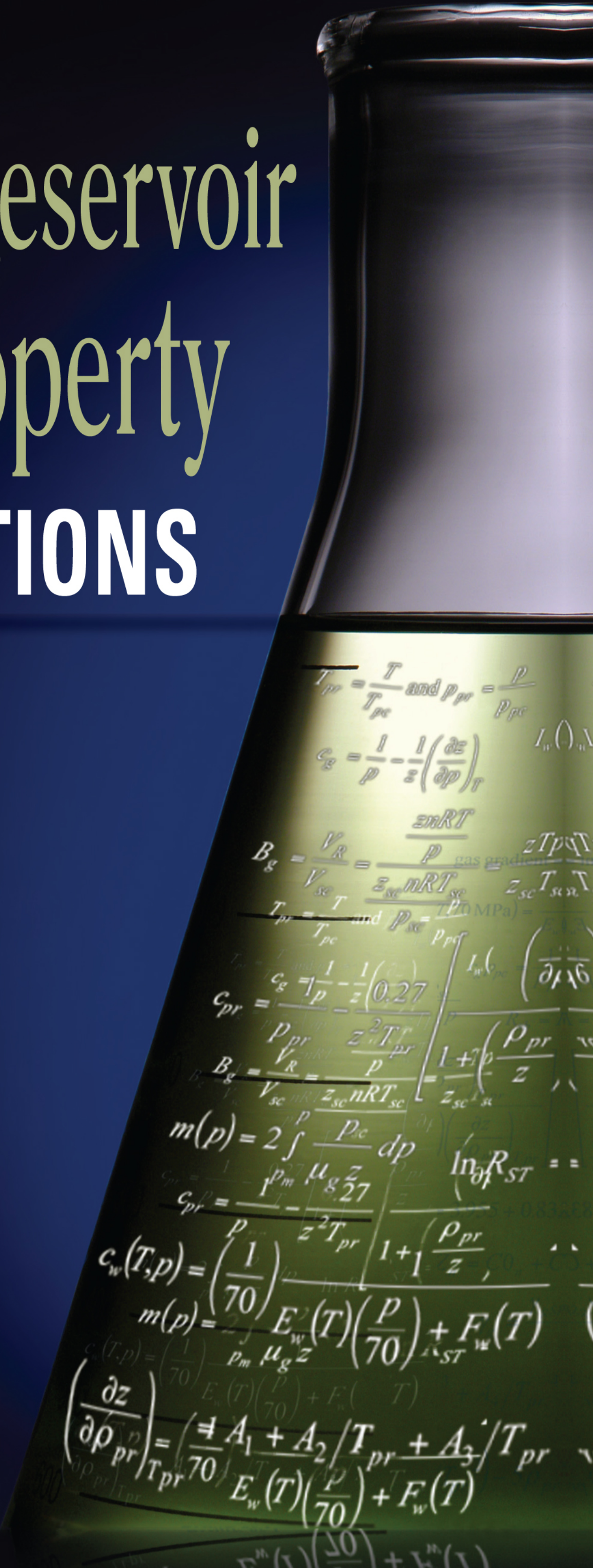
Petroleum Reservoir Fluid Property

CORRELATIONS



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We also recognize the host of petroleum engineering scientists who provided the many correlation equations that are presented in this book.

The contents of this work are the personal views and opinions of the authors and do not necessarily reflect the views or opinions of the authors' employers.

1

INTRODUCTION

This book is the result of several years of a fluid properties research project. Large sets of reservoir fluid property data were made available to the project from a variety of PVT report databases. The results of a few research projects were included. Generally the various sets of data covered the full range of conditions and properties that might be found in practice. Unfortunately, no data for naturally occurring petroleum fluids are currently available for high-pressure, high-temperature (HPHT) conditions that have recently become of interest.

The intent was not to create new correlations but to evaluate fluid property correlation equations proposed in the petroleum literature to determine the best-in-class for each fluid property.

The focus was on correlations that required readily available field data. Correlation equations that require composition or other laboratory data were not considered.

The assumption was that if composition is known, a laboratory report is available and therefore various applicable fluid properties have been measured in addition to composition.

Evaluation Criteria

The evaluations were based on *average relative error* and *average absolute relative error* as defined below.

Average relative error, ARE, %:

$$ARE = \frac{100}{n}\sum_{i=1}^{n}\frac{y_{calculated} - y_{measured}}{y_{measured}} \tag{1.1}$$

Average absolute relative error, AARE, %:

$$AARE = \frac{100}{n}\sum_{i=1}^{n}\left|\frac{y_{calculated} - y_{measured}}{y_{measured}}\right| \tag{1.2}$$

ARE is simply the arithmetic average of the relative errors. It is a measure of the bias of the correlation; a value of zero indicates a random deviation of the measured values around the correlation. If the deviations in the laboratory-measured values of a property are truly random, a correlation with a very small value of ARE will result in predicted values as accurate, or possibly more accurate, than laboratory measurements.

AARE is the arithmetic average of the absolute values of the relative errors; an indication of both the precision of the correlation and the accuracy of the data. A small value of AARE denotes a good correlation based on good data. A large value of AARE could mean a poor quality correlation (inadequate functional form). However, the situation of several different correlation equations all with large values of AARE most likely indicates poor quality data.

ARE and AARE are just more robust versions of the more familiar statistical measures, mean and standard deviation. Their use here is justified because they require fewer assumptions regarding the distribution of the measurement errors.

Evaluation Procedures

The several correlation equations proposed in the petroleum literature for each fluid property were evaluated using appropriate sets of data. These sets of data are fairly large and are based on reservoir fluid samples with worldwide origins. The results of these evaluations are presented in tabular form. The best-in-class correlation equations that are recommended are reproduced in the text.

Each data set then was sorted and sliced into subsets to determine if the better of the correlation equations were valid throughout the full range of field conditions and fluid properties that might occur. The results are presented graphically. This procedure was followed for nearly all of the fluid properties that were studied. In some cases, the evaluation of the AREs and AAREs of these subsets served to determine the best-in-class correlation equations.

The evaluations of these subsets were *not* meant to imply that one set of correlation equations should be used over part of the range of values of an independent variable and that another set of correlation equations should be used over the remainder of values of that independent variable. These evaluations were made to ensure that the correlation equations selected as best-in-class are valid across the full range of the data.

Comments

Please note that each chapter has its own nomenclature section with units for each variable specified. The units for each chapter may be different. For instance, chapter 2 uses T for temperature with units of °R; chapter 3 uses T for temperature with units of °F; and chapter 4 uses T for temperature with units of °C, unless otherwise noted.

Nomenclature

n	Number of measurements
$y_{calculated}$	Value of a fluid property calculated with correlation equation
$y_{measured}$	Value of a fluid property measured in laboratory

2

PROPERTIES OF RESERVOIR GASES

The reservoir gas properties of interest in petroleum engineering are the following:

1. Gas z-factors
2. Gas densities
3. Gas formation volume factors
4. Gas viscosities
5. Coefficient of isothermal compressibility of gas
6. Real gas pseudopressures

Gas Correlations

There are only two correlations in this chapter: gas z-factor and gas viscosity. All other gas properties are calculated from gas z-factors. In fact, the value of gas density required in the gas viscosity correlation is calculated using gas z-factor.

Two choices are available to describe a gas of interest: gas specific gravity or gas composition. The assumption is that gas composition is not available, i.e., results of a laboratory study are not available. The accuracies of the predicted fluid properties are about the same regardless of the choice between correlations based on gas specific gravity or correlations based on gas composition. Gas specific gravity can readily be calculated given gas composition.

The input data for the correlation equations in this chapter will be reservoir gas specific gravities and the compositions of nonhydrocarbons in mole fractions, along with reservoir temperatures and selected pressures. If the reservoir gas is a dry gas, i.e., no petroleum condensate at the surface, the specific gravity and compositions of the nonhydrocarbons of the surface gas can be used to represent the reservoir gas. If there is petroleum condensate with the gas at the surface, a recombination calculation given by Gold et al. can be used to calculate the specific gravity of the reservoir gas.[1] If the reservoir pressure is less than the dew point pressure of the reservoir gas, the Gold et al. procedure will give an estimate of the specific gravity of the reservoir gas but will not represent the total (gas plus condensate) fluid in the reservoir. Values of compositions of nonhydrocarbons measured with the surface gas can be used as reasonable estimates of these compositions for the reservoir gas.

The best available gas z-factor correlation equations/procedure is very accurate; the results are as accurate as normally reported in petroleum industry fluid property studies (PVT reports). The various properties calculated using these z-factors are also of good accuracy. The best available gas viscosity correlation does not have the precision of the best z-factor correlation but does give results of sufficient accuracy for use in petroleum engineering calculations.

Gas z-Factors

Gas z-factors are also called *gas compressibility factors* or *gas deviation factors*. Gas z-factors are defined as

$$z = \frac{V\text{ actual}}{V\text{ ideal}} \tag{2.1}$$

and are used to adjust the ideal gas equation to calculate real gas volumes.

$$pV = znRT \tag{2.2}$$

Standing and Katz proposed a graphical correlation of gas z-factors plotted against pseudoreduced pressures and pseudoreduced temperatures.[2]

These pseudoreduced properties are defined as

$$T_{pr} = \frac{T}{T_{pc}} \text{ and } p_{pr} = \frac{p}{p_{pc}} \tag{2.3}$$

where absolute values of pressures and temperatures are used. The pseudocritical pressures and pseudocritical temperatures for equation (2.3) must be obtained from correlations. These pseudocritical properties do not in any way reflect the true critical pressures and critical temperatures of the gas mixture; they are simply parameters used in the z-factor correlation.

Thus, obtaining values of z-factors from the Standing-Katz correlation is a two-step procedure. First, values of pseudocritical pressure and pseudocritical temperature must be obtained for the gas of interest using correlation equations. Second, equations must be used to represent the Standing-Katz chart, which was originally proposed as a graphical correlation.

Algebraic representations of the Standing-Katz graphical z-factor correlation

Takacs evaluated eight commonly used mathematical descriptions of the Standing-Katz graph.[3] Two methods, Gray-Sims and Carlile-Gillett, gave very good representations of the graph but were eliminated from this study. The first used a table lookup procedure and the second used a set of equations that have discontinuities, which could cause problems if the methods are used in calculations that require trial and error.[4]

Takacs showed that both the Hall-Yarborough and Dranchuk-Abou-Kassem (DAK) equations gave very good representations of the Standing-Katz graph.[5] However, both sets of equations require complex trial-and-error solutions for z-factors, so two other methods in which the z-factors are calculated directly were evaluated: the Papay equation and the Standing modification of the Beggs-Brill equation.[6]

Subsequent to Takacs' work, Ohirhian proposed a set of equations representing the Standing-Katz chart. These were not evaluated due to the many discontinuities among the equations.[7]

Recently Hall and Iglesias-Silva proposed modifications to the Hall-Yarborough equation that improved the procedure for very low values of pseudoreduced temperature.[8] This new equation was not evaluated because the applicable temperatures were below normal reservoir conditions.

Estimation of pseudocritical properties using gas specific gravities and nonhydrocarbon contents

Gas specific gravity is defined as the density of the gas divided by the density of dry air, both densities measured or calculated at the same temperature and pressure, usually 60°F and atmospheric pressure. Several authors have proposed equations for estimation of pseudocritical properties using gas specific gravities. The Standing and the Sutton equations require modification when nonhydrocarbon components (hydrogen sulfide, carbon dioxide, and nitrogen) are present.[9] Both recommended the Wichert-Aziz correlation for adjusting the pseudocritical properties for hydrogen sulfide and carbon dioxide; no adjustment for nitrogen is available for these correlations.[10] Elsharkawy et al. did not mention adjustments for presence of nonhydrocarbons; however, Elsharkawy and Elkamel did take the compositions of the nonhydrocarbons into account.[11] The Piper et al. correlation equations include adjustments for all three nonhydrocarbons.[12] Londono et al. did not offer any adjustments for nonhydrocarbon components.[13]

Evaluation of proposed correlation equations for gas z-factors based on gas specific gravities and nonhydrocarbon contents

Values of gas z-factors from a large set of petroleum industry fluid property reports (PVT reports) were assembled for use in evaluating the several correlations. The z-factors were measured at reservoir temperatures for various pressures. All gases were samples of reservoir gases; no surface gases, artificial gas mixtures, or pure gases were included in the data set. The gases were gas condensates, gases liberated from oils at pressures below the bubblepoint pressure of the oil, and some dry gases. The z-factors of the gas condensates were measured in constant composition expansion and constant volume depletion experiments. The z-factors of solution gas were measured in differential liberation experiments of the oils. The dry gas z-factors were measured in constant composition expansion experiments. The samples were obtained from reservoirs throughout the world.

Table 2–1 gives the ranges of the variables in the data used to evaluate the correlations that use gas specific gravity to estimate the gas z-factors. There are 6,000 lines of data that include the variables listed in the table. The ranges of temperatures, pressures, and gas specific gravities cover nearly all of the reservoir situations encountered in the petroleum industry. More than three-quarters of the gases contained some nonhydrocarbon components; 1,214 gas samples had 5 mole percent (mol%) nonhydrocarbons or more, and 145 samples contained more than 50 mol% nonhydrocarbons.

Table 2–1. Description of the data set used to evaluate z-factor correlations based on gas specific gravity (6,000 lines of data from 1,434 PVT reports of worldwide origins)

Variable	Minimum	Median	Mean	Maximum
Pressure, psia	14.7	2,750.0	3,415.2	18,050.0
Temperature, °F	35.0	220.0	217.2	418.0
Gas specific gravity	0.5590	0.8301	0.9002	2.2380
Hydrogen sulfide, mole fraction	0.0000	0.0000	0.0105	0.5137
Carbon dioxide, mole fraction	0.0000	0.0085	0.0260	0.7198
Nitrogen, mole fraction	0.0000	0.0020	0.0110	0.4954
Gas z-factor	0.4000	0.9340	0.9959	2.3500

Piper et al. used approximately one-quarter of these data to develop their correlations. Apparently none of the other researchers had access to any of these data.

Gas z-factors calculated with several correlations were compared with the laboratory-measured z-factors. The results are given in table 2–2. The evaluation criteria, average relative error and average absolute relative error, are defined in chapter 1. A positive value of ARE indicates that the calculated value is larger than the measured value.

Table 2–2. Average relative errors of several gas z-factor correlations based on gas specific gravity compared with the data set of table 2–1

Pseudocritical Properties Correlation	z-factor Correlation	Average Relative Error, %	Average Absolute Relative Error, %
Piper et al. (1999)	Hall-Yarborough (1973)	0.307	1.562
Piper et al. (1999)	Dranchuk-Abou-Kassem (1975)	0.296	1.590
Piper et al. (1999)	Brill-Beggs (1974), (1977)[†]	–0.253	1.731
Sutton (2007)*	Hall-Yarborough (1973)	–0.548	1.621
Sutton (2007)*	Dranchuk-Abou-Kassem (1975)	–0.422	1.511
Elsharkawy-Elkamel (2000)	Dranchuk-Abou-Kassem (1975)	0.010	1.821
Elsharkawy-Elkamel (2000)	Hall-Yarborough (1973)	–0.014	1.812
Standing (1977)*	Hall-Yarborough (1973)	–1.084	1.968
Standing (1977)*	Dranchuk-Abou-Kassem (1975)	–1.010	1.949
Standing (1977)*	Brill-Beggs (1974), (1977) [†]	–1.609	2.320
Elsharkawy et al. (2000)	Hall-Yarborough (1973)	1.200	2.434
Elsharkawy et al. (2000)	Dranchuk-Abou-Kassem (1975)	1.185	2.452
Sutton (1985)*	Dranchuk-Abou-Kassem (1975)	1.746	2.878
Sutton (1985)*	Hall-Yarborough (1973)	1.769	2.887
Londono et al. (2005)	Dranchuk-Abou-Kassem (1975)	–1.044	3.254
Piper et al. (1999)	Nishuimi-Saito (2005)[‡]	5.646	5.911
Londono et al. (2005)	Nishuimi-Saito (2005) [‡]	5.157	6.432
Piper et al. (1999)	Papay (1968)	15.113	16.166

* Included Wichert-Aziz correlation for hydrogen sulfide and carbon dioxide [†] From Standing (1977) [‡] Nishuimi-Saito equation with coefficients derived by Londono et al.

Table 2–2 shows that the Hall-Yarborough representation of the Standing-Katz z-factor graph gives slightly better results than the DAK equations. Figure 2–1 also shows that the Hall-Yarborough equations give slightly better results with the data set divided into 11 approximately equal-sized subsets according to gas specific gravity.

The comparisons in figure 2–1 used the Piper et al. equations to calculate the pseudocritical properties. Comparisons of the Hall-Yarborough equations with the DAK equations using other pseudocritical correlations and with the data sorted by temperatures and pressures gave similar results. Elsharkawy et al. reported a similar conclusion.[14]

An evaluation of the results of these procedures using routine statistical techniques showed that there is no difference (in a statistical sense) between the results from Hall-Yarborough and DAK. Both sets of equations require very difficult trial-and-error solutions; occasionally the Hall-Yarborough solutions did not converge using either the Hall-Yarborough procedure or the fairly robust Dake procedure.[15] Thus, the DAK equations will be used in further evaluation of the correlations for determining pseudocritical properties.

The Standing modification of the Beggs-Brill equations (SBB) gave results nearly as good as the Hall-Yarborough and DAK methods. This SBB procedure is recommended for use in situations in which trial-and-error solutions are not feasible. The Papay equation gave very poor results, as was reported by Takacs.

Table 2–2 also shows that the Piper et al. correlation equations and the Sutton (2007) correlation (his equations 10 and 11) for pseudocritical properties give the more nearly accurate predictions of z-factors. This table gives results for the entire data set. An examination of subsets of these data will show how well the correlations predict z-factors across the ranges of temperature, pressure, and gas specific gravity that might be encountered in application.

Table 2–3 gives a comparison of the methods of estimating the pseudocritical properties for four subsets of the data containing appreciable amounts of nonhydrocarbon components. The Sutton (2007), Piper et al., and Standing correlations are approximately as accurate for gases with nonhydrocarbon components as for sweet gases. The poor results for the Sutton (1985) correlation are not surprising; the gas samples in his data set had no hydrogen sulfide and limited amounts of carbon dioxide and nitrogen. Elsharkawy et al. did not mention using the Wichert-Aziz adjustments for hydrogen sulfide and carbon dioxide; however, these adjustments were tried and did not improve the results.

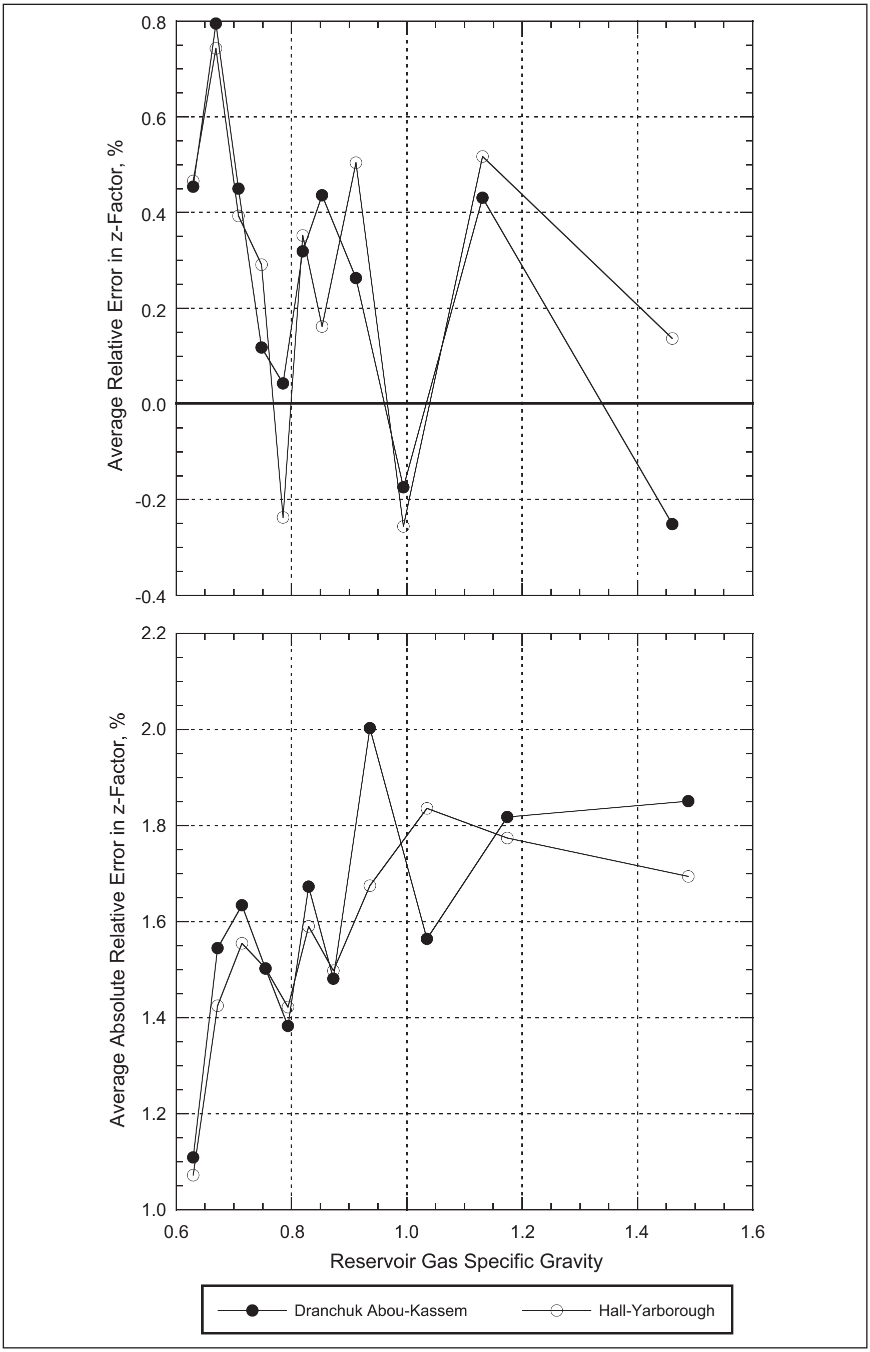

Fig. 2–1. Comparisons of relative errors in z-factors between Dranchuk-Abou-Kassem and Hall-Yarborough equations. All calculations used the Piper et al. equations to calculate the pseudocritical properties.

Table 2–3. Average relative errors of several gas z-factor correlations based on gas specific gravity compared with gases with nonhydrocarbon components from the data set of table 2–1. All calculations used Dranchuk-Abou-Kassem z-factor equations.

Correlation	**H_2S > 5 mol%***		**CO_2 > 5 mol%†**		**N_2 > 5 mol%‡**		**$H_2S + CO_2 + N_2$ > 10 mol%§**	
	Average relative error	Average absolute relative error	Average relative error	Average absolute relative error	Average relative error	Average absolute relative error	Average relative error	Average absolute relative error
Sutton (2007)	–0.7	1.9	0.1	1.7	–0.8	2.1	–0.2	1.9
Piper et al. (1999)	1.9	2.9	0.6	1.8	0.3	2.7	0.6	2.2
Standing (1977)	–0.5	2.0	–0.3	2.2	–1.5	2.6	–0.4	2.2
Elsharkawy-Elkamel (2000)	1.3	2.8	–1.0	2.8	–3.9	4.7	–2.1	4.0
Londono et al. (2005)	9.7	13.5	–1.2	6.0	–2.5	6.5	–0.1	8.4
Elsharkawy et al. (2000)	16.6	17.4	2.9	4.4	–0.1	4.7	4.8	7.9
Sutton (1985)	30.4	30.6	6.6	6.8	1.7	4.5	11.5	12.6

* 177 lines of data † 715 lines of data ‡ 336 lines of data § 681 lines of data

Figures 2–2 through 2–4 show the relative errors for Piper et al., Sutton (2007), Standing, and Elsharkawy-Elkamel correlations for 11 approximately equal-sized subsets of the data sorted and then sliced on gas specific gravity, reservoir pressure, and reservoir temperature.

The Piper et al. correlation equations give the best results across the full range of the data. The Sutton (2007) equations are essentially equivalent. The deviations of the Standing equations at higher values of gas specific gravity are not surprising; he had limited data for high specific gravity gases.

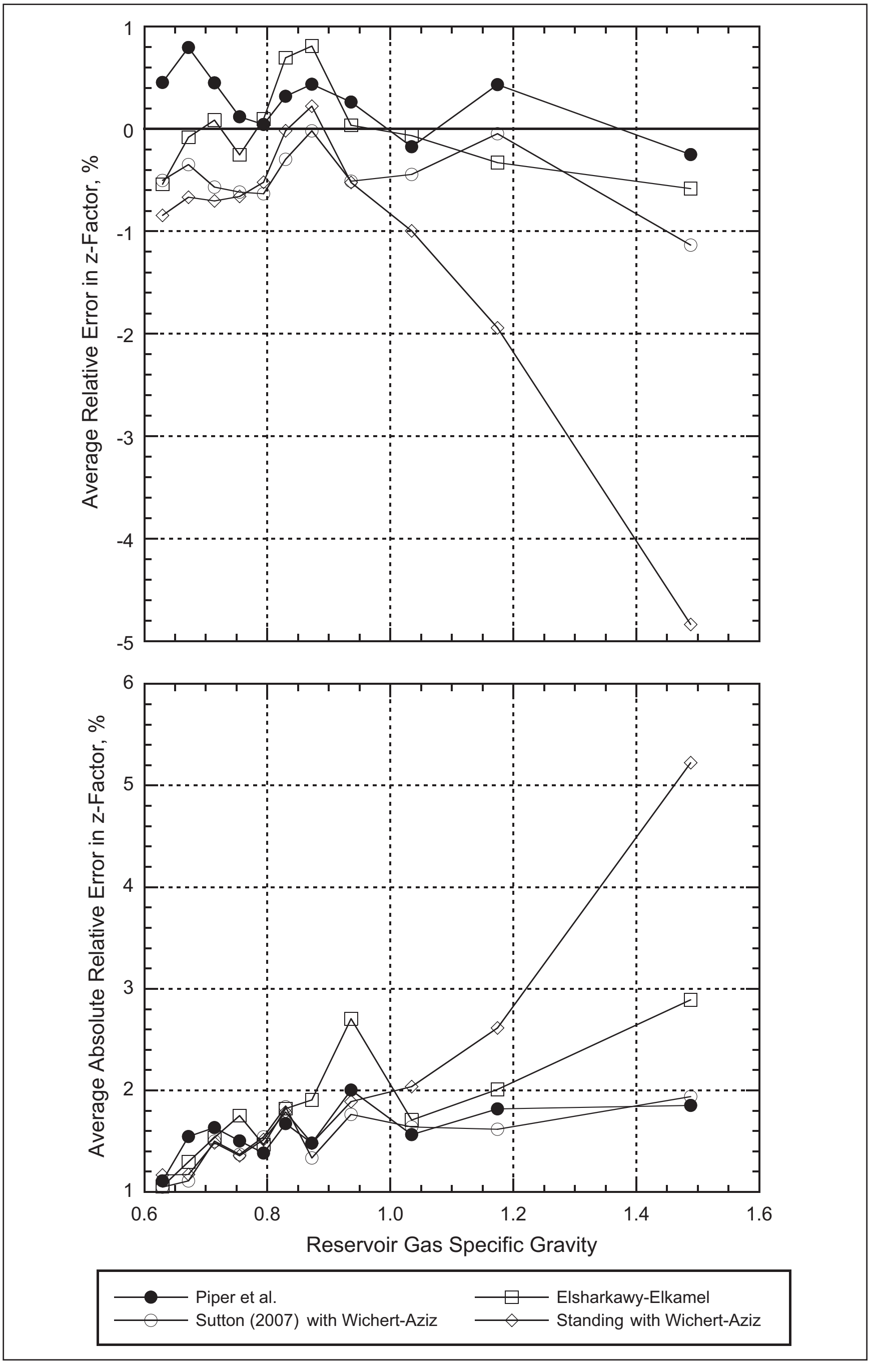

Fig. 2–2. Comparisons of relative errors in gas z-factor correlations using data of table 2–1 split by gas specific gravity. All calculations used Dranchuk-Abou-Kassem z-factor equations.

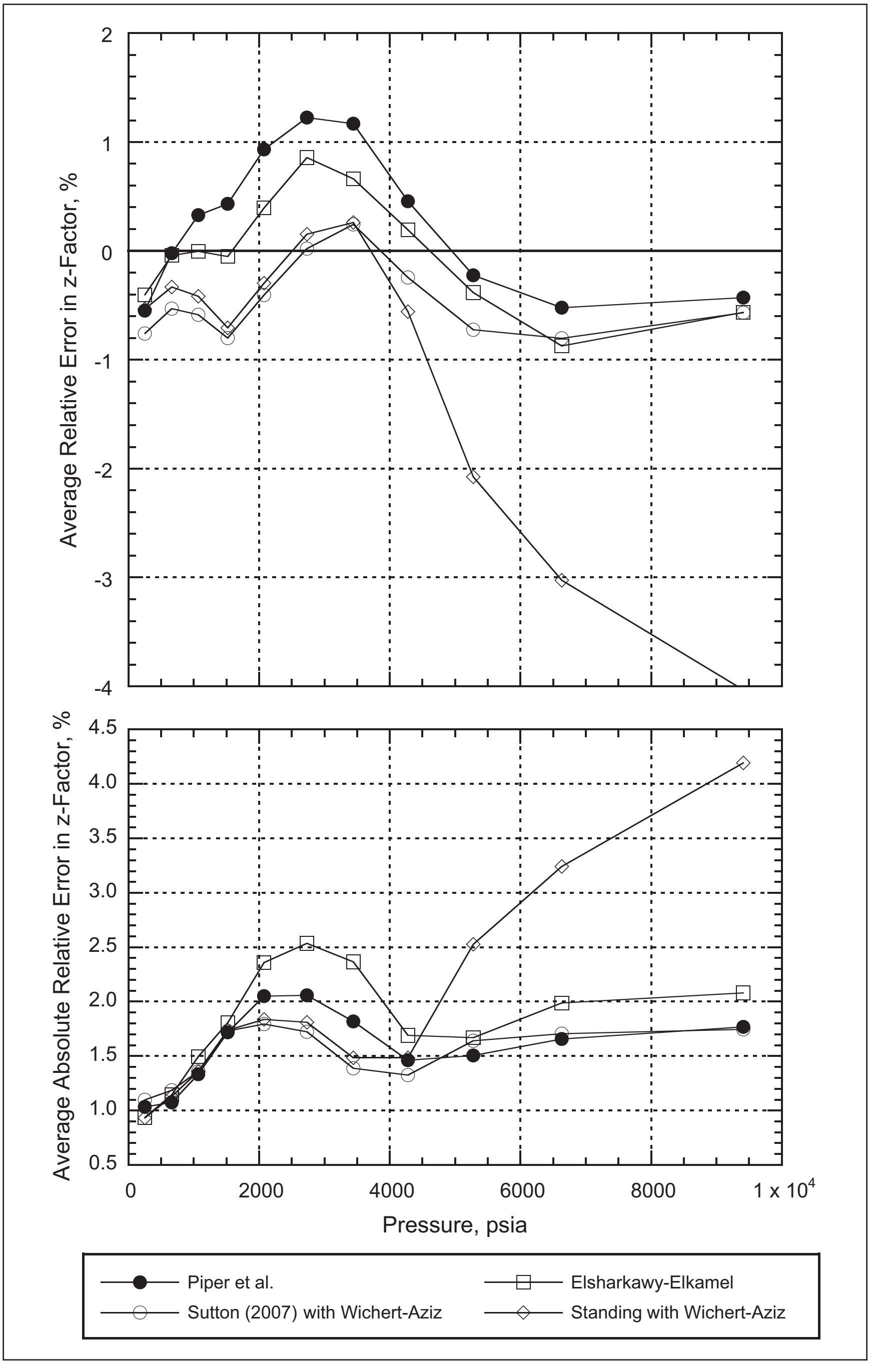

Fig. 2–3. Comparisons of relative errors in gas z-factor correlations using data of table 2–1 split by pressure. All calculations used Dranchuk-Abou-Kassem z-factor equations.

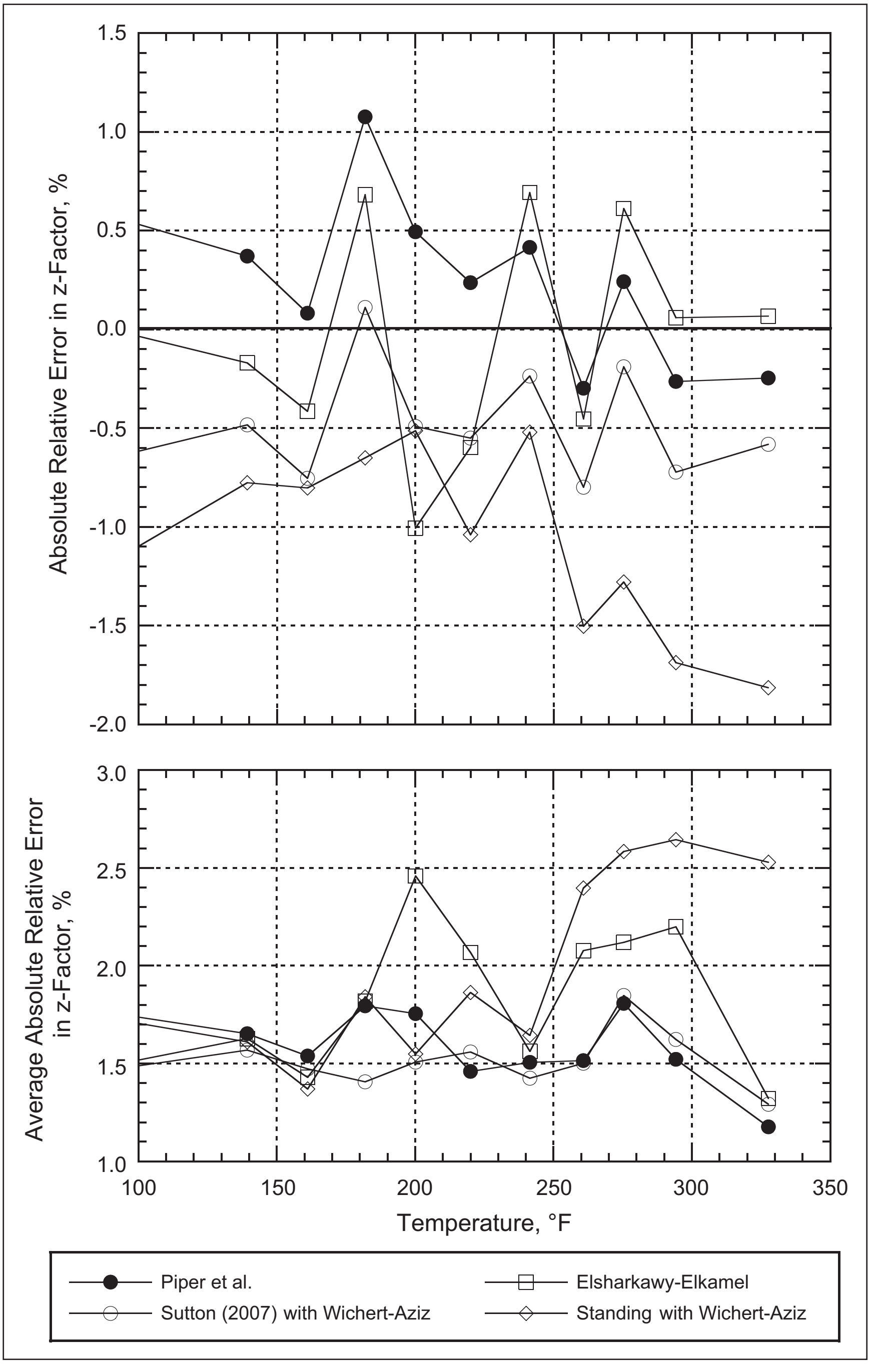

Fig. 2–4. Comparisons of relative errors in gas z-factor correlations using data of table 2–1 split by temperature. All calculations used Dranchuk-Abou-Kassem z-factor equations.

Recommended correlation equations/procedures for estimation of gas z-factors based on gas specific gravities and nonhydrocarbon contents

The recommended equations for estimation of the pseudocritical properties are given in equations (2.4)–(2.6).[16]

$$T_{pc} = \frac{K^2}{J} \text{ and } p_{pc} = \frac{T_{pc}}{J} \tag{2.4}$$

$$J = \alpha_o + \sum_{i=1}^{3} \alpha_i y_i \left(\frac{T_c}{p_c} \right)_i + \alpha_4 \gamma_g + \alpha_5 \gamma_g^{\,2} \tag{2.5}$$

$$K = \beta_o + \sum_{i=1}^{3} \beta_i y_i \left(\frac{T_c}{\sqrt{p_c}} \right)_i + \beta_4 \gamma_g + \beta_5 \gamma_g^{\,2} \tag{2.6}$$

where

i	α_i	β_i
0	1.1582E–01	3.8216E+00
1	–4.5820E–01	–6.5340E–02
2	–9.0348E–01	–4.2113E–01
3	–6.6026E–01	–9.1249E–01
4	7.0729E–01	1.7438E+01
5	–9.9397E–02	–3.2191E+00

The summations in each of equations (2.5) and (2.6) are hydrogen sulfide, carbon dioxide, and nitrogen, in that order.

The recommended equations for representation of the Standing-Katz graphical z-factor correlation are from DAK, shown in equation (2.7).[17]

$$\begin{aligned} z = 1 &+ \left(A_1 + A_2/T_{pr} + A_3/T_{pr}^{\,3} + A_4/T_{pr}^{\,4} + A_5/T_{pr}^{\,5}\right)\rho_{pr} \\ &+ \left(A_6 + A_7/T_{pr} + A_8/T_{pr}^{\,2}\right)\rho_{pr}^{\,2} \\ &- A_9\left(A_7/T_{pr} + A_8/T_{pr}^{\,2}\right)\rho_{pr}^{\,5} \\ &+ A_{10}\left(1 + A_{11}\rho_{pr}^{\,2}\right)\left(\rho_{pr}^{\,2}/T_{pr}^{\,3}\right)\text{EXP}\left(-A_{11}\rho_{pr}^{\,2}\right) \end{aligned} \tag{2.7}$$

where

$$\rho_{pr} = 0.27\left[p_{pr}/\left(zT_{pr}\right)\right] \tag{2.8}$$

where

i	A_j
1	0.3265
2	–1.0700
3	–0.5339
4	0.01569
5	–0.05165
6	0.5475
7	–0.7361
8	0.1844
9	0.1056
10	0.6134
11	0.7210

This combination of correlation equations should predict values of z-factors from measured values of gas specific gravities to a precision equal to normal experimental accuracy.

Convergence of the trial-and-error correlation procedures

The best of the z-factor correlation procedures, Hall-Yarborough and Dranchuk-Abou-Kassem, require very difficult trial-and-error calculations. The Hall-Yarbrough procedure failed to converge in about 0.5% of the lines of data listed in table 2–1 even though the relatively robust calculation procedure recommended by Dake was used.[18]

A "brute-force" procedure for solving the DAK trial-and-error calculations converged for every one of the 6,000 lines of data in the z-factor data set. This procedure is as follows. Start with a first trial value of z-factor = 1.0. Calculate the pseudoreduced density of the gas with equation (2.8) and then calculate z-factor with equation (2.7). Calculate an error function, *delta*, as the calculated z-factor minus the trial value of z-factor. If p_{pr} is less than or equal to 2.0, the next trial value of z-factor will be the previous trial value plus *delta*. If p_{pr} is greater than 2.0 and less than or equal to 3.0, the next trial value will be the previous trial value plus *delta*/2.0. If p_{pr} is greater than 3.0 and less than or equal to 6.0, the next trial value will be the previous trial value plus *delta*/3.0. If p_{pr} is greater than 6.0, the next trial value will be the previous trial value plus *delta*/5.0. Occasionally this procedure requires as many as 20 iterations to converge, especially at high pressures, but often less than five iterations are required.

Gas Densities and Pressure Gradients

Gas densities are calculated with equation (2.9).

$$\rho_g = \frac{pM}{zRT} \tag{2.9}$$

If a value of 10.732 psia cu ft/lb-mole °R is used for universal gas constant, *R*, then pressure should be in psia, temperature should be in °R, and the molecular weight of the gas can be calculated with equation (2.10).

$$\gamma_g = \frac{M}{29} \tag{2.10}$$

Values of z-factor can be obtained with equations (2.4)–(2.8). The resulting density will be in pounds per cubic feet (lb/cu ft).

Equation (2.9) is not a correlation; it is simply a rearrangement of the real gas equation, equation (2.2). Also, equation (2.10) is not a correlation; it follows from the definition of gas specific gravity. Thus, the results of the use of equation (2.9) are as accurate as the values of z-factor used in the calculations.

Gas pressure gradients are calculated with

$$\text{gas gradient} = \frac{g}{g_c}\frac{\rho_g}{144}\frac{psi}{ft} \tag{2.11}$$

where *g* and g_c have the same numerical value at sea level (in the set of units used here) and thus are not usually shown in the equation. The resulting pressure gradient has units of pounds per square inch per feet (psi/ft).

Gas Formation Volume Factors

Gas formation volume factor is defined as the volume of gas at reservoir temperature and pressure required to produce one standard volume of gas at the surface. The equation for gas formation volume factor is derived using the real gas equation, equation (2.2), to calculate the gas volumes for reservoir conditions and standard conditions.

$$B_g = \frac{V_R}{V_{sc}} = \frac{\dfrac{znRT}{p}}{\dfrac{z_{sc}nRT_{sc}}{p_{sc}}} = \frac{zTp_{sc}}{z_{sc}T_{sc}p} \tag{2.12}$$

The units of B_g resulting from the use of equation (2.12) are either reservoir cubic feet per standard cubic feet (res cu ft/scf) or cubic meters per cubic meter; appropriate conversion factors can be used to alter the units. The z-factor at standard conditions, z_{sc}, is usually assumed to be 1.0, and the z-factor at p and T can be calculated with equations (2.4)–(2.8).

Equation (2.12) is not a correlation. It was derived directly from the definition of gas formation volume factor. Thus, the results of the use of equation (2.12) are as accurate as the values of gas z-factors used in the calculations.

Gas Viscosities

Gas viscosity, also called the *coefficient of viscosity of a gas*, is a measure of the resistance to flow exerted by a gas. Viscosities in units of *centipoise* are most often used in the petroleum industry. These are called *dynamic viscosities*. The units of centipoise are grams of mass per 100 second-centimeters (g mass/100 sec cm). Sometimes viscosity is reported in *centistokes*. These are called *kinematic viscosities* and have units of square centimeters per 100 seconds (cu cm/100 sec cm). Thus, the relationship between kinematic and dynamic viscosity is

$$\text{kinematic viscosity}\,\nu = \frac{\text{dynamic viscosity}\ \mu}{\text{density}\ \rho_g} = \frac{\text{centipoise}}{\text{g/cc}} = \text{centistoke} \tag{2.13}$$

where the density in grams per cubic centimeters (g/cc) is determined at the pressure and temperature of interest.

Carr, Kobayashi, and Burrows proposed graphical correlations to estimate gas viscosities.[19] One graph allows calculation of the gas viscosity at atmospheric pressure as a function of gas molecular weight and temperature, with adjustments for nonhydrocarbon content of the gas. Another graph relates the gas viscosity at the pressure and temperature of interest to the gas viscosity at atmospheric pressure using pseudoreduced temperatures and pseudoreduced pressures.

Standing wrote equations to represent the graphical correlation for gas viscosity at 1 atmosphere pressure, and Dempsey prepared an equation to represent the graphical correlation for the ratio of gas viscosity at pressure and temperature to the gas viscosity at 1 atmosphere pressure.[20]

Two similar sets of correlation equations were proposed by Jossi, Stiel, and Thodos and by Dean and Stiel.[21] Pseudocritical pressures, pseudoreduced temperatures, pseudoreduced densities, and molecular weights of the gases are required.

Lee, Gonzalez, and Eakin proposed a set of correlation equations for gas viscosities as a function of temperature, gas molecular weight, and gas density.[22] They referenced a similar set of correlations from Lee, Starling, Dolan, and Ellington.[23]

Elsharkawy proposed corrections to the viscosity calculated with the Lee, Gonzalez, Eakin equations to account for the presence of nitrogen, carbon dioxide, and heptanes plus.[24]

Poling et al. reported a set of correlation equations that require pseudocritical temperatures and pseudocritical pressures.[25]

Londono et al. collected a large set of gas viscosity data and prepared a new set of gas viscosity correlation equations (their "implicit model").[26] They started with an equation for gas viscosity at 1 atmosphere pressure correlated with gas specific gravity and temperature. This was adjusted to reservoir conditions with another equation based on gas density with coefficients based on temperature.

Sutton proposed equations that combine the method reported by Poling et al. with the form of the Lee et al. equations.[27]

Evaluation of proposed correlation equations for gas viscosities

The data set compiled by Gonzalez et al. has extensive gas viscosity measurements for only eight naturally occurring gases.[28] These appear to be the only reservoir gas viscosity data available in the literature. Table 2–4 gives a description of these data. They were used to evaluate the gas viscosity correlation equations. All eight of these gases have gas compositions indicating that they are dry gases. No gas viscosity data for gas condensates appear to be available.

Lee, Gonzalez, and Eakin used approximately one-half of these data to develop their correlation equations.[29] Londono et al. used all of these data plus a large amount of data from pure component, binary, and ternary mixtures in preparing their correlation equations.[30] Sutton used all of these data plus some data from pure component and binary mixtures.[31] Sutton also used some "laboratory measured gas/condensate viscosity" data that apparently are not in the petroleum literature.

Table 2–4. Description of the data set used to evaluate gas viscosity correlations (243 data points for eight naturally occurring petroleum gases)

Variable	Minimum	Median	Mean	Maximum
Pressure, psia	200	1,500	2,169	8,000
Temperature, °F	100	220	227	340
Gas specific gravity	0.572	0.658	0.656	0.769
Hydrogen sulfide, mole fraction	0.000	0.000	0.000	0.000
Carbon dioxide, mole fraction	0.002	0.009	0.012	0.032
Nitrogen, mole fraction	0.000	0.005	0.016	0.052
Gas viscosity, cp	0.0125	0.0169	0.0193	0.0510

The Carr et al. gas viscosity correlation equations require values of pseudocritical temperature and pseudocritical pressure.[32] For this analysis, these were calculated from gas specific gravity and nonhydrocarbon content with the Piper et al. correlations.[33]

The Jossi et al. equations and the Dean and Stiel equations were not evaluated because each procedure has a discontinuity at a pseudoreduced temperature of 1.5. This is near the middle of the pseudoreduced temperatures expected for petroleum reservoirs at moderate to higher pressures.[34]

Both of the Lee et al. gas viscosity correlation equations and the Sutton correlation equations require values of gas density and gas molecular weight. Gas densities were calculated using z-factors from the Piper et al. and DAK equations. Gas molecular weight was calculated using equation (2.10).

The Londono et al. gas viscosity correlation equations require values of gas densities. These were calculated with z-factors from the Piper et al. and DAK correlations for this evaluation.

The Piper et al. and DAK equations were used in this evaluation because they gave the best results in the evaluation of the z-factor correlation equations/procedures.

The gas samples of the set of gas viscosity data did not have sufficient nonhydrocarbon or heptanes plus composition to allow evaluation of the Elsharkawy equations.[35]

The average relative errors for the gas viscosity correlations described above are given in table 2–5.

Table 2–5. Relative errors to published gas viscosity correlations compared to data described in table 2–4

Viscosity correlation*	Average relative error, %	Average absolute relative error, %
Lee et al. (1966)*	–1.60	2.26
Londono et al. (2005)*	–2.66	3.08
Sutton (2007)*	2.05	3.10
Poling et al. (2001)*	–0.61	3.34
Lee et al. (1964)*	–3.23	3.70
Carr et al. (1954)†	–9.81	9.81

* Piper et al. correlation was used to estimate pseudocritical properties and Dranchuk-Abou-Kassem equation was used to estimate z-factor for density calculations as needed. † Standing and Dempsey equations represented the Carr, Kobayashi, and Burrows graphs.

Figures 2–5 through 2–7 show the results of the best four correlations, with the data set sorted and divided into approximately equal subsets based on temperature, pressure, and gas specific gravity.

The data were also sorted and split by gas viscosity. The results looked very much like figure 2–6.

These figures show that the Lee, Gonzalez, and Eakin (LGE) equations give the best results with this set of data. Londono et al. reported the same results.

The correlations look fairly good across these ranges of temperatures, pressures, and gas specific gravities. However, the maximum pressure of the data set is well below pressures often encountered in petroleum reservoirs. Also, there is no way of determining how well the correlations predict gas viscosities for gas condensates. Calculations with the LGE equations were extended to very high pressures and high specific gravities. No reservoir gas data exist to compare with the results, but the calculated values were well behaved and looked reasonable.

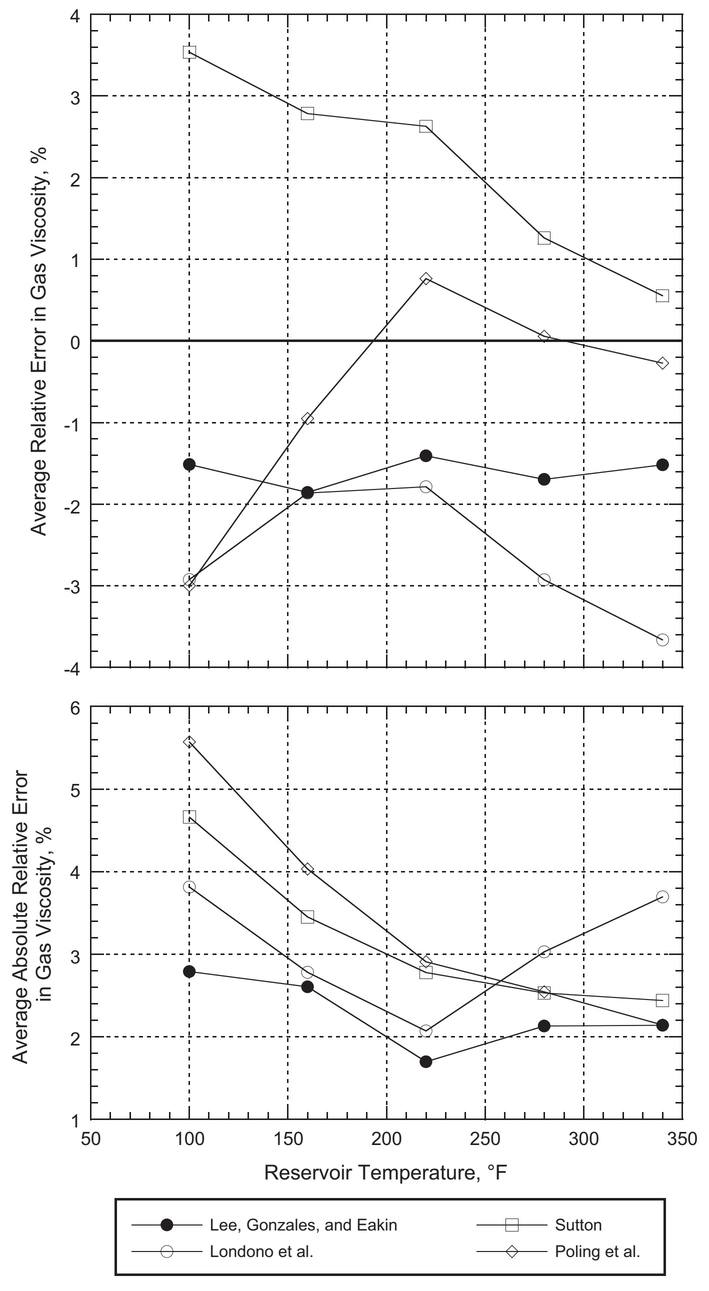

Fig. 2–5. Comparisons of relative errors in gas viscosity correlations using data of table 2–4 sorted and split by temperature. All calculations used the Piper et al. pseudocritical temperature and pseudocritical pressure equations. The Dranchuk-Abou-Kassem z-factor equations were used to obtain gas densities as required.

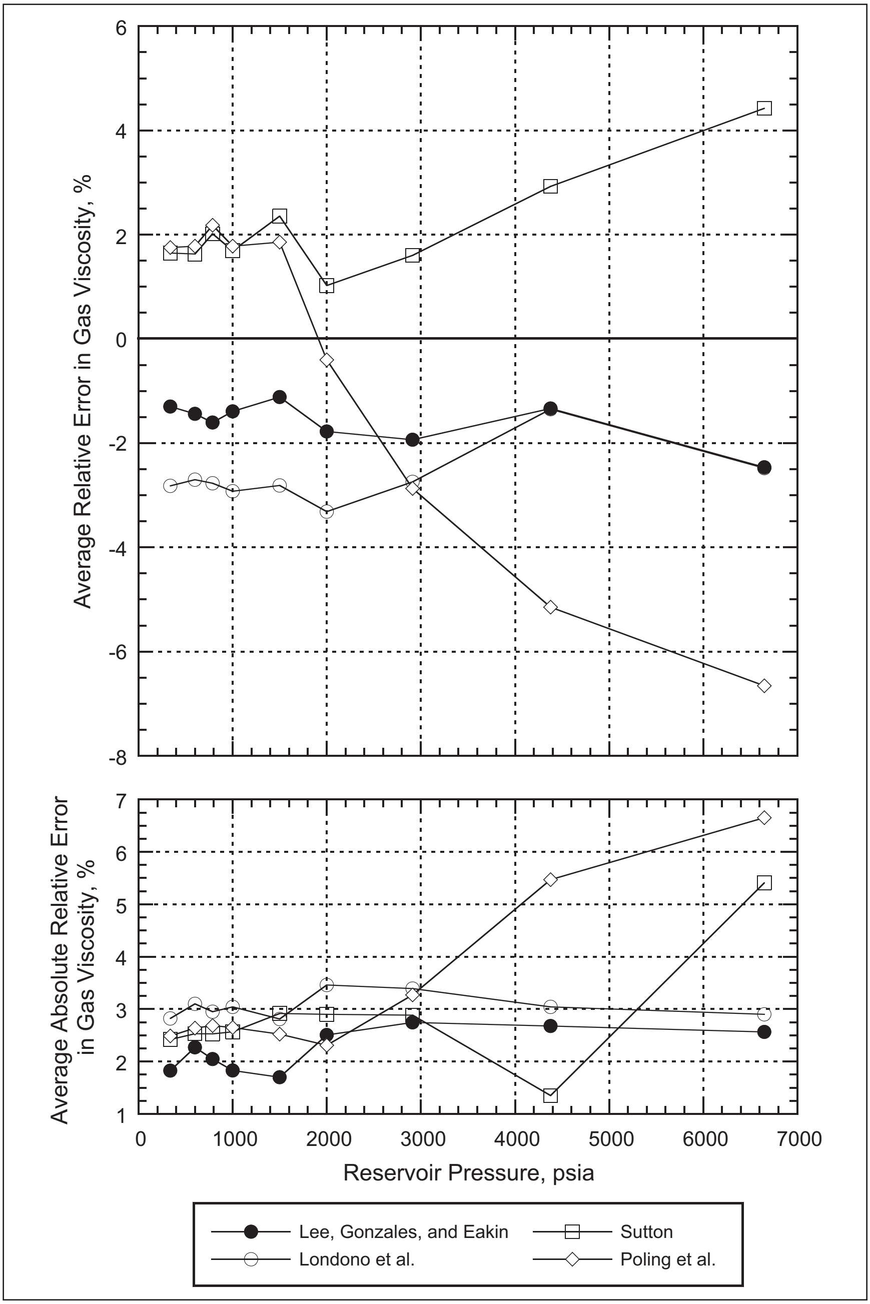

Fig. 2–6. Comparisons of relative errors in gas viscosity correlations using data of table 2–4 sorted and split by pressure. All calculations used the Piper et al. pseudocritical temperature and pseudocriticaal pressure equations. The Dranchuk-Abou-Kassem z-factor equations were used to obtain gas densities as required.

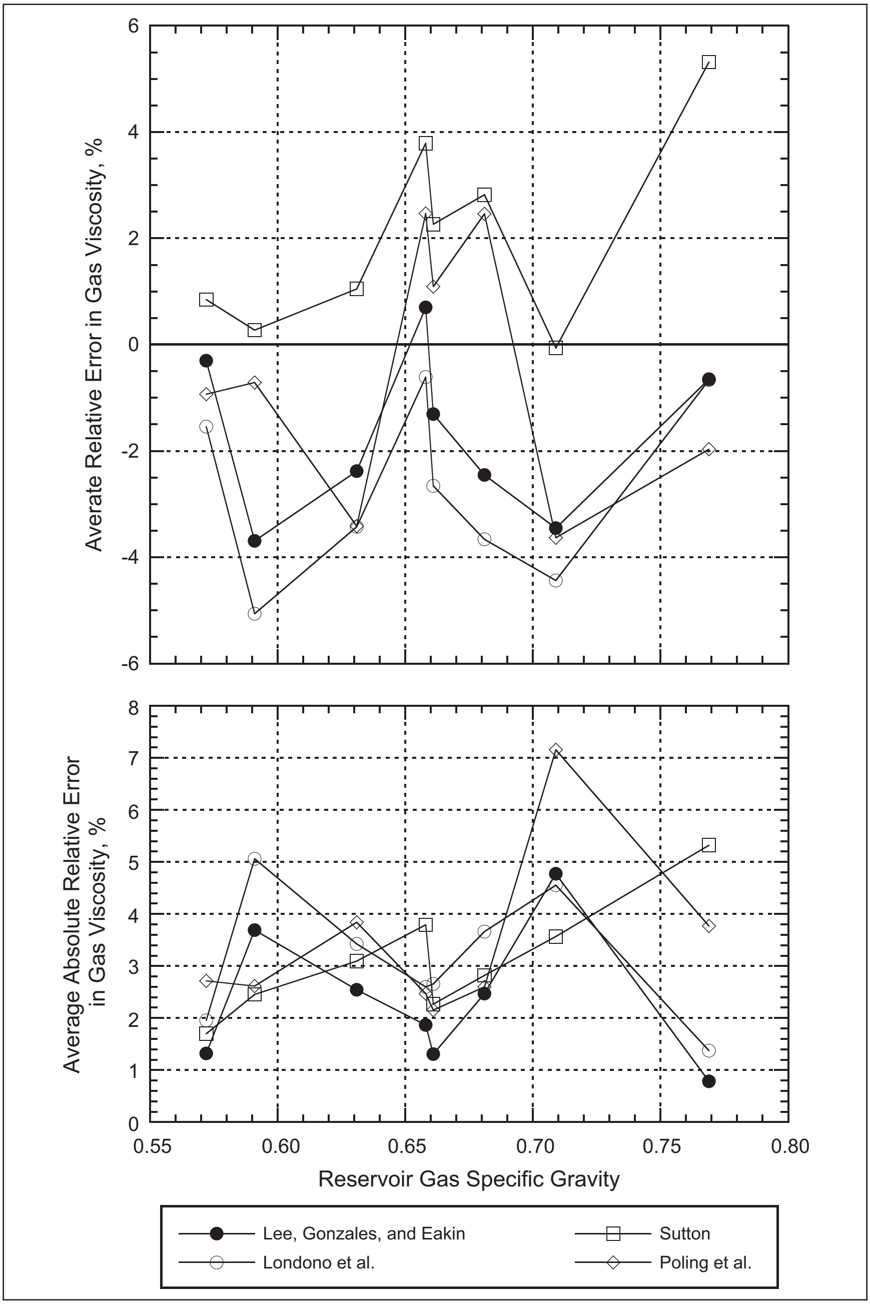

Fig. 2–7. Comparisons of relative errors in gas viscosity correlations using data of table 2–4 sorted and split by gas specific gravity. All calculations used Piper et al. pseudocritical temperature and pseudocritical pressure equations. The Dranchuk-Abou-Kassem z-factor equations were used to obtain gas densities as required.

Recommended correlation equations for estimation of gas viscosities

Thus, the LGE correlations are recommended to estimate gas viscosity.[36]

$$\mu_g = A(10^{-4})\ \text{EXP}\left(B\rho_g{}^{C}\right) \tag{2.14}$$

where

$$A = \frac{(9.379 + 0.01607M)T^{1.5}}{209.2 + 19.26M + T}\quad, \tag{2.15}$$

$$B = 3.448 + \frac{986.4}{T} + 0.01009M\quad \text{, and} \tag{2.16}$$

$$C = 2.447 - 0.2224B. \tag{2.17}$$

The gas density in equation (2.14) is in grams per cubic centimeter (g/cc), which is 62.428 multiplied by the results of equation (2.9), and the resulting gas viscosity is in centipoise.

Coefficients of Isothermal Compressibility of Reservoir Gases

The coefficient of isothermal compressibility of a gas is often called *gas compressibility*. It is the fractional change in volume of the gas as pressure changes at constant temperature. The definition is given in equation (2.18).

$$c_g = -\frac{1}{V}\left(\frac{\partial V}{\partial p}\right)_T \tag{2.18}$$

An equation for calculation of gas compressibility can be obtained by combining equation (2.18) with the real gas equation, equation (2.2), resulting in equation (2.19).

$$c_g = \frac{1}{p} - \frac{1}{z}\left(\frac{\partial z}{\partial p}\right)_T \tag{2.19}$$

This equation can be combined with the definition of pseudoreduced pressure, equation (2.3), to result in equation (2.20)[37]

$$c_g p_{pc} = \frac{1}{p_{pr}} - \frac{1}{z}\left(\frac{\partial z}{\partial p_{pr}}\right)_{T_{pr}} \tag{2.20}$$

where the product on the left side is called *pseudoreduced compressibility*.

$$c_{pr} = c_g p_{pc} \tag{2.21}$$

Mattar et al. combined equation (2.20) with equation (2.8) to arrive at equation (2.22).[38]

$$c_{pr} = \frac{1}{p_{pr}} - \frac{0.27}{z^2 T_{pr}}\left[\frac{\left(\dfrac{\partial z}{\partial \rho_{pr}}\right)_{T\,pr}}{1 + \left(\dfrac{\rho_{pr}}{z}\right)\left(\dfrac{\partial z}{\partial \rho_{pr}}\right)_{T\,pr}}\right] \tag{2.22}$$

Since the DAK equations have been selected as the most nearly accurate source of z-factors, they will be used to evaluate the derivatives in equation (2.22).

$$\begin{aligned}\left(\frac{\partial z}{\partial \rho_{pr}}\right)_{Tpr} &= A_1 + A_2/T_{pr} + A_3/T_{pr}^{\;3} + A_4/T_{pr}^{\;4} + A_5/T_{pr}^{\;5} \\ &+ 2\rho_{pr}\left(A_6 + A_7/T_{pr} + A_8/T_{pr}^{\;2}\right) - 5\rho_{pr}^{\;4} A_9\left(A_7/T_{pr} + A_8/T_{pr}^{\;2}\right) \\ &+ \frac{2A_{10}\rho_{pr}}{T_{pr}^{\;3}}\left(1 + A_{11}\rho_{pr}^{\;2} - A_{11}^{\;2}\rho_{pr}^{\;4}\right)\,\mathrm{EXP}\left(-A_{11}\rho_{pr}^{\;2}\right)\end{aligned} \tag{2.23}$$

The accuracies of the values of gas compressibility, c_g, calculated with equations (2.21)–(2.23) depend on how well the DAK equations represent the slopes of the isotherms of the graphical z-factor correlation of Standing and Katz.

Real Gas Pseudopressures

The real gas pseudopressure was defined by Al-Hussainy et al. to allow solutions to the diffusivity equation originally developed for slightly compressible liquids to be applied to flow in gas reservoirs.[39] This fluid property is used primarily in pressure transient analysis.

$$m(p) = 2\int_{p_m}^{p} \frac{p}{\mu_g z}\,dp \tag{2.24}$$

The lower limit on the integral can be taken to be 0.0 if the Lee, Gonzalez, Eakin equations are used to estimate gas viscosity. The units of pressure are pounds per square inch absolute (psia), and gas viscosity units are centipoises, so the units of real gas pseudopressure are pounds per square inch squared per centipoises (psi^2/cp). This equation is not a correlation; it is simply a part of the larger gas diffusivity equation. The integral can be evaluated numerically using values of gas viscosities and gas z-factors from the correlation equations and procedures discussed previously. The resulting values of *m(p)* are as accurate as the values of gas viscosities and gas z-factors used in the calculations.

Gas Properties at High Pressures and High Temperatures, HPHT

HPHT reservoirs are generally considered to have pressures from 10,000 psia to 30,000 psia and temperatures above 300°F. Currently no published data of gas z-factors or gas viscosities for naturally occurring petroleum gases are available for use in preparing correlations for use at HPHT conditions.

Gas densities (z-factors) at HPHT

The hydrocarbon portion of gases that are found in HPHT reservoirs is essentially pure methane, although some level of nonhydrocarbon components is usually present.

The National Institute of Standards and Technology, NIST, has prepared tables of density and viscosity values for pure methane.[40] Gas densities calculated with the Piper et al. and DAK equations [equations (2.3)–(2.8)] were tested against the NIST methane density values at temperatures of 300°F to 400°F and pressures of 10,000 to 30,000 psia. The ARE was 1.312% and the AARE was 1.312%. No particular trends in ARE or AARE with either temperature or pressure were observed.

The effect of nonhydrocarbon components is unknown. However, equations (2.3)–(2.8) predict reasonably accurate values of z-factors for fairly large values of nonhydrocarbon content, as can be seen in tables 2–2 and 2–3. Thus, these equations can be used with reasonable confidence for HPHT conditions until sufficient data are available to prepare HPHT correlations.

