**BEST TIPS OF THE MONTH**

**Featured TOTMS**

- [Gas-Liquid Separators Sizing Parameter](#)
- [Correlations for Conversion between True and Reid Vapor Pressures](#)
- [Benefits of Standby Time in Adsorption Dehydration Process](#)
- [Effect of Relative Density (Specific Gravity) on the Saturated Water Content of Sweet Natural Gases](#)

**Top 10 Tips of the Month**

1. Benefits of Standby Time in Adsorption Dehydration Process
   By Harvey Malino

2. Charts and Correlations for Estimating Methanol Removal in the Gas Sweetening Process
   By Dr. Mahmood Moshfeghian

3. Gas-Liquid Separators Sizing Parameter
   By Dr. Mahmood Moshfeghian

4. Correlations for Conversion between True and Reid Vapor Pressures
   By Dr. Mahmood Moshfeghian

5. Effect of Relative Density (Specific Gravity) on the Saturated Water Content of Sweet Natural Gases
   By Dr. Mahmood Moshfeghian

6. Finding Opportunities Checklist
   By James F. Langer, P.E.

7. Impact of Gas-Oil Ratio (GOR) on Crude Oil Pressure Drop in Gathering Systems
   By Dr. Mahmood Moshfeghian

8. Infinite Recycle Impacts on Compression Systems
   By James F. Langer, P.E.

9. Oil Dehydration Error Found in Most Designs Worldwide
   By James F. Langer, P.E.

10. What is the Impact of Light Hydrocarbons on the Natural Gas Hydrate Formation Conditions?
    By Dr. Mahmood Moshfeghian
Molecular sieves are used upstream of turbo expander units and LNG facilities to dehydrate natural gas to <0.1 ppmv water content. In the natural gas industry, the molecular sieves employ heat to drive off the adsorbed water. Figure 1 shows a typical flow schematic for a 2 tower system; Figure 2 shows a 3 tower system.

The cyclical heating/cooling of the adsorbent results in a capacity (kg water/100 kg adsorbent; lbm water/100 lbm adsorbent) decline due to a gradual loss of crystalline structure and/or pore closure. A more troublesome cause of capacity decline is contamination of the molecular sieves due to liquid carryover from the upstream separation equipment.

Figure 3 shows a generic molecular sieve capacity decline curve. A few important observations can be made from Figure 3:

The life of the adsorbent is a function of the number of cycles, not the elapsed calendar time.

The capacity decline is steep at the beginning but gradually flattens out. This assumes no step-change events such as NGL, glycol, and/or liquid amine carryover, bed support failure, etc.

Shown in this figure are “Good”, “Average” and “Poor” curves that are a function of site specific factors.

Locating one data point on Figure 3 from a performance test allows you to extrapolate the decline curve of the unit in question.

If your regeneration circuit has excess capacity over the “normal design conditions”, i.e., a design factor, you have standby time. This excess capacity allows you to reduce your online adsorption time and “turn the beds around” faster by regenerating the beds in a shorter cycle time. When you are involved in the design of an adsorption unit, it is recommended to add 10 – 20% excess regeneration capacity. Because the capacity decline curves flatten out, available standby time may be able to extend the life of a molecular sieve unit when your unit is operating on fixed cycle times. Other operating options include: running each cycle to water breakthrough; and, reducing the cycle times in discreet steps throughout the life of the adsorbent.

To illustrate the benefits of standby time, consider the following case study. A natural gas processing plant has commissioned a new 3 tower molecular sieve dehydration unit to process 11.3 x 10^6 std m3/d (400 MMscfd) prior to flowing to a deep ethane recovery unit. The unit is expected to run for 3 years before needing a recharge and the plant turnaround is based on this expectation.
Figure 1. Typical process flow diagram for a 2-tower adsorption dehydration system [1]
Figure 2. Typical process flow diagram for a 3-tower adsorption dehydration system [1]
Benefits of Standby Time in Adsorption Dehydration Process

Figure 3. A generic molecular sieve decline curves [1]
The following assumptions are made:

- 3 tower system (2 towers on adsorption, 1 on regeneration)
- External Insulation
- Tower ID = 2.9 m (9.5ft)
- Each tower contains 24630 kg [54300 lbm] of Type 4A 4×8 mesh beads
- Regeneration circuit capable of handling an extra 15% of flow
- Unit is operated on fixed time cycles
- No step-change events such as liquid carryover, poor flow distribution, etc.

The design basis and molecular sieve design summary are shown in Tables 1 and 2. The additional 15% of flow from the regeneration gas heater is well below the point at which bed lifting will occur.

The calculations presented here are valid for low pressure regeneration (less than 4100 kPaa (600 psia)). Using the concepts outlined in Chapter 18 of Gas Conditioning and Processing, Volume 2 [1]: The Equipment Modules (9th Edition) we find a design life factor, FL, of 0.6 after 3 years (1095 cycles) of operation at design conditions. This point lies slightly above the “average” life curve as seen in Figure 4.

After 12 months of operation, a Performance Test Run (PTR) is conducted. The results are shown in Table 3. The feed flow rate and temperature are slightly lower compared to the design values. A water breakthrough time of 20.9 hours is recorded. The FL is determined (using the concepts in Chapter 18) to be 0.68 after 365 cycles (one year of operation). It is important and useful to understand the equation sequence of the concepts in Chapter 18, as shown by Equations 18.5 through 18.10 to arrive at the cited value for FL. This data point is shown in Figure 5 and is seen to lie just below the generic “Average” curve. Note that the slope of the curves are starting to flatten out. Since the PTR FL is lower than the Design FL, the molecular sieves will experience water breakthrough if operated at design conditions in less than three years. Figure 6 shows the projected life factor, FL, after 3 years of service at design conditions. If the capacity decline continues to follow the same trend as seen from the PTR, water breakthrough will occur after 750 cycles or just a little over 2 years from startup if operation continues at design conditions. This is shown in Fig 7.

Because the unit has a regeneration circuit that can handle an additional 15% of flow, the complete regeneration cycle (heating, cooling, de- and re-pressurization) can be reduced to 7.0 hours. This allows the beds to turn around faster.

Using the reduced cycle time (the complete cycle time is now 21 hours vs the original 24 hours), we find an FL = 0.53. This is because less water is being adsorbed per cycle. This occurs at around the 1500 cycle mark as shown in Figure 8.

If the plant elects to take advantage of the standby time and operate at reduced cycle time immediately following the PTR, the molecular sieves should last an additional 2.7 years, resulting in a total life of 3.7 years. In this case, standby time will allow the unit to operate until the scheduled plant turnaround.
Benefits of Standby Time in Adsorption Dehydration Process

Table 1. Design basis for the case study

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Feed</th>
<th>Regen Heat</th>
<th>Regen Cool</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flow rate, $10^6$ x std m$^3$/d (MMscfd)</td>
<td>11.3 (400)</td>
<td>0.71 (25.7)</td>
<td>0.71 (25.7)</td>
</tr>
<tr>
<td>Pressure, kPaa (psia)</td>
<td>6205 (900)</td>
<td>2068 (300)</td>
<td>2068 (300)</td>
</tr>
<tr>
<td>Temperature, °C (°F)</td>
<td>30 (86)</td>
<td>288 (500)</td>
<td>30 (86)</td>
</tr>
<tr>
<td>Molecular Weight</td>
<td>20.3</td>
<td>17.1</td>
<td>17.1</td>
</tr>
<tr>
<td>Water content</td>
<td>Sat'd</td>
<td>&lt;0.1 ppmv</td>
<td>&lt;0.1 ppmv</td>
</tr>
</tbody>
</table>

Table 2. Design Summary for the case study

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Adsorption</th>
<th>Heat</th>
<th>Cool</th>
<th>Depresssure</th>
<th>Represssure</th>
</tr>
</thead>
<tbody>
<tr>
<td>Time (hrs/tower)</td>
<td>16</td>
<td>4.88</td>
<td>2.62</td>
<td>0.2</td>
<td>0.2</td>
</tr>
<tr>
<td>Flow Direction</td>
<td>Down</td>
<td>Up</td>
<td>Up</td>
<td>Down</td>
<td>Up</td>
</tr>
<tr>
<td>Pressure Drop, kPaa (psi)</td>
<td>&lt;50 (7)</td>
<td>&lt;7 (1)</td>
<td>&lt;7 (1)</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

Table 3. Results of Performance Test Run (PTR) after 12 months of operation

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Feed</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flow Rate $10^6$ x std m$^3$/d (MMscfd)</td>
<td>10.9 (385)</td>
</tr>
<tr>
<td>Pressure, kPaa (psia)</td>
<td>6205 (900)</td>
</tr>
<tr>
<td>Temperature, °C (°F)</td>
<td>28 (82)</td>
</tr>
<tr>
<td>Molecular Weight</td>
<td>20.3</td>
</tr>
<tr>
<td>Water content</td>
<td>Sat'd</td>
</tr>
<tr>
<td>Breakthrough time (h)</td>
<td>20.9</td>
</tr>
</tbody>
</table>
Benefits of Standby Time in Adsorption Dehydration Process

Figure 4. Design condition life factor [1]
Benefits of Standby Time in Adsorption Dehydration Process

Figure 5.
Performance test run (PTR) life factor [1]

Figure 6.
Projected life factor (red triangle) running at design conditions [1]
Benefits of Standby Time in Adsorption Dehydration Process

Figure 7. Projected life factor running at design conditions [1]

Figure 8. Projected life factor (red triangle) if standby time is used [1]
Conclusions

The methods presented allow the user to estimate the decline of their adsorbent based on only one performance test run for molecular sieve dehydrators using low pressure regeneration. This permits early formulation of a credible action plan.

Site-specific factors will determine your unit’s decline curve. Consequently, conducting more than one performance test is highly recommended. A poorly performing inlet separator, for example, could result in a unit exhibiting a more pronounced decline than indicated by the generic curves in Figure 3.

Standby time offers a large degree of operating flexibility because the decline curves tend to level off; always try to build in standby time in any new molecular sieve design.

Adsorption capacity is a function of the number of cycles, not calendar time.

Install a good filter coalescer or filter separator upstream of your adsorption unit to keep the contaminants out of the system.

The approach discussed in this Tip of the Month should help a facility engineer plan for the inevitable replacement of molecular sieves in their natural gas dehydration facility.

To learn more about similar cases and how to minimize operational problems, we suggest attending our G4 (Gas Conditioning and Processing) and PF4 (Oil Production and Processing Facilities) courses.

By: Harvey M. Malino

Reference

The gas-sweetening process by amines like methyl-diethanolamine (MDEA) removes a considerable amount of methanol from a sour gas stream. Moreover, if the methanol content of the sour gas is high, the sweet gas may still retain high methanol content and can cause operational troubles in the downstream processes. Provisions of purging reflux (Water Draw) of the regenerator column and its replacement with “Fresh Water” can improve methanol recovery [1, 2].

The July 2016 tip of the month (TOTM) considered the presence of methanol in the sour gas stream and determined the quantitative traces of methanol ending up in the sweet gas, flash gas and acid gas streams [2]. It simulated a simplified MDEA gas-sweetening unit by computer and studied the effect of sour gas methanol content, and the rate of replacing condensed reflux with fresh water on the sweet gas methanol content. For the sour gas temperature of 43.3 and 32.2 °C (110 and 90 °F) the tip studied three inlet gas methanol contents of 50, 250, and 500 PPM on mole basis. In each case the tip varied the rate of fresh water replacement from 0 to 100 % by an increment of 20%.

The methanol removal efficiency (MRE) on the volume basis is defined by:

\[
\text{MRE, } \% = 100 \frac{(\text{Methanol PPMv in Sour Gas}) - (\text{Methanol PPMv in Sweet Gas})}{(\text{Methanol PPMv in Sour Gas})}
\]

Table 1 presents the summary of calculated methanol removal efficiency (MRE) based on the simulation results of the July 2016 TOTM [2].

Table 1. The effect of purging and sour gas temperature on methanol removal efficiency [2]

<table>
<thead>
<tr>
<th>Temperature, °C (°F)</th>
<th>43.3 (110)</th>
<th>32.2 (90)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0 % Purge (Basic MDEA Process)</td>
<td>88.8</td>
<td>92.6</td>
</tr>
<tr>
<td>100 % Purge (Modified MDEA Process)</td>
<td>92.4</td>
<td>94.4</td>
</tr>
<tr>
<td>Improvement in Methanol Removal Efficiency</td>
<td>3.6</td>
<td>1.8</td>
</tr>
</tbody>
</table>

In continuation of the July 2016 TOTM, this tip will consider the presence of methanol in the sour gas stream and determine the quantitative traces of methanol ending up in the sweet gas, flash gas and acid gas streams. This tip simulates a simplified MDEA gas-sweetening unit by computer simulation [3, 4]. This tip also studies the effect of sour gas methanol content, temperature and the rate of replacing condensed reflux with fresh water on the sweet gas methanol content.

For the sour gas temperatures of 43.3, 32.2 and 21.1 °C (110, 90, and 70 °F) the tip studies three inlet sour gas methanol contents of 50, 250, 500 PPM on mole basis. In each case the tip varies the rate of fresh water replacement from 0 to 100 % by an increment of 20%. Similar to the September 2016 TOTM [5] and based on the computer simulation results, the tip develops simple charts and correlations to estimate the methanol removal efficiency under various operating conditions. These charts and correlations are accurate enough for facilities calculations.
Case Study

For the purpose of illustration, this tip considers sweetening of a sour gas stream saturated with water using the basic and modified MDEA processes as described in the July 2016 TOTM [2]. In addition to the two sour gas temperatures reported in the July TOTM, this tip also considers a sour gas temperature of 21.1 °C (70 °F). Table 2 presents its composition on the dry basis, gas standard volume rate, pressure, and temperatures. This tip uses ProMax [6] simulation software with the “Amine Sweetening – PR” property package to perform all of the simulations.

Results and Discussions

Figures 1 through 3 present the calculated MRE as a function of the reflux rate replacement (RRR) with fresh water for the sour gas temperatures of 43.3, 32.2, and 21.1 °C (110, 90, and 70 °F), respectively. Each figure presents MRE vs replacement rate for the three sour gas methanol contents (50, 250, and 500 PPMV).

