

IMPROVED ESTIMATION OF RESERVES IN GAS CONDENSATE RESERVOIRS

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ABSTRACT

Estimating Gas condensate reserves and predicting their performance has been an area to reconcile with in the industry today. The general method of estimating the volume of gas (reserves) in a gas reservoir is the Gas Material Balance (P/Z against Gp plot). This method basically involves computing gas compressibility factors (z) at different pressures during production from the reservoir. Condensate reservoirs exhibit two phase occurrence as reservoir pressure is depleted below dew point, hence posing a challenge to accurately estimating the compressibility factor (z factor) to be used for the plot. This may ultimately result in underestimating or overestimating the Gas-Initially-In-Place in such reservoirs. The reservoir fluid acquired from the field is used in experimental Constant Volume Depletion simulation to acquire data representing the actual reservoir depletion which is validated with observed minimal deviations and presented in the PVT reports. To replace the need for time consuming and complicated experiments, it was attempted to develop a relationship or model to determine the best estimates of the two-phase compressibility factors at different pressures. In this paper, the Peng-Robinson's equation of state is employed to determine the different phase compressibility factors and combined in proportions with respect to the overall mole fractions of the two phases as a weighting factor to obtain a general relationship for the two-phase compressibility factor of the reservoir fluid. The reserve estimates from this method were compared with the field data and other methods for possible deviations and corrections. It is conclusive that this project results in the highest accuracy in condensate reserves estimation with an average error of about 1%, whereas obtainable error using the Standing-Katz single phase z factor could be up to 8%.

Keywords: *Two Phase Z factor, Peng-Robinson EOS, CVD simulation, P/Z plot, Standing Katz, z factor*

1. INTRODUCTION

Gas condensate reservoirs are characterized by production of both surface gas and varying quantities of stock-tank oil (STO). The STO is commonly referred to as "condensate" or "distillate". Typical condensate surface yields range from 10 to 300 STB/MMscf. The added economic value of produced condensate in addition to gas production makes the recovery of condensate a key consideration in developing gas condensate reservoirs; in the extreme case of a non-existent gas market, producible condensate is the only potential source of income.

At reservoir conditions, a gas condensate reservoir contains single phase gas. The gas flows from the reservoir through the production tubing to the surface separator, and then liquid condenses from the gas. Isothermal condensation of liquids in the reservoir as pressure drops below the dew point pressure constitutes the process of retrograde condensation. Liquids condensed in the reservoir are mostly lost or unrecoverable. In a gas-condensate reservoir, the initial reservoir condition is in the single phase area to the right of the critical point. As the reservoir pressure declines, the fluid passes through the dew point and a liquid drops out of the gas, the percentage of vapour decreases, but can increase again with continued pressure decline.

A main difference between a "gas" reservoir and a "gas condensate" reservoir is that a gas reservoir will not experience two hydrocarbon phases at reservoir conditions, and hence there will be no liquid condensation ("condensate loss") in the reservoir.

Furthermore, a gas reservoir would not have significant condensable surface liquids to "lose" due to retrograde condensation. This leads to the apparent contradiction that the leaner the gas condensate, the higher the condensate recovery (as a percentage of initial in place). Surface separators typically operate at conditions of low pressure and low temperature. [1]

Another important difference between gas and gas condensate reservoirs is the loss in well deliverability experienced by gas condensate reservoirs due to the buildup of significant liquid saturations near the wellbore. Gas reservoirs will not experience such deliverability loss because liquids do not condense at reservoir temperature.

2. PROBLEM FORMULATION

Condensate reservoirs exhibit two phase phenomenon at pressures below the dew point and this affects the qualitative parameters used in characterizing the fluid system (i.e. z factor considered in this project). Therefore, it is necessary to convert the single-phase parameters to a usable one for two phase system.

Calculations of initial gas and condensate in place require the use of the real gas equation of state, with some modifications.

First, we calculate volumetrically the volume of gas condensate per volume of reservoir:

$$G = \frac{379.4P_R V}{zRT_R} \quad (1)$$

There are two basic cases of every condensate system with respect to the dew point pressure; i.e. when the reservoir pressure is above the dew point pressure (when the system is strictly single gas phase) and when the pressure below the dew point.

Case I - Reservoir Pressure above Dew point Pressure

The p/z vs. Gp form of the material balance is also used for gas condensate reservoirs when reservoir pressure is above the dew point. The cumulative gas production is modified to include the condensate production. This gas is called "wet" gas. The volume of condensate produced is converted to its gas equivalent (GE) by assuming that the condensate can be expressed in terms of an ideal gas,

$$GE = V = \frac{nRT_{SC}}{P_{SC}} = \frac{a_2 N_p \rho_o}{M_o} \frac{RT_{SC}}{P_{SC}} \quad (2)$$

In summary, the material balance for a volumetric gas condensate reservoir above the dew point is,

$$\frac{P}{z} = \frac{P_i}{z_i} \left(1 - \frac{G_{PW}}{G_w} \right) \quad (3)$$

Where, Gw is initial wet gas in place, and Gpw produced wet gas.

