PROCESS ENGINEERING DESIGN GUIDE

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CLOUGH ENGINEERING & INTEGRATED SOLUTIONS

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CEIS CLOUGH ENGINEERING & INTEGRATED SOLUTIONS

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1 PURPOSE

This document offers guidance on the safe and practical design of a wide range of hydrocarbon processing systems, equipment and utilities commonly encountered by process engineers working in the Oil and Gas sector.

Similar documents have been developed by Clough in the past. However, these have always been issued within the framework of individual projects and have inevitably consumed much of the lead engineer's time during the process-critical phase. Projects are finite, key personnel are transient and historical data frequently comes without context. Consequently, in the absence of co-ordination, many of our good technical guidelines risk being lost.

With the intent of more consistently applying design practices and improving process efficiency, a consolidated set of project guidelines are summarised in this document. Efforts have been made to incorporate engineering lessons from past projects into the design guide. Where a lesson has been taken directly from our register, a designator (☞) has been provided in the page margin.

It is the responsibility of the lead process engineer to assess whether guidance presented here is applicable to particular projects. Much of the detail in this document may not be relevant for a given application; in other cases supplementary material will be required. The level of engineering development required for a particular project (i.e. conceptual definition, FEED, tendering, detailed design) will influence how the process guidelines should be applied.

2 DEFINITIONS

2.1 ABBREVIATIONS

ACOL	Air cooler software	**LP**	Low Pressure
ALPEMA	Aluminium Plate-fin Heat Exchanger Manufacturer's Association	**LPG**	Liquefied Petroleum Gas
ANSI	American National Standards Institute	**LR**	Long Radius
API	American Petroleum Institute	**LTL**	Lower Tan Line
AS	Australian Standard	**MDMT**	Minimum Design Metal Temperature
ASME	American Society of Mechanical Engineers	**NLL**	Normal Liquid Level
BTEX	Benzene, Toluene, Ethyl Xylene, and Xylene	**NPSH**	Net Positive Suction Head
CSA	Cross-sectional Area	**P&ID**	Piping & Instrumentation Diagram
CRA	Corrosion Resistant Alloy	**PCHE**	Printed Circuit Heat Exchanger
DN	Diameter Nominal (piping)	**PHE**	Plate (and frame) Heat Exchanger
FEED	Front End Engineering Design	**PSV**	Pressure Safety Valve
ID	Internal Diameter	**SMYS**	Specified Minimum Yield Stress
HP	High Pressure	**SS**	Stainless Steel
LAH	Level Alarm High	**TASC**	S&T heat exchanger software
LAHH	Level Alarm High-High	**TEG**	Tri-ethylene Glycol
LAL	Level Alarm Low	**TEMA**	Tubular Exchanger Manufacturer's Association
LALL	Level Alarm Low-Low	**UTL**	Upper Tan Line
LFL	Lower Flammability Limit		

2.2 NOMENCLATURE

A	Area	[m^2]
c	Erosional constant	[$kg^{0.5}/m^{0.5}.s$]
C	Drag co-efficient	[-]
d	Nominal diameter of steam piping	[inches]
D_P	Particle (droplet) diameter	[meters]
g	Acceleration due to gravity	[m/s^2]
h	Heat transfer co-efficient	[W/m^2K]
K	Vessel sizing co-efficient	[-]
MTD	Mean temperature difference	[°C]
Q	Duty	[kW]
Re	Reynolds number	[-]
T	Temperature, subscripts below	[°C]
	a = ambient	
	s = surface	
u	Velocity, general	[m/s]
U	Overall heat transfer co-efficient	[W/m^2K]
v	Fluid velocity, subscripts below	[m/s]
	i = inlet	
	o = outlet	
	G = gas or vapour	
	L = liquid	
	M = mixed phase	
v_t	Terminal velocity or allowable vapour velocity	[m/s]
VOL	Volume	[m^3]
ε	Emissivity	[-]
σ	Stefan Boltzmann Constant	
μ	Viscosity	[cP]
ρ	Density, subscripts below	[kg/m^3]
	G = gas or vapour	
	L = liquid	
	M = mixed phase	

3 PROCESS LINE SIZING

3.1 GENERAL

This section is intended to provide guidance to the process engineer in selection of piping line sizes.

The criteria presented here are considered to be in line with good engineering practice; similar criteria are widely applied throughout the oil and gas industry.

Although specific guidelines are presented below, in many cases the selection of line sizes must consider overall system requirements, gas train hydraulics or contractual obligations.

For example, the application of pressure loss criteria to some vapour services may result in unnecessarily large piping. This is particularly true where relatively short lengths are involved or where pressure loss is not an economic consideration. Fuel gas piping prior to let-down is a good example of this, where a large amount of pressure will otherwise dissipated across a control valve.

Conversely, where hydraulics are particularly tight (i.e. Sawan Development), these criteria may not satisfy project drivers.

It remains the responsibility of the process engineer to recognise where other factors may influence line sizing.

3.2 LINE SIZING CRITERIA

3.2.1 Liquid

Apply the more stringent of the following criteria:

Table 3.1(a) – Liquid Line Sizing Criteria

Service	Line Sizing Criteria		
	Normal Pressure Loss (kPa/100 m)	**Velocity, Normal Maximum (m/s)**	**Notes**
Centrifugal pump suction, sub-cooled liquid	10 – 20	1.0 – 1.5	1, 2
Centrifugal pump suction, bubble-point liquid	≤ 10	≤ 1.0	1, 2
General, sub-cooled liquid	50 – 70	2.5 – 3.5	-
General, bubble-point liquid	≤ 10	1.0 – 1.5	-
Gravity flow	5	≤ 1.0	-

The piping associated with reciprocating pumps is a special case. The following typical velocity limits have been taken from API RP 14E [Ref 31]. Guidance is given in Section 6.2 with regard to piping design and routing.

Table 3.1(b) – Specific Liquid Line Sizing Criteria for Reciprocating Pumps

Pump Speed	Suction Velocity (m/s)	Discharge Velocity (m/s)	Notes
< 250 rpm	0.61	1.80	1, 2
250 – 330 rpm	0.46	1.40	1, 2
> 330 rpm	0.30	0.91	1, 2

<u>Notes</u>:

1. NPSH requirements shall be considered in selection of the final line size.
2. Full bore valves are to be employed on pump suction lines at DN 150 and below.

3.2.2 Vapour

Apply the more stringent of the following criteria:

Table 3.2 – Vapour Line Sizing Criteria

Service	Line Sizing Criteria		
	Normal Pressure Loss (kPa/100 m)	**Momentum Criteria (metric)**	**Notes**
General < 700 kPag	11 - 23	146 (Note 1)	Note 2
General, 700 - 5000 kPag	23 – 40		Note 2
General, 5000 - 7000 kPag	40 – 50		Note 2
General > 7000 kPag	50 – 70		Note 2
Compressor Suction	< 11		Note 2, 3

Notes:

1. The momentum criterion limits gas velocity to attenuate erosion and piping noise [Ref 1]. The maximum normal velocity is given by:

$$v_G < \frac{146}{\sqrt{\rho_G}}$$

   Apply to both carbon steel and corrosion resistant alloys.

2. These criteria shall not be applied to flare system piping (i.e. PSV lead and tail pipes, flare headers, etc). Refer to Section 9.1.3.
3. Applies between suction scrubber and compressor inlet nozzle only.

Additionally, API RP 14E [Ref 2] suggests that noise may be a problem where the vapour velocity exceeds 18.3 m/s (60 ft/s). However the recommended practice states that this figure should not be viewed as a rigid limit. Higher velocities are generally acceptable when the piping has been laid out (and valves selected/placed) to minimise or isolate noise.

3.2.3 Two-Phase

Two-phase process piping (solids free) is to be sized using the following erosional limits, as set out in API 14E [Ref 2].

Table 3.3 – Multiphase Line Sizing Criteria

Service	Variant	Erosional Criteria (c)		Notes
		Imperial	**Metric**	
Continuous, carbon steel	-	100	122	Notes 1, 2
Intermittent, carbon steel	-	125	152	Notes 1, 2
Continuous operation where corrosion is not anticipated or is controlled (inhibition, corrosion resistant alloys)	Onshore	120	146	Notes 1, 2, 3
	Offshore	200	250	Notes 1, 2, 3
Intermittent operation where corrosion is not anticipated or is controlled (inhibition, corrosion resistant alloys)	-	250	300	Notes 1, 2, 3

Notes:

1. Where the two-phase system is part of a critical hydraulic path, the kPa/100 meter criteria given in Section 3.2.2 shall be applied here also. Examples: (a) between gas train cooler and separator; (b) between turbo-expander and LPG absorber.
2. The maximum normal velocity is given by:

$$v_M < \frac{c}{\sqrt{\rho_M}}$$

3. The use of c = 250 for sizing of onshore two-phase piping systems is not seen as reasonable, even where piping is fabricated from corrosion-resistant alloys. This figure is only applicable to relatively short offshore flowlines and piping manifolds. For two-phase onshore applications, where corrosion is not a concern, c = 146 should be employed.

Fluid lines exhibiting anything other than mist flow must be carefully sized and routed to avoid slugging. Where this is a concern, it is the responsibility of the process engineer to designate lines as having ‘no pockets’ on the P&IDs. This may have a significant effect on the piping layout and critical equipment elevations. The requirement for ‘no pockets’ should therefore be reviewed on all major two-phase lines at the earliest possible juncture.

☞ Where slugging is anticipated, this should be clearly indicated via notes on the P&ID and the piping line list. The piping group may require basic slug data for the design of supports and stress calculations. The maximum slug velocity should generally be taken as the vapour superficial velocity at steady-state conditions. An estimate of the slug volume can usually be made by reviewing the piping layout. For example:

Figure 3.1 – Slug Volume Estimate

3.2.4 Utility System Piping

3.2.4.1 Instrument Air

Instrument air headers should be sized to limit pressure loss to approximately 10 kPa/100 meters [Ref 3].

3.2.4.2 Hot Oil

Hot oil headers should be sized to limit the normal pressure loss to 35 kPa/100 meters. Velocities should generally be limited to 3 m/s.

3.2.4.3 Steam

Table 3.4 – Steam Header Line Sizing Criteria

Service	Normal Pressure Loss (kPa/100 m)	Max Velocity (m/s)
Saturated, 0 – 2 barg	3.3% of static pressure (kPa abs) per 100 m	$15\sqrt{d}$
Saturated, 2– 17 barg	13 – 29	$12\sqrt{d}$
Saturated, 17 – 69 barg	30 – 68	$9\sqrt{d}$
Superheated, < 17 barg	5 – 17	30 – 45
Superheated, > 17 barg	2.3% of static pressure (kPa abs) per 100 m	$15\sqrt{d}$

Where d is the nominal diameter of the steam header expressed in inches.

Note that these values are given for in-plant steam headers. Short branch lines can generally be sized for twice the velocity and pressure loss criteria defined above.

3.2.4.4 Cooling Water

Table 3.5 – Cooling Water System Line Sizing Criteria

Service	Normal Pressure Loss (kPa/100 m)	Maximum Velocity (m/s)
Cooling Water Headers	10 – 15	4.0 – 5.0
Cooling Water Branch Lines	40 – 50	4.0 – 5.0

3.2.4.5 Fire water

Fire water headers and branch lines are to be based on the relevant codes and standards.

3.3 HYDRAULIC CALCULATIONS

3.3.1 Absolute Roughness

Hydraulic calculations shall employ an absolute roughness of **45 micron** for all carbon steel and stainless steel in-plant piping.

The use of a lower roughness value should only be considered in special circumstances. A roughness factor of 30 micron, for example, is often successfully applied to onshore/offshore transmission pipelines (where the mean distance between welds and flanges is greater than for in-plant piping).

3.3.2 In-Plant Hydraulic Calculations

3.3.2.1 Single Phase

In-plant hydraulics should be examined using the latest revision of Clough's validated compressible and incompressible fluid hydraulic spreadsheets. These spreadsheets include a considered approach to equivalent length estimation, which is generally based on the velocity head data given in Crane TP 410M. For additional details, refer to Reference 4.

The hydraulic spreadsheets can be easily applied to most of the relatively uncomplicated hydraulic checks executed as part of gas plant design (i.e. PSV lead and tail pipes, pump suction-side NPSH calculations, gas train hydraulic profiles).

A margin of 20% applied to the system equivalent length is recommended for preliminary hydraulic calculations. This margin may be reduced to 10% when piping isometrics have been finalised.

The use of specialist software is recommended for more complicated systems such as steam or cooling water circuits and large flare systems.

3.3.2.2 Two Phase

In-plant hydraulic calculations for two phase flow should be examined using the Hysys pipe segment module. The *Beggs & Brills* correlations are recommended; the process engineer should however perform a brief sensitivity with other flow-pressure correlations.

Estimation of the pipe system equivalent length using the Hysys module is cumbersome. A more convenient and reliable estimate can generally be made using the head loss data presented in Reference 4.

Caution should be exercised when reading vapour and liquid phase velocities from the pipe segment module output tables. Inconsistencies have been noted in the past, and the basis for the velocity calculations is not clear. The average velocity in the segment should be manually calculated by dividing the actual volumetric flow by the mixed fluid density.

Margins should be applied as per single phase (Section 3.3.2.1).

3.3.3 Transmission Pipelines

Pipesim™ shall be used for steady-state modelling of single phase transmission pipelines (export) and multiphase production delivery systems (raw gas pipelines and substantial flowlines). This applies to the sub-sea, subterranean and surfaced-laid variants of each. Pipesim™ can be used to estimate thermal and hydraulic profiles, liquid hold-up volumes and slugging propensity.

For further details refer to CEIS-ENG-EN-P-1022 (issue pending).

4 HEAT EXCHANGER DESIGN

4.1 SHELL-AND-TUBE HEAT EXCHANGERS

4.1.1 Tube-Side Fluid

The following guidelines should be used in selection of the tube-side fluid. In many cases, more than one of the following will apply. Where a conflict exists the lead process engineer should be consulted.

1. **Place higher pressure fluid on the tube-side**

   This will generally result in a less expensive design than for fabrication of a high-pressure shell.

   Example: HP process gas against treated cooling water. Used on Sawan for Inlet Trim Cooler (E-2002) where process had high design pressure (97 barg); cooling water design pressure was 11.3 barg.

2. **Place more corrosive fluid on the tube-side**

   Fabrication of tubes from a CRA is less expensive than fabrication of the entire shell from a similar material (which would include baffles, tie-in rods, etc).

   Example: HP process gas against treated cooling water. Process side material for Sawan Inlet Trim Cooler (E-2002) was Duplex SS due to elevated acid gas levels and condensing service. This fluid was therefore placed on the tube-side of the heat exchanger.

3. **Place more fouling fluid on the tube-side**

   It is generally easier to clean the tube-side of a heat exchanger (mechanically, chemically). Through proper design (i.e. multiple passes) high tube-side velocities can be achieved which limit deposition also.

4. **Place less viscous fluid on the tube-side**

   Where one fluid is particularly viscous, placement on the shell-side should be considered. Tube-to-tube maldistribution is possible when cooling viscous liquids inside of tubes.

5. **Place significantly higher flow on the tube-side**

   For a given flowrate, the shell-side pressure drop will usually be greater than for the tube-side.

   Example: Feed/Residue Gas heat exchanger, where similar design pressures are applied to both sides and shrinkage through the gas train results in an appreciably lower residue (sales gas) flow. Feed gas is also likely to be more fouling/more corrosive, further strengthening argument for placement on tube-side.

6. **Place high temperature fluid on the tube-side**

   Where one fluid operates at an elevated temperature placement on the tube-side will reduce environmental heat losses and the insulation thickness.

   Example: Hot oil or steam heated reboilers.

7. **Place condensing fluid on tube-side**

   Liquid pocketing and blanketing of the heat transfer surface may be experienced in two-phase shells. This can be addressed by placing the condensing fluid on the tube-side. Top entry / bottom exit nozzles are generally recommended for a condensing tube-side to minimise the potential for slugging within the tubes.

4.1.2 TEMA Configuration

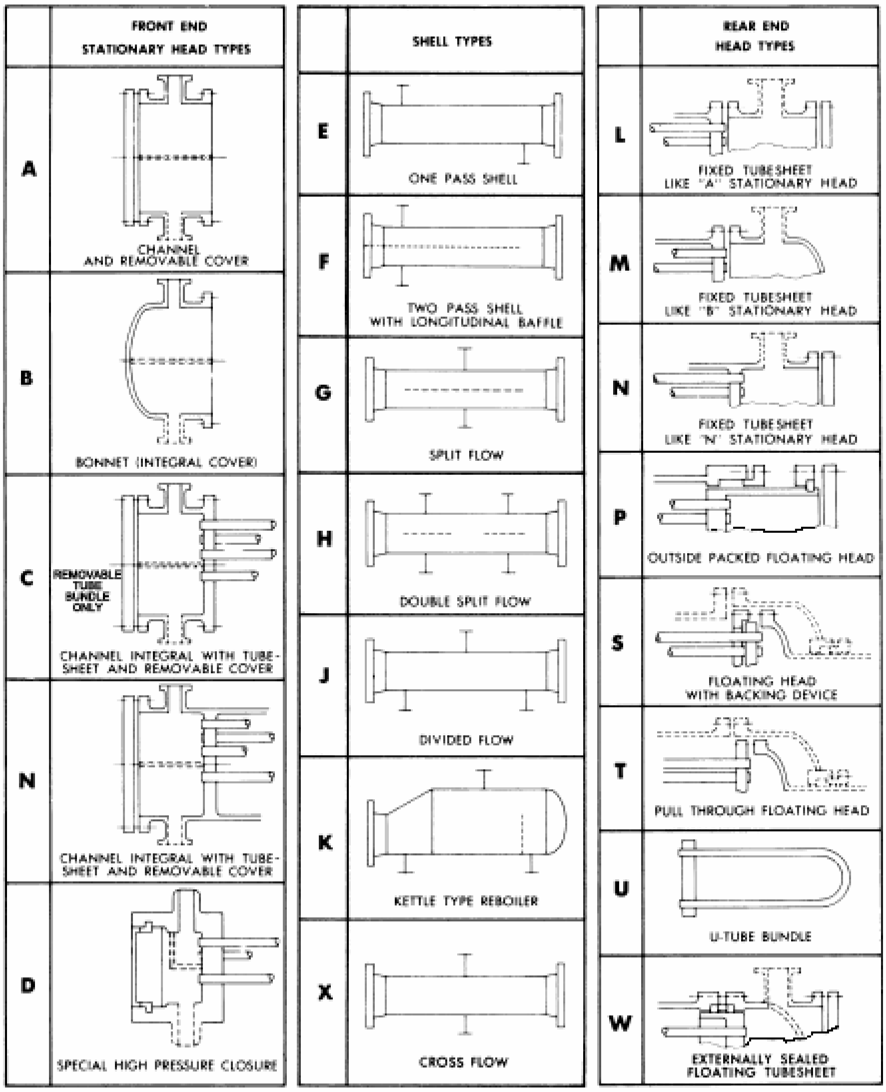
4.1.2.1 General

The following guidelines may be used for preliminary selection of a TEMA configuration in the design of shell-and-tube heat exchangers. In all likelihood, more than one configuration will be suitable for a given service. Where the process engineer has a particular requirement, this should be communicated on the equipment datasheet. In many cases, however, the supplier will be given freedom to select the final configuration in order to optimise the design.

The following guidelines are more specifically provided to aid the process engineer in:

- Ruling out elements of particular TEMA designs which are not suitable;
- Offering design guidance to the supplier by way of datasheet notes; and
- Performing a sense check of the supplier's selected configuration.

4.1.2.2 Shell-and-Tube Nomenclature



The following definitions are provided to assist the discussion given in Sections 4.1.2.3 to 4.1.2.5:

Bonnet	**:**	Head assembly used to distribute tubeside medium. It may contain both the tubeside inlet and outlet connections and a baffle. A bonnet differs from a channel in that it does not have a removable cover.
Channel	**:**	Head assembly used to distribute tubeside medium. Fitted with a removable cover to more easily access tubes than a bonnet.
Fixed Tubesheet	**:**	A non-removable tubesheet, welded to the shell. A fixed tubesheet may be present at one end (i.e. BFU construction) or both ends (i.e. BEL).
Integral Tubesheet	**:**	Tubesheet that is welded or integral to a channel cover. Where N front or read heads are applied, the tubesheet is integral to both the channel head <u>and</u> the shell.
Floating Tubesheet	**:**	The tubesheet at one end of a removal bundle (U-tube exchangers are the exception). The floating tubesheet will always have a smaller diameter than the stationary. The floating tubesheet can move freely as the bundle thermally expands and contracts.
Stationary Tubesheet	**:**	The tube sheet at one end of a removable bundle. The stationary tubesheet has a larger diameter than the floating tubesheet and is held in position between the shell flange and the bonnet or channel.
Stationary Head	**:**	Channel or bonnet style head associated with a fixed or stationary tubesheet.
Floating Head	**:**	The various head types associated with floating tubesheets.

4.1.2.3 Removable Bundle

Where the tube bundle of a heat exchanger is to be removable, the use of exchanger types incorporating two fixed, stationary or integral tube sheets is not suitable.

The requirement for a removable bundle may be due to a particular client specification or placement of a fouling fluid on the shell-side of a heat exchanger.

However - in terms of construction - fixed tube-sheet heat exchangers are extremely low cost and offer lower potential for shell-side leakage since flanged joints have been omitted.

Therefore, where we are not absolutely compelled to specify a removable bundle, careful consideration must be given to whether this is a necessary design feature. Note that although mechanical cleaning of the tube exterior (and shell) would not be possible with two fixed, stationary or integral tubesheets, if a satisfactory in-situ chemical cleaning programme can be employed then a non-removable bundle may still be acceptable.

Where the bundle must be removable:

- Front head type N is <u>not</u> suitable.
- Front head types B (least expensive) and A can be used in combination with a stationary (but not fixed) tubesheet. Front head A permits easier inspection of the tubesheet but at greater expense; the channel flange also generates an additional potential leak source. This should only be specified in preference to B where required by a particular client.
- Front head type C is suitable (with its integral but removable tubesheet).
- Rear head types L, M and N are <u>not</u> suitable.
- Rear head type U (U-tube) is acceptable, and offers a low cost solution when combined with front head types B (least expensive), A or C.

- Rear head type S (floating head with backing device) is suitable, but is considerably more expensive than a U-tube heat exchanger.
- Rear head type T (pull through floating head) is suitable, but is more expensive than both the U and S (as a larger shell is required). The bundle is however easier to remove than with an S rear head; this may be suitable where very rapid maintenance turn-around is required. This should only be specified where the shell diameter is governed by other constraints (i.e. in a kettle reboiler, K-shell) or where required by a particular client.
- Rear head types P (outside packed floating head) and W (outside packed lantern ring) allow bundle removal, but are prone to leakage. These should only be considered with shell-side fluids that are non-hazardous and non-toxic and that have moderate pressures (circa 39 barg and 300°C). Rear head types P and W should generally be avoided.

Note that where regular shell-side cleaning is envisaged, the use of triangular pattern tubes will not permit effective mechanical cleaning towards the centre of the bundle. A square pattern tube layout should therefore be considered. This may however result in a larger bundle (and a potentially larger shell).

The following discussion applies where a non-removable bundle can be justified. However, the design implications of thermal expansion have not been fully addressed. This assessment should be made by the equipment suppler:

- The use of front head type C is pointless where a fixed tubesheet is used at the rear end.
- Rear head types P, S, T and W are not applicable.
- Read head types L, M and N are all acceptable.
- Front head type B (with a fixed tubesheet) is suitable. When combined with rear head type M, this configuration offers a very cost effective design. The B-M combination permits ‘rodding’ of the tubes, but requires that the two bonnets be completely removed and the tube-side process connections unbolted.
- Front head types A (with a fixed tubesheet) and N are suitable. A-L and N-N front and rear head combinations permit ‘rodding’ of the tubes without having to unbolt process connections; maintenance is therefore easier than for a B-M exchanger (above).
- Although B-L and A-M combinations are workable designs, these should not be specified. The use of a removable channel cover (A and L) on one end only offers neither a cost advantage nor clear advantages in terms of maintenance or operability.
- Where an N type front head has been specified, an N type read head should generally be specified also.
- Where fixed or stationary tubesheets are employed and a large temperature difference exists between the shell and tube-side fluids the use of a shell expansion joint may be necessary. This requirement should be advised by the equipment supplier.
- U-tube bundles are acceptable and can be combined with A, B or N front heads. Even where a high temperature differential exists between the shell and tube side fluids, a shell expansion joint will not be required.

4.1.2.4 Acceptability of U-tubes

Where a fouling or scaling fluid exists on the tube-side of a heat exchanger, chemical or mechanical cleaning (or both) may be required to remove deposits.

All combinations of front and rear head, except U-tube, will allow ‘rodding’ or other means of mechanical cleaning of the tubes. However, front and rear head combinations such as A-L, C-L and N-N allow easier access than the generally less expensive B-M design.

If mechanical cleaning of the tube-side is required, due to a high potential for fouling or scaling (or due to the requirements of a particular client), the use of a U-tube design may not be acceptable.

Where we are not absolutely compelled to specify 'roddable' tubes by client requirements, careful consideration must be given to whether this is a necessary design feature. For most large, high-pressure heat exchangers a U-tube design is likely to be the least expensive TEMA configuration. This is particularly true where a removable bundle is required (i.e. provision of an S or T rear head can be avoided) or where a large temperature differential exists between the shell and tube-side fluids (i.e. a shell expansion joint is therefore not required).

4.1.2.5 Selection Chart

The guidelines given above have been summarised in a selection chart. Refer to Figure 4.1.

4.1.3 Multiple Shell Passes and Multiple Shells

Where two tube passes are applied, temperature crosses can be avoided by using a baffled shell (TEMA F) to maintain counter-current flow. However, where four or more tube passes are applied, a multiple shell design may be required to avoid crosses within the exchanger.

