Separation of oil and gas is a critical field processing operation.

Selecting gas/liquid separation technologies requires not only knowledge of the process conditions, but a knowledge of the characteristics of the liquid contaminants.

Selection should be made based on the:
- Droplet size
- Concentration
- Whether the liquid has waxing or fouling tendencies
Three principles used to achieve physical separation of gas and liquids or solids are:

- Momentum
- Gravity settling
- Coalescing

Any separator may employ one or more of these principles however, the fluid phases must be immiscible and have different densities for separation to occur.

Momentum force is utilized by changing the direction of flow and is usually employed for bulk separation of the fluid phases.

The gravitational force is utilized by reducing velocity so the liquid droplets can settle out in the space provided.

Gravity segregation is the main force that accomplishes the separation, which means the heaviest fluid settles to the bottom and the lightest fluid rises to the top.

However, very small droplets such as mist cannot be separated practically by gravity.

These droplets can be coalesced to form larger droplets that will settle by gravity.

The purpose of this chapter is to review the principles governing the basic separation process and associated equipment design procedure.
Gravity separators are pressure vessels that separate a mixed-phase stream into gas and liquid phases that are relatively free of each other.

In a gravity separator, gravitational forces control separation, and the efficiency of the gas/liquid separation is increased by lowering the gas velocity.

Because of the large vessel size required to achieve settling, gravity separators are rarely designed to remove droplets smaller than 250 μm.

However, an analysis of this type of separator is given because it is useful to help understand the settling mechanism of other separators.

Gravity separators are often classified by their geometrical configuration (vertical, horizontal) and by their function (two-phase/three-phase separator).

Gravity separators are classified as “two phase” if they separate gas from the total liquid stream and “three phase” if they also separate the liquid stream into its crude oil and water-rich phases.

Additionally, separators can be categorized according to their operating pressure.

- Low-pressure units handle pressures of 10 to 180 psi.
- Medium pressure separators operate from 230 to 700 psi.
- High-pressure units handle pressures of 975 to 1500 psi.
Separators are sometimes called “scrubbers” when the ratio of gas rate to liquid rate is very high.

These vessels usually have a small liquid collection section and are recommended only for the following items:

- Secondary separation to remove carryover fluids from process equipment such as absorbers and liquid dust scrubbers.
- Gas line separation downstream from a separator and where flow lines are not long.
- Miscellaneous separation where the gas–liquid ratio is extremely high.

In any case, these equipments have the same configuration and are sized in accordance with the same procedure of separators.

A "complete" separator must have the following:

1) A primary separation section to remove the bulk of the liquid from the gas
2) Sufficient liquid capacity to handle surges of liquid from the line
3) Sufficient length or height to allow the small droplets to settle out by gravity (to prevent undue entrainment)
4) A means of reducing turbulence in the main body of the separator so that proper settling may take place
5) A mist extractor to capture entrained droplets or those too small to settle by gravity
6) Proper back-pressure and liquid-level controls
Gravity separators are designed as either horizontal or vertical pressure vessels.
Figure below is a typical scheme of a three-phase horizontal separator.

The level of the gas/oil interface can vary from half the diameter to 75% of the diameter depending on the relative importance of liquid/gas separation and what purpose the separator has.
A typical configuration for a vertical three-phase separator.

The length of a horizontal separator has a greater effect on capacity than the height of a vertical type.

In the horizontal vessel, the length necessary depends on:
1) Droplet size
2) Vessel diameter
3) Gas velocity
4) Degree of turbulence
5) Droplet density
6) Gas density

Sufficient length is needed only for the velocity to become zero and for the droplet to start falling.

Increasing the length above this point accomplishes little good.
There are no simple rules for separator selection.

Sometimes, both configurations should be evaluated to decide which is more economical.

