



Training Services

Pressure Vessel Design

IPD-2004/PVD-1

Pressure Vessel Design

- Code Requirements
- Process Design Requirements
- Configuration and Sizing
- Mechanical Design Features

IPD-2004/PVD-2

Pressure Vessel Design ***(continued)***

Pressure vessel design is a two-step process

- Process engineer establishes process requirements
- Pressure vessel designer applies code and fabrication requirements

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Pressure Vessel Design ***(continued)***

- An equipment specification is usually used to transmit the process requirements
 - Design pressure and temperature
 - Diameter and length
 - Material of construction
 - Nozzle size and location
 - Details of internals

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Code Requirements

- Rules found in the codes represent hundreds of man-years of experience
- If used wisely, the code requirements can:
 - Communicate design requirements
 - Utilize know-how and technology
 - Keep equipment costs low
 - Reduce insurance costs
 - Reduce chance of legal entanglements

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Code Requirements ***(continued)***

ASME Code - Section 1

- Governs the design of vessels used for generating steam
- When steam is generated by process fluid (no flame), user may waive Section 1

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Code Requirements ***(continued)***

ASME Code - Section VIII, Division 1

- General requirement for the design of all unfired pressure vessels
- Does not apply to small vessels made of pipe and atmospheric vessels for water and injection chemicals

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Code Requirements ***(continued)***

ASME Code - Section VIII, Division 2

- For high pressure vessels
- Safety factor less than Division 1, 3 vs. 4
- Special requirements for material and fabrication quality control

Non-Code Vessels

- Code construction not required for atmospheric vessels handling water and injection chemicals

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Code Requirements (continued)

Related Codes

- API Standard 620 - Low Pressure Storage Tanks
- API Standard 650 - Welded Tanks Up to 0.5 psi
- ASME → ■ ANSI B31.3 - Petroleum Refining Piping
- ANSI B16.5 - Flanges, Valves, and Fittings
- TEMA - Heat Exchangers
- Local Codes and Standards - If More Stringent

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Process Design Requirements *Pressure*

- Normal Operating
 - Pressure at which equipment operates at design or specified conditions
- Maximum Operating
 - Highest operating pressure foreseen for specific equipment

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Process Design Requirements ***Pressure (continued)***

■ Design Pressure

- Maximum operating pressure plus a certain margin
- Normal margins are:

MAX. OPER. P, PSI	DESIGN P, PSI
Less than 25	50
25 to 250	Oper P + 25
250 to 1000	(Oper P) * (1.1)
1000 to 2000	(Oper P) + 100
More than 2000	(Oper P) * (1.05)

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Process Design Requirements ***Pressure (continued)***

■ Vacuum Design

- If necessary, vessels are designed for both internal and external pressure
- Equipment that operates under vacuum or are subject to vacuum during purging or on loss of reboiler heat may need to be specified for vacuum design
- Vacuum due to operator error normally not considered

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Process Design Requirements

Pressure (continued)

■ Effect of Pressure Drop on Design

- ASME considers design pressure at the top of the vessel
- Design conditions at the bottom should consider:
 - Liquid head
 - Upflow or downflow pressure drop
 - Hydrostatic test conditions

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Process Design Requirements

Temperature

■ Normal Operating

- Temperature at which equipment operates at design or specific condition

■ Design Temperature

- Normal operating temperature plus a margin or, if cryogenic application, could be minus a margin
- Normal margins are:

NORM. OPER. T, °F	DESIGN T, °F
Less than 200	250
More than 200	Oper T + 50

↑
28°C

121°C

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Process Design Requirements ***Temperature (continued)***

■ Minimum Design Metal Temperature (MDMT)

- Lowest value of:

- Normal operating temperature minus 25°F (4°C)
- 24 hour ambient temperature
- Auto-refrigeration temperature

■ Fractionators

- Design temperature is normally bottom T plus 50°F top to bottom
- For fractionators with large delta T's, the top to bottom design is graduated

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Process Design Requirements ***Temperature (continued)***

■ Cold Wall Design

- Internally insulated vessels allow the outside shell design T to be considerably lower than the operating T ← FCC Regen

■ Cooler Failure

- Failure of coolers upstream of equipment could require a greater increment than 50°F (28°C)

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Process Design Requirements ***Temperature (continued)***

- **Heat Exchanger Shells**
 - Design temperature of exchangers is based upon the higher of the inlet or outlet temperature
- **Flange Ratings**
 - In setting design temperature, the effect of flange rating should be considered to minimize costs

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Process Design Requirements ***Corrosion Allowance (CA)***

- Normal CA for carbon steel vessels is 1/8 inch (3-175 mm)
- Normal CA for alloy is 1/16 inch (1-5875 mm)
- When cladding or weld overlay is used to protect the base metal, the CA of the base can be nil
- When a corrosive aqueous phase is present, a layer of cement or Monel is used to protect the base metal

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Refinery Vessel Functions

- Storage
- Surge Volume
- Separation
- Reaction

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Surge Drum Orientation

- Vertical
 - Most economical orientation considering vessel cost
 - Level reacts more quickly to changes in flow than in a horizontal vessel
- Horizontal
 - Greater liquid surface area results in less change in liquid level during flow variations

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Examples of Uses

- Smooths liquid flow variations resulting from large distance pumped (e.g. from tank farm) or the type of pump (reciprocating, etc.)
- Allows operator reaction time in the event of loss of liquid feed to the unit
- Protects pump suctions from loss of liquid feed by allowing time for pump shutdown (e.g. separator liquid pumps)
- Absorbs temporary increases or “slugs” of liquid flow (e.g. short term steam injection upstream)

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Residence Time

- Major factor affecting cost and size of the vessel is residence time of the liquid phase
- Factors which determine residence time are:
 - Experience with the same services
 - Difficulty of phase separation
 - Effects of loss of level or high level on downstream equipment
 - Settling of two liquid phases
 - Properties causing foaming or emulsions
 - Source/destination of liquid phase
 - Presence of suspended solids

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Residence Time Guidelines

PROCESS SERVICE	RESIDENCE TIME
Vapor/Liquid Separation Horizontal vessel, liquid at centerline	5 minutes on liquid
Vapor/Liquid/Liquid Separation Horizontal vessel, liquid at centerline	10 minutes on liquid
Liquid Surge for Control High to low level	2-4 minutes
Critical Separator Vertical vessel	2 minutes from highest liquid level to inlet nozzle

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Residence Time Guidelines (continued)

PROCESS SERVICE	RESIDENCE TIME
Drop Leg (Boot) Controlled level	5-10 minutes
Drop Leg (Boot) Manually drained	600 minutes
Feed Surge Drum	15-30 minutes
Reflux Drums	2-5 minutes (50% full)
Product Receiver	5-20 minutes (50% full)

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Residence Time Two Liquid Phases

- $\rho_H - \rho_L > 14 \text{ lb/ft}^3$
5 Min
- $\rho_H - \rho_L < 7 \text{ lb/ft}^3$
30 Min
- Viscosity of lighter phase $> 1 \text{ Cp}$
30 Min

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Phase Separation Examples

PHASES	EXAMPLE
Vapor/Liquid	Separator Flash Drum
Liquid/Liquid	Caustic Settler Desalter
Vapor/Liquid/Liquid	Hydrotreater Separator Wet Overhead Receiver
Vapor/Solid	FCC Regenerator
Liquid/Solid	Slurry Separator

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Separation Methods

- Momentum
- Gravity Settling
- Coalescing

Separation using a combination of these methods is common.

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Momentum Separation

- Denser fluid changes direction more slowly than less dense fluid
- Often used for initial, or "bulk" separation

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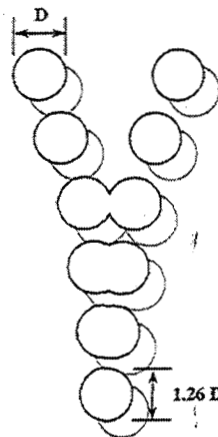
Gravity Separation

- For separation to occur, the fluids must be immiscible and of different densities
- The closer the fluid densities, the more difficult the separation
- No separation method is complete - the degree of desired separation must be defined

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Coalescing

- Increased size of droplets aids separation
- Process accelerated by coalescing medium



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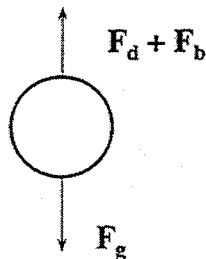
Gravity Settling Laws ***Particle Moving in a Fluid***

■ For Vessel Design

- Calculate droplet terminal velocity
- Size vessel to keep velocity of continuous phase less than particle terminal velocity to allow separation
- Size the vessel to permit the particle to escape from the continuous phase before the continuous phase (or the particle phase) is withdrawn from the vessel

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Gravity Settling Laws ***Particle Settling in a Fluid***



F_d = Drag

F_b = Buoyant Force

F_g = Gravitational Force

F = Resultant Force

a = Particle Acceleration

u = Particle Velocity

$$F = F_g - F_d - F_b = ma = m \frac{du}{dt}$$

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Gravity Settling Laws ***Particle Moving in a Fluid***

- At Steady State

$$\frac{du}{dt} = 0$$

and $u = u_t$ = terminal velocity

$$\text{So } F_g = F_d + F_b$$

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Gravity Settling Laws ***Particle Moving in a Fluid***

- Where: $F_g = \frac{mg}{g_c}$

$$F_b = \frac{m\rho g}{\rho_p g_c}$$

$$F_d = \frac{C_D u_t^2 \rho A_p}{2g_c}$$

$$A_p = \frac{\pi D_p^2}{4}$$

$$m = \frac{1}{6} \pi \rho_p D_p^3$$

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Gravity Settling Laws *Particle Moving in a Fluid*

- m = Mass of Particle
- g = Gravitational Acceleration
- g_c = Gravitation Constant
- ρ = Density of Continuous Fluid
- ρ_p = Density of Particle
- A_p = Projected Area of Particle
- D_p = Diameter of Particle
- C_D = Drag Coefficient (based upon velocity relative to continuous phase)
- μ = Viscosity of Continuous Fluid
- u_t = Terminal Velocity

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Gravity Settling Laws *Particle Moving in a Fluid*

Substitute and solve for u_t :

$$u_t = \left[\frac{4D_p g (\rho_p - \rho)}{3C_D \rho} \right]^{0.5}$$

For Heavy Droplet -
Light Continuous Phase

For Light Droplet -
Heavy Continuous Phase,
use $\rho - \rho_p$

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Gravity Settling Laws Particle Moving in a Fluid

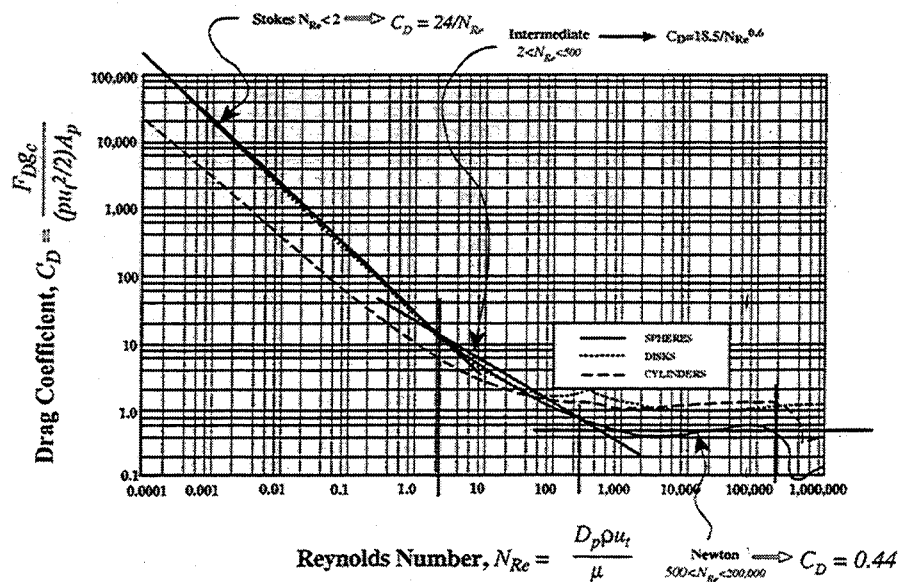
■ Limitations

- Spherical particle
- Stagnant fluid
- No interaction with other particles
- No wall effects

- The terminal velocity for a particle can now be calculated if C_D can be quantified

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Drag Coefficient Plot



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SD-R03-02

Newton's Equation

■ For: $500 < N_{Re} < 200,000$

■ Where: Particle Reynolds Number $N_{Re} = \frac{D_p \rho u_t}{\mu}$

■ Then: $C_D = 0.44$

■ And: $u_t = 1.75 \left[\frac{g D_p (\rho_p - \rho)}{\rho} \right]^{0.5}$

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Vapor Phase Separation *Newton's Equation*

■ In range of Newton's Equation, $N_{Re} > 500$ liquid droplets falling in vapor will deform

Then: $C_D \neq 0.44$

Let: $\rho_p = \rho_L$
 $\rho = \rho_v$

For: $D_p = 175 \text{ micron}$

And: $C_D = 1$

$$u_t = \left[\frac{4 D_p g (\rho_L - \rho_v)}{3 C_D \rho_v} \right]^{0.5} = \left(\frac{4(0.175)(32.2)}{3(1)(304.8)} \right)^{0.5} \left(\frac{\rho_L - \rho_v}{\rho_v} \right)^{0.5}$$

$$= 0.157 \left[\frac{\rho_L - \rho_v}{\rho_v} \right]^{0.5}$$

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Mesh Blankets

- Higher terminal velocities are achieved by using coalescing devices

$$u_t = K \left(\frac{\rho_L - \rho_V}{\rho_V} \right)^{1/2}$$

VACUUM (INCHES OF HG ABSOLUTE)	K
Less than 1"	0.17
1"	0.17
5"	0.23
10"	0.28
20"	0.32
30"	0.35

PRESSURE	
PSIA	K
15	0.35
50	0.34
100	0.32
300	0.30
500	0.28
1000	0.27

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Stokes' Equation

- For: $N_{Re} < 2$

- Where: $N_{Re} = \frac{D_p \rho u_t}{\mu}$

- Then: $C_D = \frac{24}{N_{Re}}$

- And: $u_t = \frac{(\rho_P - \rho) g D_P^2}{18\mu}$

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Stoke's Equation Applications

- Separating vapor bubbles from liquid phase
- Separating heavy liquid drops from lighter liquid phase

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Intermediate Equation

- For: $2 < N_{Re} < 500$

- Where: $N_{Re} = \frac{D_p \rho u_t}{\mu}$

- Then: $C_D = \frac{18.5}{N_{Re}^{0.6}}$

- And: $u_t = \frac{0.153 g^{0.71} D_p^{1.14} (\rho_p - \rho)^{0.71}}{\rho^{0.29} \mu^{0.43}}$

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Close Fluid Densities

The closer the fluid densities are, the more difficult they are to separate. As an example, find the terminal velocity for separating water from two fluids - one gas, one liquid. Apply Stokes equation.

