075 - 0.13 depending on mesh design. The minimum spacing of 250 mm from bottom of the mist eliminator to LSH shown in Figure 3 and Figure 4 can be used for mesh.
Vertical Vane Pack (Horizontal Flow)	0.15	0.20 (single or double pocket)	200 -250 mm thickness (typical) Actual vane pack K factor maximum limits in the range 0.15- 0.25 depending on vane design. The minimum spacing of 250 mm from bottom of the mist eliminator to LSH shown in Figure 3 and Figure 4 can be used for vanes.
Horizontal Axial Flow Cyclones	0.15-0.20	Consult technology supplier	Drainage head requirements may limit the vessel gas space K factor. For typical demisting cyclone applications, the lower vessel K-factor range results in sufficient drainage space. Drainage space should be confirmed by technology supplier. 500mm cyclone tube length (typical)

Notes:

- K factor should be based on the density of the lightest liquid phase. Refer to company guidelines.
- Refer to Table 4 for comparison of demisting technologies and selection criteria.
- See Section 5.2.4 for definition of vessel gas space.
- **The physical properties, liquid loading, specific design, and geometry of mist elimination technologies impact performance significantly and should be considered in the final design.** The K factors represent typical values for a system with a low liquid load and moderate physical properties. High liquid loads, high liquid viscosities, and low interfacial tensions can detrimentally impact performance. The final design of the system, including selection of internals by the technology supplier, should account for these effects for all liquid phases.
- Margins on these K factors are not required. Refer to the Section 4.3 Process Design Cases for recommendations on margins.

Table 6 – Typical K-factors of demisting internals and vessel gas flow area for vertical vessels

Demisting Internals	Vessel Gas Space K Factor, m/s	Demisting Internals K Factor, m/s	Comments
No demisting Device	0.07	NA	Does not remove fine mists
Horizontal Mesh pad (Vertical flow)	Typically, vessel K factor is the same as mesh K factor but will be lower if accounting for mesh support area.	0.11 standard mesh	150 mm thickness (typical) Actual mesh K factor maximum limits in the range 0.075 - 0.13 depending on mesh design
Horizontal Vane Pack (Vertical Flow)	Typically, vessel K factor is lower than the vane pack K factor due to vane pack support area. For initial sizing, use vessel K factor of 0.11.	0.13 double pocket	200 - 250 mm thickness (typical) Actual vane pack K factor maximum limits in the range 0.10 - 0.20 depending on vane design
Vertical Vane Pack (Horizontal Flow)	0.15	0.20 single or double pocket	200 - 250 mm thickness (typical) Actual vane pack K factor maximum limits in the range 0.15 - 0.25 depending on vane design Even flow distribution across the vane flow area is difficult to achieve and essential for effective operation. This design is complex and shall be based on proven applications and design techniques. See Figure 8.
Vertical Axial Flow Multi-Cyclone	0.2	Consult technology supplier	Vessel gas space K factor maximum limits are in the range 0.15-0.25 depending on the system 500mm cyclone tube height (typical) Normally an agglomerator is used upstream of the cyclones.

Notes:

- K factor should be based on the density of the lightest liquid phase. Refer to company guidelines.
- Refer to Table 4 for comparison of demisting technologies and selection criteria.
- The vessel gas space K factor applies to the full vessel diameter cross sectional area.
- The physical properties, liquid loading, specific design, and geometry of mist elimination technologies impact performance significantly and should be considered in the final design. The K factors represent typical values for a system with a low liquid load and moderate physical properties. High liquid loads, high liquid viscosities, and low interfacial tensions can detrimentally impact performance. The final design of the system, including selection of internals by the technology supplier, should account for these effects for all liquid phases.
- Margins on these K factors are not required. Refer to the Section 4.3 Process Design Cases for recommendations on margins.

5.2 Horizontal Two-Phase and Three-Phase Separator Specifications

5.2.1 General

For horizontal separators, the specifications in Section 5.1 apply with the following additions.

5.2.2 Nozzles

5.2.2.1 Inlet Nozzle/Inlet Devices

Specifications are as per Section 5.1.2.1 with the following additions:

a) Inlets can be through the head or on the vessel shell. An inlet through the head is preferred.

b) LSH spacing specification from the bottom of a non-submerged inlet device should be followed as shown in Figure 3 and Figure 4.

c) Distribution baffles or other types of baffles impart a pressure drop which helps to redistribute the flow. These baffles should be used in conjunction with the inlet device to "smooth" the velocity profile of the incoming fluids and promote good flow distribution to the gravity section of the horizontal vessel. Note that distribution baffles can also have wave dampening capability, but not all wave dampening baffles have distribution benefits. See Annex C for design considerations of separators on floating facilities.

   Distribution baffles define the boundary between the inlet section and the gravity separation section of the separator. The baffle reduces the risk of fluids short-circuiting between inlet and outlets and promotes plug flow in the gravity separation section.

   1) The distribution baffle may be a perforated plate or slats/louvres. Packings have been used; however, these should not be used for viscous or dirty fluids due to the risk of fouling.
   2) One or more distribution baffles can be used and can cover all or a portion of the vessel cross-section.
   3) The open area of the baffle(s) should be selected to balance flow distribution and pressure drop consideration. Typically, the open area range is 25 - 40%.
   4) The design of the baffle should include impact on liquid levels due to hydraulic balance.
   5) Where baffles are applied, plugging/fouling and fluid bypassing shall be considered in the size of the baffle holes or slat spacings.
      i. The hole diameter or slat spacing should be 25 - 75 mm.
      ii. For severe fouling services or those with solids, the hole diameter or slat spacing should be 50 – 75 mm.
      iii. If solids are produced, the baffles should be designed with fewer but larger openings (so as to maintain the same open area) at the bottom to allow for solids movement through the vessel as well as to not hinder cleanout.
   6) Distribution baffles shall be mechanically robust and shall be able to withstand forces arising from surges in flow rates and slugs as well as motion induced flow (See Section 7 and Annex C).
   7) CFD can be used to optimize the design of the distribution baffle.

5.2.2.2 Gas Outlet Nozzle

Specifications are as per Section 5.1.2.2 with the following addition:

The vapor outlet should be designed as depicted in Figure 5. Where the mist eliminator has a high pressure drop such as cyclones, smaller spacings can be accepted.

5.2.2.3 Liquid Outlet Nozzles

Specifications are as per Section 5.1.2.3 with the following addition:

For a horizontal vessel, the bottom liquid outlet(s) should be located at the end opposite to the inlet.

5.2.2.4 Solids (Sand) Outlet Nozzles

Specifications are as per Section 5.1.2.4.

5.2.3 Liquid Gravity Section

Specifications are as per Section 5.1.3 with the following additions.

For three-phase separators (Section 5.1.3.1):

a) The oil and water sections in the vessel shall be sized to meet required residual water in oil and oil in water specifications. The following criteria of 1) maximum velocity and 2) drop removal can be used:

   1) Maximum velocity

      The maximum bulk axial velocity of the liquid phases below NLL should be limited to avoid the turbulence inhibiting the settling of the dispersed droplets. Higher limits are allowed depending upon the required separation performance. In lieu of company specifications, the following limits can be used [5]:

      – Bulk liquid axial velocity, V_B, below NLL: 0.015 – 0.030 m/s

      with

      – Lower values for heavy or viscous oils
      – Higher values for less critical applications.

   2) Drop removal

      i. For water from oil separation, there is a general trend in performance improvement with smaller droplet removal. Targeting smaller droplets sizes may not result in reasonable design. For example, in more viscous systems, small droplet removal will not be practical and a larger droplet may be used. Recommended water from oil drop sizes targeted for removal are:

         – Water droplet separation smaller than 100 μm should normally achieve a high level of free water removal.
         – 350-500 μm droplets are commonly used for bulk separators where the outlet water content can be of the order of a few percent.
         – 1500 μm may be used with the understanding that this will result in much higher residual water in the oil outlet.

      Water droplets are modeled to settle from NLL to NIL over an effective separation length as shown in Figure 4.

ii. For dispersed oil from continuous water separation, a 150 μm oil droplet should be targeted for removal. The oil drop is modeled to rise from BOV to NIL over an effective separation length as shown in Figure 4.

b) Drop settling/rising models are described in Annex H. For the effective separation lengths shown in Figure 4, note that:

1) If a distribution baffle is installed, the start of the effective drop separation length is the baffle location.
2) If a distribution baffle is not installed, the start of the effective separation length should be determined as follows:
   - 1 vessel diameter downstream from the end of inlet device, or
   - if no inlet device and the inlet nozzle is located in the head, 1 vessel diameter downstream from the inlet end vessel tan line, or
   - if no inlet device and the inlet nozzle is located in the shell, 1 vessel diameter downstream from the inlet nozzle centerline.

c) Internals such as plate packs can be used for agglomerating small drops. The risk of fouling and plugging the internals should be evaluated. Technology supplier should be consulted for design.

d) Further verifications of developed flow patterns may be required for certain cases; CFD tools are suitable for this purpose. Such cases may include:

- Non-standard and unproven designs
- Revamps
- Performance critical services.

For solids removal (Section 5.1.3.4):

If solids settling in the separator is a requirement, then the required particle size, for settling calculations, should be specified as an additional sizing criterion.

- Specifications of settling lengths should be followed as shown in Figure 3 and Figure 4. See Annex H for particle settling model.

- For a two-phase separator, particles are assumed to settle from NLL to BOV and the highest liquid viscosity should be used in the settling model. For three-phase separators, particles are also assumed to settle from NLL to BOV.

5.2.4 Gas Gravity Section

Specifications are as per Section 5.1.4 with the following additions.

a) The gas phase flow area is defined as the vessel area above LAH, or if a foam layer is considered, the area above LAH+foam height shall be used (see Section 5.1.3.3, Figure 3, and Figure 4 for recommended spacings). More aggressive designs can be used basing the gas velocities above NLL if declining performance above the NLL is acceptable for the service.

b) For foamy oils, internals such as vanes or packing can be installed in the gas phase to help break down large foam bubbles. The risk of fouling and plugging the internals should be evaluated. Technology supplier should be consulted.

c) Although the gas space is not sized by a drop removal model, such a model can be useful for additional analysis. The ballistic model in Annex H can be used to estimate the size of drop that will settle from TOV to LAH over the effective settling length Lg,eff shown in Figure 3 and Figure 4. The lighter phase liquid properties should be used.

5.2.5 Gas Demisting Section

Specifications are as per Section 5.1.5 with the following additions:

a) A single mist eliminator is typical for horizontal separators. In some cases, dual mist eliminators can be installed where the upstream one acts as an agglomerator for the downstream one. Mesh/vanes with vanes or mesh/vanes with cyclones are the typical combinations. Consult technology supplier for design.

b) Preferred mist eliminator orientations are listed in Table *5*. See Table 4 for relative performance characteristics of commonly used mist eliminators.

c) LSH spacing requirements from the bottom of the mist eliminator should be followed as shown in Figure 3 and Figure 4.

d) The mist eliminator should be sealed by a gas box. The gas box bottom should be inclined towards the drainage point (See Figure 5).

e) If mist eliminators are sealed in the liquid phase, fouling/plugging should be taken into account in the design.

f) If a horizontal separator is equipped with a weir, the mist eliminator drains should normally extend to the oil side of the weir. Exceptions should be made when meeting the water in hydrocarbon requirement is challenging and locating the drain upstream of the weir should be considered. In this case, the downcomer should end in the hydrocarbon phase.

5.2.6 Weirs

Weirs can be used in three-phase separators to facilitate removal of the light and heavy liquid phases.

Weirs should be used instead of elevated outlet for the light liquid phase. Benefits include improved plug flow pattern and less risk of fluid short-circuiting, less pressure drop in the outlet nozzle, and reduced risk of funneling gas into the outlet due to the increased distance between nozzle and liquid level.

Both flooded (submerged) and overflow (spill-over) weirs are acceptable:

a) Flooded weirs make use of the entire length of the vessel for the light phase and can be more suitable for floating facility applications

b) Overflow weirs maintain a more constant oil level upstream of the weir.

   – Where overflow weirs are used to segregate immiscible liquid phases, consideration should be given to the hydraulics of the weir. For overflow weirs, the total liquid level upstream of the weir will be the height of the weir plus the liquid crest over the weir. The crest height can be calculated from the standard Francis Weir Equation [8, 9]. V-shaped weirs can give better flow control if overflow type operation is selected (See Annex C).

c) Weirs should be at least 150 mm higher than the highest oil-water interface level.

d) Where the watercut varies over the life of the separator, the weir support should be designed such that a higher weir can be installed. A typical weir support height of 60% of the vessel diameter is suggested to allow for increased future watercut.

5.2.7 Vessel Aspect Ratio (L_{TT}/D)

Horizontal Separators should have a minimum L_{TT}/D of 2.5 (where L_{TT} is the vessel tangent to tangent (TT) length and D the inside diameter). For three phase separation, higher ratios are usually required to meet droplet settling requirements.

5.3 Vertical Two-Phase Separator Specifications

5.3.1 General

For vertical separators, the specifications in Section 5.1 apply with the following additions.

5.3.2 Nozzles

5.3.2.1 Inlet Nozzle/Inlet Devices

Specifications are as per Section 5.1.2.1 with the following addition:

Specifications for LSH spacing from the bottom of a non-submerged inlet device should be followed as shown in Figure 6, Figure 7, and Figure 8.

5.3.2.2 Gas Outlet Nozzle

Specifications are as per Section 5.1.2.2 with the following additions:

a) The top vapor outlet should be set in the center of the top head of the vessel and the spacing from the mist eliminator should be followed as depicted in Figure 9. Where the mist eliminator has a high pressure drop such as cyclones, smaller spacings can be accepted.

b) As shown in Figure 9, the vapor outlet line should project to the edge of the mist eliminator, if applicable, at an angle of no less than 45°. At shallower angles, the flow is constricted through the mist elimination device and causes local increases in flow which can lead to premature flooding.

5.3.2.3 Liquid Outlet Nozzles

Specifications are as per Section 5.1.2.3 with the following addition:

In a vertical vessel, it is recommended that the bottom liquid outlet be located in the center of the bottom head of the vessel. When other designs are adopted, a drain in the center bottom head of the vessel should be added to properly drain the vessel. Submergence requirements per Table 3 still apply.

5.3.2.4 Solids (Sand) Outlet Nozzles

Specifications are as per Section 5.1.2.4.

5.3.3 Liquid Gravity Section

Specifications are as per Section 5.1.3.

5.3.4 Gas Gravity Section

Specifications are as per Section 5.1.4 with the following additions:

a) The vessel gas space is defined as the vessel cross-sectional area.

b) The maximum allowable K factor in the vessel gas space depends upon the selected mist elimination device. These limits should be followed as listed in Table *6*.

c) Entrainment of drops from the gas-liquid interface should not be an issue if the design methods in this document (e.g. inlet piping, inlet device, spacings) are followed. If entrainment is to be evaluated, CFD may be used to obtain velocities to be used in Annex J.

5.3.5 Gas Demisting Section

Specifications are as per Section 5.1.5 with the following additions:

a) Preferred mist eliminator orientations are listed in Table 6. See Table 4 for relative performance characteristics of commonly used mist eliminators.

b) Typical scrubbers can be designed with a single mist eliminator (Figure 6) or dual mist eliminators (Figure 7). The dual mist eliminator design has higher turndown and separation efficiency. At low gas rates, the upstream mist eliminator is within its operating range and removes liquid drops. At high gas rates beyond its operating range (i.e. flooded), the upstream mist eliminator acts as an agglomerator, increasing the size of drops that are then removed by the downstream mist eliminator. Typical combinations are mesh with vanes or mesh/vanes with cyclones.

c) Although a horizontal flowing vane mist eliminator, shown in Figure 8, is acceptable, even flow distribution across the vane flow area is difficult to achieve and is essential for effective operation. This design is complex and shall be based on proven applications and design techniques. Consult technology supplier.

d) LSH spacing requirements from the bottom of the mist eliminator due to mist eliminator drainage requirements should be followed as shown in Figure 6, Figure 7, and Figure 8.

5.3.6 Vessel Aspect Ratio (L_{TT}/D)

Vessel aspect ratio should be as per company guidelines.

6 Instrumentation

6.1 General

In addition to safety instrumentation as per API RP 14C, separators should be equipped with pressure, temperature, and level instruments for process control and troubleshooting.

The term "sensor" used here includes indicators (e.g., dial gauges, digital displays, level floats) and sensors that are connected electronically to data systems.

Instrument connections should be configured as much as possible to allow for the sensor to be removed and installed without stopping operations for maintenance, calibration, and/or replacement as needed. Instrument nozzles and any tubing between the sensor and the nozzle should be configured so that the nozzle can be flushed of the process fluids. Instrument tubing should be free draining. The specification of isolation requirements for safe and efficient operation are outside of the scope of this document and should be provided by the company. Specific recommendations for installation of different sensor types are listed in the respective sections below.

Process control instruments should have nozzles separate from the nozzles used for safety system instruments.

In cases where solids accumulation in the separator vessel is a risk, the nozzles in the lower part of the separator vessel should be positioned, whenever feasible, on the side of the vessels, rather than at the bottom. In addition, the necessary corrective measurements should be included in such a configuration, where fixed piping is installed to ensure that regular flushing of the instrumentation and nozzles can be performed. Also, where bridle configurations will be used, accommodation for the necessary flushing and isolation valves for draining operations should be made in the design of the bridle.

Insulation and potential heat tracing should be applied for instrument nozzles and impulse lines where hydrate, wax, or the freezing of fluids can create a blockage between the sensor element and the fluids(s) in the vessel. Temperatures in heat-traced liquid lines should be low enough to prevent vapor bubbles forming and creating an effective local density that is different from the liquid density in the vessel.

Effects of motion should be considered in the selection of sensor placement. Verify that sensors and nozzle connection locations remain within the range of expected locations for the target fluid of the sensor. Refer to Annex C for recommendations.

The specification of sensor accuracy requirements is not in the scope of the document. Accuracy requirements should be specified in accordance with the application on a case-by-case basis.

6.2 Pressure Sensors

Sensors to measure the pressure inside the separator vessel should use nozzles that do not carry flows into or out of the vessel. Pressure instrumentation nozzles should not be positioned in locations where the measurement could be affected by the fluid velocity. Proper shielding may mitigate this effect, when applicable.

The pressure sensor nozzle locations and positions should be selected to reduce the risk that solids, wax, or other contaminants can block the nozzle and any tubing connecting the sensor to the nozzle.

