COMPUTER APPLICATION FOR SEPARATORS DESIGN

DIEGO JORGE CRUZ GARCÍA  JUNIO 2015
ESCUELA TÉCNICA Y SUPERIOR DE INGENIEROS DE MINAS Y ENERGÍA

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COMPUTER APPLICATION FOR SEPARATORS DESIGN

Realizado por

D. Diego Jorge Cruz García

Dirigido por

D. Luis Jesús Fernández Gutiérrez del Álamo

Departamento de Energía y Combustibles
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1. Resumen & Abstract

Resumen

Programa informático desarrollado en plataforma EXCEL (VBA) y dirigido al diseño de Separadores de dos y tres fases, verticales y horizontales.

El programa de ordenador o aplicación tiene la capacidad de determinar las propiedades físicas del fluido, utilizando diferentes correlaciones sobre la base del “Black Oil Model”, con dichas propiedades el Programa predice el tipo de flujo presente. Si el tipo de flujo es “Slug Flow” el programa determinara las dimensiones del “Slug catcher” necesario. Bajo las condiciones de funcionamiento existentes el programa diseñará el separador elegido: dos o tres fases, vertical u horizontal.

Por último, la aplicación informática estimará el coste del equipo.

Abstract

Computer program developed in EXCEL (VBA) platform and aimed for the design of Two-Phase, Three-Phase, Vertical or Horizontal Separators.

The computer Program or Application has the capability to determine the fluid physical properties utilizing different correlations on the basis of the Black Oil Model, with those Properties the Program will predict the Flow Regime present. If the flow regime is Slug Flow the program will determine the necessary slug catcher dimensions. Under certain operational conditions the program will design the selected: Two-Phase or Three-Phase, Vertical or Horizontal Separator.

Finally the computer Application will estimate the cost of the equipment.
2. Preliminary considerations

From now on, the text will be written in English because all the sources: documents, correlations, Formulas, texts, slides, etc. used in this Project have been collected from the Oilfield literature from the United States of America.

Also the units used in this Project will be the ones from the Oilfield.

The conversions from SI units to Oilfield units are available in the Annex C: Conversions from SI units to Oilfield units.

The decimal separator will be (.) and the thousands separator will be (,).
COMPUTER APPLICATION FOR
SEPARATORS DESIGN

DOCUMENT I: REPORT
3. Document I: Report

3.1. Chapter I: Objectives and Scope

The objective of this project is to provide a user friendly Computer Application / Program for the preliminary design and cost estimation of the necessary equipment in an oil / gas well from the wellhead to the processing facility.

*Figure 1* shows a schematic of one of the possible configurations:

![Figure 1: Configuration Example](image)

**Source:** CNGS group

The program will allow choosing between different options in order to provide solutions for the different kind of oil / gas wells, from Heavy to light oils, from Offshore to Onshore facilities.

The Program could be used as an accurate first approximation to calculate the dimensions and cost of the required Slug Catcher and Separator for a specific well.
3.2. Chapter II: Project Development

3.2.1. Black Oil Model

The Black Oil Model is based on the assumption that hydrocarbon mixtures can be completely described using two pseudocomponents: stock tank oil and produced gas. Black Oil properties can be measured in the laboratory using a combination of Flash and Differential vaporization experiments, however, in practice, these experiments can be time consuming and expensive and required PVT properties are often estimated using empirical PVT correlations.

The calculation of these properties will be the FIRST STEP of the program.

In this project the first approximation was made with the black oil model correlations proposed in the paper *SPE 23556*, *Kartoatmodjo and Schmidt’s (1991)*.

After attending a Robert Sutton talk at The University of Tulsa where he talked about the reliability of the different sets of correlations, he shows the big error Kartoatmodjo’s viscosity correlation has for High API Gravity crudes, so it was decided to look for another paper which modifies the correlations in relation to API gravities. Was found that the paper which best fits this requirement was *SPE-28904, Reliability Analysis on PVT Correlations by Giambattista De Ghetto, Francesco Paone and Marco Villa* and those correlations have been used to calculated the PVT properties.

The required input data for all correlations are the oil, gas and water flowrates, namely, Gas Flowrate, Qg [MSCFD], Oil Flowrate qo [BPD] and Water Flowrate qw [BPD] respectively. Alternatively, instead of the the Gas Flowrate, Qg, the Produced Gas-oil ratio, R_p [scf/stb], can be specified. The correlations will be used for a specific Pressure P [psia] and Temperature T [ºF] of interest (in-situ) conditions.

Other required data are the Fluid Properties: Dissolved Gas at Bubble Point Conditions Rsb [scf/stb] or Bubble point pressure Pb [psia], Gas Specific Gravity SG [-] and the oil API gravity[ºAPI]

The Separator Temperature Tsep [ºF] and Pressure Psep [psia] for which the specific gravities are given are also necessary.
The required input data for all correlations are shown as it appears in the program in *Table 1 & Table 2*.

**Table 1: Operational Conditions Input Data**

<table>
<thead>
<tr>
<th>Operational Conditions (in-situ)</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Parameter</strong></td>
</tr>
<tr>
<td>Pressure</td>
</tr>
<tr>
<td>Temperature</td>
</tr>
<tr>
<td>Produced Gas-oil ratio</td>
</tr>
<tr>
<td>Gas flowrate</td>
</tr>
<tr>
<td>Oil Flowrate</td>
</tr>
<tr>
<td>Water Flowrate</td>
</tr>
</tbody>
</table>

**Table 2: Fluid Properties Input Data**

<table>
<thead>
<tr>
<th>Fluid Properties</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Parameter</strong></td>
</tr>
<tr>
<td>Bubble point pressure</td>
</tr>
<tr>
<td>Disolved Gas at Bubble Point Conditions</td>
</tr>
<tr>
<td>Gas Specific gravity</td>
</tr>
<tr>
<td>API</td>
</tr>
<tr>
<td>Separator Temperature</td>
</tr>
<tr>
<td>Separator Pressure</td>
</tr>
</tbody>
</table>

For specific in-situ conditions, correlations may be applied to estimate the Bubble point pressure Pb [psia], Solution Gas-Oil Ratio Rs [scf/stb], Formation Volume Factor Bo [bbl/stb], Dead Oil Viscosity VisOD [cP] and Live Oil Viscosity VisOL [cP]. These data are required for further calculations in the Program.

**The Physical Phenomena**

The physical phenomena associated with phase changes at constant temperature and increasing pressure for hydrocarbon mixtures are depicted schematically in *Figure 2*. *Figure 2-A* depicts the oil and gas phases at atmospheric pressure, representing stock tank conditions, where all the gas is in the vapor phase, and the oil is “dead”. Upon
increasing the pressure, Figure 2-B, some gas dissolves in the liquid phase. The amount of gas in solution is represented by Solution Gas-Oil Ratio $Rs \ [scf/stb]$.

Thus, at the higher pressure, less free gas is present, and the volume of the free gas is reduced due to its compressibility; at the same time, liquid phase volume increases due to the addition of dissolved gas. The ratio of the in-situ to standard conditions volumes of the oil phase is the oil Formation Volume Factor $Bo \ [bbl/stb]$. The Oil Formation Volume Factor is always greater than 1.0 at pressures above atmospheric, provided gas is present to dissolve in the liquid phase. The gas formation volume factor, denoted by $Bg \ [ft^3/scf]$, however, is much less than 1 due to the large compressibility of the gas phase. The process of free gas dissolving continues as pressure increases, as long as the pressure remains below the Bubble point pressure. The oil under these conditions is referred to as “saturated”, since it contains all the gas it can dissolve at the particular pressure and temperature. At the Bubble point pressure, the last bubble of gas dissolves, resulting in single-phase liquid with no free gas, as shown in Figure 2-C. Under these conditions, the Solution Gas-Oil Ratio attains a constant value. For wells producing above the bubble point pressure, the Solution Gas-Oil Ratio $Rs$ is equal to the producing gas-oil ratio. Further increase in the pressure results in slight compression of the oil. Oil reservoirs that exist above the bubble point pressure as referred to as “undersaturated”, since they have the capacity of dissolving more free gas if it were present Figure 2-D.
PVT Properties Calculations

The paper *SPE-28904, Reliability Analysis on PVT Correlations by Giambattista De Ghetto, Francesco Paone and Marco Villa* evaluates the reliability of the most common empirical correlations used for determining reservoir fluid properties whenever laboratory PVT data are not available.

The reliability of the correlations has been evaluated against a set of 195 crude oil samples collected from the Mediterranean Basin, and the Persian Gulf and the North Sea. About 3700 measured data points have been collected and investigated.

For all the correlations, the following statistical parameters have been calculated a) relative deviation between estimated and experimental values b) average absolute percent error c) standard deviation.

Oil samples have been divided into the following four different API gravity classes *Table 3:*

<table>
<thead>
<tr>
<th>Oil Type</th>
<th>°API</th>
</tr>
</thead>
<tbody>
<tr>
<td>Extra-heavy</td>
<td>API ≤ 10</td>
</tr>
<tr>
<td>Heavy</td>
<td>10 &lt; API ≤ 22.3</td>
</tr>
<tr>
<td>Medium</td>
<td>22.3 &lt; API ≤ 31.1</td>
</tr>
<tr>
<td>Light</td>
<td>API &gt; 31.1</td>
</tr>
</tbody>
</table>

The best correlations both for each class and for the whole range of API gravity have been evaluated for each oil-property.

The functional forms of the correlations that gave the best results for each oil property have been used for finding a better correlation with average errors reduced by 5-10 %. In particular, for extra-heavy oils, since no correlations are available in literature a special investigation has been performed and new equations are proposed.
1. **Bubble point pressure calculation**

The bubble point pressure is the pressure at which the last gas bubble dissolves in the liquid phase or the pressure at which the first gas bubble “comes out” of solution. Thus, the bubble point pressure can be solved from solution gas-oil ratio equation substituting \( R_s \) for *Disolved Gas at Bubble Point Conditions* \( R_{sb} \):

Extra-heavy oils: Standing’s Correlation

\[
P_b = 18 \left( \frac{R_{sb}}{\gamma_G} \right)^{0.83} 10^{0.0009(T)} 10^{0.0125(API)}
\]

Heavy oils: Modified Standing’s Correlation

\[
P_b = 15.7286 \left( \frac{R_{sb}}{\gamma_G} \right)^{0.7885} 10^{0.0020(T)} 10^{0.0142(API)}
\]

Medium oils: Modified Kartoatmodjo’s Correlation

\[
P_b = \left( \frac{R_{sb}}{0.09902 \cdot (\gamma_{Gcorr})^{0.2181} \cdot 10^{(7.2153 + API)/(T + 460)}} \right)^{0.9997}
\]

\[
\gamma_{Gcorr} = \gamma_G \cdot \left( 1 + 0.1595 \cdot API^{0.4078} \cdot T_{sep}^{-0.2466} \cdot \log \left( \frac{P_{sep}}{114.7} \right) \right)
\]

Light oils: Modified Standing’s Correlation

\[
P_b = 31.7648 \left( \frac{R_{sb}}{\gamma_G} \right)^{0.7857} 10^{0.0009(T)} 10^{0.0148(API)}
\]

Where:

\( P_b \): Bubble Point Pressure, Psia

\( R_{sb} \): Disolved Gas at Bubble Point Conditions, scf/stb

\( \gamma_G \): Gas Specific gravity, [-]

\( T \): in-situ Temperature, °F

\( API \): API gravity, °API
\( \gamma_{G\text{corr}} \): Gas Spacific gravity at separator conditions, [-]

\( T_{\text{sep}} \): Temperature at separator conditions, °F

\( P_{\text{sep}} \): Pressure at separator conditions, Psia

2. Solution gas-oil Ratio calculation

The solution gas-oil ratio, \( R_s \), represents the amount of gas that is dissolved in the liquid phase at a given pressure and temperature. Note that the amount is expressed in terms of scf/stbo, and must be converted properly to in-situ conditions. Also, one must realize that this dissolved gas is now part of the liquid phase. As shown schematically in Figure 3, the solution gas-oil ratio is zero at atmospheric pressure conditions; it increases with pressure and reaches its maximum value at the bubble-point pressure, and remains constant for pressures above the bubble-point pressure.

\[ Rs = \gamma_g \cdot \left( \frac{P}{10.7025} \right) \cdot 10^{(0.0169 \cdot API - 0.00156 \cdot T)} \]  

*Above bubble point Pressure \( P = P_b \)

Extra-heavy oils: Modified Standing´s Correlation

\[ Rs = \gamma_{G\text{corr}} \cdot \frac{P^{1.2057}}{56.434} \cdot 10^{(10.9267 \cdot API/(T+460))} \]

Heavy oils: Modified Vasquez-Beggs Correlation

\[ \gamma_{G\text{corr}} = \gamma_g \cdot \left( 1 + 0.5912 \cdot API \cdot T_{\text{sep}} \cdot \log \left( \frac{P_{\text{sep}}}{114.7} \right) \right) \cdot 10^{-4} \]
Medium oils: Modified Kartoatmodjo’s Correlation

\[ Rs = 0.10084 \cdot \gamma_{G_{corr}}^{0.2556} \cdot p^{0.9868} \cdot 10^{(7.4576 \cdot API/(T+460))} \]

\[ \gamma_{G_{corr}} = \gamma_G \cdot \left( 1 + 0.1595 \cdot API^{0.4078} \cdot T_{sep}^{-0.2466} \cdot \log \left( \frac{P_{sep}}{114.7} \right) \right) \]

Light oils: Modified Standing’s Correlation

\[ Rs = 0.01347 \cdot \gamma_{G_{corr}}^{0.3873} \cdot p^{1.7115} \cdot 10^{(12.753 \cdot API/(T+460))} \]

\[ \gamma_{G_{corr}} = \gamma_G \cdot \left( 1 + 0.1595 \cdot API^{0.4078} \cdot T_{sep}^{-0.2466} \cdot \log \left( \frac{P_{sep}}{114.7} \right) \right) \]

Where:

- **P**: in-situ Pressure, Psia
- **Pb**: Bubble Point Pressure, Psia
- **Rs**: Solution gas-oil Ratio, scf/stb
- **\( \gamma_G \)**: Gas Specific gravity, [-]
- **T**: in-situ Temperature, °F
- **API**: API gravity, °API
- **\( \gamma_{G_{corr}} \)**: Gas Specific gravity at separator conditions, [-]
- **T_{sep}**: Temperature at separator conditions, °F
- **P_{sep}**: Pressure at separator conditions, Psia

3. **Oil compressibility calculation**

Isothermal compressibility is the change in volume of a system as the pressure changes while temperature remains constant.

Extra-heavy oils: Modified Vasquez-Beggs Correlation

\[ Co = \frac{-889.6 + 3.1374 \cdot Rs + 20 \cdot T - 627.3 \cdot \gamma_{G_{corr}} - 81.4476 \cdot API}{P \cdot 10^5} \]
\[ Y_{Gcorr} = Y_G \cdot \left( 1 + 0.5912 \cdot API \cdot T_{sep} \cdot \log\left(\frac{P_{sep}}{114.7}\right) \right) \cdot 10^{-4} \]

Heavy oils: Modified Vasquez-Beggs Correlation

\[ C_o = \frac{-2841.8 + 2.9646 \cdot R_s + 25.5439 \cdot T - 1230.5 \cdot Y_{Gcorr} - 41.91 \cdot API}{P \cdot 10^5} \]

\[ Y_{Gcorr} = Y_G \cdot \left( 1 + 0.5912 \cdot API \cdot T_{sep} \cdot \log\left(\frac{P_{sep}}{114.7}\right) \right) \cdot 10^{-4} \]

Medium oils: Modified Vasquez-Beggs Correlation

\[ C_o = \frac{-705.288 + 2.2246 \cdot R_s + 26.0644 \cdot T - 2080.823 \cdot Y_{Gcorr} - 9.6807 \cdot API}{P \cdot 10^5} \]

\[ Y_{Gcorr} = Y_G \cdot \left( 1 + 0.5912 \cdot API \cdot T_{sep} \cdot \log\left(\frac{P_{sep}}{114.7}\right) \right) \cdot 10^{-4} \]

Light oils: Modified Labedi’s Correlation

\[ C_o = \left( 10^{-6.1646} \cdot B_o^{1.8789} \cdot API^{0.3646} \cdot T^{0.1966} \right) - \left( 1 - \frac{P_{b}}{P} \right) \cdot \left( 10^{-8.98} \cdot B_o^{3.9392} \cdot T^{1.349} \right) \]

Where:

- Co: Oil compressibility
- P: in-situ Pressure, Psia
- Pb: Bubble Point Pressure, Psia
- Rs: Solution gas-oil Ratio, scf/stb
- \( Y_G \): Gas Specific gravity, [-]
- T: in-situ Temperature, °F
- API: API gravity, °API
- \( Y_{Gcorr} \): Gas Spacific gravity at separator conditions, [-]
- Tsep: Temperature at separator conditions, °F
- Psep: Pressure at separator conditions, Psia
- Bo: Oil Formation Volume Factor, bbl/stb
4. Oil Formation Volume Factor

The oil formation volume factor is defined as the ratio of the in-situ oil volume to the standard conditions oil volume.

The correlations which estimate this property seem to be generally valid as the analysis of the range led to an average error decrease of less than the percentage point.

Kartoatmodjo’s Correlation will be used.

