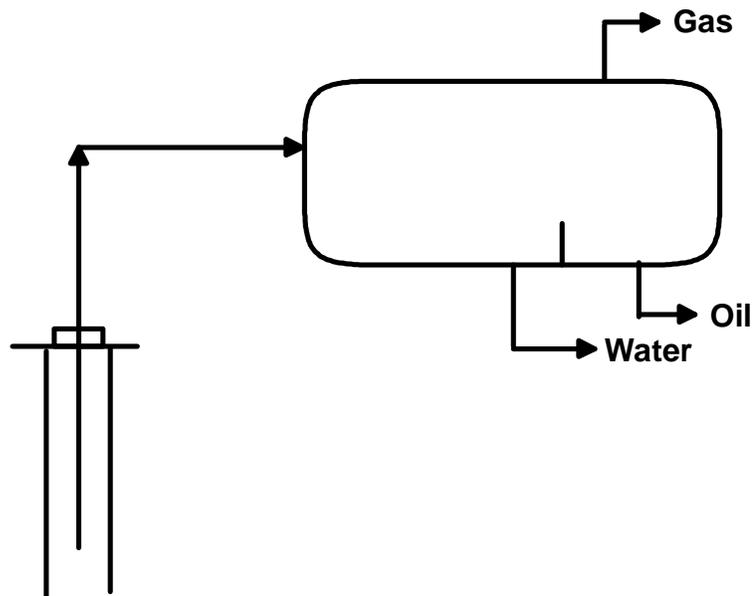


# PROGRAM GOW

(Gas - Oil – Water Calculations)



Version 6.4

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## **DISCLAIMER - AGREEMENT**

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## 1.0 INTRODUCTION

**GOWProp** calculates physical properties of the following gases and liquids:

| Gas Types  | Physical Property  |
|--|--|
| <ul style="list-style-type: none"> <li>• Natural gas</li> <li>• Nitrogen</li> <li>• Air</li> </ul> | <ul style="list-style-type: none"> <li>• Molecular weight</li> <li>• Density</li> <li>• Compressibility</li> <li>• Gas formation volume factor</li> <li>• Z-factor (gas deviation factor)</li> <li>• Viscosity</li> <li>• Thermal conductivity</li> <li>• Specific heat</li> <li>• Ideal isentropic coefficient, Cp/Cv</li> <li>• Real isentropic coefficient, k</li> <li>• Pseudo Critical properties</li> <li>• Pseudo Reduced properties</li> </ul> |

| Liquid Types  | Physical Property  |
|---|--|
| <ul style="list-style-type: none"> <li>• Oil</li> <li>• Water</li> <li>• Methanol/Water mixtures</li> <li>• Monoethylene Glycol/Water</li> <li>• Diethylene Glycol/Water</li> <li>• Triethylene Glycol/Water</li> </ul> | <ul style="list-style-type: none"> <li>• API gravity (for oil only)</li> <li>• Density</li> <li>• Compressibility</li> <li>• Formation volume factor (oil and water)</li> <li>• Solution gas-liquid ratio (oil and water)</li> <li>• Bubble point pressure (oil only)</li> <li>• Viscosity</li> <li>• Thermal conductivity</li> <li>• Surface tension</li> <li>• Specific heat</li> <li>• Pseudo Critical properties</li> <li>• Pseudo Reduced properties</li> </ul> |

Program **AGA** calculates size, flowrate or pressure drops for gas and liquid flow orifice meters based on ANSI/API-2530 - 1992 (AGA Report No. 3).

Program **ISO** calculates size, flowrate or pressure drops for gas and liquid flow orifice meters based on International Standard ISO-5167-2: 2003.

For both **AGA** and **ISO** the orifice calculations are based on sharp-edged orifices:

- Normal gas orifices
- Normal liquid orifices
- Gas restriction orifices (critical gas flow orifices)
- Liquid restriction orifices (critical liquid flow orifices)

Program **GOsep** performs flash calculations for gas/oil separators to optimize liquid recovery depending on number of separation stages and pressure/temperature at each separator stage. Maximum number of stages is set to four.

**WatGas** performs calculations for both gases with known compositions or non-compositional gases (only specific gravity known):

- Hydrate Formation Conditions
- Water Content of Natural Gases
- MeOH/Glycol Injection Requirements
- Solid CO2 Formation Conditions

## 2.0 PHYSICAL PROPERTIES

Note that standard conditions for **GOWProp** are defined as 60 °F (15.6 °C) and 14.696 psia (1.01325 Bara).

All calculations are based on customary units unless noted (Temperature = °F, Pressure = psia, Density = lb/ft<sup>3</sup>, etc.). Generally, pseudo-critical properties and pseudo-reduced properties are used throughout the program.

### 2.1 Gas Properties

For natural gas the choice of method for calculating Z-factor (compressibility factor) also determines method for calculating isothermal compressibility, formation volume factor, specific heat capacity and isentropic coefficient (ideal and real). For nitrogen and air the Redlich-Kwong equation of state is used.

#### 2.1.1 Molecular Weight

Molecular weight is only calculated for natural gas as nitrogen and air have known molecular weights (lb/lbmole):

$$M_{wtg} = 28.96247\gamma_g \quad (2.1.1)$$

where 28.96247 is molecular weight of air. Molecular weight of nitrogen here is 28.0134<sup>2</sup>.

#### 2.1.2 Density

Density is the weight of a volume of gas divided by the volume it occupies, and can be calculated as (lb/ft<sup>3</sup>):

$$\rho_g = \frac{M_{wtg} P}{RTZ} \quad (2.1.2)$$

#### 2.1.3 Compressibility

Isothermal compressibility of a gas is defined as the change in volume per unit volume per change in pressure:

$$c_g = -\frac{1}{V_M} \left( \frac{\partial V}{\partial P} \right)_T \quad (2.1.3)$$

For natural gases the gas compressibility is calculated by the AGA 8 equation of state<sup>3</sup>:

$$Z = \left( \frac{P}{RT\rho_m} \right) = 1 + B\rho_m + C\rho_m^2 + D\rho_m^3 + E\rho_m^5 + G\rho_m^2(1 + H\rho_m^2)e^{(-H\rho_m^2)} \quad (2.1.4)$$

It can be shown that:

$$c_g = \frac{1}{P \left[ 1 - \left( \frac{\partial Z}{\partial P} \right)_T \frac{P}{Z} \right]}$$

which can be combined with equation (2.1.4) to give:

$$c_g = \frac{1}{P \left[ 1 - \left( \frac{\partial Z}{\partial \rho_m} \right)_T \frac{\rho_m}{Z} \right]} \quad (2.1.5)$$

The original Redlich-Kwong equation of state<sup>4,5</sup> is used for calculating compressibility for air and nitrogen. This equation only depends on critical pressure and temperature for calculations, and therefore does not include knowledge of individual components in the gas mixture:

$$P = \frac{RT}{V_m - b} - \frac{a}{V_m^2 + V_m b} \quad (2.1.6)$$

where

$$a = \frac{0.42748R^2T_{pc}^{2.5}}{P_{pc}T^{0.5}} \quad b = \frac{0.08664RT_{pc}}{P_{pc}}$$

The compressibility equation:

$$c_g = \frac{1}{\frac{RTV_m}{(V_m - b)^2} - \frac{2aV_m + ab}{V_m(V_m + b)^2}} \quad (2.1.7)$$

**GOWProp** also has the option of calculating compressibility (and thereby Z-factor) using the Dranchuk and Abou-Kassem equation of state<sup>6</sup>:

$$Z = \left( \frac{0.27P_{pr}}{\rho_{pr}T_{pr}} \right) = 1 + A_1\rho_{pr} + A_2\rho_{pr}^2 - A_3\rho_{pr}^5 + A_4\rho_{pr}^2(1 + A_5\rho_{pr}^2)e^{(-A_5\rho_{pr}^2)} \quad (2.1.8)$$

where  $A_1, A_2, A_3, A_4$  and  $A_5$  are functions of temperature only.

For a real gas it can be show that

$$c_g = \frac{1}{P \left[ 1 + \left( \frac{\partial Z}{\partial \rho_{pr}} \right)_{T_{pr}} \frac{\rho_{pr}}{Z} \right]} \quad (2.1.9)$$

#### 2.1.4 Formation Volume Factor

Formation volume factor is the volume of gas in the reservoir needed to produce one (1) standard volume at the surface:

$$B_g = \frac{ZTP_{sc}}{T_{sc}P} \quad (2.1.10)$$

Formation volume factor is not a relevant physical property for nitrogen and air, and therefore not calculated for these gases.

### 2.1.5 Compressibility Factor

The compressibility factor or the gas deviation factor (Z-factor) is a measure of how far the gas is away from ideal conditions ( $Z = 1$  @  $P = 0$ ).

For natural gas the AGA 8 equation of state<sup>3</sup> given as equation (2.1.4) above is used.

For air and nitrogen the Redlich-Kwong equation of state<sup>4,5</sup> is used. Equation (2.1.6) can also be written as:

$$Z^3 - Z^2 + Z(A-B-B^2) - AB = 0 \quad (2.1.11)$$

where

$$A = \frac{0.42748P_{pr}}{T_{pr}^{2.5}} \quad B = \frac{0.08664P_{pr}}{T_{pr}}$$

### 2.1.6 Viscosity

Viscosity is a measure of resistance to flow exerted by a fluid. For natural gas the Lee et al. correlation<sup>4</sup> is used:

$$\mu_g = Ae^{(B\rho_g^c)} \quad (2.1.12)$$

where

$$A = \frac{(9.4 + 0.02M_{wtg})x10^{-4}T^{1.5}}{209 + 19M_{wtg} + T} \quad B = 3.5 + \frac{986}{T} + 0.01M_{wtg} \quad C = 2.4 - 0.2B$$

It should be noted that the above equations require temperatures in °R and density in g/cm<sup>3</sup>.

**GOWProp** has another option for natural gas viscosity in the Carr et al. Correlation<sup>4</sup>. The low pressure viscosity is given as:

$$\mu_{ag} = \mu_{1g} + N_{2corr} + CO_{2corr} + H_2S_{corr}$$

where

$$\mu_{1g} = [1.709x10^{-5} - 2.062x10^{-6}\gamma_g]T + 8.188x10^{-3} - 6.15x10^{-3} \log \gamma_g$$

$$N_{2corr} = y_{N_2} (8.48x10^{-3} \log \gamma_g + 9.59x10^{-3})$$

$$CO_{2corr} = y_{CO_2} (8.49x10^{-3} \log \gamma_g + 6.24x10^{-3})$$

$$H_2S_{corr} = y_{H_2S} (8.49x10^{-3} \log \gamma_g + 3.73x10^{-3})$$

and  $y_{N_2}, y_{CO_2}, y_{H_2S}$  = molefractions of  $N_2$ ,  $CO_2$  and  $H_2S$  in the vapor-phase respectively

To obtain viscosity at higher pressures the Dempsey correlation is used<sup>7</sup>:

$$\ln\left(\frac{\mu_g}{\mu_{ag}} T_{pr}\right) = a_0 + a_1 P_{pr} + a_2 P_{pr}^2 + a_3 P_{pr}^3 + T_{pr}(a_4 + a_5 P_{pr} + a_6 P_{pr}^2 + a_7 P_{pr}^3) + T_{pr}^2(a_8 + a_9 P_{pr} + a_{10} P_{pr}^2 + a_{11} P_{pr}^3) + T_{pr}^3(a_{12} + a_{13} P_{pr} + a_{14} P_{pr}^2 + a_{15} P_{pr}^3) \quad (2.1.13)$$

where  $a_0 - a_{15}$  are constants

For nitrogen and air, the Thodos et al. relations for viscosity of pure gases are used:<sup>8</sup>