Table 2. Feed composition on the dry basis, volumetric flow rate and conditions [2]

<table>
<thead>
<tr>
<th>Component</th>
<th>Mol %</th>
</tr>
</thead>
<tbody>
<tr>
<td>H₂S</td>
<td>0.10</td>
</tr>
<tr>
<td>CO₂</td>
<td>5.00</td>
</tr>
<tr>
<td>Methane</td>
<td>88.40</td>
</tr>
<tr>
<td>Ethane</td>
<td>5.00</td>
</tr>
<tr>
<td>Propane</td>
<td>1.00</td>
</tr>
<tr>
<td>n-Butane</td>
<td>0.50</td>
</tr>
<tr>
<td>MeOH, three different values</td>
<td>0.005, 0.025, 0.05</td>
</tr>
<tr>
<td>MeOH, three different values, PPMV</td>
<td>50, 250, 500</td>
</tr>
<tr>
<td>Volume Rate, 10⁶ Sm³/d (MMscfd)</td>
<td>2.837(100.2)</td>
</tr>
<tr>
<td>Temperature, °C (°F)</td>
<td>43.3, 32.2, 21.1 (110, 90, 70)</td>
</tr>
<tr>
<td>Pressure, kPaa (psia)</td>
<td>5617 (814.7)</td>
</tr>
</tbody>
</table>
Charts and Correlations for Estimating Methanol Removal in the Gas Sweetening Process

Figure 1. Methanol removal efficiency vs reflux replacement for sour gas temperature of 43.3 °C (110 °F)

Figure 2. Methanol removal efficiency vs reflux replacement for sour gas temperature of 32.2 °C (90 °F)
Figure 3. Methanol removal efficiency vs reflux replacement for sour gas temperature of 21.1 °C (70 °F)

Figure 4. Average methanol removal efficiency vs reflux replacement
Since the three curves for different sour gas methanol contents on each figure are close, the effect of the sour gas methanol content on MRE can be neglected. For each sour gas temperature, the calculated arithmetic average of MRE of the three sour gas methanol content are provided in Figure 4. This figure indicates that as the sour gas temperatures decreases the impact of the reflux rate replacement with fresh water diminishes.

A non-linear regression program was used to determine the parameters of the following correlation for the methanol removal efficiency as a function of the reflux rate replacement % (RRR).

MRE,% = A + B (RRR)^c

Where:

MRE = Methanol Removal Efficiency on the mole basis, %
RRR = Reflux Rate Replacement %

A generalized form of this correlation to cover the temperature effect can be expressed as:

MRE,% = (A1 + A2 T) + (B1 + B2 T)(RRR)(C1+C2T)

Where:

MRE% = Methanol removal efficiency on the weight basis
RRR = Reflux Rate Replacement %
T = Temperature, °C (°F)

Table 4 presents the regressed parameters of A1, A2, B1, B2, C1 and C2 of Equation 2 for temperatures in °C and °F. Similarly, the last two rows in Table 4 present the AAPE and the MAPE, respectively.

Table 4. Parameters of Equation 1 for methanol removal efficiency

<table>
<thead>
<tr>
<th>Parameter</th>
<th>T in °C</th>
<th>T in °F</th>
</tr>
</thead>
<tbody>
<tr>
<td>A1</td>
<td>101.740</td>
<td>107.017</td>
</tr>
<tr>
<td>A2</td>
<td>-0.2973</td>
<td>-0.1650</td>
</tr>
<tr>
<td>B1</td>
<td>0.0393</td>
<td>-0.1478</td>
</tr>
<tr>
<td>B2</td>
<td>0.0105</td>
<td>0.0059</td>
</tr>
<tr>
<td>C1</td>
<td>0.3506</td>
<td>0.3163</td>
</tr>
<tr>
<td>C2</td>
<td>0.00194</td>
<td>0.00108</td>
</tr>
<tr>
<td>AAPE</td>
<td>0.06</td>
<td>0.06</td>
</tr>
<tr>
<td>MAPE</td>
<td>0.17</td>
<td>0.16</td>
</tr>
</tbody>
</table>

Table 3. Parameters of Equation 1 for methanol removal efficiency

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Sour Gas Temperature, °C (°F)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>43.3 (110)</td>
</tr>
<tr>
<td>A</td>
<td>88.7872</td>
</tr>
<tr>
<td>B</td>
<td>0.5115</td>
</tr>
<tr>
<td>C</td>
<td>0.4306</td>
</tr>
<tr>
<td>AAPE</td>
<td>0.08</td>
</tr>
<tr>
<td>MAPE</td>
<td>0.14</td>
</tr>
</tbody>
</table>

AAPE = Average Absolute Percent Error
MAPE = Maximum Absolute Percent Error
The MRE predictions by Equation 2 were added to Figure 4 and is presented as Figure 5. In this figure the solid lines present the MRE prediction by Equation 2 and dashed lines with the filled symbols represents simulation results. The analysis of Figures 5 and the calculated values of AAPE and MAPE in Table 4 indicate that accuracy of the proposed correlations, compared to the simulation results, is very good for estimation of methanol removal efficiency (MRE).

Figure 5. Comparison of model prediction of average methanol removal efficiency vs reflux replacement
Conclusions

Based on the results obtained for the considered case study, this TOTM presents the following conclusions:

The impact of the sour gas methanol content on the methanol removal efficiency is small (Figures 1-3), overall only a minor impact, less than 0.5 % point.

As the sour gas temperature decreases, the methanol removal efficiency increases (Figures 1-5), overall only minor impact, less than 3 % points.

Methanol removal efficiency with MDEA sweetening can remove only 89-97% of the methanol in the sour gas feed. This may still leave more methanol than the gas spec allows. A separate water wash step may be required. The fresh water used for the water wash could be recycled as MDEA reflux purge make-up.

The tip presents three simple charts (Figures 1-3) and two correlations (Equations 1 and 2) along with their parameters (Tables 3 and 4) for estimating the average methanol removal efficiencies for the sour gas temperatures of 43.3, 32.2, and 21.1 °C (110, 90, and 70 °F), respectively.

Compared to the rigorous computer simulation, the accuracy of the proposed correlations (Equations 1 and 2) to estimate the average methanol removal efficiency is very good (Tables 3 and 4 and Figure 5) and can be used for facilities calculations.

The proposed correlations (Equations 1 and 2) and charts (Figures 4-5) are easy to use.

To learn more about similar cases and how to minimize operational troubles, we suggest attending our G6 (Gas Treating and Sulfur Recovery), G4 (Gas Conditioning and Processing), G5 (Advanced Applications in Gas Processing), and PF4 (Oil Production and Processing Facilities) courses.

PetroSkills | Campbell offers consulting expertise on this subject and many others. For more information about these services, visit our website at petroskills.com/consulting, or email us at consulting@PetroSkills.com.

By: Dr. Mahmood Moshfeghian
Accurate measurement and prediction of crude oil and natural gas liquid (NGL) products vapor pressure are important for safe storage and transportation, custody transfer, minimizing vaporization losses and environmental protection. Vapor pressure specifications are typically stated in Reid Vapor Pressure (RVP) or True Vapor Pressure (TVP). In addition to the standard procedures for their measurements, there are rigorous and shortcut methods for their estimation and conversion.

There are figures and monographs for conversion of RVP to TVP for NGLs (Natural Gas Liquids) and crude oil at a specified temperature. This tip will present simple correlations for conversion from RVP to TVP and vice versa at a specified temperature. The correlations are easy to use for hand or spreadsheet calculations. Figures generated using these correlations will be presented, too.

Chapter 5 of reference [1] presents an excellent overview of TVP and RVP including their definitions, standard procedures for their measurements and diagrams for their conversion. The proceeding two paragraphs are extracted with minor revisions from reference [1].

TVP is the actual vapor pressure of a liquid product at a specified temperature and is measured with a sample cylinder. TVP specifications must always be referenced to a temperature, which frequently falls between 30-50°C (86-122°F). TVP is difficult to measure and depends on the ratio of the vapor to liquid, V/L, in the measurement device. If V/L = 0, the vapor pressure is essentially equivalent to the bubble point of the mixture which is the highest vapor pressure value for the liquid. As V/L increases, i.e., a small amount of vapor exists at the point of measurement, the measured vapor pressure will decrease. The relationship between the measured TVP and V/L depends on the composition of the mixture. For “near pure” component mixtures, V/L has little effect on the measured vapor pressure. For mixtures with a large composition range, such as crude oil or condensate, the effect of V/L on the measured vapor pressure can be significant (See ASTM D 6377 – 10 for detail). Reference [1] lists the Standards used for TVP measurements.

Because of the difficulty in accurately measuring TVP, an alternative method of measuring vapor pressure is frequently used. This is the RVP. The RVP is a standard test set out in ISO 3007:1999, Petroleum products and crude petroleum – Determination of vapour pressure – Reid method. Another standard that applies to RVP is: ASTM D323 – 08, Standard Test Method for Vapor Pressure of Petroleum Products (Reid Method). The RVP test is applicable for crude oils, condensates, and petroleum products such as gasoline (petrol) mixtures. A liquid sample is collected in the lower 20% chamber (see Fig 5.12 of reference [1]). The 80% chamber, which is filled with air (may also contain a small amount of water vapor) at atmospheric pressure. Both chambers are cooled to 0°C (32°F). The 80% air chamber (at atmospheric pressure) is then connected to the 20% liquid chamber. The connecting valve is opened and the cylinder is heated in a water bath to 37.8°C (100°F). The pressure indicated on the gauge is the RVP.

Reference [1] provides Figure 5.14 for conversion RVP to TVP for motor gasoline (petrol) and natural gasoline (C5+NGLs) at various temperatures. Figure 5.15 of reference [1] is a nomograph that shows the approximate relationship between RVP and TVP for crude oil. It is commonly used for converting RVP to TVP at custody transfer points where the vapor pressure specification for the oil is a TVP, but the actual vapor pressure measurement is an RVP.

Vazquez-Esparragoza et al. [2] present an algorithm to calculate RVP without performing the actual test. The algorithm, based on an air-and-water free model, uses the Gas Processors Association Soave-Redlich-Kwong [3] equation of state and assumes liquid and gas volumes are additive. This algorithm can be used to predict RVP of any hydrocarbon mixture of known composition. Since the calculations are iterative, it should be incorporated into a general purpose process simulator. Vazquez-Esparragoza et al. [2] reported good agreement between predicted and experimental values.
Riazi et al. [4] presented a new correlation for predicting the RVP of gasoline and naphtha based on a TVP correlation. The input parameters for this correlation are the mid-boiling point, specific gravity, critical temperature, and critical pressure, where the critical properties may be estimated from the boiling point and specific gravity critical temperature, and critical pressure, where the critical properties may be estimated from the boiling point and specific gravity using available methods. They evaluated their proposed correlation with data collected on 50 gasoline samples from crude oils from around the world with API gravity ranges from 41 to 87, average boiling point ranges from 43 to 221 °C (110 to 430 °F) and RVP of 0.7 to 115 kPa (0.1–17 psi). The average error from their proposed correlation is about 6 kPa (0.88 psi) [4].

Development of Model for Motor Gasoline and Natural Gasoline

To develop the desired correlations for conversion of motor gasoline and natural gasoline RVP to TVP and vice versa, this tip generated 127 data points from Figure 5.14 of reference [1]. These data covered the full ranges of temperature, RVP and TVP of Figure 5.14.

RVP to TVP: This tip used these data to determine the parameters to the Equations 1 through 3. The API 2517 originally reported the same forms of equations for crude oil.

\[
A = A_1 - A_2 \ln(RVP) \quad (1a)
\]
\[
B = B_1 - B_2 \ln(RVP) \quad (2a)
\]
\[
TVP = \exp \left( A - \frac{B}{T + C} \right) \quad (3a)
\]

Similarly this tip TVP to RVP: propose the following equations for conversion from TVP to RVP.

\[
A = A_1 - A_2 \ln(TVP) \quad (1b)
\]
\[
B = B_1 - B_2 \ln(TVP) \quad (2b)
\]
\[
RVP = \exp \left( A - \frac{B}{T + C} \right) \quad (3b)
\]

Where:

- \( T \) = Temperature, °C (°F)
- \( RVP \) = Reid Vapor Pressure, kPa (psi)
- \( TVP \) = True Vapor Pressure, kPaa (psia)

Note that the values of \( A_1, A_2, B_1, \) and \( B_2 \) are different in the above two sets of equations. The value of “C” is a function of the chosen units (SI versus FPS) and is consistent.