4.1.4 Tube Velocities in the presence of Corrosion/Hydrate Inhibitors

There is a special consideration where a corrosion or hydrate inhibitor is injected upstream of the heat exchanger or directly onto the tubesheet. In order to achieve good distribution in all tubes, a certain minimum velocity must be maintained. From test, it has been established that [Ref 29]:

$$v_m < \frac{26}{\sqrt{\rho}}$$

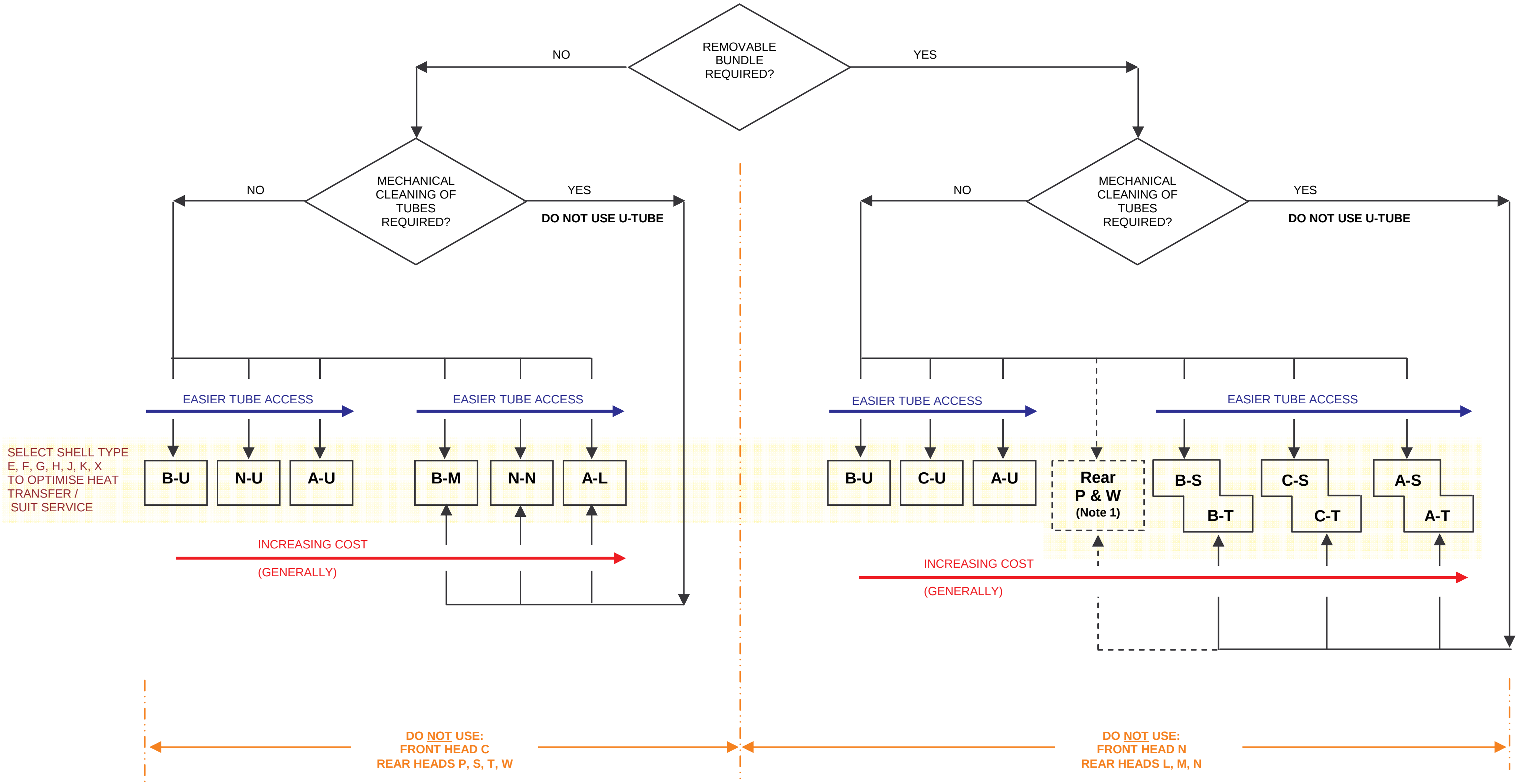
Where,

v_m is the minimum tube velocity (m/s), and

ρ is fluid density.

For gases, the average exchanger pressure and temperature should be used in determination of ρ [Ref 29].

Figure 4.1 – TEMA Configuration Selection Chart



NOTES:

1. Rear head types P & W are prone to leakage and are not commonly used in the Oil & Gas industry. Where a removable bundle is required these heads should only be considered when the shell-side fluid is non-toxic, non-hazardous and at moderate pressures (< 39 barg) and temperatures (<300 °C).

4.1.5 Design Margins

All general service shell-and-tube heat exchangers shall be furnished with a 10% excess surface area margin. This margin is to apply in the fouled (service) condition and should generally be achieved by installing additional tubes (not by lengthening tubes)[1]. This requirement should be clearly communicated on the process datasheet.

The equipment supplier shall also guarantee that both the shell and tube sides will be free of flow induced vibrations at up to 110% of the design rates.

Reboilers and condensers are a special application of shell-and-tube heat exchangers. Appropriate margins are described in Sections 4.1.7 and 4.1.8 respectively.

4.1.6 Fouling Factors

The following fouling factors are taken from Engineering Note EPI-ENG-EN-P-1003 [Ref 5]. These resistances shall be applied to both the shell and tube sides, as appropriate.

Table 4.1 – Shell-and-Tube Fouling Factors

Service	**Variant**	**Fouling Resistance (m^2.K/W)**	**Remarks**
Natural Gas	Clean	0.00018	Processed gas free of residues/ precipitates. For example, D/S of an inlet separator with high efficiency internals.
	Dirty	0.00035	Raw gas upstream of scrubbing of filtration system.
Light Hydrocarbon Reboilers	Clean	0.00035	Services protected from residues and precipitates from upstream reboiler. For example, debutaniser fed from depropaniser only.
	Dirty	0.00053	Services subject to residue/precipitates from water carry-under, etc. For example, stabiliser reboiler.
Light Hydrocarbon Processing	Condensers	0.00018	-
	Product Coolers/ Heaters	0.00018	-
	Unprocessed Condensate	0.00035	Upstream of stabilisation. For example, stabiliser pre-heater.
Heat Transfer Fluid (Hot Oil)	-	0.00035	-
Steam	Clean	0.00009	-
	Oil Bearing	0.00030	-
Steam Condensate	-	0.00090	-
Treated Cooling Water	< 50℃	0.00018	-
	> 50℃	0.00035	-
TEG	-	0.00035	-
MDEA	-	0.00035	-

[1] Provision of excess surface area by lengthening the tubes will increase the tube-side pressure loss. However, by installing 10% additional tubes instead, the tube-side restriction to flow is not increased. Also, if a tube fails, then an exchanger with 10% additional tubes will retain a larger portion of its original surface area should tube removal and plugging prove necessary.

4.1.7 Reboilers

4.1.7.1 Selection

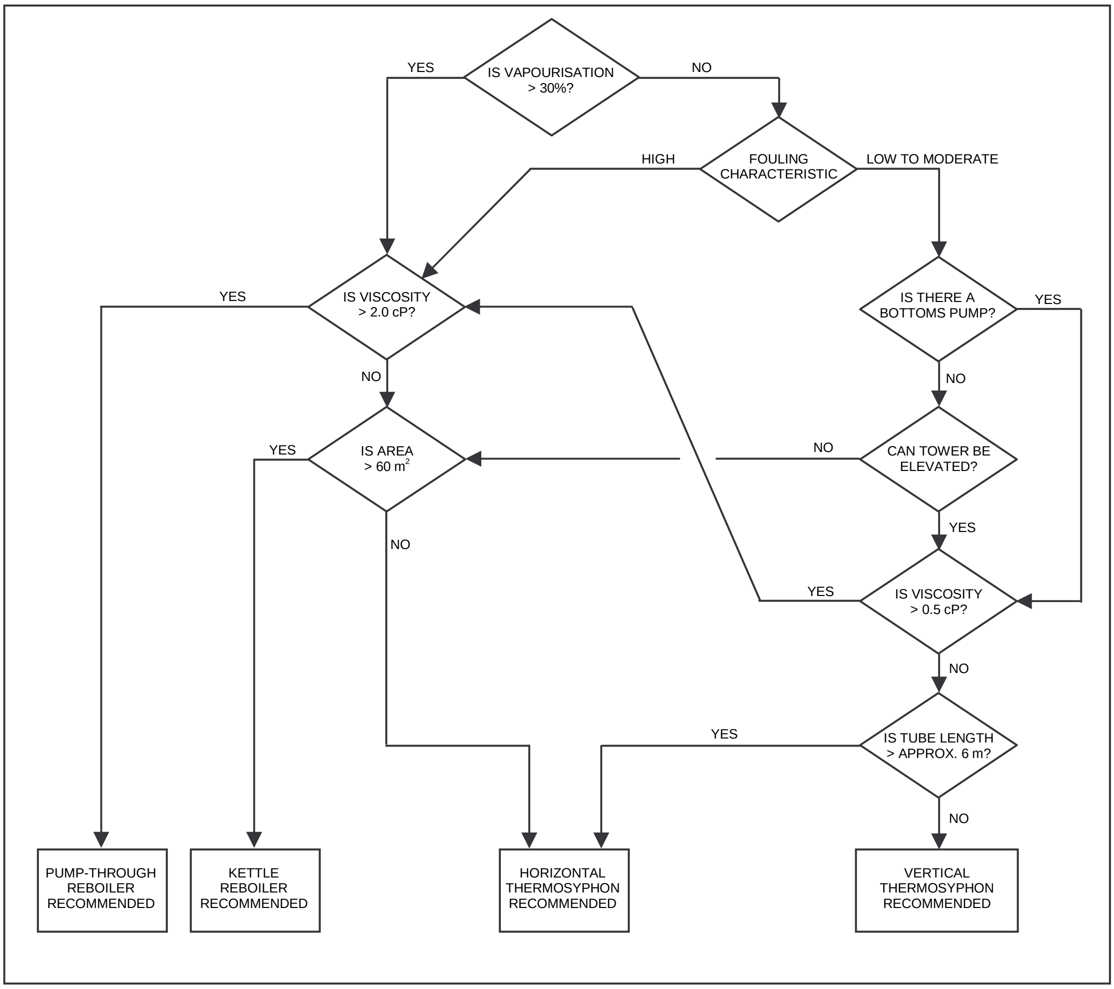
Kettle-style (K shell) heat exchangers should be employed in most light hydrocarbon applications. This type of exchanger is also routinely used to reboil stripper bottoms in amine regeneration systems. Kettle-style reboilers offer an exceptionally good turndown capability.

Vertical thermosyphon reboilers may be cost competitive in relatively small-scale, light hydrocarbon applications. As vertical thermosyphons can be mounted on the column itself, they are less space intensive and do not require a separate foundation. However, thermosyphon reboilers are limited to around 50% turndown. This limitation is based on maintenance of an acceptable flow regime in the vertical tubes.

In a similar vein, thermosyphon reboilers are generally limited to applications where vaporisation is less than 30%. Most designs are based on slug flow to assist heat transfer. At higher vapour fractions, the regime moves towards mist flow and a reduction is the heat transfer co-efficient is experienced. Kettle reboilers do not have a prescribed limit on the percentage vaporisation.

A selection guide is given below for reboilers [Ref 6]. However, this chart does not address turndown. It is noted that the vast majority of fractionation columns encountered by Clough will have a turndown requirement – expressed as a contractual obligation or inherent to production in a *declining field* – which may make use of a thermosyphon reboiler impractical or inefficient.

Figure 4.2 – Reboiler Selection Chart



4.1.7.2 Kettle Reboilers

The following shall be considered in the design of kettle-type reboilers.

- Kettle reboilers shall be furnished with 20% excess surface area by considering 120% of the controlling case flows in thermal and hydraulic design of the heat exchanger. This margin is to apply in the fouled (service) condition. The equipment supplier shall also guarantee that both the shell and tube sides will be free of flow induced vibrations at up to 120% of the design rates.

  These margins should provide adequate flexibility for column operation. The 20% margin on flow should also be considered in hydraulic design of the reboiler piping circuit (see below).

- To limit liquid entrainment, the freeboard between the liquid level (weir height) and the vapour outlet should be a minimum of 250 mm. The vapour velocity should be limited to [Ref 26]:

  $$u < 0.2\left[\frac{\rho_L - \rho_G}{\rho_G}\right]^{0.5}$$

- Square tube layouts should generally be used in boiling (and fouling) services, but come at the expense of a larger tube bundle. However, in most kettle reboilers, the depth of the tube bundle has only a second-order effect on the size of the shell. Therefore, a square tube pattern should not be significantly more expensive when applied to kettles. When the shell-side Reynolds number is low (< 2000) the use of a rotated square (45°) pattern may improve heat transfer [Ref 7].

- Shell-side pressure losses should be limited to 1-1.5 kPa.

- Multiple vapour outlet nozzles should be considered for long tube bundles, particularly where the shell-side vapours are corrosive (i.e. amine reboiler).

- Where product is drawn directly from the reboiler, sufficient residence time is to be provided downstream of the weir for accommodation of control and shutdown level settings.

  To avoid liquid short-circuiting, the NLL should be set at the level of the weir. The LAH and LAHH can therefore be accommodated above the weir, utilising the full length of the reboiler for level co-ordination.

  Based on the design product rate, a residence time of 1 minute between the NLL and LAL is reasonable where the product leaves the reboiler under level control. A further 1 minute allowance should be made between LAL and LALL. With the LALL set sufficiently above the bottom inside of the shell (see below) this should provide a total residence time of approximately 3 – 4 minutes below NLL downstream of the weir.

  The total volume requirement downstream of the weir should be communicated on the process datasheet, together with the requirement for level bridle nozzles, etc.

- Where product is drawn directly from the reboiler, a vortex breaker is to be fitted to the liquid outlet nozzle. The liquid outlet nozzle should be sized to limit the outlet velocity to less than 1 m/s. Where this is the case, the minimum liquid level (LALL) should be set 200 mm above the vortex breaker. If the outlet velocity is greater than 1.2 m/s, then the LALL should be set using the following equation:

  $$\text{LALL (mm above top of vortex breaker)} = 200 + 250(u - 1.2)$$

  Where u = liquid velocity (m/s)

- A review of the *reboiler circuit* hydraulics should be undertaken prior to purchasing the heat exchanger. The nozzle sizes generated by the equipment supplier may be smaller than the minimum line sizes (liquid supply and vapour return) required to prevent excessive backing up of liquid in the column (relative to the reboiler NLL). Minimum liquid inlet and vapour outlet nozzles sizes should therefore be provided on the process datasheet, to prevent having to swage down and up respectively.

  A total reboiler circuit pressure loss equal to 500 mm of feed liquid is seen as reasonable (feed liquid piping → reboiler → vapour return piping). With, say, 1.5 kPa across the reboiler (flange-to-flange) this would allow between 200 – 350 mm of pressure loss through supply and return piping, depending on bottoms density. At design rates this would give a liquid level in the column 500 mm above the reboiler weir (NLL).

- The vapour return piping shall contain no pockets.
- Where an unavoidable low point exists in the reboiler liquid feed piping, a maintenance drain should be provided.

4.1.7.3 Vertical Thermosyphon Reboilers

The following shall be considered in the design of vertical thermosyphon reboilers.

- In general, thermosyphon reboilers shall be furnished with 20% excess surface area by considering 120% of the controlling case flows in thermal and hydraulic design of the heat exchanger. This margin is to apply in the fouled (service) condition. The equipment supplier shall also guarantee that both the shell and tube sides will be free of flow induced vibrations at up to 120% of the design rates.

  However, it is noted that limited turndown is achievable with thermosyphons. Where low turndown is required relative to the design rate, the lead process engineer may elect to reduce the margins to those given in Section 4.1.5. Both the margin and turndown requirements shall be clearly defined on the process datasheet.

- Hydraulic design of the reboiler and associated piping is critical to performance. As a general guide, the liquid velocity in the feed piping is typically 0.7 – 2.0 m/s [Ref 29]. The allowable velocity in the return line is usually limited to [Ref 29]:

  $$v_R = \sqrt{\frac{6000}{\rho_m}}$$

  Where,

  v_R = return velocity (m/s)

  ρ_m = mixed phase density in the return line (kg/m^3)

  As a final check, the cross-sectional area of the return line should be approximately twice the total tube open area [Ref 30]. The return line should be as short as possible.

- For natural circulation reboilers, the liquid level in the column must be adequate to overcome pressure losses through the feed line, the heat exchanger and the return piping. As a general guide, the top tubesheet of a vertical thermosyphon should be aligned with the lowest operating level (LAL).

  Where a wide range of column levels is expected (i.e. LAL - LAH is a significant dimension), an associated level rise can be expected in the thermosyphon. This is detrimental to heat transfer as a zone controlled by nucleate boiling is replaced by convective transfer only.

  To address this, a manual butterfly valve (or equal) is often installed in the feed piping to compensate for excess levels.

4.1.8 Condensers

The piping around column overhead condensers (and other shell-and-tube exchangers in condensing service) should be arranged for downward process flow (i.e. top entrance nozzles, bottom outlet nozzles). Downward flow should greatly reduce the likelihood of slugging within the tubes.

Apply to both partial and total condensers.

Condensers in distillation service shall be furnished with 20% excess surface area by considering 120% of the controlling case flows in thermal and hydraulic design of the heat exchanger. This margin is to apply in the fouled (service) condition. The equipment supplier shall also guarantee that both the shell and tube sides will be free of flow induced vibrations at up to 120% of the design rates.

These margins should provide adequate flexibility for column operation.

4.1.9 Evaluation and Preliminary Sizing

Shell-and-tube evaluation (rating) and preliminary sizing (if required) should be performed using Aspentech TASC.

4.2 AIR-COOLED HEAT EXCHANGERS

4.2.1 Configuration

Air-cooled heat exchangers should - in general - be specified as **forced draft**. For most applications encountered by Clough, improvements in terms of maintainability (of the fan and bearing assembly), ease of installation and reduced motor horsepower outweigh the fairly modest reduction in the potential for hot air recirculation reported for induced draft designs.

It is noted, however, that many clients have a preference for either forced or induced draft air-coolers. The relevant client specification should be consulted.

Where induced draft air-coolers are employed, mounting of the motor assembly below the heat exchanger (with the shaft extending upwards through the tube bundle) offers an improvement to maintainability.

4.2.2 Fin Types

4.2.2.1 Onshore

For process temperatures less than 150℃, **extruded** or **'Double L'** (double-footed) fins should generally be employed. Although historically tension-wrapped fins (such as the 'Double L') are considered to be the less expensive option, experience gained from particular projects has demonstrated that the extruded type is cost competitive, if not less expensive, in some applications.

For process temperatures above 150℃, **embedded (G designation)** fins should be employed.

4.2.2.2 Offshore (and Other Aggressive Environments)

Tension-wrapped fins ('Double L 'or 'L') should <u>not</u> be used offshore, or in other locations where airborne contaminants are prevalent (i.e. remote desert locations). Sea spray is of particular concern; salt build-up between the fin and tube is possible as the extended surface is wrapped, not bonded.

Below 150℃ **extruded** fins are therefore recommended offshore. Above 150℃, **embedded** fins should be used.

4.2.3 Air-Side Control

Air-side control (generally of the heat exchanger outlet temperature) can be achieved through application of:

1. On/off fans.
2. Variable speed motors.
3. Variable pitch fans.
4. Automated louvers.
5. Bypass temperature control.

Louvers are <u>not</u> recommended; they tend to be problematic and are inefficient (i.e. form a resistance to flow even when fully open).

Most air-cooler designs include two or more fans per bay. Where this is the case, provision of a single variable pitch or variable speed fan per bay (with the other simply on/off) provides reasonable controllability.

Note that in many applications temperature control may not be required on an air-cooler. The Condensate Rundown Cooler (E-6107) installed at East Spar, for example, was fitted with constant speed (on/off) fans only. In this case, no penalty existed for cooling of the stabilised condensate below the design outlet temperature as the minimum ambient temperature on Varanus Island is relatively high. This is typical of liquid product rundown coolers.

As a general guideline, with all fans turned off natural convection effects will result cooling equal to 20-30% of the design duty [Ref 8].

4.2.4 Temperature Approach

Air-cooled heat exchangers can generally be designed to achieve cold end approaches of 10 - 12℃. Consideration may be given to specifying a slightly tighter approach (i.e. 8 – 10℃) where this would negate the use of a trim cooler / cooling water system.

4.2.5 Tube Length

Air-cooler designs which employ longer, rather than wider, bundles generally result in less expensive heat exchangers [Ref 9]. Although the equipment supplier will be given freedom to optimise the air-cooler design, it is not uncommon for projects to provide an absolute limit on the length of the heat exchanger. This may be due to shipping limitations or installation constraints (i.e. air-cooler is to be mounted on a pipe-rack of given width).

Given that less expensive designs will result where longer tubes are employed, care should be taken not to unduly limit the bundle/bay length.

Tubes are manufactured in standard lengths; 24 ft (7.3 meters) is a common, transportable length. The following table demonstrates variation in the cost of equivalent air-cooler designs relative to 24 ft tubes [Ref 9].

The data is indicative only, but the trend is clear.

Table 4.2 – Air-Cooler Relative Cost versus Tube Length

Tube length, feet (meters)	Relative Cost
10 (3.0)	1.15
14 (4.3)	1.11
18 (5.5)	1.06
24 (7.3)	1.00
30 (9.1)	0.95
36 (11.0)	0.89
40 (12.2)	0.85

4.2.6 Margins

All general service air-cooled heat exchangers shall be furnished with a 10% excess surface area margin. This margin is to apply in the fouled (service) condition and should be furnished as additional rather than longer tubes.

Condensers are a special application of air-cooled heat exchangers. Appropriate margins are described in Section 4.2.8.

4.2.7 Fouling Factors

For internal factors, refer to Section 4.1.6 (as applied to shell-and-tube heat exchangers).

A fouling factor is not routinely applied to the outer (finned) surface. Generally speaking, it is incumbent on the operations staff to maintain a clean exterior.

As noted in Section 4.2.2.2, tension-wrapped fins should not be employed offshore to avoid salt build-up between the tube and the extended surface (which cannot be readily removed). This would most definitely increase the air-side resistance. Extruded or embedded fins should be considered instead, as appropriate.

☞ Where flying insects are a concern, provision of a removable (cleanable) bug screen is usually an adequate measure. Where applicable, this will generally be outlined in the relevant client specification.

4.2.8 Condensers

The piping around column overhead condensers (and other air-coolers in condensing service) should be arranged for downward process flow (i.e. top entrance nozzles, bottom outlet nozzles). Downward flow should greatly reduce the likelihood of slugging within the tubes and header boxes.

Apply to both partial and total condensers.

Where a flooded condenser is used for process control (i.e. BassGas onshore depropaniser and debutaniser columns), a single tube pass inclined on a 1:50 grade should be considered.

Condensers in distillation service shall be furnished with 20% excess surface area by considering 120% of the controlling case flows in thermal and hydraulic design of the heat exchanger. This margin is to apply in the fouled (service) condition and should provide adequate flexibility for column operation.

4.2.9 Debottlenecking and Brownfield Projects

Air-cooled heat exchangers tend to consume relatively large plot areas. For debottlenecking and brownfield projects where space is tight, the following points should be noted:

- The capacity of existing air-cooled heat exchangers can often be increased by replacing the fans and/or motor drivers to increase air flow. This concept was successfully applied to overhead condensers in the Wesfarmers Fractionation Debottlenecking Project (circa 2001).
- Several vendors claim that higher air volumes and therefore better performance can be achieved by using fibre reinforced plastic (FRP) fans.
- For relatively small duties, it is often possible to combine several tube bundles (heat exchangers) in a single bay with common fans. This concept saves on both weight and space. However, the requirement for temperature control on one or both streams may render such a design impractical.

4.2.10 Evaluation and Preliminary Sizing

Air cooler evaluation (rating) and preliminary sizing (if required) should be performed using Aspentech ACOL.

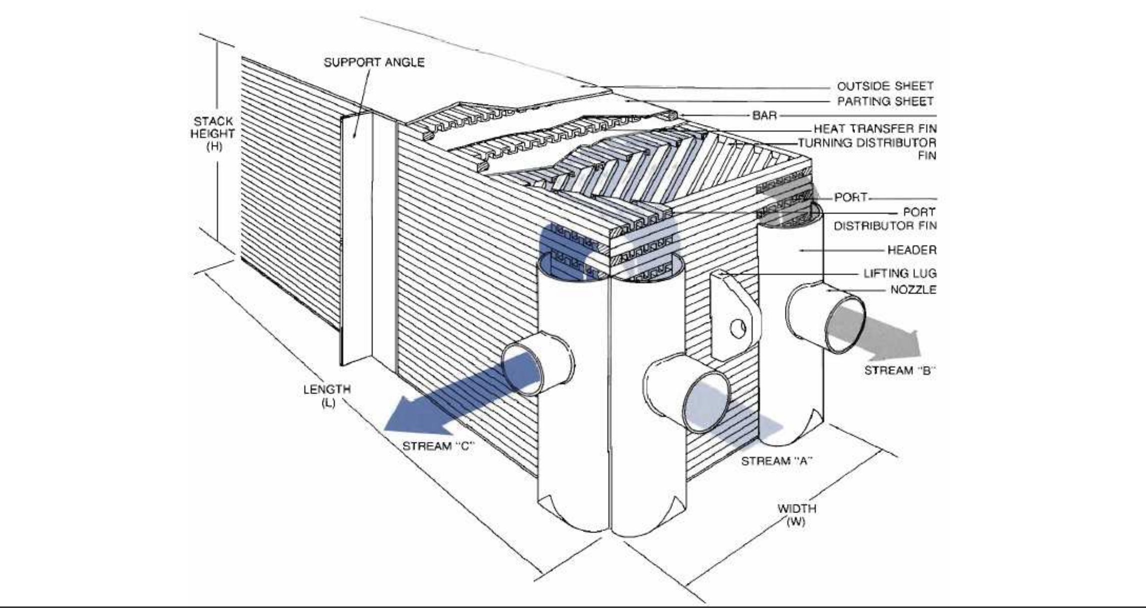
4.3 ALUMINIUM PLATE-FIN HEAT EXCHANGERS

4.3.1 Application

Brazed aluminium plate-fin heat exchangers are frequently used in low temperature gas processing services. Often referred to as a *cold-box*, they can be designed to handle upwards of six (6) streams in a single exchanger and can operate at temperatures below -200℃.

The brazed core is composed of alternating layers of corrugated fins and flat *parting* sheets. The construction is shown diagrammatically in Figure 4.3.

Figure 4.3 – Brazed Aluminium Plate-Fin Heat Exchanger



4.3.2 Margins

Plate-fin heat exchangers shall be furnished with a 10% excess surface area margin. This margin is to apply in the fouled (service) condition.

Where a plate-fin heat exchanger operates below ambient, an allowance for heat in-leak may be justified. Guidance is not provided here as plate-fin configurations can vary greatly. The lead process engineer should be consulted.

By way of example, the BassGas Development included a 5 kW allowance for heat in-leak where plate-fin heat exchangers operated below 0℃. This allowance was made on the process simulation directly (Hysys). A note was added to the process datasheet to explain why the hot stream duty exceeded the sum of the cold stream loads by this amount.

