MANUAL

GAS/LIQUID SEPARATORS - TYPE SELECTION AND DESIGN RULES

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APPENDICES

1. INTRODUCTION

1.1 SCOPE

This DEP specifies requirements and gives recommendations for the selection and design of gas/liquid separators.

Design rules for the following types of separators are given in (Section 3):

- Knock-out drum (vertical and horizontal separator)
- Wiremesh demister (vertical and horizontal separator)
- Vane-type demister (vertical and horizontal separator)
- Separators of SMS family (SMS,SVS,SMSM)
- Cyclone with tangential inlet (conventional cyclone)
- Cyclone with straight inlet and swirler ("Gasunie" cyclone)
- Vertical multicyclone separator with reversed-flow multicyclone bundle (conventional multicyclone separator)
- Separator with axial-flow multicyclone bundle (with recirculation of secondary gas; vertical and horizontal separator)
- Filter separator (horizontal two-stage separator)

NOTE The design of gas/liquid/liquid three-phase separators is excluded from the scope of this DEP; for this subject, DEP 31.22.05.12-Gen. should be consulted.

Users of this DEP should first read Section 2 ("Selection Criteria for Gas/Liquid Separators") to familiarise themselves with the general design philosophy and the characteristics of the various separators. After selection of the desired separator the design rules can be obtained from Section 3.

Further guidance for debottlenecking of existing separators is given in Appendix XI.

1.2 CHANGES SINCE PREVIOUS EDITION

This DEP is a revision of the DEP with the same number dated February 1991. A summary of the main changes since the previous edition is given below:

Addition of the following separators:

- cyclone with straight inlet and swirler ("Gasunie" cyclone);
- separator with axial-flow multicyclone bundle (with recirculation of secondary gas; vertical and horizontal separator).

Knock-out drums:

- Change of minimum gas cap height from 0.4 m to 0.3 m in horizontal knock-out drums;
- Design rules for vertical knock-out drum fitted with a Schoepentoeter;
- Design rules for flare knockout drums.

SMS, SVS and SMSM-separators:

- Modification of the liquid handling capacity criterion;
- Level control:
  - minimum value of LA(L) - LZA(LL) 0.10 rather than 0.19 m;
  - minimum value of LA(H) - LA(L) 0.35 rather than 0.36 m;
  - specification of hold-up times for control brought in line with latest SIOP-philosophy;
  - slug volumes are based on **maximum** flow conditions or have a design margin

included.

- incorporation of vessel head in control volume calculations.

Appendices

- schoepentoeter:
  - for R_v = 300 mm, calculation of W_{vo} ($40 \leq W_{vo} \leq 80$ mm)
  - in horizontal vessel: $3d_1 \leq E \leq 5d_1$
- a two-phase flowmap for vertical feedpipes (upflow) is now included in addition to a two-phase flowmap for horizontal feedpipes.
- an Appendix has been added which gives the chord length, chord area and vessel head volume relationships for control volume calculations.

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The word **shall** indicates a requirement.

The word **should** indicates a recommendation.

1.5 CROSS-REFERENCES

Where cross-references are made, the number of the section or sub-section referred to is shown in brackets. All documents referenced in this DEP are listed in (5.).

1.6 SYMBOLS AND ABBREVIATIONS

Unless explicitly stated otherwise, all symbols used in this DEP are expressed in the units given below:

A	area	m^2
Ar	Archimedes number:	

	$Ar = (\rho_L / \eta_L^2)\sqrt{\sigma^3 / \{g(\rho_L - \rho_G)\}}$	-
a	length of side, square cyclone inlet (Type 1)	m
C	specific heat	J/kg.K
D	internal diameter of vessel or large pipe	m
d	diameter of small pipe, nozzle, bubble or droplet	m
E	available clearance for Schoepentoeter	m
Eff	efficiency: $Eff = (Q_{L,out}/Q_{L,in}) * 100$	%
Fr_G	gas Froude number:	
	$Fr_G = v_G \sqrt{\rho_G / \{(\rho_L - \rho_G)gD\}}$	-
Fr_L	liquid Froude number:	
	$Fr_L = v_L \sqrt{\rho_L / \{(\rho_L - \rho_G)gD\}}$	-
f	derating factor	-
GOR	gas / oil ratio	N(ormal)m^3/m^3
g	acceleration due to gravity	m/s^2
H	height	m
h	height of vessel for liquid hold-up (to LZA(HH))	m
K	friction loss factor	-
L	length	m
LA(H)	high level pre-alarm	
LA(L)	low level pre-alarm	
LZA(HH)	high level trip	
LZA(LL)	low level trip	
M	mass flow rate	kg/s
max(x,y)	maximum of the values x and y	
MW	molecular weight	kg/kmol
NFA	net free area	-
NL	normal level	
n	number of vanes (Schoepentoeters), cyclones, etc.	-
P	pressure	Pa
p	pressure	Pa
Q	volumetric flow rate	m^3/s
R	radius of vane (Schoepentoeters)	m
R	cyclone scroll	m
R	gas constant (R = 8314 J/kmol.K)	J/kmol.K
s	width of split between cyclone bottom plate and wall	m
SMS	Schoepentoeter Mistmat Swirldeck	
SMSM	Schoepentoeter Mistmat Swirldeck Mistmat	
SVS	Schoepentoeter Vane Swirldeck	
T	absolute temperature	K

t	thickness	m
V	volume	m^3
v	velocity	m/s
W	width	m
w	width	m
X	clearance or distance	m
x	cyclone pressure drop coefficient	-
y	cyclone pressure drop coefficient	-

Greek symbols:

α	vane angle (Schoepentoeter)	degrees
α	ratio of the short and long axes of the vessel head	-
β	edge angle (Schoepentoeter)	degrees
ε	porosity (of wiremesh)	-
η	dynamic viscosity	Pa.s
κ	ratio of the specific heats (C_p/C_v)	-
λ	gas load factor:	
	$\lambda = (Q_G / A_G)\sqrt{\rho_G / (\rho_L - \rho_G)}$	m/s
ρ	density	kg/m^3
σ	gas/liquid interfacial tension	N/m
ϕ	flow parameter:	
	$\phi = Q_L / Q_G \sqrt{\rho_L / \rho_G}$	-

Subscripts:

c	cyclone
cf	filter part of candle (in filter separators)
crit50	related to droplet with 50% chance of removal in G/L separator
crit99	related to droplet with 99% chance of removal in G/L separator
ct	candle tube (in filter separators)
dp	drain pipe
feed	related to feed flow
fp	related to feed pipe
G	gas
hd	header
HH_H	related to control band between LZA(HH) and LA(H)
H_L	related to control band between LA(H) and LA(L)
in	related to inlet
L	liquid
L_LL	related to control band between LA(L) and LZA(LL)
m	demister (either vane pack or mistmat)
m	mixture

max	maximum
min	minimum
noz	nozzle
p	at constant pressure (as in C_p)
out	related to outlet
perfpl	perforated plate
sch	related to Schoepentoeter
sd	swirldeck
sonic	related to sonic velocity
st	swirltube
v	vane
v	at constant volume (as in C_v)
vb	vane box
vfb	related to distance between bottom plate and vortex finder in cyclones
ves	related to vessel
vo	vane entrance opening (in Schoepentoeters)
w	wire (of wiremesh)
wm	wiremesh
0	related to outside (nominal) diameter of inlet nozzle
1	related to feed inlet
2	related to gas outlet
3	related to liquid outlet
4	related to diameter of drip ring in cyclones
	(numbers 1 to 5 inclusive are also related to important distances/clearances in the separator vessels (see the individual layout drawings))
η	related to dynamic viscosity of liquid
ϕ	related to flow parameter

Superscript:

*	density correction (e.g. in Q^*_{max})

$$Q^*_{max} = Q_{G,max}\sqrt{\rho_G / (\rho_L - \rho_G)} \qquad m^3 / s$$

2. SELECTION CRITERIA FOR GAS/LIQUID SEPARATORS

This Section outlines various criteria and features which play a role in separator performance and selection. Table 1, at the end of this Section, summarises the relative performance of various types of separator.

2.1 DUTY

It is often necessary to separate liquid and gas phases in a certain stage of an operation or process. Since both the conditions of the wet gas stream (or more generally the gas/liquid stream) and the required efficiency may vary widely, care shall be taken in selecting a separator in order to match the specific duty. For instance, a gas/liquid separator upstream of a gas compressor would need to be very efficient, whereas in other cases a simple knock-out vessel may be sufficient if only bulk separation of the gas and liquid phases is required (e.g. upstream of a heat exchanger in which condensation will take place).

2.2 ORIENTATION

For gas/liquid separation, a vertical vessel should normally be selected for the following reasons:

- a smaller plan area is required (critical on offshore platforms);
- easier solids removal;
- liquid removal efficiency does not vary with liquid level (area in vessel available for gas flow remains constant);
- vessel volume is generally smaller.

However, a horizontal vessel should be chosen if:

- large liquid slugs have to be accommodated;
- there is restricted head room;
- a low downward liquid velocity is required (for de-gassing purposes or for foam breakdown).

2.3 COMPONENTS

Following the gas/liquid flow path through the separator, the following components are identified.

2.3.1 Feed inlet

This comprises the upstream piping, inlet nozzle and inlet device (if any).

Detailed piping requirements are given in (4.).
The diameter of the inlet nozzle is a function of the feed flow rate and pressure.
Information on the nature of the feed (two-phase flow regime, maximum drop size, foaminess, etc) is given in Appendix I.
The criterion for nozzle sizing is that the momentum of the feed shall not exceed prescribed levels. The maximum allowable inlet momentum can be increased by fitting inlet devices.
The momentum criteria are given in Appendix II.
The function of the inlet device is to initiate gas/liquid separation and distribute the gas flow evenly in the gas compartment of the vessel.
Commonly used inlet devices are the half-open pipe and the Schoepentoeter. Rules for the design of the Schoepentoeter are given in Appendix III.

2.3.2 Separator internals

After passage through the feed inlet, the gas stream will usually still contain liquid in the form of droplets. Normally a separation internal is required in the vessel to complete the separation process. Only when moderate separation efficiency is required and the droplets are relatively large can the separation internal be omitted. By selection of a sufficiently large vessel diameter the gas velocity in the vessel can be kept low and the majority of the droplets will settle by gravity. This separation mechanism is used in knock-out vessels.

Separation internals are required in all other types of gas/liquid separators. The choice of internal depends on the required duty.

Options include:

wiremesh
vane pack (either for horizontal or vertical flow)
(multi)cyclones - axial flow (e.g. "swirltubes")
- reversed flow (tangential)
filter candles

A combination can also be used, such as:

wiremesh + swirltubes (in SMS)
wiremesh + swirltubes + wiremesh (in SMSM)
vertical-flow vane + swirltubes (in SVS)
filter candles + vane pack or wiremesh (in two-stage filter separator)

Selection depends on the required efficiency, capacity, turndown, maximum allowable pressure drop and fouling tolerance.

2.3.3 Gas and liquid outlets

After completion of the gas/liquid separation process the two phases will leave the vessel via the gas and liquid outlet respectively. The nozzle sizing criteria are given in Appendix II.

2.4 DESIGN BASE

2.4.1 Gas handling capacity

The separator shall be large enough to handle the gas flow rate under the most severe process conditions.
The procedure is to first determine the highest value of the volumetric gas load factor, Q^*_{max}:

$$Q^*_{max} = Q_{G,max} \sqrt{\rho_G / (\rho_L - \rho_G)}$$

$Q_{G,max}$ is the highest envisaged gas flow rate and includes a margin for surging, uncertainties in basic data, etc. and is expressed in m^3/s (actual).
This margin is typically between 15 and 50%, depending on the application. For the recommended margin see Appendix IV.
ρ_G and ρ_L are the densities of the gas and liquid phase respectively (kg/m^3). If two immiscible liquids are present in the feed and the flow rate of the lower density liquid is at least 5% vol. of the total liquid flow rate, then the density of the lighter liquid shall be used in the above formula.

The minimum required vessel cross-sectional area for gas flow, $A_{G.min}$, is determined by the following formula:

$$A_{G,min} = Q^*_{max} / \lambda_{max}$$

where λ_{max} is the maximum allowable gas load factor, which is a measure of the gas handling capacity of the selected separator.

In a vertical vessel $A_{G,min}$ is the cross-sectional area of the vessel. When a wiremesh is used, $A_{G,min}$ is the cross-sectional area of the wiremesh which can be much smaller than the vessel's cross-sectional area. In a horizontal vessel it is the cross-sectional area above the highest liquid level LZA(HH) - see Appendix V.

If a horizontal-flow vane pack is used, a second gas load factor is used for the sizing of the separator internal. In that case, $A_{G,min}$ is not related to the vessel but is the minimum required vane face area. (In a vertical-flow vane pack or wiremesh occupying the total cross-section of the vessel the two load factors are identical). For more details on this load factor, see (3.5.3).

2.4.2 Flow parameter

Another commonly applied criterion for separator design is the flow parameter, ϕ, used to characterise the type of gas/liquid feed into the vessel or the wet gas stream approaching the separator internal.

$$\phi = (Q_L / Q_G)\sqrt{\rho_L / \rho_G}$$

where Q_L and Q_G are the volumetric flow of the liquid and gas phase respectively (m^3/s).

If ϕ is low, e.g. less than 0.01, the liquid will generally enter the separator as a combination of a film and droplets. Slugs would not be expected.

2.4.3 Efficiency

The efficiency of a gas/liquid separator, Eff, is normally defined as the ratio of the liquid flow rate separated from the gas stream and the liquid flow rate in the feed entering the separator, multiplied by 100.

$$Eff = (Q_{L,out} / Q_{L,feed}) * 100\%$$

It should be noted, however, that if the flow parameter of the feed is very low, i.e. less than 0.001, the efficiency of the separator as defined above could be relatively low, even though in absolute terms the liquid carry-over in the gas stream is still very small.

Under such conditions it is more meaningful to describe the carry-over in absolute terms (m^3/s or kg/s) as well as the efficiency in percentage terms.

In this manual only typical efficiencies are quoted for the various separators, since the efficiency is highly dependent on the liquid droplet size distribution and liquid load at the gas/liquid separation internal.

2.5 SELECTION STRATEGY

To facilitate the choice of a separator type for a given application, the performance characteristics of various separators are summarised in Table 1.

The separators are compared on the following points:

Gas handling capacity:
- max. capacity (gas load factor)
- turndown ratio (is ratio of maximum and minimum flow

Liquid removal efficiency:
- overall
- with respect to fine mist
- with respect to the possible flooding above the maximal load factor (flooding will cause a sharp decrease in efficiency)

Liquid handling capacity:
- slugs
- droplets (overloading of separation internal)

Fouling tolerance:
- sand
- sticky material

Pressure drop

The following selection strategy is suggested:
First define the mandatory requirements which the separator shall satisfy. With the aid of Table 1 a number of separators can then be ruled out. Check, using (2.2), whether there are limitations which will rule out horizontal or vertical vessels.
For each separator out of those remaining use the appropriate part of (3.). The first part of

each Section gives the profile of the separator (e.g. characteristics and typical process applications). Based on these profiles, a final choice of separator can be made.

Table 1: Performance comparison of the various G/L separators

	VKO	HKO	VW	HW	VV1	VV2	HV	SMS	SVS	SMSM	CT	CS	VRMC	VAMC	HAMC	HTFS
Gas handling																
max. capacity (λ)	B	B	C	C	D	D	D	E	E	E	E	F	E	F	F	B
turndown (max/min flow)	∞	∞	4	4	3	3	3	10	4	10	2	7	2	>6	>6	∞
Liquid removal efficiency																
overall,%	90	90	> 98	> 98	> 96	> 96	> 96	> 98	> 96	> 99	> 96	> 99	> 93	> 99	> 99	> 99
with respect to fine mist	A	A	E	E	C	C	C	E	D	F	B	E	B	E	E	F
flooding above λ_{max} (Y/N)	N	N	Y	Y	*	*	*	N	N	N	N	N	Y	N	N	Y
Liquid handling capacity																
as slugs	D	E	D	E	A	D	E	D	D	D	D	D	B	D	E	B
as droplets ($Q_{L,max}$)	D	D	D	D	B	C	C	D	D	D	D	D	B	D	D	B
Fouling tolerance																
sand	E	E	B	B	**	**	**	B	C	B	E	C	C	C	C	D
sticky material	E	E	A	A	**	**	**	A	C	A	E	C	C	C	C	A
Pressure drop	A	A	B	B	B	B	B	C	C	C	D	C	D	C	C	***

A = very low
B = low
C = moderate
D = high
E = very high
F = exceedingly high

∞ : infinite
* : if double-pocket vanepack: N; if single-pocket or no-pocket vane pack: Y
** : if double-pocket vanepack: A; if single-pocket vanepack: B; if no-pocket vanepack: C
*** : depending on the degree of fouling, ranging from C to F

VKO Vertical knock-out drum (3.1)
HKO Horizontal knock-out drum (3.2)
VW Vertical wiremesh demister (3.3)
HW Horizontal wiremesh demister (3.4)
VV1 Vertical in-line separator with horizontal flow vane pack (3.5.5)
VV2 Vertical two-stage separator with horizontal flow vane pack (3.5.6)
HV Horizontal vane-type demister (3.6)
SMS Schoepentoeter-mistmat-swirldeck separator (3.7)
SVS Schoepentoeter-vane pack-swirldeck separator (3.7)
SMSM Schoepentoeter-mistmat-swirldeck-mistmat separator (3.7)
CT Cyclone with tangential inlet (conventional cyclone) (3.8)
CS Cyclone with straight inlet and swirler ("Gasunie" cyclone) (3.9)
VRMC Vertical separator with reversed-flow multicyclone bundle (conventional multicyclone) (3.10)
VAMC Vertical separator with axial-flow multicyclone bundle (two-stage version) (3.11)
HAMC Horizontal separator with axial-flow multicyclone bundle (3.12)
HTFS Horizontal two-stage filter separator (3.13)

3. DESIGN RULES

In this Section the design rules for the various separators are given.
Unless explicitly stated otherwise, both the maximum gas and liquid flow rates and slug volume will contain a design margin or surge factor as defined in Appendix IV.
Pressure drop calculations are based on maximum gas and liquid flow rates.
Table 1 summarises the performance data contained in this Section to enable a ready comparison of the various separators.

3.1 VERTICAL KNOCK-OUT DRUM

(Figure 3.1)

3.1.1 Selection criteria

Application:

- bulk separation of gas and liquid.

Characteristics:

- unlimited turndown;
- high slug handling capacity;
- liquid removal efficiency typically 90%;

  Warning: poor removal efficiency of liquid from mist

- very low pressure drop;
- insensitive to fouling.

Recommended use:

- vessels where internals have to be kept to a minimum (e.g. flare knock-out drums);
- fouling service e.g. wax, sand, asphaltenes;
- foaming service.

Non-recommended use:

- where efficient demisting of gas is required.

Typical process applications:

- vent and flare stack knockout drums;
- production separator;
- bulk separator (e.g. upstream of gas coolers);
- flash vessel.

3.1.2 Diameter

In general, the vessel diameter, D, shall satisfy:

The gas handling capacity criterion:

$$\lambda_{max} = Q^*_{max} / (\pi D^2_{min} / 4) = 0.07 \qquad \text{m/s}$$

or $$D \geq 4.26 \sqrt{Q^*_{max}}$$

If two immiscible liquids are present in the feed and the flow rate of the lower density liquid is at least 5% vol. of the total liquid flow rate, then the physical properties of the lighter liquid shall be used in the gas handling criterion.

NOTES

1. This criterion applies under moderate conditions only. If the pressure exceeds 100 bar (abs) or the interfacial tension is below 0.005 N/m, a larger D has to be selected, in which case the Principal should be consulted.
2. In the case of flare knock-out drums higher λ-values are acceptable:
   - If the relief flow is essentially dry or if an inside battery limit knock-out drum is used for those units or plants that are expected to produce significant two-phase flow quantities, λ_{max} = 0.25 m/s .
   - If significant liquid quantities are expected during the major relief case and no inside battery limit knock-out drum is used and

     if no feed inlet device is used: λ_{max} = 0.10 m/s

     if the feed inlet device is a Schoepentoeter: λ_{max} = 0.15 m/s

   The above criteria are in line with DEP 80.45.10.10-Gen.; which should be consulted for more information on flare knock-out design.

The vessel diameter shall also satisfy the following criteria:

If liquid de-gassing is required:

$$D \geq 7608 \sqrt{Q_{L,\,max} \eta_L / (\rho_L - \rho_G)}$$

This criterion is explained in Appendix VII.

If de-foaming is required:

$$D \geq 95Q^{0.5}_{L,max} \{\eta_L/(\rho_L - \rho_G)\}^{0.14}$$

3.1.3 Height

The total vessel height (tangent to tangent) is given by:

$$H = h + X_1 + X_2 + X_3$$

where

- h = the height required for liquid hold-up (Appendix V) calculated from the bottom tangent line,
- X_1 = the clearance between the highest liquid level LZA(HH) (Appendix V) and the inlet device,
- X_2 = the height required for the feed nozzle
- X_3 = the clearance between the inlet device and the top tangent line;

for a half-open pipe inlet device:

- X_1 = 0.3 D with a minimum of 0.3 m
- X_2 = d_1, see Figure 3.1
- X_3 = 0.9 D with a minimum of 0.9 m

for a Schoepentoeter inlet device:

- X_1 = 0.05 D with a minimum of 0.15 m
- X_2 = d_1 + 0.02 m
- X_3 = 0.6 D with a minimum of 0.6 m

3.1.4 Nozzles

The feed nozzle should preferably be fitted with a half-open pipe inlet device with its opening directed downwards.
However, the use of a Schoepentoeter may also be considered, in particular if D > 2 m.
The choice between Schoepentoeter and half-open pipe will be a trade-off between the required separation efficiency of the inlet internal (in knock-out drums not a critical issue) and costs.

For the sizing of the gas and liquid outlet nozzles see Appendix II.

NOTE For flare knock-out drums, smaller nozzles than those dictated by the above-mentioned sizing rules are acceptable:

Feed nozzle:

If no inlet device is fitted, then the momentum criterion can be relaxed from 1 000 to 2 000 Pa.

If a half-open pipe is fitted, then the momentum criterion can be relaxed from 1 500 to 5 000 Pa.

If a Schoepentoeter is fitted, then the momentum criterion can be relaxed from 6 000 to 10 000 Pa.

The Schoepentoeter shall be of sturdy design to cope with the high inlet momentum (e.g. thickness of vane material 5 mm and a divided ladder construction).

Gas outlet:

The momentum criterion can be relaxed from 3 750 to 5 000 Pa.

3.1.5 Pressure drop

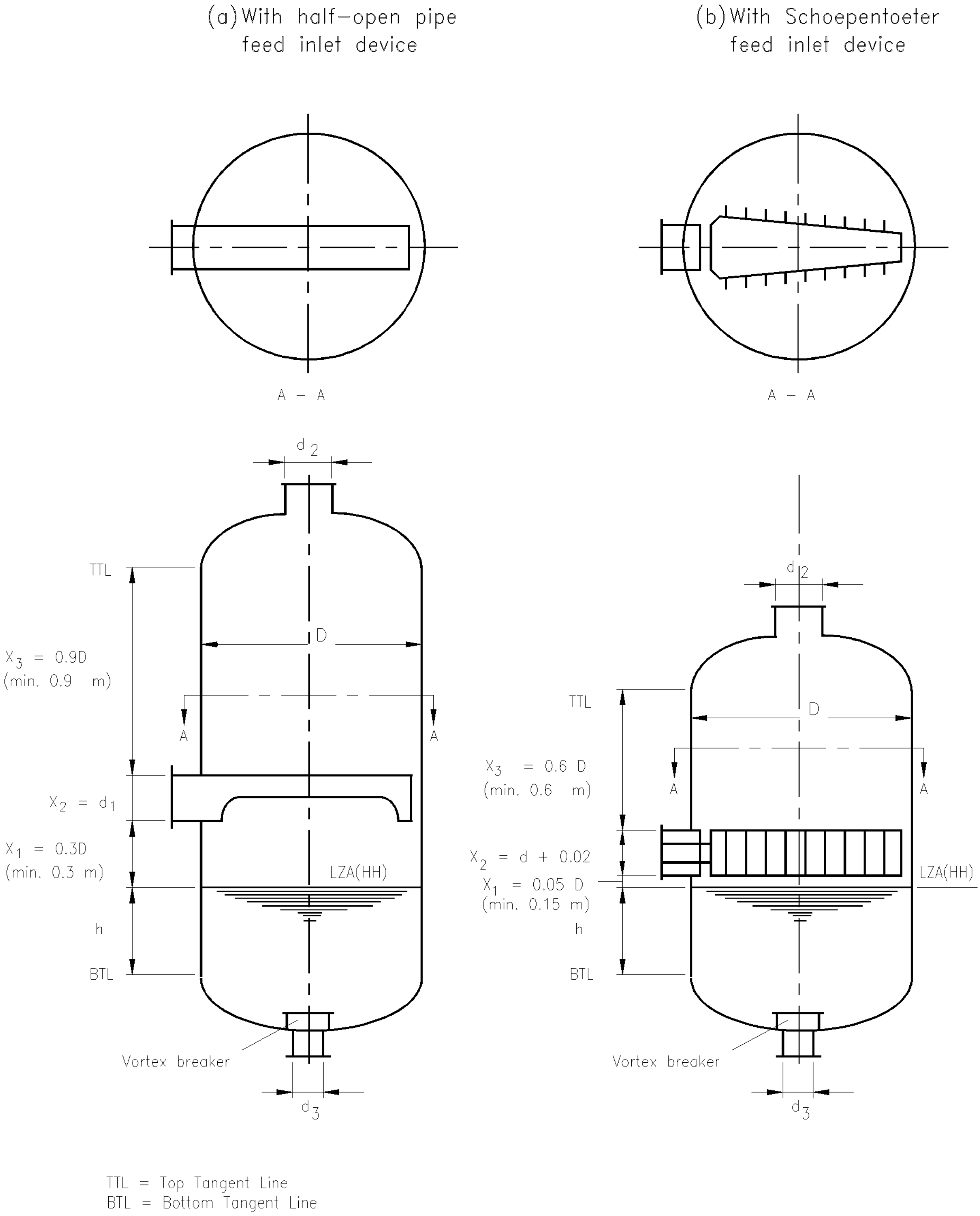
The pressure differential between inlet and vapour outlet is

$$P_{in} - P_{out} = 0.5\, \rho_m\, v^2_{m,in} + 0.22\, \rho_G\, v^2_{G,out}$$

The subscript "m" refers to the gas/liquid mixture entering the gas/liquid separators.

The definitions of ρ_m and $v_{m,in}$ are given in Appendix II.

FIGURE 3.1 THE VERTICAL KNOCK-OUT DRUM



3.2 HORIZONTAL KNOCK-OUT DRUM

(Figure 3.2)

3.2.1 Selection criteria

Application:

- bulk separation of gas and liquid.

Characteristics:

- can handle large liquid fractions;
- unlimited turndown;
- very high slug handling capacity;
- liquid removal efficiency typically 90%
  Warning: poor removal efficiency of liquid from mist;
- very low pressure drop;
- insensitive to fouling.

Recommended use:

- vessels where internals have to be kept to a minimum and where there are height limitations;
- slug catchers;
- fouling service, e.g. wax, sand, asphaltenes;
- for foaming or very viscous liquids.

Non-recommended use:

- where efficient demisting of gas is required.

Typical process applications:

- vent and flare stack knock-out drums;
- production separator for low gas/oil ratio (GOR);
- bulk separator;
- slug catcher.

3.2.2 Diameter and length

For horizontal knock-out drums, the vessel diameter is derived after considering the requirements for both gas and liquid.

Vertical cross-sectional area for gas flow

In general, the minimum vessel cross-sectional area for gas flow, $A_{G,min}$, follows from

$$\lambda_{max} = Q^*_{max} / A_{G,min} = 0.07 \quad \text{m/s}$$

in which $A_{G,min}$ is taken above the LZA(HH) liquid level (see Appendix V).

This can be rewritten as: $A_G \geq Q^*_{max} / 0.07 \quad \text{m}^2$

If two immiscible liquids are present in the feed and the flow rate of the lower density liquid is at least 5% vol. of the total liquid flow rate, then the physical properties of the lighter liquid shall be used in the gas handling criterion.

NOTES

1. This criterion applies under moderate conditions only. If the pressure exceeds 100 bar (abs) or the interfacial tension is below 0.005 N/m, a larger vessel diameter has to be selected, in which case the Principal should be consulted.

2. In the case of flare knock-out drums higher λ-values are acceptable:
   - If the relief flow is essentially dry or if an inside battery limit knock-out drum is used for those units or plants that are expected to produce significant two-phase flow quantities, λ_{max} = 0.25 m/s .
   - If significant liquid quantities are expected during the major relief case and no inside battery limit knock-out drum is used and
     if no feed inlet device is used: λ_{max} = 0.10 m/s

if the feed inlet device is a Schoepentoeter: $\lambda_{max} = 0.15$ m/s

The above criteria are in line with DEP 80.45.10.10-Gen.; which should be consulted for more information on flare knock-out design.

Liquid-full section of the vessel: the separator size

For the design of the liquid-full section of the vessel and the selection of the separator size the procedure outlined in Appendix VI shall be followed, with the restrictions that only liquid levels LZA(HH) up to 80% of the vessel diameter are allowed and the height of the gas cap shall be 0.3 m minimum. The procedure will lead to a vessel tangent-to-tangent length/diameter ratio of between 2.5 and 6.

The vessel diameter shall also satisfy the following criteria:

If liquid de-gassing is required:

$$D \geq \{4.5 * 10^7 Q_{L,max}\, \eta_L / (\rho_L - \rho_G)\} / L$$

This criterion is explained in Appendix VII.

If de-foaming is required:

$$D \geq 7\,000\ Q_{L,max}\{\eta_L / (\rho_L - \rho_G)\}^{0.27} / L$$

Note that apart from the de-gassing and de-foaming criteria the vessel diameter shall also be sufficiently large to accommodate the half-open pipe (if one is to be used as a feed inlet device).

Also, sufficient distance (> 0.15 m) shall be available between the bottom of the half-open pipe and LZA(HH).

A typical minimum diameter for a horizontal knock-out drum in non-foaming service is 1.0 m and in foaming service 1.25 m.

3.2.3 Nozzles

The diameters of the nozzles may be taken as equal to the inlet and outlet piping sizes provided that the nozzle design criteria are satisfied.

For the sizing of the nozzles see Appendix II.

The feed nozzle should be located at the top of the vessel. The use of a feed inlet device is optional.

A half-open pipe or Schoepentoeter could be used.

When a half-open pipe or Schoepentoeter is used, at least 0.15 m shall be left between the bottom of the inlet device and LZA(HH). (See Figure 3.2.). If a half-open pipe is used, its last section should be horizontal, pointing opposite to the flow direction in the vessel and with its opening directed upwards.

The gas outlet nozzle shall be located on the top of the vessel and should be fitted with a gas outlet deflector (see Figure 3.2).

NOTE For flare knock-out drums, smaller nozzles than those dictated by above-mentioned sizing rules are acceptable:

Feed nozzle:
If no inlet device is fitted, then the momentum criterion can be relaxed from 1 000 to 2 000 Pa.
If an half-open pipe is fitted, then the momentum criterion can be relaxed from 1 500 to 5 000 Pa.
If a Schoepentoeter is fitted, then the momentum criterion can be relaxed from 6 000 to 10 000 Pa.
The Schoepentoeter shall be of sturdy design to cope with the high inlet momentum (e.g. thickness of vane material 5 mm and a divided ladder construction).

Gas outlet:
The momentum criterion can be relaxed from 3 750 to 5 000 Pa.

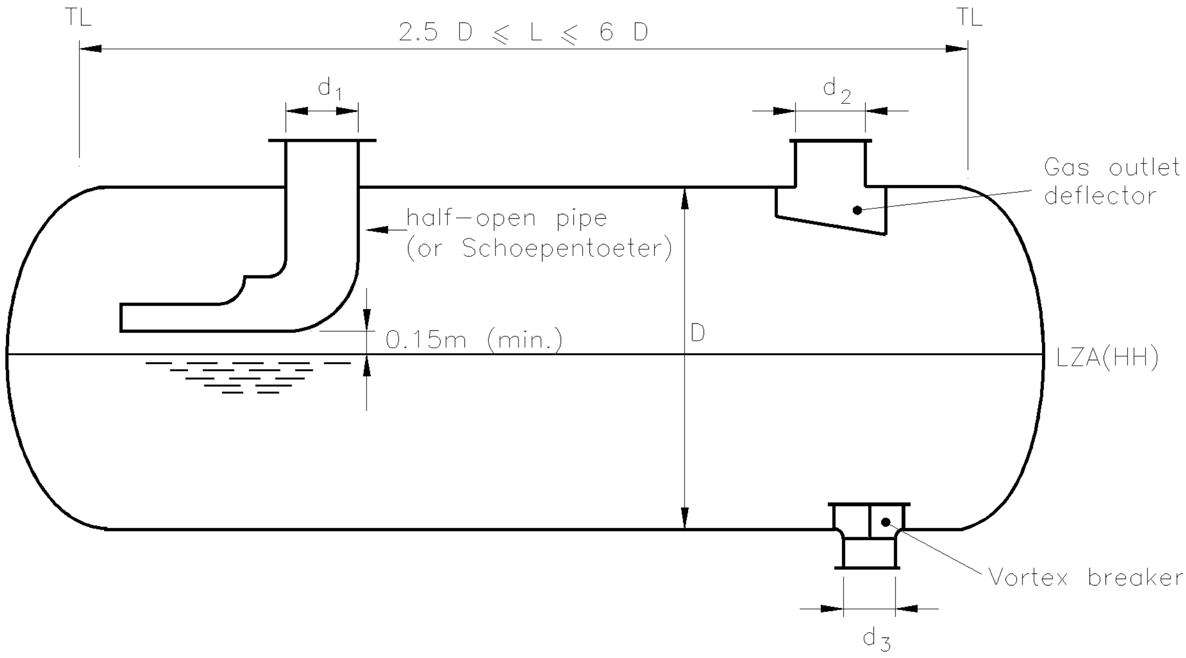
3.2.4 Pressure drop

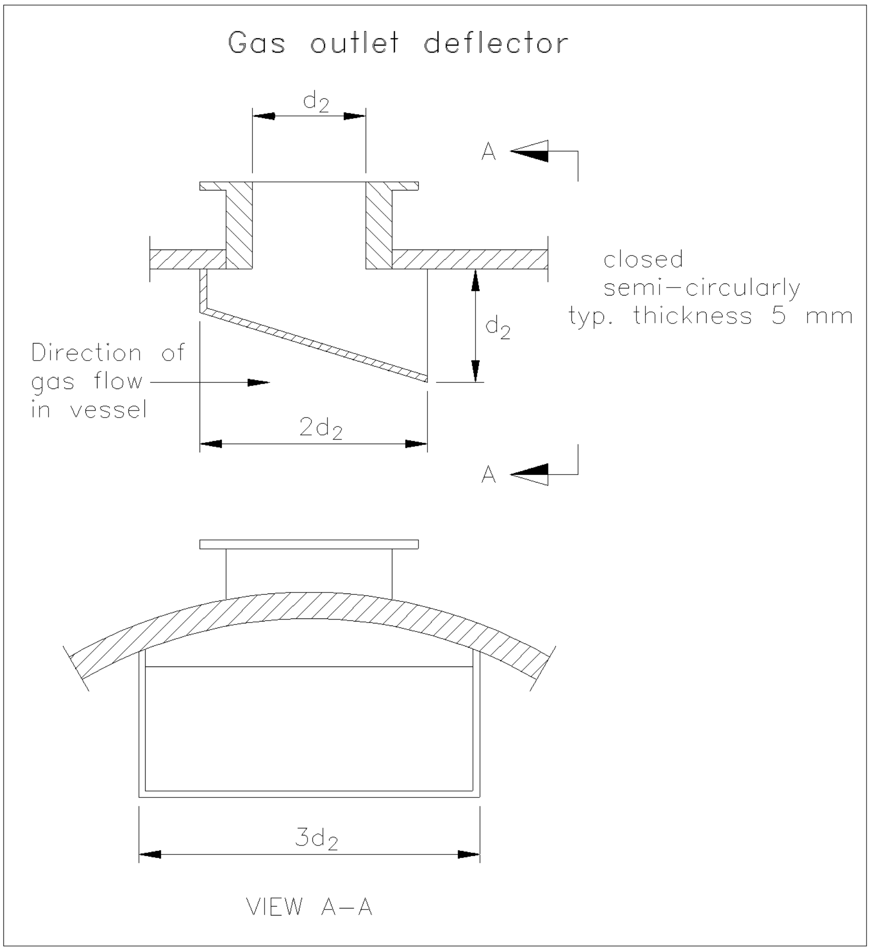
The pressure differential between inlet and vapour outlet is

$$P_{in} - P_{out} = 0.5\ \rho_m v^2_{m,in} + 0.22\ \rho_G v^2_{G,out}$$

As an example, the detailed design of a Horizontal Knockout Drum is given in Appendix X.

FIGURE 3.2 THE HORIZONTAL KNOCK-OUT DRUM





3.3 VERTICAL WIREMESH DEMISTER

(Figures 3.3 and 3.4)

3.3.1 Selection criteria

Application:

- demisting of gas.

Characteristics:

- high turndown ratio (factor 4);
- high slug handling capacity;
- liquid removal efficiency > 98%;
- sensitive to fouling;
- low pressure drop.

Recommended use:

- for demisting service with a moderate liquid load in form of droplets;
- where slug handling capacity may be required.

Non-recommended use:

- fouling service (wax, asphaltenes, sand, hydrates)
- for viscous liquids where de-gassing requirement determines vessel diameter
- for compressor suction scrubbers unless precautions are taken to prevent the possibility of loose wire cuttings entering the compressor or of the demister mat becoming clogged and thereby increasing the suction pressure drop.

Typical process applications:

- production/test separator - moderate GOR;
  - non-fouling;
- inlet/outlet scrubbers for glycol contactors;
- inlet scrubbers for gas export pipelines;
- for small diameter and/or low pressure vessels, where extra costs of e.g. vane or SMS internals cannot be justified.

3.3.2 Demister mat specifications

For the manufacturing and mounting of demister mats see Appendix VIII.

Perforated plates shall NOT be mounted downstream or upstream of the demister (in an attempt to minimise gas flow maldistribution over the wiremesh), since during operation liquid will tend to accumulate downstream of the plates resulting in a deterioration of demister performance.

3.3.3 Process considerations

3.3.3.1 Turndown

In practical terms, the turndown of a vertical demisting vessel should not become a constraint. The efficiency of the demister mat will decrease at gas flows less than around 30% of the design limit if the droplet size distribution of the liquid entrained in the gas flow remains the same. This is evident in Figure 3.4; it should be noted that this is the efficiency of the wiremesh only (the total efficiency of a wiremesh demister will be higher because of separation by the inlet device etc.).
Nevertheless, as the gas flow rate decreases (in the absence of chokes etc), the formation of droplets by dispersion upstream of the separator is less pronounced, resulting in larger droplets which are easier for the wiremesh to intercept. This effect will offset the loss in efficiency caused by the decrease of the gas flow rate.
This means in practice a turndown ratio of 4.

3.3.3.2 Efficiency

The liquid removal efficiency of a wiremesh separator is highly dependent on the liquid droplet size distribution and liquid load at the demister mat.
Figure 3.4 demonstrates the effect of average droplet size on efficiency.
For design purposes, an overall liquid removal efficiency of greater than 98% can be assumed for a correctly sized vertical demisting vessel (this includes the pre-separation by the feed inlet device).

3.3.4 Diameter

The vessel diameter, D, shall satisfy

the gas handling capacity criterion:

$$\lambda_{max} = Q^*_{max}/(\pi D^2_{min} / 4) = 0.105 f_\eta f_\phi \qquad \text{m/s}$$

or

$$D \geq 3.48 \sqrt{Q^*_{max} / (f_\eta \, f_\phi)}$$

f_η is the derating factor allowing for the viscosity of the liquid phase

$$f_\eta = (0.001/\eta_L)^{0.04} \qquad \text{if } \eta_L > 0.001 \text{ Pa.s,}$$

$$\text{if } \eta_L \leq 0.001 \text{ Pa.s, then } f_\eta = 1$$

f_ϕ is the derating factor related to the flow parameter at the face of the wiremesh, ϕ_{wm}

$$f_\phi \approx 1/(1+10\,\phi_{wm}) \qquad \text{if } \phi_{wm} \leq 0.1$$

In practice, ϕ_{wm} will not exceed 0.1.

ϕ_{wm} will be a function of the flow parameter of the feed entering the vessel and the K.O. capacity of the feed inlet device.

If a Schoepentoeter is used, assume $\phi_{wm} = 0.05\phi_{feed}$ (i.e. assumed efficiency of Schoepentoeter is 95%)

If a half-open pipe is used, assume $\phi_{wm} = 0.2\,\phi_{feed}$ (i.e. assumed efficiency of half-open pipe is 80%)

If two immiscible liquids are present in the feed and the flow rate of the lower density liquid is at least 5% vol. of the total liquid flow rate, then the physical properties of the lighter liquid shall be used in the gas handling criterion.

For a support ring for the demister mat designed in accordance with Standard Drawing S 20.030, the width of the ring can be neglected and the diameter D calculated by the above formula will be the vessel internal diameter. For other types of ring the vessel diameter is the calculated D plus twice the ring width.

NOTE The gas handling capacity criterion applies under moderate pressure conditions only.

If the pressure exceeds 100 bar (abs) or the gas/liquid interfacial tension is below 0.005 N/m, a larger vessel diameter has to be selected, in which case the Principal should be consulted.

