MANUAL

GAS/LIQUID SEPARATORS - TYPE SELECTION AND DESIGN RULES

DEP 31.22.05.11-Gen.

December 2007
(DEP Circulars 03/08 and 14/08 have been incorporated)

DESIGN AND ENGINEERING PRACTICE

PREFACE

DEPs (Design and Engineering Practice) publications reflect the views, at the time of publication, of:

Shell Global Solutions International B.V. (Shell GSI)

and/or

Shell International Exploration and Production B.V. (SIEP)

and/or

other Shell Service Companies.

They are based on the experience acquired during their involvement with the design, construction, operation and maintenance of processing units and facilities, and they are supplemented with the experience of Shell Operating Units. Where appropriate they are based on, or reference is made to, international, regional, national and industry standards.

The objective is to set the recommended standard for good design and engineering practice applied by Shell companies operating an oil refinery, gas handling installation, chemical plant, oil and gas production facility, or any other such facility, and thereby to achieve maximum technical and economic benefit from standardization.

The information set forth in these publications is provided to Shell companies for their consideration and decision to implement. This is of particular importance where DEPs may not cover every requirement or diversity of condition at each locality. The system of DEPs is expected to be sufficiently flexible to allow individual Operating Units to adapt the information set forth in DEPs to their own environment and requirements.

When Contractors or Manufacturers/Suppliers use DEPs they shall be solely responsible for the quality of work and the attainment of the required design and engineering standards. In particular, for those requirements not specifically covered, the Principal will expect them to follow those design and engineering practices which will achieve the same level of integrity as reflected in the DEPs. If in doubt, the Contractor or Manufacturer/Supplier shall, without detracting from his own responsibility, consult the Principal or its technical advisor.

The right to use DEPs is granted by Shell GSI, in most cases under Service Agreements primarily with Shell companies and other companies receiving technical advice and services from Shell GSI or another Shell Service Company. Consequently, three categories of users of DEPs can be distinguished:

1) Operating Units having a Service Agreement with Shell GSI or other Shell Service Company. The use of DEPs by these Operating Units is subject in all respects to the terms and conditions of the relevant Service Agreement.

2) Other parties who are authorized to use DEPs subject to appropriate contractual arrangements (whether as part of a Service Agreement or otherwise).

3) Contractors/subcontractors and Manufacturers/Suppliers under a contract with users referred to under 1) or 2) which requires that tenders for projects, materials supplied or - generally - work performed on behalf of the said users comply with the relevant standards.

Subject to any particular terms and conditions as may be set forth in specific agreements with users, Shell GSI disclaims any liability of whatsoever nature for any damage (including injury or death) suffered by any company or person whomsoever as a result of or in connection with the use, application or implementation of any DEP, combination of DEPs or any part thereof, even if it is wholly or partly caused by negligence on the part of Shell GSI or other Shell Service Company. The benefit of this disclaimer shall inure in all respects to Shell GSI and/or any Shell Service Company, or companies affiliated to these companies, that may issue DEPs or require the use of DEPs.

Without prejudice to any specific terms in respect of confidentiality under relevant contractual arrangements, DEPs shall not, without the prior written consent of Shell GSI, be disclosed by users to any company or person whomsoever and the DEPs shall be used exclusively for the purpose for which they have been provided to the user. They shall be returned after use, including any copies which shall only be made by users with the express prior written consent of Shell GSI. The copyright of DEPs vests in Shell GSI. Users shall arrange for DEPs to be held in safe custody and Shell GSI may at any time require information satisfactory to them in order to ascertain how users implement this requirement.

All administrative queries should be directed to the DEP Administrator in Shell GSI.

TABLE OF CONTENTS

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 the 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)
- Filter 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 consult 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.

This is a revision of the DEP of the same number dated September 2002; a summary of the main changes is given in (1.6).

1.2 DISTRIBUTION, INTENDED USE AND REGULATORY CONSIDERATIONS

Unless otherwise authorised by Shell GSI, the distribution of this DEP is confined to Shell companies and, where necessary, to Contractors and Manufacturers/Suppliers nominated by them.

Attention is drawn to the copyright and confidentiality statements on the title page and preface of this DEP and to the fact that some of the referenced documents cover Shell-proprietary items.

This DEP is intended for use in oil refineries, chemical plants, gas plants, exploration and production facilities and supply and marketing installations.

When DEPs are applied, a Management of Change (MOC) process should be implemented; this is of particular importance when existing facilities are to be modified.

If national and/or local regulations exist in which some of the requirements may be more stringent than in this DEP, the Contractor shall determine by careful scrutiny which of the requirements are the more stringent and which combination of requirements will be acceptable with regard to the safety, environmental, economic and legal aspects. In all cases the Contractor shall inform the Principal of any deviation from the requirements of this DEP which is considered to be necessary in order to comply with national and/or local regulations. The Principal may then negotiate with the Authorities concerned, the objective being to obtain agreement to follow this DEP as closely as possible.

1.3 DEFINITIONS

The **Contractor** is the party which carries out all or part of the design, engineering, procurement, construction, commissioning or management of a project or operation of a facility. The Principal may undertake all or part of the duties of the Contractor.

The **Manufacturer/Supplier** is the party which manufactures or supplies equipment and services to perform the duties specified by the Contractor.

The **Principal** is the party which initiates the project and ultimately pays for its design and construction. The Principal will generally specify the technical requirements. The Principal may also include an agent or consultant authorised to act for, and on behalf of, the Principal.

The word **shall** indicates a requirement.

The word **should** indicates a recommendation.

1.4 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)	°
α	ratio of the short and long axes of the vessel head	-
β	edge angle (Schoepentoeter)	°
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})

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

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 SUMMARY OF CHANGES SINCE PREVIOUS EDITION

The previous edition of this DEP was dated September 2002. Other than editorial changes, the following are the main changes since that edition:

Section	Change
2.3.2	Minor changes
2.4.1	Minor changes
2.5	Table I. Efficiency Filter Separator corrected
3	Amended statement about high pressure applications
3.1	Requirements for degassing and defoaming moved to App. VII. Specifications for Flare Knockout Drums removed.
3.2	Requirements for degassing and defoaming moved to App. VII. Specifications for Flare Knockout Drums removed. Deflector plate specification (thickness) updated.
3.3	Wire mesh specifications added. Requirements for degassing and defoaming moved to App. VII.
3.4	Wire mesh specifications added. Requirements for degassing and defoaming moved to App. VII.
3.5	Two stage separator with horizontal flow vanepack removed. Figure 3.6 adapted accordingly.
3.5.1	Revised recommendations for use as compressor suction drum.
3.5.7	Moved to 3.5.6. Included different gas handling capacity for vertical flowed through vane packs.
3.6.1	Correction of maximum allowable operating pressure to 70 bar.
3.6.2	Removed suggestion of diagonal vanepacks.
3.6.3	Updated equation for $A_{G,min}$. Minimum gas cap height requirement removed.
3.7	Moved to 3.10
3.8	Moved to 3.7
3.9	Moved to 3.8
3.10	Moved to 3.9
3.10	Replaced by 3.7. Heading changed into ‘Vertical Separators with Axial Flow Multicyclones Text substantially changed. SMMSM separator added. Table 2 several changes to the number of swirltubes. Obsolete specifications of drain pipes removed. Included new section about Non-Shell Vertical Axial flow Separators. Figures

	11 and 12: heights of swirldeck and primary demister adapted.
3.11	Replaced by new chapter on Horizontal Separators with Axial Flow Multicyclones.
3.12	Removed.
3.13	Moved to 3.12
4	Figure 4.1 added.
Appendix VII	Degassing and defoaming criteria extended.
Appendix VIII	Original chapter about wire mesh specifications removed. Replaced by new chapter about design implications of high operating pressures.
Appendix X	Design examples removed. This section has become obsolete with the advent of PC based vessel design tools, and the cases were not well chosen. Replaced by a new statement about dealing with separators subject to motion.

1.7 COMMENTS ON THIS DEP

Comments on this DEP may be sent to the DEP Administrator at standards@shell.com. Shell staff may also post comments on this DEP on the Surface Global Network (SGN) under the Standards folder.

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;
- head room is restricted;
- 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.

Amended per Circular 14/08

Commonly used inlet devices are a half-open pipe or a Schoepentoeter. Rules for the design of Schoepentoeters are given in DEP 31.20.20.31-Gen..

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 further internals are required in the vessel to complete the

separation process. Only when not more than bulk separation is required and the droplets are relatively large the separation internal can 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.

In flare knock-out drums internals other than a half-open pipe - or schoepentoeter inlet and an outlet deflector plate are not allowed, to prevent blockage of the gas outlet in case of mechanical failure or fouling.

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}:

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

where $Q_{G,max}$ 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:

$$A_{G,min} = Q^*_{max} / \lambda_{max} \qquad [m^2]$$

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

The gas load factor, also referred to as k-factor or Souders-Brown velocity, is defined as

$$\lambda = (Q_G / A_G)\sqrt{\rho_G / (\rho_L - \rho_G)}$$

This is a superficial gas velocity modified with a gas density scaling factor which accounts to a large extent for the effect of operating pressure.

In a vertical vessel $A_{G,min}$ is the cross-sectional area of the vessel. If 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 relative importance of the liquid load approaching the separator internal.

$$\phi = (Q_L / Q_G)\sqrt{\rho_L / \rho_G} \qquad [-]$$

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

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.

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

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.

2.6 SEPARATORS SUBJECT TO MOTION

More and more separators are installed on floaters and other installations subject to motion, e.g. Tension Leg Platforms. The performance of these separators can be adversely affected by the motion imposed by waves and wind. The accompanying sloshing will compromise liquid handling capacity, separation efficiency and the functioning of level instrumentation. Mitigation of these effects will require the installation of additional internals and other measures.

Guidelines for the design of separators which are subject to motion are being developed and will be listed in Appendix X in future. For the present, the Principal shall be consulted for the further advice.

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

	VKO	HKO	VW	HW	VV1	VV2	HV	SMS	SVS	SMSM	CT	CS	VRMC	FS
Gas handling														
max. capacity (λ)	B	B	C	C	D	D	D	E	E	E	E	E	E	C
turndown (max/min flow)	∞	∞	2	2	3	3	3	10	4	10	2	3	2	∞
Liquid removal efficiency														
overall, %	90	90	> 98	> 98	> 96	> 96	> 96	> 98	> 96	> 99	> 96	> 99	> 93	50-80
with respect to fine mist	A	A	E	E	C	C	C	E	D	F	B	D	B	E
flooding above λ_{max} (Y/N)	N	N	Y	Y	*	*	*	N	N	N	N	N	Y	Y
Liquid handling capacity														
as slugs	D	E	D	E	A	D	E	D	D	D	D	D	B	-
as droplets ($Q_{L,max}$)	D	D	D	D	B	C	C	D	D	D	D	D	B	B
Fouling tolerance														
sand	E	E	B	B	**	**	**	B	C	B	E	C	C	B
sticky material	E	E	A	A	**	**	**	A	C	A	E	C	C	A
Pressure drop	A	A	B	B	B	B	B	C	C	C	D	D	D	***

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

∞ : Infinite
* : if double-pocket vanepack: N; for a single-pocket vanepack or in case of straight vanes: Y
** : if double-pocket vanepack: A; if single-pocket vanepack: B; if straight vanes: C
*** : depending on the degree of fouling, ranging from C to E

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 vane pack (3.5.5)
VV2 Vertical two-stage separator with 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)
FS Filter separator (3.13)

3. DESIGN RULES

In this Section the design rules for the various separators are given.

Unless explicitly stated otherwise, the design is to be based on maximum gas and liquid flow rates and slug volume taking into account a design margin or surge factor as specified 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:

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

$$\text{or} \qquad D \geq 4.26 \sqrt{Q^*_{max}} \qquad [m]$$

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 the pressure is well above 90 bara, or if otherwise the surface tension is as low as 5×10^{-3} N/m or below, see also AppendixVIII.

The vessel should also be large enough to enable proper disengagement of gas from the bulk liquid. The relevant criteria are discussed in Appendix VII.

In the case of **flare knock-out drums** higher λ-values are acceptable. For more information on flare knock-out design DEP 80.45.10.10-Gen should be consulted.

3.1.3 Height

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

$$H = h + X_1 + X_2 + X_3 \qquad [m]$$

where

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

– 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 [m];
- 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 $D > 1.5$ 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.

For flare knock-out drums, smaller nozzles are acceptable. For flare knock-out design DEP 80.45.10.10-Gen should be consulted.

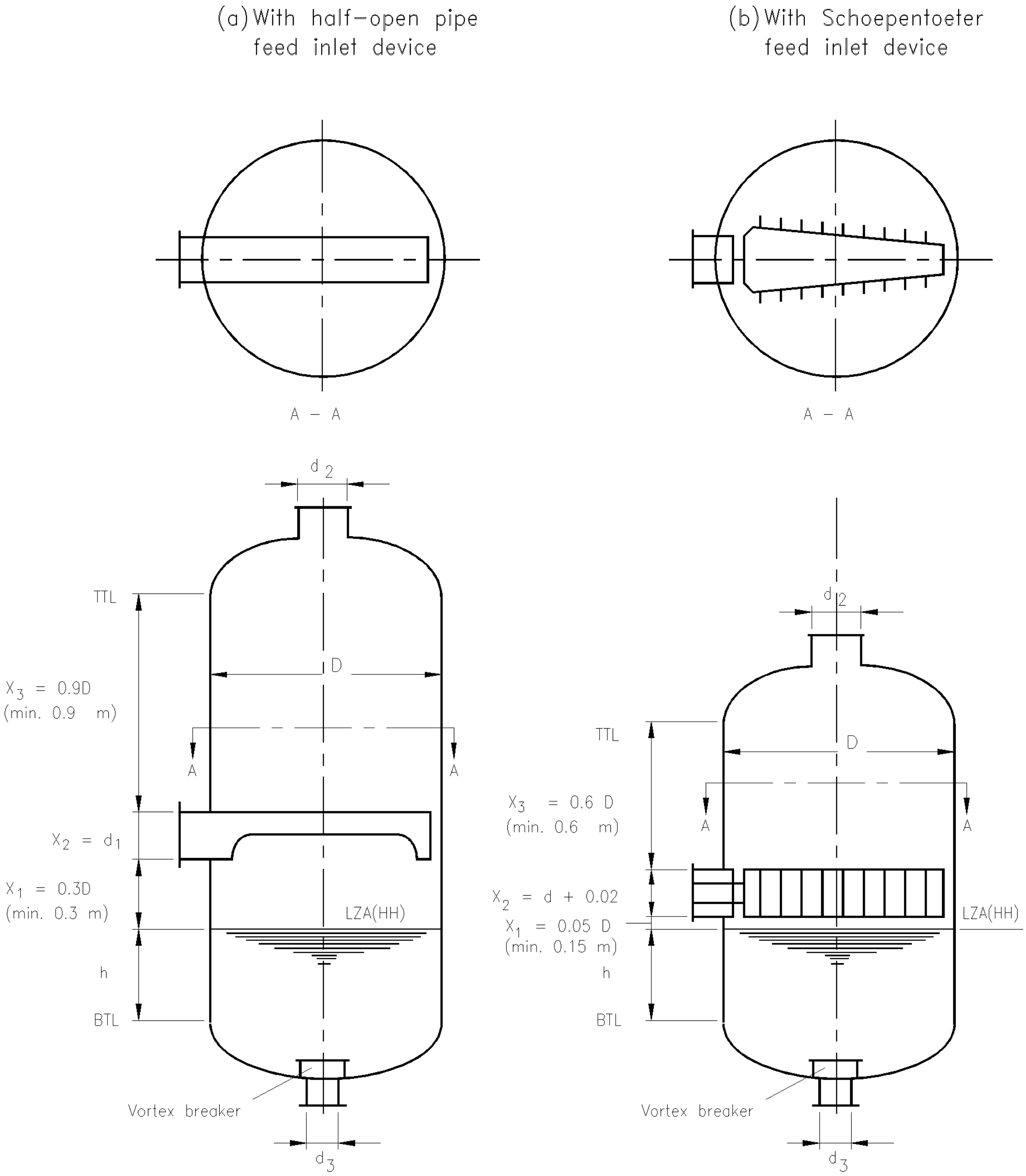
3.1.5 Pressure drop

The pressure differential between inlet and vapour outlet is

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

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. If a Schoepentoeter inlet is fitted, the Schoepentoeter pressure drop shall also be included, see Appendix II.

Figure 3.1 Vertical knock-out drum



TTL = Top Tangent Line
BTL = Bottom Tangent Line

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

$$\lambda_{max} = Q^*_{max} / A_{G,min} = 0.07 \qquad [m/s]$$

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

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

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.

If the pressure is well above 90 bara, or if otherwise the surface tension is as low as $5x10^{-3}$ N/m or below, see also AppendixVIII.

In the case of flare knock-out drums higher λ-values are acceptable. For more information on flare knock-out design DEP 80.45.10.10-Gen should be consulted.

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. This will result in a vessel tangent-to-tangent length/diameter ratio of between 2.5 and 6.

The vessel diameter shall be sufficiently large to accommodate the feed inlet device. Also, sufficient distance (> 0.15 m) shall be available between the bottom of the inlet device and LZA(HH).

The vessel should also be large enough to enable proper disengagement of gas from the bulk liquid. The relevant criteria are discussed in Appendix VII. 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 or in the head 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).

In flare knock-out drums smaller nozzles are acceptable. For the design of flare knock-out drums DEP 80.45.10.10-Gen should be consulted.

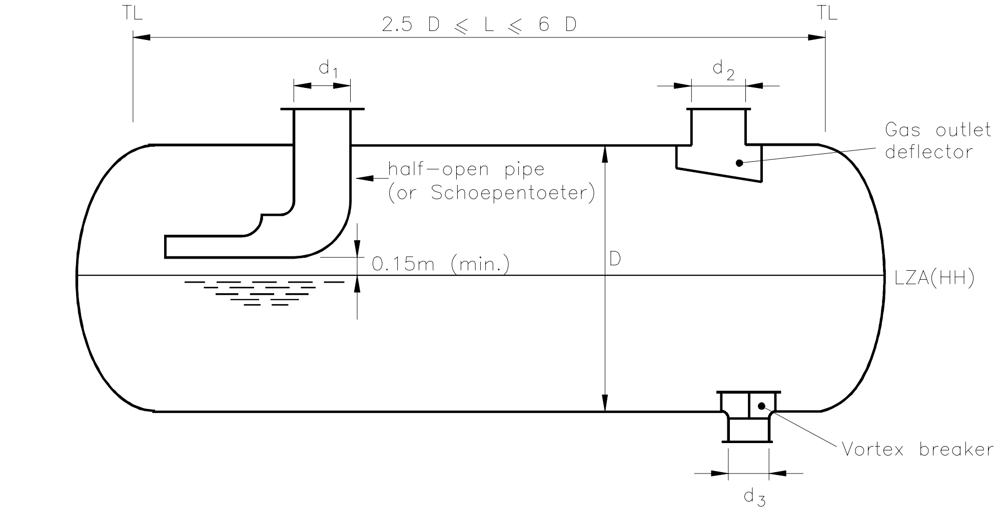
Note: For outlet deflectors in flare knockout drums additional mechanical strength requirements apply.