Proper insulation and/or heat tracing of piping connections between the vessel and the pressure measurement sensors should be provided where solids (e.g. frozen water, waxes) or condensates can impair the measurement.

6.3 Differential Pressure Sensors

Separator internal devices (e.g., baffles, packings, demisting devices) that are prone to fouling, flooding, or clogging should have differential pressure measurements across each device for monitoring purposes.

The requirements and recommendations for pressure sensors listed in Section 6.2 apply to differential pressure sensors as well.

6.4 Fluid Sample Nozzles

Some applications required sampling of fluids from the vessel. The quantity and locations should be per company guidelines.

6.5 Level Sensors

6.5.1 General

Separators with a central inlet should have level instruments in both separation sections. Level sensors should be provided for the entire measurement range. Level sensors should be placed so that their readings are not negatively affected by the presence of other internals.

Level instrumentation nozzles should not be positioned in locations where the measurement could be affected by the fluid velocity. Proper shielding may mitigate this effect, when applicable.

For level sensors located outside of the vessel, care should be taken to ensure that the upper and lower sensor connections are always in contact with the fluids of interest. For example, for a gas/liquid interface measurement, the upper sensor or tube connection should remain in contact with only the gas in the vessel, and the lower sensor or tube connection should remain in contact with only the liquid being measured (whether water or oil). Similarly, for a water/oil interface level measurement, the upper sensor or tube should remain in contact with only the oil in the vessel, and the lower sensor or tube connection should remain in contact with only the water.

For cases where a separator design calls for bridles to be used for level sensors, appropriate isolation valves should be installed between the vessel and the bridle to allow for maintenance activities. The bridle assembly should be designed with valves to vent gas and drain as much liquid as possible for servicing operations.

6.5.2 Internal Level Sensors

In general, it is preferred to use level measurement instruments with the sensing elements inside the vessel. It is acknowledged, however, that these types of level sensors are not always feasible.

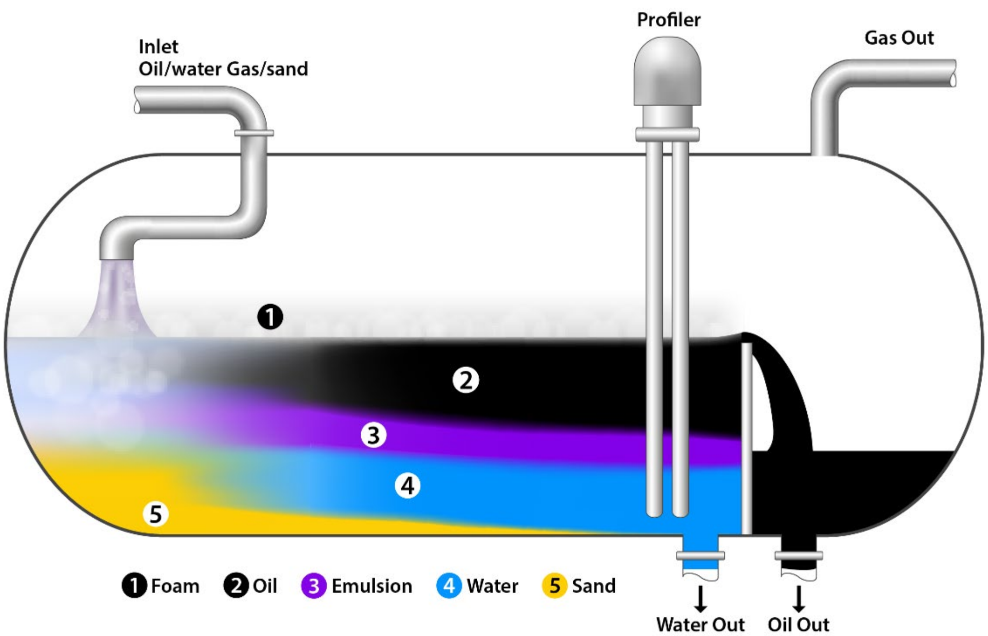
These types of sensors are capable of measuring all or some of the interface levels and the fluid type and solids in a vessel. Devices in this category are based on a wide range of technologies to sense the differences in density of the adjacent fluids. The technologies include nucleonics, radar waves, capacitance, inductance, thermal energy absorption, thermal energy diffusion, and mechanical floats. Examples of multi-interface sensor systems based on nucleonics are shown in Figure 10.

Regardless of the technology that is used (e.g., capacitance, radar, nucleonics), a level profile meter that is installed inside a vessel should be evaluated to understand its performance in such conditions as heavy oil service, separators with expected emulsion challenges, foaming challenges, or other separation issues.

Sensors to measure the fluid interface levels inside the separator vessel should use nozzles that do not carry flows into or out of the vessel. It is acknowledged that it may not be possible to remove internal level sensors without interrupting vessel operations due to the size and configuration of many internally installed devices.

Per Section 7.3, internal level sensors or profilers should be capable of withstanding the mechanical loading caused by all operational condition including slugs entering the vessel and sloshing loads caused by vessel motions. Shielding such as perforated pipe that protects the probe shaft from fluid mechanical loads should be assessed for its effect on the level measurement.

There are special requirements and recommendations for each of the available products. The sensor technology supplier should be engaged early in the design process to coordinate the installation and uses of these sensors.



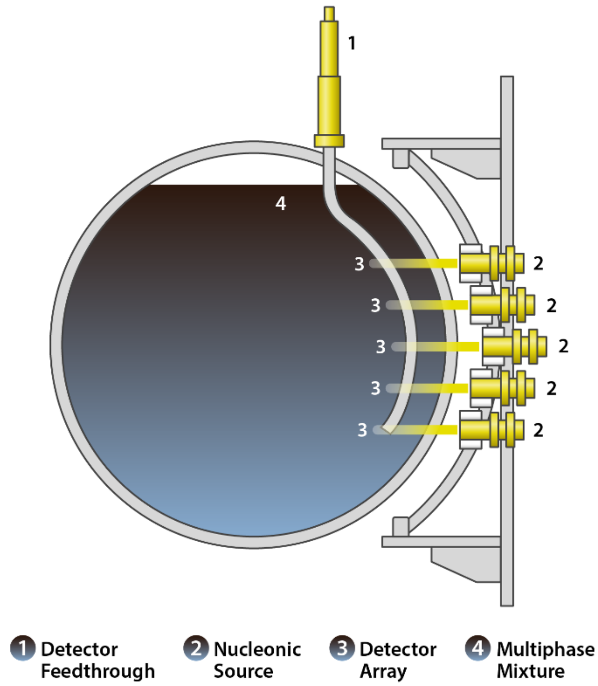


Figure 10 – Examples of Level Profilers in a Separator Vessel

Special permits are typically required by national and local regulatory agencies to transport, install, and operate nucleonic devices. The vessel designer and company should comply with all governing bodies with authority over the design, installation, and operation of a nucleonic device.

6.5.3 External mechanical and visual level gauges

Level measurements should be made inside the vessel. If external measurements are made, one nozzle is required for each different density fluid phase~~.~~

A schematic example of externally mounted level sensors is shown in Figure 11. The vessel connections are placed such that they bound the entire range of fluid interface positions being detected. Multiple connection points for each fluid between the vessel and the external standpipes should be considered to reduce the risk of the standpipe fluid being isolated from the vessel.

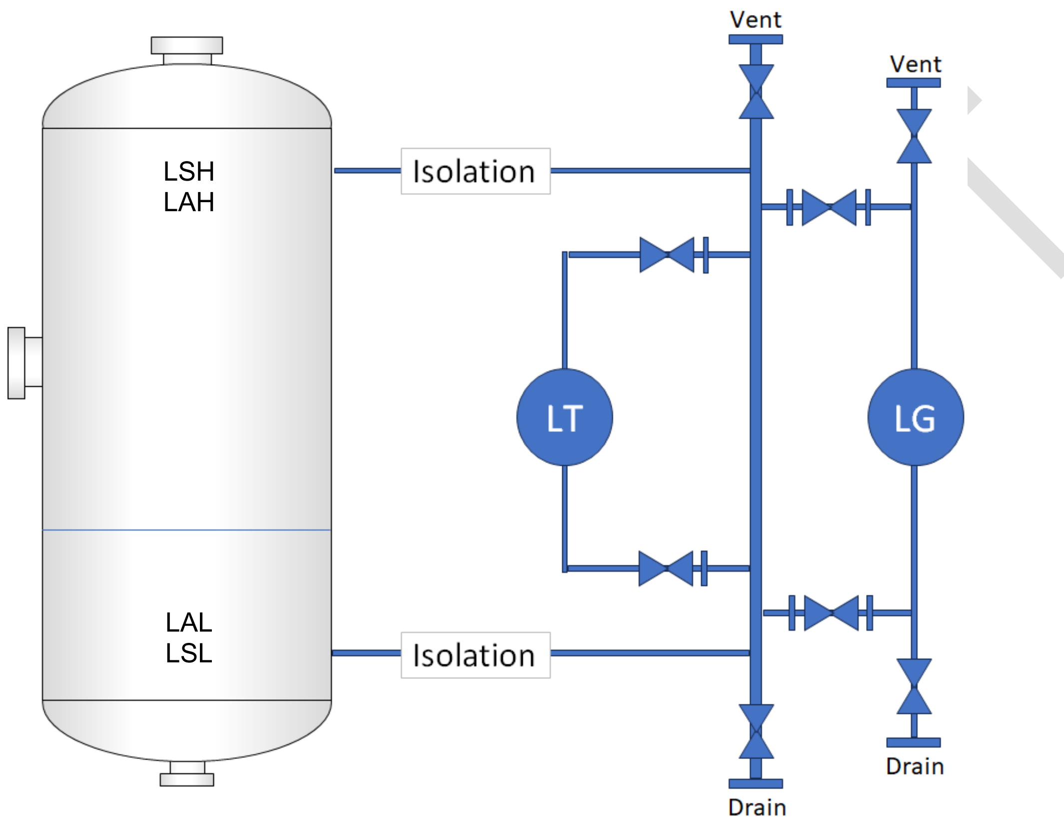


Figure 11 – Example of Two-Phase Externally Mounted Level Sensor System

6.5.4 External differential pressure sensors for liquid height

The requirements and recommendations for pressure sensors listed in Section 6.2 apply to level sensors that employ differential pressure sensors.

It is especially important to consider the issue of heat tracing and insulation of the interconnection piping because condensation in the vapor lines and bubbles in the liquid lines can be trapped with no means of detecting them. Both sets of connections for a differential pressure sensor should be installed so that the nozzles and tubing can be flushed and vented.

6.6 Temperature Sensors

Temperature sensors should be placed so that they remain in the fluid of interest under all operational conditions including cases of sloshing due vessel motions.

Temperature sensors should be placed in the location of interest. For example, those for heater control should be near the heater and in the fluid being heated. Temperature sensors for process control or trouble shooting should be placed at the relevant internal location.

Temperature sensor probe locations should be selected to reduce the risk that solids, wax, or other contaminants can thermally insulate the probe from the separator fluids.

To help ensure a representative reading, the temperature sensor should be long enough to extend into the vessel and to avoid stagnant zones.

Temperature sensors should be placed in thermowells, as seen in Figure 12, to facilitate the removal of the sensor without stopping the process flow. It is recommended that the thermowell be placed outside the path of high velocity jets or, otherwise the thermowell design should be designed to prevent mechanical vibrations.

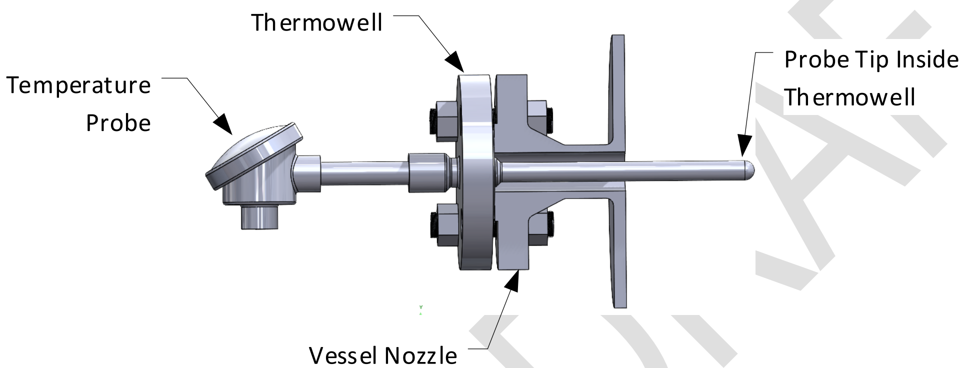


Figure 12 – Temperature Sensor in a Thermowell

7 Mechanical Design Considerations

7.1 General

The vessel which acts as a separator is almost always a pressure vessel. While there are well established mechanical design codes for pressure vessels, the process and mechanical designs are related and as such should not be performed as separate exercises. At the very least, the process designer must consider the areas of the vessel design which might influence the process performance. In addition, separator internals are subjected to numerous forces both from operational loads as well as from some occasional loads. The mechanical design of internal components shall be sufficient to withstand all applicable loads and combinations of loads. The following are general steps in establishing the parameters to consider when designing the mechanical aspects of the separator.

7.2 Vessel Design

The separator vessel should be designed to an industry recognized mechanical design code, such as the ASME Boiler and Pressure Vessel Code, or similar. The design code should be appropriate for the service and should comply with all local regulatory requirements. To provide the pressure vessel designer with the required information, the process designer should specify the following:

- Maximum and Minimum Operating Pressures
- Maximum and Minimum Operating Temperatures
- Maximum and Minimum Design Pressures
- Maximum and Minimum Design Temperatures
- Maximum Liquid Level
- Maximum Liquid Specific Gravity or Liquid Density

The maximum design pressure is the highest internal pressure that a separator can be continuously subjected to coincident with a specified design temperature range. The maximum design pressure is defined at the highest part of the separator in its installed position and is greater than or equal to the sum of the maximum operating pressure plus a suitable margin required between maximum operating pressure and the pressure relief device set pressure. The maximum design pressure plus suitable margins for static liquid head are used to determine separator component thicknesses required to meet the applicable mechanical design standard.

The minimum design pressure is the lowest internal pressure that a separator can be continuously subjected to coincident with a specified design temperature range. The minimum design pressure defines the external pressure load used for mechanical design, and normally is used to specify if the pressure vessel will be designed for a full vacuum, a partial vacuum, or no vacuum.

The maximum design temperature is the highest temperature that a separator can be continuously subjected to coincident with a specified design pressure range.

The minimum design temperature is the lowest temperature that a separator can be continuously subjected to coincident with a specified design pressure range. The minimum design temperature specification is necessary for evaluation of brittle fracture risks including achieving compliance with pressure vessel design and construction codes.

In establishing these values, the process designer should evaluate all relevant conditions to determine the worst-case conditions to be used for design. Selected conditions should account for normal operations as well as start-up, shutdown, transient, and maintenance conditions. Examples of often overlooked cases are a depressurizing event that could lead to extremely low metal temperatures, or a steam out for vessel cleaning that could result in high temperatures and vacuum conditions, and similar cases. These cases shall be clearly identified for the benefit of the pressure vessel designer.

The process designer can provide the various conditions in the following format:

Condition #1: P1 @ T1
Condition #2: P2 @ T2
Condition #3: P3 @ T3

The process designer can also add a descriptor to each condition:

Condition #1 (Max Flow): 1250 psig @ 180 ^{o}F.

7.3 Internal Component Design

7.3.1 General

The process designer should provide the internal component designer with the information required to complete the mechanical design of the internal components.

All applicable operating cases and combination of operating cases should be evaluated to determine the worst-case loads on the separator's internal components. Distributed loads, which are the most common load type encountered, should be applied to the area of fluid contact. Point loads, which are less common, should be applied to the point at which the load contacts the component. Dead loads should be applied at the center of gravity of the internal component.

The reactions, moments, stresses, and deflections resulting from the applied force(s) should be calculated. Verify that all forces are applied in the correct direction. Multiple calculations may be required to determine the governing stress and/or limiting deflections.

As discussed in Section 4.10 and Annex G, the flow into and within a separator is rarely uniformly distributed across the area through which it is flowing. As a result of flow maldistribution, the mechanical loads on the internal components may not be evenly distributed. Flow maldistribution should be included in the mechanical design of internal components.

For relatively simple geometries, simplified methods of calculations may be acceptable; however, verify that the maldistribution of all mechanical loads is properly determined. For complex geometries and fluid distributions, the use of CFD to determine the loads coupled with FEA to determine the mechanical response is advised.

Removable internals should be capable of being extracted through dedicated nozzles or through the separator manways without the need for cutting, welding, or other hot work. In addition, removable components should meet any applicable safety requirements which limit the weight of individual components.

7.3.2 Operational Cases

The mechanical design of separator internal components should consider, as a minimum, the operational cases defined below, as appropriate. While a separator may not need to meet its performance requirements during non-typical operating events, the internal components should be designed to mechanically withstand such events.

The designer should review the intended application for the applicable cases below and add any other appropriate cases.

a) Operating flow

The forces imparted onto internal components due to operating flow should be determined and included in the design. This includes the momentum flux of mixed phases, partially mixed phases, or individual fluids as appropriate to each internal component under consideration. All flow cases should be evaluated (i.e., min, max, normal, etc.). In some instances, the worst case may be an off-design operating mode such as recycling flow for compressor anti-surge control.

b) Startup/Shutdown flow
Flows, pressures, and temperatures outside the range of normal operating cases can be encountered during startup and shutdown events. Startup events may include fluids with different densities and viscosities than normal production fluids (e.g., completion fluids.) Off-design conditions resulting from startup and shutdown events should be determined and included in the design, even if these conditions are only temporary.

c) Trips
Unanticipated events that can trigger a shutdown can also result in off-design flows, pressures, temperatures, and such. Off-design conditions resulting from trips should be determined and included in the design, even if these conditions are only temporary.

d) Slugs
Liquid slugs often occur and can subject the separators internals to extreme loads. When slugs are possible, the corresponding load cases should be determined and included in the design.

e) Pressure relief valve and depressurization events
Pressure relief and depressurization events can result in extremely high flow rates through a separator. In addition, depending on the location of the PRV, flow could possibly travel through the separator in the opposite direction to normal operating flows. In such cases, the load case and direction of the resulting forces on the internal components should be determined and included in the design.

f) Other cases
There are numerous other cases, both operational and non-operational, which can cause off-design flow cases, with some potentially much greater than the normal operating condition. Any abnormal conditions along the separator life cycle can have an impact on the design. When applicable, these cases should be determined and included in the design. Examples of such cases include, but are not limited to, hot vapor bypass, anti-surge recycling, pigging/scraping operations, pressure settle out, steam out for cleaning, and other non-typical cases.