At low pressures:

$$\mu_{ag} = \frac{4.610 T_{pr}^{0.618} - 2.04 e^{-0.449 T_{pr}} + 1.94 e^{-4.058 T_{pr}} + 0.1}{\varepsilon}$$

where

$$\varepsilon = \frac{T_{pc}^{1/6}}{M_{wtg}^{1/2} P_{pc}^{2/3}}$$

and at given pressure and temperature:

$$\left[ (\mu_g - \mu_{ag}) \varepsilon + 1 \right]^{0.25} = 1.023 + 0.2336 \rho_{pr} + 0.5853 \rho_{pr}^2 - 0.4076 \rho_{pr}^3 + 0.0933 \rho_{pr}^4 \quad (2.1.14)$$

### 2.1.7 Thermal Conductivity

Thermal conductivity for gases at low pressures can be approximated by:

$$k_{ha} = 0.1 M_{wtg}^{(-0.637 + 6.46 \times 10^{-4} T)}$$

This equation is based on the thermal conductivity chart for natural gases at low pressures given in GPSA Data Book<sup>9</sup>. Thermal conductivity at elevated pressures is based on data from Thodos et al.<sup>8</sup>:

$$\rho_{pr} < 2.4383: \quad (k_h - k_{ha}) Z_c^5 \lambda = 2.662 \times 10^{-6} \rho_{pr} e^{0.528 \rho_{pr}} \quad (2.1.15)$$

$$\rho_{pr} \geq 2.4383: \quad (k_h - k_{ha}) Z_c^5 \lambda = 1.54 \times 10^{-6} \rho_{pr} e^{1.118 \rho_{pr}} \quad (2.1.16)$$

where

$$\lambda = \frac{M_{wtg}^{0.5} T_{pc}^{1/6}}{P_{pc}^{2/3}}$$

### 2.1.8 Specific Heat capacity

Specific heat capacity for natural gas at low pressures is based on the charts given by Buthod<sup>10</sup> and can be expressed as:

$$C_{pga} = 0.62 + 2.2 \times 10^{-4} \times T + 4.14 \times 10^{-7} \times T^2 + \gamma_g (-0.213 + 1.986 \times 10^{-4} \times T - 2.648 \times 10^{-7} \times T^2)$$

For air the low pressure specific heat capacity is based on curve-fitting the values<sup>2</sup>:

$$C_{Pga} = 6.94 + 1.286 \times 10^{-4} T + 5.728 \times 10^{-7} T^2$$

Similarly, the low pressure specific heat capacity of nitrogen can be calculated as:

$$C_{Pga} = 6.95 + 1.396 \cdot 10^{-4} T + 5.768 \cdot 10^{-7} T^2$$

To calculate the heat capacity at higher pressures the Z-factor equations given in section 2.1.5 are used. The definitions of heat capacity at constant pressure and volume are given by:

$$C_{Pg} = \left( \frac{\partial H}{\partial T} \right)_P$$

$$C_{Vg} = \left( \frac{\partial U}{\partial T} \right)_V$$

Based on these definitions the following equations are derived and used in this program for calculating heat-capacities of real gases:

For the AGA-8 and Redlich-Kwong equations of state (used in **AGA** and **ISO**):

$$C_{Pg} = C_{Vg} - T_{pr} \frac{\left( \frac{\partial P}{\partial T} \right)_V^2}{\left( \frac{\partial P_{pr}}{\partial V_{pr}} \right)_T} \quad (2.1.17)$$

For Dranchuk/Abou-Kassem equation:

$$C_{Pg} = C_{Vg} - RT_{pr} \frac{\left( \frac{\partial P_{pr}}{\partial T_{pr}} \right)_{V_{pr}}^2}{\left( \frac{\partial P_{pr}}{\partial V_{pr}} \right)_{T_{pr}}} \quad (2.1.18)$$

### 2.1.9 Isentropic Coefficient

Both the ideal and the real isentropic coefficients are calculated. The ideal isentropic coefficient:

$$\kappa_i = \frac{C_{Pg}^o}{C_{Vg}^o}$$

The real gas isentropic coefficient used for AGA-8 and Redlich-Kwong equations (used in **AGA** and **ISO**):

$$\kappa_r = \frac{C_{Pg}}{C_{Vg}} \left( \frac{\partial P}{\partial V} \right)_T \frac{V_m}{P} \quad (2.1.19)$$

The isentropic coefficient for the Dranchuk/Abou-Kassem equation of state can be expressed as:

$$\kappa_r = - \left( \frac{\partial P_{pr}}{\partial V_{pr}} \right)_{T_{pr}} - \frac{RT_{pr}}{C_{Vg}} \left( \frac{\partial P_{pr}}{\partial T_{pr}} \right)_{V_{pr}}^2 \frac{1}{P_{pr} \rho_{pr}} \quad (2.1.20)$$

### 2.1.10 Critical Properties

Critical properties or rather pseudo-critical properties for natural gas (pressure and temperature) are given by Standing<sup>11</sup>. The gravity of the hydrocarbon portion of the gas is given by:

$$\gamma_{g1} = \frac{\gamma_g - 0.967y_{N_2} - 1.52y_{CO_2} - 1.18y_{H_2S}}{1 - y_{N_2} - y_{CO_2} - y_{H_2S}}$$

Surface gas:

$$P_{pc} = 677 + 15.0\gamma_{g1} - 37.5\gamma_{g1}^2$$

$$T_{pc} = 168 + 325\gamma_{g1} - 12.5\gamma_{g1}^2$$

Reservoir gas (gas in equilibrium with crude oil):

$$P_{pc} = 706 + 51.7\gamma_{g1} - 11.1\gamma_{g1}^2$$

$$T_{pc} = 187 + 330\gamma_{g1} - 71.5\gamma_{g1}^2$$

If the gas contains N<sub>2</sub>, CO<sub>2</sub>, or H<sub>2</sub>S the pseudo-criticals of the gas mixture can be calculated as:

$$P_{pc} = (1 - y_{N_2} - y_{CO_2} - y_{H_2S}) \times P_{pc} + 493y_{N_2} + 1071y_{CO_2} + 1306y_{H_2S}$$

$$T_{pc} = (1 - y_{N_2} - y_{CO_2} - y_{H_2S}) \times T_{pc} + 227y_{N_2} + 548y_{CO_2} + 677y_{H_2S}$$

These mixture pseudo-criticals are then adjusted for CO<sub>2</sub> and H<sub>2</sub>S content by the Wichert-Aziz correlations<sup>12</sup>:

$$\varepsilon = 120[(y_{CO_2} + y_{H_2S})^{0.9} - (y_{CO_2} + y_{H_2S})^{1.6}] + 15(y_{H_2S}^{0.5} - y_{H_2S}^4)$$

$$T_{pc} = T_{pc} - \varepsilon \quad (2.1.21)$$

$$P_{pc} = P_{pc} \frac{T_{pc} - \varepsilon}{T_{pc} + y_{H_2S}(1 - y_{H_2S})\varepsilon} \quad (2.1.22)$$

For air the critical properties are<sup>2</sup>:

$$T_{pc} = 238.9 \text{ (}^\circ\text{R)} \quad (2.1.23)$$

$$P_{pc} = 546.7 \text{ (Psia)} \quad (2.1.24)$$

### 2.1.11 Reduced Properties

The calculated reduced properties are actually the pseudo-reduced properties and are given by:

$$P_{pr} = P/P_{pc} \quad (2.1.25)$$

$$T_{pr} = (T + 459.67)/T_{pc} \quad (2.1.26)$$

## 2.2 Liquid Properties

**AGA** and **GOWProp** standard conditions for liquids are defined in API 2530<sup>1</sup> :

- Base Temperature ( $T_{Base}$ ) = 60 °F
- Base Pressure ( $P_{Base}$ ) = 14.696 Psia (101.325 kPa)

For **ISO** base conditions (Standard Conditions) for liquids are defined in ISO-5024<sup>2</sup> :

- Base Temperature ( $T_{Base}$ ) = 59 °F (15 °C)
- Base Pressure ( $P_{Base}$ ) = 14.69595 Psia (101.325 kPa)

Correlations involving Vasquez and Beggs<sup>13</sup> methods of calculating physical properties require use of a normalized separator gas gravity at 100 psia. However, the correlations included here do not take this normalized gas gravity into account. Instead, the gas gravity related to air = 1.0 is used. To calculate the normalized gas gravity and use this as gas gravity for input to the program:

$$\gamma_{gNorm} = \gamma_{gSep} \left[ 1 + 5.912 \times 10^{-5} API x T_{Sep} \log \frac{P_{Sep}}{14.696} \right]$$

where

- $\gamma_{gNorm}$  = normalized separator gas gravity (air=1.0)
- $\gamma_{gSep}$  = separator gas gravity (air=1.0)
- $T_{Sep}$  = separator temperature, °F
- $P_{Sep}$  = separator pressure, psia

### 2.2.1 API Gravity

The API gravity is defined as (Degrees API):

$$API = \frac{141.5}{\gamma_o} - 131.5 \quad (2.2.1)$$

### 2.2.2 Density

Oil density is here normally calculated by the McCain formation volume factor method<sup>14</sup> :

$$\rho_o = \frac{1}{B_o} (62.3663 \gamma_o + 0.0136 R_{so} \gamma_o) \quad (2.2.2)$$

where 62.3663 = density of water in lb/ft<sup>3</sup> at 14.696 psia, 60 °F<sup>2</sup>.

**GOWProp** also has the following options for calculating oil density:

Apparent liquid density method:

This method is based on adjusting the gas free crude by the apparent liquid density of solution gas. Standing<sup>11</sup> gives the apparent liquid density of a natural gas when dissolved in oil as:

$$\rho_{lapp} = 38.52 \times 10^{-0.0033 API} + [94.75 - 33.93 \log API] \log \gamma_g$$

The oil density at 14.696 psia and 60 °F is then:

$$\rho_{oa} = \frac{62.3663\gamma_o \times 5.614584 + R_{so} \times 0.0765\gamma_g}{5.614584 + \frac{R_{so} \times 0.0765\gamma_g}{\rho_{lapp}}}$$

where 0.0765002 = density of air in lb/ft<sup>3</sup> at 14.696 psia, 60 °F<sup>2</sup>.  
5.614584 = conversion factor: 1 STB = 5.614584 ft<sup>3</sup>

The numerator gives the total mass of one barrel of oil (oil+gas) at saturated conditions, lb<sub>m</sub>. The denominator gives the volume of the same barrel of oil (oil+gas) at standard conditions, ft<sup>3</sup>. The density at standard conditions must be adjusted for correct pressure and temperature.

The temperature correction is<sup>11</sup> :

$$\Delta\rho_T = \frac{P}{1000} \left( 0.167 + 16.181 \times 10^{-0.0425\rho_{oa}} \right) - 0.01 \left( \frac{P}{1000} \right)^2 \left( 0.299 + 263 \times 10^{-0.0603\rho_{oa}} \right)$$

The pressure correction is:

$$\Delta\rho_P = (0.0133 + 152.4(\rho_{oa} + \Delta\rho_T)^{-2.45})(T - 60) - (8.1 \times 10^{-6} - 0.0622 \times 10^{-0.0764(\rho_{oa} - \Delta\rho_T)})(T - 60)^2$$

and the density at actual temperature and pressure is:

$$\rho_o = \rho_{oa} + \Delta\rho_T - \Delta\rho_P \quad (2.2.3)$$

Katz formation volume factor method:

This method requires knowledge of the dissolved gas gravity, and can be approximated by (air =1.0)<sup>15</sup> :

$$\rho_{gDis} = 0.25 + 0.02API - R_{so}(-0.0369 + 0.36API) \times 10^{-5}$$

Density of saturated crudes can be calculated as<sup>16</sup> :

$$\rho_o = \frac{1}{B_o} \left( 62.3663\gamma_o + \frac{0.0765\rho_{gDis}R_{so}}{5.614584} \right) \quad (2.2.4)$$

For pressures higher than the bubble point pressure, the density is given by<sup>14</sup> :

$$\rho_o = \rho_{ob} e^{[c_o(P-P_b)]} \quad (2.2.5)$$

Water density in **GOWProp** is calculated as:

$$\rho_w = \frac{\rho_{wa}}{B_w} \quad (2.2.6)$$

and the density at atmospheric conditions is adjusted for salt-content according to <sup>14</sup> :

$$\rho_{wa} = 62.3663 + 0.4386 \times C + 1.6 \times 10^{-3} \times C^2$$

where C = salt concentration (NaCl) in parts per million (PPM) by weight. PPM = % x 10000.