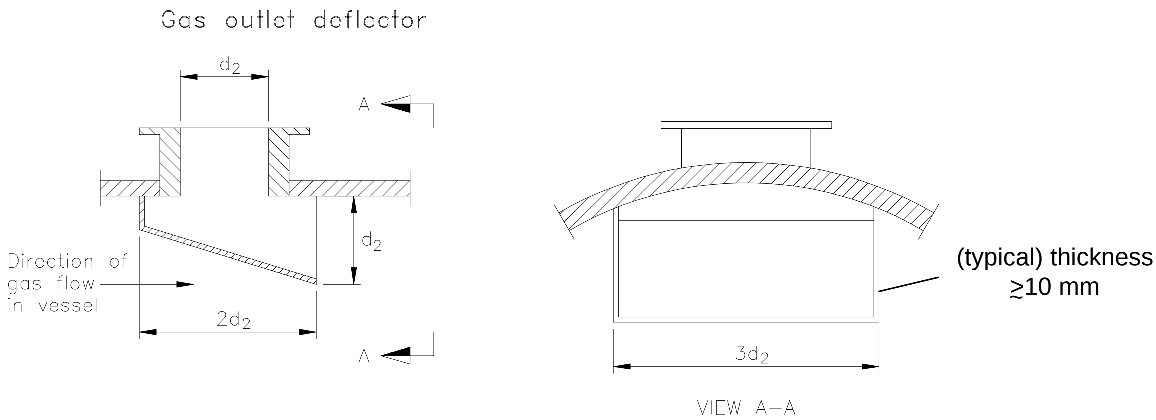
3.2.4 Pressure drop

The pressure differential between inlet and vapour outlet is

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

Figure 3.2 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.
- for compressor suction scrubbers, in non-fouling service, provided that precautions are taken to prevent the disengagement of loose wire cuttings.

Non-recommended use:
- fouling service (wax, asphaltenes, sand, hydrates)
- for viscous liquids where de-gassing requirement determines vessel diameter

Typical process applications:
- production/test separator (non-fouling, moderate GOR);
- 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 specification and installation

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.

The thickness of a horizontal mat in a vertical vessel is normally 0.1 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 support 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.

In a vertical vessel perforated plates shall NOT be mounted upstream of the demister (in an attempt to minimise gas flow maldistribution over the wiremesh), since during operation

liquid will accumulate between the plate and the mat, 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 40 % 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 dispersion in the flow line 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:

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

or

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

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

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

$f_\phi \approx 1/(1+10\,\phi_{wm})$ 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.

If the pressure is well above 90 bara, or if otherwise the surface tension is as low as $5x10^{-3}$ N/m or below, see also AppendixVIII.

The vessel should also be large enough to enable proper disengagement of gas from the bulk liquid. The relevant criteria are discussed in Appendix VII.

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:

$$H = h + X_1 + X_2 + X_3 + t_{wm} + X_4 \qquad [m]$$

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 [m]
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.

Amended per Circular 14/08

For the design of Schoepentoeters, see DEP 31.20.20.31-Gen.

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

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

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

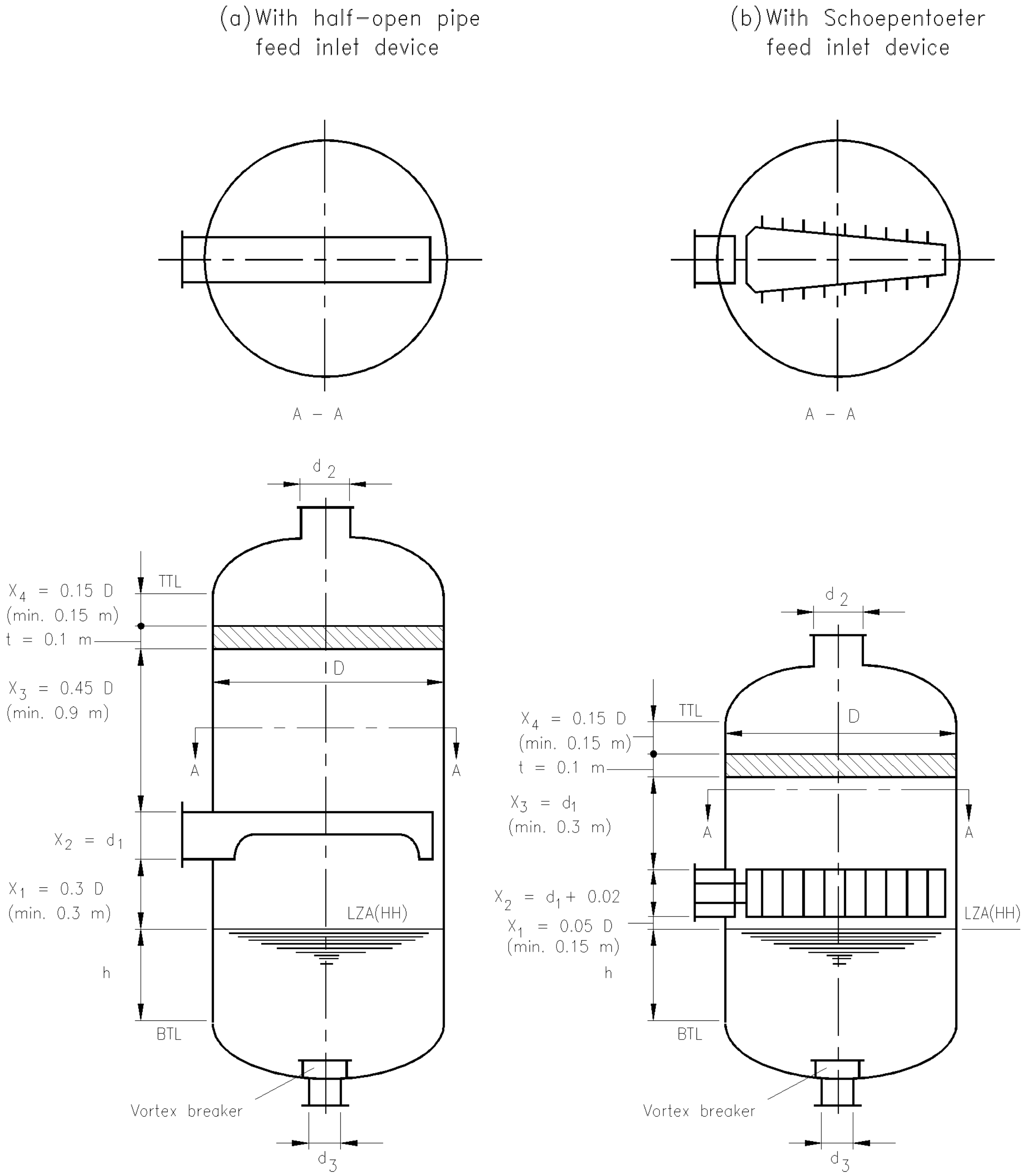
$$\Delta p_{wm} = 200\,(\rho_L - \rho_G)\,\lambda^2\,t_{wm} \qquad [Pa]$$

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

(≈ 22 mm at maximum gas load conditions if mat thickness t_{wm} = 0.1 m)

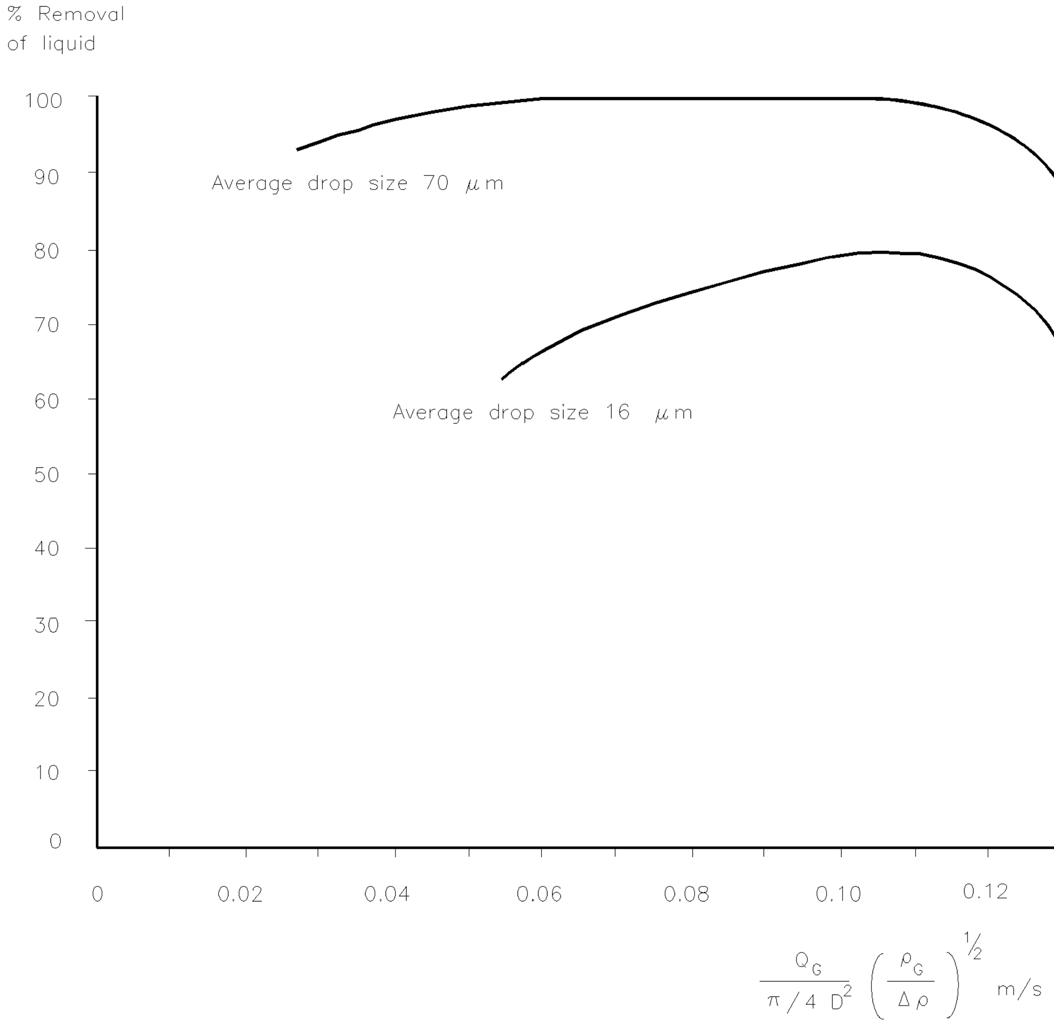
NOTE: This is the wet pressure drop. The dry pressure drop across the mistmat is 50 % of this value. In the above recommended formula for the pressure drop an averaged correction for the liquid loading of the mistmat has been taken into account.

Figure 3.3 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 x 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

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.

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 vertical demister mat shall extend from the top of the vessel to 0.10 m below LZA(LL). The area between the mat and the bottom of the vessel shall allow free passage of the full liquid flow (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.

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.

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:

$$A_{G,min} = Q^*_{max} / \lambda_{max} \qquad [m^2]$$

and $$\lambda_{max} = 0.09\, f_\eta f_\phi \qquad [m/s]$$

so $$A_G \geq Q^*_{max} / (0.09\, f_\eta f_\phi) \qquad [m^2]$$

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

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

$f_\phi \approx 1/(1+10\,\phi_{wm})$ 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.

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.

If the pressure is well above 90 bara, or if otherwise the surface tension is as low as $5x10^{-3}$ N/m or below, see also AppendixVIII.

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 should also be large enough to enable proper disengagement of gas from the bulk liquid. The relevant criteria are discussed in Appendix VII.
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 head 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.

Amended per Circular 14/08

For the design of Schoepentoeters, see DEP 31.20.20.31-Gen.

The gas outlet shall be located on the top of the vessel and should 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

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

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

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

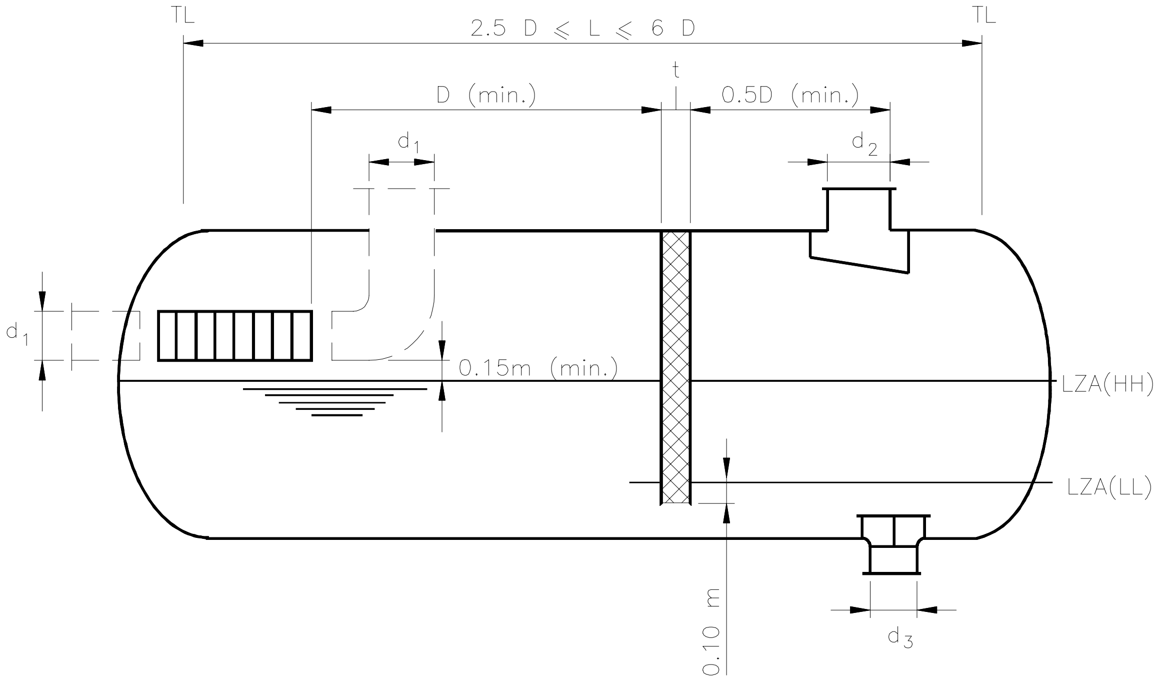
$$= 20\,000\,\lambda^2\, t_{wm} \qquad \text{[mm process liquid]}$$

(≈ 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 formula recommended above for the pressure drop an averaged correction for the liquid loading of the mistmat has been taken into account.

Figure 3.5 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 VERTICAL 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 (straight or single-pocket vanes only);
– not suitable for pressures above 70 bar;
– robust design;
– sensitive to liquid slugs (in-line separator cannot handle slugs at all).

Recommended use:

– typically for demisting service;
– in-line separator to be used only with relatively low flow parameter ($\phi_{feed} < 0.01$);
– two-stage separator to be used if $\phi_{feed} \geq 0.01$;
– attractive for slightly fouling service (straight or single-pocket vanes only);
– may be used where demister mats may become plugged, e.g. waxy crudes, sulphur recovery units.

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 70 bar, due to the consequent sharp decline in liquid removal efficiency and insufficient turndown.

Typical process applications:

– demisting vessels with slightly fouling service.

– compressor suction scrubbers – where vane packs are preferred to demister mats since their construction is more robust. In view of the limited separation efficiency of vanepacks, this application should be limited to operating pressures well below 50 bar.

3.5.2 Vane pack specification

Until recently, the vertical vane-type demister vessels available on the market were normally equipped with a horizontally flowed-through vane pack. Vertically flowed-through vane packs are also available, and these are preferred in two-stage vane separators because they make better use of the available space and are less prone to gas flow maldistribution.

Basically, there are three types of vane elements: the straight (no-pocket), single-pocket and double-pocket type. In vertically flowed-through vane packs only double-pocket vanes are applied. In Appendix IX more information is given on the various vane designs and some general guidelines are given for the installation of vane packs.

3.5.3 Determination of vane pack face area

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

$$A_v = Q^*_{max} / \lambda_{v,max} \qquad [m^2]$$

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

$\lambda_{v,max}$ 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.

In vanepacks different re-entrainment mechanisms can occur depending on physical properties. The relevant regime can be established from the Archimedes number. The Archimedes number is a dimensionless physical property group and is defined as:

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

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)

  $$\lambda_{v,max} = c_1\{g\,\sigma / (\rho_L - \rho_G)\}^{0.24} (\sigma/\eta_L)^{0.04} / (1 + 25\phi_V) \qquad [m/s]$$

  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:** $Ar \leq 225$

  $$\lambda_{v,max} = c_2 (\sigma/\eta_L) / (1+25\phi_V) \qquad [m/s]$$

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

For horizontal flowed through vanepacks $c_1 = 1.75$ and $c_2 = 0.14$.

For vertical flowed through vanepacks the gas handling capacity is lower, reflected in lower values for the proportionality constants, i.e. $c_1 = 0.95$ and $c_2 = 0.08$. The difference is partly due to the different vane configuration, for the other part due to the different flow direction.

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 vane-type demister 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 % may be assumed for a correctly sized demister.

3.5.5 In-line separator with horizontal flow vane pack

(Figure 3.6a)

Warning: This type shall only be used if ϕ_{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:

$$h_v = \sqrt{1.5\ A_v} \qquad [m]$$

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

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

h_v should be adjusted to fall within the following range:

$$0.30 \le h_v \le 1.5 \qquad [m]$$

The vane pack width:

$$w_v = A_v / h_v \qquad [m]$$

The width of the mist extractor box:

$$w_{vb} = w_v + 0.1 \qquad [m]$$

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).

$$\text{Typically, } h_{vb} = h_v + 0.3 \qquad [m]$$

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 0.20 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 back of the vane pack.
   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:

1. $D \geq 0.2 + \sqrt{(w^2_{vb} + t^2_{vb})}$ [m] (vane box requirement)
2. $D \geq 0.6$ [m] (vessel accessibility requirement)
3. The vessel should also be large enough to enable proper disengagement of gas from the bulk liquid. The relevant criteria are discussed in Appendix VII.

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

$$H = h + X_1 + h_{vb} + X_2 \qquad [m]$$

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:

$$d_1 \geq 0.43 \sqrt{A_v} \qquad [m]$$

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:

$$d_2 \geq d_1 \qquad [m]$$

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).

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

where

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

in which $K_v = 15$ for single-pocket vanes
or $K_v = 10$ for double-pocket vanes

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

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

3.5.6 Vertical demister with vertical flow vane pack

(Figure 3.6c)

In recent years, BM/KCH 628 vertical flow vane packs with hollow vanes were retrofitted as a replacement of mist mats in vertical wiremesh demisters because it was not feasible to install a horizontal flow vane pack. Following positive experience with these vane packs they are now considered as equally acceptable.

In liquid sulphur service these vane-packs are designed with short drain pipes without a hydraulic seal. This is possible because the pressure drop over the vane-pack is low and density of sulphur is high.