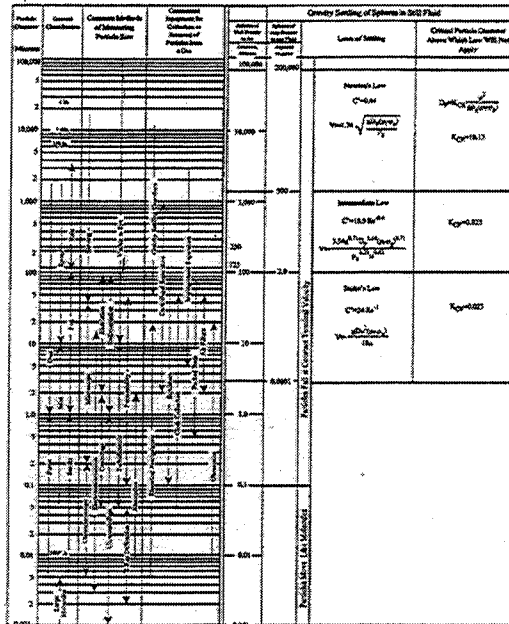
	CASE A	CASE B	
Droplet diameter	175	175	microns (0.00057 ft)
Water density	62	62	lb/ft ³
Continuous density	0.5	45	lb/ft ³
Viscosity	0.5	0.5	Cp (lb/ft · sec = Cp/1488)
g	32.2	32.2	ft/sec ²
u _t	0.108	0.030	ft/sec

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Selection of the Dispersed Particle Size to be Separated

- Consider effect upon downstream equipment, effluent quality, overall economics
- Consider difficulty and cost of separating small(er) particles
- Dispersed particles vary in size but generally follow a bell, or normal curve - use of an "average" size works well
- Optimal particle size is difficult to determine

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Gravity Setting Laws and Particle Characteristics

SD-P01-08
IPD-2004/PVD-47

Dispersed Particle Sizes for Separation

■ Default particle sizes used for gravity separation are:

- Liquid from vapor separation: 175 microns (0.00057 feet)
- Vapor from liquid separation: 175 microns (0.00057 feet)
- Dispersed liquid from continuous liquid: 125 microns (0.00041 feet)

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Use for Design

■ Vertical Vessel

- Maximum downward velocity = 0.075 fps or 4.5 fpm

■ Horizontal Vessel

- Upward gas bubble velocity = 0.075 fps or 4.5 fpm
- Settling water droplet velocity = 0.035 fps or 2.1 fpm

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Vessel Configuration: Separators

Process Conditions

High Vapor Rate
Low Liquid Rate
One Liquid Phase

Low Vapor Rate
High Liquid Rate
One Liquid Phase

Vessel Configuration

Vertical Vessel

Horizontal Vessel
(Vertical vessel may
be considered)

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Vessel Configuration: Separators

Process Conditions

**No Vapor Phase
Two Liquid Phases**

**Vapor Phase Present
Two Liquid Phases
Low Heavy Liquid
Phase Rate**

Vessel Configuration

Horizontal Vessel

**Horizontal with Boot
(Drop leg)**

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Vessel Configuration: Separators

Process Conditions

**Vapor Phase Present
Two Liquid Phases
Low Light Liquid
Phase Rate**

**Vapor Phase Present
Two Liquid Phases
Liquid Phases Equal
Rates**

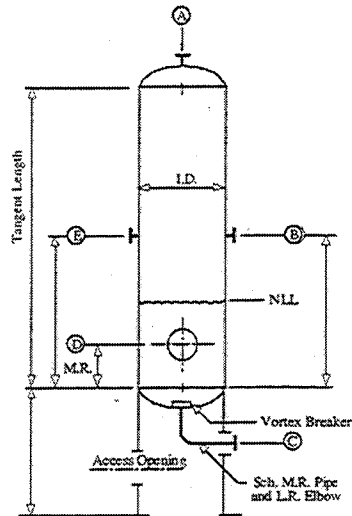
Vessel Configuration

**Horizontal with
Baffled Outlet**

**Horizontal with 2
Baffled Outlets**

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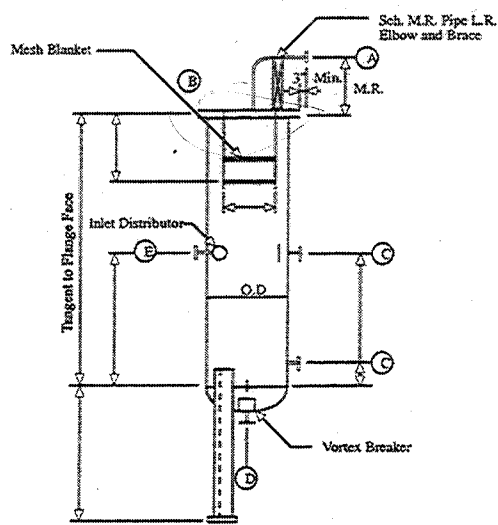
Vapor/Liquid Separator



Example: Separator, Flash Drum

EDS-R00-3443
IPD-2004/PVD-53

Small Vertical Separators

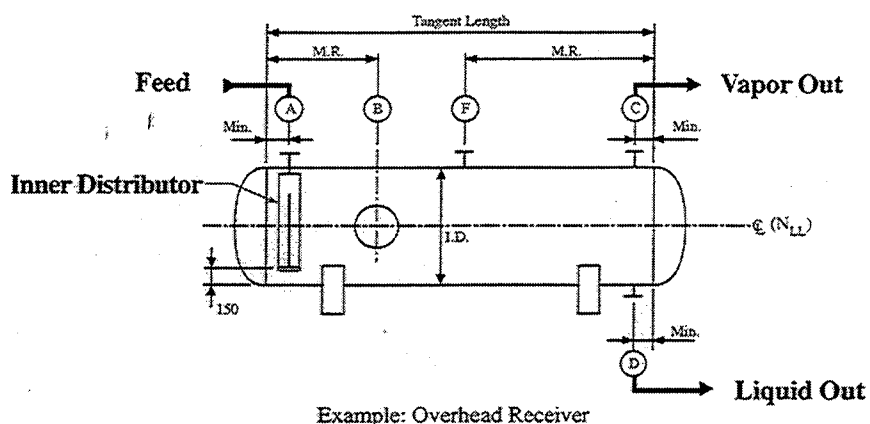


ID of 30 inch or less
Elevation of 5 feet or less

Example: Reciprocating Compressor Suction Drum

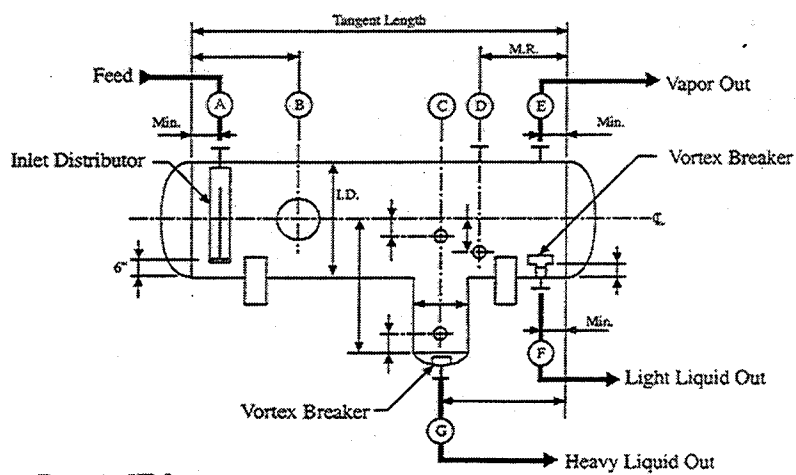
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Types of Vessels Vapor/Liquid Separator



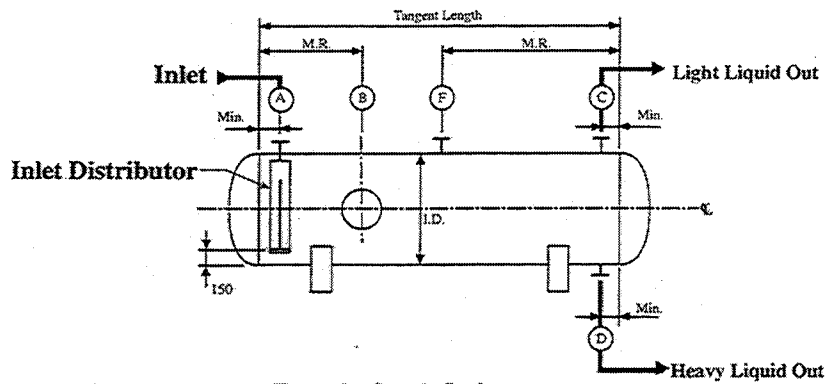
SD-R02-05
IPD-2004/PVD-55

Types of Vessels Vapor/Liquid/Liquid Separator Low Heavy Liquid Rate



SD-R01-07
IPD-2004/PVD-56

Types of Vessels *Liquid/Liquid Separator*

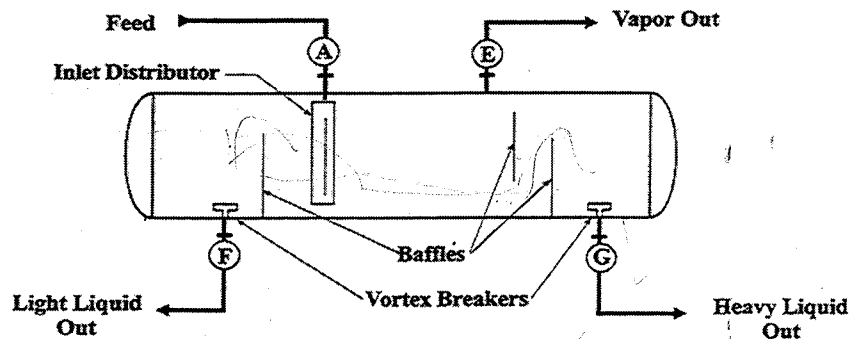


Example: Caustic Settler

Note: Use side inlet and horizontal distributor if liquid volume flows are approximately equal.

SD-R03-05
IPD-2004/PVD-57

Types of Vessels *Vapor/Liquid/Liquid Separator* *Equivalent Light and Heavy Liquid Rates*



SD-R00-44
IPD-2004/PVD-58

Horizontal Separator Sizing Example

■ Problem

- Design a reactor products separator for a naphtha hydrotreating process unit
- The relevant portion of the heat and material balance is reproduced below; then complete the vessel sketch

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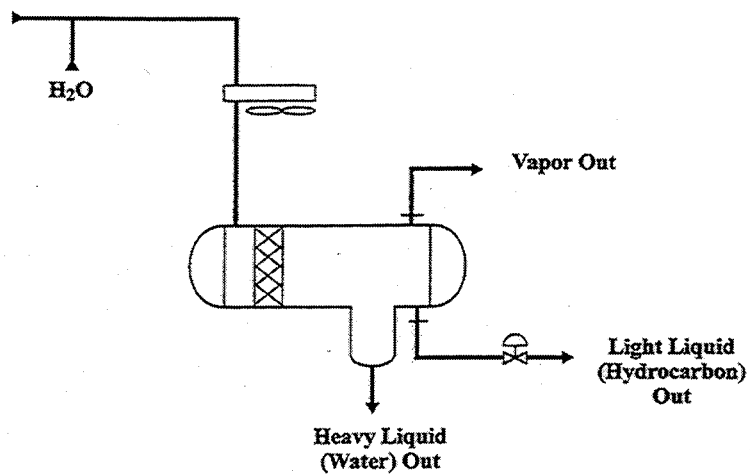
Horizontal Vessel Sizing Example

Tabulate data and calculate volumetric flows

	VAPOR	HYDROCARBON LIQUID	WATER LIQUID	TOTAL LIQUID
Flow, lb/h	693	58586	2250	60836
, ft ³ /s	1.54	0.35	0.01	0.36
, gal/min	—	158	5	163
Density, lb/ft ³	0.125	46.34	61.55	
Viscosity, c _p	0.01	0.42	0.51	
, lb/ft sec (x 10 ⁻⁴)	0.06	2.82	3.42	

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Flow Schematic



SD-R01-10
IPD-2004/PVD-61

Horizontal Vessel Sizing Example

Sizing Steps

- Select L/D
- Determine Liquid Residence Time
- Determine Normal Liquid Level
- Calculate Diameter to Satisfy Residence Time

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Horizontal Vessel Sizing Example

Sizing Steps

- Calculate diameter for each phase separation
 - Liquid from vapor
 - Vapor from liquid
 - Liquid from liquid
- If a liquid coalescing blanket is used, calculate diameter for 3 ft/m liquid velocity
- Select governing (largest) diameter
- Round up to nearest multiple of 6" or 100mm

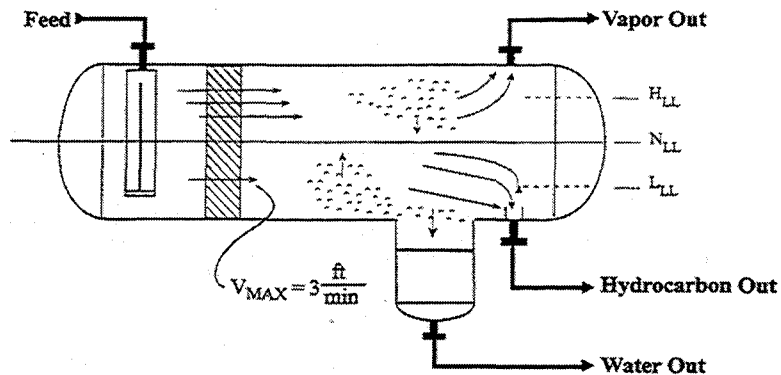
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Liquid Coalescing Blanket