7.3.3 Mechanical Loads Resulting From Operational Flow Cases

As flow travels through a separator, it encounters the various internal components installed within the separator. These internal components interact with the flow to direct it in such a way as to facilitate separation. In doing so, the flowing fluids impart some of their energy onto these components. These components should be designed to withstand the resulting loads. The designer should review the intended application for the applicable loads below and add any other appropriate loads.

Common loads which result from the operational process within a separator are presented here.

a) Differential Pressure Loads
One of the most common loads imparted onto the internal components is a result of the pressure loss across that component. The drop in pressure as flow passes through, or around, an internal component results in a pressure load onto the component. For components whose operation generates differential pressure (such as mesh pads and cyclone decks) the differential pressure loads caused by all operating conditions should be calculated and used in determining the worst-case design load for each internal component.

b) Hydrostatic Pressure Loads
Baffles that retain fluids, in particular liquids, result in a hydrostatic load being imparted onto the baffle plate, like a dam holding back water. The hydrostatic force increases with liquid height, resulting in an increasing force profile onto the baffle. For components whose operation results in hydrostatic pressure (such as weir plates or buckets) the hydrostatic pressure loads caused by all operating conditions should be calculated and used in determining the worst-case design load for each internal component.

It should be noted that often the worst-case hydrostatic pressure load may occur during the hydro test of the pressure vessel. In addition, the direction of the hydrostatic load resulting from a hydro test may not be the same as the loads during normal operation of the separator.

c) Momentum Loads
Internal components are subjected to loads resulting from the change in momentum that occurs when the flowing fluid encounters something in its path. This is especially true of inlet devices and internal components within separators that undergo movement. When the momentum of the incoming mixed phase fluid impacts onto the inlet device, forces are applied to that device. These forces can be significant and shall be evaluated and accounted for per Section 7.3.5.

The loads described above are a result of the fluid streams interacting with the internal components and are therefore dependent upon the properties of the fluid streams. For components upstream of the gravity settling section of a separator, such as inlet devices and distribution baffles, these fluid streams are multi-phase streams. As such, the calculation of loads must consider the mixture properties of the multi-phase flow streams. This requires the mixture density and mixture velocity of the fluids at the actual separator pressure and temperature be calculated and used in all estimates. The equations for determining the density and velocity of a multi-phase fluid mixture are provided in Section 5.1.2.1. If solids or other fouling material are anticipated, then the designer should consider higher velocities and pressure drops resulting from partially blocked flow areas.

7.3.4 Other Mechanical Load Cases

In addition to the mechanical loads which result from operating the separator, there are other mechanical loads which should be determined and included in the design. Common load cases include:

a) Dead load, or the weight of the internal components.

b) Maintenance loads, including using internal components, clips, etc., as support for scaffolding and/or personnel during separator maintenance events.

c) PWHT of dissimilar materials. Should the vessel have non-removable components which differ in material from the shell, the difference in expansion rates of the materials can result in mechanical loads.

d) Thermal expansion between the pressure boundary of the vessel and the internal components.

e) Transportation, lifting, handling, erection, and installation loads, especially when large vertical vessels are shipped lying down. This situation can result in unexpected loads on internal components in directions in which those components may not be adequately supported unless temporary shipping supports are provided. During the installation of separators, loads can often be imparted onto internal components in unexpected directions.

f) On floating oil production facilities, loads on internals caused by cyclic movements of the facility (wave and wind effects), specifically bulk liquid movements ("sloshing") in separators. Fatigue effects of cyclic loading should also be evaluated.

   Separators that undergo significant movement, such as with separators on FPSO facilities, encounter especially complicated loadings from cyclic, sloshing type flows. These types of separators require special internal components to mitigate the wave and bulk liquid motions in the fluid that results from the motion of the separator. It is strongly advised that CFD and FEA be used in the design of internal components in these types of separators. Annex C includes additional requirements for separators that undergo motion.

g) Internal components may undergo flow induced phenomenon, such as lift, drag, and flow induced vibrations. Flow induced loads should be estimated based on calculations or CFD when possible and by sound engineering practice otherwise.

   Flow induced vibration is an often-overlooked phenomena in the design of internal components. It occurs when the turbulence from a flowing fluid interacts with a structural component causing vibration. When the frequency of the fluid vortices aligns with the natural frequency of a structural component, catastrophic failure can occur. Flow induced vibration can be an issue in separators, especially for the inlet devices on large diameter inlet nozzles. While there are no simple checks to determine if flow induced vibration is likely, the process designer should advise the mechanical designer of the potential for flow induced vibration. When it is deemed a potential risk, analysis should be performed to determine the vortex shedding frequencies of the flow. In addition, the internal component should be designed such that its mechanical natural frequencies do not align with the vortex shedding frequencies of the fluid flow.

7.3.5 Minimum Material Thickness

The internal components should be designed with a material thickness sufficient to ensure that they will not fail under the worst-case loading conditions(s), including their own weight. The minimum material thickness for all internals should be selected to withstand all the design loads cases. If corrosion allowance is applicable, it should be added to the minimum material thickness, per 7.3.6. A common base thickness for structural components such as clips, baffles, support rings, tube sheets, vane/cyclone boxing, etc. is the greater of the minimum material thickness plus any applicable corrosion allowance, or 6mm.

7.3.6 Corrosion Allowance

Corrosion allowance should be added for materials susceptible to corrosion. In cases in which corrosion allowance is applied, an internal component exposed to the process fluid on only one side (not common) should add one times the corrosion allowance to the minimum material thickness. In all other cases in which corrosion allowance is applied, two times the corrosion allowance should be added to the minimum material thickness.

The corrosion allowance for internal components should be specified by the company and should be applied to all sides of any applicable component in contact with the process fluids (i.e., wetted surfaces). In the case of non-removable internals, applicable internal components should have the same corrosion allowance as the vessel. In the case of removable internals, a corrosion allowance consistent with the expected life span of applicable internal components should be established.

Corrosion allowance should be applied to weld joints.

When structural members made from materials susceptible to corrosion are used, the fully corroded section properties should be determined and used in any calculations as deemed necessary by the designer.

Baffles, partitions, stiffeners, and other internal components should be designed to avoid trapping fluids that cannot be drained or purged. In addition, these components should be installed to minimize corrosion traps.

7.3.7 Materials of construction

Materials of construction should be as specified by the company. If a material is not specified, the material should be one that is compatible with the process fluids and operating conditions of the separator.

When practical, corrosion resistant materials should be used for all removable internal components, especially for components that are difficult to inspect.

Pressure retaining parts subjected to mechanical or adhesive wear should be fitted with wear pads.

7.3.8 Welding

All internal components that are welded to a pressure retaining parts should use full thickness welds. Weld sizes should be sufficient to withstand all loads as defined within this document.

If internal components, which are welded to a pressure retaining component, are made from dissimilar materials to the vessel material, mounting pads made in the same material as the vessel shell should be provided to prevent possible degradation during the welding process. Mounting pads should comply with all requirements of this document.

7.3.9 Bolting of internal components

Bolting design criteria and the extent to which bolting stress calculations are required shall be agreed upon with the company. Unless otherwise noted by the company, all internal bolting should be double-nutted and with no more than two turns of thread protruding from the nut. The minimum bolt size should be M10 or equivalent.

Stainless steel bolting should use studs/bolts and nuts of different hardness to reduce the chance of galling.

7.3.10 Gaskets for Internal Components

The company shall specify gasket materials for internal components based on fluid compatibility and susceptibility to explosive decompression.

7.3.11 Galvanic corrosion prevention.

Electrical insulation gaskets should be provided where different materials are joined and galvanic corrosion can occur. All bolting between such materials should have insulation sleeves and washers.

Consult the company in cases where dissimilar materials are welded together. In such cases the potential for galvanic corrosion and its impact on the design should be evaluated.

7.3.12 Access for Inspection and Maintenance.

All internal components should be designed for full visual inspection with a minimum of dismantling. Access to critical areas should be ensured, with the inclusion of hand grips, brackets, internal manways/hatches between levels/compartments, etc. Personal access openings should have a minimum size of DN600.

7.3.13 Special considerations

The company should specify whether the service requires special metallurgical considerations, resulting from the presence of H2S, CO2, wet CO2, or similar constituents. In such cases the company should identify all material requirements for the pressure vessel and internal components.

When materials with special metallurgical properties are required for a separator, those materials should include material test reports indicating compliance with the required standard. For example, the presence of H2S often requires that the pressure vessel materials comply with NACE MR0175/ISO 15156 or NACE MR0103/ISO 17945, depending on the application. In these cases, these materials should include material test reports indicating compliance with the appropriate NACE/ISO standard.

When certified materials are required for the pressure vessel, the company should specify whether the material requirements extend to the internal components. If so, the company should indicate whether these requirements apply to all internal materials, or only those which are welded to the pressure envelope.

Regardless of the company's requirements, where internal components are welded to a pressure vessel for which special metallurgical properties are required, the design, fabrication, installation, and inspection of the internal components shall be completed in such a way as to ensure conformance of the vessel to the specific requirements (e.g. welding of internals does not compromise the compliance of the vessel to the required standard.)

Annex A

(informative)

Separator Orientation

Selection of horizontal or vertical orientation for a separator depends upon several considerations such as listed in Table A.1. Horizontal separators are typically used for liquid dominated systems with the main advantages being surge handling capacity and higher oil water separation efficiency. Gravity separation of liquid drops or gas bubbles is not opposed by the drag of the continuous phase. Horizontal separators provide more surface area for liquid degassing that is especially needed for high liquid rates. In addition, separation internals are more easily arranged and accommodated. Three-phase separators are typically horizontal.

Vertical separators are more suitable for high GVF streams such as flowing from gas condensate fields or from production separators. An advantage of vertical separators is the reduced footprint although height can be an issue. Compressor scrubbers are typically vertical.

Table A.1 – Relative Design Characteristics Between Horizontal and Vertical Separators

	Horizontal	**Vertical**
Recommended for Two-Phase Application	Yes	Yes
Recommended for Three-Phase Application	Yes	See Note 1
Separation	Handle high liquid rates more efficiently. More surface area for settling and degassing. Should be used for 3-phase separation.	Efficient for high GVF flows.
Vessel Solids Cleaning	More difficult to clean.	Easier to clean.
Level/Motion Handling	Liquid Level control more critical. Much more affected by motion.	Liquid Level control less critical.[2] Less affected by motion.[2]
Foaming	Has more gas-liquid interfacial surface area to disengage foam	Has less gas-liquid interfacial surface area to disengage foam
Surges	Larger volumes/often better for high flowrates. Smaller level rise. Level affects gas separation.	Easier and less expensive to add surge capacity for low-rate applications. Level does not affect gas capacity.
Space/Access	Larger plot space required. More adaptable to skid mounting. Easier to install and service.	Occupies less plot space (not if extends into second deck), but tall and not as easily adapted to skid mounting. Scaffolding often necessary to perform activities.

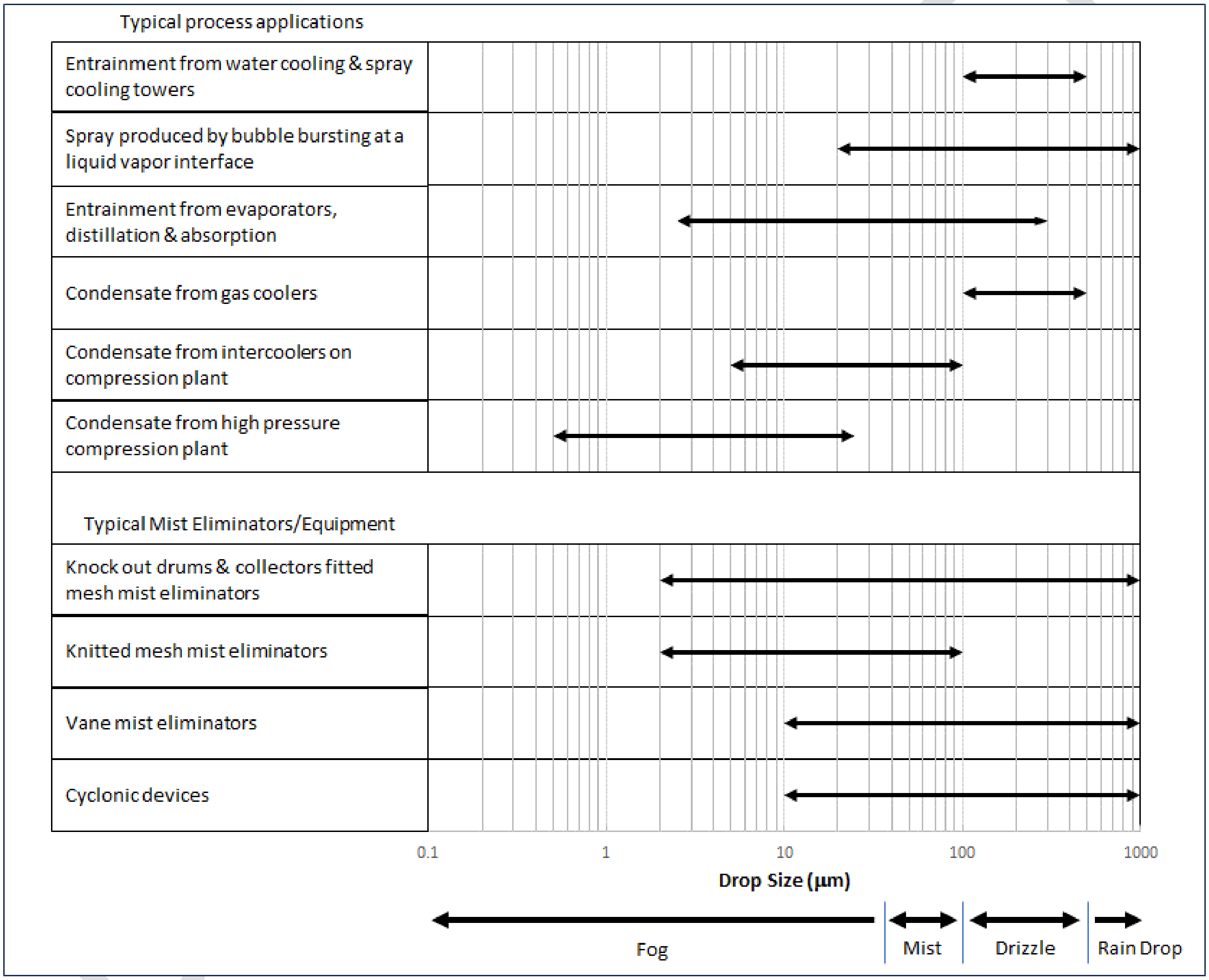
Notes
1. Generally, not used for three-phase separation due to limited liquid-liquid separation length. Exceptions include high GVF streams with light hydrocarbons that are easily separated from water.
2. Large diameter vertical vessels (>3000mm) should be checked for motion effects.

Annex B

(informative)

Droplet Sizes

Proper selection of mist eliminators depends on many parameters including the size of the droplets to be removed. Figure B.1 shows droplet sizes formed by various processes as well as the operating range of typical mist eliminators.



Note: Smaller drops can be removed by equipment not covered in this document.

Figure B.1 – Typical droplet sizes

Annex C

(informative)

Separation on Floating Production Facilities (Motion Impacts)

C.1 Floating Production Facilities and Wave Induced Motions

For horizontal and vertical vessels that are installed on offshore floating production facilities, the designer should be aware that specific design considerations and requirements apply. Specific design tools such as computational fluid dynamics (CFD) and finite element analysis (FEA) are typically used. These specifics along with some background information are presented in this section.

A floating production facility is defined as an offshore facility installation for the production and processing of hydrocarbons which is not supported by the seabed but instead floats on the surface. Examples of these are FPSO vessels, converted oil tankers, spar platforms, and semi-submersible platforms. As the surface of the sea is in continuous motion, so are the floating structures and the separators. This is particularly relevant during bad weather scenarios (storm conditions) which are expected to occur during the operational life of such facilities.

The floating structure motions, as shown in Figure C.1, cause accelerations that put the fluids inside separators in motion and this has to be accounted for when designing the separators. Each floating facility design has different motion responses to certain waves. The motion response depends on:

- Design of Facility
- Mooring System
- Wave Height
- Wave Period
- Wave Heading

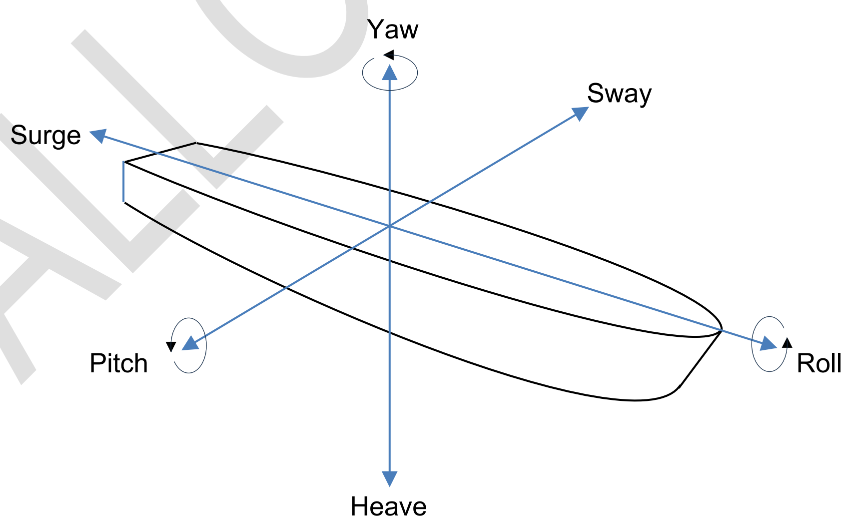


Figure C.1 – Illustration of the linear and angular motions of a floating facility

The motion dampening design therefore needs to be evaluated individually for each floating facility. Depending upon the facility, typical minimum motion data are surge/sway amplitudes and frequencies or pitch/roll angles and frequencies.

The position of the separator needs to be carefully evaluated in conjunction with motion characteristics. Typically, horizontal separators should be located near the center of gravity of the floating facility and along the axis of the minimum movement to minimize impact of motion.