Recently some z-factor data for four gases with carbon dioxide contents of 0, 5, 10, and 20 mol%, pressures up to 20,000 psia, and temperatures of 300°F and 400°F were presented in graphical form.[41] The hydrocarbon portions of these gases were 96 mol% methane, with 3 mol% ethane and 1 mol% propane. The Piper et al. and DAK equations [equations (2.3)–(2.8)] gave z-factors that fit these data to within 1% for pressures from 3,000 to 20,000 psia.

Recommended correlation equations/procedures for estimation of gas z-factors at HPHT

Thus equations (2.3)–(2.8) are recommended for estimating gas z-factors at HPHT until additional HPHT data using naturally occurring petroleum gases become available.

Gas viscosities at HPHT

Viswanathan modified the LGE gas viscosity equations for HPHT conditions by fitting them to the NIST pure methane gas viscosity values.[42] The equations produce values that fit the NIST pure methane gas viscosity values (300°F to 400°F and 10,000 to 30,000 psia) with an ARE of –0.707% and an AARE of 0.707%. The data were sorted and split by temperature and also by pressure, and no particular trends in ARE or AARE with either were noted.

Again, the effect of nonhydrocarbon components is not known. However, the densities required in the Viswanathan viscosity correlation equations are calculated with the Piper et al. and DAK equations, so any nonhydrocarbons present in the gas should not seriously affect the results. These equations can be used with some confidence until sufficient data are available to prepare correlations based on naturally occurring petroleum gases for use at HPHT.

Recommended correlation equations for estimation of gas viscosities at HPHT

The recommended equations for HPHT use are given in equations (2.25)–(2.28).

$$\mu_g = A(10^{-4})EXP\left(B\rho_g^C\right) \tag{2.25}$$

where

$$A = \frac{(5.0512 - 0.2888M)T^{1.832}}{-443.8 + 12.9M + T}, \tag{2.26}$$

$$B = -6.1166 + \frac{3084.9437}{T} + 0.3938M \text{ , and} \tag{2.27}$$

$$C = 0.5893 + 0.1563B. \tag{2.28}$$

Gas density in equation (2.25) should be in units of grams per cubic centimeter, and the resulting gas viscosities are in centipoise.

Nomenclature

Latin

A_1 through A_{11}	Constants in DAK equation, equation (2.7)
A, B, and C	Constants in the LGE and modified LGE equations, equations (2.14) and (2.25)
B_g	Gas formation volume factor, res vol/std vol
c_g	Coefficient of isothermal compressibility (gas compressibility), 1/psi
c_{pr}	Pseudoreduced coefficient of isothermal compressibility, $c_g p_{pc}$, dimensionless
g	Acceleration due to gravity, ft/sec^2
g_c	Gravitational constant, 32.2 lb$_m$ ft/lb$_f$, sec^2

J	Equation (2.5), °R/psia
K	Equation (2.6), °R/psia$^{1/2}$
M	Molecular weight (molar mass), lb/lb mole
$m(p)$	Real gas pseudopressure, psia^2/cp
n	Mass of gas, lb moles
p	Pressure, psia
p_c	Critical pressure, psia
p_{pc}	Pseudocritical pressure, psia
p_{pr}	Pseudoreduced pressure, p/p_{pc}, dimensionless
p_{sc}	Standard (base) pressure, psia
R	Universal gas constant, 10.732 psia cu ft/lb mole °R
T	Temperature, °R
T_c	Critical temperature, °R
T_{pc}	Pseudocritical temperature, °R
T_{pr}	Pseudoreduced temperature, T/T_{pc}, dimensionless
T_{sc}	Standard (base) temperature, °R
V	Volume, cu ft
V_{actual}	Actual (real) volume at some p and T, cu ft
V_{ideal}	Volume calculated with ideal gas equation at some p and T, cu ft
V_R	Volume at reservoir conditions, cu ft
V_{sc}	Volume at standard (base) conditions, cu ft
z	Compressibility factor (z-factor), dimensionless
z_{sc}	Compressibility factor (z-factor) at standard conditions, dimensionless, usually taken to have a value of 1.0

Greek

$\alpha_0, \alpha_1, \alpha_2, \alpha_3, \alpha_4, \alpha_5$	Constants in Piper et al. equation, equation (2.5)
$\beta_0, \beta_1, \beta_2, \beta_3, \beta_4, \beta_5$	Constants in Piper et al. equation, equation (2.6)
γ_g	Gas specific gravity

μ	Dynamic viscosity, centipoise
μ_g	Gas dynamic viscosity, centipoise
ν	Kinematic viscosity, centistoke
ρ_g	Gas density, lb/cu ft; except equations (2.13), (2.14), and (2.25) require g/cc
ρ_{pr}	Pseudoreduced gas density, dimensionless, defined by equation (2.8)

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3

PROPERTIES OF BLACK OILS

The reservoir fluid properties of interest in petroleum engineering include the following:

1. Bubblepoint pressures
2. Solution gas-oil ratios at pressures less than bubblepoint pressure
3. Coefficients of isothermal compressibility
4. Reservoir oil densities
5. Formation volume factors
6. Reservoir oil viscosities

Correlation equations that have been proposed for estimating values of these reservoir fluid properties have been evaluated using large sets of data from service company fluid property laboratory studies on black oil samples with worldwide origins.

Field Data for Use in Black Oil Correlations

The basic premise is that in the application of the correlations, the compositions of the oils are not known, i.e., a laboratory study is not available. Thus, field data are required as input to the correlation equations to estimate values of the fluid properties of interest. The correlations rely on readily available data: (1) stock-tank oil gravity, °API, (2) first-stage separator gas specific gravity, and (3) solution gas-oil ratio at bubblepoint pressure, standard cubic feet per stock-tank barrels (scf/STB), all of which are routinely measured as a part of the sales process; (4) reservoir temperature, and (5) selected values of reservoir pressure.

Values of solution gas-oil ratio at the bubblepoint pressure should be obtained from production test data or early production data measured prior to the reservoir pressure declining below the bubblepoint pressure of the reservoir oil. Field-reported values of gas production are usually sales gas metered from the first-stage separator. Estimates of the volume of gas vented from the stock tank must be added to the volume of separator gas in order to obtain values of solution gas-oil ratio at the bubblepoint pressure, which should represent the total gas-in-solution initially in the reservoir. A method of determining the quantity of stock-tank gas will be discussed.

Sometimes the bubblepoint pressure of the reservoir oil can be estimated from field-measured average reservoir pressures.[1] If the necessary pressure measurements are available, bubblepoint pressures estimated in this way will improve the accuracy of the results of almost all of the fluid property correlations.

Many correlation equations require the specific gravity of the separator gas as an independent variable. Some correlations and equations give better results if the weighted average of the total surface gas is used. A method of estimating the specific gravity of the stock-tank gas for use in calculating the weighted-average specific gravity of the total surface gas will be discussed.

Solution gas-oil ratios at bubblepoint pressures from field data

Solution gas-oil ratio at the bubblepoint pressure must be estimated by adding the volume of gas from the separator and the volume of gas from the stock tank.

$$R_{sb} = R_{SP} + R_{ST} \tag{3.1}$$

The producing gas-oil ratio from the separator, R_{SP}, is obtained from early separator gas and stock-tank oil production data (i.e., while reservoir pressure is still above the bubblepoint pressure of the oil). The producing gas-oil ratio from the stock-tank, R_{ST}, is estimated using a correlation. The best available correlation is given as equation (3.2).[2]

$$\ln R_{ST} = 3.955 + 0.83Z - 0.024Z^2 + 0.075Z^3 \tag{3.2a}$$

where
$$Z = \sum_{n=1}^{3} Z_n \tag{3.2b}$$

and
$$Z_n = C0_n + C1_n VAR_n + C2_n VAR_n^2 \tag{3.2c}$$

n	*VAR*$_n$	*C0*$_n$	*C1*$_n$	*C2*$_n$
1	$\ln p_{SP}$	–8.005	2.7	–0.161
2	$\ln T_{SP}$	1.224	–0.5	0
3	API	–1.587	0.0441	-2.29×10^{-5}

Equation (3.2) requires values of separator pressure and separator temperature. Often this information is not available. If so, an option is to use equation (3.3).[3]

$$R_{sb} = 1.1618 R_{SP} \tag{3.3}$$

Evaluation of proposed correlation equations for solution gas-oil ratios at bubblepoint pressures from field data

The available correlation equations for stock-tank gas-oil ratio have been evaluated using the data set described in table 3–1. Most of the data of table 3–1 were used in the preparation of both equation (3.2) and equation (3.3).

Table 3–1. Separator/stock-tank data set for use in evaluating solution gas-oil ratio and weighted-average surface-gas specific gravity correlation equations has a wide range of values of the independent variables (881 lines of data from 881 reservoir fluid studies with worldwide origins).

Laboratory measurement	Minimum	Median	Mean	Maximum
Separator pressure, p_{SP}, psig	12	80	130	950
Separator temperature, T_{SP}, °F	35	80	92	194
Stock-tank oil gravity, *API*, °API	6.0	38.8	36.2	56.8
Separator gas-oil ratio, R_{SP}, scf/STB	8	520	559	1,817
Stock-tank gas-oil ratio, R_{ST}, scf/STB	2	54	70	527
Surface gas specific gravity, γ_g	0.566	0.881	0.879	1.292
Separator gas specific gravity, γ_{gSP}	0.561	0.815	0.837	1.237
Stock-tank gas specific gravity, γ_{gST}	0.581	1.285	1.256	1.598

Evaluations of the published correlations for this property are given in table 3–2.

Table 3–2. Average errors in estimates of solution gas-oil ratio at bubblepoint pressure from field data for this study are within expected experimental error.

	Predicted solution gas-oil ratios from field data	
Correlation	**ARE, %**	**AARE, %**
Valko-McCain (2003) equations (3.2) and (3.1)	0.0	5.2
Valko-McCain (2003) equation (3.3)	0.0	9.9
Rollins et al. (1990)[a]	9.9	11.8
Hanafy et al. (SPE 37295, 1997)[b]*	17.1	19.8

[a] Rollins, J.B., McCain, W.D., Jr., and Creeger, J.T. "Estimation of Solution GOR of Black Oils." *Journal of Petroleum Technology*. Vol. 42 (January 1990) 92–94. *Transactions*. AIME. Vol. 289. [b] Hanafy, H.H., Macary, S.M., El Nady, Y.M., Bayomi, A.A., and El Batanony, M.H. "Empirical PVT Correlations Applied to Egyptian Crude Oils Exemplify Significance of Using Regional Correlations." Paper SPE 37295. Presented at the SPE International Symposium on Oilfield Chemistry, Houston (1997). * Author restricted the correlation to a specific geographical area.

The data of table 3–1 were sorted according to separator pressures, separator temperatures, and separator producing gas-oil ratios to determine if the results of equations (3.1) and (3.2) are valid across the data set. The results, shown in table 3–3, seem to be consistent throughout the ranges of separator conditions and producing gas-oil ratios normally expected in the oil field.

Table 3–3. The solution gas-oil ratios at bubblepoint pressures from field data predicted with equations (3.1) and (3.2) are consistent across the data set.

Data set	Number of lines of data	ARE, %	AARE, %
All data	881	0.0	5.2
$p_{SP} \le 50$ psig	359	–0.4	4.3
50 psig $< p_{SP} \le 100$ psig	291	–1.9	5.0
$p_{SP} > 100$ psig	231	3.0	6.9
$T_{SP} \le 80$°F	472	1.0	5.3
$T_{SP} > 80$°F	409	–1.2	5.1
$R_{SP} \le 300$ scf/STB	294	1.3	9.6
300 scf/STB $< R_{SP} \le 700$ scf/STB	287	–0.8	3.8
$R_{SP} > 700$ scf/STB	300	–0.5	2.1

Weighted-average specific gravities of surface gases

Weighted-average surface-gas specific gravities may be calculated with equation (3.4).

$$\gamma_g = \frac{\gamma_{gSP} R_{SP} + \gamma_{gST} R_{ST}}{R_{SP} + R_{ST}} \tag{3.4}$$

A method of estimating the stock-tank producing gas-oil ratio, R_{ST}, was given as equation (3.2). Unfortunately, the specific gravity of stock-tank gas is rarely measured in the field. Apparently only one publication has addressed the problem of estimating the specific gravities, γ_{gST}, of stock-tank gases.[4]

$$\gamma_{gST} = 1.219 + 0.198Z + 0.0845Z^2 + 0.03Z^3 + 0.003Z^4 \tag{3.5a}$$

where
$$Z = \sum_{n=1}^{5} Z_n \tag{3.5b}$$

and
$$Z_n = C0_n + C1_n VAR_n + C2_n VAR_n^2 + C3_n VAR_n^3 + C4_n VAR_n^4 \tag{3.5c}$$

n	VAR_n	$C0_n$	$C1_n$	$C2_n$	$C3_n$	$C4_n$
1	$\ln p_{SP}$	–17.275	7.9597	–1.1013	2.7735×10^{-2}	3.2287×10^{-3}
2	$\ln R_{SP}$	–0.3354	–0.3346	0.1956	-3.4374×10^{-2}	2.08×10^{-3}
3	API	3.705	–0.4273	1.818×10^{-2}	-3.459×10^{-4}	2.505×10^{-6}
4	γ_{gSP}	–155.52	629.61	–957.38	647.57	–163.26
5	T_{SP}	2.085	-7.097×10^{-2}	9.859×10^{-4}	-6.312×10^{-6}	1.4×10^{-8}

Use of equation (3.5) requires knowledge of separator pressures and temperatures. If this information is not available, equation (3.6) can be used.[5]

$$\gamma_g = 1.066\gamma_{gSP} \tag{3.6}$$

Evaluation of proposed correlation equations for weighted-average surface-gas specific gravities

These two methods were evaluated with the data set described in table 3–1, and the results are given in table 3–4. Only 618 of the lines of data could be used because several of the laboratory reports did not report the stock-tank gas specific gravity.

Table 3–4. Average errors in estimates of weighted-average surface-gas specific gravities are well within expected experimental error.

	Predicted weighted-average surface-gas specific gravities	
Correlation	**ARE, %**	**AARE, %**
Valko-McCain (2003) equations (3.5), (3.2), and (3.4)	–0.7	2.2
Valko-McCain (2003) equation (3.6)	0.0	3.8

Most of the data described in table 3–1 were used to prepare correlation equations (3.5) and (3.6).[6]

The data were separated into groups based on separator pressures, separator temperatures, and separator producing gas-oil ratios. Table 3–5 shows that equations (3.1), (3.2), and (3.5) give good results across the ranges of separator conditions and producing gas-oil ratios of the full set of data.

Table 3–5. The weighted-average surface-gas specific gravities estimated with equations (3.5), (3.2), and (3.4) are consistent across the data set.

		Predicted weighted-average surface-gas specific gravities	
Data set	**Number of lines of data**	**ARE, %**	**AARE, %**
All data	618	–0.7	2.2
$p_{SP} \leq 50$ psig	274	–0.6	1.9
50 psig < $p_{SP} \leq 100$ psig	245	–1.2	2.4
$p_{SP} > 100$ psig	99	0.4	2.4
$T_{SP} \leq 80$°F	389	–0.2	2.0
$T_{SP} > 80$°F	229	–1.4	2.5
$R_{SP} \leq 300$ scf/STB	170	–1.5	3.7
300 scf/STB < $R_{SP} \leq 700$ scf/STB	205	–0.5	2.1
$R_{SP} > 700$ scf/STB	243	–0.3	1.3

Bubblepoint Pressures at Reservoir Temperatures

A value of bubblepoint pressure at reservoir temperature is required for nearly all of the oil property correlation equations. Sometimes a value of bubblepoint pressure can be deduced for an oil if pressure-production data are available for that particular reservoir.[7] However, the usual situation is that a value of bubblepoint pressure must be estimated with correlation equations.

Evaluation of proposed correlation equations for bubblepoint pressures at reservoir temperatures

Table 3–6 gives statistics of the data set used to evaluate published correlation equations for bubblepoint pressures at reservoir temperatures. The data include the full range of values of the independent variables normally associated with black oil reservoirs.

Comparisons of the measured bubblepoint pressures from the data described in table 3–6 with the results of calculations with 24 sets of correlation equations from the literature are shown in table 3–7.

Table 3–6. The data set used to evaluate bubblepoint pressure correlations has a wide range of values of independent variables (1,794 lines of data from 1,794 reservoir fluid studies with worldwide origins).

Laboratory measurement	Minimum	Median	Mean	Maximum
Solution gas-oil ratio at p_b, R_{sb}, scf/STB	10	530	596	2,438
Bubblepoint pressure, p_b, psia	82	2,134	2,237	7,750
Reservoir temperature, T_R, °F	60	180	185	342
Stock-tank oil gravity, *API*, °API	6.0	37.0	35.6	63.7
Separator gas specific gravity, γ_{gSP}	0.555	0.801	0.830	1.685

Table 3–7. A comparison of published bubblepoint pressure correlation equations with data set of table 3–6 reveals the more reliable correlations.

	Predicted bubblepoint pressures	
Correlation	**ARE, %**	**AARE, %**
Valko-McCain (2003)	0.0	10.9
Velarde et al. (1999)	1.3	12.5
McCain et al. (1998)	3.7	12.5
Labedi (1990)	–0.1	12.6
Standing (1977)*	–2.2	12.6
Lasater (1958)*	–1.5	13.5
Levitan-Murtha (1999)[a]	4.2	13.8
De Ghetto et al. equation (A.4) (1994)[b]	3.3	13.9
Al-Shammasi (1999)[c]	–1.6	14.3
Vazquez-Beggs (1980)	7.1	14.5
Omar-Todd (1993)[d]*	5.1	15.4
Kartoatmodjo-Schmidt (1994)	4.4	15.7
Dindoruk-Christman (2004)[e]*	0.9	16.1
Glaso (1980)	4.7	16.7
Farshad et al. (1996)[f]*	–5.6	17.7
Al-Marhoun (1988)[g]*	9.1	18.0
Malallah et al. (2005)[h]	–17.4	20.4
Dokla-Osman (1992)[i]*	0.1	21.8
Almehaideb (1997)[j]*	–0.8	22.3
Macary-El Batanony (1993)[k]*	11.9	23.1
Khairy et al. (1998)[l]*	5.0	23.2
Hanafy et al. (SPE 37439, 1997)*	10.2	28.7
Petrosky-Farshad (1998)[m]*	17.7	37.6
Yi (2000)[n]	42.3	45.0

[a] Levitan, L.L. and Murtha, M. "New Correlations Estimate P_b, FVF." *Oil & Gas Journal*. Vol. 97 (March 8, 1999) 70–76. [b] De Ghetto, G., Paone, F., and Villa, M. "Reliability Analysis on PVT Correlations." Paper SPE 28904. Presented at the European Petroleum Conference, London (1994). [c] Al-Shammasi, A.A. "Bubble Point Pressure and Oil Formation Volume Factor Correlations." Paper SPE 53185. Presented at the SPE Middle East Oil Show, Bahrain (1999). [d] Omar, M.I. and Todd, A.C. "Development of New Modified Black Oil Correlations for Malaysian Crudes." Paper SPE 25338. Presented at the SPE Asia Pacific Oil and Gas Conference, Singapore (1993). [e] Dindoruk, B. and Christman, P.G. "PVT Properties and Viscosity Correlations for Gulf of Mexico Oils." *Society of Petroleum Engineers Reservoir Evaluation and Engineering*. Vol. 7, No. 6 (2004) 427–437. [f] Farshad, F.F., LeBlanc, J.L., Gruber, J.D., and Osorio, J.G. "Empirical PVT Correlations for Columbian Crude Oils." Paper SPE 36105. Presented at the Fourth Latin American and Caribbean Petroleum Engineering Conference, Port-of-Spain (1996). [g] Al-Marhoun, M.A. "PVT Correlations for Middle East Crude Oils." *Journal of Petroleum Technology*. Vol. 40 (May 1988) 650–666. [h] Malallah, A., Gharbi, R., and Algharaib, M. "Accurate Estimation of the World Crude Oil PVT Properties Using Graphical Alternating Conditional Expectation." *Energy & Fuels*. 2006. [i] Dokla, M.E. and Osman, M.E. "Correlation of PVT Properties for UAE Crudes." *Society of Petroleum Engineers Formation Evaluation*. Vol. 7 (March 1992) 41–46. [j] Almehaideb, R.A. "Improved PVT Correlations for UAE Crude Oils." Paper SPE 37691. Presented at the SPE Middle East Oil Conference and Exhibition, Bahrain (1997). [k] Macary, S.M. and El Batanony, M.H. "Derivation of PVT Correlations for Gulf of Suez Crude Oils." *Sekiyu Gakkaishi*. Japan. Vol. 36, No. 6 (1993). [l] Khairy, M., El-Tayeb, S., and Hamdallah, M. "PVT Correlations Developed for Egyptian Crudes." *Oil & Gas Journal*. Vol. 98, No.18 (May 4, 1998) 114–116. [m] Petrosky, G.E. and Farshad, F.F. "Pressure-Volume-Temperature Correlations for Gulf of Mexico Crude Oils." *Society of Petroleum Engineers Reservoir Evaluation and Engineering*. Vol. 1 (October 1998) 416–420. [n] Yi, X. "Using Wellhead Sampling Data to Predict Reservoir Saturation Pressure." Paper SPE 59700. Presented at the SPE Permian Basin Oil and Gas Recovery Conference, Midland (2000). *Author restricted the correlation to a specific geographical area.

Correlation equations with low values of ARE indicate that the calculations describe a line that bisects the data. The relatively high values of AARE for all correlations in table 3–7 are a reflection of the precision of the laboratory-reported values of bubblepoint pressures. Involved in this precision is the method of controlling and measuring temperature during the laboratory procedures. A good correlation can give results centered on the data, but a correlation cannot be more precise than the underlying data.

Valko and McCain used essentially all of the data described in table 3–6.[8] Velarde et al. and McCain et al. each used approximately 40% of these data, and Vazquez and Beggs used approximately one-third of these data.[9] Hanafy et al. used less than 10% of these data to prepare their correlation equations.[10]

The Valko-McCain correlation equations appeared to give the best results in the comparison of table 3–7; however, an examination of how well these equations stand up across the data set is important. The 1,794 lines of data were sorted and sliced into 16 approximately equal subsets, and the ARE and AARE for each of these subsets were calculated for several correlations. Calculated values of ARE and AARE may be found in figures 3–1, 3–2, and 3–3 for sorting on stock-tank oil gravity, solution gas-oil ratio at bubblepoint pressure, and reservoir temperature.

The resulting graphs would be impossibly complicated if all correlations were included. The Standing, Glaso, and Lasater equations were selected because they seem to be the most popular in the industry today.[11] The Labedi equations were selected because of the very low ARE achieved on the entire data set.[12] Note that the results with the Valko-McCain equations are consistently close to zero throughout the full range of the data. The calculations with the Standing equations are remarkably well behaved. The low ARE for the full data set for the Labedi correlation equation is a result of the averaging of some large negative values of ARE at one end of the spectrum of data with large positive ARE values at the other end.

These graphical comparisons are not meant to indicate that one set of correlation equations should be used for one part of the spectrum of the independent variables and another set used for another part of the spectrum. Rather, a single set of correlation equations should be selected for use across the total distribution of possible independent variables.

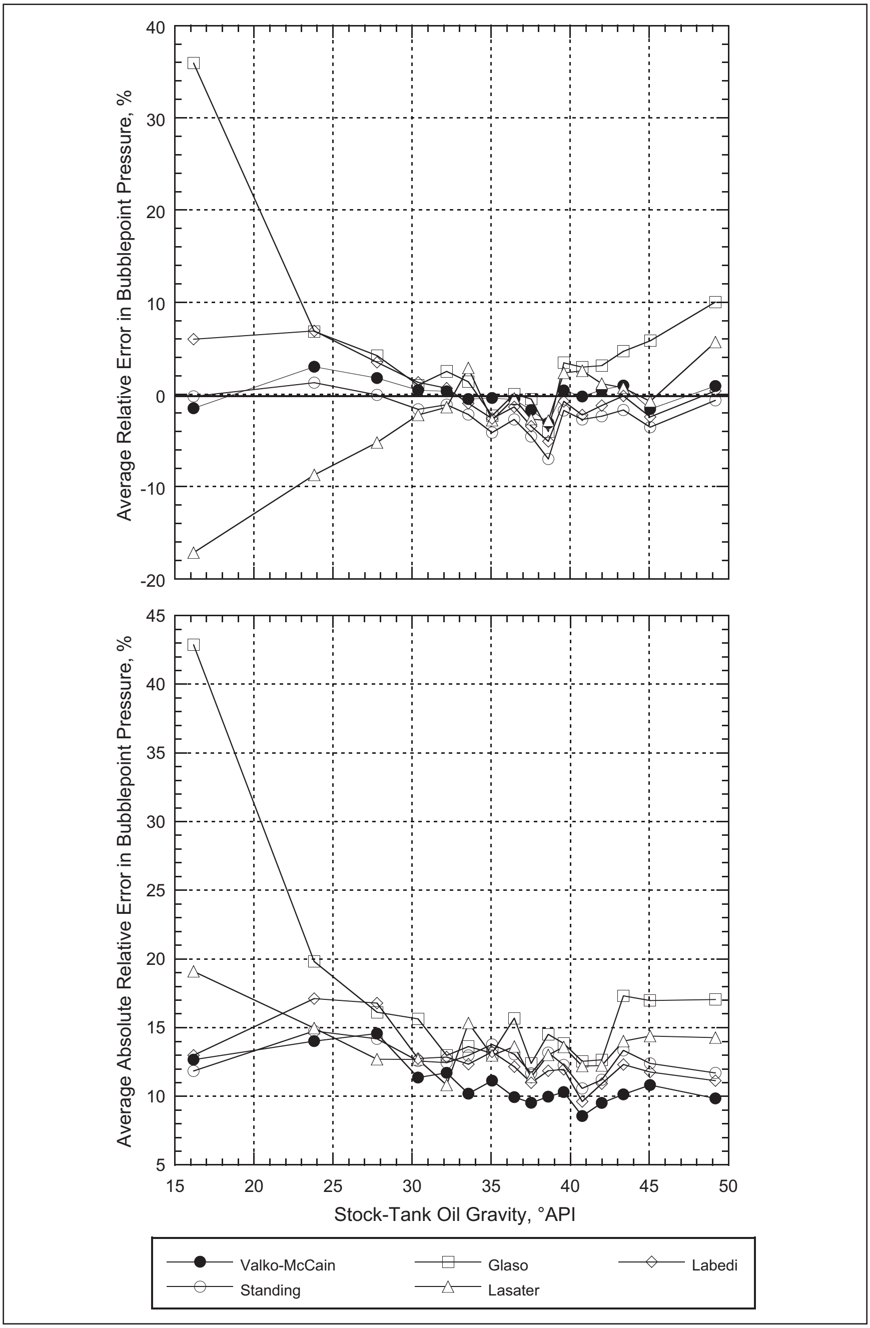

Fig. 3–1. Average relative errors and average absolute relative errors of subsets of the data described in table 3–6 sorted on stock-tank oil gravity show the quality of selected correlation equations for estimating bubblepoint pressures across the full range of this indpendent variable.

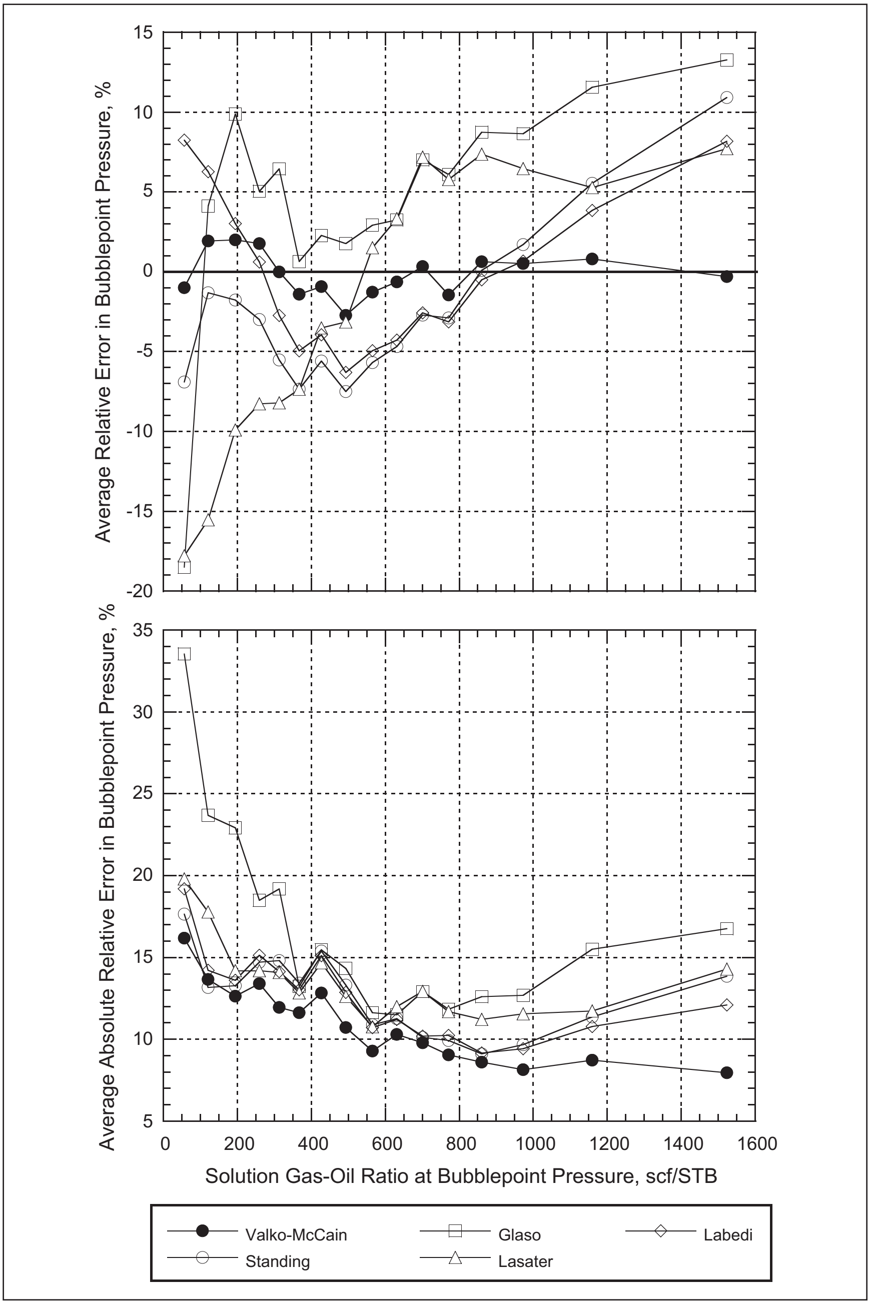

Fig. 3–2. Average relative errors and average absolute relative errors of subsets of the data described in table 3–6 sorted on solution gas-oil ratio at bubblepoint pressure show the quality of selected correlation equations for estimating bubblepoint pressures across the full range of this independent variable.

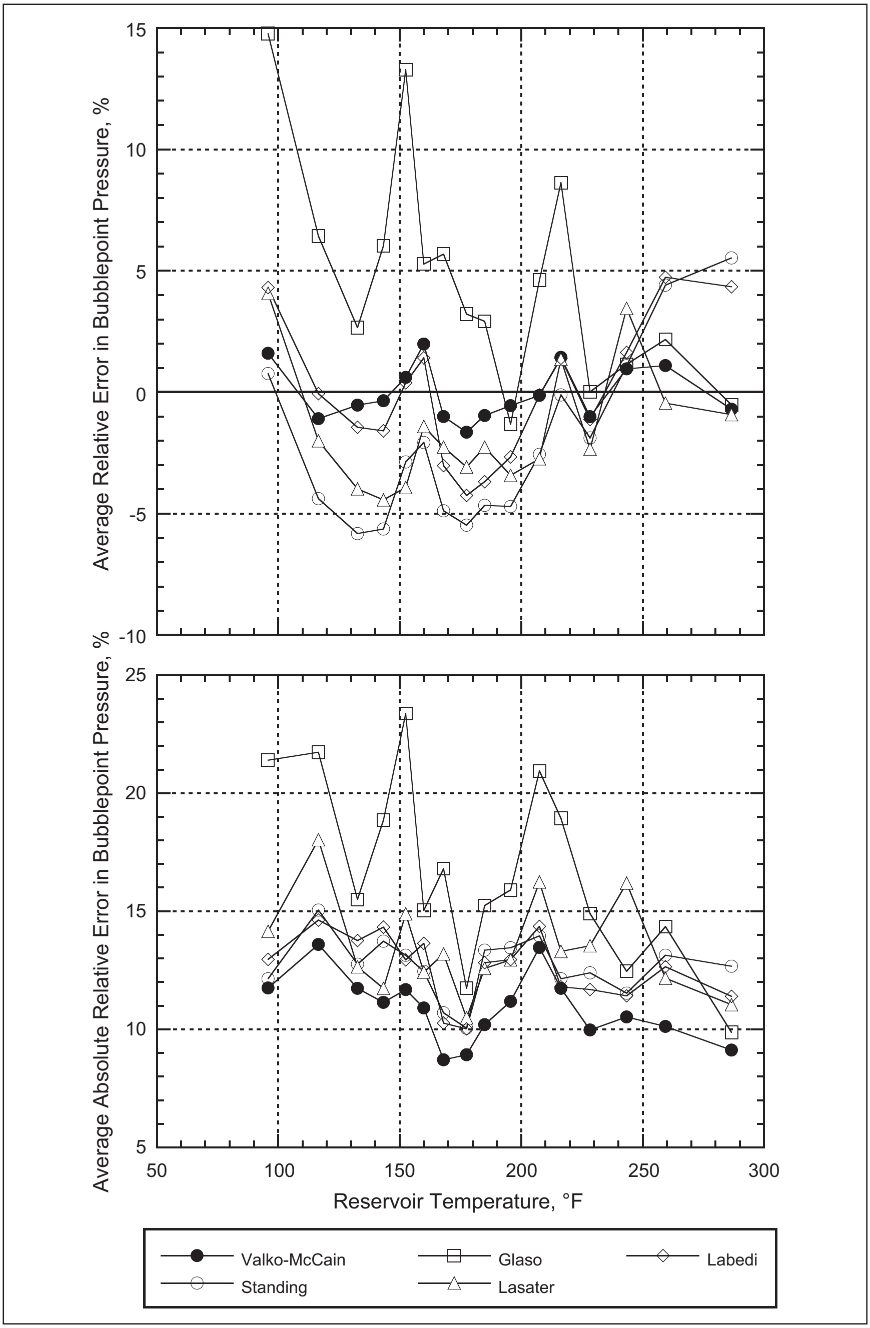

Fig. 3–3. Average relative errors and average absolute relative errors of subsets of the data described in table 3–6 sorted on reservoir temperature show the quality of selected correlation equations for estimating bubblepoint pressures across the full range of this independent variable.

Recommended correlation equations for estimation of bubblepoint pressure at reservoir temperature

The recommended equation for estimating bubblepoint pressures is given as equation (3.7).[13]

$$\ln p_b = 7.475 + 0.713Z + 0.0075Z^2 \tag{3.7a}$$

where

$$Z = \sum_{n=1}^{4} Z_n \tag{3.7b}$$

and

$$Z_n = C0_n + C1_n VAR_n + C2_n VAR_n^2 + C3_n VAR_n^3 \tag{3.7c}$$

n	VAR_n	$C0_n$	$C1_n$	$C2_n$	$C3_n$
1	$\ln R_{sb}$	–5.48	–0.0378	0.281	–0.0206
2	API	1.27	–0.0449	4.36×10^{-4}	-4.76×10^{-6}
3	γ_{gSP}	4.51	–10.84	8.39	–2.34
4	T_R	–0.7835	6.23×10^{-3}	-1.22×10^{-5}	1.03×10^{-8}

Necessity of having correlations specific to a particular geographical area

There have been many discussions about the necessity of having fluid property correlations specific to particular geographical regions. This question has not been fully answered. However, some of the data described in table 3–6 could be identified by sampling location. Table 3–8 shows a comparison of calculations using equation (3.7) with subsets of the data that could be referenced to particular geographical areas.

The average absolute relative errors are approximately the same for each of the geographical subsets of data as well as for the entire data set. This indicates that, at least for bubblepoint pressures, regional correlations are not necessary.

Table 3–8. The recommended bubblepoint pressure correlation equation, equation (3.7), can be used in various geographical areas with adequate accuracy.

	Predicted bubblepoint pressure	
Geographical area	**Number of lines of data**	**AARE, %**
Worldwide	1,794	10.9
Middle-East, Al-Marhoun (1988)[a]	157	9.8
North Sea, Glaso (1980)	17	11.1
Egypt, Hanafy (1999)[b]	125	12.2
Malaysia, Omar and Todd (1993)[c]	93	11.8
USA, Katz (1942)[d]	53	12.2

[a] Al-Marhoun, M.A. "PVT Correlations for Middle East Crude Oils." *Journal of Petroleum Technology*. Vol. 40 (May 1988) 650–666.

[b] Hanafy, H.H.A. "Developing New PVT Correlations for the Egyptian Crude Oils." Thesis, Master of Science, Al-Azhar University, Cairo (1999).

[c] Omar, M.I. and Todd, A.C. "Development of New Modified Black Oil Correlations for Malaysian Crudes." Paper SPE 25338. Presented at the SPE Asia Pacific Oil and Gas Conference, Singapore (1993). [d] Katz, D.L. "Prediction of the Shrinkage of Crude Oils." API Drilling & Production Practices. (1942) 137–147.

Solution Gas-Oil Ratios of Reservoir Oils

A value of solution gas-oil ratio at bubblepoint pressure is required for several of the oil correlations. This variable will be the starting point for the correlations mentioned in this section. A discussion of techniques for determining a value may be found near the start of this chapter.

Evaluation of proposed correlation equations for solution gas-oil ratios at reservoir pressures less than bubblepoint pressure

A description of the set of data used to evaluate the published correlation equations for solution gas-oil ratios may be found in table 3–9. There were 718 laboratory reports, each with several lines of differential liberation data at pressures equal to and below the bubblepoint pressure. Solution gas-oil ratios were calculated with a combination of these differential liberation data and separator test data using the procedure of McCain.[14]

Table 3–9. Statistical summary of data set used to evaluate solution gas-oil ratio and oil formation volume factor correlations at pressures equal to and below bubblepoint pressure (5,359 lines of data from 718 reservoir fluid studies with worldwide origins). Data at pressures below 400 psia were eliminated due to very large errors for some of the correlations at low pressures.

Laboratory measurement	**Minimum**	**Median**	**Mean**	**Maximum**
Bubblepoint pressure, p_b, psia	415	2,151	2,278	7,750
Reservoir temperature, T_R, °F	70	183	187	320
Stock-tank oil gravity, API, °API	11.6	38.4	37.2	56.2
Separator gas specific gravity, γ_{gSP}	0.5548	0.7835	0.7931	1.4720
Solution gas-oil ratio at p_b, R_{sb}, scf/STB	43.0	542	598	1,808
Oil formation volume factor at p_b, B_{ob}, res bbl/STB	1.012	1.333	1.358	2.042

The data described in table 3–9 cover the full range of conditions and properties expected in black oil reservoirs. The samples for the laboratory work that provided the data for table 3–9 had worldwide origins. The results of calculations with various correlation equations from the literature compared with these values of solution gas-oil ratios are given in table 3–10.

Table 3–10. Evaluation of published solution gas-oil ratio correlation equations for reservoir pressures equal to and below bubblepoint pressure reveals the more reliable correlations.

	Predicted solution gas-oil ratios	
Correlation	**ARE, %**	**AARE, %**
Velarde et al. (1999), p_b from data	1.94	4.50
Velarde et al. (1999); p_b from equation (3.7)	2.15	8.81
Casey-Cronquist (1992),* p_b from data	–6.54	9.51
Casey-Cronquist (1992),* p_b from equation (3.7)	–5.41	12.62
Kartoatmodjo-Schmidt (1994)	–10.81	19.06
Levitan-Murtha (1999)[a]	–16.22	19.44
Glaso (1980)[†]	–11.99	19.47
Standing (1977)*	–8.64	19.62
Vazquez-Beggs (1980)	–17.78	22.31
Hanafy et al. (SPE 37439, 1997)*	–16.51	29.20
Elsharkawy-Alikhan (1997)[b]	–12.67	30.62
Carlson (1997); p_b from equation (3.7)	–19.17	32.62

[a] Levitan. L.L. and Murtha, M. “New Correlations Estimate P_b, FVF.” Oil & Gas Journal. Vol. 97 (March 8, 1999) 70–76.

[b] Elsharkawy, A.M. and Alikhan, A.A. “Correlations for Predicting Solution Gas/Oil Ratio, Oil Formation Volume Factor, and Undersaturated Oil Compressibility.” Journal of Petroleum Science and Engineering. Vol. 17 (1997) 291–302.

* Author restricted correlation to a specific geographical area.

[†] Rearrangement of bubblepoint pressure equation—not suggested by author.

Only three of the sets of correlation equations (Velarde et al., Casey-Cronquist, and Carlson) permit entering a bubblepoint pressure from an outside source.[15] All the other correlations require that the bubblepoint pressure predicted by that particular set of equations be used. So the Velarde et al. and Casey-Cronquist correlations were evaluated twice. The correlations were evaluated first with values of bubblepoint pressures from the laboratory reports and then with values of bubblepoint pressures from the best available bubblepoint pressure correlation.[16] It is obvious that an accurate value of bubblepoint pressure, if available, will improve the quality of the results.

Velarde et al. used approximately one-third of these data in preparing their correlation equations; Vazquez-Beggs apparently used most of these data as well as additional data in their research.[17] The other authors apparently did not use any of these data in preparing their correlations.

In order to determine if the recommended equations are valid throughout the full range of possible reservoir conditions and oil properties, the data described in table 3–9 were sorted into 10 groups of approximately 500 lines of data each. The groups were sorted by reservoir temperatures, reservoir pressures, and stock-tank oil gravities. Comparisons of these subsets of the data with five different correlations are given in figures 3–4, 3–5, 3–6, and 3–7. The Velarde et al. correlations are clearly the best and are effective throughout the full range of the data. The Standing and Vazquez-Beggs equations were selected for these comparisons as they seem to be the most popular in the petroleum industry.[18] The use of a rearrangement of the Glaso bubblepoint pressure equation for calculation of solution gas-oil ratios at pressures below bubblepoint pressure is a misapplication of Glaso's work but is included here because this seems to be a common practice in the industry.[19]

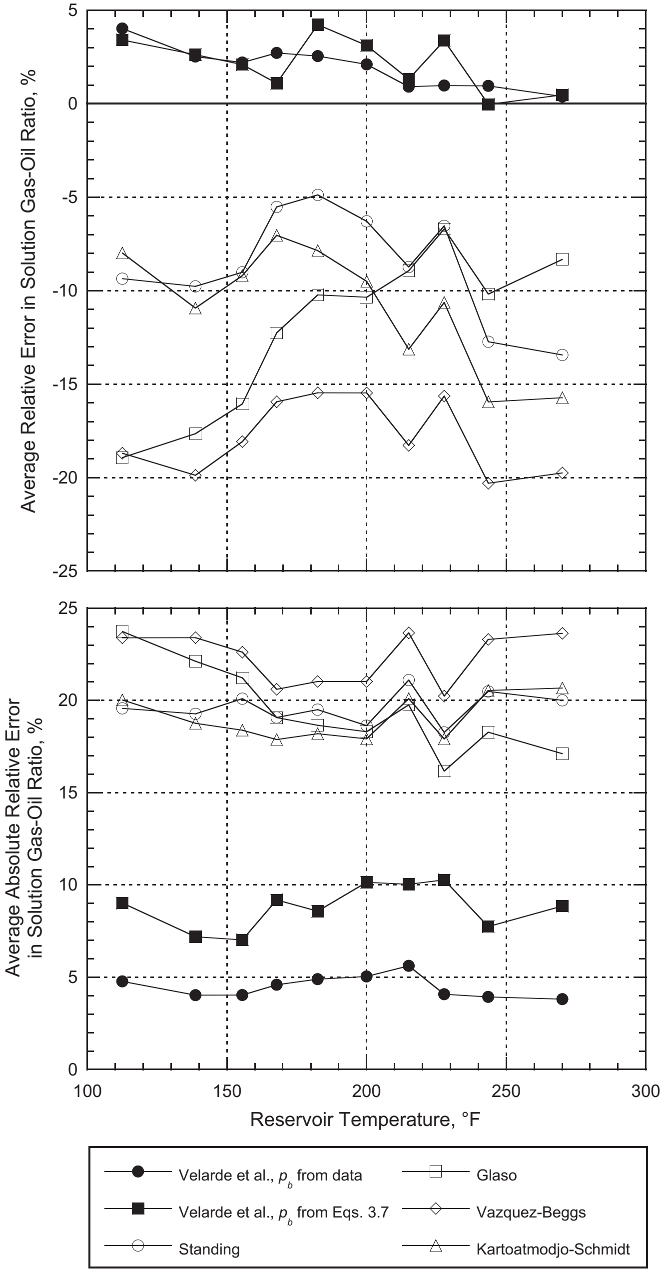

Fig. 3–4. Average relative errors and average absolute relative errors of subsets of the data described in table 3–9 sorted on reservoir temperature show the quality of selected correlation equations for estimating solution gas-oil ratios at pressures less than bubblepoint pressure across the full range of this independent variable.

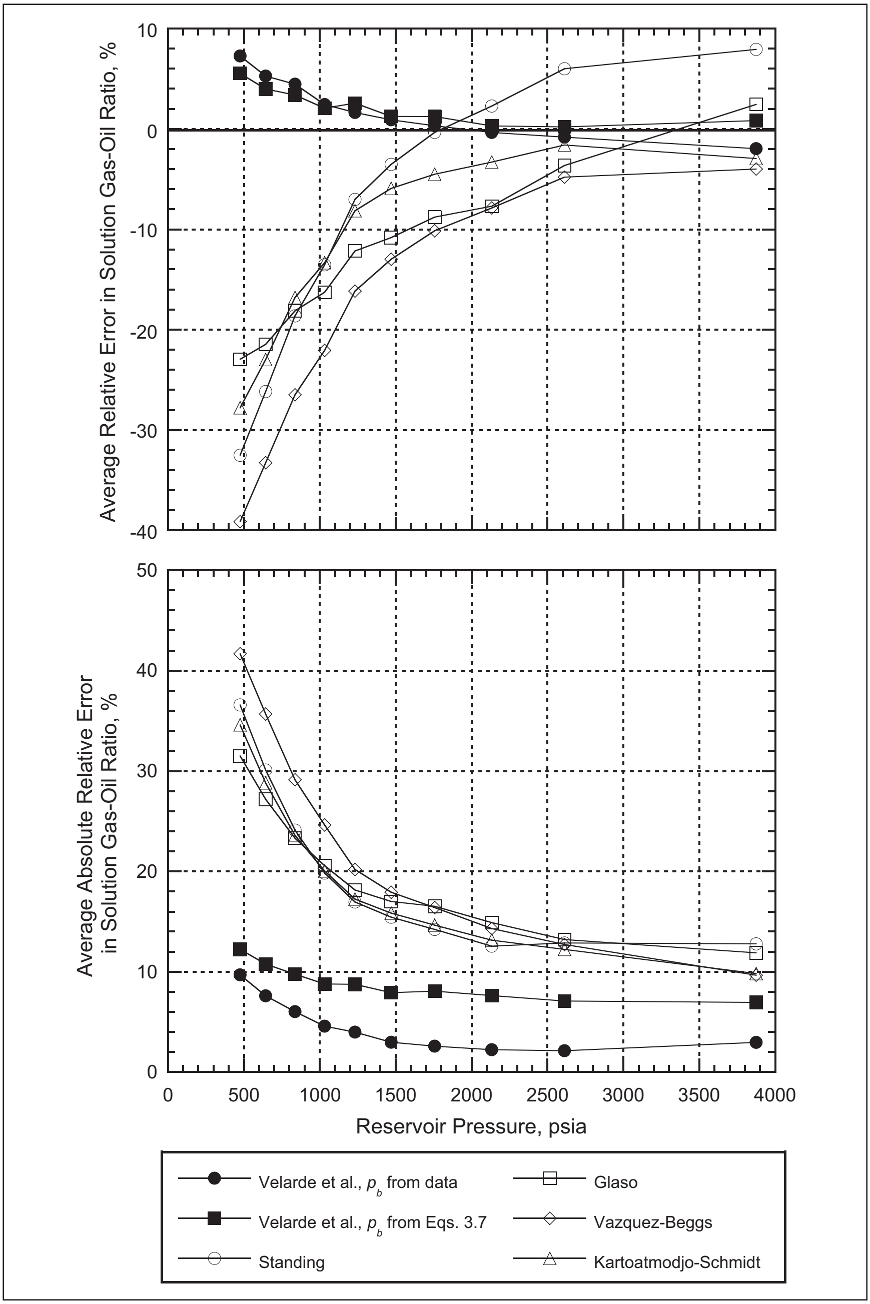

Fig. 3–5. Average relative errors and average absolute relative errors of subsets of the data described in table 3–9 sorted on reservoir pressure show the quality of selected correlation equations for estimating solution gas-oil ratios at pressures less than bubblepoint pressures across the full range of this independent variable.

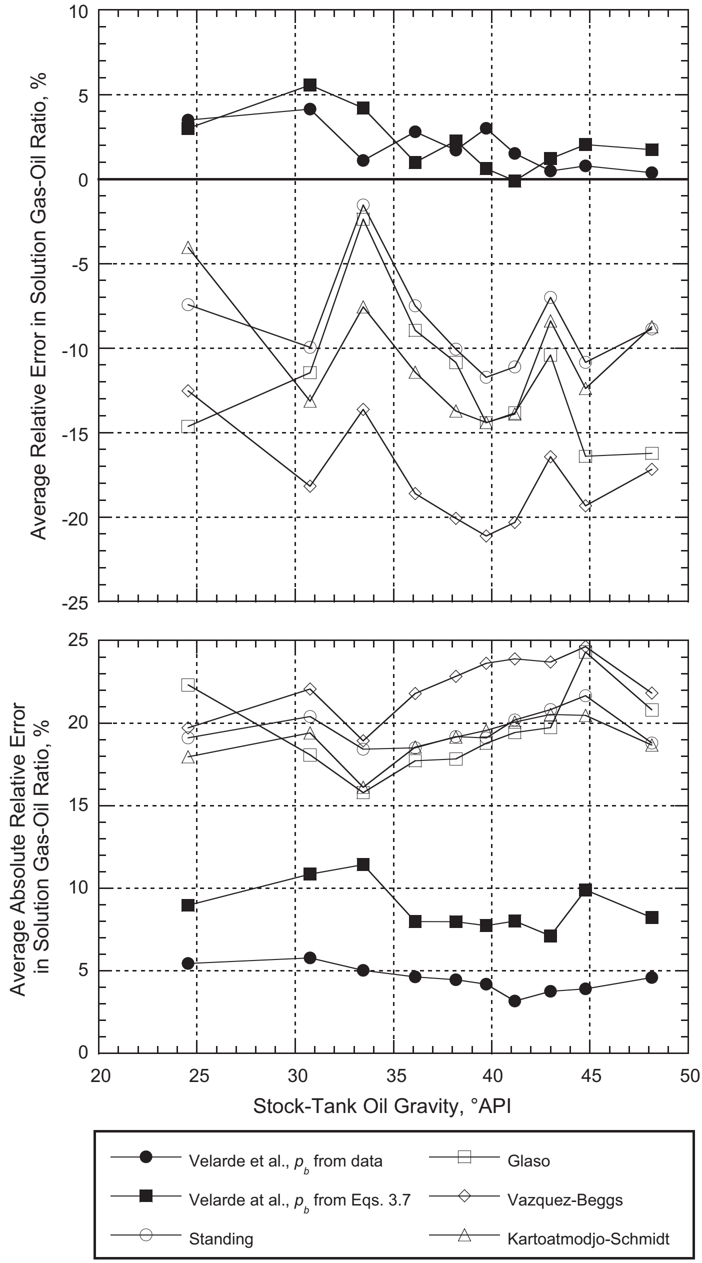

Fig. 3–6. Average relative errors and average absolute relative errors of subsets of the data described in table 3–9 sorted on stock-tank oil gravity show the quality of selected correlation equations for estimating solution gas-oil ratios at pressures less than bubblepoint pressure across the full range of this independent variable.

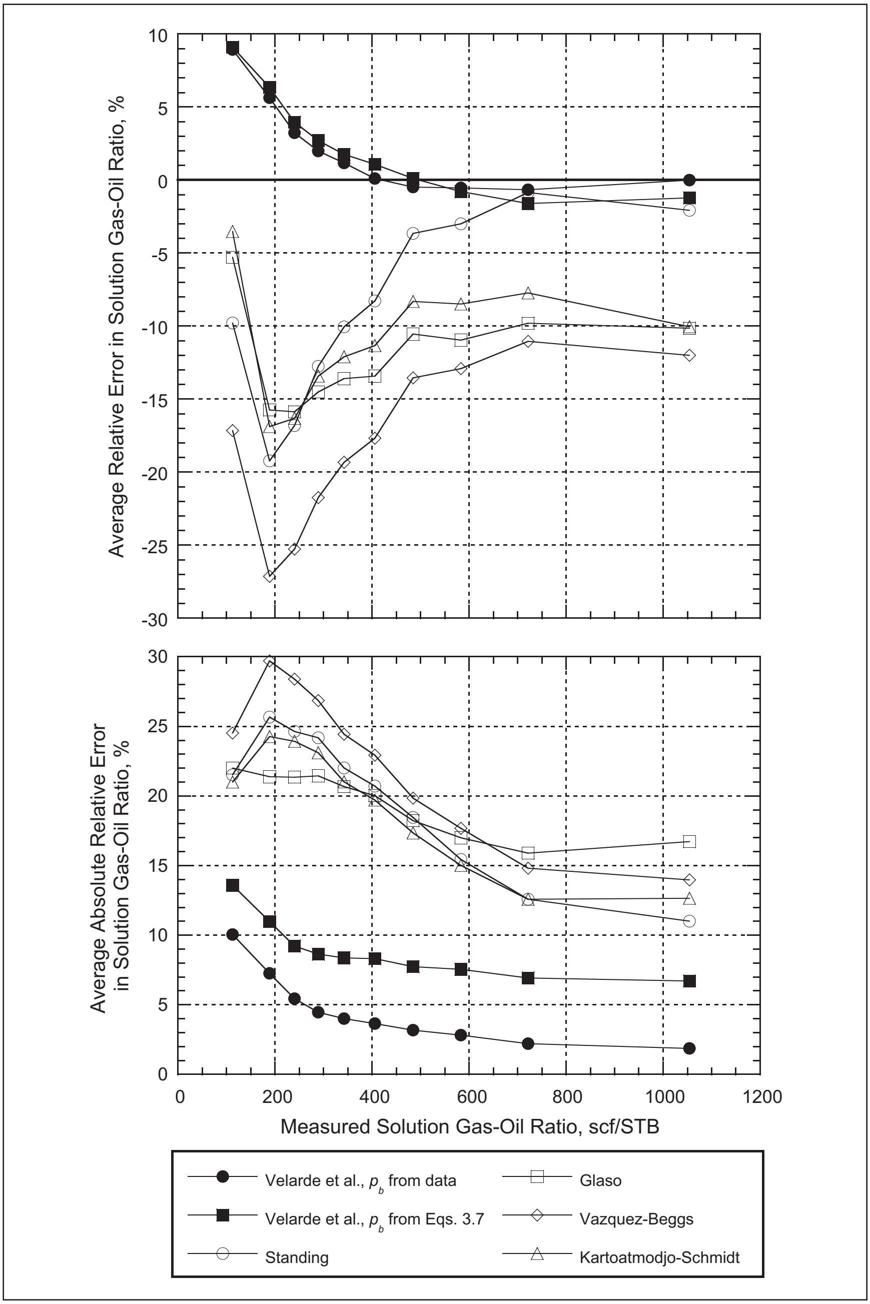

Fig. 3–7. Average relative errors and average absolute relative errors of subsets of the data described in table 3–9 sorted on measured solution gas-oil ratio show the quality of selected correlation equations for estimating solution gas-oil ratios at pressures less than bubblepoint pressure across the full range of this dependent variable.

Recommended correlation equations for estimation of solution gas-oil ratios at reservoir pressures less than bubblepoint pressure

The recommended correlation equations for solution gas-oil ratios, R_s, at reservoir pressures less than bubblepoint pressure using field data are given below.[20]

$$R_s = (R_{sb})(R_{sr}) \tag{3.8a}$$

where

$$R_{sr} = a_1 p_r^{a_2} + (1 - a_1) p_r^{a_3} \ , \tag{3.8b}$$

$$p_r = (p - 14.7)/(p_b - 14.7) \ , \tag{3.8c}$$

$$a_1 = A_0 \gamma_{gSP}^{A_1} \ API^{A_2} T^{A_3} (p_b - 14.7)^{A_4} \ , \tag{3.8d}$$

$$a_2 = B_0 \gamma_{gSP}^{B_1} \ API^{B_2} T^{B_3} (p_b - 14.7)^{B_4} \ , \tag{3.8e}$$

and

$$a_3 = C_0 \gamma_{gSP}^{C_1} \ API^{C_2} T^{C_3} (p_b - 14.7)^{C_4} \ . \tag{3.8f}$$

$A_0 = 9.73 \times 10^{-7}$	$B_0 = 0.022339$	$C_0 = 0.725167$
$A_1 = 1.672608$	$B_1 = -1.004750$	$C_1 = -1.485480$
$A_2 = 0.929870$	$B_2 = 0.337711$	$C_2 = -0.164741$
$A_3 = 0.247235$	$B_3 = 0.132795$	$C_3 = -0.091330$
$A_4 = 1.056052$	$B_4 = 0.302065$	$C_4 = 0.047094$

Coefficients of Isothermal Compressibility of Reservoir Oils

The *coefficient of isothermal compressibility of oil* is usually called *oil compressibility*.

At pressures above bubblepoint pressure, oil compressibility simply represents the expansion of the reservoir oil as reservoir pressure decreases. At pressures below bubblepoint pressure, the oil compressibility includes the change in volume of the oil (which is often a decrease in oil volume due to the release of solution gas) plus the increase of volume of the released solution gas in the reservoir.

Thus, the defining equations for this independent variable are different depending on whether the reservoir pressure is above or below bubblepoint pressure. This is the only oil fluid property that has a discontinuity at the bubblepoint pressure. Just below the bubblepoint pressure, the value of oil compressibility could be three to four times greater than the value just above the bubblepoint pressure.

Coefficients of isothermal compressibility of oil at reservoir pressures greater than bubblepoint pressure

The defining equations for coefficients of isothermal compressibility of oils (oil compressibilities), shown here as equation (3.9), are very simple. They are simply the fractional change of volume or density as pressure changes at constant temperature.

$$c_o = -\frac{1}{V}\left(\frac{\partial V}{\partial p}\right)_T \qquad c_o = -\frac{1}{B_o}\left(\frac{\partial B_o}{\partial p}\right)_T \qquad c_o = \frac{1}{\rho_o}\left(\frac{\partial \rho_o}{\partial p}\right)_T \tag{3.9}$$

However, there are some complexities in application. Three different applications for this property are found in petroleum engineering calculations:

1. Use of oil compressibilities in estimating certain fluid properties at pressures above bubblepoint pressure
2. Use of oil compressibilities in the black oil material balance equation at pressures above bubblepoint pressure
3. Use of oil compressibilities in single-phase (oil) pressure transient analysis

Application 1 can be illustrated with the coefficient of isothermal compressibility equation written in terms of oil density.

$$c_o = \frac{1}{\rho_o}\left(\frac{\partial \rho_o}{\partial p}\right)_T \tag{3.10a}$$

$$c_o \int_{p_b}^{p} dp = \int_{\rho_{ob}}^{\rho_o} \frac{d\rho_o}{\rho_o} \tag{3.10b}$$

$$\rho_o = \rho_{ob} \,\mathrm{EXP}\left[c_o\left(p - p_b\right)\right] \tag{3.10c}$$

The integration is from the bubblepoint pressure to some selected pressure higher than bubblepoint pressure. This will permit values of oil densities at higher pressures to be calculated from a known value of oil density at the bubblepoint pressure.

Placing the oil compressibility, c_o, outside the integral sign in equation (3.10b) implies that it is constant as pressure changes. This is not correct, as can be seen in figure 3–8. Performing the integration as indicated in equation (3.10b) implies that the oil compressibility is some weighted average of oil compressibilities from the bubblepoint pressure to the higher pressure. This oil compressibility will be called c_{ofb}, i.e., weighted-average oil compressibility *from bubblepoint pressure to a higher pressure* of interest.

Application 2 can be illustrated with the material balance equation for undersaturated oil reservoirs.[21]

$$c_o = \frac{B_o - B_{oi}}{B_{oi}\left(p_i - p\right)} \tag{3.11a}$$

$$c_e = \frac{S_o c_o + S_w c_w + c_f}{S_o} \tag{3.11b}$$

$$N B_{oi} c_e\left(p_i - p\right) = N_p B_o - W_e + B_w W_p \tag{3.11c}$$

The pressure changes in both equations (3.11a) and (3.11c) are from initial reservoir pressure, p_i, to some lower pressure of interest; both pressures above the bubblepoint pressure. Thus, this oil compressibility is some weighted-average from initial reservoir pressure to some selected lower pressure. This oil compressibility will be called c_{ofi}, i.e., average oil compressibility *from initial pressure* to a lower pressure of interest.

Application 3 can be illustrated with the continuity equation for radial flow, which is the starting point of the derivation of equations for pressure transient analysis.[22]

$$\frac{1}{r}\frac{\partial}{\partial r}\left(\frac{r\rho k_r}{\mu}\frac{\partial p}{\partial r}\right) = \frac{\partial}{\partial t}\left(\phi\rho\right) \tag{3.12a}$$

For single phase oil, the derivative on the right-hand side can be replaced using equations (3.12b) and (3.12c).

$$c_o = \frac{1}{\rho_o}\frac{\partial \rho_o}{\partial p} \tag{3.12b}$$

$$\frac{1}{\rho_o}\frac{\partial \rho_o}{\partial t} = c_o \frac{\partial p}{\partial t} \tag{3.12c}$$

This results in equation (3.12d).

$$\frac{\partial^2 p}{\partial r^2} + \frac{1}{r}\frac{\partial p}{\partial r} = \frac{\phi \mu c_o}{k_r}\frac{\partial p}{\partial t} \tag{3.12d}$$

The use of oil compressibility in the final equation did not require integration. Thus, the derivative in the oil compressibility equation, equation (3.12b), must be determined with the slope of a tangent line to the isothermal density-pressure curve at the pressure of interest. This oil compressibility will be called c_o.

These three applications require very different values of oil compressibilities. Figure 3–8 shows values of oil compressibilities calculated using the constant composition expansion data reported in a well-known sample reservoir fluid study.[23] A value of 5,000 psig was used as initial pressure for the reservoir for the c_{ofi} calculations. This figure shows that the values of oil compressibilities used in the three applications are 20%–30% different depending on the pressure of interest.

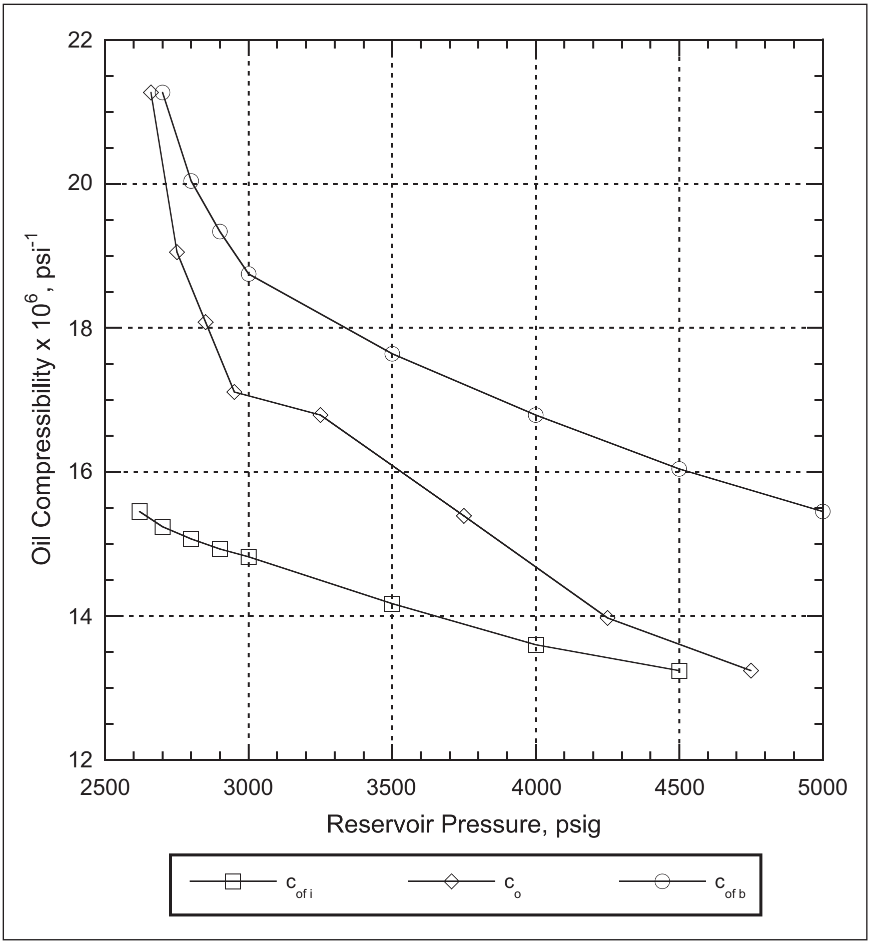

Fig. 3–8. A comparison of values of oil compressibilities from initial pressure, c_{ofi}, instantaneous oil compressibilities, c_o, and oil compressibilities from bubblepoint pressure, c_{ofb}, calculated using data from a well-known sample laboratory fluid study shows the large differences among these properties. *Source: Data from Core Laboratories, Inc., 1994.*

Evaluation of proposed correlation equations for coefficients of isothermal compressibility of oil at reservoir pressures greater than bubblepoint pressure

Several correlation equations reported in the petroleum literature were evaluated with the data set summarized in table 3–11. The 3,051 lines of data were all at or above the bubblepoint pressures of each of the 399 laboratory reports. The underlying samples were of worldwide origin.