The vessel diameter shall also satisfy the following criteria:

If liquid de-gassing is required:

$$D \geq 7608 \sqrt{Q_{L,max}\, \eta_L / (\rho_L - \rho_G)}$$

This criterion is explained in Appendix VII.

If de-foaming is required:

$$D \geq 95Q^{0.5}_{L,max} \{\eta_L/(\rho_L\text{-}\rho_G)\}^{0.14}$$

3.3.5 Height

Let h be the height of vessel required for liquid hold-up (Appendix V). Then the total tangent-to-tangent vessel height (see Figure 3.3) is:

$$H = h + X_1 + X_2 + X_3 + t_{wm} + X_4$$

where

t_{wm} = thickness of demister mat, usually 0.1 m

and either (with Schoepentoeter as inlet device):

X_1 = 0.05 D with a minimum of 0.15 m
X_2 = d_1 + 0.02 m
with d_1 = internal diameter of inlet nozzle
X_3 = d_1 with a minimum of 0.3 m
X_4 = 0.15 D with a minimum of 0.15 m

or (with half-open pipe as inlet device):

X_1 = 0.3 D with a minimum of 0.3 m
X_2 = d_1
X_3 = 0.45 D with a minimum of 0.9 m
X_4 = 0.15 D with a minimum of 0.15 m

3.3.6 Nozzles

If the vessel diameter is less than 0.5 m, the feed nozzle should be fitted with a half-open pipe inlet device, with the opening directed downwards.

For vessel diameters of 0.5 m and larger and inlet nozzle sizes of 0.15 m and larger, a Schoepentoeter inlet device is recommended.

For the sizing of the feed nozzle, see Appendix II.

For the design of Schoepentoeters, see Appendix III.

The gas outlet should be located on the top of the vessel and be fitted with a gas outlet deflector (see also Figure 3.2 for details of the deflector).

For the sizing of the gas and liquid outlet nozzles see Appendix II.

3.3.7 Pressure drop

The pressure differential between inlet and vapour outlet is

$$P_{in} - P_{out} = 0.5\rho_m v^2_{m,in} + 0.22\rho_G v^2_{G,out} + \Delta p_{wm}$$

It is recommended that the pressure drop across the mistmat (Δp_{wm}) be calculated as follows:

$$\Delta p_{wm} = 200\ (\rho_L\text{-}\rho_{G)} \lambda^2\ t_{wm}$$

$$= 20\ 000\ \lambda^2\ t_{wm} \qquad \text{mm process liquid}$$

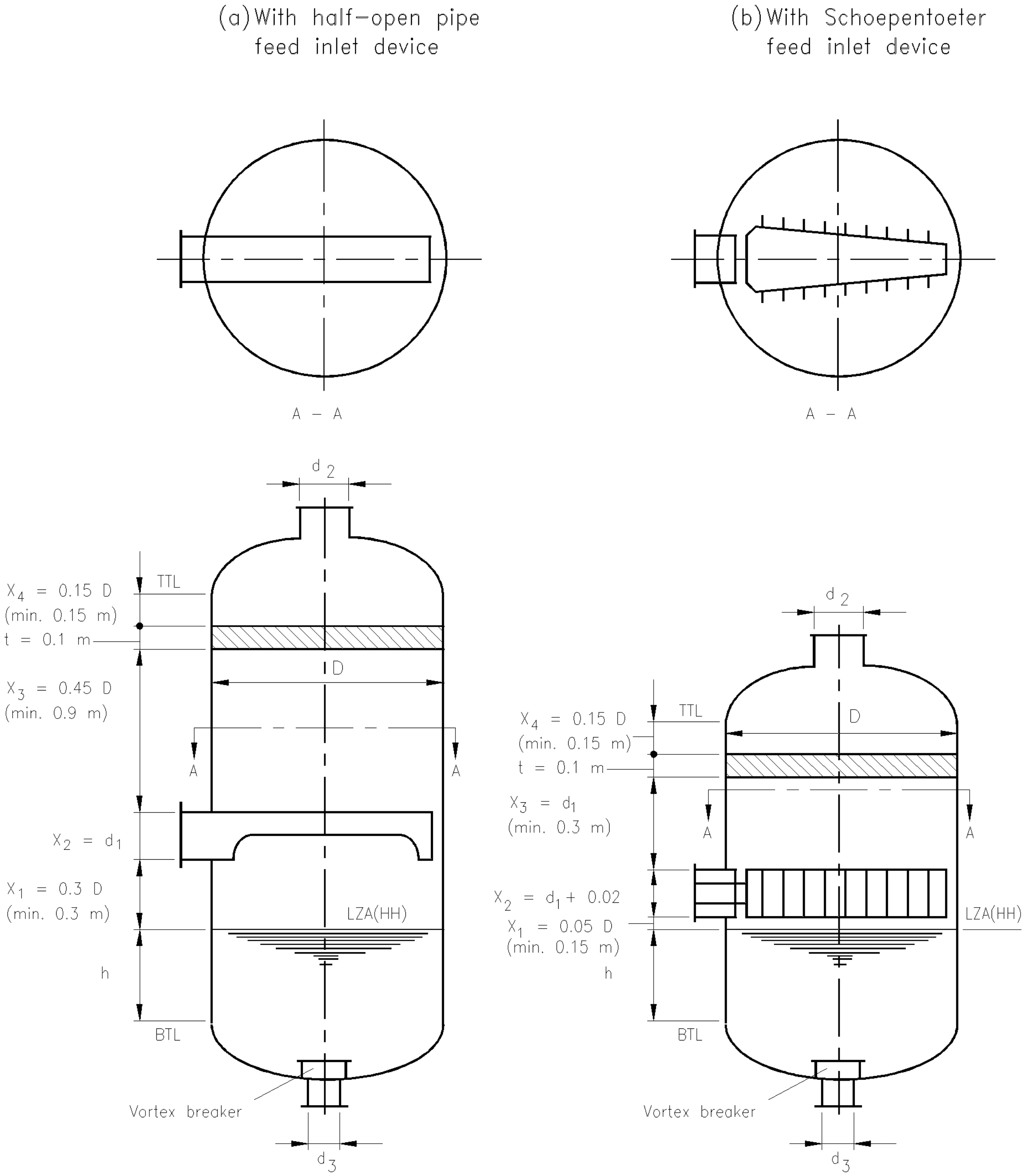
(≈ 22 mm at maximum gas load conditions if mat thickness

t_{wm} = 0.1 m)

NOTE This is the wet pressure drop.

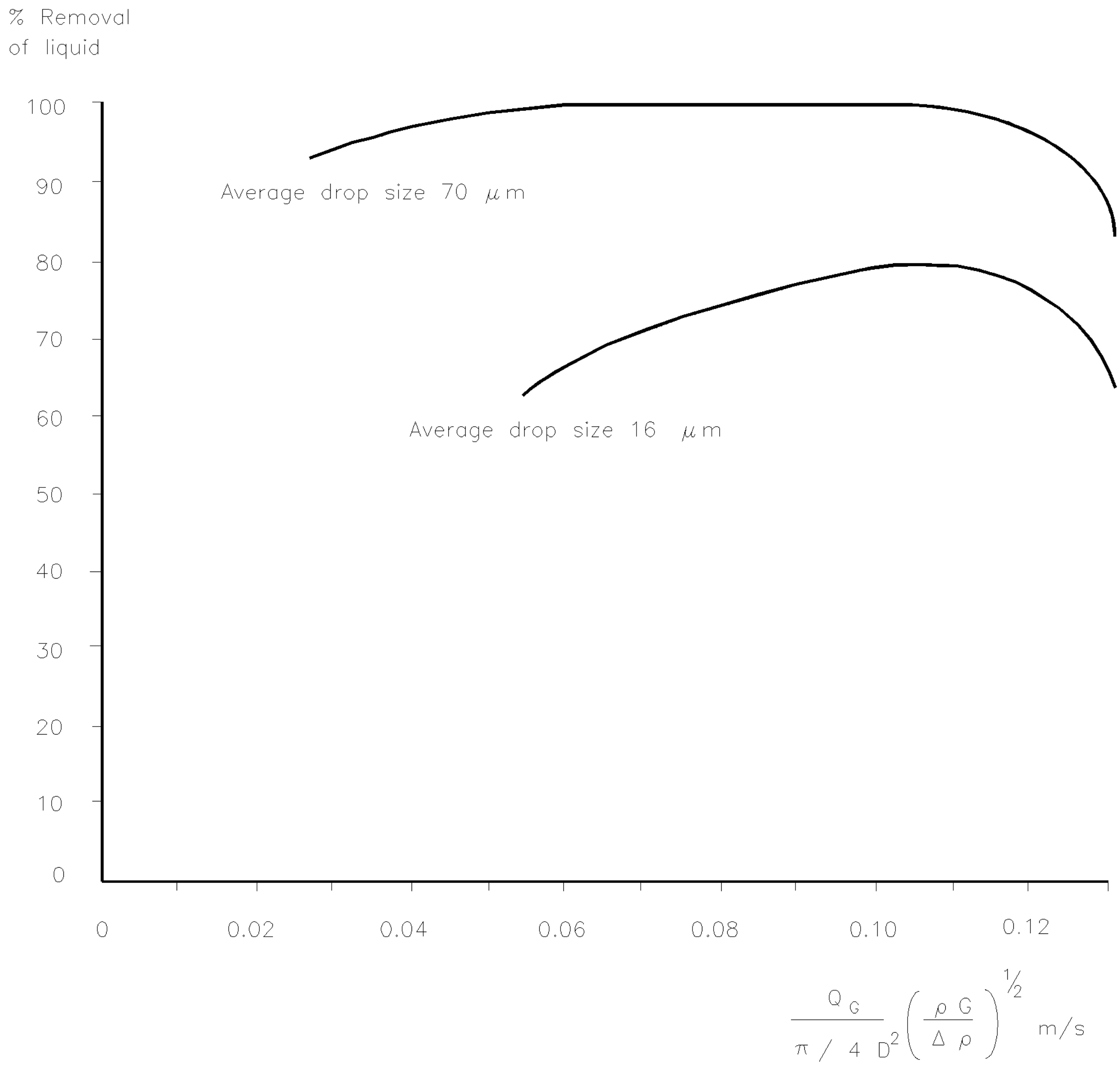
The dry pressure drop across the mistmat is 50% of this value. In the above recommended formula for the pressure drop an averaged correction for the liquid loading of the mistmat has been taken into account.

FIGURE 3.3 THE VERTICAL WIREMESH DEMISTER



Note: TTL = Top Tangent Line
BTL = Bottom Tangent Line

FIGURE 3.4 TYPICAL EFFICIENCY CURVES FOR DEMISTER MATS



System: Air / spindle oil (kin. viscosity = $13 \times 10^{-6}\ m^2/s$)

Mat: 100 mm thick, mounted horizontally
Wire diameter = 0.23 mm

3.4 HORIZONTAL WIREMESH DEMISTER

(Figure 3.5)

3.4.1 Selection criteria

Application:

- demisting of gas where a high liquid handling capacity is required.

Characteristics:

- high turndown ratio (factor 4);
- very high slug handling capacity;
- liquid removal efficiency > 98%;
- sensitive to fouling;
- low pressure drop.

Recommended use:

- typically for demisting service with a high liquid load and a low GOR;
- applied where slug handling capacity may be required;
- for viscous liquids where liquid de-gassing requirement determines vessel diameter;
- in situations where head room is restricted;
- for foaming liquids.

Non-recommended use:

- Fouling service (wax, asphaltenes, sand).

Typical process applications:

- production/test separator for low GOR
- applications with height limitations.

3.4.2 Demister mat specifications

For the manufacturing of the mistmat see Appendix VIII.

Note that the vertical mistmat in a horizontal vessel shall have a thickness of at least 10% of the vessel diameter, with a minimum of 0.15 m (as opposed to the normal 0.1 m thickness for a horizontal mistmat in a vertical vessel).

Mounting

The vertical demister mat shall extend from the top of the vessel to 0.10 m above the bottom. The area between the mat and the bottom of the vessel shall allow free passage of liquid (Figure 3.5).

The distance between the Schoepentoeter or the horizontal section of the half-open pipe and the front face of the demister mat shall be at least D.

The distance between the downstream side of the outlet nozzle and the rear face of the demister mat shall be at least 0.5D. Both distances are also indicated in Figure 3.5.

3.4.3 Diameter and length

For horizontal wiremesh demisters, the vessel diameter is derived after considering the requirements for both gas and liquid.

Vertical cross-sectional area for gas flow

The minimum vessel cross-sectional area for gas flow, $A_{G,min}$, is taken above the LZA(HH) liquid level (see Appendix V) and is computed as follows:

$$A_{G,min} = Q^*_{max} / \lambda_{max}$$

and $$\lambda_{max} = 0.09\, f_\eta f_\phi \quad \text{m/s}$$

so $A_G \geq Q^*_{max} / (0.09\ f_\eta f_\phi)$

where

f_η is the derating factor accounting for the viscosity of the liquid phase

$f_\eta = (0.001/\eta_L)^{0.04}$ if $\eta_L > 0.001$ Pa.s,

or $f_\eta = 1$ if $\eta_L \leq 0.001$ Pa.s,

f_ϕ is the derating factor related to the flow parameter at the face of the wiremesh, ϕ_{wm}

$f_\phi \approx 1/(1+10\ \phi_{wm})$ if $\phi_{wm} \leq 0.1$

In practice, ϕ_{wm} will not exceed 0.1

ϕ_{wm} will be a function of the flow parameter of the feed entering the vessel and the K.O. capacity of the feed inlet device.

If a Schoepentoeter is used, assume $\phi_{wm} = 0.05\phi_{feed}$

If a half-open pipe is used, assume $\phi_{wm} = 0.2\ \phi_{feed}$

If two immiscible liquids are present in the feed and the flow rate of the lower density liquid is at least 5% vol. of the total liquid flow rate, then the physical properties of the lighter liquid shall be used in the gas handling criterion.

NOTE The gas handling capacity criterion applies for moderate pressure conditions only.

If the pressure exceeds 100 bar (abs) or the interfacial tension is below 0.005 N/m, a larger vessel diameter has to be selected, in which case the Principal should be consulted.

WARNING: The gas handling capacity of the horizontal wiremesh demister is lower than that of its vertical counterpart, the main reason being that the gravity force is less effective in preventing the droplets from migrating through the mistmat and reaching the downstream face.

Liquid-full section of the vessel: the separator size

For the design of the liquid-full section of the vessel and the selection of the separator size the procedure outlined in Appendix VI shall be followed, with the restrictions that only liquid levels LZA(HH) up to 60% of the vessel diameter are allowed and that the height of the gas cap shall be 0.6 m minimum.

The procedure will lead to a vessel tangent-to-tangent length/diameter ratio of between 2.5 and 6.

The vessel diameter shall also satisfy the following criteria:

If liquid de-gassing is required:

$$D \geq \{4.5*10^7 Q_{L,max}\eta_L/(\rho_L-\rho_G)\} / L$$

This criterion is explained in Appendix VII.

If de-foaming is required:

$$D \geq 7\ 000\ Q_{L,max}\{\eta_L/(\rho_L-\rho_G)\}^{0.27} / L$$

Note that apart from the de-gassing and de-foaming criteria the vessel diameter shall also be sufficiently large to accommodate the Schoepentoeter. Also, sufficient distance ($\geq$ 0.15 m) shall be available between the bottom of the Schoepentoeter and LZA(HH).

A typical minimum diameter for a horizontal wiremesh demister is 1.3 m in non-foaming service and 1.6 m in foaming service.

3.4.4 Nozzles

A Schoepentoeter inlet device is recommended for horizontal wiremesh demisters.

The feed nozzle may be located at the vessel front or vessel top, as indicated in Figure 3.5. For process purposes, the top location is slightly preferable.
In both cases the distance between the Schoepentoeter and the mistmat shall be at least one vessel diameter.
For the sizing of the feed nozzle, see Appendix II.
For the design of Schoepentoeters, see Appendix III.

The gas outlet shall be located on the top of the vessel and be fitted with a gas outlet deflector (See also Figure 3.2 for details of the deflector).

For the sizing of the gas and liquid outlet nozzles, see Appendix II.

3.4.5 Pressure drop

The pressure differential between inlet and vapour outlet is

$$P_{in} - P_{out} = 0.5\rho_m v^2_{m,in} + 0.22\rho_G v^2_{G,out} + \Delta p_{wm}$$

It is recommended that the pressure drop across the mistmat (Δp_{wm}) be calculated as follows:

$$\Delta p_{wm} = 200\ (\rho_L - \rho_G)\lambda^2\ t_{wm}$$

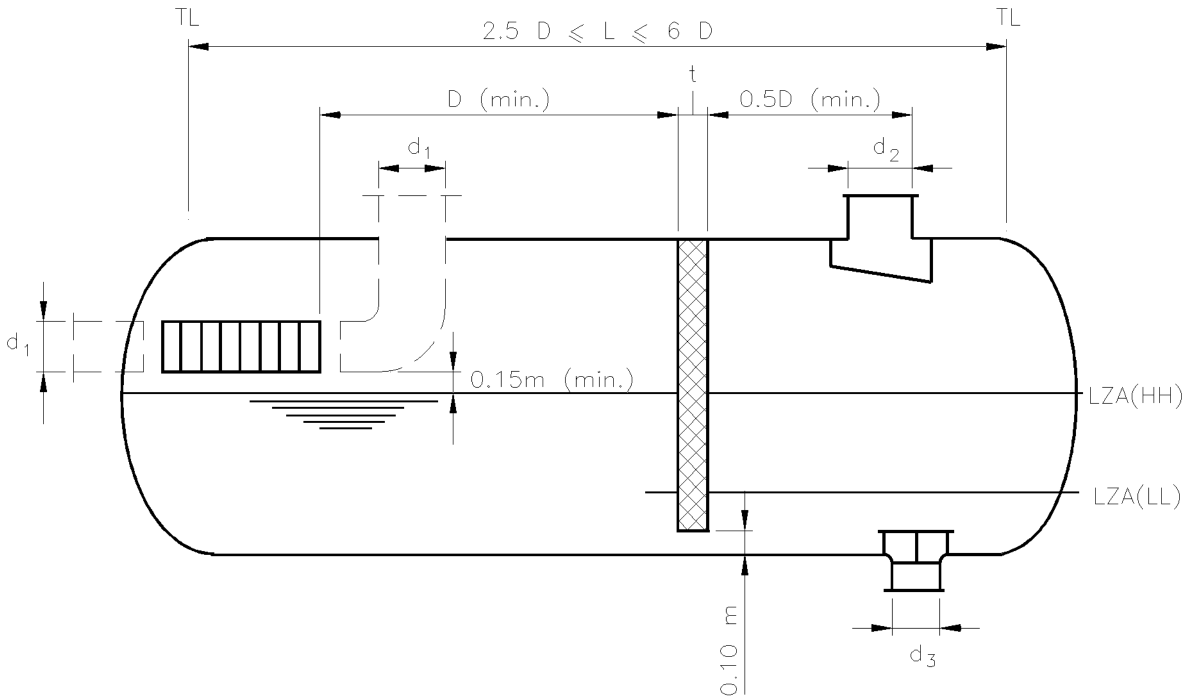
$$= 20\ 000\ \lambda^2\ t_{wm} \qquad \text{mm process liquid}$$

(≈ 24 mm at maximum gas load conditions if mat thickness t_{wm} = 0.15 m)

NOTE This is the wet pressure drop.

The dry pressure drop across the mistmat is 50% of this value. In the above recommended formula for the pressure drop an averaged correction for the liquid loading of the mistmat has been taken into account.

FIGURE 3.5 THE HORIZONTAL WIREMESH DEMISTER



3.5 VERTICAL VANE-TYPE DEMISTER

Types:

1. IN-LINE SEPARATOR WITH HORIZONTAL FLOW VANE PACK (Figure 3.6a)
2. TWO-STAGE SEPARATOR WITH HORIZONTAL FLOW VANE PACK (Figure 3.6c)

3.5.1 Selection criteria

Application:

- demisting of gas.

Characteristics:

- liquid removal efficiency > 96%;
- moderate turndown ratio (factor 3);
- suitable for slightly fouling service (if without double-pocket vanes);
- robust design;
- sensitive to liquid slugs (in-line separator cannot handle slugs).

Recommended use:

- typically for demisting service;
- in-line separator to be used only with relatively low flow parameter (ϕ_{feed} < 0.01);
- two-stage separator to be used if $\phi_{feed} \geq 0.01$;
- attractive for slightly fouling service (if without double-pocket vanes);
- may be used where demister mats may become plugged, i.e. waxy crudes.

Non-recommended use:

- heavy fouling service (heavy wax, asphaltenes, sand, hydrates);
- for viscous liquids where de-gassing requirement determines vessel diameter;
- the in-line vertical flow vane pack separator shall not be used where liquid slugging may occur or where $\phi_{feed} \geq 0.01$;
- if pressure exceeds 100 bar (abs), due to the consequent sharp decline in liquid removal efficiency.

Typical process applications:

- compressor suction scrubbers - where vane packs are preferred to demister mats since their construction is more robust;
- demisting vessels with slightly fouling service.

3.5.2 Vane pack specification

The vertical vane-type demister vessels available on the market are normally equipped with a horizontally flowed-through vane pack. Basically there are three types of vane elements: the no-pocket (straight), single-pocket and double-pocket type.
The vane elements used in horizontally flowed-through vane packs are mostly of the single-pocket or double-pocket type.

In Appendix IX more information is given on the various types of vanes and also a few general guidelines are given for the manufacturing and mounting of vane packs.

3.5.3 Determination of vane pack face area

The vane pack area face, A_v, shall be calculated from

$$A_v = Q^*_{max} / \lambda_{v,max}$$

where $\lambda_{v,max}$ is the maximum allowable gas load factor based on the face area of the vane pack.

It is defined as the density-corrected maximum allowable gas velocity in the vane pack (disregarding the thickness of the vanes).

$$\lambda_{v,max} = v_{G,max,v} \sqrt{\rho_G / (\rho_L - \rho_G)}$$

$\lambda_{v,max}$ is not fixed, but is a function of the flow parameter and of physical properties.

In case two immiscible liquids are present and the flow rate of the "minority" liquid exceeds 5% vol. of the total liquid flow rate, $\lambda_{v,max}$ shall be calculated for both liquid phases and the lowest value shall be used for the sizing of the vane pack.

For the calculation of $\lambda_{v,max}$ two possibilities have to be considered, depending on the Archimedes number.

The Archimedes number is a dimensionless physical property group and is defined as:

$$Ar = (\rho_L / \eta_L^2) \sqrt{\sigma^3 / \{g(\rho_L - \rho_G)\}}$$

where g is the acceleration due to gravity and σ is the gas/liquid interfacial tension.

possibility 1: Ar > 225 (this normally will be the case)

$$\lambda_{v,max} = 1.75\{g\,\sigma / (\rho_L - \rho_G)\}^{0.24} (\sigma/\eta_L)^{0.04} / (1 + 25\phi_v)$$

where

ϕ_v is the flow parameter at the vane face and is derived from the feed flow parameter by taking into account the separation efficiency of the feed internal.

In a correctly designed separator, ϕ_v will not exceed 0.01.

possibility 2: $Ar \leq 225$

$$\lambda_{v,max} = 0.14(\sigma/\eta_L) / (1+25\phi_v)$$

The second possibility usually applies when the liquid viscosity is relatively high (e.g. liquid sulphur or glycols).

3.5.4 Process considerations

3.5.4.1 Turndown

Since the demisting efficiency of the vane pack is related to the centrifugal forces induced by the oscillatory gas flow path between the plates of the vane pack, the mist removal efficiency of a vane pack decreases at lower gas flows (assuming the droplet size distribution remains the same).
The efficiency of the vane pack mist eliminators tends to decline at around 50% of the design throughput. However, a turndown to 30% of the design gas rate can typically be achieved with acceptable efficiencies, especially if it is taken into account that at lower gas flow rates the size of the liquid droplets tends to be larger which facilitates gas/liquid separation.

3.5.4.2 Efficiency

The liquid removal efficiency of a separator is highly dependent on the liquid droplet size distribution and liquid load at the vane pack.
For design purposes, an overall liquid removal efficiency of greater than 96% can be assumed for a correctly sized vertical vane-type demister.

3.5.5 In-line separator with horizontal flow vane pack

(Figure 3.6a)

Warning: This type shall only be used if $\phi_{feed} < 0.01$ and slugging is not expected.

The design philosophy is first to determine the vane face area required for the G/L separation. The rules for this are given in (3.5.3). Subsequently the vessel and nozzle dimensions are determined such that the gas flow maldistribution over the vane pack is kept within acceptable limits.

3.5.5.1 Layout of vane pack (Figure 3.6b)

The vane pack shall be enclosed in a box with a gas-tight connection to the outlet nozzle.

This connection should be bolted.

There should be a clearance of at least 0.1 m between the vessel wall and the mist extractor box to allow installation, removal, attachments and inspection.

Similarly, the distance between the top of the vane box and the top seam weld of the vessel should be at least 0.1 m.

To facilitate the installation of the vane box, the vessel should be top-flanged if the vessel diameter is less than 1.2 m.

Initially, take the vane height, h_v, as:

$$h_v = \sqrt{1.5\ A_v} \qquad \text{m}$$

where the vane area, A_v, follows from the formulae given in (3.5.3).

If $\phi_v < 0.01$, even if not exactly known, take as default $\phi_v = 0.01$ in the formulae.

h_v should be adjusted to fall within the following range:

$$0.30 \le h_v \le 1.5 \qquad \text{m}$$

The vane pack width, $w_v = A_v / h_v$

The width of the mist extractor box, $w_{vb} = w_v + 0.1$ m

The height of the vane pack box, h_{vb}, shall include a margin to obtain sufficient coverage of the vanes in order to prevent vapour by-passing the demister. Also, sufficient height shall be available to allow proper draining of the separated liquid (Figure 3.6a).

Typically, $h_{vb} = h_v + 0.3$ m

Liquid shall be drained from the vane pack to the bottom compartment of the vessel via drain pipes having a minimum diameter of 0.05 m. At least one pipe for each metre of vane pack width shall be used.

The drain pipe(s) shall extend at least 0.10 m below LZA(LL) for sealing purposes.

The depth of the vane box, t_{vb}, is dependent on the type of vane selected and is normally between 0.30 and 0.45 m.

NOTE In t_{vb}, a margin of 2 times 0.05 m should be included to allow for the distance from the perforated plates to the vane elements (assuming perforated plates both upstream and downstream of the vane pack).

To promote even gas flow over the vane pack face the following measures shall be taken:

1. The face of the vane pack shall be perpendicular to the centre lines of both the feed inlet nozzle and the gas outlet nozzle.
2. The cross-section of the inlet nozzle shall be at least 15% of A_v (in most cases the momentum criterion for the inlet nozzle is then also satisfied (see (3.5.5.4))).
3. The outlet nozzle diameter shall not be smaller than the feed nozzle diameter.
4. A perforated plate shall be installed at the front of the vane pack.
   A perforated plate at the back is optional.
   The recommended net free area (NFA) of the perforated plate is about 20 %.
   The holes shall be evenly distributed over the plate and the nominal hole size should be about 12 mm.

3.5.5.2 Vessel diameter

The minimum vessel diameter of the in-line separator is determined by the following two requirements:

$$D \geq 0.2 + \sqrt{(w^2_{vb} + t^2_{vb})} \quad \text{m} \quad \text{(vane box requirement)}$$

and $D \geq 0.6$ m (vessel accessibility requirement)

3.5.5.3 Vessel height

Let h be the height of vessel required for liquid hold-up (Appendix V).

Then the total vessel height H (tangent to tangent) is

$$H = h + X_1 + h_{vb} + X_2$$

where

X_1 is the distance between the vane box and LZA(HH)
$X_1 \geq 0.5$ m.
X_2 is the distance between the vane box and the top tangent line
$X_2 \geq 0.1$ m

3.5.5.4 Nozzles

The following requirements shall be satisfied for the feed nozzle:

1. The cross-section of the inlet nozzle and the last section of the feed pipe over a length of 4 diameters shall be at least 15% of A_v.

   This means that

   $d_1 \geq 0.43 \sqrt{A_v}$

2. $\rho_G v^2_{G,in} \leq 3750$ Pa

   Normally, if requirement 1 is met, requirement 2 is also satisfied.

   The diameter of the gas outlet nozzle, d_2, shall satisfy the following requirement:

   $d_2 \geq d_1$

For the sizing of the liquid outlet nozzle, see Appendix II.

3.5.5.5 Pressure drop

The pressure differential between inlet and vapour outlet is the sum of the pressure drops across the nozzles, the vane pack and the perforated plate(s).

$$P_{in} - P_{out} = 0.5\rho_m v^2_{m,in} + 0.22\rho_G v^2_{G,out} + \Delta p_v + n\Delta p_{perfpl}$$

where

$$\Delta p_v = K_v(\rho_L - \rho_G)\,\lambda^2_v$$

in which $K_v = 15$ for single-pocket vanes

or $K_v = 10$ for double-pocket vanes

and n = number of perforated plates

$$\Delta p_{perfpl} = 0.8(1 - NFA)\,(\rho_L - \rho_G)\lambda^2_v / NFA^2$$

in which NFA = the net free area of the perforated plate (in fraction)

3.5.6 Two-stage separator with horizontal flow vane pack

(Figure 3.6c)

This two-stage separator can be used when limited slugging is expected and/or the feed flow parameter is higher than 0.01.

If double-pocket vanes are used, the separator is suitable for clean service only.

A Schoepentoeter shall be fitted as the feed inlet device for primary gas/liquid separation. This primary separation shall limit the flow parameter to a maximum value of 0.01.

The design philosophy is first to determine the minimum vessel diameter required for the reduction of the flow parameter down to 0.01, with the restriction that the diameter shall be sufficiently large for accessibility and de-gassing/de-foaming performance. Then, a suitable vane pack configuration is dimensioned to fit the selected vessel diameter. If the vessel diameter is insufficient for the required vane area it has to be increased further. In such a case, however, it might be worthwhile to select another separator, such as the SVS (or SMS in the case of non-fouling service).

3.5.6.1 Vessel diameter

The vessel diameter shall satisfy

1. **the primary separation criterion** (i.e. separation by Schoepentoeter):

$$\lambda_{max} = Q^*_{max} / (\pi D^2_{min} / 4) = 0.1 + 0.008\rho_G^{-0.14}\, \phi_{feed}^{-0.76}$$

giving $$D \geq 1.13\sqrt{Q^*_{max} / \lambda_{max}}$$

(λ_{max} should be low enough to have sufficient separation efficiency of the Schoepentoeter to bring the flow parameter just upstream of the vane pack down to 0.01).

If **no slugs** are expected, the above criterion shall **not exceed**:

$$\lambda_{max} = 0.20 \quad \text{m/s}$$

If **slugs** are expected:

$$\lambda_{max} = 0.10 \quad \text{m/s (to prevent overloading of the vane pack)}$$

NOTE Here, λ is the load factor based on the **vessel's** cross-sectional area.

If two immiscible liquids are present in the feed and the flow rate of the lower density liquid is at least 5% vol. of the total liquid flow rate, then the physical properties of the lighter liquid shall be used in the gas handling criterion.

2. **the de-gassing criterion:**

$$D \geq 7608\sqrt{Q_{L,max}\, \eta_L / (\rho_L - \rho_G)}$$

3. **the de-foaming criterion:**

$$D \geq 95Q^{0.5}_{L,max}\{\eta_L / (\rho_L - \rho_G)\}^{0.14}$$

4. **the accessibility criterion:**

$$D \geq 0.6 \text{ m}$$

5. **the criterion for even flow distribution over the vane face:**

$$D \geq \{\, 2\, (t_{vb} + \sqrt{t^2_{vb} + 0.6\pi A_v}\,)\} / \pi$$

3.5.6.2 Layout of vane pack

The vessel diameter calculated above is used to determine the vane pack dimensions.

It is usually advantageous to minimise the vane pack height so that the amount of separated liquid at the bottom of the vane pack is also minimised and thus the chance of local re-entrainment is eliminated. Therefore, as a starting value for the design, the maximum possible width of the vane pack is calculated.

Under those conditions the minimum spacing between the vane box and the vessel wall is assumed to be 0.1 m (spacing required for installation, removal, attachments and inspection).

$$W_v = \sqrt{(D-0.2)^2 - t^2_{vb}} - 0.10 \qquad \text{m}$$

The depth of the vane box, t_{vb}, is dependent on the type of vane selected and ranges typically between 0.30 and 0.45 m.

NOTE In t_{vb}, a margin of 2 times 0.05 m should be included to allow for the distance from the perforated plates to the vane elements.

The corresponding vane height, h_v, is calculated using the required vane pack face area, A_v, as determined in (3.5.3) with $\phi_v = 0.01$:

$$h_v = A_v / w_v$$

The dimensions w_v and h_v should then be adjusted to fulfil the condition:

$$0.3 \leq h_v \leq 1.5 \qquad \text{m}$$

If $h_v < 0.3$ m, a smaller value for w_v is selected and the corresponding h_v is then recalculated.

If $h_v > 1.5$ m , the vessel diameter should be increased to keep h_v within the limits.

The remaining vane pack dimensions are calculated as follows:

The width of the vane pack box , w_{vb}, is:

$$w_{vb} = w_v + 0.10 \qquad \text{m}$$

The height of the vane pack box, h_{vb}, shall include a margin to obtain sufficient coverage of the vanes in order to prevent vapour by-passing the demister.

Also, sufficient height shall be available to allow proper draining of the separated liquid (Figure 3.6c).

Typically: $h_{vb} = h_v + 0.30$ m

The vane pack box shall have a gas-tight connection to the outlet nozzle.

This connection should be bolted.

Liquid shall be drained from the vane pack to the bottom compartment of the vessel via drain pipes having a minimum diameter of 0.05 m. At least one pipe for each metre of vane pack width shall be used.

The drain pipe(s) shall extend at least 0.10 m below LZA(LL) for sealing purposes.

To promote even gas flow over the vane pack face the following measures shall be taken:

1. The vane area shall be perpendicular to the centre lines of both the feed inlet nozzle and the gas outlet nozzle.
2. A perforated plate shall be installed at the front of the vane pack.
   A perforated plate at the back is optional.
   The recommended net free area (NFA) of the perforated plate is 20%.
   The holes shall be evenly distributed over the plate and the hole size should be about 12 mm nominal.
3. The cross-sectional vessel area available for upward gas flow at the bottom of the demister box (i.e. the cross-section of the segment cutoff by the front of the vane box) shall be at least 30% of the installed vane pack face area, A_v. If necessary, the vessel diameter has to be increased to satisfy this requirement.

If t_{vb} is much smaller than D (which is normally the case) the above-mentioned requirement for D can be approximated as:

$$D \geq \{2\ (t_{vb} + \sqrt{t^2{}_{vb} + 0.6\pi A_v}\ )\} / \pi$$

(The height h of the vane pack face area has to be checked. If necessary D has to be increased further in order to keep $h_v \leq 1.5$ m).

3.5.6.3 Height

Let h be the height required for liquid hold-up (Appendix V). Then the total vessel height (tangent to tangent) is (see Figure 3.6.c):

$$H = h + X_1 + X_2 + X_3 + h_{vb} + X_4$$

where

X_1 is the distance between LZA(HH) and the Schoepentoeter.
$X_1 = 0.05D$ with a minimum of 0.15 m.
$X_2 = d_1 + 0.02$ m
X_3 is the distance between the Schoepentoeter and the vane box.
$X_3 = 0.5D$ with a minimum of 0.3 m
X_4 is the distance between the vane box and the top tangent line.
$X_4 \geq 0.1$ m.

3.5.6.4 Nozzles

The feed nozzle shall be fitted with a Schoepentoeter.

For the design of Schoepentoeters, see Appendix III.

For the sizing of the feed nozzle see Appendix II.

A side gas outlet rather than a top outlet should be used in order to minimise gas maldistribution and pressure drop.

The following criteria shall be satisfied for the diameter of the gas outlet nozzle, d_2:

1. $d_2 \geq 0.43\sqrt{A_v}$
2. $\rho_G v^2{}_{G,out} \leq 3750$ Pa

For the sizing of the liquid outlet nozzle see Appendix II.

3.5.6.5 Pressure drop

The pressure differential between inlet and vapour outlet is the sum of the pressure drops across the nozzles, the vane pack and the perforated plate(s).

(The pressure drop across the Schoepentoeter is negligible.)

$$P_{in} - P_{out} = 0.5\rho_m v^2{}_{m,in} + 0.22\rho_G v^2{}_{G,out} + \Delta p_v + n\Delta p_{perfpl}$$

where

$$\Delta p_v = K_v(\rho_L - \rho_G)\ \lambda^2{}_v$$

in which $K_v = 15$ for single-pocket vanes

or $K_v = 10$ for double-pocket vanes

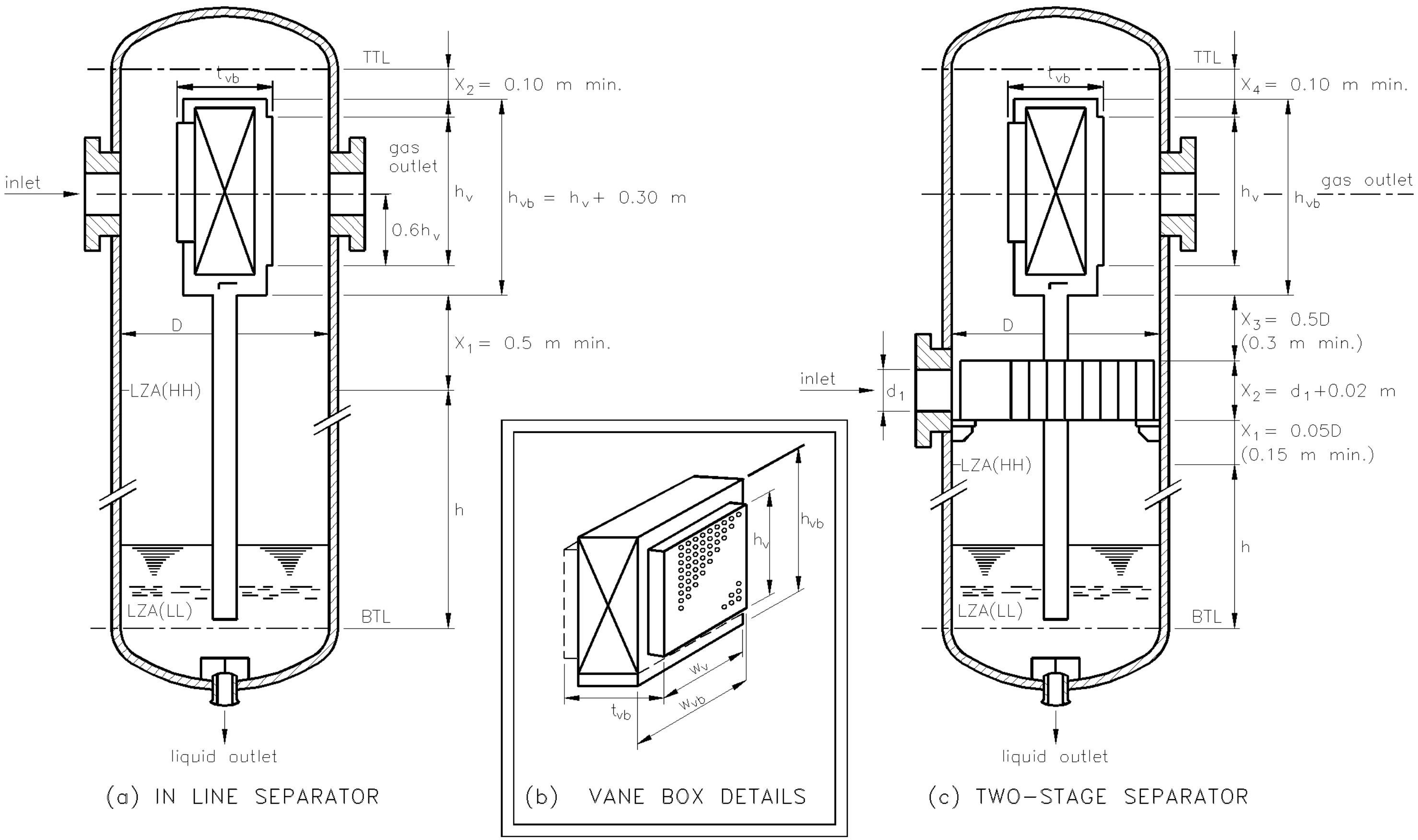
and n = number of perforated plates

$$\Delta p_{perfpl} = 0.8(1 - NFA)\ (\rho_L - \rho_G)\lambda^2{}_v / NFA^2$$

in which NFA = the net free area of the perforated plate (in fraction)

The detailed design of a Vertical Vane-type Demister is given as an example in Appendix X

FIGURE 3.6 LAYOUT OF THE VERTICAL VANE-TYPE SEPARATOR



3.6 HORIZONTAL VANE-TYPE DEMISTER

(Figure 3.7)

3.6.1 Selection criteria

Application:

- demisting of gas where a high liquid handling capacity is required.

Characteristics:

- liquid removal efficiency > 96%;
- moderate turndown ratio (factor 3);
- suitable for slightly fouling service (if without double-pocket vanes);
- high slug handling capacity;
- robust design.

Recommended use:

- typically for demisting service with a high liquid load and a low GOR and where very high efficiency is not required;
- attractive for slightly fouling service (if single-pocket vanes are used) and may be used where demister mats may become plugged, i.e. waxy crudes.

Non-recommended use:

- heavy fouling service (heavy wax, asphaltenes, sand, hydrates);
- if pressure exceeds 100 bar (abs).