Below bubble point pressure

\[ \text{Bob} = 0.98496 + 0.0001 \cdot f^{1.5} \]

\[ f = R_s^{0.755} \cdot Y_{\text{Gcorr}}^{0.25} \cdot Y_{\text{Go}}^{-1.5} + 0.45 \cdot T \]

\[ Y_{\text{Gcorr}} = Y_G \cdot \left(1 + 0.1595 \cdot API^{0.4078} \cdot T_{sep}^{-0.2466} \cdot \log\left(\frac{P_{sep}}{114.7}\right)\right) \]

\[ Y_{\text{Go}} = \frac{141.5}{131.5 + API} \]

Above bubble point pressure

\[ Bo = \text{Bob} \cdot e^{Co \cdot (Pb - P)} \]

Where:

- \( Bo \): Oil Formation Volume Factor, bbl/stb
- \( Bob \): Oil Formation Volume Factor at the bubble point pressure, bbl/stb
- \( Co \): Oil compressibility
- \( P \): in-situ Pressure, Psia
- \( Pb \): Bubble Point Pressure, Psia
- \( Rs \): Solution gas-oil Ratio, scf/stb
- \( Y_G \): Gas Specific gravity, [-]


\( \gamma_{Go} \): Oil Specific gravity, [-]

T: in-situ Temperature, °F

API: API gravity, °API

\( \gamma_{Gcorr} \): Gas Specific gravity at separator conditions, [-]

Tsep: Temperature at separator conditions, °F

Psep: Pressure at separator conditions, Psia

5. Gas Formation Volume Factor

The gas formation volume factor is used to relate the volume of gas, as measured at reservoir conditions, to the volume of the gas as measured at standard conditions. This gas property is then defined as the actual volume occupied by a certain amount of gas at a specified pressure and temperature, divided by the volume occupied by the same amount of gas at standard conditions. In an equation form, the relationship is expressed as:

\[
B_g = 0.0283 \cdot \frac{z \cdot T}{p}
\]

Where:

\( B_g \): Gas Formation Volume Factor, ft³/scf

Z: Compressibility factor, [-]

T: in-situ Temperature, °F

P: in-situ Pressure, Psia

6. Oil Density

The prediction of live oil density, at in-situ pressure and temperature, can be determined by an overall mass balance on the gas and the liquid phases, with a basis of 1 stbo. Below bubble point pressure:

\[
DenOb = \frac{\gamma_o (62.4) + \gamma_{Go} (0.0764) R_s}{5.614} B_o
\]
Above the bubble point pressure, oil density can be calculated in a manner similar to the oil formation volume factor:

\[ y_{GD} = \frac{12.5 + API}{50} - 3.5715 \cdot 10^{-6} \cdot API \cdot Rs \]

\[ DenO = DenOb \cdot e^{co \cdot (P_s - P)} \]

Where:

\[ DenO \]: Oil Density, lb/ft³

\[ DenOb \]: Oil Density at Bubble point Pressure, lb/ft³

\[ Bo \]: Oil Formation Volume Factor, bbl/stb

\[ Co \]: Oil compressibility

\[ P \]: in-situ Pressure, Psia

\[ Pb \]: Bubble Point Pressure, Psia

\[ Rs \]: Solution gas-oil Ratio, scf/stb

\[ y_{GO} \]: Oil Specific gravity, [-]

\[ y_{GD} \]: Dissolved Gas Specific gravity, [-]

\[ API \]: API gravity, °API

7. **Dead Oil Viscosity**

Dead oil viscosity is the one where all the gas is in the vapor phase and the oil is “dead”.

**Extra-heavy oils**: Modified Egbogah-Jack’s Correlation

\[ VisOD = \left(10^{1.0296 - 0.012619 \cdot API - 0.61748 \cdot Log(T)}\right) - 1 \]

**Heavy oils**: Modified Egbogah-Jack’s Correlation

\[ VisOD = \left(10^{2.06492 - 0.0179 \cdot API - 0.70226 \cdot Log(T)}\right) - 1 \]
Medium oils: Modified Kartoatmodjo’s Correlation

\[ \text{VisOD} = 220.15 \cdot 10^9 \cdot T^{-3.5560 \cdot \log(API)} \cdot 12.5428 \cdot \log(T) - 45.787 \]

Light oils: Modified Egbogah-Jack’s Correlation

\[ \text{VisOD} = \left( 10^{1.67083 - 0.017628 \cdot API - 0.61304 \cdot \log(T) \right) - 1 \]

Where:

- \( \text{VisOD} \): Dead oil Viscosity, cp
- \( T \): in-situ Temperature, °F
- \( \text{API} \): API gravity, °API

8. **Live Oil Viscosity**

Live Oil Viscosity provides a measure of a fluid’s internal resistance to flow.

The viscosity of the gas saturated oil is found as a function of dead oil viscosity and solution gas-oil ratio. Undersaturated oil viscosity is determined as a function of gas saturated oil viscosity and pressure above bubble point pressure (P > Pb).

**Gas saturated oil Viscosity P < Pb**

Extra-heavy oils: Modified Kartoatmodjo’s Correlation

\[ \text{VisOL} = 2.3945 + 0.8927 \cdot f + 0.001567 \cdot f^2 \]

\[ f = (-0.0335 + 1.0785 \cdot 10^{-0.000845 \cdot Rs}) \cdot \text{VisOD}^{0.5798 + 0.3432 \cdot y} \]

\[ y = 10^{-0.00081 \cdot Rs} \]

Heavy oils: Modified Kartoatmodjo’s Correlation

\[ \text{VisOL} = -0.6311 + 1.078 \cdot f - 0.003653 \cdot f^2 \]

\[ f = (0.2478 + 0.6114 \cdot 10^{-0.000845 \cdot Rs}) \cdot \text{VisOD}^{0.4731 + 0.5158 \cdot y} \]

\[ y = 10^{-0.00081 \cdot Rs} \]

Medium oils: Modified Kartoatmodjo’s Correlation
Light oils: Modified Labedi's Correlation

$$ VisOL = 0.0132 + 0.9821 \cdot f - 0.005215 \cdot f^2 $$

$$ f = (0.2038 + 0.8591 \cdot 10^{-0.000845 \cdot Rs}) \cdot VisOD^{0.3855+0.5664 \cdot y} $$

$$ y = 10^{-0.00081 \cdot Rs} $$

Where:

- \( VisOL \): Live oil viscosity, cp
- \( VisOD \): Dead oil Viscosity, cp
- \( Rs \): Solution gas-oil Ratio, scf/stb

Undersaturated oil viscosity \( P > Pb \)

Extra-heavy oils: Modified Labedi's Correlation

$$ VisOL = VisOLS \cdot \left[ \left(1 - \frac{P}{Pb}\right) \cdot \left(10^{-2.19 \cdot VisOD^{1.055} \cdot Pb^{0.3132}} \right) \right] $$

Heavy oils: Modified Kartoatmodjo's Correlation

$$ VisOL = 0.9886 \cdot VisOLS + 0.002763 \cdot (P - Pb) \cdot $$

$$ \cdot (-0.01153 \cdot VisOLS^{1.7933} + 0.0316 \cdot VisOLS^{1.5939}) $$

Medium oils: Modified Labedi's Correlation

$$ VisOL = VisOLS \cdot \left[ \left(1 - \frac{P}{Pb}\right) \cdot \left(10^{-2.19 \cdot VisOD^{1.055} \cdot Pb^{0.3132}} \right) \right] $$

Light oils: Modified Labedi's Correlation

$$ VisOL = VisOLS \cdot \left[ \left(1 - \frac{P}{Pb}\right) \cdot \left(10^{-2.19 \cdot VisOD^{1.055} \cdot Pb^{0.3132}} \right) \right] $$

Where:

- \( VisOL \): Live oil viscosity, cp
**VisOls**: Live gas saturated oil viscosity, cp

**VisOD**: Dead oil Viscosity, cp

P: in-situ Pressure, Psia

Pb: Bubble Point Pressure, Psia

*VBA CODE FOR PVT CALCULATIONS AVAILABLE IN: Annex B.1: DeGhetto Module*

**Gas-Liquid flow Variables Calculations**

A phase refers to the solid, liquid, or vapor state of matter. A multiphase flow is the flow of a mixture of phases such as gases (bubbles) in a liquid, or liquid (droplets) in gases, and so on

Phases in Production System: Gas, Oil (Liquid), Water (Liquid) and Solid

Gas–liquid flows occur in many applications. The motions of bubbles in a liquid as well as droplets in a conveying gas stream are examples of gas–liquid flows. Gas–liquid flows in pipes can assume several different configurations ranging from bubbly flow to annular flow, in which there is a liquid layer on the wall and a droplet laden gaseous core flow.

Gas-Liquid flow Variables: Superficial Velocity, (m/s)

Superficial velocity of a phase is the velocity, which would occur if only that phase flow alone in the pipe.

For this calculations the diameter of the pipe is needed as an input

Superficial Liquid velocity:

\[ V_{SL} = \frac{q_L}{A} \]

\[ q_L = (q_o + q_w) \cdot \frac{5.615}{86400} \]

\[ A = \frac{\pi}{4} \cdot \left(\frac{d}{12}\right)^2 \]
Where:

- $V_{SL}$: Superficial Liquid velocity, ft/s
- $q_L$: Liquid Flowrate, BPD
- $q_o$: Oil Flowrate, BPD
- $q_w$: Water Flowrate, BPD
- $A$: Cross sectional area, $ft^2$
- $d$: Pipe diameter, in

Superficial Gas velocity:

$$V_{SG} = \frac{q_g}{A}$$

$$q_g = \left(q_g \cdot 10^3\right) \cdot \frac{Bg}{86400}$$

$$A = \frac{\pi}{4} \cdot \left(\frac{d}{12}\right)^2$$

Where:

- $V_{SG}$: Superficial gas velocity, ft/s
- $q_g$: Gas Flowrate, MSCFD
- $A$: Cross sectional area, $ft^2$
- $B_g$: Gas Formation Volume Factor, ft3/scf
- $d$: Pipe diameter, in

If $q_g$ is unknown the program will calculate $V_{SG}$ as follows:

$$V_{SG} = \frac{q_L}{A}$$

$$q_L = (q_o + q_w) \cdot (R_p - R_s) \cdot \frac{B_g}{86400}$$
\[ A = \frac{\pi}{4} \cdot \left( \frac{d}{12} \right)^2 \]

Where:

\( V_{SL} \): Superficial Liquid velocity, ft/s

\( q_g \): Gas Flowrate, MSCFD

\( A \): Cross sectional area, \( ft^2 \)

\( Bg \): Gas Formation Volume Factor, ft3/scf

\( Rs \): Solution gas-oil Ratio, scf/stb

\( R_P \): Produced Gas-oil ratio, scf/stb

\( q_L \): Liquid Flowrate, BPD

\( q_o \): Oil Flowrate, BPD

\( q_w \): Water Flowrate, BPD

\( d \): Pipe diameter, in

*VBA CODE FOR GAS PROPERTIES CALCULATIONS AVAILABLE IN:

Sheet2 (BlackOilMode) code

**Gas Properties Calculations**

In dealing with gases at a very low pressure, the ideal gas relationship is a convenient and generally satisfactory tool. At higher pressures, the use of the ideal gas equation-of-state may lead to errors as great as 500%, as compared to errors of 2–3% at atmospheric pressure.

Basically, the magnitude of deviations of real gases from the conditions of the ideal gas law increases with increasing pressure and temperature and varies widely with the composition of the gas. Real gases behave differently than ideal gases. The reason for this is that the perfect gas law was derived under the assumption that the volume of molecules is insignificant and that no molecular attraction or repulsion exists between them. This is not the case for real gases. Numerous equations-of-state
have been developed in the attempt to correlate the pressure-volume-temperature variables for real gases with experimental data. In order to express a more exact relationship between the variables p, V, and T, a correction factor called the gas compressibility factor, gas deviation factor, or simply the z-factor, must be introduced into the ideal gas relationship to account for the departure of gases from ideality.

1. Gas compressibility Factor

The gas compressibility factor \( Z \) is a dimensionless quantity and is defined as the ratio of the actual volume of n-moles of gas at T and p to the ideal volume of the same number of moles at the same T and P.

Studies of the gas compressibility factors for natural gases of various compositions have shown that compressibility factors can be generalized with sufficient accuracies for most engineering purposes when they are expressed in terms of the following two dimensionless properties: Pseudo-reduced pressure and Pseudo-reduced temperature.

\[
Tr = \frac{T}{T_c}
\]

\[
Pr = \frac{P}{P_c}
\]

\[
P_c = 756.8 - 131 \cdot \gamma_G - 3.6 \cdot \gamma_G^2
\]

\[
T_c = 169.2 + 349.5 \cdot \gamma_G - 74 \cdot \gamma_G^2
\]

Where:

\( Tr \): Pseudo-reduced temperature, [-]

\( Pr \): Pseudo-reduced pressure, [-]

\( T \): in-situ Temperature, °R

\( P \): in-situ Pressure, Psia

\( P_c \): Pseudo-critical pressure, Sutton 1985 Correlation, [-]

\( T_c \): Pseudo-critical Temperature, Sutton 1985 Correlation, [-]

\( \gamma_G \): Gas Specific gravity, [-]
The Z factor will be calculated by The Hall-Yarborough Method.

Hall and Yarborough (1973) presented an equation-of-state that accurately represents the Standing and Katz z-factor chart. The proposed expression is based on the Starling-Carnahan equation-of-state. The coefficients of the correlation were determined by fitting them to data taken from the Standing and Katz z-factor chart. Hall and Yarborough proposed the following mathematical form:

\[
z = \frac{0.06125 \cdot Pr \cdot t \cdot \exp(-1.2(1 - t)^2)}{y}
\]

Where:

- \(Z\): Compressibility factor, [-]
- \(t = 1/Tr\)
- \(Tr\): Pseudo-reduced temperature, [-]
- \(Pr\): Pseudo-reduced pressure, [-]
- \(y\): the reduced density that can be obtained as the solution of the following equation by the Newton’s Method

\[
f(y) = a + b + c + d = 0
\]

\[
a = -0.06125 \cdot Pr \cdot t \cdot \exp(-1.2 \cdot (1 - t)^2)
\]

\[
b = \left(\left((-y + 1)y + 1\right)y + 1\right) \cdot y/(1 - y)^3
\]

\[
c = -((4.58 \cdot t - 9.76) \cdot t + 14.76 \cdot t \cdot y^2
\]

\[
d = ((42.4 \cdot t - 242.2) \cdot t + 90.7) \cdot t \cdot y^{2.18 + 2.82 \cdot t}
\]

2. **Gas Density**

Gas density is defined as the mass per unit of volume. Considering the molecular weight of the gas as the product of specific gravity and air molecular weight:

\[DenG = 2.7 \times \frac{Pr \cdot y_g}{z \cdot T}\]

Where:
\textit{DenG}: Gas density, lb/ft³

\textit{Z}: Compressibility factor, [-]

\textit{T}: in-situ Temperature, °R

\textit{P}: in-situ Pressure, Psia

\textit{γ}_G: Gas Specific gravity, [-]

3. Gas Viscosity

Viscosity is a measure of a fluids internal resistance to flow. The viscosity of a natural gas, expected to increase with both pressure and temperature. It is usually several orders of magnitude smaller than that of oil or water. Gas is much more mobile in the reservoir than either oil or water.

The most commonly used correlation to calculate gas viscosity is Lee et al. (1966)

\[ M = γ_G \cdot 28.97 \]

\[ X = 3.448 + 0.01009 \cdot M + \frac{986.4}{T} \]

\[ K = \frac{(9.379 + 0.01607 \cdot M)T^{1.5}}{209.2 + 19.26 \cdot M + T} \]

\[ y = 2.447 - 0.2224 \cdot X \]

\[ VisG = 10^{-4} \cdot K \cdot exp(X \cdot DenG^y) \]

Where:

\textit{VisG}: Gas viscosity, cp

\textit{DenG}: Gas density, lb/ft³

\textit{γ}_G: Gas Specific gravity, [-]

\textit{Z}: Compressibility factor, [-]

\textit{T}: in-situ Temperature, °R

\textit{P}: in-situ Pressure, Psia
3.2.2. Flow Pattern Prediction

The fundamental difference between single-phase flow and gas-liquid two-phase flow is the existence of flow patterns or flow regimes in two-phase flow. The term flow pattern refers to the geometrical configuration of the gas and the liquid phases in the pipe. When gas and liquid flow simultaneously in a pipe, the two phases can distribute themselves in a variety of flow configurations. The flow configurations differ from each other in the spatial distribution of the interface, resulting in different flow characteristics, such as velocity and holdup distributions.

The existing flow pattern in a given two-phase flow system depends on the variables like: Operational parameters, namely, gas and liquid flow rates, geometrical variables, including pipe diameter and inclination angle and the physical properties of the two phases, i.e., gas and liquid densities, viscosities, and surface tension.

Determination of flow patterns is a central problem in two-phase flow analysis.

In the past, there has been a lack of agreement between two-phase flow investigators on the definition and classification of flow patterns. Some investigators detailed as many flow patterns as possible, while others try to define a set with minimum flow patterns. The disagreement was mainly caused by the complexity of the flow phenomena and to the fact that the flow patterns were usually determined subjectively by visual observations. Also, the flow patterns are dependent on the inclination angle, and usually they were reported for either one inclination or a narrow range of inclination angles.

In recent years, there has been a trend to define an acceptable set of flow patterns. On the one hand, the set must be minimal, but on the other hand, it must include acceptable definitions with minor changes.

Also, it must apply to all the range of inclination angles. An attempt to define an acceptable set of flow patterns has been made by Shoham (1982). The definitions are based on experimental data acquired over the entire range of inclination angles, namely horizontal flow, upward and downward inclined flow, and upward and
downward vertical flow. Figure 4 shows the flow patterns existing in horizontal and near-horizontal pipes and Figure 5 shows the flow patterns existing in vertical and sharply inclined pipes.

Source: Mechanistic modeling of gas-liquid two-phase flow in pipes

Figure 4: Flow patterns existing in horizontal and near-horizontal pipes

Following are the definitions and classifications of the flow.

**Horizontal and Near-Horizontal Flow**

The existing flow patterns in these configurations can be classified as Stratified flow (Stratified-Smooth and Stratified-Wavy), Intermittent flow (Slug flow and Elongated-Bubble flow), Annular flow and Dispersed-Bubble flow

**Stratified Flow (ST).** This flow pattern occurs at relatively low gas and liquid flow rates. The two phases are separated by gravity, where the liquid-phase flows at the bottom of the pipe and the gas phase on the top. The Stratified flow pattern is subdivided into Stratified-Smooth (SS), where the gas-liquid interface is smooth, and Stratified-Wavy (SW), occurring at relatively higher gas rates, at which stable waves form at the interface.
**Intermittent Flow (I):** Intermittent flow is characterized by alternate flow of liquid and gas. Plugs or slugs of liquid, which fill the entire pipe cross-sectional area, are separated by gas pockets, which contain a stratified liquid layer flowing along the bottom of the pipe. The mechanism of the flow is that of a fast moving liquid slug overriding the slow moving liquid film ahead of it. The liquid in the slug body may be aerated by small bubbles, which are concentrated towards the front of the slug and at the top of the pipe. The Intermittent flow pattern is divided into **Slug (SL)** and **Elongated-Bubble (EB)** patterns. The flow behavior of Slug and Elongated-Bubble patterns are the same with respect to the flow mechanism. The Elongated-Bubble pattern is considered the limiting case of Slug flow, when the liquid slug is free of entrained bubbles. This occurs at relatively lower gas rates when the flow is calmer. At higher gas flow rates, where the flow at the front of the slug is in the form of an eddy, the flow is designated as Slug flow.