4.3.3 Fouling Factors

As permanently brazed units, aluminium plate-fin heat exchangers are only suitable for clean services. A fouling factor of 0.00018 m^2K/W should therefore be applied to all passes.

4.3.4 Temperature Approaches

Temperature approaches of 1.5℃ can be achieved in single phase applications. Slightly wider minimum approaches are applicable in two-phase service; typically 2.8℃.

Corrected mean temperature differences as low as 3 - 6℃ are common.

4.3.5 Differential Pressures

☞ The process engineer must give careful consideration to differential pressure allowances for plate-fin heat exchangers. For warm passes in auto-refrigeration service, cooling across the heat exchanger will often be supplemented by expansion cooling across a downstream JT valve. In this particular case, the supplier may be offered a relatively high differential allowance (say 1 – 2 bar). The differential will result in adiabatic cooling within the pass, in addition to cooling through heat exchange. The aggregate cooling will be the same.

Where surplus pressure is not available, care must be taken not to be unduly onerous. As much of the pressure loss across plate-fin heat exchangers is experienced through the nozzles and header boxes, a low allowance may require a significant number of oversized connections. The Gas/Gas Heat Exchanger on Lakshmi, for example, included a raw gas (cold) pass upstream of the inlet compression system. With an allowance of less than 50 kPa for 150 MMscfd at a relatively low operating pressure (circa 25 barg), this pass required 6 x 200 mm NB inlet nozzles and 4 x 200 mm NB outlet nozzles. The associated header arrangement, pipework and support structure was complicated and expensive.

In many applications, it may not be possible to alleviate this problem; the process engineer should however be aware that a relatively low pressure loss allowance is likely to result in a large number of piping connections.

4.3.6 Limitations

4.3.6.1 Design Temperatures

Plate-fin heat exchangers are generally limited to operation below +100°C. Aluminium retains good mechanical strength down to -269°C.

4.3.6.2 Temperature Differences between Adjacent Streams

Due to the nature of aluminium plate-fin heat exchanger construction, where all components are metallurgically bonded, the temperature difference between streams must be limited to prevent excessive internal stresses.

It is generally accepted that the maximum permissible difference between streams is 50°C [Ref 10].

However, in more severe cases such as two-phase flows, transient and/or cyclic conditions, this temperature difference should be lower; typically 20-30°C [Ref 10].

The equipment supplier should be consulted where an application approaches these limits.

4.3.6.3 Mercury Attack

Liquid mercury is extremely corrosive to aluminium alloys. Where plate-fin heat exchangers operate below -39°C, gaseous mercury will deposit[2] on the aluminium walls as a solid. The solid form of mercury is not corrosive to aluminium. However, when the heat exchanger warms up during maintenance activities the mercury deposits will melt and pool at low points within the core. It is then possible for aluminium and magnesium (particularly from the heat affected zones near welds) to dissolve in the liquid mercury.

Where mercury is present in the produced fluids it should be removed using a non-regenerable guard bed at the front-end of the plant.

Plate-fin heat exchangers should also be designed as ‘free draining’, to prevent liquid mercury pooling. This applies to both the core and headers, and should be communicated on the process datasheet.

Where the use of an aluminium heat exchanger is proposed and the presence of mercury is not addressed in the client’s design basis, confirmation should be sought. This should be initiated by the process engineer.

[2] Deposition in this context is the reverse of sublimation.

☞ Note that elemental mercury can be found in both hydrocarbon vapour and liquid phases. The process engineer should therefore consider hydrocarbon liquid carryover as a possible source of mercury.

4.3.6.4 Corrosion in Wet Gas Passes / Free-Draining Design

H_2S and CO_2 are not corrosive to aluminium where the water dew-point is lower than the cold-end temperature. This is the case in the vast majority of applications, where the gas streams are dehydrated.

The Lakshmi project however incorporated a pass in which untreated gas was used to provide additional cooling of the sales gas stream (traditional gas/gas/liquid heat exchanger + one). The raw gas stream entered at its water dew-point but warmed through the core; this was deemed to be acceptable.

However, in such an application, it would be possible for water to condense in the wet pass during shutdown or blowdown (given the intimate heat transfer between streams). Wet gas passes should therefore be designed as **free-draining**; the equipment supplier can usually accommodate this requirement.

4.3.6.5 Fire

Because of its relatively low melting point, aluminium is more susceptible to fire damage than steel equipment, piping and structures. For this reason many clients are reticent to use plate-fin heat exchangers in offshore applications (where the density of equipment and piping is high, egress more difficult and the consequences of equipment failure are dire).

Other types of compact heat exchanger (i.e. stainless steel printed-circuit heat exchangers) should be considered for offshore applications.

Note that the onshore plate-fin exchanger installed as part of the Lakshmi project was fitted with manual fire water deluge facilities. These consisted of a remote manual isolation valve and a DN 50 connection to the fire water ring main which swaged down to two DN 25 piping loops (with nozzles) that circled the core. However this provision was not made on earlier or even more recent construction projects (including BassGas) which underwent full HAZID analysis and HAZOP review as part of the design process.

4.3.7 Start-Up

The rate of cool-down should be limited to approximately 1°C/minute to limit internal stresses as the core contracts [Ref 11].

4.3.8 Strainers

To prevent particulate matter entering the heat exchanger, in-line strainers should be installed at the inlet to each pass. The distance between the strainers and the inlet flange is to be minimised.

☞ Cone strainers are typically installed in this service. As a minimum, the mesh should be capable of removing particles larger than 0.18 mm diameter (80 mesh Tyler standard; Ref 12). The strainer open area should be 200% of the pipe cross-sectional area as a minimum.

Where the vapour stream may routinely contain particulates, filter vessels should be installed. Cone strainers should not be used with regular removal in mind.

Provision of pressure gauge tappings upstream and downstream of the strainers may be a sensible design feature (to enable checking the strainer differential using portable gauges).

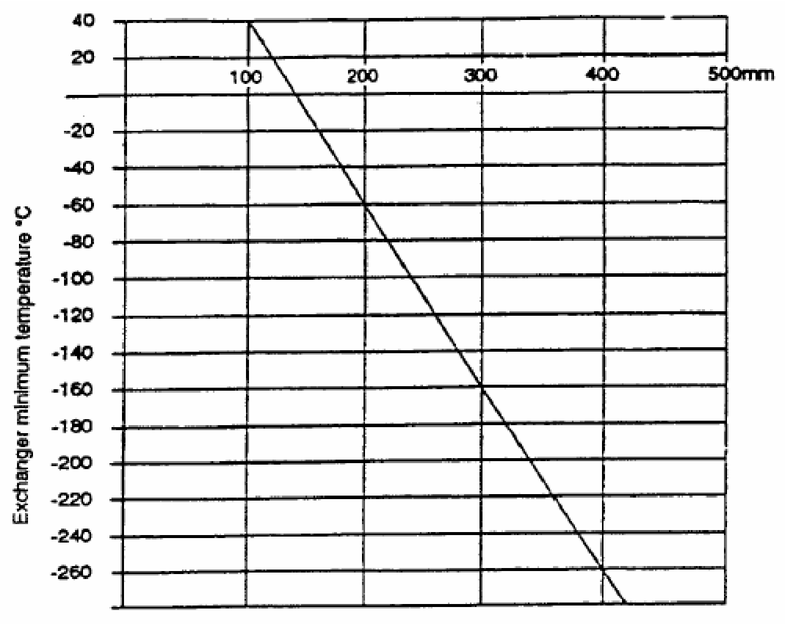
4.3.9 Insulation

ALPEMA [Ref 13] recommend the following insulation thicknesses for aluminium plate-fin heat exchangers. The chart given overleaf applies to both:

- *Cold box* applications, where it represents the thickness of insulation between the exchanger and the steel box wall; and

- *Stanchion* (*pedestal*) or *frame mounted* heat exchangers, where it represents the thickness of spray-on insulation.

Figure 4.4 – Recommended Plate-fin Insulation Thickness



Expanded perlite or rockwool is typically used to fill the void space in cold boxes. For perlite, a density in the range of 60 to 80 kg/m^3 is normally used. When packing with rockwool care must be taken to avoid damage to the heat exchanger connections. Prior to start-up, a continuous dry oxygen-free nitrogen purge is to be connected to the cold box.

For stanchion or frame mounted units, spray-on polyurethane foam is typically used. After application of the insulant, the heat exchanger must be sealed with a waterproof coating (jacket).

4.3.10 Run-Back Condensers (Dephlegmators)

Run-back condensers are a special application of plate-fin heat exchangers where the unit is mounted directly above a distillation column, usually bolted to the overhead (vapour outlet) flange. By mounting the exchanger directly above the column, condensed liquids run back to reflux the distillation process directly without the need for an intermediate reflux drum and pumps. Where applicable, this can provide a cost-effective means of refluxing.

Clough employed plate-fin heat exchangers in De-ethaniser Condenser services on both the Wesfarmers Upgrade and BassGas Development projects. In both cases cooling was achieved against colder process fluids from within the LPG extraction process.

The following should be considered in the specification and design of run-back condensers:

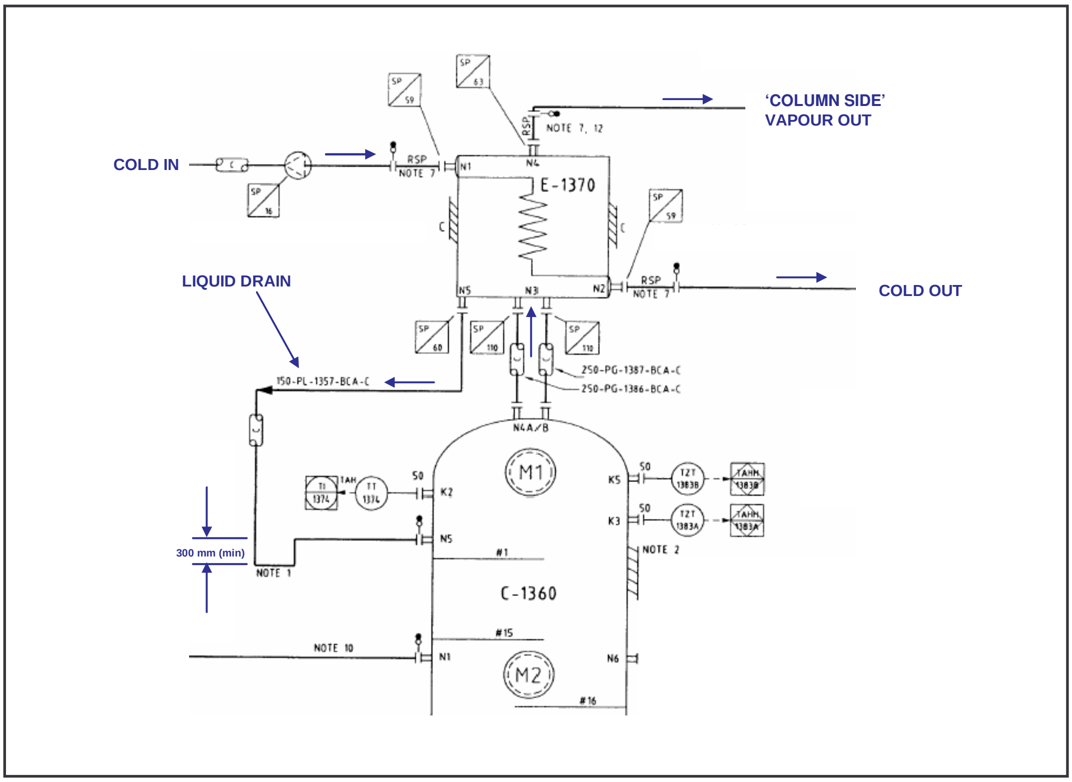
- Run-back condensers are only suitable for use in partial condensing services.
- The run-back service should be clearly indicated on the process datasheet, as the unit will require specialised design features. A brief description of the application should be provided.
- In addition to the margin applied to the heat transfer surface area, the process datasheet should call for a 10% margin on the ‘column side’ cross-sectional area (to allow free drainage and prevent re-entrainment of condensed liquids).
- Mass transfer will occur within the ‘column side’ of the run-back condenser, as the condensed liquid contacts vapour flowing upwards through the unit. Therefore – strictly speaking – the ‘column side’ of the run-back condenser represents more than one equilibrium stage. Credit should <u>not</u> be taken for this mass transfer in design of the associated column to reduce the number of installed trays. The plate-fin supplier can however be asked to provide an indication (not a guarantee) of the expected mass transfer (i.e. number of equilibrium stages). This may be useful when reviewing the actual system performance.

- The number and size of 'column side' inlet nozzles will be determined by the equipment supplier. The diameter of the associated column should be communicated on the heat exchanger datasheet; this will enable the plate-fin manufacturer to develop a suitable support nozzle template.
- The heat exchanger should be provided with nozzle(s) for condensed liquid drainage from the bottom of the heat exchanger, which are separate from the 'column side' vapour inlet nozzles. The process datasheet should clearly indicate that the exchanger design should allow condensed liquids to drain through this nozzle (or nozzles) and back to the heat exchanger under gravity flow.

  A piping seal leg should be provided in the drain line. The height of the seal should be such that the associated liquid head prevents vapour passing upwards through the drain line. A minimum leg of 300 mm should be adequate; however this height is to be reviewed based on system hydraulics.

  A typical run-back configuration, including the seal leg, is shown below.

Figure 4.5 – Run-Back Condenser



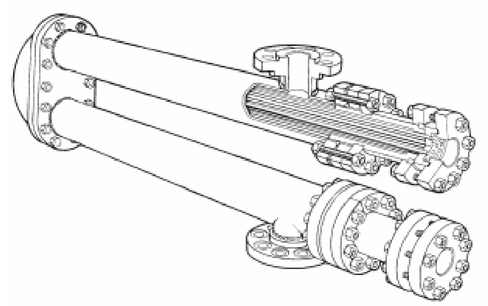
4.4 OTHER HEAT EXCHANGERS

4.4.1 Brown Fintube

Plenty Uniquip manufacture a range of finned-tube heat exchangers (hair-pin and bundle types) which may be competitive with shell-and-tube heat exchangers where surface area requirements are relatively small. Brown Fintube heat exchangers should be considered where:

- The required surface area is less than 10-15 m^2.
- The service is clean, especially the shell-side.

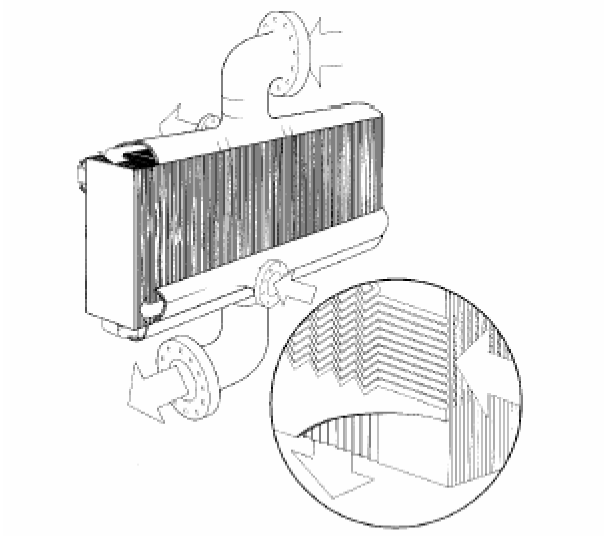
Figure 4.6 – Brown Fintube (Multitube) Heat Exchanger



4.4.2 Printed Circuit Heat Exchangers

Printed circuit heat exchangers are another form of compact exchanger, similar in design to the aluminium plate-fin units discussed in Section 4.3. However, they are fabricated from a series of chemically-etched stainless steel parting sheets, with 2.0 mm (approximate) semi-circular passages. The plates are diffusion bonded and are impossible to mechanically clean. Therefore, printed circuit heat exchangers should not be considered in fouling services.

Figure 4.7 – Printed Circuit Heat Exchanger



Although considerably more expensive than an equivalent plate-fin heat exchanger, fire safety concerns linked to the relatively low melting point of aluminium makes the use of plate-fin heat exchangers unpopular in offshore applications (refer to Section 4.3.6.5). PCHEs have therefore found a niche in offshore processing where:

- The use of a compact multi-pass heat exchanger offers layout advantages over several shell-and-tubes; and particularly when
- Stainless steel would have been employed for construction of the shell-and-tube exchanger(s) anyway, i.e. to mitigate corrosion.

The small etched passages promote good mixing within the heat exchanger. Direct glycol injection can therefore be affected upstream of the unit - for hydrate suppression - with confidence that the inhibitor will be distributed and intimately mixed with the process stream. Dedicated nozzles can be provided for inhibitor injection. This concept was employed on the Kakap Gas Compression Project.

PCHEs are not suitable for use as run-back condensers, as the small diameter passages do not permit condensed liquids to drain back to the associated column against the overhead vapour flow.

It is noted that the PCHE vendor, Heatric, does not apply fouling factors in unit design. Instead a percentage over-surface for fouling is allowed, typically 10%. Fouling will be low for PCHEs as these are only suitable for clean services.

4.4.3 Plate Frame Heat Exchangers

Plate frame heat exchangers consist of an arrangement of gasketed pressed metal plates (heat transfer surface), aligned on two carrying bars, secured between two covers by compression bolts.

PHEs offer the following advantages:

- Plot area requirements are small compared to shell-and-tube heat exchangers;
- Plates can be added or removed should service conditions change;
- The units can be easily disassembled for cleaning; and
- Relatively high velocities on both the hot and cold sides lead to low deposition and fouling.

However, relatively low upper design temperature and pressure limits exist due to the nature of construction. The design pressure of plate frame heat exchangers is generally limited to 20 barg. The upper design temperature is generally a function of gasket selection. Typical gasket material temperature limits are presented below [Ref 16]:

Table 4.4 – Plate Heat Exchanger Gasket Limits (Typical)

Gasket Material	Temperature Limit
Natural rubber, styrene, neoprene	70°C
Nitrile	100°C
Resin-cured butyl, viton	150°C
Ethylene/propylene, silicone	150°C
Compressed non-asbestos fibre	200°C

The application of fouling factors approximately 50-times lower than those presented in Section 4.1.6 for shell-and-tube heat exchangers can generally be justified. The following fouling factors have been taken from the GPSA Engineering Data Book [Ref 16] for comparison with those given for conventional heat exchangers:

Table 4.5 – Plate Heat Exchanger Fouling Allowances

Service	PHE Fouling Factors ($m^2.°C/W$)
Process fluids, general	0.000007
Steam	0.000002
Cooling water (treated)	0.000002 – 0.00001

Excess surface area is generally not included as additional plates can be added. It should be noted that the low fouling factors presented above require that minimum velocities are maintained to limit deposition.

5 SEPARATION EQUIPMENT

5.1 VESSEL ORIENTATION

For gas/liquid separation, a vertical vessel should normally be utilised where:

- A small footprint is desirable;
- The ratio of vapour to liquid is high;
- A consistently high liquid removal efficiency is required; and
- A large liquid surge volume or slugging allowance is not required.

It is noted that the liquid removal efficiency of vertical separators is not affected by the liquid level, as the cross-sectional area for vapour flow (circular) remains constant. When a high efficiency internal device is employed, this attribute results in *consistently* high removal efficiencies. This accounts for their frequent application as compressor suction scrubbers and low-temperature separators (in hydrocarbon dew-point control) where some – usually intermittent – variability in liquid levels will be experienced, but minimisation of liquid carry-over remains critical.

Horizontal vessels should be considered where:

- Large liquid slugs must be accommodated (i.e. vessel-style inlet slugcatchers);
- The ratio of vapour to liquid is low;
- The separation of two immiscible liquid phases is required;
- Headroom is limited (i.e. some inter-deck offshore applications); and
- A low downward velocity of liquid is desirable (i.e. for degassing or foam breakdown).

In some applications, the choice between a vertical or horizontal vessel is not clear cut. With the exception of vessels requiring high liquid residence times (or interfacial areas) for three-phase separation or where large surge or slugging volumes are required, the use of either orientation may be possible.

Vertical separators are rarely fabricated with aspect ratios (tan-tan length divided by internal diameter) considerably in excess of 3.5:1. As a general guide, where there is uncertainty, assume vertical orientation and estimate the vapour-controlled internal diameter, then the tan-tan height using the guidelines given in the subsequent sections. If the aspect ratio is appreciably greater than 3.5:1, increase the vessel diameter by one or two standard head sizes and recalculate the tan-tan height for each. If the aspect ratio is still unusually high then sizing of the vessel is liquid-controlled and a horizontal vessel should be considered.

5.2 GENERAL

5.2.1 Margins

A 10% margin should be applied to both the vapour and liquid flow for vessel design purposes, unless a ‘global’ design margin has been previously added or otherwise included.

5.2.2 Nozzle Sizing

The following momentum-based criteria should be employed [Ref 27].

For inlet nozzles:

$$v_i < \frac{60}{\sqrt{\rho_M}} \qquad \text{(metric)}$$

For vapour outlet nozzles:

$$v_i < \frac{75}{\sqrt{\rho_M}} \qquad \text{(metric)}$$

For liquid outlet nozzles:

$$v_i < 1 \text{ m/s}$$

☞ Process nozzles will in almost all applications carry the same ANSI flange rating as the connected piping. Past experience has shown however that where liquid slugging is anticipated, a higher rating may be required to withstand the associated forces.

Instrument connections shall be made using DN 50 nozzles as a minimum.

5.2.3 Manways

Manways shall be DN 500 as a minimum. The requirements (specifications) of particular clients should however be reviewed before finalising manway sizes. Where internals are to be installed through a manway, consideration should be given to installing a larger manway (DN 550 or DN 600) after consulting the internals supplier and the lead mechanical engineer.

☞ Care should be taken when making allowances for the installation of a manway in the upper section of a column (i.e. between the top tray or top of packing and the upper tan-line, adjacent to the liquid distributor and/or meshpad). Past experience has shown that the vertical span required to accommodate the manway whilst maintaining practical clearances from the internal devices can be easily underestimated. Where this level of definition is required, particularly for EPC work, a workable arrangement should be detailed early in the project in consultation with the mechanical group and the internals supplier.

5.3 VERTICAL SEPARATORS

5.3.1 Internals

Vapour-liquid separation in vertical separators can be improved by installing a high-efficiency internal device. The internals listed below are those most commonly employed on recent Clough projects, and are given in order of increasing expense:

- Wire mesh demisters or meshpads;
- Vane packs; and
- Cyclone-type devices (i.e. swirl tubes or swirl decks).

Where sizing of a vertical vessel is vapour-controlled, installation of a suitable vane pack will require a smaller vessel diameter than for an equivalent mesh pad. The smaller vessel will be lighter and less expensive. Therefore, in high-pressure services (and particularly where exotic materials are required for vessel construction), the additional cost of a vane pack is often comfortably offset by the reduced cost of the associated vessel.

☞ Cyclone-type devices, such as Shell's Swirldeck (which is often combined with a meshpad or vane pack above the deck and a proprietary vane-type inlet device below), come at significant expense. However, this style of high-efficiency device may still offer a cost advantage where a single unit replaces two filter coalescers or equal. For example, the Sawan Field Development included 2 x 100% filter coalescers upstream of the amine contactor in each train. A standby filter was required due to the potential for element plugging. The coalescers were of the dual compartment style; each required 4 x level bridles and transmitters, and 2 x level control valve/SDV sets. It was noted retrospectively that provision of a single swirl deck – generally regarded as resistant to particulates - would have greatly simplified the piping, isolation, and instrumentation arrangements. The total installed cost of the system is likely to have been lower also.

The use of a swirl deck may also be appropriate where normal liquid carryover rates give an unacceptably high loss of consumable (i.e. recovery of direct-injected glycol from a cold vapour stream). It is noted, however, that Clough can provide several project references in which vane packs were successfully used to limit glycol losses for direct injection systems (both TEG and MEG).

Therefore, for most high-pressure applications less than 100 barg, a vane pack internal device will provide the most economic *complete* separator design. Where we are not compelled to provide anything more complicated, vane packs should be our first choice. However, at operating pressures above 100 barg, the efficiency of vane packs generally decline.

Mesh pads should be considered in:

- Low-pressure applications, where an increase in the vessel diameter does not greatly affect the cost of the vessel.
- Services in which diameter of the vessel is not governed by the selected internal or is not vapour controlled; or
- Applications where the diameter of the vessel is governed by another internal (i.e. mesh pad mounted above structured packing in a glycol contactor, to limit mechanical losses).

Where possible, the use of mesh pads should be avoided in reciprocating compressor suction scrubbers. The pulsing service may lead to deterioration of the pad and wire migration into the compressor causing damage.

5.3.2 Internal Diameter

5.3.2.1 General

With the exception of vessels sized for droplet separation, many of the guidelines given in the following sections should be treated as preliminary only (for estimation). In almost all applications, the internal diameter of the vessel will be confirmed as adequate by the internals vendor for compliance with warranties related to liquid carryover.

5.3.2.2 Meshpad Demisters

The internal diameter of vertical vessels fitted with a meshpad demister can be estimated using the following equation [Ref 17]:

$$v_t = K\left(\frac{\rho_L - \rho_G}{\rho_G}\right)^{0.5}$$

Where v_t is the allowable velocity through the demister. The K-factor is pressure dependant; the following values should be employed:

Table 5.1 – Factor K for Mesh Pad Sizing

Operating Pressure	**K Factor (m/s)**
Atmospheric	0.110
2000 kPag (300 psig)	0.101
4100 kPag (600 psig)	0.091
6200 kPag (900 psig)	0.082
10,300 kPag (1500 psig)	0.064

For compressor suction scrubbers, turbo-expander inlet scrubbers, for liquids with a high foaming tendency (i.e. amine) or where the liquid is a consumable (i.e. TEG, amine) the K-factors given above should be multiplied by 0.7.

The minimum meshpad area is then estimated by dividing actual volumetric flow of vapour by allowable velocity (v_t).

The available mesh pad area will be reduced by the support ring and, where required, support beams. The mesh pad (vessel) diameter must be increased to allow for the loss of area due to the support hardware. The required incremental increase to the mesh pad diameter is described in Table 5.2.

Table 5.2 – Support Hardware Allowances for Mesh Pad Installations

Computed Diameter (mm)	Support Ring Width (mm)	Support Beam		Increase in Diameter (mm)
		Number	Width (mm)	
< 450	32	0	-	64
450 – 1800	38	0	-	76
1800 – 3600	51	1	100	178
3600 - 5400	76	2	100	280

After adjustment for accommodation of support hardware, the next larger standard semi-elliptical head size should be selected. A list of standard sizes is given in Attachment A.

5.3.2.3 Vane Packs

The internal diameter of vertical vessels fitted with a vane pack demister can be estimated using the following equation [Ref 18]:

$$v_t = (0.12)\left(\frac{\rho_L - \rho_G}{\rho_G}\right)^{0.5}$$

Where v_t is the allowable velocity upwards through the vessel cross-section (circular).