Typical process applications:

- production separator where GOR is low and the service is slightly fouling.

3.6.2 General

Normally the horizontal vane-type demister is equipped with horizontal flow vane packs having vanes of either the single-pocket or double-pocket type. The required vane pack area is calculated with the equations given in (3.5.3).

The vane-type demister is suitable for slightly fouling service if single-pocket vanes are used. Double-pocket vanes are acceptable in clean service only.

The demister shall be equipped with a Schoepentoeter as feed inlet device in which the primary gas/liquid separation will take place.

The design philosophy is first to size the demister for a proper primary gas/liquid separation and an adequate liquid handling capacity. This will determine the minimum vessel cross-sectional area for gas flow and the minimum liquid control volume.

Then the vane pack is dimensioned with sufficient vane area for proper demisting. Normally a vane pack has its face perpendicular to the vessel axis. Sometimes, however, this layout will not create sufficient vane area (for instance in the case of a retrofit in a undersized vessel). Under those circumstances it is possible to use a diagonal vane pack where the angle between the vane face and the vessel axis can be taken down to 10 degrees. To minimise maldistribution over the vane area this diagonal vane pack shall be equipped with one or more perforated plates.

3.6.3 Diameter and length

For horizontal vane-type demisters, the vessel diameter is derived after considering the requirements for both gas and liquid.

Vertical cross-sectional area for gas flow

The minimum vessel cross-sectional area for gas flow, $A_{G,min}$, is derived from

$$\lambda_{max} = Q^{*}_{max} / A_{G,min} = 0.1 + 0.008\ \rho_G^{-0.14}\ \phi_{feed}^{-0.76}$$

(λ_{max} should be sufficiently low efficiency for the Schoepentoeter to have sufficient separation to bring the flow parameter just upstream of the vane pack down to 0.01.)

$A_{G.min}$ is taken above the LZA(HH) liquid level (see Appendix V).

giving: $A_G \geq Q^{*}_{max} / \lambda_{max}$

If two immiscible liquids are present in the feed and the flow rate of the lower density liquid is at least 5% vol. of the total liquid flow rate, then the physical properties of the lighter liquid shall be used in the gas handling criterion.

If **no slugs** are expected, above criterion shall **not exceed**:

$\lambda_{max} = 0.15$ m/s

If **slugs** are expected:

$\lambda_{max} = 0.10$ m/s (to prevent overloading of the vane pack)

Liquid-full section of the vessel: separator size

For the design of the liquid-full section of the vessel and the selection of the separator size the procedure outlined in Appendix VI shall be followed with the restrictions that only liquid levels LZA(HH) up to 50% of the vessel diameter are allowed and the height of the gas cap shall be 0.8 m minimum.

This procedure will lead to a tangent-to-tangent length/diameter ratio between 2.5 and 6.

The following criteria shall also be satisfied:

If liquid de-gassing is required:

$$D \geq \{4.5*10^7 Q_{L,max}\eta_L/(\rho_L-\rho_G)\} / L$$

This criterion is explained in Appendix VII.

If de-foaming is required:

$$D \geq 7\,000\ Q_{L,max}\{\eta_L/(\rho_L-\rho_G)\}^{0.27} / L$$

Note that apart from the de-gassing and de-foaming criteria, the vessel diameter shall also be sufficiently large to accommodate the Schoepentoeter.
Also, sufficient distance (> 0.15 m) shall be available between the bottom of the Schoepentoeter and LZA(HH).

A typical minimum diameter for a horizontal vane-type demister is 1.5 m in non-foaming service and 1.9 m in foaming service.

3.6.4 Vane pack

The distance between the Schoepentoeter and front face of the vane pack shall be at least D.

The distance between the downstream side of the outlet nozzle and the rear face of the vane pack shall be at least 0.5D. Both distances are also indicated in Figure 3.7.

The horizontal flow vane pack used in the horizontal separator has a similar performance as the horizontal flow vane pack in the vertical separator. Therefore for the determination of the required vane pack face area, A_v, the formulae given in (3.5.3) can be used (with the assumption that ϕ_v = 0.01).

The vane height, h_v, shall be between 0.3 m and 1.5 m.

The vanes are positioned in a box with a liquid sump. The vane pack has basically the same layout as the one employed in the vertical vane-type demisters equipped with horizontal flow vane packs (see Figure 3.6b for vane box details).

The distance between the lower end of the free surface of the vanes and LZA(HH) shall be at least 0.25 m.

Measures shall be taken to prevent gas bypassing the vane pack.

One method is to have the vanes positioned in a gas-tight box around the gas outlet. Liquid drainage from the sump of the vane box is via drain pipes (typically 0.05 m diameter drain pipe per metre of vane box width) and via typically a 0.05 m drain pipe in the vane box housing; the drain pipes shall extend at least 0.10 m below LZA(LL) level. This gas-tight box method is shown in Figure 3.7.

Another method is to place the vanes in a box which has no direct connection with the gas outlet. In this case gas bypassing is prevented by blanking off the cross-sectional vessel area adjacent to the vane box and attaching a baffle to the underside of the vane box extending at least 0.10 m below LZA(LL).

Also in this case, liquid drainage from the sump of the vane box is via drain pipes (typically one 0.05 m diameter drain pipe per metre).

3.6.5 Nozzles

The feed nozzle shall be fitted with a Schoepentoeter.

For the design of Schoepentoeters, see Appendix III.

For the sizing of the feed nozzle, see Appendix II.

The feed nozzle may be located at the vessel front or vessel top, as indicated in Figure 3.7. For process purposes, the top location is the preferred option.

In both cases the distance between the Schoepentoeter and the vane pack shall be at least one vessel diameter.

The gas outlet shall be located on the top of the vessel and be fitted with a gas outlet deflector (See also Figure 3.2 for details of the deflector).

For the sizing of the gas and liquid outlet nozzles see Appendix II.

3.6.6 Pressure drop

The pressure differential between inlet and vapour outlet is the sum of the pressure drops across the nozzles, the vane pack and the perforated plate(s).

(The pressure drop across the Schoepentoeter is negligible.)

$$P_{in} - P_{out} = 0.5\rho_m v^2_{m,in} + 0.22\rho_G v^2_{G,out} + \Delta p_v + n\Delta p_{perfpl}$$

where

$$\Delta p_v = K_v(\rho_L - \rho_G)\,\lambda^2_v$$

in which K_v = 15 for single-pocket vanes

or K_v = 10 for double-pocket vanes

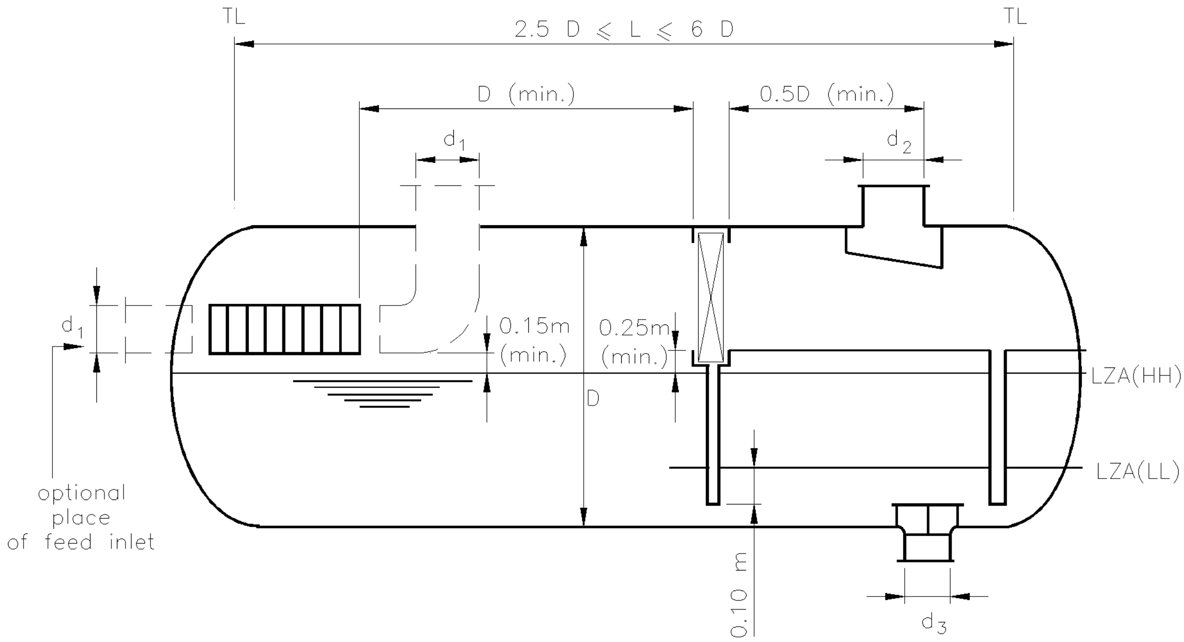
and n = number of perforated plates

$$\Delta p_{perfpl} = 0.8(1 - NFA)\,(\rho_L - \rho_G)\lambda^2_v / NFA^2$$

in which NFA = the net free area of the perforated plate (in fraction)

FIGURE 3.7 THE HORIZONTAL VANE-TYPE DEMISTER

(Only the gas-tight box option is shown, see main text)



3.7 VERTICAL SEPARATOR WITH SWIRLTUBE DEMISTER DECK

Types:

1. SCHOEPENTOETER-MISTMAT-SWIRLDECK SEPARATOR (SMS)
   (Figures 3.8a and 3.8b)
2. SCHOEPENTOETER-VANE PACK-SWIRLDECK SEPARATOR (SVS)
   (Figure 3.9a)
3. SCHOEPENTOETER-MISTMAT-SWIRLDECK-MISTMAT SEPARATOR (SMSM)
   (Figure 3.9b)

3.7.1 Selection criteria

Application:

- demisting of gas where a high gas handling capacity is required.

Characteristics:

- compact;
- high gas handling capacity (maximum allowable vessel gas load factor 0.25 m/s);
- SMS and SMSM:
  - high efficiency (>98% and >99% respectively);
  - sensitive to fouling;
  - very high turndown ratio (factor 10).

  SVS:
  - less efficient (but still >96%);
  - less sensitive to fouling;
  - high turndown ratio (factor 4).

Recommended use:

- because of compactness, suitable for offshore industry or in general for high pressure conditions;
- for debottlenecking of existing separators;
- SVS attractive for slightly fouling service and may be used where demister mats may become plugged, i.e. waxy crudes.

Non-recommended use:

- SMS and SMSM: - fouling service;
  SVS: - heavy fouling service;
- for viscous liquids where de-gassing requirement determines vessel diameter;
- if insufficient head room is available for a vertical vessel.

Typical process applications:

SMS: production separator, compressor suction drum, inlet scrubber for glycol contactors;
SMSM: cold separator (removal of glycol/HC liquid mixture), inlet scrubber for gas export pipelines;
SVS: well head separator, centrifugal compressor suction drum.

3.7.2 General

The SMS, SVS and SMSM separators are Shell proprietary high-capacity gas/liquid separators having the following internals:

- a Schoepentoeter as feed inlet device for the primary gas/liquid separation;
- a mistmat (SMS and SMSM) or a vane pack (SVS) acting as a coalescer;
- a swirldeck consisting of a number of swirltubes of standard size;
- a second mistmat (in the case of the SMSM; for demisting of the secondary gas only) downstream of the swirldeck to further improve the gas/liquid separation efficiency.

In this Section the various internals of the SMS, SVS and SMSM will be specified and subsequently design rules will be given for the three types.

Purchasing policy

For the SMS, SVS and SMSM separators the current SIOP procedure is to buy the vessel and total package of internals separately.

The internals shall be obtained only from a Manufacturer approved by the Principal.

Vessel attachments for the SMS, SVS and SMSM separators will be specified by the Manufacturer, but the Principal or Contractor shall specify the Schoepentoeter and swirldeck.

3.7.3 Schoepentoeter

This feed inlet device is used in all three types.
Its function is the bulk separation of the gas and liquid entering the separator.
For the detailed design see Appendix III.

In the particular case of SMS, SVS or SMSM separators a catcher cap is not allowed unless the remaining nozzle cross-sectional area is still sufficiently large to satisfy the momentum criterion (see Appendix II).

3.7.4 Coalescing mistmat

Both in the SMS and SMSM a mistmat is mounted between the Schoepentoeter and the swirldeck.

Under design conditions (i.e. the gas load factor λ = 0.25 m/s; see 3.7.8) the mistmat will be flooded.
Under those conditions, the mistmat will act as a coalescer: it will increase the droplet size of the mist passing through the mistmat. This increase in droplet size will facilitate the gas/liquid separation in the swirltubes. When the load factor drops below 0.10 m/s, the separation efficiency of the swirltubes will decrease. However, under those conditions the mistmat will function as an efficient separator. Due to this, both the SMS and SMSM have an excellent turndown ratio (10).

Manufacturing and mounting

For the manufacturing and mounting of the mistmat see Appendix VIII. Purchase of the mistmat should be left to the approved internals Manufacturer since the design is complicated due to the passage of the various drainpipes. The sub-supplier of the mistmat shall be subject to approval of the Principal.

3.7.5 Coalescing vane pack

In the SVS a vane pack rather than a mistmat is used as coalescer medium. For this application, a simple no-pocket vane pack is used. Due to its open structure it is less sensitive to fouling than a mistmat but its coalescing capacity is less as well.

Also, at low gas load factors, it will not take over the separator function from the swirldeck because it is, like the swirltubes, an inertial separator. Due to these factors both the turndown ratio and the efficiency of an SVS are lower than those of the SMS.

Manufacturing and mounting

The vane pack size shall match the vessel cross-sectional area and shall have an open area as large as possible.

The vane pack should have drain pipes extending at least 0.1 m below LZA(LL) to prevent accumulation of liquid on the mounting plate between the vane pack and the vessel wall.

In small vessels, the drain pipes are normally 25 mm diameter (at least one for every nine swirltubes) and in large vessels the drain pipes are 50 mm diameter (at least one for every 36 swirltubes).

If the vessel is not top-flanged, the vane pack shall be designed such that its parts can pass through the manhole.

3.7.6 Swirldeck

The SMS, SVS and SMSM all have a swirldeck consisting of swirltubes.

The swirltube:

The demister swirltube is an axial cyclone.
Schematic drawings are presented in Figure 3.10 of the swirltubes used in the SM(V)S and SMSM respectively.
The swirltube is, in essence, a 0.11 m ID stainless steel tube with a swirler at the inlet and longitudinal slits in the tube wall. Liquid is separated on the vanes of the swirler by impaction of droplets and on the tube wall by the centrifugal forces induced by the swirling gas flow.
Re-entrainment of this liquid is prevented by draining the film via the slits to the liquid collection chamber outside the tube. To ensure the proper functioning of the swirltube it is essential that some gas is also bled through these slits. This gas leaves the liquid collection chamber via the secondary outlets at the top and, in the case of the SMSM, passes the second mistmat. The main fraction of the gas leaves the swirltube via the primary gas outlet at the top. Drainpipes guide the liquid, collected in the space between the tubes and on the upper cover of the swirldeck, to below the liquid level.
Scaling-up of a separator equipped with a swirldeck is done simply by increasing the number of swirltubes proportional to the gas flow in the separator.
The swirltube has a weight of about 4 kg and has a wall thickness of 2 mm .
The weight of the swirldeck (including the swirltubes) is roughly 8 kg per swirltube.

The swirldeck (swirltray):

Depending on the application and the accessibility of the vessel, swirltubes can be combined in different ways to form a demister swirl tray.

Presently the following three types of swirl trays are used:

Integral swirl tray	(Figure 3.11a)
Swirl tray composed of boxes	(Figure 3.11b)
Swirl tray composed of banks	(Figure 3.11c)

The integral swirl tray is recommended for new vessels up to 1.2 m ID because it is easy to install and maintain. For this tray a full diameter top flange and sufficient hoisting height above the vessel are required.

The layout of this type of SMS is given in Figure 3.8a.

For larger diameter vessels and for existing vessels (e.g. for a revamp) the swirl tray should be composed of boxes.

Square boxes (L=0.28 m; B=0.28 m; H=0.51 m) are used as shown in Figure 3.11b, each box comprising four swirltubes and one drain pipe. Boxes are built together inside the vessel with panels at the edges. For the installation of these boxes a manhole with a minimum inside diameter of 20" is required.

This type of construction is relatively expensive and requires a relatively large vessel size. The alternative, however, is the use of a top flange (in conjunction with an integral swirl tray), which can be even more expensive.

In very large vessels (above about 3 metres ID) it is advantageous for the swirl tray to be composed of banks. Normally, a double row of swirltubes per bank (B=0.28 m; H=0.51 m) is used as shown in Figure 3.11c. This method of installation is less expensive than the installation with boxes, but a manhole of at least 26" ID is required.

The layout of an SMS with either a box-type or bank-type swirl tray is presented in Figure 3.8b.

The maximum number of swirltubes that can be fitted in a vessel is given as a function of the vessel ID in Table 2.

If the vessel cross-section is larger than that required for gas handling purposes, then the cross-sectional area shall not be filled up completely with swirltubes.

The maximum number of swirltubes to be installed is then determined by the criterion that

$\lambda_{st} \geq 0.5$ m/s.

The liquid separated off by the swirldeck (depositing on the bottom of the liquid collection box and on the top cover of the swirldeck) has to be transported to the liquid compartment via drainpipes. Each of these drainpipes should be fitted with a vertical plate in the entrance to act as a vortex breaker.

In the case of drainage of the liquid from the bottom of the swirldeck, the requirement is:

$d_{dp} \geq 1.5(n_{st,dp}Q_{L,max} / n_{st})^{0.5}$ with a **minimum of 0.025 m**

where

$n_{st,dp}$ is the number of swirltubes (typically four) served by the drain pipe

$Q_{L,max}$ is the maximum liquid flow rate in the feed,

n_{st} is the number of swirltubes in the vessel

and D is the vessel diameter.

If more than six swirltubes are used, these vertical drain pipes are combined in a horizontal header with diameter d_{hd}, for which:

$d_{hd} \geq 1.3(n_{st,hd}Q_{L,max} / n_{st})^{0.4}$ with a **minimum of 0.05 m**

where $n_{st,hd}$ is the total number of swirltubes served by the header.

The header ends in a vertical drain pipe of the same diameter extending at least 0.10 m below LZA(LL).

If it is not necessary to maintain a liquid level in the vessel, the drainpipe may be equipped with a flapper valve which opens in response to the weight of the liquid accumulated in the drain pipe. This latter option is subject to the approval of the Principal.

If horizontal collection headers are used, to avoid maldistribution of the gas over the swirltubes, a space of at least 0.5 m shall be present between the bottom of the swirldeck and the top of the coalescer (vane or mistmat) underneath the vane pack. This increased space will permit the lowering of the collection headers, so that their couplings with the drain piping from the swirldeck will no longer block the entrance of the swirltubes, resulting in a better gas distribution over the tubes.

If horizontal collection headers are not used, a space of 0.3 m is sufficient.

The liquid collected on the top of the swirldeck is transported to the liquid compartment of the vessel by separate drainpipes also extending at least 0.10 m below LZA(LL). In small vessels the drain pipes are normally 25 mm diameter (at least one for every nine swirltubes) and in large vessels the drain pipes are 50 mm diameter (at least one for every 36 swirltubes).

Table 2: Sizing of SMS (or SVS/SMSM) Separators

D (m)	Top flange (Y/N)	Type of swirldeck	Number of swirltubes	Q^*_{max} (m^3/s)	λ_{max} (m/s)
0.21	Y	Integral	1	0.0064	0.185
0.45	Y	Integral	4	0.0256	0.161
0.50	Y	Integral	5	0.0320	0.163
0.65	Y	Integral	9	0.0576	0.174
0.70	Y	Integral	12	0.0768	0.200
0.85	Y	Integral	16	0.102	0.180
0.90	Y	Integral	21	0.134	0.211
0.95	Y	Integral	24	0.154	0.217
1.05	Y	Integral	29	0.186	0.214
1.10	Y	Integral	32	0.205	0.216
1.15	Y	Integral	37	0.237	0.228
1.20	Y	Integral	44	0.282	0.249
1.30	N	Box	52	0.333	0.251
1.40	N	Box	57	0.365	0.237
1.45	N	Box	61	0.390	0.236
1.50	N	Box	68	0.435	0.246
1.60	N	Box	76	0.486	0.242
1.70	N	Box	88	0.563	0.248
1.80	N	Box	100	0.640	0.252

For large diameters calculate number of swirltubes as $30 * D^2$ and round up to a multiple of four.

3.7.7 Secondary mistmat

In the SMSM a secondary mistmat is employed downstream of the swirldeck to demist the fraction of the gas stream leaving the swirldeck via the secondary gas outlets.

The gross cross-sectional area of this mistmat shall at least be equal to the cross-sectional area of the swirldeck. The swirltube primary outlets shall penetrate neatly through the mistmat with a tight fit. In the SMSM the primary outlets have a horizontal slit 5 mm wide at the height of the secondary mistmat in order to catch the liquid film creeping up the inner wall of the primary gas outlet of the swirltubes. This slit shall be covered by the secondary mistmat (See also Figure 3.10). The mistmat, support and hold-down grid are stainless steel. The grids are of special design to ensure the mistmat flatness and positioning. For further details on the manufacturing and mounting of the mistmat see Appendix VIII.

3.7.8 Diameter

The gas handling capacity criterion:

$$\lambda \leq Q^*_{max} / (\pi D^2_{min} / 4) = 0.25 \quad \text{m/s}$$

or $$D \geq 2.26\sqrt{Q^*_{max}}$$

If two immiscible liquids are present in the feed and the flow rate of the lightest liquid is at least 5% vol. of the total liquid flow rate, then the physical properties of the lighter liquid shall be used in the gas handling criterion.

If the calculated D_{min} < 1.5 m, then D_{min} has to be recalculated using Table 2 in which the standard SMS separator vessel diameters are listed together with their Q^*_{max} and the maximum number of swirltubes (pitch 0.14 m) that can be fitted in the vessel.

The procedure is to select a value of Q^*_{max} that is at least as high as the value used in above formula. The required D_{min} is then the associated vessel diameter.

The smallest separator feasible has a diameter of 0.21 m and holds one swirltube.

If the gas handling capacity criterion is met, then the swirltube load factor, $\lambda_{st,}$ is computed as follows:

$$\lambda_{st} \leq Q^*_{max} / (A_{st} n_{st}) \leq 0.67 \qquad \text{m/s}$$

where n_{st} is the number of swirltubes in the swirldeck

and A_{st} is the swirltube cross-sectional area (= 0.0095 m^2)

or $Q^*_{max} / n_{st} \leq 0.0064$ m^3/s

It may be necessary to install a larger amount of swirltubes in the separator (with a corresponding increase of the vessel diameter) to allow for future increases of the gas flow rate.

It is recommended that the following criterion be used to determine the maximum allowable number of "active" swirltubes to be installed in the separator:

$$\lambda_{st} \geq 0.5 \qquad \text{m/s}$$

If more swirltubes are installed then the surplus tubes shall be blinded off.
To avoid gas bypass this blinding off shall take place at the inlet AND at the PRIMARY and SECONDARY gas outlets of these swirltubes.

NOTE Under special circumstances (e.g. retrofit of SMS in existing vessels) it is possible to install more swirltubes in the vessel than indicated in Table 2 by using a 0.13 m rather than a 0.14 m pitch. The consequence of this is that if the criterion for λ_{st} is maintained, λ_{max} will approach 0.30 m/s.
To check whether this is permissible from a process point of view, the Principal should be consulted.

The following criteria shall also be satisfied:

If liquid de-gassing is required:

$$D \geq 7608\sqrt{Q_{L,max}\eta_L / (\rho_L - \rho_G)}$$

This criterion is explained in Appendix VII.

If de-foaming is required:

$$D \geq 95Q^{0.5}_{L,max} \{\eta_L/(\rho_L - \rho_G)\}^{0.14}$$

The liquid handling capacity criterion

There should be a sufficient number of swirltubes in the swirldeck to keep the liquid load per swirltube below the maximum allowable level.

The swirltubes can handle up to 0.00015 m^3/s of liquid per tube.

The following calculation procedure should be followed:

1. Start with the minimum diameter required for gas handling, de-foaming and de-gassing
2. Calculate λ
3. Calculate the corresponding Schoepentoeter efficiency, Eff_{sch}

for $\lambda \leq 0.1$ m/s:$Eff_{sch} \geq 95\%$

for $0.1 < \lambda \leq 0.25$ m/s: $Eff_{sch} = 95 - 113(\lambda - 0.1)\rho_G^{0.14}/\varphi^{0.24}_{feed}\%$

with as (conservative) lower limit: $Eff_{sch} = 0\%$

4. Calculate the liquid load to the swirldeck, $Q_{L,sd}$

   $Q_{L,sd} = (1-Eff_{sch}/100)*Q_{L,max}$

5. Calculate the maximum number of swirltubes which can be held by the selected vessel diameter

   See Table 2 for $D \leq 1.8$ m; $n_{st} = 30D^2$ for $D > 1.8$ m

6. Calculate the liquid load per swirltube, $Q_{L,st}$

   If $Q_{L,st} \leq 0.00015$ m^3/s, the selected diameter satisfies the liquid handling criterion, otherwise repeat steps 2 - 6 with a stepwise increased vessel diameter until the criterion is met.

If for liquid handling such a high number of swirltubes has to be selected that λ_{st} drops below 0.5 m/s, then the choice of a separator of the SMS-family is probably not optimal and the Principal should be consulted.

3.7.9 Height

Let h be the height required for liquid hold-up (up to LZA(HH), see Appendix V), then the total vessel height, H, (tangent to tangent) is:

$$H = h + X_1 + X_2 + X_3 + X_4 + X_5 + 0.51 + X_6 \quad m$$

All the terms on the right of the above equation are indicated in the drawings of the SMS, SVS and SMSM separator (Figures 3.8a, 3.8b, 3.9a and 3.9b respectively).

d_1 = internal diameter of the inlet nozzle
X_1 = 0.05 D with a minimum of 0.15 m
X_2 = d_1 + 0.02 m
X_3 = d_1 with a minimum of 0.3 m
X_4 = 0.1 m in case of mistmat (SMS(M))
X_4 = 0.2 m in case of vane pack (SVS)
X_4 = 0.2 D
X_5 ≥ 0.5 m if horizontal collection headers are used underneath the swirldeck
X_5 ≥ 0.3 m if horizontal headers are not used
X_6 = 0.15 D in case of SMS or SVS
X_6 = 0.15 D + 0.3 m in case of SMSM

The distance between LZA(HH) and the bottom of the swirldeck shall be sufficiently large to accommodate the liquid head in the drainpipes caused by the pressure drop between the gas compartment below the coalescing vane pack or the coalescing mistmat and the liquid compartment of the swirldeck, $(p_{in} - p_{L,out})$.

Recommendation: $X_1 + X_2 + X_3 + X_4 + X_5 > 1.5(p_{in} - p_{L,out})/\{g(\rho_L - \rho_G)\}$

In SMS and SMSM: $p_{in} - p_{L,out} = 0.5\Delta p_{st} + \Delta p_{wm}$

In SVS: $p_{in} - p_{L,out} = 0.5\Delta p_{st} + \Delta p_v$

For the calculation of the various pressure drops, see (3.7.11)

Onder normal operating conditions, if the pressure drop over the coalescing device does not become excessively high due to hydrate formation for instance, the requirement for the drainpipe length is always fulfilled.

3.7.10 Nozzles

The feed nozzle shall be fitted with a Schoepentoeter if $D \geq 0.5$ m.

For the design of Schoepentoeters, see Appendix III.

For the sizing of the feed nozzle, see Appendix II.

If $D < 0.5$ m, then a half-open pipe should be used (in which case the separator would no longer strictly be classified as an SMS, SVS or SMSM separator).

For the sizing of the gas and liquid outlet nozzles see Appendix II.

3.7.11 Pressure drop

The pressure differential between inlet and vapour outlet is basically the sum of the pressure drops over the nozzles, the coalescing medium and the swirldeck.

The pressure drop across both the Schoepentoeter and the secondary mistmat (in the case of the SMSM) is negligible.

$$P_{in} - P_{out} = 0.5\rho_m v^2_{m,in} + 0.22\rho_G v^2_{G,out} + \Delta p_{wm} \text{ (or } \Delta p_v) + \Delta p_{sd}$$

where $\Delta p_{wm} = 20(\rho_L - \rho_G)\,\lambda^2$ Pa (in SMS and SMSM)

$= 2\,000\,\lambda^2$ mm process liquid

($\approx$ 125 mm at maximum gas load conditions).

NOTE This is the wet pressure drop; the dry pressure drop across the mistmat is about 50% of this value.

In the above expression for the pressure drop, an averaged correction for the liquid loading has been taken into account.

$\Delta p_v = 4(\rho_L - \rho_G)\,\lambda^2$ Pa (in SVS)

$= 400\,\lambda^2$ mm process liquid

($\approx$ 25 mm at maximum gas load conditions)

$\Delta p_{sd} = 8.4(\rho_L - \rho_G)\,\lambda_{st}2$ Pa

$= 840\,\lambda_{st}^2$ mm process liquid

($\approx$ 375 mm at maximum gas load conditions)

NOTE In the **last** equation the **SWIRLTUBE** load factor is used whereas in the **other** equations the **VESSEL** load factor is used.

For the pressure difference between the gas compartment below the swirldeck and the liquid compartment of the swirldeck, $p_{in} - p_{L,out}$ holds:

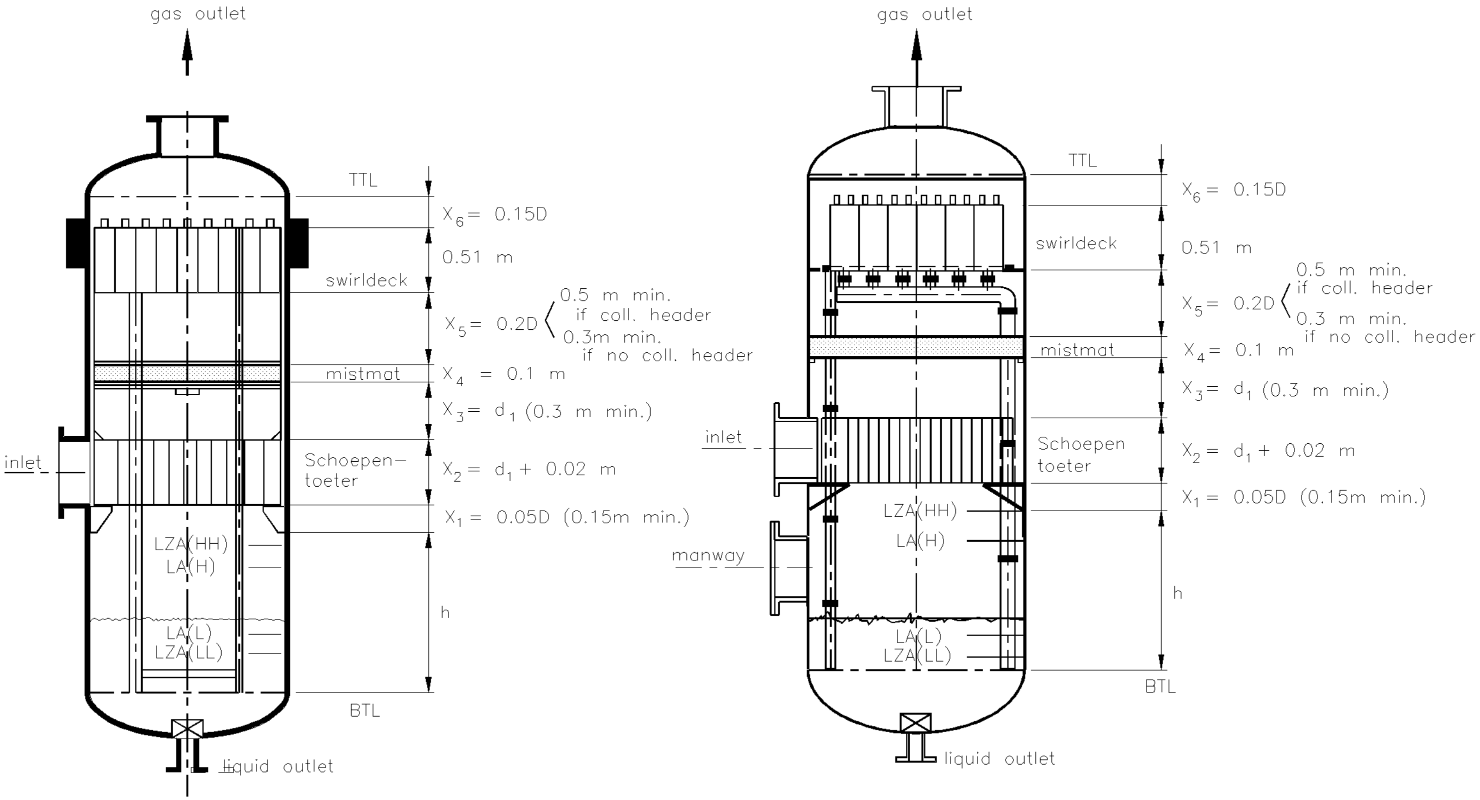
$$p_{in} - p_{L,out} = 0.5\Delta p_c$$

Pressure drop measurement

To monitor the condition of the internals (e.g. degree of fouling), it is recommended to install a pressure differential measurement across the coalescing medium and the swirldeck.

An example of the detailed design of an SVS Separator is given in Appendix X.

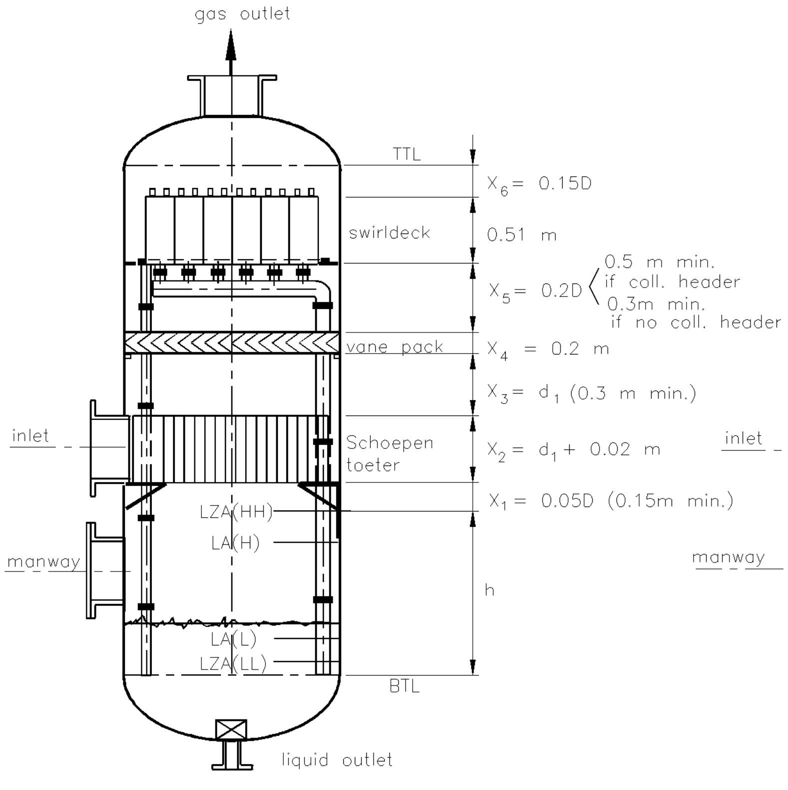
FIGURE 3.8 THE SMS SEPARATOR



(a) D ⩽ 1.2 m
vessel is top–flanged and
an integral swirl tray is used

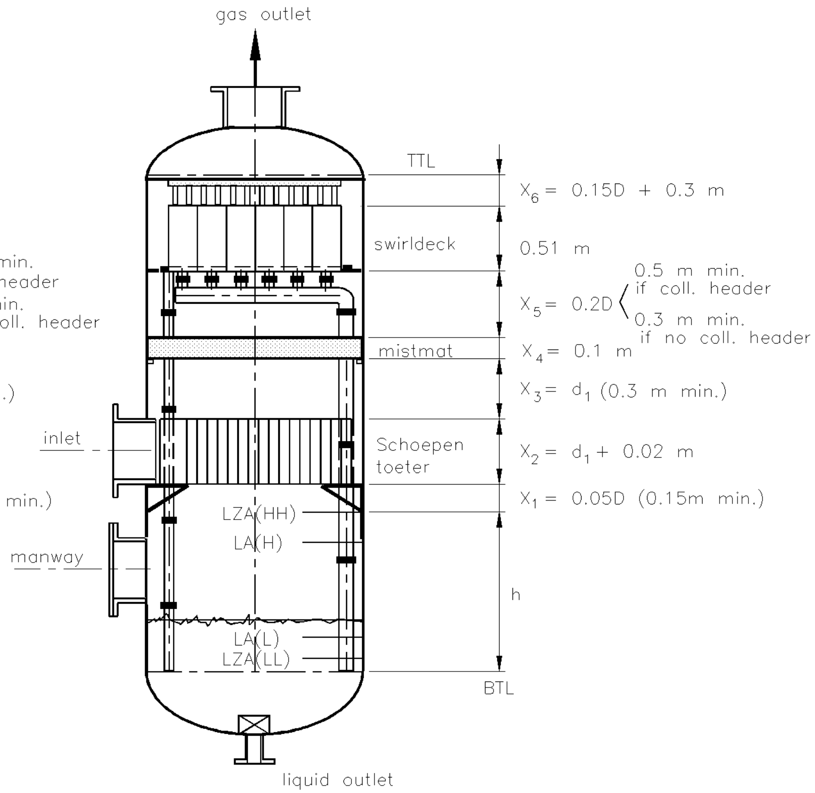
(b) D > 1.2 m
swirl tray either of box or bank type

FIGURE 3.9 THE SVS AND SMSM SEPARATORS



(a) SVS

(drain pipes of vane pack not shown)



(b) SMSM

FIGURE 3.10 SCHEMATIC OUTLINE OF A SINGLE SWIRLTUBE -WITHOUT AND WITH SECONDARY DEMISTING (SMSM)

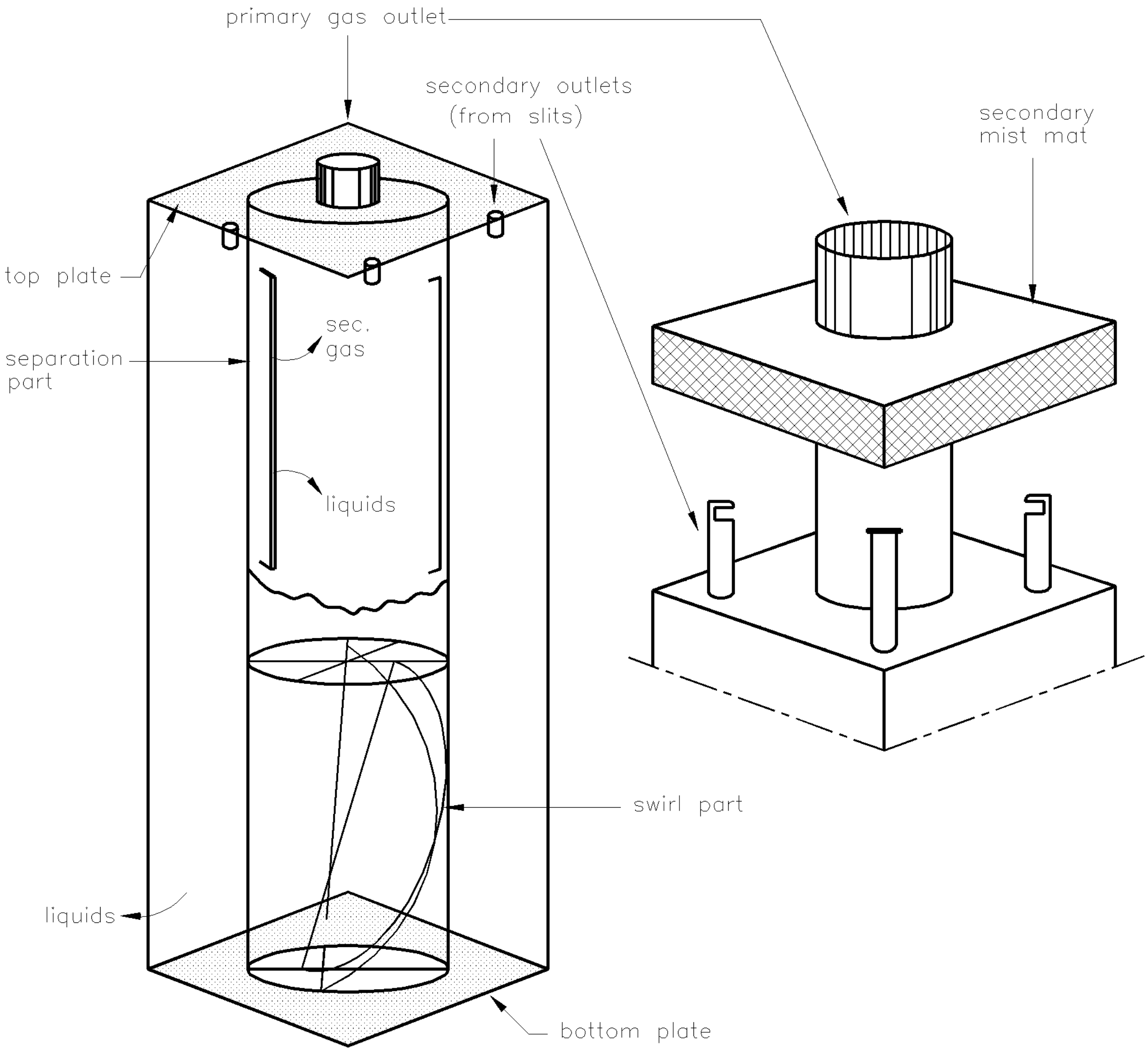
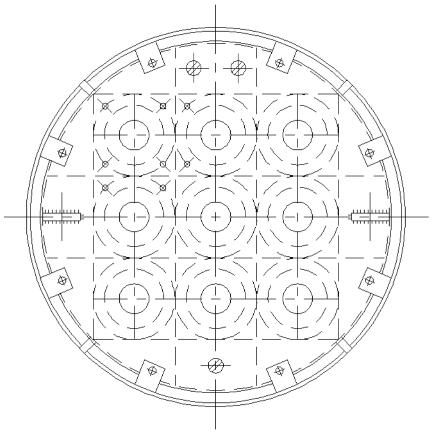
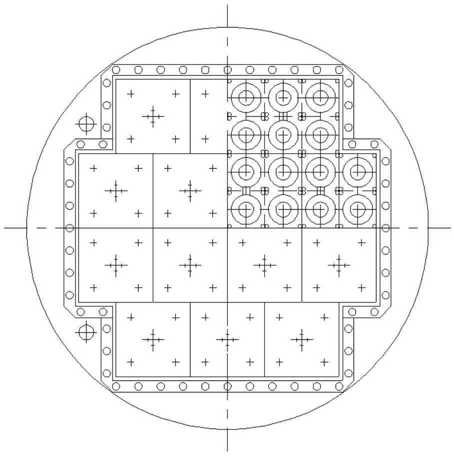


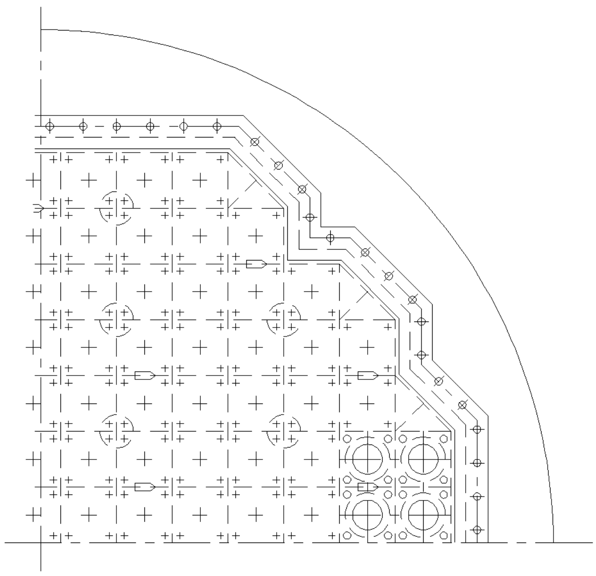
FIGURE 3.11 THE VARIOUS TYPES OF SWIRL TRAYS (SCHEMATIC)



(a) integral



(b) composed of boxes



(c) composed of banks

3.8 CYCLONE WITH TANGENTIAL INLET (CONVENTIONAL CYCLONE)

(Figure 3.12)

3.8.1 Selection criteria

Application:

- demisting of gas in fouling service

Characteristics:

- liquid removal efficiency > 96%;
- insensitive to fouling;
- limited turndown ratio (factor 2);
- high pressure drop.