Water density at atmospheric conditions in **AGA** and **ISO** is calculated based on an equation given by Dickey <sup>17</sup> :

$$\rho_{wa} = 62.7519 - 3.5374 \times 10^{-3} \times T - 4.8193 \times 10^{-5} \times T^2 \quad (2.2.7)$$

Density of pure methanol and glycols are based on regression of curves given by Gallant <sup>18</sup> with temperatures in °C:

$$\rho_{MeOH} = 1000 \left( \frac{0.8089 - 0.00246T}{1 - 0.0021T} \right) \quad (2.2.8)$$

$$\rho_{MEG} = 1000 \left( \frac{1.1253 - 0.00155T}{1 - 8.07 \times 10^{-4}T} \right) \quad (2.2.9)$$

$$\rho_{DEG} = 1000 \left( \frac{1.1355 - 0.00114T}{1 - 3.615 \times 10^{-4}T} \right) \quad (2.2.10)$$

$$\rho_{TEG} = 1000 \left( \frac{1.1441 - 5.869 \times 10^{-4}T}{1 - 2.018 \times 10^{-4}T} \right) \quad (2.2.11)$$

For different water contents these densities are adjusted by interpolating with density of water given as <sup>17</sup>:

$$\rho_w = 62.3663 - 0.004(T - 32) - 5.619 \times 10^{-5} (T - 32)^2 \quad (2.2.12)$$

### 2.2.3 Compressibility

Oil compressibility equal to or above the bubble point is given by Vasquez and Beggs <sup>13</sup> :

$$c_o = \frac{-1433 + 5R_{sob} + 17.2T - 1180\gamma_g + 12.6xAPI}{Px10^5} \quad (2.2.13)$$

At pressures below the bubble point oil compressibility is defined as:

$$c_o = -\frac{1}{B_o} \left( \frac{\partial B_o}{\partial P} \right)_T + \frac{B_g}{B_o} \left( \frac{\partial R_{so}}{\partial P} \right)_T$$

In this program, however, it is chosen to use the equation given by McCain, Rollins and Villena Lanzi <sup>19</sup> :

$$c_o = 7.573 - 1.45 \ln P - 0.383 \ln P_b + 1.402 \ln(T + 460) + 0.256 \ln API + 0.449 \ln R_{sob} \quad (2.2.14)$$

Water compressibility is a source of drive energy for reservoirs producing above the bubble point. An equation has been developed based on the figures given by Dodson and Standing<sup>20</sup>.

This has a wider temperature range than other published correlations<sup>14, 21</sup> :

$$C_{w1} = 10^{-6} \times (A + BT + CT^2)$$

where

$$A = 3.861 - 8.628 \cdot 10^{-5} P - 7.33 \cdot 10^{-9} P^2$$

$$B = -10.643 \cdot 10^{-3} - 2.154 \cdot 10^{-7} P + 1.054 \cdot 10^{-10} P^2$$

$$C = 4.029 \cdot 10^{-5} + 1.187 \cdot 10^{-9} P - 3.259 \cdot 10^{-13} P^2$$

The correction for dissolved gas-content can be calculated as:

$$\text{Corr}_g = 1 + 8.9 \cdot 10^{-3} R_{\text{swBrine}}$$

The salinity correction factor used is based on a figure given by Eichelberger<sup>22</sup> :

$$\frac{R_{\text{swBrine}}}{R_{\text{sw}}} = e^{-0.2022 \frac{\text{NaCl}}{10000} T^{-0.2945}}$$

Combining the equations above gives the compressibility of water above the bubble-point as:

$$C_w = C_{w1} \times \text{Corr}_g \quad (2.2.15)$$

The compressibility below the bubble-point is defined as:

$$C_w = -\frac{1}{B_w} \left( \frac{\partial B_w}{\partial P} \right)_T + \frac{B_g}{B_w} \left( \frac{\partial R_{\text{sw}}}{\partial P} \right)_T$$

The first term of this equation is given as equation 2.2.15. McCain<sup>14</sup> suggests calculating the partial derivative of solution-gas-water-ratio with respect to pressure at constant temperature with one of the equations given in section 2.2.4 and estimating the formation volume factor of gas with a gas gravity of 0.63. However, in this program the compressibility both above and below the bubble point for water is calculated by equation (2.2.15).

For the other liquids a general liquid compressibility factor is calculated<sup>2</sup> :

$$Z_l = 0.269T_{\text{pr}} - 0.5163T_{\text{pr}}^2 + 0.3521T_{\text{pr}}^3 - 0.0461 \quad (2.2.16)$$

## 2.2.4 Formation Volume Factor

Formation volume factor for saturated oil is the reservoir volume in barrels (measured at reservoir conditions) that is occupied by 1 STB of oil and its dissolved gas. Three different methods are available here:

1) Vasquez and Beggs method<sup>13</sup> :

$$B_o = 1 + AxR_{so} + B(T - 60) \frac{API}{\gamma_o} + CxR_{so}(T - 60) \frac{API}{\gamma_o} \quad (2.2.17)$$

For API <=30 deg : A =  $4.677 \cdot 10^{-4}$   
 B =  $1.751 \cdot 10^{-5}$   
 C =  $-1.811 \cdot 10^{-8}$

For API > 30 deg:  $A = 4.670 \cdot 10^{-4}$   
 $B = 1.100 \cdot 10^{-5}$   
 $C = 1.337 \cdot 10^{-9}$

2) Standing's method <sup>11</sup> :

$$B_o = 0.9759 + 12 \cdot 10^{-5} C_n^{1.2} \quad (2.2.18)$$

where

$$C_n = R_{so} \left( \frac{\gamma_g}{\gamma_o} \right)^{0.5} + 1.25T$$

3) Glaso's method is recommended for North Sea Oils <sup>23</sup> :

$$\log(B_o - 1) = -6.585 + 2.913 \log B_{ob}^* - 0.2768 (\log B_{ob}^*)^2 \quad (2.2.19)$$

and  $B_{ob}^* = R_{so} \left( \frac{\gamma_g}{\gamma_o} \right)^{0.526} + 0.968T$

For an undersaturated liquid the formation volume factor is calculated by:

$$B_o = B_{ob} e^{c_o(P_b - P)} \quad (2.2.20)$$

Two methods are available for water:

1) McCain's method <sup>14</sup> :

$$B_w = (1 + \Delta w_T)(1 + \Delta w_P) \quad (2.2.21)$$

where

$\Delta w_T$  is change in volume due to reduction in temperature to 60 °F, and

$\Delta w_P$  is change in volume due to pressure reduction to atmospheric pressure (14.696 psia):

$$\Delta w_T = -1.0001 \times 10^{-2} + 1.334 \times 10^{-4} T + 5.506 \times 10^{-7} T^2$$

$$\Delta w_P = -1.953 \times 10^{-3} PT - 1.728 \times 10^{-13} P^2 T + 3.589 \times 10^{-7} P - 2.253 \times 10^{-10} P^2$$

2) Keenan and Keyes <sup>24</sup> :

The formation volume factor of water is:

$$B_w = B_{w1} + R_{swBrine}/R_{sw} \quad (2.2.22)$$

where  $B_{w1}$  is formation volume factor of pure water

Based on Keenan and Keyes figures for the formation volume factor of pure water the following equation is developed:

$$B_{w1} = A + B \times P$$

where

$$A = 0.996 + 6.854 \times 10^{-6}T + 1.009 \times 10^{-6}T^2$$

$$B = 3.275 \times 10^{-6} - 6.567 \times 10^{-9}T + 3.259 \times 10^{-11} T^2$$

### 2.2.5 Solution Gas-Liquid Ratio

The solution gas-oil ratio is the volume of gas (standard volume) dissolved in one stock tank barrel of oil at a given reservoir pressure and temperature. Four methods are available:

1) Vasquez and Beggs method<sup>13</sup> :

$$R_{so} = A \gamma_g P^B e^{C \left[ \frac{API}{T+460} \right]} \quad (2.2.23)$$

For API <=30 deg : A = 0.0362  
 B = 1.0937  
 C = 25.724

For API > 30 deg : A = 0.0178  
 B = 1.187  
 C = 23.931

2) Standing's method<sup>11</sup> :

$$R_{so} = \gamma_g \left[ \frac{\frac{P}{18.2} + 1.4}{10^{(0.00091T - 0.0125API)}} \right]^{1/0.83} \quad (2.2.24)$$

3) Glaso's method<sup>23</sup> :

$$R_{so} = \gamma_g \left[ \frac{P_b^* \cdot API^{0.989}}{T^{0.172}} \right]^{1/0.816} \quad (2.2.25)$$

where

$$P_b^* = 10^{[2.887 - 1.819(4.285 - \log P)^{0.5}]}$$

3) Lasater<sup>25</sup> :

The molefraction of gas can be approximated by:

$$\gamma_g = \frac{0.0119 + 0.341 \left( \frac{P \gamma_g}{T + 460} \right)}{1 + 0.257 \left( \frac{P \gamma_g}{T + 460} \right)}$$

The solution gas-oil ratio from Lasater's correlation is then:

$$R_{so} = \frac{379.482 \cdot 62.3663 \cdot 5.614584 \gamma_o}{M_{wto}} \frac{y_g}{1 - y_g} \quad (2.2.26)$$

where 379.482 is the volume of an ideal gas (ft<sup>3</sup>/lbmole) at 60°F and 14.696 psia

Effective molecular weight from Lasater's graph:

$$M_{wto} = 794.2 - 19.43API + 0.137API^2$$

Gas-water ratio can be calculated according to:

1) McCain method <sup>14</sup> :

$$R_{sw1} = A + BP + CP^2$$

where

$$A = 8.1584 - 6.123 \times 10^{-2}T + 1.917 \times 10^{-4}T^2 - 2.165 \times 10^{-7}T^3$$

$$B = 0.0101 - 7.442 \times 10^{-5}T + 3.06 \times 10^{-7}T^2 - 2.949 \times 10^{-10}T^3$$

$$C = -(9.025 - 0.13T + 8.534 \times 10^{-4}T^2 - 2.341 \times 10^{-6}T^3 + 2.37 \times 10^{-9}T^4) \times 10^{-7}$$

To adjust for salinity:

$$R_{sw} = R_{sw1} \times 10^{\left[ \frac{-0.084 \text{NaCl}}{1000} T^{-0.286} \right]} \quad (2.2.27)$$

2) Dodson and Standing <sup>20</sup> :

The figure given by Dodson and Standing can be expressed as:

$$R_{sw1} = A + BP + CP^2$$

where

$$A = 2.58472 - 2.9689 \times 10^{-3}T - 1.7265 \times 10^{-5}T^2$$

$$B = 0.010447 - 4.8458 \times 10^{-5}T + 1.3474 \times 10^{-7}T^2$$

$$C = -(8.8865 + 3.9753 \times 10^{-2}T - 1.0399 \times 10^{-4}T^2) \times 10^{-7}$$

To adjust for salinity:

$$R_{sw} = R_{sw1} \times R_{swBrine} \quad (2.2.28)$$

## 2.2.6 Bubble Point Pressure

Bubble point pressure is defined as the point at which the first bubble of gas is formed in the liquid. It is calculated from the same correlations as solution gas-oil ratio, but now with gas-oil ratio at saturation:

1) Vasquez and Beggs <sup>13</sup> :

$$P_b = \left\{ \frac{R_{so}}{A\gamma_g e^{\left[ \frac{C \cdot API}{T+460} \right]}} \right\}^{1/B} \quad (2.2.29)$$

where A, B, and C are given in section 2.2.5.