Table 1 presents the optimized values of \( A_1, A_2, B_1, B_2, \) and C for three sets of data in FPS (Foot-Pound-Second) and SI (System International). The data set labeled “All” included 127 data points covering all of the data of combined motor gasoline and natural gasoline. The data set “Motor Gasoline” and “Natural Gasoline” cover 76 and 51 data points for motor gasoline and natural gasoline, respectively. Table 1 also presents the Average Absolute Percent Deviation (AAPD), the Maximum Absolute Percent Deviation (MAPD), Average Absolute Deviation (AAD), and the number of data points (NP) for each data set. The error analysis of Table 1 indicates that the accuracy of the proposed correlations is good. Their accuracy is as good as the quality of tabular data generated from Figure 5.14.
Table 1. The optimized parameters for motor gasoline and natural gasoline

<table>
<thead>
<tr>
<th>Parameters</th>
<th>FPS</th>
<th>TVP Calculation</th>
<th>RVP Calculations</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>ALL</td>
<td>Motor Gasoline</td>
<td>Natural Gasoline</td>
</tr>
<tr>
<td>A₁</td>
<td>9.4674</td>
<td>9.4286</td>
<td>9.4964</td>
</tr>
<tr>
<td>A₂</td>
<td>-0.9445</td>
<td>-0.9658</td>
<td>-0.9324</td>
</tr>
<tr>
<td>B₁</td>
<td>5211.0</td>
<td>5217.6</td>
<td>5172.2</td>
</tr>
<tr>
<td>B₂</td>
<td>16.014</td>
<td>17.820</td>
<td>5.359</td>
</tr>
<tr>
<td>C</td>
<td>459.67</td>
<td>459.67</td>
<td>459.67</td>
</tr>
<tr>
<td>AAPD</td>
<td>1.23</td>
<td>1.06</td>
<td>0.98</td>
</tr>
<tr>
<td>MAPD</td>
<td>4.00</td>
<td>3.46</td>
<td>3.15</td>
</tr>
<tr>
<td>AAD, psi</td>
<td>0.271</td>
<td>0.135</td>
<td>0.333</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>SI</th>
<th>ALL</th>
<th>Motor Gasoline</th>
<th>Natural Gasoline</th>
</tr>
</thead>
<tbody>
<tr>
<td>A₁</td>
<td>9.5745</td>
<td>9.4947</td>
<td>9.6269</td>
</tr>
<tr>
<td>A₂</td>
<td>-0.9445</td>
<td>-0.9658</td>
<td>-0.9324</td>
</tr>
<tr>
<td>B₁</td>
<td>2912.19</td>
<td>2917.76</td>
<td>2879.21</td>
</tr>
<tr>
<td>B₂</td>
<td>8.8969</td>
<td>9.001</td>
<td>2.9768</td>
</tr>
<tr>
<td>C</td>
<td>273.15</td>
<td>273.15</td>
<td>273.15</td>
</tr>
<tr>
<td>AAPD¹</td>
<td>1.23</td>
<td>1.06</td>
<td>0.98</td>
</tr>
<tr>
<td>MAPD²</td>
<td>4.00</td>
<td>3.46</td>
<td>3.15</td>
</tr>
<tr>
<td>AAD³, kPa</td>
<td>1.867</td>
<td>0.930</td>
<td>2.300</td>
</tr>
<tr>
<td>NP⁴</td>
<td>127</td>
<td>76</td>
<td>51</td>
</tr>
</tbody>
</table>

¹AAPD = Avg Absolute Percent Deviation
²MAPD = Maximum Absolute Percent Deviation
³AAD = Average Absolute Deviation
⁴NP = Number of data Points (NP)
Correlations for Conversion between True and Reid Vapor Pressures

Figure 1a. TVP as a function of RVP and temperature for motor gasoline and natural gasoline (typical C5+ NGLs)
Figure 1b. TVP as a function of RVP and temperature for motor gasoline and natural gasoline (typical C5+ NGLs)
Correlations for Conversion between True and Reid Vapor Pressures

Figures 1a (SI) and 1b (FPS) present the tabular data (generated from Figure 5.14 of reference [1]) in the form of "legends". This figure also presents the predicted TVP by the proposed correlations (Equations 1 through 3 and the corresponding parameters listed in Table 1) as a function of temperature and RVP in the form of "continuous lines" and "broken lines" for motor gasoline and natural gasoline, respectively.

Development of Model for Crude Oil

To develop the desired correlations for conversion of crude oil RVP to TVP and vice versa, this tip generated 196 data points for Figure 5.15 of reference [1] using the correlations reported in API 2517. The API 2517 equations are the same as shown in Equations 1 through 3. This data covered the full ranges of temperature, RVP and TVP of Figure 5.15. For Equations 1a, 2a, and 3a (T, °F, RVP, psi and TVP, °C, and B2, B1, A2, A1 Table 2 presents the API 2517 FPS parameters of RVP to TVP:psia). This tip determined the SI (T, °C, RVP, kPa and TVP, kPaa) corresponding parameters. Table 2 also presents the SI parameters.

Table 2. API 2517 parameters for crude oil (For use with Equations 1a, and TVP calculations 2a and 3a)

<table>
<thead>
<tr>
<th>Parameter</th>
<th>FPS (API 2517)</th>
<th>SI (This Tip)</th>
</tr>
</thead>
<tbody>
<tr>
<td>A1</td>
<td>12.82</td>
<td>16.62</td>
</tr>
<tr>
<td>A2</td>
<td>0.9672</td>
<td>0.9675</td>
</tr>
<tr>
<td>B1</td>
<td>7261</td>
<td>5339</td>
</tr>
<tr>
<td>B2</td>
<td>1216</td>
<td>675.7</td>
</tr>
<tr>
<td>C</td>
<td>459.67</td>
<td>273.15</td>
</tr>
</tbody>
</table>

Figures 2a (SI) and 2b (FPS) present the predicted TVP by Equations 1a, 2a and 3a and the corresponding parameters listed in Table 2 as a function of temperature and RVP for crude oil.

TVP to RVP: To cover the full ranges of Figure 5.15 of reference [1], this tip proposes the following equations for conversion of TVP to RVP.

\[
A = A_1 - A_2 \ln(TVP) - A_3 (T + C)
\]

\[
B = B_1 - B_2 \ln(TVP) - B_3 \left[ \ln(TVP) \right]^2
\]

\[
RVP = \exp \left( A - \frac{B}{T + C} \right)
\]

(1c)
(2c)
(3c)

Where:

\[ T \text{ = Temperature, °C (°F)} \]
\[ RVP \text{ = Reid Vapor Pressure, kPa (psi)} \]
\[ TVP \text{ = True Vapor Pressure, kPaa (psia)} \]

Table 3 presents the optimized values of A1, A2, A3, B1, B2, B3, and C for RVP calculation in FPS and SI for the proposed Equations 1c through 3c. Table 3 also presents the Average Absolute Percent Deviation (AAPD), the Maximum Absolute Percent Deviation (MAPD), Average Absolute Deviation (AAD), and the Number of data Points (NP). The error analysis of Table 3 indicates that the accuracy of the proposed correlations is good and can be used for the estimation purposes.
**Correlations for Conversion between True and Reid Vapor Pressures**

**Table 3.** The optimized parameters for crude oil TVP conversion to RVP (For use with Equations 1c, 2c & 3c)

<table>
<thead>
<tr>
<th>Parameter</th>
<th>FPS</th>
<th>SI</th>
</tr>
</thead>
<tbody>
<tr>
<td>A_1</td>
<td>15.2051</td>
<td>13.1085</td>
</tr>
<tr>
<td>A_2</td>
<td>-2.0857</td>
<td>-2.0857</td>
</tr>
<tr>
<td>A_3</td>
<td>0.0224</td>
<td>0.0403</td>
</tr>
<tr>
<td>B_1</td>
<td>1424.2</td>
<td>45.61</td>
</tr>
<tr>
<td>B_2</td>
<td>-696.74</td>
<td>-385.14</td>
</tr>
<tr>
<td>B_3</td>
<td>0.9051</td>
<td>-0.5028</td>
</tr>
<tr>
<td>C</td>
<td>459.67</td>
<td>273.15</td>
</tr>
<tr>
<td>AAPD^1</td>
<td>0.73</td>
<td>0.73</td>
</tr>
<tr>
<td>MAPD^2</td>
<td>3.55</td>
<td>3.55</td>
</tr>
<tr>
<td>AAD^3</td>
<td>0.057 psi</td>
<td>0.391 kPa</td>
</tr>
<tr>
<td>NP^4</td>
<td>196</td>
<td>196</td>
</tr>
</tbody>
</table>

1AAPD = Avg Absolute Percent Deviation  
2MAPD = Maximum Absolute Percent Deviation  
3AAD = Average Absolute Deviation  
4NP = Number of data Points (NP)

**Conclusions**

For converting RVP to TVP of motor gasoline and natural gasoline, this tip presented simple correlations similar to the ones reported in API 2517 for crude oil RVP to TVP. This tip determined the correlation parameters by regressing the data generated from available diagrams. These correlations were also extended for easy conversion from TVP to RVP. Tables 1, 2, and 3 present the correlation parameters in SI and FPS system of units. Tables 1 and 3 presents the accuracy of the proposed correlations against the data generated from diagrams. Tables 1 and 3 indicate that the accuracy of these correlations is as good as the quality of the data in the original diagram and can be used for easy conversion of RVP to TVP or vice versa. These correlations are easy to use for hand or spreadsheet calculations and should be used for estimation purposes. For accurate measurements, correction procedures outlined in ASTM D 6377–10 and other guidelines should be consulted. Several organizations are currently working to improve the accuracy of TVP estimation from RVP and/or VPCR(x) (ASTM D6377) measurement techniques. In all cases, Federal and State Laws and Regulations should be followed for safety and environmental protection.

To learn more about similar cases and how to minimize operational problems, we suggest attending our G4 (Gas Conditioning and Processing), G5 (Advanced Applications in Gas Processing), and PF4 (Oil Production and Processing Facilities) courses.

PetroSkills offers consulting expertise on this subject and many others. For more information about these services, visit petroskills.com/consulting, or email us at consulting@PetroSkills.com.

By: Dr. Mahmood Moshfeghian
Correlations for Conversion between True and Reid Vapor Pressures

References


Appendix

Figure 2a. TVP as a function of RVP and temperature for crude oil
Figure 2b. TVP as a function of RVP and temperature for crude oil
In the past Tips of the Month (TOTM), we discussed the phase behavior and water content of lean sweet, sour natural gases and acid gases–water systems. Specifically, in the November 2007 [1], February 2014 [2], and September 2014 [3] Tips of the Month (TOTM), we discussed the phase behavior of water-saturated sour gases and acid gases. We also demonstrated the accuracy of shortcut and rigorous calculation methods. In the April 2015 TOTM [4] we introduced correlation and graphs to estimate the water content of sour gases.

In this TOTM, we will study the effect of relative density (Specific Gravity, SG) on the saturated water content of sweet natural gases. The results of this study include the water content of sweet natural gases as a function of relative density in the range of 0.60 to 0.80. Four temperatures of 4.4, 23.9, 37.8 and 149 °C (40, 75, 100, and 300 °F) were considered. For each temperature, the saturated water content was calculated for pressures of 1724, 3448, 6897, and 13793 kPaa (250, 400, 100 and 2000 psia).

In this TOTM, we will study the effect of relative density (Specific Gravity, SG) on the saturated water content of sweet natural gases. The results of this study include the water content of sweet natural gases as a function of relative density in the range of 0.60 to 0.80. Four temperatures of 4.4, 23.9, 37.8 and 149 °C (40, 75, 100, and 300 °F) were considered. For each temperature, the saturated water content was calculated for pressures of 1724, 3448, 6897, and 13793 kPaa (250, 400, 100 and 2000 psia).

**Water Content Calculation**

The dry gas compositions of the four mixtures studied in this study are presented in Table 1. The Soave-Redlich-Kwong equation of state (SRK EOS) [5] in the ProMax [6] was used to predict the water content of these gas mixtures at different pressures and temperatures. A simplified process diagram for this study is presented in Figure 1.

**Table 1.** The compositions of the four gases with given relative density (Specific Gravity, SG)

<table>
<thead>
<tr>
<th>Component</th>
<th>Gas A</th>
<th>Gas B</th>
<th>Gas C</th>
<th>Gas D</th>
</tr>
</thead>
<tbody>
<tr>
<td>CH₄</td>
<td>92.67</td>
<td>88.51</td>
<td>86.05</td>
<td>73.50</td>
</tr>
<tr>
<td>C₂H₆</td>
<td>5.29</td>
<td>6.02</td>
<td>6.06</td>
<td>13.40</td>
</tr>
<tr>
<td>C₃H₈</td>
<td>1.38</td>
<td>3.18</td>
<td>3.39</td>
<td>6.90</td>
</tr>
<tr>
<td>iC₄</td>
<td>0.18</td>
<td>0.46</td>
<td>0.84</td>
<td>0.80</td>
</tr>
<tr>
<td>nC₄</td>
<td>0.34</td>
<td>0.85</td>
<td>1.36</td>
<td>2.40</td>
</tr>
<tr>
<td>nC₅</td>
<td>0.14</td>
<td>0.98</td>
<td>2.30</td>
<td>3.00</td>
</tr>
<tr>
<td>SG</td>
<td>0.603</td>
<td>0.652</td>
<td>0.693</td>
<td>0.790</td>
</tr>
</tbody>
</table>

The feed dry-gas at the specified pressure and temperature was saturated with water first and passed through a separator. The water content and specific gravity of vapor leaving the separator were recorded.

**Results and Discussion**

Figure 2 through 5 present the saturated water content of sweet natural gases as a function of the dry gas relative density and pressures of 1724 to 13793 kPaa (250 to 2000 psia) for temperatures of 4.4, 23.9, 37.8 and 149 °C (40, 75, 100, and 300 °F), respectively.
Effect of Relative Density (Specific Gravity) on the Saturated Water Content of Sweet Natural Gases

**Figure 2.** Variation of saturated water content of sweet natural gas with the dry gas relative density and pressure at 4.4 °C (40 °F)

**Figure 3.** Variation of saturated water content of sweet natural gas with the dry gas relative density and pressure at 23.9 °C (75 °F)
Effect of Relative Density (Specific Gravity) on the Saturated Water Content of Sweet Natural Gases

**Figure 4.** Variation of saturated water content of sweet natural gas with the dry gas relative density and pressure at 37.8 °C (100 °F)

**Figure 5.** Variation of saturated water content of sweet natural gas with the dry gas relative density and pressure at 149 °C (300 °F)
While the results presented in the above diagrams are in agreement with Figure 6.1 of Reference [7], it is not in agreement with the suggested correction factor in the inset of Figure 20-3 of the GPSA data book [8] and the results of Reference [9]. For the gases with relative density of 0.6 up to 0.8, the GPSA figure indicates a correction factor of 1 to about 0.97 should be multiplied by the water content of saturated water content of sweet gas with relative density of 0.6. This study indicates that for the range of 0.6 to 0.8 relative density, the water content is not a function of the hydrocarbon gas composition. To verify the result of this study, the saturated water content of several gas mixtures were compared with the results of EzThermo [10] software. The SRK EOS in the EzThermo software was developed and regressed to predict the properties of sweet synthetic natural gas and natural gas compositions [11]. The same is also valid for the SRK EOS in ProMax. The comparison of these two software is presented in Table 2. This table indicates an excellent agreement between these two software.