If the separator has a permanent tilt, this also has to be taken into consideration in its design.

C.2 Design considerations

C.2.1 General

Floating structure motions cause the formation of waves on the gas-liquid interface, and on the oil-water interface. To reduce this effect, some type of motion dampening internals should be placed within the liquid section of vessels.

Several different motion conditions are defined for the floating production unit. Typically for the process design, the 1- or 10-year storm motion data should be used. For mechanical integrity, the 10 - or 100-years storm data should be used.

Company specifications should be followed.

C.2.2 Horizontal vessels

Floating structure motions cause the formation of waves on the gas-liquid interface, and on the oil-water interface. To reduce this effect, motion dampening internals may be required. Figure C.2 depicts an example configuration. Most commonly, perforated baffles are used for this purpose. CFD simulations can be used to verify the optimum number and location of the motion dampening baffles.

The horizontal spacing between the motion dampening baffles will depend on the motion parameters for that specific vessel. Typically, the distance between the baffles will be 2-3 m. Typical net free area (NFA) of baffles is 20-36%. Lower NFA have better dampening effect, but should be evaluated against risk of clogging. For larger diameter vessels (D ≥ 3-4 m), a longitudinal baffle should also be evaluated.

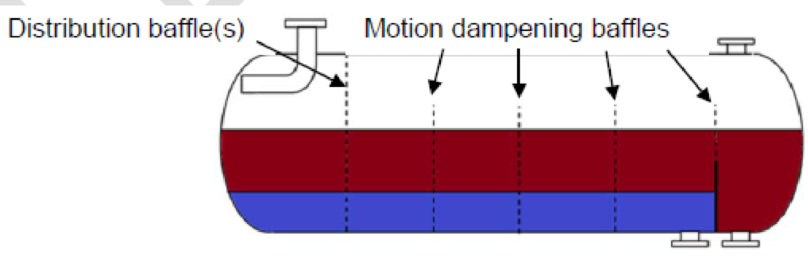


Figure C.2 – Horizontal vessel with motion dampening baffles

Ring baffles can be considered to prevent liquid from sloshing up against vessel wall. Typical location is between normal and high liquid level. The width of the ring baffle can be established using CFD.

Figure C.3 shows liquid behavior with and without a ring baffle.

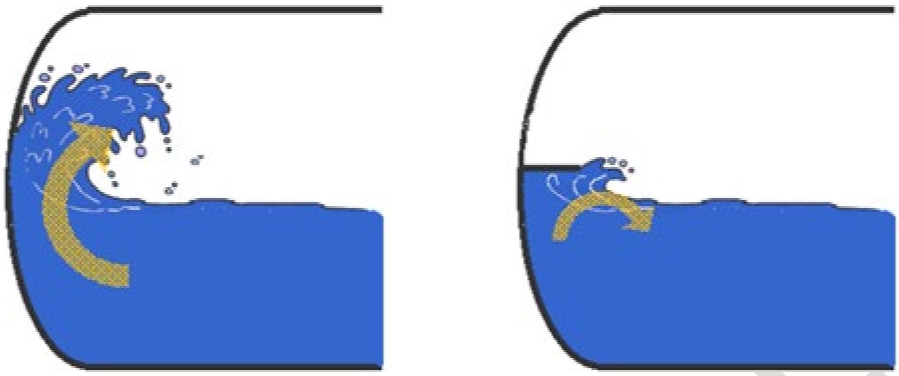


Figure C.3 – Liquid behavior in a horizontal vessel with and without a ring baffle

C.2.3 Vertical vessels

Vertical vessels have generally a smaller liquid surface area and consequently the liquid motion will be less severe. However, the level does tilt and should be considered in the level settings. A ring baffle can be considered to prevent liquid from splashing up against the vessel wall as shown in Figure C.4. The width of the ring baffle should be 10% of vessel diameter. Vertical vessels with diameters larger than 3 m should be evaluated for the need of other motion dampening baffles.

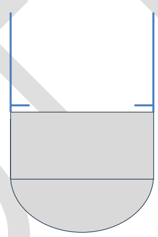


Figure C.4 – Vertical vessel with ring baffle

C.2.4 Controllability During Wave Induced Motions

The location of instrumentation should be such that the risk of spurious alarms and shutdowns is minimized. In order to achieve this, the following guidelines apply:

a) The distance between the liquid levels should take into account the motion parameters and will depend on vessel length, diameter and baffle design. As a starting point for estimating liquid displacement, the minimum distance between liquid levels can be estimated using Equation C.1 (see Figure C.5):

$$\text{Minimum liquid displacement} = \left(\frac{L_B}{2}\right) \cdot \tan\left(\frac{\pi\alpha}{180}\right) \tag{C.1}$$

where

a = roll / pitch angle (deg.)

L_B = section length between baffles (m). If no baffles included, use L_{TT} (vessel TT length).

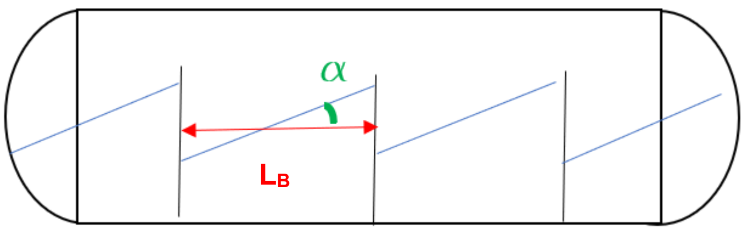


Figure C.5 – Illustration of liquid surface behavior in a baffled vessel under motion

b) For horizontal vessels, the operating level (e.g. LAL, NLL, LAH in a two-phase separator) instruments may be placed toward the center of the vessel or in a baffled section closest to the outlets. If the instruments are located towards the vessel center to minimize the effects of motion on them, the levels at the outlets should be estimated and accounted for in the level settings (see Figure C.6 and Figure C.7 as examples). In some cases, more than one level transmitter may be used to calculate the average control level. The safety shutdown level instruments should be placed per company guidelines.

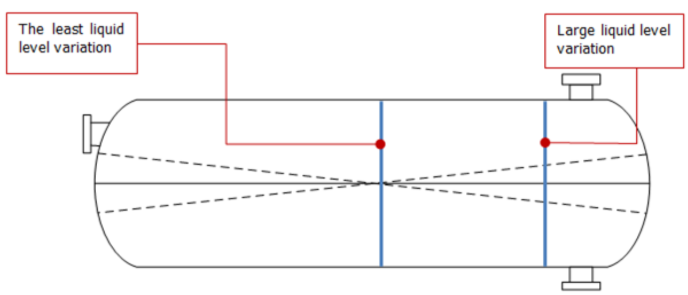


Figure C.6 – Side view of liquid level in a horizontal vessel under motion

Figure C.7 shows the effect of bridle distance and level reading with motion.

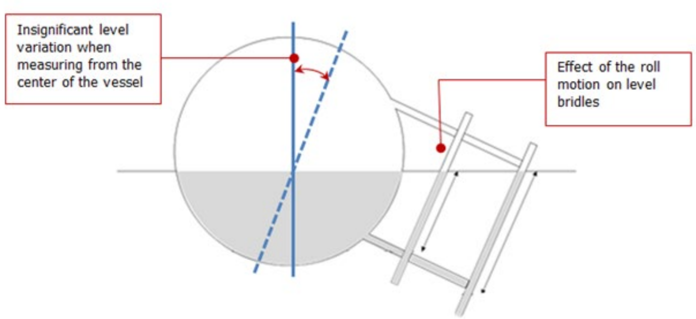


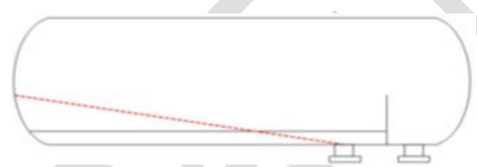
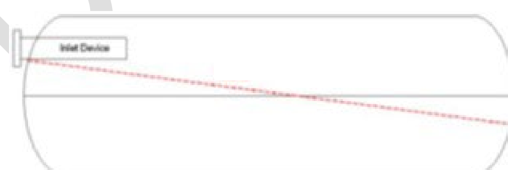
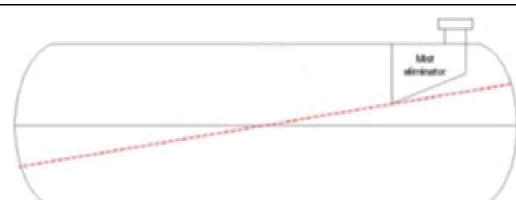
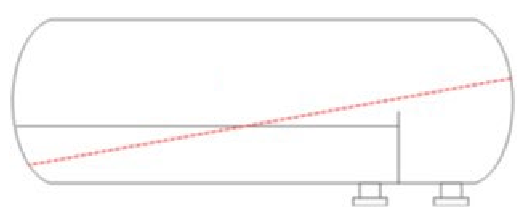
Figure C.7 – End view of horizontal vessel and level bridles under motion

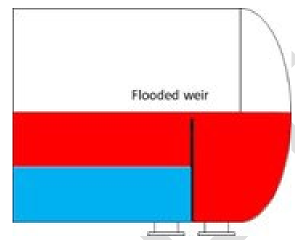
C.2.5 Internals design considerations for vessels subject to motions

For vessels subject to motion, the following general guidelines apply as shown in Table C.1.

Note that the illustrations of the interface with motion in Table C.1 are shown without motion dampening baffles. If baffles are included, the interface is better illustrated in Figure C.5.

Table C.1 – General guidelines for horizontal vessel under motion.

Verify sufficient distance between liquid outlets and respective trip levels (i.e. LSL, LISL) to prevent unwanted phase in the outlets.	
Ensure sufficient distance between gas-liquid interface and inlet device to prevent submerging inlet that can lead to separation issues such as liquid entrainment or foaming.	
Ensure sufficient distance between gas-liquid interface and the demisting device to prevent liquid overload to demisting device and / or insufficient drain margins for the demister device	
For 3-phase separators, ensure sufficient distance from interface level to top weir plate to prevent water overflowing weir.	

For floating facilities, a submerged / flooded weir is recommended. A V-shaped weir plate should be considered for separators subject to motion (Figure C.8).	

V-shaped weirs are recommended to ensure continuous flow over the weir during motion and to limit water spilling over the weir during roll motion. Figure C.8 illustrates the overflow behavior of a V-shaped weir compared to a flat weir during motion.

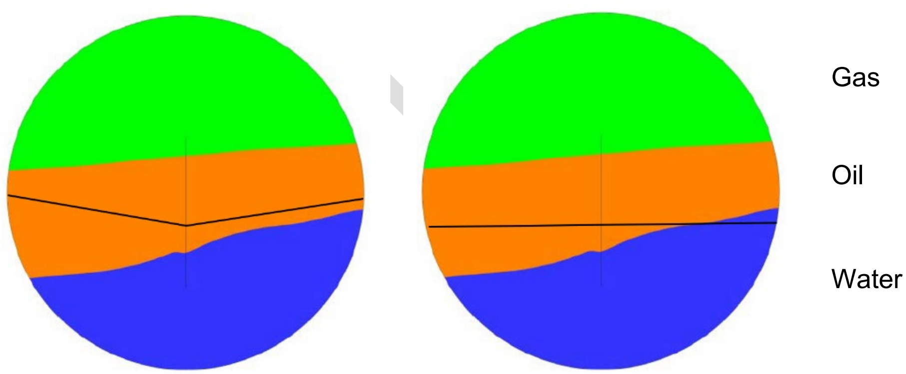


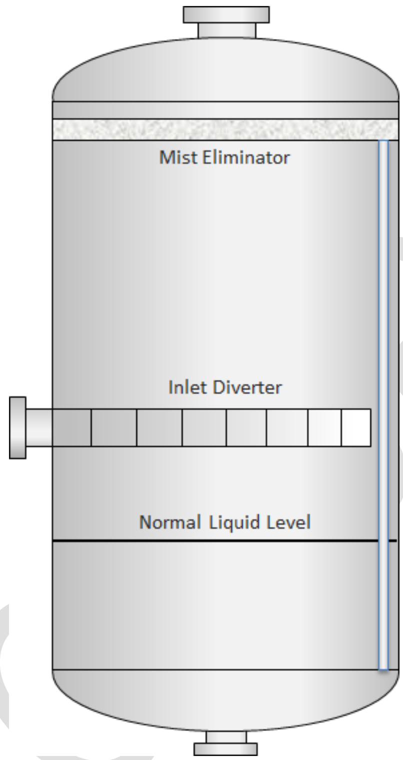
Figure C.8 – Comparison of liquid overflow with a V-shaped weir and a flat weir under motion

Annex D

(informative)

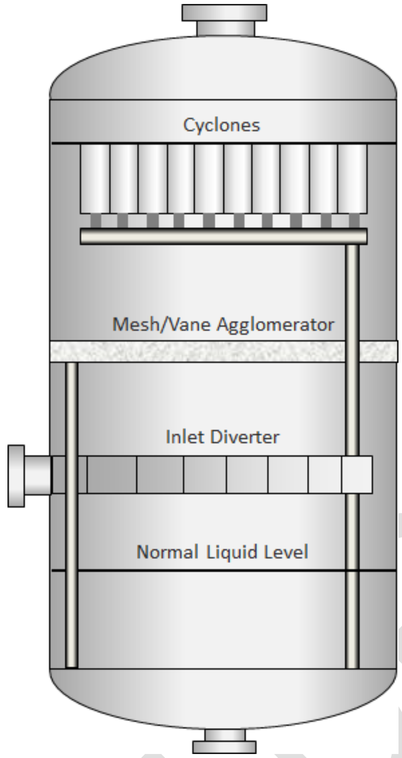
Typical Configurations

Some typical configurations of vertical and horizontal separators are shown in Figure D.1 to Figure D.10.



Note: For mesh, drainage pipe is not needed.

Figure D.1 – Typical scrubber



Note: Agglomerator drainpipe per technology supplier.