Table 3–11. Statistical summary of database used to evaluate oil compressibility correlations at pressures greater than bubblepoint pressure (3,051 lines of data from 399 reservoir fluid studies with worldwide origins). Data within 200 psi of the bubblepoint pressure were eliminated due to extremely high errors caused by round-off errors for pressures this close to bubblepoint pressure.

Laboratory measurement	Minimum	Median	Mean	Maximum
Reservoir pressure, p, psia	414.7	3,714.7	3,786.5	14,915.0
Bubblepoint pressure, p_b, psia	120.7	2,085.7	2,346.3	7,854.0
Reservoir temperature, T_R, °F	70.0	190.0	192.7	320.0
Stock-tank oil gravity, API, °API	11.6	37.7	36.4	57.7
Separator gas specific gravity, γ_{gSP}	0.561	0.791	0.804	1.798
Solution gas-oil ratio at p_b, R_{sb}, scf/STB	12.0	567.0	636.9	1,808.0
Oil compressibility, c_{ofb},1/psi x 10^6	3.6	11.9	13.4	50.3

Often the authors of proposed oil compressibility correlation equations do not specify whether the resulting values were c_o or c_{ofb} (the assumption is that none were c_{ofi}). If so, both cases were tried in the evaluation process, and the one with the lowest ARE and AARE was selected for each correlation. The results of the evaluation comparing calculated results with the data described in table 3–11 are given in table 3–12.

Table 3–12. Comparison of published correlations for the coefficients of isothermal compressibility of oils for reservoir pressures above the bubblepoint pressure reveals the most reliable correlations.

	Predicted coefficients of isothermal compressibility of oils	
Correlation	**ARE, %**	**AARE, %**
Spivey et al. (2007) c_{ofb} equations*	–0.1	6.8
Spivey et al. (2007) c_o equations†	2.1	8.4
Al-Marhoun (2003)[a]*	–3.9	12.6
Dindoruk-Christman (2004)[b]*	–5.2	12.7
De Ghetto et al. "Light Oil" °API > 31.1 (1994)[c]*	–10.8	14.8
Petrosky-Farshad (1998)[d]†‡	–3.8	15.5
Labedi (1990)†	–12.6	17.2
Almehaideb (1997) [e]†‡	–12.4	19.4
Whitson-Brule (2000)[f]†	–13.4	20.2
Hanafy et al. (SPE 37439, 1997)/Calhoun (1953)[g]†‡	–2.3	24.6
Vazquez-Beggs (1980)*	–6.8	26.2
Hanafy et al. (SPE 37439, 1997)*	22.2	26.4
Elsharkawy-Alikhan (1997)[h]†	–4.1	31.3
Kartoatmodjo-Schmidt (1994) †	–25.7	32.0
Ahmed equation 4–46 (1989)*	11.6	39.9
Farshad et al. (1996)[i]*‡	46.0	48.6

[a] Al-Marhoun, M.A. "The Coefficient of Isothermal Compressibility of Black Oils." Paper SPE 81432. Presented at the SPE Middle East Oil Show, Bahrain (2003).

[b] Dindoruk, B. and Christman, P.G. "PVT Properties and Viscosity Correlations for Gulf of Mexico Oils." *Society of Petroleum Engineers Reservoir Evaluation and Engineering*. Vol. 7, No. 6 (2004) 427–437.

[c] De Ghetto, G., Paone, F., and Villa, M. "Reliability Analysis on PVT Correlations." Paper SPE 28904. Presented at the European Petroleum Conference, London (1994).

[d] Petrosky, G.E. and Farshad, F.F. "Pressure-Volume-Temperature Correlations for Gulf of Mexico Crude Oils." *Society of Petroleum Engineers Reservoir Evaluation and Engineering*. Vol. 1 (October 1998) 416–420.

[e] Almehaideb, R.A. "Improved PVT Correlations for UAE Crude Oils." Paper SPE 37691. Presented at the SPE Middle East Oil Conference and Exhibition, Bahrain (1997).

[f] Whitson, C.H. and Brule, M.R. *Phase Behavior*. Richardson, TX: Society of Petroleum Engineers (2000) 35–37.

[g] Calhoun, J.C., Jr. *Fundamentals of Reservoir Engineering*. Norman, OK: University of Oklahoma Press (1953) 35.

[h] Elsharkawy, A.M. and Alikhan, A.A. "Correlations for Predicting Solution Gas/Oil Ratio, Oil Formation Volume Factor, and Undersaturated Oil Compressibility." *Journal of Petroleum Science and Engineering*. Vol. 17 (1997) 291–302.

[i] Farshad, F.F., LeBlanc, J.L., Gruber, J.D., and Osorio, J.G. "Empirical PVT Correlations for Columbian Crude Oils." Paper SPE 36105. Presented at the Fourth Latin American and Caribbean Petroleum Engineering Conference, Port-of-Spain (1966).

* Correlation results compared to values of c_{ofb} calculated from the data set.

† Correlation results compared to values of c_o calculated from the data set.

‡ Author restricted the correlation to a specific geographical area.

Spivey et al. used approximately 90% of the data of table 3–11 to prepare their correlation equations; Vazquez and Beggs apparently used most of these data as well as additional data in their work.[24] Apparently the other authors did not have access to the data set described in table 3–11.

Figures 3–9, 3–10, 3–11, and 3–12 show the evaluations of four correlation equations using the data set sorted and sliced into 10 approximately equal increments for each of the independent variables. Calculations with the Whitson-Brule equation were compared with values of c_o calculated with the data set; results from the other three correlations were compared with values of c_{ofb} from the data set.

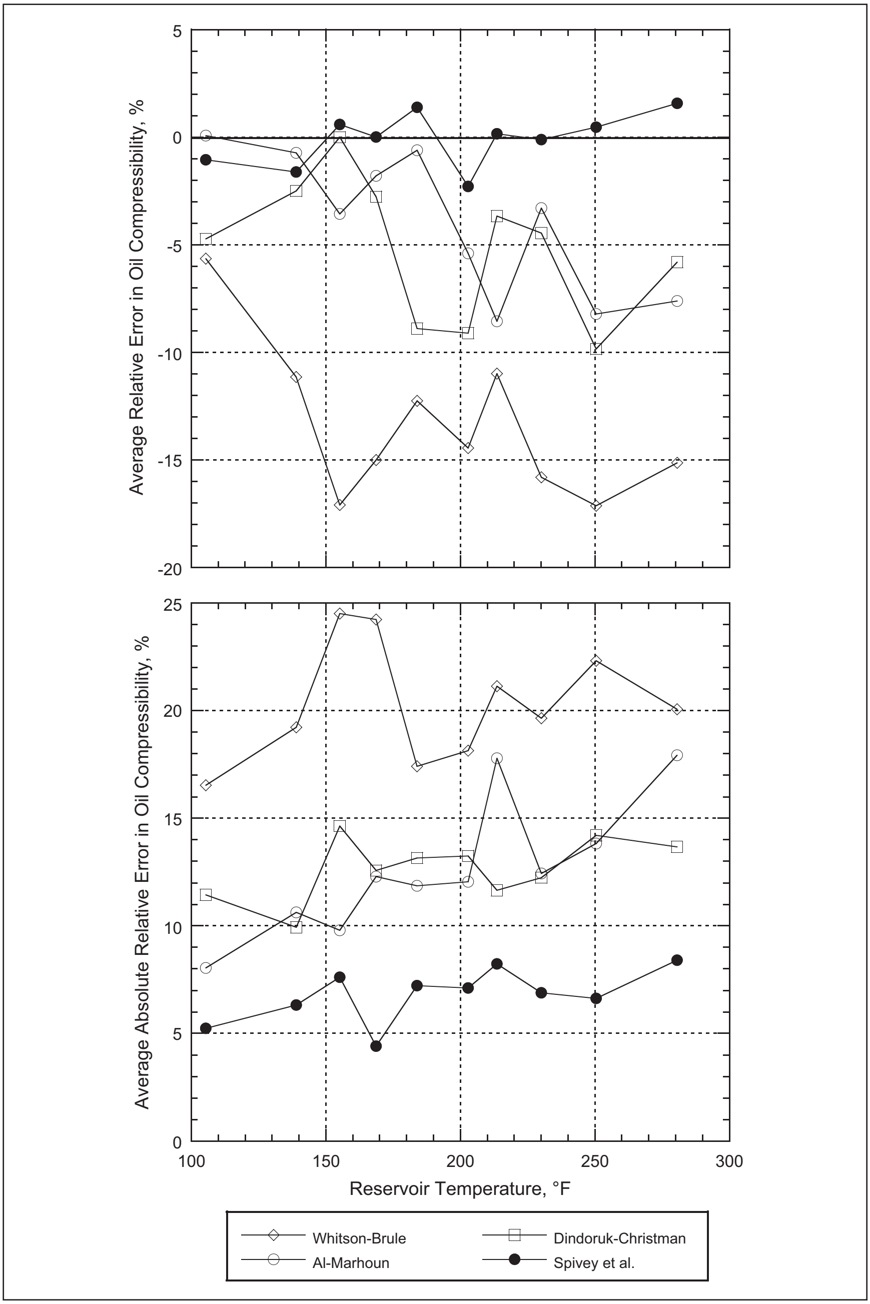

Fig. 3–9. Average relative errors and average absolute relative errors of subsets of the data described in table 3–11 sorted on reservoir temperature show the quality of selected correlation equations for estimating oil compressibilities for pressures greater than bubblepoint pressure across the full range of this independent variable.

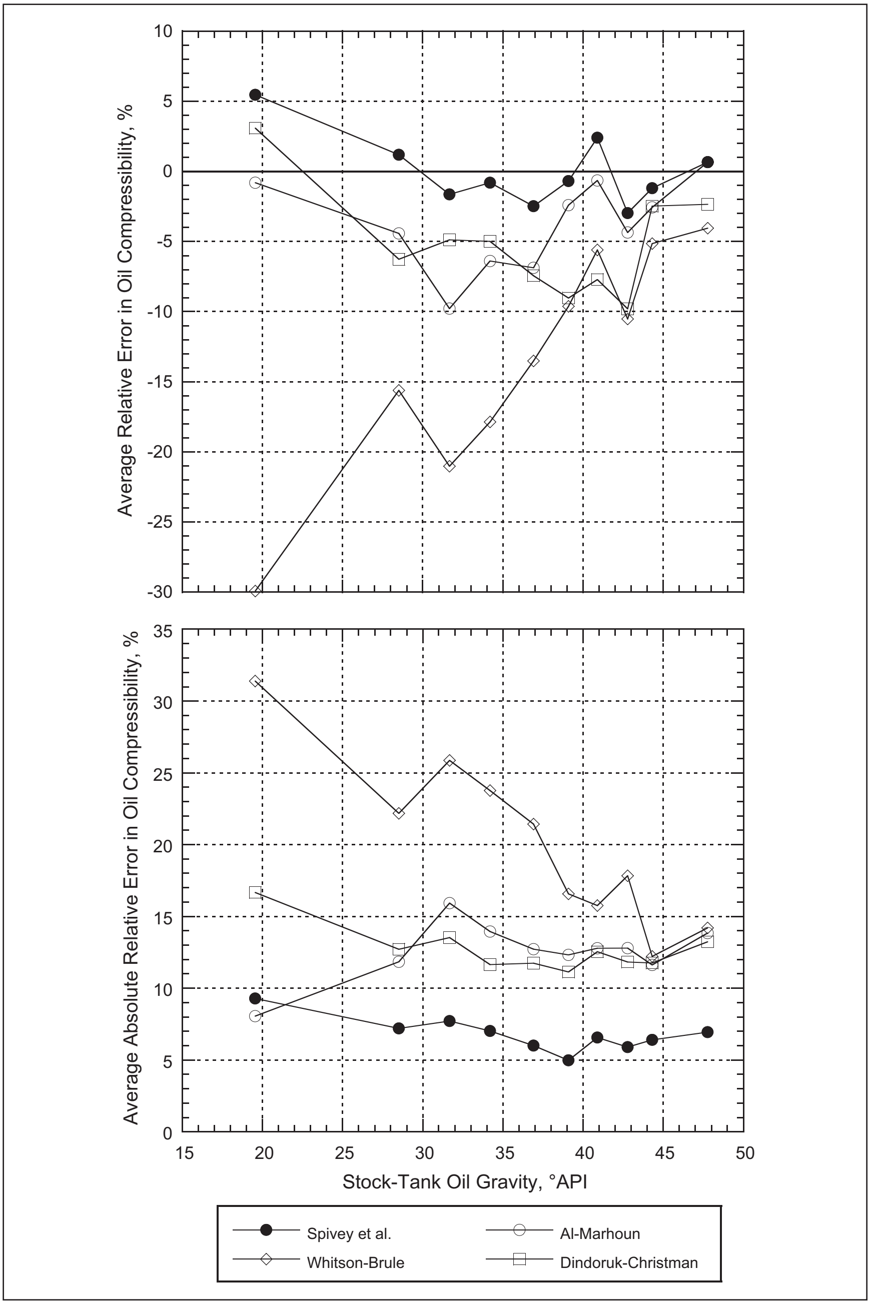

Fig. 3–10. Average relative errors and average absolute relative errors of subsets of the data described in table 3–11 sorted on stock-tank oil gravity show the quality of selected correlation equations for estimating oil compressibilities for pressures greater than bubblepoint pressure across the full range of this independent variable.

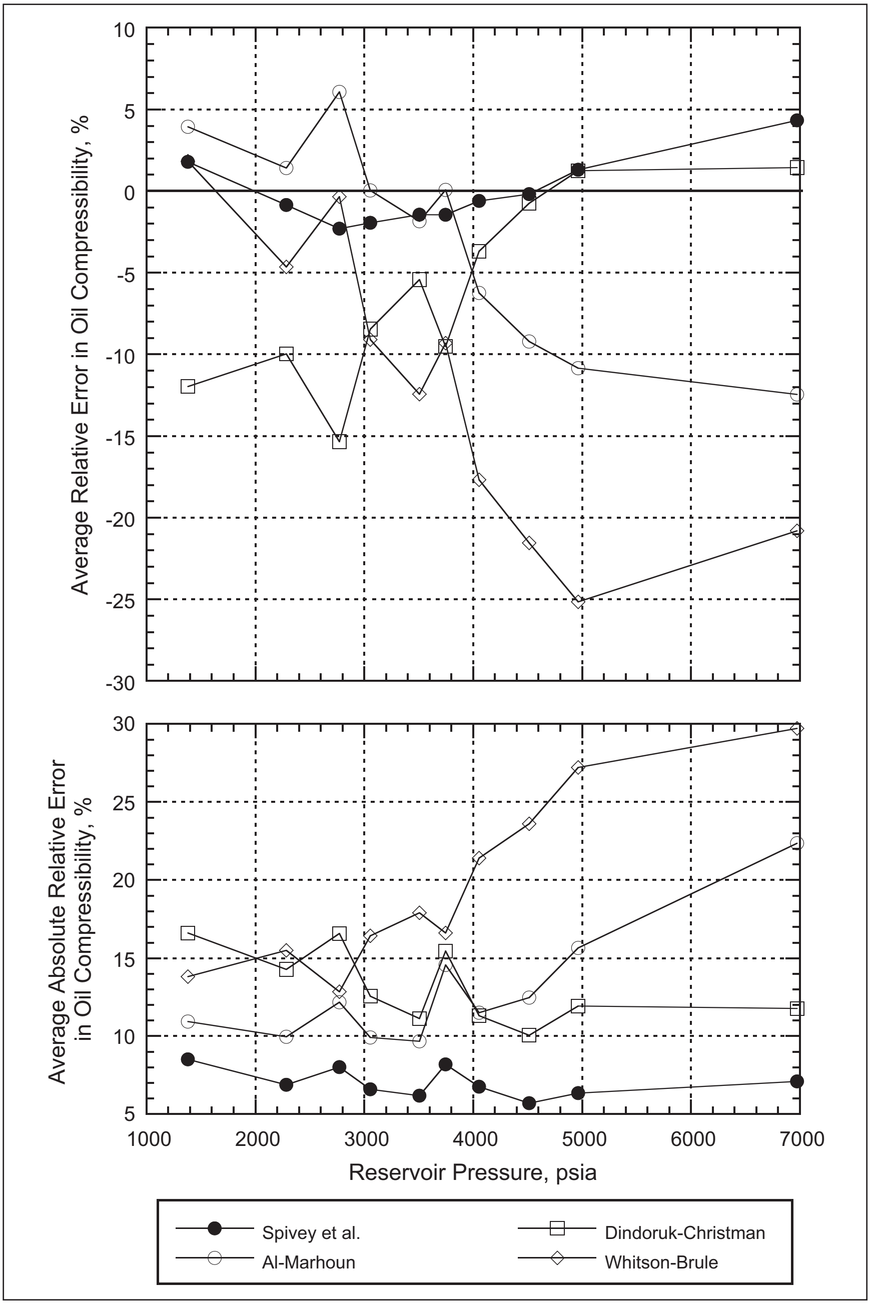

Fig. 3–11. Average relative errors and average absolute relative errors of subsets of the data described in table 3–11 sorted on reservoir pressure show the quality of selected correlation equations for estimating oil compressibilities for pressures greater than bubblepoint pressure across the full range of this independent variable.

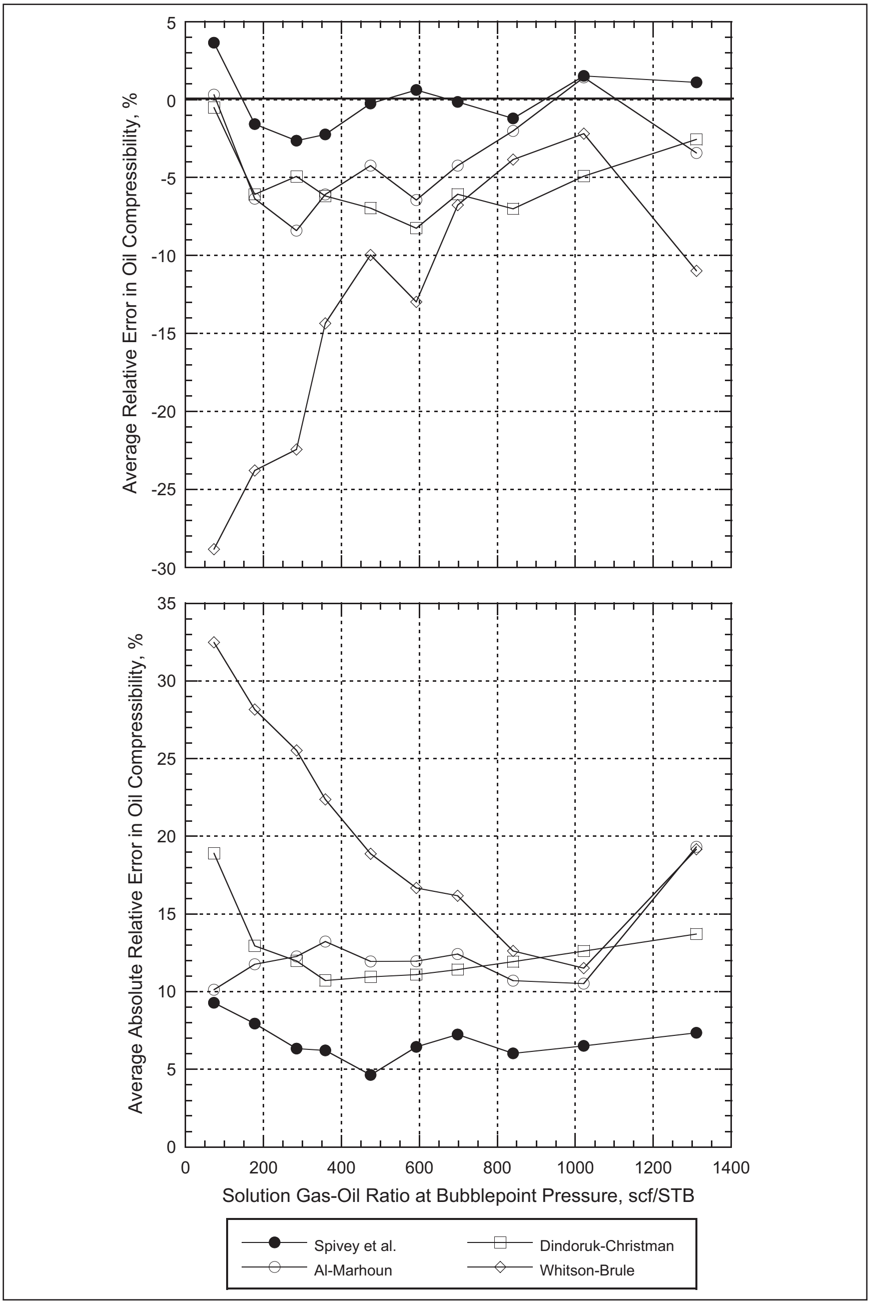

Fig. 3–12. Average relative errors and average absolute reltative errors of subsets of data described in table 3–11 sorted on solution gas-oil ratio at bubblepoint pressure show the quality of selected correlation equations for estimating oil compressibilities for pressures greater than bubblepoint pressure across the full range of this independent variable.

Recommended correlation equations for estimation of coefficients of isothermal compressibility of oil at reservoir pressures greater than bubblepoint pressure

The recommended equations for the three versions of oil compressibility are given below.[25]

Coefficients of isothermal compressibility for use from bubblepoint pressure to some higher reservoir pressure, c_{ofb}, can be estimated with equation (3.13a).

$$\ln\left[c_{of\,b}\right] = 2.434 + 0.475Z + 0.048Z^2 - \ln\left(10^6\right) \tag{3.13a}$$

where

$$Z = \sum_{n=1}^{6} Z_n \tag{3.13b}$$

and

$$Z_n = C0_n + C1_n VAR_n + C2_n VAR_n^{\;2} \tag{3.13c}$$

n	***VAR$_n$***	***C0$_n$***	***C1$_n$***	***C2$_n$***
1	ln API	3.011	–2.6254	0.497
2	ln γ_{gSP}	–0.0835	–0.259	0.382
3	ln p_b	3.51	–0.0289	–0.0584
4	ln p/p_b	0.327	–0.608	0.0911
5	ln R_{sb}	–1.918	–0.642	0.154
6	ln T_R	2.52	–2.73	0.429

Coefficients of isothermal compressibility for use from initial reservoir pressure to some lower reservoir pressure, both pressures above the bubblepoint pressure, c_{ofi}, can be estimated with equation (3.14).

$$c_{of\,i} = \frac{\left[(p_b - p)c_{of\,b}(p) - (p_b - p_i)\,c_{of\,b}(p_i)\right]}{p_i - p} \tag{3.14}$$

where c_{ofb} (p) is oil compressibility from the bubblepoint pressure, p_b, to the pressure of interest, p, and c_{ofb} (p_i) is oil compressibility from the bubblepoint pressure, p_b, to the initial pressure, p_i .

Coefficients of isothermal compressibility, tangent at reservoir pressures above bubblepoint pressure, c_o , can be estimated with equations (3.15a)–(3.15c).

$$c_o = c_{of\,b}(p) + (p - p_b)\frac{\partial c_{of\,b}}{\partial p} \tag{3.15a}$$

The derivative in equation (3.15a) may be obtained using equation (3.13), and the value of Z for equation (3.15b) should be calculated using equation (3.13b).

$$\frac{\partial c_{of\,b}}{\partial p} = c_{of\,b}(p) \times (0.475 + 0.096Z)\frac{\partial Z}{\partial p} \tag{3.15b}$$

$$\frac{\partial Z}{\partial p} = \frac{-0.608 + 0.1822 \ln \frac{p}{p_b}}{p} \tag{3.15c}$$

Coefficients of isothermal compressibility of oil at reservoir pressures less than bubblepoint pressure

Martin defined the coefficient of isothermal compressibility at pressures below the bubblepoint pressure as equation (3.16a).[26]

$$c_o = -\frac{1}{B_o}\left[\left(\frac{\partial B_o}{\partial p}\right)_T - B_g\left(\frac{\partial R_S}{\partial p}\right)_T\right] \tag{3.16a}$$

This equation takes into account the change in the volume of oil as well as the change in the volume of free gas, both referred to oil volume, as pressure changes.

Equation (3.16a) can be written in the form of equation (3.16b), which is convenient for use in calculating values of c_o from laboratory data, since a plot of B_o vs. R_s is linear, resulting in a derivative that is constant for a particular reservoir fluid study.

$$c_o = -\frac{1}{B_o}\left(\frac{\partial R_S}{\partial p}\right)_T\left[\left(\frac{\partial B_o}{\partial R_S}\right)_T - B_g\right] \tag{3.16b}$$

Ramey suggested that the derivatives of B_o and R_s for equation (3.16a) be evaluated using the best available correlation equations for each independent variable rather than using a separate correlation for oil compressibility, c_o.[27] Only one correlation equation has been proposed in the petroleum literature for this fluid property; McCain et al.[28]

Evaluation of proposed correlation equations for coefficients of isothermal compressibility at reservoir pressures less than bubblepoint pressure

The data set described in table 3–9 was used to calculate values of c_o at pressures below bubblepoint pressure for comparison with the McCain et al. equation and the Ramey suggestion. Equation (3.16b) was used to determine values of c_o from the data.

Equation (3.16a) was used to calculate values of c_o from the correlations. The solution gas-oil ratio derivative, $\partial R_s/\partial p$, was obtained with equations (3.8a)–(3.8f). The formation volume factor derivative, $\partial B_o/\partial p$, was obtained with equation (3.21b), with the necessary derivative of reservoir oil density, $\partial \rho_{oR}/\partial p$, coming from equations (3.19a)–(3.19g). Oil formation volume factor, B_o, can be obtained with equation (3.21b), and gas formation volume factor, B_g, can be obtained with equation (2.12).

The results of the evaluations are given in table 3–13.

Table 3–13. Performance of correlation equations for coefficients of isothermal compressibility of oils at reservoir pressures below bubblepoint pressure reveals the preferred procedure.

	Predicted coefficients of isothermal compressibility of oils	
Correlation	**ARE, %**	**AARE, %**
Equation (3.16a) with equations (3.17a)–(3.17h)	3.85	10.25
McCain et al. (1988)	16.02	18.41

Figures 3–13, 3–14, and 3–15 show how this procedure holds up across the distribution of independent variables. The independent variables were sorted and sliced into subsets with approximately 450 lines of data in each subset.

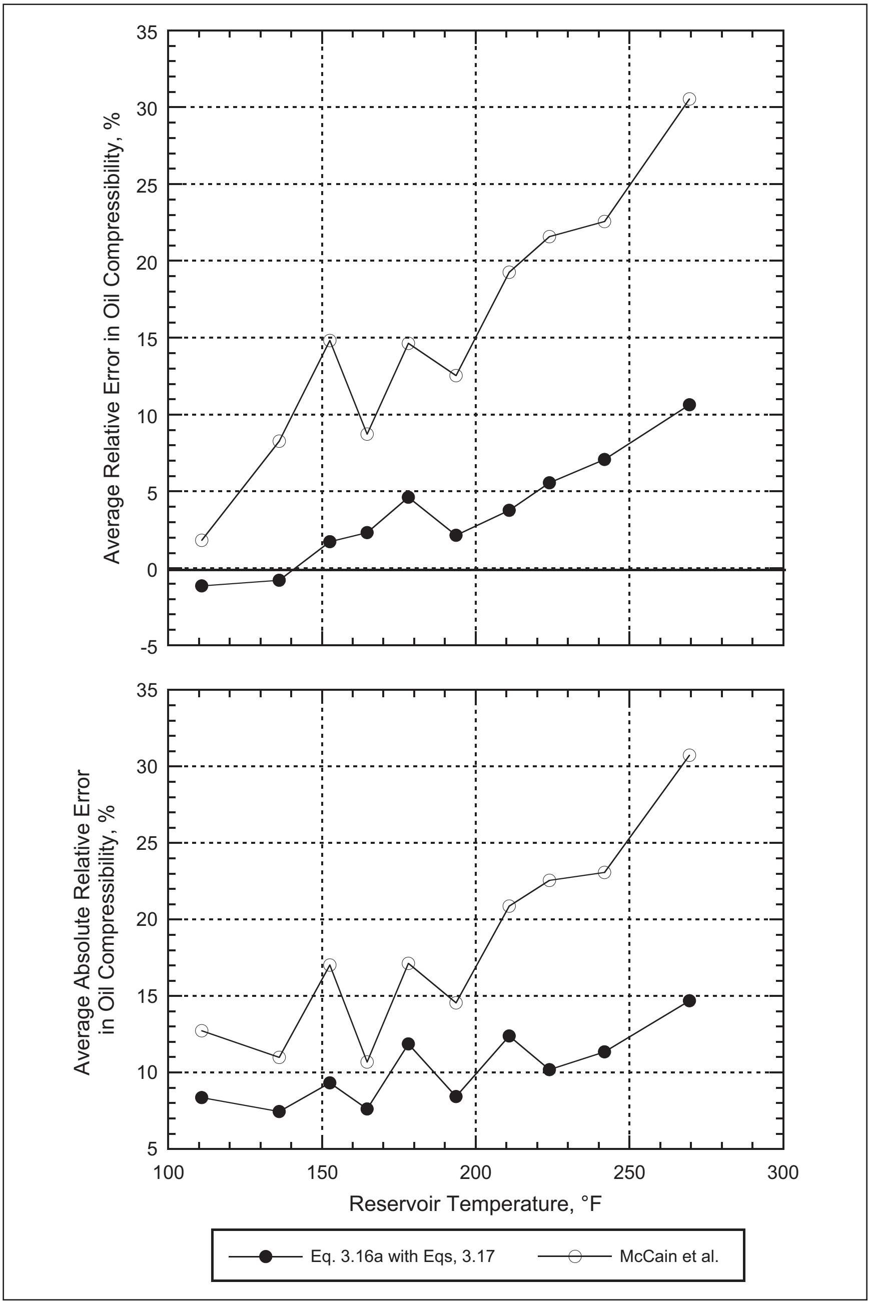

Fig. 3–13. Average relative errors and average absolute relative errors of subsets of the data described in table 3–9 sorted on reservoir temperature show the quality of selected correlation equations for estimating oil compressibilities for pressures less than bubblepoint pressure across the full range of this independent variable.

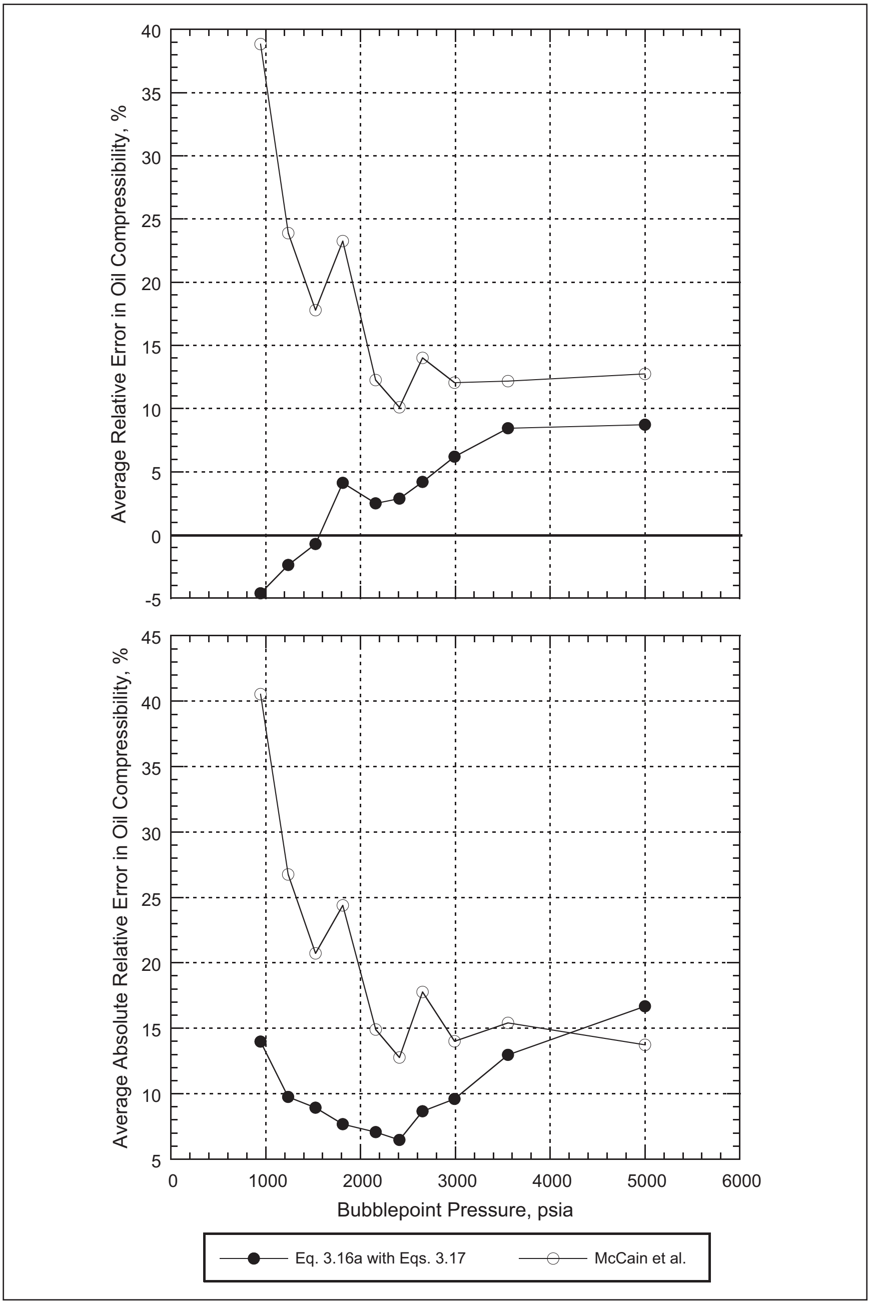

Fig. 3–14. Average relative errors and average absolute relative errors of subsets of the data described in table 3–9 sorted on bubblepoint pressure show the quality of selected correlation equations for estimating oil compressibilities for pressures less than bubblepoint pressure across the full range of this independent variable.

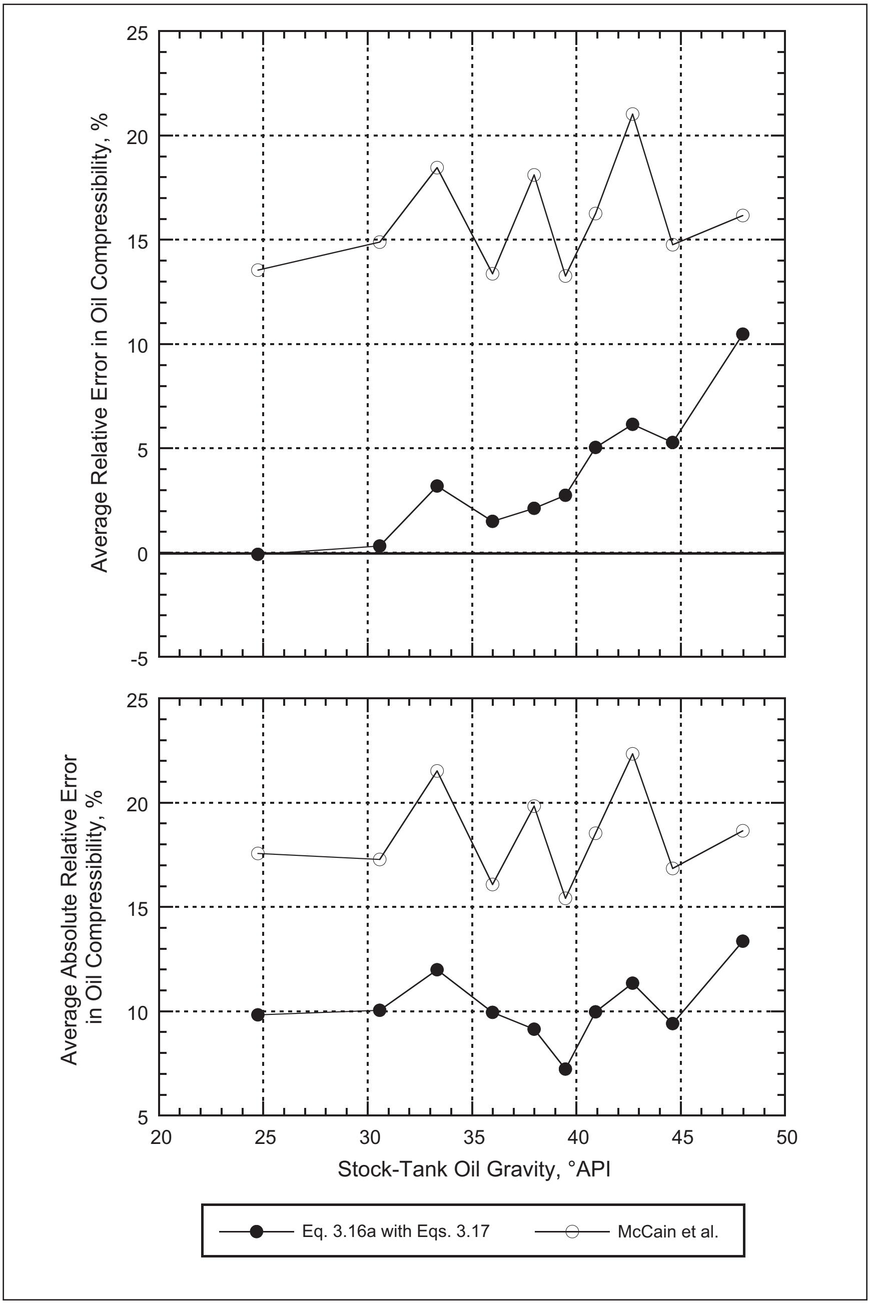

Fig. 3–15. Average relative errors and average absolute relative errors of subsets of the data described in table 3–9 sorted on stock-tank oil gravity show the quality of selected correlation equations for estimating oil compressibilities for pressures less than bubblepoint pressure across the full range of this independent variable.

Recommended correlation equations for estimation of coefficients of isothermal compressibility of oil at reservoir pressures less than bubblepoint pressure

Equation (3.16a) is recommended. The derivatives for use in this equation are taken from the recommended correlation equation, equation (3.17a).

$$\left(\frac{\partial R_s}{\partial p}\right)_T = R_{sb}\frac{\left(a_1 a_2 p_r^{(a_2-1)} + (1-a_1)a_3 p_r^{(a_3-1)}\right)}{\left(p_b - 14.7\right)} \tag{3.17a}$$

Equation (3.17a) was derived using equations (3.8a)–(3.8f). The values of the coefficients a_1, a_2, and a_3 are calculated with the constants given with equations (3.8a)–(3.8f).

$$\left(\frac{\partial B_o}{\partial p}\right)_T = \frac{1}{\rho_{oR}^{\;2}}\left[0.01357\gamma_g\rho_{oR}\left(\frac{\partial R_s}{\partial p}\right) - \left(\rho_{STO} + 0.01357R_s\gamma_g\right)\left(\frac{\partial \rho_{oR}}{\partial p}\right)_T\right] \tag{3.17b}$$

Equation (3.17b) was derived using equation (3.21b).

The derivative of reservoir oil density with respect to pressure, which is required in equation (3.17b), was derived using equation (3.19a).

$$\left(\frac{\partial \rho_{oR}}{\partial p}\right)_T = \left(\frac{\partial \rho_{po}}{\partial p}\right)_T + \left(\frac{\partial \Delta\rho_p}{\partial p}\right)_T - \left(\frac{\partial \Delta\rho_T}{\partial p}\right)_T \tag{3.17c}$$

The first derivative on the right-hand side of equation (3.17c) was obtained using equation (3.19b).

$$\left(\frac{\partial \rho_{po}}{\partial p}\right)_T = \gamma_g\left(\frac{\partial R_s}{\partial p}\right)_T\left[\frac{73.71 - 4600\gamma_{STO}/\rho_a}{\left(73.71 + R_s\gamma_g/\rho_a\right)^2}\right] \tag{3.17d}$$

Values of ρ_a for use in equation (3.17d) may be estimated with good accuracy using equation (3.17e).

$$\rho_a = 38.52\left(10^{-0.00326\,API}\right) + \left(94.75 - 33.93\log API\right)\log\gamma_g \tag{3.17e}$$

The second derivative on the right-hand side of equation (3.17c) was obtained using equation (3.19d).

$$\begin{aligned}\left(\frac{\partial \Delta\rho_p}{\partial p}\right)_T &= 10^{-3}\left[0.167 + 16.181\left(10^{-0.0425\rho_{po}}\right)\right]\\ &\quad -10^{-3}\left[1.5835\left(10^{-0.0425\rho_{po}}\right)p\left(\frac{\partial \rho_{po}}{\partial p}\right)_T\right]\\ &\quad -10^{-8}\left[0.598p + 526p\left(10^{-0.0603\rho_{po}}\right)\right]\\ &\quad +10^{-8}\left[36.52p^2\left(10^{-0.0603\rho_{po}}\right)\left(\frac{\partial \rho_{po}}{\partial p}\right)_T\right]\end{aligned} \tag{3.17f}$$

The third derivative on the right-hand side of equation (3.17c) was obtained using equation (3.19f).

$$\left(\frac{\partial \Delta\rho_T}{\partial p}\right)_T = \left[-1.4313\rho_{bs}^{-1.951}(T-60)^{0.938} - 0.0008638\left(10^{-0.0161\rho_{bs}}\right)(T-60)^{0.475}\right]\left(\frac{\partial \rho_{bs}}{\partial p}\right)_T \tag{3.17g}$$

where

$$\left(\frac{\partial \rho_{bs}}{\partial p}\right)_T = \left(\frac{\partial \rho_{po}}{\partial p}\right)_T + \left(\frac{\partial \Delta\rho_p}{\partial p}\right)_T. \tag{3.17h}$$

Values of oil formation volume factors, B_o, for use in equation (3.16a) can be obtained with either equation (3.21a) or (3.21b) as appropriate. Values of gas formation volume factors, B_g, for use in equation (3.16a) may be calculated with equation (2.12) with z-factors, z, obtained with equations (2.4)–(2.8). This procedure requires values of the specific gravities of the reservoir gas, γ_{gR}, which can be estimated at pressures of interest using equation (3.25).

Densities of Reservoir Oils

There is a change in slope at the bubblepoint pressure of a plot of reservoir oil density versus reservoir pressure. This slope is negative for pressures less than bubblepoint pressure and positive for pressures greater than bubblepoint pressure. A set of correlation equations will be used for estimating values of this property at pressures less than bubblepoint pressure. An equation based on the definition of the coefficient of isothermal compressibility, c_{ofb}, will be used for calculating values of this property at pressures greater than bubblepoint pressure. However, care must be taken in the selection of these equations such that the relationship is continuous, i.e., the values of oil density coincide at the bubblepoint pressure.

Evaluation of proposed correlation equations for reservoir oil densities at reservoir pressures equal to bubblepoint pressure

Statistics of the data set used to evaluate reservoir oil density correlation equations for reservoir pressures equal to bubblepoint pressures are shown in table 3–14.

Table 3–14. Statistics of data set used to evaluate oil density correlation equations at bubblepoint pressures (756 lines of data from 756 reservoir fluid studies with worldwide origins)

Laboratory measurement	Minimum	Median	Mean	Maximum
Bubblepoint pressure, p_b, psia	96.7	2,154.7	2,197.7	6,714.7
Reservoir temperature, T_R, °F	74.0	180.0	181.9	327.0
Solution gas-oil ratio at p_b, R_{sb}, scf/STB	10.0	597.0	643.9	1,975.0
Stock-tank oil gravity, API, °API	11.4	39.2	37.8	55.0
Separator gas specific gravity, γ_{gSP}	0.5560	0.8240	0.8276	1.2370
Reservoir oil density, ρ_{oR}, lb/cu ft	31.12	42.97	43.75	58.20

An evaluation using the data set described in table 3–14 of published correlation equations for estimating reservoir oil densities for reservoir pressures equal to bubblepoint pressures may be found in table 3–15.

Table 3–15. Evaluation of published reservoir oil density correlation equations at bubblepoint pressures reveals the most reliable correlations.

	Predicted reservoir oil densities	
Correlation	**ARE, %**	**AARE, %**
Standing (1977)/Witte (1987)/McCain-Hill (1995)	0.39	0.88
Standing (1977)/Witte (1987)	0.61	1.03
Standing (1977)*	1.14	1.39
Labedi (1990)	–0.24	1.53
Ahmed equation (4.32) (1989)†	–1.17	1.66
Ahmed equation (4.35) (1989)‡	0.34	2.10
Sutton (2008)	0.13	2.12
Hanafy et al. (SPE 37439, 1997)	–0.06	4.24

* Author restricted the correlation to a specific geographical area. † Author combined Standing B_{ob} equation (Standing, 1977) with McCain ρ_{oR} equation (McCain, 1973). ‡ Author used Standing $\Delta\rho_p$ and $\Delta\rho_T$ equations (Standing, 1977).

The Witte and McCain-Hill procedures are modifications of Standing's original set of correlation equations.[29] Witte modified the Standing equation, which determines the density difference due to thermal expansion from 60°F to reservoir temperature, $\Delta\rho_T$, because Standing did not have sufficient data at the lower densities and higher reservoir temperatures to properly determine the density difference in that region. The Witte equation is given in the McCain-Hill paper.[30] McCain and Hill modified the calculation of the apparent liquid density of the surface gas, ρ_a.

Labedi and Sutton proposed procedures that are entirely different than those proposed by Standing.[31]

Witte used less than 10% of the data of table 3–13 to develop his equation for $\Delta\rho_T$; McCain and Hill used approximately 90% of these data in developing their equation for ρ_a. Standing, Labedi, Sutton, and Hanafy et al. apparently did not use any of these data.

The data set described in table 3–14 was sorted and divided into eight subsets of just less than 100 lines of data each for each of the important independent variables. In addition, the data set was sliced on reservoir oil density from the data set. The results are in figures 3–16 through 3–19.

The successive alterations to the original Standing equations did produce some improvement; however, all three versions give good results. The calculated values of reservoir oil densities using these equations are approximately as precise as the underlying data.

The common wisdom that the Standing correlation equations do not give adequate results at low reservoir oil densities and high reservoir temperatures is not verified by figures 3–18 and 3–19.

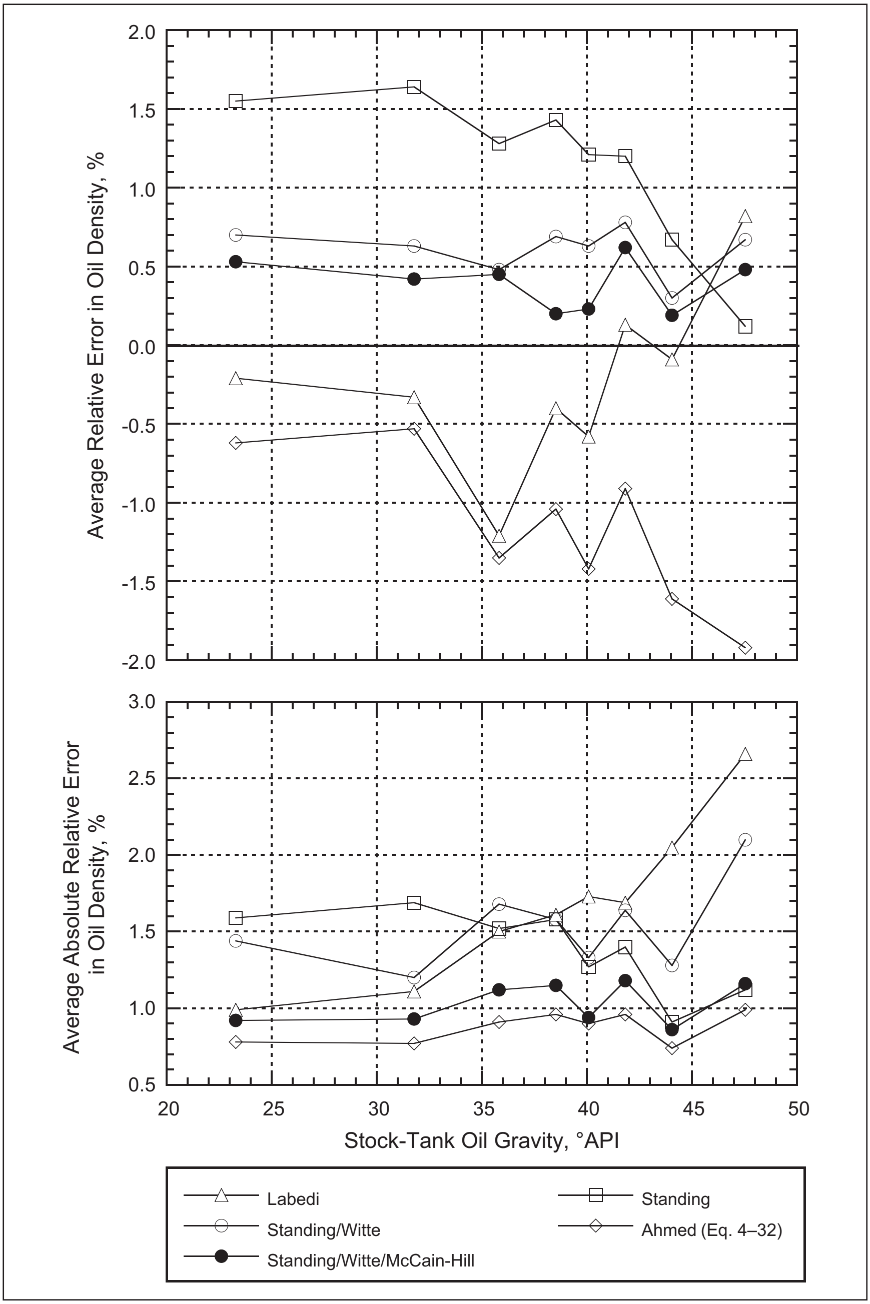

Fig. 3–16. Average relative errors and average absolute relative errors of subsets of the data described in table 3–14 sorted on stock-tank oil gravity show the quality of selected correlation equations for estimating reservoir oil densities at bubblepoint pressure across the full range of this independent variable.

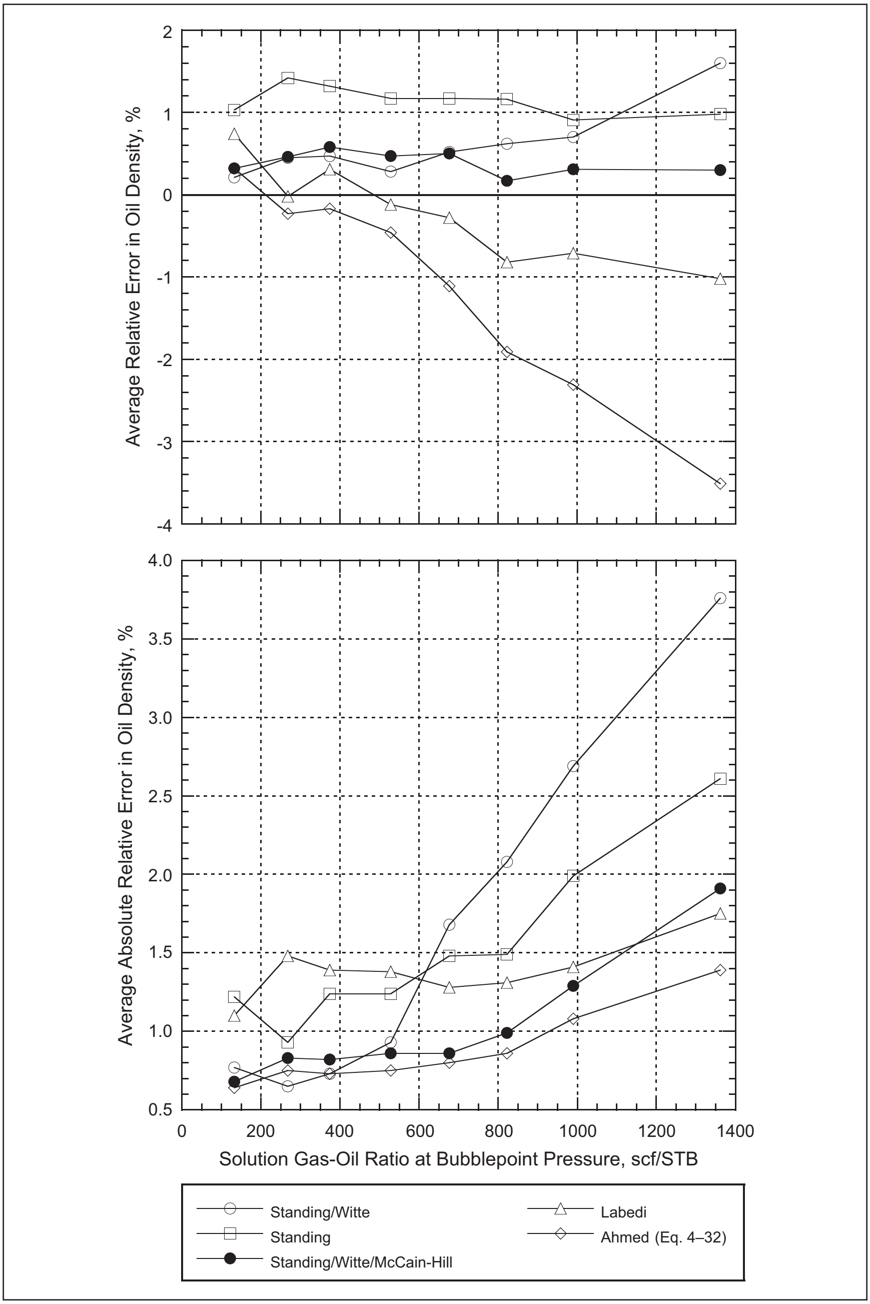

Fig. 3–17. Average relative errors and average absolute relative errors of subsets of the data described in table 3–14 sorted on solution gas-oil ratio at bubblepoint pressure show the quality of selected correlation equations for estimating reservoir oil densities at bubblepoint pressure across the full range of this independent variable.

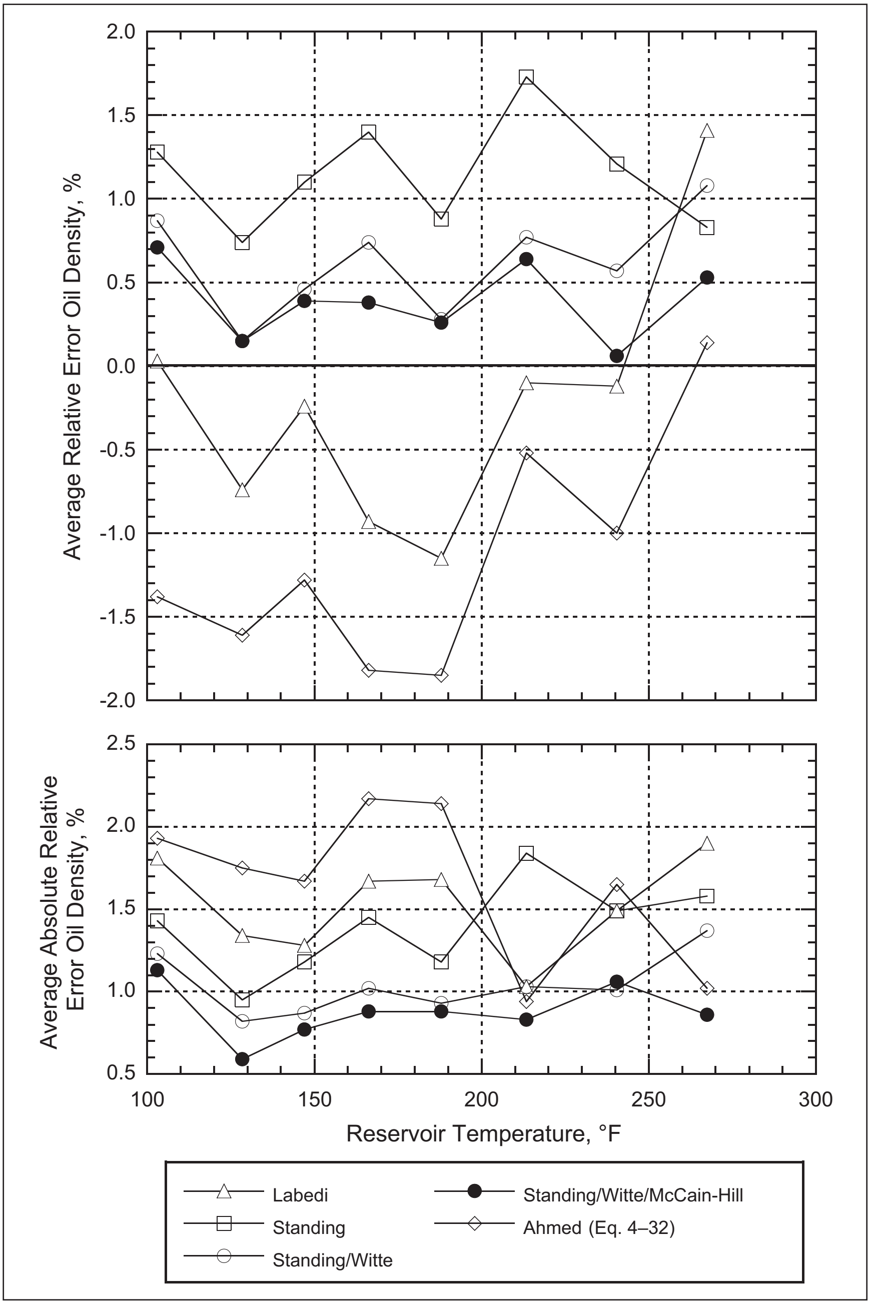

Fig. 3–18. Average relative errors and average absolute relative errors of subsets of the data described in table 3–14 sorted on reservoir temperature show the quality of selected correlation equations for estimating reservoir oil densities at bubblepoint pressure across the full range of this independent variable.

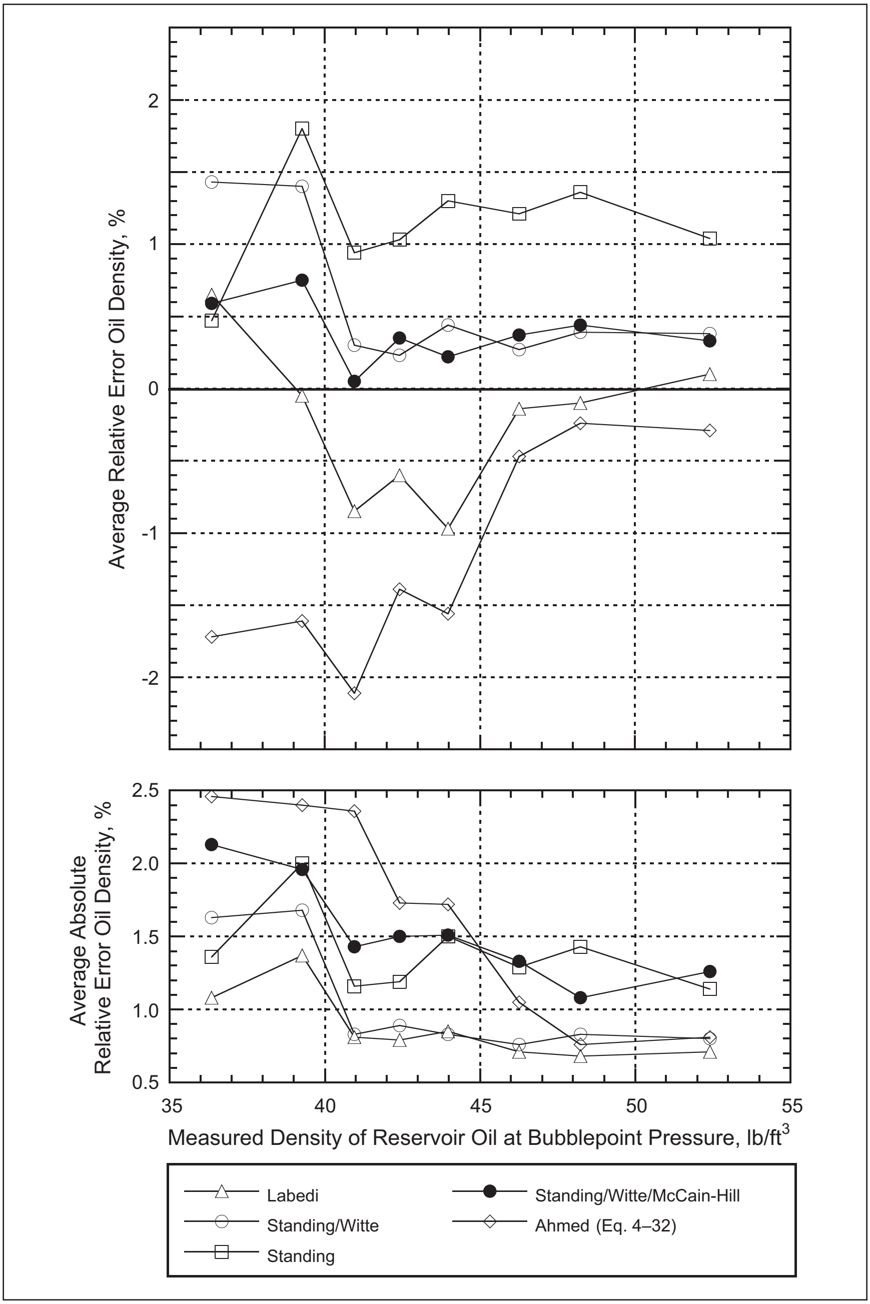

Fig. 3–19. Average relative errors and average absolute relative errors of subsets of the data described in table 3–14 sorted on measured reservoir oil density at bubblepoint pressure show the quality of selected correlation equations for estimating reservoir oil densities at bubblepoint pressure across the full range of this dependent variable.

Recommended correlation equations for estimation of reservoir oil densities at bubblepoint pressure

Recommended equations are given next.[32]

$$\rho_{oRb} = \rho_{po} + \Delta\rho_p - \Delta\rho_T \tag{3.18a}$$

$$\rho_{po} = \frac{R_{sb}\gamma_g + 4600\gamma_{STO}}{73.71 + R_{sb}\gamma_g / \rho_a} \tag{3.18b}$$

$$\rho_a = a_0 + a_1\gamma_{gSP} + a_2\gamma_{gSP}\rho_{po} + a_3\gamma_{gSP}\rho^2{}_{po} + a_4\rho_{po} + a_5\rho^2{}_{po} \tag{3.18c}$$

a_0	–49.8930
a_1	85.0149
a_2	–3.70373
a_3	0.0479818
a_4	2.98914
a_5	–0.0356888

Use of equations (3.18b) and (3.18c) requires a trial-and-error procedure. Equation (3.18d) can be used to calculate a first trial value of ρ_{po}. Successive substitutions for new trial values will converge readily.

$$\rho_{po} = 52.8 - 0.01R_{sb} \tag{3.18d}$$

The adjustment from standard pressure to reservoir pressure may be calculated with equation (3.18e).

$$\begin{aligned}\Delta\rho_p = {} & \left[0.167 + 16.181\left(10^{-0.0425\rho_{po}}\right)\right]\left(\frac{p_b}{1000}\right) \\ & -0.01\left[0.299 + 263\left(10^{-0.0603\rho_{po}}\right)\right]\left(\frac{p_b}{1000}\right)^2\end{aligned} \tag{3.18e}$$

The adjustment from standard temperature to reservoir temperature may be calculated with equations (3.18f) and (3.18g).

$$\rho_{bs} = \rho_{po} + \Delta\rho_p \tag{3.18f}$$

$$\begin{aligned}\Delta\rho_T = {} & \left(0.00302 + 1.505\,\rho_{bs}^{\;-0.951}\right)(T-60)^{0.938} \\ & -\left[0.0216 - 0.0233\left(10^{-0.0161\rho_{bs}}\right)\right](T-60)^{0.475}\end{aligned} \tag{3.18g}$$

Then the reservoir oil density at bubblepoint pressure may be calculated with equation (3.18h).

$$\rho_{oRb} = \rho_{bs} - \Delta\rho_T \tag{3.18h}$$

Note that the value of the gas specific gravity in equation (3.18b) should be the weighted-average specific gravity of the total surface gas. Thus, either equation (3.4) or equation (3.6) is usually necessary with equations (3.5) and (3.2) used to obtain the stock-tank gas properties. Equation (3.18c) requires the specific gravity of the separator gas.

Evaluation of proposed correlation equations for reservoir oil densities at reservoir pressures less than bubblepoint pressure

Statistics of the data set used to evaluate reservoir oil density correlations for reservoir pressures equal to and less than bubblepoint pressures are given in table 3–16.

Table 3–16. Statistics of data set used to evaluate reservoir oil density correlations at pressures equal to and less than bubblepoint pressures (6,475 lines of data from 745 reservoir fluid studies with worldwide origins)

Laboratory measurement	Minimum	Median	Mean	Maximum
Bubblepoint pressure, p_b, psia	139.7	2,114.7	2,213.3	7,750.0
Reservoir temperature, T_R, °F	70.0	182.0	185.4	320.0
Solution gas-oil ratio at p_b, R_{sb}, scf/STB	13.2	520.0	580.2	1,808.0
Stock-tank oil gravity, *API*, °API	11.6	38.3	37.0	56.2
Separator gas specific gravity, γ_{gSP}	0.5548	0.7840	0.7963	1.4720
Reservoir oil density, ρ_{oR}, lb/cu ft	32.57	45.57	45.67	59.89

McCain and Hill showed that the reservoir oil density equations given above, equations (3.18a)–(3.18h), which originally were developed for use at bubblepoint pressures, can be used at pressures below the bubblepoint pressure.[33] Simply substitute a selected value of pressure, p, for the bubblepoint pressure, p_b, in equation (3.18e) and substitute the solution gas-oil ratio at that pressure, R_s, for the solution gas-oil ratio at the bubblepoint, R_{sb}, in equations (3.18b) and (3.18d). Values of solution gas-oil ratio at pressures less than bubblepoint pressure may be estimated with equations (3.8a)–(3.8f).