**Annular Flow (A):** Annular flow occurs at very high gas flow rates. The gas-phase flows in a core of high velocity, which may contain entrained liquid droplets. The liquid flows as a thin film around the pipe wall. The interface is highly wavy, resulting in a high interfacial shear stress. The film at the bottom is usually thicker than that at the top, depending upon the relative magnitudes of the gas and liquid flow rates. At the lowest gas flow rates, most of the liquid flows at the bottom of the pipe, while aerated unstable waves are swept around the pipe periphery and wet the upper pipe wall occasionally. This flow occurs on the transition boundary between Stratified-Wavy, Slug and Annular flow. It is not Stratified-Wavy because liquid is swept around and wets the upper pipe wall with a thin film. It is also not Slug flow because no liquid bridging of the pipe cross section is formed. As a result, the frothy waves are not accelerated to the gas velocity but move slower than the gas phase. Also, it is not fully developed Annular flow, which requires a stable film around the pipe periphery. This flow pattern is designated sometimes as a “Proto Slug” flow. Based on the definitions and mechanisms of Slug and Annular flows, this regime is termed as Wavy-Annular (WA) and classified as a subgroup of Annular flow. The difference between Slug flow and Wavy-Annular flow is more distinguishable in upward inclined flow. During Slug flow, back flow of the liquid film between slugs is observed, whereas in Wavy-Annular flow, the liquid moves forward uphill with
frothy waves superimposed on the film. These waves move much slower than the gas-phase.

**Dispersed-Bubble Flow (DB):** At very high liquid flow rates, the liquid-phase is the continuous phase, in which the gas-phase is dispersed as discrete bubbles. The transition to this flow pattern is defined either by the condition where bubbles are first suspended in the liquid or when gas pockets, which touch the top of the pipe, are destroyed. When this happens, most of the bubbles are located near the upper pipe wall. At higher liquid rates, the gas bubbles are dispersed more uniformly in the entire cross-sectional area of the pipe. Under Dispersed-Bubble flow conditions, as a result of high liquid flow rates, the two phases are moving at the same velocity, and the flow is considered homogeneous no-slip.

**Vertical and Sharply Inclined Flow**

![Flow patterns existing in vertical and sharply inclined pipes](image)

Source: Mechanistic modeling of gas-liquid two-phase flow in pipes

*Figure 5: Flow patterns existing in vertical and sharply inclined pipes*

In this range of inclination angles, the Stratified regime disappears and a new flow pattern is observed, namely, Churn flow. Usually, the flow patterns are more symmetric around the pipe axis and less dominated by gravity. The existing flow patterns are Bubble flow, Slug flow, Churn flow, Annular flow, and Dispersed-Bubble flow.

**Bubble Flow (B):** In bubble flow the gas-phase is dispersed into small discrete bubbles, moving upwards in a zigzag motion, in a continuous liquid-phase. For
vertical flow, the bubble distribution is approximately homogeneous through the pipe cross section. Bubble flow occurs at relatively low liquid rates and is characterized by slippage between the gas and the liquid phases, resulting in large values of liquid holdup.

**Slug Flow (SL):** The Slug flow regime in vertical pipes is symmetric around the pipe axis. Most of the gas-phase is located in a large bullet shape gas pocket termed “Taylor bubble” with a diameter almost equal to the pipe diameter. The flow consists of successive Taylor bubbles and liquid slugs, which bridge the pipe cross section. A thin liquid film flows downward between the Taylor bubble and the pipe wall. The film penetrates into the following liquid slug and creates a mixing zone aerated by small gas bubbles.

**Churn (CH):** This flow pattern is characterized by an oscillatory motion of the liquid-phase. Churn flow is similar to slug flow but looks much more chaotic with no clear boundaries between the two phases. It occurs at higher gas flow rates, where the liquid slugs bridging the pipe become shorter and frothy. The slugs are blown through by the gas-phase, and then they break, fall backwards, and merge with the following slug. As a result, the bullet-shaped Taylor bubble is distorted and churning occurs.

**Annular Flow (A):** As in the horizontal case, the flow is characterized by a fast moving gas core with entrained liquid droplets and a slow moving liquid film flowing around the pipe wall. The flow is associated with a wavy interfacial structure, which results in a high interfacial shear stress. In vertical flow, the liquid film thickness around the pipe wall is approximately uniform.

**Dispersed-Bubble Flow (DB):** Similar to the horizontal flow case, Dispersed-Bubble flow in vertical and sharply inclined pipes occurs at relatively high liquid flow rates, under which conditions the gas-phase is dispersed as discrete bubbles into the continuous liquid-phase. For this flow pattern, the dominant liquid-phase carries the gas bubbles, and no slippage takes place between the phases.
Flow Pattern Map

The earlier approach for predicting flow patterns has been the empirical approach. Determination of the flow patterns was carried out mainly by visual observations.

Usually, the data is mapped on a two-dimensional plot, and the transition boundaries between the different flow patterns were determined. Such a map is called a Flow Pattern Map.

Two examples of this type of maps are shown in Figure 6 and Figure 7.

As a SECOND STEP the Program has the capability to predict the Flow Regime using Flowpatn 1.0.

If the predicted flow Pattern results in Slug Flow a slug catcher is necessary and the following step will be the design of it. If the predicted flow Pattern is not Slug Flow the program will skip the slug catcher design and will continue with the Separator design.
3.2.3. **Slug Catcher Design**

Slug flow in a transportation pipeline *Figure 8* can cause many problems in design and operation processes, which include kinetic force on fittings and vessels, pressure cycling, control instability, and inadequate phase separation.

Slugging greatly affects the design of receiving facilities.

In gas condensate systems, larger lines result in more liquid being retained in the pipeline at low rates.

When the flow rate is increased, much of the liquid can be swept out, potentially overwhelming the liquid handling capability of the receiving facilities.

The facilities can be flooded and damaged if the slugs are larger than the slug catcher capacity. Therefore, quantifying the slug size, frequency, and velocity is necessary prior to equipment design.

*Source*: Flow Assurance course slides

*Figure 7: Vertical Flow Pattern Map*
Slugging Types

Hydrodynamic slugs or Natural Slugging

Formed from the stratified flow regime due to instability of waves at certain flow rates.

Hydrodynamic slugs are initiated by the instability of waves on the gas/liquid interface in stratified flow under certain flowing conditions.

The gas/liquid interface is lifted to the top of the pipe when the velocity difference between gas phase and liquid phase is high enough.

Once the wave reaches the top of pipe, it forms a slug. The slug is pushed by the gas and so travels at a greater velocity than the liquid film, and more liquid is then swept into the slug.

Terrain-induced slugs or Severe Slugging

Caused by accumulation and periodic purging of liquid in elevation changes along the flow line, particularly at low flow rates.

It is initiated at dips of pipeline. Low spot fills with liquid and flow is blocked lately pressure builds up behind the blockage and when the pressure becomes high enough, gas blows liquid out of the low spot as a slug.

Operationally induced slugs

Formed in the system during operation transfer between a steady state and a transient state; for example, during start-up or pigging operations.
**Pipeline Shut-in and Start-up:** liquid accumulation in low points may produce big liquid slugs and a high pressure requirement at start-up. Partially shut-in or start-up can cause chain reactions in a network. Can be very long.

**Pipeline Packing/Depacking:** when pressure or inlet flow in pipeline increases, more gas/oil is stored in pipeline – pipeline packing. The opposite is pipeline depacking (or drafting).

Some problems may occur: low temperature can result due to high pressure drop (Joule-Thomson effect) combined with heat transfer with environment, change in liquid hold-up can occur and/or big liquid slugs may be created. Flow regime may change.

**Pigging:** pigging is a term used to describe a mechanical method for removing contaminants and deposits within the pipe or to clean accumulated liquids in the lower portions of hilly terrain pipelines using a mechanized plunger or pigs. Pigs *Figure 9* are run through pipelines for a variety of reasons, including: Liquid inventory control, Maintenance and data logging, pipeline cleaning and dewaxing or inhibitor application. For the Slug Catcher Design the program will take into account Natural Slugging and Pigging Derived Slug flows.

**Source:** Flow Assurance course slides

*Figure 9: Pigs*
**Slug Catchers**

A slug catcher is a piece of process equipment (typically a pressure vessel or set of pipes) located at the outlet of production flow lines or pipelines, prior to the remaining production facilities.

The most common function is to match two flows that are not identical in time but are expected to average out over the long run. Take a feed surge drum.

The purpose is to maintain sufficient inventory to feed the process and to maintain sufficient void capacity to continue receiving feed as it arrives.

Clearly the tank must be large enough to accommodate any normal discrepancies between input and output over a reasonable period of time. Between the upper and lower bound, the exact value of the level does not matter.

*Figure 10* shows the typical facility structure. In this case the outlet flow rate is controlled in a desired level. A flow controller at the outlet, properly tuned, will maintain a steady flow to the process. Usually located directly upstream of the primary production separator.

![Figure 10: Facility schematic](image)

Source: Flow Assurance course slides

**Types of Slug Catchers**

**Horizontal Pressure Vessel:** incoming production fluids strike some type of impingement device, which is designed to reduce the momentum of the fluid.
The liquid drops to the lower portion of the vessel; gas bubbles entrained in the liquid evolve out as the liquid moves towards the vessel liquid outlet.

A vortex breaker at the outlet prevents evolved gas from re-entraining into the liquid.

A mist eliminator is typically installed to aid in removing liquid droplets from the gas. As shown in Figure 11

Source: Flow Assurance course slides

**Figure 11: Horizontal Pressure Vessel Slug Catcher**

**Vertical Pressure Vessel:** Usually has less footprint as compare with horizontal separators. Footprint is the horizontal area required for the installation of particular equipment. It is critical on offshore application

**Finger Type** also known as a harp or multi-pipe, is a manifold of parallel pipes (also called “bottles”) installed downward incline.

In a particular Bottle: liquid slug enter in the pipe or bottle and drains downward toward liquid manifold, gas pocket (coming behind the liquid slug) flows up towards the gas manifold.

The use of symmetric headers is preferred to encourage equal flow splitting between the various slug catcher vessels or pipes.
Orientation of the tee so that the inlet flow is through the “branch” connection and outlet flow exits through the two sides of the tee increases the chance that even flow splitting of liquids and gas will occur.

Even with symmetric inlet piping it is common practice to provide equalizing lines for the gas and liquid to flow between the various slug catcher pipes or vessels.

A real Slug Catcher from Total Indonesia is shown in Figure 12.

Source: Panoramio

**Figure 12: Slug Catcher Finger Type**

**Principles:**

As a **THIRD STEP** the program will design a slug catcher if the Flow Pattern shows that it is necessary.

**Slug surge capacity:** sufficient liquid volume capacity to handle the design slug size.

**Liquid capacity:** sufficient liquid volume capacity for level control and to permit gas bubble separation from the bulk liquid stream.
Gas capacity: sufficient cross-sectional area to permit liquid droplet separation from the bulk gas stream.

For design propose the program will assume that the primary function of the slug catcher is Slug surge capacity or Storage.

**Slug Catcher Design Procedure:**

If the program’s previous step, Flow pattern prediction, ends with Slug Flow then:

For Natural Slugging:

\[
V_s = V_M
\]

\[
V_M = V_{SL} + V_{SG}
\]

\[
HLS = \frac{1}{1 + \left(\frac{V_M}{28.41}\right)^{1.39}}
\]

\[
LS_{mean} = \exp\left(-2.663 + 5.441 \cdot \sqrt{\ln(d)} + 0.059 \cdot \ln(V_M)\right)
\]

\[
LS_{max} = \exp\left(3.09 \cdot 0.5 + \ln(\text{LS}_{mean}) - \left(\frac{0.5 \cdot 2}{2}\right)\right)
\]

\[
V_{slug} = LS_{max} \cdot \left(\frac{\pi}{4}\right) \cdot \left(\frac{d}{12}\right)^2 \cdot HLS
\]

\[
t_{slug} = \frac{LS_{max}}{V_M}
\]

Where:

\( V_s \): Slug Velocity, ft/s

\( V_M \): Mixture Velocity, ft/s

\( V_{SG} \): Superficial gas velocity, ft/s

\( V_{SL} \): Superficial Liquid velocity, ft/s

\( HLS \): Slug Liquid Holdup, [-]

\( LS_{mean} \): Mean Slug length, ft
**LMax:** Max Slug length, ft

d: Pipe Diameter, in

**Vslug:** Slug Volume, ft³

**tslug:** Slug Duration, s

For Pigging:

\[ V_s = V_M \]

\[ V_M = V_{SL} + V_{SG} \]

\[ T_{transient} = \frac{L}{V_M} \]

\[ V_{slug} = (HL \cdot A \cdot L) \cdot (1 - 0.02) - (Vsl \cdot A \cdot T_{transient}) \]

\[ t_{slug} = \frac{v_{slug}}{A \cdot V_M} \]

Where:

**Vs:** Slug Velocity, ft/s

**V_M:** Mixture Velocity, ft/s

**V_{SG}:** Superficial gas velocity, ft/s

**V_{SL}:** Superficial Liquid velocity, ft/s

**T_{transient}:** Transient time of the Pig, s

**L:** Pipe length, ft

**HL:** liquid holdup, [-]

**A:** Pipe cross sectional area, ft²

**V_{slug}:** Slug Volume, ft³

**t_{slug}:** Slug Duration, s
Once Natural Slugging and Pigging Slug duration and Slug volume have been calculated Surge Volume will be calculated as follows and the bigger one between Natural Slugging and Pigging will be chosen for further calculations.

Slug surge volume is the volume required in the slug catcher to accommodate the rising liquid level due to the slug. Separator flowrate will be assumed to be 0 in order to account for the worst scenario.

\[
SurgeVolume = (slugflowrate - Separatorflowrate) \cdot tslug
\]

\[
slugflowrate = \frac{V_{slug}}{tslug}
\]

Where:

- **SurgeVolume**: Surge volume, ft³
- **Slug Flowrate**: Slug Flowrate, ft³/s
- **Separator Flowrate**: Separator Flowrate, ft³/s (assume = 0)
- **V_{slug}**: Slug Volume, ft³
- **tslug**: Slug Duration, s

The Slug type which provides the biggest surge volume has been chosen, using this value slug catcher dimensions will be calculated for both pressure vessel and finger type.

**Pressure Vessel**:

\[
V = SurgeVolume \cdot 1.5
\]

\[
d = \left( \frac{4V}{\pi \left( \frac{L}{d} \right)} \right)^{\frac{1}{3}} \times 12
\]

L/d or aspect ratio will be assumed to be 4

The final diameter need to be the smallest nominal diameter which is larger than the one calculated.
Nominal Vessel Diameter 16, 20, 24, 30, 36, 42, 48, 54, 60, 66, 72, 78, 84, 90 or 96

\[ L = \left( \frac{L}{d} \right) \times \frac{d_{Nominal}}{12} \]

Where:

\( V \): Slug Catcher Vessel Volume, ft\(^3\)

\( L/d \): aspect ratio, [-]

\( d_{Nominal} \): Nominal diameter, in

\( L \): Slug Catcher Vessel Length, ft

**Finger Type:**

\[ V = SurgeVolume \cdot 1.25 \]

\( d_{bottle} = d_{pipe} \)

\[ L = 500 \]

\[ V_{bottle} = \frac{\pi \cdot \left( \frac{d_{bottle}}{12} \right)^2 \cdot L}{4} \]

\[ Nbottles = \frac{V}{V_{bottle}} \]

Where:

\( V \): Slug Catcher Finger Type Volume, ft\(^3\)

\( d_{bottle} \): Bottle diameter, in

\( L \): bottles length, ft

\( V_{bottle} \): Bottle volume, ft\(^3\)

\( Nbottles \): Number of bottles, [-]

*VBA CODE FOR SLUG CATCHER DESIGN PROCEDURE AVAILABLE IN:
Annex B.4: Sheet4 (Slug Catcher) code*
3.2.4. Separators Design

Produced wellhead fluids are complex mixtures of different compounds of hydrogen and carbon, all with different densities, vapor pressures, and other physical characteristics. As a well stream flows from the hot, high-pressure petroleum reservoir, it experiences pressure and temperature reductions. Gases evolve from the liquids and the well stream changes in character. The velocity of the gas carries liquid droplets, and the liquid carries gas bubbles. The physical separation of these phases is one of the basic operations in the production, processing, and treatment of oil and gas. In oil and gas separator design, we mechanically separate from a hydrocarbon stream the liquid and gas components that exist at a specific temperature and pressure. Proper separator design is important because a separation vessel is normally the initial processing vessel in any facility, and improper design of this process component can "bottleneck" and reduce the capacity of the entire facility.

In oil and gas separator design, we mechanically separate from a hydrocarbon stream the liquid and gas components that exist at a specific temperature and pressure.

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 components.

Separators are sometimes called "gas scrubbers" when the ratio of gas rate to liquid rate is very high. They all have the same configuration and are sized in accordance with the same procedure.

Separators are designed in horizontal, vertical, or spherical configurations.

**Horizontal Separators**

*Figure 13* is a schematic of a horizontal separator. The fluid enters the separator and hits an inlet diverter causing a sudden change in momentum. The initial gross separation of liquid and vapor occurs at the inlet diverter. The force of gravity causes the liquid droplets to fall out of the gas stream to the bottom of the vessel where it is collected. This liquid collection section provides the retention time required to let
entrained gas evolve out of the oil and rise to the vapor space. It also provides a surge volume, if necessary, to handle intermittent slugs of liquid. The liquid then leaves the vessel through the liquid dump valve. The liquid dump valve is regulated by a level controller. The level controller senses changes in liquid level and controls the dump valve accordingly.