Figure D.2 – Scrubber for high efficiency, high turndown

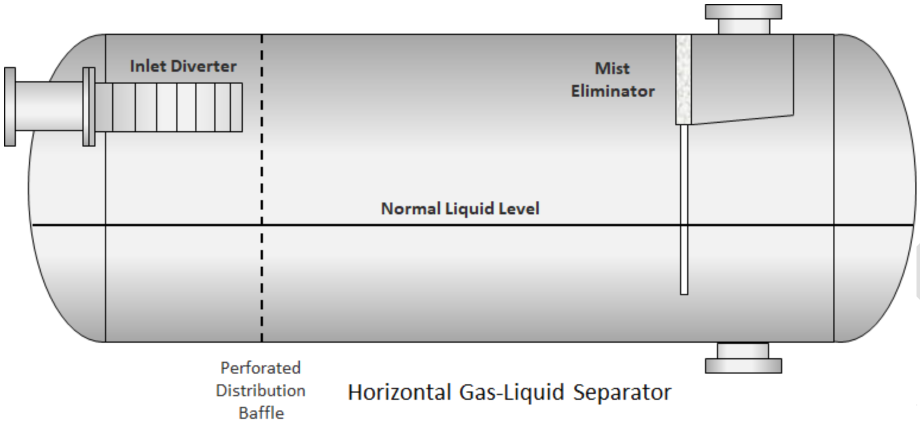


Figure D.3 – Typical gas-liquid separator

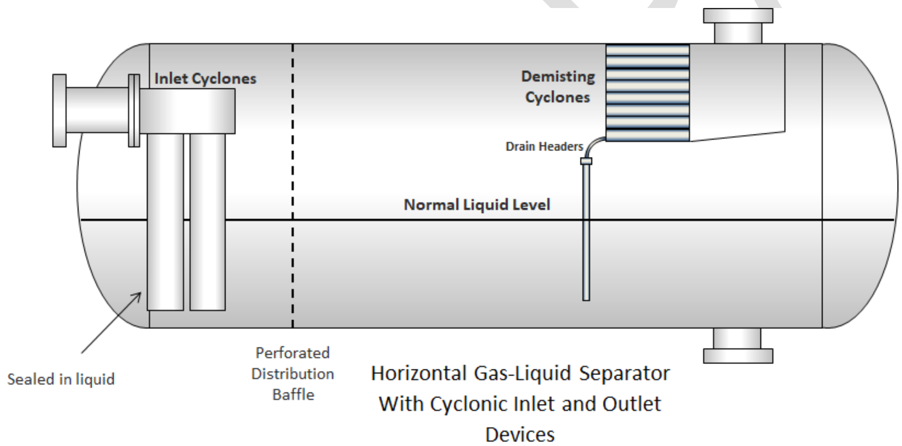


Figure D.4 – High gas rate gas-liquid separator

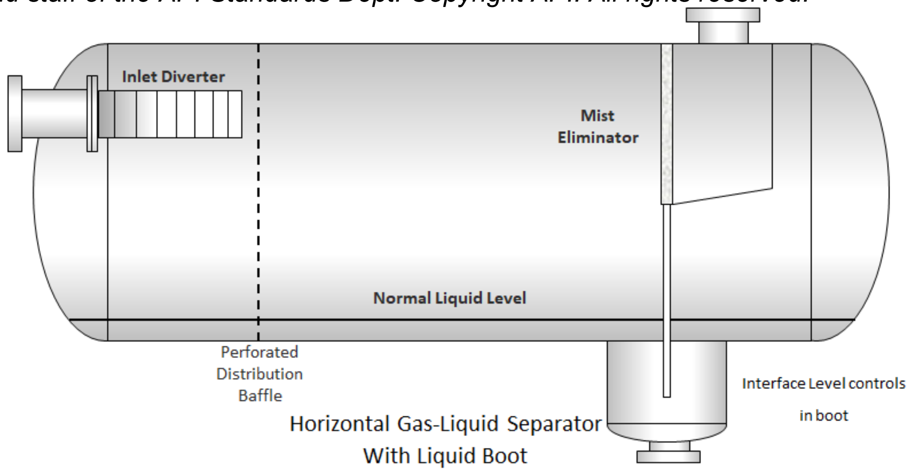


Figure D.5 – Low liquid rate gas-liquid separator

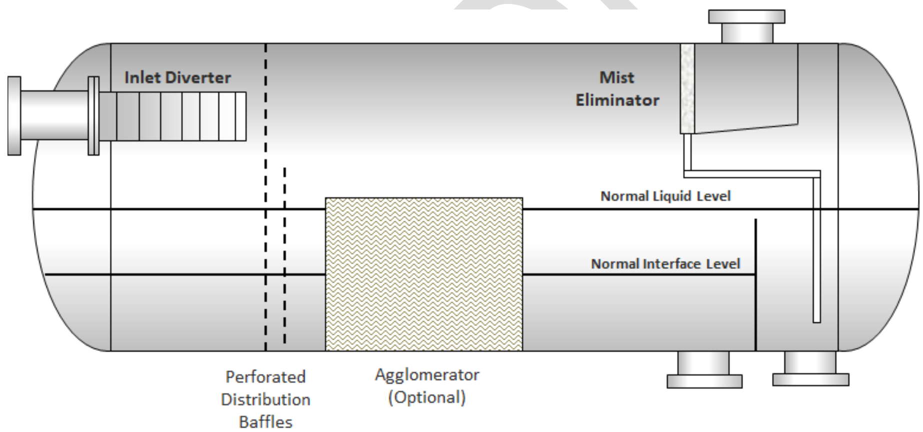


Figure D.6 – Typical gas-liquid liquid separator with flooded weir

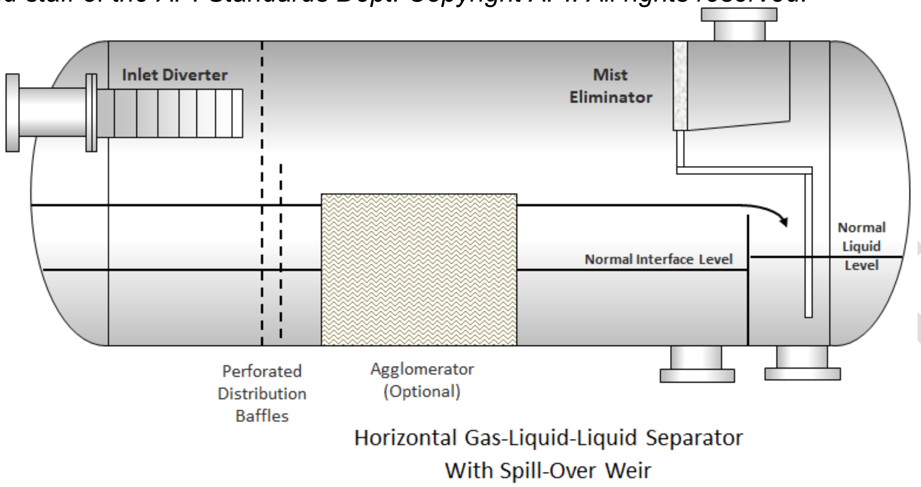


Figure D.7 – Typical gas-liquid liquid separator with spill-over weir

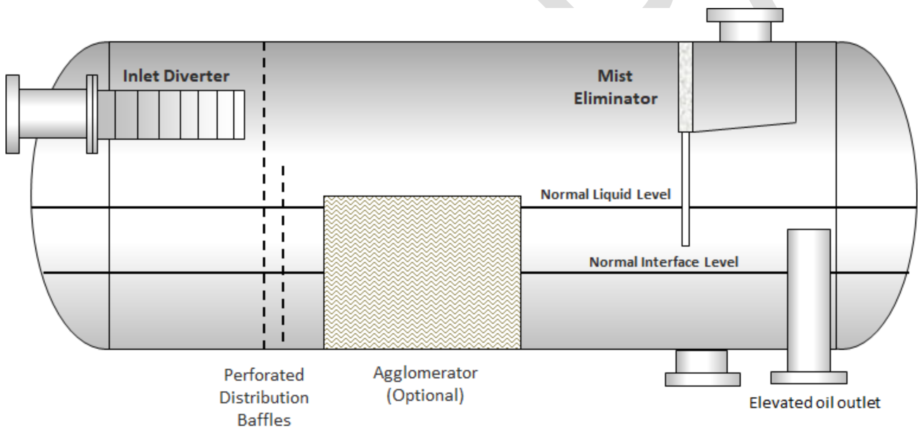
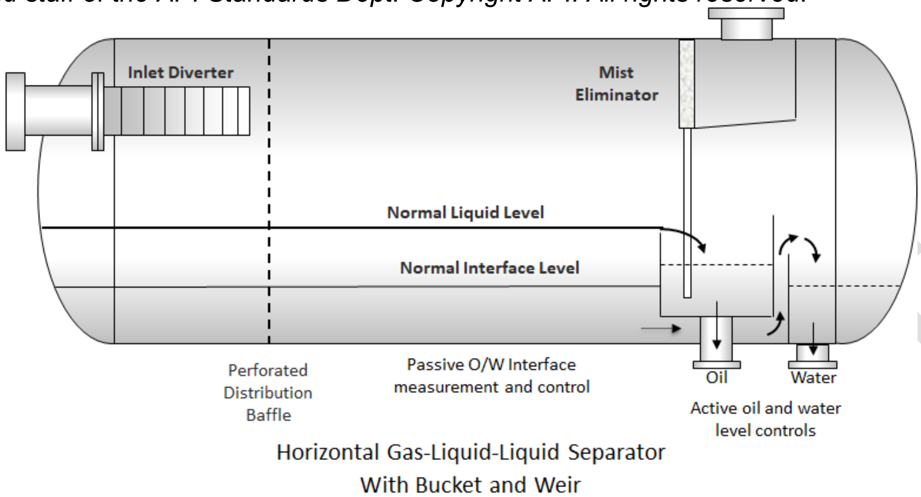


Figure D.8 – Gas-liquid liquid separator with elevated oil outlet



Note: Not recommended when oil and water densities are very close or when significant changes in oil and water densities during the operating cycle are expected.

Figure D.9 – Gas-liquid liquid separator with bucket and weir for passive oil-water interface control

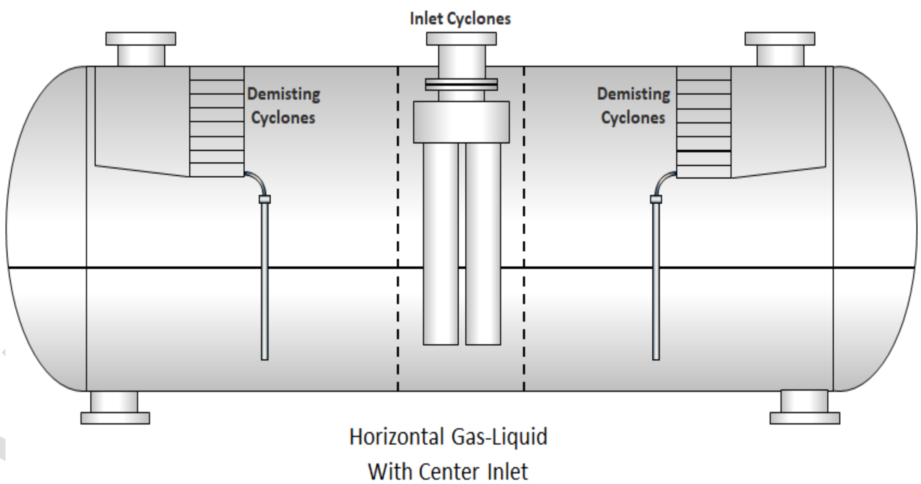


Figure D.10 – Gas-liquid liquid separator with center inlet and dual outlets for high gas rates

Annex E

(informative)

Sizing Calculation Examples

E.1 General

Three sizing examples are given below: 1) scrubber with an IVD inlet and mesh pad mist eliminator, 2) scrubber with an IVD inlet, mesh pad agglomerator, and cyclone mist eliminator, and 3) horizontal three-phase separator with an IVD inlet device and vane pack mist eliminator. In all cases, the actual ID of the nozzles should be selected so that the maximum values of inlet momentum and velocities fall within the recommended values.

E.2 Example 1: Scrubber with an IVD inlet device and mesh pad mist eliminator

The process conditions for a simple mesh scrubber are given in Table E.1.

Table E.1 – Process conditions for scrubber example sizing

Operating Pressure	barg	50
Operating Temperature	°C	45
Gas Flowrate	kg/hr	200,000
Gas Density	kg/m³	39.39
Gas Viscosity	cP	0.013
HC Flowrate	kg/hr	5,000
HC Density	kg/m³	706.06
HC Viscosity	cP	0.36
HC Surface Tension	dyne/cm	15.24
Water Flowrate	kg/hr	800
Water Density	kg/m³	993.77
Water Viscosity	cP	0.59
Water Surface Tension	dyne/cm	68.60

The volumetric flow rates are calculated as:

$$Q_g = (200{,}000/39.39)/3600 = 1.41\ m^3/s$$
$$Q_o = (5000/706.06)/3600) = 0.002\ m^3/s$$
$$Q_w = (800/993.77)/3600 = 0.0002\ m^3/s.$$

GVF is 99.85% so that a vertical scrubber is an appropriate application.

The pertinent sizing requirements are:

– Inlet Nozzle/Inlet Device (Table 1 – IVD): $\rho_m V_m^2 \leq$ 8000 Pa
– Liquid Control (Table 3): 0.150 m/30 s between levels

- Mesh K factor (Table 6): 0.11 m/s
- Mesh thickness: 0.150 m
- Liquid Outlet Nozzle (Section 5.1.2.3): velocity ≤ 2 m/s
- Gas Outlet Nozzle (Table 2): $\rho_g V_g^2 \leq$ 4500 Pa and velocity ≤ 20 m/s

Sizing Steps

1. The mixture density, ρ_m, calculated from Equation 3 is 40.47 kg/m^3. Table E.2 shows the calculated inlet momentum flux for 3 nozzle diameters.

Table E.2 – Inlet momentum flux for three nozzle IDs

Inlet ID (m)	Inlet Momentum Flux $\rho_m V_m^2$ (Pa)
0.356	8187
0.406	4799
0.457	2996

An inlet nozzle ID of 0.406 m is selected.

2. Table E.3 shows the calculated gas outlet momentum flux and velocity for 3 nozzle diameters.

Table E.3 – Gas outlet nozzle sizing

Gas Outlet ID (m)	Gas Outlet Momentum Flux $\rho_m V_m^2$ (Pa)	Gas Outlet Velocity (m/s)
0.406	4657	10.9
0.457	2907	8.60
0.508	1907	6.96

A gas outlet ID of 0.457 m is selected.

Similarly, a 0.051 m ID liquid nozzle is selected to yield a liquid velocity of 1.1 m/s. In practice, a 0.051 m ID nozzle is the smallest nozzle size that would be selected. Refer to company guidelines.

3. The maximum gas velocity, V_g, through the demister from Equation 1 is calculated to be 0.453 m/s. The vessel flow area is simply Q_g/V_g = 3.12 m^2. The vessel ID is then 1.992 m. For this example, the ID is rounded up to 2.0 m.

4. The total liquid, $Q_o + Q_w$, flow rate is 0.002+0.0002 = 0.0022 m^3/s. With such a low flow rate, the 0.150 m minimum spacing dictates the level settings that are listed in Table E.4. The differential control time between levels is 117 s.

Table E.4 – Level settings

Level	Height (m)
LSH	0.750
LAH	0.600
NLL	0.450
LAL	0.300
LSL	0.150
BTL	0.000

5. Per Figure 6, we have the following heights:

LSH	0.750 m	
		+ 0.500 m
Bottom of IVD	1.250 m	
		+ 0.406 m
Top of IVD	1.656 m	
		+ 0.900 m
Bottom of Mesh	2.556 m	
		+ 0.150 m
Top of Mesh	2.706 m	

The difference between vessel radius and gas outlet nozzle radius is (2.0 – 0.457)/2 = 0.771 m. Per Figure 9, this difference is the distance from the edge of the gas outlet nozzle to the top of the mesh pad.

The head height is 2.0/4 = 0.5 m for a 2-1 elliptical head. This head height is an approximate value for the distance from the edge of the outlet nozzle to the tan line due to the head curvature. (Exact calculations can be made.) The tan line is then 0.771 – 0.5 = 0.271 m from the top of the mesh pad. The total tan-tan height is then 2.706 + 0.271 = 2.977 m. Rounding up, the selected TT height is 3.0 m.

The ΔP across the mesh is relatively low so that the drainage head requirement is met.

In summary, the mesh scrubber parameters are (see also Figure E.1):

ID	2 m
TT Length	3 m
NLL Height	0.45 m

Inlet Nozzle ID	0.406 m
Gas Outlet Nozzle ID	0.457 m
Liquid Outlet Nozzle ID	0.051 m
Inlet Device:	IVD
Mist Eliminator:	Mesh

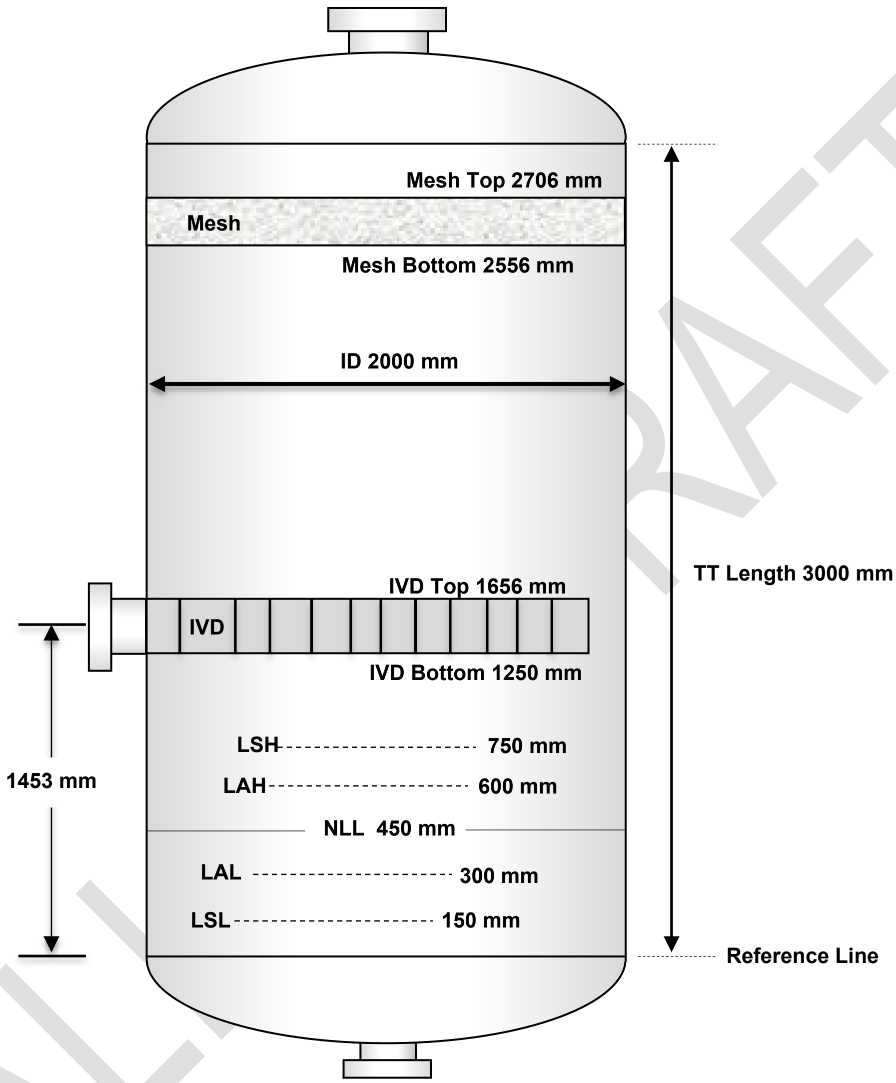


Figure E.1 – Example 1 scrubber dimensions

E.3 Example 2: Scrubber with an IVD inlet device, mesh pad agglomerator, and demisting cyclones

The process conditions for this example are the same as for Example 1.

The pertinent sizing requirements are:

- Inlet Nozzle/Inlet Device (Table 1 – IVD): $\rho_m V_m^2 \leq$ 8000 Pa
- Liquid Control (Table 3): 0.15 m/30 s between levels
- Vessel K factor (Table 6): 0.2 m/s
- Mesh agglomerator thickness: 0.15 m
- Cyclone thickness: 0.5
- Liquid Outlet Nozzle (Section 5.1.2.3): velocity ≤ 2 m/s
- Gas Outlet Nozzle (Table 2): $\rho_g V_g^2 \leq$ 4500 Pa and velocity ≤ 20 m/s

Sizing Steps

1. Inlet nozzle ID size is the same as in Example 1, 0.406 m.
2. Although the gas outlet nozzle ID can be smaller (0.406 m or even 0.355 m) than in Example 1 because of the cyclones, the same size will be used, 0.457 m. The liquid outlet nozzle ID size is also the same, 0.051 m.
3. In this example, the vessel K factor, per Table 6, is specified as 0.2 m/s. The maximum gas velocity, V_g, through the vessel from Equation 1 is calculated to be 0.82 m/s. The vessel flow area is simply Q_g/V_g = 1.71 m^2. The vessel ID is then 1.477 m. For this example, the ID is rounded up to 1.5 m.
4. Liquid levels are the same as in Example 1.
5. Per Figure 7, we have the following heights:

LSH	0.750 m	
		+ 0.500 m
Bottom of IVD	1.250 m	
		+ 0.406 m
Top of IVD	1.656 m	
		+ 0.750 m
Bottom of Mesh	2.406 m	
		+ 0.150 m
Top of Mesh	2.556 m	
		+ 0.500 m
Bottom of cyclones	3.056 m	
		+ 0.500 m
Top of cyclones	3.556 m	

In this example, it is assumed that a 0.05 m wide ring around the vessel perimeter is needed to support the cyclones. The difference between the support ring inside radius and gas outlet nozzle radius is (1.5 - 0.1 - 0.457)/2 = 0.471 m. Per Figure 9, this difference is the distance from the edge of the gas outlet nozzle to the top of the cyclones.

The head height is 1.5/4 = 0.375 m for a 2-1 elliptical head. This head height is an approximate value for the distance from the edge of the outlet nozzle to the tan line due to the head curvature. (Exact calculations can be made.) The tan line is then 0.471 – 0.375 = 0.096 m from the top of the cyclones. The total tan-tan height is then 3.556 + 0.096 = 3.652 m. Rounding up, the selected TT height is 3.7 m.

The ΔP across the mesh/cyclones is higher than mesh alone, but the 2^+ m of height between the bottom of the cyclones and LSH should satisfy the drainage head requirement. However, the ΔP should be checked by the cyclone technology supplier.