Amended per Circular 14/08

The maximum liquid load to the vanepack is limited to a feed flow parameter $\phi < 0.01$. In case a schoepentoeter is used as inlet device the flow parameter can be higher, provided that the separation efficiency of the schoepentoeter is sufficient to reduce ϕ to 0.01 just upstream of the vanepack. This means that for hydrocarbon systems the vessel gas load factor should be limited to

Amended per Circular 14/08

$$\lambda_{max} = Q^*_{max} / A_{G,min} = 0.11 + 0.0095\,\phi_{feed}^{-0.75} \quad \text{[m/s]}$$

For aqueous systems this can be relaxed to:

Amended per Circular 14/08

$$\lambda_{max} = Q^*_{max} / A_{G,min} = 0.11 + 0.0125\,\phi_{feed}^{-0.75} \quad \text{[m/s]}$$

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

$$\lambda_{max} = 0.15 \quad \text{[m/s]}$$

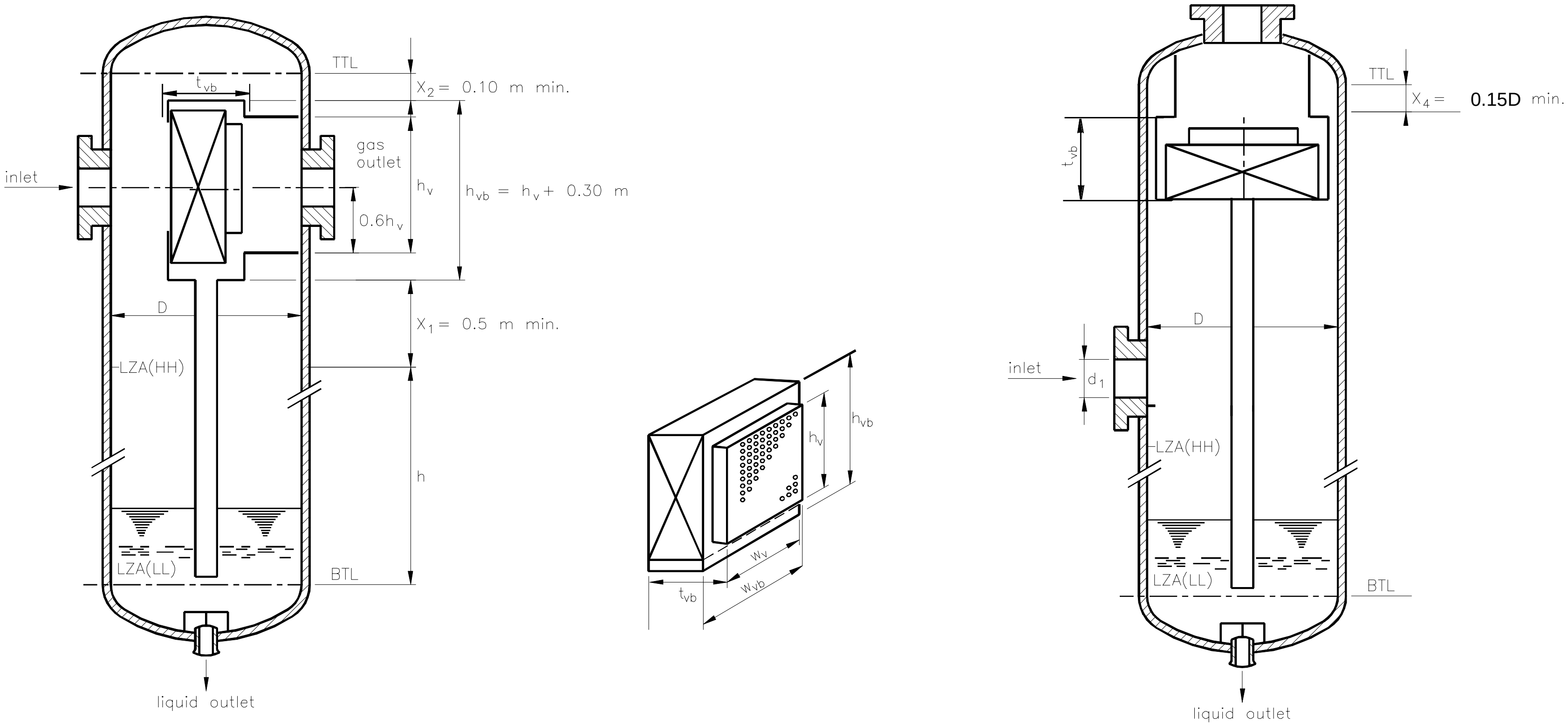
If **slugs** are expected:

Amended per Circular 14/08

$$\lambda_{max} = 0.10 \quad \text{[m/s]}$$

(to avoid flooding of the vanepack).

Figure 3.6 Layout of vertical vane-type separator



(a) IN-LINE SEPARATOR

(b) VANE BOX DETAILS

(c) VERTICAL FLOW 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 70 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 with either single-pocket or double-pocket vanes. 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. Subsequently the vane pack is dimensioned with sufficient vane area for proper demisting.

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

For hydrocarbon systems the minimum vessel cross-sectional area for gas flow, $A_{G,min}$, is derived from:

Amended per Circular 14/08

$$\lambda_{max} = Q^*_{max} / A_{G,min} = 0.11 + 0.0095\ \phi_{feed}^{-0.75} \qquad [m/s]$$

For aqueous systems this can be relaxed to:

Amended per Circular 14/08

$$\lambda_{max} = Q^*_{max} / A_{G,min} = 0.11 + 0.0125\ \phi_{feed}^{-0.75} \qquad [m/s]$$

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

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

$$A_G \geq Q^*_{max} / \lambda_{max} \qquad [m^2]$$

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.

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

$$\lambda_{max} = 0.15 \qquad [m/s]$$

If **slugs** are expected:

$$\lambda_{max} = 0.10 \qquad [m/s]$$

(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.

Note that 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).

The vessel should also be large enough to enable proper disengagement of gas from the bulk liquid. The relevant criteria are discussed in Appendix VII. 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 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 vane height, h_V, shall be between 0.3 m and 1.5 m.

Measures shall be taken to prevent gas bypassing the vane pack. The vanes are to be 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 is shown in Figure 3.7.

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

3.6.5 Nozzles

The feed nozzle shall be fitted with a Schoepentoeter.

Amended per Circular 14/08

For the design of Schoepentoeters, see DEP 31.20.20.31-Gen.

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.)

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

where

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

in which K_v = 15 for single-pocket vanes

or K_v = 10 for double-pocket vanes

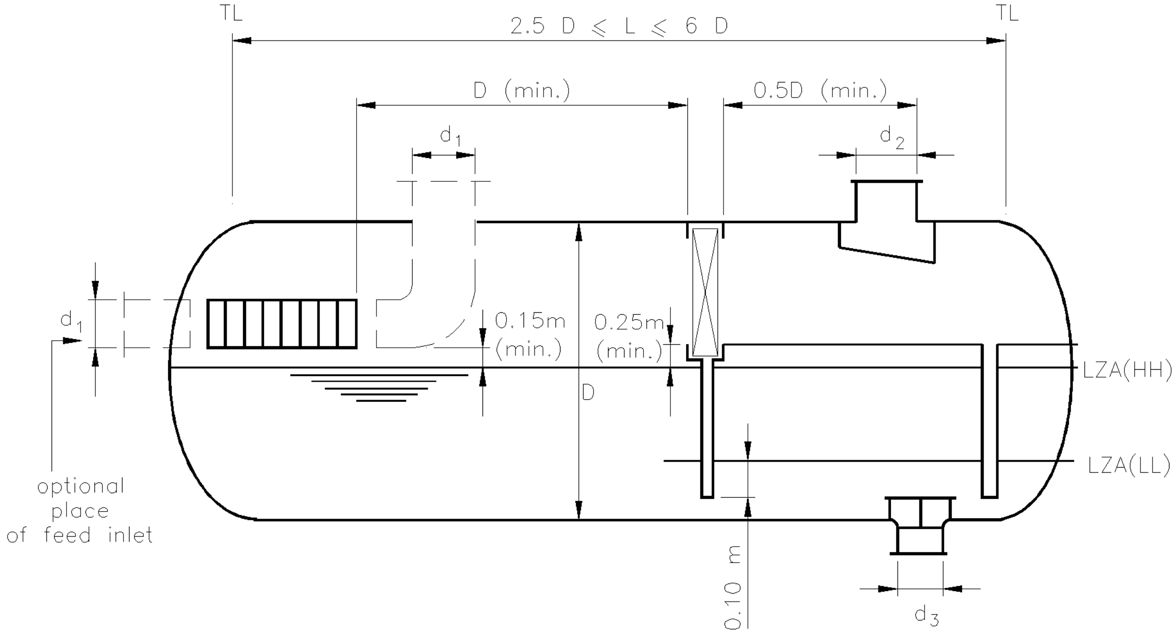
and n = number of perforated plates

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

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

Figure 3.7 Horizontal vane-type demister

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



3.7 CYCLONE WITH TANGENTIAL INLET (CONVENTIONAL CYCLONE)

(Figure 3.8)

3.7.1 Selection criteria

Application:

– demisting of gas in fouling service

Characteristics:

– liquid removal efficiency > 96 %;
– robust 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 applications:

– in gas plants:	LTX Separator
– in oil refineries:	Thermal Gasoil Units (TGU); Visbreaker Units (VBU);
– in chemical plants:	Thermoplastic Rubber Plants.

3.7.2 General

The cyclone contains the following basic elements:

– tangential inlet.
In Figure 3.8 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 (e.g cyclones used in Thermal Crackers), the Principal should be consulted.

3.7.3 Diameter

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

(See also Figure 3.8)

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

Type 1 (rectangular inlet):

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

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

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

– If erosive material is present in feed:

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

– If the liquid has a foaming tendency:

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

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 "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.

$D \geq 2.8a$ [m]

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

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

Type 2 (circular inlet):

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

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

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

– If erosive material is present in feed:

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

– If the liquid has a foaming tendency:

preferably $d_1 \leq 0.356\ Q_G^{0.5}$ [m]
(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:

$D \geq 3.5d_1$ [m]

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

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

3.7.4 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.

3.7.5 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 3.8).

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.

3.7.6 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).

3.7.7 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 3.8).

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:**

$s = 0.025D$ (with a minimum of 0.01 m) [m]

– Under **foaming** conditions:

$s = 0.05D$ (with a minimum of 0.02 m) [m]

– Under **coking** conditions:

$0.05 \leq s \leq 0.1$ [m]

3.7.8 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:

$$H = h + X_1 + H_{vfb} + X_2 + X_3 + X_4 \qquad [m]$$

All the terms on the right hand side of the above equation have been indicated in Figure 3.8.

$X_1 \geq 0.1 + 0.5D \tan(10°)$ [m]
$X_2 \geq 0.2a$ (cyclone type 1) or $\geq 0.2d_1$ (cyclone type 2) [m]
$X_3 = a$ (cyclone type 1) or $= d_1$ (cyclone type 2) [m]
$X_4 = 0.1a$ (cyclone type 1) or $= 0.1d_1$ (cyclone type 2) [m]

3.7.9 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.

3.7.10 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).

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

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

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.

3.7.11 Liquid drain sealing

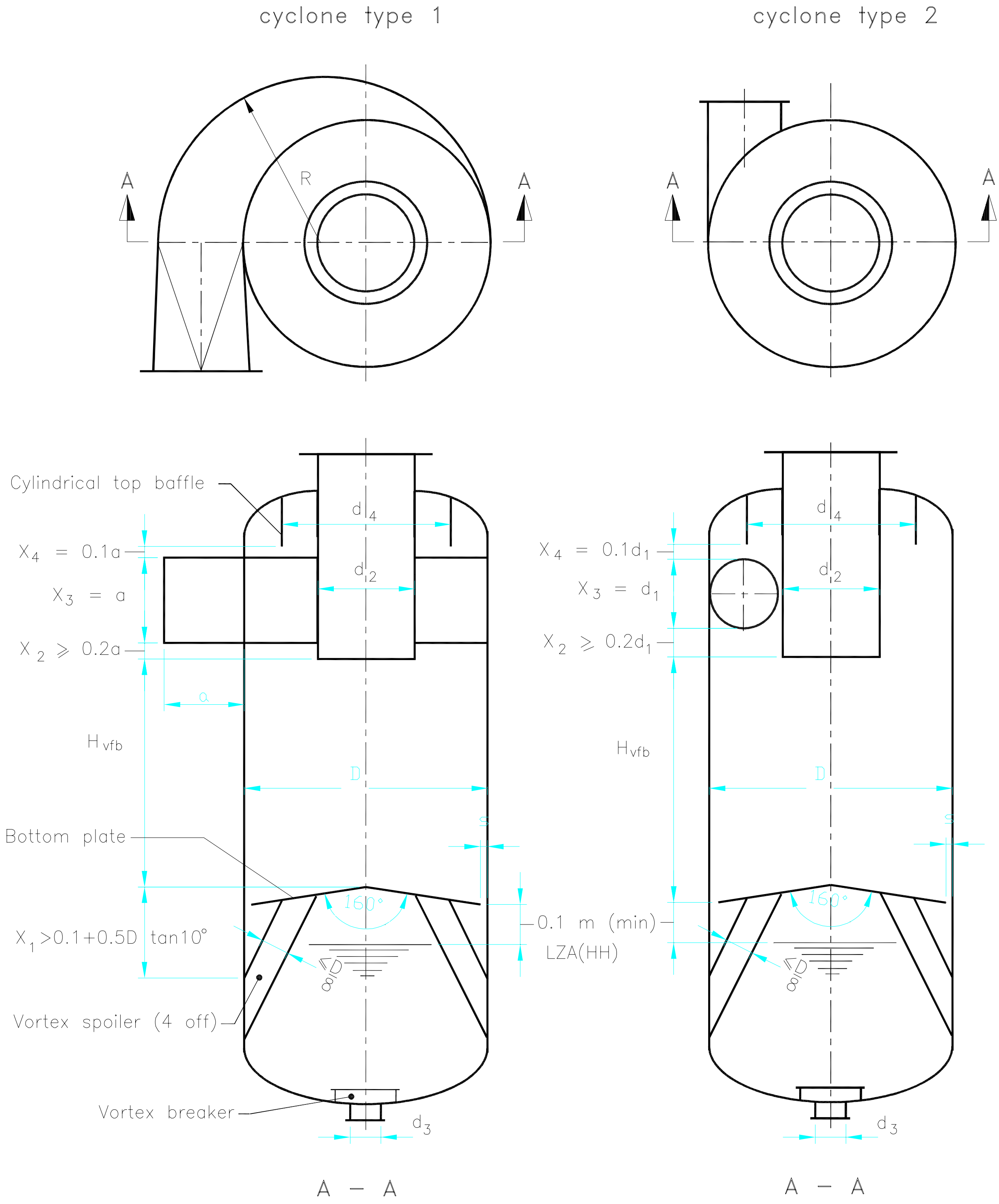
For satisfactory operation of any cyclone, it is essential that the liquid drain is sealed, i.e. there shall be no flow of gas in either direction of the drain.

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 3.8 Cyclone with tangential inlet

Amended per Circular 03/08



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

(Figure 3.9)

3.8.1 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 (98 % to 99 %);
- very high gas handling capacity (maximum allowable gas load factor 0.9 m/s);
- turndown ratio of 3;
- 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;
- in case of high liquid loads, i.e. higher than 3 % by volume.

Typical process application:

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

3.8.2 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 3.9 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 under pressure 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".

3.8.3 Diameter

The diameter shall satisfy the following critera:

1. The gas handling capacity criterion:

For pressures up to 100 bar:

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

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

For pressures above 100 bar, further derating to λ = 0.6 to 0.7 is required. 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

Amended per Circular 03/08

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

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.

3.8.4 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

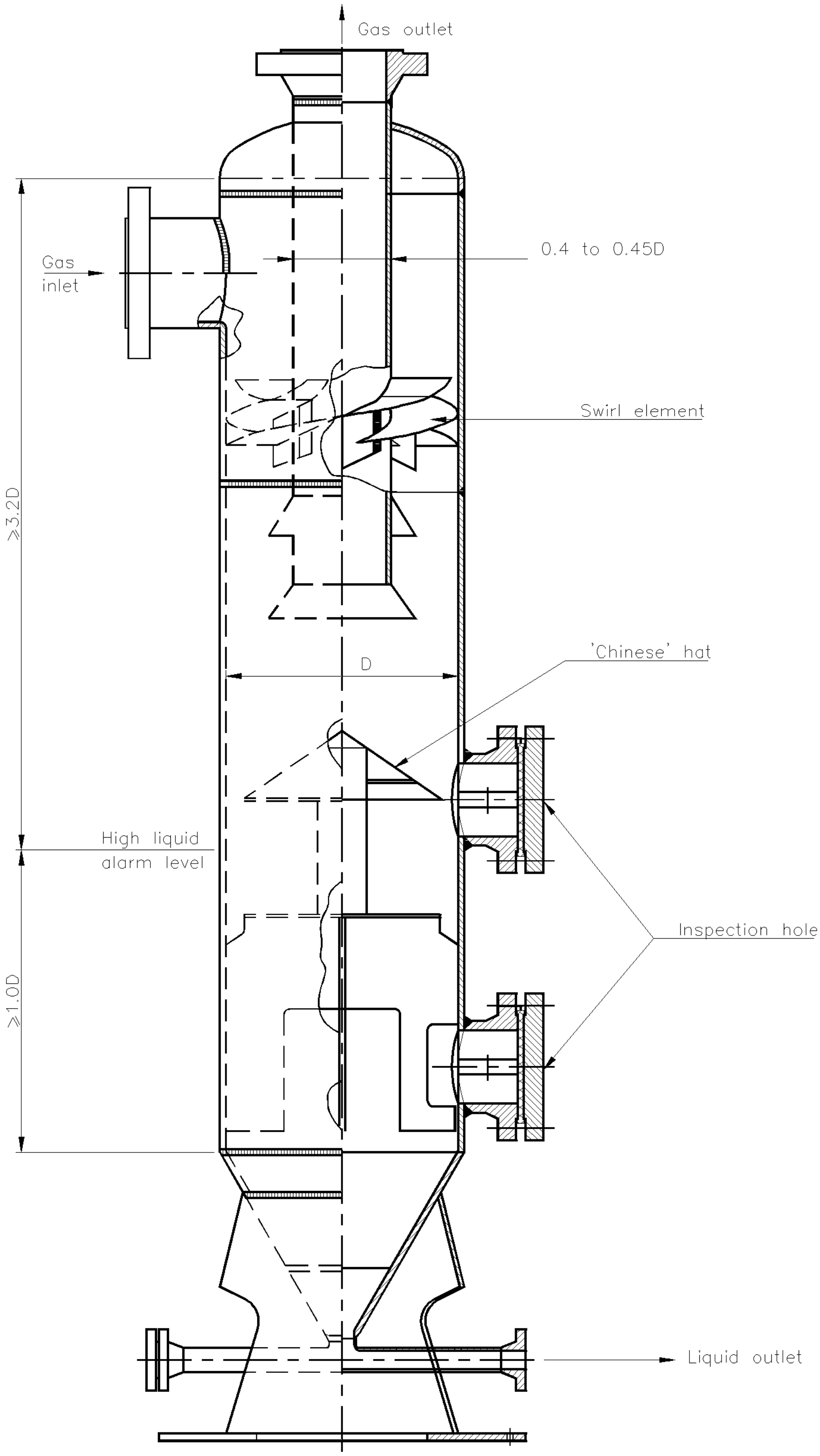
3.8.5 Droplet removal performance

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:

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

$$d_{crit99} = \sqrt{2} * d_{crit50} \qquad [\mu m]$$

Figure 3.9 Cyclone with straight inlet and swirler ("Gasunie cyclone")



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

(Figure 3.10)

3.9.1 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:

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

3.9.2 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.

3.9.3 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 **actual** $m^3/s \le Q_{G,c} \le$ 0.007 **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.

3.9.4 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.

3.9.5 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.

3.9.6 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:

$$n_c = Q_{G,max} / 0.007 \quad [-]$$

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 should also be large enough to enable proper disengagement of gas from the bulk liquid. The relevant criteria are discussed in Appendix VII. In practice, these two criteria will almost always be satisfied because of the limitations of the feed flow parameter (ϕ_{feed} < 0.003).

3.9.7 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 a vessel with gas outlet at the top (Figure 3.10a):

$$H = h + 0.6 + 0.15 + d_0 + 0.15 + 0.2D \qquad [m]$$

- for a vessel with gas outlet at the vessel side (0 3.10b):

$$H = h + 0.6 + 0.15 + d_0 + 0.2D + d_0 + 0.7D \qquad [m]$$

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

3.9.8 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:

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

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.