For compressor suction scrubbers, turbo-expander inlet scrubbers, for liquids with a high foaming tendency (i.e. amine) or where the liquid is a consumable (i.e. TEG) the K-factor (0.12) should be multiplied by 0.7.

The minimum vessel CSA is then estimated by dividing actual volumetric flow of vapour by allowable velocity (v_t).

The corresponding diameter is the minimum requirement. The next larger standard semi-elliptical head size should then be selected. A list of standard sizes is given in Attachment A.

An estimate of the vane pack face area can be made using the following equation:

$$\rho_G . u^2 = 29.8 \text{ kg/ms}^2$$

Where u is the average vapour velocity through the vane pack face area (rectangular). The face area can then be estimated by dividing the actual vapour volumetric flow by velocity u.

An estimate of suitable vane pack dimensions can then be made by using the following criteria:

(1) Vane box aspect ratios (height / width) of 1.2 are typical. Calculate the dimensions based on this aspect ratio and then perform a sense check using criteria (2) and (3).

(2) Vane pack heights are typically in the range of 300 – 1500 mm.

(3) Vane box width should not exceed vessel ID minus 200 mm (i.e. minimum 100 mm from the vessel wall on each side).

5.3.2.4 Droplet Separation

The following equations can be used to set the vessel diameter where:

- A demisting device is <u>not</u> fitted;
- A device is fitted to the vessel but client specifications dictate that the diameter be set to ensure that all droplets larger than a critical size be separated under gravity in the vessel mid-section; or

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- A device is fitted but the lead process engineer elects to size the vessel to ensure that all droplets larger than a critical size be separated under gravity.

Alternatively, these equations can be used to calculate the separable droplet size for a vessel of given diameter. This is often a useful sense check of vendor's diameter or where the vessel size has been estimated using the equations given in Sections 5.3.2.2 and 5.3.2.3.

As a broad guide, where an internal device is not employed the vessel diameter should allow for separation of all liquid droplets 100-130 micron and larger. It is noted, however, that for the type of work executed by Clough the installation of a vessel without a demisting internal is considered unusual. This should only be contemplated where the vapour rate is extremely low and the vertical vessel size is patently liquid-controlled. Services fitting this description are, in fact, likely to achieve gravity separation of droplets much smaller than the 100-130 micron range.

When a demisting internal is installed, the bulk separation of 350 – 500 micron droplets upstream of the device should provide additional confidence in the overall separator performance.

The following equations are based on methodology outlined in API RP 521 [Ref 19]. The quadratic equation (determined by curve-fit) replaces the chart given in the API standard for determining the drag co-efficient (C) from the product CRe^2.

$$CRe^2 = \frac{\left(0.13x10^8\right)\rho_G D_P^3 \left(\rho_L - \rho_G\right)}{\mu^2}$$

Where D_P is the critical droplet size (diameter) in meters. The drag co-efficient (C) can then be estimated using the following relations:

$$a = \log\left(CRe^2\right)$$

$$\log(C) = (0.0843)a^2 - (1.007)a + 2.6718$$

The maximum vapour velocity (v_t) upwards through the vessel is then determined using the following relation:

$$v_t = 1.15\left[\frac{gD_p\left(\rho_L - \rho_G\right)}{\rho_G C}\right]^{0.5}$$

The minimum vessel CSA is then estimated by dividing actual volumetric flow of vapour by allowable velocity (v_t).

The corresponding diameter is the minimum requirement. The next larger standard semi-elliptical head size should then be selected. A list of standard sizes is given in Attachment A.

5.3.3 Level Co-ordination

The guidelines given overleaf are suitable for most general purpose applications, where the vessel operates under automated level control and reasonably prompt action from competent, alert operations staff can be assumed between the pre-alarm and trip levels.

Where a range of liquid rates are expected, residence time calculations should generally be based on the upper limit of the 'normal' design range. For example, a number of mass balance cases may exist – all at the facility design rate - for variations in feed composition, ambient temperatures, field life chronology, production mode, etc. Level settings based on the highest of these should provide an operable design.[3]

[3] It is noted that the level control valve may actually be sized for a higher rate to prevent flooding of the vessel in some upset scenario (i.e. excessive cooling and liquid condensation upstream of the vessel due to failure of temperature controller).

Figure 5.3 – Level Co-ordination Guidelines for Vertical Two-Phase Separators

LAHH

(high level trip)

The largest of the following criteria should be applied:

- The vertical span between LAH and LAHH should be a minimum of 100 mm.
- The span should allow a reasonable residence time for operator intervention as the pre-alarm is generated and the level continues to rise:

  (a) Where operations staff can take action in the control room to limit the liquid inflow without resulting in an immediate loss of gas production, a **30 - 60 second** allowance should provide an operable design. For example, partial closure of the control valve on the outlet of a slugcatcher to limit liquid level rise in a downstream flash vessel.

  (b) Where normal liquid production is not expected, a **30 - 60 second** allowance at the selected design rate should be applied.

  (c) Where normal liquid production is expected in a service critical to gas production and operations staff have little control over the rate of liquid inflow then a **2 – 3 minute** allowance should be provided for an operator to intervene in the field. A means of manual level control should generally be provided where this allowance has been deemed appropriate (i.e. hand-wheel, control valve bypass).

LAH

(high level pre-alarm)

The largest of the following criteria should be applied:

- The vertical span between LAL and LAH should be a minimum of 300 mm.
- For most separators operating under level control only, an allowance for **1 - 2 minutes** residence time between LAL and LAH should be adequate for controllability.
- For applications where all or a significant part of the liquid leaves the vessel under flow control (i.e. reflux drum) provision of **3 - 4 minutes** residence time between LAL and LAH should be adequate for controllability.

NLL

Where inlet slugging is anticipated, the expected liquid slug volume should be accommodated between the NLL and LAH. If the volume of the slug is not known, 2 – 5 seconds of flow with the maximum feed (vapour plus liquid) velocity and 100% liquid filling of the pipe can be assumed. However, attempts should be made to more rigorously determine the slug volume. Accommodation of the slugging volume between NLL and LAH with levels set using the residence time criteria given above may not be possible; the LAH should be increased in this case.

LAL

(low level pre-alarm)

The largest of the following criteria should be applied:

- The vertical span between LAL and LALL should be a minimum of 100 mm.
- The span should allow a reasonable residence time for operator intervention as the pre-alarm is generated and the level continues to fall:

  (a) Where a shutdown valve is provided or where gas blow-by from the vessel does not carry serious consequences, a **30 - 60 second** allowance should be sufficient for action from the control room. This allowance is suitable for most vapour-liquid separators where closure of the SDV will not result in a loss of gas production (i.e. compressor suction scrubber).

  (b) Where maintenance of a liquid level is critical to gas production, a **2 – 3 minute** allowance should be sufficient for an operator to intervene in the field. A means of manual level control should generally be provided where this allowance has been deemed appropriate (i.e. hand-wheel, control valve bypass).

LALL

(low level trip)

- To be located as low as reasonably practicable within the vessel.
- For vertical vessels with design pressures in excess of 2000 kPag, the centreline of the lower bridle nozzle can generally be located 150 - 200 mm from the lower tangent line (this should however be confirmed by the lead mechanical engineer). Where magnetic float-style level transmitters are employed, LALL can be set at the bridle nozzle centreline.

Lower Tan Line

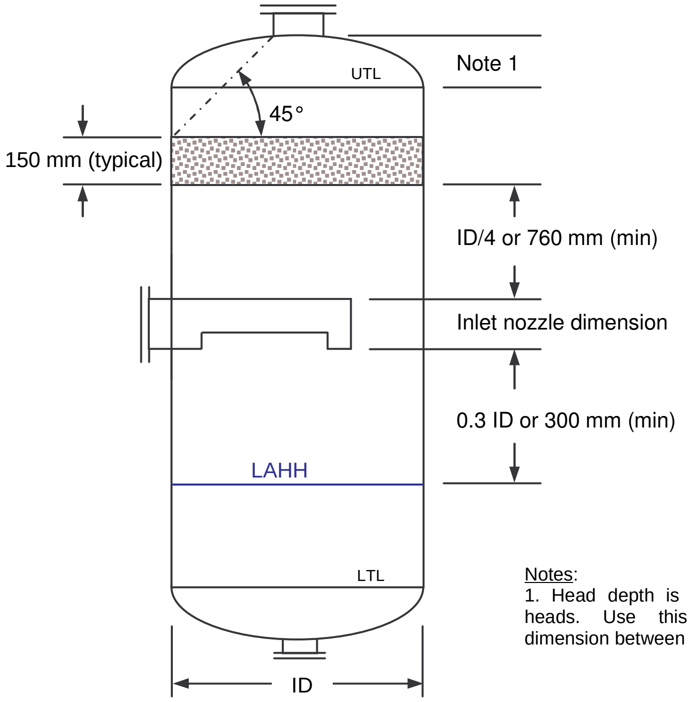
5.3.4 Vessel Height

5.3.4.1 General

The following recommendations are based on widely accepted industry guidelines and design values quoted by internals vendors on past Clough projects. These should be used for preliminary vessel sizing and study-level estimates. However, for detailed design and EPC projects, the selected internals supplier should be asked to advise or confirm that all allowances above LAHH are adequate for compliance with process guarantees.

5.3.4.2 Meshpad Demisters

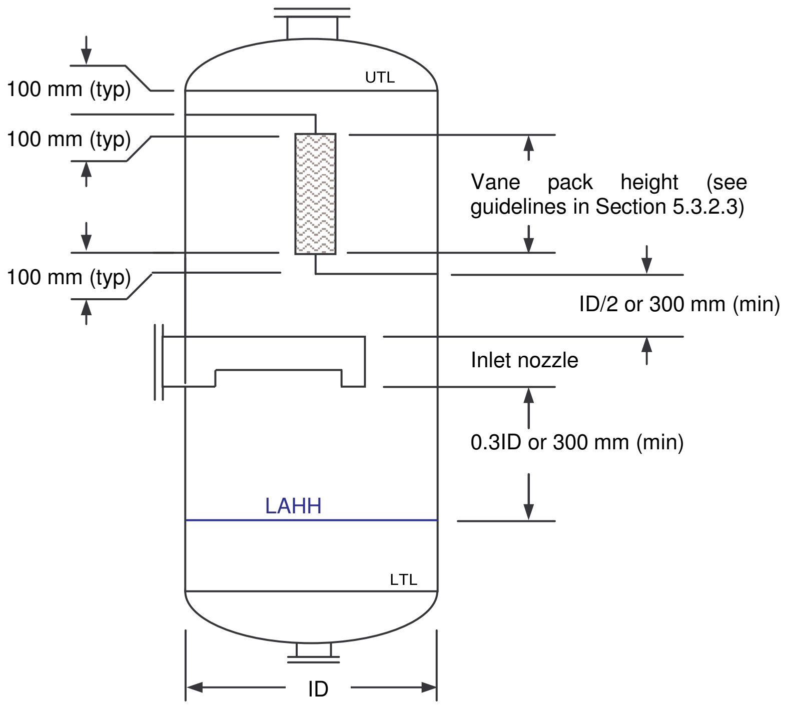
Figure 5.4



Notes:
1. Head depth is ID/4 for 2:1 semi-elliptical heads. Use this to calculate minimum dimension between mesh pad and UTL.

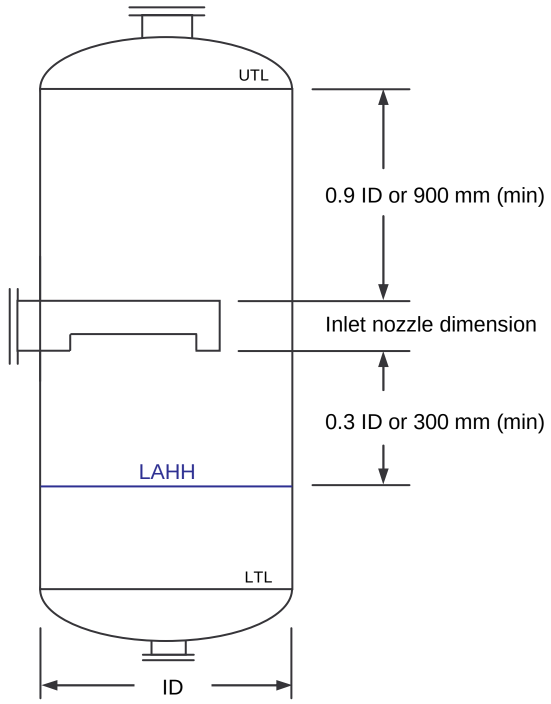
5.3.4.3 Vane Packs

Figure 5.5



5.3.4.4 Gravity Only Separators

Figure 5.6



5.3.5 Downcomers

Vane packs (and several other internal types) include a downcomer to return coalesced liquids from the vane box to the vessel sump.

The following should be noted:

- The pressure loss across the vane pack must be less than the downcomer static head, to prevent syphoning. This requirement should be easily met by the internals supplier.
- A seal leg should be included in the downcomer, typically 300 mm, to prevent vapour bypassing when the sump is empty.
- Downcomers should be DN 25 mm (minimum).
- The downcomer should extend to 50 mm below the LALL.

5.3.6 Process Guarantees

For most vapour-liquid separation services, including compressor suction scrubbers, the internals vendor should be required to guarantee the following:

1. 100% removal of liquid droplets 10 micron, and greater.
2. Total liquid carryover of less than 0.1 USGAL/MMscf.
3. Separator pressure loss of less than 7.5 kPa.

Additionally, for vane packs:

4. The vane pack pressure loss shall be less than the downcomer static head.

These specifications are typical and most internals suppliers will not view them as unduly stringent.

Prior to issuance of the internals datasheet, the process engineer should review the implications of liquid carryover at 0.1 USGAL/MMscf. In most dew-point control applications, carryover at this rate will have negligible effect on the hydrocarbon dew-point of the product vapour stream. However, it is not uncommon for clients to include an operational limit on consumable losses within their technical specifications (i.e. TEG, amine). Where this is the case, a tighter limit on the carryover rate may be required.

For example, liquid carryover from the upper mist elimination device fitted to the TEG Contactor at the Lakshmi Onshore Terminal was limited to 0.07 USGAL/MMscf due to contractual requirements. Use of a standard device with a higher total liquid carryover was not tolerable as equilibrium losses were reasonably high.

A 300 mm thick Yorkmesh Style 82 mesh pad was supplied by Koch-Otto York capable of meeting this specification.

5.3.7 Vessel Sizing

An approved and validated calculation spreadsheet (*Vertical Separator Rev A*) is available to assist the process engineer with preliminary sizing of vertical vessels. Separators fitted with meshpads, vanepacks and those reliant on gravity-only droplet separation can be sized with this tool.

The spreadsheet is consistent with recommendations made in Sections 5.3.1 to 5.3.6 of this document.

5.4 HORIZONTAL SEPARATORS

5.4.1 Normal Residence Times

5.4.1.1 Two-Phase Separators

The following minimum residence times are recommended in API 12J when setting the normal liquid level (NLL) in horizontal two-phase separators [Ref 20]. The following residence times are defined as the volume below NLL divided by the design liquid flowrate. Note that these residence time allowances are made for *separability*, not *controllability* (i.e. minimum allowances between pre-alarms and trip levels). This is an important distinction to make in vessel sizing.

Note that two-phase separators may still contain two immiscible liquid phases (i.e. hydrocarbon condensate and water). However, as no attempt is made to separate the liquid phases, relatively low residence times are acceptable at the NLL.

Table 5.3 – Normal Retention Times for Two-Phase Separators

Oil/Condensate Gravity	**Retention Time @ NLL (minutes)**
Above 35° API	1
20 – 35° API	1 – 2
10 – 20° API	2 – 4

These are given as minimum requirements. In smaller vessels, the NLL may be driven upwards to accommodate pre-alarms and trip levels (refer to Section 5.4.2).

5.4.1.2 Three-Phase Separators

The following retention times are recommended in API 12J when setting the normal liquid level and normal interface level in three-phase separators [Ref 20]. It is usual practise to allow equal residence times for the hydrocarbon and aqueous phases.

Table 5.4 – Retention Times for Various Three-Phase Separation Services

Service	Retention Time @ normal level (minutes)	Notes
Water/Hydrocarbon above 35° API	3 – 5	Note 1, 3
Water/Hydrocarbon below 35° API:		Note 1, 3
(a) and above 38°C	5 – 10	
(b) between 27 – 38°C	10 – 20	
(c) between 16 – 27°C	20 – 30	
Glycol/Hydrocarbon Separators:		
Below 24°C (cold)	20 – 60	Note 2, 3
Above 24°C	20	Note 2, 3
Amine/Hydrocarbon Separators	20 – 30	Note 3

It is difficult to predict the efficiency of separation vessels designed using the retention time guidelines of API 12J. There is no generally accepted methodology for estimating residual *oil-in-water* and *water-in-oil* in the public domain. The actual performance of a separator will vary depending on the inlet water cut, emulsifying tendencies, upstream treating and settling time.

However, some vendors provide rough estimates of water in oil separation as a function of oil (hydrocarbon) gravity where the vessel design is in line with API 12J (retention times given above). These are listed below:

API Gravity	Inlet Water	Outlet Water
> 35° API	10%	1 – 4%
20 – 35° API	10%	4 – 8%
10 – 20° API	10%	8 – 10%

The figures given above are indicative and should not be used where a process warranty is required. Generally speaking, where we are required to guarantee a low residual *water-in-oil* content (or vice versa) a high-efficiency separation device will be installed downstream of the three-phase separator, with a performance warranty provided by a specialist supplier (i.e. hydrocyclones, air floatation). The information is provided for mass balance purposes only.

The following relationships are generally true, but can be altered by individual fluid characteristics. Water-in-oil separation rates increase as the percentage of inlet water increases up to around 50%. Likewise, water and oil separate more slowly when the percentage of water decreases (below 50%). Temperature increases can also increase separation rates. These factors should be considered if the values given above are extrapolated to other conditions.

Notes for Table 5.4:

1. The recommendations of API 12J do not take account of emulsions or foaming. Where there are reasons to suspect that foaming may be a problem or where the client has indicated that stable emulsions may form, the lead process engineer should be consulted. Production separators, test separators, inlet separators and multiphase slugcatchers are all examples of vessels which may suffer poor performance where emulsions (and the like) are anticipated. For further details and cautionary notes on flow assurance issues, refer to Section 9.4.

2. For direct glycol injection systems, provision of a plate-pack (or equal) device in glycol/hydrocarbon separator is recommended. The intent of this device is to assist with glycol droplet coalescence in the hydrocarbon phase in order to limit consumable losses to the hydrocarbon product. Credit for improved phase separation should not be taken in setting of the retention time. Note that installation of a ‘trash screen’ upstream of the plate-pack may be an associated requirement.

3. The residence times given above must be fully accommodated within the *separation* section of the vessel. Any residence time provided downstream of the weir does not contribute to phase separation.

5.4.2 Level Co-ordination

For two-phase separators, pre-alarm and shutdown (trip) levels must be located either side of the normal liquid level for controllability and operability. In three-phase applications, the full set of pre-alarm and trip levels may be required for both the overall level and the liquid-liquid interface. This should however be reviewed for particular applications.

Refer to Figure 5.7 overleaf. Although the guidelines have been mapped onto a particular three-phase separator configuration, the overall (gas-liquid interface) level guidelines are applicable to horizontal two-phase separators also.

It is noted that some applications of horizontal vessels may require more than the highest general allowance for *surge time*[4] indicated overleaf (3 – 4 minutes), particularly where the drum is to provide buffering volume. The following surge volumes are provided as general guidance. Consideration must however be given to the requirements and dynamics of the particular application.

Table 5.5 – Recommended Surge Times For Specific Buffering Applications

Service	Surge Time (minutes LAL - LAH)
Process train feed drum	10 – 30
Furnace feed drum	10 – 20
Glycol surge vessel	10

5.4.3 Vortex Avoidance

As noted in Figure 5.7, consideration must be given to vortex avoidance when setting minimum levels.

Vortex breakers should be fitted to all liquid outlet nozzles. It is incumbent on the process engineer to justify not using a vortex breaker in particular applications.

The liquid outlet nozzle should be sized to limit the outlet velocity to less than 1 m/s. Where this is the case, the minimum liquid level (LALL) should be set 200 mm above the vortex breaker. If the outlet velocity is greater than 1.2 m/s, then the LALL should be set using the following equation:

$$\text{LALL (mm above top of vortex breaker)} = 200 + 250(u - 1.2)$$

Where u = liquid velocity (m/s)

4 In this context, *surge time* is equal to the volume contained between LAL and LAH divided by the volumetric feed rate.

Figure 5.7 – Level Co-ordination for Horizontal Separators

Provision of LAHH(int) is an operability issue, not a safety concern. Where appropriate, consideration may be given to replacing LAHH(int) with a control system initiated action at LAH(int). In this case, the dimension between LAH(int) and the top of the weir should be 100 mm.

Should a LAHH(int) be installed, similar allowances should be made to LAH – LAHH (see right margin).

The largest of the following criteria should be applied:

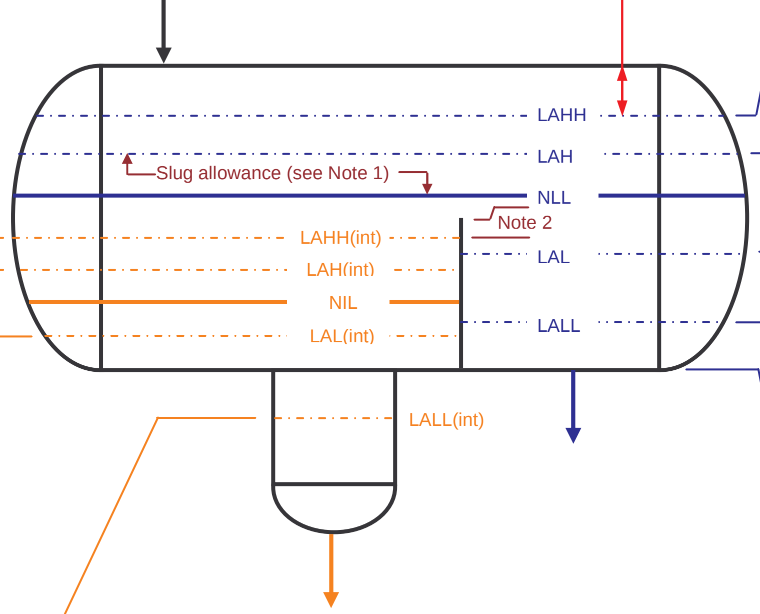
- The vertical span between LAL(int) and LAH(int) should be a minimum of 300 mm.
- For most separators operating under level control only, an allowance for **1 – 2 minutes** residence time should be adequate for controllability.
- For applications in which a significant portion of the liquid leaves under flow control provision of **3 – 4 minutes** residence time should be made.

The largest of the following criteria should be applied:

- The vertical span between LAL(int) and LALL(int) should be a minimum of 100 mm.
- The span should allow reasonable residence time for operator intervention as the pre-alarm is generated and the level continues to fall:
  - (a) Where a shutdown valve is provided or where gas blow-by (or light-phase carry-under) do not have serious consequences, a **30 - 60 second** allowance should be sufficient for action from the control room.
  - (b) Where a low-level trip will result in a loss of production, a **2 – 3 minute** allowance should be sufficient for an operator to intervene in the field. A means of manual level control should generally be provided where this allowance has been deemed appropriate (i.e. hand-wheel, control valve bypass).

Vapour space:

- Minimum 300 mm.
- Droplet knockout and accommodation of mist elimination internal (if required) to be considered also; refer to Section 5.4.6.



The largest of the following criteria should be applied:

- The vertical span between LAH and LAHH should be a minimum of 100 mm.
- The span should allow reasonable residence time for operator intervention as the pre-alarm is generated and the level continues to rise:
  - (a) Where operations staff can take action to limit the liquid inflow without resulting in an immediate loss of production, a **30 – 60 second** allowance should provide an operable design.
  - (b) Where normal light liquid production is not expected, a **30 – 60 second** allowance at the selected design rate should be applied.
  - (c) Where normal liquid production is expected in a service critical to production and operations staff have little control over the rate of liquid inflow then a **2 – 3 minute** allowance should be provided for an operator to intervene in the field. A means of manual level control should generally be provided where this allowance has been deemed appropriate (i.e. hand-wheel, control valve bypass).

The largest of the following criteria should be applied:

- The vertical span between LAL and LAH should be a minimum of 300 mm.
- For most separators operating under level control only, an allowance for **1 – 2 minutes** residence time should be adequate for controllability.
- For applications where all or a significant part of the liquid leaves under flow control provision of **3 – 4 minutes** residence time should be made.

The largest of the following criteria should be applied:

- The vertical span between LAL and LALL should be a minimum of 100 mm.
- The span should allow a reasonable residence time for operator intervention as the pre-alarm is generated and the level continues to fall:
  - (a) Where a shutdown valve is provided or where gas blow-by from the vessel does not carry serious consequences, a **30 - 60 second** allowance should be sufficient for action from the control room.
  - (b) Where maintenance of a liquid level is critical to gas production, a **2 – 3 minute** allowance should be sufficient for an operator to intervene in the field. A means of manual level control should generally be provided where this allowance has been deemed appropriate (i.e. hand-wheel, control valve bypass).

To be located as low as reasonably practicable within the vessel.

For vessels up to approx. 2.4 m ID, the centreline of the lower bridle nozzle can generally be located 200 mm from the inside bottom (this should be confirmed by the lead mechanical engineer). Where magnetic float-style level transmitters are employed the LALL can be set at the bridle nozzle centreline. However, consideration must also be given to vortex avoidance (refer to Section 5.4.3).

<u>Notes</u>:

1. Where inlet slugging is anticipated, the expected liquid surge volume should be accommodated between the NLL and LAH. If the volume of the slug is not known, 2 – 5 seconds of flow with the maximum feed (vapour plus liquid) velocity and 100% liquid filling of the pipe can be assumed. However, attempts should be made to more rigorously determine the slug volume.
2. The dimension between the highest interface level – LAH(int) or LAHH(int) - and the top of the overflow weir is to be a minimum of 100 mm (refer to Section 5.4.5).

5.4.4 Vessel Boot

The use of a boot is recommended where the heavy-phase volumetric flow is small compared with the hydrocarbon flow. In this situation, the vertical spans between co-ordinated levels [i.e. LALL(int) – LAL(int)] are likely to default to the minimum dimensions given in Section 5.4.2. Where these provide very large residence times, accommodation of the default levels within the main section of vessel is generally an inefficient use of internal diameter. By installing a boot some (or all) of the heavy-phase levels can be set below the inside bottom of the vessel, resulting in a smaller, lighter and less expensive separator. An increase to the saddle height may be required. However, this should come at modest expense.