Figure 3–20 shows that the results are valid even at very low pressures.

An evaluation of published correlations for estimating reservoir oil densities for reservoir pressures equal to and below bubblepoint pressures is in table 3–17. Labedi suggested that his procedure is valid at pressures below bubblepoint pressures. Unfortunately, an evaluation was not possible because the data set did not have values of the pressures of the first-stage separators that are required in his equations.[34] Apparently Sutton did not intend for his correlation to be used at pressures less than bubblepoint pressure.[35]

Table 3–17. Evaluation of published reservoir oil density correlation equations at reservoir pressures equal to and less than bubblepoint pressures reveals the most reliable correlations.

	Predicted reservoir oil densities	
Correlation	**ARE, %**	**AARE, %**
Standing (1977)/Witte (1987)/McCain-Hill (1995)	–0.01	1.40
Standing (1977)/Witte (1987)	–0.59	1.58
Standing (1977)*	0.24	1.67
Ahmed equation 4–32 (1989)[†]	–0.53	1.63
Ahmed equation 4–35 (1989)[‡]	–0.76	2.27
Hanafy et al. (SPE 37439, 1997)*	–1.04	6.33

* Author restricted the correlation to a specific geographical area. [†] Author combined Standing B_{ob} equation (Standing, 1977) with McCain ρ_{oR} equation (McCain, 1973). [‡] Author used Standing $\Delta\rho_p$ and $\Delta\rho_T$ equations (Standing, 1977).

McCain and Hill used approximately 30% of the data described in table 3–16 to evaluate the possibility of using these equations at pressures less than the bubblepoint pressure. Standing, Witte, Labedi, Ahmed, and Hanafy did not use any of these data.

The data described in table 3–16 were sorted and sliced into eight subsets of approximately 800 lines of data based on reservoir oil density, stock-tank oil gravity, solution gas-oil ratio, and reservoir temperature. Values of ARE and AARE for these subsets of the data are given in figures 3–20 through 3–24.

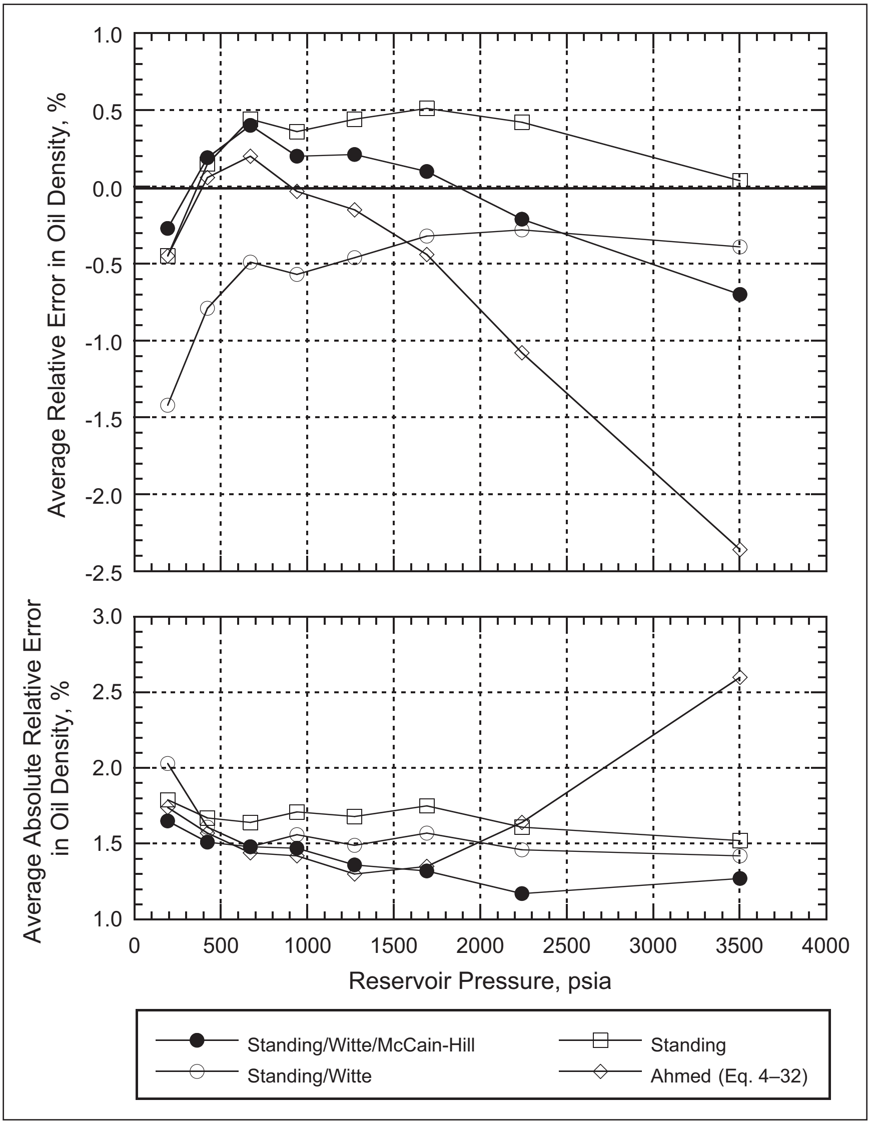

Fig. 3–20. Average relative errors and average absolute relative errors of subsets of the data described in table 3–16 sorted on reservoir pressure show the quality of selected correlation equations for estimating reservoir oil densities for pressures equal to and less than bubblepoint pressure across the full range of this independent variable.

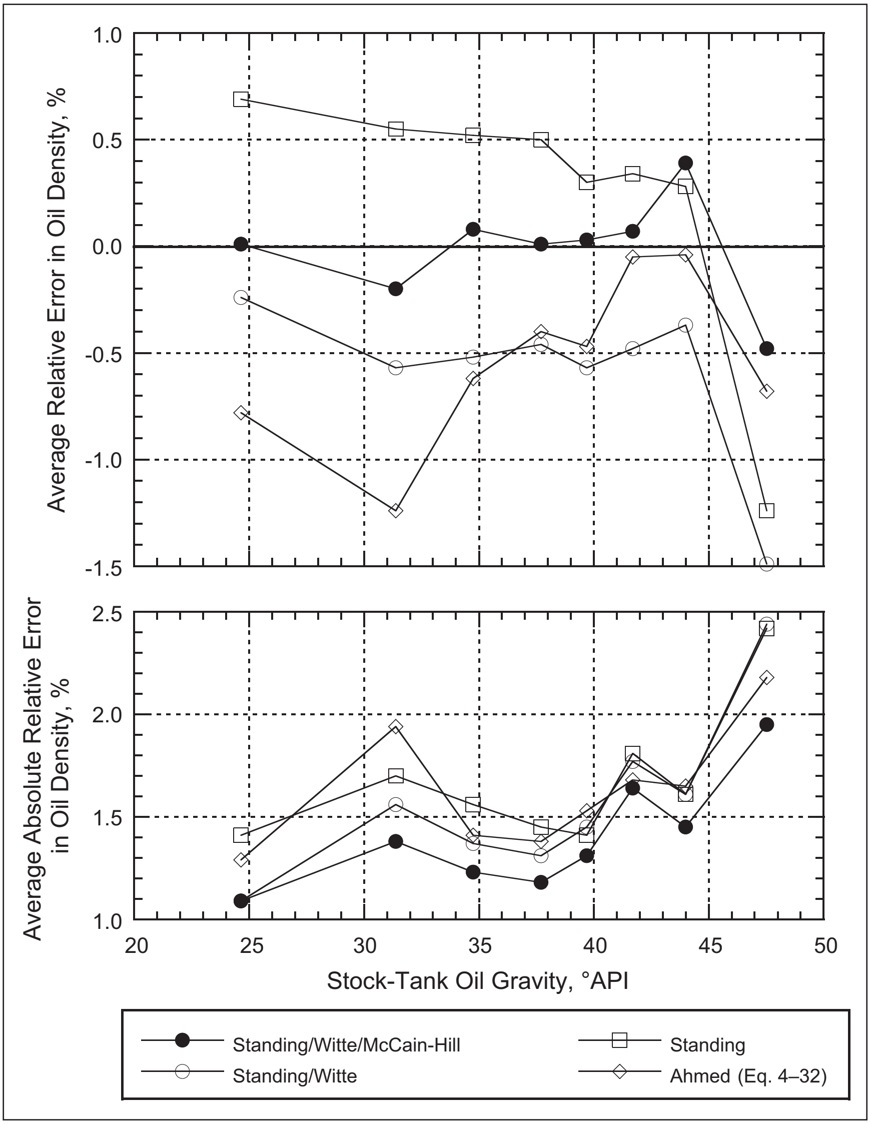

Fig. 3–21. Average relative errors and average absolute relative errors of subsets of the data described in table 3–16 sorted on stock-tank oil gravity show the quality of selected correlation equations for estimating reservoir oil densities for pressures equal to and less than bubblepoint pressure across the full range of this independent variable.

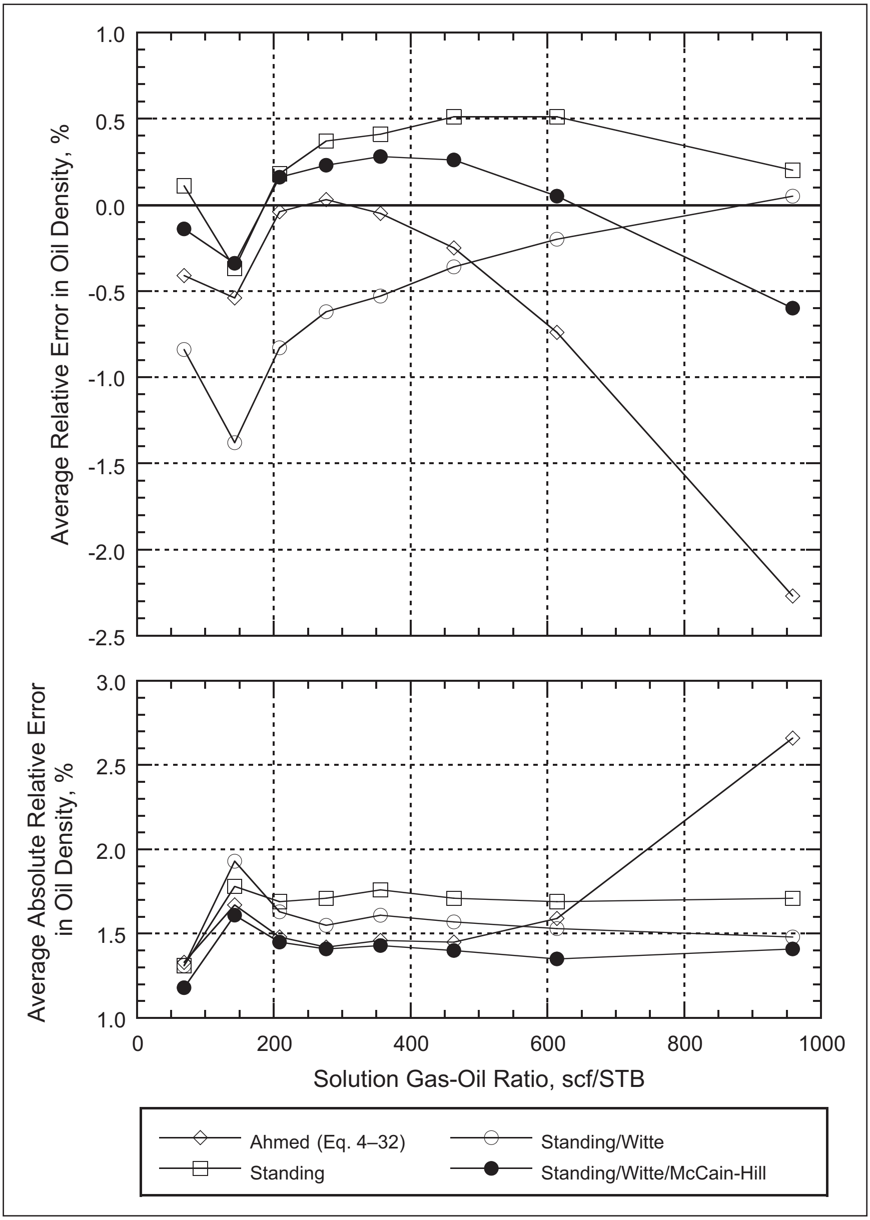

Fig. 3–22. Average relative errors and average absolute relative errors of subsets of the data described in table 3–16 sorted on solution gas-oil ratio show the quality of selected correlation equations for estimating reservoir oil densities for pressures equal to and less than bubblepoint pressure across the full range of this independent variable.

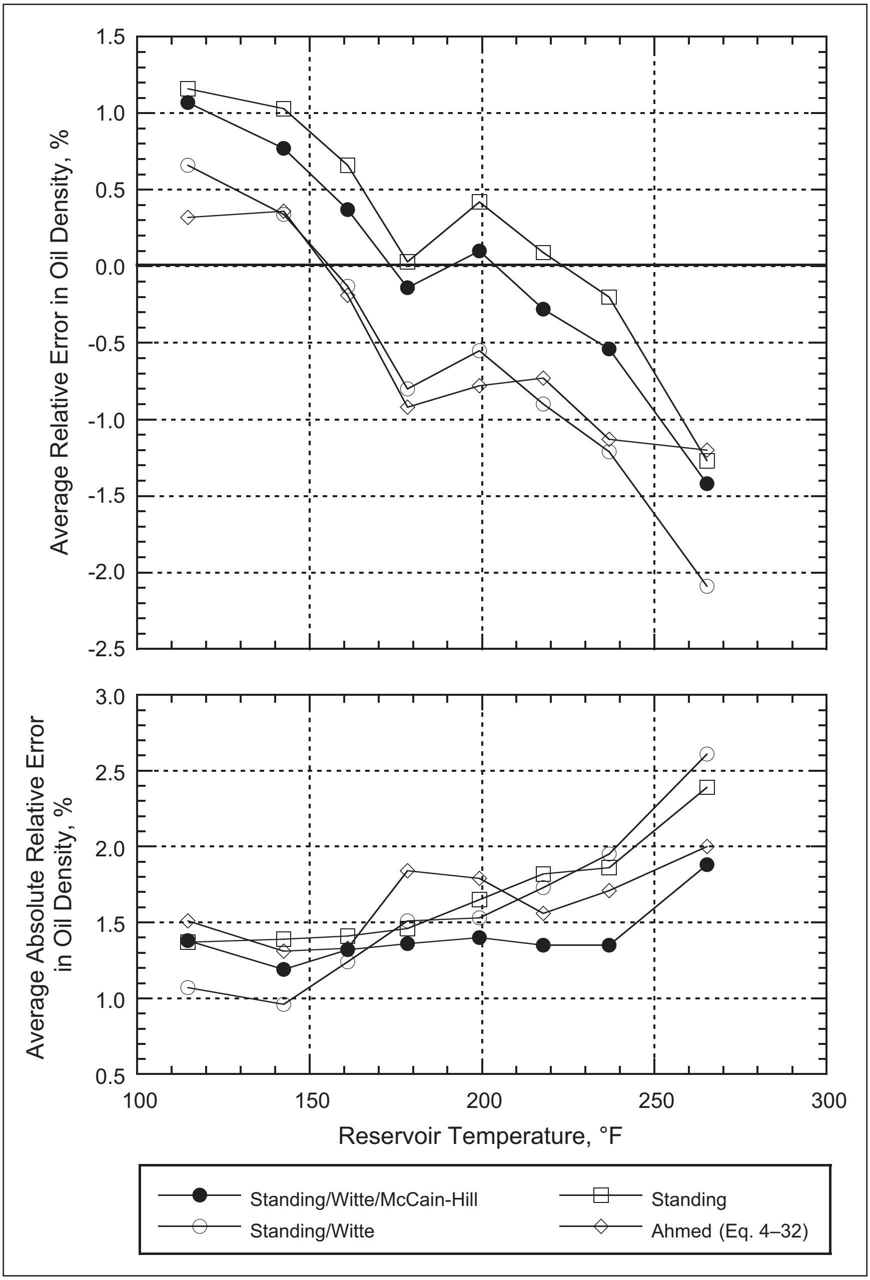

Fig. 3–23. Average relative errors and average absolute relative errors of subsets of the data described in table 3–16 sorted on reservoir temperature show the quality of selected correlation equations for estimating reservoir oil densities for pressures equal to and less than bubblepoint pressure across the full range of this independent variable.

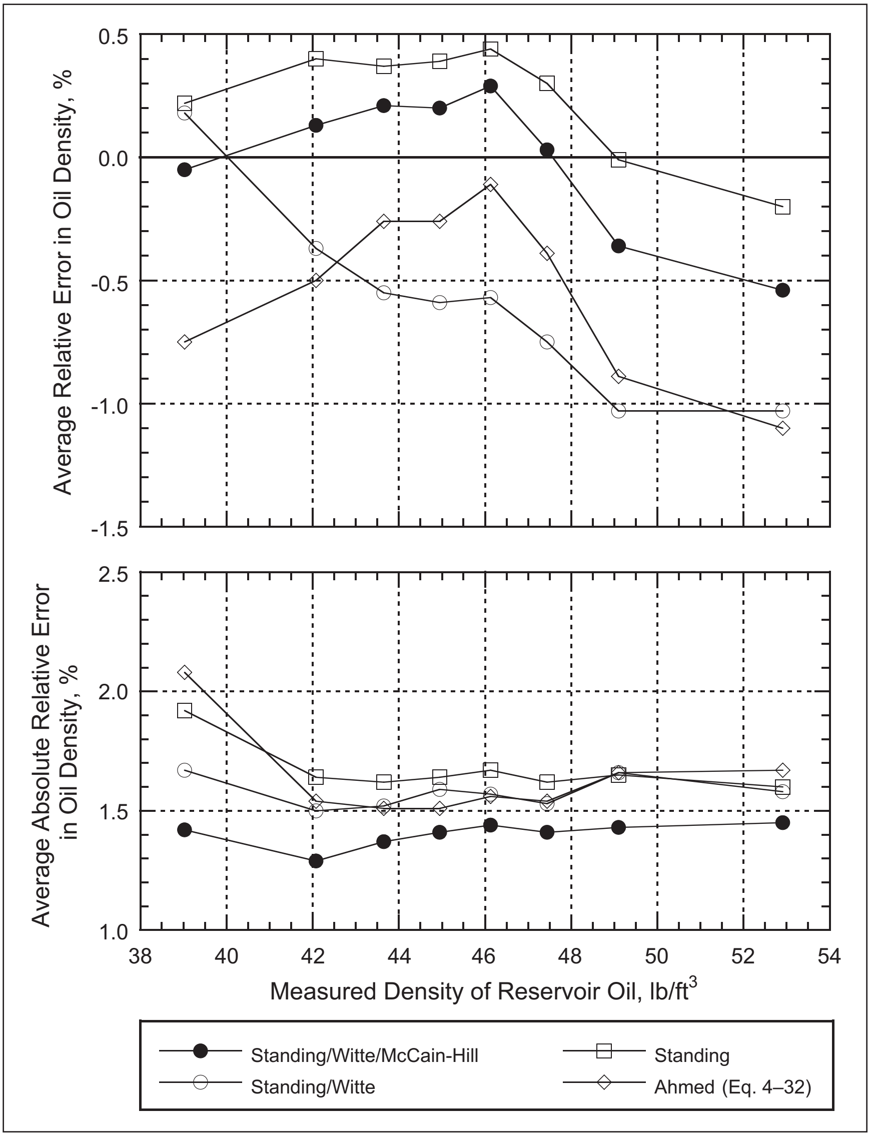

Fig. 3–24. Average relative errors and average absolute relative errors of subsets of the data described in table 3–16 sorted on measured reservoir oil density show the quality of selected correlation equations for estimating reservoir oil densities for pressures equal to and less than bubblepoint pressure across the full range of this dependent variable.

Recommended correlation equations for estimation of reservoir oil densities at reservoir pressures less than bubblepoint pressure

The recommended equations, equations (3.19a) and (3.19b), are listed below.[36]

$$\rho_{oR} = \rho_{po} + \Delta\rho_p - \Delta\rho_T \tag{3.19a}$$

$$\rho_{po} = \frac{R_s\gamma_g + 4600\gamma_{STO}}{73.71 + R_s\gamma_g/\rho_a} \tag{3.19b}$$

Equation (3.18c) may be used to estimate values of the apparent liquid density, ρ_a. The values of stock-tank oil gravity, separator gas specific gravity, and weighted-average total surface gas specific gravity determined for use at bubblepoint pressure can be assumed to remain constant as reservoir pressure declines. This is not precisely true but gives adequate results with the recommended equations.

The trial-and-error solution between equations (3.19b) and (3.18c) may be solved with successive substitutions using equation (3.19c) to determine a first trial value of ρ_{po}.

$$\rho_{po} = 52.8 - 0.01R_s \tag{3.19c}$$

Then density can be adjusted from atmospheric pressure to pressure, *p*, of interest.

$$\begin{aligned}\Delta\rho_p = &\left[0.167 + 16.181\left(10^{-0.0425\rho_{po}}\right)\right]\left(\frac{p}{1000}\right)\\ &-0.01\left[0.299 + 263\left(10^{-0.0603\rho_{po}}\right)\right]\left(\frac{p}{1000}\right)^2\end{aligned} \tag{3.19d}$$

$$\rho_{bs} = \rho_{po} + \Delta\rho_p \tag{3.19e}$$

Next adjust from 60°F to temperature, *T*, of interest.

$$\begin{aligned}\Delta\rho_T = &\left(0.00302 + 1.505\rho_{bs}^{-0.951}\right)(T-60)^{0.938}\\ &-\left[0.0216 - 0.0233\left(10^{-0.0161\rho_{bs}}\right)\right](T-60)^{0.475}\end{aligned} \tag{3.19f}$$

$$\rho_{oR} = \rho_{bs} - \Delta\rho_T \tag{3.19g}$$

Values of solution gas-oil ratios, R_s, at reservoir temperatures and selected reservoir pressures may be obtained using equations (3.8a)–(3.8f).

The selected correlation equations are not very much more accurate than the original Standing equations.[37] The values of reservoir oil densities at pressures less than the bubblepoint pressure calculated with the recommended procedure are approximately as precise as the underlying data.

Evaluation of the recommended equation for reservoir oil densities at reservoir pressures greater than bubblepoint pressure

Equation (3.20) can be used to estimate reservoir oil densities at pressures above the bubblepoint pressure of the reservoir oil.

$$\rho_{oR} = \rho_{oRb} \text{ EXP}\left[c_{ofb}\left(p - p_b\right)\right] \tag{3.20}$$

This equation is not a correlation; it follows directly from the definition of the coefficient of isothermal compressibility of oil. However, correlations are required for c_{ofb}, p_b, and ρ_{oRb}. The recommended correlation equations for these variables, equations (3.13), (3.7), and (3.18a)–(3.18h), were used in an evaluation of equation (3.20). The use of equations (3.18a)–(3.18h) to estimate values of ρ_{oRb} ensures that the results are continuous at the bubblepoint pressure.

Table 3–18 describes the statistics of the data set used to evaluate this equation. There were 399 reservoir fluid studies, each with a separator test and several lines of constant composition expansion data at pressures equal to and above the bubblepoint pressure. The "laboratory data" for use in evaluating equation (3.20) were determined by first calculating the reservoir oil density at bubblepoint pressure for each reservoir fluid study using data from the separator test with a rearrangement of equation (3.21a). Then the reservoir oil densities for pressures above bubblepoint pressure for each reservoir fluid study were calculated by dividing the reservoir oil density at bubblepoint pressure by the relative oil volumes from the constant composition expansion at the reported pressures.

Table 3–18. Statistics of data set used to evaluate reservoir oil density and oil formation volume factor correlation equations at reservoir pressures greater than bubblepoint pressure (3,061 lines of data from 399 reservoir fluid studies with worldwide origins)

Laboratory measurement	Minimum	Median	Mean	Maximum
Bubblepoint pressure, p_b, psia	120.7	2,084.7	2,346.3	7,854.0
Pressure, p, psia	127.7	3,514.7	3,561.6	14,915.0
Temperature, T, °F	70.0	187.0	192.7	320.0
Stock-tank oil gravity, API, °API	11.6	37.6	36.4	57.7
Separator gas specific gravity, γ_{gSP}	0.561	0.784	0.804	1.798
Solution gas-oil ratio at p_b, R_{sb}, scf/STB	12.0	594.0	636.9	1,808.0
Oil formation volume factor at p_b, B_{ob}, res bbl/STB	1.007	1.346	1.372	2.044
Reservoir oil density, ρ_{oR}, lb/cu ft	32.16	45.32	45.71	60.62

The average relative error of all 3,061 lines of data was –1.44%, and the average absolute relative error was 2.10% for equation (3.20). The data set was sliced by stock-tank oil gravities, solution gas-oil ratios at bubblepoint pressures, reservoir temperatures, and reservoir pressures, each into eight segments of approximately equal size. The resulting average relative errors and average absolute relative errors for all slices were amazingly consistent, with average relative errors varying from –2.53% to –0.15% and average absolute relative errors varying from 0.98% to 2.97%. No trends of the errors with these variables were observed across the full range of the data set.

Oil Formation Volume Factors

There is a change in slope at the bubblepoint pressure of a plot of oil formation volume factors versus reservoir pressures. This slope is positive for pressures less than bubblepoint pressure and negative for pressures greater than bubblepoint pressure. An equation based on a material balance of the mass of oil leaving the reservoir with the mass of total surface fluids will be used for estimating values of this property at pressures equal to and less than bubblepoint pressure. An equation based on the definition of the coefficient of isothermal compressibility, c_{ofb}, will be used for calculating values of this property at pressures greater than bubblepoint pressure. The selection of these equations ensures that the relationship with pressure is continuous, i.e., the values of oil formation volume factor coincide at the bubblepoint pressure.

Oil formation volume factors at reservoir pressure equal to bubblepoint pressure

McCain derived an equation for calculating oil formation volume factors at bubblepoint pressures.[38] This equation, equation (3.21a), is recommended for estimating oil formation volume factors at bubblepoint pressure.

$$B_{ob} = \frac{\rho_{STO} + 0.01357 R_{sb} \gamma_g}{\rho_{oRb}} \tag{3.21a}$$

This is not a correlation; rather it is the result of a material balance equating the mass of oil leaving the reservoir to the mass of the stock-tank oil plus the mass of the total surface gas. The advantage of its use is that values of oil formation volume factors, oil densities, and solution gas-oil ratios will be consistent with each other.

In application, correlation equations must be used to determine values of the bubblepoint pressure, p_b, and values of the density of the reservoir oil at that bubblepoint pressure, ρ_{oRb}. Equations (3.7) and (3.18a)–(3.18h) are recommended.

Note that the value of the gas specific gravity in equation (3.21a) should be the weighted-average specific gravity of the total surface gas. Thus, either equation (3.4) or equation (3.6) is usually necessary with equations (3.5) and (3.2) used to obtain the stock-tank gas properties.

Evaluation of proposed correlation equations for oil formation volume factors at bubblepoint pressure

The 718 reservoir fluid studies in the data set described in table 3–9 each had a separator test from which a value of oil formation volume factor at the bubblepoint was obtained. These values were used to evaluate published correlation equations of oil formation volume factors at bubblepoint pressures. The evaluation is given in table 3–19.

Table 3–19. Evaluation of published oil formation volume factor correlation equations for reservoir pressures equal to bubblepoint pressures reveals the most reliable correlations.

	Predicted oil formation volume factors	
Correlation	**ARE, %**	**AARE, %**
Equation (3.21a), p_b from data, ρ_{oRb} from equations (3.18a)–(3.18h)	–0.24	1.49
Al-Marhoun (1992)*	0.47	1.53
Equation (3.21a), p_b from equation (3.7), ρ_{oRb} from equations (3.18a)–(3.18h)	–0.34	1.57
Al-Shammasi (1999)[a]*	0.16	1.60
Kartoatmodjo-Schmidt (1994)	0.17	1.67
Al-Marhoun (1988)[b]*	–0.14	1.82
Standing (1977)†	–0.05	1.84
Sutton (2008)*, p_b from equation (3.7)	–0.05	1.90
Ahmed (1989), p_b from equation (3.7)	0.60	2.03
Omar-Todd (1993)[c]*†	0.22	2.44
Glaso (1980)*	–1.31	2.51
Casey-Cronquist (1992)†, p_b from data	1.52	2.55
Levitan-Murtha (1999)[d]	0.94	2.56
Almehaideb (1997)[e]†	0.45	3.33
Khairy et al. (1998)[f]*†‡	1.94	4.17
Vazquez-Beggs (1980), p_b from data	–1.39	4.36
Macary-El-Batanony (1993)[g]*†	9.02	9.04
Hanafy et al. (SPE 37439, 1997)†	9.47	9.52
Carlson (1997), p_b from data	–8.47	9.98
Elsharkawy-Alikhan (1977)[h]*†	37.43	37.43

[a] Al-Shammasi, A.A. "Bubble Point Pressure and Oil Formation Volume Factor Correlations." Paper SPE 53185. Presented at the SPE Middle East Oil Show, Bahrain (1999).
[b] Al-Marhoun, M.A. "PVT Correlations for Middle East Crude Oils." *Journal of Petroleum Technology*. Vol. 40 (May 1988) 650–666.
[c] Omar, M.I. and Todd, A.C. "Development of New Modified Black Oil Correlations for Malaysian Crudes." Paper SPE 25338. Presented at the SPE Asia Pacific Oil and Gas Conference, Singapore (1993).
[d] Levitan, L.L. and Murtha, M. "New Correlations Estimate Pb, FVF." *Oil & Gas Journal*. Vol. 97 (March 8, 1999) 70–76.
[e] Almehaideb, R.A. "Improved PVT Correlations for UAE Crude Oils." Paper SPE 37691. Presented at the SPE Middle East Oil Conference and Exhibition, Bahrain (1997).
[f] Khairy, M., El-Tayeb, S., and Hamdallah, M. "PVT Correlations Developed for Egyptian Crudes." *Oil & Gas Journal*. Vol. 98, No. 18 (May 4, 1998) 114–116.
[g] Macary, S.M. and El-Batanony, M.H. "Derivation of PVT Correlations for Gulf of Suez Crude Oils." *Sekiyu Gakkaishi*. Japan. Vol. 36, No. 6 (1993).
[h] Elsharkawy, A.M. and Alikhan, A.A. "Correlations for Predicting Solution Gas/Oil Ratio, Oil Formation Volume Factor, and Undersaturated Oil Compressibility." *Journal of Petroleum Science and Engineering*. Vol. 17 (1997) 291–302.
* Author restricted correlation to B_{ob} only.
† Author restricted the correlation to a specific geographical area.
‡ Modified second term in authors' equation 2 to 0.000705341.

Oil formation volume factors at reservoir pressures less than bubblepoint pressure

Equation (3.21a) was derived for use in calculating oil formation volume factors when the reservoir pressure is equal to the bubblepoint pressure of the reservoir oil. The question is whether this equation can be used at pressures less than bubblepoint pressure with reservoir oil densities, solution gas-oil ratios, and oil formation volume factors at pressures below bubblepoint pressure substituted for the values at bubblepoint pressure.

The values of stock-tank oil density and weighted average surface gas specific gravity determined for use at bubblepoint pressure may be assumed constant for the lower pressures.

Evaluation of proposed correlation equations for oil formation volume factors at reservoir pressures less than bubblepoint pressure

The data set described in table 3–9 was used to evaluate published correlation equations of oil formation volume factors at bubblepoint pressures and below. Each of the 718 reservoir fluid studies had a separator test and several lines of differential liberation data for pressures below bubblepoint pressure. Values of oil formation volume factors for use in evaluating published correlation equations were calculated with a combination of these differential liberation and separator test data using the procedure of Amyx et al.[39]

None of these data were used in preparing any of the published oil formation volume factor correlation equations.

Table 3–20 shows that this equation may be used with good results for reservoir pressures below the bubblepoint pressure.

Table 3–20. Evaluation of published oil formation volume factor correlation equations for reservoir pressures equal to and less than bubblepoint pressures reveals the most reliable correlations.

	Predicted oil formation volume factors	
Correlation	**ARE, %**	**AARE, %**
Equation (3.21b), ρ_{oR} from equations (3.19a)–(3.19g), R_s from equations (3.8a)–(3.8f), p_b from data	0.27	1.53
Equation (3.21b), ρ_{oR} from equations (3.19a)–(3.19g), R_s from equations (3.8a)–(3.8f), p_b from equation (3.7)	0.16	2.25
Casey-Cronquist (1992)*, p_b from data	1.31	3.09
Kartoatmodjo-Schmidt (1994)	–0.79	3.32
Standing (1977)*	–1.89	3.50
Casey-Cronquist (1992)*, p_b from equation (3.7)	1.38	3.52
Levitan-Murtha (1999)[a]	–0.97	3.72
Vazquez-Beggs (1980)	–2.19	4.23
Ahmed (1989), p_b from equation (3.7)	1.22	4.43
Glaso (1980) †	–4.06	5.12
Hanafy et al. (SPE 37439, 1997)*	4.48	7.23
Carlson (1997), p_b from equation (3.7)	–8.51	9.37

[a] Levitan, L.L. and Murtha, M. "New Correlations Estimate P_b, FVF." *Oil & Gas Journal.* Vol. 97 (March 8, 1999) 70–76.

* Author restricted correlation to a specific geographical area. † Author indicated correlation was for use at bubblepoint pressure only.

The data described in table 3–9 were sorted and sliced into 10 subsets with more than 500 data points in each subset. Values of ARE and AARE were calculated for subsets sorted by reservoir temperatures, stock-tank oil gravities, reservoir pressures, and measured oil formation volume factors. The results may be seen in figures 3–25 through 3–28.

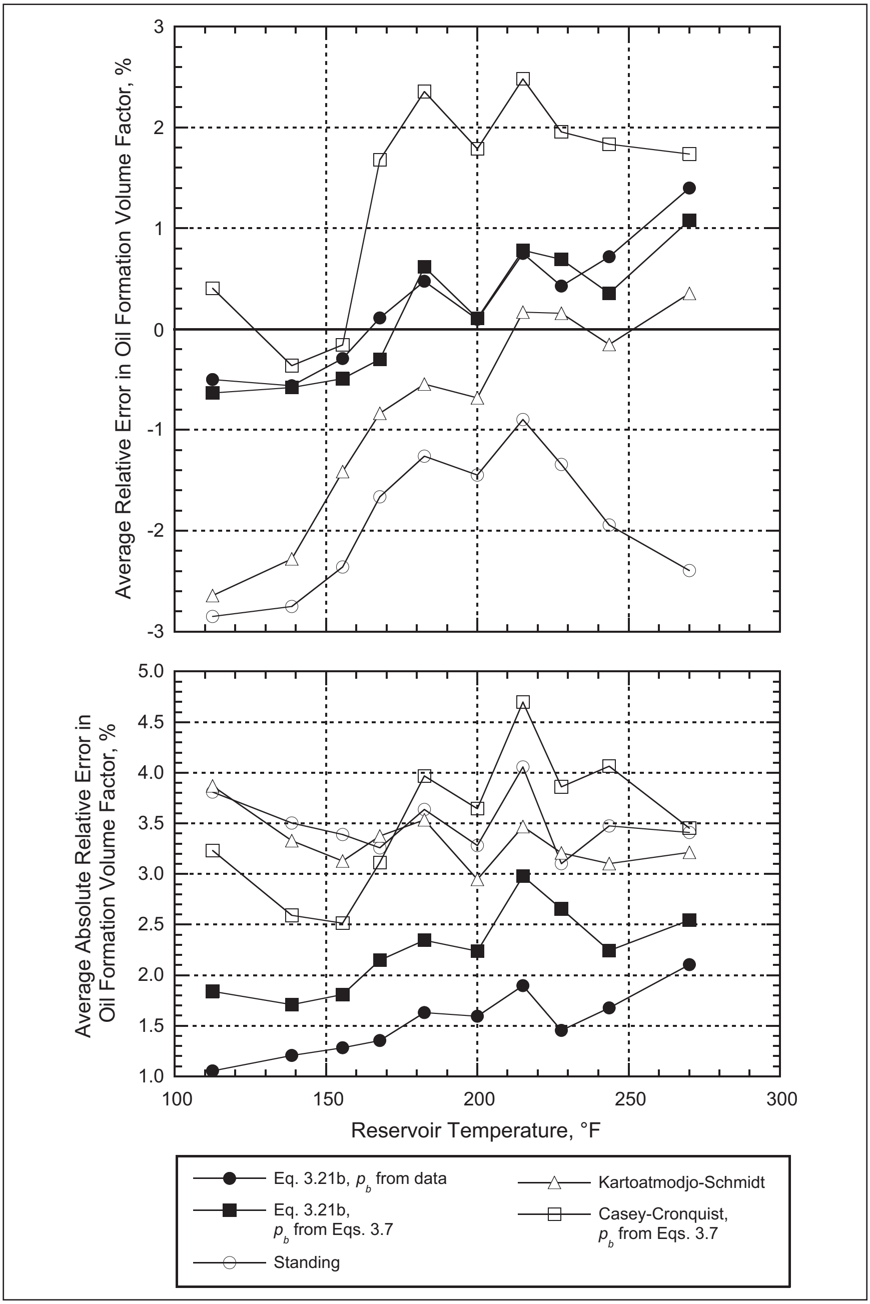

Fig. 3–25. Average relative errors and average absolute relative errors of subsets of the data described in table 3–9 sorted on reservoir temperature show the quality of selected correlation equations for estimating oil formation volume factors for pressures equal to and less than bubblepoint pressure across the full range of this independent variable.

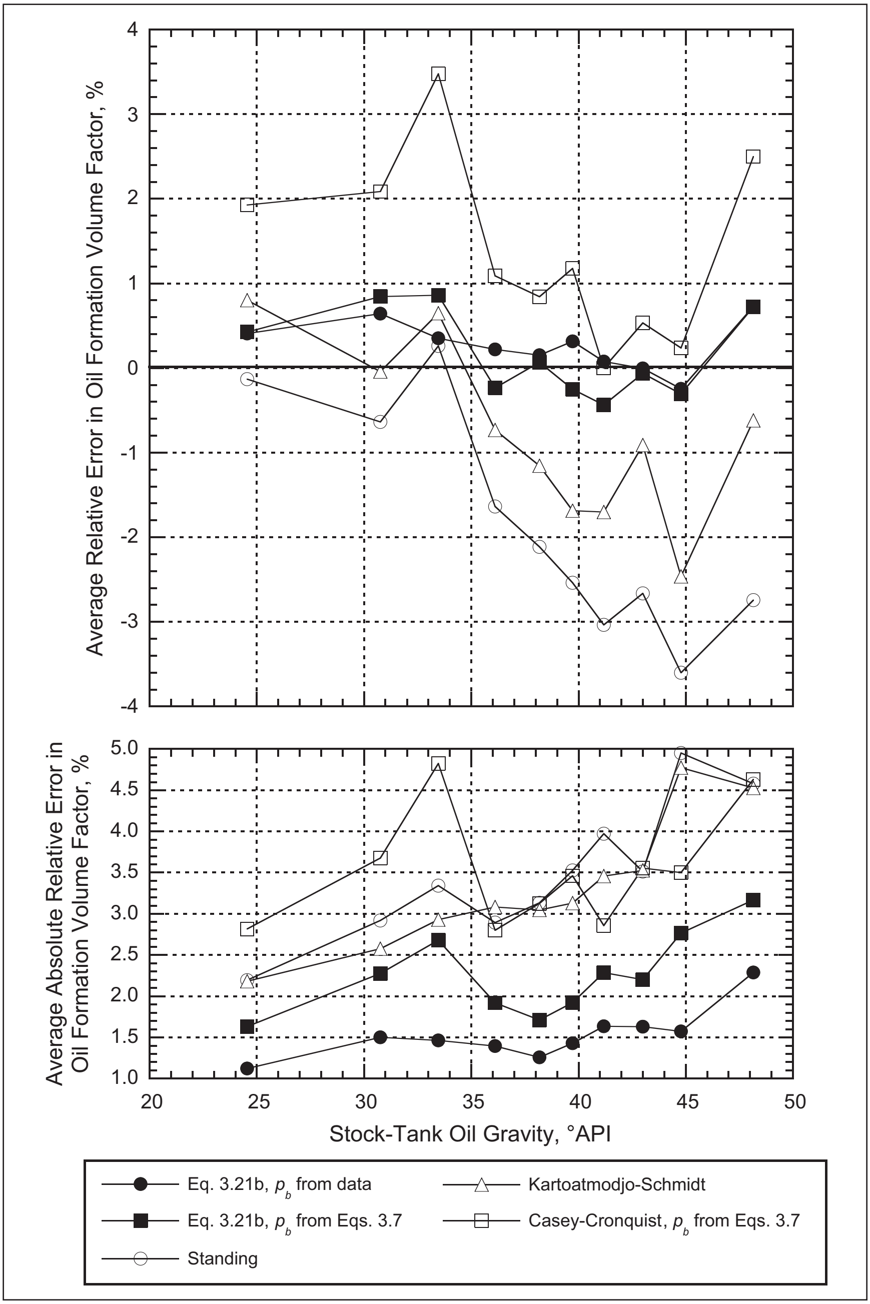

Fig. 3–26. Average relative errors and average absolute relative errors of subsets of the data described in table 3–9 sorted on stock-tank oil gravity show the quality of selected correlation equations for estimating oil formation volume factors for pressures equal to and less than bubblepoint pressure across the full range of this independent variable.

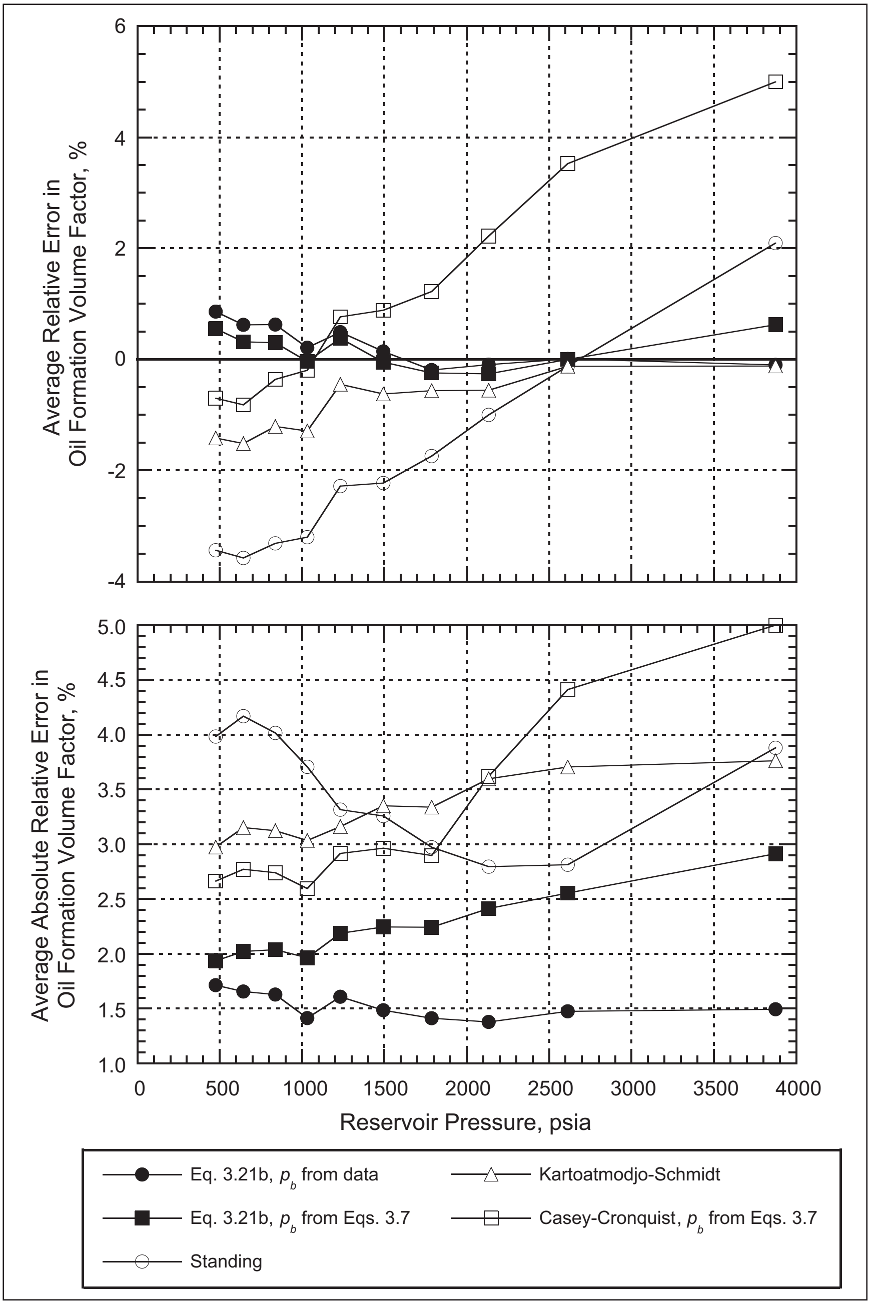

Fig. 3–27. Average relative errors and average absolute relative errors of subsets of the data described in table 3–9 sorted on reservoir pressure show the quality of selected correlation equations for estimating oil formation volume factors for pressures equal to and less than bubblepoint pressure across the full range of this independent variable.

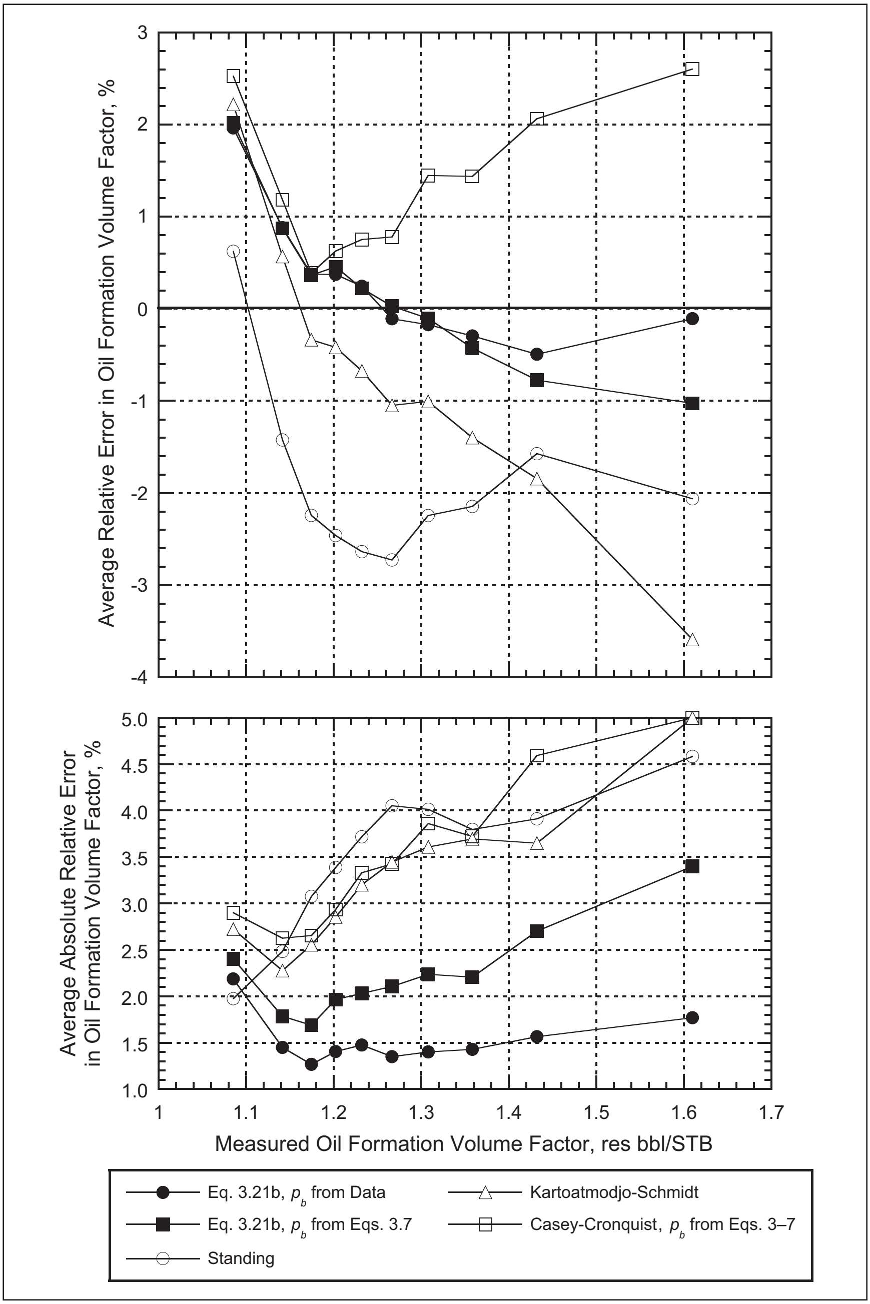

Fig. 3–28. Average relative errors and average absolute relative errors of subsets of the data described in table 3–9 sorted on measured oil formation volume factor show the quality of selected correlation equations for estimating oil formation volume factors for pressures equal to and less than bubblepoint pressure across the full range of this dependent variable.

Recommended equation for estimation of oil formation volume factors at pressures less than bubblepoint pressure

The recommended equation is given in equation (3.21b).

$$B_o = \frac{\rho_{STO} + 0.01357 R_s \gamma_g}{\rho_{oR}} \tag{3.21b}$$

Solution gas-oil ratios, R_s, and reservoir oil densities at pressures below the bubblepoint pressure, ρ_{oR}, are obtained using equations (3.8a)–(3.8f) and (3.19a)–(3.19g). Equations (3.8a)–(3.8f) require a value of bubblepoint pressure; equation (3.7) is recommended. The surface gas specific gravity should be the weighted average of the specific gravities of the separator and stock-tank gases. Thus, either equation (3.4) or (3.6) is usually necessary with stock-tank properties obtained with equations (3.2) and (3.5).

The most important factor affecting values of solution gas-oil ratios, oil formation volume factors, and reservoir oil densities is gas leaving solution from the oil in the reservoir as reservoir pressure declines below the original bubblepoint pressure. Thus, these three properties are interrelated for reservoir pressures below original bubblepoint pressure. The use of equations (3.21a) and (3.21b) preserve these relationships. These equations are not correlation equations but are the result of a rigorous material balance derivation.

Oil formation volume factors at reservoir pressures greater than bubblepoint pressure

The equation normally used for estimating values of oil formation volume factors at reservoir pressures greater than the bubblepoint pressure of the reservoir oil is given in equation (3.22).

$$B_o = B_{ob}\, EXP\left[c_{ofb}\left(p_b - p\right)\right] \tag{3.22}$$

This equation follows directly from the definition of coefficients of isothermal compressibility as discussed previously; i.e., it is not a correlation.

Values of B_{ob}, c_{ofb}, and p_b can be obtained using the correlation equations previously recommended, equations (3.21a), (3.13), and (3.7).

In addition, Ahmed presented two equations and Al-Marhoun one equation that do not require the use of coefficients of isothermal compressibility.[40]

Evaluation of proposed correlation equations for oil formation volume factors at reservoir pressures greater than bubblepoint pressure

An evaluation of these four techniques is given in table 3–21.

Table 3–21. Evaluation of published oil formation volume factor correlation equations for reservoir pressures above bubblepoint pressure

	Predicted oil formation volume factors	
Correlation	**ARE, %**	**AARE, %**
Equation (3.22), B_{ob} from equation (3.21a), c_{ofb} from equation (3.13)*	1.58	2.23
Al-Marhoun (1992)*	1.49	2.31
Ahmed equations 4–65, 4–69 (1989)*	1.49	2.70
Ahmed equations 4–65, 4–70 (1989)*	0.75	2.90

* All correlations evaluated with p_b calculated with equation (3.7).

The statistics of the data set used in the evaluation of these correlations are listed in table 3–18. Each of the 399 reservoir fluid studies had a separator test and several lines of constant composition expansion data at pressures above bubblepoint pressure. Values of oil formation volume factors at pressures above bubblepoint pressure for use in evaluating published correlations were calculated with these data using the procedure of Craft and Hawkins.[41]

None of these data were used in the preparation of any of these correlation equations.

The data of table 3–18 were sorted into subsets of approximately equal size according to reservoir temperatures, stock-tank oil gravities, reservoir pressures, and measured values of oil formation volume factors. The results of ARE and AARE calculations for these subsets are given in figures 3–29 through 3–32.

Also, Labedi offered a set of equations from which oil formation volumes factors at pressures above bubblepoint pressure can be estimated.[42] Unfortunately, some of the data required as input to his equations were not included in the data set used in this evaluation. Thus, an evaluation of his work for pressures above bubblepoint pressure was not possible.

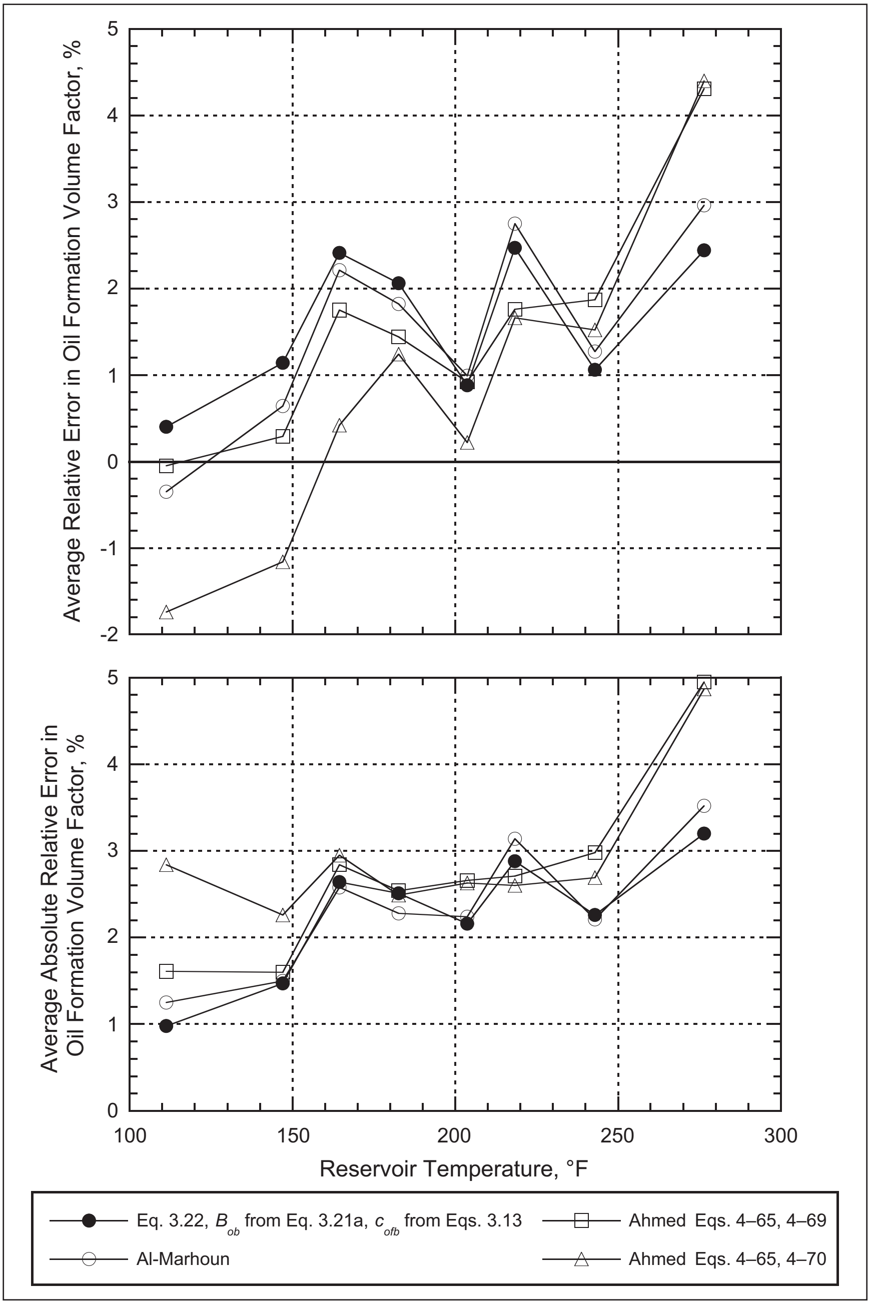

Fig. 3–29. Average relative errors and average absolute relative errors of subsets of the data described in table 3–18 sorted on reservoir temperature show the quality of selected equations for estimating oil formation volume factors for pressures greater than bubblepoint pressure across the full range of this independent variable.

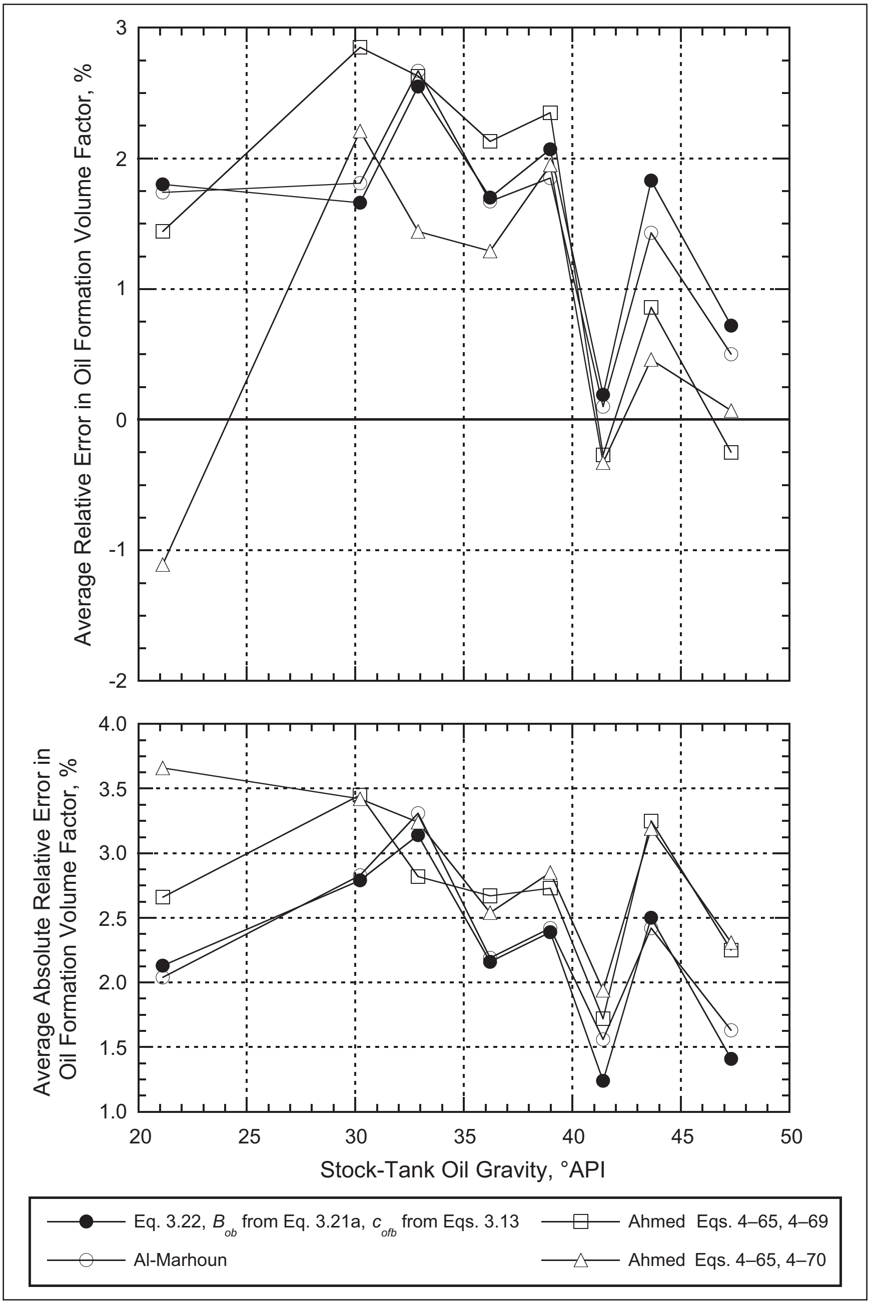

Fig. 3–30. Average relative errors and average absolute relative errors of subsets of the data described in table 3–18 sorted on stock-tank oil gravity show the quality of selected equations for estimating oil formation volume factors for pressures greater than bubblepoint pressure across the full range of this independent variable.

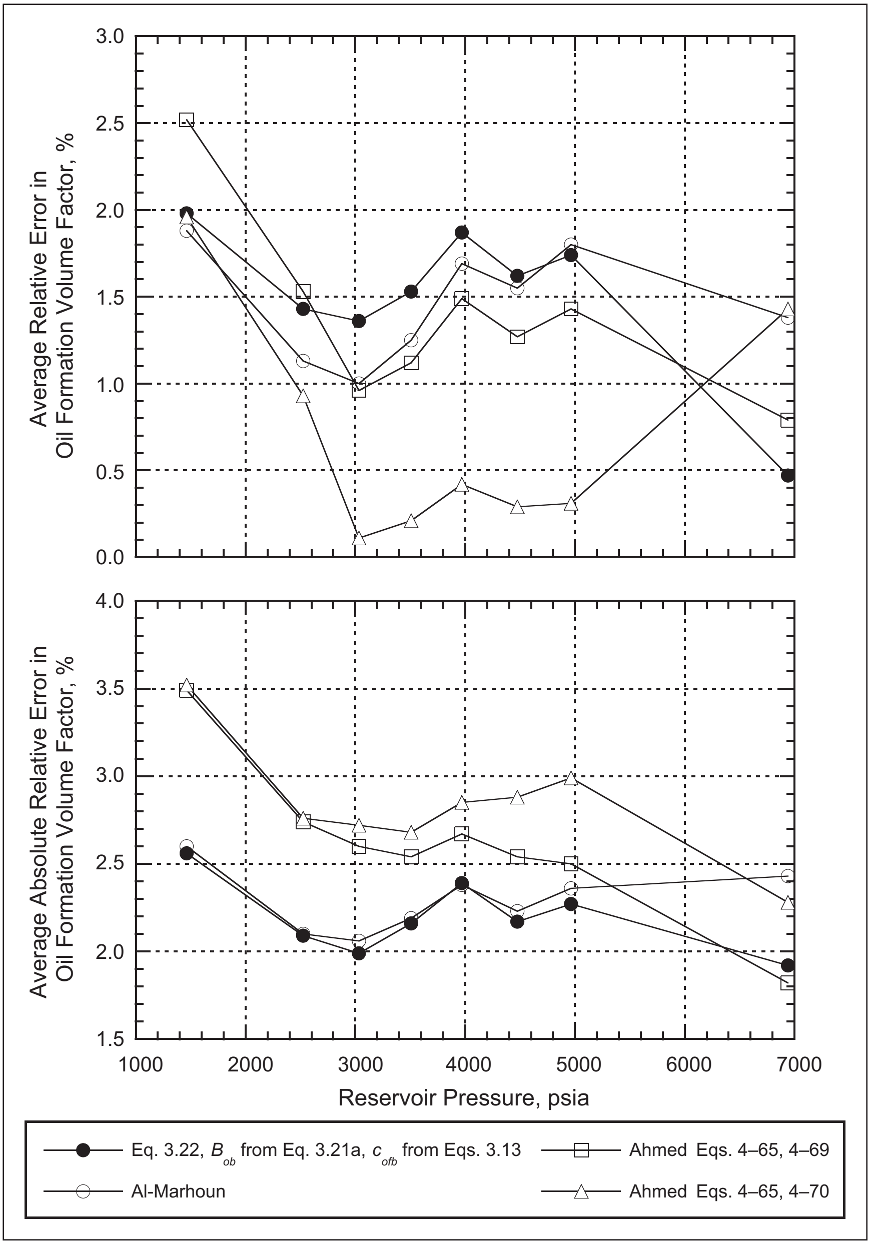

Fig. 3–31. Average relative errors and average absolute relative errors of subsets of the data described in table 3–18 sorted on reservoir pressure show the quality of selected equations for estimating oil formation volume factors for pressures greater than bubblepoint pressure across the full range of this independent variable.