3.9.9 Pressure drop

The pressure drop across the multicyclone separator is:

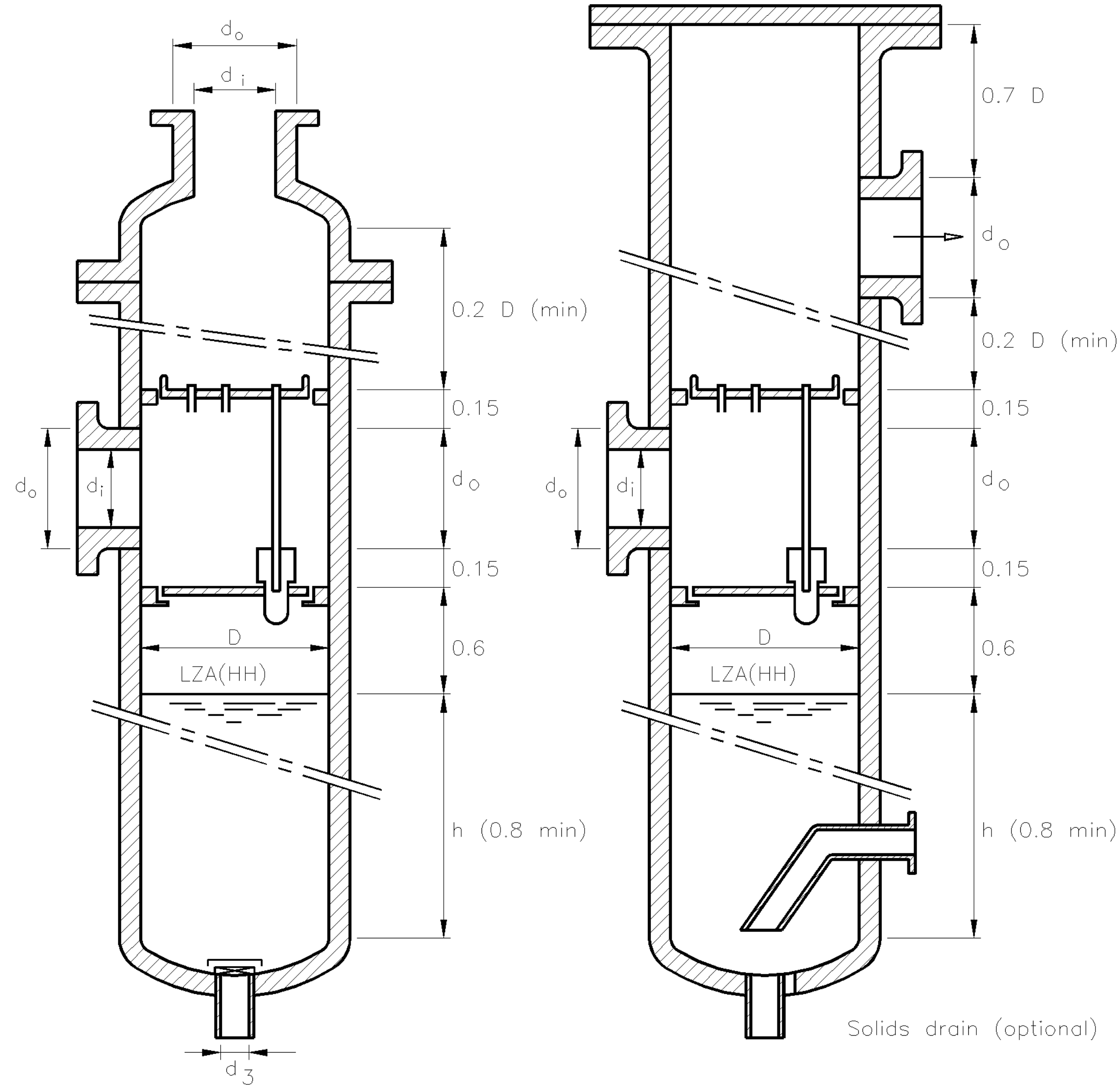
$$P_{in} - P_{out} = K_c \rho_G v^2_{G,c} \qquad [Pa]$$

Typically, K_c = 17 for 50 mm 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 3.10 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.

3.10 VERTICAL SEPARATOR WITH COALESCING MISTMAT/ VANEPACK AND AXIAL FLOW MULTICYCLONE DEMISTER DECK

Vertical separators with swirltube demister decks were developed by Shell Research in cooperation with NAM about twenty years ago. By extending a vertical wiremesh demister with a downstream multicyclone deck, a much higher gas handling capacity can be achieved. In this separator concept the wiremesh is used in two capacities: at normal flow rates it functions as a coalescer to enhance the efficiency of the downstream swirldeck, at turndown it regains its function as a demister and takes over the separation from the swirldeck. From this so-called SMS separator, the SVS, SMSM and SMMSM variants were developed for dedicated applications. These alternatives and their specific application window is discussed in (3.10.1).

Apart from Shell, an increasing number of vendors offer different embodiments of multicyclone separator internals. In cases where it is considered to apply these internals please refer to sections 3.10.2. Regardless of the type of multicyclone internals that is used, the vessel shall be sized for a maximum gas load factor - including design margin - of 0.25 m/s.

3.10.1 Shell Swirltube Separators

Types:

1. SCHOEPENTOETER-MISTMAT-SWIRLDECK SEPARATOR (SMS)
(Figures 3.11a and 3.11b)
2. SCHOEPENTOETER-VANE PACK-SWIRLDECK SEPARATOR (SVS)
(Figure 3.12a)
3. SCHOEPENTOETER-MISTMAT-SWIRLDECK-MISTMAT SEPARATOR (SMSM)
(Figure 3.12b)
4. SCHOEPENTOETER-MISTMAT-MISTMAT-SWIRLDECK-MISTMAT SEPARATOR (SMMSM)

3.10.1.1 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 SM(M)SM:
  - high efficiency;
  - sensitive to fouling;
  - high turndown ratio (factor 10).
– SVS:
  - 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. Can 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;

Typical process applications:

– SMS: production separator, compressor suction drum,;
– SVS: well head separator, centrifugal compressor suction drum.

- SMSM: inlet scrubber for glycol contactors, inlet scrubber for gas export pipelines;
- SMMSM: cold separator (removal of glycol/HC liquid mixture), inlet scrubber for glycol contactors, amine contactors

3.10.1.2 General

The SMS, SVS, and SM(M)SM 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), a high performance, double mistmat (SMMSM) 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 and SMMSM, for demisting of the secondary gas to further improve the gas/liquid separation efficiency.

In this Section the various internals of the SMS, SVS,SMSM and SMMSM will be specified and relevant design rules will be given for each type.

Purchasing policy

For the SMS, SVS and SM(M)SM separators the vessel and total package of internals should be purchased separately.

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

Vessel attachments for the SMS, SVS and SM(M)SM separators may be specified by the Manufacturer, but the Principal or the Contractor shall specify the Schoepentoeter and swirldeck.

3.10.1.3 Schoepentoeter

This feed inlet device is used in all three types. Its function is to distribute the gas flow over the cross section of the vessel, and the bulk separation of the gas and liquid entering the separator

Amended per Circular 14/08

For the detailed design see DEP 31.20.20.31-Gen.

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).

3.10.1.4 Coalescing mistmat

Both in the SMS and SMSM a standard 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).

To avoid ambiguity it is recommended to size the primary mistmat for a minimum gas load factor (at design) of 0.15 m/s.

Manufacturing and mounting

The manufacturing and mounting of the mistmat are the same as those of the wiremesh demister, see section 3.3. Purchase of the mistmat should be left to the approved internals Manufacturer since the design is complicated due to the passage of the various drain pipes. The sub-supplier of the mistmat shall be subject to approval of the Principal.

3.10.1.5 Coalescing vane pack

In the SVS a vane pack is used as coalescer instead of a mistmat. 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 lower.

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.

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

3.10.1.6 Double primary mistmat

The SMMSM separator is provided with two different primary mistmats.

In Low Temperature Separators the liquid enters in the form of a fine mist, and the separation is further complicated by the simultaneous presence of hydrocarbons and glycol. This service requires a more sophisticated coalescer design to maintain sufficient coalescence efficiency and to suppress the interference between the different fluids. For this purpose two primary mistmats are used with different specifications, one optimized to handle the glycol and the other to coalesce the hydrocarbons. This coalescer design has also been used successfully in applications with water/condensate mixtures, e.g. inlet scrubbers of gas dehydration plants.

The procurement of these so-called 9797-type mistmats should be left to the approved internals Manufacturer. For the requisitioning the density, viscosity and surface tension of the prevailing fluids shall be specified.

3.10.1.7 Swirldeck

The SMS, SVS and SM(M)SM separators all have a swirldeck consisting of swirltubes.

The swirltube:

The demister swirltube is an axial cyclone. Schematic drawings are presented in Figure 3.11.

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 on the swirler and on the tube wall as a result of 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 SM(M)SM, passes the second mistmat. The main fraction of the gas leaves the swirltube via the primary gas outlet at the top. Drain pipes guide the liquid, collected in the space between the tubes and on the upper cover of the swirldeck, to below the liquid level in the bottom of the vessel.

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 swirldeck:

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

Presently the following three types of swirl decks are used:

- Integral swirldeck (Figure 3.14a)
- Swirldeck composed of boxes (Figure 3.14b)
- Swirldeck composed of banks (Figure 3.14c)

The integral swirldeck is recommended for new vessels up to 1.2 m ID because it is easy to install and maintain. For this deck 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 3.14a.

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

Boxes are used as shown in Figure 3.14b, each box comprising two to six 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 deck), which can be even more expensive.

In very large vessels (above about 3 metres ID) it is advantageous for the swirldeck to be composed of banks. Normally, a double row of swirltubes per bank is used as shown in Figure 3.14c. This method of installation is less expensive than the installation with boxes, but a manhole with an ID of at least 550 mm is required.

The layout of an SMS with either a box-type or bank-type swirldeck 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 drain pipes. Each of these drain pipes should be fitted with a vertical plate in the entrance to act as a vortex breaker.

If more than six swirltubes are used, these vertical drain pipes are combined in a horizontal 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 drain pipe may be equipped with a flapper valve (e.g. when the feed contains a small amount of, or no, liquid) 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 and to facilitate installation, 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). This space may be reduced if the header is placed below the primary mistmat(s).

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 drain pipes also extending at least 0.10 m below LZA(LL).

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	56	0.365	0.237
1.45	N	Box	60	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	104	0.640	0.252

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

3.10.1.8 Secondary mistmat

In the SM(M)SM a secondary mistmat is installed 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. (See also Figure 3.13). The support and hold-down grid are stainless steel. The grids are of special design to ensure the mistmat flatness and positioning.

In the case of an SMSM separator the mistmat is a standard Shell stainless steel mistmat (cf. section 3.3). In the case of an SMMSM separator the secondary mistmat is also a 'special' 9797-type mistmat.

3.10.1.9 Vessel diameter

The gas handling capacity criterion:

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

or $$D \geq 2.26\sqrt{Q^*_{max}} \qquad [m]$$

Note that this is including design margin. 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 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:

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

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

If the pressure is well above 90 bara, or if otherwise the surface tension is as low as 5×10^{-3} N/m or below, see also Appendix VIII.

The vessel should also be large enough to enable proper disengagement of gas from the bulk liquid. The relevant criteria are discussed in Appendix VII.

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:

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

If more swirltubes are installed then the surplus tubes shall be blinded off. To avoid gas bypassing this blinding off shall take place at the inlet AND at the PRIMARY

and SECONDARY gas outlets of these swirltubes. Blinding at the inlet per box or bank is also acceptable.

Furthermore, there should be a sufficient number of swirltubes in the swirldeck to keep the liquid load per swirltube below the maximum allowable level. To achieve this, the liquid concentration in the feed should be less than approximately 5 % vol. For higher liquid loads, the Principal should be consulted.

3.10.1.10 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:

$$H = h + X_1 + X_2 + X_3 + X_4 + X_5 + 0.36 + X_6 \qquad [m]$$

All the terms on the right of the above equation are indicated in the drawings of the SMS, SVS and SMSM separator (Figures 3.11a, 3.11b, 3.12a and 3.12b respectively).

d_1 = internal diameter of the inlet nozzle [m]
X_1 = 0.5 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 = 0.48 m in case of an SMMSM
X_5 = 0.2 D [m]
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 [m], with a maximum of 0.4 m.
X_6 = 0.15 D + 0.3 m in case of SMSM [m]

The distance between LZA(HH) and the bottom of the swirldeck shall be sufficiently large to accommodate the liquid head in the drain pipes 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.

Under 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 drain pipe length is always fulfilled.

3.10.1.11 Nozzles

The feed nozzle shall be fitted with a Schoepentoeter if D ≥ 0.5 m.

Amended per Circular 14/08

For the design of Schoepentoeters, see DEP 31.20.20.31-Gen.

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.

3.10.1.12 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.

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

where at maximum gas load conditions

Δp_{wm} ≈ 125 [mm process liquid]
Δp_{v} ≈ 25 [mm process liquid]
Δp_{sd} ≈ 700 - 1200 [mm process liquid] depending on the application.

Pressure drop measurement

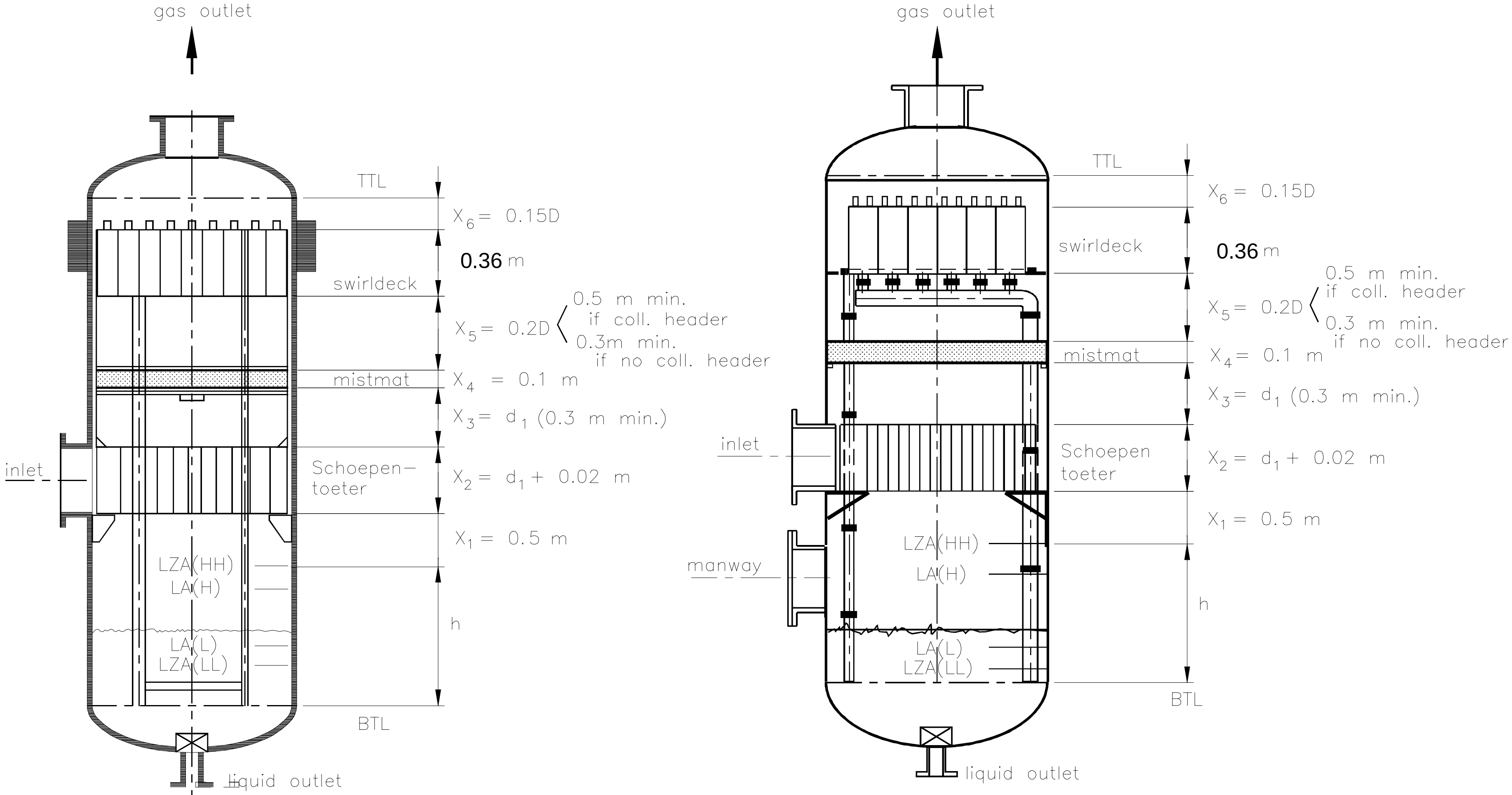
To monitor the condition of the internals (e.g. degree of fouling), it is recommended to install a pressure differential measurement across the coalescing medium and the swirldeck.

3.10.2 Non-Shell Vertical Axial Flow Multicyclone Separators

Apart from Shell, an increasing number of vendors offer different embodiments of multicyclone separator internals. Shell Global Solutions International B.V. and Shell Global Solutions US maintain and update knowledge on third party capabilities and on design criteria such as maximum gas load factor for the cyclones.

For applications of multicyclone separator internals within the US, Shell Global Solutions US should be contacted to provide consulting advice regarding the technical adequacy of the particular product for the envisaged operation on a case-by-case basis. For applications outside the US, Shell Global Solutions International B.V. should be contacted to provide such advice.