Recommended use:

- typically for use in a fouling (e.g. coke-formation) environment and where a high demisting efficiency is still required.

Non-recommended use:

- if high pressure drop cannot be tolerated.

Typical process application:

- in oil refineries: Thermal Gasoil Units (TGU); Visbreaker Units (VBU);
- in chemical plants: Thermoplastic Rubber Plants.

3.8.2 General

The cyclone contains the following basic elements:

- tangential inlet.

  In Figure 3.12 two options are given for this. Although type 1 generates a more pronounced swirl and therefore a slightly higher separation efficiency, for constructional reasons it shall not be used if the difference between the operating pressure and the pressure outside the cyclone body exceeds 5 bar.

  In both types rectangular inlets are also acceptable with the height being larger than the width. This will lead to more compact cyclones, particularly in low pressure (below 5 bar) applications where a flat rather than domed top cover can be used.

- cylindrical baffle at the top of the cyclone (drip ring).

  This baffle will improve the liquid removal efficiency because it stops the liquid film creeping over the ceiling of the cyclone to the vortex finder.

  However, if coking is expected (TGU or VBU), a drip ring shall NOT be used

- vortex finder

- bottom (dollar) plate with vortex spoilers.

  This plate is essential to shield the liquid-filled bottom compartment from the gas stream in the cyclone, in order to avoid liquid pick-up by the gas.

  In this Section, general design rules are given for cyclones.

  For design details of special-duty cyclones (such as the type of cyclone used in Thermal Crackers (TC), for instance) the Principal should be consulted.

3.8.3 Diameter

The diameter of the cyclone shall be related to the size of the inlet as given below.

(See also Figure 3.12)

(The sizing of the cyclone inlet is based on the **ACTUAL** flow rates, i.e. **EXCLUDING** a design margin).

Type 1 (rectangular inlet):

$a \geq 0.0007\{Q_G(\rho_L\text{-}\rho_G)/\eta_G\}^{0.333}$

$a \geq 0.128\ \rho_G^{0.25}\ Q_G^{0.5}$ (equivalent to $\rho_G v^2_{G,in} \leq$ 3750 Pa)

$a \geq Q_L^{0.5}$ (equivalent to $v_{L,in} \leq$ 1 m/s)

If erosive material is present in feed:

$a \geq 0.2\ Q_G^{0.5}$ (equivalent to $v_{G,in} \leq$ 25 m/s)

If the liquid has a foaming tendency:

preferably* $a \leq 0.316\ Q_G^{0.5}$ (equivalent to $v_{G,in} \leq$ 10 m/s)

The smallest value of "a" satisfying the above equations should be taken. Note that in these equations the coefficient in the right hand side is not dimensionless. Therefore only SI units shall be used.

$D \geq 2.8a$

$D \geq 0.652\ Q_G^{0.5}$ (mean gas velocity in cyclone body ≤ 3 m/s)

The smallest value of D satisfying the above equations shall be taken.

* NOTE If this criterion conflicts with the preceding inlet sizing criteria (e.g. in high pressure applications) it shall be overruled by the other criteria.

Type 2 (circular inlet):

$d_1 \geq 0.0007\{Q_G(\rho_L - \rho_G)/\eta_G\}^{0.333}$

$d_1 \geq 0.144\rho_G^{0.25}Q_G^{0.5}$ (equivalent to $\rho_G v^2_{G,in} \leq$ 3750 Pa)

$d_1 \geq 1.13\ Q_L^{0.5}$ (equivalent to $v_{L,in} \leq$ 1 m/s)

If erosive material is present in feed:

$d_1 \geq 0.226\ Q_G^{0.5}$ (This is equivalent to $v_{G,in} \leq$ 25 m/s)

If the liquid has a foaming tendency:

preferably $d_1 \leq 0.356\ Q_G^{0.5}$ (This is equivalent to $v_{G,in} \geq$ 10 m/s)

(the criterion $10 \geq v_{G,in} \leq$ 25 m/s holds for Thermal Cracker cyclones for instance)

NOTE If this criterion conflicts with the preceding inlet sizing criteria (e.g. in high pressure applications) it shall be overruled by the other criteria.

The smallest value of d_1 satisfying the above equations should be taken. Note that again in these equations the coefficient on the right hand side is not dimensionless. Therefore only SI units shall be used.

The diameter of the cyclone is then given by:

$D \geq 3.5d_1$

$D \geq 0.652\ Q_G^{0.5}$ (mean gas velocity in cyclone body ≤ 3 m/s)

The smallest value of D satisfying the above equations shall be taken.

3.8.4 Distance between vortex finder and bottom plate

The distance between the lower end of the vortex finder and the top of the bottom plate,

H_{vfb}, is a function of the ratio of the liquid and gas mass flow rates.

There is a tendency for the averaged mist droplet size in the feed to decrease if the above ratio decreases. Then a larger distance is required between the vortex finder and the bottom plate to achieve proper demisting.

$$H_{vfb}/D \geq 0.7 - 0.5 \log_{10}(M_L/M_G)$$

$$H_{vfb}/D \geq 0.5$$

M_L and M_G are the mass flow rates of liquid and gas respectively in kg/s.

3.8.5 Drip ring

The drip ring (anti-creep baffle) is concentric to the vortex finder and has as minimum diameter $d_4 = (D + d_2) / 2$ (see also Figure 3.12).

The vertical distance between the top of the inlet nozzle and the lower end of the drip ring should be 0.1a (cyclone type 1) or $0.1d_1$ (cyclone type 2).

If coking is expected (in Thermal Cracker service for instance) a drip ring shall NOT be used.

3.8.6 Vortex finder

The diameter of the vortex finder is normally equal to that of the outlet nozzle, d_2.

Its lower end shall extend at least 0.2a (cyclone type 1) or at least $0.2d_1$ (cyclone type 2) below the bottom of the feed inlet nozzle. In order to reduce liquid re-entrainment it is recommended to select a vortex finder with an entrance diameter larger than that of the outlet nozzle (by say 0.3 m).

3.8.7 Bottom (dollar) plate

It is important that the liquid level is always below the bottom plate. Therefore the lower end of the (conical) bottom plate shall be at least 0.1 m above LZA(HH).

For proper drainage of liquid settled on the bottom plate to the liquid compartment of the cyclone, a bottom plate top angle of 160° is specified. The upper surface of the bottom plate shall have no protrusions. If a manhole is required in the bottom plate then it shall be installed flush with the top of the bottom plate.

Vortex spoilers shall be present underneath the bottom plate (see Figure 3.12).

It is recommended to use four liquid vortex spoilers with a maximum width of D/8. In TC cyclones the practice is to minimise the width of the vortex spoilers as much as possible in order to reduce coke formation.

The gap, s, between the bottom plate and the cyclone wall shall satisfy the following:

Under **non-foaming, non-coking conditions:**

$s = 0.025D$ m, with a minimum of 0.01 m

Under **foaming** conditions:

$s = 0.05D$ m, with a minimum of 0.02 m

Under **coking** conditions:

$0.05 \leq s \leq 0.1$ m

3.8.8 Height

Let h be the height required for liquid hold-up (up to LZA(HH)), then the cyclone height from BTL to the lower end of the drip ring at the top of the cyclone is:

$$H = h + X_1 + H_{vfb} + X_2 + X_3 + X_4$$

All the terms on the right hand side of the above equation have been indicated in Figure 3.12.

$X_1 \geq 0.1 + 0.5D \tan(10°)$ m

$X_2 \geq 0.2a$ (cyclone type 1) or $\geq 0.2d_1$ (cyclone type 2)

$X_3 = a$ (cyclone type 1) or $= d_1$ (cyclone type 2)

$X_4 = 0.1a$ (cyclone type 1) or $= 0.1d_1$ (cyclone type 2)

3.8.9 Nozzles

The sizing of the feed nozzle has already been given in (3.8.3).

Bends in the inlet piping are only permitted if they are in a horizontal plane and the curvature is in the same direction as the cyclone vortex.

The flow area of the gas outlet should be at least equal to that of the inlet.

For the sizing of the liquid outlet nozzle see Appendix II.

3.8.10 Pressure drop

Pressure differentials between inlet, gas outlet and liquid outlet are as follows:

(pressure drop is based on **MAXIMUM** flow rates, i.e. **INCLUDING** the design margin).

$$P_{in} - P_{G,out} = x\rho_{G,in}v^2_{G,in}$$

$$P_{in} - P_{below\ bottom\ plate} = y\rho_{G,in}v^2_{G,in}$$

x and y are a function of both the cyclone geometry and the feed gas/liquid ratio.

It is recommended to use the following typical (slightly conservative) values:

For both cyclone types: x = 8 and y = 2.

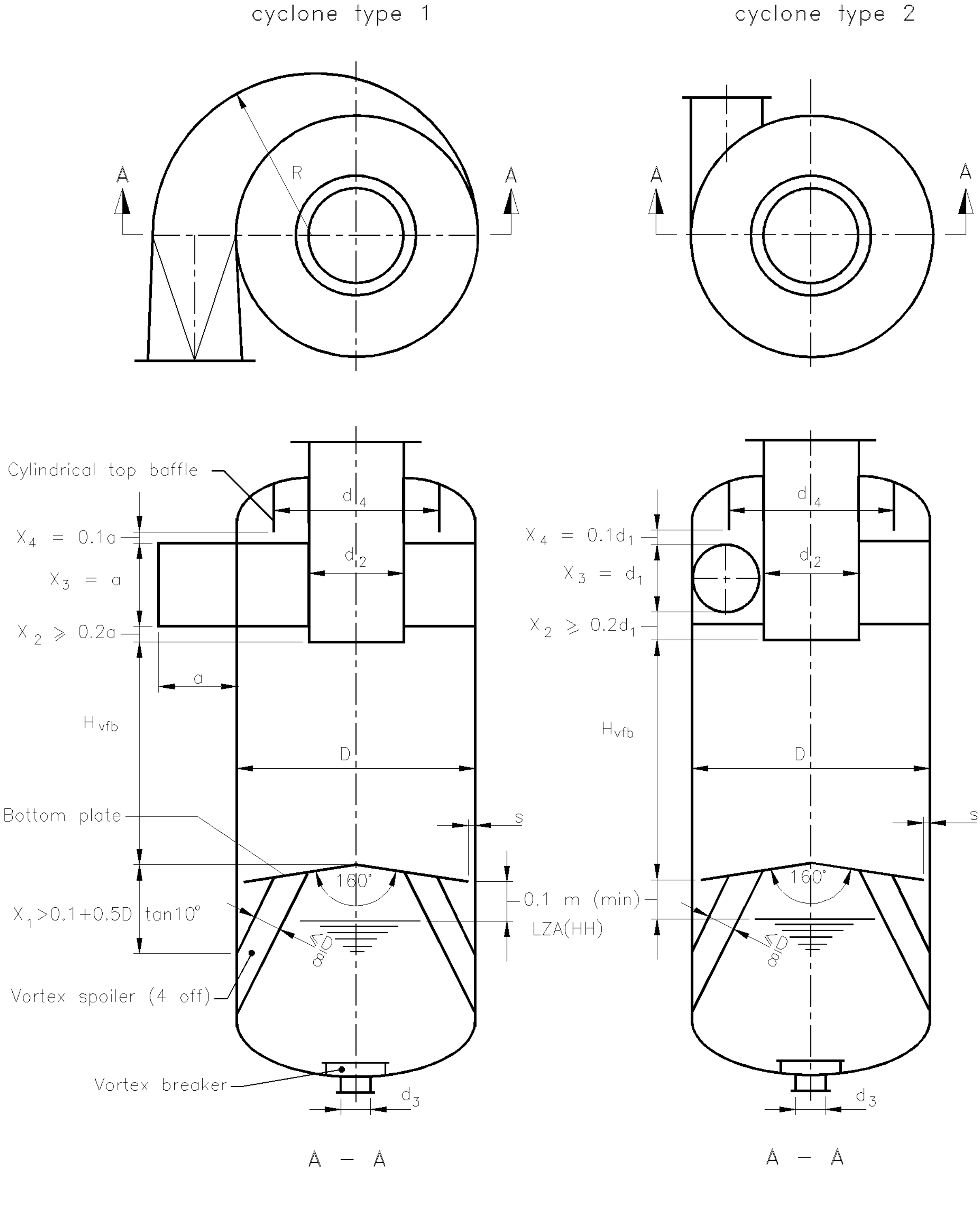
3.8.11 Liquid drain sealing

For satisfactory operation of any cyclone, it is essential that the liquid drain is sealed, i.e. there shall be no flow of gas in either direction of the drain.

Liquid from the cyclone is sometimes drained via a dip leg to a receiving vessel. In such cases it is essential that a pressure balance is calculated to ensure **that the liquid level always remains below the bottom plate.**

If the level rises above this plate, severe entrainment will result.

FIGURE 3.12 CYCLONE WITH TANGENTIAL INLET



3.9 CYCLONE WITH STRAIGHT INLET AND SWIRLER ("GASUNIE" CYCLONE)

(Figure 3.13)

3.9.1 Selection criteria

Application:

- demisting of gas where a high gas handling capacity and a high liquid removal efficiency is required.

Characteristics:

- very compact separator;
- high liquid removal efficiency (> 99%);
- very high gas handling capacity (maximum allowable gas load factor 0.9 m/s);
- high turndown ratio (factor 7);
- high pressure drop;
- suitable for slightly fouling service (e.g. low sand loading);
- slug handling capacity.

Recommended use:

- where there is little plot space available (e.g. in offshore industry or in general for high-pressure conditions);
- as retrofits of existing vessels where capacity debottlenecking and/or improved separation efficiency are required.

Non-recommended use:

- if a low pressure drop is essential;
- if insufficient head room is available.

Typical process application:

- wellhead separators;
- compressor suction and interstage scrubbers;
- cold separators;
- inlet separators to adsorption plants.

3.9.2 General

This type of cyclone has been developed by the Gasunie, the Dutch gas distribution company. It will be supplied as a complete package by the Vendor (a licensee of Gasunie) based on the Gasunie-proprietary design. The rules given below can be used to verify the proprietary Vendor's design.

A schematic layout of this cyclone is presented in Figure 3.13 and its working principle is described below.

The gas with entrained liquid enters the cyclone via a straight side inlet and is brought into rotation by a swirl element. Due to this rotation the entrained liquid is separated from the gas and driven to the cyclone wall. Subsequently it drains via the wall to the liquid compartment of the cyclone.
The cleaned gas leaves the cyclone via the vortex pipe or vortex finder. (The vortex pipe is the central pipe attached to the gas outlet).

Several measures have been taken to promote the separation of the liquid from the gas and minimise re-entrainment of the separated-off liquid in the gas.

1. Vanes present on the vortex pipe promote the gas/ liquid separation by guiding the liquid to the cyclone wall.
2. The lower end of the vortex finder is shaped such that creep of liquid into the vortex finder due to local underpressure is minimised.
3. A conical cover plate ("Chinese hat") is present underneath the vortex finder which prevents re-entrainment of liquid from the liquid compartment into the gas flow.
   LZA(HH) shall be below the "Chinese hat".

3.9.3 Diameter

The diameter shall satisfy the following critera:

1. The gas handling capacity criterion:

$\lambda \leq Q^{*}_{max} / (\pi D^{2}_{min} / 4) = 0.9$ m/s

or $D \geq 1.19 \sqrt{Q^{*}_{max}}$

If two immiscible liquids are present in the feed and the flow rate of the lightest liquid is at least 5% vol. of the total liquid flow rate, then the physical properties of the lighter liquid shall be used in the gas handling criterion.

2. Pressure drop criterion

$D \geq \{3360 \rho_G Q^2_{G,max} / (\pi^2 \Delta P_{max}\}^{0.25}$

ΔP_{max} is the maximum allowable pressure drop allowed by the process conditions downstream of the cyclone.

In general, the pressure criterion will determine the cyclone diameter

If the calculated cyclone diameter is larger than 1.5 m several separators of this type should be used in parallel.

3.9.4 Height

The distance from the bottom tangent line to LZA(HH) shall be at least 1.0 D.

The distance from LZA(HH) to the top tangent line shall be at least 3.2 D.

Consequently the total minimum height (from bottom to top tangent line) is 4.2 D

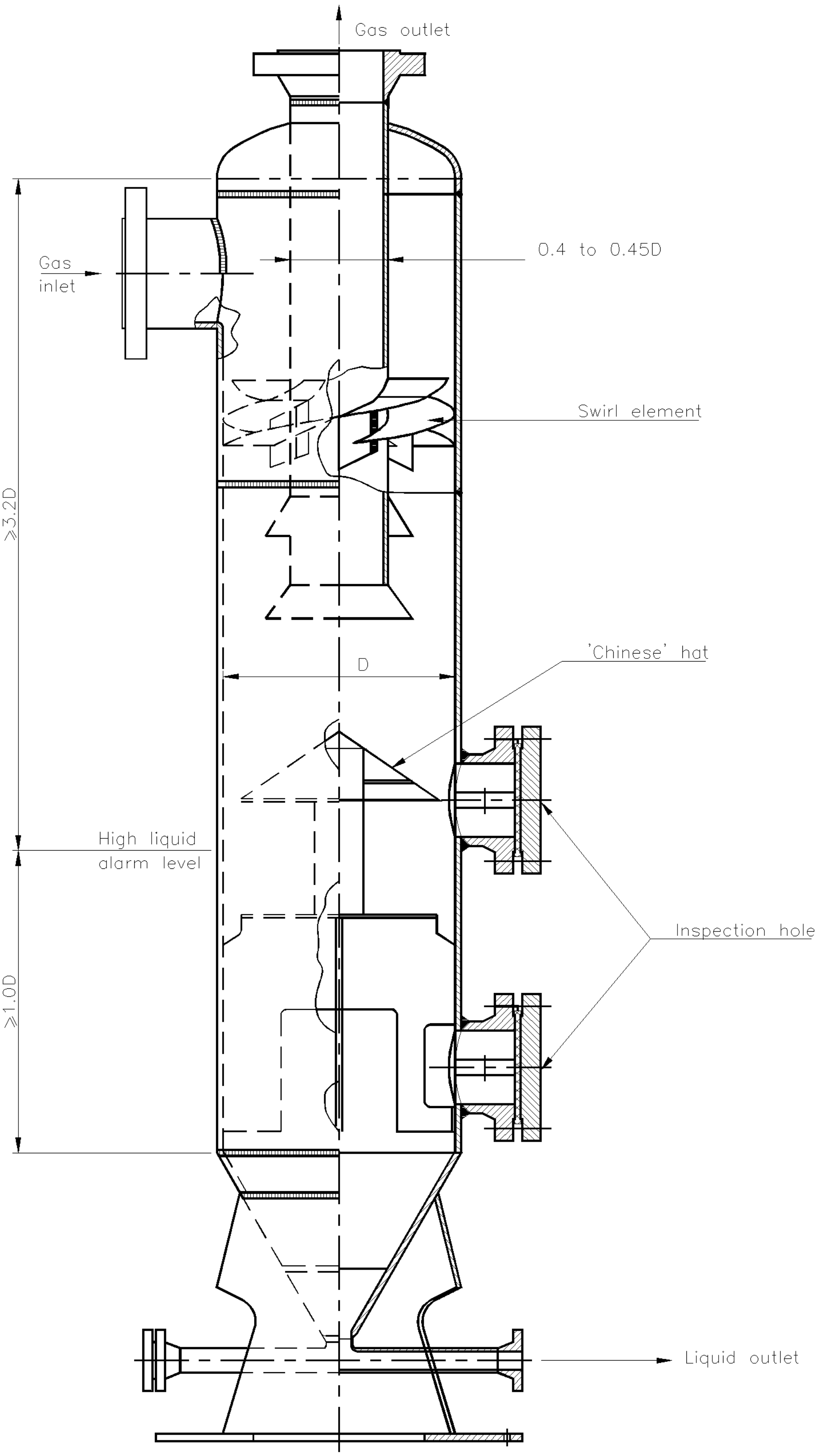
3.9.5 Droplet removal performance

From the cyclone diameter, gas flow rate and density difference between the gas and liquid phase, the droplet removal performance of the cyclone can be assessed by the calculation of d_{crit50} and d_{crit99} (the diameter of liquid droplets in the feed with a chance of removal in the cyclone of respectively 50% and 99%) with the following relationships:

$$d_{crit50} = 223 * \sqrt{D^3 / (Q_{G,max} (\rho_L - \rho_G))} \quad \mu m$$

$$d_{crit99} = \sqrt{2} * d_{crit50} \quad \mu m$$

FIGURE 3.13 CYCLONE WITH STRAIGHT INLET AND SWIRLER ("GASUNIE CYCLONE")



3.10 VERTICAL SEPARATOR WITH REVERSED-FLOW MULTICYCLONE BUNDLE (CONVENTIONAL MULTICYCLONE)

(Figure 3.14)

3.10.1 Selection criteria

Application:

- demisting and dedusting of gas in slightly fouling service and high pressure.

Characteristics:

- liquid removal efficiency > 93%;
- suitable for slightly fouling service (e.g. low sand loading);
- high pressure drop;
- compact separator;
- sensitive to high liquid loading or slugs.

Recommended use:

- typically for use in a slightly fouling environment where the gas pressure is higher than 100 bar (abs) and a compact separator is required.

Non-recommended use:

- low gas pressure;
- heavy fouling service (high sand loading will cause erosion);
- high liquid loading;
- slugs;
- when high liquid removal efficiency is required.

Typical process application:

- wellhead separators;
- primary scrubbers under slightly fouling service and when the liquid loading is low;
- compressor suction scrubbers if sand is present in the feed.

3.10.2 General

Normally the multicyclone separator is supplied as a complete package of vessel and internals based on a Vendor-proprietary design. The rules given below can be used to verify the proprietary Manufacturer's design.

A vertical multicyclone separator is a vertical vessel in which an array of parallel small cyclones are fitted between a top and bottom plate. In this way a chamber is created which is shielded from the top and bottom compartment of the vessel.
The feed flows directly into this compartment and enters the cyclones via their tangential inlets.
The gas/liquid separation takes place in these cyclones. Subsequently the cleaned gas flows to the upper vessel compartment and the separated liquid is drained to the bottom compartment.

3.10.3 Cyclones

Normally, the cyclones are tangential (reversed flow) cyclones with a 2" diameter and with two tangential, diametrically-opposed inlets.
The advantage of using a multicyclone rather than a monocyclone separator is that the same vessel can be used for a wide variety of flow conditions. For prolonged periods of relatively low flow rates, the separation efficiency is maintained by blinding off some of the multicyclones.
Scaling up of a multicyclone separator is done simply by increasing the number of cyclones proportionally to the gas flow in the separator.

The multicyclone bundles are Manufacturer-proprietary devices. For information on approved Manufacturers, the Principal should be consulted.

Unlike other types of gas/liquid separators the volumetric gas handling capacity of the cyclones is insensitive to gas pressure. At high gas pressures (between 90 and 180 bar) field experience has shown that it is necessary to satisfy the following requirements for each cyclone in order to maintain a liquid removal efficiency of at least 93%:

$$0.003 \textbf{ actual } m^3/s \le Q_{G,c} \le 0.007 \textbf{ actual } m^3/s$$

$$\phi_{feed} < 0.003$$

slugs are **not** allowed

NOTES

1. $Q_{G,c}$, the volumetric gas flow rate per cyclone, is in units of actual m^3/s.
2. If the upper flow limit is complied with, erosion is also minimised.
3. If the reversed-flow multicyclone is to be used at a much lower pressure than 90 bar, the prescribed upper limit for both the gas flow rate and flow parameter is conservative. In such cases, the Principal should be consulted.
4. If cyclones have to operate for a prolonged period at a gas flow rate lower than 0.003 m^3/s, it is recommended to temporarily plug a number of cyclones at their top and bottom outlets, thereby obtaining actual loadings of Q_G > 0.005 actual m^3/s. Such plugs are supplied by the multicyclone Manufacturers.
5. To increase the liquid handling capacity of a reversed-flow multicyclone separator, Manufacturers also offer a two-stage separator in which the first stage has a liquid knock-out function. However, since the pressure drop across this type of (reversed-flow) cyclones is relatively high, additional vessel height is required since the drain pipe from the bundle has to be sufficiently long to accommodate the liquid pushed up into the pipe in the liquid compartment by the pressure difference. An alternative solution to overcome the liquid back-up problem is to increase the number of cyclones. The drawback of this is that apart from a possible decrease of liquid removal efficiency, a larger vessel diameter is required. Other types of separator might then be preferred.

   Because of the above reasons the two-stage reversed-flow multicyclone separator is not attractive.

3.10.4 Reversed-flow multicyclone bundle

The multicyclone bundle layout should satisfy the following requirements:

1. Cyclone-cyclone pitch should be at least 80 mm.
2. All cyclone inlets shall be at the same level. Vertical staggering shall not be employed to save space.
3. To minimise maldistribution of the feed over the cyclones, it is recommended (particularly if the vessel diameter is larger than 1.5 m) to have a cyclone-free sector facing the feed inlet of the vessel.
4. Top and bottom cover shall be gas-tight.
5. It is recommended to have the multicyclone bundle removable in one piece.

To monitor the condition of the multicyclone bundle (leakage of top/bottom plate and/or cyclone erosion), it is recommended to install a pressure differential measurement across the multicyclone bundle. Loss of pressure drop will indicate deterioration of the bundle.

3.10.5 Vessel accessibility

Because of the risk of cyclone erosion, regular inspection of the internals is required.
The design of the separator should therefore enable the removal of the complete cyclone bundle and the vessel should be equipped with a top flange or a full-diameter top cover.

3.10.6 Diameter

The vessel diameter, D, of the multicyclone separator shall be large enough to hold sufficient cyclones in order to meet **the gas handling capacity criterion.**

The number of cyclones required, n_c, is given by:

$$n_c = Q_{G,max} / 0.007$$

with n_c rounded up to the nearest integer.

$Q_{G,max}$ is the maximum **actual** volumetric flow rate and shall include the appropriate design margin (see Appendix IV).

The denominator of the right hand side of the above equation is the maximum actual flow rate per cyclone (m^3/s).

The vessel diameter shall also satisfy the following criteria:

If liquid de-gassing is required:

$$D \geq 7608 \sqrt{Q_{L,max} \eta_L / (\rho_L - \rho_G)}$$

This criterion is explained in Appendix VII.

If de-foaming is required:

$$D \geq 95 Q^{0.5}_{L,max} \{\eta_L / (\rho_L - \rho_G)\}^{0.14}$$

In practice, these two criteria will almost always be satisfied because of the limitations of the feed flow parameter ($\phi_{feed} < 0.003$).

3.10.7 Height

Let h be the height required for liquid hold-up (up to LZA(HH)), then the total vessel height, H (tangent to tangent or bottom tangent to flange face if a full vessel flange is fitted), is typically

for vessel with gas outlet at the top (Figure 3.14a):

$$H = h + 0.6 + 0.15 + d_0 + 0.15 + 0.2D \qquad \text{m}$$

for vessel with gas outlet at the vessel side (Figure 3.14b):

$$H = h + 0.6 + 0.15 + d_0 + 0.2D + d_0 + 0.7D \qquad \text{m}$$

In the above equations, d_0 is the outer diameter of the inlet nozzle.

3.10.8 Nozzles

The size of the feed nozzle and of the gas outlet nozzle may be taken to be equal to the pipeline size, but it shall be ensured that:

$$\rho_G v^2_{G,noz} \leq 3750 \qquad \text{Pa}$$

Bends in the inlet piping are only permitted if they are in the horizontal plane and the curvature is in the same direction as the vortex of the cyclones.

For the sizing of the liquid outlet nozzle see Appendix II.

3.10.9 Pressure drop

The pressure drop across the multicyclone separator is:

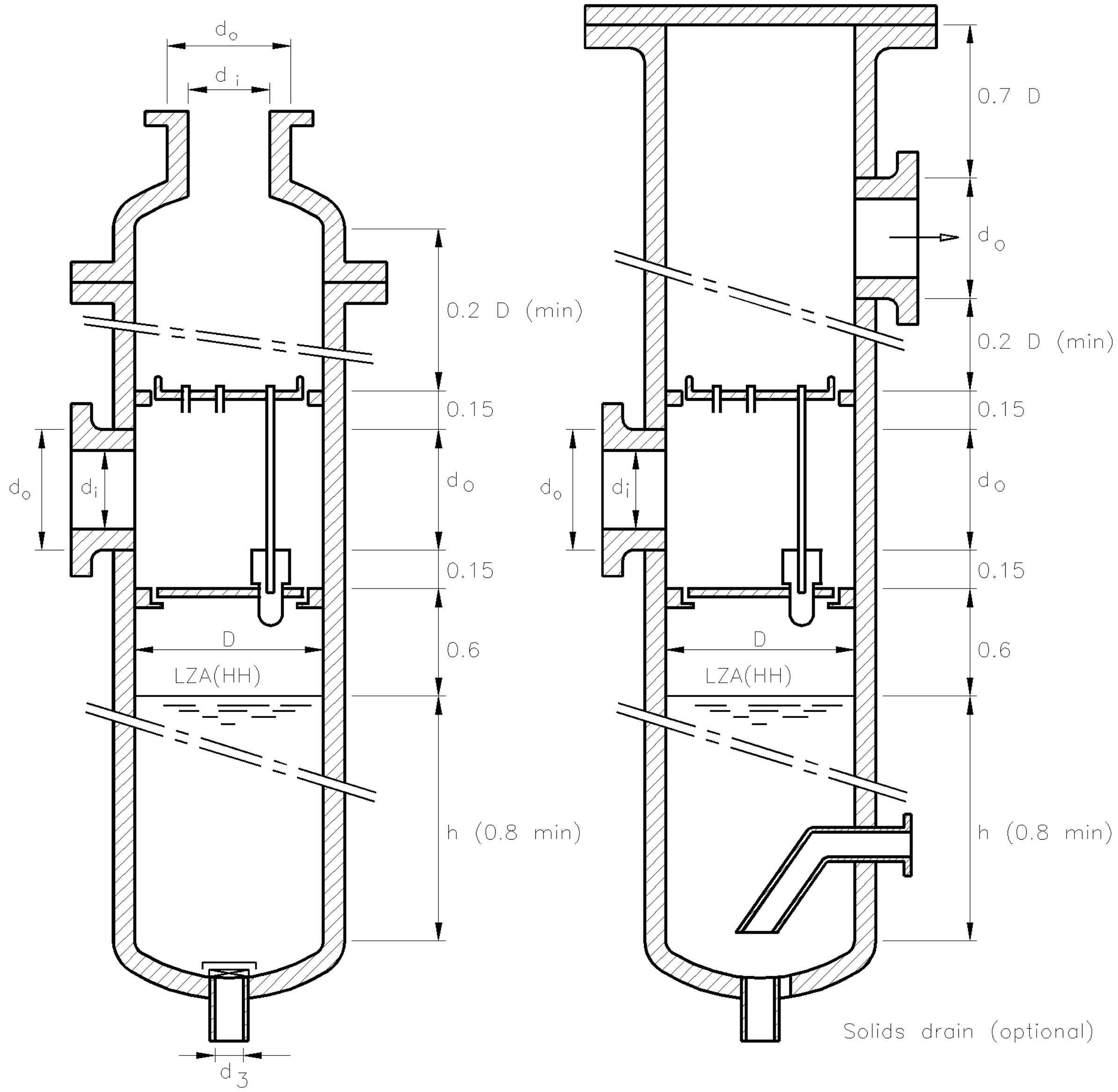
$$P_{in} - P_{out} = K_c \rho_G v^2_{G,c}$$

Typically, K_c = 17 for 2 inch diameter cyclones.

Thus, at maximum gas flow (assuming ρ_G = 120 kg/m^3), the pressure drop is 0.24 bar.

The pressure difference between the vapour inlet and the space below the cyclones is about half the pressure drop across the separator.

FIGURE 3.14 THE VERTICAL SEPARATOR WITH REVERSED FLOW MULTI-CYCLONE BUNDLE (CONVENTIONAL MULTI-CYCLONE)



(a) Separator with top gas outlet

(b) Separator with side gas outlet

In the separators only one cyclone of the multicyclone bundle is shown.

3.11 VERTICAL SEPARATOR WITH AXIAL-FLOW MULTICYCLONE BUNDLE (WITH RECIRCULATION OF SECONDARY GAS)

(Figure 3.15)

3.11.1 Selection criteria

Application:

- demisting of gas where a high gas handling capacity and a high liquid removal efficiency is required.

Characteristics:

- very compact separator;
- high liquid removal efficiency (> 99%);
- very high gas handling capacity;
- high turndown ratio (at least factor 6);
- suitable for slightly fouling service (e.g. low sand loading).

Recommended use:

- where there is little plot space available (e.g. in offshore industry or in general for high-pressure conditions);
- as retrofits of existing vessels where capacity debottlenecking and/or improved separation efficiency are required.

Non-recommended use:

- when a very low pressure drop is essential (e.g. in a high-vacuum environment);
- for viscous liquids where de-gassing requirement determines vessel diameter;
- if insufficient head room is available for a vertical vessel.

Typical process application:

- wellhead separators;
- compressor suction and interstage scrubbers;
- cold separators;
- inlet separators to adsorption plants.

3.11.2 General

Vertical separators of this type are equipped with a bundle of axial-flow cyclones.

For the gas/liquid separation in the cyclones use is made of a secondary gas stream which is recirculated.

Normally the axial-flow multicyclone bundle, including drainpipes etc. is supplied as a complete package by the Manufacturer and is based on the Manufacturer-proprietary design. The rules given below can be used to verify the proprietary Manufacturers' design.

The cyclones can be installed either horizontally or vertically.

The layout of the separator may be similar to that of the in-line separator with horizontal flow vane pack (with horizontal cyclones) or that of the two-stage separator with vane pack (with horizontal or vertical cyclones).

The recommended configuration for the vertical multicyclone separator is that of a two-stage separator equipped with vertical cyclones. Its layout is shown in Figure 3.15a and sizing rules are given below.

3.11.3 Cyclones

The working principle of vertical axial-flow cyclones with recirculation of secondary gas is shown in Figure 3.15b.

The cyclones have a swirler in the inlet and longitudinal slits in the cyclone wall.
The centrifugal forces induced by the swirling gas flow will cause gas/liquid separation in the cyclone tubes. The separated-off liquid is removed from the cyclone tubes through the slits

with the aid of a secondary gas flow and is transported to the liquid collection chamber outside the tubes. From there it is transported via drainpipes to the liquid compartment of the separator.
The secondary gas is recirculated to the central axis of the cyclone (recirculation driven by the underpressure on the central axis of the cyclone) and leaves the cyclone tube via the hollow tube of the swirler and the gas outlet at the top of the cyclone.

In order to achieve the high performance as indicated in the separator profile in (3.11.1) the cyclone geometry shall meet at least the following requirements:

1. the swirler shall be aerodynamically shaped;
2. the top of the swirler tube shall be fitted with vanes to minimise re-entrainment of separated-off liquid into the gas stream.

In general, the ID of the cyclone tubes of this type of multicyclone is 2 inch. However, other IDs such as 3 inch are also used.

The multicyclone bundles are Manufacturer-proprietary devices. For information on approved Manufacturers, the Principal should be consulted.

As in the case of the reversed-flow multicyclones, the volumetric gas handling gas capacity is insensitive to the gas pressure.

For 2 inch ID cyclones:

if 0.005 **actual** $m^3/s \leq Q_{G,c} \leq$ 0.030 **actual** m^3/s, then the liquid removal efficiency of the cyclones will be greater than 99%

For 3 inch ID cyclones:

if 0.014 **actual** $m^3/s \leq Q_{G,c} \leq$ 0.084 **actual** m^3/s, then the liquid removal efficiency of the cyclones will be greater than 99%.

3.11.4 Axial-flow multicyclone bundle

The multicyclone bundle layout should satisfy the following requirements:

1. It is acceptable to have the bundle built up in units, provided that each unit is equipped with a separate drainage system.
2. The drain pipe(s) of the multicyclone bundle shall extend at least 0.10 m below LZA(LL) for sealing purposes.
3. For the installation of the bundle in the separator bolting shall be employed, not welding.

3.11.5 Diameter

The vessel diameter, D, of the multicyclone separator shall be large enough to hold sufficient cyclones in order to meet **the gas handling capacity criterion**.

The minimum number of cyclones required, n_c, is given by:

$$n_c = Q_{G,max} / 0.03$$

with n_c rounded up to the nearest integer.

$Q_{G,max}$ is the maximum **actual** volumetric flow rate and shall include the appropriate design margin (see Appendix IV).
The denominator of the right hand side of the above equation is the maximum actual flow rate per cyclone (m^3/s).

The vessel diameter shall also satisfy the following criteria:

If liquid de-gassing is required:

$$D \geq 7608 \sqrt{Q_{L,max} \eta_L / (\rho_L - \rho_G)}$$

This criterion is explained in Appendix VII.

If de-foaming is required:

$D \geq 95Q^{0.5}_{L,max} \{\eta_L/(\rho_L-\rho_G)\}^{0.14}$

3.11.6 Height

Let h be the height required for liquid hold-up (Appendix V). Then the total vessel height (tangent to tangent) is (see Figure 3.15).

$H = h + X_1 + X_2 + X_3 + h_c + X_4$

where

X_1 is the distance between LZA(HH) and the feed inlet device (either Schoepentoeter or half-open pipe)

Schoepentoeter: $X_1 = 0.05D$ with a minimum of 0.15 m

Half-open pipe: $X_1 = 0.3D$ with a minimum of 0.3 m

X_2 is the height required for the inlet nozzle

Schoepentoeter: $X_2 = d_1 + 0.02$ m

Half-open pipe: $X_2 = d_1$

d_1 is the internal diameter of the inlet nozzle

X_3 is the distance between the feed inlet device and the multicyclone bundle

Schoepentoeter: $X_3 = d_1$ with a minimum of 0.3 m.

Halfopen pipe: $X_3 = 0.45$ D with a minimum of 0.9 m

The distance between LZA(HH) and the bottom of the multicyclone bundle shall be sufficiently large to accommodate the liquid head in the drainpipes caused by the pressure drop between the gas compartment below the multicyclone bundle and the liquid compartment of the bundle, $(p_{in} - p_{L,out})$

Recommendation: $X_1 + X_2 + X_3 > 1.5(p_{in} - p_{L,out})/\{g(\rho_L - \rho_G)\}$

(For the calculation of $(p_{in} - p_{L,out})$, see (3.11.8))

h_c is the height of the multicyclone bundle.

if the cyclone diameter is 2 inch, then typically h_c is 0.40 m

if the cyclone diameter is 3 inch, then typically h_c is 0.50 m

X_4 is the distance between the top of the multicyclone bundle and the top tangent line

$X_4 = 0.15$ D with a minimum of 0.15 m

3.11.7 Nozzles

The feed nozzle shall be fitted with a Schoepentoeter if $D \geq 0.5$ m

For the design of Schoepentoeters, see Appendix III. For the sizing of the feed nozzle, see Appendix II.