Case II - Reservoir Pressure below Dew point Pressure

When the reservoir pressure drops below the dew point pressure, condensate drops out of the reservoir gas. This violates several of the basic assumptions implicit in the traditional gas material balance equation.

One of the earlier solutions to this problem was the Constant Volume Depletion experiment - a laboratory experiment designed to duplicate or model closely the reservoir depletion of a volumetric gas condensate reservoir, the experiment is a constant volume depletion experiment (CVD).

The standard experimental program for the analysis of a gas-condensate fluid includes

- (1) Recombined wellstream compositional analysis through C7+
- (2) Constant composition expansion (CCE), and
- (3) Constant volume depletion (CVD).

The CCE and CVD data are measured in a high-pressure visual cell where the dew point pressure is determined visually. Total volume/pressure and liquid-dropout behaviour is measured in the CCE experiment.

The CCE data for a gas condensate usually include total relative volume, Vrt, defined as the volume of gas or of gas-plus-oil mixture divided by the dew point volume. Z factors are reported at pressures greater than and equal to the dew point pressure. With relevance to this study they could be used to determine dew point z factor, Zd of the condensate fluid system.

Phase volumes defining retrograde behaviour are measured in the CVD experiment together with Z factors and produced-gas compositions through C7+. Optionally, a multistage separator test can be performed as well as TBP analysis or simulated distillation of the C7 into single-carbon-number cuts from C7 to about C20+.

Knowing the cumulative moles removed and its volume occupied as a single-phase gas at the removal pressure, we can calculate the equilibrium gas Z factor from

$$z = \frac{P\Delta V_g}{\Delta n_g RT} \quad (4)$$

The CVD experiment provides data that can be used directly by the reservoir engineer. From the data obtained, a “two-phase” Z factor is also reported. This is calculated assuming that the gas-condensate reservoir depletes according to the material balance for a dry gas and that the initial condition of the reservoir is at dew point pressure

Where G_{pw} = cumulative wellstream (wet gas) produced and G_w = Initial wet gas in place.

It is defined that the term G_{pw}/G_w equals n_p/n reported in the CVD report. From Eq. 3, the only unknown at a given pressure is Z_2 , and the two-phase Z factor is then given by

$$z_2 = \frac{P}{P_d/z_d \left(1 - \frac{n_p}{n}\right)} \quad (5)$$

3. SCOPE OF STUDY

This study is based on the Gas MBE application using Peng-Robinson equation of state (PREOS) derived two phase z factors and using existing data for liquid and vapour phase compositions (from flash) and CVD z factors in the PVT Reports by Flo-petrol for OBEN Field wet gases with appreciable amount of wet gas and condensate recovery.

Objectives

The objective of this project is to accurately or near accurately determine the amount of Gas Initially in Place (GIIP) in gas condensate reservoirs. In achieving this, the project focuses on the use of simple systematic and linear model by using overall mole fractions as a weighting factor. The project also applies Fortan and Matlab programs to automate the generation of z factor and simplify the calculations..

4. METHODOLOGY

4.1 Introduction

In this study, a novel mathematical-based approach is proposed to develop reliable model for prediction of compressibility factor of natural gas condensate using the Peng Robinson equation of state [2] based on Slot-Petersen [3] binary interaction parameters and adopting some principles from the Zudkevitch and Joffe's [4] approach to solving the Redlich Kwong EOS.

$$\begin{aligned} P/Z &= (P/Z_i)_i [1 - G_p/G] \\ Z_{2ph} &= Z_vV + Z_lL \end{aligned} \quad (6)$$

The Peng-Robinson equation was developed in 1976 at The University of Alberta in order to satisfy the following goals:

- The parameters should be expressible in terms of the critical properties and the acentric factor.
- The model should provide reasonable accuracy near the critical point, particularly for calculations of the compressibility factor and liquid density.
- The mixing rules should not employ more than a single binary interaction parameter, which should be independent of temperature pressure and composition.
- The equation should be applicable to all calculations of all fluid properties in natural gas processes.