Source: Arnold K., and Stewart, M. Jr.: Surface Production Operations

Figure 13: Horizontal separator schematic

The gas flows over the inlet diverter and then horizontally through the gravity settling section above the liquid. As the gas flows through this section, small drops of liquid that were entrained in the gas and not separated by the inlet diverter are separated out by gravity and fall to the gas-liquid interface.

Some of the drops are of such a small diameter that they are not easily separated in the gravity settling section. Before the gas leaves the vessel it passes through a coalescing section or mist extractor. This section uses elements of vanes, wire mesh, or plates to coalesce and remove the very small droplets of liquid in one final separation before the gas leaves the vessel.

The pressure in the separator is maintained by a pressure controller. The pressure controller senses changes in the pressure in the separator and sends a signal to either open or close the pressure control valve accordingly. By controlling the rate at which gas leaves the vapor space of the vessel the pressure in the vessel is maintained. Normally, horizontal separators are operated half full of liquid to maximize the surface area of the gas liquid interface.
Vertical Separators

*Figure 14* is a schematic of a vertical separator. In this configuration the inlet flow enters the vessel through the side. As in the horizontal separator, the inlet diverter does the initial gross separation. The liquid flows down to the liquid collection section of the vessel. Liquid continues to flow downward through this section to the liquid outlet. As the liquid reaches equilibrium, gas bubbles flow counter to the direction of the liquid flow and eventually migrate to the vapor space. The level controller and liquid dump valve operate the same as in a horizontal separator.

The gas flows over the inlet diverter and then vertically upward toward the gas outlet. In the gravity settling section the liquid drops fall vertically downward counter to the gas flow. Gas goes through the mist extractor section before it leaves the vessel. Pressure and level are maintained as in a horizontal separator.

![Vertical Separator Schematic](image)

**Source:** Arnold K., and Stewart, M. Jr.: Surface Production Operations

**Figure 14:** Vertical separator schematic

Spherical Separators

A typical spherical separator is shown in *Figure 15*. The same four sections can be found in this vessel. Spherical separators are a special case of a vertical separator where there is no cylindrical shell between the two heads. They may be very efficient from a pressure containment standpoint but because they have limited liquid surge
capability and they exhibit fabrication difficulties, they are not usually used in oil field facilities.

**Figure 15: Spherical separator schematic**

**Scrubbers**

A scrubber is a two-phase separator that is designed to recover liquids carried over from the gas outlets of production separators or to catch liquids condensed due to cooling or pressure drops. Liquid loading in a scrubber is much lower than that in a separator. Typical applications include: upstream of mechanical equipment such as compressors that could be damaged, destroyed or rendered ineffective by free liquid; downstream of equipment that can cause liquids to condense from a gas stream; upstream of gas dehydration equipment that would lose efficiency, be damaged, or be destroyed if contaminated with liquid hydrocarbons; and upstream of a vent or flare outlet.

Vertical scrubbers are most commonly used. Horizontal scrubbers can be used, but space limitations usually dictate the use of a vertical configuration.

**Horizontal vs. vertical vessel selection**

Horizontal separators are smaller and less expensive than vertical separators for a given gas capacity. In the gravity settling section of a horizontal vessel, the liquid droplets fall perpendicular to the gas flow and thus are more easily settled out of the
gas continuous phase. Also, since the interface area is larger in a horizontal separator than a vertical separator, it is easier for the gas bubbles, which come out of solution as the liquid approaches equilibrium, to reach the vapor space. Horizontal separators offer greater liquid capacity and are best suited for liquid-liquid separation and foaming crudes.

Thus, from a pure gas/liquid separation process, horizontal separators would be preferred. However, they do have the following drawbacks, which could lead to a preference for a vertical separator in certain situations:

Horizontal separators are not as good as vertical separators in handling solids. The liquid dump of a vertical separator can be placed at the center of the bottom head so that solids will not build up in the separator but continue to the next vessel in the process. As an alternative, a drain could be placed at this location so that solids could be disposed of periodically while liquid leaves the vessel at a slightly higher elevation.

In a horizontal vessel, it is necessary to place several drains along the length of the vessel. Since the solids will have an angle of repose of 45° to 60°, the drains must be spaced at very close intervals. Horizontal vessels require more plan area to perform the same separation as vertical vessels. While this may not be of importance at a land location, it could be very important offshore.

Smaller, horizontal vessels can have less liquid surge capacity than vertical vessels sized for the same steady-state flow rate. For a given change in liquid surface elevation, there is typically a larger increase in liquid volume for a horizontal separator than for a vertical separator sized for the same flow rate. However, the geometry of a horizontal vessel causes any high level shutdown device to be located close to the normal operating level. In a vertical vessel the shutdown could be placed much higher, allowing the level controller and dump valve more time to react to the surge. In addition, surges in horizontal vessels could create internal waves that could activate a high level sensor.

It should be pointed out that vertical vessels also have some drawbacks that are not process related and must be considered in making a selection. These are:
The relief valve and some of the controls may be difficult to service without special ladders and platforms.

The vessel may have to be removed from a skid for trucking due to height restrictions.

Overall, horizontal vessels are the most economical for normal oil-gas separation, particularly where there may be problems with emulsions, foam, or high gas-oil ratios. Vertical vessels work most effectively in low Rs applications. They are also used in some very high RS applications, such as scrubbers where only fluid mists are being removed from the gas.

**VESSEL INTERNALS**

**Inlet Diverters**

There are many types of inlet diverters. Two main types are baffle plates and centrifugal diverters shown in *Figure 16* and *Figure 17*, respectively. A baffle plate can be a spherical dish, flat plate, angle iron, cone, or just about anything that will accomplish a rapid change in direction and velocity of the fluids and thus disengage the gas and liquid. The design of the baffles is governed principally by the structural supports required to resist the impact-momentum load. The advantage of using devices such as a half sphere or cone is that they create less disturbance than plates or angle iron, cutting down on re-entrainment or emulsifying problems.

![Image of baffle plates](image1.png)

**Source:** Arnold K., and Stewart, M. Jr.: Surface Production Operations

**Figure 16:** Baffle plates
Centrifugal inlet diverters use centrifugal force, rather than mechanical agitation, to disengage the oil and gas. These devices can have a cyclonic chimney or may use a tangential fluid race around the walls. Centrifugal inlet diverters are proprietary but generally use an inlet nozzle sufficient to create a fluid velocity of about 20 ft/s. Centrifugal diverters work well in initial gas separation and help to prevent foaming in crudes.

Source: Arnold K., and Stewart, M. Jr.: Surface Production Operations

Figure 17: Centrifugal inlet diverter
Wave Breakers

In long horizontal vessels it is necessary to install wave breakers, which are nothing more than vertical baffles spanning the gas-liquid interface and perpendicular to the flow.

Defoaming Plates

Foam at the interface may occur when gas bubbles are liberated from the liquid. This foam can be stabilized with the addition of chemicals at the inlet. Many times a more effective solution is to force the foam to pass through a series of inclined parallel plates or tubes to aid in coalescence of the foam bubbles.

Vortex Breaker

It is normally a good idea to include a simple vortex breaker as shown in Figure 18 to keep a vortex from developing when the liquid control valve is open. A vortex could suck some gas out of the vapor space and re-entrain it in the liquid outlet.

Source: Arnold K., and Stewart, M. Jr.: Surface Production Operations

Figure 18: Typical vortex breakers
Mist Extractor

Mist extractors can be made of wire mesh, vanes, centrifugal force devices, or packing. Wire mesh pads, Figure 19, are made of finely woven mats of stainless steel wire wrapped into a tightly packed cylinder.

The liquid droplets impinge on the matted wires and coalesce. The effectiveness of wire mesh depends largely on the gas being in the proper velocity range. If the velocities are too high, the liquids knocked out will be re-entrained. If the velocities are low, the vapor just drifts through the mesh element without the droplets impinging and coalescing.

The construction is often specified by calling for a certain thickness and mesh density. Experience has indicated that a properly sized wire mesh eliminator can remove 99% of 10-micron and larger droplets. Although wire mesh eliminators are inexpensive they are more easily plugged than the other types.

Vane eliminators Figure 20 force the gas flow to be laminar between parallel plates that contain directional changes. Figure 21 shows a vane mist extractor made from angle iron. In vane eliminators, droplets impinge on the plate surface where they

![Figure 19: Wire mesh mist extractor](image)

![Figure 20: Vane mist extractor](image)

Source: Arnold K., and Stewart, M. Jr.: Surface Production Operations
coalesce and fall to a liquid collecting spot.

They are routed to the liquid collection section of the vessel. Vane-type eliminators are sized by their manufacturers to assure both laminar flow and a certain minimum pressure drop.

Source: Arnold K., and Stewart, M. Jr.: Surface Production Operations

Figure 21: Angle iron vane mist extractor

Some separators have centrifugal mist eliminators Figure 22 that cause the liquid drops to be separated by centrifugal force. These can be more efficient than either wire mesh or vanes and are the least susceptible to plugging. However, they are not in common use in production operations because their removal efficiencies are sensitive to small changes in flow. In addition, they require relatively large pressure drops to create the centrifugal force. To a lesser extent, random packing is sometimes used for mist extraction, as shown in Figure 23. The packing acts as a coalescer.

Source: Arnold K., and Stewart, M. Jr.: Surface Production Operations
POTENTIAL OPERATING PROBLEMS

Foamy Crudes

The major cause of foam in crude oil is the appearance of impurities, other than water, which are impractical to remove before the stream reaches the separator. Foam presents no problem within a separator if the internal design assures adequate time or sufficient coalescing surface for the foam to "break."

Foaming in a separating vessel is a threefold problem:

Mechanical control of liquid level is aggravated because any control device must deal with essentially three liquid phases instead of two.

Foam has a large volume-to-weight ratio. Therefore, it can occupy much of the vessel space that would otherwise be available in the liquid collecting or gravity settling sections.

In an uncontrolled foam bank, it becomes impossible to remove separated gas or degassed oil from the vessel without entraining some of the foamy material in either the liquid or gas outlets.

Comparison of foaming tendencies of known oil to a new one, about which no operational information is known, provides an understanding of the relative foam problem that may be expected with the new oil as weighed against the known oil. A related amount of adjustment can then be made in the design parameters, as compared to those found satisfactory for the known case.

It should be noted that the amount of foam is dependent on the pressure drop to which the inlet liquid is subjected, as well as the characteristics of the liquid at separator conditions. In some cases, the effect of temperature may be significant.

Foam depressants often will do a good job in increasing the capacity of a given separator. However, in sizing a separator to handle particular crude, the use of an effective depressant should not be assumed because characteristics of the crude and of the foam may change during the life of the field. Also, the cost of foam depressants for high rate production may be prohibitive. Sufficient capacity should be provided in the separator to handle the anticipated production without use of a
foam depressant or inhibitor. Once placed in operation, a foam depressant may allow more throughput than the design capacity.

**Paraffin**

Separator operation can be adversely affected by an accumulation of paraffin. Coalescing plates in the liquid section and mesh pad mist extractors in the gas section are particularly prone to plugging by accumulations of paraffin. Where it is determined that paraffin is an actual or potential problem, the use of plate-type or centrifugal mist extractors should be considered. Manways, handholes, and nozzles should be provided to allow steam, solvent, or other types of cleaning of the separator internals. The bulk temperature of the liquid should always be kept above the cloud point of the crude oil.

**Sand**

Sand can be very troublesome in separators by causing cutout of valve trim, plugging of separator internals, and accumulation in the bottom of the separator. Special hard trim can minimize the effects of sand on the valves. Accumulations of sand can be alleviated by the use of sand jets and drains.

Plugging of the separator internals is a problem that must be considered in the design of the separator. A design that will promote good separation and have a minimum of traps for sand accumulation may be difficult to attain, since the design that provides the best mechanism for separating the gas, oil, and water phases probably will also provide areas for sand accumulation. A practical balance for these factors is the best solution.

**Liquid Carryover and Gas Blowby**

Liquid carryover and gas blowby are two common operating problems. Liquid carryover occurs when free liquid escapes with the gas phase and can indicate high liquid level, damage to vessel internals, foam, improper design, plugged liquid outlets, or a flow rate that exceeds the vessel's design rate.

Gas blowby occurs when free gas escapes with the liquid phase and can be an indication of low liquid level, vortexing, or level control failure.
SEPARATORS DESIGN PROCEDURE

Settling

In the gravity settling section the liquid drops will settle at a velocity determined by equating the gravity force on the drop with the drag force caused by its motion relative to the gas continuous phase.

The drag force is one of the most investigated forces, aiming at the prediction of the drag coefficient, CD, as function of the particle Reynolds number. A compilation of four correlations has been selected for the program. The correlations of Sciller Naumann, Ishii and Zuber, Ihme et al. and the one from the paper SPE 56645:

Sciller Naumann

\[ Cd = \frac{24}{Re} \cdot (1 + 0.15 \cdot Re^{0.687}) \]

Ishii and Zuber

\[ Cd = \frac{24}{Re} \cdot (1 + 0.1 \cdot Re^{0.75}) \]

Ihme

\[ Cd = \frac{24}{Re} + 5.48 \cdot Re^{-0.573} + 0.36 \]

SPE 56645

\[ Cd = \frac{24}{Re} + \frac{3}{\sqrt{Re}} + 0.34 \]

Where:

Cd: drag coefficient, [-]

Re: Reynolds number, [-]

*VBA CODE FOR SLUG CATCHER DESIGN PROCEDURE AVAILABLE IN:
Annex B.5: Cd module

Equating drag and buoyant forces, the terminal settling velocity is given by:
\[ Vt = 0.0119 \cdot \left( \frac{\text{Deno} - \text{DenG}}{\text{DenG}} \right) \cdot \left( \frac{\text{dOG}}{\text{Cd}} \right)^{0.5} \]

Where:

\textit{Deno}: Oil density, lb/ft³

\textit{DenG}: Density of the gas, lb/ft³

\textit{dOG}: Diameter of Oil Droplet in Gas Phase, microns

\textit{Cd}: drag coefficient, [-]

*AN ITERATIVE PROCESS SHOULD BE CARRIED ON IN ORDER TO CALCULATE \(Vt\) AND \(Cd\), REFER TO: Annex B.6: \(Cd\) calculation

**Drop Size**

The purpose of the gas separation section of the vessel is to condition the gas for final polishing by the mist extractor. From field experience it appears that if 100 micron drops are removed in this section, the mist extractor will not become flooded and will be able to perform its job of removing those drops between 10 and 100 micron diameter.

The gas capacity design equations in this section are all based on 100 micron removal. In some cases, this will give an overly conservative solution.

Flare or vent scrubbers are designed to keep large slugs of liquid from entering the atmosphere through the vent or relief systems. In vent systems the gas is discharged directly to the atmosphere and it is common to design the scrubbers for removal of 300 to 500 micron droplets in the gravity settling section. A mist extractor is not included because of the possibility that it might plug creating a safety hazard. In flare systems, where the gas is discharged through a flame, there is the possibility that burning liquid droplets could fall to the ground before being consumed.

The recommended Parameters are:

Diameter of Oil Droplet in Gas Phase \(dOG = 100\) [microns]

Diameter of Water Droplet in Oil Phase \(dOW = 500\) [microns]
Retention Time

To assure that the liquid and gas reach equilibrium at separator pressure a certain liquid storage is required. This is defined as "retention time" or the average time a molecule of liquid is retained in the vessel assuming plug flow. The retention time is thus the volume of the liquid storage in the vessel divided by the liquid flow rate.

The Recommended retention times are shown in Table 4 and 5:

### Table 4: Residence Time, Two-Phase Separators

<table>
<thead>
<tr>
<th>API Gravity</th>
<th>Minutes</th>
</tr>
</thead>
<tbody>
<tr>
<td>&gt; 35 API</td>
<td>1</td>
</tr>
<tr>
<td>20 – 30 API</td>
<td>1 - 2</td>
</tr>
<tr>
<td>10 – 20 API</td>
<td>1 - 4</td>
</tr>
</tbody>
</table>

### Table 5: Residence Time, Three-Phase Separators

<table>
<thead>
<tr>
<th>API Gravity</th>
<th>Minutes</th>
</tr>
</thead>
<tbody>
<tr>
<td>&gt; 35 API</td>
<td>1</td>
</tr>
<tr>
<td>20 – 30 API</td>
<td>1 - 2</td>
</tr>
<tr>
<td>10 – 20 API</td>
<td>1 - 4</td>
</tr>
</tbody>
</table>

3.2.4.1. Two-phase horizontal separators

For sizing a horizontal separator it is necessary to choose a seam to seam vessel length and a diameter. This choice must satisfy the conditions for gas capacity that allow the liquid drops to fall from the gas to the liquid volume as the gas traverses the effective length of the vessel. It must also provide sufficient retention time to allow the liquid to reach equilibrium.

**Gas Capacity Constrain**

\[
k = \frac{(Den_G \cdot Cd)}{((Den_o + Den_w) - Den_G) \cdot dOG)}
\]
\[ d_{\text{Leff}} = \frac{(420 \cdot (T + 460) \cdot Z \cdot \left( \frac{q_g}{1000} \right) \cdot k)}{p} \]

Where:

- \( d \): Vessel Internal diameter, in
- \( \text{Leff} \): effective length, ft
- \( D_{\text{eno}} \): Oil density, lb/ft\(^3\)
- \( D_{\text{enG}} \): Density of the gas, lb/ft\(^3\)
- \( D_{\text{enw}} \): Density of the water, lb/ft\(^3\)
- \( d_{\text{OG}} \): Diameter of Oil Droplet in Gas Phase, microns
- \( C_d \): drag coefficient, [-]
- \( Z \): Compressibility factor, [-]
- \( T \): in-situ Temperature, °R
- \( q_g \): Gas Flowrate, MSCFD
- \( p \): in-situ Pressure, Psia

**Liquid Capacity Constrain**

\[ d_{2\text{Leff}} = 1.42 \cdot (q_w \cdot trw + q_o \cdot tro) \]

Where:

- \( d \): Vessel Internal diameter, in
- \( \text{Leff} \): effective length, ft
- \( q_o \): Oil Flowrate, BPD
- \( q_w \): Water Flowrate, BPD
- \( trw \): Water retention time, min
- \( tro \): Oil retention time, min
The biggest Leff will determine if liquid or gas constrain governs. The seam to seam length will be calculated using this value.