**Horizontal Separators**

Horizontal separators are used most commonly in the following conditions

1) Large volumes of gas and/or liquids
2) High-to-medium gas/oil ratio (GOR) streams
3) Foaming crudes
4) Three-phase separation

**Advantages**

1) Require smaller diameter for similar gas capacity as compared to vertical vessels
2) No counter-flow (gas flow does not oppose drainage of mist extractor)
3) Large liquid surface area for foam dispersion generally reduces turbulence
4) Larger surge volume capacity
Disadvantages

1) Only part of shell available for passage of gas
2) Occupies more space
3) Liquid level control is more critical
4) More difficult to clean produced sand, mud, wax, paraffin, etc

These separators are used in the following conditions:
1) Small flow rates of gas and/or liquids
2) Very high GOR streams or when the total gas volumes are low
3) Plot space is limited
4) Ease of level control is desired

Advantages

1) Liquid level control is not so critical.
2) Have good bottom-drain and clean-out facilities.
3) Can handle more sand, mud, paraffin, and wax without plugging.
4) Less tendency for reentrainment.
5) Has full diameter for gas flow at top and oil flow at bottom.
6) Occupies smaller plot area.
Separation equipment employs one or more of the following mechanisms:

1) gravity settling
2) centrifugal force
3) Impingement
4) electrostatic precipitation
5) sonic precipitation
6) Filtration
7) adhesive separation
8) Adsorption
9) thermal

Vapor/liquid separation is usually accomplished in three stages:

1) Primary separation
2) Secondary separation
3) Mist elimination
In the gravity-settling section of the separators, the liquid drops will settle at a velocity determined by equating the gravity force ($F_B$) on the drop with the drag force ($F_D$) caused by its motion relative to the vapor continuous phase.

When the drag force is equal to the buoyancy (gravity) force, the droplet acceleration is zero so that it moves at a constant velocity. This velocity is the terminal or free settling velocity.

$$F_B = \left( \frac{\pi}{6} \right) D_d^3 (\rho_L - \rho_V) \left( \frac{g}{g_c} \right)$$  \hspace{1cm} (1)

$D_d$ is drop diameter, ft
$\rho_L$ is liquid density, lbm/ft$^3$
$\rho_V$ is vapor density, lbm/ft$^3$
$g$ is gravitational constant, 32.174 ft/s$^2$
$g_c$ is conversion factor, 32.174 lbm-ft/s$^2$-lbf.
Also, the drag force on the droplet is given by

\[ F_D = C_D \left( A_p \right) \left( \frac{\rho_V V_d^2}{2g_c} \right) \]  

(2)

- \( C_D \) is drag coefficient, dimensionless
- \( A_p \) is projected drop area, ft\(^2\); = \( \frac{\pi}{4} D^2 \) (area of circle, not sphere)
- \( V_d \) is drop velocity, ft/sec.

Therefore, the terminal settling velocity of the liquid droplets \( (V_t) \) can be determined by equating Equations (1) and (2) as follow:

\[ U_T = \sqrt{\frac{4}{3}} D \rho_L (\rho_L - \rho_V) g / (C_D \rho_V) \]  

(3)

The droplet-settling velocity equation considers the escape of a drop from the continuous phase (e.g., the escape of an oil drop from the gas phase).

For this purpose, the droplet-settling velocity must be greater than the superficial upward bulk vapor velocity, \( U_V \).

Typically, the allowable vapor velocity is set between 0.75 \( U_t \) and \( U_t \) (Svrcek and Monnery, 1993).
Equation (3) can be rearranged as a Sauders and Brown (1934) type equation as follows:

$$U_T = K \left( \frac{\rho_L - \rho_V}{\rho_V} \right)^{1/2} \text{ ft/s } \tag{4}$$

where

$$K = \sqrt[3]{\frac{4gD_d}{3C_d}} \tag{5}$$

In practice $K$ depends primarily on the
1) type of mist extractor present
2) separator geometry
3) flow rates
4) fluid properties

Therefore, $K$ is usually determined from experiments.

Maximum terminal velocities calculated using the $K_{SB}$ factors are for separators normally having a wire-mesh mist extractor and should allow all liquid droplets larger than 10 μm to settle out of the gas.

If no mist extractor is present, multiply $K_{SB}$ by 0.5.
DESIGN CONSIDERATIONS

The following factors must be determined before beginning separator design.