- Coalescing blankets are often used to accelerate the separation of two liquid phases; because of the often close densities, gravity separation only may result in a large vessel
- When removing water, blankets are always used by UOP
- Liquid velocities over 3 feet per minute may re-entrain droplets from the blanket

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Vessel Configuration



SD-R02-12
IPD-2004/PVD-66

Horizontal Vessel Sizing Example

- Calculate the diameter based on residence time .
 - Select $L/D = 3$ (the most common, economical value)
 - For settling water phase from light hydrocarbon phase, use 10 minute residence time
 - Set normal liquid level (N_{LL}) at centerline

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Horizontal Vessel Sizing Example

- Calculate liquid volume

$$V_L = Q_L t = \left(\frac{0.36 \text{ ft}^3}{\text{s}} \right) \left(\frac{60 \text{ s}}{\text{min}} \right) (10 \text{ min}) = 216 \text{ ft}^3$$

- If vessel is 1/2 full, then full vessel volume is:

$$V_V = \frac{V_L}{F} = \frac{216 \text{ ft}^3}{0.50} = 432 \text{ ft}^3$$

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Horizontal Vessel Sizing Example

- V_V is a cylinder with $L/D = 3$; ignoring the head volume:

$$V_V = \frac{\pi D^2 L}{4} = \frac{3\pi D^3}{4}$$

$$D = \left[\frac{4V_V}{3\pi} \right]^{1/3} = \left[\frac{4(432 \text{ ft}^3)}{3\pi} \right]^{1/3}$$

$$D = 5.7 \text{ feet}$$

Round up to 6 feet

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Check Vapor/Liquid Separation

- U_{Des} for droplet/vapor separation is 3 ft/s

$$\begin{aligned}U_{Des} &= 0.157 \left(\frac{\rho_L - \rho_V}{\rho_V} \right)^{0.5} \\&= 0.157 \left(\frac{46.34 - 0.125}{0.125} \right)^{0.5} \\&= 3.0 \text{ f/s}\end{aligned}$$

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Check Vapor/Liquid Separation (continued)

- Horizontal gas velocity is:

$$\begin{aligned}U_h &= Q/CSA \\&= \frac{1.54 \text{ ft}^3/\text{s}}{(1/2)(\pi) \left(\frac{6}{2} \right)^2} \\&= 0.108 \text{ f/s}\end{aligned}$$

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Check Vapor/Liquid Separation (continued)

- Distance needed for droplet to fall three feet
(1/2 D)

$$\begin{aligned} L &= (H/U_{Des})V_h \\ &= \left(\frac{3f}{3f/s} \right) (0.108 f/s) \\ &= 0.1 f \end{aligned}$$

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Check for Water/Hydrocarbon Separation

- U_{Des} for water/hydrocarbon separation is
0.035 f/s
- Horizontal liquid velocity is:

$$\begin{aligned} U_h &= Q/CSA \\ &= \frac{0.36 \text{ ft}^3/\text{s}}{(1/2)(\pi)\left(\frac{6}{2}\right)^2} \\ &= 0.025 \text{ f/s} \end{aligned}$$

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Check for Water/Hydrocarbon Separation (continued)

- Distance needed for droplet to separate in three feet (1/2 D)

$$L = (H/U_{Des})N_h$$

$$= (3/0.035)(0.025)$$

$$= 2.2f$$

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Horizontal Vessel Sizing Example Drop Leg (Boot) Sizing Steps

- Determine residence time
- Calculate diameter based on phase separation
- Calculate diameter based on process velocity limits (if any)
- Calculate diameter based on residence time (UOP standard length = 3.5 feet)

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Horizontal Vessel Sizing Example

Drop Leg (Boot) Sizing Steps

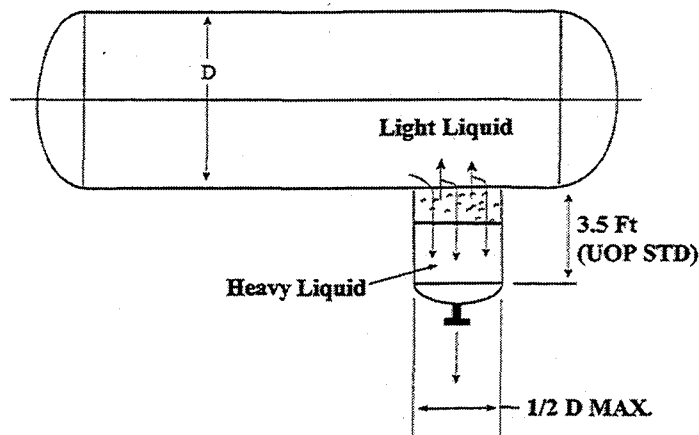
(continued)

- Select governing (largest) diameter
- If D_{DL} greater than half vessel diameter, then set $D_{DL} = 1/2$ vessel diameter and recalculate L_{DL}
- Round up to the next pipe size

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Horizontal Vessel Sizing Example

(continued)



SD-R01-14
IPD-2004/PVD-76

Horizontal Vessel Sizing Example

Vessel Drop Leg (Boot)

- Size the drop leg for 10 minute residence time for the water phase
- Size drop leg for separation of dispersed hydrocarbon liquid phase from a continuous water phase

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Horizontal Vessel Sizing Example

Vessel Drop Leg (Boot)

(continued)

- Calculate the drop leg diameter based upon the 10 minute residence time and a drop leg length of 3.5 feet

$$V_{DL} = Q_{CHL} t = \frac{\pi D_{DL}^2 L}{4}$$

$$\left(\frac{5 \text{ gal}}{\text{min}} \right) (10 \text{ min}) \left(\frac{\text{ft}^3}{7.48 \text{ gal}} \right) = \frac{3.5 \pi D_{DL}^2}{4}$$

$$D_{DL}^2 = 2.43$$

$$D_{DL} = 1.56 \text{ ft} = 18.7 \text{ inches}$$

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Horizontal Vessel Sizing Example ***Vessel Drop Leg (Boot)*** ***(continued)***

- Use Stokes' Law to determine the terminal velocity of the dispersed liquid droplet

$$\begin{aligned}
 U_{DLL} &= \frac{g D_{PL}^2 (\rho_{HL} - \rho_{LL})}{18 \mu_{HL}} \\
 &= \frac{(32.2)(0.00041)^2 (61.55 - 46.34)}{(18)(0.000342)} \\
 &= 0.013 \text{ f/s}
 \end{aligned}$$

IPD-2004/PVD-79

Horizontal Vessel Sizing Example ***Vessel Drop Leg (Boot)*** ***(continued)***

- The velocity of the continuous heavy liquid phase (water) is the volumetric flow rate divided by the cross-sectional area of the drop leg

$$\begin{aligned}
 U_{CHL} &= \frac{Q_{CHL}}{A_{DL}} = \frac{Q_{CHL}}{0.785 D_{DL}^2} \\
 &= \frac{0.01 \text{ ft}^3/\text{s}}{(0.785) \left(\frac{18.7 \text{ in}}{12 \text{ in/f}} \right)^2} \\
 &= 0.0052 \text{ f/s}
 \end{aligned}$$

IPD-2004/PVD-80

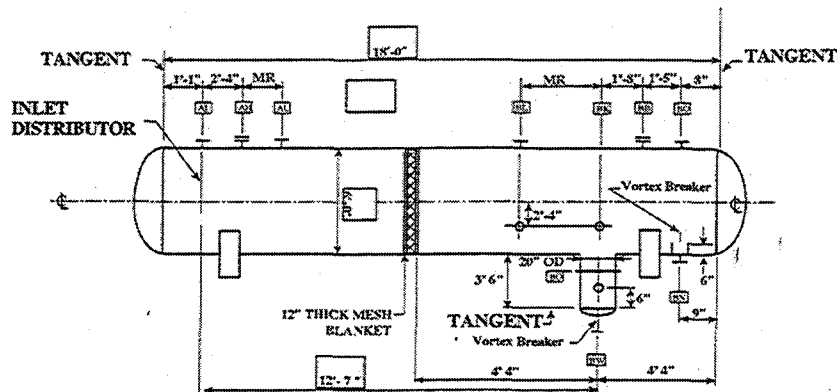
Horizontal Vessel Sizing Example

Vessel Drop Leg (Boot)

- Round to the next larger standard pipe size, 20 inch outside diameter
- Drop leg diameter is less than half of the 6 foot vessel diameter
 - Length of the drop leg is 3 feet, 6 inches
 - Since the drop leg is less than 30 inches outside diameter and the vessel is internally lined, the drop leg will be flanged to the vessel

IPD-2004/PVD-81

Reactor Products Separator



PROVIDE 18\" THICK E 316L WELD
OVERLAY ON THE
BOTTOM HALF OF THE VESSEL 316L
OVERLAY IS CORROSION ALLOWANCE

NOZZLES IN LINED SECTION
SHALL BE LINED AND FACED.

SD-R03-15
IPD-2004/PVD-82

Vertical Separator Sizing Without Disentraining Device

- Calculate Diameter Based on Phase Separation in Vapor Phase
- Calculate Diameter Based on Phase Separation in Liquid Phase
- Select Governing (Largest) Diameter

IPD-2004/PVD-83

Vertical Separator Sizing Without Disentraining Device (continued)

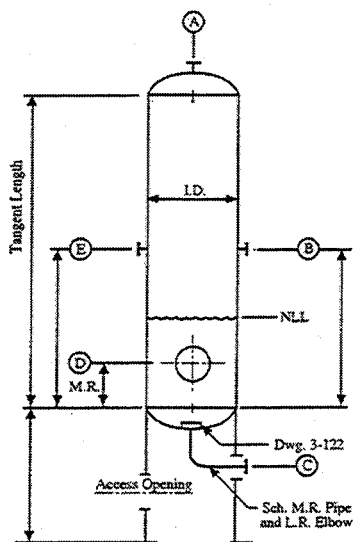
- Determine the Liquid Depth
 - The distance between H_{LL} and L_{LL} is determined from the required residence time (surge volume)
 - Provide 6" from vessel TL to L_{LL}
 - Set N_{LL} —and center of float (COF) on the level glass — at the N_{LL}

IPD-2004/PVD-84

Vertical Separator Sizing Without Disentraining Device (continued)

- **Determine the Required Tangent Length**
 - Determine the distance from the H_{LL} to the inlet nozzle -- use 1' 6" (minimum)
 - For critical services, where moisture carryover is a major concern, provide 2 minutes of surge volume, with a 1' 6" minimum
 - Determine the distance from the inlet nozzle to the top TL -- use 3' 0" (minimum)
 - Add liquid depth + H_{LL} to inlet + inlet to TL + inlet nozzle size to determine the overall TL
 - If TL is too long, consider a larger vessel diameter

IPD-2004/PVD-85



Example: Separator, Flash Drum

EDS-301-3443

Vapor/Liquid Separator

IPD-2004/PVD-86

Process Data Summary

Vapor	0.125 lb/ft ³	695 lb/h	1.54 ft ³ /s
Liquid			
HC	46.4 lb/ft ³	58586 lb/h	0.35 ft ³ /s
H ₂ O	62.3 lb/ft ³	2250 lb/h	0.01 ft ³ /s
		60636	0.36

IPD-2004/PVD-87

Diameter Based on Residence Time

- Size for 10 minute residence time
- Normal liquid level = 8 feet
- Vol = 0.35 ft³/s x 60 s/m x 10 m = 210 ft³
- $D = \left(\frac{210/8}{\pi/4} \right)^{0.5} = 5.9 \text{ ft}$ Use 6 ft diameter

IPD-2004/PVD-88

Diameter Based on Residence Time *(continued)*

- Inlet nozzle ID = 12 inch
- $L = \text{Max LL} + \text{clearance to feed nozzle} + \text{distance above feed nozzle}$
- $L = (NLL + \frac{1}{2}NLL) + \frac{1}{2} \text{ nozzle ID} + 1.5 \text{ ft} + \frac{1}{2} \text{ nozzle ID} + 3.0 \text{ ft}$
- $L = (8 + \frac{1}{2}8 + \frac{1}{2}(1) + 1.5 + \frac{1}{2}(1) + 3.0 = 18.5 \text{ ft}$
- Round to 18 feet

IPD-2004/PVD-89

Diameter Based on Gas Velocity

$$U_{\text{MAX}} = 0.157 \left(\frac{\rho_L - \rho_V}{\rho_V} \right)^{0.5}$$

$$U_{\text{MAX}} = 3.0 \text{ ft/s}$$

$$A_{\text{MIN}} = \frac{1.54 \text{ ft}^3/\text{s}}{3 \text{ ft/s}} = 0.51 \text{ ft}^2$$

$$D = \left(\frac{0.51}{\pi/4} \right)^{0.5} = 0.8 \text{ ft}$$

IPD-2002/PVD-90

Diameter Based on Liquid Velocity

$$U_{\text{Des}} = 0.075 \text{ ft/s}$$

$$A = \frac{0.36 \text{ ft}^3/\text{s}}{0.075 \text{ f/s}} = 4.8 \text{ ft}^2$$

$$D = \left(\frac{4.8}{\pi/4} \right)^{0.5} = 2.5 \text{ ft}$$

IPD-2004/PVD-91

Process Data Summary (continued)

■ Summary of Calculated Diameters

Residence Time	6 ft diameter
Vapor Velocity	0.8 ft diameter
Liquid Velocity	2.5 ft diameter

IPD-2004/PVD-92

UOP Manways and Auxiliary Nozzles

■ Manways (18" min ID)

- Vertical Vessels
 - Provide on top head
 - Flange top head if vessel diameter $\leq 30''$
- Horizontal Vessels
 - Provide on the side at or below the centerline; place above the centerline if vessel is lined
 - Flange vessel head if diameter $\leq 30''$
 - Locate near one end

IPD-2004/PVD-93

UOP Manways and Auxiliary Nozzles ***(continued)***

■ Ventilation

- Required for safety on horizontal vessels $> 10'$ diameter
- Locate on top, at the end opposite the manway
- Minimum sizes

$\geq 10'$ diameter	4" nozzle
$> 15'$ diameter	6" nozzle
$> 25'$ diameter	8" nozzle