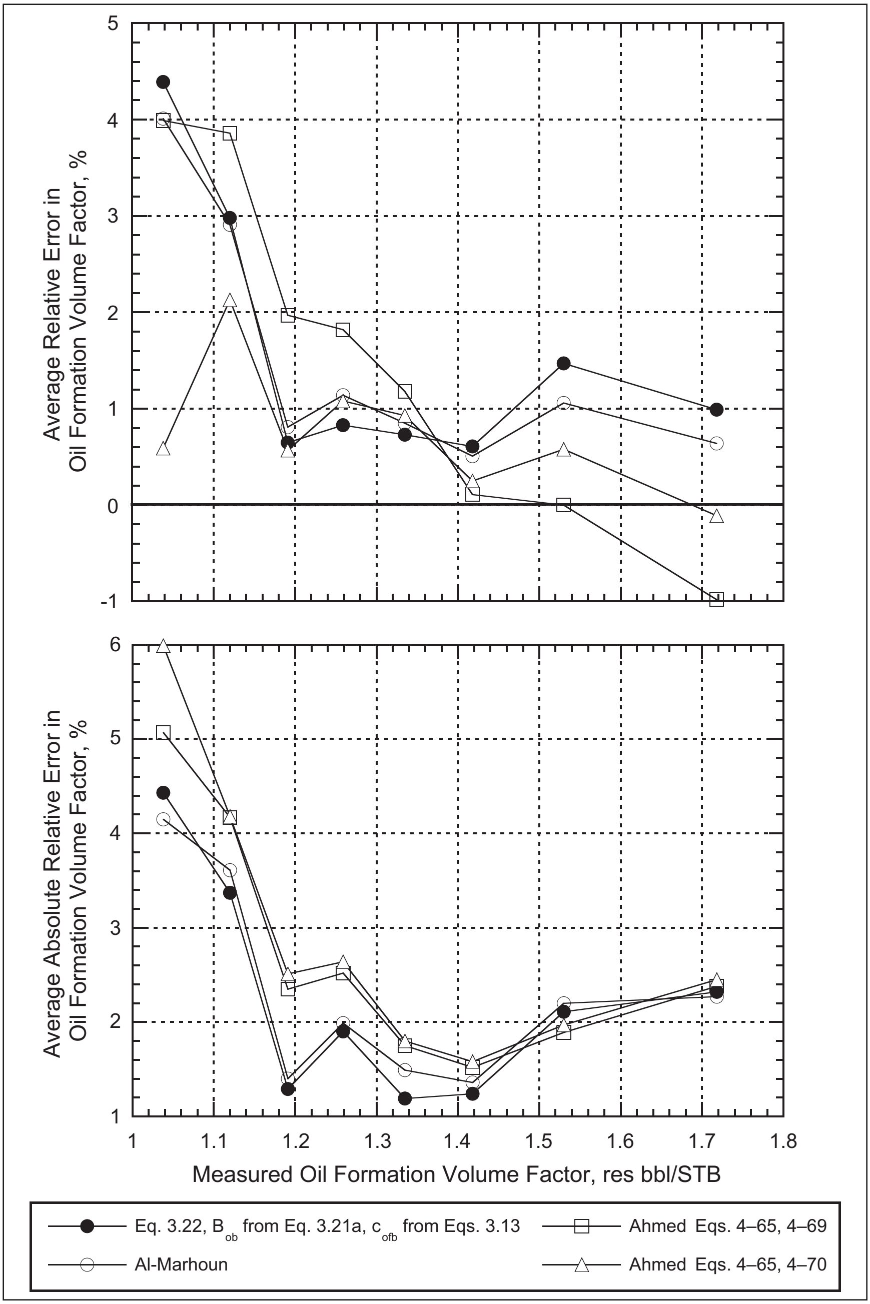

Fig. 3–32. Average relative errors and average absolute relative errors of subsets of the data described in table 3–18 sorted on measured oil formation volume factor show the quality of the selected equations for estimating oil formation volume factors for pressures greater than bubblepoint pressure across the full range of this dependent variable.

Recommended equation for estimation of oil formation volume factors at reservoir pressures greater than bubblepoint pressure

Although the ARE values reported in figures 3–29 through 3–32 have mixed results, equation (3.22) was consistently best on AARE. Further, use of equation (3.22) with B_{ob} from equation (3.21a) ensures that the values of oil formation volume factors are continuous as reservoir pressure decreases through the bubblepoint pressure.

The expansion of the oil in the reservoir as pressure decreases from initial pressure to bubblepoint pressure is the controlling factor for the changes in the values of reservoir oil densities, oil formation volume factors, and oil compressibilities. Thus, there are relationships among these properties for pressures greater than bubblepoint pressure. The use of equation (3.20) for oil density and equation (3.22) for oil formation volume factor, both tied to oil compressibility, equation (3.13), preserves these relationships. Note that equations (3.20) and (3.22) are not correlations; they follow directly from the definition of oil compressibility.

Viscosities of Reservoir Oils

There is a change in slope at the bubblepoint pressure of a plot of reservoir oil viscosity versus reservoir pressure. This slope is negative for pressures less than bubblepoint pressure and positive for pressures greater than bubblepoint pressure. Two sets of correlation equations will be used for estimating values of this property for pressures less than bubblepoint pressure and pressures greater than bubblepoint pressure. Care must be taken in the selection of these equations such that the relationship is continuous, i.e., the values of oil viscosity coincide at the bubblepoint pressure.

Evaluation of proposed correlation equations for reservoir oil viscosities at reservoir pressures equal to bubblepoint pressure

Statistics of the data set used to evaluate reservoir oil viscosity correlation equations for reservoir pressures equal to bubblepoint pressure are shown in table 3–22.

An evaluation of the published correlation equations for reservoir oil viscosities at reservoir pressures equal to the bubblepoint pressure of the reservoir oil is given in table 3–23.

Table 3–22. Statistics of data set used to evaluate reservoir oil viscosity correlation equations for reservoir pressures equal to bubblepoint pressures (338 lines of data from 338 reservoir fluid studies with worldwide origins)

Laboratory measurement	**Minimum**	**Median**	**Mean**	**Maximum**
Bubblepoint pressure, p_b, psia	120.7	1,839.7	1,993.9	5,311.7
Reservoir temperature, T_R, °F	107.0	205.0	208.3	320.0
Stock-tank oil gravity, API, °API	15.8	34.0	33.3	57.7
Separator gas specific gravity, γ_{gSP}	0.6000	0.8297	0.8584	1.3033
Solution gas-oil ratio at p_b, R_{sb}, scf/STB	12.0	430.0	511.1	1,716.0
Reservoir oil density, ρ_{oR}, lb/cu ft	34.27	45.1	45.46	57.32
Reservoir oil viscosity, μ_{ob}, cp	0.12	0.632	4.71	149.30

Table 3–23. Evaluation of published reservoir oil viscosity correlation equations at bubblepoint pressures reveals the most reliable correlation.

	Predicted saturated oil viscosity	
Correlation	**ARE, %**	**AARE, %**
Beggs-Robinson (1975)	–11.24	27.49
Dindoruk-Christman (2004)[a]*	–9.36	30.22
Hanafy et al. (SPE 37439, 1997)*	–4.63	30.91
Ng-Egbogah (1983)[b] with Beggs-Robinson (1975)	1.34	30.95
Almehaideb (1997)[c]*	–4.63	31.05
Petrosky-Farshad (1995)*	2.86	31.25
Elsharkawy-Gharbi (1999)[d]	–7.43	31.63
Abu-Khamsin-Al-Marhoun (1991)*	2.17	32.80
Al-Khafaji et al. (1987)[e]	–10.58	33.10
Bergman (1992)[f]	11.93	35.39
Kartoatmodjo-Schmidt (1994)	–10.51	36.39
Elsharkawy-Alikhan (1999)[g]*	18.27	38.69
Bergman-Sutton (2007)[h]	21.50	40.41
Standing (1977)	–14.63	46.65
Aziz et al. (1972)[i]	–47.78	50.50
Naseri et al. (2005)[j]*	–60.94	67.22
Khan et al. (1987)[k]*	45.51	69.02
Labedi (1992)[l]	110.4	142.80

[a] Dindoruk, B. and Christman, P.G. "PVT Properties and Viscosity Correlations for Gulf of Mexico Oils." *Society of Petroleum Engineers Reservoir Evaluation and Engineering*. Vol. 7, No. 6 (2004) 427–437. [b] Ng, J.T.H. and Egbogah, E.O. "An Improved Temperature-Viscosity Correlation for Crude Oil Systems." Paper 83-34-3. Presented at Petroleum Society of CIM 34th Annual Technical Meeting, Banff (1983). [c] Almehaideb, R.A. "Improved PVT Correlations for UAE Crude Oils." Paper SPE 37691. Presented at the SPE Middle East Oil Conference and Exhibition, Bahrain (1997). [d] Elsharkawy, A.M. and Gharbi, R.B.C. "Comparing Classical and Neural Regression Techniques in Modeling Crude Oil Viscosity." *Advances in Engineering Software and Workstations*. Vol. 32 (2001) 215–224. [e] Al-Khafaji, A.H., Abdul-Majeed, G.H., and Hassoon, S.F. "Viscosity Correlation for Dead, Live, and Undersaturated Crude Oils." *Journal of Petroleum and Gas Exploration Research*. Vol. 6, No. 2 (1987) 1–6. [f] Whitson, C.H. and Brule, M.R. *Phase Behavior*. Richardson, TX: SPE (2000) 35–37. [g] Elsharkawy, A.M. and Alikhan, A.A. "Models for Predicting the Viscosity of Middle East Crude Oil." *Fuel*. Vol. 78 (1999) 891–903. [h] Bergman, D.F. and Sutton, R.P. "An Update to Viscosity Correlations for Gas-Saturated Crude Oils." Paper SPE 110195. Presented at the SPE Annual Technical Conference and Exhibition, Anaheim (2007); and Whitson, C.H. and Brule, M.R., 2000. [i] Aziz, K., Govier, G.W., and Forarasi, M. "Pressure Drop in Wells Producing Oil and Gas." *Journal of Canadian Petroleum Technology*. Vol. 11 (1972) 38–42. [j] Naseri, A., Nikazar, M., and Mousavi Dehghani, S.A. "A Correlation Approach for Prediction of Crude Oil Viscosities." *Journal of Petroleum Science and Engineering* (2005) 163–174. [k] Khan, S.A., Al-Marhoun, M.A., Duffuaa, S.O., and Abu-Khamsin, S.A. "Viscosity for Saudi-Arabian Crude Oils." Paper SPE 15720. Presented at the Fifth Middle East Oil Show, Manama, Bahrain (1987). [l] Labedi, R. "Improved Correlations for Predicting the Viscosity of Light Crudes." *Journal of Petroleum Science and Engineering*. Vol. 8 (1992) 221–234. * Author restricted correlation to a specific geographical area.

Beggs and Robinson used approximately one-half of the data described in table 3–22 in preparation of their correlation equations.[43] Hanafy et al. used approximately one-half of the table 3–22 data in preparing their correlation equations.[44] The other researchers apparently did not use any of these data.

Equations (3.23a)–(3.23e) are recommended for use at reservoir pressures equal to bubblepoint pressure with R_{sb} substituted for R_s in equations (3.23d) and (3.23e).[45]

Evaluation of proposed correlation equations for reservoir oil viscosities at reservoir pressures equal to and less than bubblepoint pressure

Statistics of the data set used to evaluate reservoir oil viscosity correlation equations for reservoir pressures equal to and less than bubblepoint pressures are given in table 3–24.

Table 3–24. Statistics of data set used to evaluate reservoir oil viscosity correlation equations at reservoir pressures equal to and less than bubblepoint pressures (1,123 lines of data from 184 reservoir fluid studies with worldwide origins)

Laboratory measurement	Minimum	Median	Mean	Maximum
Bubblepoint pressure, p_b, psia	120.7	1,669.7	1,809.0	5,311.7
Pressure, p, psia	41.7	1,114.7	1,288.2	5,311.7
Reservoir temperature, T_R, °F	107.0	193.0	197.4	320.0
Stock-tank oil gravity, API, °API	15.8	35.0	34.3	57.7
Separator gas specific gravity, γ_{gSP}	0.6506	0.8579	0.8908	1.3033
Solution gas-oil ratio at p_b, R_{sb}, scf/STB	12.3	378.0	349.6	1,534.0
Reservoir oil density, ρ_{oR}, lb/cu ft	34.27	46.32	46.48	57.42
Reservoir oil viscosity, μ_{oR}, cp	0.167	0.728	4.2	157.6
Solution gas-oil ratio, R_s, scf/STB	4.0	287.0	456.8	1,534.0

An evaluation of the published correlation equations for reservoir oil viscosities at reservoir pressures equal to and less than bubblepoint pressures is shown in table 3–25.

Beggs and Robinson and Kartoatmodjo and Schmidt used essentially all of the data described in table 3–24 in preparation of their correlation equations.[46] It is not known if the other researchers had access to these data.

Table 3–25. Evaluation of published reservoir oil viscosity correlation equations for reservoir pressures equal to and less than bubblepoint pressures reveals the most reliable correlations.

Correlation	**Predicted saturated oil viscosity**	
	ARE, %	**AARE, %**
Beggs-Robinson (1975)	–5.63	27.60
Dindoruk-Christman (2004)[a]*	–9.18	29.90
Ng-Egbogah (1983)[b] with Beggs-Robinson (1975)	2.86	30.50
Petrosky-Farshad (1995)*	2.47	31.15
Elsharkawy-Gharbi (2001)[c]	–7.43	31.63
Almehaideb (1997)[d]*	7.39	34.36
Hanafy et al. (SPE 37439, 1997)*	10.06	34.78
Kartoatmodjo-Schmidt (1994)	–7.62	35.69
Al-Khafaji et al. (1987)[e]	–7.63	35.70
Bergman (1992)[f]	13.46	36.44
Abu-Khamsin-Al-Marhoun (1991)[g]*	18.99	40.04
Elsharkawy-Alikhan (1999)[h]*	21.46	40.45
Bergman-Sutton (2007)[i]	21.12	40.81
Standing (1977)*	–8.13	41.60
Aziz et al. (1972)[j]	–41.10	44.06
Khan et al. (1987)[k]*	15.41	45.04
Naseri et al. (2005)[l]*	–67.28	68.55
Labedi (1992)[m]	58.37	83.65

[a] Dindoruk, B. and Christman, P.G. "PVT Properties and Viscosity Correlations for Gulf of Mexico Oils." *Society of Petroleum Engineers Reservoir Evaluation and Engineering*. Vol. 7, No. 6 (2004) 427–437.

[b] Ng, J.T.H. and Egbogah, E.O. "An Improved Temperature-Viscosity Correlation for Crude Oil Systems." Paper 83-34-43. Presented at Petroleum Society of CIM 34th Annual Technical Meeting, Banff (1983).

[c] Elsharkawy, A.M. and Gharbi, R.B.C. "Comparing Classical and Neural Regression Techniques in Modeling Crude Oil Viscosity." *Advances in Engineering Software and Workstations*. Vol. 32 (2001) 215–224.

[d] Almehaideb, R.A. "Improved PVT Correlations for UAE Crude Oils." Paper SPE 37691. Presented at the Middle East Conference and Exhibition, Bahrain (1997).

[e] Al-Khafaji, A.H., Abdul-Majeed, G.H., and Hassoon, S.F. "Viscosity Correlation for Dead, Live and Undersaturated Crude Oils." *Journal of Petroleum and Gas Exploration Research*. Vol. 6, No. 2 (1987) 1–6.

[f] Whitson, C.H. and Brule, M.R. *Phase Behavior*. Richardson, TX: Society of Petroleum Engineers (2000) 35–37.

[g] Abu-Khamsin, S.A. and Al-Marhoun, M.A. "Development of a New Correlation for Bubblepoint Oil Viscosity." *Arabian Journal for Science & Engineering*. Vol. 16, No. 2A (1991) 99–106.

[h] Elsharkawy, A.M. and Alikhan, A.A. "Models for Predicting the Viscosity of Middle East Crude Oil." *Fuel*. Vol. 78 (1999) 891–903.

[i] Bergman, D.F. and Sutton, R.P. "An Update to Viscosity Correlations for Gas-Saturated Crude Oils." Paper SPE 110195. Presented at the SPE Annual Technical Conference and Exhibition, Anaheim (2007); and Whitson, C.H. and Brule, M.R., 2000.

[j] Aziz, K., Govier, G.W., and Forarasi, M. "Pressure Drop in Wells Producing Oil and Gas." *Journal of Canadian Petroleum Technology*. Vol. 11 (1972) 38–42.

[k] Khan, S.A., Al-Marhoun, M.A., Duffuaa, S.O., and Abu-Khamsin, S.A. "Viscosity for Saudi-Arabian Crude Oils." Paper SPE 15720. Presented at the Fifth Middle East Oil Show, Manama, Bahrain (1987).

[l] Naseri, A., Nikazar, M., and Mousavi Dehghani, S.A. "A Correlation Approach for Prediction of Crude Oil Viscosities." *Journal of Petroleum Science and Engineering*. (2005) 163–174.

[m] Labedi, R. "Improved Correlations for Predicting the Viscosity of Light Crudes." *Journal of Petroleum Science and Engineering*. Vol. 8 (1992) 221–234.

* Author restricted the correlation to a specific geographical area.

The data of table 3–24 were sorted and split into eight segments each with approximately 140 lines of data. These were done for stock-tank oil gravities, solution gas-oil ratios, reservoir temperatures, reservoir oil pressures, reservoir oil densities, and measured values of reservoir oil viscosities. The five best correlation equations listed in tables 3–23 and 3–25 were selected for evaluation. Values of ARE and AARE were determined for each of these segments. The results are given in figures 3–33 through 3–38.

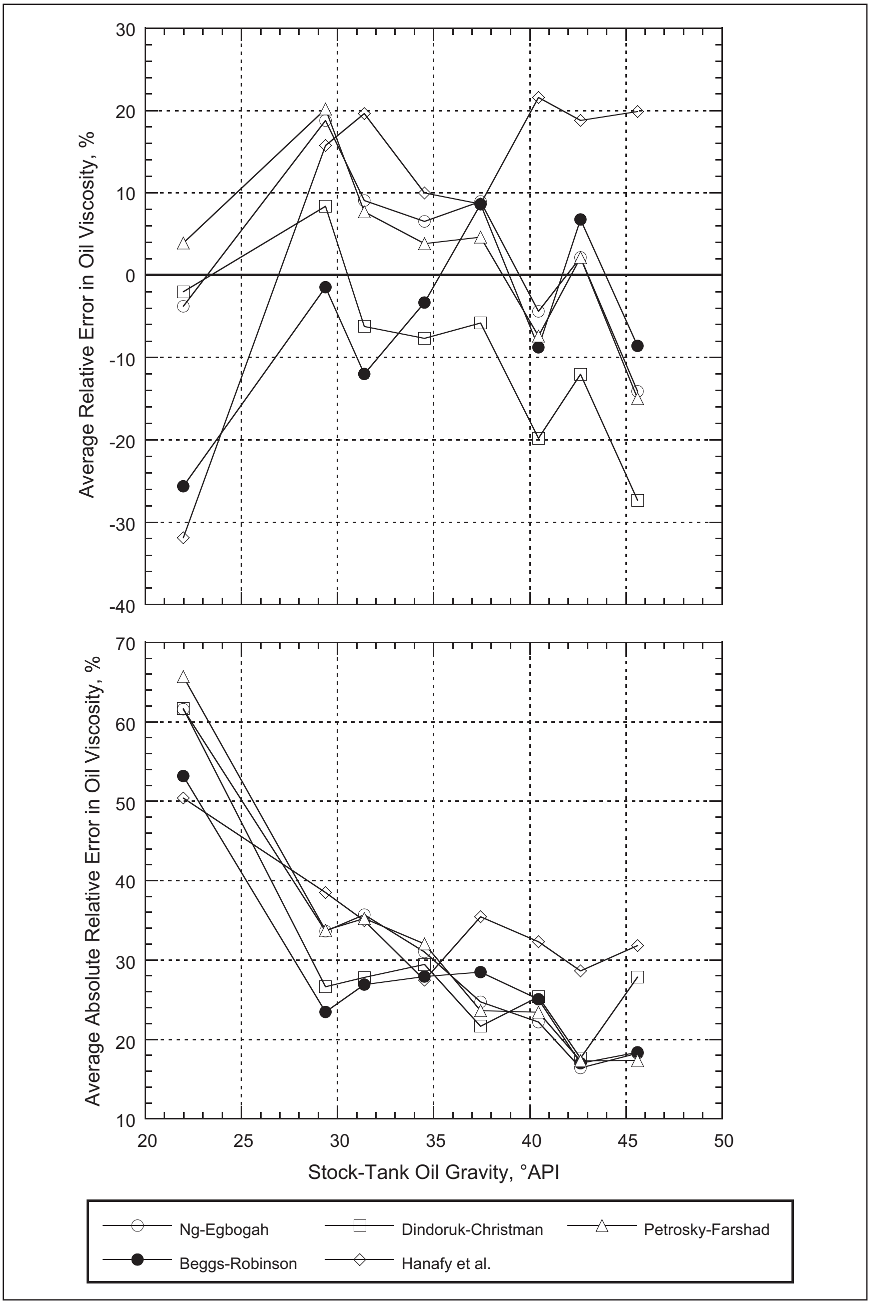

Fig. 3–33. Average relative errors and average absolute relative errors of subsets of the data described in table 3–24 sorted on stock-tank oil gravity show the quality of selected correlation equations for estimating reservoir oil viscosities for pressures equal to and less than bubblepoint pressure across the full range of this independent variable.

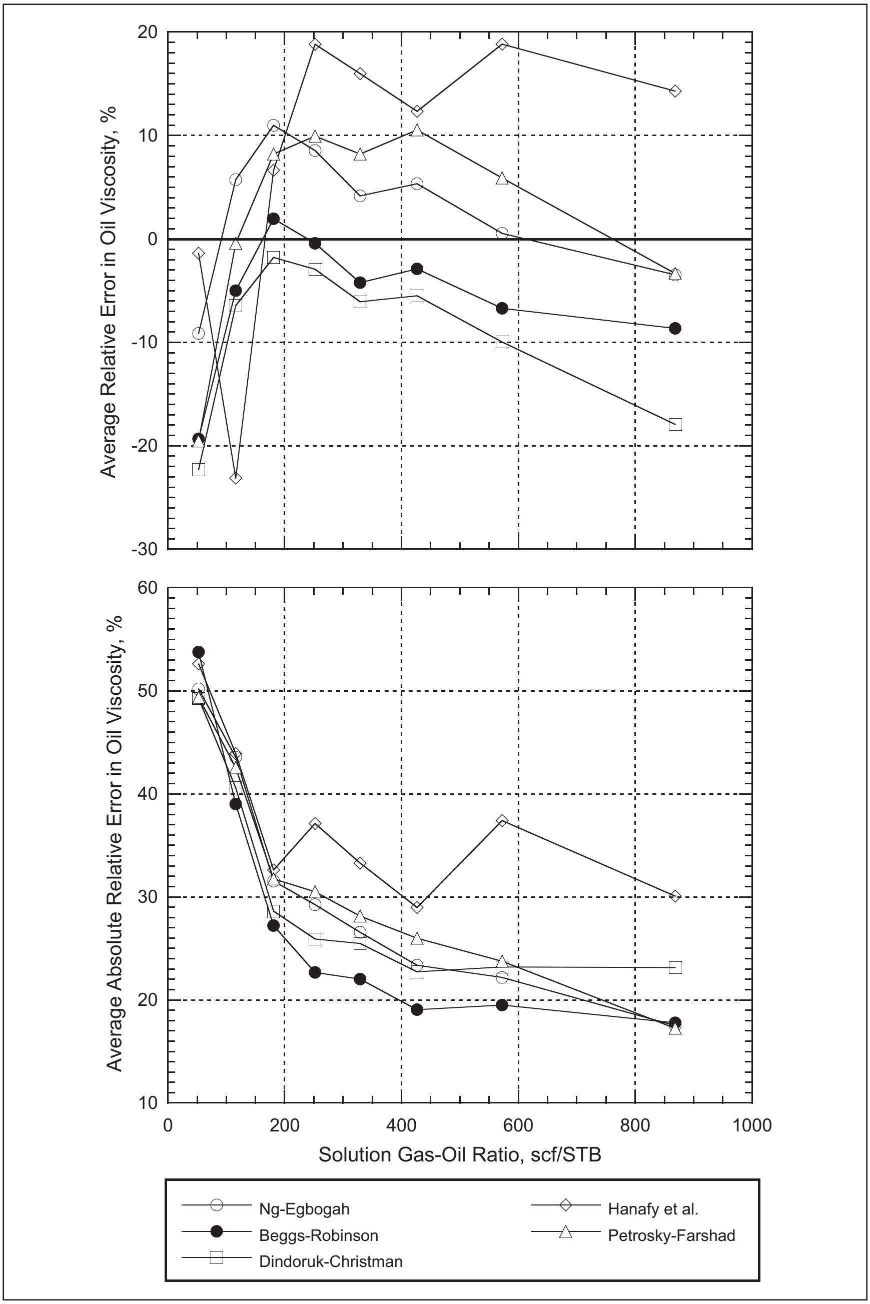

Fig. 3–34. Average relative errors and average absolute relative errors of subsets of the data described in table 3–24 sorted on solution gas-oil ratio show the quality of selected correlation equations for estimating reservoir oil viscosities for pressures equal to and less than bubblepoint pressure across the full range of this independent variable.

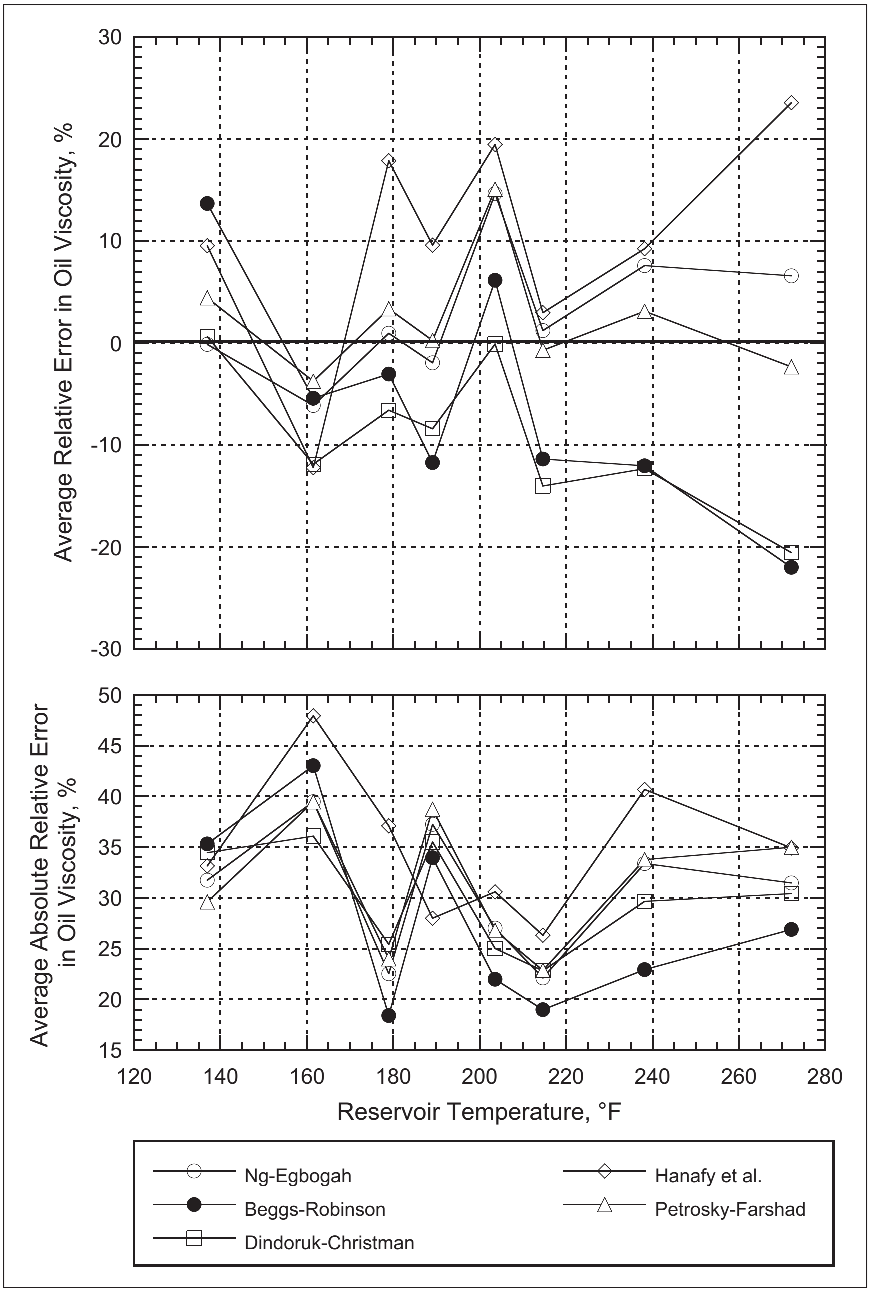

Fig. 3–35. Average relative errors and average absolute relative errors of subsets of the data described in table 3–24 sorted on reservoir temperature show the quality of selected correlation equations for estimating reservoir oil viscosities for pressures equal to and less than bubblepoint pressure across the full range of this independent variable.

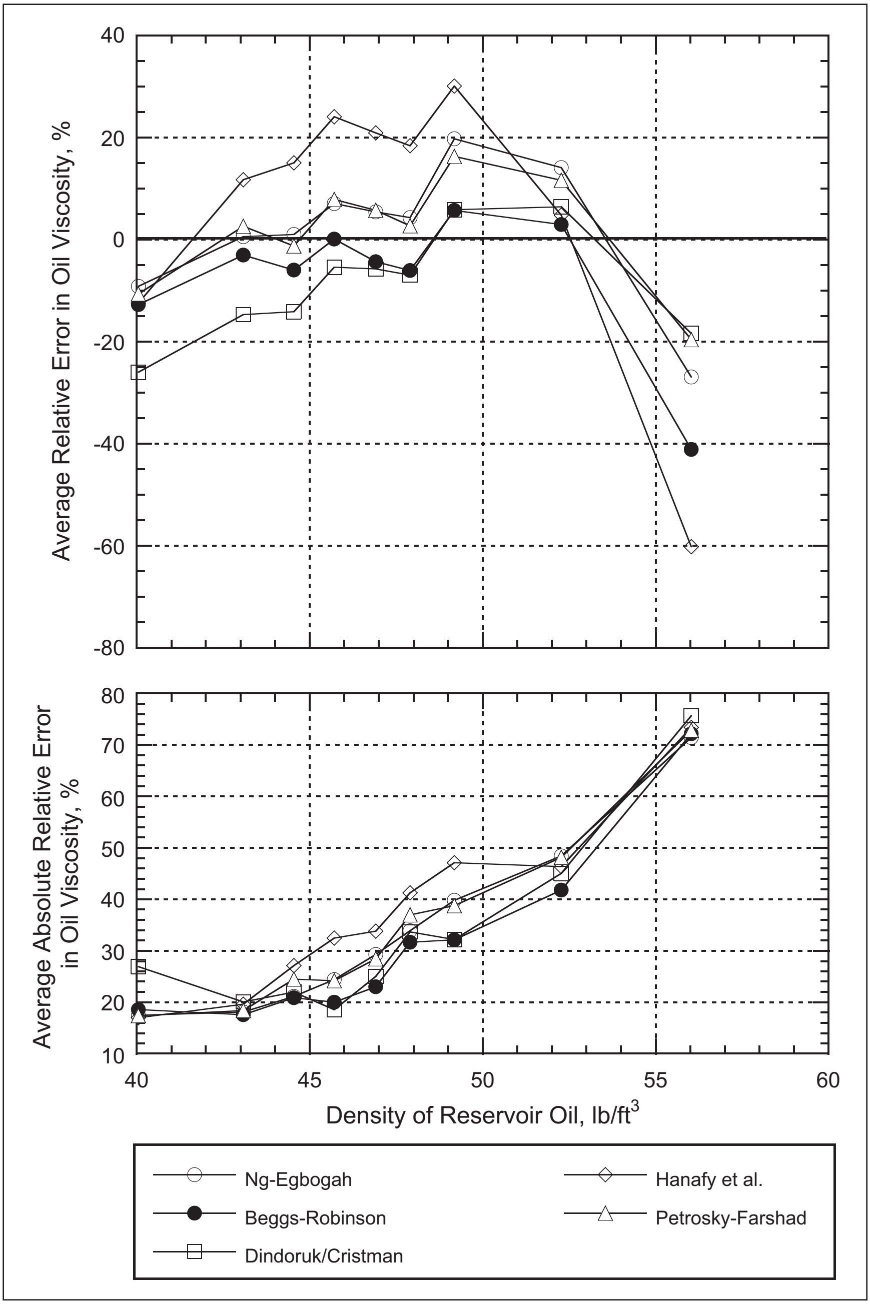

Fig. 3–36. Average relative errors and average absolute relative errors of subsets of the data described in table 3–24 sorted on reservoir oil density show the quality of selected correlation equations for estimating reservoir oil viscosities for pressures equal to and less than bubblepoint pressure across the full range of this independent variable.

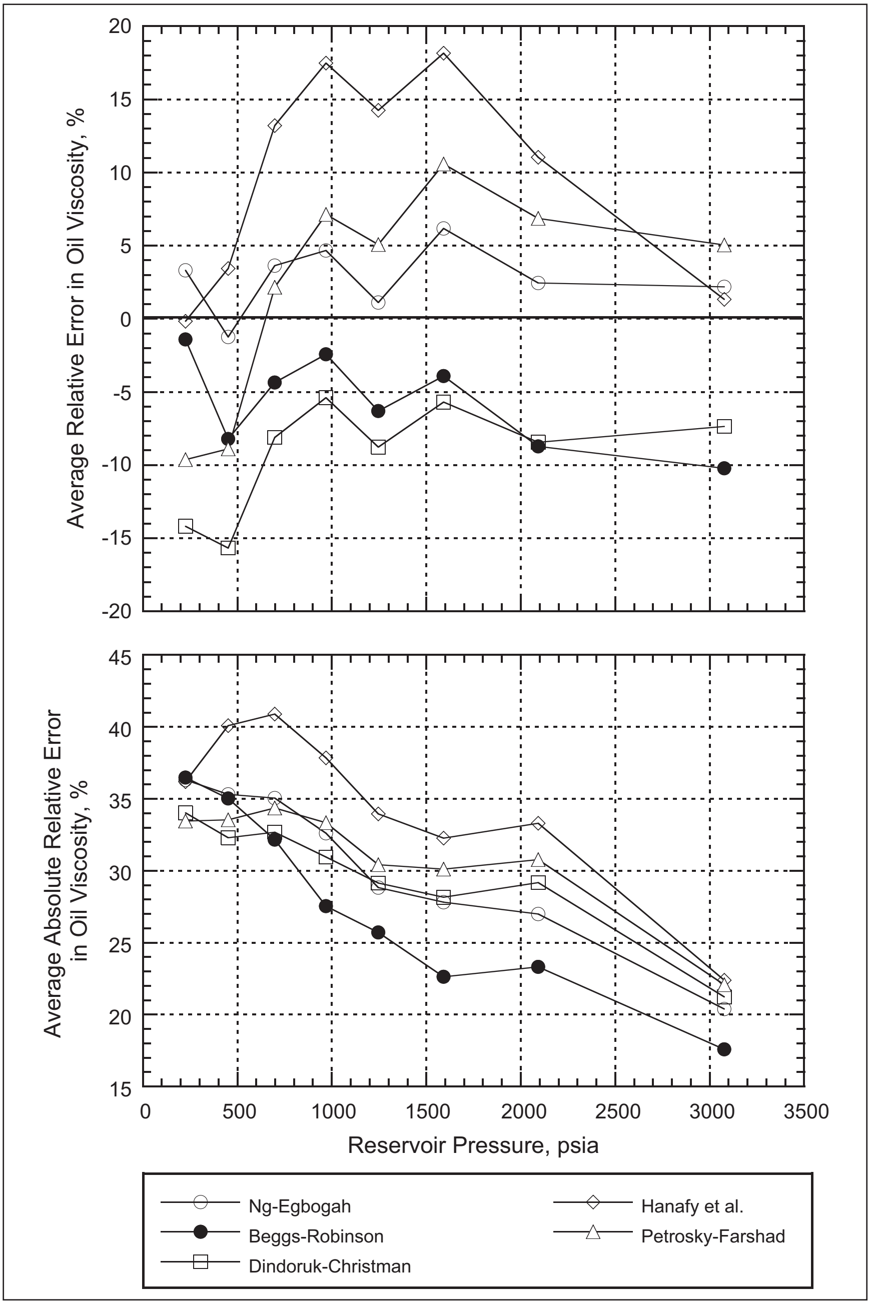

Fig. 3–37. Average relative errors and average absolute relative errors of subsets of the data described in table 3–24 sorted on reservoir pressure show the quality of selected correlation equations for estimating reservoir oil viscosities for pressures equal to and less than bubblepoint pressure across the full range of this independent variable.

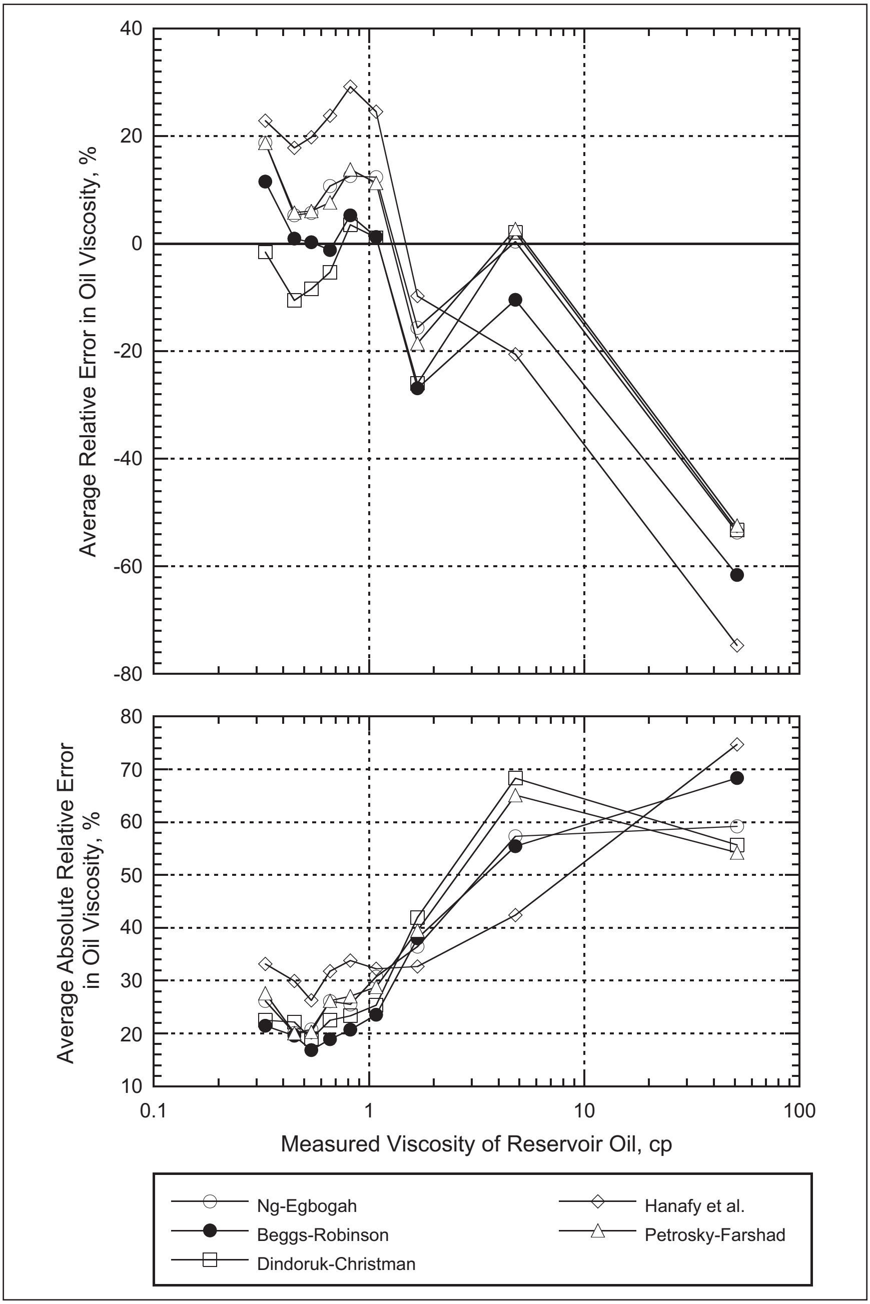

Fig. 3–38. Average relative errors and average absolute relative errors of subsets of the data described in table 3–24 sorted on measured reservoir oil viscosity show the quality of selected correlation equations for estimating reservoir oil viscosities for pressures equal to and less than bubblepoint pressure across the full range of this dependent variable.

Although the values of ARE in these figures do not indicate a preference for any of the correlation equations, the AARE values clearly indicate the Beggs-Robinson formulations for both dead oil viscosity and viscosity at pressure to be better than the others.

Recommended correlation equations for estimation of reservoir oil viscosities at reservoir pressures equal to and less than bubblepoint pressure

Thus, the recommended equations for oil viscosities for use with pressures equal to and less than the bubblepoint pressure are given in equations (3.23a)–(3.23e).[47]

$$\mu_{oD} = 10^{C} - 1 \tag{3.23a}$$

where

$$C = \left(10^{3.0324 - 0.02023\,API}\right)\left(T^{-1.163}\right) \tag{3.23b}$$

and

$$\mu_{oR} = A\mu_{oD}{}^{B} \tag{3.23c}$$

where

$$A = 10.715\left(R_s + 100\right)^{-0.515}, \tag{3.23d}$$

$$B = 5.44\left(R_s + 150\right)^{-0.338}, \tag{3.23e}$$

and where values of R_s can be obtained at the pressure of interest using equations (3.8a)–(3.8f).

Evaluation of proposed correlation equations for reservoir oil viscosities at reservoir pressures greater than bubblepoint pressure

Statistics of the data set used to evaluate reservoir oil viscosity correlation equations at reservoir pressures greater than the bubblepoint pressure of the reservoir oil are given in table 3–26.

Table 3–26. Statistics of data set used to evaluate oil viscosity correlation equations at reservoir pressures greater than bubblepoint pressure (1,981 lines of data from 186 reservoir fluid studies with worldwide origins)

Laboratory measurement	**Minimum**	**Median**	**Mean**	**Maximum**
Bubblepoint pressure, p_b, psia	120.7	1,689.7	1,821.4	5,311.7
Reservoir pressure, p, psia	120.7	2,714.7	2,749.1	7,514.7
Reservoir temperature, T_R, °F	107.0	193.0	197.5	320.0
Stock-tank oil gravity, API, °API	15.8	35.0	34.4	57.7
Separator gas specific gravity, γ_{gSP}	0.6506	0.8579	0.8949	1.9161
Solution gas-oil ratio at p_b, R_{sb}, scf/STB	12.0	384.0	461.2	1,534.0
Reservoir oil density, ρ_{oR}, lb/cu ft	34.27	46.49	46.95	58.30
Reservoir oil viscosity, μ_{oR}, cp	0.17	0.84	10.18	279.83

An evaluation of published reservoir oil viscosity correlation equations is given in table 3–27.

Table 3–27. Evaluation of published oil viscosity correlation equations for reservoir pressures greater than bubblepoint pressure

	Predicted undersaturated oil viscosity	
Correlation	**ARE, %**	**AARE, %**
Petrosky-Farshad (1995)*	–18.40	32.89
Bergman-Sutton (2006)[a]	–18.58	32.99
Dindoruk-Christman (2004)[b]*	–17.20	34.69
Vazquez-Beggs (1980)	–13.61	35.09
De Ghetto et al. (1994)[c]	–11.52	35.87
Hanafy et al. (SPE 37439, 1997)*	–6.27	36.34
Almehaideb (1997)[d]*	10.32	36.75
Elsharkawy-Alikhan (1999)[e]*	3.96	37.86
Elsharkawy-Gharbi (2001)[f]	–11.80	40.09
Kartoatmodjo-Schmidt (1994)	–21.60	41.82
Khan et al. (1987)[g]*	0.77	44.49
Abdul-Majeed et al. (1990)[h]	1.34	45.89
Standing (1977)*	–17.15	55.09
Al-Khafaji et al. (1987)[i]	46.21	76.95
Labedi (1992)[j]	53.79	87.81

[a] Bergman, D.F. and Sutton, R.P. "Undersaturated Oil Viscosity Correlation for Adverse Conditions." Paper SPE 103144. Presented at the SPE Annual Technical Conference and Exhibition, San Antonio (2006). [b] Dindoruk, B. and Christman, P.G. "PVT Properties and Viscosity Correlations for Gulf of Mexico Oils." *Society of Petroleum Engineers Reservoir Evaluation and Engineering*. Vol. 7, No. 6 (2004) 427–437. [c] De Ghetto, G., Paone, F., and Villa, M. "Reliability Analysis on PVT Correlations." Paper SPE 28904. Presented at the European Petroleum Conference, London (1994). [d] Almehaideb, R.A. "Improved PVT Correlations for UAE Crude Oils." Paper SPE 37691. Presented at the SPE Middle East Oil Conference and Exhibition, Bahrain (1997). [e] Elsharkawy, A.M. and Alikhan, A.A. "Models for Predicting the Viscosity of Middle East Crude Oil. *Fuel*. Vol. 78 (1999) 891–903. [f] Elsharkawy, A.M. and Gharbi, R.B.C. "Comparing Classical and Neural Regression Techniques in Modeling Crude Oil Viscosity." *Advances in Engineering Software and Workstations*. Vol. 32 (2001) 215–224. [g] Khan, S.A., Al-Marhoun, M.A., Duffuaa, S.O., and Abu-Khamsin, S.A. "Viscosity for Saudi-Arabian Crude Oils." Paper SPE 15720. Presented at the Fifth Middle East Oil Show, Manama, Bahrain (1987). [h] Abdul-Majeed, G., Kattan, R.R., and Salman, N.H. "New Correlation for Estimating the Viscosity of Undersaturated Crude Oils." *Journal of Canadian Petroleum Technology*. Vol. 29 (1990) 80–85. [i] Al-Khafaji, A.H., Abdul-Majeed, G.H., and Hassoon, S.F. "Viscosity Correlation for Dead, Live and Undersaturated Crude Oil." *Journal of Petroleum Research*. Vol. 6, No. 2 (1987) 1–6. [j] Labedi, R. "Improved Correlations for Predicting the Viscosity of Light Crudes." *Journal of Petroleum Science and Engineering*. Vol. 8 (1992) 221–234. * Author restricted correlation to a specific geographical area.

Vazquez and Beggs, and Kartoatmodjo and Schmidt used essentially all of the data described in table 3–26 in preparation of their correlation equations.[48] It is not known if the other researchers had access to these data.

The data set described in table 3–26 was sorted and sliced into 10 subsets of approximately equal size, approximately 200 lines of data each. The results of calculating ARE and AARE on these subsets for several of the better correlation equations are shown in figures 3–39 through 3–44. Several of the correlation equations depend on values of oil viscosities at bubblepoint pressures. Equations (3.23a)–(3.23e) were selected to determine these values since they represent the best available procedures for use at bubblepoint pressures.

Recommended correlation equations for estimation of reservoir oil viscosities at reservoir pressures greater than bubblepoint pressure

Although there does not appear to be a favorite when examining the ARE figures, the Petrosky-Farshad equations appear to have an edge in the AARE figures.[49] Thus, the recommended correlation equations are given in equations (3.24a) and (3.24b).

$$\mu_{oR} = \mu_{ob} + 1.3449x10^{-3}\left(p - p_b\right) \cdot 10^{A} \tag{3.24a}$$

where

$$\begin{aligned} A = {} & -1.0146 + 1.3322\left[\log\left(\mu_{ob}\right)\right] \\ & -0.4876\left[\log\left(\mu_{ob}\right)\right]^2 - 1.15036\left[\log\left(\mu_{ob}\right)\right]^3 \end{aligned} \tag{3.24b}$$

Values of oil viscosities at bubblepoint pressures for use in equations (3.24a) and (3.24b) are calculated with equations (3.23a)–(3.23e) using values of solution gas-oil ratios at bubblepoint pressures. This ensures the continuity of the values of reservoir oil viscosities as reservoir pressure declines through the bubblepoint pressure.

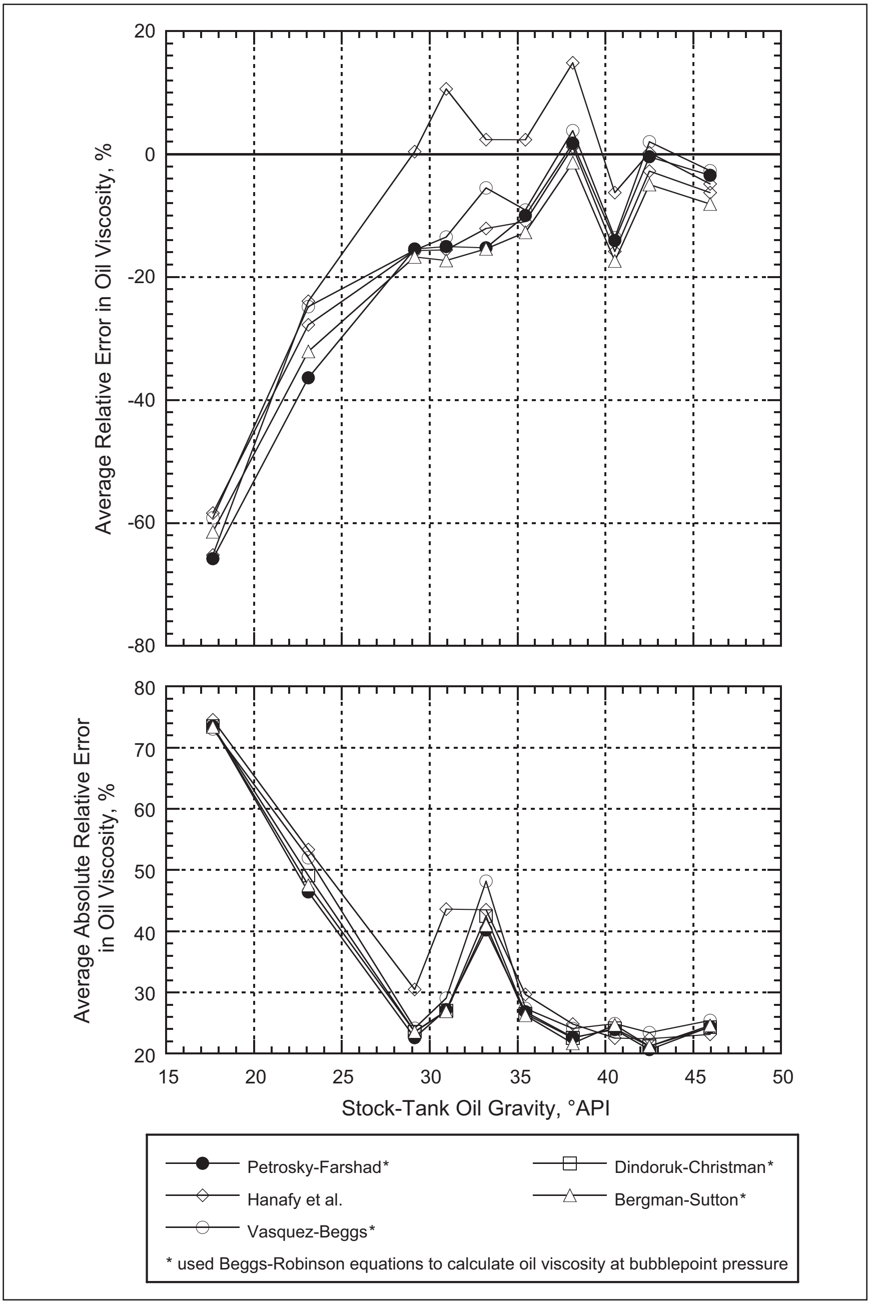

Fig. 3–39. Average relative errors and average absolute relative errors of subsets of the data described in table 3–26 sorted on stock-tank oil gravity show the quality of selected correlation equations for estimating reservoir oil viscosities for pressures greater than bubblepoint pressure across the full range of this independent variable.

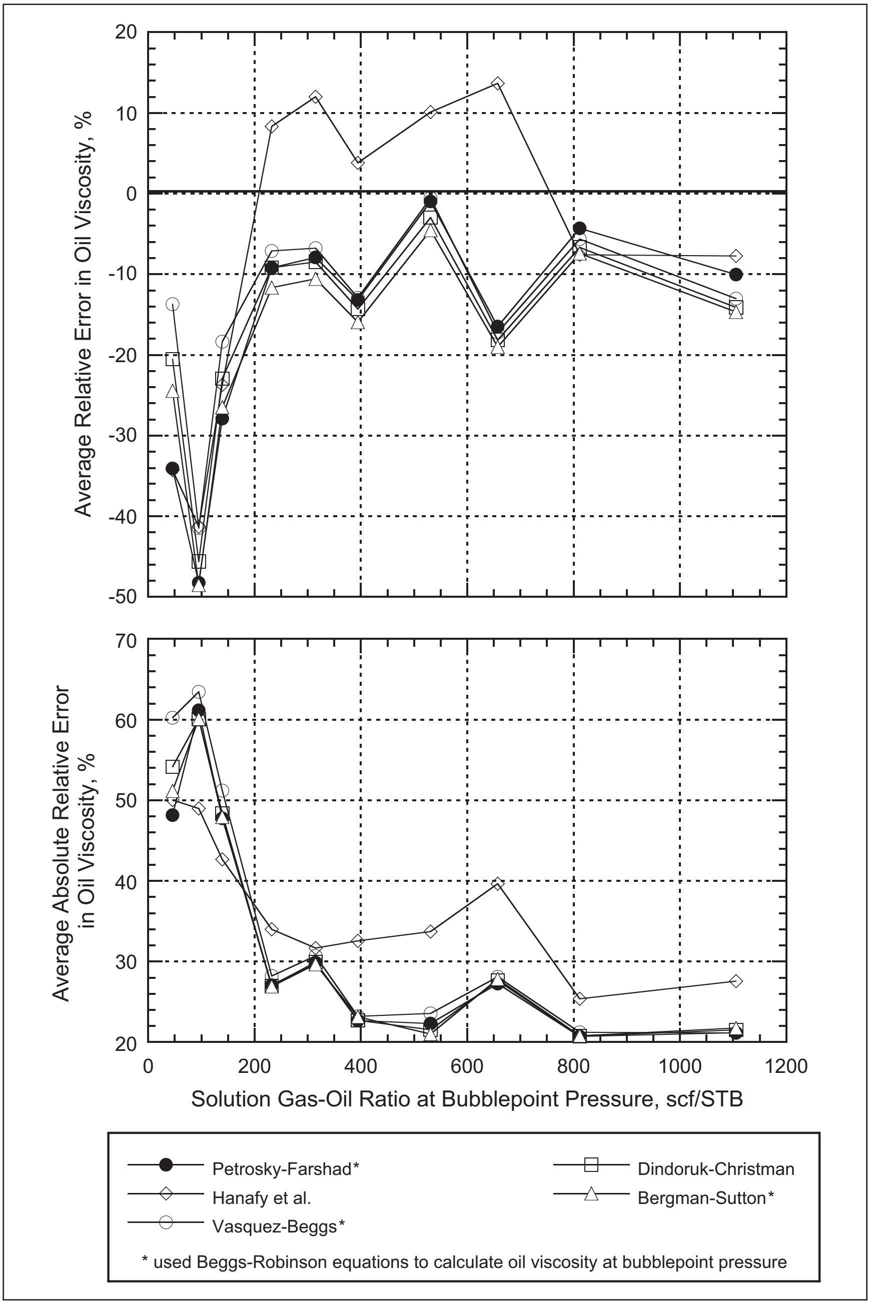

Fig. 3–40. Average relative errors and average absolute relative errors of subsets of the data described in table 3–26 sorted on solution gas-oil ratio at bubblepoint pressure show the quality of selected correlation equations for estimating reservoir oil viscosities for pressures greater than bubblepoint pressure across the full range of this independent variable.

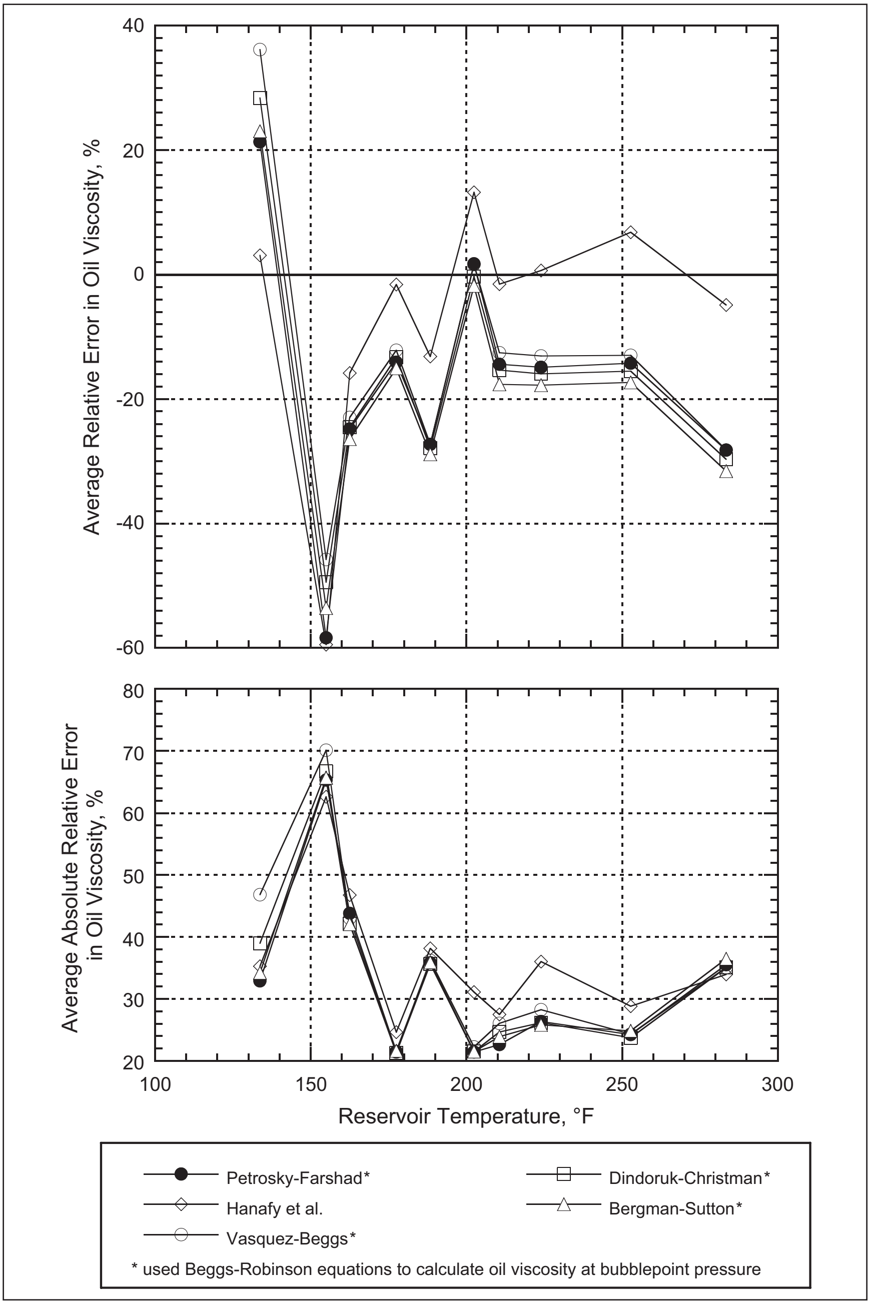

Fig. 3–41. Average relative errors and average absolute relative errors of subsets of the data described in table 3–26 sorted on reservoir temperature show the quality of selected correlation equations for estimating reservoir oil viscosities for pressures greater than bubblepoint pressure across the full range of this independent variable.

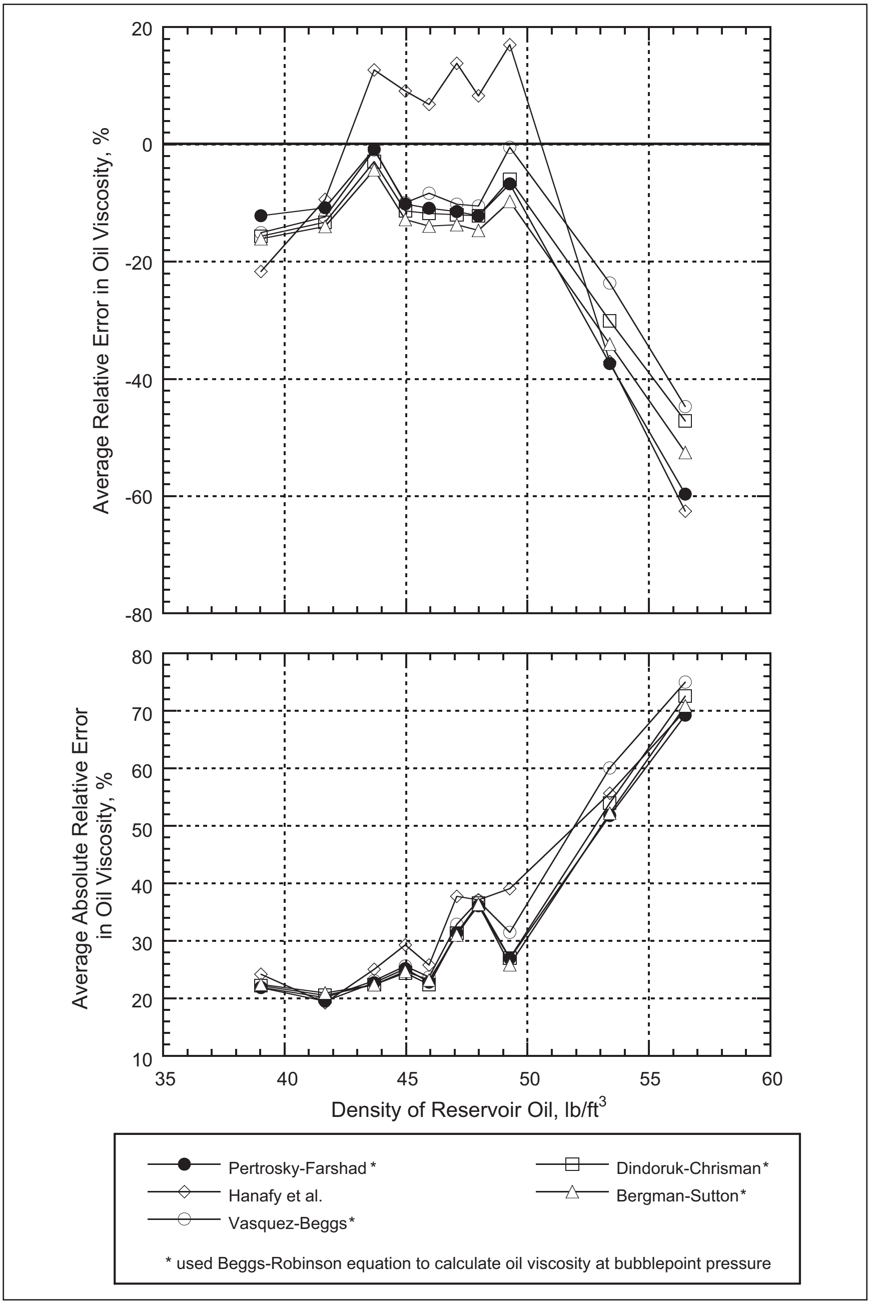

Fig. 3–42. Average relative errors and average absolute relative errors of subsets of the data described in table 3–26 sorted on reservoir oil density show the quality of selected correlation equations for estimating reservoir oil viscosities for pressures greater than bubblepoint pressure across the full range of this independent variable.

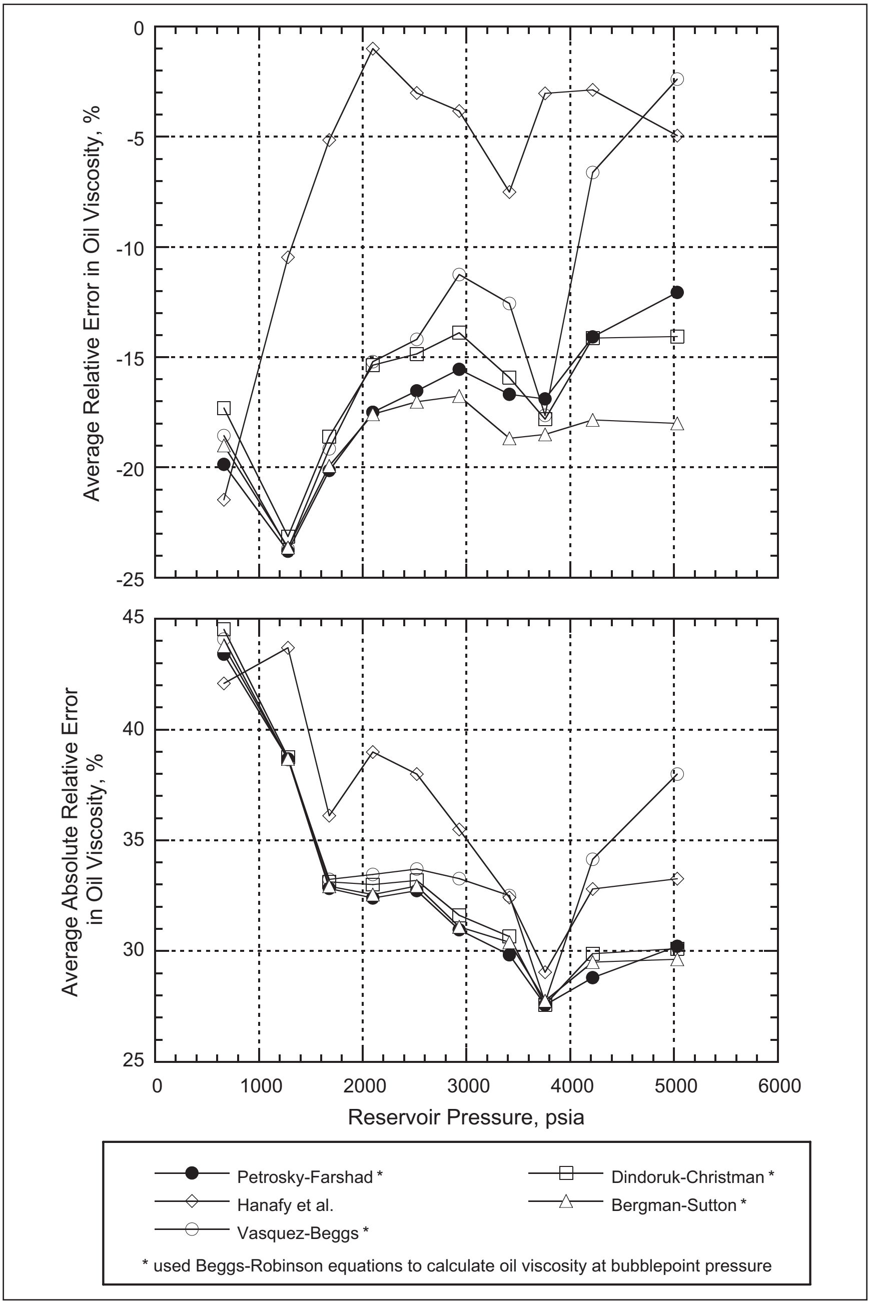

Fig. 3–43. Average relative errors and average absolute relative errors of subsets of the data described in table 3–26 sorted on reservoir pressure show the quality of selected correlation equations for estimating reservoir oil viscosities for pressures greater than bubblepoint pressure across the full range of this independent variable.