2) Standing's method<sup>11</sup> :

$$P_b = 18.2(C_N - 1.4) \quad (2.2.30)$$

where

$$C_N = \left( \frac{R_{sob}}{\gamma_g} \right)^{0.83} \times 10^{(0.00091T - 0.0125API)}$$

3) Glaso<sup>23</sup> :

$$P_b = 10^{(1.7669 + 1.7447 \log P_b^*)} - 0.03022 \times (\log P_b^*)^2 \quad (2.2.31)$$

where

$$P_b^* = \left( \frac{R_{sob}}{\gamma_g} \right)^{0.816} \times \frac{T^{0.172}}{API^{0.989}}$$

4) Lasater<sup>25</sup> :

$$P_b = 5.535 \times 10^{-3} + 2.365\gamma_g + 1.519\gamma_g^2 + 5.698\gamma_g^3 \frac{T + 460}{\gamma_o} \quad (2.2.32)$$

where

$$\gamma_g = \frac{\frac{R_{sob}}{379.482}}{\frac{R_{sob}}{379.482} + 62.3663 \times 5.614584 \frac{\gamma_o}{M_{wto}}}$$

For other liquids water vapor-pressure is calculated instead of the bubble-point pressure<sup>8</sup>:

$$P_{Vw} = 0.019337e^{\left( \frac{18.3036 - \frac{3816.44}{T+460}}{1.8 - 46.13} \right)} \quad (2.2.33)$$

$$P_{VMeOH} = 0.019337e^{\left( \frac{18.5875 - \frac{3626.55}{T+460}}{1.8 - 34.29} \right)} \quad (2.2.34)$$

$$P_{VMEG} = 0.019337e^{\left( \frac{20.2501 - \frac{6022.18}{T+460}}{1.8 - 28.25} \right)} \quad (2.2.35)$$

$$P_{VDEG} = 0.019337e^{\left(17.0326 - \frac{4122.52}{\frac{T+460}{1.8} - 122.5}\right)} \quad (2.2.36)$$

For triethylene-glycol a regression analysis is performed on the curve given by Gallant<sup>18</sup>:

$$P_{VTEG} = e^{\left(\frac{-54.284 + 0.1153 \frac{T+460}{1.8}}{1 + 0.004 \frac{T+460}{1.8}}\right)} \quad (2.2.37)$$

Linear interpolation is used to adjust water content.

### 2.2.7 Viscosity

Four methods are available for calculating viscosity of crude oil below the bubble point:

#### 1) Beggs/Robinson and Ng/Egbogah<sup>26, 27</sup> :

This method uses the Beggs/Robinson method for calculating the effect of solution gas-oil-ratio,  $R_{so}$ , on the dead oil viscosity:

$$\mu_o = A\mu_{Dead}^B \quad (2.2.38)$$

where

$$A = 10.715(R_{so} + 100)^{-0.515}$$

$$B = 5.44(R_{so} + 150)^{-0.338}$$

$$\mu_{Dead} = 10^{\left[10^{(1.8653 - 0.02509 API - 0.5644 \log(T))}\right]} - 1$$

#### 2) Beal, Chew and Connally<sup>28, 29</sup> :

The Chew and Connally equation for saturated oil is the same as equation (2.2.38):

$$\mu_o = A\mu_{Dead}^B \quad (2.2.39)$$

where

$$A = 10^{\left[R_{so} (2.2 \times 10^{-7} R_{so} - 7.4 \times 10^{-4})\right]}$$

$$B = \frac{0.68}{10^{8.62 \times 10^{-5} R_{so}}} + \frac{0.25}{10^{1.1 \times 10^{-3} R_{so}}} + \frac{0.062}{10^{3.74 \times 10^{-3} R_{so}}}$$

$$\mu_{Dead} = \left(0.32 + \frac{1.8 \times 10^7}{API^{4.53}} \left[\frac{360}{T + 200}\right]^C\right)$$

$$\text{where } C = 10^{(0.43 + 8.33 / API)}$$

#### 3) Beggs and Robinson<sup>27</sup> :

$$\mu_o = A\mu_{Dead}^B \quad (2.2.40)$$

where A and B are given above under equation (2.2.38), and

$$\mu_{Dead} = 10^{\left[ \frac{3.032 - 0.0202API}{T^{1.163}} \right]} - 1$$

4) Glaso + Beggs and Robinson<sup>23, 27</sup> :

Glaso gives the viscosity of dead oil specifically for North Sea oils. As he does not give any correlation for live oil viscosity, it is chosen here to use the Beggs and Robinson method for live oil combined with the dead oil viscosity correlation of Glaso:

$$\mu_{Dead} = \frac{3.141 \times 10^{10}}{T^{3.444}} (\log API)^{[1.0313 \cdot \log T - 36.447]} \quad (2.2.41)$$

Two correlations are included for calculating the viscosity above the bubble point:

1) Vasquez and Beggs<sup>13</sup> :

$$\mu_o = \mu_{ob} \left( \frac{P}{P_b} \right)^M \quad (2.2.42)$$

where

$$M = 2.6P^{1.187} e^{(-11.513 - 8.98 \times 10^{-5} P)}$$

2) Beal<sup>28</sup> :

Standing<sup>11</sup> gives a curve-fit of Beal's figure as:

$$\mu_o = \left[ 0.024\mu_{ob}^{1.6} + 0.38\mu_{ob}^{0.56} \right] \bullet 0.001(P - P_b) + \mu_{ob} \quad (2.2.43)$$

Water viscosity can be calculated by the following methods:

1) Brill and Beggs<sup>16</sup> :

The correlation is based on a figure by van Wingen<sup>30</sup>:

$$\mu_{w1} = e^{(1.003 - 1.479 \times 10^{-2} T + 1.982 \times 10^{-6} T^2)}$$

With quantities of salt the viscosity should be adjusted as<sup>16</sup>:

$$\mu_w = \mu_{w1} \left( 2.083 \times 10^{-2} \frac{NaCl}{10000} + 1 \right) \quad (2.2.44)$$

2) McCain<sup>14</sup> :

The viscosity at atmospheric pressure is given by:

$$\mu_{wa} = AT^{-B}$$

where

$$A = 109.574 - 8.406 \frac{NaCl}{10000} + 0.3133 \left( \frac{NaCl}{10000} \right)^2 + 8.722 \times 10^{-3} \left( \frac{NaCl}{10000} \right)^3$$

$$B = 1.1217 - 2.64 \times 10^{-2} \frac{NaCl}{10000} + 6.795 \times 10^{-4} \left( \frac{NaCl}{10000} \right)^2 + 5.47 \times 10^{-5} \left( \frac{NaCl}{10000} \right)^3 - 1.556 \times 10^{-6} \left( \frac{NaCl}{10000} \right)^4$$

For higher pressures the viscosity is adjusted as:

$$\mu_w = \mu_{wa} (0.9994 + 4.03 \times 10^{-5} P + 3.106 \times 10^{-9} P^2) \quad (2.2.45)$$

3) van Velzen, Cardozo and Langenkamp<sup>8</sup> :

Reid, Prausnitz and Sherwood<sup>8</sup> recommend using this method for liquids up to  $T_{pr} = 0.75$ :

$$\log \mu_{wa} = 658.25 \left( \frac{1}{T} - \frac{1}{283.16} \right)$$

where

$\mu_{wa}$  is in cP and T is in K

Viscosity of methanol and glycols are calculated according to the following equations based on Gallant's charts<sup>18</sup>:

$$\mu_{MeOH} = 0.8389 e^{(-0.02T + 1.282 \times 10^{-4} T^2 - 5.973 \times 10^{-7} T^3)} \quad (2.2.46)$$

$$\mu_{MEG} = 56.561 e^{(-0.0545T + 2.785 \times 10^{-4} T^2 - 6.277 \times 10^{-7} T^3)} \quad (2.2.47)$$

$$\mu_{DEG} = 112.589 e^{(-0.0543T + 1.096 \times 10^{-4} T^2 - 1.866 \times 10^{-7} T^3)} \quad (2.2.48)$$

$$\mu_{TEG} = 1162.55 e^{(-0.064T + 3.226 \times 10^{-4} T^2 - 6.766 \times 10^{-7} T^3)} \quad (2.2.49)$$

## 2.2.8 Thermal Conductivity

Based on the chart given in GPSA Engineering Data Book<sup>9</sup>, the following correlation is developed for the thermal conductivity of oil at low pressures:

$$K_{hoa} = 5.04 \times 10^{-2} - 3.1 \times 10^{-4} T - 2.61 \times 10^{-7} T^2 + (2.814 \times 10^{-2} + 2.833 \times 10^{-4} T + 2.78 \times 10^{-7} T^2) \gamma_o$$

For higher pressures a method described in the API Technical Data Book is used<sup>31</sup>:

$$k_{ho} = k_{hoa} \frac{C_2}{C_1} \quad (2.2.50)$$

where

$$C = 17.77 + 0.65 P_{pr} - 7.764 T_{pr} - \frac{2.054 T_{pr}^2}{e^{0.2 P_{pr}}}$$

and  $C_1$  and  $C_2$  represent the standard condition ( $T = 60$  °F,  $P = 14.696$  psia) and elevated pressure conditions respectively.

Thermal conductivity of formation water is based on tables of thermal conductivity of water versus temperature and pressure<sup>32</sup>. A correlation developed from these tables is:

$$K_{hwa} = 0.56 + 2.785 \times 10^{-5} P + 4.64 \times 10^{-8} P^2 + T(1.891 \times 10^{-3} + 2.0 \times 10^{-6} P - 7.493 \times 10^{-9} P^2 + 8.638 \times 10^{-12} P^3) + T^2(6.974 \times 10^{-5} + 9.2 \times 10^{-9} P - 3.767 \times 10^{-11} P^2 + 4.234 \times 10^{-14} P^3)$$

where P is pressure in bara, and T is temperature in °C. Thermal conductivity is here in W/m·K.

By assuming all dissolved salts is in the form of Na<sup>+</sup> -ions and Cl<sup>-</sup> -ions, the following equation can be used for adjusting the thermal conductivity at 20°C for the amount of salts <sup>8</sup>:

$$k_{hw20} = k_{hwa20} - 9.347 \times 10^{-8} \text{ NaCl}$$

where NaCl is concentration of salt in PPM.

To adjust for calculations at other temperatures with thermal conductivity in W/m·K:

$$K_{hw} = k_{hw20} \frac{k_{hwa}}{k_{hwa20}} \quad (2.2.51)$$

Other liquid thermal conductivities are calculated as follows based on Gallant's charts <sup>18</sup>:

Methanol (MeOH):

$$k_{hMeOH} = k_{hw} e^{[-(8.308 \times 10^{-3} + 2.485 \times 10^{-5} T)(100 - Wt\%_w)]} \quad (2.2.52)$$

Monoethylene glycol (MEG):

If Wt%<sub>MEG</sub> ≤ 80 then

$$k_{hMEG} = k_{hw} - AWt\%_{MEG} + BWt\%_{MEG}^2 \quad (2.2.53)$$

where

$$A = \frac{0.65035 + 0.01085T}{1 + 4.0353 \times 10^{-3} T} \quad B = \frac{1.4286 + 3.799 \times 10^{-2} T}{1 + 4.2745 \times 10^{-3} T}$$

For Wt%<sub>MEG</sub> > 80 linear interpolation with equation (2.2.53) and water is performed.