To generate the required estimates for compressibility factors, it would involve;

- Solving the equations for two phase hydrocarbon fluids incorporating the flash calculations
- Solving the PREOS and the required binary interaction coefficients

- Defining and correlating the fugacity –and fugacity coefficient equations
- Accounting for the effects of the heavy hydrocarbon plus fractions on the phase equilibria predictions

4.2 The Peng-Robinson Equation of State for Mixtures

As written, the P-R equation is intended for description of the PVT behaviour of pure compounds.

However, it can also be used for mixtures of compounds by using "mixture-averaged" values for the equation parameters, as adopted from Ahmed T. [5-7].

$$P = \frac{RT}{V-b} - \frac{a(T)}{V(V+b)+b(V-b)} \quad (7)$$

Here, R is the gas constant, P is the absolute pressure, T is the absolute temperature, v is the molar volume, and b and a(T) are given by:

$$a(T) = 0.45724 \frac{R^2 T_c^2}{P_c} \alpha(T) \quad (8)$$

$$b(T) = 0.07780 \frac{RT_c}{P_c} \quad (9)$$

$$[\alpha(T)]^{1/2} = 1 + m \left(1 - \sqrt{\frac{T}{T_c}}\right) \quad (10)$$

$$m = 0.37464 + 1.54226\omega - 0.26992\omega^2 \quad (11)$$

Here, T_c is the critical (absolute) temperature, P_c is the critical (absolute) pressure, and ω is the Pitzer acentric factor. Thus the P-R equation has in effect just three compound-specific experimentally-determined physical properties: T_c , P_c , and ω .

Note that if the compressibility factor $z = Pv/(RT)$, then (1) can be written as a cubic equation in z :

$$z^3 - (1 - B)z^2 + (A - 3B^2 - 2B)z - (AB - B^2 - B^3) = 0$$

$$\text{Here, } B = Pb/(RT) \text{ and } A = aP/(RT)^2.$$

Let the values of parameters $a_{ii}(T)$ and b_i be the pure-component values of $a(T)$ and b , respectively, for the i (th) compound in a mixture.

Also, let x_i be the mole fraction of component i in the vapour phase of the mixture.

Then "mixing" rules are applied to compute the mixture-averaged values of $a(T)$ and b for a mixture of n different compounds as follows:

$$b_m(T) = \sum_{i=1}^n x_i b_i \quad (12)$$

$$a_m = \sum \sum x_i x_j \sqrt{a_i a_j \alpha_i \alpha_j} (1 - k_{ij}) \quad (13)$$

Where K_{ij} is the binary interaction parameter

The corresponding values of a and b for the liquid phase will be obtained by replacing x with y_i ie, mole the mole fraction of component i in the liquid phase of the mixture

NB: It is to be noted that Equations of States (EOS) are not perfect;

EOS provides self-consistent fluid properties;

- Density (oil or gas) trends are correctly predicted with pressure, temperature, and compositions (and all derived properties).
- Same phase equilibrium model for gas and liquid phases (material balance consistency).

However, predicted fluid property values may differ substantially from data. EOS are routinely “calibrated” to selected & limited experimental data, After “calibration” EOS predictions beyond range of data can be used with confidence, and EOS are extensively used in reservoir simulation.

4.3 Equation of State (EOS) calibration

Minimization of squared differences between experimental and predicted fluid properties

$$\sum_{i=1}^N \left(g_{i(\text{predicted})} - g_{i(\text{experimental})} \right)^2 = \min \quad (14)$$

These Properties (g_i) include:

- Densities, saturation pressures, Relative amounts of gas and liquid phases Compositions, etc.
- Accomplished by changing within certain limits selected EOS parameters.
- Minor adjustments (1 to 2%) of binary interaction parameters (k_{ij}) can change saturation pressures by 20 to 30%
- Different properties of the C7+fraction affect liquid dropout and densities. These properties include;
 1. Molecular weight (uncertainty is +/-10%)
 2. Specific gravity
 3. Critical properties and acentric factors which are highly dependent on correlations and cannot be easily measured.

4.4. Binary Interaction Parameter

The parameter k_{ij} is an empirically determined correction factor (called the binary interaction coefficient) that is designed to characterize any binary system formed by component i and component j in the hydrocarbon mixture. These binary interaction coefficients are used to model the intermolecular interaction through empirical adjustment of the **(α) m** term as represented mathematically.

They are dependent on the difference in molecular size of components in a binary system and they are characterized by the following properties:

- The interaction between hydrocarbon components increases as the relative difference between their molecular weights increases: $k_{i,j+1} > k_{i,j}$
- Hydrocarbon components with the **same** molecular weight have a binary interaction coefficient of zero:

$$k_{i,i} = 0$$
- The binary interaction coefficient matrix is symmetric: $k_{j,i} = k_{i,j}$

Slot-Petersen (1987) [3] and Vidal and Daubert (1978) [8] presented a theoretical background to the meaning of the interaction coefficient and techniques for determining their values. Graboski and Daubert (1978) [9] and Soave (1972) [10] suggested that no binary interaction coefficients are required for hydrocarbon systems.