The nominal diameter of the boot shall be a minimum of 300 mm. Standard semi-elliptical head sizes should be employed where possible (Attachment A). A minimum boot depth of 600 mm is recommended.

To satisfy mechanical and economic considerations, boot diameters should not exceed the values given in Table 5.6.

Table 5.6 – Maximum Boot Diameters

Vessel Diameter (mm)	Boot Diameter (mm)
< 1000	½ vessel diameter
1000 – 1500	500 mm
> 1500	⅓ vessel diameter

5.4.5 Weirs and Stand-Pipes

Where liquid-liquid separation is required, a weir may be used to segregate the separation section from the aft end (light liquid only).

The weir height should, as a minimum, be set 100 mm above the highest interface level, i.e. LAH(int) or LAHH(int). Refer to Figure 5.7.

Credit should not be taken for volume downstream of the weir when setting the normal interface and normal liquid levels based on the retention time guidelines for phase separation. Credit may however be taken for volume on both sides of the weir (as appropriate) when setting high pre-alarms and high shutdown levels.

Fully welded weirs are recommended, to prevent leakage of the heavy phase.

A minimum length of 1000 mm should be provided between the weir and the aft tan-line. If the weir is welded then accommodation of the liquid outlet nozzle between its heat affected zone and that of the rear weld line may be difficult if a smaller dimension is used.

The use of a stand-pipe may be considered instead of a weir. The stand-pipe height should be set at least 100 mm above the highest interface level, i.e. LAH(int) or LAHH(int).

5.4.6 Vapour Space

5.4.6.1 Minimum Dimension

The minimum vapour space (LAHH to inside top of vessel) is set at 300 mm for general service applications. Consideration must however be given to droplet setting criteria (Section 5.4.6.2), installation of an inlet device (Section 5.4.6.4) and accommodation of mist elimination internals, where required (Section 5.4.6.3).

5.4.6.2 Droplet Settling

For general service applications, the average velocity through the vapour space should be low enough to permit entrained 300 – 500 micron droplets to settle to the highest normal liquid surface (LAH).

For two-phase separators, the light liquid droplets should settle over 80% of the length between the inlet nozzle and the vapour outlet nozzle.

For three-phase separators fitted with a weir, the area of the vapour space should be set by the more stringent of:

- Droplets of the heaviest liquid phase should settle over 80% of the distance between the inlet nozzle and the weir.
- Droplets of the lightest liquid phase should settle over 80% of the distance between the inlet nozzle and the vapour outlet nozzle.

The settling velocity can be calculated using the equations given in Section 5.3.2.4. The settling time is then equal to the vertical drop divided by the settling velocity. The vertical settling time should then be taken as the period over which droplets must settle in compliance with the above guidelines, whilst travelling horizontally at the average vapour rate.

This process is iterative. Liquid levels should be set using a preliminary vessel internal diameter; droplet settling should then be checked for the available vapour space above LAH. An upward or downward adjustment to the vessel internal diameter and/or tan-tan length should be considered depending on the result.

5.4.6.3 Mist Elimination Internals

If excessive liquid carryover cannot be tolerated, a mesh pad can be fitted within horizontal vessels.

However, sizing of the vapour space for 300 – 500 micron droplet settling is still recommended (refer to Section 5.4.6.2).

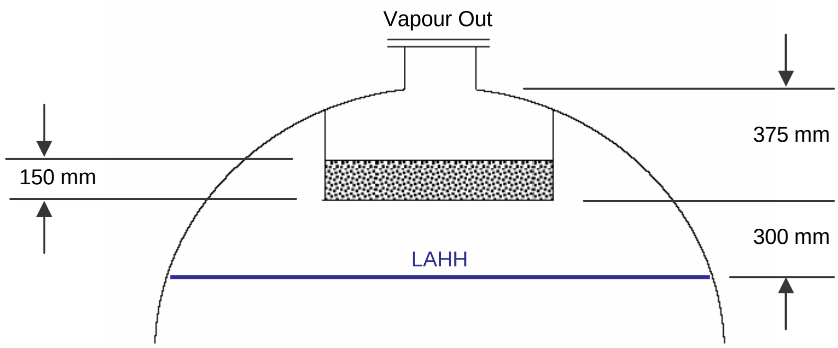
A vertically mounted meshpad will afford greater face area and should be considered where the vapour rate is relatively high and liquid carryover is a particular concern. The internals supplier shall warrant that the liquid carryover will not exceed the limits given in Section 5.3.6.

For mesh pads mounted vertically, the following requirements are also noted:

- In three-phase separators, the mesh pad should be mounted upstream of the weir.
- The mesh pad should extend at least 75 mm below the normal liquid level (to prevent vapour bypassing).

Even where vapour rates are relatively low (i.e. droplet settling criteria are easily met) or in services where liquid carryover is not a particular concern, provision of a mesh pad may still be necessary, generally as a requirement of particular clients. A smaller horizontally mounted mesh pad should be considered in this case. Refer to Figure 5.8.

Figure 5.8 – Horizontal Mesh Pad (for a Horizontal Separator)



In this case, the minimum distance between the LAHH and the bottom of the mesh pad assembly should be taken as 300 mm. Other dimensions given in Figure 5.8 are typical and can be used for preliminary sizing.

Where limiting liquid carryover is not critical, the vessel fabricator can be asked to supply the mesh pad. The process engineer need not solicit a liquid carryover warranty in this case. The following specifications should however be included on the process datasheet as minimum requirements:

1. A minimum face area, determined using the methodology presented in Section 5.3.2.2 for horizontally mounted mesh pads.
2. A minimum mesh pad thickness of 150 mm (6 inches).
3. A minimum mesh pad bulk density of 112 kg/m^3 (7 lb/ft^3).

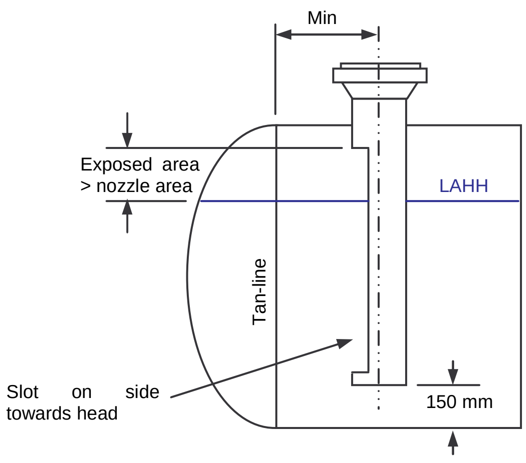
It may also be prudent to communicate any assumptions made with regard to the depth of the mesh pad housing (i.e. 375 mm, as shown in Figure 5.8). If growth in this dimension cannot be tolerated, then the limit should be included on the process datasheet. The supplier should be asked to confirm its acceptability from a fabrication perspective.

Several styles and configuration of vane pack are also available for installation in horizontal vessels (i.e. V-pack). Guidance on the spatial requirements of these devices should be sought from a reputable supplier.

5.4.6.4 Inlet Device

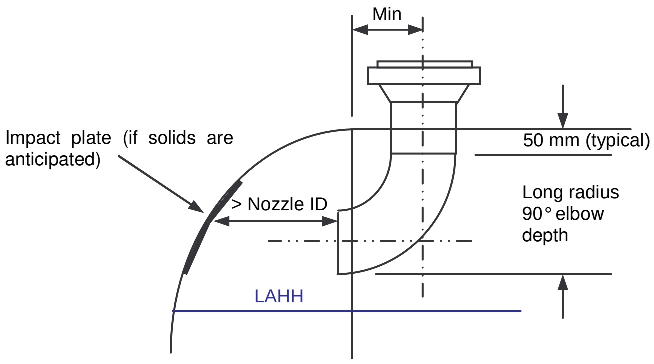
Careful consideration must also be given to the spatial requirements of the inlet device. Where a vertical open-pipe device is used, the slot should extend above LAHH such that the open area (available for vapour flow) at this level is at least equal to the inlet nozzle area. A typical detail is given below:

Figure 5.9 – Open Pipe Inlet Device (Typical)



If an internal elbow is used, the device should be mounted completely above LAHH. For large inlet nozzle sizes, this may consume significant vertical space within the vessel. Where this presents a problem, consideration may be given to installing a smaller inlet nozzle at each end of the vessel and a central vapour outlet connection. This detail should only be used in two-phase separation services. A typical detail is given overleaf.

Figure 5.10 – Internal Elbow Inlet Device (Typical)



Inlet Nozzle	LR Elbow Depth
DN 100	210 mm
DN 150	313 mm
DN 200	415 mm
DN 250	518 mm
DN 300	619 mm
DN 400	813 mm
DN 500	1016 mm
DN 600	1219 mm

5.4.7 Vessel Sizing

Horizontal vessel sizing is an iterative process. A set of approved and validated spreadsheets, namely *Three Phase Horizontal Separator Rev 0* and *Two Phase Horizontal Separator Rev 0*, are available to assist vessel design.

The process engineer should commence by selecting a preliminary vessel internal diameter, from the list of standard head sizes given in Attachment A. Horizontal separators are typically fabricated with aspect ratios in the range of 3:1 to 6:1. In the first instance, set a preliminary tan-tan length by assuming an aspect ratio of 4:1 – 5:1.

The acceptability of this size should then be checked against the sizing criteria (as appropriate) given in the previous sections.

An upward or downward adjustment should then be made to the internal diameter and/or the vessel length depending on the result of the preliminary check.

6 PUMPS

6.1 CENTRIFUGAL PUMPS

6.1.1 Net Positive Suction Head (NPSH)

The margin between $NPSH_A$ and $NPSH_R$ should be the greater of 1.0 meter or 50% of the calculated frictional suction loss. This margin can be reduced once line hydraulics are finalised, i.e. suction piping isometrics have been issued for construction.

6.1.2 Design Pressure

The design pressure of centrifugal pumps should be a minimum of (a) 150% of the design head plus the maximum suction pressure (i.e. suction side PAHH) or (b) the design pressure of the discharge system or (c) the design pressure of the suction system, whichever is greater. This will conservatively account for the characteristic head rise at low throughputs for most centrifugal pumps. This should however be reviewed based on the performance curve of the selected pump.

6.1.3 Design Margins

The difference between the normal and rated capacity for general service pumps should be a minimum of 10%.

A margin of 20% is recommended for reflux pumps to assist with column operations and system controllability.

A design margin is not required for pumps of nominal capacity (i.e. flare drum pumps or make-up pumps).

6.1.4 Pump Circuit Hydraulics

The following pressure drop allowances are recommended when preliminarily sizing control valves in pump circuits. The maximum of the following should be employed:

- 70 kPa at the rated flow;
- 50% of the system frictional pressure loss at the normal flow plus 10% of the elevation difference; or
- 10% of the system frictional loss at the rated flow plus 10% of the elevation difference.

Final control valve pressure drop allowances should be determined using the performance curve of the selected pump and detailed hydraulic calculations considering normal, rated and minimum flows.

6.1.5 Minimum Flow Recycle

☞ Where centrifugal pumps are applied, a minimum flow recycle may be required for stable operation at low flows. An instrumented/controlled recycle should only be provided where the process turndown requirement is beyond the capability of the pump(s). Provision of a manual recycle is generally adequate where recycling is for start-up purposes only, or where the balance of system operation is manual also (i.e. minor utility system).

A continuous recycle is sometimes applied where rapid changes in the forward flow are expected. This feature may be applied to firewater pumps, for example, by installing an RO on the pump discharge returning water to the reservoir. This ensures that when the pump starts automatically on a low header pressure signal, the minimum stable flow will be satisfied regardless of the demand. However, this feature requires that the pump and motor be oversized, with a total capacity equal to the recycle plus the largest process demand.

6.2 POSITIVE DISPLACEMENT PUMPS

6.2.1 Net Positive Suction Head (NPSH)

The minimum margin on NPSH is to be 5 meters if the acceleration head loss is not calculated.

If the acceleration head loss can be calculated, the minimum NPSH margin may be relaxed to 1 meter. Acceleration losses can be estimated using the equation given in Section 12 (Equation 12-17) of the GPSA Engineering Databook [Ref 21].

6.2.2 Design Pressure

As a minimum, the design pressure of positive displacement pumps shall be the greater of the discharge system design pressure or the suction system design pressure.

Relief valves are generally required on the discharge of positive displacement pumps to protect the downstream system against over-pressure. Reciprocating pumps, for example, are often capable of generating a blocked discharge head well in excess of the system design pressure.

The relief valves should be:

- Routed back to the suction vessel where possible;
- Sized for the maximum pump capacity; and
- Must be shown on the P&ID, regardless of whether they are supplied by Clough or the pump vendor.

Where pressure protection is within the scope of the equipment supplier, the vendor may chose to offer a device integral to the pump body. Such devices may not be immediately obvious on the pump drawings. Provision of a suitable device and confirmation of its capacity should be sought through the rotating equipment group.

6.2.3 Design Margins

A volumetric design margin is generally not required for positive displacement pumps.

6.2.4 Piping Design Guidelines

The following piping design guidelines for reciprocating pump systems have been taken from API RP 14E [Ref 31]:

- Discharge piping should be as short as possible and should be one or two nominal pipe sizes larger than the pump discharge connection;
- To minimise pulsations, the velocity in the discharge piping should not exceed three times the velocity in then suction piping; and
- A suitable pulsation dampener, or specific provision for later addition of a dampener, should be fitted as close to the pump cylinder as possible.

6.3 PUMP SEALS

Pumps in hydrocarbon, hot oil and amine service are generally fitted with double mechanical seals to reduce the potential for leakage.

Pumps handling volatile hydrocarbons generally employ a pressurised barrier fluid (i.e. Seal Plan 53 of API 610). In this case, the seal system is maintained above the operating pressure. A low barrier pressure signals seal failure. The barrier fluid for Seal Plan 53 must be compatible with the process, as the normal leakage path will be into the casing.

Amine and hot oil pumps generally have an unpressurised barrier fluid (i.e. Seal Plan 52 of API 610). A high barrier fluid pressure signals seal failure. Plan 52 seals must have a vent to a safe location due to the potential for leakage into the seal system; this should be to the LP flare where possible.

Pump seal instrumentation should preferably be routed to the Process Control System (PCS) rather than the Safety Instrumented System (SIS). This results in a more cost effective solution with no appreciable reduction in safety. Pump shutdowns initiated by seal failure are then affected by the PCS.

6.4 MOTOR DRIVERS

For preliminary estimation purposes, a list of standard electric motor sizes has been appended as Attachment B.

The list provides preferred motor sizes between 1.1 and 1000 kW. However, it should be noted that for motors above approximately 150 kW the ratings are nominal and that manufacturers do not normally publish standard design data, but rather design motors to suit specific duties.

7 SELECTION OF DESIGN PRESSURES AND TEMPERATURES

7.1 DESIGN PRESSURES

7.1.1 Maximum Design Pressure

The system design pressure should be at least 111% of, or 100 kPa greater than, the maximum operating pressure, whichever provides the wider margin. This recommendation will prevent simmering of conventional relief valves and should provide a reasonable band for controllability. The maximum operating pressure should consider start-up, shutdown, maintenance, allowances for controllability, extreme operating conditions such as turndown and extreme ambient temperatures.

Through application of pilot-operated relief valves, manufacturer data suggests that operating pressures may approach 95% of the system design pressure. However, this should only be considered in exceptional circumstances.

Where other considerations do not exist, the guidelines given above can be used to set the system design pressure. It is noted however that an over-riding requirement may set the design pressure at a higher value. This may include:

- Contractual requirements;
- For consistency with an upstream or downstream system;
- To limit or negate a particular over-pressure scenario (i.e. centrifugal compressor or pump discharge). For centrifugal pumps, specific recommendations are made in Section 6.1.2.

7.1.2 Minimum Design Pressure

Equipment should be provided with a minimum design pressure of atmospheric pressure where the formation of a vacuum is not possible. It is noted that vessels, etc may be cleaned with steam during maintenance activities (steamed out) and that condensing steam may pull a vacuum within the system. Where this is the case, a minimum design pressure equal to ‘full vacuum’ should be specified.

Other than for thin-walled vessels, design for full vacuum is unlikely to have a significant impact on wall thickness calculations.

Most small bore piping is suitable for full vacuum service. However, to ensure that steam out operations can be executed safely, all associated piping should be specified for ‘full vacuum’ (FV) on the line list.

7.2 DESIGN TEMPERATURES

7.2.1 Maximum Design Temperature

The maximum design temperature of equipment and piping should be at least 10°C above the maximum anticipated operating temperature. The maximum operating temperature should consider start-up, shutdown, maintenance (i.e. steam out), allowances for controllability, extreme operating conditions such as turndown and extreme ambient temperatures.

For outdoor equipment exposed at least partially to sunlight and capable of undergoing extended periods of pressurised zero flow (i.e. circa 24 hours), the maximum design temperature should be at least as high as the *black body* temperature.

Where guidance is not provided, the black body temperature can be estimated using the following equation:

$$K = \sigma T_s^4 + \frac{h(T_s - T_a)}{\varepsilon}$$

Where,

K = Solar radiation intensity, W/m^2

σ = Stefan Boltzmann constant, 5.6697×10^{-8} W/m^2K^4

T_s = Surface (black body) temperature, K

T_a = Ambient temperature, K

ε = Surface emissivity

h = Surface heat transfer co-efficient, W/m^2K

The hot spot temperature calculated is normally an overestimate since conduction of heat along surfaces is ignored and radiation losses are usually assumed to be only from the receiving surface.

The value of emissivity generally ranges from 0.8 to 1.0. These are typical values for metal surfaces that are painted or have suffered from corrosion or the deposition of dirt.

Depending on latitude, the peak solar radiation intensity usually lies between 900 and 1100 W/m^2. A project specific value should be employed.

Correlations for estimating the convective heat transfer co-efficient (h) are readily available in literature. Two equations with general applicability are given below:

$h = 4.5 + 4.1u$ for wind speeds up to 4.5 m/s

$h = 7.75u^{0.75}$ for wind speeds above 4.5 m/s

Where,

u = wind velocity, m/s

A minimum wind speed of 2 m/s (4 knots) is general acceptable for use in determining the local black body temperature.

7.2.2 Minimum Design Temperature

The **minimum design temperature** for an item of equipment should be the lesser of the minimum ambient temperature or 10°C below the minimum operating temperature. The minimum operating temperature should consider start-up, shutdown, maintenance, allowances for controllability, extreme operating conditions such as turndown and ambient extremes. The minimum design temperature should be given on the process datasheet, and shall be coincident with the maximum design pressure.

Where a pressure vessel or item of equipment designed to AS1210 will experience temperatures colder than the minimum design temperature (above) during controlled depressurisation (blowdown), a **minimum design metal temperature (MDMT)** shall also be given on the process datasheet. The MDMT shall be the lowest of the following:

(a) The lowest internal vapour or liquid temperature resulting from blowdown to 40% of the design pressure;

(b) 50°C warmer than the lowest internal vapour or liquid temperature resulting from blowdown to atmospheric pressure; or

(c) The lowest one day mean ambient temperature per AS1210.

The selected MDMT must be taken as coincident with the design pressure, for material selection purposes.

However, it is extremely important to note that specification of an MDMT using the AS1210 code requirements (above) does not permit *fast* repressurisation of equipment. The various code relaxations recognise the fact that although the equipment may experience extremely cold temperatures during blowdown, these will be coincident with reduced internal pressures and stresses only. If a system with significant metal mass is repressurised immediately following blowdown then the basis for the relaxations is void. In the worst possible scenario, brittle fracture of the equipment could be experienced as internal stresses are rapidly re-introduced.

Full provision for fast repressurisation should only be made where this is required by particular clients.

It may however be appropriate to make this provision on certain small items of gas train equipment, such as filters, where rapid maintenance turn-around is required.

For equipment (i.e. finger style slugcatchers) and pipelines designed to AS2885 the MDMT shall be determined by blowdown to 30% of the **specified minimum yield stress (SMYS)**. The relationship between pressure and stress is given below:

$$P = \frac{2 \text{ x } 0.3 \text{ x } S \text{ x } FET \text{ x } t}{D}$$

Where,

P = Internal pressure, MPaa

S = Specified minimum yield stress (SMYS)

FET = 0.6 for slugcatchers

t = Wall thickness of fingers without corrosion allowance, mm

D = Internal diameter of fingers, mm

[HOLD 1 – ASME Section VIII guidelines]

[HOLD 2 – ANSI B31.3 guidelines]

[HOLD 3 – ANSI B31.8 guidelines]

The process engineer should give careful consideration to how the MDMT is applied. Large systems are likely to include stagnant sections of piping well outside of the blowdown path which remain relatively warm during depressurisation. Conversely, regions which experience high velocities during depressurisation - such as the blowdown nozzle, piping and valves – will experience temperatures very near to that of the internal fluid. For vapour-only systems with a low volume and a high metal mass, it is often reasonable to limit the application of low temperature materials to the blowdown path itself. For example, the BassGas project included a sales gas pig launcher with an MDMT of -29℃ (carbon steel) which required a stainless steel nozzle insert and vent piping, with a locally applied MDMT of -100℃, for this very reason.

Blowdown analyses shall be performed using the Hysys depressurisation module. In general, blowdown operations for minimum temperature calculations should commence from the lower of (a) the minimum operating temperature or (b) the minimum ambient temperature where the system contains a substantial amount of non-condensable vapour. For systems containing only negligible quantities of non-condensable vapour (i.e. depropaniser or debutaniser systems) the blowdown analysis should commence from the normal operating temperature. In both cases, commencement of the blowdown operation from PAHH is reasonable for minimum temperature calculations.

Within the depressurisation module, the *PV Work Term Contribution* (entered under the 'Options' tab in the module) should be set at 0.4 for systems containing substantial quantities of liquid and 0.98 for vapour-only sections of plant[5].

Heat transfer between the system metal mass and the internal fluid shall be modelled using Hysys but the resulting fluxes and co-efficients should be carefully scrutinised prior to acceptance. In general, heat transfer co-efficients from the vessel wall to the contents will be in the order of 100 W/m^2K or less. Heat transfer from the ambient environment to the vessel wall should be neglected.

Credit for heat input via tracing or other means should not be taken when determining minimum temperatures.

[5] Previous Hysys releases required that the process engineer estimate the isentropic efficiency of the blowdown process. Note that the PV Work term is <u>not</u> equivalent to isentropic efficiency. Numerically, it is less conservative when determining minimum temperatures.

8 PIPING AND VALVE ARRANGEMENTS

8.1 ISOLATION

8.1.1 Standard Isolation

8.1.1.1 Single Valve Isolation

Single valve isolation should generally be applied for the isolation of equipment in **ANSI Class 600 systems and lower** which operate at near-ambient temperatures.

A single valve does not constitute *positive isolation*; where vessel entry is required an acceptable means of positive isolation will also be required (i.e. spectacle blind, drop-out spool). Inasmuch, provision of an isolating block valve on all connections to unspared item of equipment is not likely to improve maintainability.

If it is inevitable that maintenance of an equipment item will result in a loss of production, it is generally acceptable for a larger section of the gas train to be isolated for controlled depressurisation. Therefore, where possible, a sectional approach to valved isolation should be taken. The omission of unnecessary isolation valves will reduce the number of flanges (leakage sources) in the system.

8.1.1.2 Double Valve Isolation

Single valve isolation may not be adequate for high pressure systems - typically ANSI Class 900 and above - where even a small valve opening will result in a high leakage rate. To improve the reliability of isolation between a high pressure system which remains in operation and a low pressure or atmospheric system, the application of two series isolation valves is recommended.

Where maintenance of an item of gas train equipment may be required on a frequent basis (i.e. filters, scraper receivers) provision of double block isolation should also be considered for ANSI Class 600 systems.

In most circumstances, an appropriately located shutdown valve can be used as one of the two valves in double valve isolation. Note that the instrument air supply line can be closed and tagged, or simply disconnected, to give assurance that the SDV will not be inadvertently opened. The SDV must be fail-closed.

Although the detail is inherently more safe than single valve isolation, double block isolation does not constitute *positive isolation*; where vessel entry is required an acceptable means of positive isolation will also be required (i.e. spectacle blind, drop-out spool). Inasmuch, provision of two isolating block valves on all connections to an unspared item of equipment is not likely to improve maintainability.

If it is inevitable that maintenance of an equipment item will result in a loss of production, it is generally acceptable for a larger section of the gas train to be isolated for controlled depressurisation. Therefore, where possible, a sectional approach to valved isolation should be taken, with double block valves at the system's periphery. The omission of unnecessary isolation valves will reduce the number of flanges (leakage sources) in the system.

Control valves should only be provided with double valve isolation (upstream and downstream) in ANSI Class 900 service (and above) and where continued safe operation is possible whilst the valve is maintained. If the downstream block valve is located at a piping specification break to ANSI Class 600 or lower, then the second downstream block valve can be omitted.

8.1.1.3 Double Block and Bleed

This detail includes a small bore bleed valve between the duplicated isolation valves. The bleed should in general be executed as a DN 20 drain on the bottom of the line (although particular clients may have a requirement for DN 25 or DN 50 valves). However, in fouling services the valve should be mounted on the top or side of the line, with piping directed to grade. In this arrangement the isolation of equipment is better ensured than for single valve isolation, and offers an advantage over double block in that leakage across each valve can be checked.

Operating procedures should ensure that the bleed valve remains closed even when the equipment is isolated.

Double block and bleed isolation should be considered in any case requiring double valve isolation (refer to Section 8.1.1.2). Realistically, because valve tightness is better assured, most operating companies will have a preference for this arrangement. Where direction is not provided, a clear philosophy should be established prior to entering into a lump-sum EPC contract.

The risks associated with the use of double block and bleed installations should be noted. The drain valves are generally small bore; these are easily blocked by debris. Full closure of the valve may not be possible after venting if a partial blockage occurs. Furthermore, a fully blocked drain/vent valve can be inadvertently left open with obvious safety implications when the line is repressurised.

Double block and bleed isolation is not considered to be positive isolation.

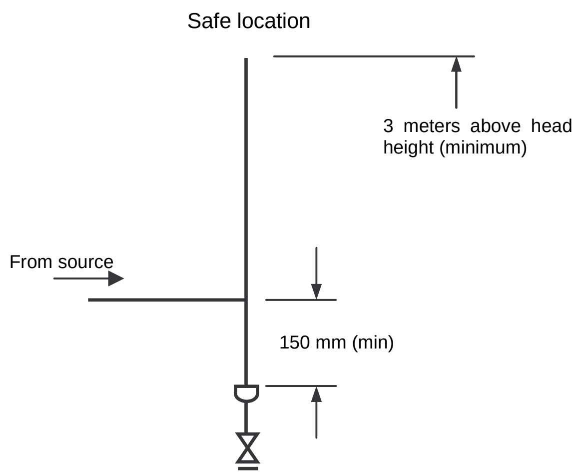
8.1.1.4 Double Block and Bleed to Safe Location

This should be considered where double block and bleed isolation is required but the process fluid is toxic. This arrangement should also be considered for large valves with a significant inter-block volume (i.e. between isolation valves for a scraper receiver).