Diethylene glycol (DEG):

If Wt%<sub>DEG</sub> ≤ 80 then

$$k_{hDEG} = k_{hw} - AWt\%_{DEG} + BWt\%_{DEG}^2 \quad (2.2.54)$$

where

$$A = \frac{0.4726 + 2.51116 \times 10^{-2} T}{1 + 0.0119T} \quad B = \frac{-5.2113 + 0.2704T}{1 + 0.0551T}$$

For Wt%<sub>DEG</sub> > 80 linear interpolation with equation (2.2.54) and water is performed.

Triethylene glycol (TEG):

$$k_{hTEG} = A - BWt\%_{TEG} \quad (2.2.55)$$

where

$$A = 132.63 + 0.10835T \quad B = 0.75024 + 2.2792 \times 10^{-3} T$$

For Wt%<sub>TEG</sub> > 80 or Wt%<sub>TEG</sub> < 20 linear interpolation with equation (2.2.55) and water is performed.

## 2.2.9 Surface Tension

Two correlations are available for estimating surface tension of crude oils:

1) Baker and Swerdloff method <sup>33</sup>:

At atmospheric pressure their empirical plot can be approximated by:

T < 68°F:

$$\sigma_a = \sigma_{68} = 39 - 0.2571API \quad (2.2.56)$$

T > 100°F:

$$\sigma_a = \sigma_{100} = 37.5 - 0.2571API \quad (2.2.57)$$

Other temperatures:

$$\sigma_a = \sigma_{68} - (T - 68) \frac{\sigma_{68} - \sigma_{100}}{32} \quad (2.2.58)$$

At higher pressures the surface tension based on Baker's figures is calculated as:

$$\sigma_o = \sigma_a e^{-0.00073P} \quad (2.2.59)$$

2) API method <sup>31</sup>:

At low pressures:

$$\sigma_a = 673.7 \frac{\left[ \frac{T_{pc} - T + 460}{T_{pc}} \right]^{1.23}}{\frac{T_{ba}^{0.333}}{\gamma_o}} \quad (2.2.60)$$

The mean average boiling point can be calculated as <sup>34</sup>:

$$T_{ba} (^{\circ}R) = \left[ \frac{M_{wto}}{4.567 \times 10^{-5} \gamma_o^{-1.0164}} \right]$$

The moleweight,  $M_{wto}$ , is supplied by the user or calculated according to a figure given by Campbell <sup>35</sup>:

$$M_{wto} = 588.7268 \gamma_o^{5.5721}$$

The Baker and Swerdloff figure is used for calculating the high pressure surface tension:

$$\sigma_o = \sigma_a e^{-0.00073P} \quad (2.2.61)$$

For gas-water systems an equation is developed based on Katz et. al. Chart of surface tension <sup>36</sup>:

$$T < 74^{\circ}F: \quad \sigma_w = \sigma_{74} = \frac{74.367 + 1.928 \times 10^{-2} P}{1 + 4.492 \times 10^{-4} P} \quad (2.2.62)$$

T > 280°F and P < 4845.87 psia:

$$\sigma_w = \sigma_{280} = 52.8 - 6.518 \times 10^{-3} P \quad (2.2.63)$$

For other pressures:

$$\sigma_w = 25.52 - 1.689 \times 10^{-3} P + 1.652 \times 10^{-7} P^2 \quad (2.2.64)$$

T > 74°F and T < 280°F:

$$\sigma_w = \sigma_{w74} + (T - 74) \frac{\sigma_{w280} - \sigma_{w74}}{206} \quad (2.2.65)$$

Surface tension of methanol and glycols is calculated as <sup>17</sup>:

$$\sigma_{MeOH} = 26.039 - 0.05T \quad (2.2.66)$$

$$\sigma_{MEG} = 51.75 - 4.9 \times 10^{-2} T \quad (2.2.67)$$

$$\sigma_{DEG} = 48.94 - 4.9 \times 10^{-2} T \quad (2.2.68)$$

$$\sigma_{TEG} = 48.5 - 4.9 \times 10^{-2} T \quad (2.2.69)$$

For different concentrations of water, the above equations, (2.2.66 – 2.2.69) are linearly interpolated with water surface tension given as <sup>31</sup>:

$$\sigma_w = 78.46 - 8.209 \times 10^{-2} T$$

### 2.2.10 Specific Heat Capacity

Specific heat for oil is calculated according to <sup>31</sup>:

$$C_o = A + B(T+460) + C(T+460)^2 \quad (2.2.70)$$

where

$$A = -1.17 + (0.02372 + 0.02491\gamma_o) \frac{T_{ba}^{0.333}}{\gamma_o} + \frac{1.1498 - 0.04654 \frac{T_{ba}^{0.333}}{\gamma_o}}{\gamma_o}$$

$$B = 10^{-4} \left( 1.0 + 0.8246 \frac{T_{ba}^{0.333}}{\gamma_o} \right) \left( 1.1217 - \frac{0.2763}{\gamma_o} \right)$$

$$C = -10^{-8} \left( 1.0 + 0.8246 \frac{T_{ba}^{0.333}}{\gamma_o} \right) \left( 2.9027 - \frac{0.7096}{\gamma_o} \right)$$

For water, equations are developed from tables of specific heat at given temperatures and pressures <sup>32</sup>:

$$C_{wa} = 4.2102 - 1.882 \times 10^{-4} P - (1.428 \times 10^{-3} + 5.256 \times 10^{-7} P) T + (1.448 \times 10^{-5} - 6.684 \times 10^{-9} P) T^2$$

where P = bara, T = °C, and C<sub>wa</sub> = kJ/kg.K. A correlation is developed for calculating the effect of salt (ionic effect) on the specific heat based on a table given in 'Chemical Engineers Handbook' <sup>37</sup>:

$$S_{Corr} = 4.1833 \left[ \left( 1.67 \times 10^{-2} - 1.1298 \times 10^{-3} \ln T \right) NaCl - \left( 3.76 \times 10^{-4} + 2.736 \times 10^{-5} T \right) \frac{NaCl^2}{1 + 0.182T} \right]$$

The specific heat is calculated as:  $C_w = C_{wa} \cdot S_{Corr}$  (2.2.71)

Specific heat of methanol and glycols is calculated from <sup>17</sup>:

$$C_{MeOH} = 0.541 + 7.085 \times 10^{-4} T + 3.123 \times 10^{-6} T^2 \quad (2.2.72)$$

$$C_{MEG} = C_w - A \cdot \text{Wt\%}_{MEG} - B \cdot (\text{Wt\%}_{MEG})^2 \quad (2.2.73)$$

where

$$A = \frac{5.7829 e^{(-5.2734 \times 10^{-3} T)}}{1000} \quad B = 10^{-5} \frac{-0.6778 + 0.0227T}{1 + 6.95 \times 10^{-3} T}$$

$$C_{DEG} = C_w - A \cdot \text{Wt\%}_{DEG} - B \cdot (\text{Wt\%}_{DEG})^2 \quad (2.2.74)$$

where

$$A = 10^{-3} \frac{5.3066 - 0.01534T}{1 + 1.387 \times 10^{-3} T} \quad B = 10^{-5} \frac{-0.2061 + 0.01534T}{1 + 1.678 \times 10^{-3} T}$$

$$C_{\text{TEG}} = C_w - A \cdot \text{Wt\%}_{\text{TEG}} - B \cdot (\text{Wt\%}_{\text{TEG}})^2 \quad (2.2.75)$$

where

$$A = 10^{-3} \frac{4.1223 - 8.339 \times 10^{-3} T}{1 + 2.859 \times 10^{-3} T} \quad B = 10^{-5} \frac{1.0786 + 2.189 \times 10^{-2} T}{1 + 5.46 \times 10^{-3} T}$$

For different concentrations of water, the above equations (2.2.72-2.2.75) are linearly interpolated with water specific heat given as:

$$C_w = 1.0152 - 3.617 \times 10^{-4} T + 2.171 \times 10^{-6} T^2 - 2.983 \times 10^{-9} T^3$$

### 2.2.11 Critical and Pseudo-Reduced Properties

Oil pseudo-critical pressure and temperature <sup>9, 31, 34</sup>:

$$P_{pc} = 3.12281 \times 10^9 T_{ba}^{-2.3125} \gamma_o^{2.30221} \quad (2.2.76)$$

$$T_{pc} = 24.2787 T_{ba}^{0.58848} \gamma_o^{0.3596} \quad (2.2.77)$$

The critical properties for water are taken from GPSA's 'Engineering Data Book' <sup>9</sup>:

$$P_{pc} = 3208 \quad (\text{psia}) \quad (2.2.78)$$

$$T_{pc} = 1165.27 \quad (^\circ \text{R}) \quad (2.2.79)$$

MeOH critical properties <sup>9</sup>:

$$P_{pc} = 1153.9 \quad (\text{psia}) \quad (2.2.80)$$

$$T_{pc} = 923.67 \quad (^\circ \text{R}) \quad (2.2.81)$$

MEG critical properties <sup>9</sup>:

$$P_{pc} = 1116.2 \quad (\text{psia}) \quad (2.2.82)$$

$$T_{pc} = 1161.27 \quad (^\circ \text{R}) \quad (2.2.83)$$

DEG critical properties <sup>9</sup>:

$$P_{pc} = 676.0 \quad (\text{psia}) \quad (2.2.84)$$

$$T_{pc} = 1226.07 \quad (^\circ \text{R}) \quad (2.2.85)$$

TEG critical properties <sup>9</sup>:

$$P_{pc} = 479.2 \quad (\text{psia}) \quad (2.2.86)$$

$$T_{pc} = 1287.67 \quad (^\circ \text{R}) \quad (2.2.87)$$

Pseudo-reduced temperature and pressure are calculated as:

$$P_{pr} = P/P_{pc} \quad (2.2.88)$$

$$T_{pr} = (T + 460)/T_{pc} \quad (2.2.89)$$

### 3.0 ORIFICE CALCULATIONS

#### 3.1 AGA-3 (API 2530-1992, Part 3)

##### 3.1.1 Orifice Meters with Flange Taps

The fundamental orifice meter flow equation is <sup>1</sup>:

$$Q_m = 359.072 C_d (FT) E_v Y_1 d^2 \sqrt{\rho_{P_1} h_w} \quad (3.1.1)$$

where

|              |   |   |
|--------------|---|---|
| $Q_m$        | = | mass flow rate (lb <sub>m</sub> /h)   |
| $C_d(FT)$    | = | coefficient of discharge for flange tapped orifice meters = $F_c + F_{sl}$        |
| $F_c$        | = | orifice calculation factor  |
| $F_{sl}$     | = | orifice slope factor  |
| $E_v$        | = | velocity of approach factor   |
| $Y_1$        | = | expansion factor (upstream tap)   |
| $d$          | = | orifice plate bore diameter, inches   |
| $\rho_{P_1}$ | = | density of fluid at upstream flowing conditions, lb <sub>m</sub> /ft <sup>3</sup> |
| $h_w$        | = | orifice differential pressure, inches H <sub>2</sub> O at 60 °F                   |