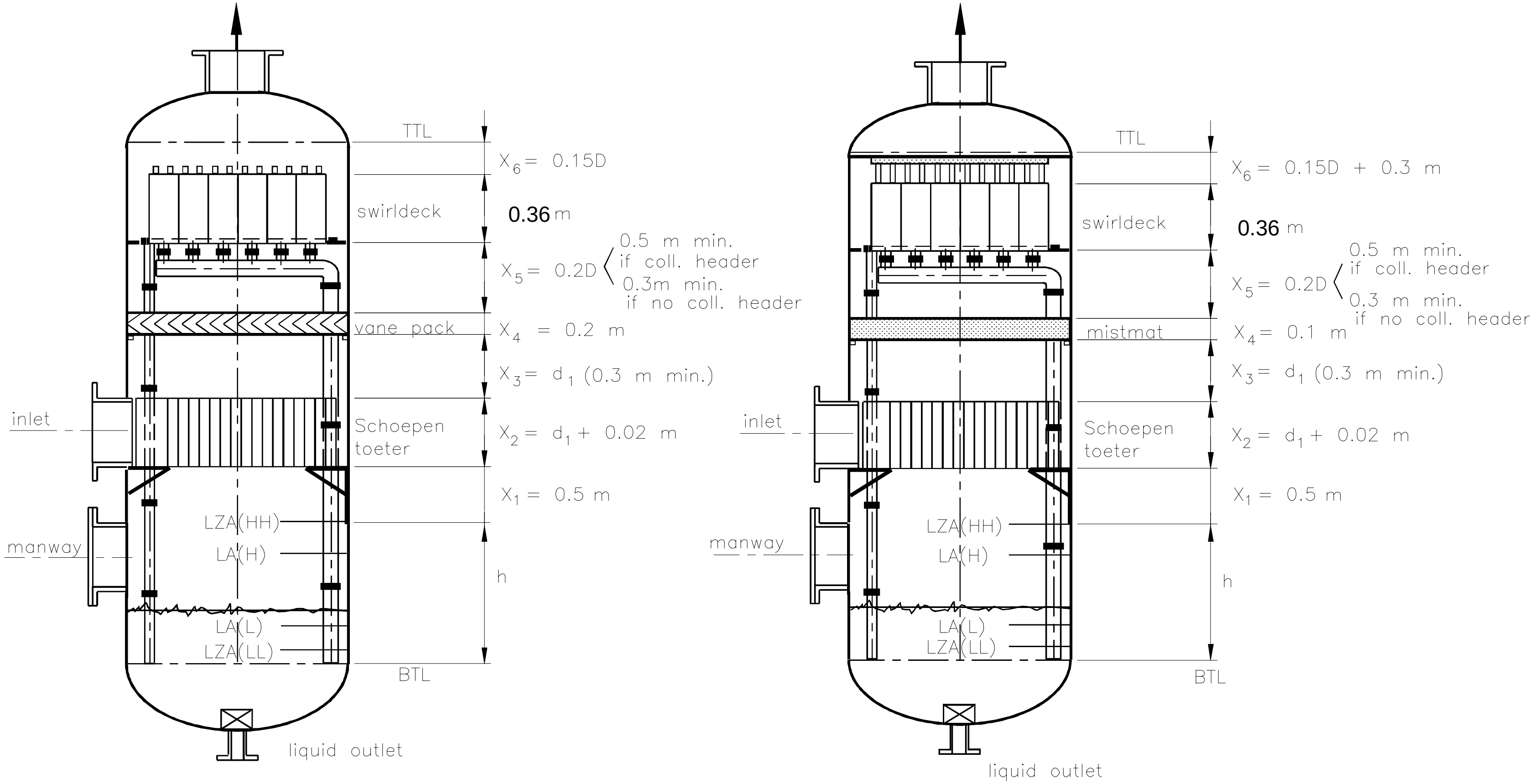
Figure 3.11 SMS separator



(a) $D \leqslant 1.2$ 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 3.12 SVS and SMSM separators



(a) SVS

(drain pipes of vane pack not shown)

(b) SMSM

Figure 3.13 Schematic outline of a single swirltube – Without and with secondary demisting (SMSM)

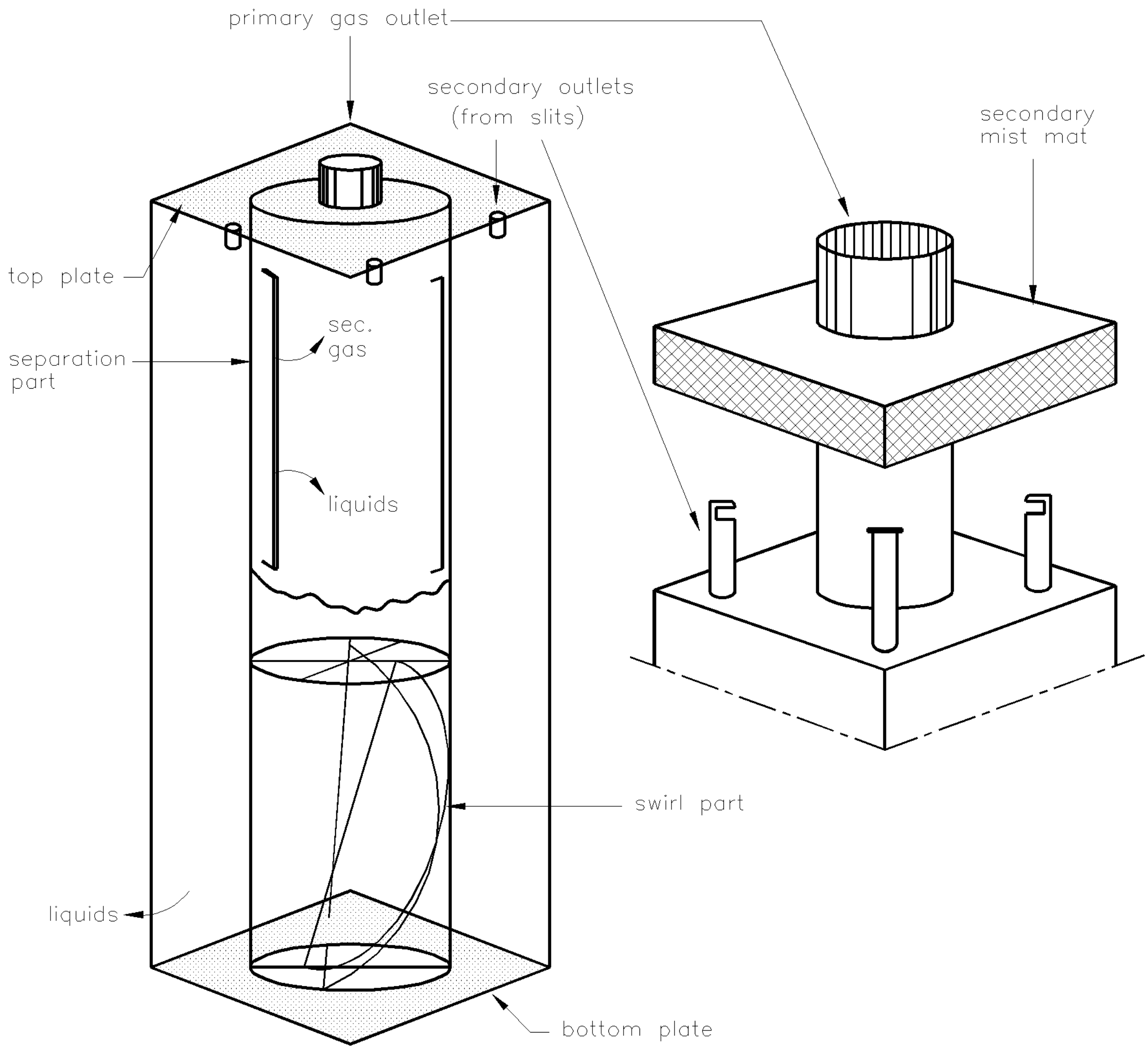
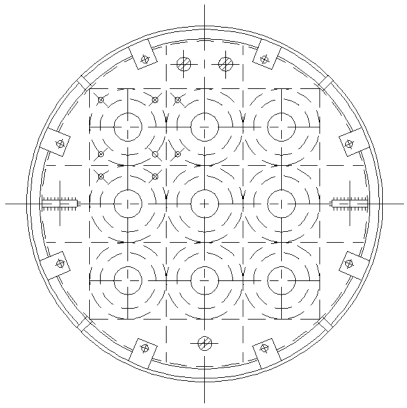
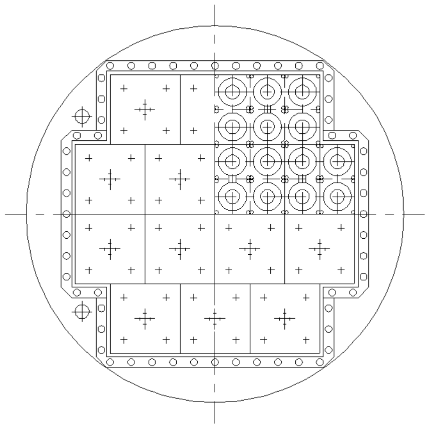


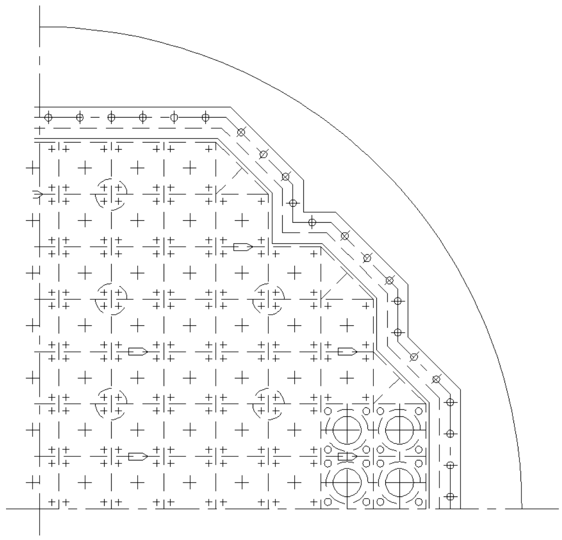
Figure 3.14 Various types of swirl decks (schematic)



(a) integral



(b) composed of boxes



(c) composed of banks

3.11 HORIZONTAL SEPARATOR WITH COALESCING MISTMAT AND AXIAL FLOW MULTICYCLONE DEMISTER DECK

3.11.1 Shell Swirltube separators

Shell swirltubes must be installed vertically. This complicates their installation in a horizontal vessel, because special precautions are required to ensure an equal distribution of the gas over all swirltubes. Furthermore it minimizes the head available for draining of the separated liquid.

If it is considered to use Shell swirltubes in a horizontal vessel the Principal should be consulted.

3.11.2 Other Horizontal Axial Flow Multicyclone Separators

Apart from Shell, an increasing number of vendors offer different embodiments of multicyclone separator internals. Embodiments of axial flow cyclones which recirculate the secondary gas to the inlet can make an installation in a horizontal orientation easier. Also here the available drain height remains an important issue. Regardless of the type of multicyclone separator internals that is used, the vessel shall be sized for a maximum gas load factor – including design margin – of 0.2 m/s.

Shell Global Solutions International B.V. and Shell Global Solutions US maintain and update knowledge on third party capabilities and on design criteria such as the maximum gas load factor for the cyclones.

For applications of multicyclone separator internals within the US, Shell Global Solutions US should be contacted to provide consulting advice regarding the technical adequacy of the particular product for the envisaged operation on a case-by-case basis. For applications outside the US, Shell Global Solutions International B.V. should be contacted to provide such advice.

3.12 FILTER SEPARATOR

(Figure 3.15)

3.12.1 Selection criteria

Application:

– after-cleaning of already demisted gas when maximum liquid removal efficiency is required.

Characteristics:

– liquid removal efficiency 50 % to 80 %;
– high pressure drop;
– unsuitable in case of high liquid loading or slugs;
– sensitive to fouling by solid particles or 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.
– a prefilter is required if 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;
– extra protection of downstream compressors, sensitive treating processes etc.

3.12.2 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 are to be used to verify the proprietary Manufacturer's design.

3.12.3 The single stage filter separator

(Figure 3.15)

The single stage filter separator is either a horizontal or a vertical separator. It consists of two compartments, a vapour distribution compartment and a filter compartment with a set of parallel filter candles. In order to be able to remove the candles as a bundle, the filter candle compartment shall be top-flanged.

3.12.3.1 The filter compartment

The function of this compartment is to coalesce fine mist. For this purpose, a bundle of filter candles is installed. 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"). The candles are grouped in either a square or triangular configuration.

The gas flow passes from the inside to the outside of the candles. In a fouling environment or if 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 may be used with the flow from the outside to the inside.

Liquid will be knocked out and is drained from the separator on level control via the tube sheet.

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 re-entrainment of the coalesced droplets in the filter compartment around the candles, the candles should fill not more than 35 % of the cross section of the vessel and the vessel load factor λ should be $\leq$ 0.1 m/s.
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 50 mm.

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

3.12.3.2 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:

$$\rho_G v^2_{G,noz} \leq 4500 \qquad \text{[Pa]}$$

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

3.12.3.6 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):

$$P_{in} - P_{out} = 0.5\rho_m v^2_{m,in} + 0.22\rho_G v^2_{G,out} + \Delta p_{cf} \qquad \text{[Pa]}$$

where

Δp_{cf} ranges typically from 0.05 bar (clean cartridges) to 0.7 bar (fouled cartridges just before replacement)

3.12.4 Horizontal two-stage filter separator

(Figure 3.16)

The horizontal two-stage filter separator consists 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 is a multicyclone bundle, a mistmat or a vane pack. This means that for sizing the vessel (diameter) the design rules for the respective demister devices shall be applied.

This separator has been installed at many locations and it is generally claimed to remove ≥ 99 % of the inlet liquids. In recent troubleshooting campaigns in the USA and The Netherlands it transpired that this was not more than about 50 %, and therefore application of this separator is no longer recommended. If total liquid removal is absolutely required, it is recommended to apply other technologies. For more information, the Principal should be consulted.

Figure 3.15 **Vertical filter separator**

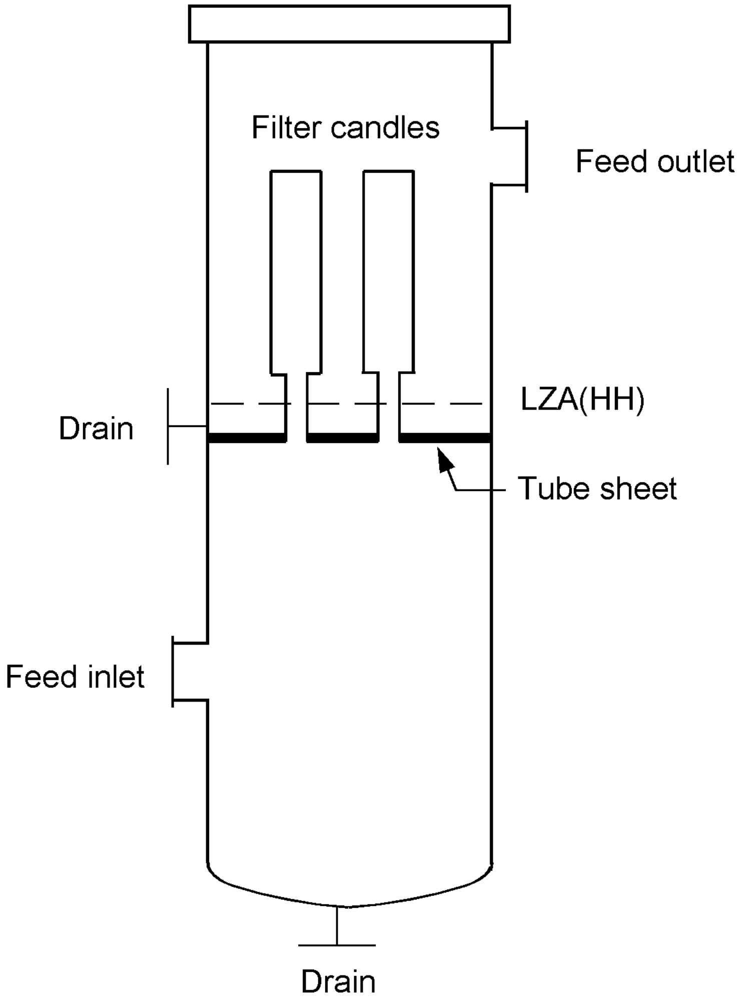
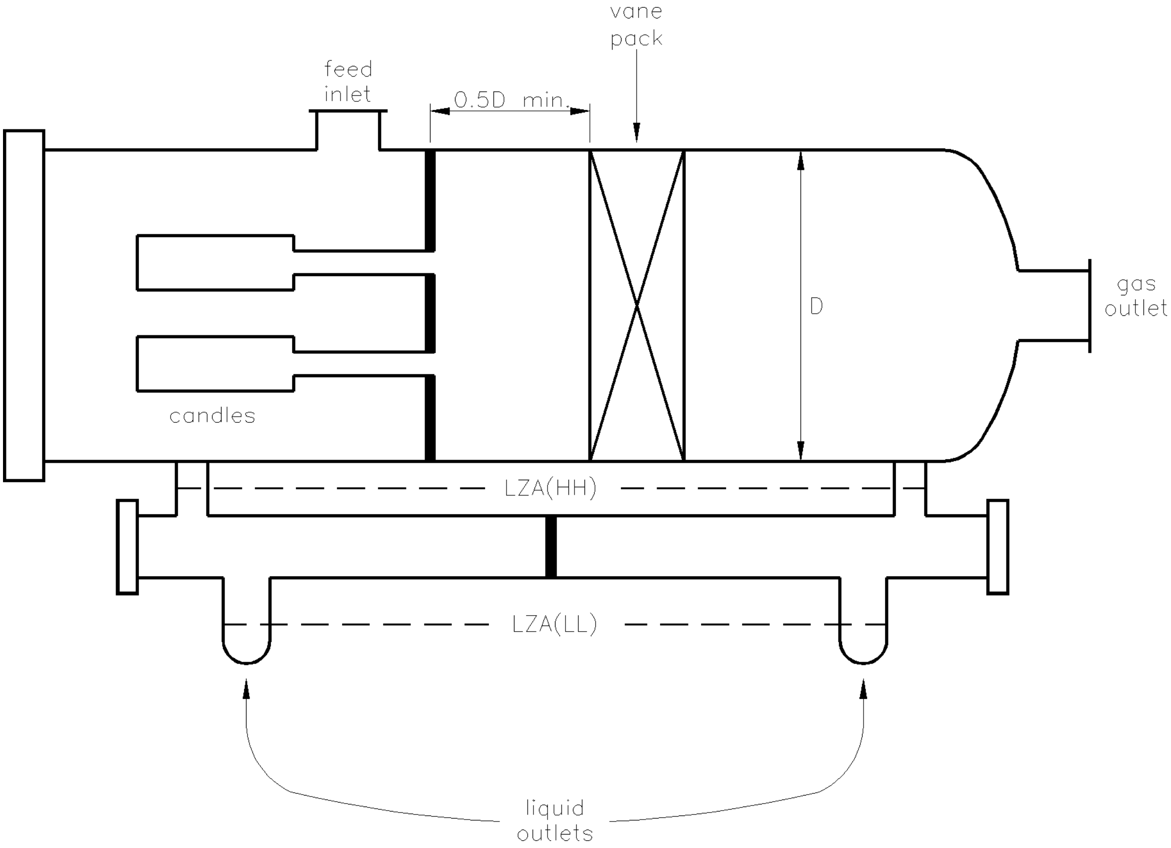


Figure 3.16 Horizontal two-stage filter separator (with vane pack as demister)



4. CONNECTING 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 shall 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.
   - 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.

d. Piping layout and location/design of any reducer shall be such that no pockets exist where liquids can accumulate and increase the risk of slug flow.

If the above conditions cannot be satisfied, liquid removal will be less efficient.

The observations listed above are illustrated in Figure 4.1.