Seam-to-Seam Length and Slenderness Ratio

The seam-to-seam length of the vessel should be determined from the geometry once an effective length has been determined. Allowance must be made for the inlet diverter and mist extractor. For screening purposes the following approximation has been proven useful:

For gas capacity

\[ L_{ss} = \frac{Leff}{0.75} \]

For liquid capacity

\[ L_{ss} = Leff + \frac{d}{12} \]

Where:

- \( L_{ss} \): Seam to seam length, ft
- \( Leff \): Effective length, ft
- \( d \): Vessel Internal diameter, in

Slenderness Ratio

\[ SlenRatio = \frac{12 \cdot L_{ss}}{d} \]

A table of possible results will be computed, the final options will depend on the maximum and minimum slenderness ratio chosen by the user.

Recommended Maximum and minimum Slenderness Ratios for Horizontal separators:

- Minimum Slenderness Ratio SIR min = 3
- Maximum Slenderness Ratio SIR Max = 5
3.2.4.2. Two-phase vertical separators

In vertical separators, a minimum diameter must be maintained to allow liquid drops to separate from the vertically moving gas. The liquid retention time requirement specifies a combination of diameter and liquid volume height. Any diameter greater than the minimum required for gas capacity can be chosen.

Gas Capacity Constrains

\[ k = \frac{(DenG \cdot Cd)}{((Deno + Denw) - DenG) \cdot dOG)} \]

\[ d^2 = \frac{(5040 \cdot (T + 460) \cdot Z \cdot \left( \frac{qg}{1000} \right) \cdot k)}{P} \]

Where:

- \( d \): Vessel Internal diameter, in
- \( Deno \): Oil density, lb/ft\(^3\)
- \( DenG \): Density of the gas, lb/ft\(^3\)
- \( Denw \): Density of the water, lb/ft\(^3\)
- \( dOG \): Diameter of Oil Droplet in Gas Phase, microns
- \( Cd \): Drag coefficient, [-]
- \( Z \): Compressibility factor, [-]
- \( T \): in-situ Temperature, ºR
- \( qg \): Gas Flowrate, MSCFD
- \( P \): in-situ Pressure, Psia

Liquid Capacity Constrains
Where:

\[ d^2 h = \frac{(q_w \cdot trw + q_o \cdot tro)}{0.12} \]

Where:

\[ d: \] Vessel Internal diameter, in
\[ h: \] High of the liquid volume, in
\[ q_o: \] Oil Flowrate, BPD
\[ q_w: \] Water Flowrate, BPD
\[ trw: \] Water retention time, min
\[ tro: \] Oil retention time, min

Seam-to-Seam Length and Slenderness Ratio

The seam-to-seam length of the vessel should be determined from the geometry once a diameter and height of liquid volume are known.

For screening purposes the following approximation has been proven useful. Use the larger of the two values:

\[ L_{ss} = \frac{h + 76}{12} \]
\[ L_{ss} = \frac{h + d + 40}{12} \]

Where:

\[ L_{ss}: \] Seam to seam length, ft
\[ h: \] High of the liquid volume, in
\[ d: \] Vessel Internal diameter, in

Slenderness Ratio

As with horizontal separators, the larger the slenderness ratio, the less expensive the vessel.
A table of possible results will be computed, the final options will depend on the maximum and minimum slenderness ratio chosen by the user.

Recommended Maximum and minimum Slenderness Ratios for Vertical separators:

Minimum Slenderness Ratio $SIR_{\text{min}} = 1.5$

Maximum Slenderness Ratio $SIR_{\text{Max}} = 3$

*VBA CODE FOR TWO-PHASE VERTICAL SEPARATOR DESIGN AVAILABLE IN: Annex B.8: TwoPhaseVertical module

OIL AND WATER SEPARATION

The separator design concepts that have been presented related to the two-phase separation of liquid and gas and are applicable to the separation of gas that takes place in three-phase separators, gas scrubbers, and any other device in which gas is separated from a liquid phase.

"Three-phase separator" and "free-water knockout" are terms that are used to describe pressure vessels that are designed to separate and remove the free water from a mixture of crude oil and water. Because flow normally enters these vessels either directly from a producing well or a separator operating at a higher pressure, the vessel must be designed to separate the gas that flashes from the liquid as well as separate the oil and water.

The term "Three-phase separator" is normally used when there is a large amount of gas to be separated from the liquid, and the dimensions of the vessel are determined by the gas capacity equations. "Free-water knockout" is generally used when the amount of gas is small relative to the amount of oil and water, and the dimensions of the vessel are determined by the oil/water separation.

The basic design aspects of Three-phase separation are identical to those discussed for Two-phase separators. The only additions are that more concern is placed on liquid-liquid settling rates; and that some means of removing the free water must be added. Water removal is a function of the control methods used to maintain
separation and removal from the oil. Several control methods are applicable to
Three-phase separators. The shape and diameter of the vessel will, to a degree,
determine the types of control used.

**Horizontal Separators**

Three-phase separators are designed as either horizontal or vertical pressure vessels. *Figure 24* is a schematic of a horizontal separator. The fluid enters the separator and hits an inlet diverter. This sudden change in momentum does the initial gross separation of liquid and vapor. In most designs, the inlet diverter contains a downcomer that directs the liquid flow below the oil/water interface.

![Horizontal three-phase separator schematic](image)

**Source:** Arnold K., and Stewart, M. Jr.: Surface Production Operations

*Figure 24: Horizontal three-phase separator schematic*

This forces the inlet mixture of oil and water to mix with the water continuous phase in the bottom of the vessel and rise through the oil/water interface. This process is called "water-washing," and it promotes the coalescence of water droplets which are entrained in the oil continuous phase. The inlet diverter assures that little gas is carried with the liquid, and the water wash assures that the liquid does not fall on top of the gas/oil or oil/water interface, mixing the liquid retained in the vessel and making control of the oil/water interface difficult.

The liquid collecting section of the vessel provides sufficient time so that the oil and emulsion form a layer or "oil pad" at the top. The free water settles to the bottom. *Figure 24* illustrates a typical horizontal separator with an interface controller and
The weir maintains the oil level and the level controller maintains the water level. The oil is skimmed over the weir. The level of the oil downstream of the weir is controlled by a level controller that operates the oil dump valve.

The produced water flows from a nozzle in the vessel located upstream of the oil weir. An interface level controller senses the height of the oil/water interface. The controller sends a signal to the water dump valve thus allowing the correct amount of water to leave the vessel so that the oil/water interface is maintained at the design height.

The gas flows horizontally and out through a mist extractor to a pressure control valve that maintains constant vessel pressure. 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. The most common configuration is half full, and this is used for the design equations in this section.

*Figure 25* shows an alternate configuration known as a "bucket and weir" design. This design eliminates the need for a liquid interface controller. Both the oil and water flow over weirs where level control is accomplished by a simple displacer float. The oil overflows the oil weir into an oil bucket where its level is controlled by a level controller that operates the oil dump valve. The water flows under the oil bucket and then over a water weir. The level downstream of this weir is controlled by a level controller that operates the water dump valve.

*Source:* Arnold K., and Stewart, M. Jr.: Surface Production Operations

*Figure 25: Three-Phase separator bucket and weir design.*
The height of the oil weir controls the liquid level in the vessel. The difference in height of the oil and water weirs controls the thickness of the oil pad due to specific gravity differences. It is critical to the operation of the vessel that the water weir height is sufficiently below the oil weir height so that the oil pad thickness provides sufficient oil retention time. If the water weir is too low and the difference in specific gravity is not as great as anticipated, then the oil pad could grow in thickness to a point where oil will be swept under the oil box and out the water outlet. Normally, either the oil or the water weir is made adjustable so that changes in oil/water specific gravities or flow rates can be accommodated.

**Vertical Separators**

*Figure 26* shows a typical configuration for a vertical three-phase separator. Flow enters the vessel through the side as in the horizontal separator; the inlet diverter separates the bulk of the gas. A downcomer is required to transmit the liquid through

![Diagram of a vertical three-phase separator](image)

**Source:** Arnold K., and Stewart, M. Jr.: Surface Production Operations

*Figure 26: Vertical three-phase separator schematic*

the oil/gas interface so as not to disturb the oil skimming action taking place.

A chimney is needed to equalize gas pressure between the lower section and the gas section.
The spreader or downcomer outlet is located at the oil-water interface. From this point as the oil raises any free water trapped within the oil phase separates out. The water droplets flow countercurrent to the oil. Similarly, the water flows downward and oil droplets trapped in the water phase tend to raise countercurrent to the water flow.

Sometimes a cone bottom three-phase separator is used. This is a design that would be used if sand production was anticipated to be a major problem. The cone is normally at an angle to the horizontal of between 45° and 60°. Produced sand may have a tendency to stick to steel at 45°. If a cone is installed it could be part of the pressure containing walls of the vessel, or for structural reasons, it could be installed internal to the vessel cylinder. In such a case, a gas equalizing line must be installed to assure that the vapor behind the cone is always in pressure equilibrium with the vapor space.

*Figure 27* shows the three different methods of control that are often used on vertical separators.

![Diagram of different liquid level control schemes](source)

**Source:** Arnold K., and Stewart, M. Jr.: Surface Production Operations

**Figure 27: Liquid level control schemes**

The first is strictly level control. A regular displacer float is used to control the gas-oil interface and regulate a control valve dumping oil from the oil section. An interface float is used to control the oil-water interface and regulate a water outlet.
control valve. Because no internal baffling or weirs are used, this system is the easiest to fabricate and handles sand and solids production best.

The second method shown uses a weir to control the gas/oil interface level at a constant position. This results in a better separation of water from the oil as all the oil must rise to the height of the oil weir before exiting the vessel. Its disadvantages are that the oil box takes up vessel volume and costs money to fabricate. In addition, sediment and solids could collect in the oil box and be difficult to drain, and a separate low level shut-down may be required to guard against the oil dump valve failing to open.

The third method uses two weirs, which eliminates the need for an interface float. Interface level is controlled by the height of the external water weir relative to the oil weir or outlet height. This is similar to the bucket and weir design of horizontal separators. The advantage of this system is that it eliminates the interface level control. The disadvantage is that it requires additional external piping and space.

Horizontal vs. Vertical Selection

The benefits of each type of design were described earlier. As in two-phase separation, it is also true for three-phase separation that the flow geometry in a horizontal vessel is more favorable from a process standpoint.

However, there may be non-process reasons to select a vertical vessel for a specific application.

**VESSEL INTERNALS**

Most of the vessel internals have been discussed. Two common internals not discussed are coalescing plates and sand jets. It is possible to use various plate or pipe coalescer designs to aid in the coalescing of oil droplets in the water and water droplets in the oil.

**Coalescing Plates**

The installation of coalescing plates in the liquid section will cause the size of the water droplets entrained in the oil phase to increase, making gravity settling of these drops to the oil-water interface easier.
Thus, the use of coalescing plates or the use of free-flow turbulent coalesce will often lead to the ability to handle a given flow rate in a smaller vessel. However, because of the potential for plugging with sand, paraffin, or corrosion products, the use of coalescing plates should be discouraged, except for instances where the savings in vessel size and weight are large enough to justify the potential increase in operating costs and decrease in availability.

**Sand Jets and Drains**

In horizontal three-phase separators, one worry is the accumulation of sand and solids at the bottom of the vessel. If allowed to build up, these solids upset the separator operations by taking up vessel volume. Generally, the solids settle to the bottom and become well packed.

To remove the solids, sand drains are opened in a controlled manner, and then high-pressure fluid, usually produced water, is pumped through the jets to agitate the solids and flush them down the drains. The sand jets are normally designed with a 20 ft/s jet tip velocity and aimed in such a manner to give good coverage of the vessel bottom.

To prevent the settled sand from clogging the sand drains, sand pans or sand troughs are used to cover the outlets. These are inverted troughs with slotted side openings.

**EMULSIONS**

Emulsions can be particularly troublesome in the operation of three-phase separators. Over a period of time an accumulation of emulsified materials and/or other impurities usually will form at the interface of the water and oil phases. In addition to adverse effects on the liquid level control, this accumulation will also decrease the effective oil or water retention time in the separator, with a resultant decrease in water-oil separation efficiency. Addition of chemicals and/or heat often minimizes this difficulty.

Frequently, it is possible to appreciably lower the settling time necessary for water/oil separation by either the application of heat in the liquid section of the separator or the addition of de-emulsifying chemicals.
3.2.4.3. Three-phase horizontal separators

For sizing a horizontal three-phase separator it is necessary to specify a vessel diameter and a seam-to-seam vessel length. The gas capacity and retention time considerations establish certain acceptable combinations of diameter and length.

Specific gravity difference

\[ \text{Delta}SG = \frac{\text{Denw} - \text{Deno}}{62.4} \]

Where:

\( \text{Delta}SG \): Specific gravity difference, [-]

\( \text{Deno} \): Oil density, lb/ft\(^3\)

\( \text{Denw} \): Density of the water, lb/ft\(^3\)

Gas Capacity Constrain

\[ k = \frac{(\text{DenG} \cdot \text{Cd})}{((\text{Deno} + \text{Denw}) - \text{DenG}) \cdot \text{dOG})} \]

\[ d^2 \text{Leff} = \frac{(420 \cdot (T + 460) \cdot Z \cdot \left( \frac{qg}{1000} \right) \cdot k)}{p} \]

Where:

\( d \): Vessel Internal diameter, in

\( \text{Leff} \): effective length, ft

\( \text{Deno} \): Oil density, lb/ft\(^3\)

\( \text{DenG} \): Density of the gas, lb/ft\(^3\)

\( \text{Denw} \): Density of the water, lb/ft\(^3\)

\( \text{dOG} \): Diameter of Oil Droplet in Gas Phase, microns

\( \text{Cd} \): drag coefficient, [-]

\( Z \): Compressibility factor, [-]
T: in-situ Temperature, °R

$q_g$: Gas Flowrate, MSCFD

$P$: in-situ Pressure, Psia

**Maximum oil pad thickness**

$$h_{max} = \frac{0.00128 \cdot tro \cdot DeltaSG \cdot dWO^2}{Viso}$$

Where:

$h_{max}$: Maximum oil pad thickness, [-]

$DeltaSG$: Specific gravity difference, [-]

$dWO$: Diameter of Water Droplet in Oil Phase, microns

$Viso$: Oil viscosity, cp

$tro$: Oil retention time, min

**Maximum diameter for oil pad thickness constrain**

$$\frac{Aw}{A} = \frac{0.5 \cdot qw \cdot trw}{tro \cdot qo + trw \cdot qw}$$

Where:

$Aw$: water cross sectional area, ft2

$A$: Cross sectional area, ft2

$q_o$: Oil Flowrate, BPD

$q_w$: Water Flowrate, BPD

$trw$: Water retention time, min

$tro$: Oil retention time, min

**Coefficient Beta $\beta$ for a cylinder half filled with water**
hoDGivenAwA function has been created to calculate the Beta coefficient.

(Refer to ThreePhaseHorizontal module)

Maximum diameter

\[ d_{\text{max}} = \frac{h_{\text{max}}}{\beta} \]

Where:

- \( h_{\text{max}} \): Maximum oil pad thickness, [-]
- \( \beta \): Beta coefficient

Liquid Capacity Constrain, retention time constrain

\[ d^2 \text{Leff} = 1.42 \cdot (q_w \cdot \text{trw} + q_o \cdot \text{tro}) \]

Where:

- \( d \): Vessel Internal diameter, in
- \( \text{Leff} \): effective length, ft
- \( q_o \): Oil Flowrate, BPD
- \( q_w \): Water Flowrate, BPD
- \( \text{trw} \): Water retention time, min
- \( \text{tro} \): Oil retention time, min

The biggest \( \text{Leff} \) will determine if liquid or gas constrain governs. The seam to seam length will be calculated using this value.

Seam-to-Seam Length and Slenderness Ratio

The seam-to-seam length of the vessel should be determined from the geometry once an effective length has been determined. Allowance must be made for the inlet diverter and mist extractor. For screening purposes the following approximation has been proven useful:

For gas capacity
\[ L_{ss} = \frac{L_{eff}}{0.75} \]

For liquid capacity

\[ L_{ss} = L_{eff} + \frac{d}{12} \]

Where:

- \( L_{ss} \): Seam to seam length, ft
- \( L_{eff} \): Effective length, ft
- \( d \): Vessel Internal diameter, in

Slenderness Ratio

\[ SlenRatio = \frac{12 \cdot L_{ss}}{d} \]

A table of possible results will be computed, the final options will depend on the maximum and minimum slenderness ratio chosen by the user.

Recommended Maximum and minimum Slenderness Ratios for Horizontal separators:

- Minimum Slenderness Ratio \( SIR_{min} = 3 \)
- Maximum Slenderness Ratio \( SIR_{Max} = 5 \)

*VBA CODE FOR THREE-PHASE HORIZONTAL SEPARATOR DESIGN AVAILABLE IN: [Annex B.9: ThreePhaseHorizontal module](#)*

### 3.2.4.4. Three-phase vertical separators

As with vertical two-phase separators, a minimum diameter must be maintained to assure adequate gas capacity. In addition, vertical three-phase separators must maintain a minimum diameter to allow the 500 micron water droplets to settle. The height of the three-phase separator is determined from retention time considerations.
Specific gravity difference

\[ \text{Delta}SG = \frac{\text{Denw} - \text{Deno}}{62.4} \]

Where:

\( \text{Delta}SG \): Specific gravity difference, [-]

Deno: Oil density, lb/ft³

Denw: Density of the water, lb/ft³

Gas Capacity Constrain

\[ k = \frac{(\text{DenG} \cdot \text{Cd})}{((\text{Deno} + \text{Denw}) - \text{DenG}) \cdot \text{dOG})} \]

\[ d^2g = \frac{(5040 \cdot (T + 460) \cdot Z \cdot \left( \frac{\text{qg}}{1000} \right) \cdot k}{p} \]

Where:

\( d_g \): Minimum diameter for gas constrain, in

Deno: Oil density, lb/ft³

DenG: Density of the gas, lb/ft³

Denw: Density of the water, lb/ft³

dOG: Diameter of Oil Droplet in Gas Phase, microns

Cd: drag coefficient, [-]

Z: Compressibility factor, [-]

T: in-situ Temperature, ºR

qg: Gas Flowrate, MSCFD

P: in-situ Pressure, Psia
Liquid Capacity Constrain

\[ d^2l = \frac{6690 \cdot Viso \cdot qo}{DeltaSG \cdot dWO^2} \]

Where:

- \(dl\): Minimum diameter for liquid constrain, in
- \(DeltaSG\): Specific gravity difference, [-]
- \(dWO\): Diameter of Water Droplet in Oil Phase, microns
- \(Viso\): Oil viscosity, cp
- \(tro\): Oil retention time, min

Choose the larger diameter between liquid and gas constrain.