IPD-2004/PVD-94

UOP Manways and Auxiliary Nozzles (continued)

■ Vents and Drains

- Generally required on the vessel or on piping near the vessel
- Locate to completely vent and drain the vessel (no pockets)
- Minimum sizes

≤ 15' diameter	2" nozzle
15' - 20' diameter	3" nozzle
> 20' diameter	4" nozzle

IPD-2004/PVD-95

Vortex Breakers

- Provide vortex breakers on liquid outlets to eliminate possible vapor entrainment (especially objectionable on a pump suction)
- When two liquid phases are present, raise the light phase vortex breaker to prevent possible inflow of heavy phase liquid
 - For water, it is typically raised 6 inches

IPD-2004/PVD-96

Inlet Distributors

- Inlet distributors change the liquid direction to provide momentum separation
- Horizontal Vessels
 - Subcooled liquids use an open ended pipe from the top or bottom of the vessel; distributor exit located to always remain beneath liquid
 - Bubble point liquids (e.g. small amount of vapor) use a vertical slotted pipe the same size as the inlet
 - Two liquids of approximately equal volume – use a horizontal slotted inlet pipe near the normal liquid interface (i.e. the vessel centerline); pipe is the same size as the inlet

IPD-2004/PVD-97

Inlet Distributors ***(continued)***

- Horizontal Vessels (continued)
 - Mixed Phases -- use a vertical slotted inlet one pipe size larger than the inlet pipe
 - Slot area is 2 to 3 times the distributor area
 - If there are two approximately equal liquid phases, locate the inlet near the center of the vessel and use two slots facing the shell

IPD-2004/PVD-98

Inlet Distributors ***(continued)***

■ **Horizontal Vessels (continued)**

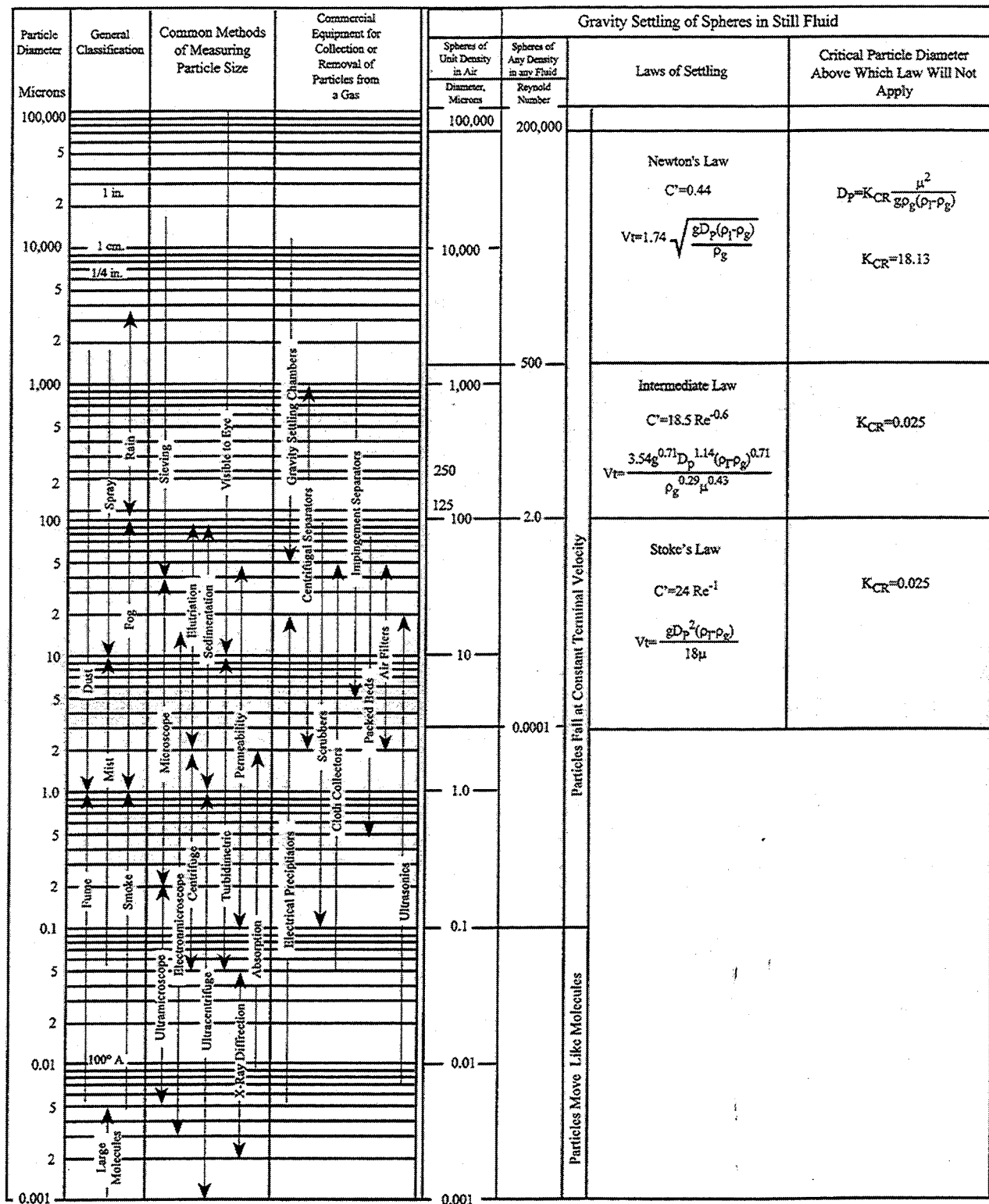
- Unless otherwise noted:
 - Slots begin 6" from the inlet nozzle
 - Slots end at the end of the distributor
 - Distributor ends 6" from the vessel bottom
 - Slot area is twice the distributor area
 - Use one slot
 - Use intermittent rather than very narrow (< 3/16" (5mm)) slots
 - Slot to face the nearest head (to maximize vessel gravity separation length and residence time and provide additional momentum separation)

IPD-2004/PVD-99

Inlet Distributors ***(continued)***

- **Vertical vessels use a tee style distributor, except on small vessels with flanged heads**

IPD-2004/PVD-100



INTRODUCTION

The design and analysis of modern pressure vessels is a complex task requiring the pressure vessel designer to have a broad and in-depth understanding of various specialized fields such as applied mechanics, strength of materials, fatigue, fracture mechanics, plasticity, creep, stress rupture, metallurgy, heat transfer and computational methods. The pressure vessel designer must also maintain an up-to-date working knowledge of the various codes and fabrication processes as they relate to vessel materials and construction. He must also provide coordination between process and project engineers and the vessel fabricator in order to properly communicate the requirements of the vessel process design.

This presentation will provide an overview of the various facets of pressure vessel design in order to provide process and project engineers with additional background to communicate with the pressure vessel designer. The design code referred to herein is the ASME (American Society of Mechanical Engineers) Boiler and Pressure Vessel Code which UOP utilizes as a minimum requirement. However, it should be recognized that other country and local codes may ultimately govern in design where their requirements are more stringent.

CODE REQUIREMENTS

ASME Code

- **Section I**

All vessels used for the purpose of generating steam with a process fluid or within a process heater will be designed according to the ASME Code, Section I, Boiler Code (latest edition). In some instances where steam is generated by a process stream without the use of direct fire, the vessels may also be designed according to Section VIII of the ASME Code, but it is not mandatory.

- **Section VIII, Division 1**

All process vessels should be designed according to the ASME Code, Section VIII, Unfired Pressure Vessel Code (latest edition) and/or the codes which govern in the particular country in

which the unit is to be installed. This does not apply to small vessels made of pipe and atmospheric vessels handling water and injection chemicals.

- **Section VIII, Division 2**

In cases of vessels operating at pressures above 1000 psig and less than 900°F, consideration should be given to designing to the ASME Code, Section VIII, Division 2. This division of the code is similar to the ASME Code, Section III, Nuclear Code and permits using a higher design stress level than Section VIII. However, there are special requirements on the design, fabrication and material testing which can offset much of the savings due to the higher allowable stresses permitted by Division 2.

Non-Code Vessels

Many atmospheric vessels handling water and injection chemicals need not be code vessels. There is a nominal savings due to not having to be fabricated in a code shop and not requiring a code stamp.

Other Related Codes

- API Standard 620, Large Low Pressure Storage Tanks
Pressure 0.5 to 15 psig
- API Standard 650, Welded Storage Tanks
Pressures up to 0.5 psig
- ANSI B31.3, Petroleum Refining Piping
- ANSI B16.5, Steel Pipe Flanges, Flanged Valves and Fittings
- TEMA, Heat Exchangers

Foreign Codes and Standards

The requirements of foreign codes and standards vary considerably, but usually any savings or additional costs due to the calculated wall thicknesses are compensated for by variations in quality and fabrication control requirements.

DESIGN CONDITIONS AND REQUIREMENTS

Design Conditions

The operating temperatures, pressures, flowing quantities, and densities determined by the process design engineer is the basic criteria used in setting the vessel design conditions. Additionally, climatic conditions relating to wind and earthquake must also be taken into consideration. With this information, the following guidelines can be utilized in setting vessel design requirements.

- **Pressure**

- Normal Design - The design pressure should, normally, be a minimum of 25 psi or 10% above the maximum operating pressure, whichever is greater. Minimum design pressure is normally 50 psig.
- Very Large Vessels - In cases of very large vessels operating at low pressures, the 25 psi increment is sometimes used for design above operating pressure rather than the minimum of 50 psig.
- High Pressure - At operating pressures above 1000 psig, the design conditions are taken as a minimum of 100 psi or 5% above operating pressure, whichever is greater. A pilot operated relief valve is required for this lower design margin to be used, however, so as to obtain bubble tightness at 95% of set pressure.
- Vacuum - Vessels operating under any condition of vacuum should be designed for full vacuum.
- Possible Vacuum - Vessels not operating under a vacuum, but subject to a vacuum condition due to failure of a control or heat source, should be designed for full vacuum or provided with a vacuum breaking device. Designing for vacuum usually only requires the addition of stiffening rings to the shell for most vessels designed for a reasonable internal pressure, although some vessels may also require an increase in wall thickness.

- Liquid Filled Vessels - Vessels operating liquid full, where a sudden drainage or other circumstance such as flow stoppage and subsequent cooling could produce a vacuum condition, should be designed for vacuum.
- Hydrostatic Test - In some instances, the design pressure at the bottom of the vessel may be dictated by the static head of liquid required for hydrostatic testing of the vessel.

- **Temperature**

- Normal Design - The design temperature should be about 50°F above the normal operating temperature or 250°F minimum. Consideration should be given to failure of coolers ahead of vessels, which could require a greater increment than 50°F.
- High Temperature - Reactors operating at high temperatures and high pressures, where the allowable stress drops rapidly with each temperature increment, are usually designed for the maximum expected operating temperature at end of run. This practice is usually conservative if the reactions are endothermic. Consideration should be given to the accuracy of the temperature-sensing element so as not to exceed the design temperature without sensing it.
- Flange Rating Consideration - In setting the design temperature of vessels, the effect on flange rating should always be taken into account so as not to increase cost unduly. Pressure and temperature limitations are as indicated in ANSI B16.5 (latest edition).
- Cold Wall Design - When temperatures are well above levels which can be reasonably designed for and, in some instances, to reduce alloy and thickness requirements in high pressure vessels, internally insulated (cold wall) vessels are used where the outside shell is designed for a temperature lower than the operating temperature.
- Minimum Design Metal Temperature - The ASME Code requires that the Minimum Design Metal Temperature (MDMT) be specified. The MDMT will be the lowest of the normal operating temperature minus 25°F (14°C), the lowest 24-hour mean ambient temperature, or the result of an auto-refrigeration calculation based upon the process fluid.

- **Corrosion Allowance**

- Normal Corrosion Allowance - The normal corrosion allowance on carbon steel vessels in non-corrosive service is 1/8 inch.
- Cladding - Vessels in severely corrosive service are usually alloy clad or have "weld-overlay" for protection. Where cladding or weld-overlay is used to protect against corrosion, no corrosion allowance is added to the base metal.
- Cement Lining - Where corrosive aqueous phases are handled at low temperatures, the metal is sometimes protected with a layer of "acid proof cement" or monel. The monel lining is easier to inspect than the metal behind the cement lining.

- **Minimum Wall Thickness**

- 0.25" for carbon steel vessels with 1/16" corrosion allowance.
- 0.3125" for carbon steel vessels with 1/8" corrosion allowance.
- 0.1875" for stainless steel vessels with 1/16" corrosion allowance.
- Wall thickness must be great enough to assure structural stability.

Setting Basic Dimensions

The size and shape of most process vessels is determined by the process design engineer. By training and experience, he is the logical person to make such decisions. The diameter and height of a distillation tower are basically set by the distillation design calculation. The volume of accumulator drums, reactors, and settling vessels are similarly dictated by process calculations. Anyone who has made such calculations, however, realizes that there is considerable leeway in the final choice of dimensions. For example, an accumulator drum must have a certain volume, but a great number of length-to-diameter ratios can be employed to give that volume. Other factors being equal, it is necessary to choose the most economical ratio. A length to diameter ratio of 3:1 or 4:1 is usually most economical. Other factors, such as plant layout, appearance and foundation conditions must also be considered. Above all, the process design engineer must set vessel dimensions so standard components can be employed throughout. The use of standard heads, for example, will avoid extra charges for special patterns.

If possible, the process design engineer should consult with the vessel designer during the early stages of vessel planning so that the most economical design can be realized. Engineers concerned with the detailed mechanical design and fabrication of a vessel are not in a position to change basic dimensions, since these dimensions are required for the process. Therefore, it is imperative that the basic dimensions be set intelligently by the process design engineer so that they will fulfill both process requirements and the requirements of economical design.

Residence Time

The determination of a residence time for the liquid phase should consider the following factors:

- A recommendation based on experience with the process
- The effects of loss of level or over filling on upstream or downstream equipment
- The difficulty in separating phases. Fluid properties, emulsifying or foaming tendencies, upstream mixing, surface active agents and suspended solids can affect phase separation in varying degrees.