The general practice is to express the flow in ft<sup>3</sup>/h at some defined standard condition. For gas this can be done by using the orifice flow constant:

$$Q_v = C^* \sqrt{P_1 h_w} \quad (3.1.2)$$

where

|                      |   |   |
|----------------------|---|---|
| $Q_v$                | = | volume flow rate at standard conditions, ft <sup>3</sup> /h   |
| $C^*$                | = | composite orifice flow constant   |
| $P_1$                | = | absolute flowing pressure upstream tap, psia  |
| $h_w$                | = | orifice differential pressure, inches H <sub>2</sub> O at 60 °F   |
| $C$                  | = | $F_n(F_c + F_{sl})Y_1F_{pb}F_{tb}F_{tf}F_{gr}F_{pv}$  |
| $F_n$                | = | numeric conversion factor = $338.196 \times E_v \times D^2 \beta^2$   |
| $F_c$                | = | $0.5961 + 0.0291 \times \beta^2 - 0.2290 \times \beta^8 + (0.0433 + 0.0712 \times e^{-8.5/D} - 0.1145 \times e^{-6.0/D}) \times [1 - 0.23 \times (19,000 \times \beta / R_{eD})^{0.8}] \times [\beta^4 / (1 - \beta^4)] - 0.0116 \times \{2 / [D \times (1 - \beta)] - 0.52 \times (2 / [D \times (1 - \beta)])^{1.3}\} \times \beta^{1.1} \times \{1 - 0.14 \times (19,000 \times \beta / R_{eD})^{0.8}\}$ |
| $F$                  | = | $0.000511 \times (1,000,000 \times \beta / R_{eD})^{0.7} + \{0.0210 + 0.0049 \times (19,000 \times \beta / R_{eD})^{0.8}\} \times \beta^4 \times (1,000,000 \times \beta / R_{eD})^{0.33}$  |
| $Y_1$                | = | $1 - (0.41 + 0.35 \times \beta^4) \times (x_1/k)$   |
| $F_{pb}$             | = | pressure base factor = $14.73/P_b$  |
| $F_{tb}$             | = | temperature base factor = $T_b/519.67$  |
| $F_{tf}$             | = | flowing temperature factor = $(519.67/T)^{0.5}$   |
| $F_{gr}$             | = | real gas relative density factor = $(1/SPGG)^{0.5}$   |
| $F_{pv}$             | = | supercompressibility factor = $(Z_b/Z)^{0.5}$   |
| $E_v$                | = | $(1 - \beta^4)^{-0.5}$  |
| $\beta$              | = | diameter ratio, (d/D)   |
| $d$                  | = | $d_r \{1 + \alpha_1 (T - T_r)\}$  |
| $D$                  | = | $D_r \{1 + \alpha_2 (T - T_r)\}$  |
| $\alpha_1, \alpha_2$ | = | linear coefficient of thermal expansion for orifice material and pipe material respectively, here taken from Miller (Ref. 1)  |
| $x_1/k$              | = | acoustic ratio  |
| $x_1$                | = | ratio of differential pressure to absolute pressure at upstream tap = $(P_1 - P_2)/P_2 = h_w/27.707/P_1$  |
| $k$                  | = | isentropic coefficient  |

and  $d_r$ ,  $D_r$  and  $T_r$  refer to measurements at a reference temperature of 68 °F.

Variables without subscripts,  $Z$ ,  $T$ ,  $P_1$ , and  $P_2$  refer to actual flowing conditions.

If pipe diameter is less than 2.8 inches, the orifice calculation factor is modified as follows:

$$F_c^* = F_c + 0.003 \times (1 - \beta) \times (2.8 - D)$$

For liquids the composite orifice flow constant reduces to:

$$C^* = F_n(F_c + F_{sl})F_{pb}F_{tb}F_{tf}F_{gt}$$

### 3.1.2 Orifice Meters with Pipe Taps

Equation (3.1.2) above is also used for pipe taps, but the definition of  $C^*$  = orifice flow constant is different:

$$Q_v = C^* \sqrt{P_1 h_w} \quad (3.1.2)$$

$$C^* = \text{orifice flow constant} = F_b F_r Y_1 F_{pb} F_{tb} F_{tf} F_{gr} F_{pv}$$

$$F_b = \text{basic orifice factor} = 338.178 d^2 K_0$$

$$F_r = \text{Reynolds number factor} = 1 + E/R_{eD}$$

$$K_0 = \text{coefficient of discharge at infinite Reynolds number} \\ = K_e / [1 + 15 \times E / 1,000,000 / d]$$

$$K_e = \text{coefficient of discharge when Reynolds number} = 1,000,000 \times d / 15 \\ = 0.5925 + 0.0182/D + (0.44 - 0.06/D) \times \beta^2 + (0.935 + 0.225/D) \times \beta^5 + 1.35 \times \beta^{14} \\ + (1.43/D^{0.5}) \times (0.25 - \beta)^{2.5}$$

$$E = d \times (905 - 5000 \times \beta + 9000 \times \beta^2 - 4200 \times \beta^3 + 875/D)$$

The other factors  $Y_1$ ,  $F_{pb}$ ,  $F_{tb}$ ,  $F_{tf}$ ,  $F_{gr}$ , and  $F_{pv}$  are defined above.

From  $K_0$  and  $K_e$  above it can be shown that the discharge coefficient at any Reynolds number is:

$$K = K_0 \times (1 + E/R_{eD})$$

For liquids  $C^*$  reduces to:  $C^* = F_b F_r F_{pb} F_{tb} F_{tf} F_{gt}$

where  $F_{gt}$  = gravity temperature factor for liquids, only included for oil calculations

## 3.2 ISO-5167 (ISO-5167-2: 2003)

### 3.2.1 Orifice Meters

The rate of mass flow is related to the pressure differential according to <sup>38</sup> :

$$Q_m = C E \varepsilon \frac{\pi}{4} d^2 \sqrt{2 \Delta P \rho_1} \quad (3.2.1)$$

where

$$\Delta P = P - P_v = \text{upstream pressure minus vapor pressure of liquid, Pa}$$

$$Q_m = \text{mass flow rate (kg/s)}$$

$$C = \text{discharge coefficient} = \frac{\alpha}{E}$$

$$E = \text{velocity of approach factor} = (1 - \beta^4)^{-0.5}$$

$$d = \text{orifice diameter at actual flowing conditions, m}$$

$$\rho_1 = \text{density of flowing fluid measured at upstream tap, kg/m}^3$$

$$\alpha = \text{flow coefficient} = CE$$

$$\beta = d/D$$

The discharge coefficient, C, and the expansion factor are determined empirically<sup>38</sup>:

$$C = 0.5961 + 0.0261\beta^2 - 0.216\beta^8 + 0.000521 \times [10^6 \times \beta / R_{eD}]^{0.7} + (0.0188 + 0.006A)\beta^{3.5} \times [10^6 / R_{eD}]^{0.3} \\ + (0.043 + 0.080e^{-10L_1} - 0.216e^{-7L_1})(1 - 0.11A) \times (\beta^4 / (1 - \beta^4)) - 0.031(M'_2 - 0.8M'_2)^{1.1} \beta^{1.3}$$

If  $D < 71.12$  mm (2.8 in), the following term must be added to the above equation:

$$+ 0.011(0.75 - \beta)(2.8 - D/25.4) \quad (D \text{ in millimeter})$$

where  $R_{eD}$  = pipe Reynolds number

$$M'_2 = 2L'_2 / (1 - \beta)$$

$$A = (19000\beta / R_{eD})^{0.8}$$

For corner tappings:  $L_1 = L'_2 = 0$

For D and D/2 tappings:  $L_1 = 1.0$  and  $L'_2 = 0.47$

For flange tappings:  $L_1 = L'_2 = 25.4/D$  (D in millimeter)

The expansion factor is given as:

$$\varepsilon = 1 - (0.351 + 0.256\beta^4 + 0.93\beta^8) \{1 - [p_2/p_1]^{1/\kappa}\} \quad (3.2.2)$$

For liquids the expansion factor = 1.0. Compressibilities are adjusted for liquid density calculations.

### 3.3 Restriction Orifices (Critical Flow Orifices)

Restriction orifice calculations in this program are performed according to R. W. Miller's "Flow Measurement Engineering Handbook"<sup>2</sup>.

#### 3.3.1 Gas Restriction Orifices

Restriction orifices are calculated based on critical flow in the orifice. This means that sonic velocity exists at the orifice throat, and further decrease in the downstream pressure will not increase the mass flow rate. For critical flow the basic mass flow rate equation is:

$$Q_m = 1335.485Cd^2Y_{CR}\sqrt{Z\rho F_{TP}P} \quad (3.3.1)$$

where

$Q_m$  = mass flow rate, lb<sub>m</sub>/h

C = critical discharge coefficient

- d = orifice diameter at flowing conditions, in
- $Y_{CR}$  = critical flow function
- $F_{TP}$  = total pressure correction factor to adjust for difference between static pressure read at the pipe wall (Manometer) and total pressure of the fluid
- and Z,  $\rho$ , and P are measured at flowing upstream conditions

For  $\beta$ -ratios less than 0.5 the total pressure correction factors are approximated by:

$$F_{TP} = \{1 - k/2[2/(k+1)]^{[(k+1)/(k-1)]} \times \beta^{4-1}\} \quad (3.3.2)$$

$$Y_{CR} = \{k/Z \times [2/(k+1)]^{[(k+1)/(k-1)]}\}^{0.5} \quad (3.3.3)$$

where

- k = isentropic coefficient at flowing conditions

Assuming steady isentropic flow, critical flow (choked flow) occurs when:

$$P_2/P_1/F_{TP} \leq [2/(k+1)]^{[k/(k-1)]}$$

where

- $P_1$  = pressure upstream orifice
- $P_2$  = pressure downstream orifice

By assuming sharp-edged orifices with plate thickness to bore diameter between 1 and 6 the discharge coefficient is a constant given as <sup>2</sup> :

$$C = 0.83932$$

### 3.3.2 Liquid Restriction Orifices

Liquid choked flow occurs if a cavitation barrier exists within an orifice. Only upstream pressure increases can increase flowrates. Thick square-edged orifices are used as they are inexpensive. The sizing and flowrate equation used for liquids is:

$$Q_m = CE \frac{\pi}{4} d^2 \sqrt{2\Delta P \rho_1} \quad (3.3.4)$$

where

- $\Delta P$  =  $P - P_v$  = upstream pressure minus vapor pressure of liquid, Pa
- $\rho_1$  = 1.0 = expansion factor for liquids

Other factors in equation (3.3.4) are defined above.

Assuming square-edged orifices with plate thickness to bore diameter less than 6 with a minimum of 0.125 in (3 mm), the liquid restriction orifice constant is <sup>2</sup> :

$$C = 0.6 \quad (3.3.5)$$

## 4.0 WATER/GAS PHASE BEHAVIOR CALCULATIONS

Knowing the water content of natural gases is essential to the design and operation of production, dehydration and transmission systems. Water may condense in production and gathering systems. This may result in hydrate formation and plugging of flow systems and damage to internals of production equipment. Condensed water may form water slugs which will tend to decrease flow efficiency and increase the pressure drop in a line. Presence of free water in pipeline systems may also cause corrosion. If carbon dioxide and/or hydrogen sulfide are present, the gases may form carbonic acid and sulfurous acid respectively if dissolved in water.

### 4.1 Hydrate Calculations

Gas hydrates form when water attaches to molecules of methane, ethane, propane, iso-butane, carbon-dioxide, hydrogen-sulfide, and to a certain extent n-butane. Hydrates form easily with methane, carbon-dioxide, and hydrogen-sulfide. Heavier hydrocarbons tend to inhibit hydrate-formation. It is therefore more difficult to form hydrates of iso-butane and n-butane. Recent research<sup>39</sup> also shows that benzene, methylcyclopentane and cyclohexane may form hydrates in presence of methane or nitrogen.