In summary, the mesh/cyclone scrubber parameters are (see also Figure E. 2):

ID	1.5 m
TT Length	3.7 m
NLL Height	0.45 m
Inlet Nozzle ID	0.406 m
Gas Outlet Nozzle ID	0.457 m
Liquid Outlet Nozzle ID	0.051 m
Inlet Device:	IVD
Mist Eliminator:	Mesh agglomerator plus cyclones.

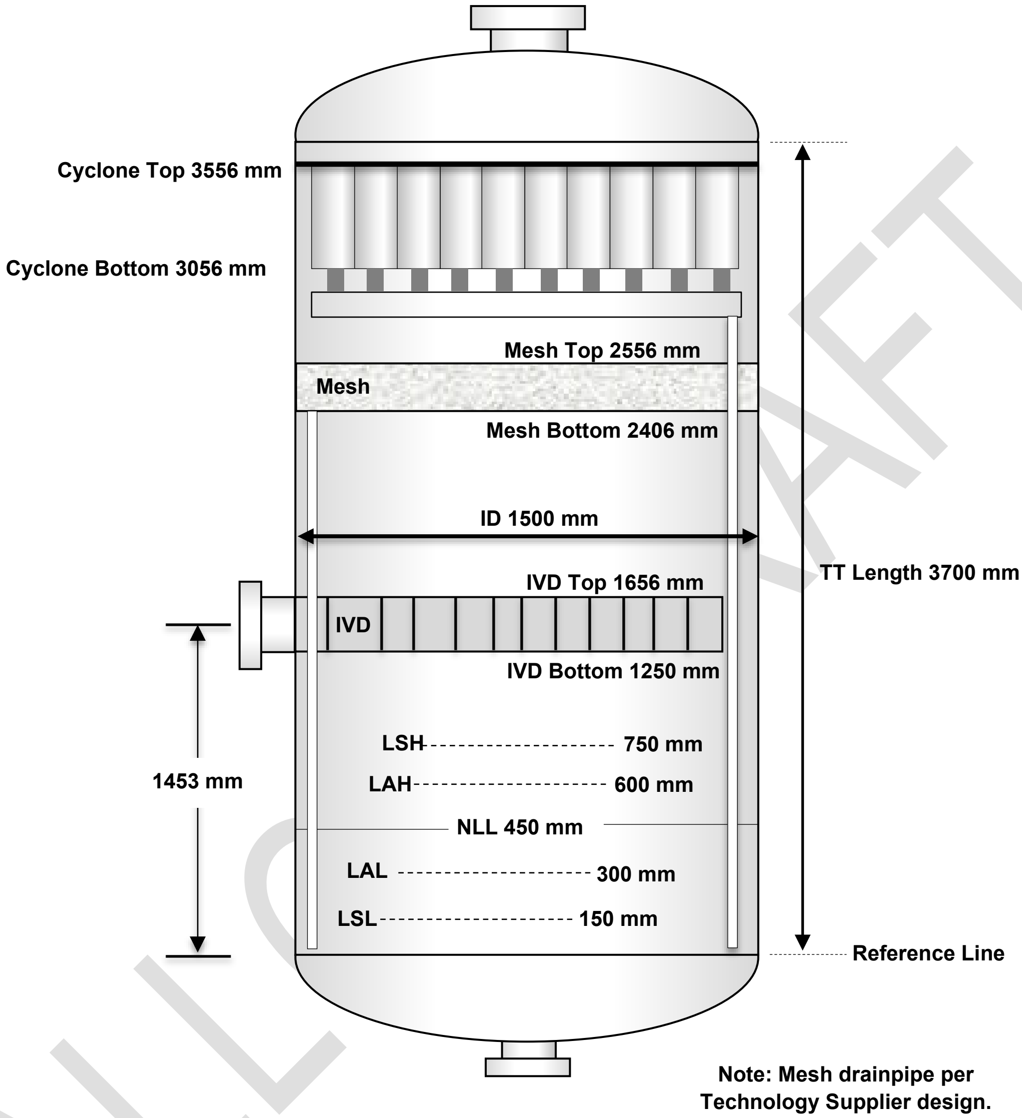


Figure E.2 – Example 2 scrubber dimensions

E.4 Example 3: Horizontal Three-Phase Separator with IVD inlet device, vane mist eliminator, and flooded weir

The process conditions for a horizontal three-phase separator are given in Table E.5.

Table E.5 – Process conditions for horizontal three-phase separator example sizing

Operating Pressure	barg	8.81
Operating Temperature	°C	60
Gas Flowrate	kg/hr	59135
Gas Density	kg/m³	9.5
Gas Viscosity	cP	0.0125
HC Flowrate	kg/hr	307878.9
HC Density	kg/m³	774.6
HC Viscosity	cP	1.24
HC Surface Tension	dyne/cm	20
Water Flowrate	kg/hr	270993.8
Water Density	kg/m³	974
Water Viscosity	cP	0.41
Water Surface Tension	dyne/cm	70
Oil/Water Interfacial Tension	dyne/cm	15

The volumetric flow rates are calculated as:

Qg = (59135/9.5)/3600 = 1.729 m^3/s
Qo = (307878.9/774.6)/3600) = 0.1104 m^3/s
Qw = (270993.8/974)/3600 = 0.0773 m^3/s.

The inlet watercut is 41%. In this case, an outlet water cut on the order of 5-10% suffices downstream process requirements. The gas volume fraction is 91%.

The pertinent sizing requirements are:

- Inlet Nozzle/Inlet Device (Table 1 – IVD): $\rho_m V_m^2 \leq$ 8000 Pa
- Liquid Control (Table 3): 0.15 m/30 s between levels
- Vessel K factor above LAH (Table 5): 0.15 m/s
- Vane K factor (Table 5): 0.20 m/s
- Vane thickness: 0.20 m
- Liquid Outlet Nozzle (Section 5.1.2.3): oil nozzle velocity ≤ 2 m/s, water nozzle velocity ≤ 1 m/s
- Gas Outlet Nozzle (Table 2): $\rho_g V_g^2 \leq$ 4500 Pa and velocity ≤ 20 m/s
- Bulk liquid velocity (Section 5.2.3) below NLL of 0.02 m/s.
- Water in oil drop size (Section 5.2.3) target 350 – 500 μm

- Oil in water drop size (Section 5.2.3) target of 150 μm.

Sizing Steps

1. The mixture density, ρ_m, calculated from Equation 3 is 92.5 kg/m^3. Table E.6 shows the calculated inlet momentum flux for 3 nozzle diameters.

Table E.6 – Inlet momentum flux for three nozzle IDs

Inlet ID (m)	Inlet Momentum Flux $\rho_m V_m^2$ (Pa)
0.508	8269
0.558	5647
0.610	3988

Although the 0.558 nozzle size has less than 8000 Pa inlet momentum flux, a larger nozzle is selected due to the high GVF and for better separation. An inlet nozzle ID of 0.610 m is selected.

2. Table E.7 shows the calculated gas outlet momentum flux and velocity for 3 nozzle diameters.

Table E.7 – Gas outlet nozzle sizing

Gas Outlet ID (m)	Gas Outlet Momentum Flux $\rho_m V_m^2$ (Pa)	Gas Outlet Velocity (m/s)
0.305	5335	23.7
0356	2880	17.4
0.406	1688	13.3

Although a 0.356 m ID nozzle satisfies the limits, a gas outlet ID of 0.406 m is selected for potential transient flows.

Similarly, a 0.305 m ID oil nozzle is selected to yield a liquid velocity of 1.51 m/s; and a 0.356 m ID water nozzle for a velocity of 0.78 m/s.

3. The height of a vortex breaker is typically ½ of the nozzle ID. For the oil nozzle, this height is 0.153. Similarly, for the water nozzle, ½ the nozzle ID is 0.178 m. An LSL and LISL of 0.2 m are chosen to maintain submergence.

4. For this example, an L_{TT}/D of 4 will be selected and an oil compartment length of 2.0 m.

   As a first pass, use a vessel ID of 4.0 m. L_{TT} is 16 m.

5. Determine the liquid levels. An NLL of 2.8 m is calculated to yield a bulk velocity of 0.02 m/s. LAH is then at 2.95 m and LSH is 3.1 m, both of which meet the 30 s time difference between levels.

6. Determine vessel gas space. With a LAH of 2.95 m, the K factor above this level is calculated to be 0.073 m/s which is less than the 0.15 m/s specification.

7. Determine gas entrainment velocity. Section 5.1.4 recommends that the gas velocity above LAH be limited to that given in Table J.1. The limit is calculated to be 3.0 m/s while the calculated gas velocity above LAH is 0.66 m/s. Entrainment from the liquid surface should not occur.

8. Determine the vane pack height. For simplicity in this example, it is assumed that the vane pack will occupy a chordal area. In reality, a support ring and beams are typically needed to contain the vanes. In addition, the vanes may not occupy the entire horizontal space due to minimum vane sizes. With the above in mind, the bottom of the vane pack is calculated to be at 3.48 m.

9. Determine maximum LSH for mist eliminator drainage. With the vane pack bottom at 3.48 m, the maximum LSH is 3.48 – 0.25 = 3.23 m. The minimum spacing between the bottom of the vanes and LSH is 0.25 m per Figure 4. It has been assumed that the ΔP across the vanes is small enough so that the 0.25 m spacing also satisfies the 55% drainage requirement of Figure 4. However, the ΔP from the technology supplier should be used in the final design. The initial LSH is 3.1 m so there is enough room for drainage.

10. Determine maximum LSH for the inlet device. For a head inlet, it is assumed that the inlet nozzle top tangent is 10% of the vessel ID from TOV. The designer should determine the actual nozzle location based on their requirements. The height of the inlet nozzle centerline is then $0.9\times4.0 - 0.610/2 = 3.30$ m. The bottom of the nozzle/IVD is then at 2.99 m. Per Figure 4, the maximum LSH is 2.99 m – 0.15 m = 2.84 m. The initial LSH is 3.1 m so there is not enough space below the inlet device.

    At this point, return to Step 3 to select another D and L_{TT} and repeat Steps 4-10 until the requirements are met. For this example, the vessel ID will be increased and the L_{TT}/D of 4 will be maintained. As the inlet nozzle spacing requirement is nearly met, a small change in vessel ID is needed. For a vessel ID of 4.2 m, L_{TT} of 16.8 m, and a 2 m oil compartment length, the following results.

    - An NLL of 2.7 m is calculated to yield a bulk velocity of 0.02 m/s. LAH is then at 2.85 m and LSH is 3.0 m, both of which meet the 30 s time difference between levels.
    - NIL is selected to be 1.35 m which yields a water velocity of 0.02 m/s below this level. The oil velocity between NIL and NLL is also 0.02 m/s. Note that the oil and water velocities being equal is coincidental and is not a design criterion.
    - Vessel gas space K factor above LAH is 0.05 m/s ($\leq$ 0.15 m/s).
    - Vessel gas space velocity is 0.45 m/s ($\leq$ 3.0 m/s).
    - Bottom of vane is 3.68 m so the maximum LSH for drainage is 3.43 m.
    - Bottom of the inlet nozzle/IVD is at 3.17 m so the maximum LSH is 3.02 m.
    - LSH of 3.00 m satisfies the spacing requirements.

11. Determine drop sizes. Use an L_{IVD}/D_{in} of 4.5 so that the IVD is 2.74 m long. The end of the IVD from the inlet tan line will depend on the internal connecting piping and flanges. For this example, assume that the end of the inlet device is 2.5 m from the tan line and that a perforated baffle is located 0.30 m away at 2.80 m from the tan line.

    The weir is located at 14.8 m from the inlet tan line. The effective length, $L_{wio, eff}$, for settling water drops in the oil phase is 14.8 – 2.80 = 12.0 m. NIL is at 1.35 m and NLL at 2.70 for a difference of 1.35 m. The ballistic model in Annex H is used to determine the size of a water drop that will fall 1.35 m before traveling 12.0 m horizontally. A 165 μm water drop size is calculated. Section 5.2.3 recommends targeting 350 – 500 μm drop sizes for bulk separation.

    The water outlet nozzle ID is 0.356. It is assumed that the nozzle tangent is 0.3 m from the weir. Per Figure 4, the downstream end of the effective oil in water separating length, $L_{oiw, eff}$, is $2\times D_w$ + 0.3 or 1.01 m from the weir. $L_{oiw, eff}$, is then 12.0 – 1.01 = 11.0 m. NIL is 1.35 m. The ballistic model in Annex H is used to determine the size of an oil drop that will rise 1.35 m before traveling 11.0 m horizontally. A 100 μm oil drop size is calculated. Section 5.2.3 recommends targeting a 150 μm drop size for deoiling purposes. If degassing is required, an 85 μm bubble size is calculated with a target of 200 μm.

In summary, the horizontal three-phase separator (with flooded weir) parameters are (see also Figure E.3):

ID	4.2 m
TT length	16.8 m
Tan-Weir length	14.8 m
NIL height	1.35 m
Weir (flooded) height	1.80 m
NLL height	2.700 m
Inlet Nozzle ID	0.610 m
Gas Outlet Nozzle ID	0.406 m
Oil Outlet Nozzle ID	0.305 m
Water Outlet Nozzle ID	0.356 m
Inlet Device:	IVD
Mist Eliminator:	Vanes

Table E.8 shows the liquid levels.

Table E.8 – Liquid Levels

Level	H (m)	Adjacent Level Differential Control Time (s)
LSH	3.00	92
LAH	2.85	95
NLL	2.70	679
LAL	0.80	33
LSL	0.20	
BOV	0.00	
Weir	1.80	
LISH	1.65	120
LIAH	1.50	117
NIL	1.35	645
LIAL	0.35	60
LISL	0.20	
BOV	0.00	

Note that in this example, the vessel gas space K factor is low enough such that a vane mist eliminator may not be needed.

In the case of a spill-over weir, the height of the gas-oil interface is controlled by the height of the weir. The control levels LSL, LAL, and NLL are below the weir height. Depending upon the design, LAH and LSH can be either below or above the weir.

Proceeding as before, the vessel size is estimated to be 4.0 m ID x 16 m TT with a 2 m long oil compartment. The gas-oil interface is calculated to be at 2.8 m to yield a bulk liquid velocity of 0.02 m/s. NIL is at 1.35 m and the weir is at 2.7 m. A crest height of 0.1 m is calculated using the Francis weir equation.

The liquid control levels are given in Table E.9 along with the differential control times. In this case, all control levels are below the weir and satisfy the minimum differential control times of 30 s.

Table E.9 – Liquid Control Levels

Oil Compartment		
Level	H (m)	Adjacent Level Differential Control Time (s)
LSH	2.00	30
LAH	1.65	33
NLL	1.25	33
LAL	0.80	32
LSL	0.20	
BOV	0.00	
Weir	2.7	
Gravity Separation Section		
LISH	1.65	109
LIAH	1.50	107
NIL	1.35	592
LIAL	0.35	56
LISL	0.20	
BOV	0.00	

The maximum level below the inlet vane device is calculated to be 2.84 m. With the gas-oil interface height at 2.8 m, the required spacing from the bottom of the inlet vane device is satisfied.

The bottom of the mist eliminator is calculated to be at 3.48 m so that the maximum LSH for drainage is 3.23 m. The LSH of 2 m is below this maximum level.

The calculated drop/bubble sizes of 175 μm water drop, 105 μm oil drop, 90 μm gas bubble are within the recommended limits.

In summary, the horizontal three-phase separator (with spill-over weir) parameters are (see also Figure E.4):

ID	4.0 m
TT length	16.0 m
Tan-Weir length	14.0 m
NIL height	1.35 m
Weir height	2.70 m
Oil Level	2.80 m
NLL height	2.700 m
Inlet Nozzle ID	0.610 m

Gas Outlet Nozzle ID	0.406 m
Oil Outlet Nozzle ID	0.305 m
Water Outlet Nozzle ID	0.356 m
Inlet Device:	IVD
Mist Eliminator:	Vanes