To provide the modified PR EOS with a consistent procedure for determining the binary interaction coefficient k_{ij} , the following computational steps are proposed:

Step 1. Calculate the binary interaction coefficient between methane and the heptanes-plus fraction from:

$$k_{c1-c7+} = 0.00189T - 1.167059$$

where the temperature T is in °R.

Step 2. Set: $k_{CO_2-N_2} = 0.12$
 $k_{CO_2\text{-hydrocarbon}} = 0.10$

$$k_{N2\text{-hydrocarbon}} = 0.10$$

Step 3. Adopting the procedure recommended by Petersen (1989)[11], calculate the binary interaction coefficients between components heavier than methane (e.g., C2, C3) and the heptanes-plus fraction from:

$$k_{Cn - C7+} = 0.8 k_{C(n-1) - C7+}$$

where n is the number of carbon atoms of component Cn;

e.g: Binary interaction coefficient between C2 and C7+ is $k_{C2 - C7+} = 0.8 k_{C1 - C7+}$

Binary interaction coefficient between C3 and C7+ is $k_{C3 - C7+} = 0.8 k_{C2 - C7+}$

Step 4. Determine the remaining kij from:
$$k_{ij} = k_{C1-C7+} \left[\frac{(M_j^5) - (M_i^5)}{(M_{C7+}^5) - (M_i^5)} \right] \quad (15)$$

where M is the molecular weight of any specified component.

For example, the binary interaction coefficient between propane C3 and butane C4 is:

$$k_{C3-C4} = k_{C3-C7+} \left[\frac{(M_{C4}^5) - (M_{C3}^5)}{(M_{C7+}^5) - (M_{C3}^5)} \right] \quad (16)$$

4.5 Solving the Peng Robinson equation of state (step by step)

The liquid and vapour phase mole fractions x_i , y_i , and the binary interaction coefficient k_{ij} , are used to solve the PR-EOS and determine the liquid phase Z_L factor and the vapour phase z_v factor. The binary interaction parameter is such that every two components, either molecules of different substances (e.g. C1 and C4) or those of same compound (e.g. C1 and C1) are all considered.

Therefore every multicomponent system results has k_{ij} existing in an $n \times n$ matrix order, where n is the number of components i.e. for a two component system there exist a 2×2 matrix order for k_{ij} values including, k_{11} , k_{12} , k_{21} , and k_{22} .

Take a_m and b_m from (12) and (13)

$$\mathbf{B} = \mathbf{P}b_m / (\mathbf{RT}) \text{ and } \mathbf{A} = \mathbf{a}_m \mathbf{P} / (\mathbf{RT})^2$$

Same process is carried out for the liquid phase using the flashed fractional compositions (y_i) and the values are computed and then substituted into the general equation;

$$z^3 - (1 - \mathbf{B})z^2 + (\mathbf{A} - 3\mathbf{B}^2 - 2\mathbf{B})z - (\mathbf{A}\mathbf{B} - \mathbf{B}^2 - \mathbf{B}^3) = 0$$

Using the Newton Raphson iterative method, solve the above Equation and obtain Z_L and Z_v , i.e., smallest and largest roots respectively, for liquid and vapour phases.

4.6. Flash Calculations

Flash calculations are an integral part of all reservoir and process engineering calculations. They are required whenever it is desirable to know the amounts (in moles) of hydrocarbon liquid and gas coexisting in a reservoir or a vessel at a given pressure and temperature. These calculations are also performed to determine the composition of the existing hydrocarbon phases.

Given the overall composition of a hydrocarbon system at a specified pressure and temperature, flash calculations are performed to determine:

- Moles of the gas phase n_v
- Moles of the liquid phase n_L
- Composition of the liquid phase x_i
- Composition of the gas phase y_i

The computational steps for determining n_L , n_v , y_i , and x_i of a hydrocarbon mixture with a known overall composition of z_i and characterized by a set of equilibrium ratios K_i are summarized in the following steps:

Step 1. Calculation of n_v : Equation can be solved for n_v by using the *Newton-Raphson iteration techniques*.