Common residence times used for sizing vessels are as follows:

- For a horizontal vessel with a vapor phase and one liquid phase, 5 minutes based on the liquid phase rate, $\frac{1}{2}$ full, is usually adequate.
- For a horizontal vessel with a vapor phase and two liquid phases, 10 minutes based on the total liquid rate, $\frac{1}{2}$ full, is usually suitable for most conditions. When the difference in densities of the liquid phases is low (7 lb/ft^3 or less) and/or the lighter liquid viscosity is high (1 centipoise or more), 30 minutes or more may be necessary. When the difference in densities of the liquid phases is high (14 lb/ft^3 or more), 5 minutes may be adequate. Experience with the system is especially important in liquid/liquid separations.

- For liquid surge time from high to low level, 2-4 minutes is normally adequate for most conditions. For vertical separators upstream of compressors or other critical equipment, a minimum of two minutes surge time is provided between the bottom of the inlet nozzle and the maximum controlled level. For horizontal vessels, consideration should be given to the location of the normal liquid level. If the vapor rate is small and the liquid rate is large, it may be advantageous to set the normal liquid level above the vessel centerline to maximize the liquid residence time. If the vapor rate is large and the liquid rate is small, it may be advantageous to set the normal liquid level below the vessel centerline to minimize the vapor velocity.
- For drop legs with a level controlled outlet, 5-10 minutes is a suitable range for the drop leg capacity.
- For drop legs with a manually controlled outlet, 600 minutes of drop leg capacity is adequate for one draining during an eight-hour shift.

Phase Separation

Designing for phase separation assumes that a droplet falling under the influence of gravity will accelerate until the frictional drag force balances the gravitational force. The droplet then falls at a constant velocity, which is defined as the terminal or free settling velocity. This relationship is shown in Figure IX-1. The curve has been divided into three sections where the drag coefficient can be considered a linear relationship relative to Reynolds number. There are three laws for determining the terminal velocity, which is the basis of the criteria recommended for sizing vessels for phase separation. These laws have limits based on the particle Reynolds number as follows:

Particle Reynolds No.	Law	C_D	Terminal Velocity Equation
< 2	Stokes'	$\frac{24}{N_{Re}}$	$\frac{(\rho_P - \rho)gD_P^2}{18\mu}$
2 to < 500	Intermediate	$\frac{18.5}{(N_{Re})^{0.6}}$	$\frac{0.153g^{0.71}D_P^{1.14}(\rho_P - \rho)^{0.71}}{\rho^{0.29}\mu^{0.43}}$
500 to < 200,000	Newton's	0.44	$1.75\left[\frac{gD_P(\rho_P - \rho)}{\rho}\right]^{0.5}$

The terms in these equations are as follows:

ρ = density of the continuous phase

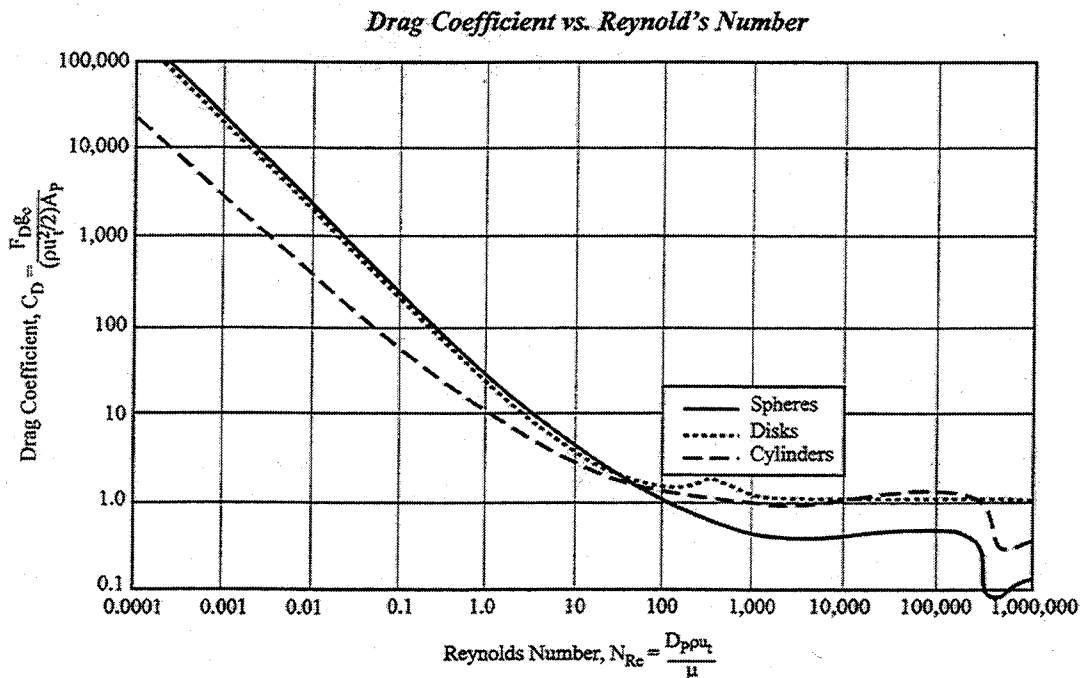
ρ_P = density of the droplet

D_P = diameter of the droplet

μ = viscosity of the continuous phase

g = gravitational constant

Figure IX-1



Design conditions for receivers, separators, and drums fall into the categories of the Stokes' and Intermediate Laws. Stokes' Law is used for disentraining vapor from the liquid phase and gravity settling the dispersed liquid phase from the continuous liquid phase; the Intermediate Law is used for gravity settling liquid from the vapor phase. The droplet size or diameter used in these laws is often difficult to determine. In selecting the droplet diameter, the effect of entrainment on downstream equipment, effluent quality, and overall economics should be considered.

The following droplet diameters are used in these guidelines in lieu of more definitive sizes:

- 175 microns for gravity settling liquid from the vapor phase
- 175 microns for disentraining vapor from the liquid phase
- 125 microns for gravity settling the dispersed liquid phase from the continuous liquid phase

In range of Newton's equation, $N_{Re} > 500$, liquid droplets falling in vapor will deform and C_D will not be 0.44. It is reasonable to set $C_D = 1$ as the curve indicates for cylinders and discs. ρ_L is substituted for ρ_p and ρ_V for ρ and a typical droplet diameter of 175 microns is selected, then the terminal velocity becomes:

$$u_t = 0.157 \left[\frac{\rho_L - \rho_V}{\rho_V} \right]^{0.5}$$

Mesh blankets are frequently used to enhance coalescing of the liquid drops so that separation can be achieved at higher vapor velocities. The following relationship applies when mesh blankets are used.

$$u_t = K \left[\frac{\rho_L - \rho_V}{\rho_V} \right]^{0.5}$$

PSIA	K
15	0.35
50	0.34
100	0.32
300	0.30
500	0.28
1000	0.27

GUIDELINES FOR SIZING VERTICAL VESSELS

Surge/Storage Vessels

- Choose a desirable length/diameter (L/D) ratio. Usually 3:1 is a good choice.
- Determine the liquid residence time required and percent liquid full at normal liquid level.
- Calculate the liquid holdup volume required to satisfy the liquid residence time. This is the volumetric liquid flow rate, Q_L , multiplied by the liquid residence time, T .

$$V_L = Q_L T$$

- Calculate the volume of the vessel, V_V . This is simply the liquid volume, V_L , divided by the percent liquid full, F .

$$V_V = V_L / F$$

- Since the volume of the vessel equals the volume of a cylinder, set the vessel volume, V_V , equal to the volume of a cylinder.

$$V_V = \frac{\pi D^2 L}{4}$$

- Substitute for L the D determined by the L/D ratio and solve for the vessel diameter, D . For a L/D ratio of 3, the equation is as follows:

$$V_V = \frac{\pi D^2 (3D)}{4} \text{ and } D = \left[\frac{4V_V}{3\pi} \right]^{1/3}$$

- Round off diameters and tangent lengths to the next larger standard increment. Standard head sizes usually are available in 6 inch or 100 millimeter increments.

Phase Separation Vessels without Demisting Blanket

The method used sizes the vessel diameter so that the vapor velocity is equal to or less than the terminal velocity of a liquid droplet falling through the continuous vapor phase. The liquid velocity must also be equal to or less than the terminal velocity of a bubble rising through the continuous liquid phase. Refer to Figure IX-2 for a sketch of the system and a definition of terms. Subscripts DL and CV refer to the dispersed liquid and continuous vapor phases.

- Calculate the terminal velocity of the liquid droplet, V_{DL} , using the Intermediate equation.

$$V_{DL} = \frac{0.153 g^{0.71} D_{PL}^{1.14} (\rho_L - \rho_V)^{0.71}}{\rho_V^{0.29} \mu_V^{0.43}}$$

- The vapor velocity rising in the vessel, V_{CV} , is equal to the vapor volumetric flow rate, Q_{CV} , divided by the vessel cross-sectional area, and the cross-sectional area is the area of a circle with diameter D .

$$V_{CV} = \frac{Q_{CV}}{A} = \frac{4Q_{CV}}{\pi D^2} = V_{DL}$$

- Substitute for V_{DL} and solve for D .

$$D = 2.9 \left[\frac{Q_V \rho_V^{0.29} \mu_V^{0.43}}{g^{0.71} D^{1.14} (\rho_L - \rho_V)^{0.71}} \right]^{0.5} \quad (1)$$

- Similarly, the terminal velocity of the bubble rising through the continuous liquid phase can be calculated using Stokes' equation.

$$V_{DV} = \frac{g D_P^2 (\rho_L - \rho_V)}{18 \mu}$$

- The liquid velocity flowing downward in the vessel, V_{CL} , is equal to the liquid volumetric flow rate, Q_L , divided by the vessel cross-sectional area, and the cross-sectional area is that of a circle with diameter D .

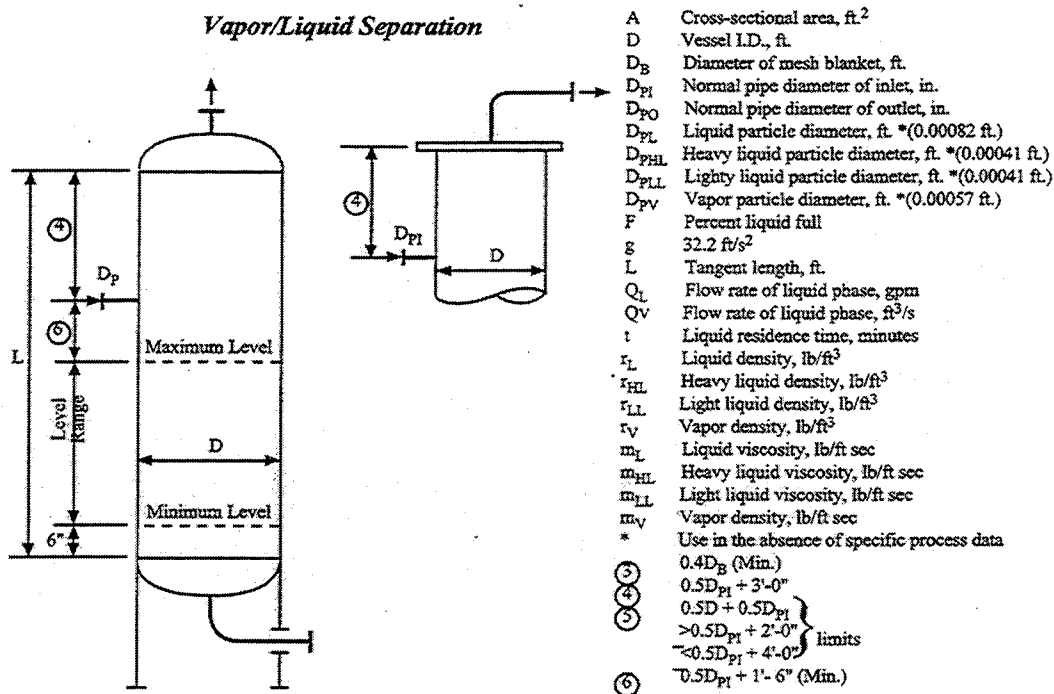
$$V_{CL} = \frac{Q_L}{A} = \frac{4Q_L}{\pi D^2} = V_{DV}$$

- Substitute for V_{DV} and solve for D .

$$D = 0.23 \left[\frac{Q_L \mu_L}{g D_{PV}^2 (\rho_L - \rho_V)} \right]^{0.5} \quad (2)$$

- Based on the largest calculated vessel diameter from Equations 1 and 2 above, determine whether an ellipsoidal or a flanged flat top head is suitable. Vessels smaller than 30 inches in diameter should have a flanged flat top head. Vessels with a diameter of 30 inches can have either type of top head. Vessels larger than 30 inches in diameter should have ellipsoidal heads.
- Determine the liquid level range for the vessel based on residence time considerations.
- Calculate the vessel tangent length based on the type of top head, the liquid level range, and any necessary surge time above the maximum controlled level by using the equations in the sketches in Figure IX-2.
- Round off diameters and tangent lengths to the next larger standard increment. Standard head sizes usually are available in 6 inch or 100 millimeters increments.

Figure IX-2



Note: Flowing data are at operating conditions.

EDS-R00-3421

GUIDELINES FOR SIZING HORIZONTAL VESSELS

Vessel Shell

Phase separation in a horizontal vessel should account for the fact that the forces acting on the droplet to be separated are not acting in the same plane. The traditional drag/gravity/buoyant forces are acting in a vertical direction, but the continuous fluid movement and its drag forces are acting in a horizontal direction. Our method takes into consideration the direction of these forces, as well as the vessel geometry in determining the vessel size.

Focus on the separation of a liquid droplet from a continuous vapor phase. Refer to Figure IX-3 for a sketch of the system and a definition of terms. The droplet must fall through the vertical distance of the vapor space to the continuous liquid phase before the horizontal forces of the vapor carry the droplet out of the vessel. To assure that the separation is maintained at all operating conditions, we will investigate the separation when the liquid level is at its highest. The highest liquid level will be no more than 0.25 diameter above the centerline.