### Table 2. Comparison of the ProMax [6] and EzThermo [10] predicted saturated water content results at 37.8 °C (100 °F)

<table>
<thead>
<tr>
<th>P, psia</th>
<th>P, kPa</th>
<th>Specific Gravity</th>
<th>lb/m/MMSCF</th>
<th>kg/10⁶ Sm³</th>
<th>Percent Deviation</th>
</tr>
</thead>
<tbody>
<tr>
<td>500</td>
<td>3448</td>
<td>0.603</td>
<td>101.7</td>
<td>1626.9</td>
<td>-0.36</td>
</tr>
<tr>
<td>500</td>
<td>3448</td>
<td>0.652</td>
<td>101.6</td>
<td>1625.6</td>
<td>-0.49</td>
</tr>
<tr>
<td>500</td>
<td>3448</td>
<td>0.693</td>
<td>101.5</td>
<td>1624.0</td>
<td>-0.60</td>
</tr>
<tr>
<td>500</td>
<td>3448</td>
<td>0.790</td>
<td>100.9</td>
<td>1614.0</td>
<td>-0.55</td>
</tr>
<tr>
<td>1000</td>
<td>6897</td>
<td>0.603</td>
<td>57.4</td>
<td>918.2</td>
<td>-0.06</td>
</tr>
<tr>
<td>1000</td>
<td>6897</td>
<td>0.652</td>
<td>57.4</td>
<td>917.9</td>
<td>-0.32</td>
</tr>
<tr>
<td>1000</td>
<td>6897</td>
<td>0.693</td>
<td>57.3</td>
<td>916.9</td>
<td>-0.55</td>
</tr>
<tr>
<td>1000</td>
<td>6897</td>
<td>0.790</td>
<td>56.8</td>
<td>908.5</td>
<td>-0.21</td>
</tr>
<tr>
<td>1500</td>
<td>10345</td>
<td>0.603</td>
<td>43.1</td>
<td>690.1</td>
<td>0.07</td>
</tr>
<tr>
<td>1500</td>
<td>10345</td>
<td>0.652</td>
<td>43.2</td>
<td>691.3</td>
<td>-0.31</td>
</tr>
<tr>
<td>1500</td>
<td>10345</td>
<td>0.693</td>
<td>43.2</td>
<td>691.6</td>
<td>-0.68</td>
</tr>
<tr>
<td>1500</td>
<td>10345</td>
<td>0.790</td>
<td>42.8</td>
<td>684.3</td>
<td>-0.49</td>
</tr>
<tr>
<td>2000</td>
<td>13793</td>
<td>0.603</td>
<td>36.4</td>
<td>582.6</td>
<td>0.33</td>
</tr>
<tr>
<td>2000</td>
<td>13793</td>
<td>0.652</td>
<td>36.6</td>
<td>585.5</td>
<td>-0.19</td>
</tr>
<tr>
<td>2000</td>
<td>13793</td>
<td>0.693</td>
<td>36.7</td>
<td>587.5</td>
<td>-0.66</td>
</tr>
<tr>
<td>2000</td>
<td>13793</td>
<td>0.790</td>
<td>36.7</td>
<td>586.4</td>
<td>-0.40</td>
</tr>
</tbody>
</table>
Conclusions

Figures 2 through 5 cover wide ranges of pressures and temperatures commonly encountered in the gas processing operation. The analysis of Figures 2 through 5 indicates that the gas relative density has minimal effect on the saturated water content of sweet natural gases. This conclusion is valid for the following ranges:

- Sweet gas relative density in the range of 0.6 to 0.8
- Temperature range of 4.4 to 149 °C (40 to 300 °F)
- Pressure range of 1724 to 13793 kPaa (250 to 2000 psia)

Additional experimental water content data are being taken and analyzed by the GPA Research Committee to update the relative density correction factor as presented in [8]. From the data that are currently available, it appears that this correction is minimal if any for gas compositions, and temperature and pressure ranges that typically occur in the oil and gas facilities. This will be confirmed once the results from the GPA Research Committee have been published.

To learn more about similar cases and how to minimize operational problems, we suggest attending our G4 (Gas Conditioning and Processing), G5 (Advanced Applications in Gas Processing), G6 (Gas Treating and Sulfur Recovery), PF49 (Troubleshooting Oil and Gas Facilities), and PF81 (CO2 Surface Facilities) courses.

PetroSkills offers consulting expertise on this subject and many others. For more information about these services, visit our website at http://petroskills.com/consulting, or email us at consulting@PetroSkills.com.

By: Dr. Mahmood Moshfeghian

References

http://www.jmcampbell.com/tip-of-the-month/2014/02/acid-gas-water-content/
GPA Research Report RR-42, Predicting Synthetic Gas and Natural Gas Thermodynamic Properties Using a Mod
This Tip of the Month (TOTM) discusses places to look for opportunities to increase production quickly. I’ll discuss the concepts, techniques, and later where to look for and find opportunities….to be able to answer the **Key Question: “Are we getting the most from our Asset?”**

Do you know how your asset makes money? Is it simply oil and gas price?

Do the prices of oil and gas change daily?

Do you adjust your operating conditions based on the prices of oil and gas?

Does fuel gas have any value?

**Benchmarking Against Physics™**

What are your process Key Performance Indicators?

Energy Consumed per boe (barrel of oil equivalent) produced?

Theoretical Energy Required per boe produced?

Design Energy required per boe produced?

**Where are there gaps?**

How do you evaluate the performance of your:

- Pumps
- Compressors
- Processes
  - Oil / Water Dehydration
  - Gas Dehydration
  - Gas Treating

Looking at the Technical Limit Diagram (Choke Model) shown in Figure 1, where could there be opportunities?

How can we use the current equipment to increase production?

What’s the cost of one day of downtime?

What’s the monetary value of the annual flare volume?

How can we increase inflow from the wells to increase production?

- Can we reduce the inlet pressures to the facilities to increase reservoir inflow?
- Are we holding backpressure on the facilities, or are we floating the sales gas pipeline?
- Are there any excessive pressure drops in the flowlines/equipment?
- Scale
- Remove choke internals

By James F. Langer, P.E.
Technical Limit Diagram / Choke Model

Absolute Limit is the maximum theoretical Reservoir limit

Production Limit is today’s *maximum economic* limit

Reservoir

Wet flood
Gas SO
Water SO
Side Tracks
Addt’l Wells

Stimulation

A Lift
SC
Bean Ups
GLO

BP Red.
Chemical
Twinned Lines

Comp Opt

Vessel Mod
Process Opt
Chemical

Drill Red
Commprr,
Pumps

Wells

Gathering

Facilities

Export

Uneconomic work with current technology
Possible with add’t’l *economic* work/investment
Possible with current status

Economic Limit

Integrated Capacity

Deferred
Sales

Current Production

*Figure 1.* Technical Limit Diagram or Choke Model
Compressors

Energy Efficiency:
- What’s the position of the recycle valve? Are you wasting energy?
- Reciprocating: Adjust Pockets / re-cylinder
- Centrifugal: Consider re-wheeling
- Can you reduce the suction temperature of the compressors to increase available HP
- Infinite Recycling of LPG
  - Is this occurring in your compression system?
- What’s causing downtime of your machines?
- Could your upstream separators be carrying over liquids?

Heat Exchangers

- Have the overall heat transfer coefficients become lower?
- Have the pressure drops increased?

Pumps

- What’s the position of your centrifugal pump discharge control valve?
- Do you have too much HP, consider re-wheeling
- Are you shearing and creating a lot of emulsion requiring higher chemical costs?

Specifications

- Challenge sales specifications.
- Are we over treating?
- How’s our crude vapor pressure (RVP)?
- Are we over stabilizing?
- How’s our crude BS&W?
- Are we over treating?

Reduce OPEX

Chemical Costs: Demulsifier
- Are we over dosing?
- Are we adjusting the chemical injection rate

Heat Exchangers

- Have we identified our fixed and variable costs?
- What’s the cost of fuel? Is fuel free?
- What’s the cost of flaring?

Downtime

- How many times have we repaired this pump this year? Or over its life to date?
- What’s the cost of downtime in $/hr? How many $ in revenue has we lost this quarter?
- How long does it take to repair? Do we need a new seal design? More spare parts?

Capital Projects

What could we add to increase production?
- Is inlet compression profitable?
  (Lease / Purchase)

Flare

Is the flare smokey?

This means its rich gas containing propane / butanes worth more than methane. It’s a place to look for process changes to prevent the loss of these higher valued components.
Finding Opportunities Checklist

Wells

- What’s the effect of increasing the tubing size?
- What’s the effect of optimizing the gas lift?
- Do all injection wells need the same maximum pressure for gas lift or water injection? Can we segregate the system?
- Do the wells need stimulation?
- Do the wells need water shutoff?
- Manifolds – Are there leaking check valves allowing backflow from an HP well into an LP well?

CASE 1: Different Molecules have different values in the Gas and Oil. Are you maximizing the value of your products?

Offshore in the Gulf of Mexico, some contracts allow higher vapor pressure in the sales crude oil during the winter months.

KEY QUESTION: How can you take advantage of this opportunity?

Generally, selling molecules as oil has been more valuable than selling the molecules as gas. So how can you send more of smaller LPG molecules into the sales crude oil to increase the number of crude oil barrels sold?

As you learned about vapor pressure and phase diagrams in G4 (Gas Conditioning and Processing course) you simply have to increase the pressure of the final separation, and decrease its temperature. In winter months by lowering the oil dehydration train temperature, and increasing its pressure you increase the vapor pressure in the sales oil tank, and increase the number of barrels sold. You can generate millions of dollars per year, by simply adjusting a temperature control setpoint and a pressure control setpoint. For Zero CAPEX.

CASE 2: High Gas Prices

For a short period in 2000 a unique opportunity presented itself. Gas prices spiked!

KEY QUESTION: How do we take advantage of this opportunity?

Simply send your molecules into the gas sales by dropping the oil dehydration treater pressure and raising its temperature. Many Operators in the GOM took advantage of this opportunity. Some just sat and watched!

Conclusion

To learn more about similar cases and how to minimize operational problems, we suggest attending our G4 (Gas Conditioning and Processing), G5 (Advanced Applications in Gas Processing), PF3 (Concept Selection and Specification of Production Facilities in Field Development Projects), and PF49 (Troubleshooting Oil & Gas Processing Facilities) courses.

PetroSkills offers consulting expertise on this subject and many others. For more information about these services, visit our website at petroskills.com/consulting, or email us at consulting@PetroSkills.com.

By: James F. Langer, P.E.
In the December 2014 tip of the month (TOTM) [1], we discussed troubleshooting of gas-liquid separators for removal of liquids from the gas stream leaving the separator. There are two methods for sizing gas-liquid separators: 1. Droplet settling theory method, 2. Souders-Brown approach. Historically the Souders-Brown equation has been employed as it can provide reasonable results and is easy to use, but has shortcomings in terms of quantifying separator performance. References [2-4] provide the details on the droplet settling theory methods which can be used to more accurately quantify separator performance. The Souders-Brown method is limited in that it is based on the average droplet size, but cannot quantify the amount of liquid droplets exiting the gas gravity section.

In this TOTM, we will focus on the application of Souders-Brown approach in gas-liquid separators and present diagram, simple correlations and tables to estimate the Souders-Brown equation constant, Ks (the so-called sizing parameter). We will consider both vertical and horizontal gas-liquid separators. Knowing the actual gas flow rate through the vessel, one can use Ks parameter to determine the maximum allowable gas velocity through the vessel and determine the required separator diameter. One can also use the appropriate value of Ks to size the mist extractor in the vessel. The performance of a gas-liquid separator is highly dependent on the value of Ks; therefore, the choice of appropriate Ks-values is important.

**Gas Gravity Separation Section**

The gas gravity separation section of a separator has two main functions:

1. Reduction of entrained liquid load not removed by the inlet device
2. Improvement / straightening of the gas velocity profile.

Most mist extractors have limitations on the amount of entrained liquid droplets that can be efficiently removed from the gas, thus the importance of the gas gravity section to remove the liquids to an acceptable level upstream of the mist extractor. This is particularly important for separators handling higher liquid loads. For scrubber applications with low liquid loadings, the Ks-values will be primarily dependent on the mist extractor type, and the gas gravity separation section becomes less important. For the higher liquid load applications, i.e. conventional separators, there are two approaches for sizing the gravity separation section to remove liquid droplets from the gas:

1. The Souders-Brown approach (Ks Method)
2. Droplet settling theory

**The Souders-Brown Approach**

If we consider a spherical liquid droplet with a diameter of $D_p$ in the gas phase two forces as shown in Figure 1 act on it. The drag force, $F_D$, is exerted by flow of gas and gravity force, $F_G$, is exerted by the weight of droplet. The drag force acts to entrain the liquid droplet while the gravity force acts to pull it down and separating it from the gas phase.

**Figure 1.** Schematic of the forces acting on a liquid droplet in the gas phase [5]
Assuming plug flow with no eddies or disturbances, a single droplet and ignoring the end effect, at equilibrium (free fall or terminal velocity), these two forces are equal.

\[ FD = FG \quad (1) \]

As presented in the Appendix, substitution of expressions for the drag and gravity forces in Equation 1, the maximum allowable gas velocity, \( V_{G_{\text{max}}} \), which prevents entrainment of liquid is obtained.

\[ V_{G_{\text{max}}} = K_S \sqrt{\frac{\rho_L - \rho_G}{\rho_G}} \]

Equation 2 is called Souders-Brown [6] equation and \( K_S \) is referred to as the design or sizing parameter. The terms \( \rho_G \) and \( \rho_L \) are the gas phase and liquid phase densities, respectively.

Once the maximum gas velocity, \( V_{G_{\text{max}}} \), through the vessel is determined by Equation 2, one can calculate the minimum vessel diameter, \( D_{\text{min}} \) by Equation 3.

\[ D_{\text{min}} = \sqrt{\frac{4}{\pi}} q_a / (F_G V_{G_{\text{max}}}) \]

Where:

- \( F_G \) = Fraction of cross section area available for gas flow (\( F_G = 1 \) for vertical separators and is a function of liquid height for horizontal separators)
- \( q_a \) = Gas flow rate at the actual flowing condition

The Design Parameter, \( K_S \)

The design parameter, \( K_S \), in the Souders-Brown equation is an empirical parameter and is a key factor for the sizing the gas-liquid separators vessel diameter as well as for determination of the mist extractor diameter. Its value depends on several factors including:

- Pressure
- Fluid properties (note temperature has a large impact on fluid properties)
- Separator geometry
  - Vessel length and the liquid level (for horizontal separators)
- Steadiness of flow
- Inlet device design and performance
- Relative amounts of gas and liquid
- Most importantly – mist extractor type and design (e.g. mesh pad, vane pack, multi-cyclone)

There are several sources that one can look up the \( K_S \) values for different applications. In the following sections, we will discuss three sources.
API 12 J

The API 12J [7] recommends ranges of Ks values for vertical and horizontal gas-liquid separators. These values are presented in Table 1. The equivalent of API 12J for the North Sea region is NORSOK P-100. Per API 12J, “the maximum allowable superficial velocity, calculated from the above factors, is for separators normally having a wire mesh mist extractor. This rate should allow all liquid droplets larger than 10 microns to settle out of the gas. The maximum allowable superficial velocity or other design criteria should be considered for other type mist extractor. Mist extractor manufacturer’s recommended minimum distances upstream and downstream of the wire mesh between gas inlet and outlet nozzles should be provided for full utilization of the mist extractor. These values assume separators are equipped with standard mesh pad mist extractors” [7].