In applying this iterative technique:

- Assume any arbitrary value of n_v between 0 and 1, e.g. n_v could be 0.5, $n_v = 0.5$. A good assumed value may be calculated from the following relationship, providing that the values of the equilibrium ratios are accurate:

$$n_v = A/(A-B) \quad (17)$$

$$A = \sum_i [z_i(K_i - 1)] \quad (18)$$

$$B = \sum_i [z_i(K_i - 1)/K_i] \quad (19)$$

- Evaluate the function $f(n_v)$ as given by Equation 3.26 using the assumed value of n_v
- If the absolute value of the function $f(n_v)$ is smaller than a preset tolerance, e.g., 10^{-15} , then the assumed value of n_v is the desired solution.
- If the absolute value of $f(n_v)$ is greater than the preset tolerance, then a new value of n_v is calculated from the following expression

$$(n_v)_n = n_v - f(n_v)/f'(n_v) \quad (20)$$

$$\text{Where } f(n_v) = \sum_i \left[\frac{z_i(K_i - 1)}{n_v(K_i - 1) + 1} \right] \quad (21)$$

$$f' = -\sum_i \left[\frac{z_i(K_i - 1)^2}{(n_v(K_i - 1) + 1)^2} \right] \quad (22)$$

where $(n_v)_n$ is the new value of n_v to be used for the next iteration.

- The above procedure is repeated with the new values of n_v until convergence is achieved.

Step 2. Calculation of n_L : Calculate the number of moles of the liquid phase from;

$$n_L = 1 - n_v \quad (23)$$

Also;

Wilson's Correlation [12] is sufficient in determining the K values at low pressures.

$$K_i = \frac{P}{P_{ci}} \exp \left[5.37(1 + \omega_i) \left(1 - \frac{T_{ci}}{T} \right) \right] \quad (24)$$

Whitson and Torp (1983) [13-14] reformulated Wilson's equation to yield accurate results at higher pressures.

Wilson's equation was modified by incorporating the convergence pressure into the correlation, to give:

$$K_i = \left(\frac{P_{ci}}{P_k} \right)^{A-1} \frac{P_{ci}}{P} \exp \left[5.37A(1 + \omega_i) \left(1 - \frac{T_{ci}}{T} \right) \right] \quad (25)$$

$$A = 1 - \left(\frac{P_{ci}}{P_k} \right)^{0.7} \quad (26)$$

where p = system pressure, psig
 P_k = convergence pressure, psig
 T = system temperature, °R

This project is provided with CVD data which are used to infer the mole fraction of the vapour phase and the liquid phase thus: In the laboratory, the removed gas (well-stream) is brought to atmospheric conditions, where the amount of surface gas and condensate are measured. Surface compositions y_g and x_o of the produced surface volumes from

the reservoir gas are measured, as are the volumes ΔV_o and ΔV_g , densities ρ_o and ρ_g and oil molecular weight M_o . From these quantities, we can calculate the moles of gas removed, Δn_g . [15]

$$\Delta n_g = \frac{\Delta V_o \rho_o}{M_o} + \frac{\Delta V_g}{379} \quad (27)$$

These data are reported as cumulative well-stream produced, n_g , relative to the initial moles n .

$$\left(\frac{n_g}{n} \right)_k = \frac{1}{n} \sum_{j=1}^k (\Delta n_g)_j \quad (28)$$

Then, we have that;

$$\overline{n_g} = \left(\frac{\Delta n_g}{n} \right)_k = \left(\frac{n_p}{n} \right)_k - \left(\frac{n_p}{n} \right)_{k-1} \quad (29)$$

Where $\overline{n_g}$ = vapour phase mole fraction (of the produced fluid in cvd).

The result here would then be further iterated and modified to be used in the required relation.

5. RESULTS AND DISCUSSION

- The calculated z factors from this project is to be compared with those obtained from conventional two phase z factors from constant volume depletion and constant composition expansion experiments in the lab and the deviation is observed.
- The equations are programmed and automated to generate the two phase z factor values and maintain accuracy.
- The p/z plot is then observed and correlated with other sources of reserves estimates available and the deviations are observed and adjusted.
- It is worthy of note that setting the k as zero and using the exact values yield very slight difference in the results for z factors of about a maximum 0.0001 tolerance
- Inaccuracy of the k values in determining actual mole fraction of the phase reflects a notable deviation from experimental values; also, since it is known that the vapour mole fraction is 1.000 at the dew point, z factor here is calculated strictly by using $z = PV/nRT$ for that initial point.
- The estimation of reserves based on this method is comparing the estimates of GIIP based on two phase z factors obtained from this project and those obtained from the standing –Katz chart based on the Hall-Yarborough[16-17] empirical representation, and the two phase z factors for the limited Rayes et al correlation[18]
- This project have provided its result with a very reliable source of data from PVT experiments actually performed to determine accurately the composition of the gas produced by flashing from several pressures, when the composition in mole percent of all the component in their several phases are not correct this process will yield erroneous results.
- When the phase mole fractions (weighting factors) becomes difficult to estimate due to inaccuracy in K -value calculations; the Vapour phase z -factor (Z_v) is still a very good approximation of the Z -2phase in order to conveniently estimate the reserves using the P/Z plot.