Retention time constrain

\[ ht = ho + hw \]

\[ ht = \frac{tro \cdot qo + trw \cdot qw}{0.12 \cdot d^2} \]

Where:

- \(ht\): High of oil pad + high from water to interface, in
- \(d\): Vessel Internal diameter, in
- \(q_o\): Oil Flowrate, BPD
- \(q_w\): Water Flowrate, BPD
- \(trw\): Water retention time, min
- \(tro\): Oil retention time, min

Seam-to-Seam Length and Slenderness Ratio

\[ L_{ss} = \frac{ht + 76}{12} \]
Lss = \frac{ht + d + 40}{12}

Where:

Lss: Seam to seam length, ft

h: High of the liquid volume, in

d: Vessel Internal diameter, in

Choose the bigger Seam to Seam length

Slenderness Ratio

\[ SlenRatio = \frac{12 \cdot Lss}{d} \]

A table of possible results will be computed, the final options will depend on the maximum and minimum slenderness ratio chosen by the user.

Recommended Maximum and minimum Slenderness Ratios for Vertical separators:

Minimum Slenderness Ratio SIR min = 1.5

Maximum Slenderness Ratio SIR Max = 3

*VBA CODE FOR THREE-PHASE VERTICAL SEPARATOR DESIGN AVAILABLE IN: Annex B.10: ThreePhaseVertical module

The FOURTH STEP of the program is the design of the Chosen separator as we have seen before.
3.3. Chapter III: References


Pereyra Eduardo (2015), Flow Assurance Course Slides, The University of Tulsa, Tulsa, OK, USA.
COMPUTER APPLICATION FOR SEPARATORS DESIGN

DOCUMENT II: ECONOMIC ANALYSIS
4. Document II: Economic Analysis

4.1. Capital Cost Estimation

The **LAST STEP** of the program is to estimate the cost of the designed equipment as follows:

For capital cost estimation, we distinguish the following terms:

<table>
<thead>
<tr>
<th>Term</th>
<th>Symbol</th>
<th>Definition</th>
</tr>
</thead>
<tbody>
<tr>
<td>Purchased equipment cost</td>
<td>$C_p$</td>
<td>Free on board cost of equipment at manufacturer’s site</td>
</tr>
<tr>
<td>Base equipment cost</td>
<td>$C_p^0$</td>
<td>Purchased equipment cost for equipment made of the most common material and operating at near ambient pressures</td>
</tr>
<tr>
<td>Bare module cost</td>
<td>$C_{BM}$</td>
<td>Equipment cost including labor and installation; freight, insurance, and taxes; construction overhead; and contractor expenses</td>
</tr>
<tr>
<td>Total module cost</td>
<td>$C_{TM}$</td>
<td>Total capital cost, including contingency and fees, to making small-to-moderate expansions or alterations to an existing facility</td>
</tr>
<tr>
<td>Grass roots cost</td>
<td>$C_{GR}$</td>
<td>Total cost for a completely new facility in which construction is started on essentially undeveloped land</td>
</tr>
</tbody>
</table>

The program determines the base equipment cost for each piece of equipment using the equation:

$$ \log_{10} C_p^0 = K_1 + K_2 \log_{10}(A) + K_3 [\log_{10}(A)]^2 $$

Where:

- $C_p^0$: Base equipment cost, $\}$
- $A$: capacity or size parameter
- $K_1$ through $K_3$ vary with the type of equipment.
To calculate the total module cost from the base equipment cost, the bare module method is used. The equation used by the program is

\[ C_{BM} = C_p^0 F_{BM} = C_p^0 (B_1 + B_2 F_M F_p) \]

Where:

- \( C_{BM} \): bare module cost, $
- F_{BM} \): bare module factor
- \( F_M \): material factor
- \( F_p \): pressure factor
- \( B_1 \) and \( B_2 \) depend on the equipment type.

To find the total module cost, 18% is generally added to the bare module cost for contingency costs and fees:

\[ C_{TM} = 1.18 \sum_{i=1}^{n} C_{BM,i} \]

Where:

- \( C_{BM} \): bare module cost, $
- C_{TM} \): total module cost, $

Finally, if a grass roots plant is being built, an additional auxiliary facilities cost is added to the total module cost. This auxiliary facilities cost, which accounts for site development, auxiliary buildings, and off-sites and utility construction.

\[ C_{GR} = 1.5 C_{TM} \]

Where:

- \( C_{GR} \): Grass roots cost, $
- C_{TM} \): total module cost, $

**SPECIFIC EQUIPMENT:**

All capital costs are based on 2001 prices

CEPCI (2001) = 397

CEPCI (2014) = 577

Bare module cost should be multiply by the factor \( \left( \frac{577}{397} \right) \)

**Horizontal separators:**

\( K_1 = 3.5565, K_2 = 0.3776, K_3 = 0.0905 \)

\( A = \text{volume (m}^3\text{)} \ (0.1 < A < 628) \)

\( B_1 = 1.49, B_2 = 1.52 \)

\( F_M = \text{provided in Table 6} \)

\[
F_p = \frac{(P + 1)D}{2[850 - 0.6(P + 1)]} + 0.00315 \frac{0.0063}{0.0063}
\]

Where:

\( P: \text{Pressure, bar} \)

\( d: \text{diameter, m} \)

or \( F_p = 1 \), whichever is greater

**Vertical separators:**

\( K_1 = 3.4974, K_2 = 0.4485, K_3 = 0.1074 \)

\( A = \text{volume (m}^3\text{)} \ (0.3 < A < 520) \)

\( B_1 = 2.25, B_2 = 1.82 \)

\[
F_p = \frac{(P + 1)D}{2[850 - 0.6(P + 1)]} + 0.00315 \frac{0.0063}{0.0063}
\]
Where:

\[ P: \text{Pressure, bar} \]

\[ d: \text{diameter, m} \]

or \( F_p = 1 \), whichever is greater

**Table 6: FM correction factor**

<table>
<thead>
<tr>
<th>Material</th>
<th>Correction factor ( F_M )</th>
</tr>
</thead>
<tbody>
<tr>
<td>Carbon steel</td>
<td>1.0</td>
</tr>
<tr>
<td>Stainless steel (low grades)</td>
<td>2.1</td>
</tr>
<tr>
<td>Stainless steel (high grades)</td>
<td>3.2</td>
</tr>
<tr>
<td>Monel</td>
<td>3.6</td>
</tr>
<tr>
<td>Inconel</td>
<td>3.9</td>
</tr>
<tr>
<td>Nickel</td>
<td>5.4</td>
</tr>
<tr>
<td>Titanium</td>
<td>7.7</td>
</tr>
</tbody>
</table>

*VBA CODE FOR COST ESTIMATION AVAILABLE IN: Annex B.11: Sheet7 (Cost)*

**Slug Catcher cost estimation**

For the slug catcher:

If the slug catcher chosen is a Pressure Vessel type the program will calculate the cost as if the slug catcher is a separator.

If the slug catcher chosen is finger type the program will calculate the cost as follows:

\[ CBM = W \cdot l \cdot Nbottles \cdot 0.0005 \cdot CSprice \cdot FI \]

Where:

\[ CBM: \text{bare module cost, } \$

\[ W: \text{Pipe Weight, lb/ft} \]
l: Pipe length, ft

Nbottles: Number of Bottles, [-]

CSprice: Carbon steel price, $/ton

FI: installation Factor
COMPUTER APPLICATION FOR
SEPARATORS DESIGN

DOCUMENT III: ANNEXES
DOCUMENT III: ANNEXES

ANNEX A: COMPUTER APPLICATION / PROGRAM
5. **Document III: Annexes**

**Annex A: Computer Application / Program Download**

The Computer application can be downloaded in the following link:

https://drive.google.com/folderview?id=0B-04Bf8F-migYmdKV1NCLW5ZVlU&usp=sharing

*Note: for the computer application to work the whole folder has to be downloaded and saved in the computer

**Annex A.1: Computer Application / Program practical example**

**FIRST STEP: Black Oil Model**
SECOND STEP: FlowPatn 1.0

THIRD STEP: Slug Catcher Design
FOURTH STEP: Separator Design
LAST STEP: Capital Cost Estimation
DOCUMENT III: ANNEXES

ANNEX B: VBA CODE
Annex B: VBA code

Annex B.1: DeGhetto Module

'*------------------------------------------------------------------*
'                    Bubble Point Pressure [Pb]
'------------------------------------------------------------------'*

Public Function PbFnc(API As Double, Rsb As Double, T As Double, Tsep As Double, SG As Double, Psep As Double) As Double
If API < 10 Then
    "************** Extra Heavy **************"
    Range("correlation1") = "Standing"
    PbFnc = 18 * ((Rsb / SG) ^ 0.83) * ((10 ^ (0.00091 * T)) / (10 ^ (0.0125 * API)))
ElseIf 10 < API And API < 22.3 Then
    "*********** Heavy **********"
    Range("correlation1") = "Modified Standing"
    PbFnc = 15.7286 * ((Rsb / SG) ^ 0.7885) * ((10 ^ (0.002 * T)) / (10 ^ (0.0142 * API))
ElseIf 22.3 < API And API < 31.1 Then
    "*********** Medium **********"
    Range("correlation1") = "Modified Kartoatmodjo"
    Dim SG1
    SG1 = SG * (1 + 0.1595 * API ^ 0.4078 * Tsep ^ -0.2466 * Log10(Psep / 114.7))
    PbFnc = (Rsb / (0.09902 * SG1 ^ 0.2181 * 10 ^ (7.2153 * API / (T + 460)))) ^ 0.9997
Else
    "*********** Light **********"
    Range("correlation1") = "Modified Standing"
    PbFnc = 15.7648 * ((Rsb / SG) ^ 0.7857) * ((10 ^ (0.0009 * T)) / (10 ^ (0.0148 * API))
End If

'*------------------------------------------------------------------*
'                       Dissolve Gas Solution Gas Oil Ratio[Rs]
'------------------------------------------------------------------'*

Public Function RsFnc(SG As Double, P As Double, Pb As Double, API As Double, T As Double, Tsep As Double, Psep As Double) As Double
If P > Pb Then
    ' Above Bubble point use Pb instead of p'
    P = Pb
Else
End If
If API < 10 Then
    "************** Extra Heavy **************"
    Range("correlation2") = "Modified Standing"
    RsFnc = SG * ((P * 10 ^ (0.0169 * API - 0.00156 * T)) / (10.7052)) ^ 1.1128
ElseIf 10 < API And API < 22.3 Then  '*******************  Heavy *******************
    Range("correlation2") = "Modified Vasquez-Beggs"

Dim SGcorr1

SGcorr1 = SG * (1 + 0.5912 * API * Tsep * Log10(Psep / 114.7) * 10 ^ -4)

RsFnc = (SGcorr1 * (P ^ 1.2057) * 10 ^ ((10.9267 * API) / (T + 460))) / (56.434)

ElseIf 22.3 < API And API < 31.1 Then  '*******************  Medium *******************
    Range("correlation2") = "Modified Kartoatmodjo"

Dim SGcorr2

SGcorr2 = SG * (1 + 0.1595 * API ^ 0.4078 * Tsep ^ -0.2466 * Log10(Psep / 114.7))

RsFnc = (0.10084 * SGcorr2 ^ 0.2556 * P ^ 0.9868 * 10 ^ (7.4576 * API / (T + 460)))

Else
    '*******************  Light *******************
    Range("correlation2") = "Modified Kartoatmodjo"

Dim SGcorr3

SGcorr3 = SG * (1 + 0.1595 * API ^ 0.4078 * Tsep ^ -0.2466 * Log10(Psep / 114.7))

RsFnc = (0.01347 * SGcorr3 ^ 0.3873 * P ^ 1.1715 * 10 ^ (12.753 * API / (T + 460)))

End If

End Function

'*****************************************************************
'                            Compressability factor [Co]
'*******************************************************************

Public Function CoFnc(P As Double, T As Double, Rs As Double, API As Double, SG As Double, Tsep As Double, Psep As Double, Pb As Double, Bo As Double) As Double

Dim SGcorr As Double

SGcorr = SG * (1 + 0.5912 * API * Tsep * Log10(Psep / 114.7) * 10 ^ -4)

If API < 10 Then  '*******************  Extra Heavy *******************
    Range("correlation4") = "Modified Vasquez-Beggs"

CoFnc = (-889.6 + 3.1374 * Rs + 20 * T - 627.3 * SGcorr - 81.4476 * API) / (P * 10 ^ 5)

ElseIf 10 < API And API < 22.3 Then  '*******************  Heavy *******************
    Range("correlation4") = "Modified Vasquez-Beggs"

CoFnc = (-2841.8 + 2.9646 * Rs + 25.5439 * T - 1230.5 * SGcorr + 41.91 * API) / (P * 10 ^ 5)

ElseIf 22.3 < API And API < 31.1 Then  '*******************  Medium *******************
    Range("correlation4") = "Modified Vasquez-Beggs"

CoFnc = (-705.288 + 2.2246 * Rs + 26.0644 * T - 2080.823 * SGcorr - 9.6807 * API) / (P * 10 ^ 5)
Else

' **************** Light ******************

Range("correlation4") = "Modified Labeled's"

'---------------------------------- From BoFnc-----------------------------------

Dim f As Double

Dim SGo As Double

Dim SGcorr2 As Double

Dim Bovalue As Double

SGo = 141.5 / (131.5 + API)

SGcorr2 = SG * (1 + 0.1595 * API ^ 0.4078 * Tsep ^ -0.2466 * Log10(Psep / 114.7))

f = Rs ^ 0.755 * SGcorr ^ 0.25 * SGo ^ -1.5 + 0.45 * T

Bovalue = 0.98496 + 0.0001 * f ^ 1.5

'---------------------------------------------------------------------

CoFnc = (10 ^ (-6.1646) * BoFnc ^ (1.8789) * API ^ (0.3646) * T ^ (0.1966)) -

(1 - (Pb / P)) * (10 ^ (-8.98) * BoFnc ^ (3.9392) * T ^ (1.349))

End If

End Function

'***********************************************************

Public Function BoFnc(P As Double, Pb As Double, Rs As Double, API As Double, T As Double, SG As Double, Tsep As Double, Psep As Double, Co As Double) As Double

Range("correlation3") = "Kartoatmodjo's"

Dim f As Double

Dim Bovalue As Double

Dim SGo As Double

Dim SGcorr As Double

SGo = 141.5 / (131.5 + API)

SGcorr = SG * (1 + 0.1595 * API ^ 0.4078 * Tsep ^ -0.2466 * Log10(Psep / 114.7))

f = Rs ^ 0.755 * SGcorr ^ 0.25 * SGo ^ -1.5 + 0.45 * T

Bovalue = 0.98496 + 0.0001 * f ^ 1.5

If P > Pb Then

BoFnc = Bovalue * Exp(Co * (Pb - P))

Else

BoFnc = Bovalue

End If

End Function

'***********************************************************

Formation Volume Factor [Bo]
Public Function DenOFnc(API As Double, SG As Double, Rs As Double, Bo As Double, T As Double, P As Double, Pb As Double, Tsep As Double, Psep As Double, Co As Double) As Double

    Range("correlation5") = "Mass Balance"
    Dim SGgd As Double
    Dim SGo As Double
    Dim SGcorr As Double
    Dim DenOValue As Double

    SGo = 141.5 / (131.5 + API)
    SGgd = (12.5 + API) / 50 - 3.5715 * 10 ^ -6 * API * Rs
    SGcorr = SG * (1 + 0.1595 * API ^ 0.4078 * Tsep ^ -0.2466 * Log10(Psep / 114.7))
    DenOValue = (SGo * 62.4 + (SGgd * (0.0764) * Rs) / 5.614) / Bo

    If P > Pb Then
        DenOFnc = DenOValue * Exp(Co * (P - Pb))
    Else
        DenOFnc = DenOValue
    End If

End Function

Public Function VisODFnc(API As Double, T As Double, SG As Double, Tsep As Double, Psep As Double) As Double

    If API < 10 Then
        Range("correlation6") = "Egbogah Jack's"
        VisODFnc = (10 ^ (10 ^ (1.90296 - 0.012619 * API - 0.61748 * Log10(T)))) - 1
    ElseIf 10 < API And API < 22.3 Then
        Range("correlation6") = "Egbogah Jack's"
        VisODFnc = (10 ^ (10 ^ (2.06492 - 0.0179 * API - 0.70226 * Log10(T)))) - 1
    Else
        Range("correlation6") = "Egbogah Jack's"
        VisODFnc = (10 ^ (10 ^ (2.06492 - 0.0179 * API - 0.70226 * Log10(T)))) - 1
    End If

End Function
ElseIf 22.3 < API And API < 31.1 Then
          '************** Medium ******************'
          Range("correlation6") = "Modified Kartoatmodjo"
          Dim SGcorr
          SGcorr = SG * (1 + 0.1595 * API ^ 0.4078 * Tsep ^
                          -0.2466 * Log10(Psep / 114.7))
          Else
          '************** Light ******************'
          Range("correlation6") = "Egbogah Jack's"
          VisODFnc = (10 ^ (10 ^ (1.67083 - 0.017628 * API - 0.61304 * Log10(T)))) - 1
          End If
End Function

  '****************************************************************
  ********************
  '                                Live Oil Viscosity [VisO]
  ************************************************************************************