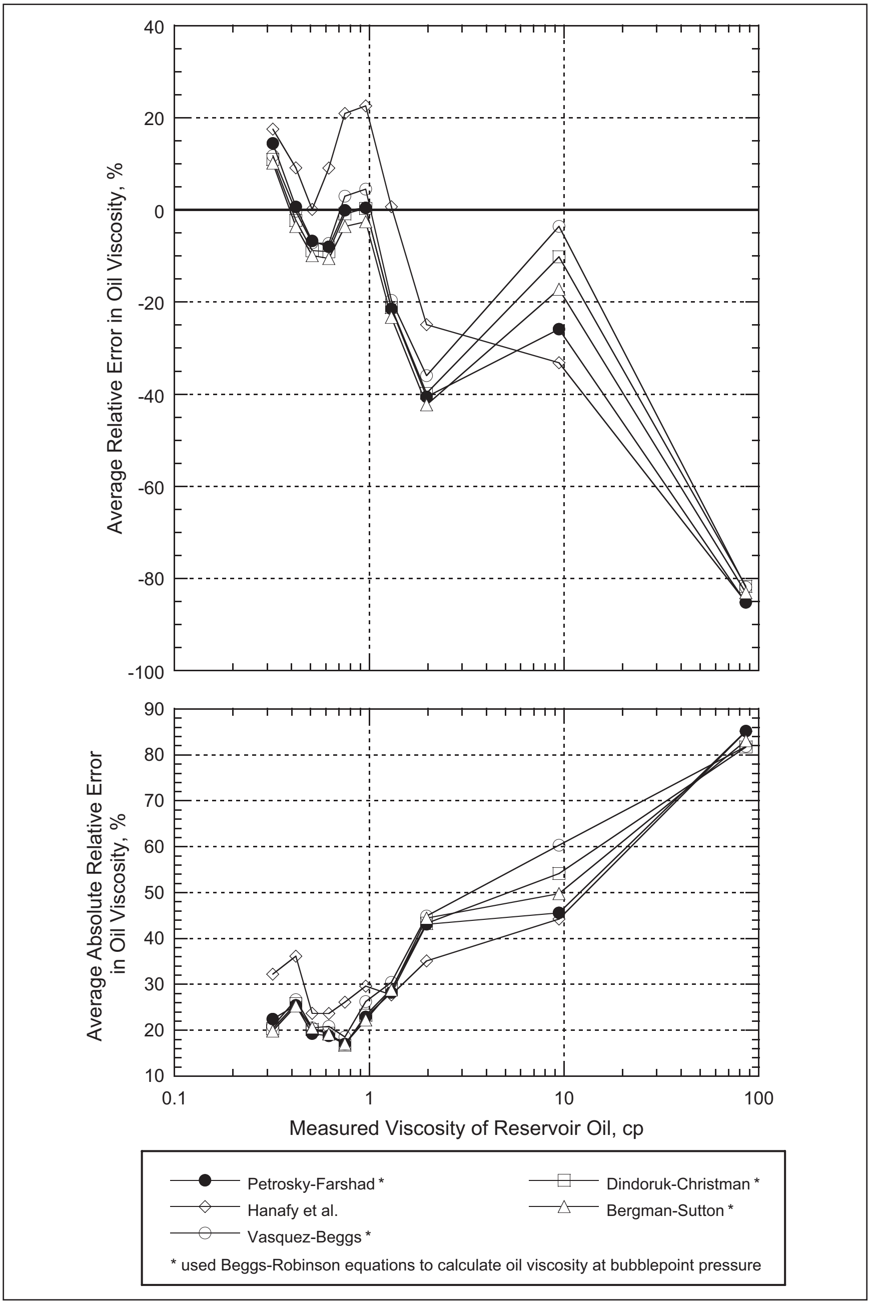

Fig. 3–44. Average relative errors and average absolute relative errors of subsets of the data described in table 3–26 sorted on measured reservoir oil viscosity show the quality of selected correlation equations for estimating reservoir oil viscosities for pressures greater than bubblepoint pressure across the full range of this dependent variable.

Comments on reservoir oil viscosity correlation equations

The very large values of average absolute relative error in tables 3–23, 3–25, and 3–27 do not occur because the researchers proposing the correlation equations are incompetent. Rather, they reflect the quality of the reservoir oil viscosity data. Most of these data were measured using an instrument called a *rolling-ball viscosimeter*. This device is simple and easy to use, and thus the procedure is relatively inexpensive. Unfortunately, in order to produce accurate results with this equipment, the tested fluid must be a Newtonian fluid. Reservoir oils with high densities, and thus high viscosities, have components (resins and asphaltenes) that cause the oils to be non-Newtonian. Thus, the resulting data are very inaccurate, and any correlation equation prepared using these poor quality data cannot be any better than the quality of the data. The same results can be observed in figures 3–33 through 3–44.

Figures 3–38 and 3–44 show that the better correlations fit the measured oil viscosities reasonably well for oils with viscosities below approximately 1 cp. Thus, the measured values of oil viscosities below approximately 1 cp are most likely reasonably accurate. Reservoir oils with larger viscosities apparently have very large measurement errors, and these poor results contribute to inaccurate correlation equations. Further, the use in engineering calculations of oil viscosities higher than approximately 1 cp as measured with the rolling-ball instrument can result in inaccurate results.

The relationships among oil density, oil formation volume factor, and oil compressibility at pressures above the bubblepoint pressure have already been discussed. Values of these properties are controlled primarily by the expansion of the reservoir oil as reservoir pressure declines for pressures above the bubblepoint pressure. This expansion should also affect reservoir oil viscosities at pressures above the bubblepoint pressure. So a correlation of reservoir oil viscosities should take these relationships into account, most likely reservoir oil density. Abu-Khamsin and Al-Marhoun are the only researchers that have recognized this.[50] Although their work was somewhat successful, their equation was devised only for use at bubblepoint pressure. Future work should explore the relationships of reservoir oil viscosity with reservoir oil density or even oil compressibility at pressures above the bubblepoint pressure.

Free Gas in Reservoir at Reservoir Pressures Less Than Bubblepoint Pressure

McCain and Hill reported a reasonably accurate correlation equation to predict the specific gravities of gases evolved in the reservoir from black oils at pressures below the bubblepoint pressures of the reservoir oils.[51] These values of free gas specific gravities can be used with the gas correlations discussed in chapter 2 to estimate values of the properties of reservoir free gas associated with black oils. No other correlations for the specific gravities of reservoir free gas have been published.

The equation is given in equation (3.25).

$$\frac{1}{\gamma_{gR}} = \frac{a_1}{p} + \frac{a_2}{p^2} + a_3 p + \frac{a_4}{\sqrt{T}} + a_5 T + a_6 R_{sb} + a_7 API + \frac{a_8}{\gamma_{gSP}} + a_9 \gamma_{gSP}^2 \quad (3.25)$$

a_1	–208.0797
a_2	22,885
a_3	–0.000063641
a_4	3.38346
a_5	–0.000992
a_6	–0.000081147
a_7	–0.001956
a_8	1.081956
a_9	0.394035

Evaluation of recommended correlation equation for specific gravity of free gas in the reservoir

This equation was evaluated with the data set described in table 3–28. Notice that the minimum pressure is approximately 300 psig; this was the minimum reservoir pressure specified for equation (3.25) by McCain and Hill.[52] Apparently use of this equation for pressures less than 300 psig can result in large errors in predicted free gas specific gravities.

Table 3–28. Statistics of data set used to evaluate the specific gravities of free gases in the reservoir at reservoir pressures below bubblepoint pressures (2,453 lines of data from 313 reservoir fluid studies with worldwide origins)

Laboratory measurement	**Minimum**	**Median**	**Mean**	**Maximum**
Reservoir gas specific gravity, γ_{gR}	0.559	0.776	0.788	1.530
Reservoir pressure, p, psia	311.0	1,214.7	1,488.1	7,200.0
Reservoir temperature, T_R, °F	70.0	186.0	183.2	327.0
Solution gas-oil ratio at p_b, R_{sb}, scf/STB	101.0	738.0	721.8	2,424.0
Stock-tank oil gravity, *API*, °API	11.6	39.6	37.5	63.0

The evaluation led to an ARE of –0.17% and an AARE of 3.76%. McCain and Hill said that use of the resulting values of reservoir free gas specific gravities in the correlation equations given in chapter 2 gave values of gas z-factors and gas formation volume factors within ±2% of measured values of these properties.

The data of table 3–28 were sorted into 10 subsets, each with approximately 250 lines of data. The values of ARE and AARE calculated with these subsets of data are shown in figures 3–45, 3–46, and 3–47.

The data were also sorted on reservoir pressure and reservoir temperature. The ARE and AARE were consistent across the full ranges of these conditions and plotted very much like figure 3–45.

The large increase in AARE at high values of solution gas-oil ratios in figure 3–46 and the large increases in ARE and AARE at high values of reservoir gas specific gravities in figure 3–47 indicate that equation (3.25) should not be used to estimate free gas specific gravities for volatile oils.

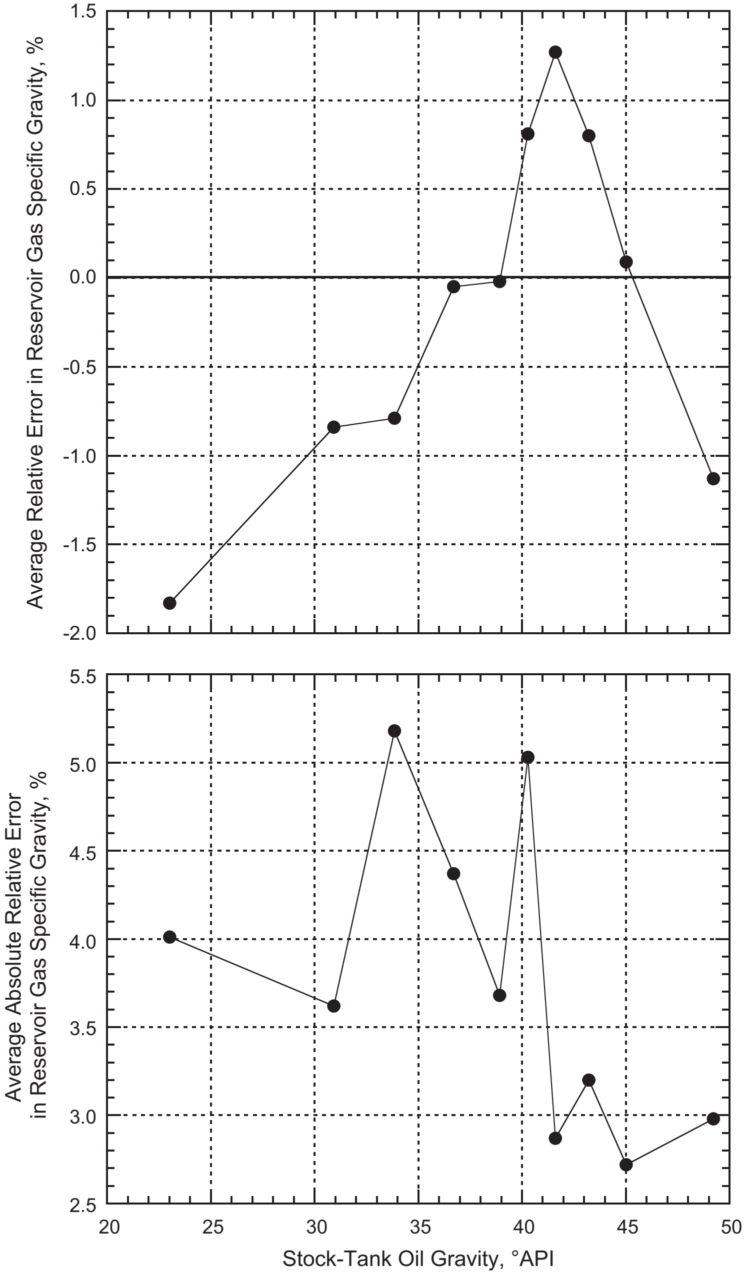

Fig. 3–45. Average relative errors and average absolute relative errors of subsets of the data described in table 3–28 sorted on stock-tank oil gravity show the quality of selected correlation equations for estimating reservoir gas specific gravities across the full range of this independent variable.

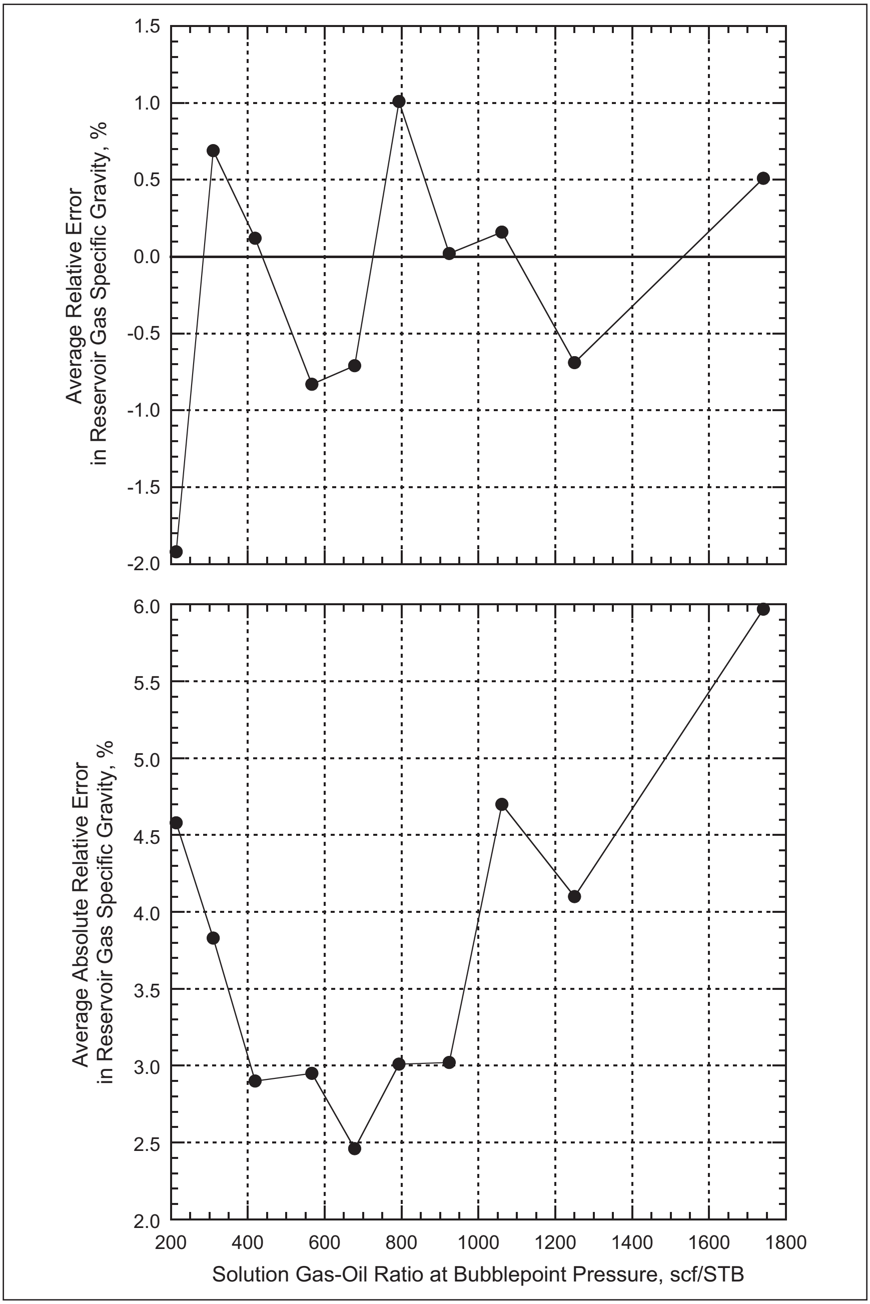

Fig. 3–46. Average relative errors and average absolute relative errors of subsets of the data described in table 3–28 sorted on solution gas-oil ratio at bubblepoint pressure show the quality of selected correlation equations for estimating reservoir gas specific gravities across the full range of this independent variable.

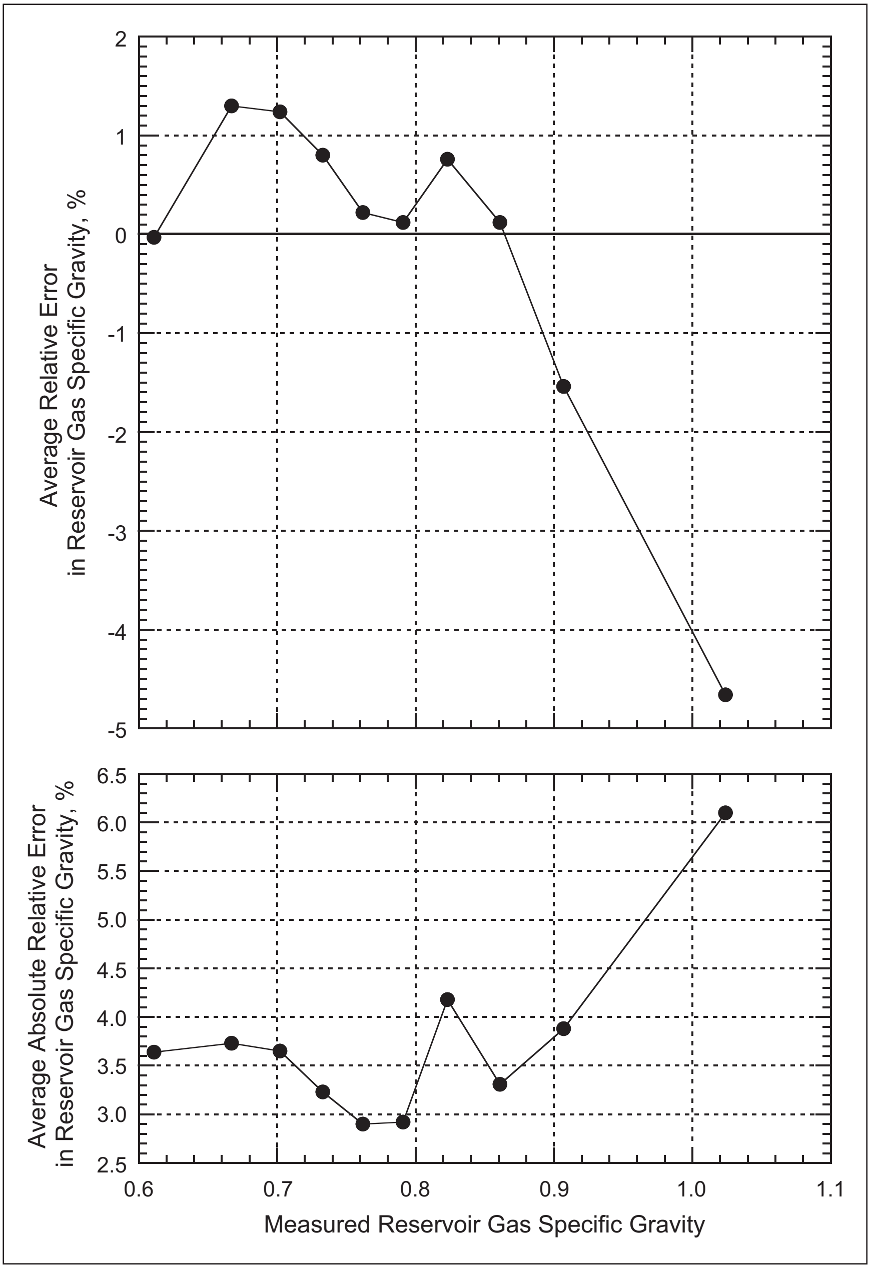

Fig. 3–47. Average relative errors and average absolute relative errors of subsets of the data described in table 3–28 sorted on measured reservoir gas specific gravity show the quality of selected correlation equations for estimating reservoir gas specific gravities across the full range of this dependent variable.

Nomenclature

Latin

a_1–a_9	Constants or coefficients in various equations
A	Coefficient in equation (3.23d) and also in equation (3.24b)
A_0–A_4	Constants in equation (3.8d)
API	Stock-tank oil gravity, °API
B	Coefficient in equation (3.23e)
B_0–B_4	Constants in equation (3.8e)
B_g	Gas formation volume factor, res bbl/scf, equation (2.12)
B_o	Oil formation volume factor, res bbl/STB, equation (3.21b)
B_{ob}	Oil formation volume factor at bubblepoint pressure, res bbl/STB, equation (3.21a)
B_{oi}	Oil formation volume factor at initial reservoir pressure, res bbl/STB
B_w	Water (brine) formation volume factor, res bbl/STB, equation (4.36)
c_e	Effective compressibility, 1/psi, equation (3.11b)
c_f	Pore volume compressibility, 1/psi
c_o	Coefficient of isothermal compressibility, 1/psi, equations (3.9) and (3.16a)
c_{ofb}	Weighted-average oil compressibility from bubblepoint pressure to a higher pressure of interest, 1/psi, page 56
c_{ofi}	Weighted-average oil compressibility from initial reservoir pressure to a lower pressure of interest
c_w	Coefficient of isothermal compressibility of water (brine), 1/psi, equation (4.32)
C	Coefficient in equation (3.23b)
C_0–C_4	Constants in equation (3.8f)
$C0_n$–$C4_n$	Constants in various equations
k_r	Permeability in radial direction, md
N	Original oil in place, STB

N_p	Cumulative oil production, STB
p	Pressure, psia
p_b	Bubblepoint pressure, psia, equation (3.7a)
p_i	Initial reservoir pressure, psia
p_r	Pressure ratio, (p – 14.7)/(p_b – 14.7), psig/psig, equation (3.8c)
p_{SP}	Separator pressure, psia
r	Radial distance, ft
R_s	Solution gas-oil ratio, scf/STB, equation (3.8a)
R_{sb}	solution gas-oil ratio at bubblepoint pressure, scf/STB, equation (3.1)
R_{sr}	Reduced solution gas-oil ratio, R_s/R_{sb}, equation (3.8b)
R_{SP}	Separator producing gas-oil ratio, scf of SP gas/STB
R_{ST}	Stock-tank producing gas-oil ratio, scf of ST gas/STB, equation (3.2a)
S_o	Oil saturation, fraction of pore space
S_w	Water (brine) saturation, fraction of pore space
t	Time, hours
T	Temperature, °F
T_R	Reservoir temperature, °F
T_{SP}	Separator temperature, °F
V	Volume
VAR_n	Independent variable in various equations
W_e	Cumulative water (brine) encroached into reservoir from aquifer, res bbl
W_p	Cumulative water (brine) production, STB
Z	Summation of transforms of independent variables in various equations
Z_n	Transforms of independent variables in various equations

Greek

$\Delta\rho_p$	Adjustment to oil density due to pressure, lb/cu ft, equations (3.18e) and (3.19d)
$\Delta\rho_T$	Adjustment to oil density due to temperature, lb/cu ft, equations (3.18g) and (3.19f)
γ_g	Weighted average surface gas specific gravity, equation (3.4)
γ_{gR}	Reservoir free gas specific gravity, equation (3.25)
γ_{gSP}	Separator gas specific gravity
γ_{gST}	Stock-tank gas specific gravity, equation (3.5a)
γ_{STO}	Stock-tank oil specific gravity
μ	Viscosity, cp
μ_{ob}	Oil viscosity at bubblepoint pressure, cp
μ_{oD}	Dead oil viscosity, cp, equation (3.23a)
μ_{oR}	Oil viscosity at reservoir conditions, cp, equation (3.23c) or (3.24a)
ρ	Oil density, lb/cu ft
ρ_a	Apparent liquid density of total surface gases, lb/cu ft, equation (3.17e) or (3.18c)
ρ_{bs}	Fake density used in oil density calculations, lb/cu ft, equations (3.18f) and (3.19e)
ρ_o	Oil density, lb/cu ft
ρ_{ob}	Oil density at bubblepoint pressure, lb/cu ft
ρ_{oR}	Oil density at reservoir conditions, lb/cu ft, equation (3.19a) or (3.20)
ρ_{oRb}	Oil density at reservoir bubblepoint pressure, lb/cu ft, equation (3.18a)
ρ_{po}	Density of pseudoliquid created by combining (by calculation) the stock-tank liquid with the total surface gases, equations (3.18b) and (3.19b)
ρ_{STO}	Density of stock-tank oil, lb/cu ft
ϕ	Porosity, fraction

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4

PROPERTIES OF OILFIELD BRINES

The brine properties of interest in petroleum reservoir engineering include the following:

1. Brine density
2. Brine formation volume factor
3. Coefficient of isothermal compressibility of brine
4. Solution gas-water ratio
5. Brine viscosity

Brine Correlations

The five field properties of interest listed above can be calculated from three independent correlations based on lab measurements: (1) density of methane-free brine, (2) solubility of methane in brine, and (3) brine viscosity. Because the density, formation volume factor, and compressibility are interrelated, the authors prefer to use a single correlation to calculate all three quantities, thereby eliminating potential inconsistencies.

The next three sections compare correlations for brine density, brine compressibility, and methane solubility with available laboratory data reported in the literature. The following section gives a step-by-step procedure for calculating the reservoir brine properties of interest: brine density, brine formation volume factor, brine compressibility, and solution gas-water ratio. The final three sections compare correlations for brine viscosity with selected laboratory data from the literature, discuss extrapolation of data beyond the range of data on which the correlations are based, and give a step-by-step procedure for estimating viscosities for oilfield brines.

Brine Density

Volumetric properties database

A large number of studies of volumetric properties of sodium chloride brines have been published over the last 80 years. Published data have included specific volume, relative volume, compressibility, density, and the difference in density between that of brine and pure water.

Spivey et al. compiled a database of published data on volumetric properties of sodium chloride brines for temperatures from 32°F to 530°F (0°C to 277°C), pressures from 14.5 to 29,000 psia (0.1 to 200 MPa), and salinities from 0% to 26% sodium chloride by weight (0 to 6 g-mol/kg H_2O).[1] They then used this database to develop a new correlation for estimating brine density, specific volume, and coefficient of isothermal compressibility.

Table 4–1. Brine volumetric properties databas

Reference	**n**	**Temperature range (°C)**	**Pressure range (MPa)**	**NaCl salinity range (g-mol/kg H_2O)**	**Quantity measured**
Freyer (1931)[a]	6/6	20	0.1	0.17–5.4	c_b
Adams (1931)[b]	30/30	25	0.1–200	0.0–5.7	v_b/v_{b0}
Gibson–Loeffler (1941), (1949)	174/175	25–85	0.1–100	0.9–5.7	v_b
Ellis (1966)	32/32	25–200	2.03	0.1–1	ρ_b
Millero (1970)[c]	83/84	0–55	0.1	0.01–1.0	$\rho_b – \rho_w$
Rowe-Chou (1970)	0/144	22–175	1–30.4	0–5.7	dv/dp
Hilbert (1979)	357/404	20–300	10–200	0.02–5.7	v_b
Rogers et al. (1982)[d]	54/54	75–200	2	0–4.4	ρ_b
Gehrig et al. (1983)	36/36	300	10–200	1.1–4.3	v_b
Gates–Wood (1985)[e]	40/40	25	0.1–40	0.05–5.0	$\rho_b – \rho_w$
Osif (1988)	253/254	93–132	6–138	0–3.7	v_b/v_{b0}
Majer et al. (1988)[f] Set I	250/250	50–275	0.1–40	0.056–5.0	$\rho_b – \rho_w$
Majer et al. (1988) Set II	163/163	48–277	0.6–33	0.0026–3.1	$\rho_b – \rho_w$
Mironenko et al. (2001)[g]	90/90	0–20	0.1	0.009–6.0	ρ_b
Combined	**1,568/1,762**	**0–300**	**0.1–200**	**0.0–6.0**	

Note: n = number of points used in fit/number of points in database. [a] Freyer, E.B. “Sonic Studies of the Physical Properties of Liquids. II. The Velocity of Sound in Solutions of Certain Alkali Halides and Their Compressibilities.” *Journal of the American Chemical Society*. Vol. 53 (April 1931) 1,313–1,320. [b] Adams, L.H. “Equilibrium in Binary Systems Under Pressure. I. An Experimental and Thermodynamic Investigation of the System, NaCl-H_2O, at 25°.” *Journal of the American Chemical Society*. Vol. 53 (October 1931) 3,769–3,813. [c] Millero, F.J. “The Apparent and Partial Molal Volume of Aqueous Sodium Chloride Solutions at Various Temperatures.” *Journal of Physical Chemistry*. Vol. 74, No. 2 (1970) 356–362. [d] Rogers, P.S.Z., Bradley, D.J., and Pitzer, K.S. “Densities of Aqueous Sodium Chloride Solutions from 75 to 200°C at 20 bar.” *Journal of Chemical & Engineering Data*. Vol. 27, No. 1 (1982) 47–50. [e] Gates, J.A. and Wood, R.H. “Densities of Aqueous Solutions of NaCl, MgCl2, KCl, NaBr, LiCl, and $CaCl^2$ from 0.05 to 5.0 mol kg-1 and 0.1013 to 40 MPa at 298.15 K.” *Journal of Chemical & Engineering Data*. Vol. 30 (1985) 44–49. [f] Majer, V., Gates, J.A., Inglese, A., and Wood, R.H. “Volumetric Properties of Aqueous NaCl Solutions from 0.0025 to 5.0 mol kg, 323 to 600 K, and 0.1 to 40 MPa.” *Journal of Chemical Thermodynamics*. Vol. 20 (1988) 949–968. [g] Mironenko, M.V., Boitnott, G.E., Grant, S.A., and Sletten, R.S. “Experimental Determination of the Volumetric Properties of NaCl Solutions to 253 K.” *Journal of Physical Chemistry*. B. Vol. 105, No. 41 (October 18, 2001) 9,909–9,912.

The Spivey et al. database has been updated to include additional data to extend the range of temperatures to 572°F (300°C).

Table 4–1 summarizes the data included in the updated database of brine volumetric properties. In addition to the data used in developing the Spivey et al. correlation equations, the database also includes data from Ellis and Gehrig et al., as well as additional data from Hilbert for temperatures up to 572°F (300°C).[2]

Figure 4–1 shows the distribution of data in the brine density database by pressure and temperature. Note that the database includes only data points for pressures above the vapor pressure curve for pure water.

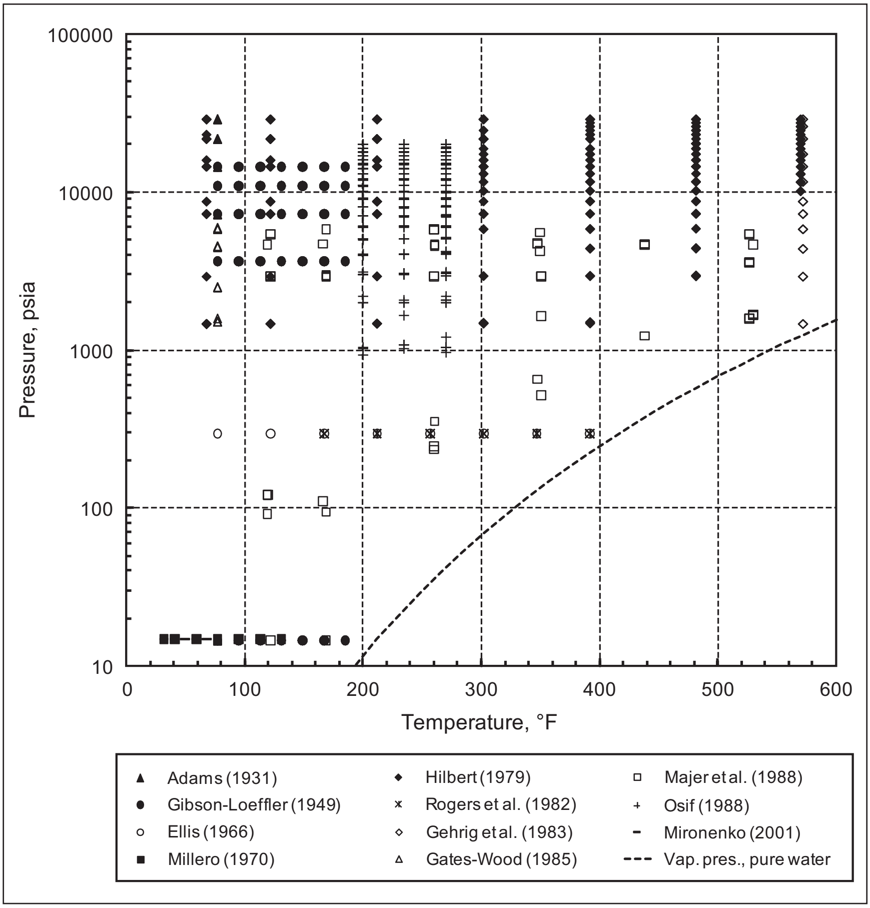

Fig. 4–1. Distribution of data points in updated brine density database, pressure vs. temperature

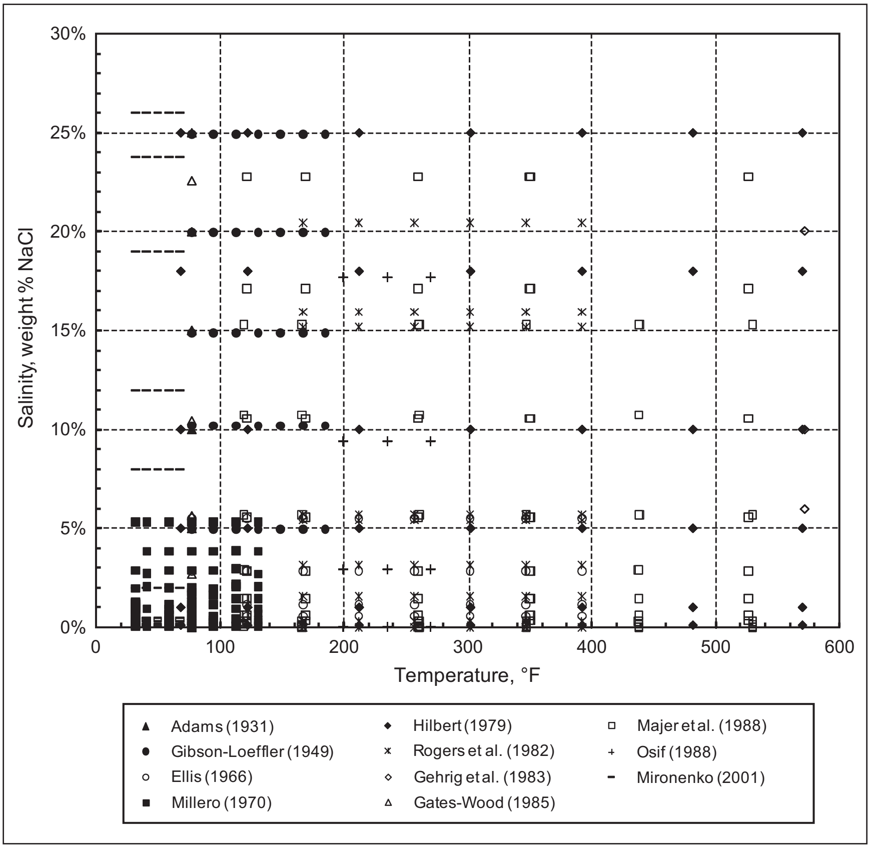

Fig. 4–2. Distribution of data points in updated brine density database, salinity vs. temperature

Figure 4–2 shows the distribution of data in the brine density database by salinity and temperature. Most of the data were obtained at temperatures below 212°F and salinities less than 6% sodium chloride by weight.

Figure 4–3 shows the distribution of data in the brine density database by pressure and salinity.

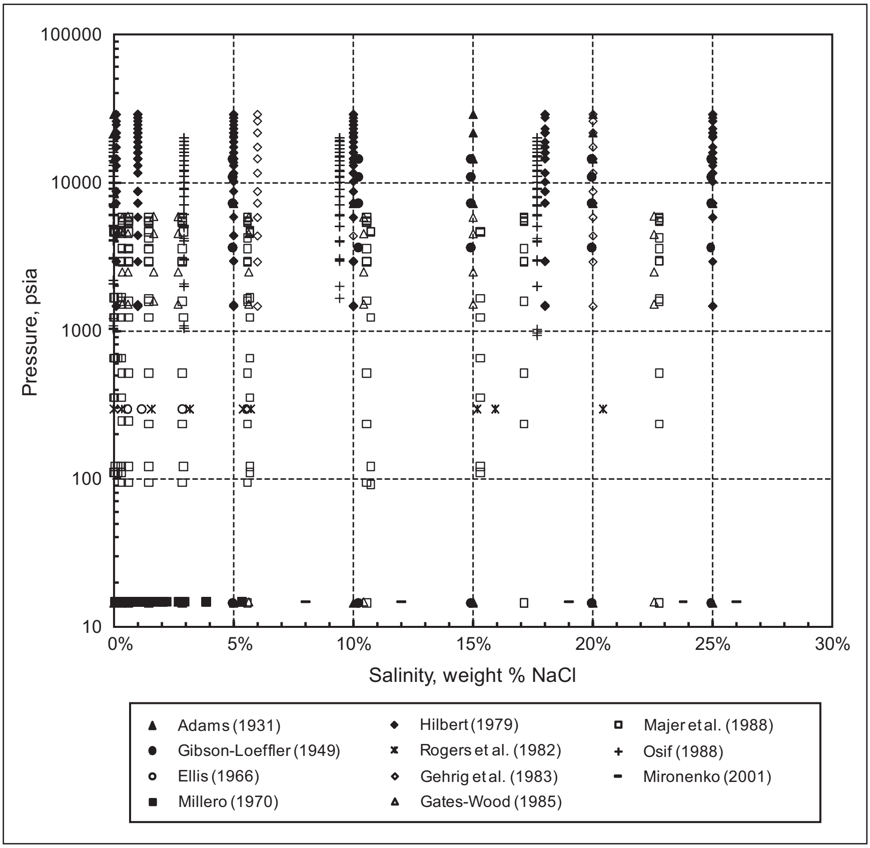

Fig. 4–3. Distribution of data points in updated brine density database, pressure vs. salinity

Comparison of correlations for brine density

Few correlations of brine density have been presented specifically for oilfield use. Six correlations and equations of state were selected for evaluation against the brine density database.

McCain presented a simple brine formation volume factor correlation by fitting polynomials to graphical correlations for the change in volume due to temperature (as a function of temperature), and the change in volume due to pressure (as a function of both pressure and temperature).[3] Density was obtained from the brine specific gravity as a function of total dissolved solids, also using a polynomial fit. The graphical correlations from which the polynomial fits were obtained cover a range of temperatures from 90°F to 260°F and pressures from 1,000 to 5,000 psia.

Rowe and Chou presented a correlation based on their compressibility measurements to calculate the specific volumes of brine as a function of temperature, pressure, and salinity.[4] Rowe and Chou claimed their specific volume correlation was within 0.15% for temperatures from 0°C to 150°C (32°F to 302°F), pressures from 1 to 350 kg/cm^2 (14.5 to 4,930 psia; 0.1 to 34MPa), and sodium chloride concentrations from 0 to 25 wt% (0 to 5.7 g-mol/kg H_2O).

Kemp et al. presented a density correlation for both single salts and salt mixtures, for drilling fluid applications.[5] They gave coefficients for NaCl, NaBr, KCl, KBr, $CaCl_2$, $CaBr_2$, $ZnCl_2$, and $ZnBr_2$. However, the authors did not give a range of validity for their correlation, noting that while extensive data are available for sodium chloride, data for other salts cover much narrower ranges of pressure and temperature.

Rogers and Pitzer presented an equation of state for predicting volumetric properties of aqueous sodium chloride solutions at temperatures from 32°F to 572°F (0°C to 300°C), pressures from 0 to 14,500 psia (0 to 100 MPa), and sodium chloride concentrations up to 24.3 wt% (5.5 g-mol/kg H_2O).[6]

Archer presented an equation of state similar to that of Rogers and Pitzer using a larger database, incorporating additional data published after 1982.[7] Archer's equation of state covered the same range of temperatures and pressures as Rogers and Pitzer, with better fit of all but one of the data sets studied by Rogers and Pitzer. Archer and Carter later extended this equation of state to lower temperatures.[8]

The Rogers-Pitzer and Archer equations of state both rely on the use of a complex equation of state for pure water, and both are cumbersome to implement. Further, both the Rogers-Pitzer and the Archer equations of state predict negative compressibilities when extrapolated to pressures above 22,000 psia (150 MPa) at high temperatures.

Spivey et al. presented a correlation to estimate volumetric properties of sodium chloride brines for temperatures from 32°F to 527°F (0°C to 275°C), pressures from 14.5 to 29,000 psia (0.1 to 200 MPa), and salinities from 0 to 26 wt% sodium chloride (0 to 6 g-mol/kg H_2O).[9] This correlation was later modified to extend the temperature range to 572°F (300°C).

Table 4–2 shows the results of the evaluation of correlations for brine density using the database summarized in table 4–1.

Table 4–2. Results of evaluation of brine density correlations against brine density database in table 4–1

Correlation	ARE	AARE
Spivey et al. (modified)	–0.009%	0.026%
Archer (1992)	–0.045%	0.066%
Rogers-Pitzer (1982)	–0.047%	0.073%
Kemp et al. (1989)	0.149%	0.250%
Rowe-Chou (1970)	0.517%	0.237%
McCain (1990)	2.501%	2.745%

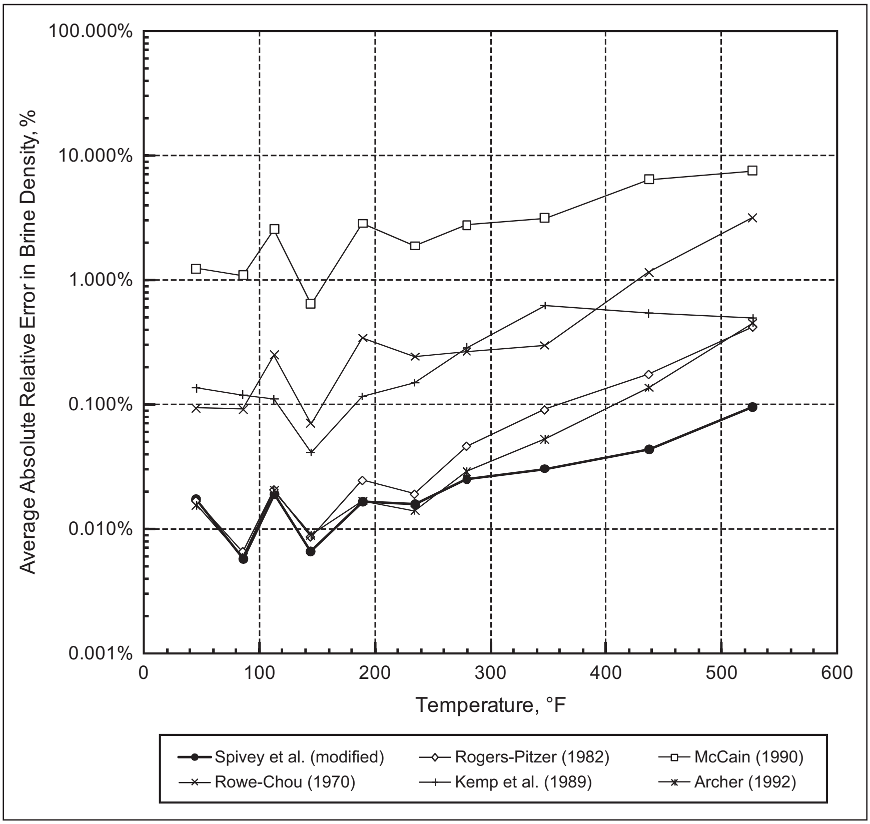

Fig. 4–4. Average absolute relative error in brine density as a function of temperature for various correlations, based on the brine density database in table 4–1

Figure 4–4 shows the average absolute relative error (AARE) in density as a function of temperature for several density correlations, for data in the brine density database in table 4–1. Because the error for different correlations varies by as much as a factor of 100, the AARE is shown on a logarithmic scale, and the average relative error (ARE) is not shown. The modified Spivey et al. correlation performs as well as or better than the other correlations throughout the range of temperatures in the database. Below 302°F (150°C), the Rogers-Pitzer and Archer equations of state perform almost as well as the modified Spivey et al. correlation. Above 302°F (150°C), the modified Spivey et al. correlation performs significantly better than either the Rogers-Pitzer or the Archer equation of state.

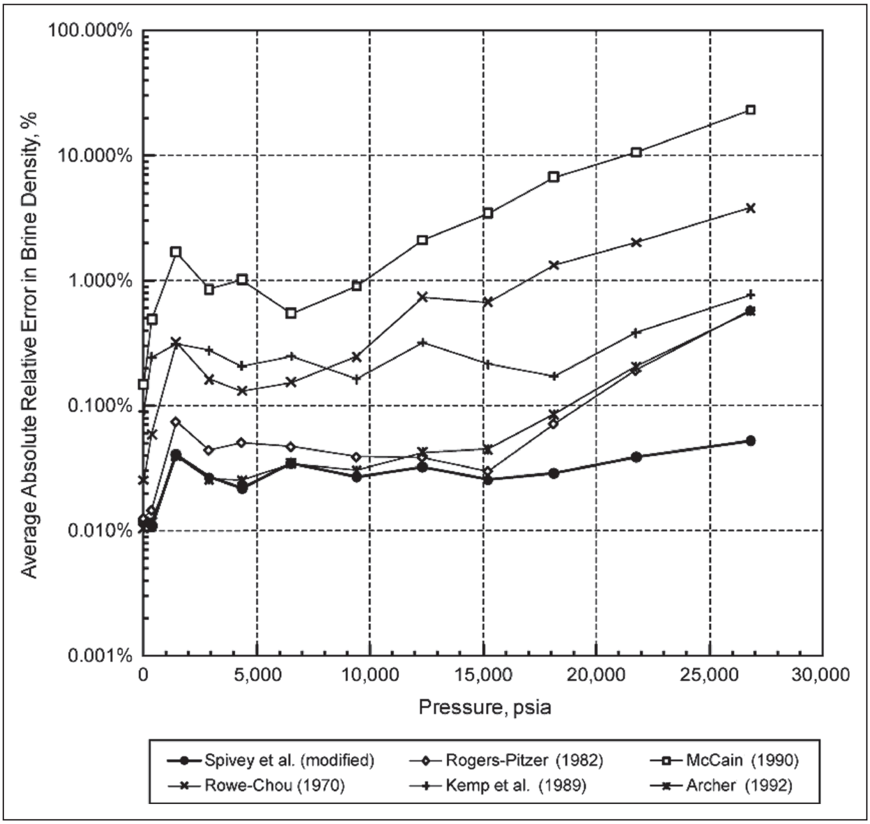

Fig. 4–5. Average absolute relative error in brine density as a function of pressure for various correlations, based on the brine density database in table 4–1

Figure 4–5 shows the AARE in brine density as a function of pressure for data in the brine density database in table 4–1. Again, the modified Spivey et al. correlation performs as well as or better than any of the other correlations. The Rogers-Pitzer and Archer equations of state perform well at pressures below 14,500 psia (100 MPa) but exhibit increasingly large errors as pressure approaches 29,000 psia (200 MPa). The AARE for the modified Spivey et al. correlation (0.052%) is more than an order of magnitude lower than that for the Rogers-Pitzer (0.58%) and Archer (0.57%) equations of state for pressures approaching 29,000 psia (200 MPa).

Figure 4–6 shows the AARE in density as a function of sodium chloride content for data in the brine density database in table 4–1. The Rogers-Pitzer and Archer equations of state have slightly lower values of AARE for salinities less than 0.5 wt%. Otherwise, the modified Spivey et al. correlation performs as well as or better than all the other correlations throughout the range of salinities in the database.

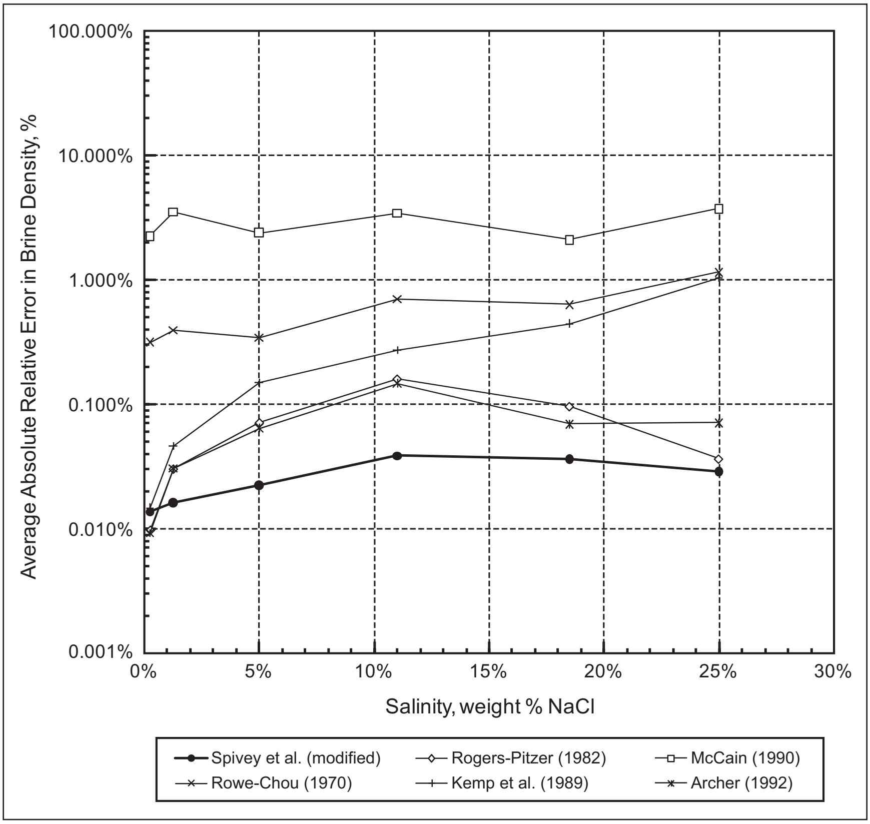

Fig. 4–6. Average absolute relative error in density as a function of salinity for various correlations, based on the brine density database in table 4–1

Coefficient of Isothermal Compressibility of Brine

The coefficient of isothermal compressibility of brine is often called simply *brine compressibility*.

Brine compressibility database

The brine compressibility database is a subset of the brine volumetric properties database, comprising only those data sets where the coefficient of isothermal compressibility of brine (brine compressibility) was measured and reported separately. Most of the data are from one of two sources: Gibson and Loeffler, and Rowe and Chou.[10]

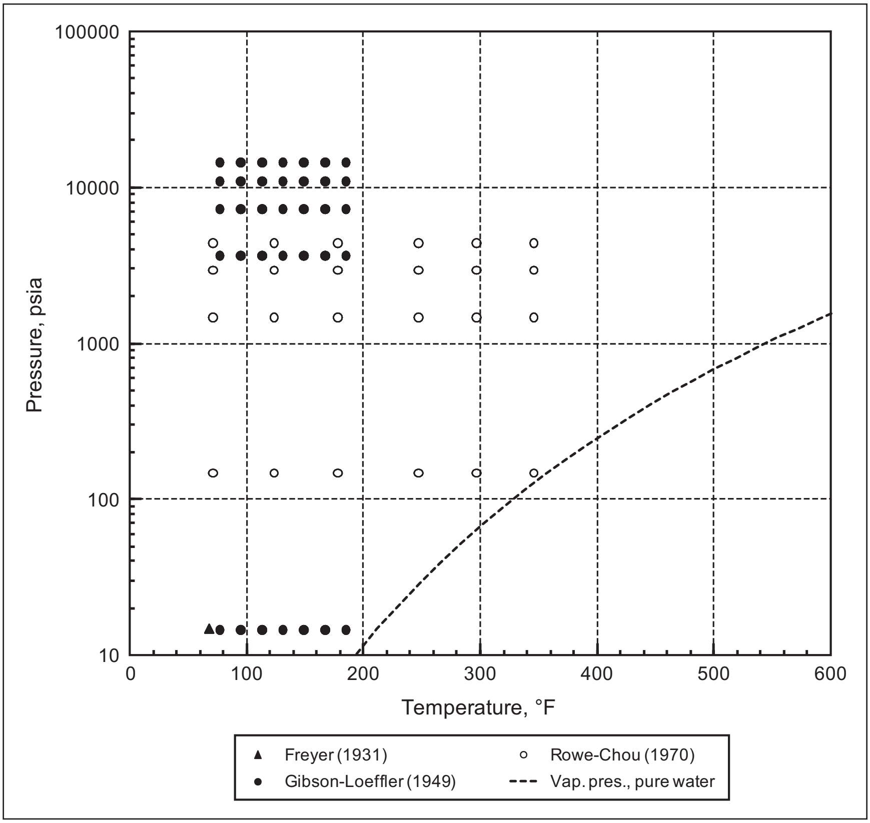

Fig. 4–7. Distribution of data points in the brine compressibility database, pressure vs. temperature

Figure 4–7 shows the distribution of data points in the brine compressibility database by pressure and temperature. Compared to the complete brine density database, the compressibility database has only a limited range of data. The Gibson-Loeffler data set includes compressibility data for pressures up to 14,500 psia (100 MPa), but only for temperatures up to 185°F (85°C).[11] The Rowe-Chou data set, on the other hand, includes compressibility data for temperatures up to 347°F (175°C), but only includes pressures up to 4,930 psia (34 MPa).[12]

Figure 4–8 shows the distribution of data points in the compressibility database by salinity and temperature. The database has good coverage over the range of salinities from 0 to 25 wt% sodium chloride, but the coverage in temperature is limited.

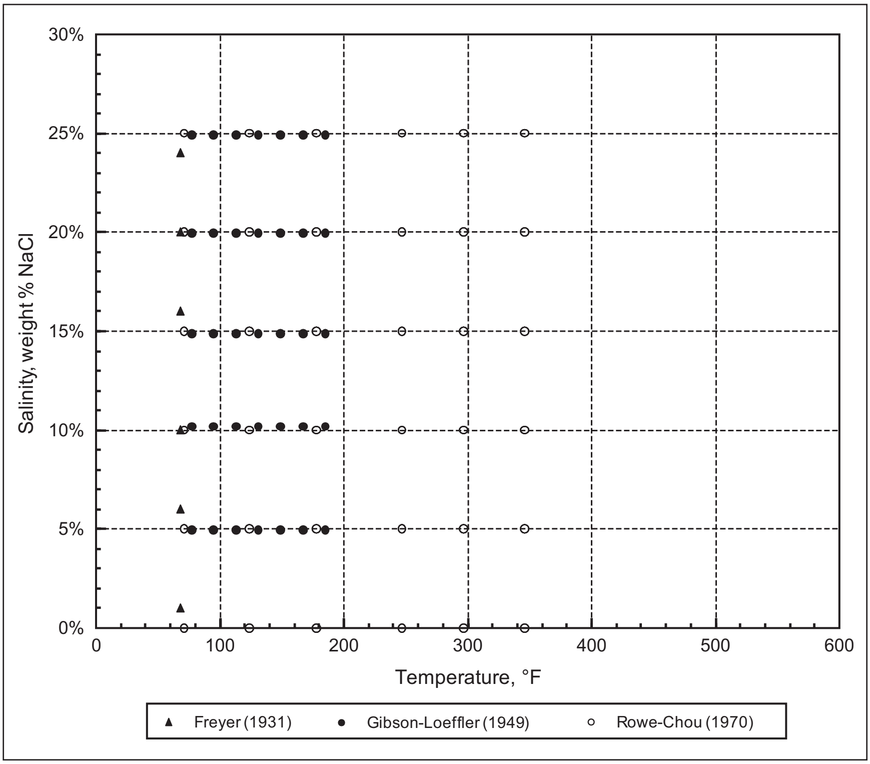

Fig. 4–8. Distribution of data points in the brine compressibility database, salinity vs. temperature

Figure 4–9 shows the distribution of data points in the brine compressibility database by pressure and salinity. The database has good coverage over the range of salinities from 0 to 25 wt% sodium chloride, but the coverage in pressure is limited.

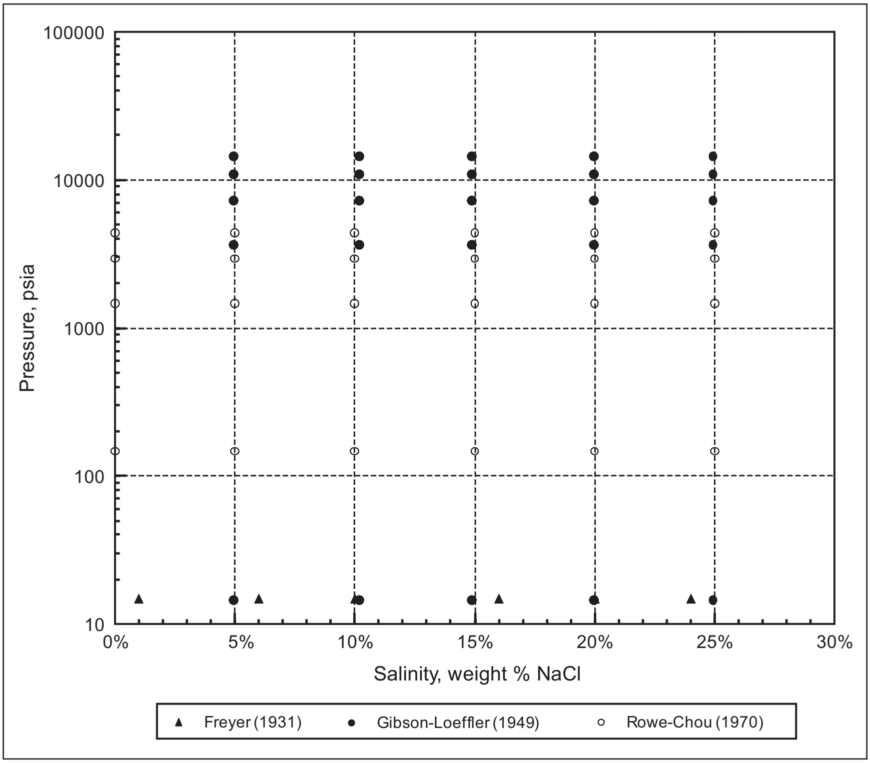

Fig. 4–9. Distribution of data points in the brine compressibility database, pressure vs. salinity

Comparison of correlations of brine compressibility

Five methods for estimating brine compressibility were chosen for evaluation against the compressibility database. Two of these methods, one by Osif, the other by Rowe and Chou, were correlations developed specifically for estimating compressibility. The third and fourth methods use the Rogers-Pitzer and Archer equations of state, respectively, to calculate the brine compressibility. The fifth method uses the modified Spivey et al. density correlation to calculate the brine compressibility.

Osif presented a correlation for compressibility that covered a relatively wide range of pressures, from 1,000 to 20,000 psia (6.9 to 138 MPa), but a limited range of temperatures, from 200°F to 270°F (93.3°C to 132°C).[13] Osif warned that his correlation equation should not be used outside of this range of temperatures. Osif's data likewise covered a limited range of salinities, from 0 to 200 g/L (0 to 17.7 wt%, 0 to 3.67 g-mol/kg H_2O).

Rowe and Chou presented a correlation for brine compressibility.[14] The Rowe and Chou compressibility correlation was developed for temperatures from 32°F to 347°F (0°C to 175°C), pressures from 14.5 to 4,930 psia (0.1 to 34 MPa), and sodium chloride concentrations from 0 to 25 wt% (0 to 5.7 g-mol/kg H_2O).

Table 4–3 shows the results of the evaluation of correlations for brine compressibility.

Table 4–3. Results of evaluation of brine compressibility correlations against brine compressibility data subset of brine density database in table

Correlation	ARE	AARE
Spivey et al. (modified)	0.16%	1.23%
Archer (1992)	0.23%	1.33%
Osif (1988)*	–0.1%	1.57%
Rogers-Pitzer (1982)	–0.55%	2.19%
Osif (1988)	2.42%	2.97%
Rowe-Chou (1970)	8.27%	9.81%

*Temperature range restricted to 200°F to 270°F

Figure 4–10 shows the AARE in brine compressibility as a function of temperature for several correlations. As in the case of brine density, because of the range of the magnitude of error, the ARE is not shown. The modified Spivey et al. brine compressibility correlation is the best correlation across the entire temperature range. The Rogers-Pitzer and Archer equations of state, as well as the modified Spivey et al. correlation, show a much lower AARE in compressibility for temperatures between 104°F and 122°F (40°C and 50°C). This is due to the fact that only the higher quality Gibson-Loeffler data set contains brine compressibility data points within this range of temperatures. Although Osif recommended his correlation be used only for temperatures from 200°F to 270°F (93.3°C to 132°C), the Osif correlation gives excellent results for temperatures from 122°F to 302°F (50°C to 150°C), with a maximum AARE of 2.05%. For temperatures below 122°F (50°C), the Osif correlation gives an AARE that approaches 10%.

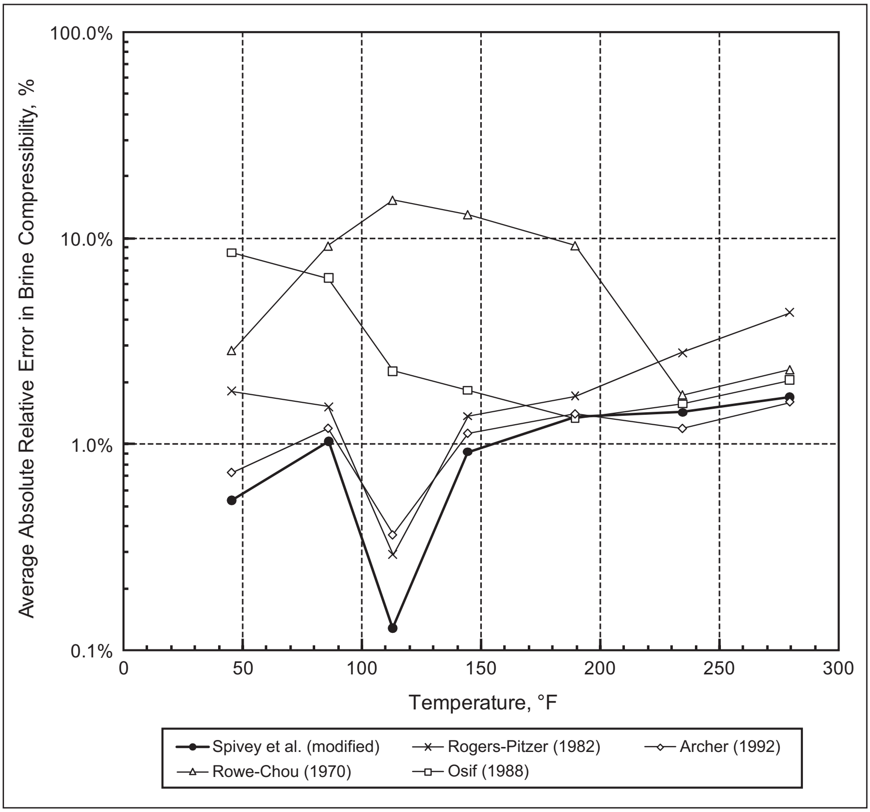

Fig. 4–10. Average absolute relative error in brine compressibility as a function of temperature for various correlations, based on the compressibility data subset of the brine density database in table 4–1

Figure 4–11 shows the AARE in brine compressibility as a function of pressure. Only the high-quality Gibson-Loeffler data set contains brine compressibility data for pressures higher than 5,000 psia. Thus, the AARE for pressures lower than 5,000 psia reflects lower quality data rather than accuracy of the correlations.

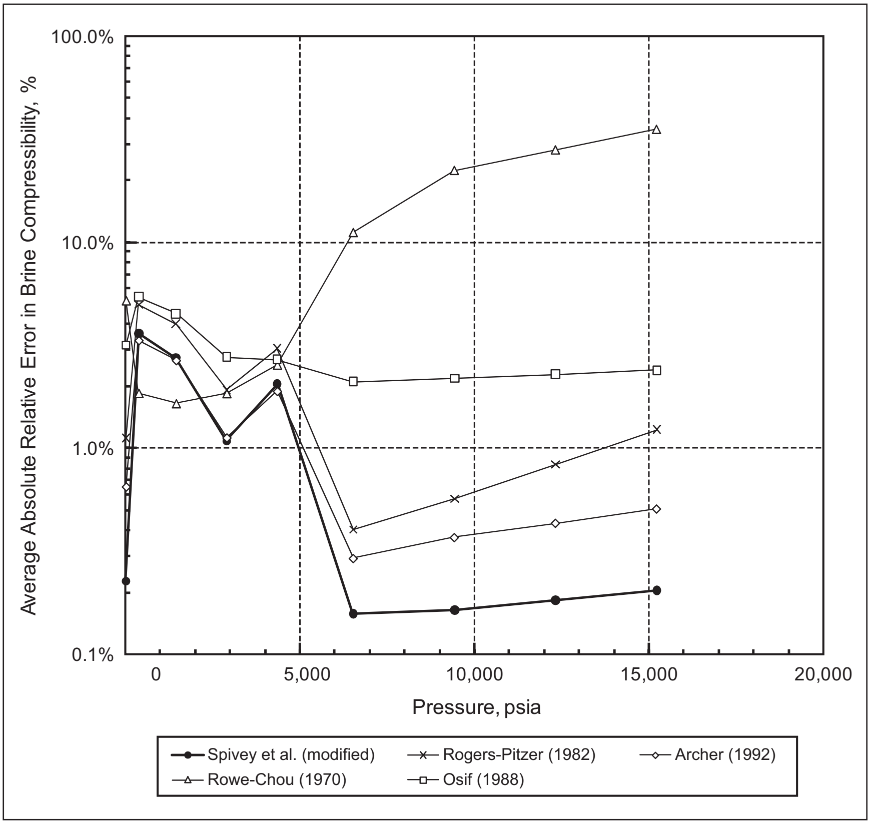

Fig. 4–11. Average absolute relative error in brine compressibility as a function of pressure for various correlations, based on the compressibility data subset of the brine density database in table 4–1

Figure 4–12 shows the AARE in brine compressibility as a function of salinity for several compressibility correlations. Again, the modified Spivey et al. correlation shows a lower AARE across the entire range of salinities in the database.