5.1 Results

5.1.1. CASE 1 (OBEN 25, BHT= 158°F)

Table 1: The Z-factor Results for OBEN 25

Pressure (psig)	Z (reported)	Z _v (PREOS)	Z _L (PREOS)	V (overall vapour phase mole fraction corrected)	Z _{two-phase} (This project)	Z _{1ph} (single- phase from the Standing- katz [19,20] chart)	Z _{rayes} (Rayes correlation for 2 phase z factor)
3700	0.848	0.848(set)	-	1.000000	0.848000	0.806	0.781319
3587	0.844	0.849172	1.806350	0.994654	0.854289	0.796	0.771561
3397	0.835	0.839848	1.713030	0.994509	0.844642	0.780	0.755240
3117	0.826	0.828028	1.575190	0.997300	0.830045	0.759	0.731381
2844	0.821	0.819178	1.440420	0.997067	0.820975	0.741	0.708341
2550	0.820	0.813288	1.294850	0.986063	0.819981	0.729	0.683774
2254	0.823	0.811991	1.147800	0.968681	0.822508	0.726	0.659298
1966	0.832	0.815996	1.004220	0.916438	0.831724	0.733	0.635731
1691	0.844	0.825288	0.866626	0.548812	0.843939	0.750	0.613456
1426	0.860	0.839634	0.733545	0.862754	0.824919	0.776	0.592202
1091	0.887	0.865371	0.564571	0.937135	0.846461	0.820	0.56563
601	0.933	0.916960	0.315777	0.933024	0.901732	0.895	0.527359
0	1.000	0.997837	0.872266	0.983349	0.995746	0.990	0.500000

Percentage Error in reserves estimation (referenced to report)

At (P/Z) dimensionless = 0.00

$$\text{For This project} = \left| \frac{(Gp/G)_{\text{calculated}} - (Gp/G)_{\text{correct}}}{(Gp/G)_{\text{correct}}} \right| \times 100\% = \left| \frac{0.9954 - 1}{1} \right| = 0.46\%$$