**Table 1.** API 12 J recommended range of Ks values for vertical and horizontal separators [7]

<table>
<thead>
<tr>
<th>Type</th>
<th>Height or Length, ft (m)</th>
<th>Typical $K_s$ range, ft/sec</th>
<th>Typical $K_s$ range, m/s</th>
</tr>
</thead>
<tbody>
<tr>
<td>Vertical</td>
<td>5 (1.52)</td>
<td>0.12 to 0.24</td>
<td>0.037 – 0.073</td>
</tr>
<tr>
<td></td>
<td>10 (3.05)</td>
<td>0.18 to 0.35</td>
<td>0.055 – 0.107</td>
</tr>
<tr>
<td>Horizontal</td>
<td>10 (3.05)</td>
<td>0.40 to 0.50</td>
<td>0.122 to 0.152</td>
</tr>
<tr>
<td>Other Lengths</td>
<td></td>
<td>0.40 to 0.50 (L/10)$^{0.56}$</td>
<td>0.122 to 0.152 (L/3.05)$^{0.56}$</td>
</tr>
</tbody>
</table>

Campbell Book

The $K_s$ method, Equation 2, is an empirical approach to estimate the maximum allowable gas velocity to achieve a desired droplet separation. For vertical separators with no mist extractor devices, Chap 11, Vol 2 of the Gas Conditioning and Processing book presents $K_s$ as a function of pressure and liquid droplet size [5]. This dependency of $K_s$ on pressure and droplet size is presented in Figure 2 [5]. Note for each droplet size a range of $K_s$ –values are given for a specified pressure. For horizontal separators, the sizing depends on (in addition to the droplet size, density of gas and liquid phases, and gas velocity) separator effective length, $L_{ef}$, and the depth available for gas flow, $h_{ef}$ (i.e. liquid level) in the separators.

**Figure 2.**

$K_s$ as a function of pressure and liquid droplet size for vertical separators with no mist extractor devices [5]
Sizing of the horizontal separators are more complicated. Referring to Figure 3, the effective \( L_e \) may be defined in terms of separator actual length and diameter like \( L_e = L - D \). Therefore, the Souders-Brown parameter for horizontal separators, \( K_{SH} \), can be estimated in by Equation 4 in terms of \( K_{SV} \) (read from Figure 2) for vertical separator [3].

\[
K_{SH} = K_{SV} \left( \frac{L_e}{D} \right) \quad \text{or} \quad K_{SH} = K_{SV} \left[ \frac{(L_e/D)}{(h_e/D)} \right]
\]

(4)

If the calculated value of \( K_{SH} \) by Equation 4 is greater than the maximum value of 0.7 ft/sec (0.21 m/s), it should be set equal to this maximum value.

The horizontal separator sizing is a trial-and-error procedure. Normally, the \( L_e/D \) and \( h_g/D \) are assumed and \( K_{SH}, V_{gmax}, D \) are calculated by Equations 4, 2, and 3, respectively. The effective length and actual lengths are calculated by Equation 5.

\[
L_e = \sqrt{\frac{4i q_L}{\pi D^2 F_L}} \quad \text{and} \quad L = L_e + D
\]

(5)

Where:
- \( D \) = Diameter
- \( F_L \) = Fraction of cross section area occupied by liquid (function of liquid height in horizontal separator)
- \( q_L \) = Liquid actual volume flow rate
- \( t \) = Residence time per API 12J [7]

If the calculated \( L/D \) is outside of its recommended range (normally \( 3 < L/D < 6 \)), the liquid height in the vessel is changed and the calculations are repeated. For detail of calculations procedure refer to chapter 11 of reference [5].

**K\text{S} Correlations**

The curves for different droplet sizes shown in Figure 2 are fitted to a 3rd order polynomial (for droplet sizes of 100, 150, and 300 microns). The correlation is in the form of Equation 6 and its regressed coefficients \( a, b, c, \) and \( d \) are presented in Tables 2A and 2B for field (FPS) and System International (SI) units, respectively.

\[
K_S = a + bP + cP^2 + dP^3
\]

(6)
In Table 2, each droplet size in micron (µ) is preceded by letter L or U representing the lower and upper curve, respectively. The pressure is in psi and Ks is in ft/sec for FPS (kPa and m/s in SI).

The last row of Table 2 provides the average absolute percent deviation (AAPD) of the predicted Ks by the proposed correlation from the corresponding values of Figure 2 values.

The two curves for 500 micron droplet size in Figure 2 were divided into 4 and 2 segments based on pressure range for the lower and upper curves, respectively. Each segment was fitted to a linear equation in the form of Equation 7 and its regressed coefficients e and f are presented in Tables 3A and 3B for FPS and SI units, respectively.

\[
K_S = e + fP
\]

(7)

**Table 2A** (FPS). Regressed coefficients for Equation 6 (P in psi and Ks in ft/sec)

Droplet Size: 100 – 300 microns

<table>
<thead>
<tr>
<th>Coefficient</th>
<th>L 100 µ</th>
<th>U 100 µ</th>
<th>L 150 µ</th>
<th>U 150 µ</th>
<th>L 300 µ</th>
<th>U 300 µ</th>
</tr>
</thead>
<tbody>
<tr>
<td>a</td>
<td>0.044882</td>
<td>0.051678</td>
<td>0.072564</td>
<td>0.078829</td>
<td>0.161458</td>
<td>0.18108</td>
</tr>
<tr>
<td>b</td>
<td>7.24E-05</td>
<td>8.13E-05</td>
<td>0.000117</td>
<td>0.000141</td>
<td>0.00024</td>
<td>0.000273</td>
</tr>
<tr>
<td>c</td>
<td>-5.5E-08</td>
<td>-7.8E-08</td>
<td>-9.4E-08</td>
<td>-1.2E-07</td>
<td>-1.8E-07</td>
<td>-2.1E-07</td>
</tr>
<tr>
<td>d</td>
<td>1.58E-11</td>
<td>2.15E-11</td>
<td>2.74E-11</td>
<td>3.61E-11</td>
<td>4.82E-11</td>
<td>5.79E-11</td>
</tr>
<tr>
<td>AAPD</td>
<td>0.36</td>
<td>0.41</td>
<td>0.49</td>
<td>0.31</td>
<td>0.29</td>
<td>0.24</td>
</tr>
</tbody>
</table>

**Table 2B** (SI). Regressed coefficients for Equation 6 (P in kPa and Ks in m/s in SI).

Droplet Size: 100 – 300 microns

<table>
<thead>
<tr>
<th>Coefficient</th>
<th>L 100 µ</th>
<th>U 100 µ</th>
<th>L 150 µ</th>
<th>U 150 µ</th>
<th>L 300 µ</th>
<th>U 300 µ</th>
</tr>
</thead>
<tbody>
<tr>
<td>a</td>
<td>0.0137</td>
<td>0.015799</td>
<td>0.022096</td>
<td>0.024005</td>
<td>0.049173</td>
<td>0.055196</td>
</tr>
<tr>
<td>b</td>
<td>3.18E-06</td>
<td>3.57E-06</td>
<td>5.19E-06</td>
<td>6.2E-06</td>
<td>1.06E-05</td>
<td>1.21E-05</td>
</tr>
<tr>
<td>c</td>
<td>-3.5E-10</td>
<td>-4.5E-10</td>
<td>-6E-10</td>
<td>-7.6E-10</td>
<td>-1.1E-09</td>
<td>-1.4E-09</td>
</tr>
<tr>
<td>d</td>
<td>1.45E-14</td>
<td>2.01E-14</td>
<td>2.53E-14</td>
<td>3.32E-14</td>
<td>4.54E-14</td>
<td>5.41E-14</td>
</tr>
<tr>
<td>AAPD</td>
<td>0.36</td>
<td>0.40</td>
<td>0.49</td>
<td>0.32</td>
<td>0.30</td>
<td>0.24</td>
</tr>
</tbody>
</table>
Gas-Liquid Separators Sizing Parameter

**Table 3A** (FPS). Regressed coefficients for Equation 7 (P in psi and Ks in ft/sec)

<table>
<thead>
<tr>
<th>Segment Number</th>
<th>Pressure Range, psi</th>
<th>L 500 μ</th>
<th>AAPD</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Low</td>
<td>High</td>
<td>e</td>
</tr>
<tr>
<td>1</td>
<td>100</td>
<td>200</td>
<td>0.27812</td>
</tr>
<tr>
<td>2</td>
<td>200</td>
<td>300</td>
<td>0.30911</td>
</tr>
<tr>
<td>3</td>
<td>300</td>
<td>400</td>
<td>0.380839</td>
</tr>
<tr>
<td>4</td>
<td>400</td>
<td>1500</td>
<td>0.402</td>
</tr>
</tbody>
</table>

**Table 3B** (SI). Regressed coefficients for Equation 7 (P in kPa and Ks in m/s in SI).

<table>
<thead>
<tr>
<th>Segment Number</th>
<th>Pressure Range, kPa</th>
<th>L 500 μ</th>
<th>AAPD</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Low</td>
<td>High</td>
<td>e</td>
</tr>
<tr>
<td>1</td>
<td>690</td>
<td>1380</td>
<td>0.084863</td>
</tr>
<tr>
<td>2</td>
<td>1380</td>
<td>2070</td>
<td>0.094220</td>
</tr>
<tr>
<td>3</td>
<td>2070</td>
<td>2760</td>
<td>0.116152</td>
</tr>
<tr>
<td>4</td>
<td>2760</td>
<td>10345</td>
<td>0.1225</td>
</tr>
</tbody>
</table>

Droplet Size: 500 microns
Mist Extractors

The mist extractor is the final gas cleaning device in a conventional separator. The selection, and design to a large degree, determine the amount of liquid carryover remaining in the gas phase. The most common types include wire mesh pads (“mesh pads”), vane-type (vane “packs”) and axial flow demisting cyclones. Figure 4 shows the location and function of a typical mist extractor in a vertical separator. Mist extractor capacity is defined by the gas velocity at which re-entrainment of the liquid collected in the device becomes appreciable. This is typically characterized by a $K_s$–value, as shown in Equation 2. Mesh pads are the most common type of mist extractors used in vertical separator applications. The primary separation mechanism is liquid impingement onto the wires, followed by coalescence into droplets large enough to disengage from the mesh pad. References [1-5] provide mesh pad examples. Table 4 provides a summary of mesh pad characteristics and performance parameters.

![Gas-Liquid Separators Sizing Parameter](image)

**Figure 4.** Typical mist extractor in a vertical separator [5]
Table 4. Mesh pads Ks and performance parameters [3, 5, 8]

<table>
<thead>
<tr>
<th>Description</th>
<th>$K_s$, m/s, (ft/sec)</th>
<th>Separable droplet size, 90% removal, microns</th>
<th>Liquid Load Before Capacity Deteriorates, L/min/m$^2$, (gal/min/ft$^2$)</th>
</tr>
</thead>
<tbody>
<tr>
<td>“Standard” mesh pad</td>
<td>0.107 (0.35)</td>
<td>5</td>
<td>31.5 (0.75)</td>
</tr>
<tr>
<td>“High capacity” mesh pad</td>
<td>0.12 (0.4)</td>
<td>8 – 10</td>
<td>63 (1.5)</td>
</tr>
<tr>
<td>“High efficiency” co-knit mesh pad</td>
<td>0.07 (0.22)</td>
<td>2 – 3</td>
<td>21 (0.5)</td>
</tr>
</tbody>
</table>

Notes:

1. Flow direction is vertical (upflow).

2. Assume mesh pad Ks – value decline with pressure as shown in Table 5. Table 5 was originally developed for mesh pads, but is used as an approximation for other mist extractor types. [9].

3. If liquid loads reaching the mesh pad exceed the values given in Table 4, assume capacity (Ks) decreases by 10% per 42 L/min/m$^2$ (1 gal/min/ft$^2$). [2-4].

4. These parameters are approximate.

Table 5. Mesh pad Ks deration factors as a function of pressure [5]

<table>
<thead>
<tr>
<th>Pressure, kPaa</th>
<th>Pressure, psia</th>
<th>$K_s$ factor, % of design value</th>
</tr>
</thead>
<tbody>
<tr>
<td>100</td>
<td>14.7</td>
<td>100</td>
</tr>
<tr>
<td>500</td>
<td>73</td>
<td>94</td>
</tr>
<tr>
<td>1 000</td>
<td>145</td>
<td>90</td>
</tr>
<tr>
<td>2 000</td>
<td>290</td>
<td>85</td>
</tr>
<tr>
<td>4 000</td>
<td>580</td>
<td>80</td>
</tr>
<tr>
<td>8 000</td>
<td>1 160</td>
<td>75</td>
</tr>
</tbody>
</table>
Vane packs, like mesh pads, capture droplets primarily by inertial impaction. The vane bend angles force the gas to change direction while the higher density liquid droplets tend to travel in a straight-line path, and impact the surface of the vane where they are collected and removed from the gas flow. Table 6 provides vane pack performance characteristics [3, 5, 8].

In the case of demisting cyclones, the vendor should be consulted in regards to performance for the current operations of interest.