In this program the hydrate calculations are based on Katz' method for predicting hydrate conditions<sup>9, 35</sup>. Empirically determined vapor-solid equilibrium constants are utilized in predicting hydrate formation conditions. These equilibrium constants are determined by<sup>9</sup>:

$$K_i = y_i/X_{Si} \quad (4.1.1)$$

where  $y_i$  = molefraction of hydrocarbon-component 'i' in the gas on a water-free basis  
 $X_{Si}$  = molefraction of hydrocarbon-component 'i' in the solid on a water-free basis

Equation (4.1.1) is assumed to be a function of a pressure and temperature only. The conditions for initial formation of hydrates are similar to dew-point calculations for multi-component gas mixtures. The basic equation is:

$$\sum(y_i/K_i) = 1.0 \quad (4.1.2)$$

which has to be solved for  $K_i$ .

The vapor-solid equilibrium curves used in this work are taken from the API Technical Data Book<sup>31</sup>, except the data for iso-butane<sup>40</sup> and n-butane<sup>41</sup>.

To use the vapor-solid equilibrium curves for finding  $K_i$  in equation (4.1.2), the values for  $K_i$  must be transformed into a computer-readable form. This is done by developing equations describing the curves. A least-square fit technique has been used to develop the equations for the vapor-solid equilibrium curves. These equations are listed in paper SPE 15306<sup>42</sup>.

### 4.2 Water Content Calculations

Water content calculations are based on a method developed by Campbell<sup>35</sup>. The method applies the standard physical chemistry equations and the Eykman Molecular Refraction (EMR) combination rules. For water content calculations the approach here uses fugacity coefficient as one of the key factors:

$$y = K \times (f_w/f)^Z \quad (4.2.1)$$

where  $y$  = mole fraction water in vapor phase  
 $K$  = ideal liquid K-value  
 $f_w$  = fugacity of water at given temperature and pressure  
 $f$  = fugacity of gas at given temperature and pressure  
 $Z$  = Z-factor (compressibility factor) from EMR approach

The fugacity coefficients are calculated from a generalized fugacity coefficient chart for pure vapors. As the pressure increases above reduced pressure greater than about 6.0 ( $P_r > 6.0$ ), the fugacity coefficient ( $f/P$ ) tends to increase. This is true for all equations of state. The increase in fugacity also causes an increase in water content of the gas according to equation (4.2.1) above. However, in real life water content will decrease with increasing pressure.

### 4.3 Inhibitor Injection Quantities

Methanol and glycols may be injected into a flowing gas stream in quantities sufficient to depress the hydrate temperature to the minimum system temperature or lower. If free water is present in these lines the risk of hydrate formation is large during shut-downs and start-ups due to gas coming into contact with free water condensed in the line. The possibilities of injection in a shut-down or start-up situation are small as there are no flow in the line. Therefore, injection is usually performed before a shut-down or during start-up. Current research in hydrate-prevention is concentrated on Threshold Hydrate Inhibitors (THI)<sup>43</sup> or kinetic inhibitors<sup>44</sup> other than methanol and glycols. Within a few years these inhibitors should be on the market, and the large amount of methanols and glycols used to prevent hydrate formation today will be replaced by smaller quantities of chemicals, thus easing operational handling and logistics in the field.

However, in this program the calculations are based on injection of methanol or glycols. The simple Hammerschmidt equation is used in this program to calculate the rate of inhibitor necessary to prevent hydrate formation in a pipeline. This approach will most likely calculate a conservative amount of inhibitor necessary for injection. It should be noted that these calculations do not take into account the partitioning of inhibitor into the liquid hydrocarbon phase, only the gas phase and the liquid water phase. The Hammerschmidt equation is:

$$W = 100 \times d \times M_w / (K_i + d \times M_w) \quad (4.3.1)$$

where  $W$  = weight % inhibitor in liquid water phase

$d$  = °F depression of hydrate point

$M$  = moleweight inhibitor

$K_i$  = constant, MeOH = 2335, glycols = 4000

Note that one important assumption is made for inhibitor quantities calculations. The highest temperature occurs at the highest pressure, and the lowest temperature coincides with the lowest pressure in a given pipe segment.

### 4.4 CO<sub>2</sub> Solid Formation

For CO<sub>2</sub> solid formation predictions only a simple approach is used in this program, namely the curves given in GPSA Data Book<sup>9</sup>. This will only give an estimate of the concentration at which CO<sub>2</sub> may solidify.

## 5.0 SEPARATOR CALCULATIONS

Equilibrium ratios (K-values) are used for calculating compositions of gas and liquid phases at given temperatures and pressures. Normally an equation of state (EOS) is used for predicting equilibrium ratios, and is a function of composition, pressure and temperature. Based on the fact that compositional effects on equilibrium ratios are small below about 1000 psia, Standing<sup>45</sup> developed a correlation for calculating equilibrium ratios based on data reported by Katz and Hachmuth<sup>46</sup>. The correlation gives the following equation for each component:

$$K = (1/P) \times 10^{(a + c \times F)} \quad (5.0.1)$$

where

- F =  $b \times (1/T_b - 1/T)$
- K = equilibrium ratio
- a, b, c = correlating parameter
- P = pressure, psia
- T = temperature, deg R
- T<sub>b</sub> = boiling point, deg R
- X = mole fraction in liquid phase
- y = mole fraction in vapor phase

### 5.1 Flash Calculations

In a flash operation a mixture is separated in a vapor- and a liquid-fraction in equilibrium with each other. The program performs isothermal flash calculations, i.e. the mixture is flashed at a given temperature. The main equations are:

$$\text{Total mass-balance:} \quad F = V + L \quad (5.1.1)$$

$$\text{Component mass-balance:} \quad z_i \times F = y_i \times V + X_i \times L \quad (5.1.2)$$

$$\text{At equilibrium:} \quad K_i = y_i/X_i \quad (5.1.3)$$

Combining the equations above gives:

$$z_i \times F = K_i \times X_i \times V + X_i \times (F - V) = [(K_i - 1) \times V + F] \times X_i \quad (5.1.4)$$

which, when solved for  $X_i$  gives:

$$X_i = z_i \times F / [(K_i - 1) \times V + F] = z_i / [1 + (K_i - 1) \times C] \quad (5.1.5)$$

where  $C = V/F$  which can vary between 0 and 1

The sum of mole-fractions for the liquid and vapor phases must both equal 1.0:

$$\sum(y_i) = 1.0$$

$$\sum(X_i) = 1.0$$

and 
$$\sum(y_i) - \sum(X_i) = 0 \quad (5.1.6)$$

Combining equation (5.1.3) with equation (5.1.5), gives:

$$y_i = K_i z_i / [1 + (K_i - 1) \times C] \quad (5.1.7)$$

and equation (5.1.6) can be written as:

$$\sum \frac{(K_i - 1) z_i}{1 + (K_i - 1) C} = 0 \quad (5.1.8)$$

There may be a problem in solving this equation when  $C = 0$  (no vapor phase) or  $C = 1$  (all vapor phase).

## 5.2 Oil Density

Oil density is used for calculating the volume of reservoir and stock tank oils. Standing<sup>45</sup> gives a method for calculating densities for pseudo-liquids containing methane and ethane. The method is based on using an apparent liquid density system. The correlation for the pseudo-density uses weight percent methane, ethane, and density of propane plus fraction. For further study of this matter you are referred to Standing<sup>45</sup>.

## 5.3 Separator Calculations

An oil and gas separator must be capable of accepting rapid changes in flow and gas/oil ratio. In many instances wells will "unload" slugs of liquid and alternately produce abnormal quantities of gas. These separators must therefore be conservatively designed and have ample liquid surge capacity.

Normally, simulations of separation processes are based on 1 mole of feed since the physical properties (gas/oil ratio, etc.) are independent of molar quantity of feed. For a series of separators, the gas is normally commingled from each separator into a single gas gathering system, while the liquid phases are flashed consecutively through the series of separators. For a two-phase separation process (high pressure and stock tank), the first flash results in  $L_1$  moles of liquid and  $V_1$  moles of gas. The  $L_1$  moles of liquid are then the feed mixture to the 2nd stage, resulting in  $L_2$  moles of liquid and  $V_2$  moles of gas. In general, the number of liquid moles,  $N_{st}$ , resulting from  $m$  number of flashes:

$$N_{st} = L_1 \times L_2 \times \dots \times L_m \quad (5.3.1)$$

Assuming 1 mole of feed, the number of moles of gas from the 1st stage is  $V_1$ . In the second stage,  $L_1$  moles are being flashed, resulting in  $L_1 V_2$  moles of gas. In general, the number of moles of gas produced at any stage:

$$N_{gj} = (L_{m-1} \times L_{m-2} \times \dots \times L_1) V_m \quad (5.3.2)$$

The total number of moles of gas for an  $M$  stage flash:

$$N_{gt} = V_1 + L_1 V_2 + \dots + (L_{m-1} \times L_{m-2} \dots \times L_1) V_m \quad (5.3.3)$$

The stock tank oil composition is the result from the final flash separation. The total separator gas composition is the mole weighted composition of each stage where the total moles for each component,  $N_{it}$ :

$$N_{it} = y_{i1} N_{g1} + y_{i2} N_{g2} + \dots + y_{ij} N_{gj} \quad (5.3.4)$$

The mole fraction of a separator gas component is:

$$y_i = N_{it} / N_{gt} \quad (5.3.5)$$

From these compositions the API gravity of the stock tank oil, the separator gas gravity and gross heating value can be calculated. Formation volume factor,  $B_o$  and the gas/oil ratio,  $R_s$ , are calculated based on volumes of separator oil and gas and stock tank conditions (standard conditions) of the same.

## 6.0 LIMITATIONS

### 6.1 Physical Properties

Natural gas-properties are limited to:  $0.55 < \gamma_g < 2.0$

$$1.05 < T_{pr} < 3.0$$

$$0 < P_{pr} < 30$$

Liquid properties should be within:  $0.55 < \gamma_g < 1.4$

$$14.696 \text{ psia} < P < 6000 \text{ psia}$$

$$0.74 < \gamma_g < 1.0$$

$$20 \text{ Scf/bbl} < R_{so} < 2000 \text{ Scf/bbl}$$

$$0 \% < \text{Salt} < 25 \%$$

It is assumed that all dissolved solids are expressed as equivalent NaCl-concentration.

### 6.2 Orifice Limitations

For specifications regarding tolerances and restrictions to the orifices themselves, length of pipe preceding and following the orifice, and tapping points, please check the standards.<sup>1, 38</sup>

The beta ratio (orifice to tube diameter ratio) for **AGA** is limited as follows:

Flange taps:  $0.15 < \beta < 0.7$

Pipe taps:  $0.20 \leq \beta \leq 0.67$

The limitations on pipe diameters or nominal bore lines, D, and orifice diameters are:

$$1.667 \text{ in} \leq ID \leq 29.25 \text{ in}$$

$$2.0 \text{ in} \leq D \leq 30 \text{ in}$$

$$0.25 \text{ in} \leq d \leq 21.5 \text{ in}$$

Limitations of use for **ISO-5167** are:

Corner taps and D and D/2 taps:  $R_{eD} \geq 5000$  for  $0.1 \leq \beta \leq 0.56$

$$R_{eD} \geq 16000\beta^2 \text{ for } \beta > 0.56$$

Flange taps:  $R_{eD} \geq 5000$  and  $R_{eD} \geq 170\beta^2 D$

For all tapping types:  $d \geq 12.5 \text{ mm}$

$$0.1 \leq \beta \leq 0.75$$

$$50 \text{ mm} \leq D \leq 1000 \text{ mm}$$

## 6.3 Water - Natural Gas Phase Behavior

### 6.3.1 Hydrate Calculations

The following limitations apply to the hydrate calculations in this program:

- 1) The program is intended as a 'work-tool' in preliminary design.
- 2) Compositional calculations are limited to pressures between 100 and 2000 psia
- 3) The program should not be used where H<sub>2</sub>S concentration exceed 20 percent. Caution should be used whenever H<sub>2</sub>S-containing gases are present.
- 4) Large errors may be expected for gases rich in propane (> 30 %) or isobutane.
- 5) No phase calculation routines are built into the program. It is therefore recommended that the program **NOT** be used whenever composition changes may occur, i.e., multi-phase flow.
- 6) The program does not check if free water is present or not, a pre-requisite for hydrate formation. However, it is assumed that the gas is saturated with water (at its dew-point), and water will condense out before it is being treated.
- 7) The program does not account for the effects of (a) water salinity and (b) hydrate formation inhibitors, which also affect hydrate formation conditions.