$$\text{For Single phase Z} = \left| \frac{1.01895 - 1}{1} \right| = 1.9\%$$

$$\text{For Rayes correlation} = \left| \frac{1.138 - 1}{1} \right| = 13.8\%$$

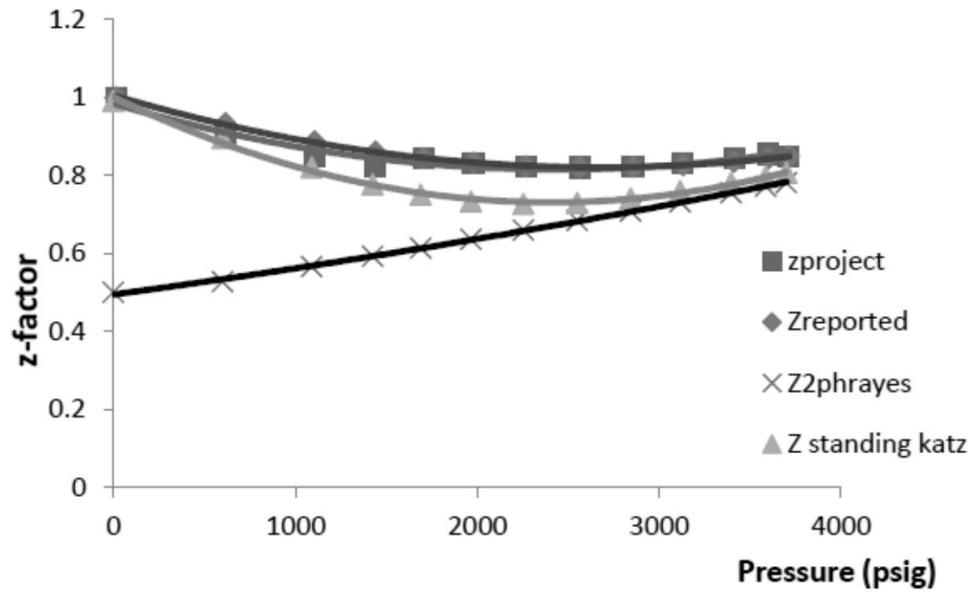


Fig.1. Z factor against Pressure OBEN 25

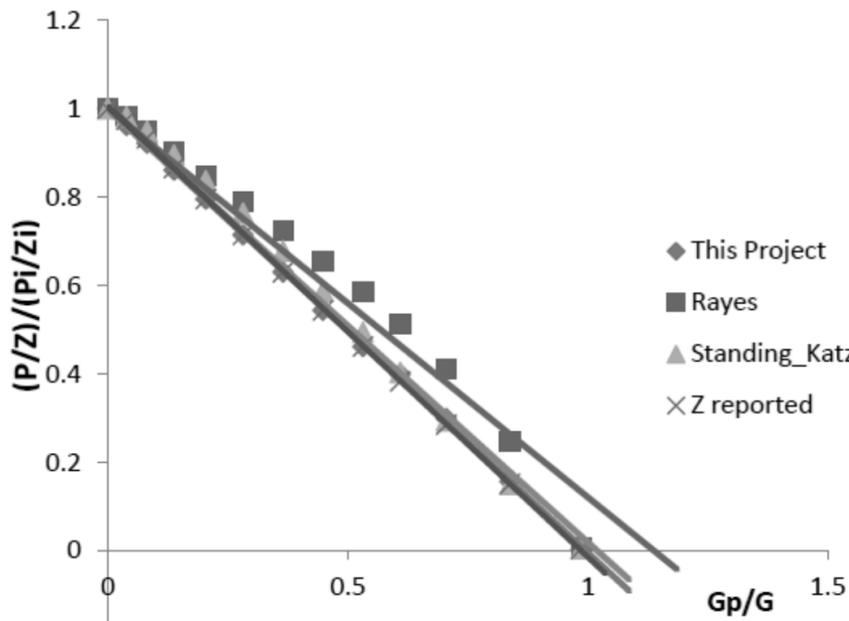


Fig.2. P/Z against against Gp Plot (DIMENSIONLESS FORM) OBEN 25

5.1.2. CASE 2 (OBEN 23, BHT= 143°F)

Table 2: The Z-factor Results for OBEN 23

Pressure (psig)	Z (reported)	Z _v (PREOS)	Z _L (PREOS)	V (overall vapour phase mole fraction corrected)	Z _{two-phase} (This project)	Z _{1ph} (single-phase from the Standing-katz chart)	Z rays (Rays correlation for 2 phase z factor)
3100	0.847	0.847	-	1.00	0.847	0.869586	0.863524
3070	0.843	0.844051	1.61549	0.998641	0.845100	0.868187	0.861278
2885	0.832	0.839148	1.52027	0.989722	0.846148	0.860511	0.847506
2635	0.82	0.834655	1.39132	0.974991	0.848576	0.853009	0.829097
2369	0.811	0.832969	1.25375	0.952726	0.852861	0.849119	0.809765
2060	0.808	0.835626	1.09345	0.911760	0.858377	0.850451	0.787638
1775	0.812	0.843004	0.945098	0.811067	0.862293	0.857490	0.767544
1480	0.83	0.856003	0.790995	0.600010	0.830000	0.870480	0.747063
1170	0.862	0.875659	0.628411	0.944756	0.862000	0.889767	0.725888
840	0.901	0.903081	0.454557	0.995360	0.901000	0.915614	0.703740
580	0.932	0.928990	0.316969	0.995130	0.926010	0.939066	0.686576
325	0.964	0.957634	0.181461	0.991931	0.951372	0.964140	0.669985

Percentage Error in reserves estimation (referenced to report)

At (P/Z) dimensionless = 0.00

$$\text{For This project} = \left| \frac{(Gp/G)_{\text{calculated}} - (Gp/G)_{\text{correct}}}{(Gp/G)_{\text{correct}}} \right| \times 100\% = \left| \frac{1.018223 - 1}{1} \right| = 1.8\%$$