Figure 4.1a Piping upstream of a separator: Ideal configuration

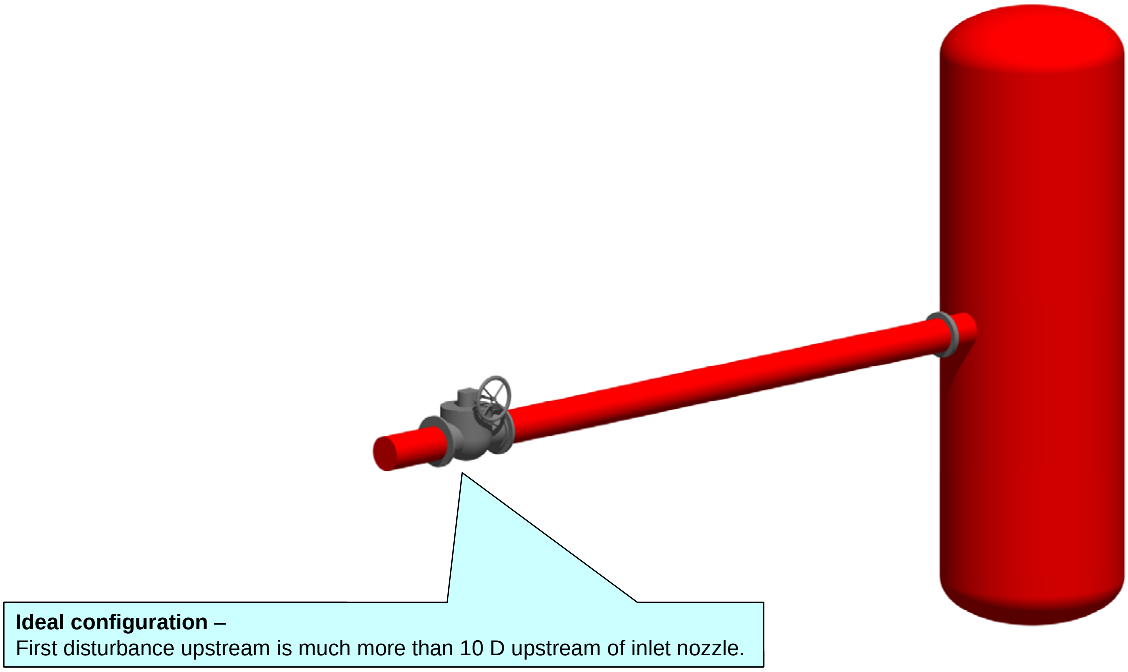


Figure 4.1b Piping upstream of a separator: Bend in horizontal plane

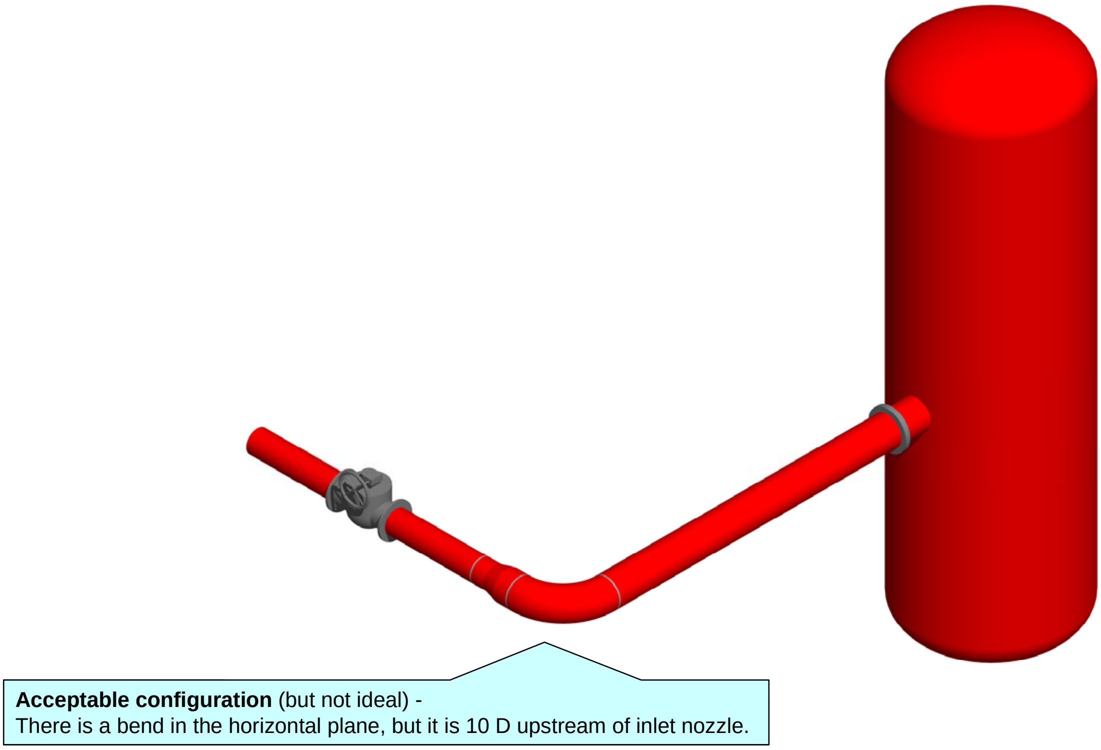


Figure 4.1c Piping upstream of a separator: Bend in horizontal plane

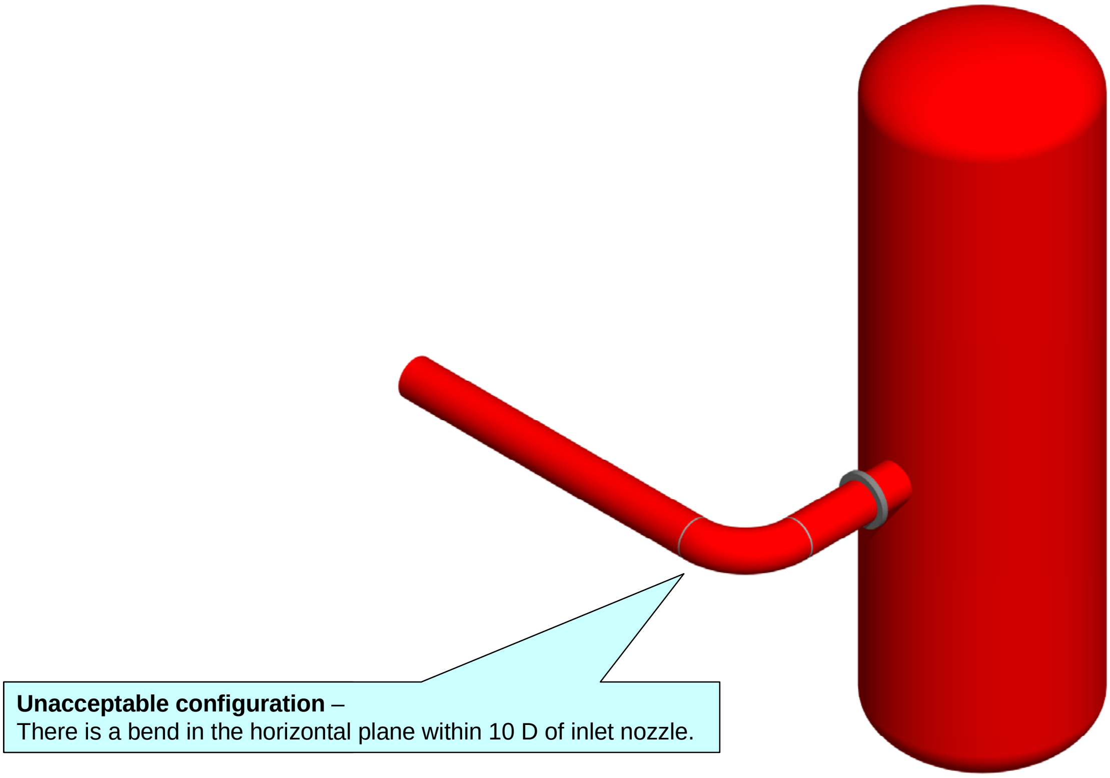


Figure 4.1d Piping upstream of a separator: Bend in vertical plane

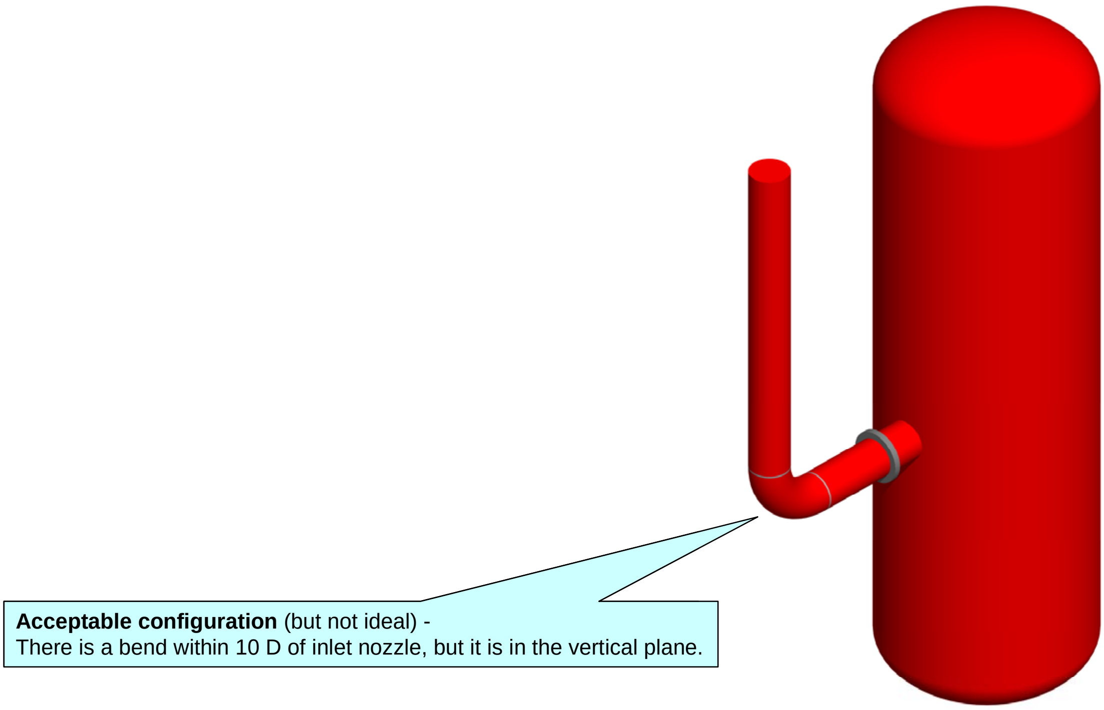


Figure 4.1e Piping upstream of a separator: Two bends in vertical plane

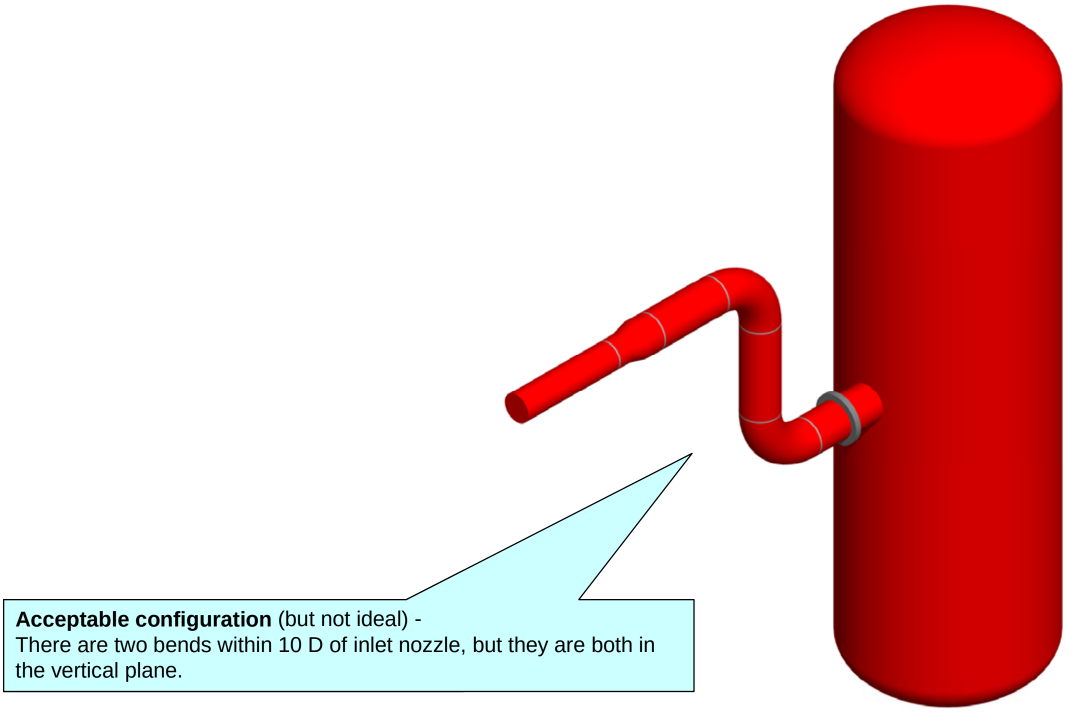
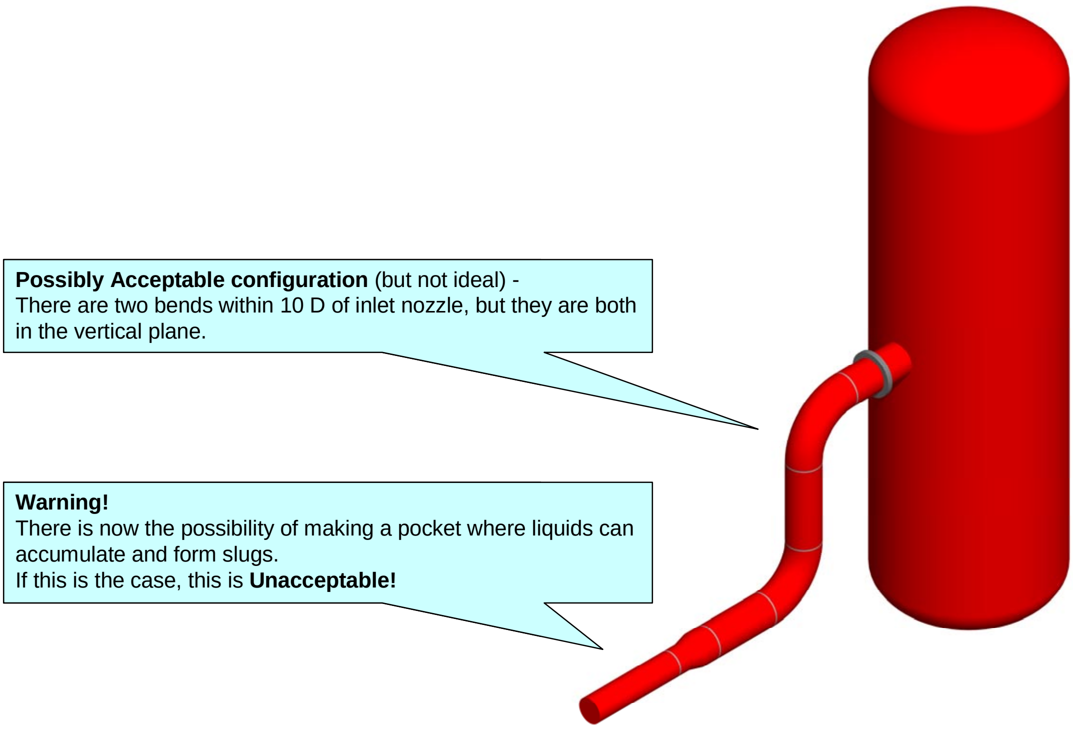


Figure 4.1f Piping upstream of a separator: Two bends in vertical plane



5. REFERENCES

In this DEP, reference is made to the following publications:

NOTES: 1. Unless specifically designated by date, the latest edition of each publication shall be used, together with any amendments/supplements/revisions thereto.

Amended per Circular 14/08

2. The DEPs and most referenced external standards are available to Shell staff on the SWW (Shell Wide Web) at http://sww.shell.com/standards/.

Amended per Circular 14/08

SHELL STANDARDS

Internals for columns	DEP 31.20.20.31-Gen.
Liquid/liquid and gas/liquid/liquid (three-phase) separators – Type selection and design rules	DEP 31.22.05.12-Gen.
Design of pressure relief, flare and vent systems	DEP 80.45.10.10-Gen.

DATA/REQUISITION SHEETS

Requsition sheet – Wiremesh demister	DEP 31.22.05.93-Gen.
Data/Requisition sheet – Schoepentoeter (vane-type) inlet device	DEP 31.20.41.93-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

Amended per Circular 14/08

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 Figures 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, 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:

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

– liquid Froude number:

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

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.

$$v_G = Q_G / (\pi d^2_{fp} /4) \qquad [m/s]$$

$$v_L = Q_L / (\pi d^2_{fp} /4) \qquad [m/s]$$

and the averaged liquid density ρ_L is defined as:

$$\rho_L = M_L/Q_L \qquad [kg/m^3]$$

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:

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

The smallest drops will generally have a diameter five to ten times smaller than $d_{p,max}$. 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).

At high pressures and in case of low surface tension fluids, fine mists can prevail which require additional measures. Further information can be obtained via the Principal.

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-under of gas or to carry-over of liquid. The latter will happen in particular when foam reaches the gas/liquid separation internal and/or the gas outlet, but even before this happens carry-over will increase due to the acceleration of the gas flow and due to re-entrainment of mist formed by bursting bubbles. Foaming 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 mm to 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 pre-filter.

Figure I.1 Two-phase flow map for horizontal feed pipes

Conditions:

$\rho_G = 8\ kg/m^3$ $\rho_L = 860\ kg/m^3$ $\sigma = 0.03\ N/m$

$\eta_G = 1.2*10^{-5}\ Pa.s$ $\eta_L = 1.6*10^{-4}\ Pa.s$ $d_{fp} = 0.50\ m$

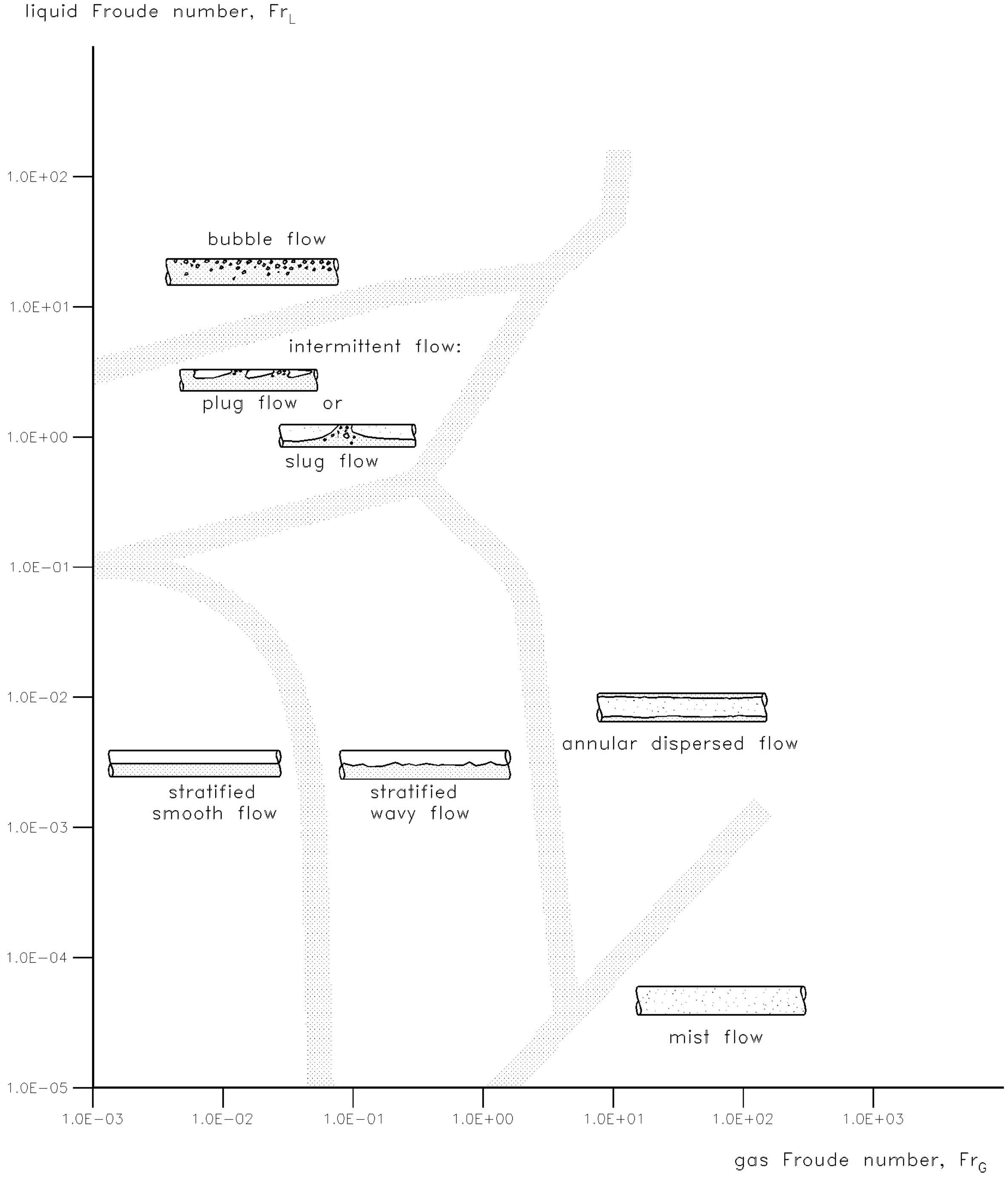
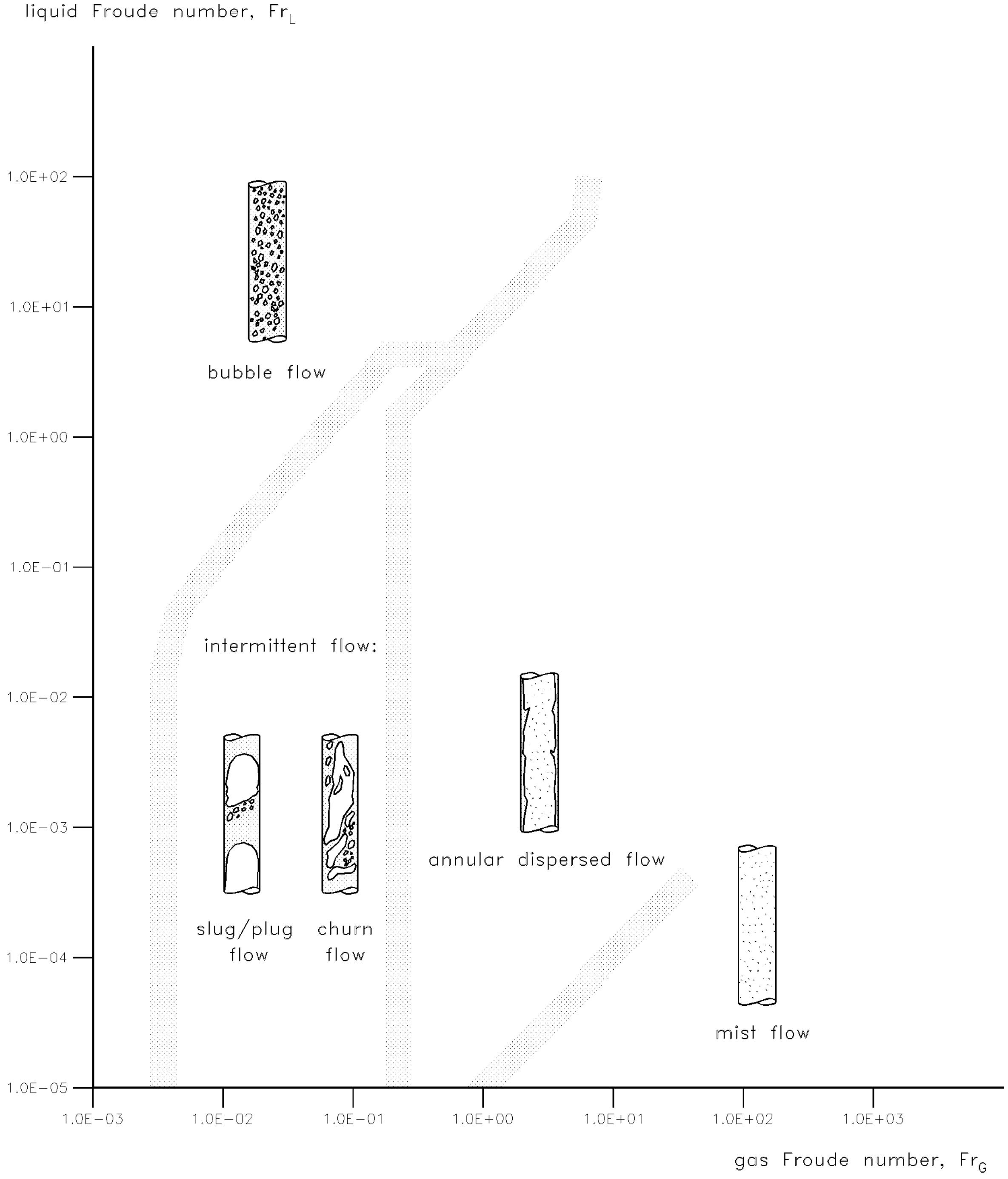


Figure I.2 Two-phase flow map for vertical feed pipes (upflow)

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 **maximum** flow rates, **including** the appropriate design margin.

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:

$$\rho_m v^2_{m,in} \leq 1\,400 \qquad [\text{Pa}]$$

where:

ρ_m is the mean density of the mixture in the feed pipe

$$\rho_m = (M_G + M_L)/(Q_G + Q_L) \qquad [\text{kg/m}^3]$$

and $v_{m,in}$ is the velocity of the mixture in the inlet nozzle

$$v_{m,in} = (Q_G + Q_L)/(\pi\, d_1^2 / 4) \qquad [\text{m/s}]$$

– If a **half-open pipe** is used as inlet device:

$$\rho_m v^2_{m,in} \leq 2100 \qquad [\text{Pa}]$$

– If a **Schoepentoeter** is used as inlet device:

$$\rho_m v^2_{m,in} \leq 8000 \qquad [\text{Pa}]$$

The pressure drop across the Schoepentoeter approximates to:

$$\Delta p_{sch} = 0.08 * \rho_m v^2_{m,in} \qquad [\text{Pa}]$$

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. The following velocity limits shall also be satisfied:

- To prevent erosion:

$$v_{G,in} \leq 70 \qquad [\text{m/s}]$$

- To prevent choking or damage due to vibrations:

$$v_{G,in} \leq 0.8 v_{sonic,G}$$

where $v_{sonic,G}$ is the sonic velocity if only gas is present (presence of liquid ignored)

$$v_{sonic,G} = \sqrt{\frac{\kappa R T}{MW_G}} \qquad [\text{m/s}]$$

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 substantial liquid is indeed present.

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:

$$\rho_G v^2_{G,out} \leq 4500 \qquad [Pa]$$

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.

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.

APPENDIX III DESIGN OF SCHOEPENTOETER (VANE-TYPE) INLET DEVICE

Amended per Circular 14/08

This remaining contents of this appendix have been deleted. For the design of Schoepentoeters, see DEP 31.20.20.31-Gen.

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 shall be supplied by the Principal.

Typical values are:

IN EXPLORATION AND PRODUCTION APPLICATIONS:

1.	**Offshore service**	**Design margin**
	Separator handling natural-flowing production from:	
	a) Direct Vertical Access (DVA) wells on their 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) DVA wells on their own platform	1.4
	b) Wells on another platform, or well jacket	1.5
	c) Subsea wells	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 REFINERIES AND CHEMICAL PLANTS:

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.

The level gauge (LG) span shall extend from 0.15 m above the vessel bottom (horizontal vessel) or 0.15 m above BTL (vertical vessel). For a horizontal vessel this may require special provisions. The level trip (LZA) range should preferably extend from 10 % to 90 % of the LG range, the level control (LCA) range should preferably extend from 20 % to 80 % of the LG range.

The following minimum spacings are recommended:

- **LZA(LL)** (low level trip) should be at least 0.1 m above the level of the lower LG nozzle.
- **LA(L)** (low level pre-alarm) should be either at least 0.10 m above LZA(LL) or, if required, should be located so that there is sufficient liquid hold-up time between the two levels for operator intervention (to be specified, but typically 1 min to 2 min for action in the control room and 5 min for action outside the control room).
- **LA(H)** (high level pre-alarm). The minimum distance between LA(H) and LA(L) shall be 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. The following generic control times should be met:
  - 3 min on total product outflow for separator trains;
  - 5 min on total product outflow for columns to furnaces or other columns;
  - 10 min on total product outflow for feed surge drums.

  The Normal Level (NL) should be positioned so that the control time is equally distributed between LA(L) and NL and between NL and LA(H); this ensures optimum operation, also in combination with SSVC.

  If slugs are expected, they should be accommodated between NL and LA(H).

  If the volume of the slug to be expected is not known, the volume should be taken as 2 s to 5 s of flow with the maximum feed (gas + liquid) velocity and 100 % liquid filling of the pipe.

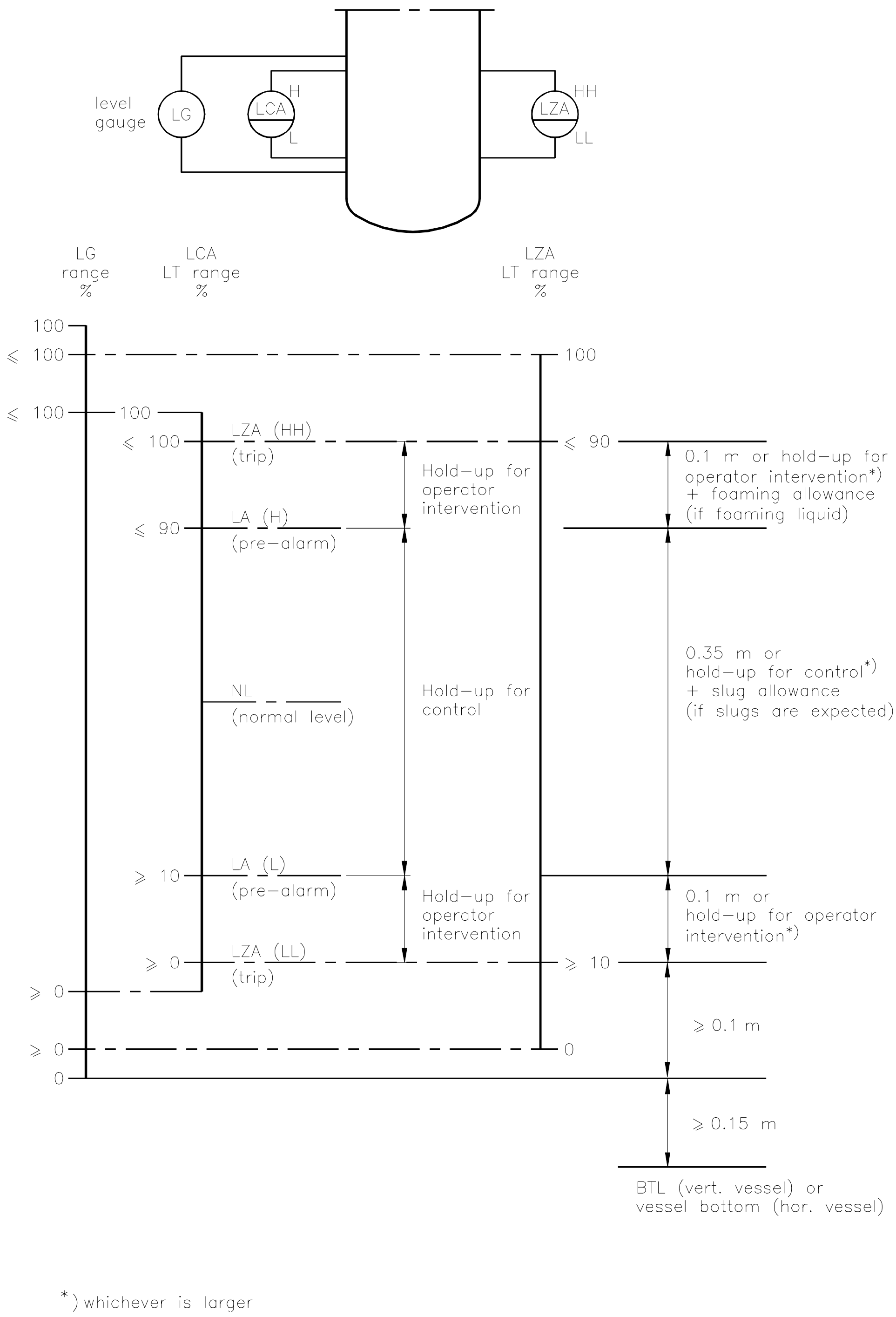
  In general, the slug size is influenced by the layout of the upstream piping. If in doubt, the Principal should be consulted for an estimate of the slug size.

  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 so that there is sufficient liquid hold-up time between the two levels for operator intervention (to be specified, but typically 1 min to 2 min for action in the control room and 5 min for action outside the control room).

  If the liquid has a foaming tendency, the distance between LA(H) and LZA(HH) shall 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. If operator action to prevent a trip is not realistically possible, the pre-alarm should be discarded.

Figure V.1 Liquid level control in a gas/liquid separator



NOTES
1. 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.
2. The control times for operator invention recommended in this Appendix are the result of a compromise between vessel size and operational robustness. If weight and space limitations are governing, e.g. in the case of offshore applications, preference may be given to smaller holdups at the cost of shorter control times.
3. The fail-safe direction is different for the high and low level trip, which is usually a reason to split up the high and low level trip range and use two transmitters. If there is only one trip range it is preferred to make this identical to the control range. This makes it easier to compare measurements, and allows dual use in the event of instrument failure.

APPENDIX VI VESSEL GEOMETRICAL RELATIONSHIPS

CROSS SECTION: 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.

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 as well.

In this Appendix the relationships will be given which enables this type of calculations, both graphically and via equations.

A cross-section of a horizontal G/L separator with diameter D is schematically shown in Figure VI.1.

The positions of the liquid and gas phases 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.

$A^*_L = A_L /(\pi D^2/4)$ $A^*_G = A_G /(\pi D^2/4)$

$(A^*_G + A^*_L = 1)$ $h^*_1 = h_1/D$

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, taking into account the vessel diameter D, the relationship between the cross-sectional areas of the various phases and their boundaries in the vessel is obtained.

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)

$A^* = 0.5*(\varphi - \sin\varphi)/\pi$

$h^* = 0.5*\{1 - \cos(\varphi/2)\}$

- **$h^* \rightarrow A^*$**

  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 $A^* = 0.5*(\varphi - \sin\varphi)/\pi$.

- **$A^* \rightarrow h^*$**

  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 boundaries, 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: α =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

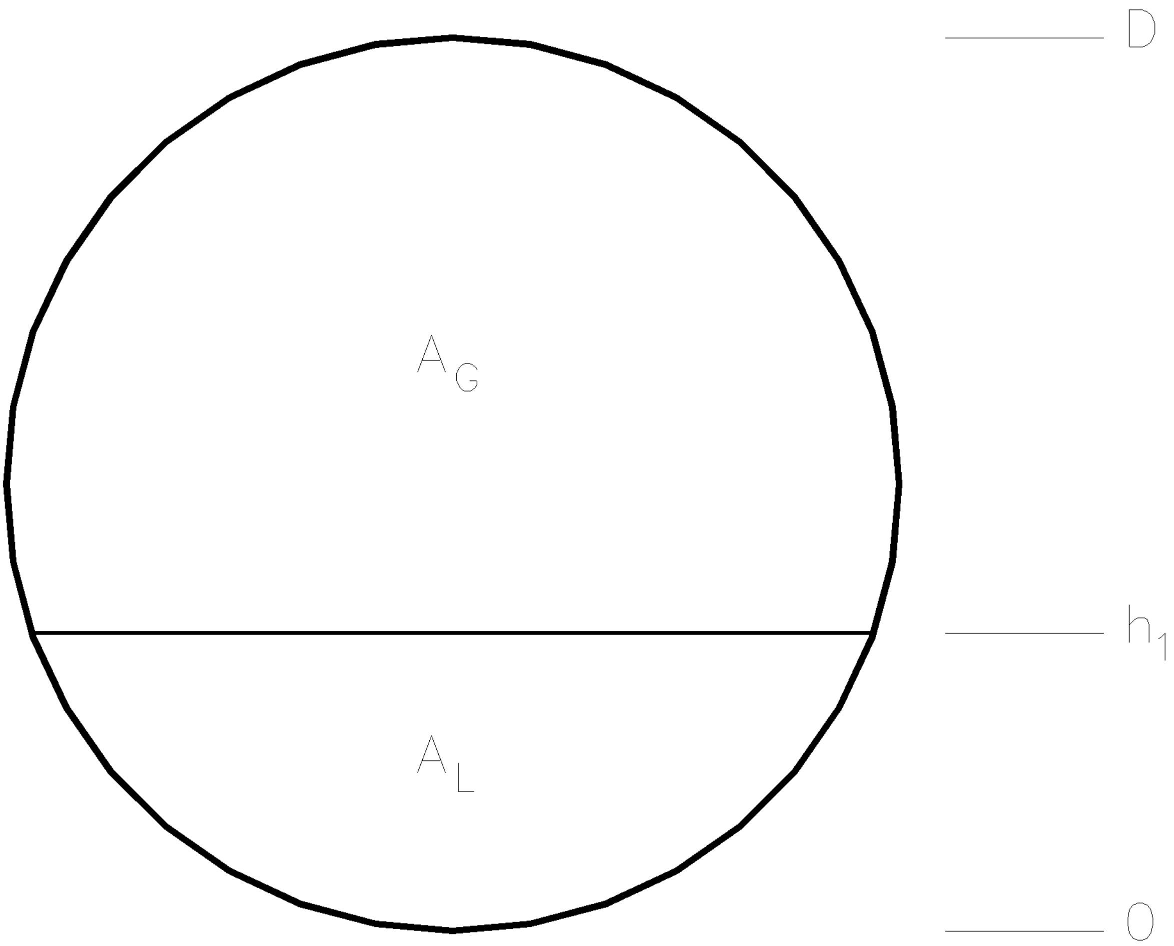
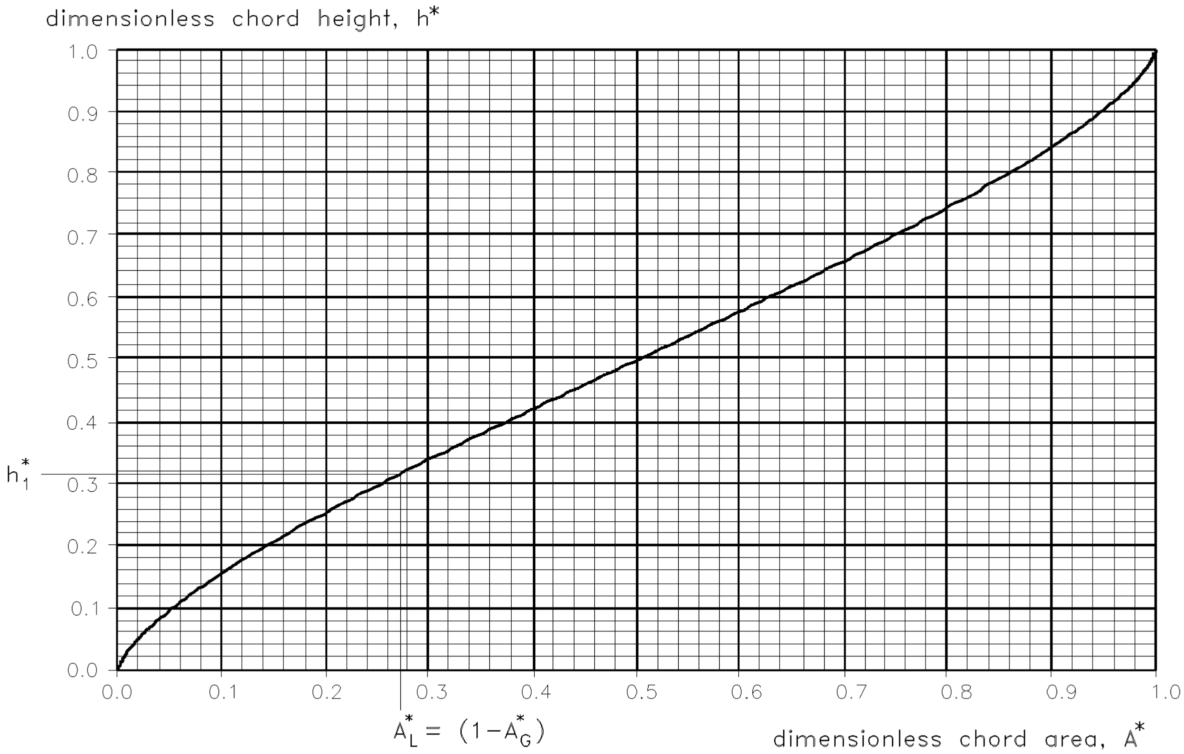
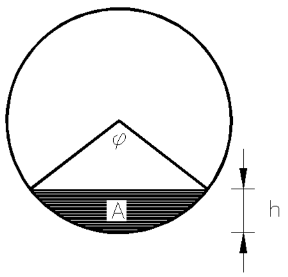


Figure VI.2 Relationship between the dimensionless chord area, A*, and dimensionless chord length, H*, of a vessel cross-section





$$A^* = A/A_{ves} = 0.5*(\varphi - \sin\varphi)/\pi$$

$$h^* = h/D = 0.5*\{1-\cos(\varphi/2)\}$$

$\varphi = 2 \text{ acos } (1\text{-}2h/D)$

APPENDIX VII SIZING OF SEPARATOR VESSELS

VII.1 VERTICAL VESSEL

To limit *liquid carry-over,* 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:

$$v_L \leq Q_{L,max}/(\pi D^2/4) = 2.2 * 10^{-8}(\rho_L - \rho_G)/\eta_L \qquad [m/s]$$

or

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

When the liquid has a **foaming** tendency the downward velocity of the liquid has to be adapted accordingly. A first check can be obtained by applying the **de-foaming criterion:**

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

Actually the degassing criterion, which is based on Stokes law, applies for low gas concentrations (< 5%), and the defoaming criterion should be considered in case of higher gas concentrations.