The safe location can be to flare or a well-located atmospheric vent. Many closed drain systems use the flare drum as the liquids receiver; this is considered to be a safe location also.

A connection to flare should always be employed in preference to an atmospheric vent. Where this is not practical, the height of the atmospheric vent should only be finalised following a dispersion analysis. As a minimum, the vent should be located 3 meters above the most onerous of the following: head height or the height of any adjacent equipment, buildings or structures [Ref 22]. The vent should be fitted with some means of draining liquids. The prevention of rain ingress using a 180° return bend at the top of the vent is not acceptable. Likewise, rain caps are not recommended as gas dispersion will be more effective when the fluid is able to jet upward freely. A small bore drain valve fitted to the base of a liquid leg is an effective means of removing collected liquids prior to venting. Refer to Figure 8.1.

Figure 8.1 – Typical Atmospheric Vent Detail



Where an atmospheric vent is employed, the inter-block bleed valve(s) shall be executed as a vent mounted on top of the piping. If the bleed valve is connected to the flare or closed drain system then a connection can be made on the top or bottom of the line. Bottom connections should be avoided in fouling services to prevent bleed valve blockage.

8.1.2 Positive Isolation

Isolation is considered *positive* when a physical separation can be guaranteed between systems.

Positive isolation shall be achieved using spectacle blinds, spades or removable spool pieces. Although a sectionalised approach to installing isolation valves is encouraged in Section 8.1.1 (standard isolation), a more conservative view should be taken for positive isolation. A means of positive isolation should be provided on all connections to equipment which can be physically entered or opened. These devices should be located as near as practical to the process nozzles (preferably at the nozzle flange itself). Apply to:

- Pressure vessels (including separators, columns and filters);
- Heat exchangers;
- Pumps; and
- Tanks.

Positive isolation need not be provided where the duration of the maintenance activity will not be significantly longer than the act of installing positive isolation itself (i.e. strainer and scraper receiver cleaning) and where confined entry into a vessel or tank is not required. Additionally, positive isolation is not required upstream and downstream of control valves as they can be removed and replaced with blind flanges in around the same time it would take to swing spectacle blinds.

The following guide should be used to select an appropriate means of positive isolation:

Table 8.1 – Positive Isolation Selection Guide

Means	Advantages	Recommended for	Exceptions
Spectacle Blind	1. Complete means of positive isolation installed on piping, not in stores. 2. Lifting of spade, spacer or spool not required. 3. Piping alignment is maintained (one bolt is held in place).	1. General service. 2. Pump suction and discharge lines (where maintenance of piping alignment is critical).	1. Where the process fluid operates below 0°C. 2. Where a removable section of piping is required for maintenance anyway. 3. Clough project piping specifications generally give size limits for the application of spec blinds. These limits should be reviewed as part of selection process. Consider spade and spacer or removable piping spool instead.
Spade and Spacer	1. Low heat ingress compared with spectacle blinds.	1. Where a spectacle blind would be employed, but the operating temperature is below 0°C. 2. Where required by project piping specifications for large bore piping.	1. Where a removable section of piping is required for maintenance anyway. Consider removable piping spool instead.
Removable Piping Spool	1. Highest degree of positive isolation is assured.	1. Tubeside connections on shell-and-tube heat exchangers, where the tube bundle is removable. Likewise, plate-and-frame heat exchangers to allow plate inspection. 2. Where process piping is connected to a manway or filter closure. 3. Where ventilation is required (see below). 4. Relief valve inlet lines (see below).	-

The installation of a relatively thin *temporary blind* between a set of flanges is, in fact, a further means of establishing positive isolation. This requires that adequate flexibility exists in the associated piping to allow for spreading of the flanges. Most references indicate that adequate flexibility can be provided in piping systems up to DN 150, assuming that the requirement is effectively communicated to the piping designers. However, many of our clients will view temporary blinds as an inferior measure; for this reason they are not recommended.

Spectacle blinds should therefore be installed in most general service applications; this will not impose an unreasonable cost on the project.

Where provision has been made to positively isolate an item of equipment, the following should also be present:

- A means of safe depressurisation;
- Provision for drainage of liquids (as appropriate); and
- Connections for purging (with nitrogen) to ensure that the environment is gas-free.

These connections need not be provided on the item of equipment (within the positive isolation). They must however be appropriately situated within the system’s isolation valves, such that the equipment can be made safe for positive isolation/entry. For example, depressurisation and drainage facilities are not typically included on shell-and-tube heat exchangers. Preparation for maintenance will be made using facilities outside of drop-out spools on the tubeside.

The introduction of nitrogen (to gas-free a system) through a vent, drain or instrument connection is acceptable where cyclic purging is employed.

As was noted in Table 8.1, the use of at least one (1) removable piping spool is recommended for ventilation on equipment where physical entry is possible (i.e. pressure vessels, tanks). Ventilation will generally be achieved by sweeping air through the vessel from the manway (ventilation inlet) and out through a nozzle. The ventilation outlet should be the nozzle most remote from the manway, and should be DN 150 mm as a minimum.

Where a large pressure vessel is fitted with suitably located manways at either end (i.e. fractionation column) this requirement is waved.

In many applications, pressure relief valve inlet piping will be connected directly to an equipment-mounted nozzle. Where the only gas path to (and from) the flare system is via a single PSV, positive isolation need not be provided as the relief valve can be physically removed and replaced with a blind flange; refer to Figure 9 of API RP 520 Part II [Ref 23].

Where multiple relief valves are installed and/or the arrangement includes a manual vent connection, the removal and replacement of valves for positive isolation may not be practical or safe. Consideration should then be given to providing a drop-out spool at the equipment nozzle. This is the only acceptable means of positive isolation in this application. Spectacle blinds and spades/spacers shall not be installed in relief valve piping.

8.2 DRAIN AND VENT CONNECTIONS

Where required for maintenance, valved drain connections shall be provided at low-points and valved vent connections shall be provided at high points in piping and equipment systems.

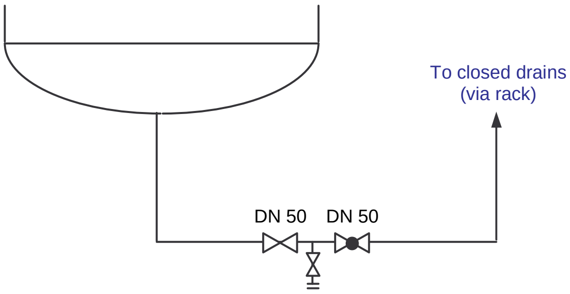
Local drains and vents connected to piping for maintenance activities should consist of a single block valve and should be closed with a blind flange, cap or threaded plug (in accordance with the piping specification). The block valve should generally be executed as a DN 20 gate valve. Local vents and drains are, by definition, intended for the local release of a fluid inventory and are not hard-piped to a closed system for disposal. Where necessary, a hose connected to the local drain point can be used to affect transfer to a drum.

Local vents and drains may be employed for nitrogen purging.

Equipment drains connected to a closed drain system should generally consist of a DN 50 block valve followed by a DN 50 globe valve (for throttling). Both valves are to be rated for the upstream pressure; low temperatures experienced during drainage (or inadvertent gas blow-by) should be considered in specification of the globe valve and downstream piping. Consideration should be given to the requirement for positive isolation of the drain line.

Provision of a DN 20 local piping drain may also be required on the equipment drain line to ensure that the vessel and piping is truly liquid-free prior to positive isolation/entry. Refer to Figure 8.2:

Figure 8.2 – Typical Vessel Maintenance Drain Detail



Primary equipment vents should be connected to the flare system where possible, for safe depressurisation. These vents should generally consist of a DN 25 block valve followed by a DN 25 globe valve (for throttling) and are usually installed as part of the relief valve arrangement (utilising a common connection to flare). Both valves are to be rated for the upstream pressure; low temperatures experienced during venting should be considered in specification of the globe valve and downstream piping. A DN 20 local vent connection should also be provided where the primary maintenance vent is connected to the flare, to enable final depressurisation to atmospheric pressure.

Note that the capacities of manual globe valves will vary depending on manufacturer. This makes estimation of the full-open venting rate difficult. Where the venting rate is higher than anticipated, the built up back-pressure against the globe valve may exceed the flange rating of the low pressure disposal system, typically Class 150. This has obvious safety implications. Where the globe valve C_V in not accurately known, one of the following measures is recommended:

(a) Installation of a restriction orifice upstream of the globe valve to limit the flow and back-pressure to a value safely within the downstream design pressure (example: Sawan Gas Field Development), or

(b) Rating of the relatively small bore piping downstream of the globe valve to the upstream design pressure. Full rating should be maintained until the vent tail-pipe ties into a significantly larger header or sub-header.

Where the requirement for a maintenance vent or drain has been identified by the process engineer or piping designers, the detail should be reflected on the P&ID.

However, vents and drains for hydrostatic testing should not be shown on the P&IDs. These connections should not be fitted with a block valve, and are to be closed with a blind flange or plug only (as appropriate). Where threaded plus are employed, these must be welded closed as PTFE tape degrades in hydrocarbon service.

8.3 SAMPLE CONNECTIONS

8.3.1 General

Consideration should be given to sampling requirements at the bid stage for EPC contracts. Even where the most basic provision is made on a gas plant, 6 - 8 connections may be required. If we are obliged to provide sample cabinets, looped systems, connections to flare, sample coolers, etc this may come at considerable expense. The number of sample points and installed cost of the full system will be considerably higher in refinery applications.

This discussion has been divided into two sections: basic requirements and example sampling stations.

Where guidance has not been provided by a particular client, the first section outlines a set of minimum facilities for sampling. It is not uncommon for P&IDs issued for tender to indicate where sampling is required but with extremely simplistic detail. If we make a costed allowance for complete sample stations where our competitors do not, this will affect Clough's competitiveness. Conversely, if realistic costs are not captured then this will affect project profitability at a later stage.

The allowance made in our bid should be clearly stated in the corresponding submission. If basic provisions have been made then the detail should be described together with an indication that Clough is willing to develop a more replete system collaboratively prior to award. Where we decide that a more considered approach is necessary, we should submit a clarification forcing other bidders to make similar allowances.

The second section offers a set of example sample station details for particular applications.

As a matter of course, consultancy work that progresses the design to P&ID-level engineering (i.e. FEED) should include a more considered approach to sampling (refer to Section 8.3.3).

8.3.2 Basic Requirements

8.3.2.1 Location

As a guide:

- Sample connections should be provided on all plant feed streams. Obtaining representative multiphase samples is difficult. Therefore, sample points should be provided on the various outlet streams of the slugcatcher, inlet separator or equal.
- Sample connections should be provided on all product streams (i.e. stabilised condensate, propane, butane, sales gas, etc).

  Most gas plants will include an analyser (gas chromatograph) on the sales gas stream; this is generally executed as part of the metering package. The analyser will require a looped system, with a connection to flare for disposal of the sample gas. The supplier should be asked to make provision for manual sampling with a pressurised bomb within the package to prevent duplication of facilities.
- If gas treatment is performed in several steps, consideration should be given to installing sample points downstream of each major unit operation (i.e. downstream of a dehydration, downstream of amine contactor, etc).

Most 'turn-key' contracts will require that a 72-hour performance test be performed prior to handover. The sample points given above should reasonably fulfil our contractual obligations by enabling an overall mass-balance review.

In refinery applications, it is recommended that all product sample points be located together at the battery limit as a large number of samples will be required per shift.

8.3.2.2 Facilities

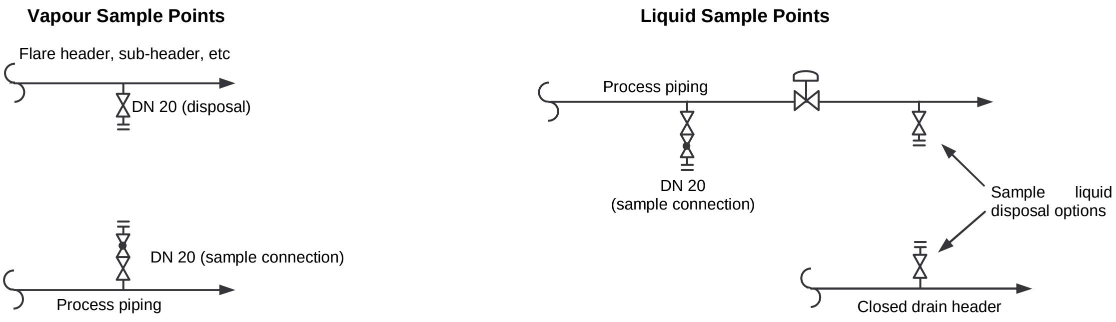
Where a sample point is required, the most basic provision on the process piping is a DN 20 block valve followed by a DN 20 globe valve or needle valve. Multiple block valves may be required in high pressure services. The arrangement should be closed with a blind flange or plug (in accordance with the piping specification). This provision does not constitute a safe sampling point, but does allow for the tie-in of temporary (or permanent) facilities at a later stage. This is a reasonable allowance where the client (operator) has not clearly outlined his requirements. The limitations of this allowance should however be described in our bid.

For vapour sample points, a looped system will generally be required for filling of a pressurised bomb. Therefore, a connection should be provided on a nearby part of the flare system also. This should consist of a DN 20 block valve and a suitable closure. The disposal block valve must carry the same rating as the DN 20 process connections, not the disposal (flare) system.

Connections should be made for looping of liquid samples also. If the sample connection can be made upstream of a control valve then provision of a DN 20 block valve (and closure) on the downstream side will permit reintroduction of the sample stream. Otherwise a connection should be provided on a nearby part of the closed drain system. The disposal block valve must carry the same rating as the DN 20 process connections, not the disposal system.

Vapour phase sample connections should be located on the top of the process piping. Liquid phase samples should be taken from the bottom of the line. Sample connections should not be situated at dead ends. The minimum facilities are shown diagrammatically overleaf:

Figure 8.3 – Basic Sampling Facilities



8.3.3 Examples of Full Sampling Stations

A series of examples for full sampling stations have been appended as Attachment D.

9 SYSTEM GUIDELINES

9.1 FLARE SYSTEMS

9.1.1 Relief Valves

Relief valves shall be sized in accordance with the recommended practices given in API RP 520 – Part I [Ref 24].

Conventional relief valves should be used where the back-pressure does not exceed 10% of the set pressure and the maximum operating pressure is less than 90% of the set pressure.

Where the back-pressure is greater than 10% of the set pressure, a balanced-bellows or pilot operated PSV should be employed.

According to manufacturer data, pilot operated relief valves can also be used in applications where a margin of only 5% exists between the operating and set pressures. The likely range of variability in a controlled pressure (normal ± 1%, for example) should however be considered when operating very near to the set pressure.

9.1.2 Flare Tips

A high pressure (HP) tip should, in general, be employed to minimise the relief header size and to achieve more complete combustion of emergency relief vapours.

For the peak flaring case, a differential pressure of 500 kPa across the HP flare tip is a reasonable design allowance.

Where local venting is shown to be unacceptable, the use of a supplementary low pressure flare should be considered where:

- continuous disposal of low pressure vapours is required (i.e. vapour effluent from a TEG regeneration package, especially where BTEX is present);
- back-flow from a high-pressure flare could cause mechanical damage to equipment with very low design pressures (i.e. out-breathing from field erected tanks); and/or
- multiple items of equipment with design pressures substantially below 1000 kPag require pressure protection (refer to Section 9.1.4).

For the peak flaring case, a differential pressure of 3 - 7 kPa across the LP flare tip is generally reasonable. This should however be reviewed for particular applications.

For facilities requiring both an LP and HP flare, manufacturers can generally provide a combined tip for mounting on a single stack.

9.1.3 Piping and Headers

9.1.3.1 Relief Valve Inlet and Pilot Piping

Sizing of relief valve inlet piping shall limit non-recoverable losses to 3% of the set pressure where conventional (or any on-body sensing) relief devices are employed.

Where this cannot be practically achieved, the device should be specified as an off-body pilot relief valve (with remote sensing). Although remote sensing may eliminate valve chatter, the relieving capacity will be reduced by the additional pressure drop in the inlet piping. In this case, the set pressure *and* the expected relief valve inlet pressure should be provided on the process datasheet so that appropriate derating can be considered in orifice and body selection.

The following points, taken from API RP 520 – Part II, are also noted:

- Relief piping pressure losses should be evaluated at the rated valve capacity.
- The nominal size of the inlet piping shall be at least equal to the inlet flange of the relief valve.
- Isolation valves shall be full bore, and shall be capable of being locked or car-sealed open.

- Where gate valves are used for isolation they should be installed with stems mounted horizontally to prevent the gate falling inadvertently. Where this cannot be accommodated, inclined mounting should be considered; however the stem angle should not exceed 45° (measured from the horizontal).
- For off-body pilot relief valves, the remote sensing line should measure static pressure where the velocity is low. For flowing pilots, remote sensing lines shall be sized to limit the pressure loss to 3% of the set pressure based on the maximum flow rate of the pilot at 110% of the set pressure. The manufacturer should be consulted for recommendations. For non-flowing pilots, a remote sensing line with a flow area of 45 square millimetres is sufficient.
- Inlet (lead) piping should free-drain back to the source with no pockets.

9.1.3.2 Relief Valve Tail Piping

Back-Pressure

Relief valve tail piping should be sized to limit the total back-pressure to 10% of the set pressure where conventional relief valves are installed. Where this is not practical, and a balanced-below or pilot operated PSV is installed, back-pressures of 30 - 50% are acceptable. Refer to API RP 520 – Part I for the effect of back-pressure on relief valve performance and capacity.

Additionally, sizing of the tail pipe shall ensure that the back-pressure does not exceed the lower of:

(a) The limit of ANSI Class 150 flanges (typical of flare systems) at the piping design temperature; or

(b) The outlet pressure limits given in Tables 2 to 29 of API Standard 526. Note that for spring-loaded pressure relief valves (i.e. conventional, balanced-bellows) the outlet pressure limitation is substantially lower than Class 150 flange ratings for ‘Q’ orifices and larger.

Velocity

The tail pipe velocity should not exceed 0.75 Mach. This should be evaluated at the lowest reasonable back-pressure in the flare system. In many cases, the lowest system back-pressure will occur when only that valve relieves.

Arrangement

The following points, taken from API RP 520 – Part II, are also noted:

- Relief piping pressure losses should be evaluated at the rated valve capacity.
- The nominal size of the discharge piping shall be at least equal to the outlet flange of the relief valve.
- Isolation valves shall be full bore, and shall be capable of being locked or car-sealed open.
- Where gate valves are used for isolation they should be installed with stems mounted horizontally to prevent the gate falling inadvertently. Where this cannot be accommodated, inclined mounting should be considered; however the stem angle should not exceed 45° (measured from the horizontal).
- Outlet (tail) piping should contain no pockets and is to slope towards the header or local sub-header at 1:200 (refer to Section 9.1.3.4).

9.1.3.3 Blowdown Valves

Blowdown valve inlet piping should be sized to limit the velocity as follows:

$$v_G < \frac{152}{\sqrt{\rho_G}}$$

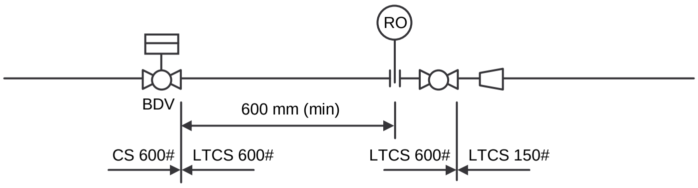
Where,

v_G = vapour velocity, m/s

ρ_G = vapour density, kg/m^3

Where low temperatures requiring a material break are anticipated, the BDV should be located a minimum of 600 mm upstream of the orifice. The temperature material break should be located immediately downstream of the BDV; the pressure material break should be taken downstream of the manual isolation valve. An example detail is shown below:

Figure 9.1 – Example Blowdown Valve Detail



The outlet pipe velocity should not exceed 0.75 Mach. This should be evaluated at the lowest reasonable back-pressure in the flare system. In many cases, this will occur when blowdown through only that valve is executed (for maintenance perhaps).

The outlet piping should contain no pockets and is to slope towards the header or local sub-header at 1:200 (refer to Section 9.1.3.4).

9.1.3.4 Collection System

Flare collection piping shall be sloped towards the flare drum. The following slopes are recommended:

Sub-headers	1:200
Main header	1:500

Header piping between the drum and the stack shall slope back towards the flare drum at 1:500 also.

Collection piping shall contain no pockets.

The velocity in the header should not exceed approximately 0.5 Mach for emergency (peak) flaring cases.

For more normal or continuous flaring loads, a velocity of approximately 0.2 Mach is a reasonable design figure.

The velocity guidance given here can be applied to both HP and LP flares, though back-pressure considerations are likely to govern LP flare design.

9.1.4 Equipment Design Pressures

Equipment connected to high pressure flare systems should have a minimum design pressure of 1000 kPag. Equipment with design pressures lower than 1000 kPag should be connected to an LP flare (or equal).

This requirement is largely based on economic considerations. Even where a non-conventional relief valve is employed, the back-pressure should never exceed approximately 50% of the set pressure because the capacity of the valve is based on critical flow. The consequence is that the design pressure of the lowest pressure system may govern the size of the entire flare system.

9.1.5 Radiation

9.1.5.1 Limits

The following acceptable exposure times at given heat radiation levels can be used in selection of the flare stack height and finalisation of the sterile radius. These values are applicable to appropriately clothed personnel and include a value of 1.0 kW/m^2 for the contribution of solar radiation.

Table 9.1 – Acceptable Flare Radiation Levels

Incidental Radiation (personnel) [kW/m^2]	Acceptable Exposure Time	Recommended Application
1.6	Continuous	At any point where appropriately clothed personnel may be present or uncontrolled public access can be gained during normal or continuous flaring. Generally applied as the radiation limit at the plant boundary for the normal or continuous flaring rate.
3.2	10 minutes	At any point where appropriately clothed personnel may be present or uncontrolled public access can be gained during peak flaring. Generally applied as the radiation limit at the plant boundary for the peak flaring rate.
4.7	2 minutes	Heat intensity in areas where emergency actions lasting several minutes may be required by personnel without shielding but with appropriate clothing.
6.3	20 seconds*	Generally applied as the radiation limit at the boundary of the flare exclusion zone.
9.5	Several seconds only*	Maximum allowable radiation level at any ground level location, i.e. at the base of the flare stack.

*For on-site personnel in emergency situations, a reaction time of 3 – 5 seconds can be reasonably assumed. Perhaps 5 seconds more time would elapse before the average individual could seek cover or depart the area, resulting in a total exposure period ranging from 8 – 10 seconds. It is generally accepted that the ground level radiation limits given above between the base of the flare stack and the boundary of the sterile area will afford adequate time for trained staff to take action.

Ground (or platform) level radiation during flaring is affected by the coincident wind velocity. Distortion of the flame at high wind speeds will tend to lower the radiation isopleths on the leeward side, leading to higher radiation levels in accessible areas. This affect must be considered in design of the flare and layout of the facility. Application of the structural design wind speed (maximum gusts) in this regard is not reasonable. Where guidance is not provided, maximum and minimum wind speeds of 15 and 2 m/s are recommended.

The guidance given below is provided to further assist plant layout and review:

- The peak radiation intensity on carbon steel structures without special protection or specifically designed for higher levels is generally given as 15 kW/m^2.
- Hand rails can become too hot to handle without gloves if exposed to heat radiation levels above 3.2 kW/m^2 at low wind speeds.
- High voltage links and transformers should not be located where they can be exposed to heat radiation intensities greater than 2 kW/m^2.
- 1.6 kW/m^2 is an acceptable continuous heat radiation level for hardy plants, including crops.

The radius of the flare exclusion zone should be set based on the radiation guidelines given above. Some companies – including many of our clients – set minimum requirements on the size of flare exclusion zones. Given that safe working distances will vary depending on the size of the peak relieving case, an arbitrary minimum requirement will not be set here. Equipment containing hydrocarbons, including the flare drum and pumps, should however be located at least 50 meters from the flare or outside of the exclusion zone, whichever is further.

When flaring continuously, equipment surface temperatures at a given radiation intensity are readily estimated by considering a steady state 'hot spot' heat balance. In this approach heat input by radiation is balanced against heat loss due to convection, re-radiation from the surface and reflected radiation. The hot spot temperature calculated is normally an overestimate since conduction of heat along surfaces is ignored and radiation losses are usually assumed to be only from the receiving surface. The following equation is applicable:

$$K = \sigma T_s^4 + \frac{h(T_s - T_a)}{\varepsilon}$$

Where,

K = radiation intensity, W/m^2

σ = Stefan Boltzmann constant, 5.6697×10^{-8} W/m^2K^4

T_s = Surface temperature, K

T_a = Ambient temperature, K

ε = Surface emissivity

h = Surface heat transfer co-efficient, W/m^2K

The value of emissivity generally ranges from 0.8 to 1.0. These are typical values for metal surfaces that are painted or have suffered from corrosion or the deposition of dirt.

Correlations for estimating the convective heat transfer co-efficient (h) are readily available in literature. Two equations with general applicability are given below:

$h = 4.5 + 4.1u$ for wind speeds up to 4.5 m/s

$h = 7.75u^{0.75}$ for wind speeds above 4.5 m/s

Where,

h = Surface heat transfer co-efficient, W/m^2K

u = wind velocity, m/s

In general, heavy equipment and structural members have sufficient heat capacity to ensure that the steady state temperature is only reached after a period of one hour or more at a high rate of flaring.

9.1.5.2 Estimation

A reliable preliminary estimate of ground-level radiation can be made using the **Kaldair** software which is available to the process group (Kaldair now trade as John Zink). The calculation is composition and rate based, and accounts for the effect of stack height and wind velocity in determination of grade radiation. This software can be usefully applied to study and concept selection work where an estimate of the flare stack height and/or the radius of the exclusion zone are required.

The following flare types can be examined: high-pressure (Indair and Mardair), open pipe, steam-assisted, air-assisted and ground (enclosed) flares. The executable software also contains useful design and selection data for each type of flare.

9.1.6 Dispersion

9.1.6.1 Flammability

For both flare stacks (flame out case) and vents a flammable mixture shall not occur within 2 meters of grade or any working surface. For calculations a flammable mixture is defined as 0.5 times the lower flammable limit (LFL). Permanent ignition sources and open flames should be located outside of the 0.25 LFL contour.