- Select the tangent length/diameter (L/D) ratio.
- Determine the liquid residence time required and percent liquid full at normal liquid level. The normal liquid level is usually at the centerline, but exceptions can be considered.
- Determine the normal liquid level for vessels not liquid full. A standard level control range should be selected so that the level at maximum range of level controller does not exceed 0.75D.
- Calculate the vessel diameter to satisfy the residence time.
- Determine the terminal velocity, V_{DL} , of the droplet using the Intermediate equation.

$$V_{DL} = \frac{0.153 g^{0.71} D_{PL}^{1.14} (\rho_L - \rho_V)^{0.71}}{\rho_V^{0.29} \mu_V^{0.43}}$$

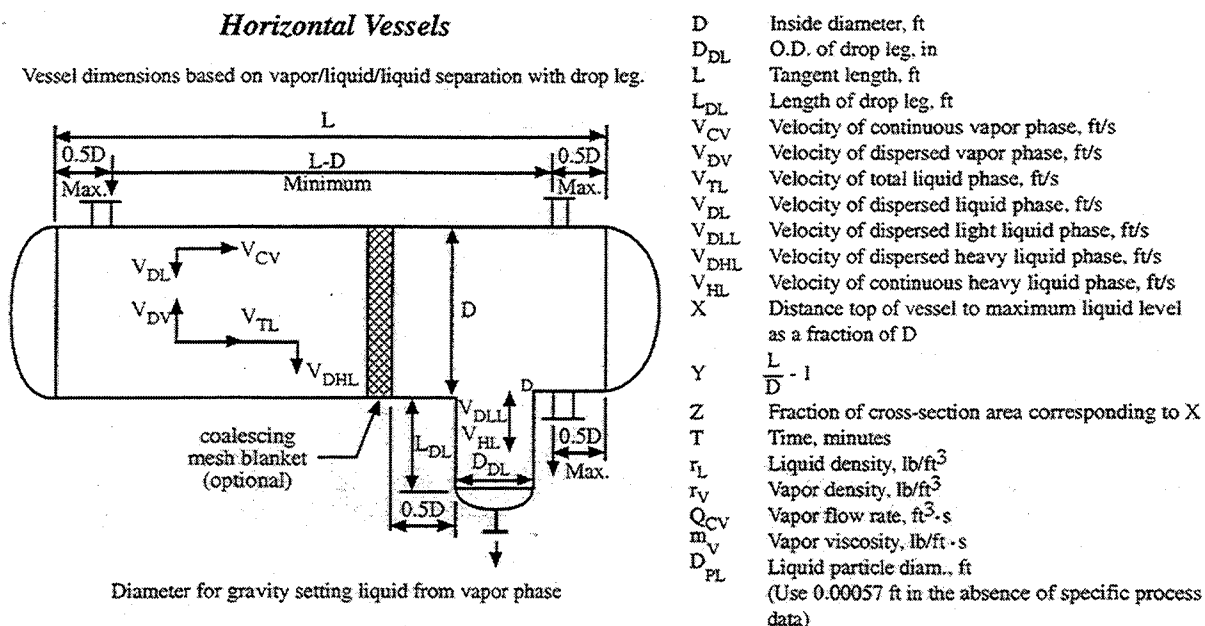
- The maximum distance the droplet must fall is the height of the vapor space from the top of the drum to the maximum liquid level. Let this distance be XD , where X is the vapor space height expressed as a fraction of D . The time, T , to drop through the vapor space is:

$$T = XD/V_{DL}$$

- The velocity of the vapor through the vessel, V_{CV} , is the volumetric flow rate of the vapor, Q_{CV} , divided by the flow area, where the flow area is expressed as a fraction, Z , of the vessel cross-sectional area, $0.785D^2$.

$$V_{CV} = \frac{Q_{CV}}{Z(0.785D^2)}$$

Figure IX-3



EDS-R00-3424

- The time for the vapor to pass through the vessel is the distance between the inlet and outlet nozzles divided by vapor velocity. We can express this distance as a factor, (Y), times the diameter, (D), where Y is (L/D)-1 if we assume that the inlet and outlet nozzles are each a maximum 0.5D from the vessel tangent. Now, if the time for the droplet to fall is set equal to the time for the vapor to pass through the vessel, we have:

$$T = \frac{XD}{V_{DL}} = \frac{YD}{V_{CV}}$$

- Substitute for V_{DL} , V_{CV} , and Y and solve for D.

$$D = \left[\frac{X}{0.785(0.153)YZ} \right]^{0.5} \left[\frac{Q_{CV} \rho_V^{0.29} \mu_V^{0.43}}{8^{0.71} D_{PL}^{1.14} (\rho_L - \rho_V)^{0.71}} \right]^{0.5}$$

- If X is 0.25, then Z is 0.20, and if we selected an L/D of 3, then Y is 2. The equation then simplifies to:

$$D = 2.28 \left[\frac{Q_{CV} \rho_V^{0.29} \mu_V^{0.43}}{g^{0.71} D_{PL}^{1.14} (\rho_L - \rho_V)^{0.71}} \right]^{0.5}$$

- Using a similar analysis, the vessel diameter can be calculated for separation of a vapor bubble from the continuous liquid phase. For this, Stokes' equation must be used in place on the Intermediate equation.
- If liquid/liquid separation is required, the calculation of the vessel diameter proceeds as in the paragraph above using the properties of the two liquid phases in Stokes' equation.
- If a vertical coalescing mesh blanket is to be installed in the vessel, calculate the diameter necessary for a maximum liquid velocity of 3 ft/m. The clearance between the mesh blanket and the drop leg should be approximately one-half of the vessel diameter.
- Size the vessel on the basis of the largest of the preceding calculated diameters.
- Round off diameters and tangent lengths to the next larger standard increment.

Vessel Drop Leg

- Determine the residence time.
- Calculate the drop leg diameter for phase separation by using Stokes' equation.
- Based on the recommended maximum drop leg velocity, calculate the drop leg diameter by using that maximum velocity.
- Calculate the drop leg diameter based on the residence time and a drop leg length of 3.5 feet (1070 millimeters).

- If the largest of the preceding calculated drop leg diameters is less than one-half of the vessel diameter, round this diameter up to the next larger pipe diameter. If the largest of the preceding calculated drop leg diameters is equal to or larger than one-half of the vessel diameter, round one-half of the vessel diameter down to the next lower whole even inch increment. Use that value as the drop leg diameter, and increase the drop leg length beyond 3.5 feet to satisfy the residence time requirement and round the length to the next largest 3-inch increment. If the drop leg length becomes larger than 6 feet, a smaller residence time or a different vessel configuration, such as a baffled separator, should be considered.
- Drop leg diameters should be specified as OD (outside diameter) in order to allow standard pipe material to be used. The minimum drop leg outside diameter is 14 inches. The minimum drop leg length is 3.5 feet.

HORIZONTAL VESSEL EXAMPLE

The following data is to be used to size a horizontal Product Separator in a naphtha hydrotreating process.

Stream	Vis cP	°API (lb/gal)	K	MW	Flowing Sp.Gr. (lb/ft ³)	°F	psig (mmHg)	bpsd @ 60°F (10 ⁶ std ft ³ /d)	Flowing GPM (ft ³ /s)	mol/h	lb/h	Btu/lb	10 ⁶ Btu/h
Reactor Section													
Total Feed to Product Separator				67.2		130	250			915.19	61529	37	2.20
Total Vapor	0.01			3.0	(0.125)	130	250	(2.11)	(2)	231.69	695	262	0.10
Hydrogen						130	250			226.07	456		
Hydrocarbon						130	250			3.67	202		
Water						130	250			1.95	35		
Total Liquid						130	250		163	683.50	60836		2.10
Hydrocarbon	0.42	50.1	11.44	104.9	0.743	130	250		158	558.62	58586	33	1.93
Water	0.51	10.0		18.0	0.987	130	250		5	124.88	2250	70	0.16

Tabulate data and calculate volumetric flows.

	Vapor	Hydrocarbon Liquid	Water Liquid
Flow, lb/h	695	58586	2250
, ft ³ /s	1.54	0.35	0.01
, gal/min	—	158	5
Density, lb/ft ³	0.125	46.4	61.55
Viscosity, cP	0.01	0.42	0.51
, lb/ft sec (x 10 ⁻⁴)	0.06	2.82	3.42

Calculate the diameter based on residence time.

Select $L/D = 3$

For settling water phase from light hydrocarbon phase, use a 10-minute residence time. Set normal liquid level at centerline.

Calculate liquid volume:

$$V_L = Q_L t = \left(\frac{0.36 \text{ ft}^3}{s} \right) \left(\frac{60 s}{min} \right) (10 \text{ min}) = 216 \text{ ft}^3$$

If vessel is $\frac{1}{2}$ full, then full vessel volume is:

$$V_V = \frac{V_L}{F} = \frac{216 \text{ ft}^3}{0.50} = 432 \text{ ft}^3$$

V_V is a cylinder with $L/D = 3$

$$V_V = \frac{\pi D^2 L}{4} = \frac{3\pi D^3}{4}$$

$$D = \left[\frac{4V_V}{3\pi} \right]^{1/3} = \left[\frac{4(432 \text{ ft}^3)}{3\pi} \right]^{1/3}$$

$$D = 5.7 \text{ feet}$$

Calculate the diameter based on liquid from vapor separation.

For normal liquid level at centerline, maximum level is $0.25D$ from top of vessel ($X = 0.25$ and $Z = 0.20$), and $L/D = 3$ (thus $Y = 2$), use the equation from page IX-16.

$$D = 2.28 \left[\frac{Q_{CV} \rho_V^{0.29} \mu_V^{0.43}}{g^{0.71} D_{PL}^{1.14} (\rho_L - \rho_V)^{0.71}} \right]^{0.5}$$

$$Q_{CV} = 1.54 \text{ ft}^3/\text{s}$$

$$\rho_V = 0.125 \text{ lb/ft}^3$$

$$\rho_L = 46.4 \text{ lb/ft}^3$$

$$\mu_V = 0.06 \times 10^{-4} \text{ lb/ft}^3$$

$$g = 32.2 \text{ ft/s}^2$$

$$D = 2.28 \left[\frac{(1.54)(0.125)^{0.29} (0.000006)^{0.43}}{(32.2)^{0.71} (0.00082)^{1.14} (46.4 - 0.125)^{0.71}} \right]^{0.5}$$

$$D = 0.68 \text{ feet}$$

Calculate the diameter based on vapor from liquid separation.

The terminal velocity of the bubble rising in the liquid, V_{DV} , is calculated with Stokes' Law.

$$V_{DV} = \frac{g D_{PV}^2 (\rho_L - \rho_V)}{18 \mu_L}$$

If X is the distance from the bottom of the vessel to the minimum level, expressed as a fraction of the diameter D , then the time, T , for the bubble to rise a distance XD is:

$$T = \frac{XD}{V_{DV}}$$

The horizontal velocity of the liquid phase is the volumetric flow of liquid divided by the flow area.

$$V_L = \frac{Q_L}{Z(0.785D^2)(7.48)(60)} = \frac{Q_L}{352.3 Z D^2}$$

where: Q_L is the liquid flow rate, gpm

Z is the fraction of cross-sectional area corresponding to X

$0.785D^2$ is the total cross-sectional area

$(7.48)(60)$ converts from ft^3/s to gpm

If $Y = LD - I$, then the time for the liquid to flow from the inlet to the outlet is:

$$T = \frac{YD}{V_L}$$

In the worst case, the time for the bubble to rise XD is equal to the time for the liquid to flow YD , and:

$$T = \frac{XD}{V_{DV}} = \frac{YD}{V_L}$$

$$XD V_L = YD V_{DV}$$

Substitute:

$$\frac{XD Q_L}{352.3 Z D^2} = \frac{YD g D_{PV}^2 (\rho_L - \rho_V)}{18 \mu_L}$$

Solve for D :

$$D = \left[\frac{18X}{352.3ZY} \right]^{0.5} \left[\frac{Q_L \mu_L}{g D_{PV}^2 (\rho_L - \rho_V)} \right]^{0.5}$$

For $X = 0.25$, then $Z = 0.20$

For $L/D = 3$, then $Y = 2$

And:

$$D = 0.18 \left[\frac{Q_L \mu_L}{g D_{PV}^2 (\rho_L - \rho_V)} \right]^{0.5}$$

$$D = 0.18 \left[\frac{(163)(0.000282)}{(32.2)(0.00057)^2 (46.4 - 0.125)} \right]^{0.5}$$

$$D = 1.75 \text{ feet}$$

This does not govern

Calculate the diameter based on liquid/liquid separation.

The method uses Stokes' Law to calculate the terminal velocity of the heavy liquid droplet and is very similar to the vapor from liquid case.

Use subscripts HL and LL for heavy liquid and light liquid.

From Stokes' Equation:

$$V_{HL} = \frac{g D_{PH}^2 (\rho_{HL} - \rho_{LL})}{18 \mu_{LL}}$$

and:

$$T = \frac{XD}{V_{HL}} = \frac{YD}{V_L}$$

Substitute and solve for D :

$$D = \left[\frac{18X}{352.3YZ} \right]^{0.5} \left[\frac{Q_L \mu_{LL}}{g D_{PH}^2 (\rho_{HL} - \rho_{LL})} \right]^{0.5}$$

For $X = 0.25$, then $Z = 0.20$

For $L/D = 3$, $Y = 2$

$$D = 0.18 \left[\frac{Q_L \mu_{LL}}{g D_{PH}^2 (\rho_{HL} - \rho_{LL})} \right]^{0.5}$$

$$D = 0.18 \left[\frac{(163)(0.000282)}{(32.2)(0.00041)^2 (61.55 - 46.4)} \right]^{0.5}$$

$$D = 4.2 \text{ feet}$$

The governing diameter, 5.7 feet, was that determined based on the liquid residence time. Round to:

$$D = 6 \text{ feet}$$

$$L = 3D = 18 \text{ feet}$$

Vessel Drop Leg (Boot)

Size the drop leg for 10-minute residence time for the water phase.

Size drop leg for separation of dispersed hydrocarbon liquid phase from a continuous water phase.