$$\text{For Single phase Z} = \left| \frac{1.02596 - 1}{1} \right| = 2.6\%$$

$$\text{For Rays correlation} = \left| \frac{1.1342 - 1}{1} \right| = 13.4\%$$

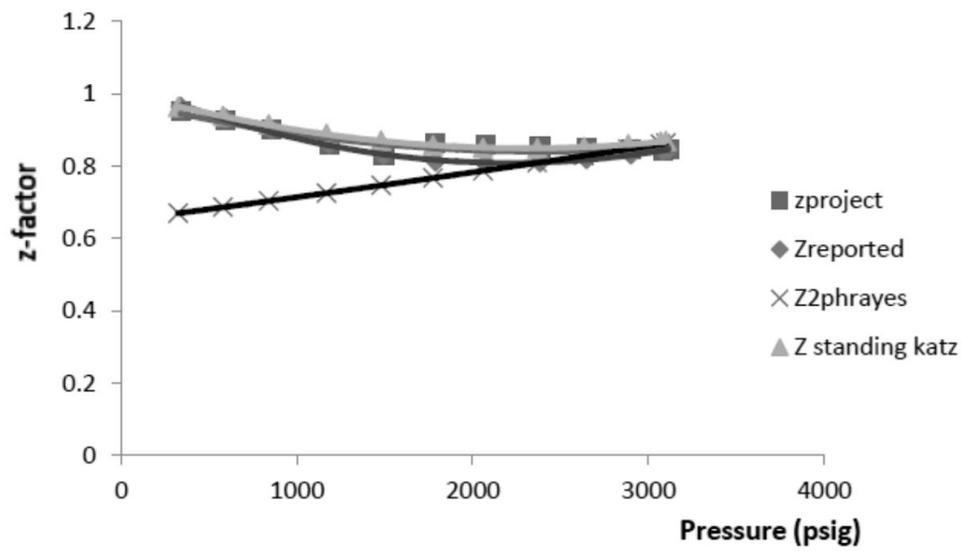


Fig.3. Z factors against Pressure OBEN 23

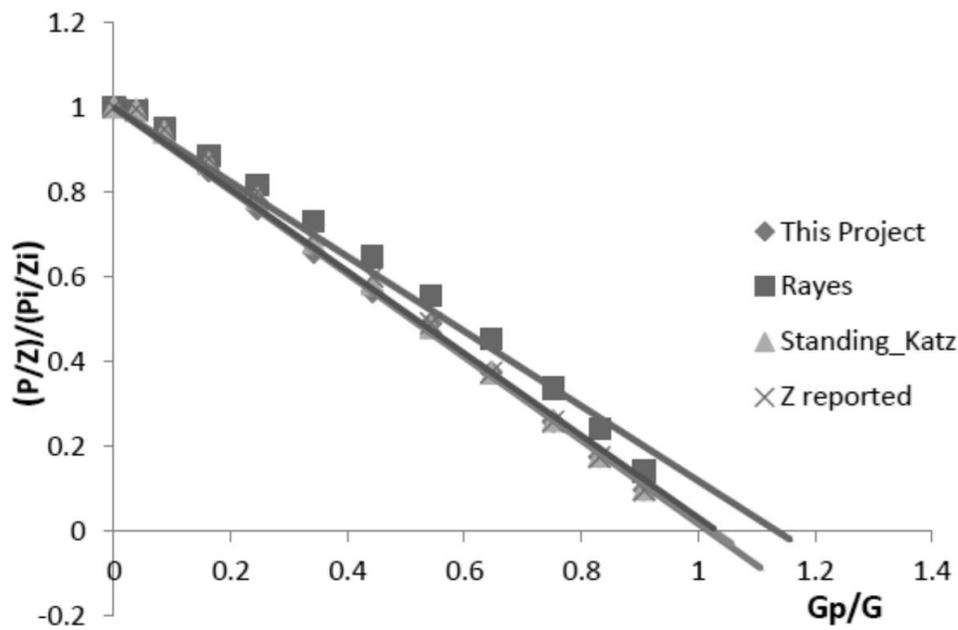


Fig.4. P/Z against against Gp Plot (DIMENSIONLESS FORM) OBEN 23

The variations in calculated error values in the different cases are majorly due to changes in compositions and nature of the fluid.

6. CONCLUSIONS

This project has been successful in determining the percentage error resulting from using the single phase z-factor in the estimation of gas condensate reserves and the level of accuracy obtainable using the two phase z factor in the p/z plot for a gas condensate system.

This project has proposed an easier two phase z factor determination program and automated it to relate the changes and depletion behaviour of the condensate system. The project has also provided a method to estimate the reserves using the general p/z plot as a means of quickly determining a reliable estimate of two phase z-factor using four basic input parameters including; pressure (P), temperature (T), composition (vapour(xi) and liquid(yi) phases from flash) and vapour-liquid equilibrium ratios (K-values); as opposed to the experimental methods that would require serious scrutiny and caution and reduce cost as well as to monitor the reservoir changes.

7. RECOMMENDATION

Accuracy might be affected by the iterative process and convergence; the correctness of the various parameters should be verified; especially the compositions (from flash), the acentric factors, the critical point values (Pc and Tc), the vapour-liquid equilibrium ratios (Ki) at the various depletion stages, which are the main sources of error. Other values from literature should be ensured correct before using this model i.e. the convergence pressure for k values binary interaction parameters etc. as a number of data used in this project results were generated from lots of trials and error k values and theory based approaches rather than an actual equilibrium ratio expert data.

The results are to be compared with all other possible sources of two phase z factor to observe the deviation to validate the model.

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