9.1.6.2 Toxicity

For both flare stacks (flame out case) and vents an acceptable toxic component concentration shall not be exceeded within 2 meters of grade or any working surface.

Guidance on acceptable design H_2S concentrations from flare and vents is given below. These requirements may be subordinate to local environmental or occupational heat and safety regulations or specific project/client requirements.

Table 9.2 – Acceptable Design H_2S Concentrations

Location	Acceptable H_2S Levels	
	Flares (flame out case)	**Vents**
Uncontrolled public access (i.e. outside of plot fence)	Continuous < 0.07 µg/m^3 (Note 1) Emergency < 10 ppmv (Note 2)	Continuous < 0.07 µg/m^3 (Note 1) Emergency < 1 ppmv
Restricted access (process area, etc)	Continuous < 10 ppmv (Note 2) Emergency < 100 ppmv (Note 3)	Continuous < 1 ppmv Emergency < 10 ppmv (Note 2)
Sterile Area	Continuous < 10 ppmv (Note 2) Emergency < 200 ppmv (Note 4)	Continuous < 10 ppmv (Note 2) Emergency < 200 ppmv (Note 4)

Notes:

1. Expressed as a yearly time weighted average.
2. Level is acceptable for exposure at 8 hr/day, 40 hr/week.
3. Level is acceptable for 1 hr exposure. Personnel not wearing breathing apparatus will experience irritation; the severity will increase with time.
4. Personnel not wearing breathing apparatus will experience severe irritation but have adequate time to leave the sterile area (approximately 10 minutes).

9.1.7 Purge Rates

Flare purging with fuel gas shall be provided to prevent air ingress into the stack piping. The purge rate should provide a velocity of 0.02 m/s in the header.

Where a *velocity seal* or equal is fitted to the flare tip lower purge rates may be acceptable. The manufacturer should be consulted for recommendations.

It is noted that many fuel gas streams will be significantly lighter than air. The less dense column of gas in the stack will result in sub-atmospheric pressures in the flare system. In order to minimise the chance of air ingress, the number of flanged connections and DN 50 and smaller connections should be minimised. Special attention should be paid to providing adequate support for small diameter branch connections.

9.1.8 Flare Drum Sizing

Supplementary to the guidelines given in Section 5.4 for horizontal separators, the following should be considered in sizing or flare drums:

- The vapour space should be adequate for settling of 500 – 600 micron liquid droplets for low-pressure flares, and up to 1000 microns where a high-pressure tip is installed (refer to Section 5.3.4.4 for guidance on calculation of droplet settling rates).
- A liquid inventory consisting of 15 minutes of relief liquids plus the largest reasonable equipment drainage volume (where the drum serves as the closed drain vessel also) should be assumed coincident with the peak vapour rate when sizing the drum.

A full treatment of flare drum sizing is given in Clause 5.4.2 of API RP 521.

9.1.9 Combustibility

Vapours with heating values lower than 200 – 300 BTU/scf may prove to be incombustible. A fuel gas supplement may be required to ensure combustion.

9.1.10 Fire Case Relief

Other than in exceptional circumstances, fire case relief for onshore facilities should be based on liquid pool fires only. Appropriate heat fluxes for wetted area and vapour-only equipment are given in API RP 521. In most situations, adequate drainage can be assumed and on-site fire fighting equipment will exist, justifying the use of the less conservative equation in Clause 3.15.2.1.1 of API RP 521.

In accordance with Clause 5.2.2 of API RP 521, when evaluating the effect of simultaneous relief from multiple fire case valves on the flare system, a pool fire diameter of 25 meters should be employed.

It should be assumed that liquid pooling is not possible on access platforms.

For pool fires, equipment elevated more than 8 meters above grade should be excluded from exposed area calculations (i.e. air cooler heat exchangers mounted high on pipe racks).

Where we are required to consider jet fires on offshore installations for particular clients, and guidance has not be provided, a heat flux of 200 kW/m^2 can be employed. This figure should however be reviewed for specific projects. For jet fires, 100% of the exposed surface area should be considered for horizontal or short vertical items of equipment. For vessels with a high vertical aspect ratio (i.e. columns) it would be reasonable to consider, say, 50% of the exposed area only.

9.2 HOT OIL SYSTEMS

9.2.1 Heat Transfer Fluid

The process heating requirements in most gas plants will be at levels well below 220°C. Mineral (or petroleum) based hot oils are suitable for operation at bulk temperatures of up to 300 - 320°C; slightly lower supply conditions are typically employed to provide a safe working margin. Combined with a substantially lower cost than synthetic products and their ease of sourcing, the use of mineral based heat transfer fluids is encouraged.

Clough has successfully employed BP Transcal N on several EPC projects. Mobiltherm 605 and Essotherm are other examples of similar mineral oils.

9.2.2 Supply and Return Conditions

By specifying the broadest possible range of supply and return conditions from heat consumers the hot oil circulation rate, pump and piping sizes, the total system inventory and the first-fill cost will all be minimised. The use of a lower return temperature will however reduce the LMTD of heat consumers, thereby increasing the required surface area. The selection of supply and return temperatures is therefore an economic consideration.

For mineral based hot oil systems, supply temperatures in the range of 270 – 290°C are generally suitable.

A return temperature of at least 20°C above the cold feed to the highest temperature consumer is recommended. For a gas plant containing glycol dehydration and condensate stabilisation, a single-tier system with a return temperature of 220 – 230°C would be appropriate.

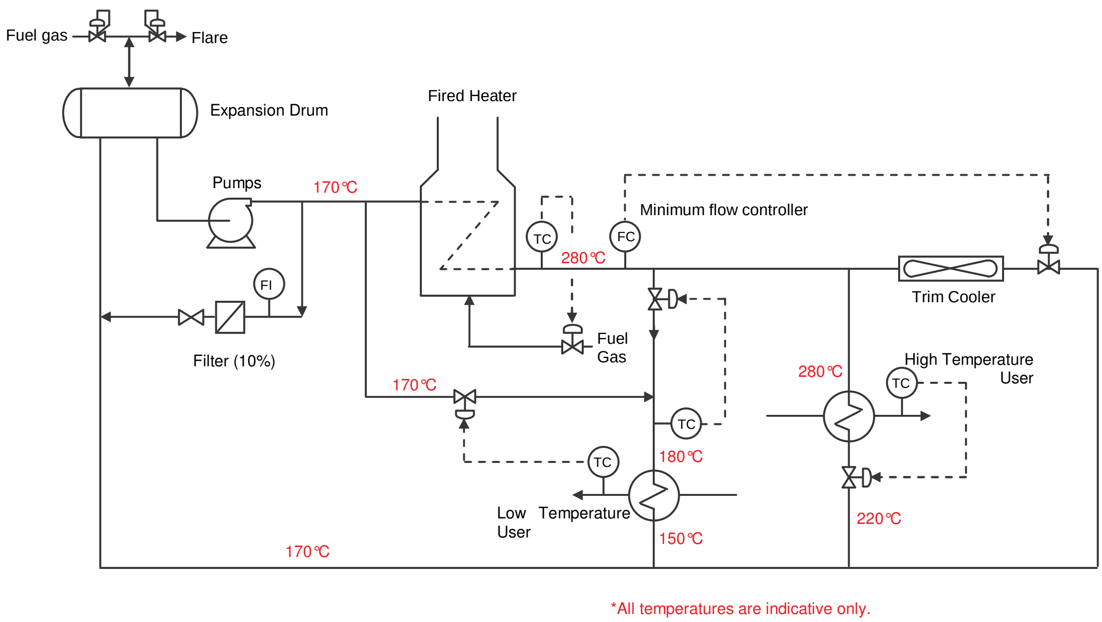
The hot oil supply temperature to amine (i.e. MDEA) stripper reboilers should generally be limited to 180 – 190°C to reduce the likelihood of degradation. This should however be confirmed by the amine system licensor or technologist. Supplying all heat consumers with hot oil at this temperature may not be workable, particularly if other equipment operates at particularly high temperatures (i.e. condensate stabiliser). A two-tier system should be considered in this case. The return temperature for the lower tier should be set above the amine stripper reboiler feed temperature; a return temperature in the range of 135 – 150°C would be typical.

9.2.3 Piping and Insulation

ANSI Class 300 piping should be used as a minimum requirement for hot oil systems. Gaskets in hot oil service are prone to leakage due to the low viscosity and low surface tension of the fluid at service conditions. The improved flange tightness for Class 300 systems will reduce the likelihood of leakage.

Non-absorbent (closed cell) insulation such as cellular glass or foam glass should be used at potential fluid leakage points (i.e. instrument connections, valve stems, flanges).

Figure 9.3 – Typical Two-Tier Hot Oil System



Alternate materials such as magnesia, silicate-bonded asbestos or calcium silicate may promote slow oxidation where a leakage goes undetected. The large internal surface area, poor heat dissipation and the oxidation reaction may cause significant temperature build-up within the insulation mass leading to sudden fires when the cladding is damaged or removed for maintenance.

Adequate drains must be provided on the hot oil system piping to ensure that the system is substantially water-free following hydrotest and flushing.

Where high points exist within the piping, provision of a vent valve for the controlled bleeding of locked vapour (steam) may prove useful whilst the system is being dried during commissioning. However, vapour bleeding through a local vent at an elevated point of the rack will be both impractical and extremely unsafe. Therefore, where a maintenance vent is provided, it should be piped to grade terminating at a height suitable for drainage into a small drum. The valve should be located at a grade-accessible height. Venting through this connection must be executed as a gradual (slow) operation; a procedure outlining safe practise should be included in the operating manual.

9.2.4 Equipment

9.2.4.1 Expansion Drum

A horizontal expansion drum should be provided on the pump suction to allow for hot oil expansion as it is heated from ambient (filling) conditions to the operating temperature. The cold level should be set at 25% of the diameter or 300 mm from the inside bottom, which ever is higher. Expansion of the entire system when heated from minimum ambient to the normal supply temperature should result in a level no greater than 80% (of the diameter) in the drum.

The height of the drum should be set to meet the NPSH requirements of the hot oil pumps.

The drum should be blanketed with dry, sweet fuel gas or nitrogen to prevent contact with air and the associated risk of oxidation. Provision of a split range controller or a set of opposed pressure regulators, with out-breathing to the flare system, is recommended.

Mineral based hot oils such as BP Transcal N have very low vapour pressures, even at circa 280℃. The expansion drum can therefore operate at or near to atmospheric pressure without boiling off light ends.

Where possible, the hot oil expansion drum should be highest point in the system. This will enable positive water removal (as steam) during commissioning. However, it is noted that for onshore applications - with piping at an elevated level within the rack – this may not be practicable.

9.2.4.2 Pumps

Centrifugal pumps are generally used in hot oil service.

The low viscosity and low surface tension of hot oils at service conditions make appropriate seal selection a challenge. Magnetic drive pumps, which do not require shaft sealing, are widely applied for this reason.

As a recirculating system, the pump head must be sufficient to overcome piping frictional losses, control valve differentials and the heater loss. Pump differentials of 7 – 10 bar are typical.

A start-up case should be included on the pump datasheet also, to ensure that the system can be started at low ambient temperature/high viscosity conditions.

The pumps shall be provided with y-type suction strainers to remove mill scale and other debris.

Refer to Section 6 for general recommendations on pump sizing and selection.

9.2.4.3 Filter

A slipstream filter should be provided to remove mill scale and other particulates from the hot oil system on a continuous basis.

10% of the total system circulation rate should be filtered. The filtered stream should be returned to the pump suction; this recirculating flow must be considered in sizing of the hot oil pumps.

A globe valve should be provided to manually regulate the filter rate. Provision of a local flow indicator is also recommended.

9.2.4.4 Trim Cooler

The provision of a small trim cooler is recommended for system start-up and operation at low loads where a reliable heat sink will not be available. Execution as an air-cooler is typical. As trim coolers in this service tend to be relatively small it may be possible to combine this bundle with that of another heat exchanger in a common bay.

The design duty of the trim cooler should be large enough to dissipate the minimum fired heater duty / WHRU heat leakage rate.

Volumetrically the trim cooler should be sized for the hot oil pump or the heat source minimum flow (fired heater or WHRU), whichever is larger.

In amine systems, it may be possible to utilise the amine reboiler as the trim cooler as long as the amine stripper overheads condenser is adequately sized to reject the trim load.

9.2.4.5 Heat Source

Fired Heaters

Fired heaters (furnaces) are widely applied in hot oil heating applications.

Design of the fired heater should ensure that the heat transfer fluid will not be subjected to temperatures in excess of the maximum allowable film temperature or the maximum allowable bulk fluid temperature. For BP Transcal N these temperatures are 340℃ and 320℃ respectively. Although typical of mineral oils, the limits of the selected fluid shall be considered.

A relatively high differential pressure allowance should be made for the hot oil fired heater; 200 - 300 kPa is reasonable. This will allow the equipment supplier to maximise velocities within the tubes, which improves heat transfer and reduces the film (skin) temperatures.

A margin of 10% should be applied on the total flow and duty of the hot oil heater. Where this margin is applied and the piping is insulated, heat losses to the environment can be ignored.

High film and bulk fluid temperatures are likely to be a concern at low flow conditions (low tube wall shear) when minimum firing is reached. The equipment supplier should therefore advise both the minimum duty and minimum allowable flow through the heater. The system design shall make provision for maintenance of this minimum flow when the process demand may be lower still (i.e. minimum flow bypass).

Fired heaters are generally impractical at less than 200 kW [Ref 25]. Therefore if the system has a load in this order, the following alternatives should be reviewed:

- Replacement of the hot oil fired heater with an electric heater;
- The use of individual electric heaters where this can be executed safely (i.e. fuel gas heater, TEG reboiler, etc); or
- The use of individual fire tubes for process heating where this can be safely executed (i.e. TEG reboiler, tank heaters, etc).

Waste Heat Recovery Units

The recovery of waste heat from gas turbine exhaust streams is an acceptable and safe means of heating hot oils and improving plant efficiency.

Temperature control of a gas turbine waste heat recovery unit (WHRU) outlet is generally achieved through provision of a full bundle bypass which is integral to the unit. Actuated dampers control the fraction of exhaust flow passing through the heat recovery bundle. The bypass should be sized for 100% of the exhaust flow to enable complete turndown. However, even in this condition, heat leak into the bundle is likely. Based on the design duty, the process engineer should conservatively assume 5% heat in-leak for preliminary design purposes. Advise should however be sought from the supplier before finalising the system design.

Gas turbine exhaust streams are typically in the range of 450 – 550°C at the design point. With the turbine operating at near to full load at high ambient temperatures the quantity of recoverable waste heat should generally be equal to or marginally greater than the shaft load (i.e. compression power or electrical demand for a GTG). However, the amount of recoverable heat reduces dramatically at:

- Low ambient conditions, due to an associated reduction in the exhaust temperature; and
- Part load or turndown conditions.

The process engineer must review operation at these conditions when selecting a hot oil system configuration. Where the waste heat demand represents a significant portion of the driven equipment (shaft) power, then a deficit can easily be experienced at low ambient temperatures. Where a deficit exists, provision of supplementary firing on the waste heat recovery unit or a separate fired heat to cover the peak heating deficit should be considered. A separate fired heat may offer advantages in terms of availability (as a supplementary-fired WHRU can seldom be operated whilst the associated turbine is maintained), operability (particularly at start-up) and flexibility. However, this is likely to come at greater capital expense.

WHRU exhaust-side pressure losses should be limited to 200 mm of water column. Excessive differentials in the exhaust system will reduce the turbine output power.

A relatively high oil-side differential pressure allowance should however be made; 200 - 300 kPa is reasonable. This will allow the equipment supplier to maximise velocities within the tubes, which improves heat transfer and reduces the film (skin) temperatures.

To maximise heat recovery, exhaust temperatures can generally be drawn down to within 30°C of the returning oil temperature. The return oil temperature should not be lower than 135°C to prevent corrosive water condensing in the exhaust system[6].

[6] Corrosive due to the presence of CO_2 and possibly H_2S and SO_2 in the exhaust.

A margin of 10% should be applied to the total flow and duty of the waste heat recovery unit. Where this margin is applied and the piping is insulated, heat losses to the environment can be ignored.

The process engineer should specify the minimum hot oil flow (turndown) on the WHRU datasheet. In many cases, this flow will be set by the minimum stable flow of the single operating hot oil pump. Where the minimum stable pump flow is not known, 50% of the rated volumetric flow can be taken as a preliminary figure. If a pump configuration other than 2 x 100% is proposed, then turndown to lower flows will be possible (i.e. by using one of 3 x 50% pumps only). If the process is capable of turndown below the stable pump flow, a controlled minimum flow bypass will be required. Design of the WHRU should ensure that the heat transfer fluid will not be subjected to temperatures in excess of the maximum allowable film temperature or the maximum allowable bulk fluid temperature in any condition, including the nominated turndown.

The following disadvantages are noted with regard to the use of waste heat recovery in hot oil systems:

- Where a single fired heater would otherwise satisfy plant availability targets, if major turbomachinery is spared then multiple waste heat recovery units will be required. The Lakshmi Field Development, for example, included five onshore gas turbines (3 x 50% compressors and 2 x 100% gas turbine generators) but the hot oil was heated by an unspared fired heater to minimise cost and complexity.
- Items of turbomachinery, particular compressors, can often be run at artificially high loads to generate additional waste heat. Gas turbines consume fuel gas with mechanical efficiencies between 25 and 35%, with the remainder of energy dissipated as heat, noise and other losses. Typically no greater than 60 - 65% of the fuel gas heating value appears as heat energy in the exhaust. Fired heaters, on the other hand, display thermal efficiencies in the order of 80 - 85%. A reliance on artificial compressor loading is therefore a less efficient use of fuel gas than supplementary heating.

9.2.4.6 Heat Consumers

When designed and operated properly, hot oil systems are considered to be a clean (low fouling) service. U-tube heat exchangers can therefore be applied where a mineral based hot oil is the tube-side fluid.

9.3 INSTRUMENT AND PLANT AIR SYSTEMS

9.3.1 General

Instrument air is required at most substantial onshore and offshore facilities for the actuation of control and emergency shutdown (ESD) valves, pressurisation of electrical instruments and the purging or cooling of essential instruments. Less frequently instrument air may be required for process use (i.e. flare ignition equipment). As instrument air is critical for the safe operation of a plant, the system should be designed to ensure that a continued supply can be maintained to essential users for a defined period when the compressors are inoperable (i.e. loss of power).

Service air is provided in a plant for air blowing purposes, operation of pneumatic equipment or tools and providing a breathable atmosphere in equipment before entry. The demand is generally intermittent. The supply of service air is not critical to the safety systems.

An instrument and service air supply system usually comprises the following elements:

- Air compressors;
- Filters and moisture separators;
- Air dryers;
- Air receiver(s); and
- Distribution piping.

Guidelines on the design and specification of air systems are given in subsequent sections.

9.3.2 Location

The air supply plant must be located in a non-hazardous area to ensure that the instrument air is free from toxic, corrosive, flammable and noxious gases. A section of the plot, located at a safe distance from the process area, should generally be set aside for utilities.

9.3.3 Air Quality

9.3.3.1 Instrument Air

Instrument air shall be free of oil and other liquids. The dried instrument air should be filtered and should contain less than 0.1 g/m^3 of solids with particle sizes less than 3 micron [Ref 3].

The water dew-point of the instrument air at operating pressure should be at least 10℃ lower than the site minimum ambient [Ref 3]. A dew-point specification of -20℃ can easily be achieved with standard dryer packages, and can be successfully applied in most applications. Vendor literature indicates that standard dryer packages are available for -40℃ and -70℃ water dew-points also, though their application would be limited to extremely cold environments.

9.3.3.2 Service (Plant) Air

There are no particular specifications for service (plant) air aside from the requirement to be generally free of particulates, moisture and oil.

9.3.4 Supply Pressure

9.3.4.1 Instrument Air

The system should be designed to supply instrument air at 7 barg to consumers in the major process area. The normal delivery pressure to the furthermost user should not be less than 5.5 barg. Header frictional losses, equipment differentials (i.e. dryers, receivers) and pressure regulator control losses should be superimposed on the above supply conditions for compressor sizing.

The available instrument air pressure should not fall below 4.2 barg at any point in the header at peak demand conditions.

9.3.4.2 Service (Plant) Air

Service air should be made available at 6 – 7 barg.

9.3.5 Air Quantity

The normal instrument air demand will comprise losses to some or all of the following elements: control valve actuators; actuated ball valves (i.e. shutdown, blowdown and sequencing valves); on-line analysers (gas chromatographs); separation gas for centrifugal compressors; (5) packaged equipment requirements; and pressurised electrical instruments (prevention of explosive atmosphere within enclosure).

The following allowances have been used on past Clough projects, and are in general alignment with industry practise. These figures can be used for preliminary estimation of the *normal* demand:

- 0.62 – 0.93 scfm (1 – 1.5 Nm^3/hr) per loop for control valves [Ref 3].
- 0.06 scfm (0.1 Nm^3/hr) for each actuated shutdown or blowdown valve (quarter-turn).
- 0.12 scfm (0.2 Nm^3/hr) for sequencing valves.
- 10 scfm (16 Nm^3/hr) can be taken as the typical demand of gas chromatographs.

Where instrument air is proposed for use as separation gas in centrifugal compressor seals, advice should be sought from the rotating equipment engineer.

The estimated air demand should be margined by 30% to allow for growth and uncertainty. The margined figure should be taken as the preliminary *normal* instrument air demand.

Most gas treatment facilities will have intermittent service (plant) air demands only. Maintenance activities requiring air, such as tool operation and cleaning, tend to be infrequent. The simultaneous requirement for tool air at multiple work-fronts in an operational gas plant is unlikely.

Where guidance is not provided, a service air demand of 70 Nm^3/hr can be assumed for operation of air-driven tools. This should allow for simultaneous operation of a blow gun (20 Nm^3/hr) and at least one of the following: drills and hammers (12 – 80 Nm^3/hr); 3” to 8” grinders (20 – 85 Nm^3/hr) and wrenches (30 – 50 Nm^3/hr).

If service air is required as the motive fluid for air-driven pumps (or similar intermittent users) then the largest of these should be taken as coincident with the 70 Nm^3/hr for plant maintenance. A relatively small diaphragm pump, for example, sized for 10 m^3/hr at 10 meters of head would require around 30 Nm^3/hr of plant air.

Where it is proposed that plant air be used as start gas for a diesel or gas engine (i.e. offshore crane) this requirement must be considered in design of the system. The start gas will be a high volume, short duration requirement; advice should be sought by the mechanical engineering group. The start gas demand may be the sizing case for the plant air receiver.

9.3.6 Design and Specification

9.3.6.1 System Configuration

Air supply systems are frequently purchased from packaged equipment suppliers, who can provide largely pre-engineered systems with standardised elements.

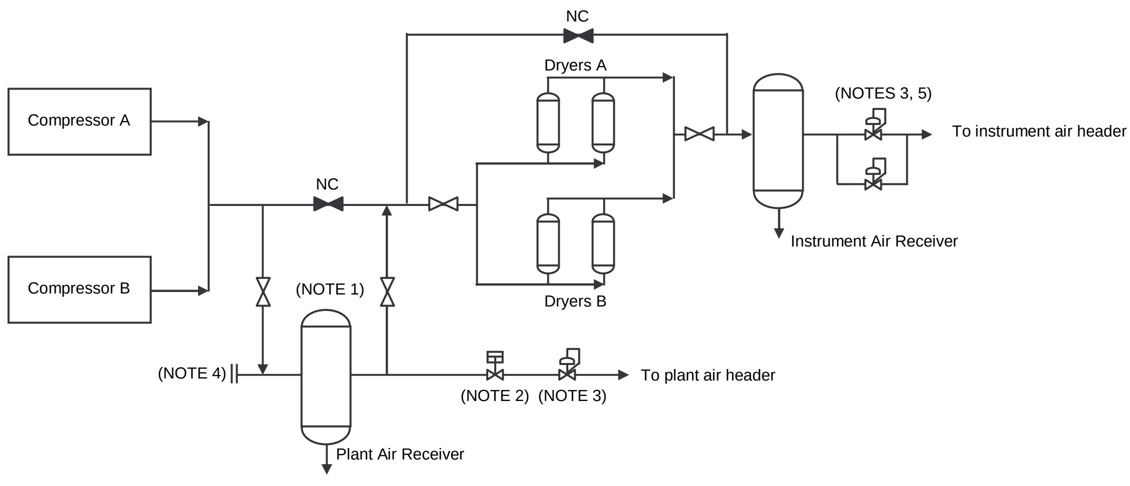
A typical scope of supply would include:

- The lead/lag air compressors complete with inlet filtration, aftercoolers and moisture separation;
- Air dryers complete with interconnecting piping, sequencing valves, after filters and the first-fill of desiccant; and
- Compressor and adsorber controls (suitable for stand-alone operation).

Depending on project drivers and procurement strategies, it may be desirable to source the receiver(s) from the package supplier also.

A typical combined instrument and plant air system configuration is shown below:

Figure 9.4 – Typical Air Plant Configuration



Notes:
1. This particular configuration uses the plant air receiver as an additional moisture separator prior to drying.
2. A shutdown valve is fitted to the plant air distribution line to ensure that all compressed air is available for instrumentation in emergency situations.
3. Optional PCVs installed on both the plant air and instrument air lines to enable higher pressure operation of the receivers (thereby reducing the buffer volume).
4. Connection for temporary compressor (to supplement inst. or plant air demand).
5. On-line spare provided for instrument air PCV.

9.3.6.2 Air Compressors

To ensure maximum reliability for the instrument air supply, at least 2 x 100% compressors should be installed. Each compressor should be sized to deliver the margined *normal* instrument air demand (refer to Section 9.3.5) plus the regeneration air requirements of the air dryers (Section 9.3.6.4). If the normal service air demand is greater than - but not significantly greater than – the instrument air demand, each compressor should be sized for this instead. If the plant air demand is significantly higher then provision of a single dedicated compressor (temporary or permanent) should be considered.

Oil injected rotary screw compressors are preferred for flowrates up to around 2500 Nm^3/hr. Above this air rate the larger and more reliable oil-free centrifugal air compressor is usually more economical.

Centrifugal air compressors should be provided with discharge non-return valves.

The vast majority of relatively small air systems will employ electrically driven compressors. Both compressors should be connected to the emergency section of the bus bar.

The air inlets for the compressors should be fitted with a wire mesh and a dust filter to prevent solids entry, which may cause mechanical damage to the units. In locations where freezing is likely, ice formation at the inlet should be prevented.

Aftercooling of the compressed air stream, usually with air, should be provided. An approach of 10°C to the maximum ambient temperature should be used for design purposes.