It is possible in the case of a viscous or foamy liquid that one of the latter equations will determine the vessel diameter. This is illustrated 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.

If the de-foaming criterion is limiting, refer to par. VII.3 for further considerations regarding foam handling.

VII.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 drain pipes 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 co-operation with the Manufacturer of the cyclones.

A sizing procedure for 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

$$L = (V_{slug} + Q_L t_{H\_L} - 2\Delta V_{hd,H\_L})/A_{H\_L} \qquad [m]$$

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:

- 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 this is not the case, increase D and return to step iii.
- 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):

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

  If this is an issue refer to par. VII.3 for further considerations regarding foam handling.

VII.3 DEFOAMING CRITERION

The defoaming criterion is based on a correlation of Barber Wijn, which relates the height of the foam layer in a flash vessel to the superficial liquid velocity, the energy in the incoming feed stream and the physical properties of the liquid. Via several simplifications the gas hold up and the inlet velocity are eliminated and the criterion is reduced to a requirement for a minimum vessel diameter. This diameter corresponds to a velocity resulting in a foam height of 1 m. If this is not available it will be necessary to increase the vessel diameter.

If the liquid is not flashing, the defoaming criterion is too conservative, and the original Barber Wijn equation should be reverted to. For further guidance the Principal should be consulted.

Figure VII.1 Diameter of the vertical wiremesh demister – Effect of flow parameter and viscosity

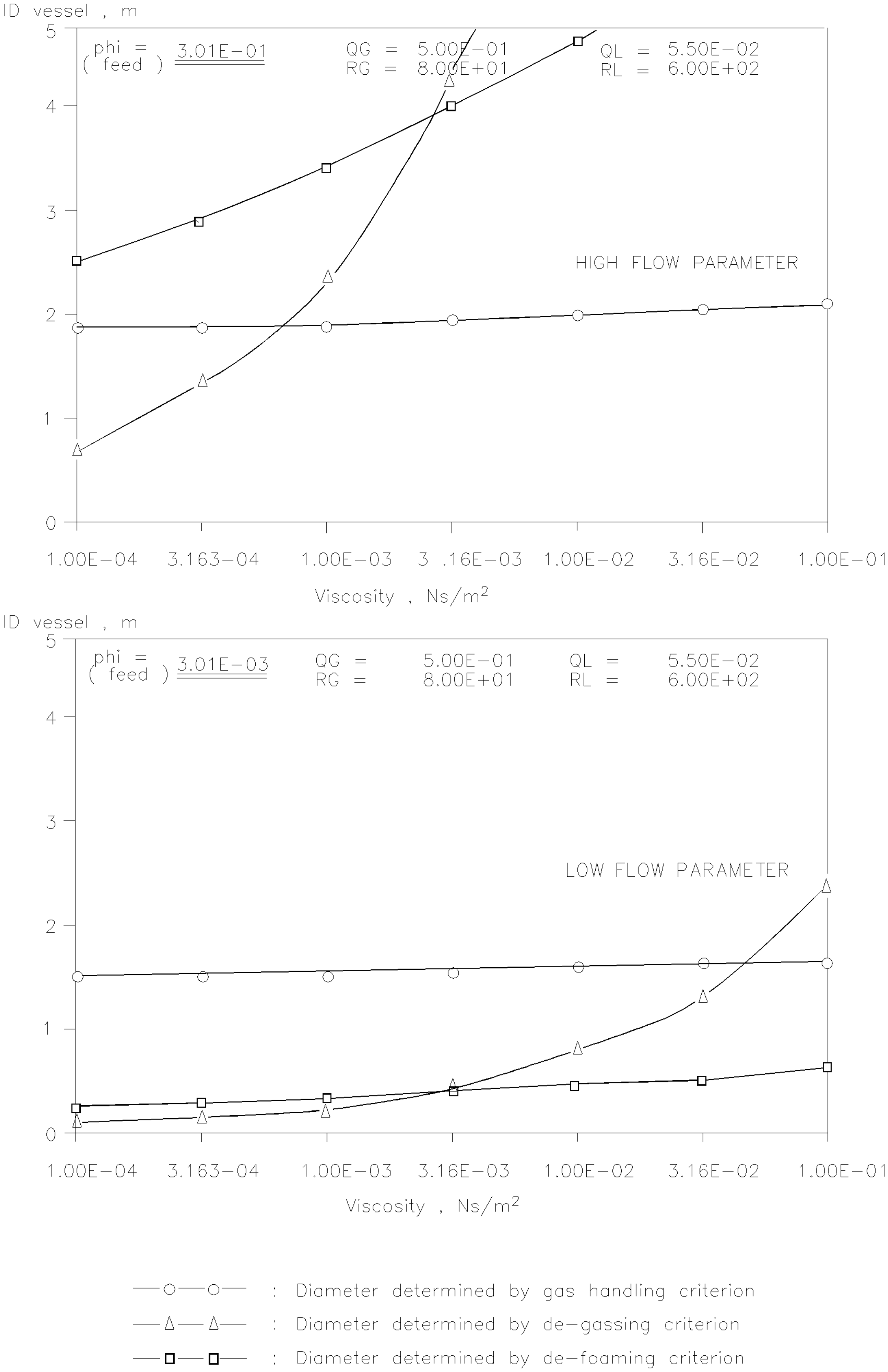


Figure VII.2: A_G (m^2) as a function of D and LZA(HH) in horizontal separators

	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	↑ NO
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	**HKO**
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	↓ YES
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	
														← YES	**HV**	→ NO		← YES	**HW**	→ NO

HKO : Horizontal knock-out drum

HW : Horizontal wiremesh demister

HV : Horizontal vane-type demister

APPENDIX VIII SEPARATION AT HIGH PRESSURE

One of the parameters determining the separation efficiency of the various separator types is the droplet size distribution of the liquid in the feed.

The movement of a droplet in a separator is determined by the resultant of the forces acting on it, on the one hand the drag force exerted by the gas, on the other hand the net gravity force. The cut-off droplet size, i.e. the diameter of a droplet that is just large enough not to be separated is related to the superficial gas velocity by the following equation:

$$\frac{\pi}{6} d_{co}{}^{3} (\rho_l - \rho_g) g = \frac{C_D \pi d_{co}{}^{2} \rho_g v_g{}^{2}}{8}$$

This can be rewritten as

$$\lambda = \sqrt{\frac{4 g d_{co}}{3 C_D}}$$

i.e. the gas load factor is proportional to $d_{co}{}^{0.5}$. The accompanying separation efficiency is equal to the cumulative volume fraction of droplets smaller than d_{co}.

In its turn the droplet size distribution depends on the physical properties of gas and liquid. The maximum stable droplet size is proportional to $(\sigma/\rho_g)^{0.6}$. The implication of this for a mixture of light hydrocarbons (C_6^-) is illustrated in Table VIII.1 below. It can be seen that the maximum droplet size decreases by almost an order of magnitude over the pressure range between 20 and 120 bar.

To maintain the same separation efficiency at elevated pressure the cutoff droplet size must decrease along with the maximum stable droplet size, which means that the gas load factor must be derated with a factor $(\sigma/\rho_g)^{0.3}$. The last column of Table VIII.1 gives the derated load factor for an open separator.

In aqueous systems and for heavier hydrocarbons the surface tension is less dependent on pressure and the effect on gas handling capacity will be less dramatic.

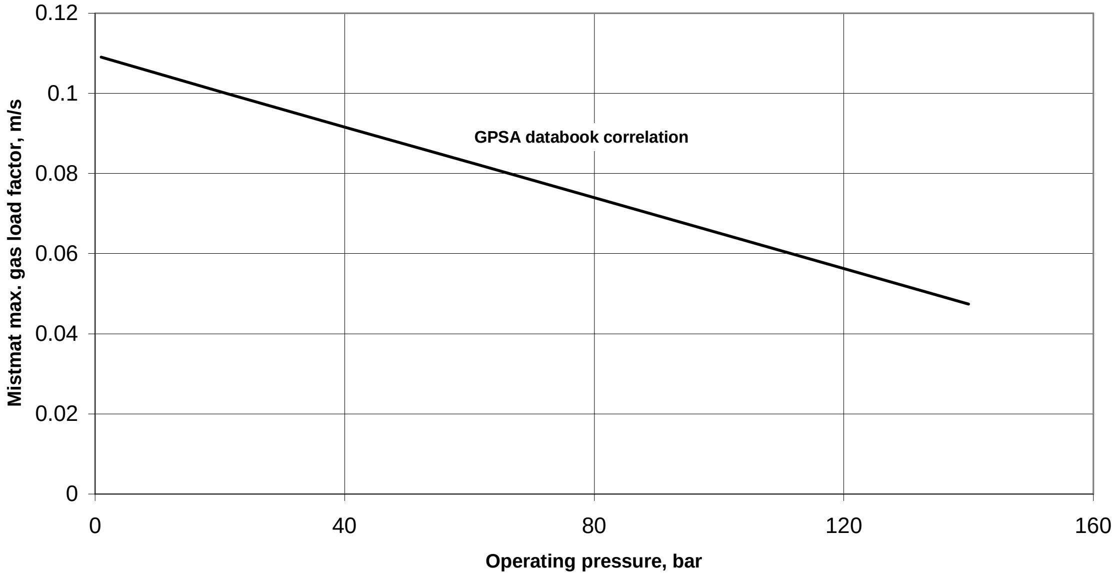
Table VIII.1 Variation of the physical properties of a C_6^- mixture, and its effect on (relative) maximum droplet size and gas handling capacity of an open separator

Pressure	Gas density	Surface tension	$(\sigma/\rho_g)^{0.6}$. (relative to properties at 20 bar)	λ
20	18.7	28.9	1.0	0.07
40	39.6	22.0	0.54	0.052
60	60.7	16.4	0.35	0.042
80	84.7	11.7	0.23	0.034
100	108.4	8.1	0.16	0.028

The GPSA databook gives a similar correlation for the derating of wiremesh demisters, which is plotted in Figure VIII.1

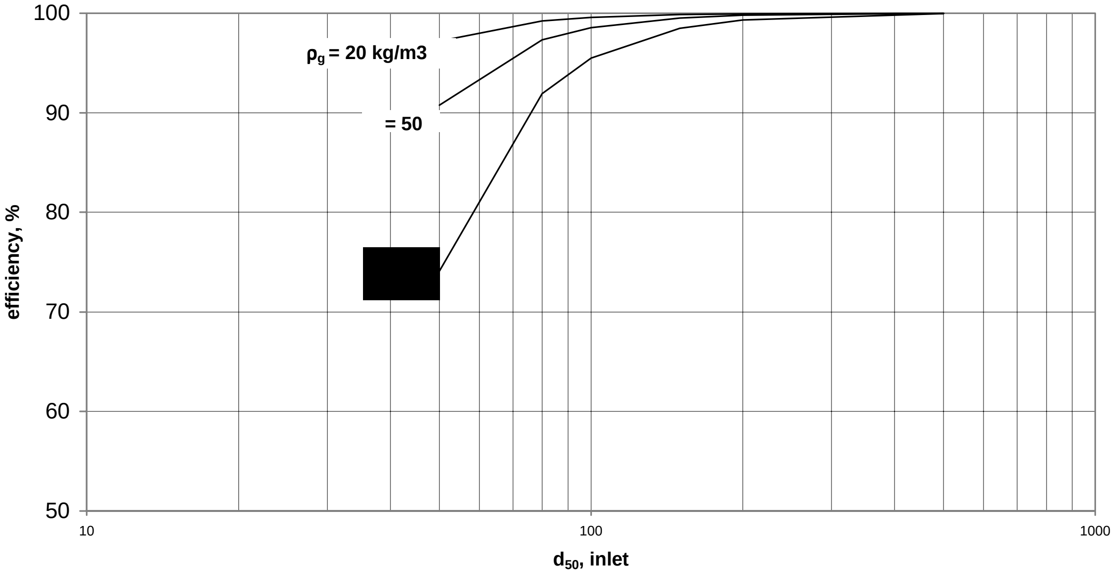
However, there is also a limit to this: at low load factors mistmats will loose efficiency for lack of impact: the droplets will not be able to hit the wires in this regime, cf. Figure 3.4. Elevated pressures require more sophisticated mistmat designs, e.g. co-knits of stainless steel and teflon wires, optimized to capture smaller droplets. Further guidance should be sought from the Principal.

Figure VIII.1 GPSA databook correlation for the maximum gas handling capacity of a wiremesh demister



Swirltube separators are also affected by high operating pressure, due to the higher gas density as well as the smaller average droplet size in the feed. The maximum tube loadfactor must be limited to < 0.6 m/s to minimize re-entrainment. Otherwise derating will be less opportune but a loss of efficiency is not always avoidable, cf. Figure VIII.2. Also here high performance primary and secondary mistmats will be required to retain maximum separation efficiency.

Figure VIII.2 Efficiency of Shell swirltube separators as function of gas density and droplet size



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 gas handling capacity.

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.). The BM/KCH hollow vane is a variation on a double-pocket vane which is more robust to fouling and suitable e.g. for liquid sulphur service.

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 robustness to fouling. In this application the low efficiency of the vane pack is less of a problem since the vane pack functions normally as a coalescer for the downstream swirldeck. Howver, the vanepack can not provide the same turndown as a coalescing mistmat.

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 0.20 m) 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 made of **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.

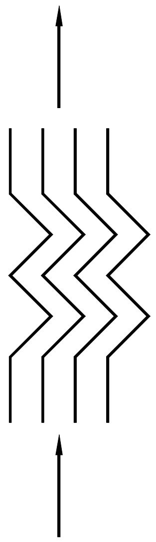
MOUNTING

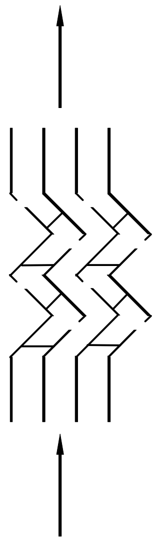
The vessel shall provide sufficient access for inspection, cleaning, maintenance and the removal and installation of the vane pack and other internals. The internals should be bolted on attachments welded to the vessel wall. The supplier of the vane pack shall provide detailed attachment drawings to 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.

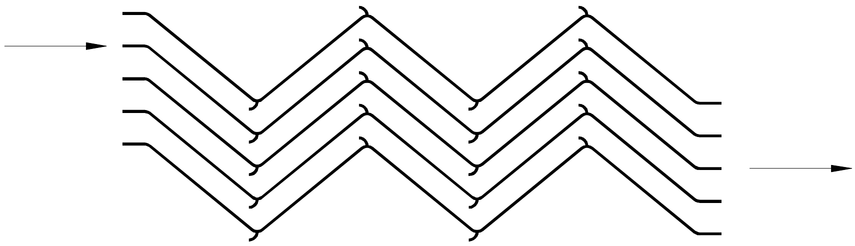
Figure IX.1 Examples of various types of vanes



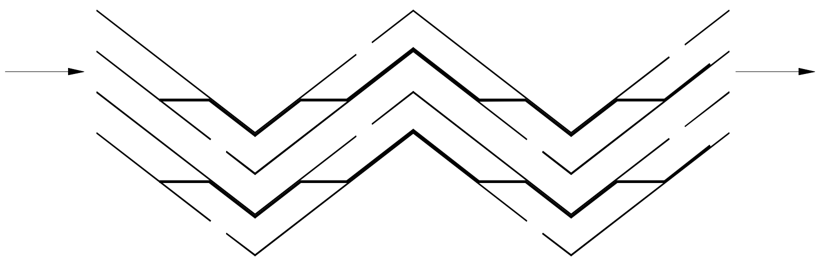


(a) straight vane without pockets; side view
used amongst others as coalescer in SVS separators

(a) hollow vane KCH 628; side view -
for use in vertical flow



(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 SLOSHING

More and more separators are installed on floaters and other installations subject to motion, e.g. Tension Leg Platforms. The performance of these separators can be adversely affected by the motion imposed by waves and wind. The accompanying sloshing will compromise liquid handling capacity, separation efficiency and the functioning of level instrumentation.

A common way to mitigate the negative impact of sloshing is to install perforated baffles to limit the liquid motion and damp the waves. However, the layout of these baffles and the selection of the net free area is critical for their success. With decreasing NFA the damping will become more effective, but in the case of oil/water mixtures the increasing fluid velocity through the holes may lead to redispersion.

Guidelines for the design of separators which are subject to motion are being developed and will be listed in Appendix X in future. For the present, the Principal shall be consulted for the further advice.

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 may be sought from the Principal.

FEED INLET

In refineries, in particular in hydrocracker plants, still many wiremesh demisters are not equipped with an inlet device. The accompanying gas flow maldistribution (‘jetting’), will lead to local overload of the mistmat. This problem can successfully be resolved by retrofitting a schoepentoeter inlet device.

If the inlet nozzle is overloaded, even with a schoepentoeter, it is recommended to install liquid collector baffles underneath the schoepentoeter. These are flat plates with an inclination of 10 degrees to the horizontal, spanning the entire cross section of the separator. These shield the sump from the gas flow in the vessel and provide surface to capture and coalesce small liquid droplets. In this way re-entrainment is suppressed.

DE-GASSING AND DEFOAMING

In separators with a relatively high liquid load gas carry-under is often caused by entrainment of gas with the liquid plunging down (‘waterfall’ effect). In this case it can be considered to install a liquid return tray or a liquid collector tray.

If an existing vessel cannot meet the de-gassing requirement (Appendix VI) due to a high liquid viscosity, the installation of a plate pack to remove the gas bubbles should be considered.

In the case of a foaming feed the existing feed inlet can be retrofitted with a cyclone type inlet device. Manufacturer-proprietary inlet devices consisting of an array of vortex tubes or cyclones are available.

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.
2. 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 option is recommended, but if insufficient space is available for the swirldeck or the resulting relatively high pressure drop of the deck is not acceptable, use option 2.

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 side outlet is applied, the Principal should be consulted.