### 6.3.2 Water Content Calculations

EMR approach is used for calculating critical temperatures and pressures. For heavy hydrocarbons (Moleweight > 200), the results are questionable. If EMR is used for calculating critical temperatures and pressures, results show that critical temperatures will fall with increasing molecular weight for hydrocarbons heavier than dodecane (C-12). The calculations should not be performed for gases with a C<sub>6</sub>+ specific gravity higher than about 0.8 (Water = 1.0).

Also note that results are questionable for gases with  $P_{pr} > 6.0$ , i.e. heavy gases or gases with large amount of condensed heavy hydrocarbons (C<sub>6</sub>+).

### 6.3.3 Inhibitor Injection Calculations

The inhibitor injection calculations use the hydrate calculation program and the water content calculation program to calculate the amount of inhibitor necessary to prevent hydrate formation in a pipeline. The limitations stated above (6.3.1 and 6.3.2) are also valid for inhibitor calculations.

In addition the program does not calculate the partitioning of inhibitor into the liquid hydrocarbon phase, only the gas phase and the liquid water phase. Normally, the amount of inhibitor in liquid hydrocarbons is small. The ratio of molefraction MeOH in liquid hydrocarbon over molefraction MeOH in aqueous phase will most likely be about .01 - .03 depending on the concentration of methanol in the liquid water phase. It is assumed that the highest temperature in a pipeline will occur at the highest pressure, and the lowest temperature will occur at the lowest pressure.

### 6.3.4 Solid CO<sub>2</sub> Formation

This program only gives an estimate of the concentration of CO<sub>2</sub> necessary to form as a solid. Detailed calculations should be carried out if results indicate operation in a range where solid CO<sub>2</sub> may form.

## 6.4 Separator Calculations

**GOsep** is limited to crude oils. It is also assumed that the crude oil is above the bubble point in the reservoir. Running the program with gas or gas/condensate compositions is not recommended, as the program may 'crash' or give erroneous results.

Also note that the equilibrium compositions calculated are based on Standing's equations which are restricted to temperatures ranging from 40 to 200 °F and pressures less than 1000 psia. Standing's equations have been tested by Glaso<sup>47</sup> for crude oils with pressures below 1,000 psia, and found that the general error from the equations are within ± 5%. These equations assume that the equilibrium ratios are independent of compositions.

## 7.0 REFERENCES

1. "Manual of Petroleum Measurement Standards Chapter 14 - Natural Gas Fluids Measurement", Section 3 - Concentric, Square-Edged Orifice Meters, Part 3 - Natural Gas Applications, API Standard 2530-1991, 3rd Ed., American Petroleum Institute, Washington, D.C. (1992).
2. Miller, R.W., "Flow Measurement Engineering Handbook", 2nd Ed., McGraw-Hill Publishing Co., New York, NY (1989) 2-30 - 2-37.
3. "Compressibility and Supercompressibility for Natural Gas and other Hydrocarbon Gases", Transmission Measurement Committee Report no. 8, AGA Catalog No. XQ 1285, Arlington, VA (1986).
4. "Phase Behavior", Reprint Series, SPE, Dallas (1981), 15, 52-63 and 211-223.
5. Edmister, W.C. and Lee, B.I., "Applied Hydrocarbon Thermodynamics", Vol. 1, 2nd Ed., Gulf Publishing Co., Houston, TX (1984) 47-49.
6. Dranchuk, P.M. and Abou-Kassem, J.H., "Calculation of Z-factors for Natural Gases Using Equation of State", J. of Canadian Pet. Tech. (July-Sept. 1975) 34-36.
7. Dempsey, J.R., "Computer Routine Treats Gas Viscosity as Variable", Oil & Gas J. (Aug. 16, 1965) 141.
8. Reid, R.C., Prausnitz, J.M., and Sherwood, T.K., "The Properties of Gases and Liquids", 3<sup>rd</sup> Ed., McGraw-Hill Book Co. Inc., New York, NY (1977) 503-506, and 536-537.
9. "Engineering Data Book", Vol II, 10th Ed., Gas Processors Suppliers Association, Tulsa, OK (1987) 13 - 42-44, 20 - 1-13, and 23 -1-65.
10. Buthod, P., "Specific Heats of Natural Gases", Oil and Gas J. (Sept. 29, 1949) 60.
11. Standing, M.B., "Volumetric and Phase Behavior of Oil Field Hydrocarbon Systems", SPE, Dallas (1977) 121-127.
12. Wichert, E. and Aziz, K., "Calculate Z's for Sour Gases", Hydrocarbon Processing (May 1971) 51, 119-122.
13. Vasquez, M. and Beggs, H.D., "Correlations for Fluid Physical Property Prediction", J. Pet. Tech. (June 1980) 968-970.
14. McCain Jr., W.D., "Reservoir-Fluid Property Correlations - State of the Art", SPE Reservoir Engineering (May 1991) 266-272.
15. Katz, D.L., "Prediction of Shrinkage of Crude Oils", Drill. and Prod. Prac., API (1940) 137-147.
16. Brill, J.P. and Beggs, H.D., "Two-Phase Flow in Pipes", 3rd Ed., Tulsa, OK (1979) 2-1- 2-67.
17. Dickey, D.S., "Practical Formulas Calculate Water Properties", Chemical Engineering (November 1991) 235-236.
18. Gallant, R.W., "Physical Properties of Hydrocarbons", 2nd Printing, Gulf Publishing Co., Houston, TX (1974) 61-72, 109-123.

19. McCain Jr., W.D., Rollins, J.B. and Villiena Lanzi, A.J., "The Coefficients of Isothermal Compressibility of Black Oils at Pressures Below the Bubblepoint", SPE Reservoir Engineering (Sept. 1988) 659-662.
20. Dodson, C.R. and Standing, M.B., "Pressure-Volume-Temperature and Solubility Relations for Natural Gas Water Mixtures", Drill. and Prod. Prac., API (1944), 173.
21. Osif, T.L., "The Effects of Salt, Gas, Temperature, and Pressure on the Compressibility of Water", SPE Reservoir Engineering (Feb. 1988) 175-181.
22. Eichelberger, W.C., Industry and Engineering Chemistry (1955) 47, 2223.
23. Glaso, O., "Generalized Pressure-Volume-Temperature Correlations", J. Pet. Tech. (May 1980) 785-795.
24. Keenan, J. A. and Keyes, F. G., "Thermodynamic Properties of Steam", John Wiley & Sons, Inc., New York, NY (1936).
25. Lasater, J.A., "Bubble Point Pressure Correlation", Trans. AIME (1958) 213, 379.
26. Craft, B.C., Hawkins, H.F. and Terry, R.E., "Applied Petroleum Reservoir Engineering", 2nd Ed., Prentice Hall, Englewood Cliffs (1991) 40-44.
27. Beggs, H.D. and Robinson, J.R., "Estimating the Viscosity of Crude Oil Systems", J. Pet. Tech. (Sept. 1975) 1140-1141.
28. Beal, C., "The Viscosity of Air, Water, Natural Gas, Crude Oil and Its Associated Gases at Oil-Field Temperatures and Pressures", Trans. AIME (1946) 165, 94-115.
29. Chew, J.N. and Connally Jr., C.A., "A Viscosity Correlation for Gas-Saturated Crude Oils", Trans. AIME (1959) 216, 23-25.
30. van Wingen, N., "Viscosity of Air, Water, Natural Gas, and Crude Oil at Varying Pressures and Temperatures", Secondary Recovery of Oil in the United States, API (1950) Ch. 6.
31. "API Technical Data Book - Petroleum Refining", Vol. I, 4th Ed., American Petroleum Institute, Washington DC (1983) 9 - 141-153.
32. Raznjevic, K., "Handbook of Thermodynamic Tables and Charts", Hemisphere Publishing Co., Washington DC (1976) 99-102.
33. Baker, O. and Swerdloff, W., "Finding Surface Tension of Hydrocarbon Liquids", Oil & Gas J. (January 2, 1956) 125.
34. Riazi, M.R. and Daubert, T.E., "Simplify Property Predictions", Hydrocarbon Processing (March 1980) 115-116.
35. Campbell, J.M., "Gas Conditioning and Processing", Vol. I, 6th Ed., Campbell Petroleum Series, Norman, OK (1984) 56-51, 60, and 137-171.
36. Katz, D.L., et. al., "Handbook of Natural Gas Engineering", McGraw-Hill Book Co., Inc., New York (1959).
37. Perry, R.H. and Chilton, C.H., "Chemical Engineers Handbook", 5th Ed., McGraw-Hill Book Co., New York, NY (1973) 3-135.

38. "Measurement of fluid flow by means of pressure differential devices inserted in circular-cross section conduits running full - Part 2: Orifice plates", International Standard ISO 5167-2: 2003, 1st Ed., International Organization for Standardization, Geneva, Switzerland (2003).
39. Tohidi, B., Danesh, A., Todd, A. C., and Burgass, R. W.: "Measurement and Prediction of Hydrate-Phase Equilibria for Reservoir Fluids", SPE Production & Facilities (May 1996) 69-76.
40. Robinson, D.B., and Ng, H.J.: "Improve hydrate predictions," Hydrocarbon Processing (Dec. 1975) 95-96.
41. Poettmann, F.H.: "Here's Butane Hydrates Equilibria," Hydrocarbon Processing (June 1984) 111-112.
42. Berge, B.K.: "Hydrate Predictions on a Microcomputer," paper SPE 15306 presented at the 1986 Symposium on Petroleum Industry Applications of Microcomputers.
43. Corrigan, A., Duncim, S. N., Edwards, A. R., and Osborne, C. G.: "Trials of Treshold Hydrate Inhibitors in the Ravenspurn to Cleeton Line", SPE Production & Facilities (November 1996) 250-255.
44. Notz, P.K., BumGardner, S.B., Schaneman, B.D., and Todd, J.L.: "Application of Kinetic Inhibitors to Gas Hydrate Problems", SPE Production & Facilities (November 1996) 256-260.
45. Standing, M.B., "A Set of Equations for Computing Equilibrium Ratio of a Crude Oil/Natural Gas System at Pressures Below 1,000 psia", J.Pet.Tech (September 1979) 1193-1195.
46. Katz, D.L. and Hachmuth, K.H., "Vaporization Equilibrium Constants in a Crude-oil Natural Gas System", Ind.Eng. Chem. (1937) 29, 1072.
47. Glaso, O. S. and Whitson, C.H. : "The Accuracy of PVT Parameters Calculated From Computer Flash Separation at Pressures Less Than 1,000 psia", JPT Forum (August 1982) 1811-1813.