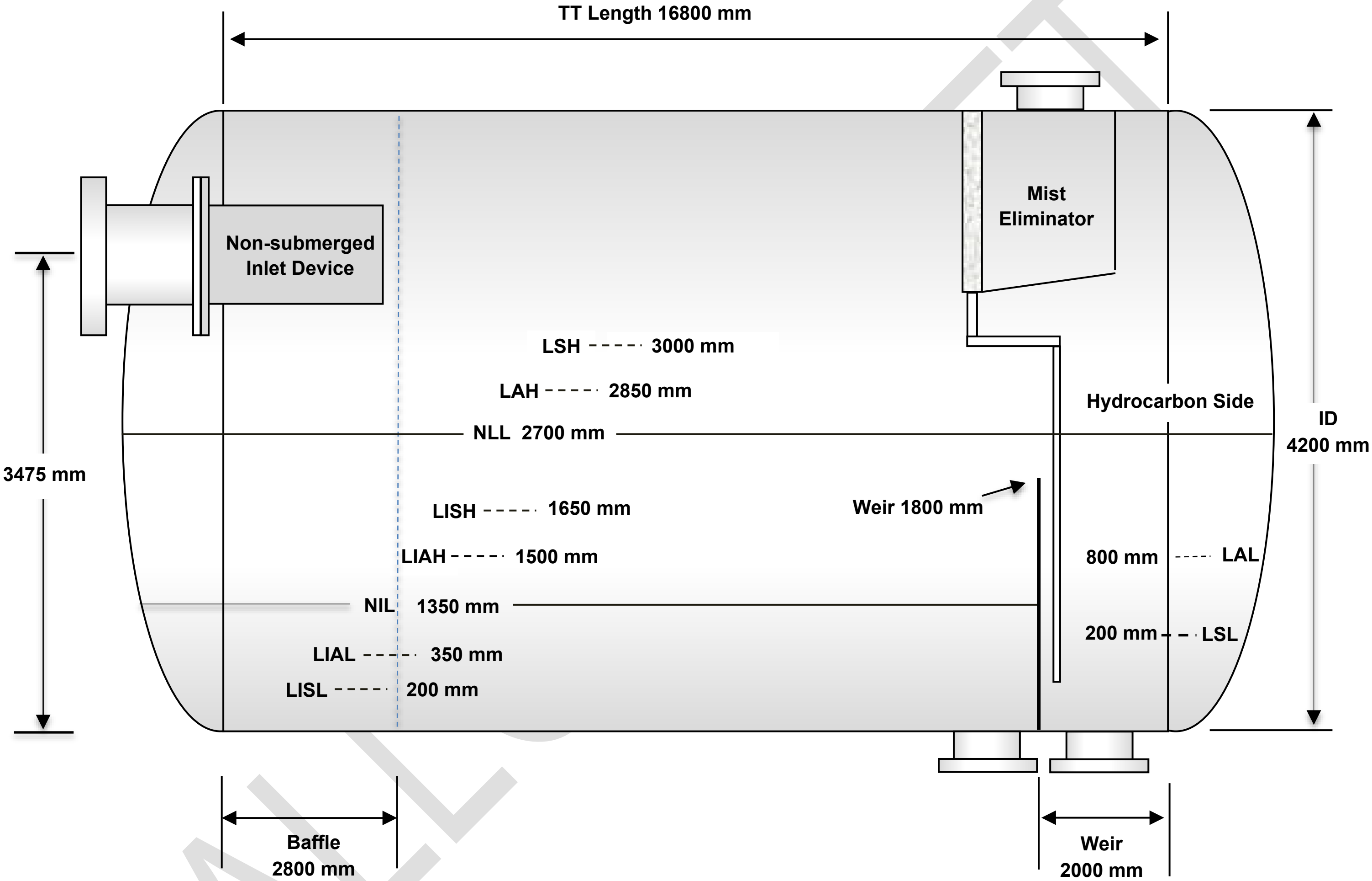


Figure E.3 – Example 3 horizontal separator (flooded weir) dimensions

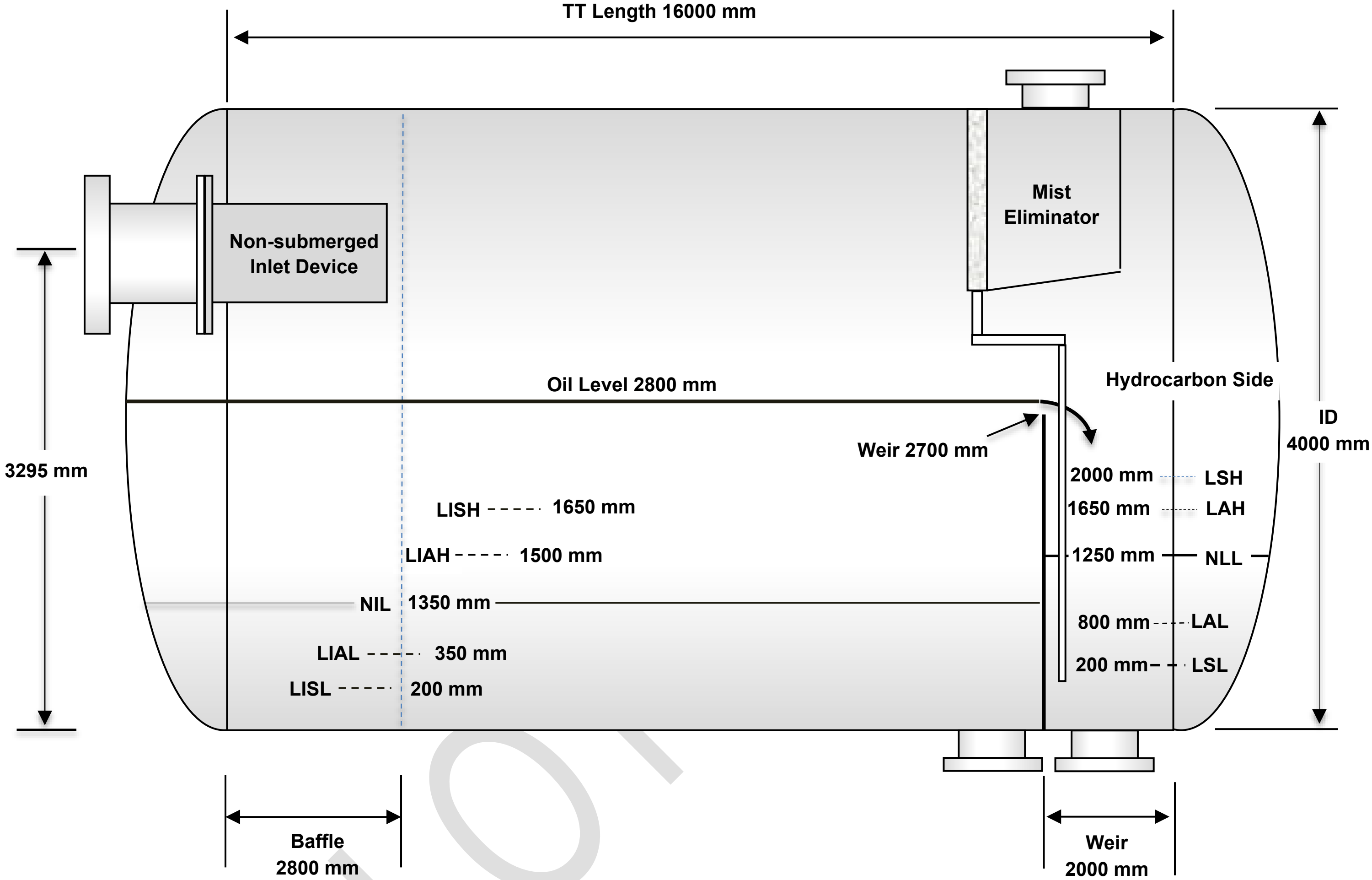


Figure E.4 – Example 3 horizontal separator (spill-over weir) dimensions

Annex F

(informative)

Datasheet

SEPARATOR PROCESS DESIGN DATA			DATE		REVISION		Page 1 or 2
CLIENT INFORMATION			**VESSEL SERVICE**				
NAME			Description				
LOCATION			Type				
VESSEL TAG			Orientation				
			Special Notes				
MECHANICAL DATA	**UNITS**						
Maximum and Minimum Design Pressure	barg						
Maximum and Minimum Design Temperature	°C						
Separator Internal Diameter, ID	mm						
Separator Length Tan-Tan	mm						
Head type(s)	-						
Manway ID(s)	mm						
Internals Material	--						
PROCESS DATA	**UNITS**	Design	Normal	Minimum	Misc 1	Misc 2	Misc 3
Pressure	barg						
Temperature	°C						
Gas Mass Rate	kg/hr						
Gas Density	kg/m³						
Gas Molecular Weight	kg/kmol						
Gas Viscosity	cP						
HC Liquid Mass Rate	kg/hr						
HC Liquid Density	kg/m³						
HC Liquid Viscosity	cP						
HC Liquid Surface Tension	dyne/cm						
Acqueous Phase Mass Rate	kg/hr						
Acqueous Phase Density	kg/m³						
Acqueous Phase Viscosity	cP						
Acqueous Phase Surface Tension	dyne/cm						
HC Liquid-Acqueous Phase Interfacial Tension	dyne/cm						
Notes on Nature of Fluids							
Liquid Slug Volume	m^3						
Notes on Nature of Liquid Slug							
Solids Removal							
Solids Density	kg/m³						
Solids Cut-off Size	microns						
Notes on Nature of Solids							
Bubble size removal from liquid	microns						

SEPARATOR PROCESS DESIGN DATA	DATE		REVISION	

Piping	**UNITS**			
Inlet Piping ID	mm			
Gas Outlet Piping ID	mm			
HC Liquid Outlet Piping ID	mm			
Water Outlet Piping ID	mm			
Maximum Allowable Vessel Pressure Drop Including Nozzles (Inlet Flange to Gas Flange)	mbar			
Notes on Inlet Feed				

Nozzle IDs	**UNITS**			
Inlet	mm			
Gas Outlet	mm			
HC Liquid Outlet	mm			
Acqueous Phase Outlet	mm			

Specified Internals	**Notes**
Inlet Distributor	
Demisting Device(s)	
Baffles	
Liquid Phase Coalescer	
Other	

Platform Motion Data	

User Required Level Spacings and Control Times

	Level	Minimum Spacing Between Levels	Minimum Time between		
High Liquid Level Alarm to High Liquid Level Trip	LAH to LSH			Two-Phase	Three-Phase (assuming weir)
Low Liquid Level Alarm to High Liquid Level Alarm	LAL to LAH				
Low Liquid Level Trip to Low Liquid Level Alarm	LSL to LAL				
BTL or BV to Low Liquid Level Trip	Ref to LSL				
High Interface Level Alarm to High Interface Level Trip	LIAH to LISH				
Low Interface Level Alarm to High Interface Level Alarm	LIAL to LIAH				
Low Interface Level Trip to Low Interface Level Alarm	LISL to LIAL				
BTL or BV to Low Interface Level Trip	Ref to LISL				

User Separator Performance Requirements

2-phase Separator			3 Phase Separator		
Liquid carryover in gas phase	Liter/Million Sm^3 gas (1.01325 bara, 15.6°C)		Liquid carryover in gas phase	Liter/Million Sm^3 gas (1.01325 bara, 15.6°C)	
			HC liquid carryover in acqueous phase	mg/Liter	
			Acqueous phase carryover in HC Liquid	BS&W %	

Annex G

(informative)

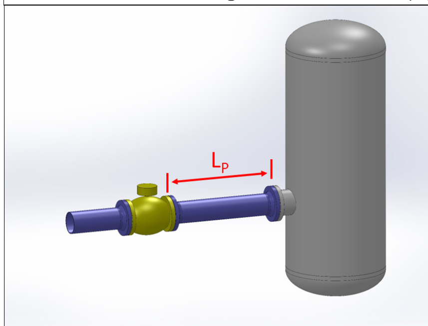
Inlet Piping

Good flow distribution in a separator's inlet nozzle is vital to the performance of a separator. Bends and/or restrictions in the piping upstream of the separator materially impact the flow within the separator. Recommendations for the inlet piping are as follows:

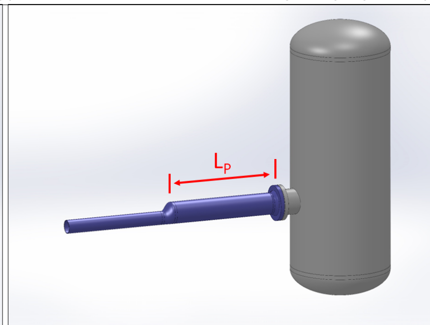
- Figure G.1, Figure G.2, and Figure G.3 provide guidance on inlet piping orientation. Lengths of inlet piping between bends or restrictions and the inlet nozzle should be followed as shown in the figures.

- Where it is not practical to maintain the recommended lengths of inlet pipe, piping bends/restrictions may be tolerated as long as the detrimental impact on the performance is accepted.

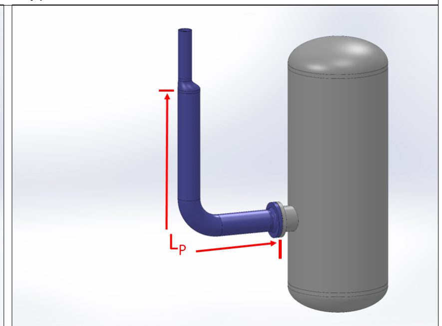
Inlet Flow Conditioning - Vertical Vessel (Typical recommendations, may vary by Company)



Valve and straight run pipe
$L_P/D_P \geq 10$

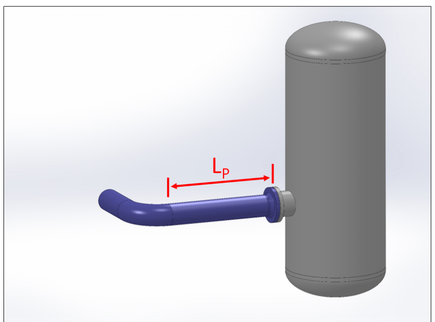


Reducer and straight run pipe
$L_P/D_P \geq 10$

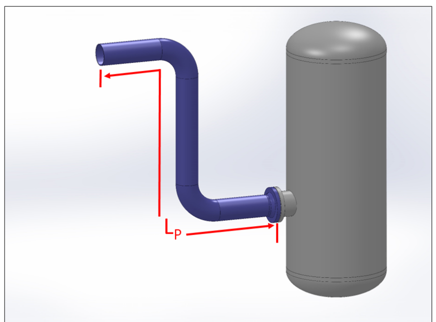


Reducer and vertical bend
$L_P/D_P \geq 10$ Given that:

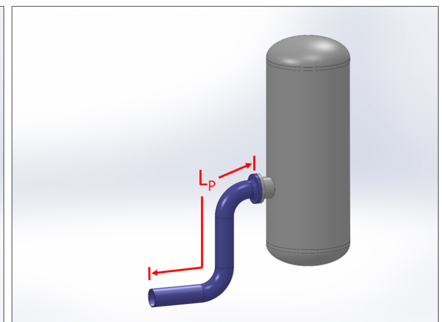
- Pipe IDs before and after reducer are less than 30% difference
- Reducer is upstream of bend



Bend in a horizontal plane
$L_P/D_P \geq 10$

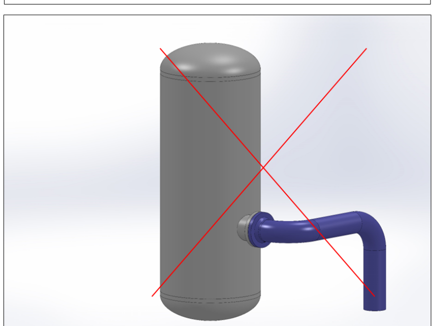


Two bends in a vertical plane
$L_P/D_P \geq 10$



Two bends in a vertical plane
$L_P/D_P \geq 10$

Note: Configuration not preferred due to potential slug generation

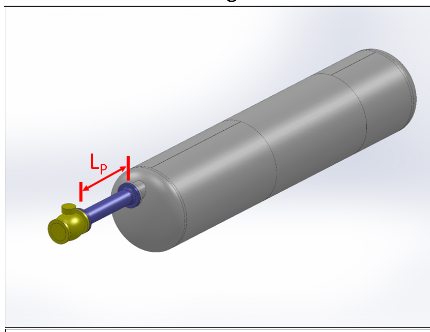


Bends in both horizontal and vertical planes are too close to inlet.

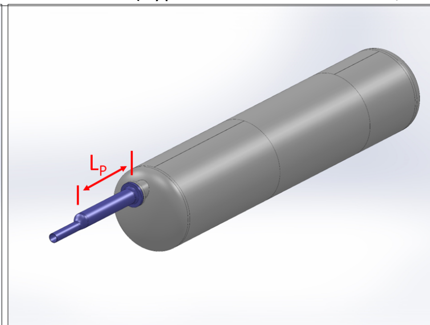
Not recommended.

Figure G.1 - Specified lengths and orientations of inlet pipe prior to inlet nozzle for vertical vessels

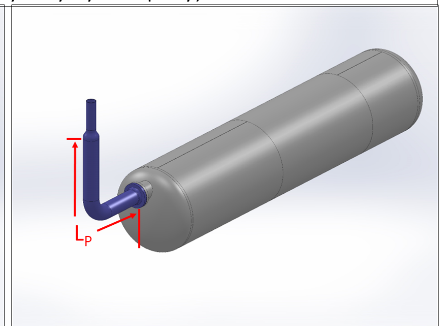
Inlet Flow Conditioning - Horizontal Vesssel - Head Inlet (Typical recommendations, may vary by Company)



Valve and straight run pipe
$L_P/D_P \geq 10$

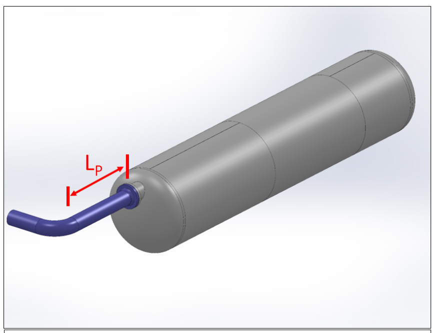


Reducer and straight run pipe
$L_P/D_P \geq 10$

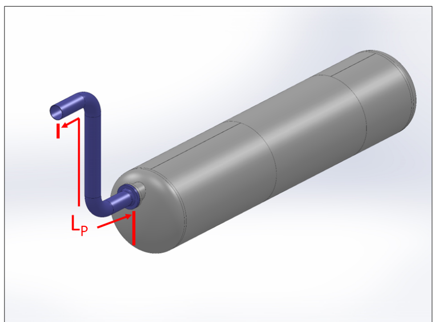


Reducer and vertical bend
$L_P/D_P \geq 10$ Given that:

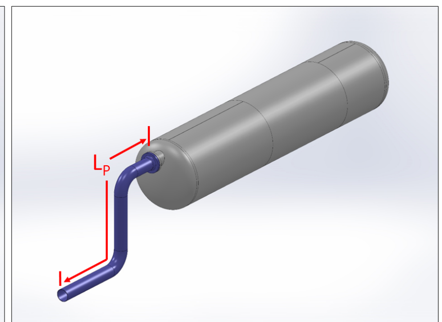
- Pipe IDs before and after reducer are less than 30% difference
- Reducer is upstream of bend



Bend in a horizontal plane
$L_P/D_P \geq 10$

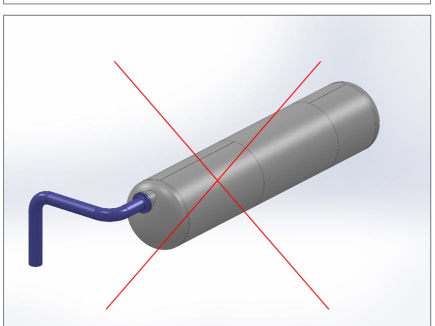


Two bends in a vertical plane
$L_P/D_P \geq 10$



Two bends in a vertical plane
$L_P/D_P \geq 10$

Note: Configuration not preferred due to potential slug generation



Bends in both horizontal and vertical planes are too close to inlet.

Not recommended.