1) Gas and liquids flow rates (minimum, average, and peak).
2) Operating and design pressures and temperatures.
3) Surging or slugging tendencies of the feed streams.
4) Physical properties of the fluids, such as density, viscosity, and compressibility.
5) Designed degree of separation (e.g., removing 100% of particles greater than 10 \( \mu m \)).

In the separator design, it is also worthwhile to clarify two definitions:

**Holdup** is defined as the time it takes to reduce the liquid level from normal (NLL) to low (LLL) while maintaining a normal outlet flow without feed makeup.

**Surge time** is defined as the time it takes for the liquid level to rise from normal (NLL) to high (HLL) while maintaining a normal feed without any outlet flow.

Holdup time is based on the reserve required to maintain good control and safe operation of downstream facilities, whereas surge time is usually based on requirements to accumulate liquid as a result of upstream or downstream variations or upsets, for e.g., slugs.
Design Procedure

TWO-PHASE SEPARATORS

VERTICAL SEPARATORS
1- Calculate the vertical terminal vapor velocity:

\[
U_T = K \left( \frac{\rho_L - \rho_v}{\rho_v} \right)^{1/2} \text{ ft/s}
\]  

(6)

Set \( U_v = 0.75U_T \) for a conservative design.

Calculate the \( K \) value from Table 1.

If there is no mist eliminator, it is recommended to use one half of the above values (2)

Or

the "theoretical" value \( K \) can be calculated from Eq. 5 if the liquid droplet size is known.

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### Table 1: Separator K values

<table>
<thead>
<tr>
<th>Mist Eliminator</th>
<th>( K ) Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>1 ( \leq P \leq 15 )</td>
<td>( K = 0.1821 + 0.0029P + 0.0460 \ln P )</td>
</tr>
<tr>
<td>15 ( \leq P \leq 40 )</td>
<td>( K = 0.35 )</td>
</tr>
<tr>
<td>40 ( \leq P \leq 5,500 )</td>
<td>( K = 0.450 )</td>
</tr>
<tr>
<td>( 0 \leq P \leq 1,500 )</td>
<td>( K = 0.35 - 0.011 \ln P )</td>
</tr>
</tbody>
</table>

\( \rho \), psia

GPPSA

\( C_s = \exp \left( \frac{1}{4} \right) \)

\( Y^2 = 0.411 \) - 2.240\( X \) + 0.0278\( X^2 \) - 1.0055 - 2\( X^3 \) + 5.2013 - 4\( X^4 \)

\( X = \frac{0.955 + 8\rho_v D_0 (\rho_L - \rho_v)}{\mu_L} \)

Notes:

\( D_0 \), ft
\( \rho \), lbm/ft\(^3\)
\( \mu \), cP

1 micron = 3.38084 \( \times 10^4 \) ft
2- Calculate the vapor volumetric flow rate:

\[ Q_v = \frac{W_v}{(3600)(\rho_v)} \quad \text{ft}^3 / \text{s} \quad (7) \]

3- Calculate the Vessel (inside) diameter:

\[ D_{VD} = \left( \frac{4Q_v}{\pi U_v} \right)^{1/2} \quad \text{ft} \quad (8) \]

If there is a mist eliminator, add 3 to 6 in. to \( D_{VD} \) to accommodate a support ring and round up to the next 6 in. increment to obtain \( D \).

If there is no mist eliminator \( D = D_{VD} \).

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4- Calculate the liquid volumetric flow rate:

\[ Q_v = \frac{W_v}{(60)(\rho_v)} \quad \text{ft}^3 / \text{min} \quad (9) \]

5- Select holdup time from Table 2 and calculate the holdup volume:

\[ V_H = (T_H)(Q_v) \quad \text{ft}^3 \quad (10) \]
# Table 2: Liquid holdup and surge times