If D < 0.5 m a half-open pipe should be used.

For the sizing of the gas and liquid outlet nozzles see Appendix II.

3.11.8 Pressure drop

The pressure differential between inlet and vapour outlet of the separator is the sum of the pressure drops across the nozzles and multicyclone bundle. (The pressure drop across the

Schoepentoeter is negligible).

$$P_{in} - P_{out} = 0.5\rho_m v^2_{m,in} + 0.22\rho_G v^2_{G,out} + \Delta p_c$$

where Δp_c is the pressure drop over the multicyclone bundle

$$\Delta p_c = K_c \rho_G v^2_c$$

with $v_c = Q_G/(A_c n_c)$

typically, $K_c = 4$ for 2 inch cyclones
typically, $K_c = 3$ for 3 inch cyclones

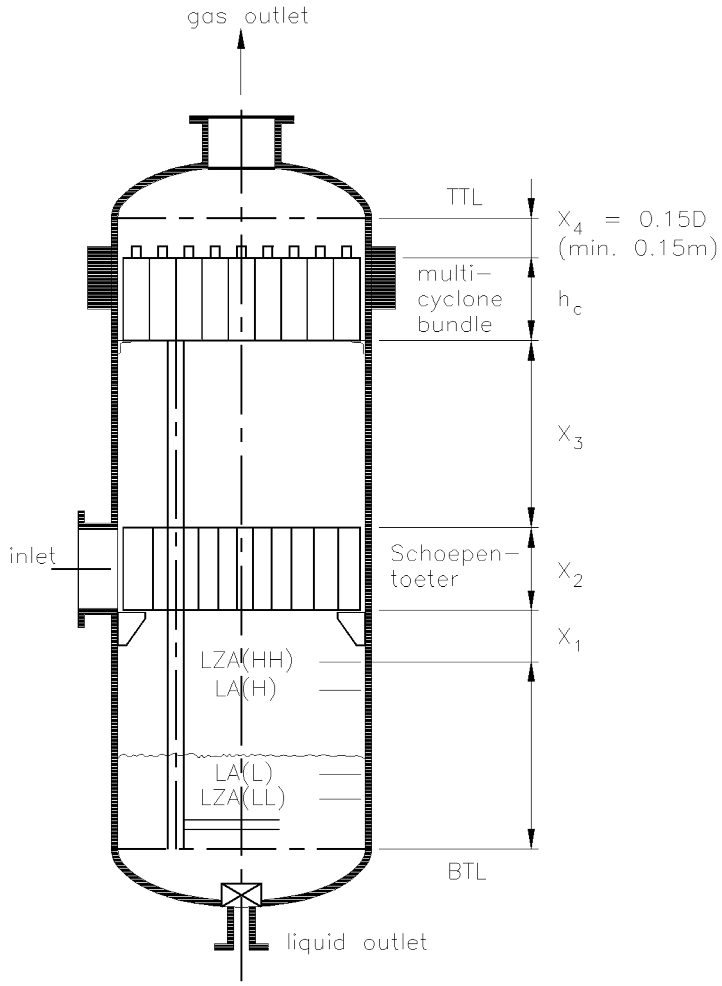
For the pressure difference between the gas compartment below the multicyclone bundle and the liquid compartment of the bundle, $p_{in} - p_{L,out}$:

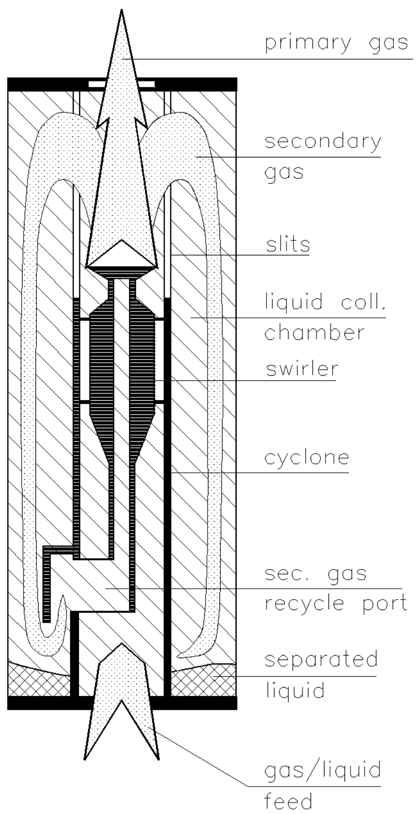
$$p_{in} - p_{L,out} = 0.5\Delta p_c$$

The height of the liquid in the drainpipes relative to the local liquid level, $\Delta h_{L,drain}$:

$$\Delta h_{L,drain} = (p_{in} - p_{L,out})/\{g(\rho_L - \rho_G)\}$$

FIGURE 3.15 THE VERTICAL SEPARATOR WITH AXIAL FLOW MULTI-CYCLONE BUNDLE (WITH RECIRCULATION OF SECONDARY GAS)





(a) Two-stage separator with Schoepentoeter as feed inlet device.

X_1, X_2, and X_3 are dependent of type of feed inlet. See (3.11.6)

(b) Working principle of the vertical cyclone.

3.12 HORIZONTAL SEPARATOR WITH AXIAL-FLOW MULTICYCLONE BUNDLE (WITH RECIRCULATION OF SECONDARY GAS)

(Figure 3.16)

3.12.1 Selection criteria

Application:

- demisting of gas where a high gas handling capacity and a high liquid removal efficiency is required.

Characteristics:

- very compact separator;
- high liquid removal efficiency (> 99%);
- very high gas handling capacity;
- high turndown ratio (at least factor 6);
- suitable for slightly fouling service (e.g. low sand loading);
- slug handling capacity.

Recommended use:

- as retrofits of existing horizontal vessels where gas handling capacity debottlenecking and/or improved separation efficiency are required.

Non-recommended use:

- when a very low pressure drop is essential (e.g. in a high-vacuum environment).

Typical process application:

- wellhead separators;
- inlet separators to gas plants;
- Hot High Pressure Separator in hydroprocessing.

3.12.2 General

This type of separator layout is recommended for separators previously fitted with wiremesh or vane pack which are now being replaced with a bundle of axial-flow multicyclones in order to upgrade the gas handling capacity. For new vessels it is recommended to select the vertical version (see 3.11) because it is easier to accommodate the required drainpipe length.

Normally the axial-flow multicyclone bundle, including drainpipes etc., is supplied as a complete package by the Manufacturer and is based on the Manufacturer-proprietary design. The rules given below can be used to verify the proprietary Manufacturer's design in the event of a retrofit.

The cyclones can be installed either horizontally or vertically and are similar to those described in (3.11.3).

It is recommended to equip the horizontal separator with horizontal multicyclones. The layout of the separator is shown in Figure 3.16a.

3.12.3 Cyclones

The working principle of the horizontal cyclone is similar to that of a vertical cyclone (as described in (3.11.3) and is shown in Figure 3.16.b.

Here too, in order to achieve the high performance as indicated in the separator profile in (3.12.1) the cyclone geometry shall meet at least the following requirements:

1. the swirler shall be aerodynamically shaped;
2. the top of the swirler tube shall be fitted with vanes to minimise re-entrainment of separated-off liquid into the gas stream.

The ID of the cyclone tubes of this type of multicyclone is generally 2 inches but other IDs, such as 3 inches, are also used.

The multicyclone bundles are Manufacturer-proprietary devices. For information on approved Manufacturers, the Principal should be consulted.

As in the case of the reversed-flow multicyclones, the volumetric gas handling gas capacity is insensitive to the gas pressure.

For 2 inch ID cyclones:

if 0.005 **actual** $m^3/s \le Q_{G,c} \le 0.030$ **actual** m^3/s, then the liquid removal efficiency of the cyclones will be larger than 99%.

For 3 inch ID cyclones:

if 0.014 **actual** $m^3/s \le Q_{G,c} \le 0.084$ **actual** m^3/s, then the liquid removal efficiency of the cyclones will be larger than 99%.

3.12.4 Axial-flow multicyclone bundle

The multicyclone bundle layout should satisfy the following requirements:

1. It is acceptable to have the bundle built up in units, provided that each unit is equipped with a separate drainage system.
2. The drain pipe(s) of the multicyclone bundle shall extend at least 0.10 m below LZA(LL) for sealing purposes
3. To guarantee proper liquid drainage these units shall not contain more than two layers of multicyclones
4. For the installation of the bundle in the separator no welding but bolting shall be employed.

3.12.5 Vertical cross-sectional area for gas flow

The minimum vertical cross-sectional area for gas flow above LZA(HH), $A_{G,min}$, is determined by the following criterion:

This area shall be sufficiently large to contain the required number of cyclones, leaving sufficient space between the bottom of the cyclone bundle and LZA(HH) to allow proper drainage of the separated-off liquid at the highest gas flow rate.

Since this application is mainly for the retrofit of multicyclone bundles in existing vessels, $A_{G,min}$ is determined by the elevation of LZA(HH).

It is recommended, that for the drainage space, h_{drain}, (distance between bottom of multicyclone bundle and LZA(HH)):

$h_{drain} > 1.5(p_{in} - p_{L,out})/\{g(\rho_L - \rho_G)\}$

(For the calculation of $(p_{in} - p_{L,out})$, see (3.12.7))

It is possible that if the minimum number of cyclones required for proper demisting are used, the pressure drop between the cyclone inlet and the liquid compartment of the cyclones is so large that it will determine $A_{G,min}$.

By installing more cyclones this pressure drop can be minimised with the overall effect that $A_{G,min}$ becomes smaller. In other words LZA(HH) can be selected at a higher elevation and a larger part of the vessel becomes available for liquid handling.

3.12.6 Nozzles

The feed nozzle should be fitted with a Schoepentoeter.

For the design of Schoepentoeters, see Appendix III.

For the sizing of the feed nozzle, see Appendix II.

The feed nozzle may be located at the vessel front or vessel top, as indicated in Figure 3.16a. For process purposes, the top location is slightly preferable.

In both cases the distance between the Schoepentoeter and the multicyclone bundle should be at least one vessel diameter.

The gas outlet should be located on the top of the vessel. The distance between the multicyclones and the gas outlet should be at least 0.5 m.

As opposed to the horizontal wiremesh and vane- pack demisters a gas outlet deflector is not mandatory. Because of the higher pressure drop across the cyclones the chance of maldistribution is much smaller.

For the sizing of the gas and liquid outlet nozzles see Appendix II.

3.12.7 Pressure drop

The pressure differential between inlet and vapour outlet of the separator is the sum of the pressure drops across the nozzles and multicyclone bundle. (The pressure drop across the Schoepentoeter is negligible).

$$P_{in} - P_{out} = 0.5\rho_m v^2_{m,in} + 0.22\rho_G v^2_{G,out} + \Delta p_c$$

where Δp_c is the pressure drop across the multicyclone bundle

$$\Delta p_c = K_c \rho_G v^2_c$$

with $v_c = Q_G/(A_c n_c)$

typically, $K_c = 4$ for 2 inch cyclones
typically, $K_c = 3$ for 3 inch cyclones

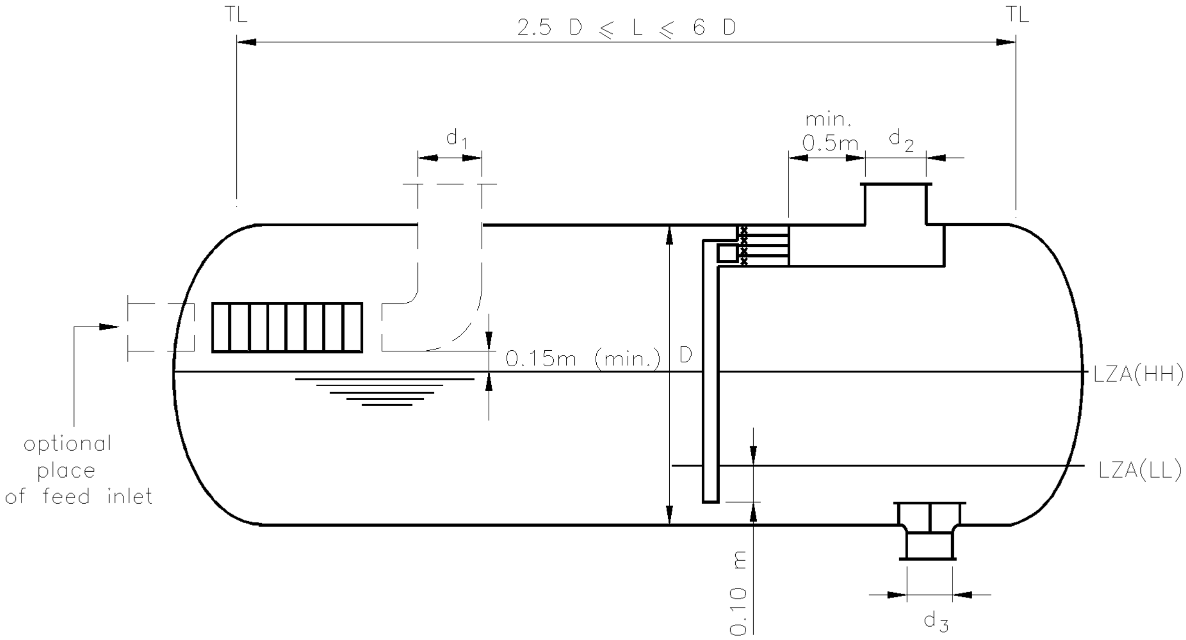
For the pressure difference between the gas compartment below the multicyclone bundle and the liquid compartment of the bundle, $p_{in} - p_{L,out}$:

$$p_{in} - p_{L,out} = 0.5\Delta p_c$$

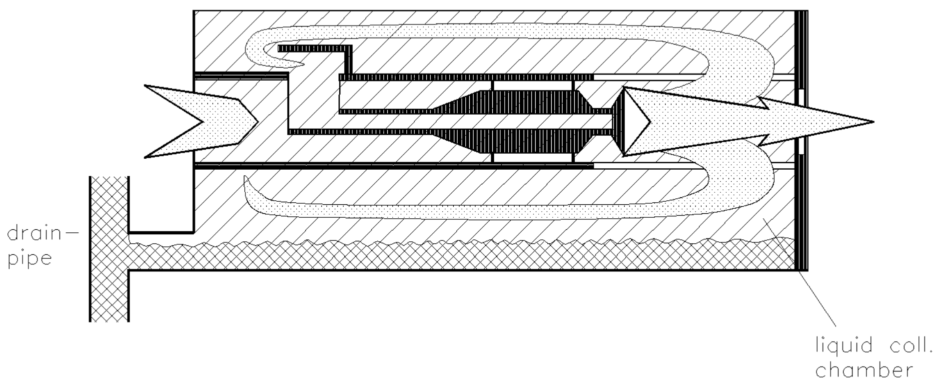
The height of the liquid in the drainpipes relative to the local liquid level, $\Delta h_{L,drain}$:

$$\Delta h_{L,drain} = (p_{in} - p_{L,out})/\{g(\rho_L - \rho_G)\}$$

FIGURE 3.16 THE HORIZONTAL SEPARATOR WITH AXIAL FLOW MULTI-CYCLONE BUNDLE (WITH RECIRCULATION OF SECONDARY GAS)



(a) Separator with horizontal cyclones



(b) Working principle of the horizontal cyclone
(for detailed description see Fig. 3.15b)

3.13 FILTER SEPARATOR

(Figure 3.17)

3.13.1 Selection criteria

Application:

- after-cleaning (liquid and solids) of already demisted gas when a very high liquid removal efficiency is required.

Characteristics:

- liquid removal efficiency > 99%;
- very high pressure drop;
- sensitive to high liquid loading or slugs;
- sensitive to fouling by sticky material.

Recommended use:

- typically as a second-line gas/liquid separator to after-clean the gas stream exiting from the first-line gas/liquid separator.
  * use filter candles with the flow from OUT to IN where solids are present;
  * use filter candles with the flow from IN to OUT where ultimate efficiency is required and NO solids are present.

Non-recommended use:

- heavy fouling (sticky material) service;
- high liquid loading;
- slugs.

Typical process application:

- last demisting stage of natural gas prior to despatch for sale.

3.13.2 General

Normally the filter separator is supplied as a complete package of vessel and internals based on a Manufacturer proprietary design. The rules given below be used to verify the proprietary Manufacturer's design.

The conventional filter separators consist of two compartments, one with a parallel set of filter candles (to coalesce the fine mist and to separate dust) followed by one with a final demisting device. This demisting device could be a multicyclone bundle, a mistmat or a vane pack.

The two-compartment filter separator is either a horizontal or a vertical separator. In the latter version, the filter candle compartment is the top compartment.

If dust is expected, it is common practice to replace the candles periodically (e.g. once a year).

In order to be able to remove the candles as a bundle, the filter candle compartment shall be top-flanged.

Conventionally, the flow direction in the candles is from the outer to the inner side. This is the preferred direction for the removal of dust, but is not ideal for coalesence.

Also, the coalescence process could be spoiled if the velocity in the outlet tubes of the filter candles is too high, because this will give rise to re-entrainment and atomisation of the coalesced liquid. For the demisting section, the demisting area shall be determined using the design rules of this manual.

If the initial gas flow direction is parallel to the face of the demister, the risk of maldistribution is present. Either the maldistribution has to be minimised by mechanical means (perforated plates shall not be used if the flow is vertical) or, if that is not possible, the demisting area shall be increased to compensate for the maldistribution effect; in the latter case the Principal should be consulted.

In alternative versions of filter separators the gas flow passes from the inside to the outside of the candles thus creating better conditions for coalescence. However, in a fouling environment or when mercury is present in the feed (e.g. natural gas), a pre-filtering stage is essential because otherwise the dirt or mercury will accumulate inside the candles and will

eventually block them. For the pre-filtering step, conventional filter candles could be used with the flow from the outside to the inside.
Also, downstream of the coalescing candles a demisting stage should be installed in order to keep the filter separator compact.
In the following paragraphs, rules are given for the most commonly used filter separator: the horizontal two-stage filter separator.
For more information on other types of filter separators, the Principal should be consulted.

3.13.3 Horizontal two-stage filter separator

(Figure 3.17)

3.13.3.1 The coalescer compartment

The function of this compartment is to remove dust and to coalesce fine mist. For this purpose, a bundle of filter candles is installed. The gas to be cleaned enters the candles peripherally and leaves them through the outlet tube.
The filter candle consists of a fibrous filter (typically with an outer and inner diameter of 4" and 3" respectively and with a length of either 36" or 72") and an outlet tube (having the same inner diameter as the filter and a length of say $1.4d_1$, where d_1 is the feed nozzle inner diameter).
The candles are grouped in either a square or triangular configuration.
The feed enters the compartment via the top of the vessel. In order to avoid mechanical damage of the filter candles, the location of the feed nozzle shall be such that the jet of the feed touches the candles only on their outlet tubes.
Some liquid will be knocked-out and flows into the liquid collection compartment underneath the filter separator.
Subsequently, the gas enters the candles peripherally, while dust (and also mercury, if present in the feed) remains in the outer layer of the candle filter. The fine mist droplets in the gas coalesce in the filter medium of the candle.
Subsequently the gas and liquid (as a film and droplets) flow via the outlet tubes to the demisting compartment.

The following requirements and recommendations apply:

1. To ensure a proper coalescing process the gas velocity into the filter part of the candles, $v_{G,cf}$, should not be too high. Recommended upper limit is 0.10 m/s ($v_{G,cf}$ is based on outer candle surface).
2. To avoid atomisation of the coalesced droplets in the outlet tubes of the candles:

   $$\rho_G v^2_{G,ct} \leq 1\,000 \qquad \text{Pa}$$

   with $v_{G,ct}$ being the gas velocity inside in the outlet tubes.
3. The coalescer compartment shall have a top flange or a full diameter top cover to enable installation of the filter candles. It should allow the filter candle bundle to be removed in one piece. If a filter carrier is used for this purpose, then the minimum distance between the filter candles and the vessel wall should be 0.05 m.

The first two requirements will determine the number of candles of a given type.

3.13.3.2 The demisting compartment

In this compartment the coalesced liquid droplets are removed from the gas stream by means of a demisting device, which is normally a vertical-flow mistmat.
However, if a compact and robust filter separator is required a vane pack could be considered instead of a mistmat.
The vane pack shall then be of the horizontal flow type, with its vane face either perpendicular to, or making a small angle (typically 10°) with, the central axis of the filter separator. Figure 3.17 shows the perpendicular configuration.
The minimum distance between the demisting device and the plate on which the candles are mounted should be 0.5 D.
For the sizing of the demisting devices and the location of the gas outlet nozzle the following rules apply:

Vertical flow wiremesh:

Use the sizing rules for the mistmat in vertical wiremesh demisters as given in (3.3.4).
The mistmat shall be fitted in a gas-tight box connected to the outlet nozzle which should be located at the top of the vessel. If a vertical flow mistmat is used, flow maldistribution over the mistmat is almost unavoidable.
It is therefore recommended to have the gas outlet nozzle as large as possible in order to minimise this flow maldistribution.

Horizontal flow vane pack:

The sizing rules given in ((3.5.3) shall be used.
To prevent liquid overloading of the vanes, the flow parameter, ϕ, of the feed to the filter separator shall be limited by: $\phi \leq 0.01$.
Perforated plates shall be used to minimise flow maldistribution.
The NFA of the perforated plates should be about 20%, with the holes evenly distributed and having a size of about 12 mm nominal.
In the case of a vane pack with its face parallel to the central axis of the filter separator, take a 10% larger area than calculated to compensate for maldistribution. In the latter case the vane pack shall be fitted in a gas-tight box connected to the outlet nozzle which should be located at the side of the vessel.

3.13.3.3 The liquid collection compartment

This compartment receives liquid both from the coalescer and the demisting compartment. The liquid collection compartment shall be divided into two separate parts to prevent gas bypassing through it. For a correct level control, each part of the liquid collection compartment should have a boot leg.
Each part of the liquid collection compartment shall have an independently operating liquid level control, with LZA(HH) located in the upper connecting pipes and LZA(LL) in the boot legs. Each part of the collection compartment shall be sufficiently large to guarantee a required specified liquid residence time (typically 2 minutes).

3.13.3.4 Diameter

The diameter of the filter separator is determined by the diameter of the coalescer compartment which in turn is determined by the number of filter candles required for proper separation.

3.13.3.5 Nozzles

The size of the feed nozzle and of the gas outlet nozzle may be taken to be equal to the pipeline size, but it shall be ensured that:

$$\rho_G v^2_{G,noz} \leq 3750 \qquad \text{Pa}$$

The inlet nozzle is located at the top of the filter separator.

The outlet nozzle is also located at the top of the vessel if a horizontal mistmat is used as demister; in which case its diameter shall be as large as possible in order to minimise flow maldistribution over the wiremesh.
In case of a horizontal flow vane pack, a horizontal gas outlet is preferred.
For the sizing of the liquid outlet nozzles see Appendix II.

3.13.3.6 Pressure drop

The pressure differential between inlet and vapour outlet of the filter separator is the sum of the pressure drops across the nozzles, filter candles and demister (including perforated plates):

$$P_{in} - P_{out} = 0.5\rho_m v^2_{m,in} + 0.22\rho_G v^2_{G,out} + \Delta p_{cf} + \Delta p_m + n\Delta p_{perfpl}$$

where

Δp_{cf} ranges typically from 0.05 bar (clean cartridges)

to 0.7 bar (fouled cartridges just before replacement)

$\Delta p_m \quad = K_m (\rho_L - \rho_G)\lambda^2{}_m$

λ_m = the gas load factor based on the demisting area

K_m = 15 for single-pocket vanes

K_m = 10 for double-pocket vanes

K_m = 20 for wet mistmat

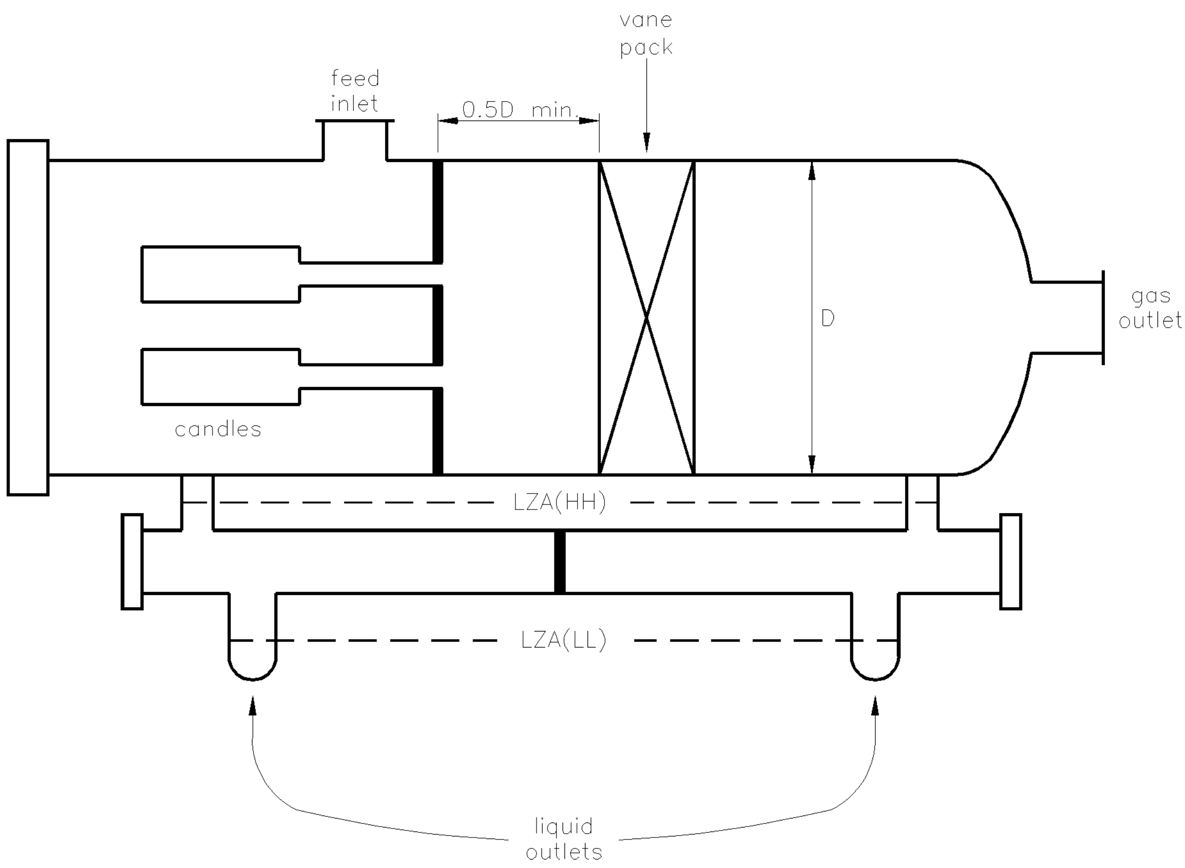
n = number of perforated plates

(**for a vertical flow mistmat, n = zero**)

$\Delta p_{perfpl} = 0.8(1 - NFA)(\rho_L - \rho_G)\lambda^2{}_v / NFA^2$

with NFA being the net free area of the perforated plate (as a fraction).

FIGURE 3.17 HORIZONTAL TWO-STAGE FILTER SEPARATOR (WITH VANE PACK AS DEMISTER)



4. PIPING REQUIREMENTS

Piping to and from the separator shall interfere as little as possible with the working of the separator. The following constraints should be observed:

a. The use of valves, pipe expansions or contractions within ten pipe diameters of the inlet nozzle should be avoided because of their tendency to generate relatively small liquid droplets.

   If a valve in the feed line near to the separator cannot be avoided, it should be of the gate or ball type, fully open in normal operation. High pressure drops which cause flashing and atomisation should be avoided in the feed pipe.

   If a pressure-reducing valve in the feed pipe cannot be avoided, it should be located as far upstream of the vessel as practicable.

b. The use of bends within ten pipe diameters of the inlet nozzle should be avoided because they will generate gas flow maldistribution in the separator.

   If bends cannot be avoided within ten pipe diameters of the inlet nozzle, the following rules should be followed:

   - For knock-out drums, wiremesh and vane-type demisters and for any other type of demister in which a tangential inlet is NOT used, a bend in the feed pipe is only permitted if this is in the vertical plane through the axis of the feed nozzle.

     If this results in a riser system just upstream of the vessel, slugging may occur in case of a high feed flow parameter. The vessel should be able to handle this.

     If for some reason a bend in the horizontal plane cannot be avoided (for instance if the demister vessel has to be installed in an existing plant) the negative effect of the horizontal bend on the efficiency of the separator can be minimised by the installation of a static mixer between the bend and the inlet nozzle or by fitting guiding vanes in the vessel. In such cases the Principal should be consulted.

   - For cyclones a bend in the feed pipe is only permitted if this is in a horizontal plane and the curvature is in the same direction as the cyclone vortex.

c. If desired, a pipe reducer may be used in the vapour line leading from the separator, but it should be situated no closer than twice the outlet nozzle diameter from the top of the vessel.

If the above conditions cannot be satisfied, some loss of liquid removal efficiency will result.

5. REFERENCES

In this DEP reference is made to the following publications:

NOTE Unless specifically designated by date, the latest edition of each publication shall be used, together with any amendments/supplements/revisions thereto.

SHELL STANDARDS

Index to DEP publications and standard specifications	DEP 00.00.05.05-Gen.
Data/Requsition sheet - Schoepentoeter (vane-type) inlet device	DEP 31.20.41.93-Gen.
Liquid/liquid and gas/liquid/liquid (three-phase) separators - Type selection and design rules	DEP 31.22.05.12-Gen.
Requsition sheet - Wiremesh demister	DEP 31.22.05.93-Gen.
Pressure relief, emergency depressuring, flare and vent systems	DEP 80.45.10.10-Gen.

STANDARD DRAWINGS

Vortex breakers for nozzles	S 10.010
Typical details of Schoepentoeter (vane feed inlet)	
Type II	S 20.015
Type III	S 20.016
Type IV	S 20.017
Vane ladder - supports and general notes	S 20.019
Type IA	S 20.020
Type IB	S 20.021
Type IC	S 20.022
Typical details of demisters	S 20.028
Typical details of demister attachments	S 20.029
Typical details of demister supports	S 20.030

APPENDIX I NATURE OF THE FEED

Flow regime in the feed pipe

The feed entering a G/L separator can be in the form of mist, stratified flow, slug flow, etc. depending on the flow rates and physical properties of the gas and liquid phases and on the feed pipe characteristics (diameter, length, vertical/ horizontal).

Use can be made of available two-phase flow maps to characterise the G/L flow regime in the feed pipe.
In Figure I.1 and I.2 two flow maps are presented. The first flow map gives the two-phase flow regimes in a horizontal pipe and the second one in a vertical pipe (upflow).

Strictly speaking the flow maps are only applicable to very long pipes with equilibrium two-phase flow. However, if the feed pipe is longer than ten pipe diameters, the flow maps still give a fair indication of the prevailing flow regime for a given set of conditions.

The transition from one flow regime to another is rather gradual, and the boundaries shown separating the different regimes should not be interpreted as sharp changes in flow pattern.

The flow maps are generalised by using as parameters the gas and liquid Froude number respectively, based on the feed pipe velocity and diameter.
The advantage of this general representation is that the flow maps become insensitive to variations in flow conditions, physical properties and feed pipe geometry. This means that the flow maps can be used for a wide range of flow conditions, physical properties and feed pipe diameters.
If it is still considered necessary for a flow map to be generated for a particular set of conditions, the Principal should be consulted.

The gas and liquid Froude numbers are defined as follows:

gas Froude number:

$$Fr_G = v_G \sqrt{\rho_G / \{(\rho_L - \rho_G) g d_{fp}\}}$$

liquid Froude number:

$$Fr_L = v_L \sqrt{\rho_L / \{(\rho_L - \rho_G) g d_{fp}\}}$$

In the above formulae v_G and v_L are the superficial gas and liquid velocity respectively in the feed pipe and d_{fp} is the inner diameter of the feed pipe.

$$v_G = Q_G / (\pi d^2_{fp} /4)$$

$$v_L = Q_L / (\pi d^2_{fp} /4)$$

and the averaged liquid density ρ_L is defined as:

$$\rho_L = M_L/Q_L$$

Within stratified-wavy and annular flow regimes it is possible that droplet formation will take place in the feedpipe, resulting in a mist when these droplets are entrained into the separator.

As a rough indication, the approximate size of the largest droplets, $d_{p,max}$, formed in the feed pipe with diameter d_{fp} is given by:

$$d_{p,max}/d_{fp} = 4.5\{\sigma/(\rho_G v^2_G d_{fp})\}^{0.6}(\rho_G/\rho_L)^{0.4}$$

The smallest drops will generally have a diameter five to ten times smaller than $d_{p,max}$.

NOTE However, much smaller droplets may be formed if the multi-phase flow has passed a sudden and significant pressure reduction (e.g. a choke with a pressure drop of more than, say, 10 bar).

Foaming tendency

For foaming to occur it is necessary for gas bubbles to be formed, and for the drainage of

the liquid films surrounding the bubbles to be retarded. Drainage of the films is slower in highly viscous liquids, but the chief causes of foaming are surface properties which are usually unpredictable. For this reason the foaming tendency is best judged on the basis of experience of similar cases. Laboratory tests may also give an indication of the foaminess of the system.
Examples of foaming systems are some crude oils, heavy residues, absorption and extraction solvents.
Foaming in the separator may lead to carry-over of liquid (when foam reaches the gas/liquid separation internal and/or the gas outlet) or to carry-under of gas. It will also upset the level control system.
It should be noted that foaming is more likely to be a problem at high liquid loads, when the flow in the inlet pipe is in the frothy or intermittent flow regimes.
Installation of G/L separation internals to combat foam can only be effective if they are of the cyclone type.
Foaming in the vessel is minimised by decreasing the downward liquid velocity, for instance by increasing the diameter of the separator vessel.
Sometimes an antifoam agent can be injected to suppress foaming.

Feeds with solids and wax (fouling service)

Sand, rust, scale or other solids present in the feed will leave the separator together with the liquid. However, solids will also settle out in the separator and tend to accumulate. For this reason care should be taken with the location of instrument connections which could become plugged. Provision should be made for cleaning the separator during shutdowns, and if necessary during operation, by the installation of a liquid (water) spray and drain.

If solids are present in the feed, consideration should be given to reducing the inlet velocity and adding an "erosion allowance" of 1 - 2 mm extra material thickness to the inlet device (if present).

Wax in the feed will be deposited on any surfaces where the velocities and temperatures are low. Also, narrow openings will tend to become plugged.

It is recommended to use empty settlers in fouling service. However, if because of high efficiency requirements the use of L/L and/or G/L separation internals cannot be avoided, the selected internal should be robust to fouling (for instance, by using a vane pack instead of a mist mat for G/L separation and a plate pack with large plate angle or large plate spacing for L/L separation). If that is not possible, the internal has to be protected adequately by a prefilter.

FIGURE I.1 TWO-PHASE FLOW MAP FOR HORIZONTAL FEED PIPES

Conditions:

$\rho_G = 8\ kg/m^3$ $\rho_L = 860\ kg/m^3$ $\sigma = 0.03\ N/m$

$\eta_G = 1.2*10^{-5}\ Pa.s$ $\eta_L = 1.6*10^{-4}\ Pa.s$ $d_{fp} = 0.50\ m$

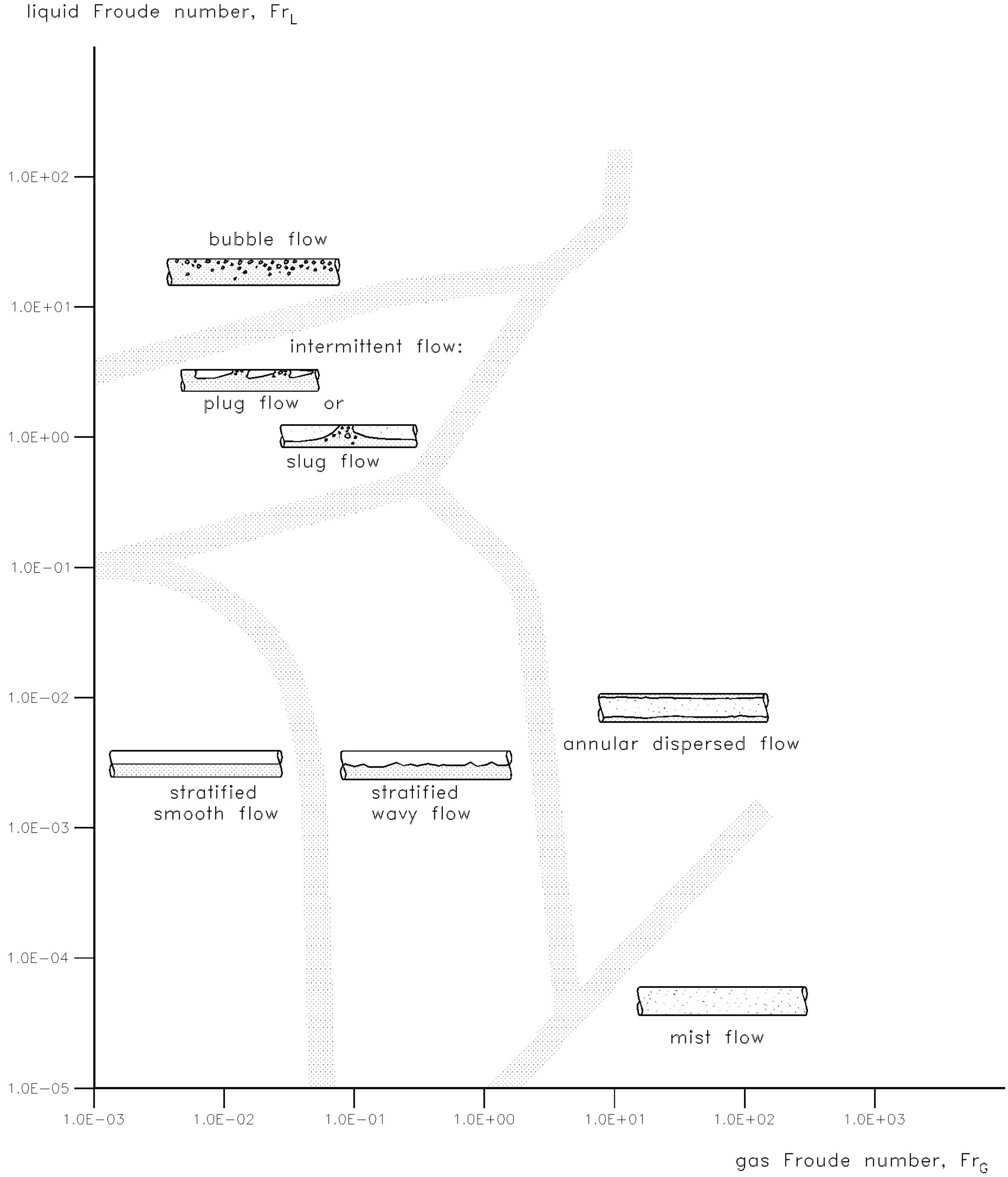
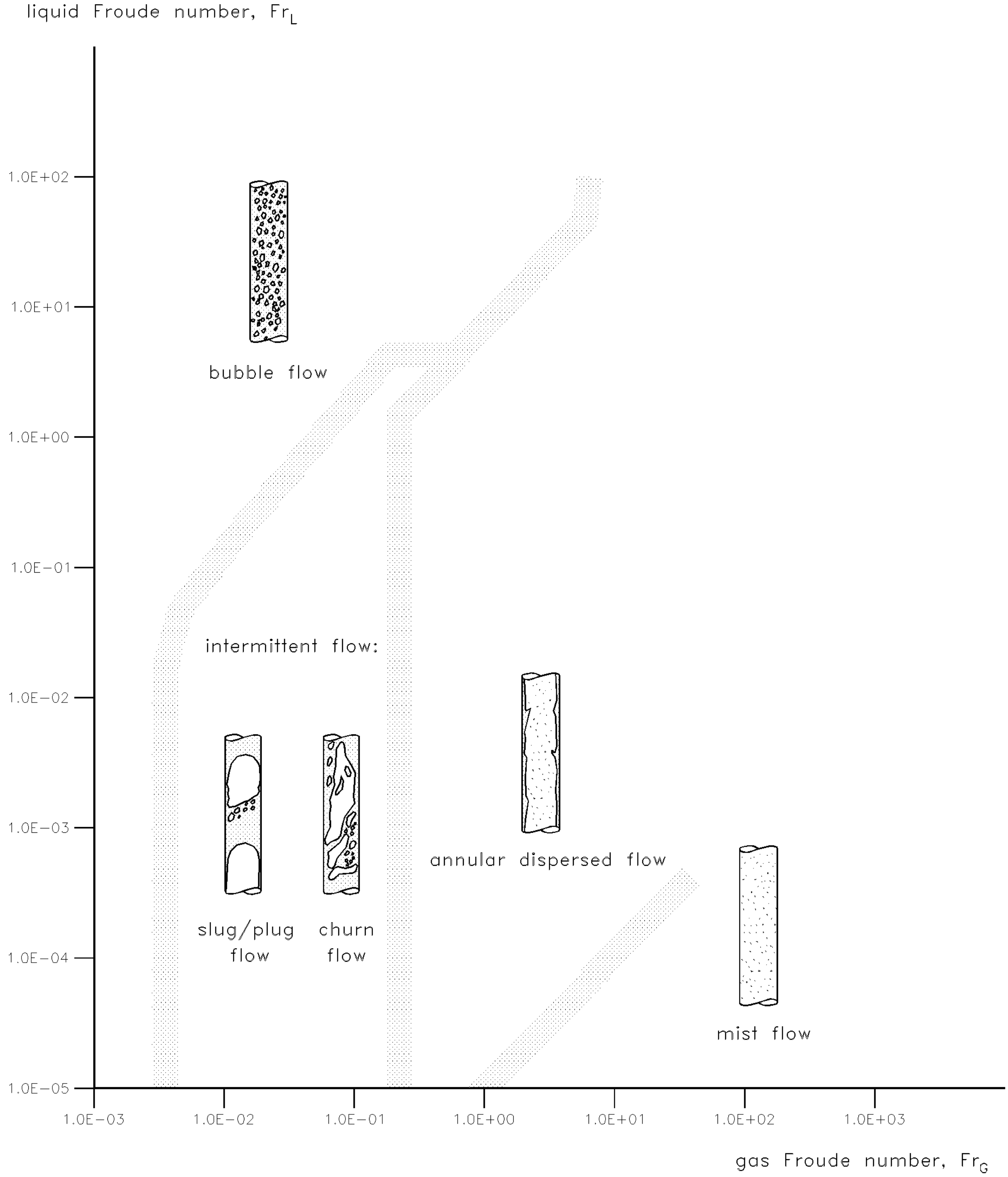


FIGURE I.2 TWO-PHASE FLOW MAP FOR VERTICAL FEED PIPES (UPFLOW)

Conditions:

$\rho_G = 8$ kg/m^3 $\rho_L = 860$ kg/m^3 $\sigma = 0.03$ N/m

$\eta_G = 1.2*10^{-5}$ Pa.s $\eta_L = 1.6*10^{-4}$ Pa.s $d_{fp} = 0.50$ m



APPENDIX II SIZING OF THE FEED AND OUTLET NOZZLES

The sizing of the nozzles shall be based on the **actual** flow rates (i.e. **excluding** the appropriate design margin).