### Table 6. Typical vane-pack characteristics [3, 5, 8]

<table>
<thead>
<tr>
<th>Vane Type</th>
<th>Flow Direction</th>
<th>Ks, m/s (ft/s)</th>
<th>Droplet Removal Efficiency</th>
<th>Liquid Load Before Capacity Deteriorates, L/min/m² (gal/min/ft²)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Simple Vane</td>
<td>Upflow</td>
<td>0.15 (0.5)</td>
<td>90+% &gt; 20 microns</td>
<td>84 (2)</td>
</tr>
<tr>
<td>Simple Vane</td>
<td>Horizontal</td>
<td>0.20 (0.65)</td>
<td>90+% &gt; 20 microns</td>
<td>84 (2)</td>
</tr>
<tr>
<td>High Capacity Vane</td>
<td>Upflow</td>
<td>0.25 – 0.35 (0.82 – 1.15)</td>
<td>95+% &gt; 10 microns</td>
<td>210 (5)</td>
</tr>
<tr>
<td>High Capacity Vane</td>
<td>Horizontal</td>
<td>0.3 – 0.35 (1.0 – 1.15)</td>
<td>95+% &gt; 10 microns</td>
<td>210 (5)</td>
</tr>
</tbody>
</table>

Notes:

Assume vane-pack Ks – value decline with pressure as shown in Table 5.

If liquid loads reaching the vane pack exceed the values given in Table 2, assume capacity Ks decreases by 10% per 42 L/min/m² (1 gal/min/ft²). [2-4]. These parameters are approximate only. The vane-pack manufacturer should be contacted for specific information.

### Conclusions

We focused on the application of Souders-Brown Approach (SBA) in gas-liquid separators and presented diagram, simple correlations and tables to estimate the SBA design parameter, Ks.

The SBA can provide reasonable results and is easy to use.

The SBA is limited in that it is based on the average droplet size, but cannot quantify the amount of liquid droplets exiting the gas gravity section and mist extractor section.

To learn more about similar cases and how to minimize operational problems, we suggest attending our PF49 (Troubleshooting Oil and Gas Facilities), PF42 (Separation Equipment Selection and Sizing), G4 (Gas Conditioning and Processing), G5 (Gas Conditioning and Processing – Special), and PF4 (Oil Production and Processing Facilities), courses.

PetroSkills offers consulting expertise on this subject and many others. For more information about these services, visit our website at [http://petroskills.com/consulting](http://petroskills.com/consulting), or email us at consulting@PetroSkills.com.

Dr. Mahmood Moshfeghian
Gas-Liquid Separators Sizing Parameter

References


PF-49, Troubleshooting Oil and Gas Processing Facilities, Bothamley, M., 2014, © PetroSkills, LLC. All Rights reserved.


Appendix

Derivation of the Souders-Brown and Stokes’ Law Equations.

If we consider a spherical liquid droplet with a diameter of, $D_p$, in the gas phase two forces as shown in Figure 1 act on it. The drag force, $F_D$, is exerted by flow of gas and gravity force, $F_G$, is exerted by weight of droplet. The drag force acts to entrain the liquid droplet while the gravity force acts to pull it down.

At equilibrium, these two forces are equal.

$$F_D = F_G \ (1)$$
The drag force is expressed as:

$$F_D = C_D A_p \rho_L v^2 / (2 g_C)$$  \hspace{1cm} (2)

The droplet projected area, $A_p$, is defined by:

$$A_p = (\pi / 4) D_p^2$$  \hspace{1cm} (2A)

The gravity force, $F_G$, is defined:

$$F_G = (\rho_L - \rho_G) V_p g / g_C$$  \hspace{1cm} (3)

The volume of spherical droplet, $V_D$, is calculated by

$$V_p = (\pi / 6) D_p^3$$  \hspace{1cm} (3A)

Substitution of Equations 3 and 4 into Equation 1 and solving for the gas maximum velocity,

$$V_{G_{\text{max}}} = \sqrt{\frac{4 g D_p}{3 C_D}} \left( \frac{\rho_L - \rho_G}{\rho_G} \right)$$  \hspace{1cm} (4)

For practical applications, the first term on the right hand side is replaced by $K_s$

$$K_s = \sqrt{\frac{4 g D_p}{3 C_D}}$$  \hspace{1cm} (5)

Therefore, the maximum gas velocity which prevents entrainment of liquid is obtained.

$$V_{G_{\text{max}}} = K_s \sqrt{\left( \frac{\rho_L - \rho_G}{\rho_G} \right)}$$  \hspace{1cm} (6)

Equation 6 is called Souder-Brown equation and $K_s$ is referred to as the design parameter.

Where:

$A_p$ = Project area of droplet
Gas-Liquid Separators Sizing Parameter

\[ C_D = \text{Drag coefficient} \]
\[ g = \text{Acceleration of gravity} \]
\[ g_c = \text{Conversion factor} \]
\[ V = \text{Gas velocity} \]
\[ V_p = \text{Volume of droplet} \]
\[ \rho_G = \text{Gas density} \]
\[ \rho_L = \text{Liquid density} \]

Once the maximum, \( V_{G\text{max}} \), gas velocity through the vessel is determined by Equation 6, one can calculate the required minimum cross sectional area of vessel for gas flow by the following equation.

\[
A_{G\text{min}} = \left( \frac{\pi}{4} \right) D_{\text{min}}^2 (F_G) = q_a / V_{G\text{max}}
\]

(7)

Solving for the minimum vessel diameter, \( D_{\text{min}} \).

\[
D_{\text{min}} = \sqrt{\left( \frac{4}{\pi} \right) q_a / (F_G V_{G\text{max}})}
\]

(8)

Where:

\[ F_G = \text{Fraction of cross sectional area available for gas flow (} F_G = 1 \text{ for vertical separators and it is a function of liquid height for horizontal separators)} \]
\[ q_a = \text{Gas flow rate at the actual flowing condition} \]

The drag coefficient, \( C_D \), is a function of Reynolds number, \( \text{Re} = (D_p V \rho_G) / \mu_G \). For Stokes’ law \( \text{Re} \):

\[
C_D = 24 / \text{Re} = \frac{24 \mu_G}{D_p V \rho_G} \quad \text{for} \quad \text{Re} < 2
\]

(9)

Substitution of \( C_D \) from Equation 9 into Equation 4 gives liquid droplet terminal velocity, \( V_T \), in the gas phase based on the Stokes’ law.

\[
V_T = \frac{g D_p^2 (\rho_L - \rho_G)}{18 \mu_G}
\]

(10)

Similarly, the terminal velocity for the other flow regimes like Intermediate and Newton can be derived based on their corresponding expressions for the drag coefficients [3].
The use of multiphase flow systems is common practice in the oil and gas industry. Multiphase flow is often encountered in the well tubing, flow lines and gathering systems. For transport of oil and gas (and water) to downstream processing facilities the preference is normally a single pipeline in which both phases are transported simultaneously for economic reasons. Even in gas pipelines where the gas enters the line as a single phase fluid, condensation of liquids can occur due to pressure and temperature changes along the line.

Modeling and simulation of a multiphase systems, even under steady-state conditions, is complex. There are a few tools designed specifically for modeling and analysis of complex multiphase systems such as PipePhase, PipeSim, OLGA, etc. [1].

In the June 2008 Tip of the Month (TOTM), we demonstrated how general-purpose process simulation programs can be used to simulate gas dominated two-phase pipelines. In the August 2008 TOTM, we discussed the value of the simple Flanigan correlation and how it can be used to model and analyze the behavior of a wet gas transmission pipeline. The results of the Flanigan correlation were compared with more rigorous calculation methods for multiphase pipelines.

In this TOTM, we will study the impact of gas-oil ratio (GOR) on pressure drop in crude oil gathering systems. Specifically, pressure drop along a gathering line for nominal pressures of 690, 3450, and 6900 kPag (100, 500, and 1000 psig) and nominal pipe size of 101.6 and 152.4 mm (4 and 6 inches) was calculated using multiphase rigorous method from commercial simulator. The calculated pressure drops are presented in graphical format as a function of the oil stock tank volume flow rate and GOR. Variation of thermo physical properties was considered.

**Case Study**

For the purpose of illustration, we considered a case study for transporting a crude oil of relative density of 0.852 (°API = 34.6) at stock tank condition combined with a gas with relative density of 0.751. The selected GORs were 0 (dead oil), 17.8, 356.5, and 891.3 Sm$^3$ of gas/STm$^3$ of oil (0, 100, 2000, and 5000 scf/STB). The compositions of oil and gas are presented in Table 1. The oil C$_6^+$ was characterized as 30 single carbon number (SCN) [2] ranging from SCN6 to SCN35 while the gas C$_6^+$ was characterized by 10 SCN ranging from SCN6 to SCN15. For details of the SCN components, see Table 3.2 on page 64 of reference [2]. The mole fraction of SCN components were determined by an exponential decay algorithm [3].

**Table 1. Feed composition at stock condition**

<table>
<thead>
<tr>
<th>Component</th>
<th>Composition, mol%</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Gas</td>
</tr>
<tr>
<td>N$_2$</td>
<td>0.272</td>
</tr>
<tr>
<td>CO$_2$</td>
<td>0.330</td>
</tr>
<tr>
<td>CH$_4$</td>
<td>75.541</td>
</tr>
<tr>
<td>Ethane</td>
<td>14.252</td>
</tr>
<tr>
<td>Propane</td>
<td>5.303</td>
</tr>
<tr>
<td>iC$_4$</td>
<td>1.344</td>
</tr>
<tr>
<td>nC$_4$</td>
<td>1.502</td>
</tr>
<tr>
<td>iC$_5$</td>
<td>0.613</td>
</tr>
<tr>
<td>nC$_5$</td>
<td>0.491</td>
</tr>
<tr>
<td>C$_6^+$</td>
<td>0.352</td>
</tr>
<tr>
<td>Total</td>
<td>100</td>
</tr>
</tbody>
</table>
The following assumptions were made:

- Steady state conditions
- The line is 1.601 km (1 mile) long with nominal size of 101.6 and 152.4mm (4 and 6 inches), onshore buried line.
- Segment lengths and elevation changes are presented in Table 2. This elevation profile is considered to be approximately equivalent to “rolling” terrain.
- Pipeline inside surface roughness of 46 microns (0.046 mm, 0.0018 inch)
- Line nominal pressure 690, 3450, and 6900 kPag (100, 500, and 1000 psig)
- The feed enters the line at 15.6 °C and (60 °F)
- The ground/ambient temperature, is 15.6 °C and (60 °F)
- Water cut is 0 (no water in the feed).
- Overall heat transfer coefficients of 2.839 W/m²-°C (0.5 Btu/hr-ft²-°F), for onshore buried line (minor effect as inlet temperature = ambient ground temperature).

**Results and Discussions**

The two phase (oil and gas) flow through the gathering line was simulated by ProMax with SRK EOS for vapor-liquid equilibria and Beggs-Brill for two phase pressure drop calculations. Figures 1A and 1B present the calculated pressure drop per unit length as a function of oil stock tank volume rate and GOR for nominal line diameter of 101.6 mm (4 inches) at nominal line pressure of 690 kPag (100 psig) in SI (international) and FPS (Engineering) system of units, respectively. Figures 1A and 1B indicate that as the GOR increases from 0 to 891 Sm³/STm³ (0 to 5000 scf/STB), the pressure drop increases considerably. Consequently, as the GOR increases, the line capacity decreases.

Figures 2A, 2B, 3A, and 3B present the results for the same line size but at nominal pressures of 3445 and 6900 kPag (500 and 1000 psig), respectively. Contrary to Figure 1, Figures 2 and 3 indicate that at these higher pressures as the GOR increases, the pressure drop decreases for low GOR value. However, for further increase of GOR the pressure drop increases considerably.

**Table 2. Line segment length and elevation change**

<table>
<thead>
<tr>
<th>Segment No</th>
<th>Length, m</th>
<th>Elevation Change, m</th>
<th>Length, ft</th>
<th>Elevation Change, ft</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>0 to 402</td>
<td>15.24</td>
<td>0 to 1320</td>
<td>50</td>
</tr>
<tr>
<td>2</td>
<td>402 to 805</td>
<td>-15.24</td>
<td>1320 to 2640</td>
<td>-50</td>
</tr>
<tr>
<td>3</td>
<td>805 to 1207</td>
<td>15.24</td>
<td>2640 to 3960</td>
<td>50</td>
</tr>
<tr>
<td>4</td>
<td>1207 to 1609</td>
<td>-15.24</td>
<td>3960 to 5280</td>
<td>-50</td>
</tr>
</tbody>
</table>
Impact of Gas-Oil Ratio (GOR) on Crude Oil Pressure Drop in Gathering Systems

**Figure 1A** (SI). Variation of pressure drop per unit length with oil stock tank volume rate and GOR at 690 kPag for 101.6 mm pipe diameter.

**Figure 1B** (FPS). Variation of pressure drop per unit length with oil stock tank volume rate and GOR at 100 psig for 4 in pipe diameter.
Impact of Gas-Oil Ratio (GOR) on Crude Oil Pressure Drop in Gathering Systems

**Figure 2A** (SI). Variation of pressure drop per unit length with oil stock tank volume rate and GOR at 3445 kPag for 101.6 mm pipe diameter

**Figure 2B** (FPS). Variation of pressure drop per unit length with oil stock tank volume rate and GOR at 500 psig for 4 in pipe diameter
Impact of Gas-Oil Ratio (GOR) on Crude Oil Pressure Drop in Gathering Systems

**Figure 3A** (SI). Variation of pressure drop per unit length with oil stock tank volume rate and GOR at 6900 kPag for 101.6 mm pipe diameter

**Figure 3B** (FPS). Variation of pressure drop per unit length with oil stock tank volume rate and GOR at 1000 psig for 4 in pipe diameter
Similar calculations were repeated for another line with nominal pipe size of 152.4 mm (6 inches) and the simulation results are presented in Figures 4 through 6. Figures 4 through 6 also demonstrate the same impact of GOR on the pressure drop, at higher pressures and low GOR, the pressure drop decreases. However, the impact of low GOR at higher pressures is less compared to the smaller line diameter.

**Figure 4A** (SI). Variation of pressure drop per unit length with oil stock tank volume rate and GOR at 690 kPag for 152.4 mm pipe diameter

**Figure 4B** (FPS). Variation of pressure drop per unit length with oil stock tank volume rate and GOR at 100 psig for 6 in pipe diameter
Impact of Gas-Oil Ratio (GOR) on Crude Oil Pressure Drop in Gathering Systems

**Figure 5A** (SI). Variation of pressure drop per unit length with oil stock tank volume rate and GOR at 3445 kPag for 152.4 mm pipe diameter.