<table>
<thead>
<tr>
<th>Services</th>
<th>Holdup Times (NLL-HLL)</th>
<th>Surge Time (NLL-LLL)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>min.</td>
<td>min.</td>
</tr>
<tr>
<td>A. Unit Feed Drum</td>
<td>10</td>
<td>5</td>
</tr>
<tr>
<td>B. Separators</td>
<td></td>
<td></td>
</tr>
<tr>
<td>1. Feed to column</td>
<td>5</td>
<td>3</td>
</tr>
<tr>
<td>2. Feed to other drum or tankage</td>
<td></td>
<td></td>
</tr>
<tr>
<td>a) with pump or through exchanger</td>
<td>5</td>
<td>2</td>
</tr>
<tr>
<td>b) without pump</td>
<td>2</td>
<td>1</td>
</tr>
<tr>
<td>3. Feed to fired heater</td>
<td>10</td>
<td>3</td>
</tr>
<tr>
<td>C. Reflux or product accumulator</td>
<td></td>
<td></td>
</tr>
<tr>
<td>1. Reflux only</td>
<td>3</td>
<td>2</td>
</tr>
<tr>
<td>2. Reflux and product</td>
<td>3+</td>
<td>2+</td>
</tr>
<tr>
<td>* based on reflux (3 min.) + appropriate holdup time of overhead product (per B-1, 2)</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

D. Column bottoms

<table>
<thead>
<tr>
<th>Services</th>
<th>Holdup Times (NLL-HLL)</th>
<th>Surge Time (NLL-LLL)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>min.</td>
<td>min.</td>
</tr>
<tr>
<td>1. Feed to another column</td>
<td>5</td>
<td>2</td>
</tr>
<tr>
<td>2. Feed to other drum or tankage</td>
<td></td>
<td></td>
</tr>
<tr>
<td>a) with pump or through exchanger</td>
<td>5</td>
<td>2</td>
</tr>
<tr>
<td>b) without pump</td>
<td>2</td>
<td>1</td>
</tr>
<tr>
<td>3. Feed to fired boiler</td>
<td>5-8</td>
<td>2-4</td>
</tr>
<tr>
<td>* based on reboiler vapor expressed as liquid (3 min.) + appropriate holdup time for the bottom product (per D-1, 2)</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

E. Compressor suction/interstage scrubber

* 3 min between NLL (HLA) and HLSO
* 10 min from bottom tangent line to HLA

F. Fuel gas knockout drum

* 20 ft slug in the incoming fuel gas line between NLL and HLSO

G. Flare knockout drum

* 20 to 30 min. to NLL

<table>
<thead>
<tr>
<th>Personnel</th>
<th>Factor</th>
<th>Instrumentation</th>
<th>Factor</th>
</tr>
</thead>
<tbody>
<tr>
<td>Experienced</td>
<td>1.0</td>
<td>Well instrumented</td>
<td>1.0</td>
</tr>
<tr>
<td>Trained</td>
<td>1.2</td>
<td>Standard instrumented</td>
<td>1.2</td>
</tr>
<tr>
<td>Inexperienced</td>
<td>1.5</td>
<td>Poorly instrumented</td>
<td>1.5</td>
</tr>
</tbody>
</table>
6- If the surge volume is not specified, select a surge time from Table 3 and calculate the surge volume:

\[ V_s = (T_s)(Q_L) \text{ ft}^3 \]  \hspace{1cm} (11)

7- Obtain low liquid level height, \( H_{LLL} \), from Table 3.

<table>
<thead>
<tr>
<th>Vessel diameter</th>
<th>Vertical LLL</th>
<th>Horizontal LLL</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>&lt; 300 psia</td>
<td>&gt; 300 psia</td>
</tr>
<tr>
<td>( \leq 4 \text{ ft} )</td>
<td>15 in.</td>
<td>6 in.</td>
</tr>
<tr>
<td>6 ft</td>
<td>15 in.</td>
<td>6 in.</td>
</tr>
<tr>
<td>8 ft</td>
<td>15 in.</td>
<td>6 in.</td>
</tr>
<tr>
<td>10 ft</td>
<td>6 in.</td>
<td>6 in.</td>
</tr>
<tr>
<td>12 ft</td>
<td>6 in.</td>
<td>6 in.</td>
</tr>
<tr>
<td>16 ft</td>
<td>6 in.</td>
<td>6 in.</td>
</tr>
</tbody>
</table>

8- Calculate the height from low liquid level to normal liquid level:

\[ H_H = \frac{V_H}{(\pi / 4)D_V^2} \text{ ft} \]  \hspace{1cm} (12)