Public Function VisOLFnc(VisOD As Double, Rs As Double, API As Double, P As Double, _
Pb As Double) As Double
  If API < 10 Then
          '************** Extra Heavy **************'
          Dim Y As Double
          Dim f As Double
          Dim VisOvalue As Double
          Y = 10 ^ (-0.00081 * Rs)
          f = (-0.0335 + 1.0785 * 10 ^ (-0.000845 * Rs)) * VisOD ^ (0.5798 + 0.3432 * Y)
          VisOvalue = 2.3945 + 0.8927 * f + 0.001567 * f ^ 2
          If P > Pb Then
            Range("correlation7") = "Modified Labedi's"
            VisOLFnc = VisOvalue - ((1 - (P / Pb)) * (((10 ^ -2.19) * (VisOvalue ^ 1.055)) _
                                   * (Pb ^ 0.3132)) / (10 ^ (-0.00288 * API))))
          Else: VisOLFnc = VisOvalue
          Range("correlation7") = "Modified Kartoatmodjo"
          End If
  ElseIf 10 < API And API < 22.3 Then
          '************** Heavy ******************'
          Dim y1 As Double
          Dim f1 As Double
          Dim VisOvalue1 As Double
          y1 = 10 ^ (-0.00081 * Rs)
\[ f_1 = (0.2478 + 0.61145 \times 10^{-0.000845 \times R_s}) \times \text{VisOD}^{0.4731 + 0.5158 \times y_1} \]
\[ \text{VisOvalue1} = -0.6311 + 1.078 \times f_1 - 0.003653 \times f_1^2 \]

If \( P > P_b \) Then

Range("correlation7") = "Modified Kartatmodjo"

\[ \text{VisOLFnc} = 0.9886 \times \text{VisOvalue1} + 0.002763 \times (P - P_b) \times (-0.01153 \times \text{VisOvalue1}^1.7933 + 0.0316 \times \text{VisOvalue1}^1.5939) \]

Else: \( \text{VisOLFnc} = \text{VisOvalue1} \)

Range("correlation7") = "Modified Kartatmodjo"
End If

ElseIf 22.3 < API and API < 31.1 Then

************** Medium **************

Dim y2 As Double
Dim f2 As Double
Dim VisOvalue2 As Double

\[ y_2 = 10^{-0.00081 \times R_s} \]
\[ f_2 = (0.2038 + 0.8591 \times 10^{-0.000845 \times R_s}) \times \text{VisOD}^{0.3855 + 0.5664 \times y_2} \]
\[ \text{VisOvalue2} = 0.0132 + 0.9821 \times f_2 - 0.005215 \times f_2^2 \]

If \( P > P_b \) Then

Range("correlation7") = "Modified Labedi's"

\[ \text{VisOLFnc} = \text{VisOvalue2} - ((1 - (P / P_b)) \times ((10^{-2.19}) \times (\text{VisOvalue2}^{1.055}) \times (P_b^{0.3132})) / (10^{0.0099 \times \text{API}})) \]

Else: \( \text{VisOLFnc} = \text{VisOvalue2} \)

Range("correlation7") = "Modified Kartatmodjo"
End If

Else

************** Light **************

Dim VisOvalue3 As Double

\[ \text{VisOvalue3} = (25.1921 \times (R_s + 100)^{-0.6487}) \times (\text{VisOD}^{2.7516 \times (R_s + 150)} \times -0.2135) \]

If \( P > P_b \) Then

Range("correlation7") = "Modified Labedi's"

\[ \text{VisOLFnc} = \text{VisOvalue3} - ((1 - (P / P_b)) \times ((10^{-2.19}) \times (\text{VisOvalue3}^{1.055}) \times (P_b^{0.3132})) / (10^{0.0099 \times \text{API}})) \]

Else: \( \text{VisOLFnc} = \text{VisOvalue3} \)

Range("correlation7") = "Modified Beggs & Robinson"
End If
Annex B.2: Sheet2 (BlackOilMode) code

Dim A As Double
A = (3.14159265 / 4) * (Range("d") / 12) ^ 2
ql = Range("qw") + Range("qo")
Range("Vsl") = (ql * 5.615 / 86400) / (A)

Dim SG As Double
SG = Range("SG")
If Range("qg_Rp") = "qg" Then
  Range("Vsg") = (Range("qg") * BgFnc(Range("T") + 460, Range("P"), SG) * 10 ^ 3 / 86400) / (A)
Else
  Range("Vsg") = (ql * (Rp - Range("Rs")) * BgFnc(Range("T") + 460, Range("P"), SG) / 86400) / (A)
End If

Annex B.3: GaslProperties module

' *******************************************************
' Hall-Yarborough z-factor
' #3 This function computes the actual z factor utilizing the reduced pressure and temperature
' utilizing the Successive Substitution Method
Public Function zFacHYFnc(Pr As Double, Tr As Double) As Double
  Dim i As Long
  Const nmax As Long = 1000
  Const Tol As Double = 0.00001
  Dim xi As Double            ' solution at i
  Dim xi1 As Double           ' solution i-1
  Dim fxi As Double
  Dim T As Double
  T = 1# / Tr
  xi1 = 0.06125 * Pr * T * Exp(-1.2 * (1 - T) ^ 2)  ' first iteration
  For i = 2 To nmax
fxi = fFnc(xi1, Pr, T)

xi = xi1 - fxi / derfFNC(xi1, Pr, T)

If (Abs((xi - xi1) / xi) < Tol) Or Abs(fxi) < Tol Then
' method converge
Exit For
End If

xi1 = xi

Next i

If i > nmax Then
' no convergence
MsgBox ("no convergence")

zFacHYFnc = -999.99
Else

zFacHYFnc = 0.06125 * Pr * T * Exp(-1.2 * (1 - T) ^ 2) / xi

End If

End Function

' #1 This function performs the derivative equations
Public Function derfFNC(Y As Double, Pr As Double, Tr As Double) As Double

Dim A As Double
Dim b As Double
Dim c As Double

A = ((((Y - 4#) * Y + 4#) * Y + 4#) * Y + 1#) / (1 - Y) ^ 4
b = -((9.16 * Tr - 19.52) * Tr + 29.52) * Tr * Y

End Function

' #2 This function actually computes the derivative equations for later use
Public Function fFnc(Y As Double, Pr As Double, Tr As Double) As Double

Dim A As Double
Dim b As Double
Dim c As Double
Dim D As Double

A = -0.06125 * Pr * Tr * Exp(-1.2 * (1# - Tr) ^ 2)
b = (((1# - Y) * Y + 1#) * Y + 1#) * Y / (1# - Y) ^ 3
c = -((4.58 * Tr - 9.76) * Tr + 14.76) * Tr * Y * Y
D = ((42.4 * Tr - 242.2) * Tr + 90.7) * Tr * Y ^ (2.18 + 2.82 * Tr)
fFnc = A + b + c + D

End Function

Public Function BgFnc(T As Double, P As Double, SG As Double) As Double
' SG = SG
' T = Temperature [°R]
' P = Pressure [psi]
Dim Z As Double ' compresibility factor
Dim Tc As Double ' critical temperature [R]
Dim Tr As Double ' reduce temperature [R]
Dim Pr As Double ' reduce pressure [psi]
Dim Pc As Double ' critical pressure [psi]
Pc = 756.8 - 131 * SG - 3.6 * SG ^ 2 ' Sutton 1985 Corelation
Tc = 169.2 + 349.5 * SG - 74 * SG ^ 2
Tr = (T) / Tc
Pr = P / Pc
Z = zFacHYFnc(Pr, Tr)
Sheet2.Range("Z") = Z
BgFnc = 0.0283 * Z * T / P

End Function

Public Function DenGFnc(T As Double, P As Double, SG As Double) As Double
' SG = SG
' T = Temperature [°R]
' P = Pressure [psi]
Dim Z As Double ' compresibility factor
Dim Tc As Double ' critical temperature [R]
Dim Tr As Double ' reduce temperature [R]
Dim Pr As Double ' reduce pressure [psi]
Dim Pc As Double ' critical pressure [psi]
Dim MW As Double
Pc = 756.8 - 131 * SG - 3.6 * SG ^ 2 ' Sutton 1985 Corelation
Tc = 169.2 + 349.5 * SG - 74 * SG ^ 2
Tr = (T) / Tc
\[
Pr = \frac{P}{P_c}
\]
\[
Z = z\text{FacHYFunc}(Pr, Tr)
\]
\[
\text{DenGFnc} = \frac{(2.7 \times P \times SG)}{(Z \times T)}
\]

End Function

Public Function VisGFnc(T As Double, P As Double, SG As Double) As Double
  ' SG = SG
  ' T = Temperature [°R]
  ' P = Pressure [psi]
  Dim Z As Double ' compresibility factor
  Dim Tc As Double ' critical temperature [R]
  Dim Tr As Double ' reduce temperature [R]
  Dim Pr As Double ' reduce pressure [psi]
  Dim Pc As Double ' critical pressure [psi]
  Pc = 756.8 - 131 * SG - 3.6 * SG ^ 2 ' Sutton 1985 Corelation
  Tc = 169.2 + 349.5 * SG - 74 * SG ^ 2
  Tr = (T) / Tc
  Pr = P / Pc
  Z = z\text{FacHYFunc}(Pr, Tr)
  Dim x As Double
  Dim Y As Double
  Dim k As Double
  Dim M As Double
  Dim pg As Double
  M = SG * 28.97
  x = 3.448 + 0.01009 * M + (986.4 / T)
  Y = 2.447 - 0.2224 * x
  k = (9.379 + 0.01607 * M) * T ^ 1.5 / (209.2 + 19.26 * M + T)
  pg = 0.0433 * SG * (P / (Z * T))
  VisGFnc = 0.0001 * k * Exp(x * pg ^ Y)
End Function
Annex B.4: Sheet4 (Slug Catcher) code

Private Sub CommandButton1_Click()
    If Sheet3.Range("Slug") = "YES" Then
        Range("slugd").Clear
        '************************///// Natural Slugging////**************************************
        Vsl = Sheet2.Range("Vsl")
        Vsg = Sheet2.Range("Vsg")
        D = Sheet2.Range("d")
        Dim VM As Double
        Dim HLs As Double
        Dim LSmean As Double
        Dim LSmax As Double
        Dim VslugNat As Double
        Dim tslugNat As Double
        VM = Vsl + Vsg
        HLs = 1 / (1 + (VM / 28.41) ^ 1.39)
        LSmean = Exp(-2.663 + 5.441 * Sqr(Log(D)) + 0.059 * Log(VM))
        LSmax = Exp(3.09 * 0.5 + Log(LSmean) - (0.5 ^ 2) / 2) 'sigma = 0.5'
        VslugNat = LSmax * ((3.14159265 / 4) * (D / 12) ^ 2) * HLs
        tslugNat = LSmax / VM
        Range("LsmaxN") = LSmax
        Range("VslugN") = VslugNat
        Range("TslugN") = tslugNat
        '******************************///// Pigging////*****************
        L = Sheet2.Range("L")
        HL = Sheet2.Range("HL")
        A = (3.14159265 / 4) * (D / 12) ^ 2
        Dim Ttransient As Double
        Dim VslugPig As Double
        Ttransient = L / VM
        VslugPig = (HL * A * L) * (1 - 0.02) - (Vsl * A * Ttransient) 'fleak = 0.02'
        tslugPig = VslugPig / (A * VM)
If VslugPig < 0 Or tslugPig < 0 Then
    Range("VslugP") = "."
    Range("TslugP") = "."
Else
    Range("VslugP") = VslugPig
    Range("TslugP") = tslugPig
End If

'******************************//// Catcher Design ////***********************
If Sheet2.Range("L") = "" Then
    slugflowrateP = 0
Else
    slugflowrateP = VslugPig / tslugPig
End If
slugflowrateN = VslugNat / tslugNat
surgeVolN = (slugflowrateN - 0) * tslugNat 'qsep=0 worst scenario'
surgeVolP = (slugflowrateP - 0) * tslugPig 'qsep=0 worst scenario'
If surgeVolN > surgeVolP Then
    surgeVol = surgeVolN
    Range("Type") = "Natural Slugging"
Else
    surgeVol = surgeVolP
    Range("Type") = "Pigging"
End If
Range("SurgeVolume") = surgeVol

'////////////////////////////////Pressure vessel/////////////////////////////////
Range("VolumePV") = surgeVol * 1.5
dPV = (((4 * Range("VolumePV") / (3.14159265 * 4)) ^ (1 / 3)) * 12 * L/D = 4
If dPV < 16 Then
    dn = 16
ElseIf dPV < 20 Then
    dn = 20
ElseIf dPV < 24 Then
    dn = 24
ElseIf dPV < 30 Then
dn = 30
ElseIf dPV < 36 Then
dn = 36
ElseIf dPV < 42 Then
dn = 42
ElseIf dPV < 48 Then
dn = 48
ElseIf dPV < 54 Then
dn = 54
ElseIf dPV < 60 Then
dn = 60
ElseIf dPV < 66 Then
dn = 66
ElseIf dPV < 72 Then
dn = 72
ElseIf dPV < 78 Then
dn = 78
ElseIf dPV < 84 Then
dn = 84
ElseIf dPV < 90 Then
dn = 90
ElseIf dPV < 96 Then
dn = 96
End If

Range("dPV") = dn
Range("LPV") = 4 * dn / 12

'//////////////////////////////// Finger Type ////////////////////////////////////////
Range("VolumeFT") = surgeVol * 1.25
Range("dFT") = D
Range("LFT") = 500              '1000> L > 250 --> 500ft'
Range("Vbottle") = (3.14159265 * Range("LFT") * (Range("dFT") / 12) ^ 2) / 4
Range("Nbottle") = Range("VolumeFT") / Range("Vbottle")
Else
    Range("slugd") = "No Slug Flow"
    Range("slugd").Interior.ColorIndex = 3
End If
End Sub

Annex B.5: Cd module

************************************************************************************
'                               Drag Coefficient [Cd]
************************************************************************************

Public Function CdFnc(Re As Double, Corr As Double) As Double

*****************************************************************************************
If (Re > 1) And (Re < 1000) Then
    If Corr = 1 Then
        'Sciller Naumann
        CdFnc = 24 / Re * (1 + 0.15 * Re ^ 0.687)
    ElseIf Corr = 2 Then
        'Ishii and Zuber
        CdFnc = 24 / Re * (1 + 0.1 * Re ^ 0.75)
    ElseIf Corr = 3 Then
        'Ihme
        CdFnc = 24 / Re + 5.48 * Re ^ -0.573 + 0.36
    ElseIf Corr = 4 Then
        'SPE 56645
        CdFnc = 24 / Re + 3 / Sqr(Re) + 0.34
    End If
Else
    If Re < 1 Then

Stokes Law

CdFnc = 24 / Re

Else

Newton's regimen

CdFnc = 0.44

End If

End If

End Function

Annex B.6: Cd calculation

************************************************************************************

Cd Calculations

************************************************************************************

************************************************************************************ Initial Guess ************************************************************************************

Cdi = 0.34

Vti = 0.0119 * (((Deno - DenG) / DenG) * (dOG / Cdi)) ^ 0.5

Vtf = Vti

Tol = 1

n = 0

************************************************************************************ Iterations ************************************************************************************

While ((Tol >= 0.0001) And (n < 100))

Re = (0.0049 * DenG * dOG * Vtf) / VisG

newCd = CdFnc(Re, Corr)

Vtf = 0.0119 * (((Deno - DenG) / DenG) * (dOG / newCd)) ^ 0.5

Tol = Abs((Vtf - Vti) / Vtf)

n = n + 1

Wend

Vt = Vtf

Cd = newCd

Range("cd") = Cd
Annex B.7: TwoPhaseHorizontal module

Public Function TwoPhaseH(tro As Double, trw As Double, Deno As Double, Denw As Double, DenG As Double, Viso As Double, Visw As Double, VisG As Double, Qo As Double, Qw As Double, Qg As Double, T As Double, P As Double, Z As Double, Dvesel As Double, dOG As Double, dWO As Double, Cd As Double)

' ***************************************************************************************
' **
' ** Internal Varibles:
' Dim dLeff As Double         ' Auxiliar varible gas capacity[-]
' Dim Leff As Double          ' Gravity settling length [ft]
' Dim LeffGas As Double       ' Gravity settling length for Gas Capacit [ft]
' Dim LeffLiq As Double       ' Gravity settling length for Liq Capacit [ft]
' Dim d2Leff As Double        ' Liquid retention constrain
' Dim Lss As Double           ' Seam to Seam length [ft]
' Dim SlenRatio As Double     ' Slenderness ratio [-]
' Dim Vec(1 To 5) As Double   
' Dim k As Double
' **
' ***************************************************************************************
' Gas constrain Value
' k = Sqr((DenG * Cd) / (((Deno + Denw) - DenG) * dOG))
dLeff = (420 * (T + 460) * Z * (Qg / 1000) * k) / P

' Liquid retention constarin
d2Leff = 1.42 * (Qw * trw + Qo * tro)

' Gravity Settling section constrain
LeffLiq = d2Leff / (Dvesel * Dvesel)
LeffGas = dLeff / (Dvesel)

' Lss for gas Constarin
If LeffGas < LeffLiq Then
' Liquid Constrain Govern
Leff = LeffLiq
Lss = Leff / 0.75

Else
  ' Gas Constrain Govern
  Leff = LeffGas
  Lss = Leff + Dvesel / 12
End If

' Slenderness ratio
SlenRatio = 12 * Lss / Dvesel

" Vector Conversion
Vec(1) = dLeff
Vec(2) = d2Leff
Vec(3) = Leff
Vec(4) = Lss
Vec(5) = SlenRatio
TwoPhaseH = Vec
End Function

Annex B.8: TwoPhaseVertical module

Public Function TwoPhaseV(tro As Double, trw As Double, Deno As Double, _
                        Denw As Double, DenG As Double, Viso As Double, _
                        Qo As Double, Qw As Double, Qg As Double, T As Double, _
                        P As Double, Z As Double, Dvesel As Double, _
                        dOG As Double, dWO As Double, Cd As Double)
  ' ***************************************************************************************
  ' **    Internal Varibles:
  Dim dmin As Double
  Dim ht As Double    ' ho + hw'
  Dim Lss1 As Double
  Dim Lss2 As Double
  Dim Vec(1 To 4) As Double
  Dim k As Double
  ' ***************************************************************************************
  ' Gas constrain Value
  k = Sqr((DenG * Cd) / ((Deno - DenG) * dOG))
\[ \text{dmin} = \frac{(5040 \times (T + 460) \times Z \times (Qg / 1000) \times k)}{P} \times 0.5 \]