1. Feed inlet nozzle

The internal nozzle diameter, d_1, may be taken equal to that of the feed pipe, but also a momentum criterion (dependent on the inlet device, if any) shall be satisfied:

If **no** inlet device is used:

$$\rho_m v^2_{m,in} \leq 1\,000 \qquad \text{Pa}$$

where ρ_m = mean density of the mixture in the feed pipe

$= (M_G + M_L)/(Q_G + Q_L)$

and $v_{m,in}$ = velocity of the mixture in the inlet nozzle

$= (Q_G + Q_L)/(\pi\, d_1^2 / 4)$

If a **half-open pipe** is used as inlet device:

$$\rho_m v^2_{m,in} \leq 1\,500 \qquad \text{Pa}$$

If a **Schoepentoeter** is used as inlet device:

$$\rho_m v^2_{m,in} \leq 6\,000 \qquad \text{Pa}$$

In High Vacuum Units or in any other unit or separator where the inlet velocity can be very high because of the low gas density, the use of a Schoepentoeter as feed inlet device is mandatory and the following velocity limit shall also be satisfied:

$$v_{G,in} \leq 0.8 v_{sonic,G}$$

$$v_{G,in} \leq 70 \qquad \text{m/s}$$

$v_{sonic,G}$ is the sonic velocity if only gas is present (presence of liquid ignored)

$$v_{sonic,G} = \sqrt{\frac{\kappa\, R\, T}{MW_G}}$$

where:

R is the gas constant (8314 J/kmol/K)
T is the absolute temperature
κ is the ratio of the specific heats (C_p/C_v)
MW_G is the mean molecular weight of the gas phase

NOTE The actual sonic velocity may be lower if liquid is indeed present.

2. Gas outlet nozzle

The diameter of the gas outlet nozzle, d_2, should normally be taken equal to that of the outlet pipe, but the following criterion shall be satisfied:

$$\rho_G v^2_{G,out} \leq 3750 \qquad \text{Pa}$$

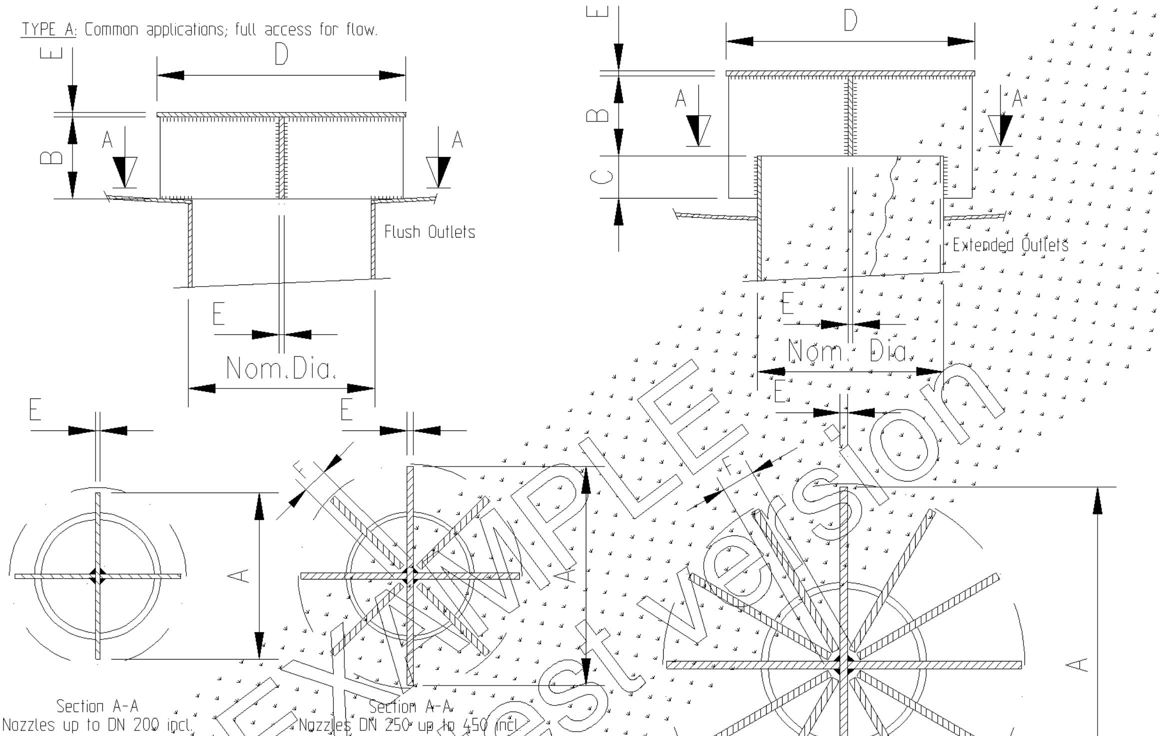
In High Vacuum Units this criterion may result in a high outlet velocity, leading to a pressure drop which is too high. In that case the gas outlet nozzle shall be sized such that the pressure drop requirements between column and downstream system are met.

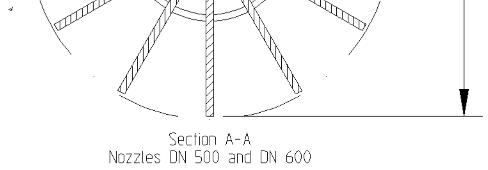
3. Liquid outlet nozzle

The diameter of the liquid outlet nozzle, d_3, shall be chosen such that the liquid velocity does not exceed 1 m/s. The minimum diameter is 0.05 m (2 in). The nozzle shall be equipped with a vortex breaker in accordance with Standard Drawing S 10.010 (specimen copy of which is shown as Figure II.1 at the end of this Appendix).

FIGURE II.1 VORTEX BREAKERS (S 10.010)

TYPE A: Common applications; full access for flow.



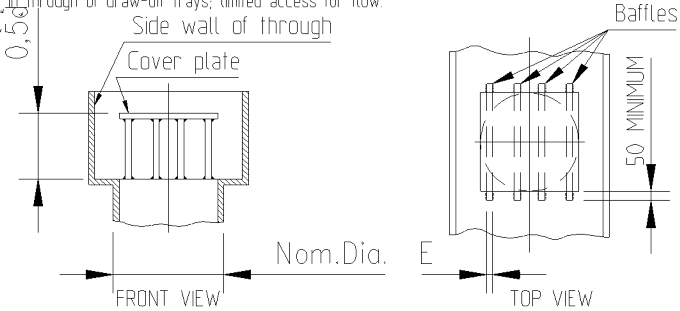


Baffle Arrangement							
Nominal Diameter	Number of Baffles	A	B	C	D	E	F
DN 40	4	70	50	25	80	4 ± .5	
DN 50	4	85	50	25	95	4 ± .5	
DN 80	4	115	50	25	125	4 ± .5	
DN 100	4	150	50	25	160	4 ± .5	
DN 150	4	215	75	25	225	6 ± .5	
DN 200	4	290	100	25	300	6 ± .5	
DN 250	8	365	125	40	375	6 ± .5	50
DN 300	8	440	150	40	450	6 ± .5	50
DN 350	8	515	175	40	525	6 ± .5	50
DN 400	8	590	200	40	600	6 ± .5	50
DN 450	8	680	225	50	690	6 ± .5	50
DN 500	12	740	250	50	750	6 ± .5	50
DN 600	12	890	300	50	900	6 ± .5	50

NOTES:

1) Dimension "E" is mandatory because thicker plate will obstruct inlet.
2) All sharp edges shall be rounded off.
3) Materials: as specified on requisition or equipment drawing.
4) all dimensions are in millimetres.

TYPE B: In through of draw-off trays; limited access for flow.



Baffle Arrangement	
Nominal Diameter	Number of baffles
100 < d ≤ 200	2
200 < d ≤ 300	3
300 < d ≤ 400	4
400 < d ≤ 500	5
500 < d ≤ 600	6

All baffles in length direction of through
For through widths above 2d apply type A.

APPENDIX III DESIGN OF SCHOEPENTOETER (VANE-TYPE) INLET DEVICE

1. INTRODUCTION

The Schoepentoeter (vane-type) is a Shell-proprietary inlet device and is commonly used for introducing gas/liquid mixtures into a vessel or column.

This Appendix provides the information for designing a Schoepentoeter and enabling the completion of a standard requisition by the Contractor.

2. DESIGN PROCEDURE OF SCHOEPENTOETER (VANE-TYPE) INLET DEVICES (TYPES I TO IV)

2.1 INTRODUCTION

Figure III.1 shows schematically the typical outline of a Schoepentoeter in a vertical vessel together with its design parameters (for simplicity not all the vanes are shown).

The geometry of the Schoepentoeter is largely standardised so that the choice of dimensions to be made by the designer is limited to the following:

a. the number of vanes per side, n_v

b. the vane angle, α, which is 8 degrees or less

c. the length of the straight part of the vanes, L_v,which shall be 75, 100, 150 or 200 mm*. The choice of L_v is also used to fix the vane spacing.

d. the radius of the vanes, R_v, which shall be 50 or 100 mm*.

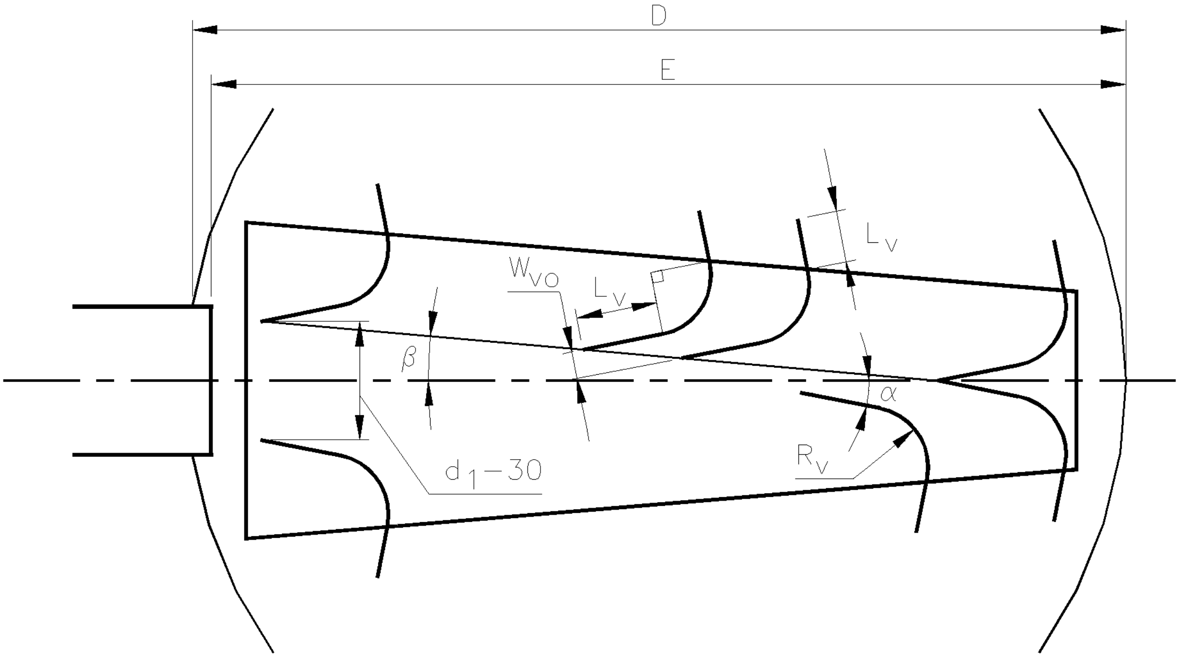
With a Schoepentoeter, it is normal to specify a protruded nozzle, although this is not essential.

Constructional details are shown in Figures III.2 to III.8 (which are specimen copies of Standard Drawings S 20.015, S 20.016, S 20.017, S 20.019, S 20.020, S 20.021, and S 20.022) included at the end of this Appendix. The Principal should be consulted if additional information is required.

* For special cases (vacuum operations) larger vanes with a larger L_V and R_V may be used.

2.2 DESIGN METHOD

FIGURE III.1 SCHEMATIC OUTLINE OF THE SCHOEPENTOETER



α = vane angle, angle made by straight part of vanes with centre line

β = edge angle, angle made by edge of the row of vanes with centre line

D = vessel inside diameter, mm

d_1 = inlet nozzle inner diameter, mm

E = available space, mm

L_v = length of straight part of vanes (normally 75, 100, 150 or 200 mm)

n_v = number of vanes per side

R_v = vane radius, mm (normally 50 or 100 mm)

t = vane material thickness, mm (normally 3 mm, but typically 5 mm for heavy duty, e.g slugs)

W_{vo} = width of vane entrance opening, mm

i) For a new vessel the inside diameter D and the inner diameter of the feed nozzle d_1 will be determined by process considerations.

For an existing vessel, D and d_1 will be known.

Schoepentoeters are not used in vessels of diameter less than 500 mm.

Schoepentoeters are only fitted on nozzles with $d_1 \geq 150$ mm.

If nozzle diameters $d_1 > 1/3$ D are encountered the Principal should be consulted.

ii) Evaluate the available space E.

For a vertical vessel, take	E = D - 50	(D ≤ 2 000 mm)
	E = D - 100	(2 000 < D ≤ 4 000 mm)
	E = D - 200	(D > 4 000 mm)

In Schoepentoeter assembly type IV, D/2 instead of D shall be taken in the expression for E and each Schoepentoeter of the type IV configuration shall be designed with the general sizing rules given below.

If $E > 5d_1$ then the available space is excessive and a shortened Schoepentoeter (Types II to IV) should be used; in which case the required space, E, should be reassessed.

NOTE For distillation columns, $E > 5d_1$ may be acceptable with the approval of the Principal.

For a horizontal vessel, $3\ d_1 \le E \le 5\ d_1$

It should be noted that the **inner** (rather than the outer) diameter of the feed nozzle is taken in designing the Schoepentoeter.

iii) Evaluate $X = (E - 270)/(d_1 - 30)$ and select L_v and R_v from the table below:

X	E (mm)	L_v (mm)	R_v (mm)
≤ 2.5	≤ 550	75	50
≤ 2.5	> 550	100	50
$2.5 < X \le 6$	> 550	100	100
$6 < X \le 20$	> 550	150	100
X > 20	> 550	200	100
< 6	> 6 000	400	300 *

* For vacuum operations only; for other applications R_v = 100 mm.

iv) Calculate number of vanes per side, n_v, from:

$n_v = (E - R_v - 70) / L_v$

rounded down to the nearest integer.

v) Evaluate $\tan \beta = (d_1 - 30) / \{ 2(n_v-1)L_v \}$

vi) Choose α = 8 degrees initially (maximum value)

vii) Evaluate W_{vo} from

$W_{vo} = L_v (\sin \alpha + \cos \alpha \tan \beta) - t$ mm

W_{vo} should be in the range of $W_{vo,\ min}$ to $W_{vo,\ max}$:

if R_v is 50 or 100 mm: $20 \le W_{vo} \le 30$ mm

if R_v is 300 mm: $40 \le W_{vo} \le 80$ mm

viii) if R_v is 50 or 100 mm:

If $W_{vo} > 30$, reduce α in steps of 1 degree (minimum value 0°) until

$W_{vo} \le 30$ mm

If $W_{vo} < 20$ mm, reduce the Schoepentoeter length, i.e. reduce n_v.

If $20 \le W_{vo} \le 30$ mm, the design is finished and n_v, R_v, L_v and α are selected.

ix) if R_v is 300 mm:

If $W_{vo} > 80$, reduce α in steps of 1 degree (minimum value 0°) until

$W_{vo} \le 80$ mm

If W_{vo} < 40 mm, reduce the Schoepentoeter length, i.e. reduce n_v.

If $40 \leq W_{vo} \leq 80$ mm, the design is finished and n_v, R_v, L_v and α are selected.

2.3 VANE HEIGHT

Each row of vanes in a Schoepentoeter is welded to a mounting strip at top and bottom. The assembly, which includes a row of vanes, is known as the vane ladder. It is not normal to specify the height of the vanes, since the height will be made sufficient to cover the inlet nozzle. However, the designer should give some consideration to the height of the vane ladders in the following cases:

2.3.1 Schoepentoeter to enter through its own nozzle

It is common for the vane ladders to be brought separately into the column, and the device assembled "in situ". Sometimes the ladders have to be brought in through their own nozzle, for example when this is the largest opening in the column wall. When this is to be done, the vane ladders shall be adapted, either by applying a "catcher cap" (Figure III.7) or by dividing the vane ladder (Figure III.8). For critical separator applications where very high liquid removal efficiency is required and the nozzle size just meets the momentum criterion, a "catcher cap" is not allowed since it would increase the inlet velocity and negatively affect the separator performance.

2.3.2 Vane height greater than 800 mm

Tall vanes are susceptible to vibration. Therefore if the vane height exceeds 800 mm, the vane ladder should be divided for greater rigidity.

2.3.3 Schoepentoeter subject to heavy load

There are circumstances in which the Schoepentoeter is subjected to a load heavier than one would normally design for. Examples are Schoepentoeters fitted to existing nozzles during revamps, or in flare system knock-out drums where exceptionally high flow rates are possible in severe relief situations. Sometimes Schoepentoeters are required to withstand liquid slugs. In these cases consideration should be given to a divided ladder construction (Figure III.8) and/or to a larger vane material thickness, typically 5 mm.

3. REQUISITIONING

Data/requisition sheet DEP 31.20.41.93-Gen. shall be used for requisitioning a Schoepentoeter. A specimen copy is included at the end of this Appendix as Figure III.9. The data to be entered shall include the **INNER** diameter of the nozzle, the number of vanes per side, the length of the straight part of the vanes, the vane radius and the vane angle.

It should also be specified via which manhole or nozzle the Schoepentoeter has to be installed.

Since the data/requisition sheet makes reference to the relevant standard drawings, it is not necessary to include the drawings with the requisition. Other sketches showing vane details should not be used in requisitioning.

The Manufacturer is required to make some checks as indicated on the standard drawings.

FIGURE III.2 S 20.015 - TYPICAL DETAILS OF SCHOEPENTOETER - TYPE II

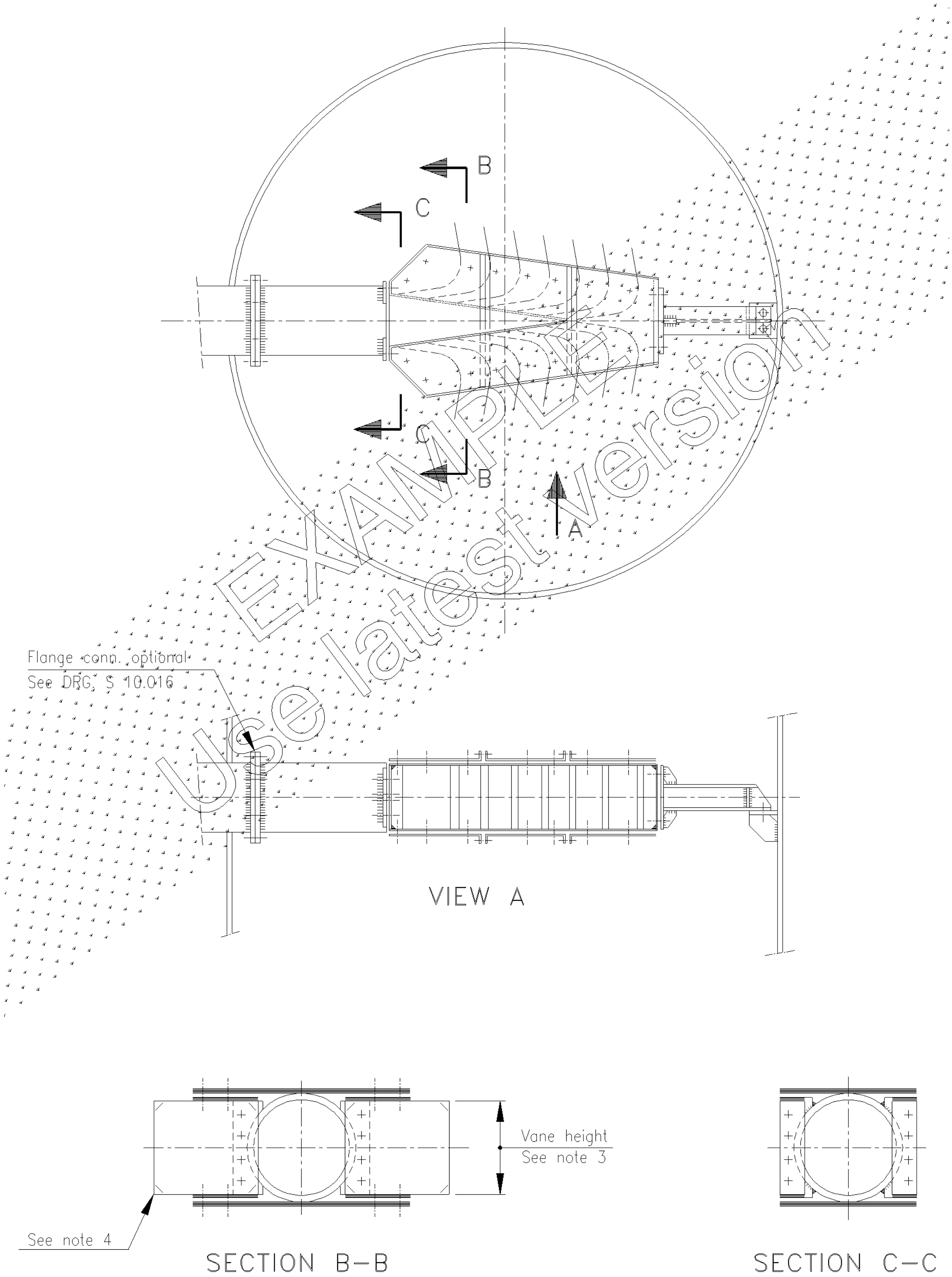


FIGURE III.3 S 20.016 - TYPICAL DETAILS OF SCHOEPENTOETER - TYPE III

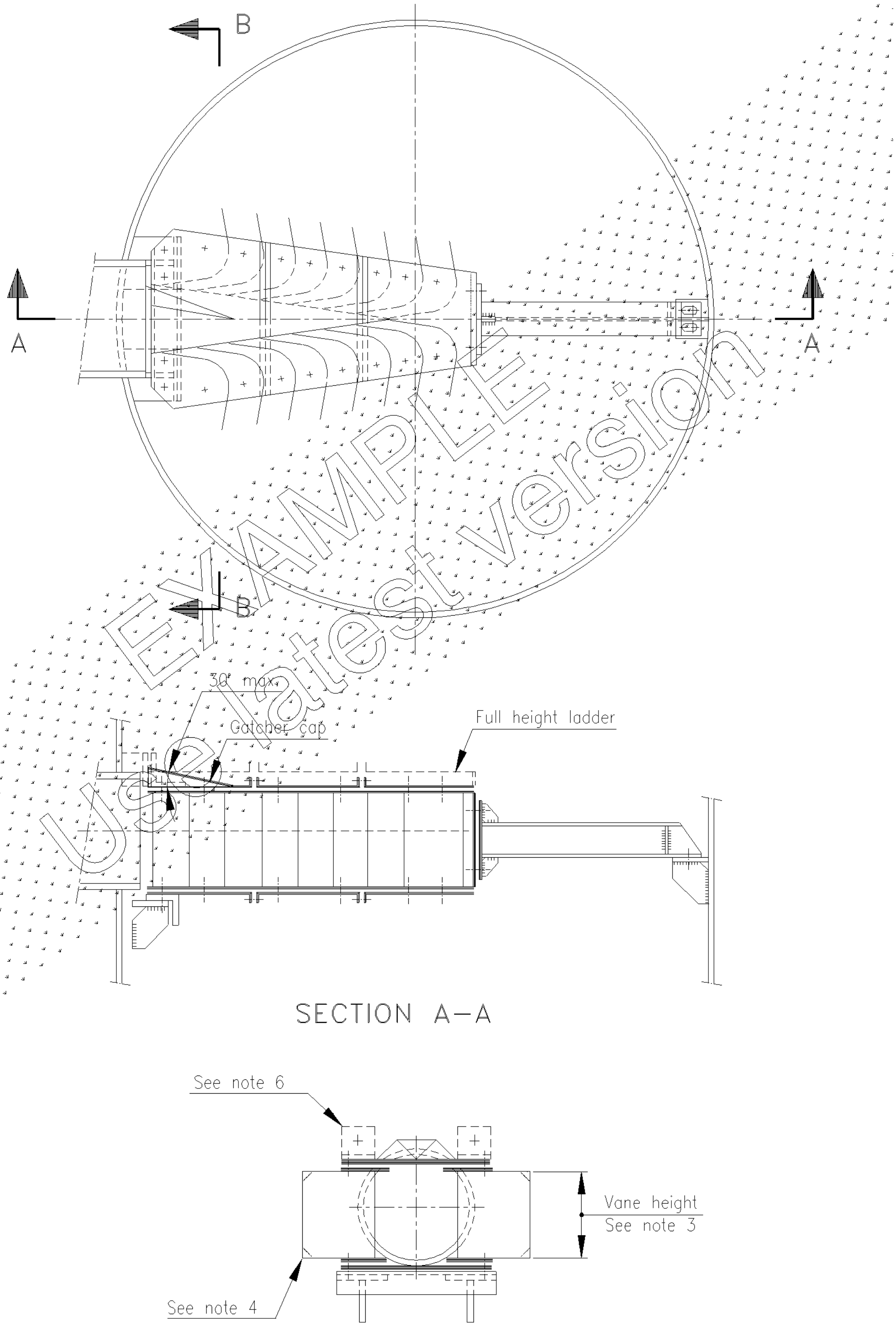


FIGURE III.4 S 20.017 - TYPICAL DETAILS OF SCHOEPENTOETER - TYPE IV

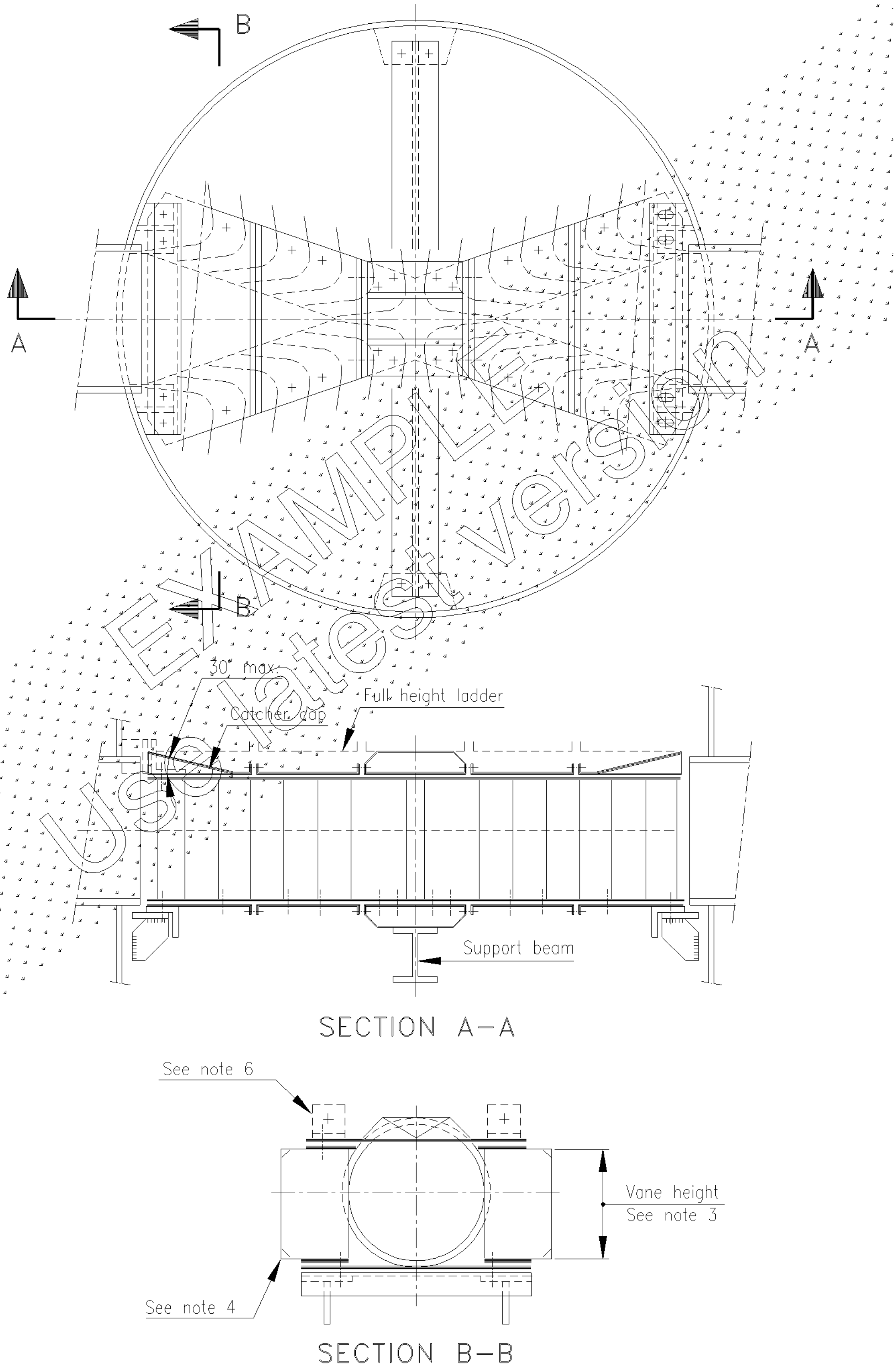
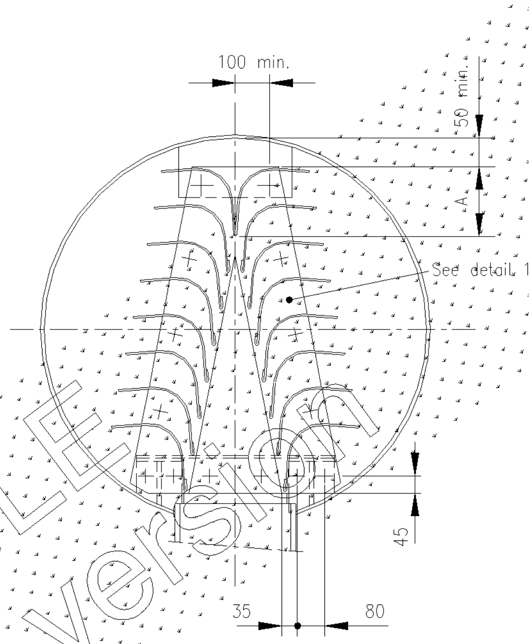
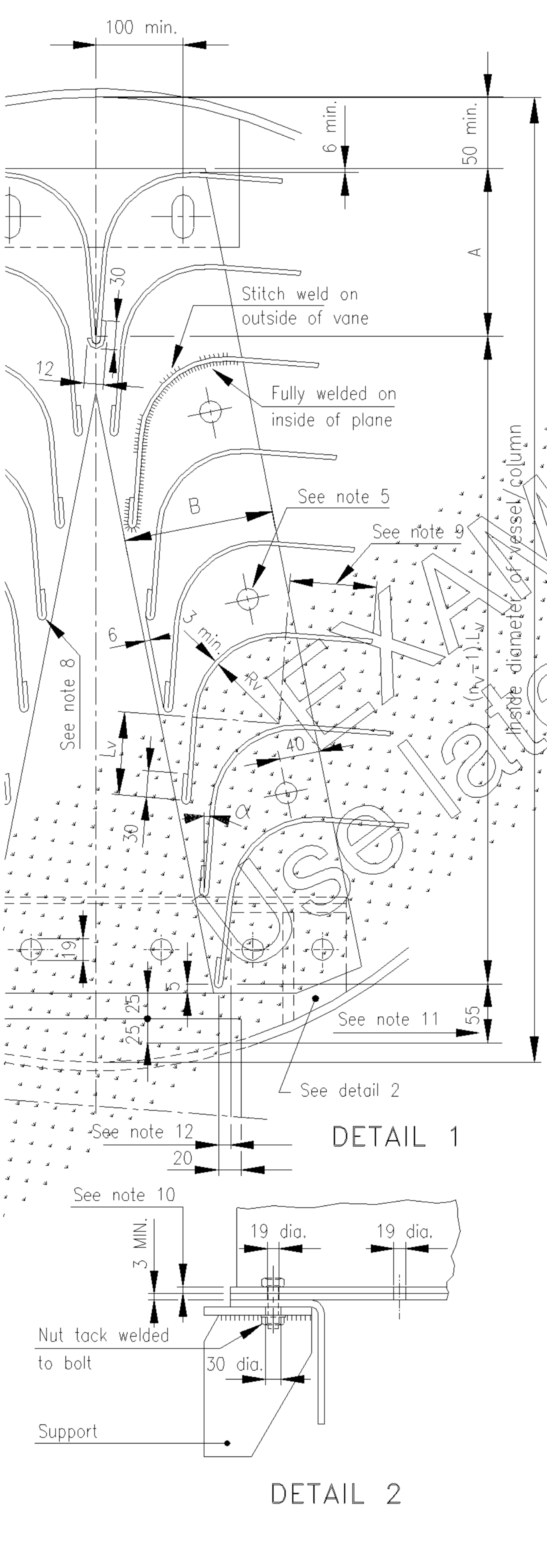


FIGURE III.5 S 20.019 - TYPICAL DETAILS OF SCHOEPENTOETER - VANE LADDER - SUPPORTS AND GENERAL NOTES





L_V = Straight length of vanes: 75, 100, 150, 200 or 400mm.

Vane radius = R_V	A (Max.)	B (Approx.)
50	65 + L_V	120
100	115 + L_V	170
300	715	435

NOTES:

1 All dimensions in mm.
2 Should any thermal expansion difference occur, clearance should be made at the supports.
3 All parts of the vane feed inlet shall pass through the inlet nozzle or manhole, if necessary, a catcher-cap or multiple section vane ladder can be applied.
See drawings S 20.021 and S 20.022 resp.
4 Corners of the vanes to be removed only if required for passing through the inlet nozzle or manhole.
5 Bolt size M16, max. spacing 250 mm.
6 If the total height of the vane feed inlet exceeds 800 mm, the top of the vane feed inlet should be guided.
7 If the height of the vanes exceeds 800 mm, the vane ladder should be divided into sections.
See drawing S 20.022.
8 The vane inlet should be folded perpendicular to the rolling direction of the vane plate, to avoid cracking. For plate ≤ 4 mm, reinforcing with a welded-on strip is not allowed.
9 The trailing straight length of the vane equals L_V, but is max. 200 mm.
10 The minimum thickness of top and bottom plate of vaneladders is 3 mm for schoepentoeters upto 2500 mm lenght and 6 mm for longer schoepentoeters.
11 For non protruding nozzles the distance of the first vane to the column wall is 30 - 55 mm, and the top and bottom plates of the schoepentoeter should be curved to give a clearance of 25 mm with the columnwall.
12 For nozzle wall thicknesses above 10 mm the distances between the first vanes of the ladders and the inner diameter shall be equal to 15 mm.