1 ft minimum

9- Calculate the height from normal liquid level to high liquid level (or high level alarm):

\[ H_S = \frac{V_S}{(\pi / 4)D_V^2} \text{ ft} \]  \hspace{1cm} (13)

6 in minimum
10- Calculate the height from high liquid level to the centerline of the inlet nozzle:

\[ H_{LIN} = 12 + d_N \text{ in.} \]  
(with inlet diverter)  \hspace{2cm} (14)

\[ H_{LIN} = 12 + d_N \text{ in.} \]  
(without inlet diverter)

Note: \( d_N \) is calculated as per Table 4.

Table 4  
Inlet nozzle sizing

\[ d_N \geq \left( \frac{4Q_m}{\pi 60 l^2/ \rho_m} \right)^{1/2}, \text{ ft} \]

\[ Q_m = Q_t + Q_v \text{ ft}^3/\text{s} \]

\[ \rho_m = \rho_t \lambda + \rho_v (1 - \lambda), \text{ lb/ft}^3 \]

\[ \lambda = \frac{Q_t}{Q_t + Q_v} \]

11- Calculate the disengagement height from the centerline of the inlet nozzle to:

\[ H_D = 0.5D_V \text{ or a minimum of} \]

\[ H_D = 36 + \frac{1}{2}d_N \text{ in.} \]  
(without mist eliminator)  \hspace{2cm} (15)

\[ H_D = 24 + \frac{1}{2}d_N \text{ in.} \]  
(with mist eliminator)
12- If there is a mist eliminator take 6 in. for the mist eliminator pad and take 1 ft. from the top of the mist eliminator to the top tangent line of the vessel.

13- Calculate the total height, \( H_T \) of the vessel:

\[
H_T = H_{LLL} + H_H + H_S + H_{LIN} + H_D + H_{ME} \quad \text{ft}
\]  

(16)

where \( H_{ME} \) is the height from step 12; if there is no mist eliminator \( H_{ME} = 0 \).
**DESIGN PROCEDURE**

1- Calculate the vapor volumetric flow rate, $Q_V$ using Eq. 7

2- Calculate the liquid volumetric flow rate, $Q_L$ using Eq. 9.

3- Calculate the vertical terminal vapor velocity, $U_T$ using Eq. 8. (K value as per Table 1 for no mist eliminator).
   Set $U_V = 0.75 U_T$ for a conservative design.

4- Select a holdup time from Table 2 and calculate the holdup volume, $V_H$ using Eq. 10.

5- If the surge volume is not specified, select the surge time from Table 2 and calculate the surge volume, $V_S$ using Eq. 11.

---

**DESIGN PROCEDURE**

6- Obtain an estimate of $L/D$ from Table 4 and initially calculate the diameter according to:

$$D = \left( \frac{4(V_H + V_S)}{\pi (0.6) L/D} \right)^{1/3} \text{ ft} \quad (17)$$

(Round to nearest 0.5 ft.)

Calculate the total cross-sectional area

$$A_T = \frac{\pi}{4} D^2 \quad (18)$$

Calculate the low liquid level height, $H_{LLL}$ using Table 3 or

$$H_{LLL} = 0.5D + 7 \text{ in.} \quad (19)$$

where $D$ in ft and round up to the nearest in.

if $D \leq 4' 0''$, $H_{LLL} = 9$ in.
7- Using \( H_{\text{LLL}} / D \), obtain \( A_{\text{LLL}} / A_r \) using Table 5 and calculate the low liquid area, \( A_{\text{LLL}} \).

8- If there is no mist eliminator pad, the minimum height of the vapor disengagement area (\( A_v \)) is the larger of 0.2D or 1 ft.

If there is a mist eliminator pad, the minimum height of the vapor disengagement area is the larger of 0.2D or 2 ft.

Hence, set \( H_v \) to the larger of 0.2D or 2 ft (1 ft if there is no mist eliminator).