**Figure 5B** (FPS). Variation of pressure drop per unit length with oil stock tank volume rate and GOR at 500 psig for 6 in pipe diameter.
Impact of Gas-Oil Ratio (GOR) on Crude Oil Pressure Drop in Gathering Systems

Figure 6A (SI). Variation of pressure drop per unit length with oil stock tank volume rate and GOR at 6900 kPag for 152.4 mm pipe diameter

Figure 6B (FPS). Variation of pressure drop per unit length with oil stock tank volume rate and GOR at 1000 psig for 6 in pipe diameter
Conclusions

The following conclusions can be made based on this case study:

The GOR has a large impact on the capacity of crude oil gathering lines. Generally, as GOR increases the pressure drop increases which lowers the line capacity.

At high pressures and low GOR, pressure drop is lower than the pressure drop for dead oil (solution gas is zero) because the viscosity of live oil is lower than viscosity of dead oil. This effect is bigger for the smaller line diameter.

By: Mahmood Moshfeghian

References

Infinite Recycle Impacts on Compression Systems

By James F. Langer, P.E.

This TOTM will discuss the phenomenon of infinite recycles and their impacts on design, troubleshooting and optimization to increase oil production simply by adjusting setpoint. As well as flaring reduction, unconventionals, identification and solutions.

**Question:** If you have an infinite recycle of LPG in your gas processing plant, how much horsepower does your compression system need?

**Answer:** Infinite

Liquefied Petroleum Gas (LPG) is composed of propane and butane. The pressure at which LPG becomes liquid, called its vapor pressure, varies depending on composition and temperature; for example, it is approximately 220 kPa (32 psi) for pure butane at 20 °C (68 °F), and approximately 2,200 kPa (320 psi) for pure propane at 55 °C (131 °F).

In Figure 1 you have the standard propane refrigeration loop. Compressor (K-100) / Cooler (E100) / Separator (V-100) / JT Valve (VLV-100)/ Chiller Load (E101) and recycle back to the compressor. Under normal operation, stream No 4 has zero flowrate and the system is a closed loop. If we were to add an additional stream of pure propane to the system at the compressor suction it creates an infinite loop since the vapor out of V100 will still be zero due to the thermodynamic limits of the phase diagram. This system requires infinite horsepower and the rest of the equipment will need to increase in size as well. The molecules are trapped in a loop.

Steady State Process simulators generally handle this situation with an unconverged solution or in some instances negative flowrates. Why is this important? Although the simulator will not allow the condition to occur. It can occur during actual operation.

**Figure 1. Propane Refrigeration Loop**
Figure 2 shows a typical three stage compression system, it is like the propane refrigeration system process line up. In a typical design where are the compressor discharge scrubber liquids sent? Generally, it's the previous stage suction scrubber. Eventually it makes its way to the dry oil tank. If the material in the scrubbers has propane and butane as a liquid and then sent to an essentially atmospheric tank the liquids will flash, and then be processed by the Oil Stock Tank Vapor Recovery compressor which will send the molecules back to the compression system. This can set up an infinite recycle or very large recycle of LPG in the compressor system.

As shown in Figure 3, the system response to this recycle of LPG from compressor scrubbers will be to load the system to its capacity, and then flare. This flaring breaks the recycle, but at a cost.

Indications that this is occurring at your offshore platform, gas plant, or unconventional tank battery is a hot flare with a smoky tail. LPG's have a significantly higher heating value than methane and the smoke is being caused by high temperatures in the flame cracking the LPG, and creating soot. The flame will be a deep red rather than orange and we feel higher radiation from flare when walking in the area.
Figure 4 shows how the flash gas composition becomes richer and richer in LPG. This can lead to infinite recycles in the last two separators of the oil stabilization system. The last stage flash has 47% LPG in the vapor flash gas. When we are in the design phase in field development planning gas compositions have a degree of uncertainty. It is wise to have multiple dispositions for these LPG streams. They need a way out: sales gas, oil or fuel.

### DESIGNED SOLUTIONS

1. Pump interstage liquids into the sales gas using a positive displacement pump. The limitation here is the maximum heating value of the sales gas. Liquids may need to be gas stripped of water to avoid hydrates.

2. Pump interstage liquids into the stock tanks oil. The limitation is crude vapor pressure and additional recycling.

3. Operate your interstage compressor discharge coolers at a high enough temperature to avoid liquid accumulation in the discharge scrubbers. This option increases the horsepower requirements of the next stage since you are sending higher temperature gas forward to the next stage of compression. This is also a method to deal with recycles and reduce flaring in operation, but will reduce the inlet oil production since you will be using additional compression HP by operating at higher temperatures.

4. Fuel gas - spike the LPG into the low-pressure fuel gas system. This is a preferred option, but the direct fire equipment must be designed to accommodate the higher heating value.

5. Change the gas composition by adding hydrocarbon dry stripping gas to the crude oil.

6. If unable to send the molecules to the gas (due to Btu limit) or oil (RVP limit) then a separate disposition like Pipeline. Pressurized trucks or a tanker is needed.

7. Add another process - in offshore consider power generator or low salinity waterflood - generation of distilled water using multi-effect evaporation. (e.g. Mechanical Flare)

Bottom line is that all the molecules need an evacuation route.
Infinite Recycle Impacts on Compression Systems

DESIGN CASE No 1: Offshore Platform 
Associated Gas Processing

The case in Figure 5 is taken from “Oilfield Processing of Petroleum-Volume 1- Natural Gas” [2]. No recycling of LPG occurs as all liquids from compressor discharge scrubbers are routed to the sales oil pipeline.

Figure 5. All compressor discharge scrubbers directly sent or pumped into the sales crude pipeline [2]
CASE STUDY No 2: NGL Recovery Plant with 50,000 bbl Floating Roof Tank Emissions Issues

The NGL recovery plant (Figure 6) processed a small amount of crude oil and was spiking liquids into the crude. An infinite recycles developed and the stock tanks were degassing large volumes of LPG. Notice the large temperature drop of the crude going from 145°F (63°C) to 102°F (39°C) as the fluid flashes into the last stage separator. The design intent was to maximize oil volume and eliminate discharge scrubbers, by utilizing the main train separators.

This recycle was broken by adding hydrocarbon dry residue gas from the NGL plant at the inlet to the Electrostatic Treater mixing with the crude oil, and stripping out the LPG breaking the LPG recycle. (Same location near DeC4 bottoms injection point) The blue and red lines show the paths for the LPG Traps.

Figure 6. Operational problems caused by infinite recycles / butane trap in an oil processing train leading to floating roof storage tank degassing.
**CASE STUDY No 3:**

An offshore FPSO (Floating Production, Storage and Onboarding) in the North Sea was in operation for two years and its production was constrained by its compression system. The flare had a smoky tail.

**Question:** What would you recommend looking for?

**Answer:** Infinite Recycles.

After some process simulations, the pressure in the last stage of oil separation was increased by 0.5 barg (7.25 psig). This broke the internal infinite recycle of LPG in the oil dehydration train sending the LPG into the crude oil rather than the flash gas to the compressor. The resulting oil was within contractapor pressure limits once cooled.

The upside was that inlet oil production was increased 30% immediately with zero capex!

Lesson- you need to visit the site to identify opportunities. And if the system is running it doesn’t mean there is no opportunity to improve.

**UNCONVENTIONAL DESIGNS**

Many unconventional plays have standard tank batteries with Tank Vapor Recovery Systems and/or Thermal Oxidizers. Numerous plays have very rich associated gas that form these infinite recycles in their flash gas compression and tank vapor recovery systems. Simply sending the liquids back to the stock tanks or water tanks will not fix this problem. The molecules must be sent forward to the gas gathering system. Also, ensure your design will function in the winter and that LPG condensation in suction piping will not occur, and that stock tanks are insulated and heated to prevent reflux condensation in the stock tanks.

**QUICK TIPS**

Do not design for a single point of operation. (i.e. 40,000 bopd and 100 MMSCFD)

To ensure that the design will work make sure to design during the early start-up phase with high or low rates. Additionally, make sure to design for extreme ambient condition variations in Summer/Winter and Day/Night. The design should also work during shutdown and composition changes over the life of the development.

**CONCLUSION**

Recycles limiting production are frequently found in oil processing facilities worldwide. However, they are easy to identify and fix with setpoint adjustments to increase production. Now that you know how to identify, prevent and take advantage of it for your company...go out and make some process improvements to your systems and make your company more profitable!

To learn more about similar cases and how to minimize operational problems, we suggest attending our G4 (Gas Conditioning and Processing), G5 (Practical Computer Simulation Applications in Gas Processing), PF3 (Concept Selection and Specification of Production Facilities in Field Development Projects), PF4 (Oil Production and Processing Facilities), and PF49 (Troubleshooting Oil & Gas Processing Facilities) courses.

PetroSkills offers consulting expertise on this subject and many others. For more information about these services, visit our website at [http://petroskills.com/consulting](http://petroskills.com/consulting) or email us at consulting@PetroSkills.com.

By: James F. Langer, P.E.

**REFERENCES**


This tip of the month (TOTM) discusses how to “avoid a costly design error found in most oil dehydration systems worldwide”.

Are you having reverse emulsion problems?
Are you having problems meeting your discharge oil in water specifications??

An area to look for immediately is the Process Flow Diagram (PFD). Normally systems work as they were designed. In cases where I have visited sites experiencing problems…the problem is the design.

Why does this occur? The world we deal with is more complex than process simulation models. They are simply thermodynamic models to predict the phase separation of multicomponent systems and perform heat and energy balances. They are useful and very well proven tools. Most designs however; do not consider the various chemicals, presented in in Table 1, that are added to a system on a routine basis.

I would encourage you to check this out at your asset using the real fluids. I perform this demonstration during the Petroskills PF-3 Course. I take a 50% water cut sample of a 45 API Gulf of Mexico Crude Oil and produced water. I shake it up and the emulsion resolves in less than 30 seconds back into oil and water. I then add 5 -10 ml of Demulsifier and Water Clarifier being used on the Platform. Shake this up and voila…instant chocolate milk emulsion. This emulsion will not resolve in days. How is the emulsion resolved? We send it to a heater treater. In class I use a coffee cup of hot water. Let it heat soak for a couple of minutes and the sample resolves into three phases as shown in Figure 1: Oil / a scrambled eggs rag layer (complex emulsion- reverse and normal) / Water. The rag layer is normally about 25% of the total sample. This rag layer is very stable, and I have had samples stay stable for over one year.

Table 1. Example of various problems and their typical treatment

<table>
<thead>
<tr>
<th>Problem</th>
<th>Treatment Chemical</th>
</tr>
</thead>
<tbody>
<tr>
<td>Hydrate Formation</td>
<td>Hydrate Inhibitor</td>
</tr>
<tr>
<td>Water Vapor</td>
<td>Dehydrator</td>
</tr>
<tr>
<td>Mineral Deposits</td>
<td>Scale Inhibitor</td>
</tr>
<tr>
<td>Chemical Corrosion</td>
<td>Corrosion Inhibitor</td>
</tr>
<tr>
<td>Bacterial Corrosion</td>
<td>Bactericide</td>
</tr>
<tr>
<td>Emulsions (Normal or Reverse)</td>
<td>Emulsion Breakers, Coagulants, Flocculants</td>
</tr>
<tr>
<td>Foaming</td>
<td>Defoamer</td>
</tr>
<tr>
<td>Paraffin</td>
<td>Paraffin Inhibitor, Solvent</td>
</tr>
</tbody>
</table>

Most of these chemicals are added to a system to improve a design's performance in oil water and water oil separation. Most of these chemicals are used in parts per million (ppm) fashion using proprietary formulations from Chemical Companies. Most assets will spend $5-$10 Million dollars per year on the use of these formulations. The ability to add the chemicals is included in most designs, but their effects are not normally included in the designs and it's left to Field Operations and the Chemical Companies to come up with a solution.

So, what's the design error? It's mixing fluids that have come into contact with both oil treating and water treating chemicals, and then recycling them back to the front of the process.

**What are these streams?**
- Oil skims from WEMCOs (induced gas floatation manufacture by Western Equipment Manufacturing Company) or Flotation Cells are routinely re-routed back into the Oil Dehydration Train.
- Hydrocyclone Oil Skims
- Closed Drain Oil Skims

![Figure 1. Oil water complex emulsion](image)
Never send these skims back into your main process. Consider sending them directly to the sales oil LACT. Most contracts allow you 3-5% BS&W. Use the contract to your advantage. You will not be able to break these emulsions with your existing process and they will continue to build. Some operators send them to the beach as hazardous waste for further processing. Other operators have a separate slop oil system for treating with higher temperatures, and then send streams to the sales LACT (Lease Automatic Custody Transfer), but never back into the process.

Every process needs a trash can for off spec product. Offshore it’s the closed drain tank / open drain tank. Never recycle this material back into the main oil dehydration system. Why? Let’s look at what fluids enter the closed and open drain tanks.

- Engine Lube Oils – these contain polymers that will generate emulsions
- Oxygenated water from rain. Oxygen will corrode your main production system.
- Oily Deck Drains from skid pans

**How to become an Expert?**

Use sources such as SPE’s One Petro. Chances are that sometime, somewhere, someone has faced the same issue you are trying to solve. In your company is it easy to publish a paper or are there controls and bureaucracy? In most instances, it’s not easy. What does it say to you about the person who has taken the time and written a paper. They want to communicate! What do you think their reaction will be if you contact them about their paper to get further information? They usually will be very happy to discuss the paper and give you even more insights / tips / advice / recommendations for your situation. It’s a way of leveraging your time, and getting to solutions quickly.

I once made a trip to an FPSO Offshore Brazil that was having water treating problems. They made the same design mistakes discussed in this TOTM. I searched One Petro and found another FPSO Offshore Brazil that faced similar issues and how they were solved [1]. The paper is listed in the references of this TOTM.