Use Stokes' Law to determine the terminal velocity of the dispersed liquid droplet.

$$V_{DLL} = \frac{g D_{PL}^2 (\rho_{HL} - \rho_{LL})}{18 \mu_{HL}}$$

The velocity of the continuous heavy phase (water) is the volumetric flow rate divided by the cross-sectional area of the drop leg.

$$V_{CHL} = \frac{Q_{CHL}}{A_{DL}} = \frac{Q_{CHL}}{0.785 D_{DL}^2}$$

Now set the velocity of the continuous water phase, V_{CHL} , equal to the terminal velocity of the hydrocarbon droplet, V_{DLL} , and solve for the diameter of the drop leg, D_{DL} .

$$V_{CHL} = V_{DLL} = \frac{Q_{CHL}}{0.785 D_{DL}^2 (7.48 \text{ gal/ft}^3)(60 \text{ s/min})} = \frac{g D_{PL}^2 (\rho_{HL} - \rho_{LL})}{18 \mu_{HL}}$$

$$D_{DL} = 2.71 \left[\frac{Q_{CHL} \mu_{HL}}{g D_{PL}^2 (\rho_{HL} - \rho_{LL})} \right]^{0.5}$$

$$D_{DL} = 2.71 \left[\frac{(5)(0.000342)}{(32.2)(0.00041)^2 (61.55 - 46.4)} \right]^{0.5}$$

$$D_{PL} = 12.4 \text{ inches}$$

Calculate the drop leg diameter based upon the 10 minute residence time and a drop leg length of 3.5 feet.

$$V_{DL} = Q_{CHL}t = \frac{\pi D_{DL}^2 L}{4}$$

$$\left(\frac{5 \text{ gal}}{\text{min}} \right) (10 \text{ min}) \left(\frac{\text{ft}^3}{7.48 \text{ gal}} \right) = \frac{3.5 \pi D_{DL}^2}{4}$$

$$D_{DL}^2 = 2.43$$

$$D_{DL} = 1.56 \text{ ft} = 18.7 \text{ in}$$

Select the largest calculated drop leg diameter, 18.7 inches, and round to the next larger standard pipe size, 20 inch outside diameter.

Length of the drop leg is 3 feet, 6 inches.

Since the drop leg is less than 30 inches outside diameter and the vessel is internally lined, the drop leg will be flanged to the vessel.

VESSEL FABRICATION

Vessel Shell

Pressure vessels for process plants are usually of cylindrical shape. Bending rollers consisting of three stacked rolls, one on top and two below, are used to form the cylinder. The length and sturdiness of the rollers will determine or limit the width of plates to be rolled. The top roller imposes the bending forces on the plate and the distance between the top and bottom rollers control the diameter of the cylinder. The rolling may be done cold on thin plates of smaller size. Prior to rolling, the ends of the plates are shaped to the desired radius in a press using a radius-forming die. By this procedure, perfectly round vessels can be made. After rolling, the edges of the plate are prepared for welding.

Plates for the fabrication of vessels are usually ordered by thickness. Plates made by a sheared-plate mill are of better quality and are available in a greater number of sizes. They can be obtained in a wide variety of thicknesses, the most common being from 3/16 inches to 1 inch in increments of 1/16 inch, 1 inch to 12 inch in increments of 1/8 inch, and 12 inch to 4 inches in increments of 3 inch. Lengths up to 800 inches are produced in the thinner plates. The maximum width available is about 195 inches, but extra charges on width may start at 100 inches. Therefore, it may prove economical to use more than one plate in certain instances for vessel sizes requiring plates over 100 inches wide. A comparison between plate width extras and the additional costs of fabrication with multiple plates must be made.

Vessel Heads

Ideally, a spherical-shaped pressure vessel is the most desirable since it withstands higher pressures for a given metal thickness. Spherical vessels are used in the storage of volatile liquids and gases, but they are not suitable for the construction of the usual process vessels. The cylindrical-shaped vessel is the next best design and is used most extensively. The heads for such vessels may be flat, ellipsoidal, elliptical dished, dished, hemispherical or conical. Flat heads can be employed economically only on small vessels (3 feet or less in diameter) or on vessels operating at low pressures. Flat heads may be welded to the vessel or installed as blind flanges on smaller vessels. Heads other than flat heads are manufactured by pressing or spinning large flat plates to the desired

shape. Sizes from twelve inches to twenty-two feet in diameter can be manufactured by spinning. Pressing is more generally confined to smaller heads and is only economical when there is a demand for a large number.

Often the process requirements dictate the style head to be used. A conical head, for instance, would be used for the bottom head of a settling drum. If no certain style is dictated, however, the choice between hemispherical, standard dished, and ellipsoidal heads is a question of economics. As the thickness required to withstand a given pressure increases in the order hemispherical, ellipsoidal, and dished, the cost of forming decreases in the same order. Although it is difficult to generalize, it may be said that in the range of low to moderate pressures, the standard dished head is used. At higher pressures (above 100 psig), it proves to be more economical to employ the ellipsoidal (elliptical dished) head with 2:1 ratio of major to minor axis. A wide assortment of standard heads is manufactured and it is always desirable to use these whenever possible. The use of a non-standard head requires the manufacture of additional dies needed in the forming process and thus entails additional expense.

Nozzles

All process vessels require nozzles for fluids, vapors or solids to enter and leave the vessel and for instrument connections, drains, vents, handholes, and manholes. These nozzles may consist of pipe couplings, forged steel nozzles, built-up nozzles, integrally reinforced nozzles or other suitable arrangements in accordance with the code requirements. Although the code permits the use of both fusion-welded and threaded nozzles, the use of threaded nozzles is generally avoided.

Cutting of holes into the shell of the vessel for the installation of nozzles weakens the vessel; therefore, reinforcement around the nozzle is required. Detailed rules for such reinforcement are given in the code. Forged steel welding necks of lengths sufficient to protrude beyond vessel insulation are mass-produced by several manufacturers. The lengths of such nozzles have been standardized and it is customary in designing vessels to adhere to these standard lengths whether the nozzle be purchased or fabricated from pipe. So-called welding-type nozzles with integral reinforcing necks are also available. These nozzles do not require the additional reinforcement necessary for welding neck nozzles.

UOP recommends a minimum nozzle size of 1-inch on pressure vessels. This ensures mechanical strength and reduces the chance of plugging of the nozzle. There are other minimum size requirements for certain services such as vents, drains, steamout, level, temperature, and manways, which will be discussed later.

Nozzles larger than half the vessel diameter require additional investigation and possibly additional reinforcement and are not commonly recommended by UOP.

According to the ASME Section VIII Pressure Vessel Code, all nozzles on vessel heads must be located inside a circle of diameter 0.8 times the inside diameter of the vessel. On very small fractionators, it is sometimes necessary to install the vapor outlet on the side rather than on the head to satisfy this requirement. Consult a flange table for the outside diameter of flanges.

To avoid interference with the vessel weld line, 2-inch nozzles and smaller are normally located with a minimum of 6 inches from the tangent line to the centerline of the nozzle. Nozzles greater than 2 inches should be located a minimum distance of the nozzle diameter plus 5 inches from the tangent line to allow for the reinforcing pad.

Generally, for tangential nozzles on horizontal vessels, the maximum permissible distance from the centerline of the vessel to the nozzle is 0.45 times the vessel diameter for diameters greater than 12 feet. For more information and a table of maximum permissible distances versus vessel diameters, refer to Figure IX-4.

Inside Diameter of Vessel	A Maximum	B Minimum	Inside Diameter of Vessel	A Maximum	B Minimum
3'-0"	1'-4½"	1½"	7'-0"	3'-3"	3"
3'-6"	1'-7"	2"	7'-6"	3'-5½"	3½"
4'-0"	1'-10"	2"	8'-0"	3'-8½"	3½"
4'-6"	2'-1"	2"	8'-6"	3'-11"	4"
5'-0"	2'-3½"	2½"	9'-0"	4'-2"	4"
5'-6"	2'-6½"	2½"	10'-0"	4'-7½"	4½"
6'-0"	2'-9½"	2½"	11'-0"	5'-1"	5"
6'-6"	3'-0"	3"	12'-0"	5'-6½"	5½"

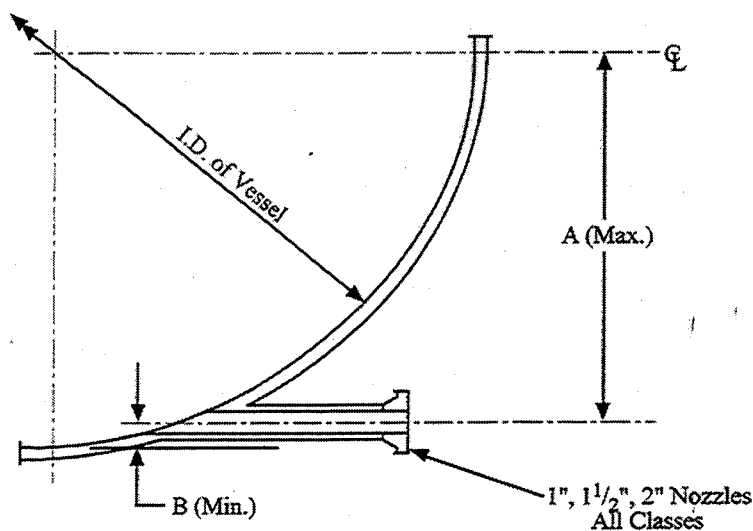
To simplify fabrication, a minimum "B" dimension of 6 inches is recommended when processing conditions will permit.

UOP practice is to specify manway sizes as inside diameter and to provide a minimum manway size of 18-inch inside diameter, including lining, if applicable. Larger size manways should be used if required to accommodate internals. On unlined horizontal vessels, the manway should normally be located on the side of the vessel at or below the horizontal center line. If the bottom half of the horizontal vessel is lined, the manway will be located on the upper side or the top of the vessel. Horizontal vessels with a diameter of 2 feet-6 inches or less can have a full flanged head.

Vertical driers, treaters, etc., usually combine the top manway with the process piping and also use it for loading solids or packing. The size should be a minimum of 18 inches inside diameter and may be larger to permit installation of internals. A manway is provided on some driers or clay towers on the side near the bottom of the vessel to unload solids directly into a truck.

Figure IX-4

*Design Limitations for
Tangential Nozzles on Horizontal Vessels*



EDS-R00-3427

For safe entry reasons, horizontal vessels 10 feet and greater in tangent length should be provided with a blanked off ventilation nozzle on top of the vessel near the end opposite the manway.

Vessel Tangent Length	Nozzle Size
10 feet through 15 feet	4 inches
Over 15 feet through 25 feet	6 inches
Over 25 feet	8 inches

A steamout connection should be provided on each vessel which is steamed out as part of the normal startup or shutdown operation. Examples of vessels which **should not** be provided a steamout connection are reactors in hydrogen-rich gas circuits, vessels containing caustic, and vessels containing materials such as molecular sieves which could be damaged by water. The steamout connection will be located near the bottom head on vertical vessels and on the head furthest from the ventilation nozzle on horizontal vessels.

The UOP recommendations for minimum sizes of vents, steamouts, and drains follows.

Vessel Diameter	Vent Steamout and Drain Size
15 feet and less	2 inches
Over 15 feet through 20 feet	3 inches
Over 20 feet	4 inches

For vertical vessels, the atmospheric vent nozzle and valve are located on the vessel, or on the piping near the vessel, at the highest point and are used for venting to atmosphere, for steamout, purging, and hydrostatic testing.

For horizontal vessels, the atmospheric vent valve is located on the vessel with a vent valve to atmosphere.

When a steamout connection is specified, the vent nozzle is located on the vessel with a vent valve to atmosphere.

The vent valve to atmosphere is not required when there is no steamout connection or when a vessel is purged and evacuated during the startup.

A depressuring connection to the pressure relief header is always required.

The drain is located on the vessel or piping near the vessel at the lowest point for draining hydrostatic test water to the sewer or hydrocarbon to a closed drain or pumpout system.

The design and pressure rating of the flanges on flanged nozzles are governed by the following codes:

- Chemical Plant and Petroleum Refinery Piping Code (ANSI B31.3)
- Steel Pipe Flanges and Flanged Fittings (ANSI B16.5 - Latest edition that has been approved by ANSI B 31.3)
- ASME Boiler and Pressure Vessel Code, Section VIII, Division 1

The nominal pressure ratings for flanges are Class 150, 300, 400, 600, 900, 1500 and 2500. The pressure temperature ratings of these classes are tabulated in ANSI B16.5. The ratings are for the maximum allowable non-shock pressures at the temperatures indicated. Interpolation between temperatures is acceptable. Markings on flanges usually consist of the manufacturer's name, material, nominal pressure class rating, temperature, size and ring joint groove number, if appropriate. Flange facings include raised face, male and female, tongue and groove, ring joint and lap joints. The raised face flange is the most commonly used.

Vessel Internals

Usually, vessel internals, such as trays and mesh blankets, are not manufactured by the vessel fabricator. The design and fabrication of these items requires a specialized technical know-how. However, support rings and bars for trays and mesh blankets will commonly be welded to the shell by the vessel fabricator. Other vessel internals such as baffles, distributors and vortex breakers may

be fabricated and installed by the vessel fabricator. It is desirable for all vessel internals to be designed so that they may be removed through the manways on the vessel.

Coalescing blankets are used to coalesce fine droplets of the dispersed phase from the continuous phase in a shorter period of time than could be accomplished in an empty vessel. If salts or corrosion products exist, coalescing blankets may plug.

Demisting blankets are used to remove entrained liquid in vapor streams where such liquid could cause damage to downstream equipment such as compressors or result in loss of product to the fuel system, etc. If excessive entrainment of liquid in the vapor outlet stream could be economically significant or detrimental to equipment, safety, or operation, the vessel is sized to allow gravity settling of liquid from the vapor phase in addition to providing a partial mesh blanket. For less critical situations, a full dimension demisting blanket which meets the vapor velocity criteria can be used.

Distributors are used to ensure that the introduction of streams into vessels does not disturb the liquid present or interfere with the separation desired. Some of the factors to consider are:

- For subcooled total liquid inlets such as cold feed surge drums or overhead systems with the final condenser at grade, the liquid inlet distributor should remain sealed under the minimum liquid level
- For bubble point liquid inlets where the possibility of small quantities of vapor may exist, such as a bubble point system, low pressure system, or overhead condensing systems with an equalizing line to the condenser rundown line, the inlet will be through the top of the vessel with an internal closed end, slotted pipe distributor. The inlet nozzle and internal distributor pipe will be the same size as the line size to the vessel. The closed end will terminate 6 inches above the bottom of the vessel. The slotted area is approximately twice the cross-sectional flow area of the distributor pipe and is provided by one slot, oriented on the vessel centerline, facing the nearest head of the vessel. The slotted section of the distributor starts approximately 6 inches from the top of the vessel and extends the length of distributor to the closed area. For small

diameter distributor pipes, intermittent larger slots should be considered in preference to very narrow slots.