Using \( H_v / D \) to obtain \( A_v / A_r \) using Table 6 and calculate \( A_v \).
9- Calculate the minimum length to accommodate the liquid holdup/surge:

\[ L = \frac{V_H + V_S}{A_T - A_V - A_{LLL}} \text{ ft} \]  \hspace{1cm} (20)

10- Calculate the liquid dropout time,

\[ \phi = \frac{H_V}{U_V} \text{ s} \]  \hspace{1cm} (21)

11- Calculate the actual vapor velocity, \( U_{VA} \)

\[ U_{VA} = \frac{Q_V}{A_V} \text{ ft / s} \]  \hspace{1cm} (22)
**DESIGN PROCEDURE**

12- Calculate the minimum length required for vapor-liquid disengagement, \( L_{\text{MIN}} \):

\[
L_{\text{MIN}} = U_{AV} \phi \quad \text{ft} \tag{23}
\]

13- If \( L < L_{\text{MIN}} \), then set \( L = L_{\text{MIN}} \) (Vapor/liquid separation is controlling). This simply results in some extra holdup.

If \( L_{\text{MIN}} > L \), then increase \( H_v \) and repeat from the step 8.

If \( L > L_{\text{MIN}} \), the design is acceptable for vapor/liquid separation.

If \( L > L_{\text{MIN}} \) (Liquid holdup is controlling), \( L \) can only be decreased and \( L_{\text{MIN}} \) increased if \( H_v \) is decreased.

\( H_v \) may only be decreased if it is greater than the minimum specified in the step 8.

---

**DESIGN PROCEDURE**

(Calculations would have to be repeated from the step 8 with reduced \( H_v \)).

Calculate \( L/D \).

If \( L/D > 6.0 \) then increase \( D \) and repeat calculations from the step 6.

If \( L/D < 1.5 \), then decrease \( D \) and repeat calculations from the step 6.

14- Calculate the thickness or the shell and heads according to Table 6.
Table 6: Wall thickness, surface area approximate vessel height.

<table>
<thead>
<tr>
<th></th>
<th>Wall Thickness (in.)</th>
<th>Surface Area (ft²)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Shell</td>
<td>( \frac{P}{2SE-1.2P} + t_l )</td>
<td>( xDL )</td>
</tr>
<tr>
<td>2:1 Elliptical Heads</td>
<td>( \frac{P}{2SE-0.2P} + t_l )</td>
<td>1.09lf</td>
</tr>
<tr>
<td>Hemispherical Heads</td>
<td>( \frac{P}{4SE-0.4P} + t_l )</td>
<td>1.57lf</td>
</tr>
<tr>
<td>Dished Heads</td>
<td>( \frac{0.885PD}{SE-0.1P} + t_l )</td>
<td>0.042lf</td>
</tr>
</tbody>
</table>

Appropriate Vessel Height

\[
W = \left( \frac{490}{f_D} \right) \left( \frac{f_l}{12} \right) (A_t + 2A_\phi)
\]

Notes:
- \( P \), design pressure, psi (typically, operating pressure + (15-30) psi or 10-15%, whichever greater)
- \( T \), design temperature, °F (typically, operating temperature +25-50°F if \( T_{in} > 200°F \), if \( T_{in} = 200°F \), 250°F)
- \( S \), allowable stress, psi (Reference 9)
- \( E \), joint efficiency, (0.6-1.0), 0.85 for spot examined joints, 1.0 for 100% x-ray joints
- \( t_l \), corrosion allowance, in, typically 0 to q in.
- \( A_\phi \), larger of \( t_l \) and \( t_l \) (to nearest 1 in.)

**DESIGN PROCEDURE**

15- Calculate the surface area of the shell and heads according to Table 6.

16- Calculate the approximate vessel weight according to Table 6.

17- Increase and decrease the diameter by 6 in. increments and repeat the calculations until L/D has ranged from 1.5 to 6.0.

18- With the optimum vessel size (minimum weight), calculate normal and high liquid levels:

\[
A_{NLL} = A_{LLL} + \frac{V_H}{L} \quad (24)
\]

With \( A_{NLL}/A_t \) obtain \( H_{NLL} \) from Table 6

\[
H_{NLL} = D - H_V \quad (25)
\]