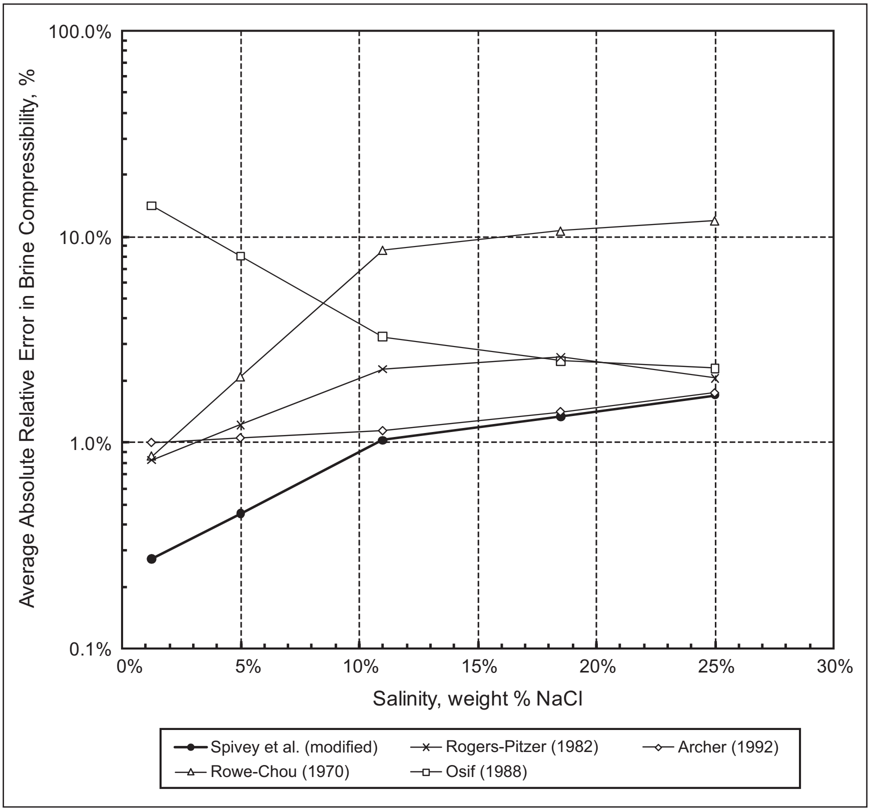

Fig. 4–12. Average absolute relative error in brine compressibility as a function of salinity for various correlations, based on the compressibility data subset of the brine density database in table 4–1

Methane Solubility in Sodium Chloride Brines

Methane solubility in sodium chloride brines database

Many studies of methane solubility in sodium chloride brines at high pressures and high temperatures have been published over the years.

Spivey et al. compiled a database of published data on solubility of methane in pure water and sodium chloride brines for temperatures from 32°F to 680°F (0 to 360°C), pressures from 50 to 28,600 psia (0.34 to 197 MPa), and sodium chloride concentrations from 0 to 26.3 wt% (0 to 6.1 g-mol/kg H_2O).[15] They then used this database to develop a correlation for estimating solubility of methane in sodium chloride brines.

Table 4–4. Methane solubility database

Reference	n	Temperature range (°C)	Pressure range (MPa)	NaCl salinity range (g-mol/kg H_2O)	Methane solubility range (g-mol/kg H_2O)
Culberson-McKetta (1951)[a]	70/71	25–171	2.2–69	0	0.018–0.43
O'Sullivan-Smith (1970)[b]	75/79	51–125	10–62	0–4	0.047–0.24
Sultanov et al. (1972)[c]	35/71	150–250	4.9–108	0	0.045–1.36
Amirijafari-Campbell (1972)[d]	8/8	38–71	4–34.5	0	0.033–0.187
Price (1979)[e]	51/71	154–292	5.2–192	0	0.070–2.56
Price et al. (1981)[f]	29/29	24–72	0.9–14	0.8–3.0	0.004–0.08
McGee et al. (1981)	30/30	191–292	10–18	0–4.3	0.14–0.23
Crovetto et al. (1982)[g]	5/7	24–157	1.3–2.2	0	0.012–0.024
Blount et al. (1979) (1982)	655/670	99–240	13–157	0–5.7	0.037–1.12
Dhima et al. (1998)[h]	4/4	71	20–100	0	0.12–0.28
Keipe et al. (2003)[i]	26/26	40–100	0.34–9.3	0	0.0028–0.086
Combined	**988/1,066**	**24–292**	**0.34–192**	**0–5.7**	**0.0028–2.56**

Note: n = number of points used in fit/number of points in database. [a] Culberson, O.L. and McKetta, J.J., Jr. "Phase Equilibria in Hydrocarbon-Water Systems III–The Solubility of Methane in Water at Pressures to 10,000 psia." *Transactions*. AIME. Vol. 192 (1951) 223–226. [b] O'Sullivan, T.D. and Smith, N.O. "The Solubility and Partial Molar Volume of Nitrogen and Methane in Water and Aqueous Sodium Chloride from 50 to 125°C and 100 to 600 Atm." *Journal of Physical Chemistry*. Vol. 74, No. 7 (April 2, 1970) 1,460–1,466. [c] Sultanov, R.C., Skripka, V.E., and Namiot, A.Y. "Solubility of Methane in Water at High Temperatures and Pressures." *Gazova Promyshlennost*. Vol. 17 (May 1972) 6–7. [d] Amirijafari, B. and Campbell, J.M. "Solubility of Gaseous Hydrocarbon Mixtures in Water." *Society of Petroleum Engineers Journal*. Vol. 12, No. 1 (February 1972) 21–27. [e] Price, L.C. "Aqueous Solubility of Methane at Elevated Pressures and Temperatures." *AAPG Bulletin*. Vol. 63 (1979) 1,527–1,533. [f] Price, L.C., Blount, C.W., MacGowan, D., and Wenger, L. "Methane Solubility in Brines with Application to the Geopressured Resource." *Proceedings of the 5th Geopressured-Geothermal Energy Conference*. Vol. 5 (1981) 205–214. [g] Crovetto, R., Fernández-Prini, R., and Japas, M.L. "Solubilities of Inert Gases and Methane in H_2O and D_2O in the Temperature Range of 300 to 600 K." *Journal of Chemical Physics*. Vol. 76, No. 2 (January 15, 1982) 1,077–1,086. [h] Dhima, A., de Hemptinne, J.-C., and Moracchini, G. "Solubility of Light Hydrocarbons and Their Mixtures in Pure Water under High Pressure." *Fluid Phase Equilibria*. Vol. 145 (1998) 129–150. [i] Kiepe, J., Horstmann, S., Fischer, K., and Gmehling, J. "Experimental Determination and Prediction of Gas Solubility Data for Methane+Water Solutions Containing Different Monovalent Electrolytes." *Industrial and Engineering Chemical Research*. Vol. 42 (2003) 5,392–5,398.

Several data sets were included in the Spivey et al. database, yet were not used in developing the Spivey et al. correlation of methane solubility in brine. The data summarized in table 4–4 were used to evaluate selected methane solubility in brine correlations.

Figure 4–13 shows the distribution of points in the database of methane solubility in water and sodium chloride brine by pressure and temperature. As with the brine density database, the database includes only data points for pressures higher than the vapor pressure for pure water at any given temperature.

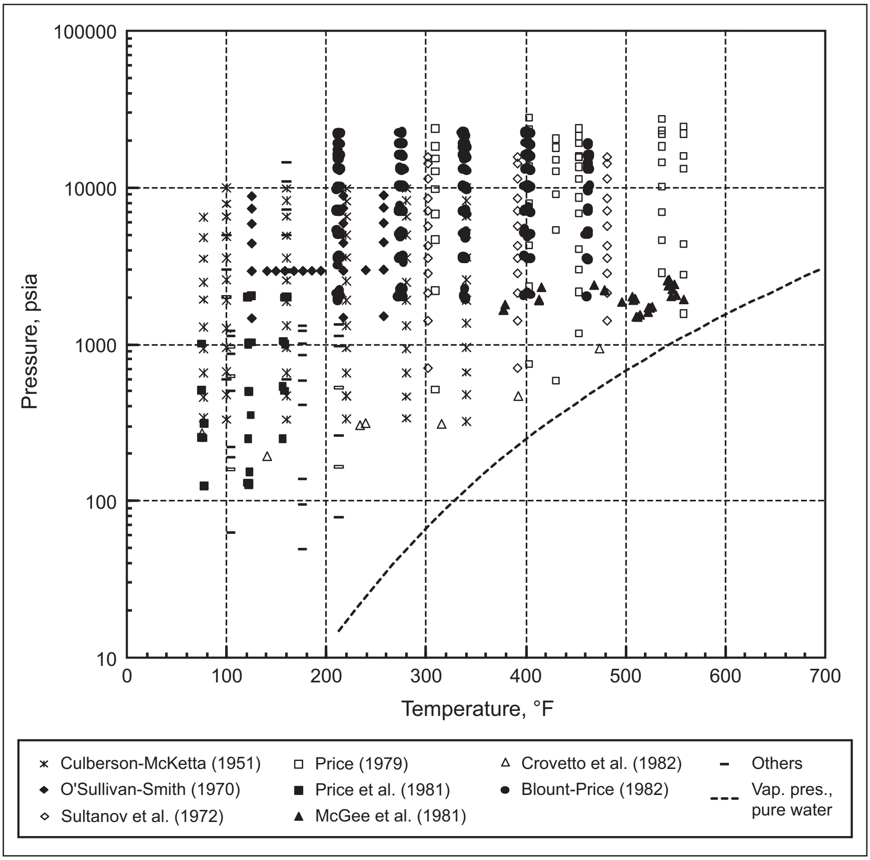

Fig. 4–13. Distribution of data points in the database of methane solubility in water and sodium chloride brine, pressure vs. temperature

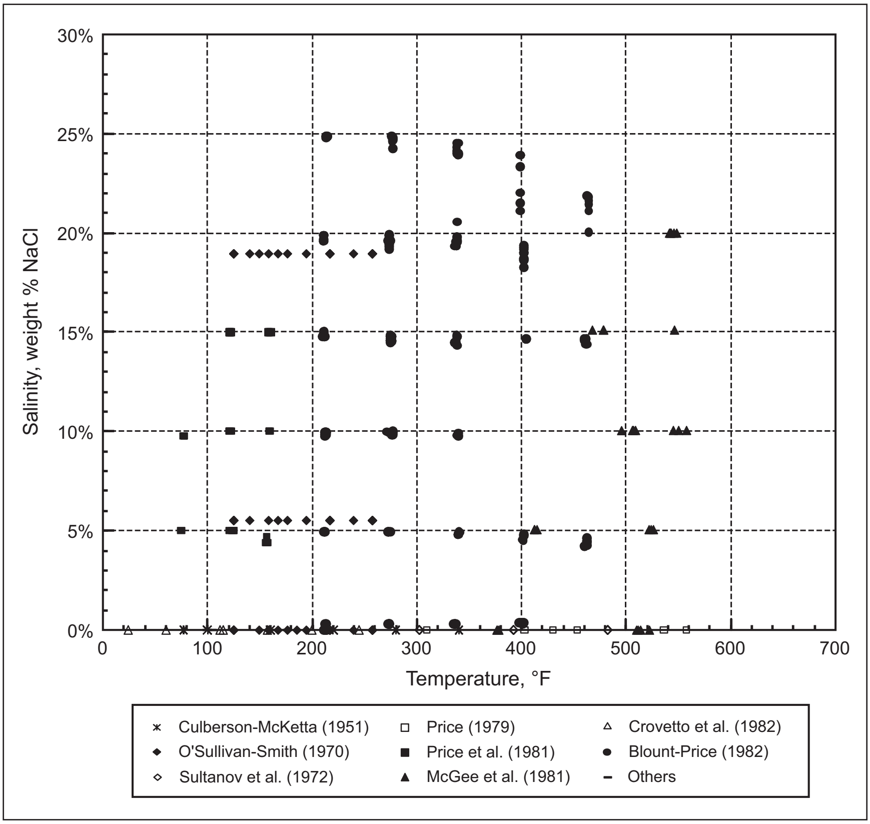

Fig. 4–14. Distribution of data points in the database of methane solubility in water and sodium chloride brine, salinity vs. temperature

Figure 4–14 shows the distribution of data in the database of methane solubility in water and sodium chloride brine by salinity and temperature. Note that the solubility database contains very little data for sodium chloride brine at temperatures higher than 527°F (250°C).

Figure 4–15 shows the distribution of data in the database of methane solubility in water and sodium chloride brine by pressure and salinity.

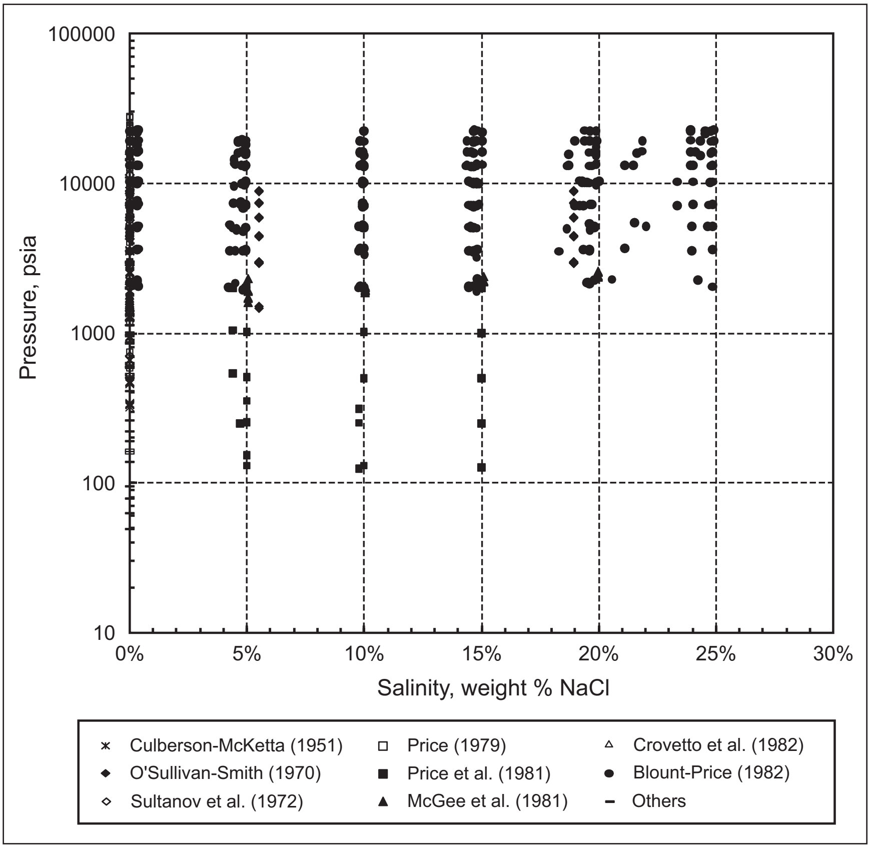

Fig. 4–15. Distribution of data points in database of methane solubility in water and sodium chloride brine, pressure vs. salinity

Comparison of methane solubility correlations

Six methane solubility correlations were chosen for evaluation against the methane solubility database.

McCain developed a set of polynomial equations in pressure and temperature to reproduce an earlier graphical correlation for methane solubility in pure water.[16] McCain also developed an equation to reproduce a graphical correlation for the effect of salinity on solubility of natural gas in water. As with any polynomial fit, McCain's correlation should not be used outside the range of data on which it is based: pressures from 1,000 to 10,000 psia (6.9 MPa to 69 MPa) and temperatures from 80°F to 340°F (26.7°C to 171.1°C).

McGee et al. measured solubility of methane in water and in sodium chloride brines for various temperatures and pressures.[17] They also presented a correlation to predict methane solubility as a function of temperature, pressure, and salinity for temperatures up to 662°F (350°C), pressures to 20,000 psia (137.9 MPa), and salinity to 25 wt% sodium chloride (5.7 g-mol/kg H_2O).

In 1992, Duan et al. proposed a method for predicting the solubility of methane in brines at temperatures from 32°F to 482°F (0°C to 250°C), pressures from 0 to 23,200 psia (0 to 1,600 bar), and sodium chloride concentration from 0 to 26 wt% (0 to 6 g-mol/kg H_2O).[18] Duan et al. claim that the method is able to predict the solubility of methane in sodium chloride brine with an estimated uncertainty of 8% throughout the ranges of temperature, pressure, and sodium chloride concentration listed above. Duan et al. also presented an equation for estimating the partial molar volume of methane in brine. Unfortunately, Duan et al. relied on a large data set presented by Blount et al. that was later found to contain a systematic error from incorrect data reduction.[19] Blount and Price published the corrected data in 1982.[20]

Spivey et al. developed an empirical correlation to estimate solubility of methane in sodium chloride brines for temperatures from 68°F to 680°F (20°C to 360°C), pressures from 130 to 29,000 psia (0.9 to 200 MPa), and salinities from 0 to 26 wt% sodium chloride (0 to 6 g-mol/kg H_2O).[21] Unfortunately, the Spivey et al. database used an invalid conversion for methane solubility for the Blount and Price data set.

In 2006, Duan and Mao updated the earlier Duan et al. correlation to use the corrected data from Blount and Price, as well as additional data.[22] They claimed the new model predicted methane solubilities in sodium chloride brines with an average accuracy of 6% over ranges of temperature from 32°F to 482°F (0°C to 250°C), pressure from 14.7 to 29,000 psia (0.1 to 200 MPa), and salinity from 0 to 26 wt% sodium chloride (0 to 6 g-mol/kg H_2O).

Although the Duan-Mao correlation gives accurate estimates of methane solubility, the calculation of the solubility of methane in pure water using the Duan-Mao method is quite complex. Accordingly, the authors have modified the Spivey et al. correlation to reflect the corrected conversion of the Blount and Price data set. The modified correlation gives accuracy comparable to that of the Duan-Mao method.

Table 4–5 shows the results of the evaluation of correlations for solubility of methane in pure water and sodium chloride brines.

Table 4–5. Results of evaluation of methane solubility correlations against methane solubility database in table 4

Correlation	ARE	AARE
Spivey et al. (modified)	–0.08%	5.90%
Duan-Mao (2006)	–1.32%	6.41%
Duan et al. (1992)	–5.86%	8.89%
Spivey et al. (2004)	–8.01%	10.42%
McGee et al. (1981)	–23.93%	26.86%
McCain (1990)	–15.64%	32.97%

Figure 4–16 shows the ARE and AARE in methane solubility in sodium chloride brine as functions of temperature. At temperatures below 100°F (37.8°C), the Duan et al. (1992) correlation gives the lowest AARE, with the Duan-Mao (2006) and modified Spivey et al. correlations both giving an AARE slightly higher than 2%. For temperatures from 100°F to 450°F (37.8°C to 232.2°C), the Duan-Mao (2006) correlation and the modified Spivey et al. correlation give very similar results. Between 450°F and 572°F (232.2°C and 300°C), the modified Spivey et al. correlation gives an AARE lower than any of the other correlations. The Duan et al. (1992) and Spivey et al. (2004) correlations perform significantly worse than either the Duan-Mao (2006) correlation or the modified Spivey et al. correlation.

Figure 4–17 shows the ARE and AARE in methane solubility in sodium chloride brine as functions of pressure. The modified Spivey et al. correlation has a consistent AARE of 5% to 7% throughout the range of pressures in the database. At pressures below 2,900 psia (20 MPa), the Duan et al. (1992) and Duan-Mao (2006) correlations perform slightly better than the modified Spivey et al. correlation. The modified Spivey et al. correlation performs slightly better than the Duan-Mao (2006) correlation for pressures from 14,500 to 29,000 psia (100 to 200 MPa).

Figure 4–18 shows the ARE and AARE in methane solubility as functions of salinity. The modified Spivey et al. correlation has a consistent AARE of 5% to 7% throughout the range of salinities in the database. For salinities below 12.5 wt% sodium chloride, the Duan-Mao (2006) correlation and the modified Spivey et al. correlation give very similar results. For salinities above 12.5 wt% sodium chloride, the modified Spivey et al. correlation gives slightly lower AARE than does the Duan-Mao correlation. The Duan et al. (1992) and Spivey et al. (2004) correlations give an AARE that increases with increasing salinity.

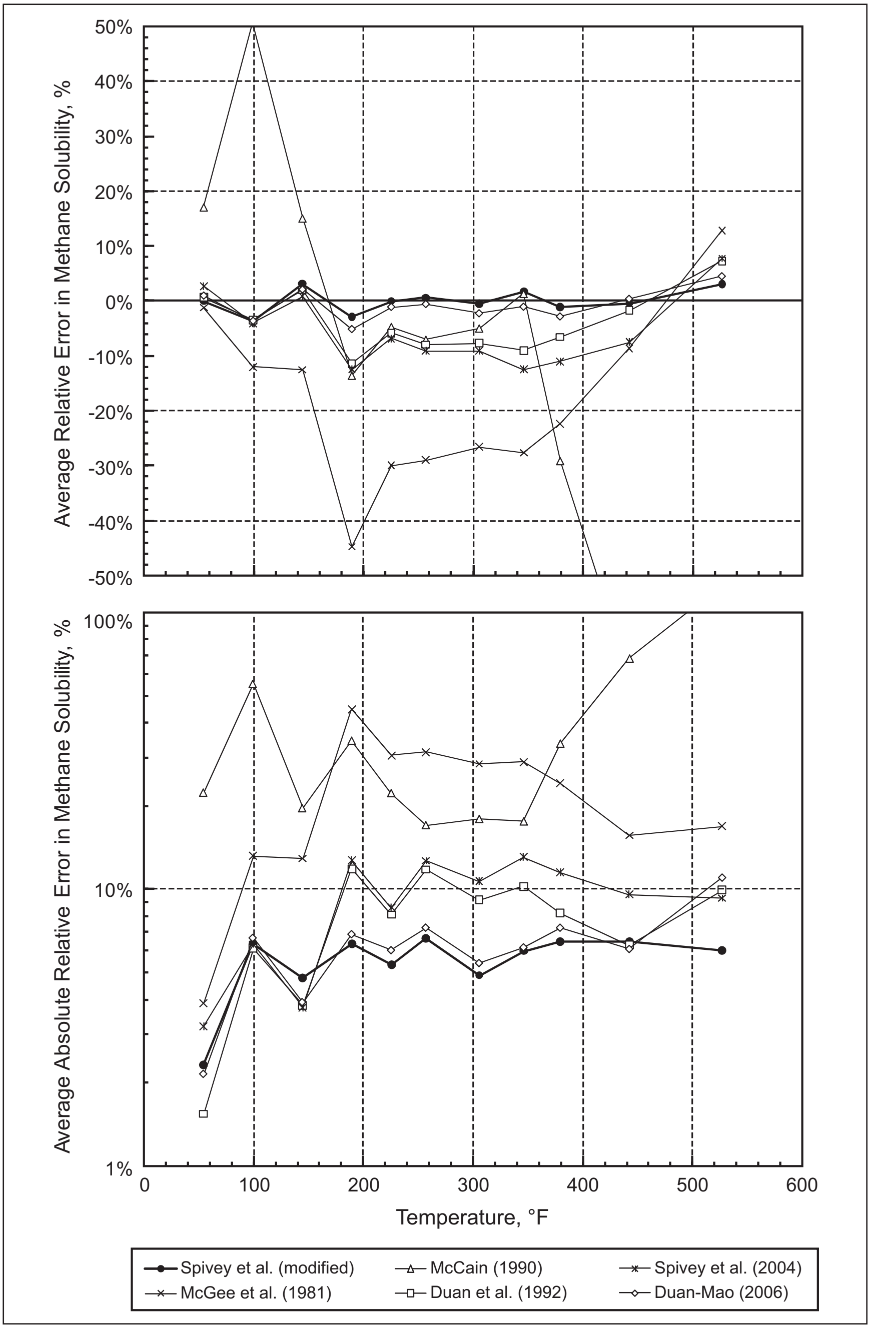

Fig. 4–16. Average relative error and average absolute relative error in methane solubility in sodium chloride brine as functions of temperature for various correlations, based on the methane solubility database in table 4–4

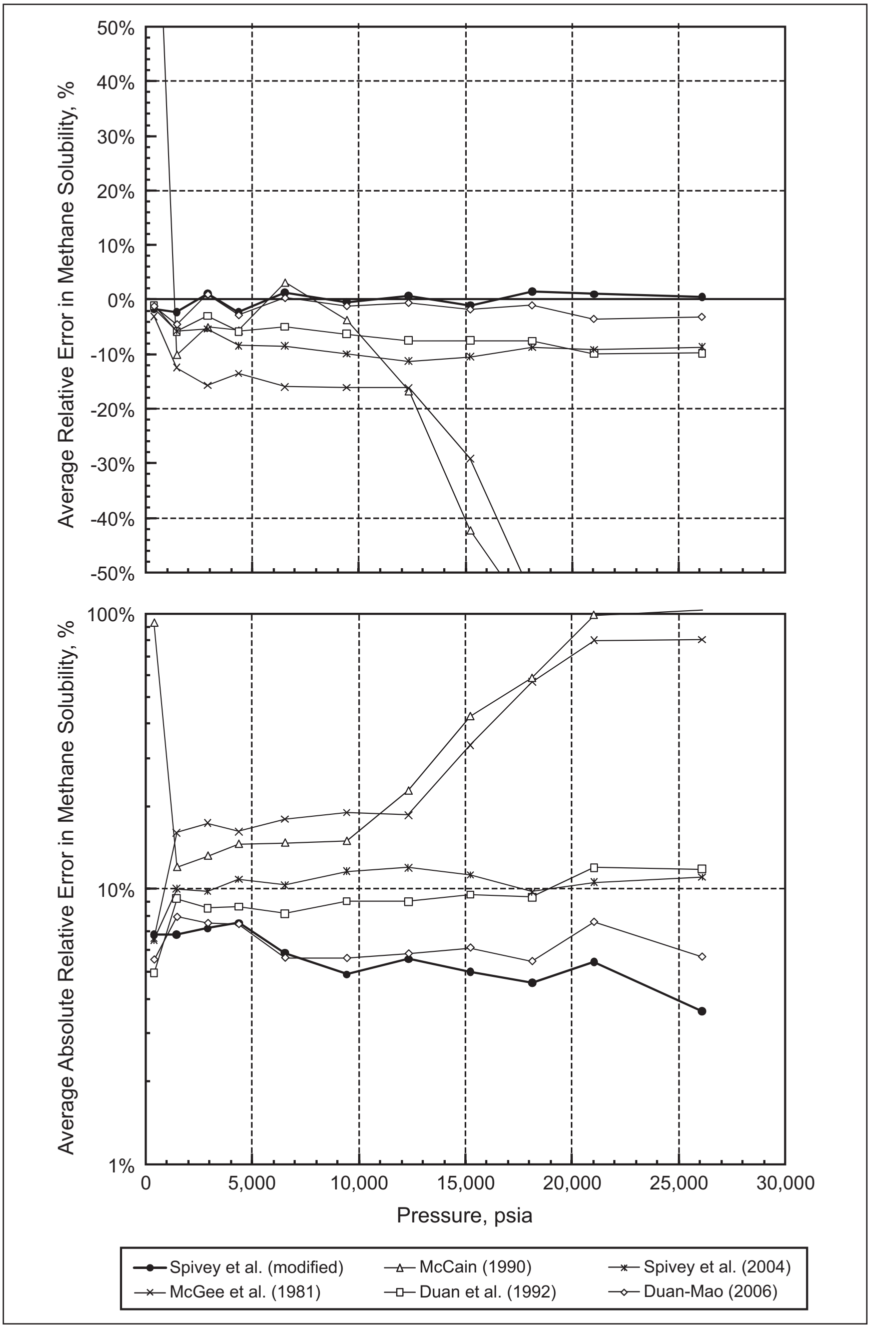

Fig. 4–17. Average relative error and average absolute relative error in methane solubility in sodium chloride brine as functions of pressure for various correlations, based on the methane solubility database in table 4–4

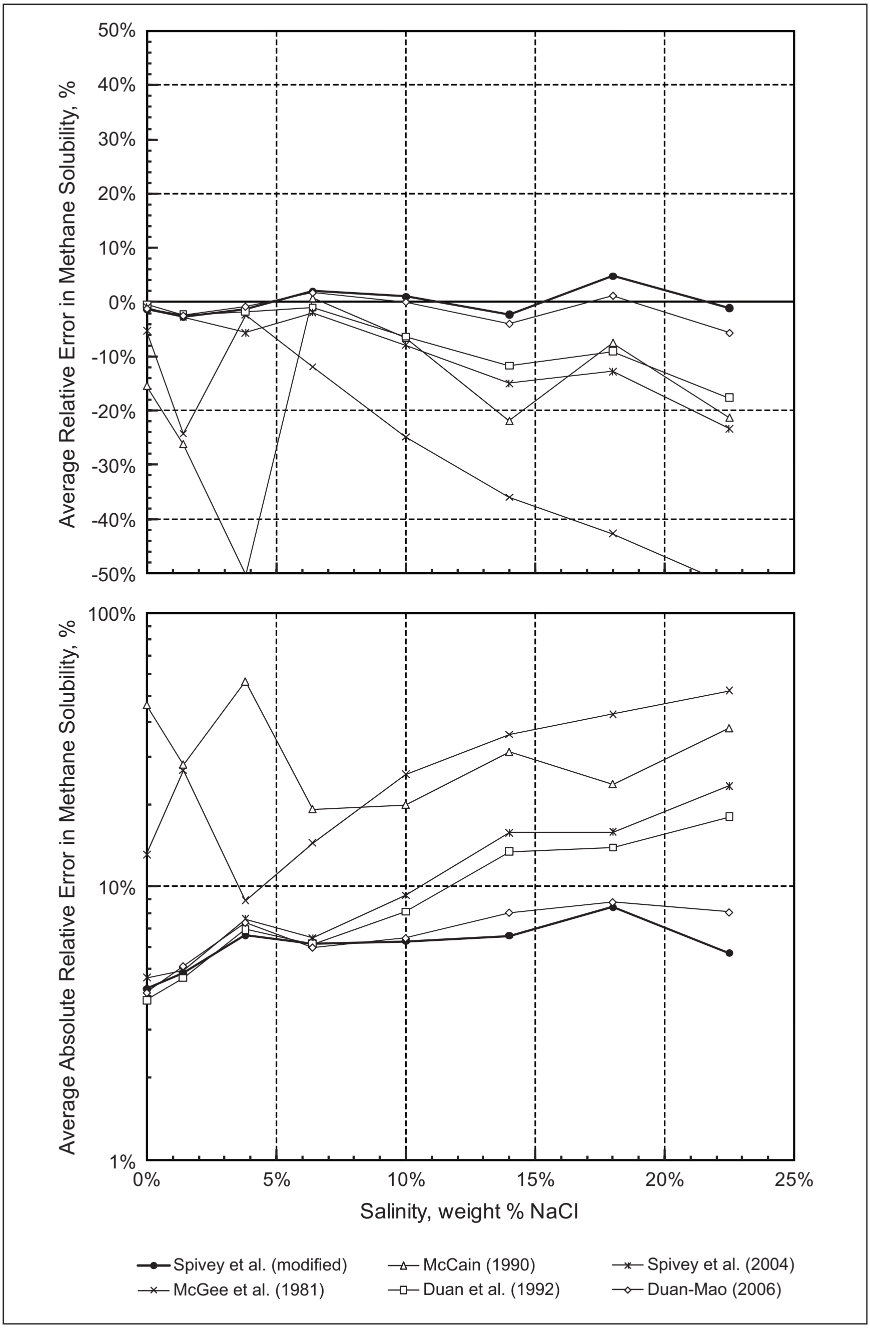

Fig. 4–18. Average relative error and average absolute relative error in methane solubility in sodium chloride brine as functions of salinity for various correlations, based on the methane solubility database in table 4–4

Computing Properties of Oilfield Brines

The modified Spivey et al. correlations are recommended for calculating properties of oilfield brines. This section shows how to calculate properties of oilfield brines using the recommended correlations.

Units. Spivey et al. used metric units in developing their brine density and methane solubility correlations. For all of the equations in this section (unless otherwise noted) temperature, T, is in °C, pressure, p, is in MPa, and salinity (sodium chloride concentration), m, is in g-mol/kg H_2O. (For clarity, the salinity is explicitly denoted with a subscript, m_{NaCl}, in equations for brine containing dissolved methane.) Density, ρ, is in g/cm^3; specific volume, v, is in cm^3/g; compressibility, c_w, is in MPa^{-1}; and methane solubility, m_{CH4}, is in g-mol/kg H_2O. Formation volume factor, B_w, is in cm^3/cm^3, and solution gas-water ratio, R_{sw}, is in cm^3/cm^3. Unit conversions between metric and oilfield units are given in a separate section below.

For most quantities, the subscripts w and b indicate properties of pure water and brine, respectively. Because of long-established practice, the superscript w is also used for certain properties of oilfield brines of particular interest in the oilfield. These properties include compressibility [equations (4.32) and (4.35)], formation volume factor [equation (4.36)], and solution gas-water ratio [equation (4.37)].

Temperature-dependent coefficients. In developing the brine density, brine compressibility, and methane solubility in brine correlations, Spivey et al. used a rational function in temperature, of the form

$$a(T) = \frac{a_1\left(\frac{T}{100}\right)^2 + a_2\left(\frac{T}{100}\right) + a_3}{a_4\left(\frac{T}{100}\right)^2 + a_5\left(\frac{T}{100}\right) + 1} \tag{4.1}$$

as a general correlation equation for temperature-dependent parameters. The coefficients a_i for use in equation (4.1) are tabulated below for various temperature-dependent parameters.

Density and compressibility of methane-free brine

Step 1. First, calculate the density of pure water at the reference pressure of 70 MPa, $\rho_w(T, 70\text{ MPa})$, in g/cm^3, at the temperature of interest, T, using equation (4.1), along with the coefficients given in table 4–6.

Table 4–6. Coefficients for equation of state for pure water for use with equation (4.1)

a_i	$\rho_w(T, 70\text{ MPa})$	$E_w(T)$	$F_w(T)$
a_1	–0.127213	4.221	–11.403
a_2	0.645486	–3.478	29.932
a_3	1.03265	6.221	27.952
a_4	–0.070291	0.5182	0.20684
a_5	0.639589	–0.4405	0.3768

Next, calculate the values of the pure water compressibility coefficients $E_w(T)$ and $F_w(T)$, from equation (4.1), with the coefficients given in table 4–6. The coefficients in table 4–6 are the same as in the original Spivey et al. correlations.

Density and compressibility of pure water. Calculate compressibility and density for pure water at the temperature and pressure of interest. Skip steps 2 and 3 if only properties for brine are required.

Step 2. Calculate the compressibility $c_w(T, p)$, in MPa^{-1}, of pure water at temperature T and pressure p, from equation (4.2).

$$c_w(T,p) = \left(\frac{1}{70}\right)\frac{1}{E_w(T)\left(\frac{p}{70}\right) + F_w(T)} \tag{4.2}$$

Step 3. Calculate the density of pure water at temperature T and pressure p.

First, calculate the values of $I_w(T, 70\text{ MPa})$ and $I_w(T, p)$ from equations (4.3) and (4.4).

$$I_w(T,70\,\text{MPa}) = \frac{1}{E_w(T)}\ln\left|E_w(T) + F_w(T)\right| \tag{4.3}$$

$$I_w(T,p) = \frac{1}{E_w(T)}\ln\left|E_w(T)\left(\frac{p}{70}\right) + F_w(T)\right| \tag{4.4}$$

Next, calculate the density of pure water $\rho_w(T, p)$ in g/cm^3 from equation (4.5).

$$\rho_w(T,p) = \rho_w(T,70\text{ MPa})\,\text{EXP}\left[I_w(T,p) - I_w(T,70\text{ MPa})\right] \tag{4.5}$$

Density and compressibility of methane-free brine. Calculate the density and compressibility of brine at the temperature, T, pressure, p, and sodium chloride concentration, m, of interest.

Step 4. Calculate the density of brine at the temperature of interest T and the reference pressure p_0 of 70 MPa, with a sodium chloride concentration m, as follows.

First, calculate the brine density coefficients $D_{m,2}(T)$, $D_{m,3/2}(\mathbf{T})$, $D_{m,1}(T)$, and $D_{m,1/2}(T)$ using equation (4.1) with the coefficients given in table 4–7. The coefficients in table 4–7 have been updated from those in the original Spivey et al. correlation.

Table 4–7. Coefficients for density of brine as a function of temperature at 70 MPa (10,153 psia) for use with equation (4.1)

a_i	$D_{m,2}(T)$	$D_{m,3/2}(T)$	$D_{m,1}(T)$	$D_{m,1/2}(T)$
a_1	-1.1149×10^{-4}	-8.878×10^{-4}	2.1466×10^{-3}	2.356×10^{-4}
a_2	1.7105×10^{-4}	-1.388×10^{-4}	1.2427×10^{-2}	-3.636×10^{-4}
a_3	-4.3766×10^{-4}	-2.96318×10^{-3}	4.2648×10^{-2}	-2.278×10^{-4}
a_4	0	0	-8.1009×10^{-2}	0
a_5	0	0.51103	0.525417	0

Next, calculate the density of brine at temperature T and the reference pressure of 70 MPa using equation (4.6).

$$\rho_b(T,70\text{ MPa},m) = \rho_w(T,70\text{ MPa}) + D_{m,2}(T)m^2 + D_{m,3/2}(T)m^{3/2} + D_{m,1}(T)m + D_{m,1/2}(T)m^{1/2} \tag{4.6}$$

Step 5. Calculate the brine compressibility coefficients $E_b(T,m)$ and $F_b(T,m)$ as follows.

First, calculate the coefficients $E_m(T)$, $F_{m,3/2}(T)$, $F_{m,1}$ (T), and $F_{m,1/2}(T)$ from equation (4.1), along with the coefficients given in table 4–8. The coefficients in table 4–8 have been updated from those in the original Spivey et al. correlation.

Table 4–8. Coefficients for brine compressibility for use with equation (4.1)

a_i	$E_m(T)$	$F_{m,3/2}(T)$	$F_{m,1}(T)$	$F_{m,1/2}(T)$
a_1	0	–0.617	0	0.0365
a_2	0	–0.747	9.917	–0.0369
a_3	0.1249	–0.4339	5.1128	0
a_4	0	0	0	0
a_5	0	10.26	3.892	0

Next, calculate the brine compressibility coefficients $E_b(T,m)$ and $F_b(T,m)$ from equations (4.7) and (4.8).

$$E_b(T,m) = E_w(T) + E_m(T)m \tag{4.7}$$

$$F_b(T,m) = F_w(T) + F_{m,3/2}(T)m^{3/2} + F_{m,1}(T)m + F_{m,1/2}(T)m^{1/2} \tag{4.8}$$

Step 6. Calculate the compressibility of methane-free brine in MPa^{-1}, at temperature T, pressure p, and salinity m, from equation (4.9).

$$c_b(T,p,m) = \left(\frac{1}{70}\right)\frac{1}{E_b(T,m)\left(\frac{p}{70}\right) + F_b(T,m)} \tag{4.9}$$

Step 7. Calculate the density of methane-free brine as follows.

First, calculate the values of $I_b(T,70\ \text{MPa},m)$ and $I_b(T,p,m)$ from equations (4.10) and (4.11).

$$I_b(T,70\,\text{MPa},m) = \frac{1}{E_b(T,m)}\ln\left|E_b(T,m) + F_b(T,m)\right| \tag{4.10}$$

$$I_b(T,p,m) = \frac{1}{E_b(T,m)}\ln\left|E_b(T,m)\left(\frac{p}{70}\right) + F_b(T,m)\right| \tag{4.11}$$

Next, calculate the methane-free brine density $\rho_b(T,p,m)$, in g/cm^3, from equation (4.12).

$$\rho_b(T,p,m) = \rho_b(T,70\,\text{MPa},m)\text{EXP}\left[I_b(T,p,m) - I_b(T,70\,\text{MPa},m)\right] \tag{4.12}$$

Methane solubility in sodium chloride brine

The solubility of methane in brine is estimated using the modified Spivey et al. correlation for methane solubility in brine as follows.

Step 8. First, calculate the vapor pressure of pure water, p_σ, from the IAWPS-95 formulation, given in equation (4.13).[23]

$$\ln\left(\frac{p_\sigma}{p_c}\right) = \frac{T_c}{T}\left(a_1\vartheta + a_2\vartheta^{1.5} + a_3\vartheta^3 + a_4\vartheta^{3.5} + a_5\vartheta^4 + a_6\vartheta^{7.5}\right), \tag{4.13}$$

where T is K, T_c = 647.096, K, p_c = 22.064 MPa, and ϑ is defined as

$$\vartheta = 1 - \frac{T}{T_c} \tag{4.14}$$

The coefficients for equation (4.13) are given in table 4–9.

Table 4–9. Coefficients for international standard water vapor-pressure equation, equation (4.13)

Coefficient	Value	Coefficient	Value
a_1	–7.85951783	a_4	22.6807411
a_2	1.84408259	a_5	–15.9618719
a_3	–11.7866497	a_6	1.80122502

Step 9. Calculate the solubility of methane in pure water at the temperature and pressure of interest.

First, calculate the methane solubility coefficients $A(T)$, $B(T)$, and $C(T)$ using equation (4.1), with the coefficients given in table 4–10.

Table 4–10. Coefficients for the modified Spivey et al. correlation for methane solubility in pure water for use in equation (4.1)

a_i	$A(T)$	$B(T)$	$C(T)$
a_1	0.0	–0.03602	0.6855
a_2	–0.004462	0.18917	–3.1992
a_3	–0.06763	0.97242	–3.7968
a_4	0	0	0.07711
a_5	0	0	0.2229

Next, calculate the solubility of methane in pure water from equation (4.15).

$$m_{CH_4,w} = \text{EXP}\left[A(T)\left\{\ln\left(p - p_\sigma\right)\right\}^2 + B(T)\ln\left(p - p_\sigma\right) + C(T)\right] \tag{4.15}$$

Step 10. Calculate the solubility of methane in brine at the temperature and pressure of interest, as follows.

Calculate coefficients $\lambda_{CH_4,Na}$ and $\zeta_{CH_4,NaCl}$ from equations (4.16) and (4.17), respectively, using the appropriate coefficients from table 4–11. The functional form for the coefficients $\lambda_{CH_4,Na}$ and $\zeta_{CH_4,NaCl}$ is similar to that used by Duan-Mao, but pressure is in MPa, and the pressure-squared term in equation (4.16) is independent of temperature.[24]

Note that temperature in equation (4.16) is Kelvin, and pressure is MPa.

$$\lambda_{CH_4,Na} = c_1 + c_2 T + \frac{c_3}{T} + c_6 p + c_{10} p^2 \tag{4.16}$$

$$\zeta_{CH_4,NaCl} = c_1 \tag{4.17}$$

Table 4–11. Coefficients for the modified Spivey et al. correlation for methane solubility in brine

Coefficient	$\frac{\mu^{l(o)}_{CH_4}}{RT}$, Eq. (4.19)	$\lambda_{CH_4,Na}$, Eq. (4.16)	$\zeta_{CH_4,NaCl}$, Eq. (4.17)
c_1	8.3143711	–0.80898	–3.89E–03
c_2	–7.2772168E–04	1.0827E–03	
c_3	2.1489858E+03	183.85	
c_4	–1.4019672E–05		
c_5	–6.6743449E+05		
c_6	7.6985890E–02	3.924E–04	
c_7	–5.0253331E–05		
c_8	–30.092013		
c_9	4.8468502E+03		
c_{10}	0	–1.97E–06	

Next, calculate the methane solubility in brine, in g-mol/kg H_2O, from equation (4.18).

$$m_{CH_4,b} = m_{CH_4,w}\,\text{EXP}[-2\lambda(T,p)m_{Na} - \zeta(T,p)m_{NaCl}^2] \tag{4.18}$$

Density of brine containing dissolved methane

The most straightforward way to calculate the density of brine containing dissolved methane is to calculate the mass and volume of a sample of brine containing 1,000 grams pure water, then calculate the density of the brine from the mass and volume. The following steps give the correct density for brines either saturated or undersaturated with methane.

Step 11. First, calculate the derivatives with respect to pressure for the terms $\left(\frac{\mu_{CH_4}^{l(o)}}{RT}\right)$, $\lambda_{CH_4,Na}$, and $\zeta_{CH_4,NaCl}$, using equations (4.19)–(4.21), along with the appropriate coefficients from table 4–11. The coefficients for the term $\left(\frac{\mu_{CH_4}^{l(o)}}{RT}\right)$ are from Duan-Mao, converted to pressures in MPa.[25] Again, note that temperature in equation (4.19) is Kelvin.

$$\frac{\partial}{\partial p}\left(\frac{\mu_{CH_4}^{l(o)}}{RT}\right)_{T,m} = c_6 + c_7 T + \frac{c_8}{T} + \frac{c_9}{T^2} \tag{4.19}$$

$$\left.\frac{\partial \lambda_{CH_4,Na}}{\partial p}\right|_{T,m} = c_6 + 2c_{10}p \tag{4.20}$$

$$\left.\frac{\partial \zeta_{CH_4,NaCl}}{\partial p}\right|_{T,m} = 0 \tag{4.21}$$

Next, calculate the partial molar volume of methane in brine, in cm^3/g-mol, using equation (4.22), from Duan.[26] Here, R is the universal gas constant, $R = 8.314467$ MPa·cm^3/ g-mol·K, and T is the temperature in Kelvin. The concentration of sodium ions is m_{Na}, in g-mole/kg H$_2$O, and is numerically equal to the salinity m_{NaCl}, since a sodium chloride brine is assumed.

$$\begin{aligned} V_{M\,CH_4(brine)} &= RT\frac{\partial}{\partial p}\left(\frac{\mu_{CH_4}^{l(o)}}{RT} + 2m_{Na}\lambda_{CH_4,Na} + m_{NaCl}{}^2\zeta_{CH_4,NaCl}\right)_{T,m} \\ &= RT\left[\frac{\partial}{\partial p}\left(\frac{\mu_{CH_4}^{l(o)}}{RT}\right)_{T,m} + 2m_{Na}\left.\frac{\partial\lambda_{CH_4,Na}}{\partial p}\right|_{T,m} + m_{NaCl}{}^2\left.\frac{\partial\zeta_{CH_4,NaCl}}{\partial p}\right|_{T,m}\right] \end{aligned} \tag{4.22}$$

Step 12. Calculate the density of brine with dissolved methane.

First, calculate the specific volume of methane-free brine, $v_{b,0}$, from the density of methane-free brine, $\rho_{b,0}$, from equation (4.12).

$$v_{b,0} = v_{b,0}\left(T, p, m_{NaCl}\right) = \frac{1}{\rho_{b,0}\left(T, p, m_{NaCl}\right)} \tag{4.23}$$

Next, calculate the density of the brine with dissolved methane, in g/cm^3, from equation (4.24).

$$\rho_b\left(T, p, m_{NaCl}, m_{CH_4}\right) = \frac{1000 + m_{NaCl}\, M_{NaCl} + m_{CH_4}\, M_{CH_4}}{\left(1000 + m_{NaCl} M_{NaCl}\right) v_{b,0} + m_{CH_4} V_{M\, CH_4\,(brine)}}, \tag{4.24}$$

where M_{NaCl} is the molecular weight of sodium chloride (58.4428 g/g-mol), and M_{CH_4} is the molecular weight of methane (16.043 g/g-mol).

If desired, the specific volume of brine with dissolved methane, in cm^3/g, may be obtained from the reciprocal of the density, as shown in equation (4.25).

$$\begin{aligned} v_b\left(T, p, m_{NaCl}, m_{CH_4}\right) &= \frac{1}{\rho_b\left(T, p, m_{NaCl}, m_{CH_4}\right)} \\ &= \frac{\left(1000 + m_{NaCl} M_{NaCl}\right) v_{b,0} + m_{CH_4} V_{M\, CH_4\,(brine)}}{1000 + m_{NaCl} M_{NaCl} + m_{CH_4}\, M_{CH_4}} \end{aligned} \tag{4.25}$$

Compressibility of brine containing dissolved methane

The equation for compressibility of brine with dissolved methane is obtained from the definition of the coefficient of isothermal compressibility and the equation for specific volume of brine with dissolved methane, equation (4.25).

The coefficient of isothermal compressibility may be defined as the fractional change in specific volume per unit change in pressure, i.e., equation (4.26).

$$c \equiv -\frac{1}{v}\left(\frac{\partial v}{\partial p}\right)_T \tag{4.26}$$

In combining equations (4.25) and (4.26), there are two cases to consider. The simpler case is for a single-phase brine system. In this situation, the amount of methane dissolved in the brine is independent of pressure, and the brine is said to be *undersaturated*. For this case, the compressibility may be calculated as follows.

Step 13. First, calculate the derivative of the methane-free brine specific volume with respect to pressure, from the methane-free compressibility as given in equation (4.27).

$$\left.\frac{\partial v_{b,0}}{\partial p}\right|_{T,m} = -v_{b,0}c_{b,0}\qquad(4.27)$$

,

where $v_{b,0}$ is the specific volume of methane-free brine from equation (4.23), and $c_{b,0}$ is the compressibility of methane-free brine from equation (4.9).

Step 14. Next, calculate the second derivatives with respect to pressure for the terms $\frac{\mu_{CH_4}^{l(o)}}{RT}$, $\lambda_{CH_4,Na}$, and $\zeta_{CH_4,NaCl}$, using equations (4.28)–(4.30), along with the appropriate coefficients from table 4–11.

$$\frac{\partial^2}{\partial p^2}\left(\frac{\mu_{CH_4}^{l(o)}}{RT}\right)_{T,m} = 0\qquad(4.28)$$

$$\left.\frac{\partial^2 \lambda_{CH_4,Na}}{\partial p^2}\right|_{T,m} = 2c_{10}\qquad(4.29)$$

$$\left.\frac{\partial^2 \zeta_{CH_4,NaCl}}{\partial p^2}\right|_{T,m} = 0\qquad(4.30)$$

Step 15. Next, calculate the derivative of the molar volume of methane dissolved in brine with respect to pressure from equation (4.31).

$$\left.\frac{\partial V_{M\,CH_4(brine)}}{\partial p}\right|_{T,m} = RT\frac{\partial^2}{\partial p^2}\left(\frac{\mu_{CH_4}^{l(o)}}{RT} + 2m_{Na}\lambda_{CH_4,Na} + m_{NaCl}{}^2\zeta_{CH_4,NaCl}\right)_{T,m}$$

$$= RT\left[\frac{\partial^2}{\partial p^2}\left(\frac{\mu_{CH_4}^{l(o)}}{RT}\right)_{T,m} + 2m_{Na}\left.\frac{\partial^2\lambda_{CH_4,Na}}{\partial p^2}\right|_{T,m} + m_{NaCl}{}^2\left.\frac{\partial^2\zeta_{CH_4,NaCl}}{\partial p^2}\right|_{T,m}\right], \tag{4.31}$$

where m_{Na} is the concentration of sodium ions in g-mol/kg H_2O (numerically equal to m_{NaCl} for a brine containing only sodium chloride).

Step 16. Calculate the compressibility of undersaturated brine, c_{wu}, in MPa^{-1}, from equation (4.32).

$$c_{wu} = -\frac{\left(1000 + m_{NaCl}M_{NaCl}\right)\left.\dfrac{\partial v_{b,0}}{\partial p}\right|_{T,m_{NaCl}} + m_{CH_4}\left.\dfrac{\partial V_{M\,CH_4(brine)}}{\partial p}\right|_{T,m_{NaCl}}}{\left(1000 + m_{NaCl}M_{NaCl}\right)v_{b,0} + m_{CH_4}V_{M\,CH_4(brine)}} \tag{4.32}$$

The second case of interest for compressibility of brine with dissolved methane is that of a two-phase system with brine in equilibrium with methane gas. In this case, the derivative with respect to pressure in equation (4.26) is taken assuming that the brine and gas phases remain in equilibrium. Thus, the change in methane concentration with pressure must be accounted for in taking the derivative in equation (4.26).

Step 17. First, calculate the derivative of the methane solubility with respect to pressure, as shown in equation (4.33).

$$\left.\frac{\partial m_{CH_4}}{\partial p}\right|_{T,m} = m_{CH_4}\left[\frac{2A(T)\ln\left(p - p_\sigma\right) + B(T)}{p - p_\sigma} - 2\frac{\partial\lambda_{CH_4,Na}}{\partial p}m_{Na}\right], \tag{4.33}$$

where p_σ is the vapor pressure of pure water calculated in Step 8, and $A(T)$ and $B(T)$ are temperature-dependent coefficients calculated in Step 9.

Step 18. Next, calculate the molar volume of methane, $V_{M\ CH_4(g)}$, in cm^3, in the gas phase from the real gas law, as shown in equation (4.34).

$$V_{M\ CH_4(g)} = \frac{zRT}{p} \tag{4.34}$$

Step 19. Next, calculate the compressibility of saturated brine, c_{ws}, in MPa^{-1}, from equation (4.35).

$$c_{ws} = -\frac{\left(1000 + m_{NaCl}M_{NaCl}\right)\left.\frac{\partial v_{b,0}}{\partial p}\right|_{T,m_{NaCl}} + m_{CH_4}\left.\frac{\partial V_{M\ CH_4}}{\partial p}\right|_{T,m_{NaCl}} + \left.\frac{\partial m_{CH_4}}{\partial p}\right|_{T,m_{NaCl}}\left(V_{M\ CH_4(brine)} - V_{M\ CH_4(g)}\right)}{\left(1000 + m_{NaCl}M_{NaCl}\right)v_{b,0} + m_{CH_4}V_{M\ CH_4(brine)}} \tag{4.35}$$

Formation volume factor of brine

Step 20. Calculate the specific volume of methane-free brine, $v_{b,0}$ at standard conditions from equation (4.23).

Step 21. Calculate the brine formation volume factor, B_w, in res cm^3/std cm^3, from equation (4.36).

$$B_w = \frac{\left(1000 + m_{NaCl}M_{NaCl}\right)v_{b,0} + m_{CH_4}V_{M\ CH_4(brine)}}{\left(1000 + m_{NaCl}M_{NaCl}\right)v_{b,0}\left(T_{sc}, p_{sc}, m_{NaCl}\right)} \tag{4.36}$$

Solution gas-water ratio of methane in brine

Step 22. Calculate the solution gas-water ratio of methane in brine, R_{sw}, in std cm^3/std cm^3, from equation (4.37).

$$R_{sw} = \frac{m_{CH_4,b}V_{M\ CH_4(g)}\left(T_{sc}, p_{sc}\right)}{\left(1000 + m_{NaCl}M_{NaCl}\right)v_{b,0}\left(T_{sc}, p_{sc}, m_{NaCl}\right)} \tag{4.37}$$

Brine Viscosity

Although several studies of viscosity of sodium chloride brines have been published, few of these have reported viscosities at typical oilfield temperatures, pressures, and salinities. On the other hand, the properties of pure water have been extensively investigated. The International Association for the Properties of Water and Steam adopted a correlation for viscosity of pure water (IAPWS-08) valid for both liquid and vapor phases at temperatures from 32°F to 1,652°F (0°C to 900°C) and pressures up to 43,500 psia (300 MPa).[27]

Brine viscosity database

Table 4–12. Brine viscosity database

Reference	n	Temperature range (°C)	Pressure range (MPa)	NaCl salinity range (g-mol/kg H_2O)	Viscosity range (cp)
Kestin et al. (1977)[a]	249	18–40.7	0.1–31.5	2.6–5.6	0.92–1.65
Kestin et al. (1978)	726	18–154	0.1–31.5	0.48–5.4	0.20–1.55
Kestin-Shankland (1984)	388	24–201.5	0.1–32.1	1.1–6.0	0.16–1.48
Combined	**1,363**	**18–201**	**0.1–32.1**	**0.48–6.0**	**0.16–1.65**

Note: n = number of points in database. [a] Kestin, J., Khalifa, H.E., Ro, S.-T., and Wakehame, W.A. "Preliminary Data on the Pressure Effect on the Viscosity of Sodium Chloride-Water Solutions in the Range 10–40°C." *Journal of Chemical and Engineering Data*. Vol. 22, No. 2 (1977).

Table 4–12 shows the database of brine viscosity data used to evaluate correlation equations for brine viscosity. While this database contains only three data sets, these three data sets represent most of the measurements of viscosity of sodium chloride brines that have been reported in the literature. Mao and Duan compiled a database of the brine viscosity measurements reported in the literature.[28] Their database includes 22 papers with a total of 2,261 data points, of which 1,363 data points were used in the present evaluation.

Three papers included in the Mao and Duan database cover ranges of temperature and pressure of interest in the petroleum industry, beyond those covered by the Kestin et al. and Kestin-Shankland papers. Semenyuk et al. measured brine viscosity for temperatures from 212°F to 673°F (100°C to 356°C) and pressures from 1,450 to 21,750 psia (1 to 150 MPa).[29] Pepinov et al. reported brine viscosities at pressures from 290 to 4,350 psia (2 to 30 MPa) and temperatures between 68°F and 662°F (20°C and 350°C).[30] Pepinov et al. reported relative viscosities of sodium chloride brine over a similar range of pressures, from 290 to 4,350 psia (2 to 30 MPa), and temperatures, from 77°F to 662°F (25°C to 350°C).[31]

Figure 4–19 shows the distribution of data in the database of brine viscosity by pressure and temperature. Note the limited number of brine measurements at higher temperatures, especially between 300°F and 400°F (149°C and 204°C).

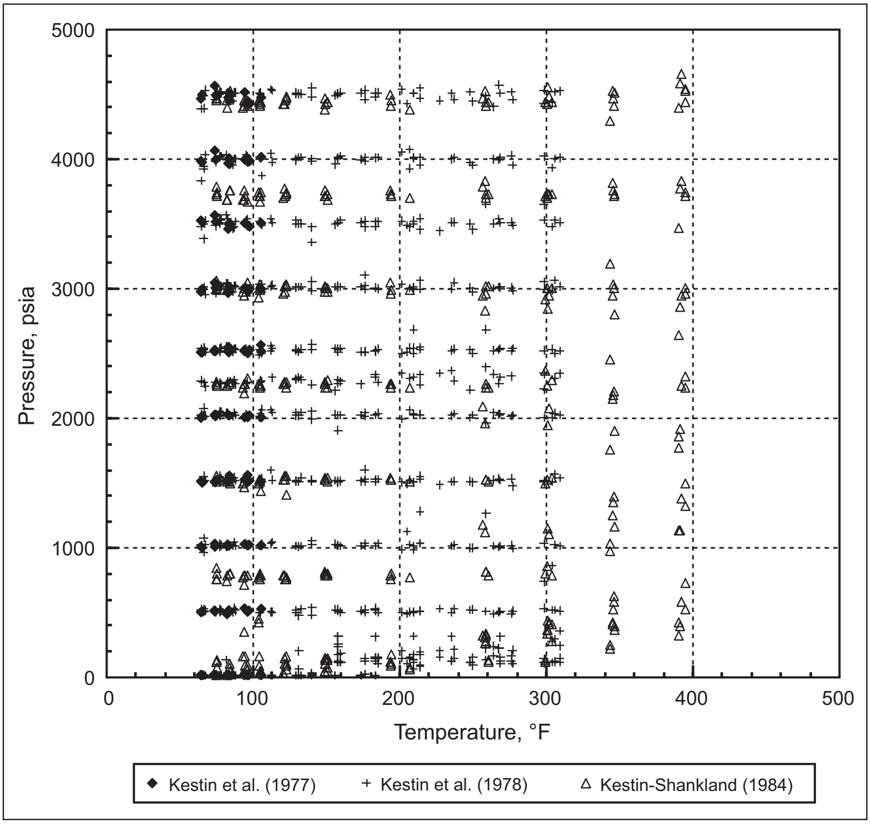

Fig. 4–19. Distribution of data points in the brine viscosity database, pressure vs. temperature

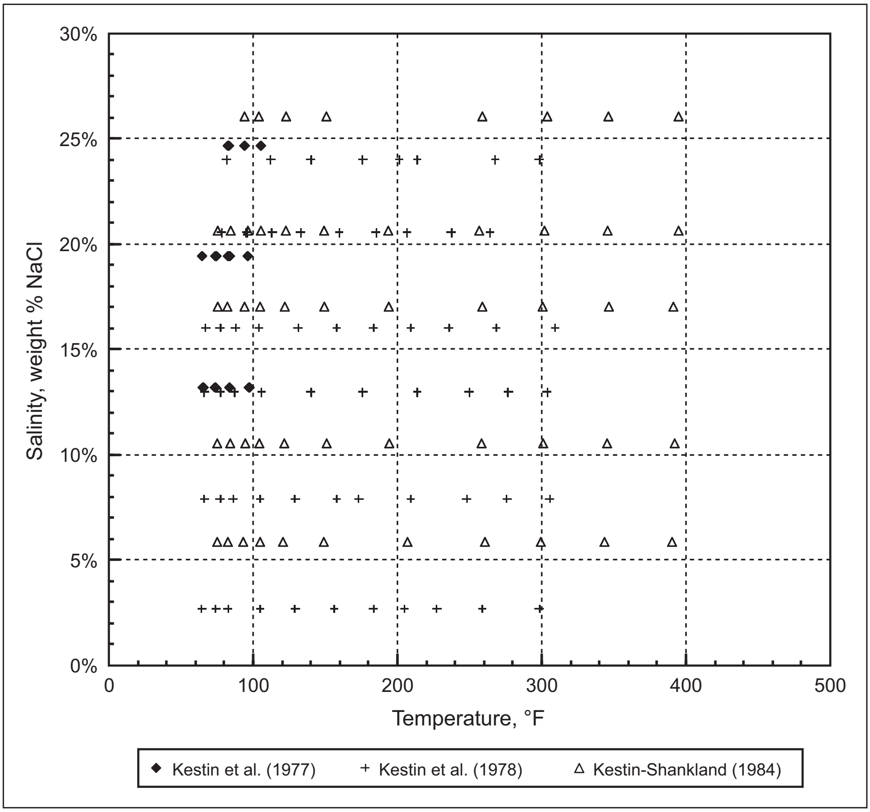

Fig. 4–20. Distribution of data points in the brine viscosity database, salinity vs. temperature

Figure 4–20 shows the distribution of data in the database of brine viscosity by salinity and temperature. Note the limited number of brine measurements at higher temperatures. Each symbol in figure 4–20 may represent multiple measurements at the same salinity and temperature, but different pressures. Thus, figure 4–20 appears to show fewer data points than figures 4–19 and 4–21.

Figure 4–21 shows the distribution of data in the database of brine viscosity by pressure and salinity. There appears to be a relatively uniform distribution of data over pressure and salinity.

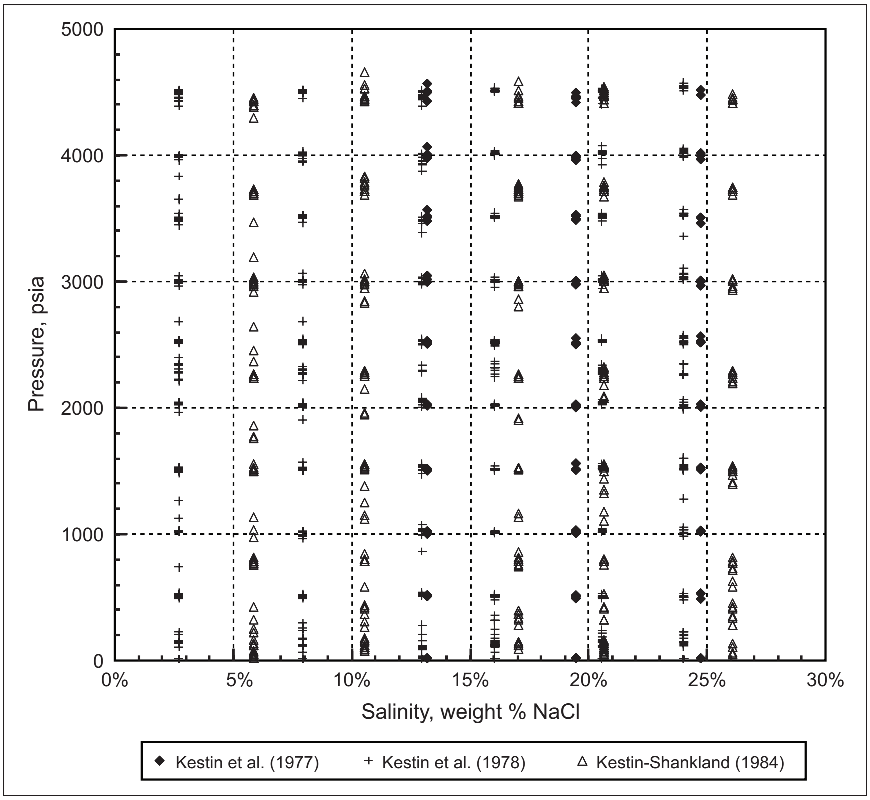

Fig. 4–21. Distribution of data points in the brine viscosity database, pressure vs. salinity

Comparison of correlations for brine viscosity

A number of correlations for viscosity of sodium chloride brines have appeared in the literature.