Here is the recommendation from SPE 90409 paper:

“Oil Reject Skid. To provide a positive path for removing oily solids and chemically stabilized emulsions from the process, an oil reject skid was installed on 135 D to receive hydrocyclone reject liquids and skimming from the surge drum, as well as skimming from the sparger IGF vessels. In the oil reject vessel (Fig. 9), larger solids are removed from the cone bottom, while oily solids and chemically stabilized emulsion separates into an oil-continuous phase that is pumped off the platform without recycling back into the process. The volume of this emulsion is low, a few barrels per day, so the impact on the BS&W of crude leaving the platform is low, increasing it by approximately 0.05 %. The clarified water is returned to one of the degasser vessels, from which it reenters the water treatment process. The oil reject skid installation has been one of the major contributors to the ability of 135 D to reliably reduce the TOG of the produced water to < 20 mg/L.”

**Conclusion**

To learn more about similar cases and how to minimize operational problems, we suggest attending our G4 (Gas Conditioning and Processing), G5 (Advanced Applications in Gas Processing), PF3 (Concept Selection and Specification of Production Facilities in Field Development Projects), and PF49 (Troubleshooting Oil & Gas Processing Facilities) courses.

PetroSkills offers consulting expertise on this subject and many others. For more information about these services, visit our website at petroskills.com/consulting, or email us at consulting@PetroSkills.com.

**Reference**

of the Month (TOTM) discussed the hydrate phase
behavior of natural gas mixtures containing high
content hydrogen sulfide, carbon dioxide, or nitrogen.
Specifically, it showed nitrogen and carbon dioxide
inhibit the hydrate formation slightly while hydrogen
sulfide enhances hydrate formation considerably. This
tip will extend the previous studies on the natural gas
hydrate formation phase behavior. Specifically, it will
study the impact of light hydrocarbons on the
formation of hydrate in a natural gas mixture.

The hydrate formation temperature of a gas depends
on the system pressure and composition. There are
several methods of calculating the hydrate formation
conditions of natural gases [3-6]. References [3-4]
present rigorous methods while [5-6] present the
shortcut methods suitable for hand calculation.
This study uses a rigorous method using the Soave-
Redlich-Kwong (SRK) equation of state [7] in ProMax

Table 1 presents the compositions (mol %) of the gas
mixtures studied. Notice that for non-hydrocarbons
(gases B, C, and D) about 18 mol % of methane is
replaced with about 20 mol % of either nitrogen, car-
bon dioxide or hydrogen sulfide. These compositions
are for a gas stream leaving a separator at 100 °F and
1000 psia (37.8 °C and 6900 kPaa) saturated with
water.

<table>
<thead>
<tr>
<th>Comp</th>
<th>A</th>
<th>B</th>
<th>C</th>
<th>D</th>
<th>E</th>
<th>F</th>
<th>G</th>
<th>H</th>
<th>I</th>
<th>J</th>
<th>K</th>
<th>L</th>
</tr>
</thead>
<tbody>
<tr>
<td>N₂</td>
<td>0.031</td>
<td>0.025</td>
<td>0.025</td>
<td>20.367</td>
<td>0.034</td>
<td>0.028</td>
<td>0.032</td>
<td>0.028</td>
<td>0.031</td>
<td>0.029</td>
<td>0.031</td>
<td>0.059</td>
</tr>
<tr>
<td>CO₂</td>
<td>0.000</td>
<td>0.000</td>
<td>20.006</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
</tr>
<tr>
<td>H₂S</td>
<td>0.000</td>
<td>20.004</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
</tr>
<tr>
<td>CH₄</td>
<td>85.620</td>
<td>68.418</td>
<td>68.166</td>
<td>68.005</td>
<td>93.073</td>
<td>76.374</td>
<td>88.966</td>
<td>77.758</td>
<td>86.132</td>
<td>77.939</td>
<td>86.298</td>
<td>84.817</td>
</tr>
<tr>
<td>C₂H₆</td>
<td>7.886</td>
<td>6.298</td>
<td>6.316</td>
<td>6.297</td>
<td>0.000</td>
<td>17.819</td>
<td>8.194</td>
<td>7.151</td>
<td>7.933</td>
<td>6.173</td>
<td>7.948</td>
<td>2.903</td>
</tr>
<tr>
<td>C₃H₈</td>
<td>3.758</td>
<td>3.002</td>
<td>3.043</td>
<td>3.027</td>
<td>4.069</td>
<td>3.352</td>
<td>0.000</td>
<td>12.665</td>
<td>3.780</td>
<td>2.417</td>
<td>3.788</td>
<td>0.628</td>
</tr>
<tr>
<td>iC₄H₁₀</td>
<td>0.595</td>
<td>0.476</td>
<td>0.491</td>
<td>0.486</td>
<td>0.641</td>
<td>0.531</td>
<td>0.618</td>
<td>0.536</td>
<td>0.000</td>
<td>12.685</td>
<td>0.600</td>
<td>0.056</td>
</tr>
<tr>
<td>nC₄H₁₀</td>
<td>0.785</td>
<td>0.629</td>
<td>0.654</td>
<td>0.644</td>
<td>0.844</td>
<td>0.700</td>
<td>0.816</td>
<td>0.706</td>
<td>0.790</td>
<td>0.362</td>
<td>0.000</td>
<td>0.000</td>
</tr>
<tr>
<td>iC₅H₁₂</td>
<td>0.372</td>
<td>0.302</td>
<td>0.322</td>
<td>0.312</td>
<td>0.396</td>
<td>0.332</td>
<td>0.387</td>
<td>0.332</td>
<td>0.374</td>
<td>0.134</td>
<td>0.375</td>
<td>0.016</td>
</tr>
<tr>
<td>nC₅H₁₂</td>
<td>0.220</td>
<td>0.179</td>
<td>0.193</td>
<td>0.186</td>
<td>0.233</td>
<td>0.196</td>
<td>0.229</td>
<td>0.195</td>
<td>0.221</td>
<td>0.071</td>
<td>0.222</td>
<td>0.008</td>
</tr>
<tr>
<td>C₆H₁₄</td>
<td>0.127</td>
<td>0.106</td>
<td>0.123</td>
<td>0.112</td>
<td>0.129</td>
<td>0.113</td>
<td>0.132</td>
<td>0.110</td>
<td>0.128</td>
<td>0.021</td>
<td>0.128</td>
<td>0.002</td>
</tr>
<tr>
<td>C₇H₁₆</td>
<td>0.486</td>
<td>0.424</td>
<td>0.528</td>
<td>0.443</td>
<td>0.462</td>
<td>0.434</td>
<td>0.505</td>
<td>0.402</td>
<td>0.489</td>
<td>0.050</td>
<td>0.490</td>
<td>0.004</td>
</tr>
<tr>
<td>H₂O</td>
<td>0.121</td>
<td>0.137</td>
<td>0.133</td>
<td>0.121</td>
<td>0.120</td>
<td>0.121</td>
<td>0.121</td>
<td>0.119</td>
<td>0.121</td>
<td>0.118</td>
<td>0.121</td>
<td>0.120</td>
</tr>
<tr>
<td>Total</td>
<td>100.0</td>
<td>100.0</td>
<td>100.0</td>
<td>100.0</td>
<td>100.0</td>
<td>100.0</td>
<td>100.0</td>
<td>100.0</td>
<td>100.0</td>
<td>100.0</td>
<td>100.0</td>
<td>100.0</td>
</tr>
</tbody>
</table>
Figure 1 presents the calculated hydrate formation and the dew point portion of the phase envelope (continuous curves) of a sweet natural gas (gas E of Table 1) containing 0 mol % C\textsubscript{2}H\textsubscript{6}. Figure 1 also presents the dew point and hydrate formation (broken curves) for gas F of Table 1 containing 17.8 mol % C\textsubscript{2}H\textsubscript{6}.

Figure 1 indicates that the presence of 17.8 mol % C\textsubscript{2}H\textsubscript{6} has a negligible effect on the hydrate formation curve. Note that the points to the left and above the hydrate curves represent the hydrate formation region. From an operational point of view, this region should be avoided. This figure also indicates that the presence of C\textsubscript{2}H\textsubscript{6} decreases the cricondenbar pressure and the cricondentherm temperature; therefore, the two-phase (gas + liquid) region within the envelope shrinks.

![Figure 1. The impact of C\textsubscript{2}H\textsubscript{6} on the hydrocarbon dew point and hydrate formation curves](image-url)
Figure 2 presents the calculated hydrate formation and the dew point portion of the phase envelope (continuous curves) of a sweet natural gas (gas G of Table 1) containing 0 mol % C$_3$H$_8$. Figure 2 also presents the dew point and hydrate formation curves (broken curves) for gas H of Table 1 containing 12.7 mol % C$_3$H$_8$. Figure 2 indicates that the presence of 12.7 mol % C$_3$H$_8$ shifts the hydrate formation curve to the right promoting the hydrate formation condition. This figure also indicates that the presence of C$_3$H$_8$ decreases the cricondenbar pressure while having little effect on the cricondentherm temperature; the two-phase (gas + liquid) region within the envelope shrinks.

![Figure 2. The impact of C$_3$H$_8$ on the hydrocarbon dew point and hydrate formation curves](image-url)
Similarly, Figure 3 presents the impact of 12.7 mol % iC$_4$H$_{10}$ on the dew point and hydrate formation curves for gases I and J of Table 1. This figure indicates that iC$_4$H$_{10}$ is like C$_3$H$_8$ is a hydrate promotor and shifts the hydrate curve to the right.

**Figure 3.** The impact of iC$_4$H$_{10}$ on the hydrocarbon dew point and hydrate formation curves
What is the Impact of Light Hydrocarbons on the Natural Gas Hydrate Formation Conditions?

Figure 4 presents the impact of 11.4 mol % nC₄H₁₀ on the dew point and hydrate formation curves for gases K and L of Table 1. This figure indicates that contrary to iC₄H₁₀, nC₄H₁₀ is a hydrate inhibitor and shifts the hydrate curve to the left. Both iC₄H₁₀ and nC₄H₁₀ lower the cricondentherm temperature and increase the cricondenbar pressure.

Figure 4. The impact of nC₄H₁₀ on the hydrocarbon dew point and hydrate formation curves.
What is the Impact of Light Hydrocarbons on the Natural Gas Hydrate Formation Conditions?

Figure 5 presents a summary of the calculated hydrate formation curves for sweet gas A of Table 1 (Continuous curve), and gases B (20 mol % H$_2$S), gas C (20 mol % CO$_2$), gas D (20 mol % N$_2$), gas F (17.8 mol % C$_2$H$_6$), gas H (12.7 mol % C$_3$H$_8$), gas J (12.7 mol % iC4H10), gas L (11.4 mol % nC$_4$H$_{10}$) (broken curves). For the cases studied, this figure clearly indicates that the impact of N$_2$ is much less than of H$_2$S and slightly less than of CO$_2$. Nitrogen, carbon dioxide, and nC$_4$H$_{10}$ depress the hydrate formation condition (shift the hydrate curves to the left). Between these three components, nC$_4$H$_{10}$ has the larger depression effect even though its mol % is smaller. While C$_2$H$_6$ has the same effect as CH$_4$ on the hydrate formation condition (no shift on the hydrate formation curve), C$_3$H$_8$, iC$_4$H$_{10}$, and H$_2$S promotes hydrate formation condition. Among these hydrate promoters, H$_2$S has the largest contribution even for only 10 mol %. Note that “Sweet Gas” refers to gas A in Table 1.

For training courses pertaining to this month’s topic, view our Gas Processing and Production & Completions engineering disciplines.

By: Dr. Mahmood Moshfeghian
What is the Impact of Light Hydrocarbons on the Natural Gas Hydrate Formation Conditions?

References

GET TO KNOW OUR EXPERTS

MR. JAMES F. LANGER is a registered professional chemical engineer in Texas and California. Jim worked for Hess as a Senior Process Engineering Advisor for 8 years. He is retired from Shell having worked 28 years as a Senior Staff Process Engineer, and Principal Technical Expert for Shell / Shell Global Solutions. He has had a global job for the past 20 years and had experience in offshore / onshore, shallow water / deepwater, heavy oil / light oil, water treating, and natural gas processing. He has been a project manager working field development projects through all the phase gates and stages. He frequently travels the globe assisting operations with process issues, and showing them how to unlock additional barrels through the application of production optimization. This has delivered over $2 billion in value to sponsors. Jim installed Shell’s smallest, most expensive gas plant. The project took 8 years and is located on Pacific Coast Highway in Huntington Beach California. Jim graduated with a BS in Chemical Engineering from UCLA and has an MBA from Pepperdine.

MR. HARVEY MALINO is an Instructor/Consultant for PetroSkills | John M. Campbell. He has more than 40 years’ experience in the chemical and hydrocarbon processing industries. During his 28 years with Union Carbide Corporation/ UOP, he held both technical and commercial positions. These included: Molecular Sieve Technical Manager- Design and Field Service; Licensing Manager for the Ethylene Oxide business; Area Sales and Marketing Manager for Southeast Asia; Business Manager for the Gas Processing Business Group; and, World Wide Sales Manager for the Gas Processing Business Group. Mr. Malino has lived and worked in New Hampshire, Maine, New York, Singapore and Chicago. He is currently based in Sullivans Island, South Carolina where he is President of his own consulting business. Mr. Malino is a registered Professional Engineer in the State of New Hampshire. He is a Senior Member of the Advisory Board of the Laurance Reid Gas Conditioning Conference in Norman, OK. Mr. Malino earned a BS in Chemical Engineering from the City College of New York; and, an MBA from Pace University in New York.

DR. MAHMOOD MOSHFEGHIAN is a Senior Technical Advisor and Senior Instructor for PetroSkills. He is the author of most Tips of the Month and develops technical software for PetroSkills. He has 40 years’ teaching experience in universities as well as for oil and gas industries. Dr. Moshfeghian joined John M. Campbell & Co. in 1990 as a part time consultant and then as full time instructor/consultant in 2005. Dr. Moshfeghian was Professor of Chemical Engineering at Shiraz University. Dr. Moshfeghian is a senior member of AIChE and has published more than 125 technical papers on thermodynamic properties and process engineering. Dr. Moshfeghian has presented invited papers at international conferences. He is a member of the Editorial Board for the International Journal of Oil, Gas, and Coal Technology. He holds a BS (74), an MS (75) and a PhD (78) in Chemical Engineering, all from Oklahoma State University.

See our website for a full list of our 200+ instructors.