9.3.6.3 Moisture Separation

To remove free water from the compressed air, water separators should be installed downstream of the aftercooler of each compressor. The design of the moisture separator should be such that the outlet air is not more than 5% oversaturated under all operation conditions [Ref 3].

9.3.6.4 Air Dryers

The desiccant used in the adsorption vessels will usually be either alumina or silica gel in bead form. If silica gel is used, a bottom layer of approximately 10% activated alumina can be provided to achieve better resistance to entrained water (assuming that gas enters at the bottom of the dryers).

For most relatively small instrument air systems, pressure swing (heatless) regeneration should be employed. Two sets of dryers should be provided, each consisting of one duty and one regenerating adsorber. Each set of dryers should have the same capacity as stated for the individual compressors. The specification should clearly indicate that the stated air capacity is purchaser's demand only, and that regeneration air losses are to be accounted for by supplier in design of the dryers (and compressors).

For preliminary estimation, pressure swing (heatless) dryers typically consume 10-15% of their design throughput as regeneration gas.

For larger air supply plants, the losses can be reduced to 2 - 3% by regenerating the beds at an elevated temperature with dry, heated air at atmospheric pressure.

Design of the beds should consider 5% oversaturation of the wet compressed air [Ref 3]. The beds shall be capable of the meeting the specifications given in Section 9.3.3.1 under all conditions.

For heatless dryers, timed switching at 5 minute intervals is typical. When regeneration at atmospheric pressure is used, the adsorption vessel should be depressurised slowly through a muffler to prevent fragmentation of the bed and to reduce exhaust noise. More frequent bed switching should be avoided to prevent excessive valve wear.

2 x 100% after filters should be installed to prevent desiccant particles from entering the instrument air supply piping (refer to Section 9.3.3.1). Consideration may be given to installing filters upstream of the adsorption beds to prevent rust settling on the desiccant.

Where a piping low point exists between the compressor aftercooler/moisture separator and the desiccant beds, an automated drain valve should be provided. This valve should drain collected water as part of the regeneration sequence.

9.3.6.5 Air Receivers

The instrument air receiver should be sized to provide the more stringent of the following buffer volumes:

1. 40 minutes at the *normal* consumption rate, or
2. 20 minutes at the *peak* consumption rate.

The first criteria should be employed for preliminary estimation. When the project moves into a more advanced phase and an estimate of the peak consumption rate is available, sizing of the receiver should be revisited.

Where the receiver operates at the header pressure, the buffer volume should be between the range of supply pressures given in Section 9.3.4.1 with an allowance for frictional losses, i.e. between (7 + 1 =) 8 barg and (5.5 + 1 =) 6.5 barg.

If pressure regulation is provided downstream of the air receiver to enable higher pressure operation of the vessel (reducing its volume) then the buffer allowances should be between:

(a) The start pressure of the lead compressor, and

(b) 5.5 barg plus frictional header losses plus the full-open differential of the regulator.

By using the start pressure of the lead compressor for buffer volume calculations, normal operation of the plant and the instrument air system is assumed up to the point of losing all electrical power (or an equivalent emergency condition). This is seen as a reasonable basis for sizing.

Where the instrument air pressure is regulated downwards from a higher receiver operating pressure, two on-line regulators should be provided. The set-point of the second regulator should be set slightly below the first, say 0.5 – 0.7 bar.

A buffer volume of circa 10 minutes at the nominated supply rate should be provided in the plant air receiver (refer to Section 9.3.5).

Both vessels should be provided with facilities for manual drainage to a suitable location.

9.3.6.6 Distribution

Instrument air and plant air should be distributed throughout the plant using dedicated headers; interconnections between these headers should not be made.

Sizing of the instrument air header should limit to the frictional losses to less than 10 kPa/100 meters (refer to Section 3.2.4.1), but should also ensure that the losses do not result in an unworkably low supply pressure to the most remote users during *normal* and *peak* supply conditions (refer to Section 9.3.4.1).

Critical consumers that must stay in operation after a total air supply failure (i.e. blowdown valves where depressurisation is staged) should be provided with a secure supply system. This usually consists of a small local buffer vessel connected to the distribution piping via a non-return valve. Buffer vessels are typically sized to provide sufficient air for operating the device, i.e. stroking the valve, say, 3 - 5 times.

9.3.6.7 System Control

The air plant should be provided with on-skid control for both the compressors and sequencing of the dryers. The unit shall operate independently of purchaser's control system but should be capable of interfacing with the DCS, communicating 'unit running', 'unit fault' or similar status and health signals.

The air compressors should be configured as lead/lag units. This will make the capacity of the second compressor – also sized for 100% of the normal demand - available as an on-line supplement during peak demand situations. During normal operation the lead compressor will run intermittently to maintain the air pressure between its start and stop pressures; the lag compressor will remain in an unloaded condition.

The following start and stop pressures for a lead/lag arrangement are typical of systems where the receivers and dryers operate at the instrument air header pressure plus frictional losses [Ref 3]:

Table 9.3 – Typical Control Settings for Lead/Lag Arrangement (no pressure regulation)

Machine	Pressure to start (barg)	Pressure to stop (barg)
Duty (lead) compressor	7	8
Standby (lag) compressor	6.5	7.5

The Lakshmi Field Development employed pressure regulators downstream of the instrument air receiver to enable higher pressure operation, leading to a reduction in the volume of the buffer vessel. Two regulators were installed to provide on-line sparing, with the first set at 7.5 barg and the second at 6.2 barg. The following control settings were employed for sequencing of the lead/lag compressors:

Table 9.4 – Example Control Settings for Lead/Lag Units (pressure regulation used)

Machine	Pressure to start (barg)	Pressure to stop (barg)
Duty (lead) compressor	10.6	11.6
Standby (lag) compressor	8.6	9.6

Where the pressure transmitter used for control purposes is located immediately downstream of the compressors, the start and stop pressures must take account losses across the dryers and associated piping/valves. An alternative arrangement would see the transmitter located on the instrument air receiver instead. This requirement should be communicated to the equipment supplier on the process datasheet.

9.4 PRODUCTION DELIVERY SYSTEMS (FLOW ASSURANCE)

The propensity of a crude or condensate to form emulsions with water in production delivery systems cannot be accurately (or even approximately) estimated by calculation. This is a flow assurance issue, which is most effectively quantified by laboratory testing. For detailed design and EPC contracts, if the client has not provided a flow assurance report for upstream production (pipeline or flowline) systems, the process engineer should raise a technical clarification. This clarification should also address other flow assurance issues (i.e. asphaltene formation, wax deposition) which may affect downstream phase separability.

With varying degrees of success, these issues can all be addressed through the injection of appropriate inhibitors or suppressants. However, chemical suppliers will only provide a meaningful guarantee if:

- They are afforded an opportunity to perform laboratory tests with representative samples;

  <u>and</u>

- They are provided with a significant amount of rely-upon data collected from previous tests.

The list of physical properties, parameters and tests given as Attachment C is indicative of data required to *know* a production fluid sample sufficiently well from a flow assurance and chemical selection perspective. EPC contractors such as Clough have no way of predicting this data. Where a contract calls for production issues to be addressed, this information must be supplied by our clients as part of the bid package. If this data cannot be made available, Clough should not take on production system risk as we can offer no greater assurance than the heavily qualified advice offered by our chemical supplier(s).

Unforseen or unquantifiable production issues will manifest in poor downstream phase separation and a range of other production issues which may limit raw fluid delivery (wax deposition requiring regular pipeline scraping, tight emulsions giving rise to high flowing viscosities, etc). These issues may compromise our ability to commission a plant in a timely manner and to achieve the required product delivery rates and specifications (i.e. BS&W).

We should establish a clear position on this issue prior to entering into a lump sum contract.

10 PRELIMINARY EFFICIENCY GUIDANCE

10.1 GENERAL

The following efficiencies and factors are provided for use by the process engineer in process-lead studies or concept selection work. The figures should generally provide a reliable basis for preliminary equipment sizing, driver selection, etc.

However, in all cases, the efficiencies should be confirmed as acceptable by equipment suppliers or discipline engineers when the project moves into a more advanced phase.

10.2 DISTILLATION DESIGN

10.2.1 System Factors

The following system factors are based on 600 mm (24 inch) tray spacings. The system factors account for the effects of foaming and derate (increase) tower area.

Table 10.1 – Fractionation System Factors

Service	System Factor (Foaming)	Max % of Flood
Amine Regenerator	0.75 – 0.85	70
Amine Contactor	0.75	70
Condensate Stabiliser (wet)	0.75 – 0.85	80 – 85
Demethaniser/De-ethaniser (top)	0.70 – 0.80	80 - 85
Demethaniser/De-ethaniser (bottom)	0.90 – 1.00	80 - 85
Depropaniser	0.80	80 - 85
Debutaniser	1.00	80 - 85
Glycol Contactor (trayed)	0.50 – 0.65	70

10.2.2 Tray Efficiencies

The tray efficiencies given below are expressed as the ratio between the number theoretical and actual trays.

Table 10.2 – Fractionation Tray Efficiencies

Service	Tray Efficiency (%)	No. Actual Trays (typical)
Condensate Stabiliser (wet, top feed)	40 – 60	16 – 24
Demethaniser	45 – 60	18 – 26
De-ethaniser	50 – 70	25 - 35
Depropaniser	80 – 90	30 – 40
Debutaniser	85 – 95	25 - 35
Glycol Contactor (trayed)	25 – 40	8 – 10

10.3 FIRED HEATERS

Fired heater thermal efficiency (based on fuel gas LHV):

- Cabin or cylindrical heaters 82 – 88%
- Salt bath heaters 70%

10.4 ROTATING EQUIPMENT

10.4.1 Process Efficiency

Centrifugal compressor adiabatic efficiency:	68% preliminary (typically 65 – 75%)
Turbo-expander adiabatic efficiency:	83% preliminary (typically 80 – 85%)
Screw compressor efficiency:	70 – 82%
Reciprocating compressor efficiency:	78 – 88%
Centrifugal pump efficiency (per stage):	65% preliminary (typically 65 – 80%)

10.4.2 Driver Efficiencies

10.4.2.1Gas Turbines

Driver efficiencies are based on the consumption of fuel gas with lower heating values in the range 31 – 43 MJ/Nm^3 (800 – 1100 BTU/scf)

Table 10.3 – Typical Gas Turbine Fuel Efficiencies

ISO Power Rating (MW)	Typical model in range	Fuel Efficiency (%)
0.8 – 1.1	Solar Saturn 20	23 - 25
3.0 – 4.1	Solar Centaur 40	25 - 28
5.7 – 6.4	Solar Taurus 60	30 - 32
9.0 – 10.5	Solar Mars 90	31 - 33
15.0 – 16.5	Solar Titan 130	33 - 36
22.0 – 26.3	Nuovo Pignone PGT25	34 - 37

The turbine ratings given above are at ISO conditions (60°F and sea level) and do not account for ambient effects, mechanical losses (i.e. gear box), power losses due to mechanical wear/fouling or air inlet/exhaust losses. For a given application, the ISO turbine rating must be derated for each of the factors given here.

10.4.2.2Naturally Aspirated Gas Engines

Table 10.4 – Typical Gas Engine Fuel Efficiencies

Rated Output (kW)	Fuel Efficiency (%)
75 – 140	28
150 - 225	29
260 - 340	30
410 - 620	31
2000 - 3500	36

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ATTACHMENT A – STANDARD SEMI-ELIPTICAL HEAD SIZES

(internal diameter - millimetres)

286 (1)	864 (1)	1292 (1)	2896 (2)
356 (1)	914 (1)	1372 (1)	3048 (2)
508 (1)	940 (1)	1524 (1)	3353 (2)
533 (1)	965 (1)	1676 (1)	3658 (2)
610 (1)	991 (1)	1829 (1)	3692 (2)
635 (1)	1016 (1)	1981 (1)	4267 (2)
686 (1)	1067 (1)	2134 (1)	4572 (2)
711 (1)	1118 (1)	2286 (1)	
737 (1)	1137 (1)	2438 (1)	
762 (1)	1219 (1)	2591 (1)	
813 (1)	1270 (1)	2731 (1)	

Notes:

1. Hot pressed heads
2. Hot spun heads.

ATTACHMENT B – PREFERED MOTOR OUTPUTS

(expressed in kW)

1.1	18.5	110	280	450	710
1.5	22	132	300	475	750
2.2	30	150	315	500	800
4.0	37	160	335	530	850
5.5	45	185	355	560	900
7.5	55	200	375	600	950
11	75	220	400	630	1000
15	90	250	425	670	

ATTACHMENT C

TYPICAL DATA AND TESTING REQUIRED TO RELIABLY QUANTIFY AND ADDRESS FLOW ASSURANCE ISSUES IN PRODUCTION DELIVERY SYSTEMS

(a) Data required from testing of individual wells:

1. Formal PVT analysis including: the reservoir fluid composition (typical breakdown is given as (i) below); liquid density and liquid viscosity data plotted against temperature (typically given in the range of 8 – 60℃); pour point temperature.
2. Wax contents, by HT GC and reported as C20+ n-alkanes.
3. Wax appearance temperature and wax melting temperature; testing method would preferably be pressurised Modulated Differential Scanning Calorimetry (MDSC). If this test is not available then pressurised Differential Scanning Calorimetry (DSC) is generally acceptable.
4. SARA analysis to identify potential for asphaltene formation.
5. Cool-down rheological characteristics of stabilised oil at two nominated shear rates and crude cool-down viscosity curves.
6. Shear stress versus shear rate profile of stabilised oil at a fixed temperature to determine gel strength of oil.
7. Effect of pressurisation with methane (soluble gas) and nitrogen (insoluble gas) upon cool-down rheology and gel strength.
8. Identification of foaming tendency at 80℃.

(b) Properties required for flow assurance modelling and operational contingencies (i.e. chemical injection). Many of the properties given below are derived from test analysis given in (a) above:

Measured fluid properties

Crude Oil	**Gas Condensate**
Saturation pressure (bubble-point)	Saturation pressure (bubble-point)
Live oil compressibility	Compressibility at dew-point pressure
Fluid density and bubble-point pressure	Fluid density at dew-point pressure
Flash GOR	Flash GOR
Stocktank oil density (API gravity)	Condensate density (API gravity)
Flash gas composition (to C_{12}+)	Flash gas composition (to C_{12}+)
Stocktank oil composition (to C_{36}+)	Stocktank oil composition (to C_{36}+)
Reservoir fluid composition (to C_{36}+)	Reservoir fluid composition (to C_{36}+)
Flash gas molecular weight	Flash gas molecular weight
Stocktank oil molecular weight	Stocktank oil molecular weight
Reservoir fluid molecular weight	Reservoir fluid molecular weight
Reservoir fluid viscosity	

Predicted PVT properties

Crude Oil

Constant composition expansion at reservoir temperature:

- Oil compressibility
- Oil relative volume

Differential liberation at reservoir temperature:

- Oil formation volume factor
- Solution GOR
- Oil density
- Oil viscosity
- Produced gas gravity
- Produced gas Z-factor

Separator test at specified temperature and pressure:

- Oil formation volume factor
- Separator shrinkage factor
- Tank, separator and total GOR
- Separator oil density
- Stocktank oil density (API gravity)

Gas Condensate

Constant composition expansion at reservoir temperature:

- Single and two-phase relative volume
- Single and two-phase fluid compressibility
- Retrograde liquid deposit

Constant volume depletion at reservoir temperature:

- Cumulative produced fluid to depletion
- Single and two-phase fluid compressibility
- Retrograde liquid deposit
- Produced gas gravity

(c) Additional testing required by injection chemical supplier to identify appropriate wax treatment (pour-point depressant and wax inhibition) and emulsion breaking treatment (demulsifier injection):

1. PPD selection by pour-point tests (continuation of programme carried out on first DST oil samples).
2. Effect of PPD on cool-down rheology.
3. Effect of PPD upon gel strength rheology (yield value).
4. Emulsion breaker selection trials.
5. Foam potential testing and anti-foam selection (if required).
6. Product compatibility testing.

To execute testing listed above, the injection chemical supplier would typically require 20 litres of oil from each well. The samples should be taken from upstream of the test separator using IATA approved containers (4 x 5 L containers, for example). Samples should be taken after well clean-up to minimise contamination with drilling and completion fluids and after three (3) tubing volumes have been displaced to ensure that any gelled oil present in the tubing is displaced prior to sampling.

(i) Typical hydrocarbon analysis (referenced in (a) above):

Hydrogen sulphide

Carbon Dioxide

Nitrogen

Methane

Nonanes

Decanes

Undecanes

Dodecanes

Ethane

Propane

iso-Butane

n-Butane

iso-Pentane

n-Pentane

Hexanes

Heptanes

Octanes

Tridecanes

Tetradecanes

Pentadecanes

Hexadecanes

Heptadecanes

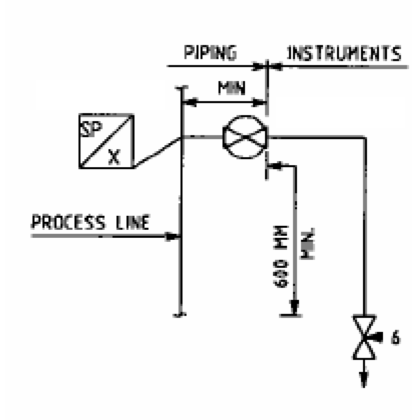
Octadecanes

Nonadecanes

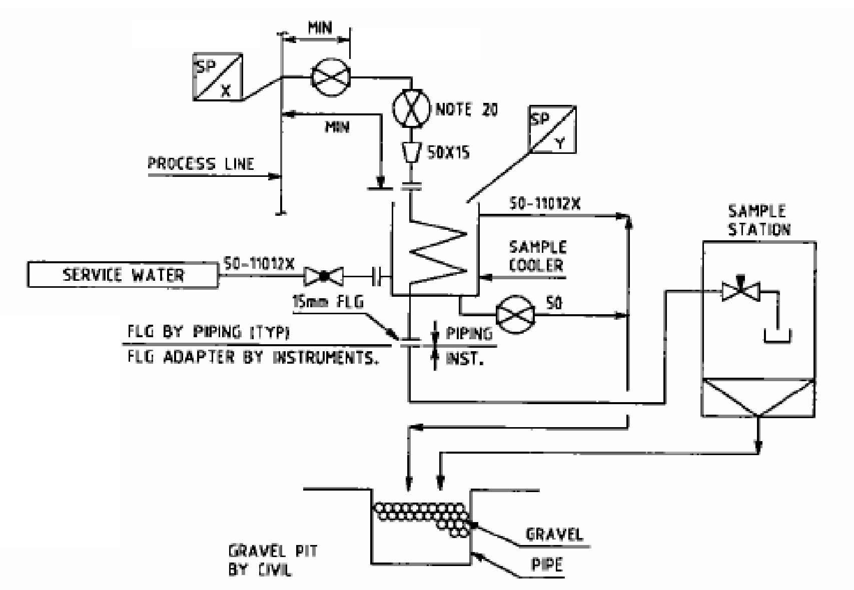
Eicosanes plus

ATTACHMENT D – EXAMPLE SAMPLE STATIONS

Type I - **For liquid samples below 65°C with atm. boiling temperatures above the sample temperature for all compositions.**



Type II - **For liquid samples which would be above 65°C without cooling and with atm. boiling temperatures above the sample temperature for all possible compositions.**



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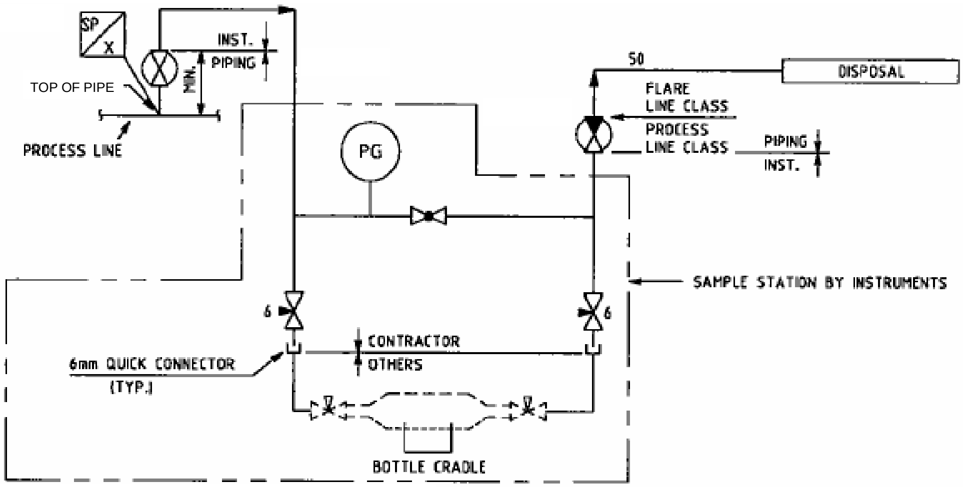
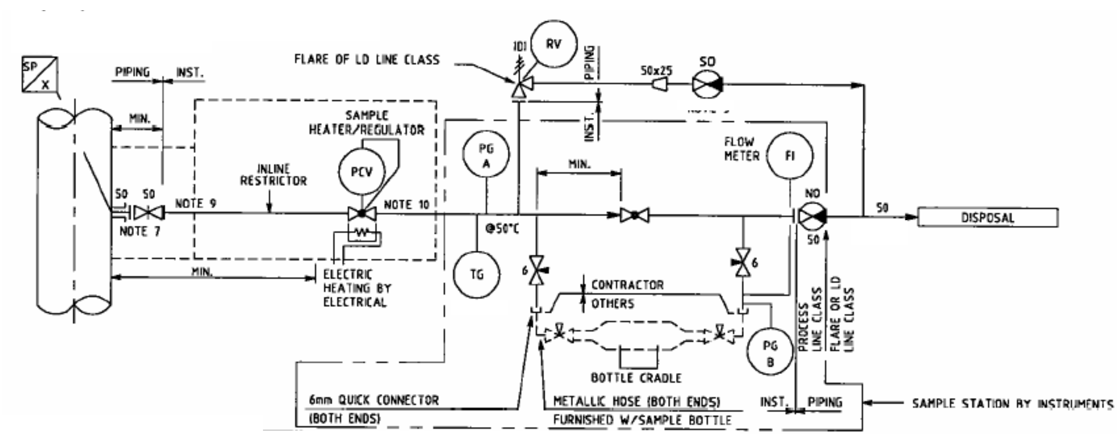
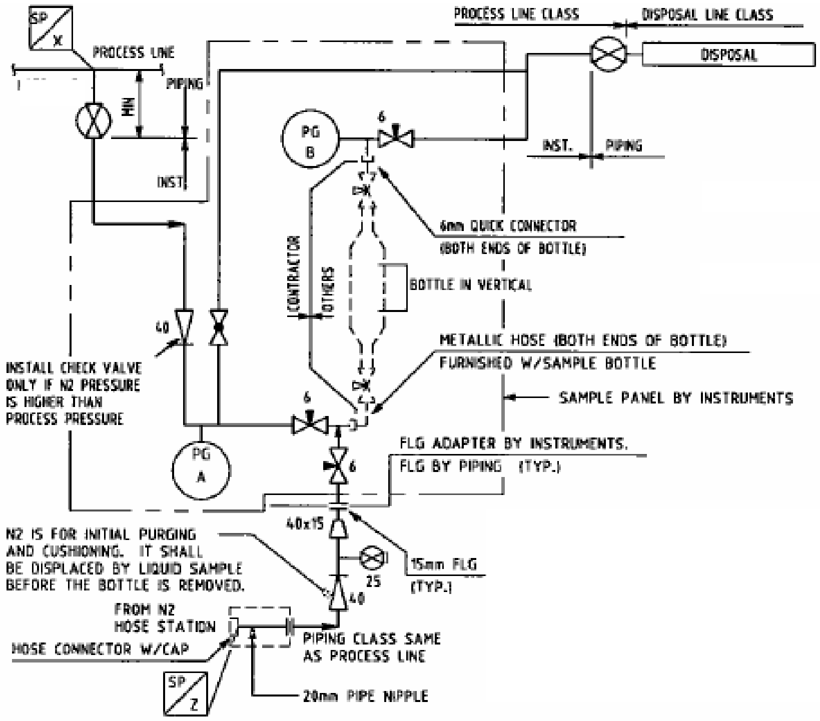
Type III - **For hydrocarbon vapours**



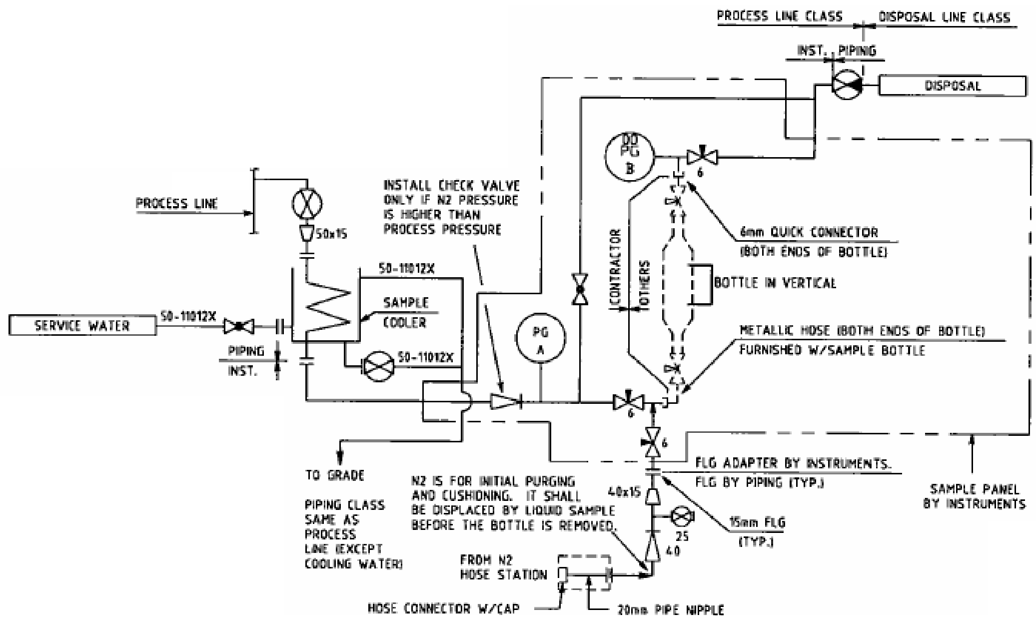
Type IV - **For light hydrocarbon liquids (C2, C3) at lower than ambient temperature**



Type V - **For liquid samples below 65℃ with atm. boiling temperatures below the sample temperature**



Type VI - **For liquid samples which would be above 65℃ without cooling and with atm. boiling temperatures below the sample temperature**



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