FIGURE III.6 S 20.020 - TYPICAL DETAILS OF SCHOEPENTOETER - TYPE IA WITH SINGLE SECTION VANE LADDER, MAX. HEIGHT 800 MM

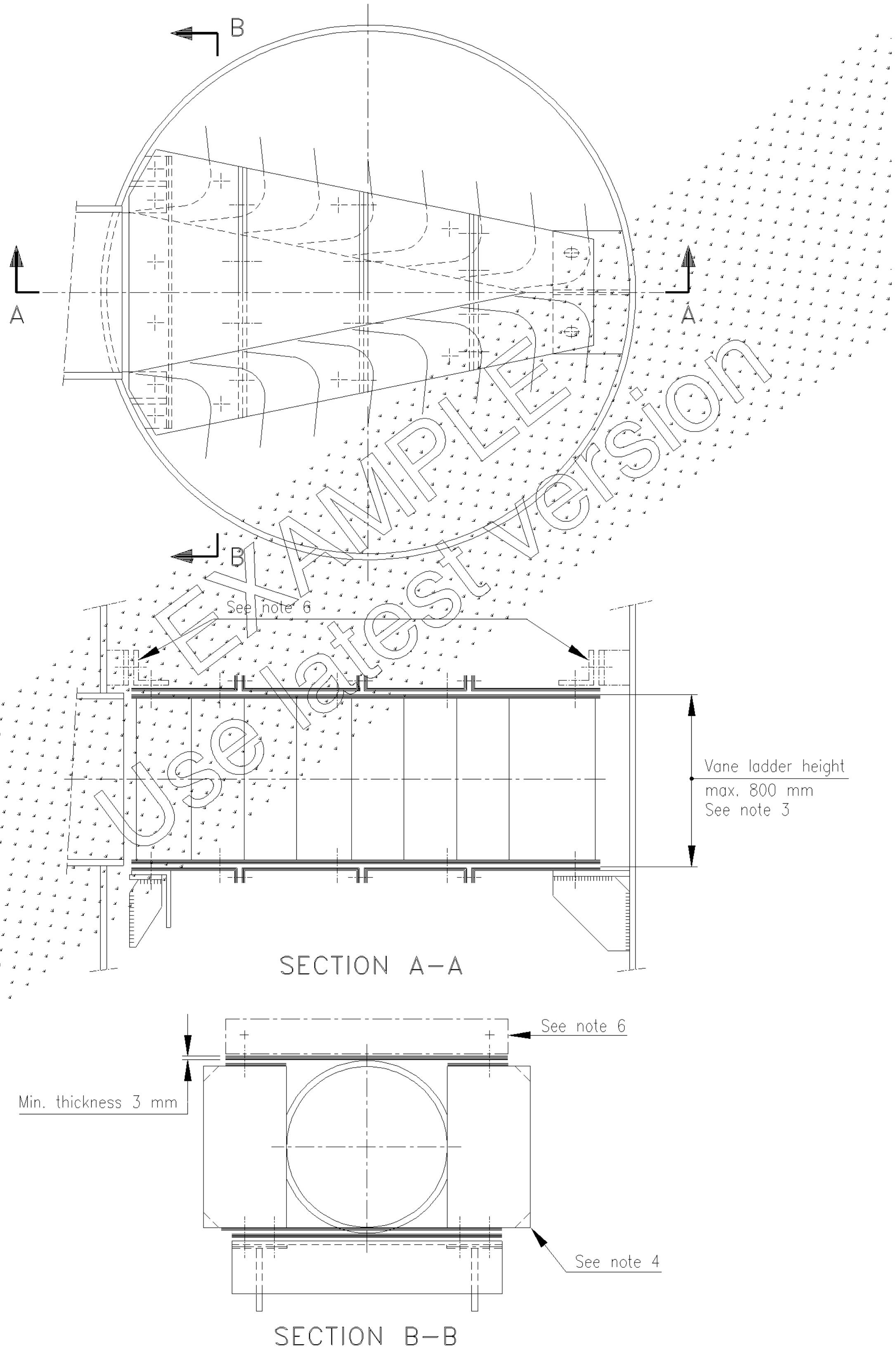


FIGURE III.7 S 20.021 - TYPICAL DETAILS OF SCHOEPENTOETER - TYPE IB WITH CATCHER CAP

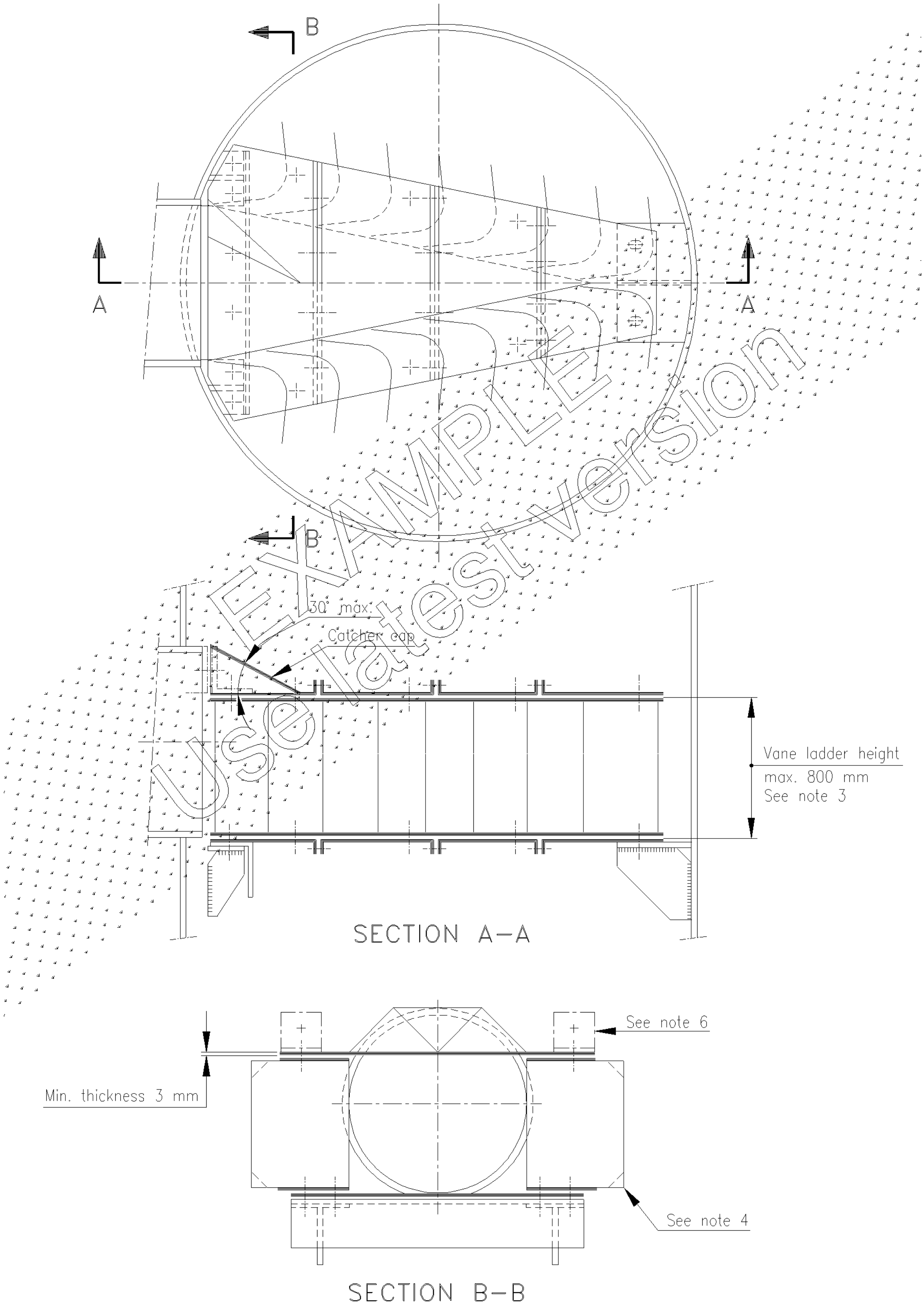
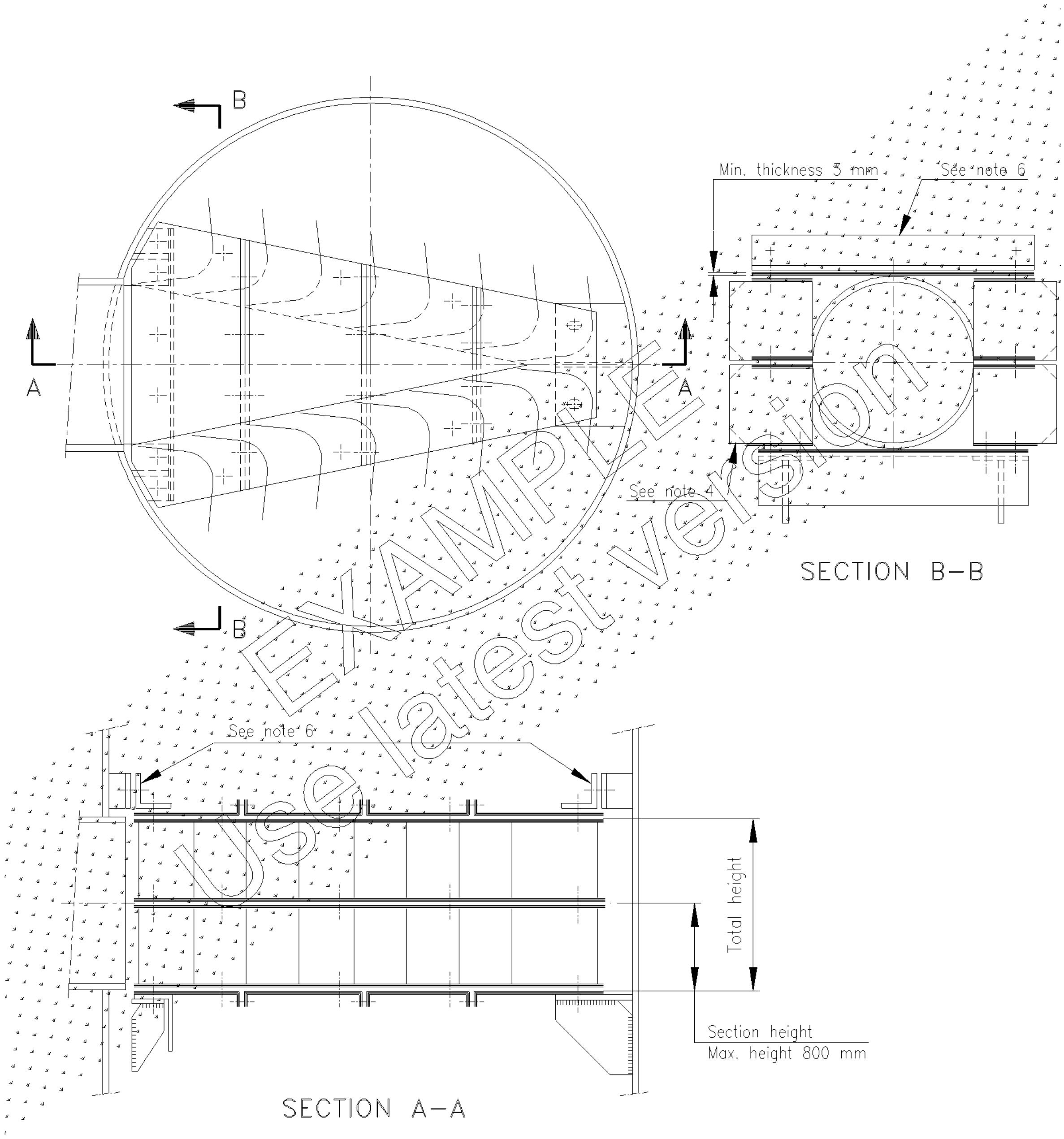


FIGURE III.8 S 20.022 - TYPICAL DETAILS OF SCHOEPENTOETER - TYPE IC WITH MULTIPLE SECTION VANE LADDER



REMARK:

Multiple section vane ladders are required:

a – To reinforce the vane ladders when their height is exceeding 800 mm in which case a single reinforcing plate would also be sufficient.

b – When the inlet nozzle or manhole restricts the installation of a single section vane ladder.

FIGURE III.9 DATA/REQUISITION SHEET DEP 31.20.41.93-GEN. FOR SCHOEPENTOETER

INFORMATION FOR USE WITHIN THE GROUP ONLY

Proj. dur.	Urg.	RAP / ROS date	T	T	D	Prefix	Project No.	Sub-group No.	Serial No.	Symbol	Card	POE Dest.	Ind.	Var. No.	Date d, m, y	OCDA	365	012	Sep. code	Status
34	36	38 – 43	1		3	4			10	18	17	18	20	21	23 – 28	29	30	31	32	33
			4	3							A									

Requisition/Indent No. 1 – 16

Sheet No. 1

47 Accounts code 82

	Reference indicator	Initial	Date
Made by			
Checked by			

Purchase action by: Central Purchasing/Local company

Authorised for purchase action	Signature	Ref. ind.	Date

Recommended supplier:

Data/Requisition for SCHOEPENTOETER (VANE-TYPE) INLET DEVICE

GENERAL		SCHOEPENTOETER, TYPE I, II, III, IV *)	
Special requirements: see DEP 31.22.05.11-Gen.		Construction requirements	To be submitted with tender
Column/vessel/separator no:		Number of vanes per side	Actual vane opening: mm
Drwg no:		Vane angle : degree	Adequacy of submitted information yes/no*)
Nozzle no:		Vane radius R_v : 50/100/300 *) mm	
Nozzle inside diameter	mm	Vane opening	
Type of Schoepentoeter: I/II/III (CA)/IV/HiFi*)		– R_v is 50 or 100 : between 20 and 30 mm	
		– R_v is 300 : between 40 and 80 mm	
Applicable Standard Drawings		Length of straight part of vanes:	
– for type I to IV: S 20.0.15/16/17/19/20/21/22*)		75/100/150/200/400*) mm	
– for type HiFi : S 20.025/26*)			
		Catcher cap allowed yes/no	
Schoepentoeter material:		SCHOEPENTOETER TYPE HiFi	
Corrosion allowance :	mm	Construction requirements	To be submitted with tender
		HiFi tray below	The supplier shall confirm in the tender that:
All parts of Schoepentoeter to pass through a		– column inside diameter mm	– vane distribution is based on nozzle inside diameter
nozzle/manhole/dome* with inside		– number of calming sections	– Schoepentoeter c/l coincides with supportbeam tray under
diameter	mm	(equals number of vanes)	– the trailing straight part of the vanes coincides with
			the c/l of panels of tray below
		Nozzle inside diameter mm	– there is no interference with other internals
			– submitted information is adequate*)

NOTES AND REMARKS	REMARKS ON REVISIONS	Var. No.	Rev. letter	Date	Initial
* The manufacturer shall check the information on equipment requisition/drwg and on the standard drwg for compatibility.					
He shall indicate in the tender all missing information.					
* Delete what is not applicable.					
PLANT :	ENG. BY :	Contractor Job No.			
CONSIGNEE:	PRINCIPAL:				

79 Estimated indent value 80	72 ROS date d, m, y 77

Estimated item value (65–70)	New item (71)	Quantity change ±	S/C (78)	From store (64–65)	OCDA (67)	Card (17/72)	Rev. letter (18/73)	Item No. (19/74–75)	Total quantity required (21–26)	Unit (27)	Additional description (30)	Equipment No.	MESC No. (55–64)
£						D							
						D							
						D							

DEP 31.20.41.93-Gen., 10/90

APPENDIX IV DESIGN MARGINS FOR SEPARATOR DESIGN

To determine the highest envisaged gas and liquid load for vessel design, design margins (surge factors) are required:

The design margins should be supplied by the Principal.

Typical values are:

In SIEP applications:

1.	**Offshore service**	**Design margin**
	Separator handling natural-flowing production from:	
	a. its own platform	1.2
	b. another platform or well jacket in shallow water	1.3
	c. another platform or well in deep water	1.4
	Separator handling gas lifted production from:	
	a. its own platform	1.4
	b. another platform or well jacket	1.5
2.	**Onshore service**	
	Separator handling natural flowing production, or gas plant inlet separator in:	
	a. flat or low rolling country	1.2
	b. hilly country	1.3
	Separator handling gas lifted production in:	
	a. flat or low rolling country	1.4
	b. hilly country	1.5

In SIOP and SIC applications:

The design margin ranges typically from 1.15 to 1.25

APPENDIX V LEVEL CONTROL

In the lower part of Figure V.1 the levels of the instrumentation nozzles are indicated for both a vertical and a horizontal vessel.

LZA(LL) (low level trip) is 0.15 m above vessel bottom (horizontal vessel) or 0.15m above BTL (vertical vessel).

LA(L) (low level pre-alarm) is either at least 0.10 m above LZA(LL) or, if required, located such that there is sufficient liquid hold-up time between the two levels for operator intervention (to be specified, but typically 60 seconds for action in the control room and 5 minutes for action outside the control room).

The minimum distance between LA(H) (high level pre-alarm) and LA(L) is 0.35 m (the standard 14" bridle). The distance between LA(L) and LA(H) shall be such that there is sufficient liquid hold-up time between the two levels for control purposes.

In SIOP the following general principles for hold-up times for control are applied:

1 minute on "short" circulation flows (i.e. back to inlet of vessel) PLUS

2 minutes on longer circulation flows (other upstream vessels within unit) PLUS

2 minutes on product to storage OR 3 minutes on product to other equipment/ vessels

OR 4 minutes on product to a furnace.

If slugs are expected, they should be accommodated between NL (normal level) and LA(H).

If the volume of the slug to be expected is not known, 2 to 5 seconds of flow with the maximum feed (gas + liquid) velocity and 100% liquid filling of the pipe should be taken as that volume.

In general the slug size is influenced by the layout of the upstream piping. In case of doubt the Principal should be consulted for an estimate of the slug size.

In practice the total volume to be provided between LA(L) and LA(H) is the sum of the required control volume and the volume of the anticipated slug.

LZA(HH) (high level trip) is either at least 0.1 m above LA(H) or located such that there is sufficient liquid hold-up time between the two levels for operator intervention (to be specified, but typically 60 seconds for action in the control room and 5 minutes for action outside the control room).

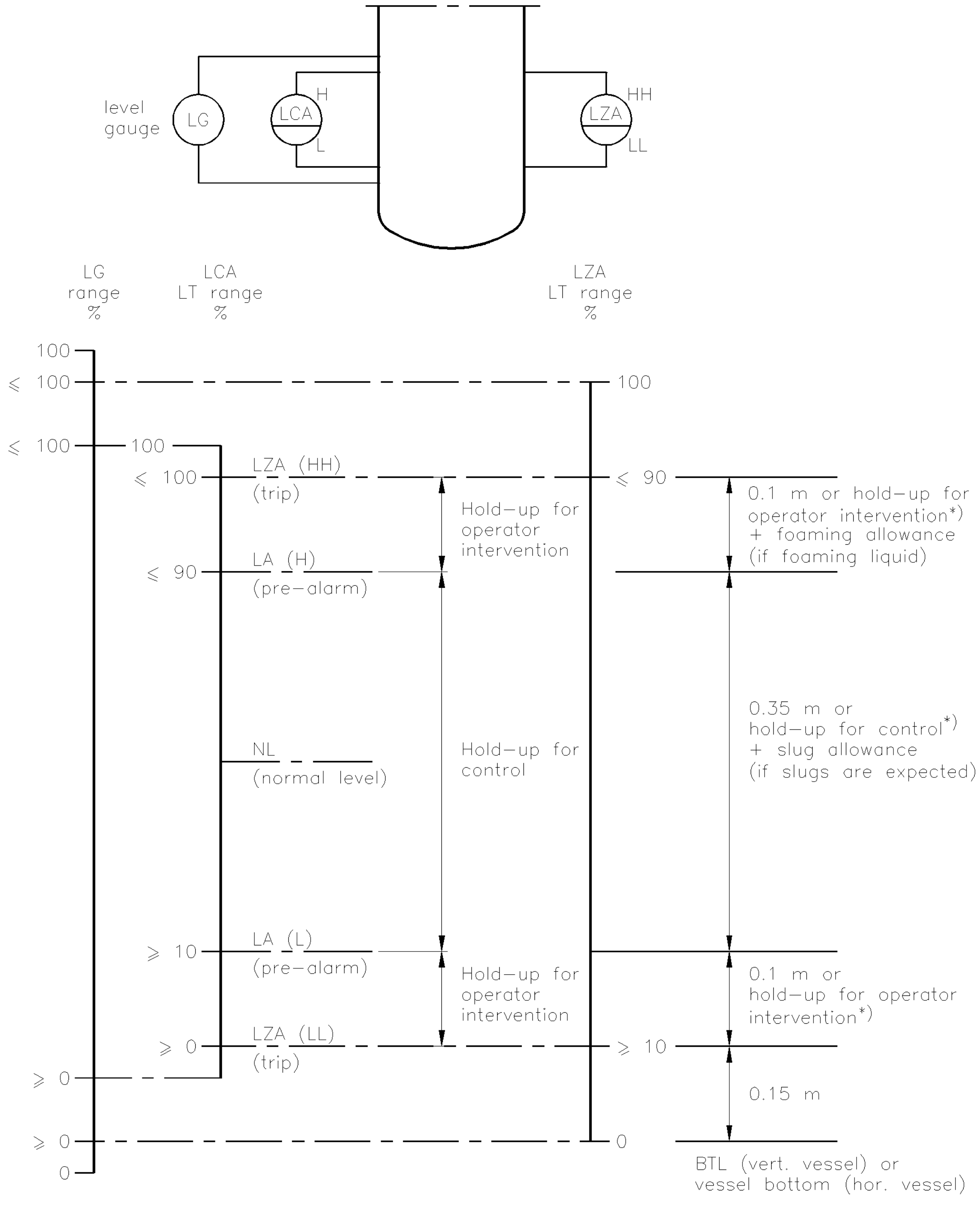
If the liquid has a foaming tendency, the distance between LA(H) and LZA(HH) has to be increased by a further 0.25 m.

The gas compartment of the vessel above LZA(HH) shall be sufficiently large to allow proper G/L separation. The sizing of this gas compartment is dependent on the type of G/L separator and is addressed in the corresponding section of this DEP.

In the case that operator action to prevent a trip is not realistically possible, the pre-alarm should be discarded.

NOTE The liquid heights given in this Appendix are for a two-phase gas/liquid separator. For the design of three-phase separators, DEP 31.22.05.12-Gen. should be consulted.

FIGURE V.1 LIQUID LEVEL CONTROL IN A GAS/LIQUID SEPARATOR



*) whichever is larger

APPENDIX VI GEOMETRICAL RELATIONSHIPS BETWEEN CHORD AREA AND CHORD HEIGHT

In the calculation procedures presented in Appendix VII for the sizing of horizontal G/L separators, frequently the boundaries have to be determined of the horizontal zones occupied by the gas and liquid phases using as input the respective cross-sectional areas.

Also the reverse calculation (calculation of the respective cross-sectional areas of the gas and liquid phases from the respective lower and upper boundaries) is often required.

In this Appendix the relationships will be given which enables this type of calculations, both graphically and via equations.

Figure VI.1 shows schematically a cross-section of a horizontal G/L separator with diameter D.

The positions of the liquid and gas phase have been indicated and have as cross-sectional area A_L and A_G respectively.

The height of the G/L interface is at h_1.

For a more general representation the areas and heights are made dimensionless by dividing them by the vessel cross-sectional area and the vessel diameter respectively.

$A^*_L = A_L /(\pi D^2/4)$ $\qquad$ $A^*_G = A_G /(\pi D^2/4)$

$(A^*_G + A^*_L = 1)$ $\qquad$ $h^*_1 = h_1/D$

A^*_L and A^*_G are now dimensionless chord areas with dimensionless chord heights, h^*_1 and $(1-h^*_1)$ respectively.

In Figure VI.2 the general relationship between a dimensionless chord area, A^* and its associated dimensionless chord height, h^*, is presented.

This graph gives directly the relationship between A^*_L (= $(1-A^*_G)$ and h^*_1. From this followsthe relationship between the cross-sectional areas of the various phases and their boundaries in the vessel, taking into account the vessel diameter D.

The Figure can also be used directly to determine the cross-sectional areas of the various G/L level control bands as a function of their upper and lower boundaries.

It also enables a control band boundary to be directly located if the cross-sectional area of the control band and the location of the other boundary is known.

Relationships

Both h* and A* can be expressed in terms of φ (see also insert in Figure VI.2)

$A^* = 0.5*(\varphi - \sin\varphi)/\pi$

$h^* = 0.5*\{1 - \cos(\varphi/2)\}$

h* → A*

Since h^* can also be expressed in terms of φ, A^* can be directly calculated from h^* via:
$\varphi = 2*\arccos(1- 2*h/D)$ and subsequently $A^* = 0.5*(\varphi - \sin\varphi)/\pi$.

A* → h*

Calculation of h* from A* requires iteration:

The iteration formula is:

$\varphi_{i+1} = \varphi_i - (2*\pi*A^*- \varphi_i +\sin\varphi_i)/(\cos\varphi_i -1)$

The start value for φ is not critical. Even the simple start $\varphi_0 = \pi$ always gives convergence.
After convergence, h* follows then from:

$h^* = 0.5 * \{1 - \cos(\varphi/2)\}$

VOLUME OF VESSEL HEADS

The interface level control band of horizontal G/L separators will include a part of the vessel heads.

The relationship for this volume, ΔV_{hd}, as a function of the vessel diameter D, and its lower and upper boundary, h_1 and h_2 respectively, is given below.

$\Delta V_{hd} = \alpha\pi\, D^3\{0.75 * (h^*_2 - h^*_1) - (h^*_2 - 0.5)^3 + (h^*_1 - 0.5)^3\} / 6$

h^*_1 and h^*_2 are the dimensionless lower and upper boundary respectively and are defined as:

$h^*_1 = h_1/D$

$h^*_2 = h_2/D$

α is the ratio of the short to the long axis of the vessel head

The most common semi-elliptical head has an α of 0.5.

In rare cases (at very high operating pressure, for instance) a semi-spherical head is specified: $\alpha = 1$.

FIGURE VI.1 SIMPLIFIED CROSS-SECTION OF A HORIZONTAL GAS/LIQUID SEPARATOR WITH THE POSITION AND CROSS-SECTIONAL AREAS OF THE TWO PHASES INDICATED

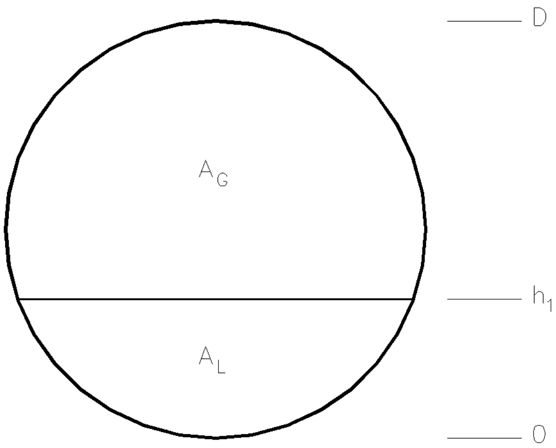
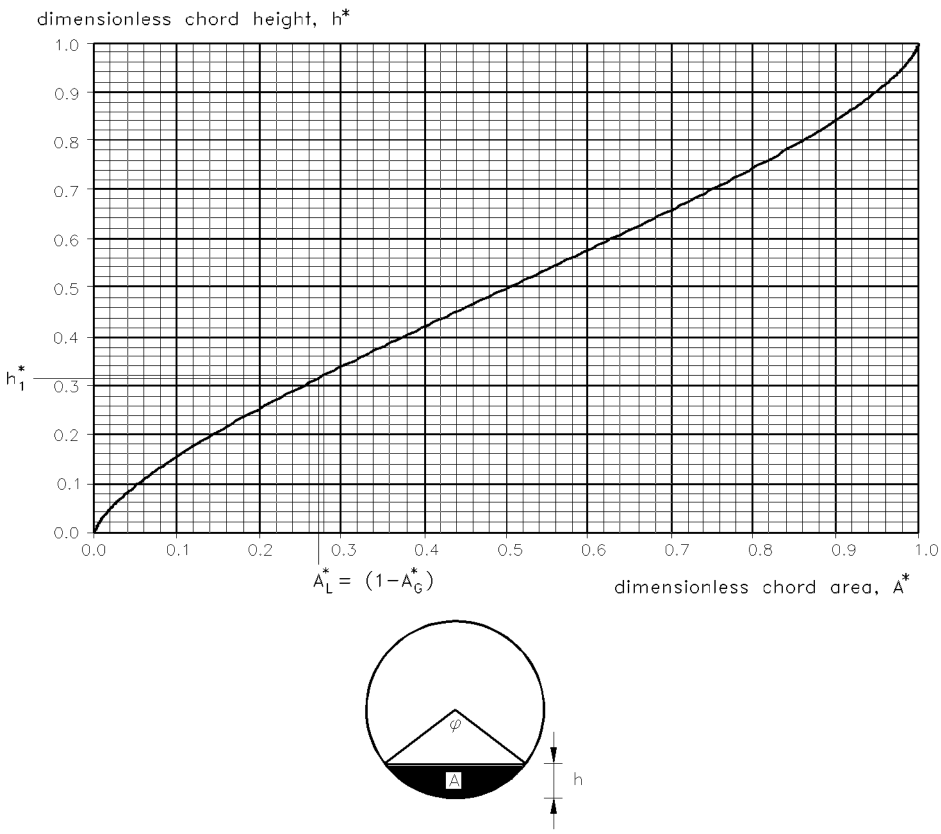


FIGURE VI.2 RELATIONSHIP BETWEEN THE DIMENSIONLESS CHORD AREA, A*, AND DIMENSIONLESS CHORD LENGTH, H*, OF A VESSEL CROSS-SECTION



$$A^* = A/A_{ves} = 0.5*(\varphi - \sin\varphi)/\pi$$
$$h^* = h/D = 0.5*\{1-\cos(\varphi/2)\}$$

APPENDIX VII SIZING OF SEPARATOR VESSELS

1. VERTICAL VESSEL

The vessel diameter, D, shall satisfy the **gas handling capacity criterion.**

This criterion is addressed for each separator individually.
See the appropriate part of (Section 3) in the main text.

Where vapour carry-under is not allowed, the vessel diameter shall satisfy the **liquid de-gassing criterion:**

In practice it can be assumed that if bubbles larger than 200 µm in size are able to escape, the carry-under will be negligible. This means that the downward velocity of the liquid shall satisfy the following requirement:

$$v_L \leq Q_{L,max}/(\pi D^2/4) = 2.2*10^{-8}(\rho_L-\rho_G)/\eta_L$$

or $$D \geq 7608 \sqrt{Q_{L,max}\eta / \rho_L - \rho_G)}$$

When the liquid has a **foaming** tendency the downward velocity of the liquid has also to be minimised by applying the **de-foaming criterion:**

$$D \geq 95Q^{0.5}_{L,max} \{\eta_L/(\rho_L-\rho_G\ )\}^{0.14}$$

It is possible in the case of a viscous or foamy liquid that one of the latter equations will determine the vessel diameter.

This has been demonstrated in Figure VII.1 for a vertical wiremesh demister.
This figure presents the minimum required vessel diameter as a function of the liquid viscosity, calculated according to three different criteria. It is seen that with a high flow parameter both the de-foaming and de-gassing criteria overrule the gas handling capacity criterion.
With a low feed flow parameter, the gas handling capacity criterion will determine the vessel diameter if the liquid viscosity is not too high (in this particular example the viscosity should be lower than 0.04 Pa.s). For higher viscosities the de-gassing criterion will determine the vessel diameter.

2. Horizontal vessel

In the case of horizontal knock-out drums, wiremesh demisters and vane-type demisters, the vessel diameter is derived straightforwardly after considering the requirements for both gas and liquid.

In the case of horizontal separators equipped with cyclones, the procedure is less straightforward, since the dimensions of the gas compartment (above LZA(HH)) are primarily determined by the minimum height of the drainpipes of the cyclone deck above LZA(HH) and the space occupied by the cyclone deck which in turn is a function of the number and diameter of cyclones and the layout of the cyclone deck .

If this type of separator is selected, It is recommended to size this vessel in cooperation with the Manufacturer of the cyclones.

A sizing procedure for the horizontal knock-out drums, wiremesh demisters and vane-type demisters is given below.
The determination of the vertical cross-sectional area for gas flow is dealt with in (3.) in the main text for each separator individually.

In general, the liquid-full section of the horizontal vessel determines the size of the separator.
The following steps detail the design approach:

i) Determine the minimum height requirements for level control (see Appendix V) including an allowance for foam or slugs (if required).
ii) Determine the minimum vessel diameter required to accommodate $A_{G,min}$ using Table VII.2, which gives A_G as a function of the vessel diameter D and LZA(HH).

In this Table it has been indicated for each of the three separators which (D-LZA(HH)) combinations are allowed (left or below the demarcation step lines in the Table).
The minimum allowable vessel diameter and maximum allowable filling degree (100*LZA(HH)/D) are also specified in Table VII.1 below for the three types of horizontal separators.

Table VII.1 Minimum diameter and maximum filling degree of the horizontal knock-out drum, wiremesh demister and vane-type demister

	min D, m (non-foaming)	min. gas cap height, m	max filling degree, %
Knock-out	1.00	0.3	80
Wiremesh	1.30	0.6	60
Vane-type	1.50	0.8	50

As a starting point it can be assumed that the liquid level is set by the minimum requirements for control.
This means LZA(HH) = 0.70 m for a non-foaming system and 0.95 m for a foaming system.
Table VII.2 gives then the corresponding D.

iii) Calculate LA(H) and LA(H).

Take as initial values: LA(L) = LZA(LL) + 0.1 m and LZA(LL) = 0.15 m

LA(H) = LZA(HH) -0.1 m (non-foaming service)

LA(H) = LZA(HH)-0.35 m (foaming service)

Calculate the vertical cross-sectional area, A_{H_L}, between LA(L) and LA(H) (i.e. the cross-sectional area of the liquid volume for control and slug requirements) with the aid of the chord area-chord height relationships (or Figure VI.2) in Appendix VI.

iv) The tangent-to-tangent length of the vessel, L, follows then from

$L = (V_{slug} + Q_L t_{H\_L} - 2\Delta V_{hd,H\_L})/A_{H\_L}$

V_{slug} is the volume of the anticipated slug (if any) to be accommodated

$t_{H\text{-}L}$ is the required control time between the levels LA(L) and LA(H).

$\Delta V_{hd,H\_L}$ is the volume of the vessel head between LA(L) and LA(H) and can be calculated with the relationships presented in the second part of Apppendix VI.

Calculate L/D

If L/D < 2.5, take L = 2.5 D; go to step vi

If 2.5 ≤ L/D ≤ 6 then go to step vi

If L/D > 6 then go to step v.

v) Increase D and recalculate LZA(HH).

LZA(HH) is determined keeping A_G constant, with the chord area-chord length relationships presented in Appendix VI or by using Figure VII.2.

In extreme cases D cannot be increased further without increasing A_G as well, because LZA(HH)/D cannot exceed the prescribed maximum value.

A_G is then determined by the maximum allowable LZA(HH) at the selected D.

Return to step iii.

vi) Check that:

LZA(HH)-LA(H) and LA(L)-LZA(LL) are sufficiently large to meet the corresponding specified control time, taking into account a foaming allowance (if required).

$t_{L\_LL} = (L*A_{L\_LL} + 2\Delta V_{hd,L\_LL})/Q_L \geq$ specified control time

For non-foaming system:

$t_{HH\_H} = (L*A_{HH\_H} + 2\Delta V_{hd,HH\_H})/Q_L \geq$ specified control time

For foaming system:

$t_{HH\_H} = (L*A_{(HH-0.25)\_H} + 2\Delta V_{hd,(HH-0.25)\_H})/Q_L \geq$ specified control time

If the control times are met, go to step vii.

If the control times are not met, increase the width of the associated control band(s) by multiplying with the ratio of the specified and calculated control times and return to step iii.

vii) Check that:

- De-gassing is sufficient (if critical in the particular application): A similar formula as for the vertical vessel is used (VII.1) taking the area for the downward liquid flow as D*L.

This is the G/L interface area when the vessel is 50% liquid-filled. Thus:

$D \geq \{4.5*10^7 Q_{L,max} \eta_L/(\rho_L-\rho_G)\} / L$

- De-foaming is sufficient (if foaming is expected):

$D \geq 7\,000\ Q_{L,max}\{\eta_L/(\rho_L-\rho_G)\}^{0.27} / L$

- The gas cap is sufficiently large to accommodate the feed inlet device (half-open pipe or Schoepentoeter).

Also, sufficient distance shall be available between the bottom of the inlet device and LZA(HH).

The gas cap above LZA(HH) should have a central height of typically $2d_1 + 0.15$ m.

If not, increase D and return to step iii.

If only the feed inlet device cannot be accommodated and the other two criteria are met, the inlet configuration could be modified (if justified for gas/liquid separation).

A vessel sizing example for a horizontal knock-out drum is given in Appendix X

FIGURE VII.1 DIAMETER OF THE VERTICAL WIREMESH DEMISTER - EFFECT OF FLOW PARAMETER AND VISCOSITY

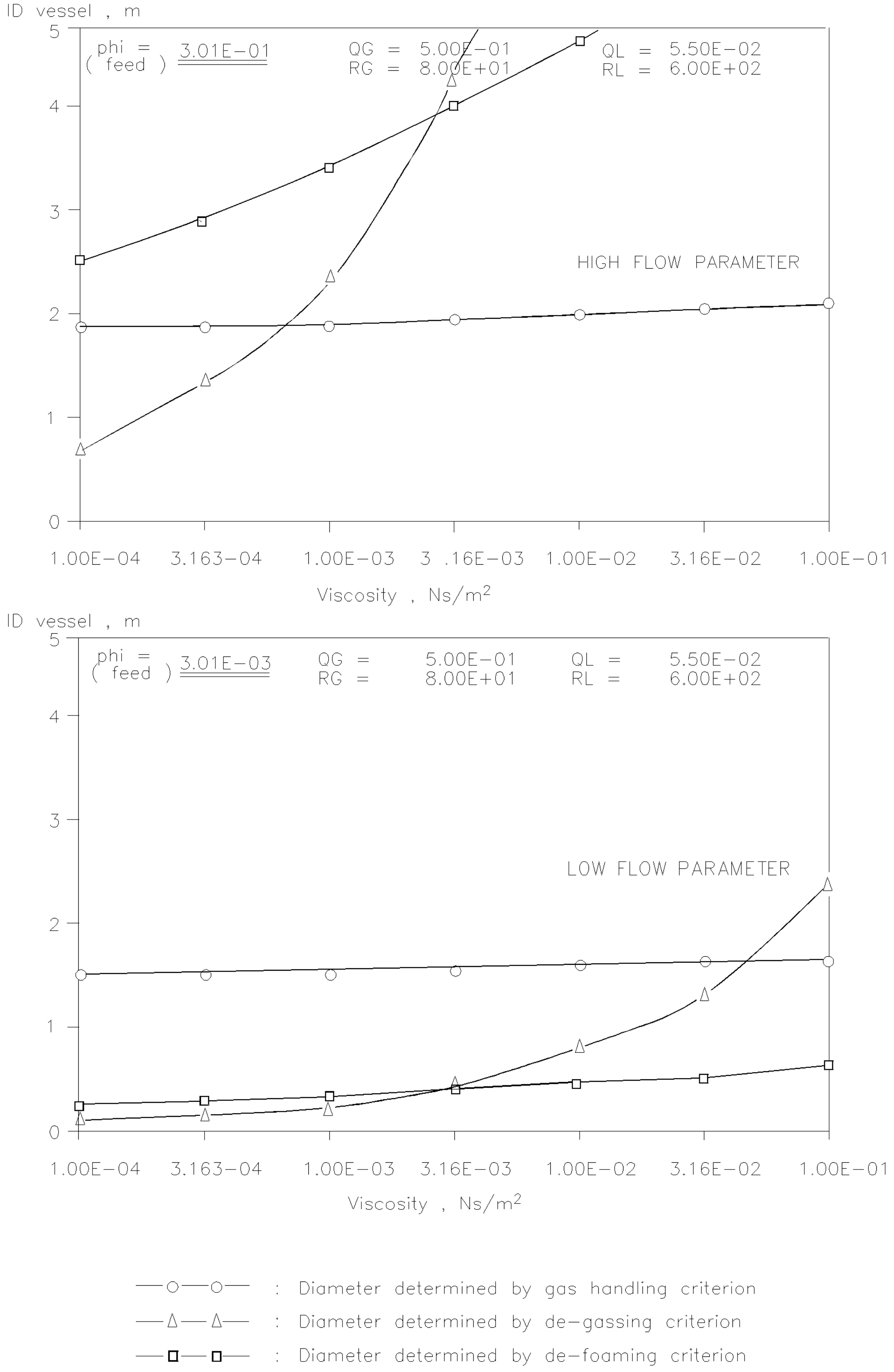


FIGURE VII.2: A_G (m^2) AS A FUNCTION OF D AND LZA(HH) IN HORIZONTAL SEPARATORS

	LZA(HH), m																		
D, m	**0.7**	**0.8**	**0.9**	**1.0**	**1.1**	**1.2**	**1.3**	**1.4**	**1.5**	**1.6**	**1.7**	**1.8**	**1.9**	**2.0**	**2.1**	**2.2**	**2.3**	**2.4**	**2.5**
1.0	0.20	0.11	0.04																
1.1	0.31	0.21	0.12	0.04															
1.2	0.45	0.33	0.22	0.12	0.05														
1.3	0.60	0.47	0.35	0.23	0.13	0.05													
1.4	0.77	0.63	0.49	0.36	0.24	0.13	0.05												
1.5	0.96	0.81	0.66	0.52	0.38	0.25	0.14	0.05											
1.6	1.16	1.01	0.85	0.69	0.54	0.39	0.26	0.15	0.05										
1.7	1.39	1.22	1.05	0.88	0.72	0.56	0.41	0.27	0.15	0.05									
1.8	1.63	1.45	1.27	1.09	0.92	0.74	0.58	0.42	0.28	0.15	0.06								
1.9	1.89	1.70	1.51	1.32	1.13	0.95	0.77	0.60	0.43	0.29	0.16	0.06							
2.0	2.16	1.97	1.77	1.57	1.37	1.17	0.98	0.79	0.61	0.45	0.30	0.16	0.06						
2.1	2.45	2.25	2.05	1.84	1.63	1.42	1.21	1.01	0.82	0.63	0.46	0.30	0.17	0.06					
2.2	2.76	2.55	2.34	2.12	1.90	1.68	1.46	1.25	1.04	0.84	0.65	0.47	0.31	0.17	0.06				
2.3	3.09	2.87	2.65	2.42	2.19	1.96	1.73	1.51	1.28	1.07	0.86	0.67	0.48	0.32	0.18	0.06			
2.4	3.43	3.20	2.97	2.74	2.50	2.26	2.02	1.78	1.55	1.32	1.10	0.88	0.68	0.50	0.33	0.18	0.06		
2.5	3.78	3.55	3.32	3.08	2.83	2.58	2.33	2.08	1.83	1.59	1.35	1.13	0.91	0.70	0.51	0.33	0.18	0.07	
2.6	4.16	3.92	3.68	3.43	3.17	2.91	2.65	2.39	2.14	1.88	1.63	1.39	1.15	0.93	0.71	0.52	0.34	0.19	0.07
2.7	4.55	4.31	4.05	3.80	3.53	3.27	3.00	2.73	2.46	2.19	1.93	1.67	1.42	1.18	0.95	0.73	0.53	0.35	0.19
2.8	4.95	4.71	4.45	4.18	3.91	3.64	3.36	3.08	2.80	2.52	2.25	1.97	1.71	1.45	1.20	0.97	0.75	0.54	0.35
2.9	5.38	5.12	4.86	4.59	4.31	4.02	3.74	3.45	3.16	2.87	2.58	2.30	2.02	1.75	1.48	1.23	0.99	0.76	0.55
3.0	5.82	5.56	5.29	5.01	4.72	4.43	4.13	3.83	3.53	3.23	2.94	2.64	2.35	2.06	1.78	1.51	1.25	1.01	0.77
3.1	6.27	6.00	5.73	5.44	5.15	4.85	4.55	4.24	3.93	3.62	3.31	3.00	2.70	2.40	2.11	1.82	1.54	1.28	1.03
3.2	6.74	6.47	6.19	5.90	5.59	5.29	4.98	4.66	4.34	4.02	3.70	3.38	3.07	2.75	2.45	2.15	1.85	1.57	1.30
3.3	7.23	6.95	6.66	6.36	6.06	5.74	5.42	5.10	4.77	4.44	4.11	3.78	3.45	3.13	2.81	2.50	2.19	1.89	1.60
3.4	7.73	7.45	7.16	6.85	6.54	6.21	5.89	5.55	5.22	4.88	4.54	4.20	3.86	3.52	3.19	2.86	2.54	2.23	1.92
3.5	8.25	7.96	7.66	7.35	7.03	6.70	6.37	6.03	5.68	5.33	4.99	4.64	4.29	3.94	3.59	3.25	2.92	2.59	2.27
3.6	8.79	8.49	8.19	7.87	7.54	7.21	6.87	6.52	6.16	5.81	5.45	5.09	4.73	4.37	4.01	3.66	3.31	2.97	2.63
3.7	9.34	9.04	8.73	8.41	8.07	7.73	7.38	7.02	6.66	6.30	5.93	5.56	5.19	4.82	4.45	4.09	3.73	3.37	3.02
3.8	9.91	9.60	9.29	8.96	8.62	8.27	7.91	7.55	7.18	6.81	6.43	6.05	5.67	5.29	4.91	4.54	4.16	3.79	3.43
3.9	10.49	10.18	9.86	9.53	9.18	8.82	8.46	8.09	7.71	7.33	6.95	6.56	6.17	5.78	5.39	5.00	4.62	4.23	3.86
4.0	11.09	10.78	10.45	10.11	9.76	9.40	9.02	8.65	8.26	7.87	7.48	7.08	6.68	6.28	5.88	5.48	5.09	4.69	4.30

↑ NO **HKO** ↓ YES

←YES **HV** → NO

←YES **HW** → NO

HKO : Horizontal knock-out drum

HW : Horizontal wiremesh demister

HV : Horizontal vane-type demister

APPENDIX VIII MISTMAT SPECIFICATIONS

MANUFACTURING

The demister mat shall be made of **knitted** wire formed to give the correct shape, and **not cut** so as to leave raw edges and/or loose pieces of wire which could become detached.