Kestin et al. (1978) presented a correlation equation along with their measurements of brine viscosity.[32]

Phillips et al. (1978) presented another correlation equation.[33] The Phillips et al. correlation equation is independent of pressure.

Phillips et al. (1980) presented a simple correlation equation to calculate the brine viscosity from the viscosity of pure water as shown in equation (4.38).

$$\mu_b(T,p,m) = \mu_r(T,m)\mu_w(T,p) \tag{4.38}$$

where $\mu_b(T,p,m)$ is the brine viscosity, $\mu_w(T,p)$ is the viscosity of pure water, and $\mu_r(T,m)$ is the relative viscosity of the brine, defined as the ratio of the brine viscosity at temperature T, pressure p, and sodium chloride concentration, m, to that of pure water at the same temperature and pressure.[34]

Meehan (1980) developed a simple correlation equation, based on an earlier graphical correlation presented by Matthews and Russell.[35] The Meehan correlation equation gives the viscosity as the product of the viscosity of brine at atmospheric pressure (the saturation pressure of water for temperatures above 212°F), $\mu^*(T,S)$, and a pressure correction factor, $f(T,p)$. Meehan stated that the estimated maximum error in the graphical correlation for $\mu^*(T,S)$ was 1% for temperatures between 40°F and 120°F, 5% for temperatures from 120°F to 212°F, and 10% for temperatures from 212°F to 400°F. According to Meehan, the estimated maximum error in graphical correlation for $f(T,p)$ was 5%. Meehan noted that additional error might be introduced by the correlation equation.

Kestin and Shankland (1984) presented two correlation equations to fit their brine viscosity data, both of which fit the data within the experimental uncertainty.[36]

Spivey et al. (2004) noted that the Kestin (1978) correlation predicted negative viscosities for high temperatures.[37] They suggested a simple equation for extrapolating the Kestin correlation equation for temperatures above 257°F (125°C). Spivey et al. did not compare the results of their extrapolation method with laboratory data for brine viscosity.

In 2009, Mao and Duan presented correlation equations for viscosity of lithium chloride, sodium chloride, and potassium chloride brines.[38] They claimed their correlation equation for viscosity of sodium chloride brine matched the available laboratory data within 1% for temperatures between 32°F and 662°F and (0°C and 350°C), pressures up to 14,500 psia (100 MPa), and sodium chloride content up to 26 wt% (6 g-mol/kg H_2O). The Mao-Duan correlation equations have the form given in equation (4.38), with a correlation equation for pure water defined as a function of temperature and density, as shown in equation (4.39).

$$\mu_w = \mu_w\left(T, \rho_w\right) \tag{4.39}$$

where ρ_w is the density of pure water at the temperature and pressure of interest.

Mao and Duan recommend using the IAPWS-IF97 correlation equation for density of pure water.[39]

Table 4–13 shows the calculated AARE and ARE for the available correlations. The Phillips et al. (1980) and the Mao-Duan (2009) correlations for relative viscosity of brine were evaluated using both the IAWPS-08 correlation equation for viscosity of pure water and the Mao-Duan correlation equation for viscosity of pure water with the modified Spivey et al. correlation equation for density of pure water. In addition, the Mao-Duan relative viscosity correlation equation was also evaluated using the Mao-Duan correlation equation for viscosity of pure water with the IAPWS-IF97 correlation equation for density of pure water.

Table 4–13. Results of evaluation of brine viscosity correlations against the brine viscosity database in table 4–12

Correlation	ARE	AARE
Mao-Duan (2009)*	–0.10%	0.64%
Mao-Duan (2009)†	–0.11%	0.64%
Mao-Duan (2009)‡	–0.11%	0.64%
Phillips et al. (1980)*	0.06%	0.73%
Phillips et al. (1980)†	0.06%	0.79%
Kestin et al. (1978)	0.25%	0.81%
Kestin-Shankland (1984), correlation (a)	0.39%	0.82%
Spivey et al. (2004)	0.17%	0.83%
Kestin-Shankland (1984), correlation (b)	0.45%	0.88%
Phillips et al. (1978)	–0.26%	1.75%
Meehan (1980)	2.50%	4.39%

* With pure water viscosity from IAPWS-08 † With pure water viscosity from Mao-Duan (2009), pure water density from modified Spivey et al. ‡ With pure water viscosity from Mao-Duan (2009), pure water density from IAPWS-IF97

The performance of six of the correlations as functions of temperature, pressure, and salinity is shown in figures 4–21 through 4–24.

The Mao-Duan relative viscosity correlation was evaluated using the Mao-Duan water viscosity correlation with the modified Spivey et al. correlation for density of pure water. Results using the IAWPS-08 correlation for pure water viscosity, as well as the Mao-Duan pure water viscosity equation with the IAWPS-IF97 pure water density equation gave almost identical results, and were omitted from the graphical comparisons.

The Phillips et al. (1980) correlation was evaluated using the Mao-Duan water viscosity correlation with the modified Spivey et al. correlation for density of pure water. Results using the IAWPS-08 correlation for pure water viscosity were almost identical and were omitted from the graphical comparisons.

The two Kestin-Shankland correlation equations, which they referred to as (a) and (b), gave very similar results, with correlation (a) giving slightly better results for the database in table 4–12. The Kestin-Shankland (b) correlation was omitted from the graphical comparisons.

The Spivey et al. correlation equation gave results very similar to those of the Kestin et al. correlation on which it was based, and it was also omitted from the graphical comparisons.

Figure 4–22 shows the ARE and the AARE as functions of temperature for the selected brine viscosity correlations.

The Mao-Duan (2009) correlation gave the best overall performance, with a maximum AARE of 1.03%, with the Phillips et al. (1980), Kestin et al. (1978), and Kestin-Shankland (1984)(a) correlations also having AAREs less than 1.5% across the entire range of temperatures.

Figure 4–23 shows the ARE and AARE as functions of pressure for the selected brine viscosity correlations.

The Mao-Duan (2009) correlation gave the best overall performance, with a maximum AARE of 0.76%, with the Phillips et al. (1980), Kestin (1978), and Kestin-Shankland (1984) (a) correlations also having AAREs less than 1.2%, across the entire range of pressures.

Figure 4–24 shows the ARE and the AARE as functions of salinity for the selected brine viscosity correlations.

Again, the Mao-Duan (2009) correlation gave the best overall performance, with a maximum AARE of 0.87%, with the Phillips et al. (1980), Kestin (1978), and Kestin-Shankland (1984)(a) correlations also having AAREs less than 1.4%, across the entire range of salinities.

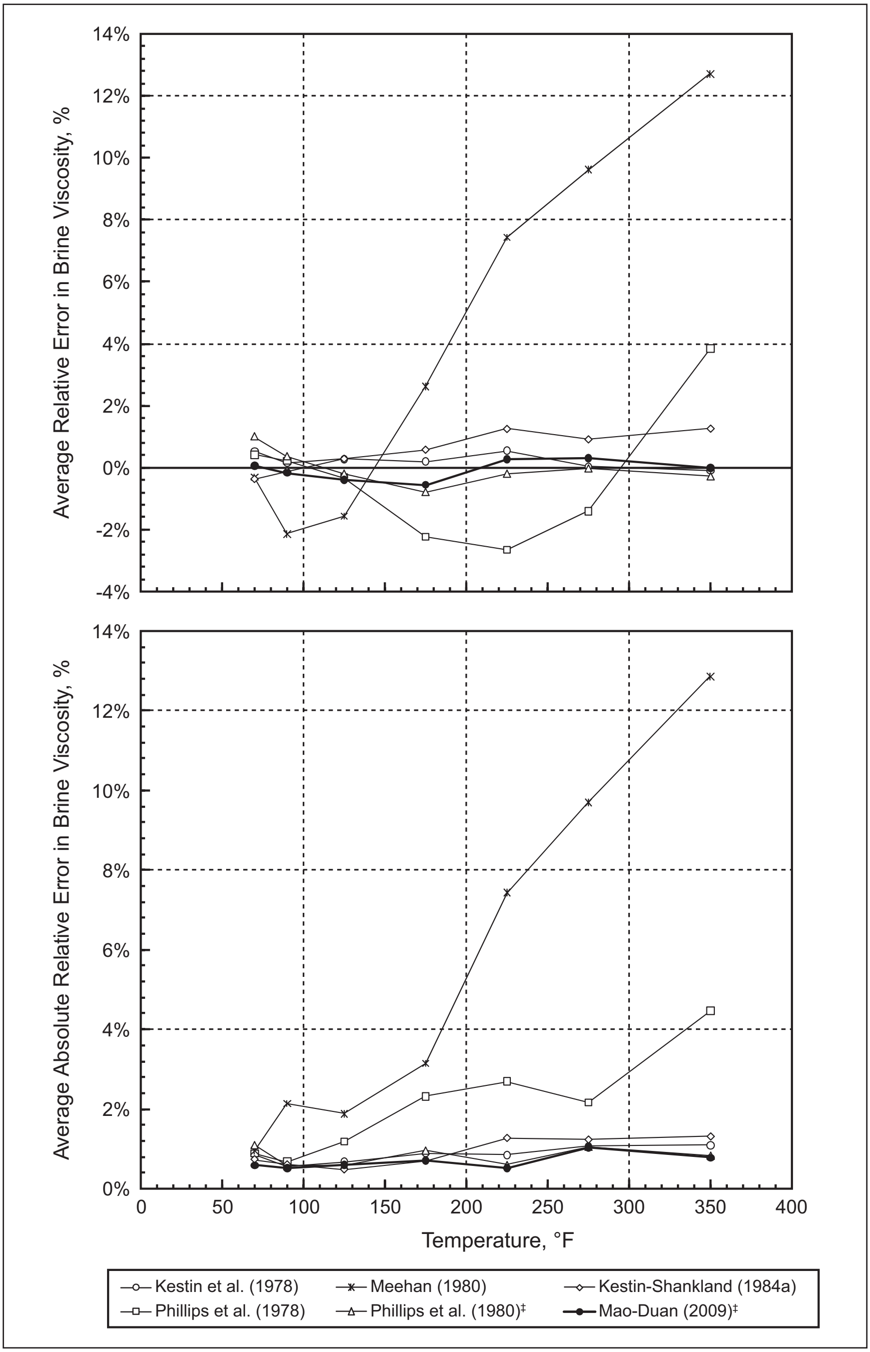

Fig. 4–22. Average relative error and absolute relative error in brine viscosity as functions of temperature for various correlations, based on the brine viscosity database in table 4–12

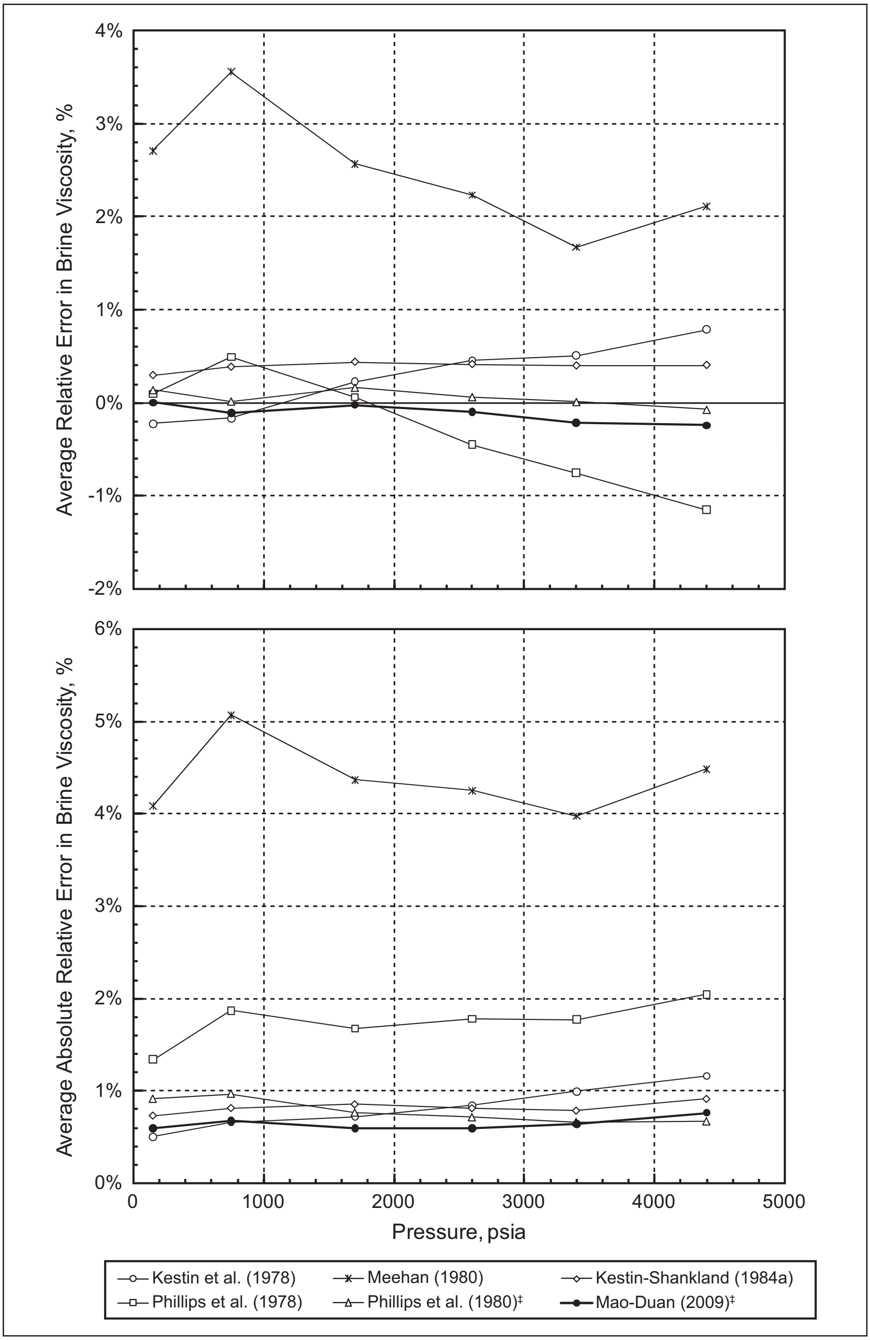

Fig. 4–23. Average relative error and average absolute relative error in brine viscosity as functions of pressure for various correlations, based on the brine viscosity database in table 4–12

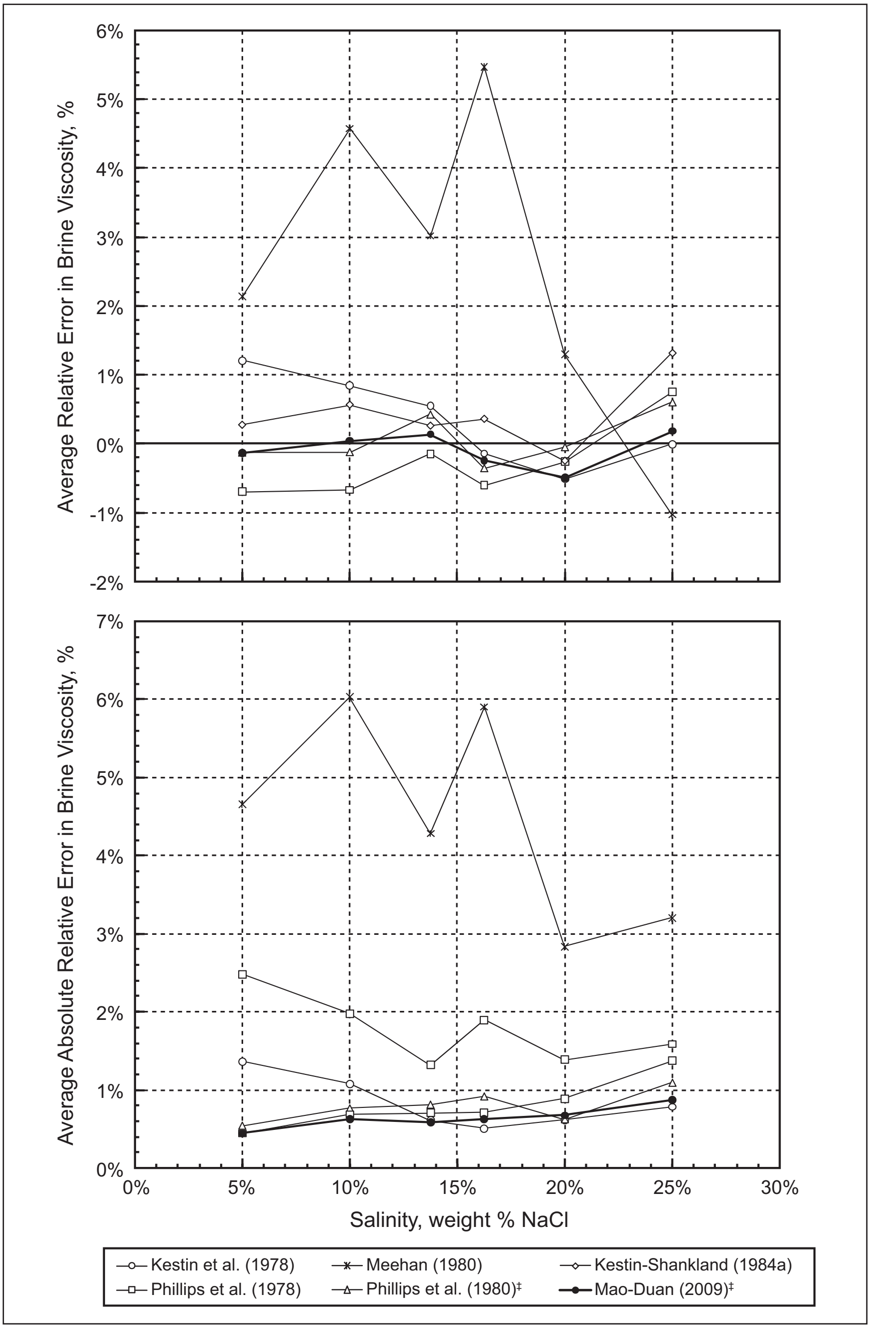

Fig. 4–24. Average relative error and average absolute relative error in brine viscosity as functions of salinity for various correlations, based on the brine viscosity database in table 4–12

Extrapolating beyond the range of experimental data

Two of the correlations, Phillips et al. (1980) and Mao-Duan (2009), have the form given in equation (4.40).

$$\mu_b(T,p,m) = \mu_r(T,m)\,\mu_w(T,p), \tag{4.40}$$

where $\mu_b(T,p,m)$ is the brine viscosity, $\mu_w(T,p)$ is the viscosity of pure water, and $\mu_r(T,m)$ is the relative viscosity of the brine.

Note that the relative viscosity is independent of pressure. Thus, the pressure dependence of brine viscosity is accounted for by the pressure dependence of pure water viscosity.

Figure 4–25 shows a comparison of the relative viscosity for brine as a function of temperature for several different values of salinity, for the Phillips et al. (1980) correlation equation and the Mao-Duan (2009) correlation equation. The Phillips et al. correlation equation used some of the Semenyuk et al. data in developing their correlation. While Mao-Duan did not use the Semenyuk data in developing their correlation equation, they did test their correlation equation against the Semenyuk data. Both correlation equations show very similar performance, with the maximum difference slightly larger than 4% at a temperature of 572°F (300°C) and a salinity of 22.6 wt% sodium chloride (5 g-mol/kg H_2O).

Thus, either the Mao-Duan (2009) correlation equation or the Phillips et al. (1980) correlation equation for relative viscosity of brine may be used, along with a suitable correlation equation for pure water viscosity, to get reasonable estimates of brine viscosity for pressures to 29,000 psia (200 MPa) and temperatures to 572°F (300°C). The Mao-Duan (2009) correlation equation for pure water, used with either the modified Spivey et al. correlation equation or the IAPWS-IF97 correlation equation for density of pure water, gives estimates of pure water viscosity within 0.5% of those given by the IAPWS-08 pure water viscosity correlation equation for pressures up to 14,500 psia (100 MPa) and temperatures from 50°F to 572°F (10°C to 300°C). It gives estimates within 2% for pressures from 14,500 to 29,000 psia (100 to 200 MPa) and temperatures from 140°F to 572°F (60°C to 300°C).

The Spivey et al. method (2004) gives viscosities that are as much as 25% lower than those estimated from the Phillips et al. and Mao-Duan equations, and is no longer recommended.

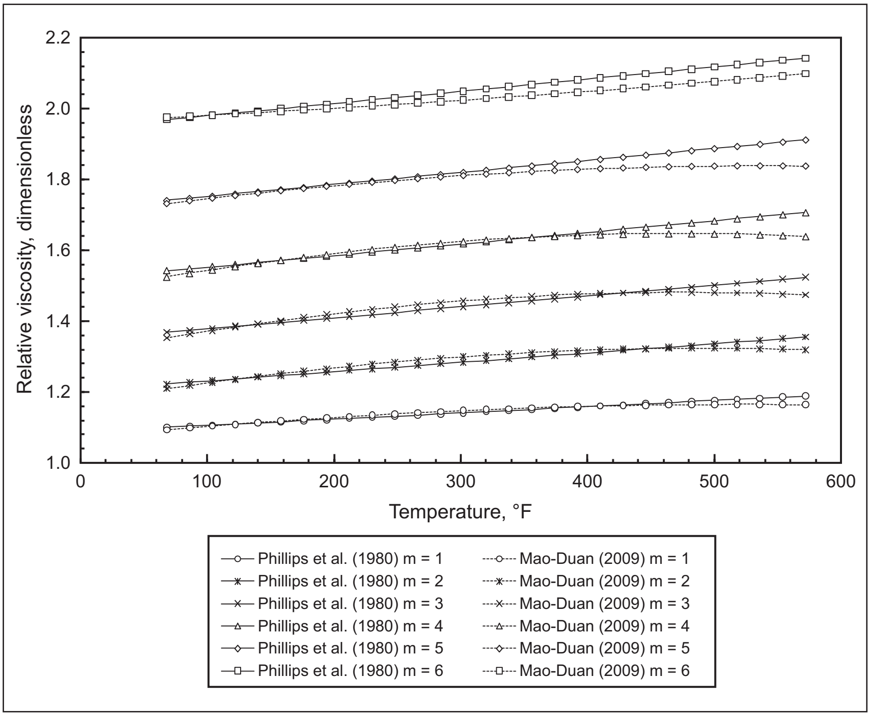

Fig. 4–25. Extrapolation of relative viscosity to high temperatures

Computing viscosity of oilfield brines

The following procedure is recommended for calculating the viscosity of oilfield brines. The density of pure water is calculated using the Spivey et al. correlation. The Mao-Duan correlation is then used to calculate the viscosity of pure water, the relative viscosity of brine, and the brine viscosity.

Step 1. Calculate the density of pure water, $\rho_w(T,p)$, using the Spivey et al. correlation, equations (4.2)–(4.5).

Step 2. Calculate the viscosity of pure water using the Mao-Duan correlation. First, calculate the natural logarithm of the water viscosity using equation (4.41).

$$\ln \mu_w(T,p) = \sum_{i=1}^{5} d_i T^{i-3} + \rho_w(T,p) \sum_{i=6}^{10} d_i T^{i-8} \tag{4.41}$$

where T is the temperature in K, $\mu_w(T,p)$ is the viscosity of pure water in Pa·s, and the coefficients d_i are given in table 4–14.

Next, calculate the viscosity of pure water, $\mu_w(T,p)$, in Pa·s, from equation (4.42).

$$\mu_w(T,p) = \mathrm{EXP}(\ln \mu_w(T,p)). \tag{4.42}$$

Step 3. Calculate the relative viscosity of brine using the Mao-Duan correlation. Calculate temperature-dependent salinity coefficients A, B, and C from equations (4.43)–(4.45).

$$A = a_o + a_1 T + a_2 T^2 \tag{4.43}$$

$$B = b_o + b_1 T + b_2 T^2 \tag{4.44}$$

$$C = c_o + c_1 T \tag{4.45}$$

where T is the temperature in K, and the coefficients a_i, b_i, and c_i are given in table 4–14.

Calculate the natural logarithm of the relative viscosity from equation (4.46).

$$\ln \mu_r(T,m) = Am + Bm^2 + Cm^3 \tag{4.46}$$

where m is the sodium chloride content in g-mol/kg H_2O.

Calculate the relative viscosity of brine, $\mu_r(T,m)$, as shown in equation (4.47).

$$\mu_r(T,m) = \mathrm{EXP}(\ln \mu_r(T,m)) \tag{4.47}$$

Step 4. Calculate the brine viscosity, in Pa·s, from equation (4.48).

$$\mu_b(T,p,m) = \mu_r(T,m)\mu_w(T,p) \tag{4.48}$$

Table 4–14. Coefficients for Mao-Duan brine viscosity correlation, equations (4.41), (4.43), (4.44), and (4.4_)

Coefficient	Value	Coefficient	Value
d_1	0.28853170×10^{7}	a_0	-0.21319213
d_2	-0.11072577×10^{5}	a_1	0.13651589×10^{-2}
d_3	-0.90834095×10^{1}	a_2	-0.12191756×10^{-5}
d_4	0.30925651×10^{-1}		
d_5	-0.27407100×10^{-4}	b_0	0.69161945×10^{-1}
d_6	-0.19283851×10^{7}	b_1	-0.27292263×10^{-3}
d_7	0.56216046×10^{4}	b_2	0.20852448×10^{-6}
d_8	0.13827250×10^{2}		
d_9	-0.47609523×10^{-1}	c_0	-0.25988855×10^{-2}
d_{10}	0.35545041×10^{-4}	c_1	0.77989227×10^{-5}

Units Conversions

Standard conversions

Table 4–15 gives conversion factors for converting between metric and oilfield units.

Table 4–15. Oilfield/metric conversion factors

Quantity	From	To	Multiply by	Divide by
Pressure	psia	MPa	6.89475729×10^{-3}	1.45037738×10^{2}
Compressibility	MPa^{-1}	psi^{-1}	6.89475729×10^{-3}	1.45037738×10^{2}
Density	g/cm^3	lb_m/ft^3	6.24279606×10^{1}	1.60184634×10^{-2}
Specific volume	cm^3/g	ft^3/lb_m	1.60184634×10^{-2}	6.24279606×10^{1}
Formation volume factor	cm^3/cm^3	bbl/STB	1	1
Solution gas-water ratio	cm^3/cm^3	scf/STB	5.61458333	1.78107607×10^{-1}
Viscosity	Pa·s	cp	1,000	1×10^{-3}

Nonstandard conversions

Temperature. Equation (4.49) is used to convert temperature in degrees Fahrenheit to degrees Celsius.

$$T(°C) = \frac{5}{9}\left[T(°F) - 32\right] \tag{4.49}$$

Equation (4.50) is used to convert from degrees Celsius to Kelvin.

$$T(K) = T(C°) + 273.15 \tag{4.50}$$

Salinity. Equation (4.51) is used to convert salinity, *S*, in weight fraction sodium chloride to sodium chloride concentration, *m*, in g-mol/kg H_2O.

$$m(g\text{–}mol/kg\ H_2O) = \frac{1000S}{M_{NaCl}(1-S)} \tag{4.51}$$

where M_{NaCl} is the molecular weight of sodium chloride, 58.4428 g/g-mol.

Nomenclature

Latin

$\boldsymbol{a(T)}$	Generic temperature-dependent parameter
$\boldsymbol{a_0 \ldots a_2}$	Coefficients in Mao-Duan relative viscosity correlation equation
$\boldsymbol{a_1 \ldots a_5}$	Coefficients in Spivey et al. temperature-dependent correlation equations
$\boldsymbol{a_1 \ldots a_6}$	Coefficients in IAPWS-95 vapor pressure correlation equation
$\boldsymbol{A}$	Coefficient in Mao-Duan relative viscosity correlation equations
$\boldsymbol{A}$	Coefficient in modified Spivey et al. methane solubility correlation equations
$\boldsymbol{b_0 \ldots b_2}$	Coefficients in Mao-Duan relative viscosity correlation equation
$\boldsymbol{B}$	Coefficient in Mao-Duan relative viscosity correlation equations
$\boldsymbol{B}$	Coefficient in modified Spivey et al. methane solubility correlation equations

c	Coefficient of isothermal compressibility, 1/psi, 1/MPa
$c_0 \ldots c_1$	Coefficients in Mao-Duan relative viscosity correlation equations
$c_1 \ldots c_{10}$	Coefficients in Duan correlation equations
C	Coefficient in Mao-Duan relative viscosity correlation equations
C	Coefficient in modified Spivey et al. methane solubility correlation equations
$d_1 \ldots d_{10}$	Coefficients in Mao-Duan pure water viscosity correlation equations
E	Coefficient in modified Spivey et al. compressibility correlation equations
F	Coefficient in modified Spivey et al. compressibility correlation equations
i	Index
I	Coefficient in modified Spivey et al. density correlation equations
m	Molality, g-mol/kg H_2O
M	Molecular weight, g/g-mol
n	Number of points
p	Pressure, psia, MPa
p_σ	Vapor pressure, MPa
p_0	Reference pressure, 70 MPa
R	Universal gas constant, 8.314467 MPa·cm^3/g-mol·K
R_{sw}	Solution gas-water ratio, cm^3/cm^3
T	Temperature, °F, °C, K
v	Specific volume, cm^3/g
V	Volume, cm^3
z	Real gas deviation factor, dimensionless

Greek

ζ	Interaction coefficient in Duan et al. methane solubility equations
ϑ	Temperature parameter in IAPWS-95 vapor pressure equation, dimensionless
λ	Interaction coefficient in Duan et al. methane solubility equations
μ	Viscosity, cp, Pa·s
$\frac{\mu_{CH_4}^{l(o)}}{RT}$	Standard chemical potential in Duan et al. methane solubility equations, unitless
ρ	Density, g/cm³

Subscripts

0	At 70 MPa reference pressure
$,0$	Methane free
$(brine)$	Dissolved in brine
b	Brine
c	Critical
CH_4	Methane
(g)	Gas phase
H_2O	Water
m	Salinity coefficient
M	Molar
Na	Sodium
$NaCl$	Sodium chloride
r	Relative
s	Saturated
sc	Standard conditions
u	Undersaturated
w	Pure water
w	Brine [equations (4.32), (4.35), (4.36), and (4.37)]

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5

IDENTIFICATION OF TYPE OF RESERVOIR FLUID

McCain gave guidelines for determining reservoir fluid type from field data.[1] Additional data have shown that some of these guidelines are incorrect.

Stock-Tank Liquid Gravity as an Indicator of Reservoir Fluid Type

Table 1 from McCain (1994) gives some guidelines for using stock-tank liquid gravities in degrees API for determining reservoir fluid type.[2] The numbers in the table are incorrect, and in fact, stock-tank liquid gravities are not valid indicators of reservoir fluid type. Figure 5–1 presents stock-tank liquid gravities and initial producing gas-liquid ratios for 2,828 different reservoir fluid studies or production tests. Both oils and gas condensates are represented on the graph.

There is obviously a trend in the gravities of the stock-tank liquids; however, the scatter is too large for the use of stock-tank liquid gravities in determining reservoir fluid type. The black oils have stock-tank liquid gravities with values ranging from less than 10°API to more than 60°API. The volatile oils have stock-tank liquid gravities from 30°API to more than 60°API. The gas condensates have stock-tank liquid gravities from slightly more than 30°API to more than 70°API. The overlaps in stock-tank liquid gravities among the three fluids are just too great for stock-tank liquid gravities to be useful in determining reservoir fluid type. One exception to this statement might be that if stock-tank oil gravity is less than 30°API, the reservoir fluid is most likely a black oil.

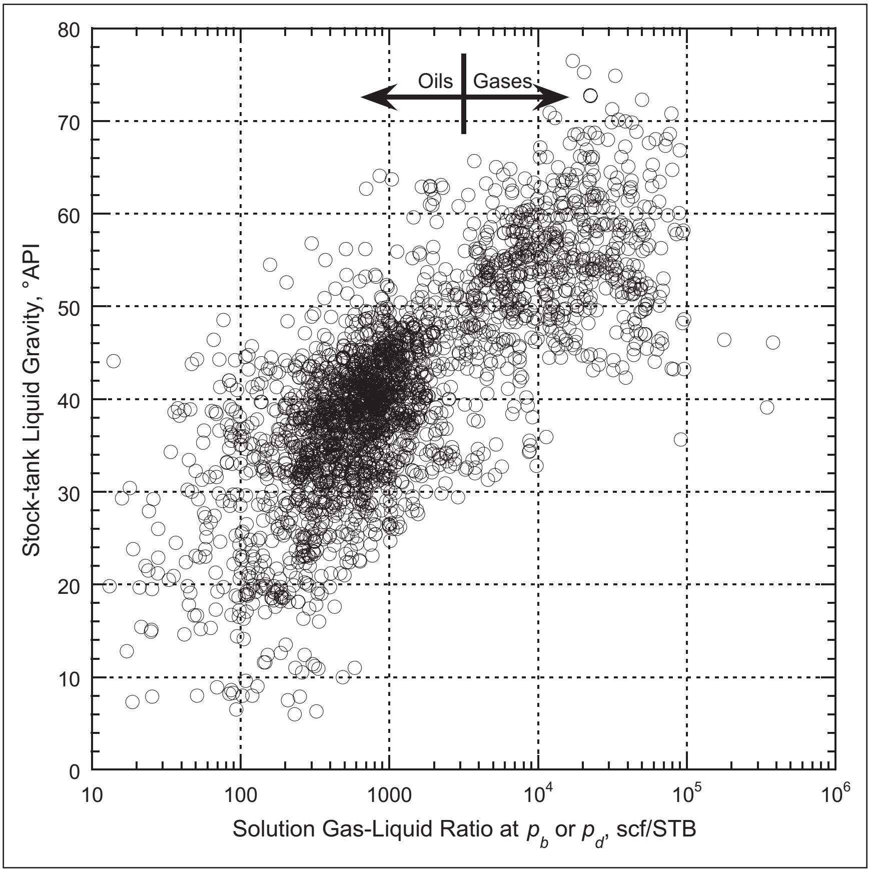

Fig. 5–1. Stock-tank oil gravity, API, is not an indicator of reservoir fluid type. Data from 2,828 reservoir fluid studies or production tests.

Stock-Tank Liquid Color as an Indicator of Reservoir Fluid Type

Table 1 from McCain (1994) also lists guidelines with respect to stock-tank liquid color.[3] While the colors listed in the table are generally correct, the color of the stock-tank liquid is not necessary for determining or confirming the reservoir fluid type.

Initial Producing Gas-Liquid Ratio as an Indicator of Reservoir Fluid Type

The term *initial* as used in this chapter, both in the text and also in the titles of the figures, means "at a time when the average reservoir pressure is above the dew point pressure or bubblepoint pressure of the reservoir fluid." The time period represented by *initial* may be a matter of days or possibly years, depending on the difference between the initial reservoir pressure and the saturation pressure of the reservoir fluid.

Also, the ratio of surface gas to stock-tank liquid is called *gas-liquid ratio* when comparing oils with gas condensates because the stock-tank liquid produced from a gas condensate is condensate rather than oil. Thus the symbol will be GLR_i. The *surface gas* includes the gas from all separators and the stock tank. A correlation procedure given in chapter 3 may be used to estimate the stock-tank gas-liquid ratio if it is not available from field data.

Volatile oils and black oils

Service company laboratory people state that the difference between volatile oils and black oils is an oil formation volume factor of 2.0 res bbl/STB when measured at bubblepoint pressure. If the measured oil formation volume factor is greater than 2.0 res bbl/STB, the produced gas is very rich and drops condensate in the gas meter, making gas volume measurement difficult. Thus, the oil is considered to be a *volatile oil* (volatile oils are associated with gas condensates rather than dry gases). Oil formation volume factors at bubblepoint pressures less than 2.0 res bbl/STB indicate black oils (black oils are associated with dry gases).

Figure 5–2 shows initial producing gas-liquid ratios (note that for oils these are the same as solution gas-oil ratios at the bubblepoint, R_{sb}) and oil formation volume factors at bubble-point pressure, B_{ob}, from 1,496 reservoir fluid samples with worldwide origins. Notice that if the initial producing gas-oil ratios are less than 1,500 scf/STB, all oil formation volume factors are less than 2.00 res bbl/STB, i.e., black oils. If the initial producing gas-oil ratios are greater than 1,900 scf/STB, all oil formation volume factors are greater than 2.00 res bbl/STB, i.e., volatile oils. Between these two values of initial producing gas-oil ratios, some oil formation volume factors are less than 2.00 res bbl/STB and some are greater. Thus, when the initial producing gas-oil ratios are between 1,500 and 1,900 scf/STB, the type of reservoir fluid cannot be determined with field production data.

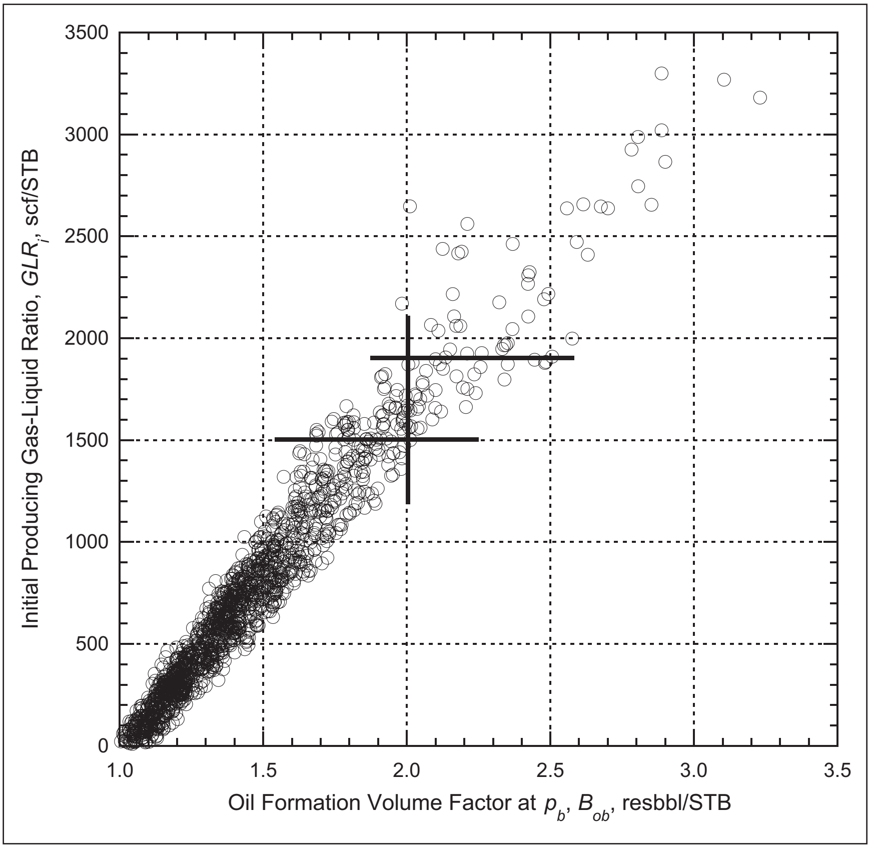

Fig. 5–2. Black oils and volatile oils—what's the difference? Data from 1,496 reservoir fluid studies.

So the transition point between volatile oils and black oils of 1,750 scf/STB for initial producing gas-oil ratio given in McCain (1994) must be changed.[4] If the initial producing gas-oil ratios are less than 1,500 scf/STB, the reservoir fluids are black oils; if greater than 1,900 scf/STB, the reservoir fluids are volatile oils. If the initial gas-producing ratios are between these values, a sample must be taken to the laboratory for determination of fluid type. When the initial producing gas-oil ratio is between 1,500 and 1,900 scf/STB, the oil should be treated as if it were a volatile oil unless determined otherwise in the laboratory.

Gas condensates and volatile oils

The other guidelines listed in table 1 of McCain (1994) with regard to initial producing gas-liquid ratios are based on substantial data and are valid.[5] In fact, figure 5–3 validates the distinct transition point of an initial producing gas-liquid ratio of 3,200 scf/STB for the difference between volatile oils and gas condensates. The data points on figure 5–3 indicate the laboratory observed dew points or bubblepoints at reservoir temperature. Virtually all the data above the horizontal dashed line exhibit dew points, i.e., the reservoir fluids are gases. Virtually all the data below this line exhibit bubblepoints, i.e., the reservoir fluids are liquids (oils). There are only two samples with bubblepoints and three samples with dew points that do not fit this pattern. Thus, an initial producing gas-liquid ratio of 3,200 scf/STB represents a precise transition from gas condensates to volatile oils.

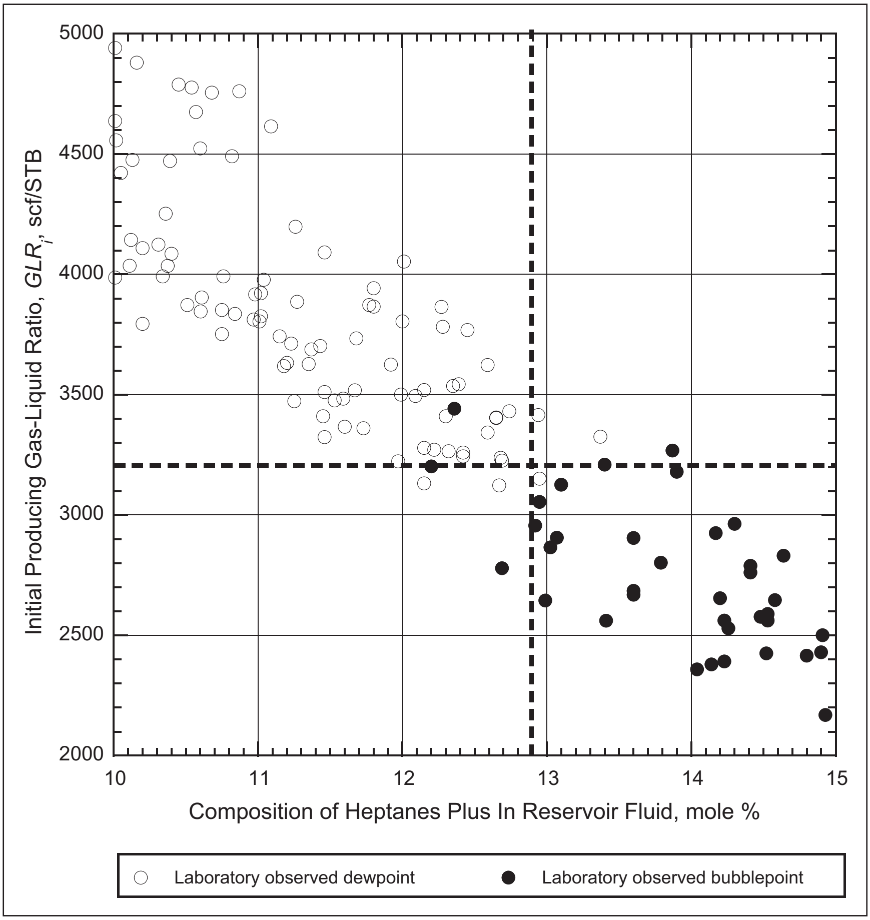

Fig. 5–3. Volatile oils and gas condensates—what's the difference? Data from 130 reservoir fluid studies.

Figure 5–3 is a subset consisting of 130 data points from the data shown in figure 5–4. Figure 5–4 shows a correlation of initial producing gas-liquid ratios against the composition of the reservoir fluid represented by the composition of the heptanes plus fraction. The full data set contains 1,451 different reservoir fluid studies (PVT reports) that used reservoir fluid samples with worldwide origins.

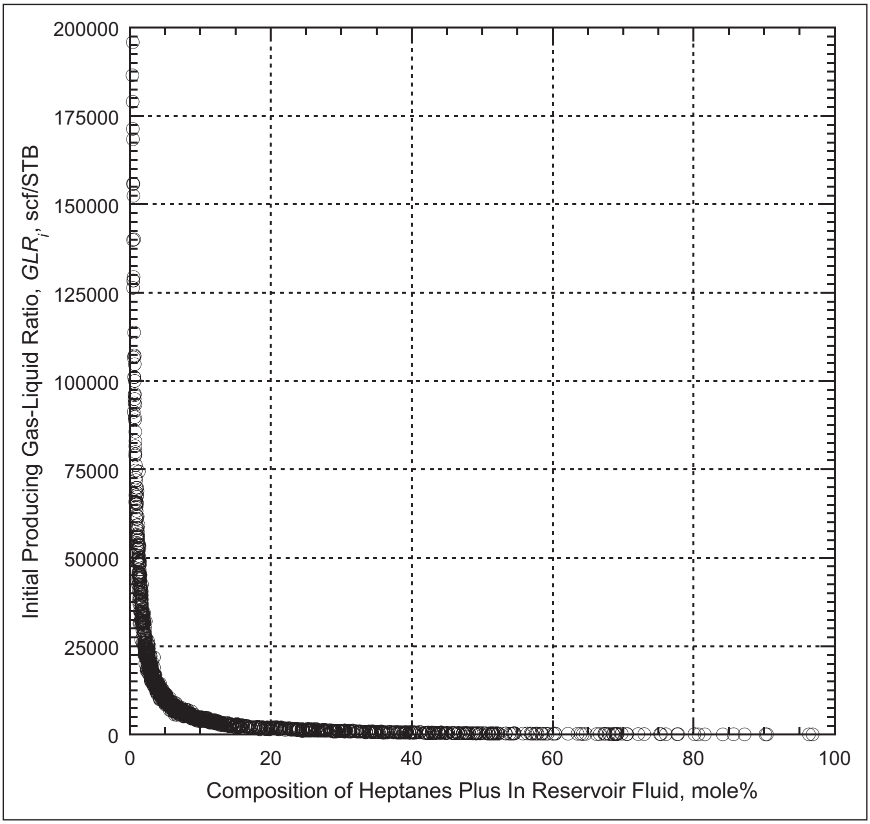

Fig. 5–4. Initial producing gas-liquid ratio correlates well with composition of heptanes plus in reservoir fluid (data from 1,441 reservoir fluid studies).

Figure 5–5 is a more useful presentation of the data of figure 5–4. Notice that reservoir fluids with initial producing gas-liquid ratios as high as 1,000,000 scf/STB exhibit dew points at reservoir temperature, i.e., they are gas condensates.

The producing gas-liquid ratios reported on figures 5–3 through 5–7 are not normalized to any standard surface facilities, i.e., number of separators, and also are not normalized to any standard operating temperatures and pressures. This lack of standardization causes most of the scatter in the data. The molecular weights of the heptanes plus fractions do have some effect; the higher molecular weights give results that tend along the bottom of the line of data points.

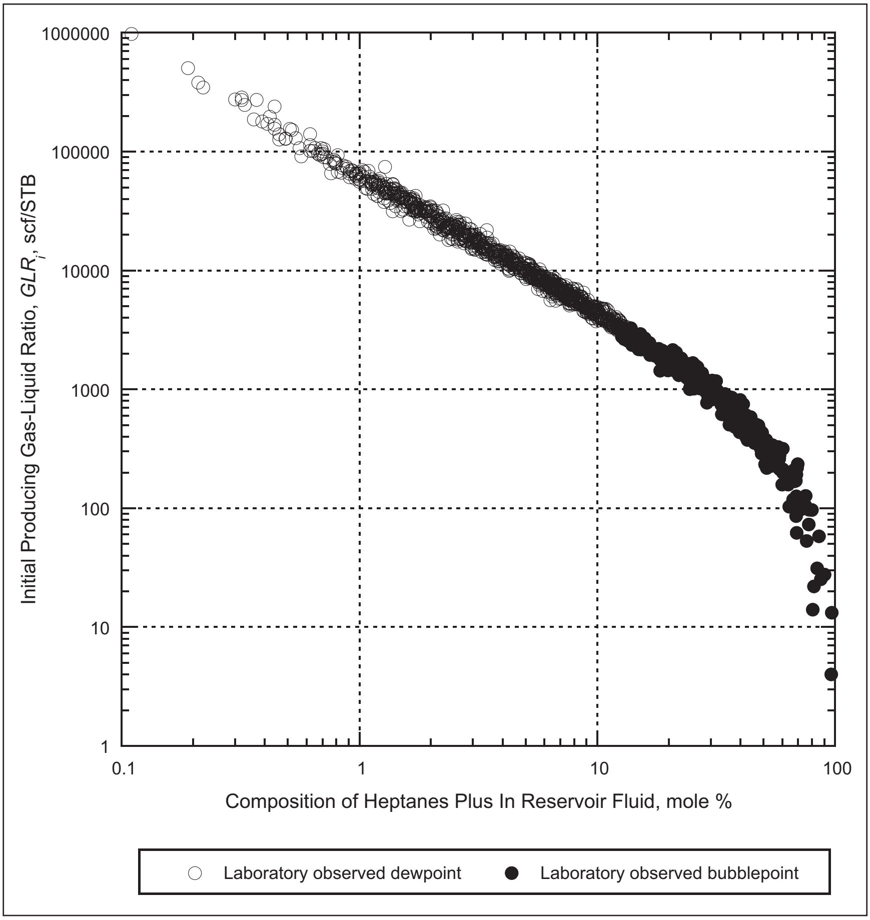

Fig. 5–5. Initial producing gas-liquid ratio correlates well with composition of heptanes plus in reservoir fluid (data from 1,451 reservoir fluid studies).

Wet gases and gas condensates

Dew points at reservoir temperatures have been observed in laboratory studies of gas condensates with initial producing gas-liquid ratios exceeding 1,000,000 scf/STB. However, at high initial producing gas-liquid ratios, the amount of condensate left in the reservoir is very small (less than about 1% of the pore volume). Apparently, nearly all gases that release condensate at the surface also release some condensate in the reservoir. Thus, these gases have dew points at reservoir temperatures and are gas condensates.

There are very few, if any, true wet gases (some condensate at the surface but no condensate in the reservoir). However, wet gas theory can be applied to gas condensates that release small amounts of condensate in the reservoir.

Rayes et al. showed that when the composition of heptanes plus in the reservoir gas is less than 4 mol%, the volume of condensate released into the reservoir is very small, and thus the gas can be treated as a wet gas (although in theory, wet gases do not release any condensate in the reservoir).[6] McCain (1994) indicated that gas condensates with initial producing ratios greater than 15,000 scf/STB can be treated as if they are wet gases.[7] However, no data was shown to confirm this value.

Figure 5–6 is another subset of the data of figure 5–4. It shows that if the initial producing gas-liquid ratios are equal to or greater than 15,000 scf/STB, the compositions of the heptanes plus in the reservoir fluids are definitely less than 4 mole percent. This confirms that gases with initial producing gas-liquid ratios of this value and higher can be treated as wet gases (although there surely will be a dew point and some retrograde condensate formed in the reservoir).

Dry gases

Figure 5–5 shows that many, if not all, reservoir gases that produce initial gas-liquid ratios much larger than 100,000 scf/STB have dew points at reservoir temperature and will release some condensate in the reservoir, and thus the gases are actually gas condensates. However, the small amount of condensate at the surface, less than10 STB/MMscf, is not enough to strongly affect the recombination calculations that are used to determine the specific gravity of the reservoir gas. Thus, the specific gravity of the surface gas can be used to calculate the properties of the reservoir gas at reservoir conditions. So, for engineering purposes, these gases can be treated as if they are dry gases (although in theory, dry gases do not release any condensate at the surface).

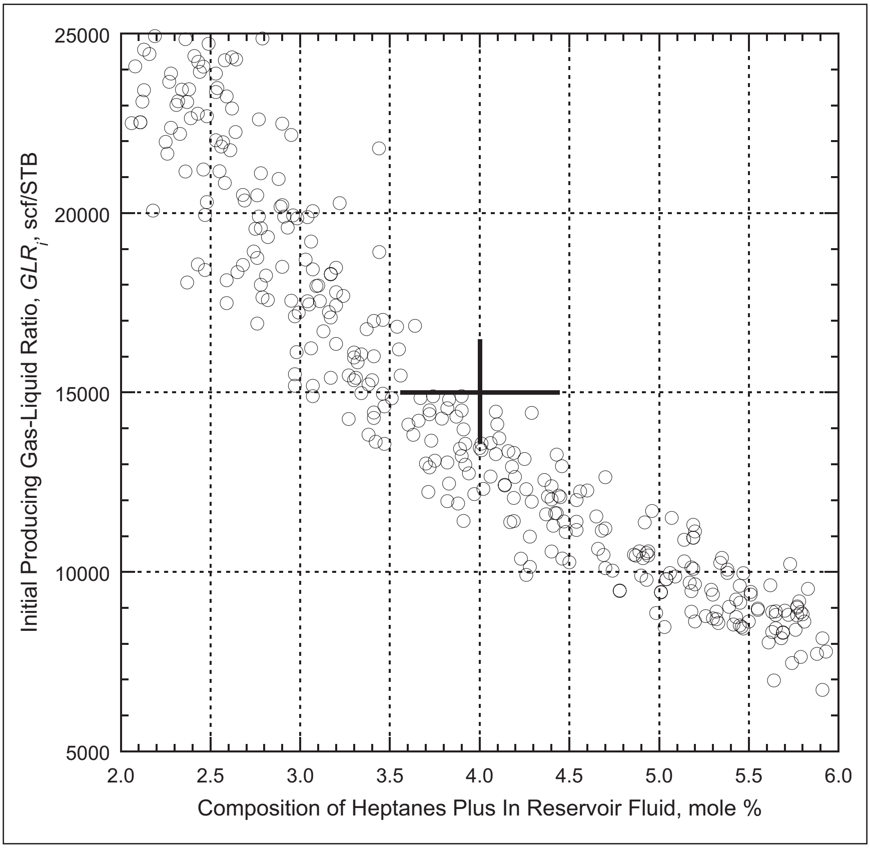

Fig. 5–6. Which gas condensates can be treated as wet gases? (Data are from 349 reservoir fluid studies.)

Summation of use of initial producing gas-liquid ratio as an indicator of fluid type

Thus, table 1 of McCain (1994) should be replaced with the table 5–1.[8]

Table 5–1. Determination of reservoir fluid type can be made with the initial producing gas-liquid ratio.

Reservoir Fluid	Initial Producing Gas-Liquid Ratio, scf/STB
Dry gas*	$100{,}000 < GLR_i$
Wet gas*	$15{,}000 < GLR_i < 100{,}000$
Gas condensate	$3{,}200 < GLR_i < 15{,}000$
Volatile oil	$1{,}900 < GLR_i < 3{,}200$
Indeterminate by GLR_i	$1{,}500 < GLR_i < 1{,}900$
Black oil	$GLR_i < 1{,}500$

* The GLR_i limits for these gases are defined for engineering applications.

Composition of Heptanes Plus as an Indicator of Reservoir Fluid Type

This book is primarily concerned with situations in which fluid property laboratory reports are not available, i.e., the composition of the reservoir fluid is not available. However, some interesting observations about the compositions of the five reservoir fluids can be determined with the data set of figures 5–4 and 5–5.

Volatile oils and black oils

Table 2 of McCain (1994) gives a precise cut-off of 20 mol% heptanes plus in the reservoir fluid between volatile oils and black oils.[9] This is not correct.

As mentioned above, there is a range of initial producing gas-liquid ratios, 1,500 to 1,900 scf/STB, in which the type of reservoir fluid is indeterminate. There is also a range of compositions of heptanes plus in which the type of reservoir fluid cannot be determined.

Figure 5–7, which is another subset of the data of figure 5–4, shows that the composition of heptanes plus in the reservoir fluid must be greater than 26.5 mol% for all of the initial producing gas-liquid ratio data to be less than 1,500 scf/STB (black oils). And the figure shows that the heptanes plus compositions must be less than 18.0 mol% for all of the initial producing gas-liquid ratios to be greater than 1,900 scf/STB (volatile oils). In the region from 18.0 mol% to 26.5 mol%, heptanes plus the initial producing gas-oil ratios vary from less than 1,500 scf/STB to greater than 1,900 scf/STB. Thus in this region, the type of oil cannot be determined using heptanes plus composition.

Gas condensates and volatile oils

The same table in McCain (1994) indicates that the transition between volatile oils and gas condensates is a composition of 12.5 mol% heptanes plus in the reservoir fluid. Additional data have been obtained and are included in figure 5–3. [This figure is similar to fig. 2 of McCain (1994) with additional data added.[10]] Figure 5–3 reveals a fairly precise transition of 12.9 mol% heptanes plus. The vertical dashed line at 12.9 mol% heptanes plus separates those reservoir fluids that exhibited bubblepoints (oils) from those fluids that exhibited dew points (gases).

Thus, if the composition of the reservoir fluid contains more than 12.9 mol% heptanes plus, the reservoir fluid is an oil. If the composition of heptanes plus in the reservoir fluid is less than 12.9 mol%, the fluid is a gas. The transition from gases to oils is very distinct; only three of the hundreds of data points do not fit this pattern.

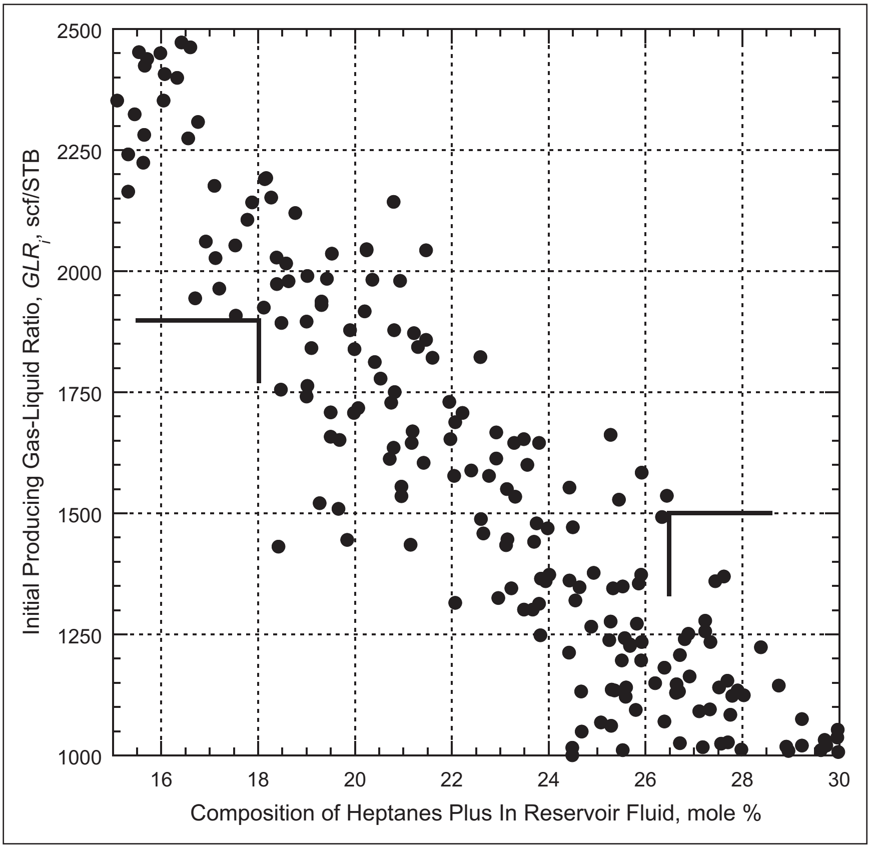

Fig. 5–7. Which composition of heptanes plus defines black oils and volatile oils? (Data are from 217 reservoir fluid studies.)

Dry gases

If the values of producing gas-liquid ratios equal to and greater than 100,000 scf/STB are accepted as indicators of gases that can be treated as dry gases, figure 5–5 indicates a corresponding composition of heptanes plus of less than approximately 0.5 mol%.

Summation of use of heptanes plus composition as indicator of fluid type

Thus, table 2 of McCain (1994) should be replaced with table 5–2.[11]

Table 5–2. Determination of reservoir fluid type can be made with composition of heptanes plus in a sample of the initial reservoir fluid.

Reservoir Fluid	**Composition of Heptanes Plus in Initial Reservoir Fluid (mol%)**
Dry gas*	$z_{C7+} < 0.5$
Wet gas*	$0.5 < z_{C7+} < 4.0$
Gas condensate	$4.0 < z_{C7+} < 12.9$
Volatile oil	$12.9 < z_{C7+} < 18.0$
Indeterminate by z_{C7+}	$18.0 < z_{C7+} < 26.5$
Black oil	$26.5 < z_{C7+}$

* The z_{C7+} limits for these gases are defined for engineering applications.

Nomenclature

B_{ob}	Oil formation volume factor, measured at bubblepoint pressure, resbbl/STB.
GLR_i	Initial producing gas-oil ratio, scf/STB; *initial* means reservoir pressure > saturation pressure, and the stock-tank *liquid* could be either oil or condensate.
R_{sb}	Solution gas-oil ratio measured at and above bubblepoint pressure, scf/STB.
z_{C7+}	Composition of heptanes plus in the reservoir fluid, mole percent (mol%).

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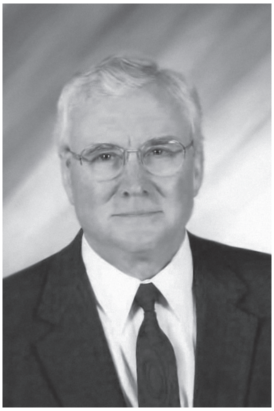
ABOUT THE AUTHORS



William D. McCain, Jr. has been a visiting professor in the Harold Vance Department of Petroleum Engineering at Texas A&M University since 1991. He also has a petroleum consulting firm, McCain Engineering.

McCain started his engineering career with Esso (now Exxon) Research Laboratories in 1956 where he assisted in research on surface processing of petroleum fluids.

He was professor and head of the petroleum engineering department at Mississippi State University from 1965 to 1976. He taught petroleum engineering at Texas A&M University from 1984 through 1987.

McCain was a consulting petroleum engineer with Cawley, Gillespie & Associates from 1987 until 1991. He was with the petroleum engineering consulting firm S.A. Holditch & Associates from 1991 until 2000, retiring as executive vice president, chief engineer, and member of the board of directors. His engineering specialties within the consulting company were properties of petroleum fluids, surface processing of petroleum, and reservoir engineering, especially for gas condensate and volatile oil fields.

He holds three U.S. patents, has more than 50 technical articles in the petroleum literature, and wrote two editions of *The Properties of Petroleum Fluids*.

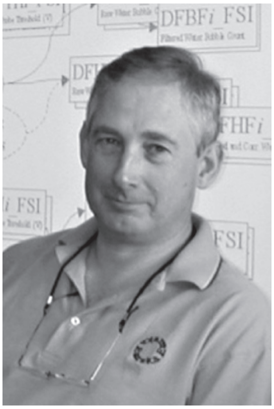
He holds a BS degree from Mississippi State University and MS and PhD degrees from Georgia Institute of Technology all in chemical engineering. He is also a graduate of the U.S. Army War College.

Dr. John P. Spivey has more than 25 years' experience in the petroleum industry, with interests in pressure transient analysis, production data analysis, reservoir engineering, continuing education, and software development. From 1984 to 1990 he worked for SoftSearch, Inc. (later Dwight's EnergyData) developing petroleum economics and engineering software.

In 1990, Dr. Spivey joined S.A. Holditch & Associates (SAH), which was purchased by Schlumberger (SLB) in 1997. While at SAH/SLB he conducted reservoir simulation, gas storage, and tight gas application studies and taught industry short courses in well testing and production data analysis. He actively participated in ongoing development of SABRE, SAH's numerical reservoir simulator, and in research in techniques for production data analysis for gas wells. He also designed and developed PROMAT, an analytical production data analysis and forecasting program, and WELLTEST, an interactive pressure transient test analysis program.

Dr. Spivey remained with Schlumberger until 2004, when he left to start his own reservoir engineering consulting company, Phoenix Reservoir Engineering, and software development company Phoenix Reservoir Software, LLC, which provides PMTx, a software package for analyzing production and production log data and forecasting future performance for multilayer low permeability reservoirs.

Since 1992, Dr. Spivey has served as a visiting assistant professor or adjunct assistant professor at Texas A&M University, teaching undergraduate and graduate classes in gas reservoir engineering and pressure transient analysis, serving on several graduate student committees. Dr. Spivey is the editor of the SPE Reprint Series Vol. 52, *Gas Reservoir Engineering*, and Vol. 57, *Pressure Transient Testing*, and coauthor of SPE Textbook Series Vol. 9, *Pressure Transient Testing*. He has published numerous papers and articles in industry journals and trade publications. Dr. Spivey has a BS in physics from Abilene Christian University, an MS in physics from the University of Washington, a PhD in petroleum engineering from Texas A&M University, and is a registered professional engineer in the state of Texas.



Dr Christopher Lenn has been employed by Schlumberger since 1990 and is currently the research director for the Fluids Department in Schlumberger Cambridge Research, Cambridge, UK. In his international career with Schlumberger he has had postings in the Middle East, the U.S. and the UK.

He entered the oil industry in 1990 with a background in engineering fluid dynamics, particularly related to multiphase flow measurement. Since joining Schlumberger he has been principally engaged in the development of cased-hole logging and monitoring technology in a variety of roles from research positions to assignments in the field and operations. While engaged in the development of novel production logging techniques for horizontal wells in the mid-1990s, he developed an interest in the issues related to the fluid properties that were needed to interpret the downhole measurements from production logging tools.

It became apparent that a wide range of oilfield fluids correlations were in use and the provenance and accuracy was not necessarily known by log interpretation practitioners. Where multiple correlations were available for the same quantity, it was not clear how to select the best-in-class correlation for the application of interest. Additionally, fluid property conversions had to be driven by readily available data. Occasionally some correlations were missing and needed development. This realization ultimately led to the collaboration with the other authors of this book.

Prior to Schlumberger, he has worked for Cranfield University (School of Mechanical Engineering) as a senior research fellow and Unilever Research (Colworth Laboratory) as a research assistant, both in the UK.

He holds a BSc degree in applied physics from London University (1976) and a PhD degree in fluid engineering from Cranfield University (1985).

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