- For total liquid systems with two immiscible liquids present in sufficient quantities to establish an interface level at or near the centerline of the vessel, the inlet will be on the side of the vessel with an internal distributor perpendicular to the major axis of the vessel. The inlet nozzle and the internal distributor will be the same size as the line size to the vessel. The horizontal pipe will be located at the normal elevation of the interface between the two liquids and will be closed at the end opposite the inlet and slotted on the side opposite the vessel outlet. The length of the horizontal pipe should be great enough to permit a slot area approximately twice the cross-sectional flow area of the distributor inlet pipe.
- Pipe lines carrying mixed phase streams are quite often sized for higher velocities than total liquid lines. If these higher velocities were carried into the vessel, the resulting turbulence would interfere with the disengaging function of the vessel. Therefore, both the inlet nozzle and the inlet distributor will be expanded one pipe size greater than the line size to the vessel in order to reduce the inlet velocity.

The inlet is located at the top of the vessel and will have a closed end, slotted internal distributor pipe terminating 6 inches above the bottom of the vessel. The slotted area is between two and three times the cross-sectional flow area of the distributor pipe. The slotted section of the distributor starts approximately 6 inches from the top of the vessel and extends the length of the distributor to the closed end. The slotted area for all vessels, except those with a vapor phase and baffles for separate ends for light and heavy liquid phases, is provided by one slot facing the nearest head of the vessel. For vessels with a vapor phase and baffles for separate ends for light and heavy liquid phases, the slotted area is provided by two slots arranged perpendicular to the major axis of the vessel.

- Vertical receivers or separators should use a tee-type distributor located close to the vessel wall.

- Vortex breakers (VBs) are used in any vessel outlet nozzle from which a pump is taking suction or if gas entrainment would present a problem in downstream equipment to which an outlet is feeding. The standard VB is made of two plates forming a cross extending down into the nozzle and overlapping the nozzle above the outlet. An extended pipe vortex breaker 6 inches long is provided if there is a drop leg on the vessel.
- For pump suction nozzles on the side of a vessel, or for very large nozzles, a grating is installed as a VB when the outlet is close to a liquid-vapor interface. A VB is **not** required for a fractionator tray drawoff nozzle.

External Appurtenances

In addition to nozzles, manholes and handholes that must be installed in the average process vessel, certain other exterior attachments are usually required. These include insulation support angles or rings, platform and ladder support lugs and brackets, davits and supports, and skirts or lugs for the support of the vessel. These items are supplied by the vessel fabricator and welded to the vessel shell before shipment. The items are usually fabricated from structural grade steel.

Vessel Supports

Most vertical vessels are supported by a skirt. The bottom nozzles must project through the skirt to avoid flanges inside the skirt, reducing the hazards associated with leaks. Generally, straight skirts are specified based on economics. There are exceptions for supports and good engineering judgement should be used. For small vessels less than 3 feet–6 inches in diameter and the support length from the base of the support to the bottom tangent line is 5 feet–0 inches or less, then support legs are provided.

Lugs are used for equipment support when good ventilation of the vessel surface is required, or a small diameter does not allow a skirt or legs to be provided.

Tabletop support is used where access to the bottom of a vessel is required. It is the most expensive type of support and is generally avoided. The vessel can be supported on the table top with lugs or a short straight or flared skirt.

Most horizontal vessels are supported on saddles, which sit on concrete supports. Saddles are normally 6-10 inches wide and are located at a distance from the vessel tangent lines no greater than 0.25 times the vessel diameter.

Insulation rings are used to support block insulation on the walls of vessels. Insulation is called for whenever heat loss or heat gain would affect the process. When insulation is required, the vessel project specification should indicate a requirement for insulation clips and rings. Insulation is required for personnel protection.

Weather shielding is sometimes specified to minimize thermal stress and differential thermal expansion of uninsulated nozzles and flanges.

Davits are a type of crane installed on fractionators and vessels to facilitate the raising of internals and for handling manway covers.

Platforms are installed at all manways and wherever required to reach equipment, such as controls, more than about 8 feet above grade. The vessel project specification indicates that ladder and platform clips are to be provided as required by the contractor.

Welding

There exists a wide variety of welding processes and techniques, but the basic types are arc, gas and resistance. These three types cover 90% of all welding done today.

- **Welding Processes**

Shielded arc welding is perhaps the most common method used in vessel fabrication. An arc is produced between a metal electrode and the base metal. The heat created causes melting of the electrode and portions of the base metal near the arc. The weld is formed by the fusion of the metal from the electrode and the base metal. The welding rods (electrodes) are coated with a flux which, when heated, produces an inert gas blanket around the arc and deposits a protective coating of slag over the weld to prevent oxidation. Arc welding can be manual or automatic. For large longitudinal and circumferential seams, automatic welding machines prove

particularly advantageous. For circumferential seams, the cylinder is moved and the torch remains stationary. The opposite procedure is followed for longitudinal seams.

TIG welding is chiefly used for welding light gages of stainless steel from less than .025 inch to 1/8 inch thick. Heavier sections may also be TIG welded, where the highest quality weld metal is required. A nonconsumable tungsten electrode is used. Inert gas is fed around the arc to protect the hot metal. Filler metal, when required, is added separately with a rod of continuous wire.

Spray-arc MIG welding is often used for high-deposition-rate welding of stainless steel from less than 3/16 inch and thicker. A continuous electrode wire, which melts and forms filler metal for the joint, is used. The metal from the melting electrode is transferred in fine droplets across the arc to the work. A protective shield of inert gas or gas mixture is used.

Shorting-arc MIG welding is usually used for welding light gage stainless from less than 1/16 inch to about 1/8 inch thick, and for out-of-position welding. Shorting-arc is similar to spray-arc welding except that lower currents and smaller diameter electrode wires are used. Molten metal is not transferred across the arc as in spray-arc, but is deposited in large size drops as the molten electrode tip makes contact with the weld puddle and the arc shorts out. This automatically occurs as much as 100 times or more per second.

Submerged-arc welding is most frequently used for welding stainless heavier than about 2 inch thick, but lighter gages to less than .078 inch can also be welded. A continuous wire electrode is used as in MIG welding, but the wire is usually of larger diameter. Instead of gas, a granular flux is used for shielding.

Acetylene gas welding is common, especially for welding thin plates and small attachments. In addition, acetylene welding is a necessity in any vessel fabrication shop for many utility purposes. A few of its uses are: flame hardening of small parts, cutting, and tack-welding sections prior to automatic arc welding. Gas welding is also used in brazing which is the joining

of metals with a nonferrous filler without melting the base metal. The filler material melts at 800°F+ and bonding is completed by capillary action.

- **Type of Electrode**

The electrode may melt (consumable) during welding and become part of the weld or remain intact (nonconsumable), depending on the welding process. Consumable electrodes are the type most commonly used in the welding of steel and stainless steel and usually are similar in composition to the base metal. Individual coated electrodes or a continuous length of wire are used depending on the welding process. The type of coating or electrode flux, and the size and composition of the electrode, will influence the weld penetration, the melt rate and the size of the finished weld bead. Nonconsumable electrodes made of tungsten can withstand the high arc heat without melting and are used in one major process (TIG) for welding stainless steel.

- **Welding Qualification**

Good welds are made by competent welders. Tests to qualify or certify welders are given in various vessel codes. The most commonly used is Section IX of the ASME Boiler and Pressure Vessel Code titled "Welding & Brazing Qualifications".

Welding of vessel seams generally requires the application of several layers of weld material. It is necessary that great care be exercised in making such "multi-pass" welds. The surface of each weld pass must be thoroughly cleaned and examined before proceeding with the next pass to insure good penetration and freedom from cracks, slag inclusion, and other weakening defects. A faulty weld can be disguised by a satisfactory final weld pass.

Postweld Heat Treatment

Residual stresses are produced around the welded seams of vessels. Such stresses weaken the vessel at these points, particularly in thicker vessels (above 13 inches thick). In these instances, and in all cases where maximum safety is required, the vessel is postweld heat treated to eliminate such residual stresses. PWHT term is commonly interchanged with stress relieving.

Stress relieving is an annealing or heat treating process, which relieves residual stresses. The operation varies, depending on the material of construction. For instance, carbon steels are heated to approximately 1100°F and then cooled slowly. Austenitic stainless steels, if required, are heated to around 2000°F and then cooled rapidly in order to prevent carbide precipitation characteristic of these metals in the range of 800-1600°F. Precipitation of the carbides in stainless steel destroys the corrosion resistance. Information of PWHT procedures can be found in the ASME Code, Section VIII, UHA-32 for stainless steel, and UCS-56 for carbon steel and low alloy steels.

A well-equipped vessel fabrication shop has large gas heated furnaces capable of receiving almost any size vessel that can be shipped in one piece. Large vessels that must be welded in the field can be stress-relieved around the welds using gas fired portable furnaces. Several low temperature stress-relieving processes are also used. These employ special oxyacetylene torches, which apply heat to the area at the welds, followed by a water spray. Because of the wide differences in physical properties of metals, it is imperative that the proper stress-relieving procedure be outlined by a competent welding engineer.

Radiographing

When the greatest confidence in the soundness of the construction of a vessel is desired, radiographing (X-raying) of all welds is done in order to detect defects. Because of the greater knowledge of weld quality, the code permits a lower safety factor in the form of a higher joint efficiency. It is particularly advisable to radiograph the large longitudinal and circumferential seams of all-important vessels. Joint efficiencies can be found in the ASME Code, UW-12. Plate thickness requiring X-ray is noted in UW-11 and UCS-57. Other inspection approaches include ultrasonic, magnaflux (magnetic particle) and dye penetrant check.

Shop and Field Fabrication

Because of the superior equipment and working conditions in a fabricator's shop, field fabrication should be kept to a minimum. Large diameter vessels (20 to 40 feet), however, must be field fabricated. Rolled sections (arcs) of the shell 7 to 8 feet wide are shipped to the job-site. When planning any vessel over 10 feet diameter by 35 feet long, railroad agents should be consulted. Sizes above this, which can be shipped, depend largely on the railroad facilities serving the area.

Vessel Inspection

The vessel codes require inspection by some public inspector, such as state, municipal, or insurance company inspectors, before the vessel can be certified as complying with the codes. Prior to this inspection, the purchaser should have inspected the vessel during all phases of its fabrication. This includes inspection of the plates and heads as they arrive at the fabricator's shop, study of mill test reports, and the following of all operations including the final hydrostatic test of the completed vessel. Mill test reports give the physical and chemical properties of the particular steel that was used by the steel mill in manufacturing the plates being used for the vessel.

PRESSURE VESSEL DESIGN

The ASME Code, Section VIII, Division 1, Article UG-27, and Appendix 1, Articles 1-1 through 1-4, provide instructions and equations for the design of pressure vessels under internal pressure. The requirements of these articles must be satisfied in the final design of a pressure vessel. However, for the limited purpose of determining an approximate wall thickness for planning and cost estimating purposes, the following basic equations are often useful.

Cylindrical Shells

The minimum thickness or maximum allowable internal working pressure of a cylindrical shell may be calculated by means of the following equations. The symbols below are used in the equations.

t = minimum required thickness of shell, in.

P = internal design pressure, psi (or maximum allowable working pressure)

R = inside radius of the shell, in.

R_o = outside radius of the shell, in.

S = maximum allowable stress value, psi

E = joint efficiency

These equations are limited to thicknesses not exceeding one-half of the inside radius or pressures not exceeding 0.385 SE.

- Based on inside radius:

$$t = \frac{PR}{SE - 0.6P} \text{ or } P = \frac{SEt}{R + 0.6t}$$

- Based on outside radius:

$$t = \frac{PR_o}{SE + 0.4P} \text{ or } P = \frac{SEt}{R_o - 0.4t}$$

Ellipsoidal Heads

Within the limitations indicated above for cylindrical shells, the following equations may be used to calculate the minimum thickness or maximum allowable internal working pressure of a formed ellipsoidal head. The symbols below are used in the equations.

- t = minimum required thickness of head after forming, in.
- P = internal design pressure, psi (or maximum allowable working pressure)
- D = inside diameter of the head skirt; or inside length of the major axis of an ellipsoidal head
- D_o = outside diameter of the head skirt; or outside length of the major axis of an ellipsoidal head
- S = maximum allowable working stress, psi
- E = joint efficiency
- h = one-half of the length of the minor axis of the ellipsoidal head, or the inside depth of the ellipsoidal head measured from the tangent line (head-bend line), in.
- K = a factor in the formulas for ellipsoidal heads depending on the head proportion $D/2h$
- $D/2h$ = ratio of the major to the minor axis of ellipsoidal heads, which equals the inside diameter of the skirt of the head divided by twice the inside height of the head

- Based on inside diameter: $t = \frac{PRK}{SE - P} \text{ or } P = \frac{2SEt}{KD + 0.2t}$

- Based on outside diameter:

$$t = \frac{PR_o K}{SE + P(K - 0.1)} \text{ or } P = \frac{SEt}{KR_o - t(K - 0.1)}$$

- Ellipsoidal head factor:
$$K = \frac{1}{6} \left[2 + \left(\frac{D}{2h} \right)^2 \right]$$

Note that $K = 1$ for a 2:1 ellipsoidal head

Stress Values

Stress values used in the equations above are as indicated in the ASME Code, Section VIII, Division 1. The values are usually one-fourth of the tensile strength, resulting in a factor of safety of 4. Designs based upon Division 2 requirements use higher stress values, resulting in a factor of safety of 3. At temperatures in the creep range, stress values are based on creep rates.

Joint Efficiencies

Table UW-12 in the ASME Code, Section VIII, Division 1, provides maximum allowable joint efficiencies for welds under various degrees of radiographic examination. Some commonly used joint efficiencies are as follows:

- **Double-welded butt joints**
 - Fully radiographed - 100%
 - Spot radiographed - 85%
 - No radiographing - 70%
- **Single-welded butt joints with backing strip**
 - Fully radiographed - 90%
 - Spot radiographed - 80%
 - No radiographing - 65%

ITEM

NAME OF VESSEL

SPECIFIC GRAVITY -



3-101.	3-122.	3-123.	3-317.	77830-A.	77831-f
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