Figure G.2 – Specified lengths and orientations of inlet pipe prior to a head inlet nozzle for horizontal vessels

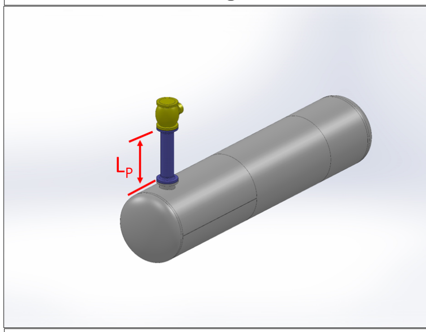
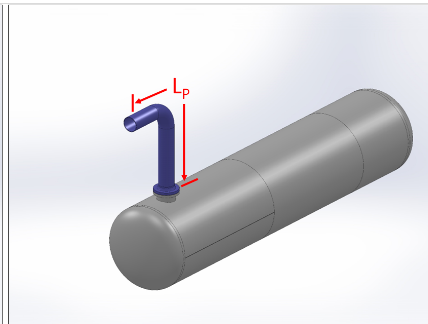
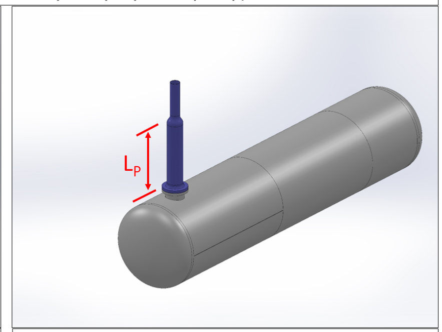
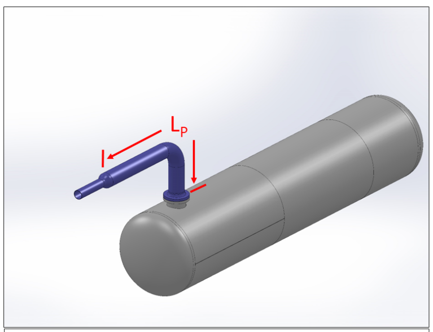
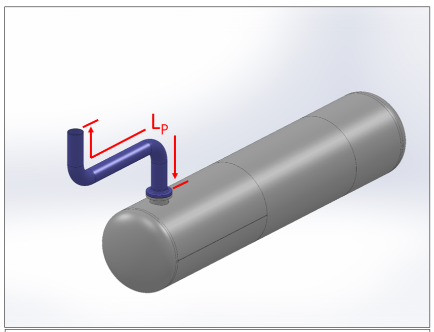
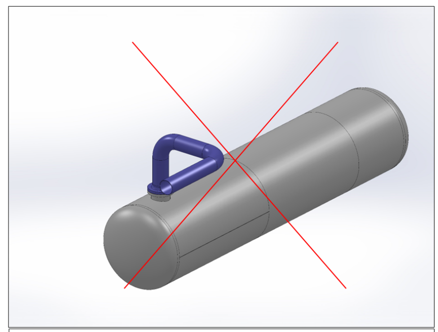
Inlet Flow Conditioning - Horizontal Vesssel - Top[1] Shell Inlet (Typical recommendations, may vary by Company).		
		
Valve and straight run pipe $L_P/D_P \geq 10$	Vertical bend and straight run pipe $L_P/D_P \geq 10$	Reducer and straight run pipe $L_P/D_P \geq 10$
		
Reducer and vertical bend $L_P/D_P \geq 10$ Given that: • Pipe IDs before and after reducer are less than 30% difference • Reducer is upstream of bend	Two bends in a vertical plane $L_P/D_P \geq 10$	Bends in both horizontal and vertical planes are too close to inlet. Not recommended.
Note 1. These recommendations also apply to center inlets on the top shell.		

Figure G.3 – Specified lengths and orientations of inlet pipe prior to a top shell inlet nozzle for horizontal vessels

Annex H

(informative)

Droplet Settling

The basic formulae of many vessel sizing criteria are based on drop settling or bubble rising. The typical force balance equation is shown in Figure H.1 describing a spherical particle (dispersed phase) falling in an upflow of fluid. If the sum of the forces, ΣF, is zero, the particle is stagnant/suspended. The particle terminal velocity V_T is then equal to the upflow velocity V_c. The force balance equation in Figure H.1 results in Equation H.1:

$$C_D \rho_c \frac{\pi d_p^2}{4} \frac{V_T^2}{2} = \left|\rho_c - \rho_d\right| \frac{g \pi d_p^3}{6} \tag{H.1}$$

where

V_T	=	Particle terminal velocity (m/s)
C_D	=	Drag coefficient of particle (dimensionless)
ρ_c	=	Continuous phase density (kg/m^3)
ρ_d	=	Particle density (kg/m^3)
g	=	Gravitational constant (m/s^2)
d_p	=	Particle diameter (m).

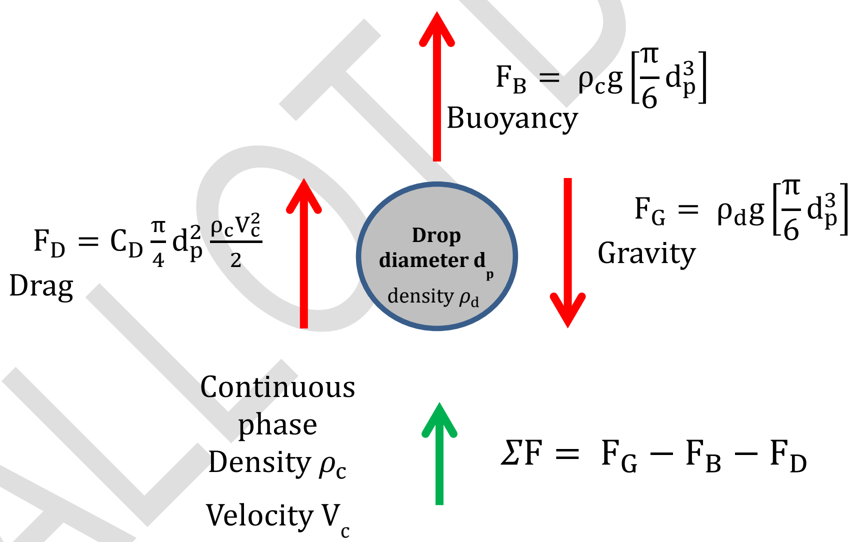


Figure H.1. Force balance on a particle settling through an upward flow

Equation H.1 assumes that the drop/bubble behaves like a rigid sphere with no interference from other drops/bubbles. We can solve for the terminal velocity as shown in Equation H.2.

$$V_T = \sqrt{\frac{4 g d_p}{3 C_D}} \cdot \sqrt{\frac{\rho_d - \rho_c}{\rho_c}} \tag{H.2}$$

The first term in Equation H.2 is typically denoted as the K factor (m/s) so that Equation H.2 can be rewritten as Equation H.3.

$$V_T = K\sqrt{\frac{\rho_d - \rho_c}{\rho_c}} \tag{H.3}$$

The unknown parameter is the drag coefficient, C_D. C_D is a function of the drop (dimensionless) Reynolds number [8]. The drop Reynolds number, Re, is defined as in Equation H.4:

$$Re = \frac{\rho_c V_T d_p}{\mu_c} \tag{H.4}$$

where

ρ_c = continuous phase density (kg/m^3)
V_T = droplet terminal velocity (m/s)
d_p = droplet (particle) diameter (m)
μ_c = continuous phase dynamic viscosity (kg/m-s)

Note that many people refer to the drop settling as Stokes flow, but Stokes flow is only for drop Reynolds numbers much less than one and is rarely the case.

Some curve-fit estimates of C_D vs the Reynolds number are listed in Table H.1 [8]. As noted in Table H.1, most gas/liquid separation applications are in the Newton's Law regime where C_D is relatively constant.

Table H.1 – Estimated value of drag coefficient depending upon the Reynolds number range [8].

Drag Coefficient C_D	**Re Regime**	**Description**
$\frac{24}{Re}$	Re < 0.1	Laminar region; Stokes Flow
$\frac{24}{Re}(1+0.14Re^{0.7})$	0.1 < Re < 1000	Intermediate range
0.445	1000 < Re < 350,000	Newton's Law region; Most liquids in g/l separations are in this regime
$0.19-\frac{80,000}{Re}$	Re > 1,000,000	Fully turbulent boundary layer

A ballistic model is typically use to calculate the maximum horizontal bulk velocity as shown in Figure H.2, an example of a droplet falling in the gas space.

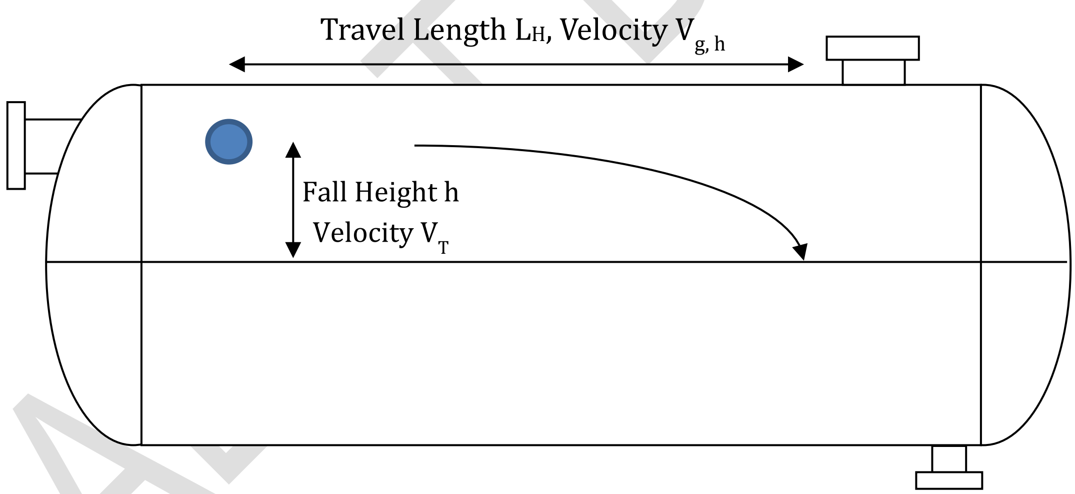


Figure H.2 – Ballistic model of drop settling.

The ballistic model assumes that the droplet will fall to the gas/liquid interface before reaching the gas outlet and is represented in Equation H.5:

$$\frac{h}{V_T}=\frac{L_H}{V_{g,h}} \tag{H.5}$$

where

h = fall height (m)
L_H = horizontal travel length (m)
V_T = droplet terminal velocity (m/s)
$V_{g,h}$ = droplet horizontal velocity (m/s).

The left-hand side of Equation H.5 is the time for the droplet to fall a distance h while the right-hand side is the time for the droplet to travel horizontally a distance L_H. Note that:

- L_H is an “effective” distance between inlet and gas outlet. See Figure 3 and Figure 4.
- Conservatively, h is distance from top of vessel to LAH or if applicable, LAH+foam layer.
- A similar ballistic model is used for oil drop rising in water or water drop settling in oil or gas bubble rising in liquid.

The horizontal velocity $V_{g,h}$ is given in Equation H.6:

$$V_{g,h} = \frac{Q_g}{A_{g,h}} \qquad (H.6)$$

where $A_{g.h}$ (m^2) is the gas space area corresponding to height h, and Q_g (m^3/s) is the gas flow rate.

To calculate the drop size that will settle out, V_T is calculated from Equation H5, then d_p from Equation H.2.

For vertical vessels, the drop has to settle against the vertical upflow velocity, $V_{g,v}$ (m/s) resulting in Equation H.7:

$$V_T = V_{g,v} \qquad (H.7)$$

where $V_{g,v}$ is given by $Q_g/A_{g,v}$ and $A_{g,v}$ (m^2) is simply the vessel cross sectional area. The drop diameter is then calculated from Equation H.2 [10].

Annex I

(informative)

Liquid Outlet Nozzle Froude Number and Submergence

Generally, the liquid outlet and drainage lines are analyzed using a Froude Number (Fr) limit. The Froude Number is defined by Equation I.1 together with Equation I.2:

$$\mathrm{Fr} = \frac{V_l}{\sqrt{g' D_l}} \tag{I.1}$$

where:

$$g' = \frac{g(\rho_H - \rho_L)}{\rho_H} \tag{I.2}$$

D_l	=	Liquid outlet nozzle ID (m)
g	=	Gravity acceleration (m/s^2)
V_l	=	Nozzle liquid velocity (m/s)
ρ_L	=	Light Phase density (kg/m^3)
ρ_H	=	Heavy Phase density (kg/m^3).

In lieu of other requirements, the nozzle may be sized for self-venting flow by using a Froude number limit of 0.31, i.e. Fr ≤ 0.31. Company specifications may allow a higher Froude number.

For larger diameter nozzles, above 24" ID, check that the trip levels (i.e. LSL and LISL) are providing sufficient submergence of the nozzles. Equation I.3 may be used to determine a minimum submergence level [11]:

$$\mathrm{Fr} \leq 1.6 \left(\frac{H_S}{D_l}\right)^2 \tag{I.3}$$

where the Froude Number, Fr, is given by Equation I.1 and

H_S = nozzle submergence (m).

Annex J

(informative)

Entrainment of Liquid

For vertical vessels, entrainment of drops from the gas-liquid interface should not be an issue if the design methods in this document (e.g. inlet piping, inlet device, spacings) are followed.

For horizontal gas flow over a liquid interface various drop entrainment mechanisms have been identified and an inception criterion for drop entrainment depending on two parameters, the liquid film Reynolds number, Re_f, defined in Equation J.1 and the viscosity number, N_μ [6], defined in Equation J.2:

$$Re_f = \frac{\rho_l V_B D_h}{\mu_l} \tag{J.1}$$

$$N_\mu = \frac{\mu_l}{\left[\rho_l \sigma \left(\frac{\sigma}{g(\rho_l - \rho_g)}\right)^{0.5}\right]^{0.5}} \tag{J.2}$$

where:

- ρ_g = gas density (kg/m^3)
- ρ_l = liquid density (kg/m^3)
- μ_l = liquid dynamic viscosity (kg/m/s)
- σ = gas-liquid interfacial tension (N/m)
- g = gravitational acceleration, (m/s^2)
- D_h = hydraulic diameter (m)
- V_B = bulk liquid axial velocity (m/s)

The hydraulic diameter is further defined in Equation J.3:

$$D_h = \frac{4A_f}{P_w} \tag{J.3}$$

where:

- A_f = liquid film flow cross sectional area (m^2)
- P_w = liquid wetted perimeter (m).

As an example, for a vessel of inside diameter D, half full of liquid, the flow area is $\pi D^2/8$ and the wetted perimeter is $\pi D/2$ so that $D_h = D$.

Gas velocities at the onset of drop entrainment developed for the five regimes are listed in Table J.1. For the turbulent regimes where $Re_f \geq 1635$, the onset velocity is a function of fluid properties only.

Typically, for horizontal separator design, the liquid flow is turbulent, and Re_f will be greater than 1635. It is recommended that the gas velocity above LAH (or if applicable, LAH+foam layer) be less than the value calculated by Table J.1 assuming $Re_f \geq 1635$ (last two rows of Table J.1). In addition, the assumption of a high Re_f will result in a lower, conservative gas velocity. If this velocity is critical, then it should be checked by calculating Re_f and using the appropriate regime of Table J.1. (For a three-phase separator, the calculations should be made using bulk liquid velocity and the light liquid properties.)

Table J.1 – Maximum gas velocity to prevent entrainment

Re_f regime	N_μ	Max gas velocity, m/s
$Re_f \leq 160$	--	$1.5\left(\frac{\sigma}{\mu_l}\right)\left(\frac{\rho_l}{\rho_g}\right)^{0.5} Re_f^{-0.5}$
$160 < Re_f < 1635$	$\leq 1/15$	$11.78\left(\frac{\sigma}{\mu_l}\right)\left(\frac{\rho_l}{\rho_g}\right)^{0.5} N_\mu^{0.8} Re_f^{-1/3}$
$160 < Re_f < 1635$	$> 1/15$	$1.35\left(\frac{\sigma}{\mu_l}\right)\left(\frac{\rho_l}{\rho_g}\right)^{0.5} Re_f^{-1/3}$
$Re_f \geq 1635$	$\leq 1/15$	$\left(\frac{\sigma}{\mu_l}\right)\left(\frac{\rho_l}{\rho_g}\right)^{0.5} N_\mu^{0.8}$
$Re_f \geq 1635$	$> 1/15$	$0.1146\left(\frac{\sigma}{\mu_l}\right)\left(\frac{\rho_l}{\rho_g}\right)^{0.5}$

Bibliography

1. Nelson, S. (2020), 0.1 gal/MMscf – A Century of Carryover, SPE Oil and Gas Facilities Savvy Separator Article, March.
2. Grave, E. and Green, S. (2016), Gas-Liquid Scrubber Performance Testing at Field-Like Conditions, AFS Spring presentation No. 2.1.2.
3. Morillo, E., van Asperen, V., and Baaren, S. (2016), Underperforming gas scrubbers, SPE Oil and Gas Facilities Savvy Separator Article, April.
4. Souders, M and Brown, G.G. (1934), Design of Fractionating Columns, Entrainment and Capacity, Industrial & Engineering Chemistry. 38 (1): 98–103.
5. Powers, M.L. (1990), Analysis of Gravity separation in Free Water Knock Outs, SPE 18205-PA.
6. Ishii, M. and Grolmes, M.A. (1975) Inception criteria for droplet entrainment in two phase concurrent film flow, AIChE Journal (Vol.21, No. 2), March.
7. IOGP S-619 Specification for Unfired, Fusion Welded Pressure Vessels – Version 2 (2022).
8. Green, D. W. and Perry, R. H. (2007), Perry's Chemical Engineer's Handbook, 8th edition, McGraw-Hill.
9. Francis, J. B. (1883), Lowell Hydraulic Experiments, D. Van Nostrand, fourth edition, New York, New York.
10. Grødal, E.O. and Realff, M.J. (1999), Optimal Design of Two and Three-Phase Separators: A Mathematical Programming Formulation, SPE 56645.
11. McDuffie, N. D. (1977), Vortex free downflow in vertical drains. AIChE Journal, 37-40.
12. API RP 14C Analysis, Design, Installation, and Testing of Safety Systems for Offshore Production Facilities
13. API Standard 521 Pressure-relieving and Depressuring Systems.
14. ASME Boiler and Pressure Vessel Code.
15. NACE MR0175/ISO 15156 Petroleum and natural gas industries — Materials for use in H2S-containing environments in oil and gas production.
16. NACE MR0103/ISO 17945 Petroleum, petrochemical and natural gas industries - Metallic materials resistant to sulfide stress cracking in corrosive petroleum refining environments.
17. Dries, H.W. and Hoffman, A.C. (2019), A correlation giving improved description of the capacity and efficiency of vane-type gas-liquid separators, AiChe J; 65:e16566.
18. Austrheim, T, Gjertsen, KL.H., and Hoffman, A.C. (2008), Experimental investigation of the performance of a large scale scrubber operating at elevate pressure on live natural gas, Fuel, June, V.87, Issue 7.
19. Austrheim, T, Gjertsen, KL.H., and Hoffman, A.C. (2007), Is the Souders-Brown equation sufficient for scrubber design? An experiment investigation at elevated pressure with hydrocarbon fluids, Chem. Eng. Sci., November, Vol. 62, Issue 21.