Demister mats are normally **stainless steel.**

The mistmat shall have:

- a free volume of at least 97% (ε = 0.97)
- a wire thickness,d_w, between 0.23 mm and 0.28 mm.

if above requirements are satisfied then the specific wire surface area
($= 4(1-\varepsilon)/d_w$) > 428 m^2/m^3

The thickness of a horizontal mat in a vertical vessel is normally 0.1 m. A vertical mistmat in a horizontal vessel shall have a thickness of at least 10% of the vessel diameter with a minimum of 0.15 m.

MOUNTING

The wire mat shall be placed between two grids having a free area of at least 97%. The mat shall be fastened in such a way that it cannot be compressed when being mounted.
To maximise the mistmat area available for demisting, the support rings should have an open structure as shown in standard drawing S 20.030.
If the rings do not have an open structure, a correction has to be incorporated in the gas handling capacity calculations (see Section 3.3.4). The effective vessel diameter in the calculations shall then be taken as the inner ring diameter. Reference is made to the requisition sheet DEP 31.22.05.93-Gen. and standard drawings S 20.028, S 20.029 and S 20.030. (Specimen copies of the standard drawings and requisition sheet are included as Figures VIII.1-4 at the end of this Appendix).

In the case of a HORIZONTAL mistmat (i.e. vertical flow) perforated plates shall **NOT** be used either upstream or downstream of the demister since, during operation, liquid will tend to accumulate downstream of the plates and cause a deterioration in demister performance.

FIGURE VIII.1 S 20.028 - TYPICAL DETAILS OF DEMISTERS

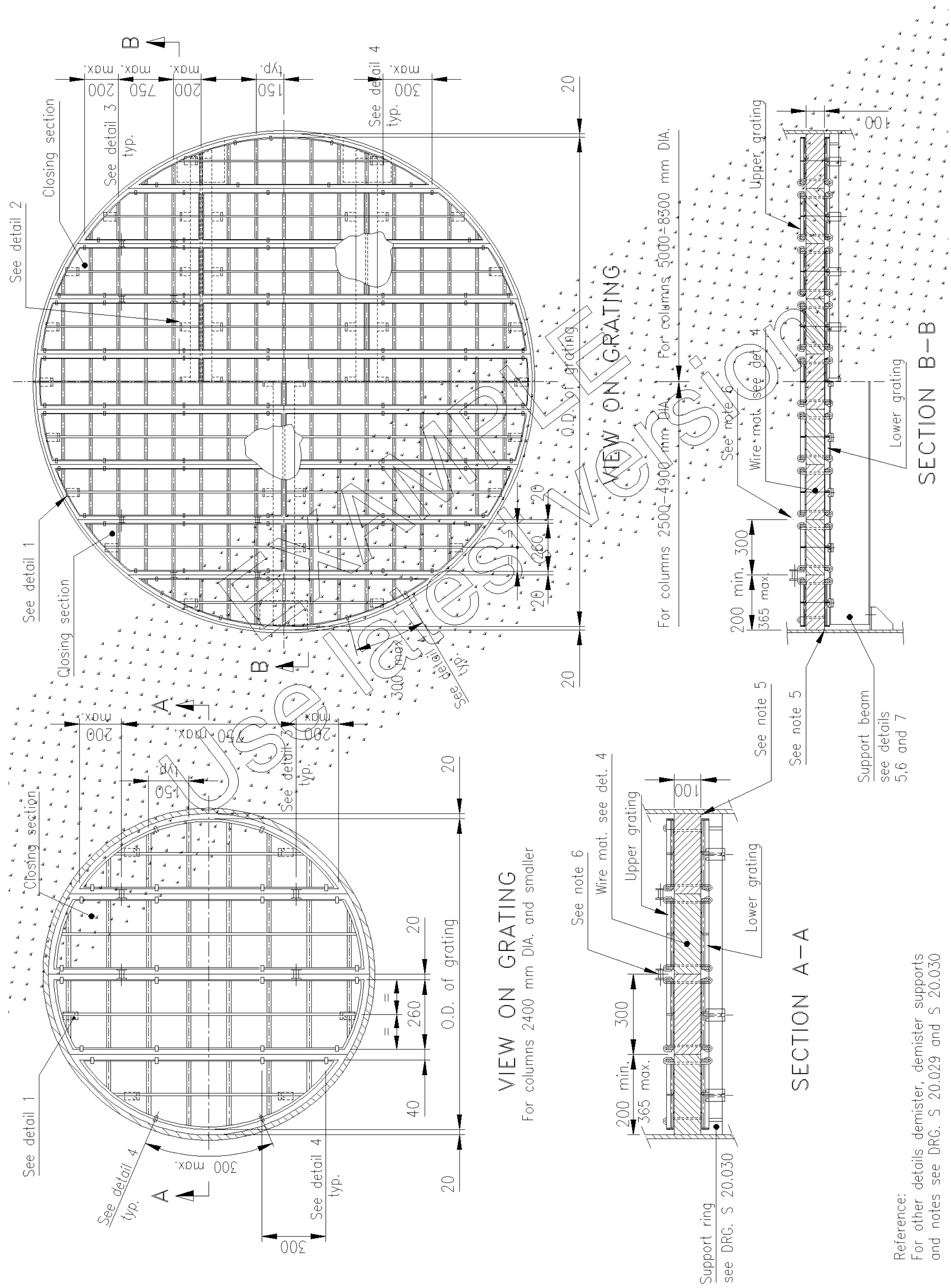
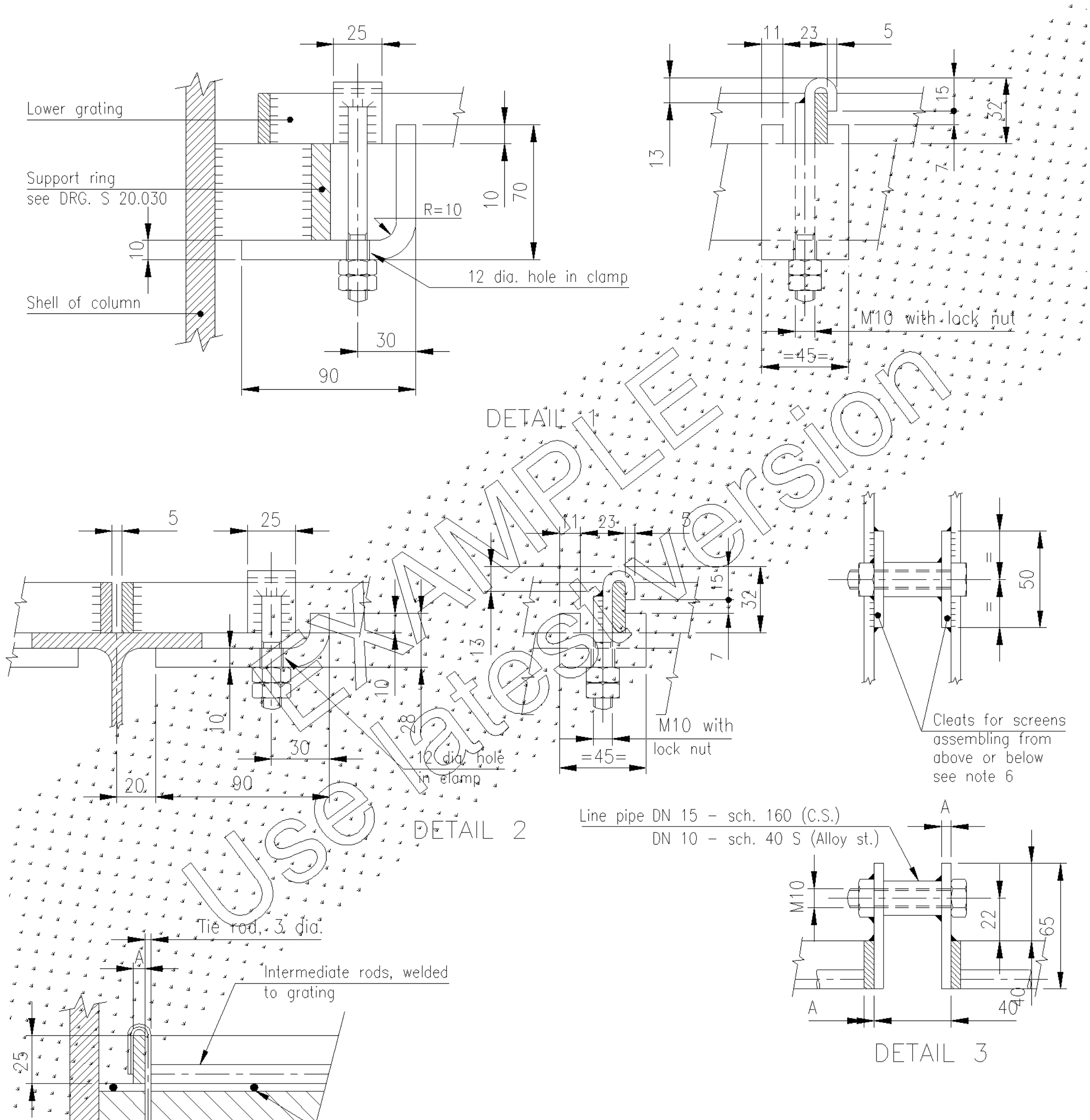
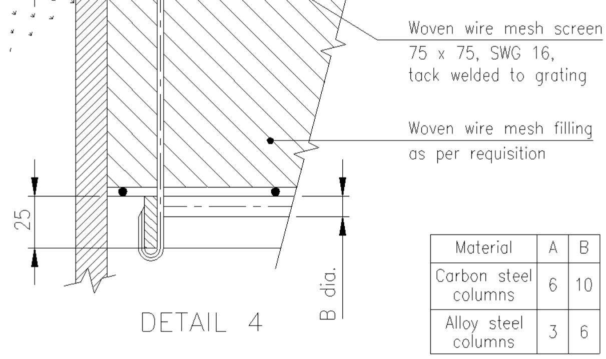


FIGURE VIII.2 S 20.029 - TYPICAL DETAILS OF DEMISTER ATTACHMENTS





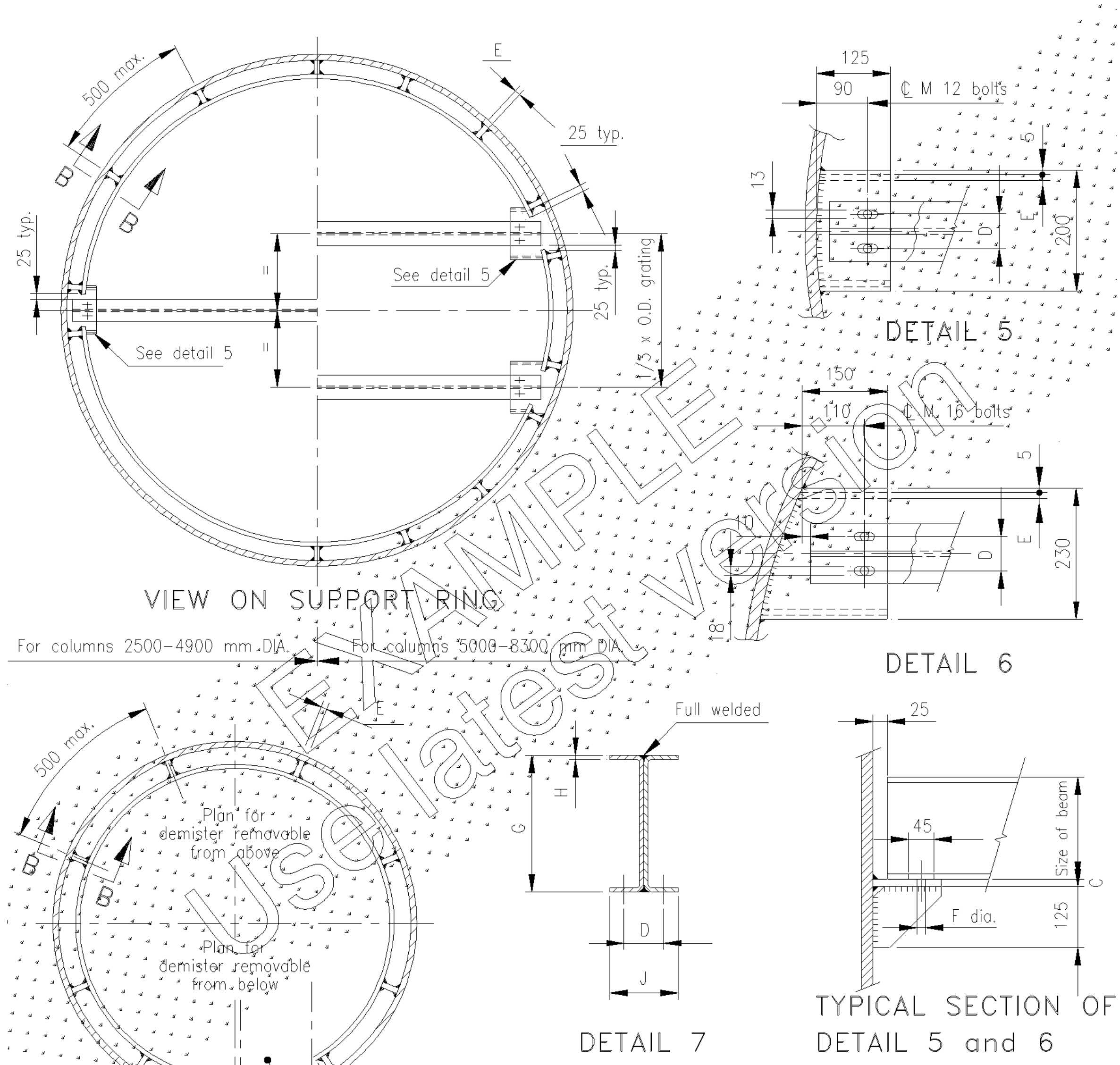
Material	A	B
Carbon steel columns	6	10
Alloy steel columns	3	6

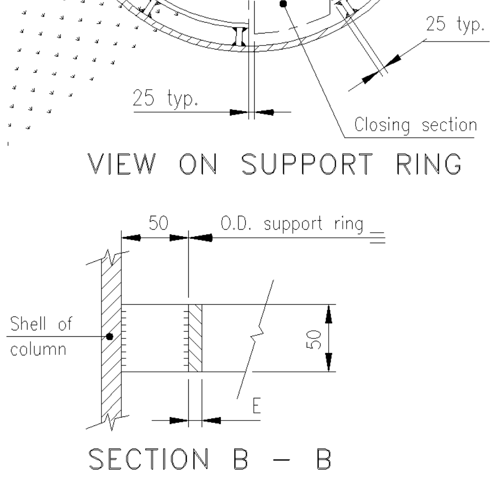
NOTES:
1. Dimensions in millimetres. (Are min. values.)
2. Material as per requisition.
3. All parts of demister to pass through a hole as specified on the requisition.
4. A total corrosion allowance of 6 mm (3 mm each side) has been added on all fixed (welded in vessel) parts. For corrosion allowance other than this figure adjust thickness proportionally.
5. Edges of closing and adjoining sections to be made smooth for easy insert.
6. Cleats required on top of gratings for screens assembling from above and on bottom for screens assembling from below. See detail 3.

Reference:
For other details demisters and demister supports see DRg. S 20.028 and S 20.030

This drawing supersede drawing S 20.013

FIGURE VIII.3 S 20.030 - TYPICAL DETAILS OF DEMISTER SUPPORTSS





Material	Nominal dia. of column	C	D	E	F	G	H	J	Size of beam
	Up to 2400	–	–		–	–	–	–	–
	2500–3700	12	50		14	180	–	91	IPE 180
Carbon steel columns	3800–4900		60	12	14	220	–	110	IPE 220
	5000–6100	16	68		18	240	–	120	IPE 240
	6200–8300		72		18	270	–	135	IPE 270
	Up to 2400	–	–		–	–	–	–	–
	2500–3700	6	50		14	180	6	90	
Alloy steel columns	3800–4900		60	6	14	220	6	110	See detail 7
	500–6100	10	70		18	240	8	120	
	6200–8300		80		18	270		135	

REFERENCE:
For notes and demister details see DRG. S 20.028 and S 20.029

FIGURE VIII.4REQUISITION SHEET DEP 31.22.05.93-GEN. FOR WIREMESH DEMISTER

INFORMATION FOR USE WITHIN THE GROUP ONLY

Requisition/Indent No: Prefix | Project No | Sub-group No | Serial No | Symbol

Sheet No , cont'd on sheet No.

Requisition for **WIRE MESH DEMISTERS**

GENERAL	CONSTRUCTION REQUIREMENTS	TO BE SUBMITTED WITH TENDER
Special requirements : see DEP 31.22.05.11-Gen.	Wire mat[2]	
Typical construction details : see std. drwgs : S 20.028/.029/.030	Wire material	Actual values for
Column/vessel/separator requisition No.[1] :	Mat diameter installed[3] mm	– mat wire thickness : mm
No.[1]	Total mat thickness 100/150 mm	– mat diameter : mm
inside diameter : mm	Wire thickness between 0.23-0.28 mm	– free volume of mat : %
Demister removable from : above/below	Free volume of mat > 97 %	– wire surface area : m^2/m^3
All parts of demister to pass through	Wire surface area > 350 m^2/m^3	– free area of grids : %
manhole/opening with inside diameter : mm	The wire mat shall be fastened in such a way that	
Location of demister in apparatus :	it cannot be compressed when being mounted.	
		Information on apparatus/requisition/drwg
NOTES AND REMARKS	Woven wire mesh screens	Adequate : yes/no*
1) The demister manufacturer shall verify that the information on	Material	
the requisition and/or drwg is adequate and compatible with	Upper/lower mesh : 75 x 75 mm	* Describe additional information required
std. drwgs.: S 20.028/.029/.030	Minimum thickness : 1.6 mm	
2) The mat shall be made of knitted wire, formed to give the correct	Grids (upper/lower grating)	
shape and shall not be cut into shape	Material :	
3) The actual mat diameter not installed shall be equal to the installed	Free area : > 97 %	
diameter + 1% or 20 mm, whichever is the largest.	Support beams	
4) The demister manufacturer shall supply attachment drawings.	Material :	
	To be supplied : yes/no	
	REMARKS ON REVISIONS	

PLANT :
CONSIGNEE:
ENG. BY :
PRINCIPAL

From sort	DCDA	Card	Rev. letter	Item No	Total quantity required	Unit	Abbreviated description	Equipment No.	MESC No.
		B				F C			
		D							
		D							

Accounts code
Made by | Checked by | Reference indicator | Initial | Date
Purchase action by: Central Purchasing/Local company
Authorised for purchase action
Recommended supplier.
ROS date
Quantity change

DEP 31.22.05.93 - Gen. 09/86

APPENDIX IX VANE PACKS

GENERAL INFORMATION

The vanes in vane packs can be either of the no-pocket (straight), single-pocket or double-pocket type (Figure IX.1).
Of the three vane types the straight vane has the lowest sensitivity to fouling, but has also the lowest liquid separation efficiency.
The double-pocket vane type has the highest sensitivity to fouling, but its efficiency above λ_{max} (where liquid re-entrainment starts) deteriorates to a lesser degree than with the other types because the separated liquid is better shielded from the gas flow.

In horizontally flowed-through vane packs the single- or double-pocket types are normally used. Horizontally flowed-through vane packs are typically used in vertical vane-type demister vessels (see 3.5).

Vertically flowed-through vane packs are equipped with either straight or double-pocket vanes. In the SVS (see 3.7) straight vanes are used because of their insensitivity to fouling. In this application the low efficiency of the vane pack is no problem since the vane pack acts as a coalescer medium.
In general the use of a vertically flowed-through vane pack with double-pocket vanes is only recommended if the service is clean. Since it has a higher gas handling capacity than mistmats and requires little height (at most about 0.20m) it is a suitable retrofitting device to upgrade undersized wiremesh demisters by installing it downstream of (i.e. above) the mistmat. If further advice is required, the Principal should be contacted.

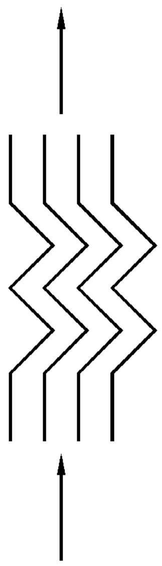
MANUFACTURING

Vane packs are usually **stainless steel.**
The Manufacturer of the vane pack shall supply to the Principal fully dimensioned and detailed drawings of the vane pack for the particular application. Such information shall be treated by the Principal as confidential.

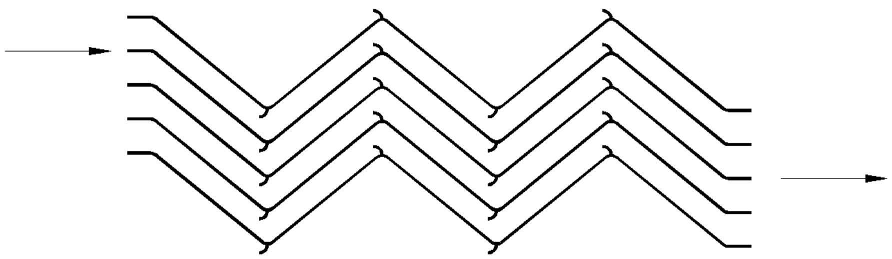
MOUNTING

The vessel shall provide sufficient access for inspection, cleaning, maintenance and the removal and installation of the vane pack and other internals. The internals should be bolted on attachments welded to the vessel wall. The supplier of the vane pack shall provide detailed attachment drawings to the vessel Manufacturer. In the case of a vertical vessel, the vessel can be provided with a full top flange or with manholes. A full top flange allows for the installation of internals as a prefabricated box; this is preferred for vessel diameters below 1.2 m.
Alternatively two manholes should be provided; one upstream and one downstream of the vane pack. The least preferable option is to provide only one manhole; in such a case it shall be located upstream of the vane pack.
A minimum vessel inner diameter of 0.6 m is recommended for vertical vessels and 1.5 m for horizontal vessels.

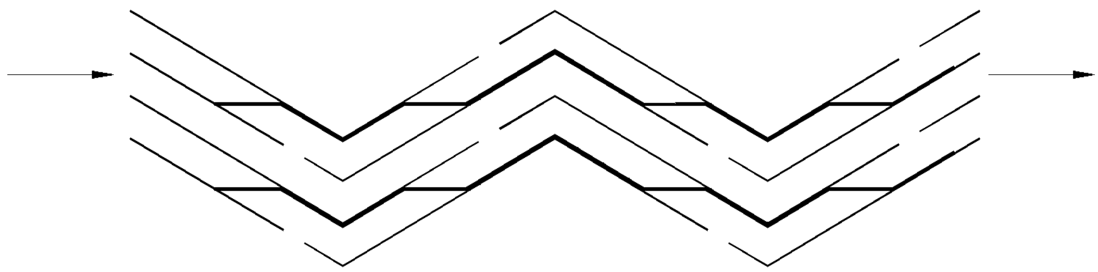
FIGURE IX.1 EXAMPLES OF VARIOUS TYPES OF VANES



(a) straight (no-pocket) vane; side view
(to be used in SVS)



(b) single-pocket vane; top view
(to be used in horizontal flow)



(c) double-pocket vane; top view
(to be used in horizontal flow)

APPENDIX X VESSEL SIZING EXAMPLES

Example 1 - **HORIZONTAL KNOCK-OUT DRUM**

Suppose a separator (with semi-elliptical heads: ratio 2:1) has to be designed to accommodate slugs.

- The liquid in the feed has a tendency to foaming and is waxy.
- High liquid removal efficiency (mist) is not required.
- The separator is to be located onshore in flat country.

Process data:

$Q_G = 0.3$ m^3/s

$Q_L = 0.03$ m^3/s

$\rho_G = 70$ kg/m^3

$\rho_L = 600$ kg/m^3

$\eta_L = 0.001$ Pa.s

- Max. slug size is 10 m^3.
- Required liquid residence time (for control purposes) is 180 seconds.
- Minimum flow time between the pre-alarm and the trip levels is 60 seconds.

Design

First, the type of separator has to be selected (2.5).

Since high liquid removal efficiency is not required and the service is fouling, a knock-out vessel is indicated.

Since large slugs have to be accommodated, a horizontal knockout drum is selected. The separator is to be located in flat country. Therefore the design margin to be applied (Appendix IV) is 1.2.

$$Q^*_{max} = 1.2 * 0.3 * \sqrt{70/(600-70)} = 0.131 \quad m^3/s$$

$$\phi_{feed} = 0.03/0.3 * \sqrt{600/70} = 0.293$$

The minimum vessel cross-sectional area for gas flow (above LZA (HH)) follows from

λ_{max} = 0.07 m/s (See 3.2.2)

$A_{G.min}$ = 0.131/0.07 = 1.87 m^2

Nozzle sizing:

(Do NOT include a design margin)

Feed inlet:

The inlet is preferably located at the top of the vessel.

If a Schoepentoeter is used as inlet device:

According to Appendix II, $\rho_m v^2_{m,in} \leq 6\,000$ Pa, therefore $d_1 \geq 0.24$ m
Take d_1 as 10".

If a half-open pipe is used as inlet device:

According to Appendix II, $\rho_m v^2_{m,in} \leq 1\,500$ Pa, therefore $d_1 \geq 0.34$ m
Take d_1 as 14".

If no inlet device is used:

According to Appendix II, $\rho_m v^2_{m,in} \leq 1\,000$ Pa, therefore $d_1 \geq 0.38$ m

Take d_1 as 16".

Gas outlet:

According to Appendix II, $\rho_{G,out} v^2_{G,out} \leq 3750$ Pa, therefore $d_2 \geq 0.23$ m

Take d_2 as 10".

Liquid outlet:

According to Appendix II, $v_{L,out} \leq 1$ m/s, therefore $d_3 \geq 0.20$ m.

Take d_3 as 8" and install a vortex breaker in accordance with Standard Drawing S 10.010.

Vessel sizing:

The procedure as outlined in Appendix VII is followed:

i) Because the liquid has a foaming tendency, the high level trip, LZA(HH), shall be at least at 0.95 m from the vessel bottom.

ii) Table VII.2 gives for $A_{g,min}$=1.87 m^2 and LZA(HH)= 0.95 m:

$D \approx 2.07$ m

iii) The relationships of Appendix VI give then with

LA(L)=0.15 + 0.10 = 0.25 and LA(H) = 0.95 - 0.35 = 0.60 m:

$A_{H-L} \approx 0.58\ m^2$

$V_{hd,H\_L} \approx 0.19\ m^3$

iv) $L \approx (10 + 1.2*0.03*180 - 2*0.19)/0.58 \approx 27.8$ m

$L/D \approx 27.8/2.07 \approx 13.4 > 6$; go to step v.

v) Take as next step D = 2.07 + 0.43 = 2.5 m.

From Table VII.2 it is derived via interpolation that if A_G is kept constant then LZA(HH) $\approx$ 1.48 m. Return to step iii.

iii) LA(H) = 1.48 -0.35 = 1.13 m; LA(L) remains 0.25 m.

The relationships of Appendix VI give then:

$A_{H\_L} \approx 1.90\ m^2$

$V_{hd,H\_L} \approx 0.82\ m^3$

iv) $L \approx (10 + 1.2*0.03*180 - 2*0.82)/1.90 \approx 7.81$ m

$L/D \approx 7.81/2.5 \approx 3.1$

$2.5 \leq L/D \leq 6$; go to step vi

vi) $t_{HH\_H} \approx (7.81*0.25 + 2*0.12)/0.036 \approx 61 > 60$ sec OK

$t_{L\_LL} \approx (7.81*0.14 + 2*0.04)/0.036 \approx 33 < 60$ sec; increase width of control band to $0.1*60/33 \approx 0.18$ m. Return to step iii.

iii) LA(L)=0.15 + 0.18 = 0.33 m. LA(H) remains 1.13 m

The relationships of Appendix VI give then :

$A_{H\_L} \approx 1.77\ m^2$

$V_{hd,H\_L} \approx 0.78\ m^3$

iv) $L \approx (10 + 1.2*0.03*180 - 2*0.78)/1.77 \approx 8.43$ m

$L/D \approx 8.43/2.5 \approx 3.4$

$2.5 \leq L/D \leq 6$; go to step vi

vi) $t_{HH\_H} \approx (8.43*0.25 + 2*0.12)/0.036 \approx 65 > 60$ sec OK

$t_{L\_LL} \approx (8.43*0.25 + 2*0.13)/0.036 \approx 66 > 60$ sec OK

go to step vii

vii) Because L = 8.43 m:

$D \geq 0.85$ m, for sufficient de-foaming.

$D \geq 0.36$ m, for sufficient de-gassing.

Since D = 2.5 m, both criteria are met.

The gas cap above LZA(HH) has a central height of 2.5 -1.48 = 1.02 m. Preferably a top feed inlet is used. The space required for this is approximately $2d_1 + 0.15$ m ≈ 0.86 m (assuming that the required elbow has a centre line radius of $1.5d_1$).
The maximum allowable $d_1 = 0.35$ m. Therefore either a Schoepentoeter or a half-open pipe can be installed. A half-open pipe has been selected.

In summary, the dimensions of the horizontal knock-out drum are as follows:

L (T/T)	=	8.43	m
D	=	2.50	m
d_1	=	14	inches (top inlet fitted with half-open pipe)
d_2	=	10	inches
d_3	=	8	inches

Pressure drop

(under maximum flow conditions; now include the design margin):

inlet nozzle	940	Pa
gas outlet nozzle	777	Pa
	_____+	
	1717	Pa

Example 2 - **VERTICAL VANE-TYPE DEMISTER**

Suppose a demisting vessel is required in a slightly fouling service.

- The liquid to be separated is slightly waxy and has a tendency to foam.
- Gas carry-under should be avoided.
- No liquid slugs are expected.
- Design margin is 25%.

Process data:

$Q_G = 0.83$ m^3/s

$Q_L = 0.0057$ m^3/s

$\rho_G = 50$ kg/m^3

$\rho_L = 700$ kg/m^3

$\eta_L = 0.002$ Pa.s

$\sigma = 0.01$ N/m

- Required liquid residence time (for control purposes) is 180 seconds.
- Minimum flow time between the pre-alarm and the trip levels is 60 seconds.
- A vertical vane-type demister is a suitable candidate.
- Horizontal-flow single-pocket vanes will be used.

Vane box depth is 0.31 m.

Perforated plates (20% NFA) will be fitted (a front and back plate)

Design (see (3.5))

First it has to be decided whether an in-line or a two-stage separator has to be used.

The criterion for this is the flow parameter of the feed.

$\phi_{feed} = 0.0257 > 0.01$, so a two-stage separator has to be used.

$Q^*_{max} = 0.288$ m^3/s

Vessel diameter:

In (3.5.6.1), a number of criteria are given for $D_{min.}$

Calculations give the following results:

sizing criterion	:	D_{min} (m)
primary separation	:	1.45
de-gassing	:	1.13
de-foaming	:	1.36
accessibility	:	0.6
even distribution	:	1.32

Consequently the vessel diameter is determined by the primary separation criterion:

D = 1.45 m

Vane pack

First the minimum vane area has to be determined:
Archimedes number is 2192 which is larger than 225.
Therefore the "low-viscosity" equation ("possibility 1") for $\lambda_{v,max}$ is selected, bearing in mind that ϕ_v is reduced to 0.01 because the vessel diameter has been determined with the primary separation criterion.

NOTE In the case that because of other criteria, e.g. de-gassing, a larger diameter would have been required than was needed for a proper primary separation only, strictly speaking $\phi_V < 0.01$. However, also in those cases it is prudent to assume $\phi_V = 0.01$).

$\lambda_{v,max} = 0.18$ m/s therefore $A_v = 1.59$ m^2

Further calculations give:

$w_v = 1.11$ m

$h_v = A_v/W_v = 1.43$ m which is within the limits of 0.3 and 1.5 m.

(If, in the first calculation step, $h_v > 1.5$ m, D would have to be increased until these limits were met.)

w_{vb} = 1.21 m

h_{vb} = 1.73 m

Nozzle sizing

(Do NOT include a design margin)

Inlet nozzle:

It is fitted with a Schoepentoeter and from the inlet momentum criterion follows that at least a 12" ID nozzle shall be used (strictly speaking this is slightly too small).

Gas outlet nozzle:

Its size is determined by the criterion

$$d_2 \geq 0.43\sqrt{A_V}$$

therefore d_2 > 0.54 m. Select a 22" ID nozzle and also install a deflector.

Liquid outlet nozzle:

d_3 > 0.085 m.

Select a 4" ID nozzle and install a vortex breaker.

Height (between bottom and top tangent line) (see Figure 3.6.c)

distance	BTL - LZA(LL)	0.15	m
	LZA(LL) - LA(L)	0.26	m
	LA(L) - LA (H)	0.78	m
	LA(H) - LZA(HH) (incl. foaming allowance of 0.25 m)	0.51	m
	X_1	0.15	m
	d_1 + 0.02	0.32	m
	X_2	0.72	m
	h_{vb}	1.73	m
	X_3	0.10	m
		____	+
		4.72	m

Pressure drop

(under max. flow conditions: now include the design margin)

inlet nozzle	5578	Pa **
gas outlet nozzle	197	Pa
vanes	318	Pa
2 perforated plates (20% NFA)	679	Pa
	____	+
	6772	Pa

** On the high side; use of Schoepentoeter will give some pressure recovery.

Example 3 - **SVS SEPARATOR**

In vessel sizing example 2 a vertical vane-type separator was selected.

An SVS separator is also possible.

In this example a SVS separator will be sized for the same process conditions. Subsequently its dimensions and pressure drop will be compared with those of the vane-type separator.

Design (see (5.7))

$Q^*_{max} = 0.288\ m^3/s$

$\phi_{feed} = 0.0257$

Vessel diameter

Gas handling capacity criterion:

First guess using the formula in 3.7.8 gives $D_{min} = 1.21$ m.

Since this is smaller than 1.5 m, Table 2 has to be consulted.

This leads to smallest standard diameter of 1.30 m.

Other sizing criteria	:	D_{min} (m)
de-gassing	:	1.13
de-foaming	:	1.36

It is concluded that the vessel diameter is determined by the de-foaming criterion.

D = 1.36 m and so $\lambda = 0.20$ m/s

The Schoepentoeter liquid removal efficiency at this λ is 48% and the swirl deck will be loaded with a liquid flow rate of maximally $(1-0.48)*1.25*0.0057 = 0.0037\ m^3/s$.

According to table 2 the swirldeck can hold at least 52 swirl tubes.

This results in a maximum liquid load per swirl tube of 0.71E-4 < 1.5E-4 m^3/s

Therefore a D of 1.36 m satisfies also the liquid handling capacity criterion.

Swirl deck

Because of the requirement $\lambda_{st} \leq 0.67$ m/s, at least 45 swirltubes are needed. In this example more swirltubes could be accommodated. In general, it is recommended to take as criterion for the maximum number of swirltubes: $\lambda_{st} \geq 0.5$ m/s. In the case of this example it is decided to install 52 swirltubes (pitch 0.14 m). $\lambda_{st} = 0.58$ m/s which is 15% lower than the maximum limit.

The swirldeck will be composed of boxes with four swirltubes each.

A top-flanged vessel is not required.

The manway shall have an inner diameter of at least 20".

Nozzle sizing

(Do NOT include a design margin)

Inlet nozzle:

It is fitted with a Schoepentoeter.

As in Example 2, select a 12" nozzle.

Gas outlet nozzle:

$\rho_G v^2_G \leq 3750$, therefore $d_1 > 0.35$ m.

Select a 14" ID nozzle (much smaller than in case 2).

Liquid outlet nozzle:

d_3 > 0.085 m.

As in Example 2, select a 4" ID nozzle + vortex breaker.

Height (between bottom and top tangent line) (see Figure 3.9a)

distance	BTL - LZA(LL)	0.15	m
	LZA(LL) - LA(L)	0.29	m
	LA(L) - LA(H)	0.88	m
	LA(H) - LZA(HH) (incl. foaming allowance of 0.25 m)	0.54	m
	X_1	0.15	m
	X_2	0.32	m
	X_3	0.30	m
	X_4	0.20	m
	X_5	0.50	m
	bottom-top swirldeck	0.51	m
	X_6	0.20	m
		____	+
		4.04	m

Pressure drop

(under max. flow conditions: now include the design margin)

inlet nozzle	5578	Pa
gas outlet nozzle	1200	Pa
vane pack	104	Pa
swirldeck	1837	Pa
	____	+
	8719	Pa

Check on drainpipe length

Under maximum flow conditions, the liquid level in the swirldeck drainpipes will rise (0.5*1837 + 104)/(9.81*(700-50)) = 0.16 m above the liquid level in the liquid compartment of the separator.
The distance between LZA(HH) and the bottom of the swirldeck is 1.47 m, which is much larger than the recommended minimum distance (1.5*0.16=0.24 m) and is therefore more than adequate to accommodate the liquid rise in the drainpipe.

Comparison of vertical vane-type and SVS separator

	vertical vane-type separator	SVS separator
vessel diameter, m	1.45	1.36
height (T/T), m	4.72	4.04
inlet nozzle diameter, inches	12	12
gas outlet nozzle diameter, inches	22	14
liquid outlet nozzle, inches	4	4
pressure difference, mbar	68	87

It can be seen that the SVS separator is smaller than the vertical vane-type separator for the same duty. It also has a higher liquid removal efficiency (typically 99 versus 96%).

However, the pressure drop across the SVS separator is higher and more expensive internals are required.

APPENDIX XI DEBOTTLENECKING TIPS

If a separator has a poor performance (e.g. due to a change in flow conditions or due to a poor design) the separator shall either be replaced or upgraded/debottlenecked.

Included below are some debottlenecking tips which both guide and serve as a reminder for the designer. This Appendix is not exhaustive. Further information is available from the Principal.

Feed inlet

Sometimes a separator of the non-cyclone type has a poor performance because of a horizontal bend in the feed piping immediately upstream of the vessel. This will cause gas maldistribution over the separation internal, resulting in local overloading.

This maldistribution can be reduced by installing a static mixer between the bend and the inlet nozzle or by fitting guiding vanes in the inlet section of the vessel.

In the case of a foaming feed the existing feed inlet should be retrofitted with a cyclone type inlet device. Presently, Manufacturer-proprietary inlet devices consisting of an array of vortex tubes or cyclones are available. Further information can be obtained via the Principal.

De-gassing

If an existing vessel cannot meet the de-gassing requirement (Appendix VI), the installation of a plate pack to remove the gas bubbles should be considered.

Perforated plates

Often the performance of a separation internal can be improved by installing a perforated plate in front of it to eliminate flow maldistribution. However, the following points should be noted:

1. DO NOT use perforated plates in vertical flow because of the danger of accumulation of liquid downstream of the perforated plates.
2. DO specify a sufficiently high NFA, mainly because of the following reasons:
    a. The associated high pressure drop will tend to shatter the liquid droplets in the gas stream and the resulting smaller droplets are more difficult to separate from the gas.
    b. If perforated plates with a too low NFA are fitted in a horizontal demister with a baffle arrangement (see the vane-type demister described in (3.6.4)), this insufficient NFA will cause a large difference in the liquid level upstream and downstream of the separation internal, which will result in either bypassing of the gas or flooding of the back of the internal (depending on the location of the liquid level control in the vessel, i.e. downstream or upstream of the perforated plate).
    The recommended NFA range is 20%.
3. Do not leave less than 5 cm between the perforated plates and the separation internal.

Installation of an additional internal or replacing of an existing internal

Often a separator can be debottlenecked by installing an additional internal downstream of the existing one. A few examples are given.

Vertical wiremesh demister

For this type of separator there are a few options:

1. Convert it into a SMS by the installation of a (Shell-proprietary) swirldeck.
   Use of a swirltube pitch of 0.13 m rather than the standard pitch of 0.14 m could be considered in order to maximise the number of swirltubes.
2. Convert it into a "SMS" by the installation of a (Manufacturer-proprietary) axial-flow multicyclone bundle.
3. Install a vertical flow double-pocket vane pack downstream of the mistmat.
   (See also Appendix IX for the double-pocket vane pack geometry)

The first two options are recommended, but if insufficient space is available for the swirldeck/multicyclone bundle or the resulting relatively high pressure drop of the deck/bundle is not acceptable, use option 3.

In the case of a top gas outlet, the distance between the gas outlet and the additional internal should be at least 0.25 times the vessel diameter or 0.25 m, whichever is the larger, in order to avoid gas maldistribution (vacuum cleaner effect). In standard vessels this normally coincides with the top tangent line. If a gas side outlet is applied, the Principal should be consulted.

Horizontal wiremesh or vane-type demister

In case of an undersized horiziontal wiremesh or vane-type demister the separation internal could be replaced with a horizontal axial-flow multicyclone bundle.
However, sufficient space shall be available between the bottom of the multicyclone bundle and the maximum liquid level for proper liquid drainage of the deck.

Maximising the vane area in a horizontal vane-type demister

An alternative to retrofitting a horizontal vanepack demister with an axial-flow multicyclone bundle is to maximise the vane pack area.

Normally the vane pack in a horizontal vane-type demister has its face perpendicular (90 degrees) to the central axis of the vessel. The gas handling capacity of this demister can be increased by replacing the vane pack with a diagonal one, thus creating a larger vane face area.
The angle between the vane face and the vessel axis can be reduced down to as little as 10 degrees. To minimize maldistribution over the vane area this diagonal vane pack shall be equipped with one or more perforated plates.