' Liq Retention Constrain (ho + hw)

\[ \text{ht} = \frac{(tro \times Qo + trw \times Qw)}{(0.12 \times Dvesel \times Dvesel)} \]

' seam to seam

\[ \text{Lss1} = \frac{(ht + 76)}{12} \]

\[ \text{Lss2} = \frac{(ht + Dvesel + 40)}{12} \]

If Lss1 > Lss2 Then

Lss = Lss1

Else

Lss = Lss2

End If

' Slenderness ratio

\[ \text{SlenRatio} = 12 \times \frac{Lss}{Dvesel} \]

' Vector Conversion

\[ \text{Vec(1)} = \text{dmin} \]

\[ \text{Vec(2)} = \text{ht} \]

\[ \text{Vec(3)} = \text{Lss} \]

\[ \text{Vec(4)} = \text{SlenRatio} \]

\[ \text{TwoPhaseV} = \text{Vec} \]

End Function

**Annex B.9: ThreePhaseHorizontal module**

Public Function ThreePhaseH(tro As Double, trw As Double, Deno As Double, Denw As Double, DenG As Double, Viso As Double, Visw As Double, VisG As Double, Qo As Double, Qw As Double, Qg As Double, T As Double, P As Double, Z As Double, Dvesel As Double, dOG As Double, dWO As Double, Cd As Double)

' ***************************************************************

' ** Internal Varibles:

Dim DeltaSG As Double ' Diference in specific gravity [-]
Dim dLeff As Double ' Auxiliar varible gas capacity[-]
Dim Leff As Double ' Gravity settling length [ft]
Dim LeffGas As Double ' Gravity settling length for Gas Capacity [ft]
Dim LeffLiq As Double ' Gravity settling length for Liq Capacit [ft]
Dim homax As Double  ' Maxun oil pad thikness
Dim AwA As Double    ' Maxun diameter for oil pad thickness constrain
Dim hoD As Double    ' Dimensionless Oil level [-]
Dim dmax As Double   ' Maxun diameter for oil pad thikness
Dim d2Leff As Double ' Liquid retention constrain
Dim Lss As Double    ' Seam to Seam length [ft]
Dim SlenRatio As Double ' Slenderness ratio [-]
Dim Vec(1 To 6) As Double
Dim k As Double

'***************************************************************************************
' Specific gravity difference
DeltaSG = (Denw - Deno) / 62.4
' Gas constrain Value
k = Sqr((DenG * Cd) / ((Deno - DenG) * dOG))
dLeff = (420 * (T + 460) * Z * (Qg / 1000) * k) / P
' Gravity settling length [ft]
Leff = dLeff / Dvesel
' Maxun oil pad thikness
homax = 0.00128 * tro * DeltaSG * dWO * dWO / Viso
' Maxun diameter for oil pad thickness constrain
AwA = 0.5 * Qw * trw / (tro * Qo + trw * Qw)
' (ho/D = Betta) from Figure 5-8 [Page 149]
hoD = hoDGivenAwA(AwA)
dmax = homax / hoD
' Liquid retention constrain
d2Leff = 1.42 * (Qw * trw + Qo * tro)
' Gravity Settling section constrain
LeffLiq = d2Leff / (Dvesel * Dvesel)
LeffGas = dLeff / (Dvesel)
' Lss for gas Constarin
If LeffGas < LeffLiq Then
' Liquid Constrain Govern
Leff = LeffLiq
Lss = Leff / 0.75
Else
  ' Gas Constrain Govern
  Leff = LeffGas
  Lss = Leff + Dvesel / 12
End If

  ' Slenderness ratio
  SlenRatio = 12 * Lss / Dvesel

  ' Vector Conversion
  Vec(1) = dLeff
  Vec(2) = d2Leff
  Vec(3) = dmax
  Vec(4) = Leff
  Vec(5) = Lss
  Vec(6) = SlenRatio
ThreePhaseH = Vec
End Function

'******************************************************************************************
'                Coefficient Betta for a cylinder half filled with water [Page 149]

Public Function hoDGivenAwA(AwA As Double)
  Dim howD As Double              '    Oil+Water dimensionless liquid level [-]
  Dim hwD As Double               '    Water dimensionless liquid level [-]
  Dim A As Double                 '    Left side interval [-]
  Dim b As Double                 '    Right side interval [-]
  Dim c As Double                 '    Solution [-]
  Dim Fa As Double                '    Function evaluated in left hand side [-]
  Dim Fb As Double                '    Function evaluated in the right hand side [-]
  Dim Fc As Double                '    Function evaluated at c [-]
  Dim Elocal As Double            '    Local Error [-]
  Dim n As Integer                '    iteration number
  Dim Tol As Double               '    Tolerance
  Dim PI As Double

  ' ********************************************************
PI = Application.PI
howD = 0.5
If (AwA <= 0#) Then
    hoDGivenAwA = 0.5
    Exit Function
End If
Tol = 0.00001
A = Tol
b = 0.5 - Tol
c = (A + b) / 2#
Fa = AwA - AwAFun(A)
Fb = AwA - AwAFun(b)
If (Fa * Fb > 0) Then
    hoDGivenAwA = -999#
    Exit Function
End If
Fc = 100
Elocal = 1
n = 0
While ((Abs(b - A) > Tol) And (Abs(Fc) > Tol) And (n < 100))
    c = (A + b) / 2#
    Fc = AwA - AwAFun(c)
    If Fa * Fc < 0 Then
        b = c
        Fb = Fc
    Else
        A = c
        Fa = Fc
    End If
    n = n + 1
Wend
hoDGivenAwA = howD - c
End Function
Private Function AwAFun(hwD)
' hwD: Water dimensionless liquid level [-]

Dim PI As Double
Dim aux As Double
Dim AwD As Double  ' Dimensionless water area [-]

PI = Application.PI()
aux = 2 * hwD - 1
AwD = (PI - Arccos(aux) + aux * Sqr(1 - aux * aux)) / PI
AwAFun = AwD

End Function

Private Function Arccos(ByVal x As Double) As Double
' Inverse Cosine
Arccos = Application.Acos(x)

End Function

Annex B.10: ThreePhaseVertical module

Public Function ThreePhaseV(tro As Double, trw As Double, Deno As Double, Denw As Double, DenG As Double, Viso As Double, Qo As Double, Qw As Double, Qg As Double, T As Double, P As Double, Z As Double, Dvesel As Double, dOG As Double, dWO As Double, Cd As Double)

' **************************************************************************************
' **    Internal Variables:
Dim DeltaSG As Double  ' Difference in specific gravity [-]
Dim dg As Double
Dim dl As Double
Dim dmin As Double
Dim ht As Double  ' ho + hw'
Dim Lss1 As Double
Dim Lss2 As Double
Dim Vec(1 To 4) As Double
Dim k As Double

' **
' **************************************************************************************
' Specific gravity difference
DeltaSG = (Denw - Deno) / 62.4
' Gas constrain Value
    k = Sqr((DenG * Cd) / ((Deno - DenG) * dOG))

dg = ((5040 * (T + 460) * Z * (Qg / 1000) * k) / P) ^ 0.5

' Liq constrain Value

dl = ((6690 * Qo * Viso) / (DeltaSG * dWO * dWO)) ^ 0.5

' Min diameter
    If dg > dl Then
        dmin = dg
    Else
        dmin = dl
    End If

' Liq Retention Constrain (ho + hw)

ht = (tro * Qo + trw * Qw) / (0.12 * Dvesel * Dvesel)

' seam to seam
    Lss1 = (ht + 76) / 12
    Lss2 = (ht + Dvesel + 40) / 12

    If Lss1 > Lss2 Then
        Lss = Lss1
    Else
        Lss = Lss2
    End If

' Slenderness ratio
    SlenRatio = 12 * Lss / Dvesel

' Vector Conversion
    Vec(1) = dmin
    Vec(2) = ht
    Vec(3) = Lss
    Vec(4) = SlenRatio

    ThreePhaseV = Vec

End Function
Annex B.11: Sheet7 (Cost)

Private Sub CommandButton1_Click()
    If Range("FM_1") = "" Then
        MsgBox ("Select Construction Material Separator")
        GoTo ref
    End If

    If Range("FM_2") = "" And Range("TYPESlugC") = "Slug Catcher - Vessel Type" Then
        MsgBox ("Select Construction Material Slug catcher")
        GoTo ref
    End If

    Dim r As Double
    Dim h As Double
    If Sheet3.Range("Slug") = "YES" Then
        ComboBox1.Visible = True
        If Range("TYPESlugC") = "Slug Catcher - Vessel Type" Then
            Range("K_1_2") = 3.5565
            Range("K_2_2") = 0.3776
            Range("K_3_2") = 0.0905
            r = Sheet4.Range("dPV") / (12 * 2)
            h = Sheet4.Range("LPV")
            Range("Volume_2") = (3.141592654 * (r ^ 2) * h) * 0.0283168 [m3]
            Range("B1_2") = 1.49
            Range("B2_2") = 1.52
            Range("P_2") = Sheet5.Range("P_s") / 14.5038 [bar]
            Range("D_2") = Sheet4.Range("dPV") / 12 * 0.3048 [m]
        Dim Fp As Double
        Dim P As Double
        Dim D As Double
        P = Range("P_2")
        D = Range("D_2")
    ```
\[
F_p = \frac{((P + 1) \cdot D)}{(2 \cdot (850 - 0.6 \cdot (P + 1))) + 0.00315} / 0.0063
\]

If \( F_p > 1 \) Then

\[
\text{Range}("F\_P\_2") = F_p
\]

Else

\[
\text{Range}("F\_P\_2") = 1
\]

End If

Dim A As Double

Dim K1 As Double

Dim K2 As Double

Dim K3 As Double

\[
K_1 = \text{Range}("K\_1\_2")
\]

\[
K_2 = \text{Range}("K\_2\_2")
\]

\[
K_3 = \text{Range}("K\_3\_2")
\]

\[
A = \text{Range}("Volume\_2")
\]

\[
\text{Range}("C\_P\_2") = 10^{(K_1 + K_2 \cdot \log_{10}(A) + K_3 \cdot (\log_{10}(A))^2)}
\]

\[
\text{Range}("CBM\_2") = \text{Range}("C\_P\_2") \cdot (\text{Range}("B1\_2") + \text{Range}("B2\_2") \cdot \text{Range}("FM\_2") \cdot \text{Range}("F\_P\_2")
\]

\[
\text{Range}("CBM\_14\_2") = \text{Range}("CBM\_2") \cdot 577 / 397
\]

ElseIf \( \text{Range}("TYPESlugC") = "Slug Catcher - Finger Type" \) Then

\[
\text{Range}("material2") = "Carbon Steel"
\]

\[
\text{Range}("FM\_2") = "."
\]

\[
\text{Range}("K\_1\_2") = "."
\]

\[
\text{Range}("K\_2\_2") = "."
\]

\[
\text{Range}("K\_3\_2") = "."
\]

\[
\text{Range}("Volume\_2") = "."
\]

\[
\text{Range}("B1\_2") = "."
\]

\[
\text{Range}("B2\_2") = "."
\]

\[
\text{Range}("P\_2") = "."
\]

\[
\text{Range}("D\_2") = "."
\]

\[
\text{Range}("F\_P\_2") = "."
\]

\[
\text{Range}("C\_P\_2") = "."
\]

\[
\text{Range}("CBM\_2") = "."
\]

\[
\text{Range}("CBM\_14\_2") = \text{Sheet2.Range}("weight") \cdot \text{Sheet4.Range}("LFT") \cdot \text{Sheet4.Range}("Nbottle") \cdot 0.0005 \cdot 2194 \cdot 1.75
\]

End If
If Range("TYPESlugC") = "" Then
    MsgBox ("Select Slug Catcher Type")
End If
Else
    Range("SlugCatcherCost").ClearContents
    Range("TYPESlugC").ClearContents
End If
If Sheet5.Range("TYPE_s") = "Three-Phase Vertical" Then
    Range("Equipment1") = "Three-Phase Vertical Separator"
    Range("K_1") = 3.4974
    Range("K_2") = 0.4485
    Range("K_3") = 0.1074
    r = Sheet6.Range("D_Draw_V") / (12 * 2)
    h = Sheet6.Range("L_Draw_V")
    Range("Volume1") = (3.141592654 * (r ^ 2) * h) * 0.0283168 \([\text{m}^3]\)
    Range("B_1") = 2.25
    Range("B_2") = 1.82
ElseIf Sheet5.Range("TYPE_s") = "Two-Phase Vertical" Then
    Range("Equipment1") = "Two-Phase Vertical Separator"
    Range("K_1") = 3.4974
    Range("K_2") = 0.4485
    Range("K_3") = 0.1074
    r = Sheet6.Range("D_Draw_V") / (12 * 2)
    h = Sheet6.Range("L_Draw_V")
    Range("B_1") = 2.25
    Range("B_2") = 1.82
ElseIf Sheet5.Range("TYPE_s") = "Three-Phase Horizontal" Then
    Range("Equipment1") = "Three-Phase Horizontal Separator"
    Range("K_1") = 3.5565
    Range("K_2") = 0.3776
    Range("K_3") = 0.0905
    r = Sheet6.Range("D_Draw_H") / (12 * 2)
    h = Sheet6.Range("L_Draw_H")
    Range("Volume1") = (3.141592654 * (r ^ 2) * h) * 0.0283168 \([\text{m}^3]\)
ElseIf Sheet5.Range("TYPE_s") = "Two-Phase Horizontal" Then
    Range("Equipment1") = "Two-Phase Horizontal Separator"
    Range("K_1") = 3.5565
    Range("K_2") = 0.3776
    Range("K_3") = 0.0905
    r = Sheet6.Range("D_Draw_H") / (12 * 2)
    h = Sheet6.Range("L_Draw_H")
    Range("Volume1") = (3.141592654 * (r ^ 2) * h) * 0.0283168 ' [m3]
    Range("B_1") = 1.49
    Range("B_2") = 1.52
End If

Range("P_1") = Sheet5.Range("P_s") / 14.5038 '[bar]
Range("D_m") = Sheet6.Cells(1, 20) / 12 * 0.3048 ' [m]
P = Range("P_1")
D = Range("D_m")
Fp = (((P + 1) * D) / (2 * (850 - 0.6 * (P + 1))) + 0.00315) / 0.0063
If Fp > 1 Then
    Range("F_P") = Fp
Else
    Range("F_P") = 1
End If
K1 = Range("K_1")
K2 = Range("K_2")
K3 = Range("K_3")
A = Range("Volume1")
Range("C_P") = 10 ^ (K1 + K2 * Log10(A) + K3 * (Log10(A)) ^ 2)
Range("CBM") = Range("C_P") * (Range("B_1") + Range("B_2") * Range("FM_1") + Range("F_P"))
Range("CBM_14") = Range("CBM") * 577 / 397
Range("Total") = 1.18 * (Range("CBM_14") + Range("CBM_14_2"))
Range("Grass") = 1.5 * Range("Total")
ref: need to select material'
End Sub
DOCUMENT III: ANNEXES

ANNEX C : CONVERSIONS FROM SI UNITS TO OILFIELD UNITS
Annex C: Conversions from SI units to Oilfield units

Conversions from SI units to Oilfield units, are shown in Table 7

Table 7: Conversions from SI units to Oilfield units

<table>
<thead>
<tr>
<th>SI</th>
<th>Oilfield</th>
</tr>
</thead>
<tbody>
<tr>
<td>Mass 1 kg</td>
<td>2.2046225 lbm</td>
</tr>
<tr>
<td>Length 0.3048 m</td>
<td>1 ft</td>
</tr>
<tr>
<td>Area 4.046.873 m²</td>
<td>1 acre = 43,560 ft²</td>
</tr>
<tr>
<td>Volume 1 m³</td>
<td>6.2898106 bbl</td>
</tr>
<tr>
<td></td>
<td>1 bbl = 5.614583 ft³</td>
</tr>
<tr>
<td>Temperature 1 K</td>
<td>1.8 °F</td>
</tr>
<tr>
<td></td>
<td>°C = (°F - 32)/1.8</td>
</tr>
<tr>
<td></td>
<td>°F = 1.8 °C + 32</td>
</tr>
<tr>
<td>Pressure 6.894757 kPa</td>
<td>1 psi</td>
</tr>
<tr>
<td></td>
<td>1 MPa = 145.03774 psi</td>
</tr>
<tr>
<td></td>
<td>101.325 kPa = 1 atm = 14.69595 psi</td>
</tr>
<tr>
<td></td>
<td>1 bar = 100 kPa = 14.503774 psi</td>
</tr>
<tr>
<td>Dynamic viscosity</td>
<td>1 mPa·s = 1 cp</td>
</tr>
<tr>
<td>Density 1000 kg/m³</td>
<td>62.42797 lbm/ft³</td>
</tr>
<tr>
<td></td>
<td>8.345405 lbm/gal</td>
</tr>
<tr>
<td>Water density @ 60°F/1 atm 999.04 kg/m³</td>
<td>62.368 lbm/ft³</td>
</tr>
<tr>
<td>Energy 1.055056 kJ</td>
<td>1 btu</td>
</tr>
<tr>
<td></td>
<td>1 kWh = 3412.14 btu</td>
</tr>
<tr>
<td></td>
<td>1 btu = 778.169 ft·lbf</td>
</tr>
<tr>
<td>Power 745.700 W</td>
<td>1 hp = 550 ft·lbf/s</td>
</tr>
<tr>
<td>Molecular weight of air 28.9625 kg/kmol</td>
<td>28.9625 lbm/lbmol</td>
</tr>
<tr>
<td>Permeability 1 mm²</td>
<td>1013.25 md</td>
</tr>
</tbody>
</table>

The following link provides a useful unit conversion tool powered by OilZone®.

http://www.oilzonetools.com/oil_gas_units.html