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COMPARATIVE ANALYSIS OF THREE EQUATIONS OF STATE IN THE PREDICTION OF LIQUID

PROPERTIES

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by

Goodness Imonighavwe

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The thesis of Goodness Imonighavwe was reviewed and approved* by the following:

Michael A. Adewumi

Professor of Petroleum and Natural Gas Engineering

Thesis Advisor

Turgay Ertekin Professor of Petroleum and Natural Gas Engineering Head of Department of Energy and Mineral Engineering

Eugene Morgan

Assistant Professor of Petroleum and Natural Gas Engineering

Sanjay Srinivasan Professor of Petroleum and Natural Gas Engineering Interim Associate Department Head for Graduate Education

*Signatures are on file in the Graduate School.

ABSTRACT

Hydrocarbon Phase Behavior modelling is extremely important in the characterization of fluids because this helps in obtaining fluid properties that are important in the design and optimization of various production processes and facilities. These fluid properties were initially obtained through laboratory measurements until the introduction of cubic equations of state which provided fairly accurate results in less time. A major drawback of cubic equations of state is measurement of liquid densities. This is extremely important when dealing with reservoir fluids such as gas condensates which exhibit a retrograde behavior of splitting into gas and liquid phases during isothermal expansion. The goal of this study is to develop a tool that will be able to detect when a hydrocarbon mixture undergoes a phase split and also be able to provide a more accurate prediction of condensate properties than the presently used Equations of state. This tool is developed by the incorporation of a newly developed equation of state into flash calculations and the predictions made are not only compared to experimental values but also to the predictions made by the popular two parameter, Peng-Robinson (PR) and three parameter Patel-Teja's (PR) equations of state.

The new equation of state proved to be more accurate in the prediction of the composition, density and viscosity of the liquid phase when compared to Peng-Robinson and It also showed great accuracy in the predictions of these properties at higher temperatures, where Patel-Teja failed to excel.

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LIST OF SYMBOLS

ROMAN SYMBOL	DEFINITION
a	Attraction Parameter term
A	Dimensionless Constant
b	Repulsive term parameter, co-volume
В	Dimensionless Constant
c	Third Parameter for three parameter EOS
С	Dimensionless Constant
f	fugacity
Р	Pressure (psia)
P _c	Critical Pressure (psia)
R	Universal Gas Constant
т	Absolute temperature (Rankine)
T_c	Critical Temperature
V	Molar Volume
Ζ	Compressibility Factor
Z _c	Critical Compressibility Factor

GREEK SYMBOL DEFINITION

α	Temperature dependent function
ζ _c	Predicted Critical Compressibility
ω	Acentric Factor
Ω	Dimensionless Coefficient

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CHAPTER 1

INTRODUCTION

Pressure (P), temperature (T), composition (x or y) and volume (V) are essential in order to accurately predict thermophysical or transport properties of hydrocarbon mixtures. These properties are of great importance in the industry because they are required in the design and operation of various equipment needed in the production, processing and transportation of petroleum. According to Assel et al, a relatively minuscule improvement in the prediction of these properties can translate to improvement in operational efficiency and thereby lead to savings in the cost of operations. In order to obtain these properties, an enormous amount of PVT data has been measured in the laboratory, however this process presents several challenges:

- 1) The process of designing the experiments can be extremely cumbersome.
- 2) The procedure is extremely expensive due to cost of purchasing the equipment.
- 3) There is difficulty in replicating all the conditions observed in the reservoir.
- 4) The possibility of human or precision error exists.

The above disadvantages highlight why it is impossible to generate PVT data for all fluids and has therefore led to the development of predictive models such as corresponding states theory (CST), equation of state (EOS) and activity coefficient model (ACM).

CST models are based on a theorem proposed by Van der Waals (1873) which establishes the relationship between reduced temperature (T_r), reduced pressure (P_r) and the compressibility factor. A limitation of the CST model is its applicability to only simple fluid calculations hence highlighting its failure when handling complex hydrocarbon systems (Nwankwo, 2014). Several modifications have been made to this model in order tackle this problem; the most notable is the introduction of the acentric factor by Pitzer (1939). The acentric factor is a measure of deviation from spherical symmetry in a molecule.

ACM uses activity coefficients data in place of fugacity coefficients for liquid phase property predictions. They are commonly applied to systems with complex polar fluids. Examples of such equations of state include the Martin-Stanford equation (1974) and the Han-Starling equation (1972). The drawback of this model is that its predictive capability has only been successful at low pressure and for polar fluid systems. The systems of interest in this study include but are not limited to polar hydrocarbon systems.

Equation of state (EOS) is a thermodynamic equation that attempts to express the relationship that exists between pressure, temperature, volume and composition in an analytical form. These equations can be classified into three types: Empirical, theoretical and semi-theoretical equations of sate. Empirical equations of state constitute of high order polynomials that require fitting to large amounts of experimental data (Nwankwo, 2014). The primary shortcoming of this kind of EOS is that it is limited to the fluid systems of interest and cannot be extrapolated to other fluids. The theoretical EOS also known as the Virial-Type EOS is based on either statistical mechanics or kinetic theory which involves intermolecular forces. Examples of such equations include Carnahan and Starling (1972), Beret and Prausnitz (1975) and Gironi and Marrelli (1976). Its downside is its inability to properly represent phase equilibriums. The semi-theoretical equations combine theory with a limited amount of experimental data. This type of EOS has shown to be good at representing both liquid and

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vapor behavior when combined with the appropriate thermodynamic relations. The advantage of EOS over Activity Coefficient Models is that you do not have to identify reference states (solid, liquid or gas). For ACM, different reference states have different activity coefficients and therefore a problem arises when you have components in a phase that is neither solid, liquid or gas (for example supercritical fluids). For these reasons our study will focus on Equations of State, specifically Semi-theoretical EOS.

The first equation of state capable of making vapor-liquid equilibria calculations was developed by Van der Waals (1873). Since then, an enormous amount of EOS has been developed ranging from simple expressions that have one or two constants to intricate expressions which contain as many as 50 constants (Patel, 1980). The more complex EOS such as Benedict-Webb-Rubin EOS (1940) are used for high precision work but are generally not utilized for thermodynamic calculations such as multi-component vapor-liquid equilibrium ratios. This is because they require extremely complicated manipulations, extremely long iterative calculations and a large amount of computer storage. In addition, the constants needed for these complex EOS at different PVT conditions require a lengthy process in order to be established. These issues have therefore made shorter equations of state more appealing. Many of these simpler equations are cubic in volume meaning the volumes can be calculated analytically at specified pressure and temperature. These equations can be classified as two or three parameter EOS.

1.1 Problem Statement

Inaccurate predictions of liquid properties has been a major drawback of some of the most prominent equations of state currently used in the oil and gas industry. Although certain modifications such as Peneloux volume shift have been implemented by several equations of state to improve accuracy, there is still a lot of room for improvement. A new equation of state was recently developed with the goal of further improving the predictions of liquid densities but this equation was only applied to single component systems which is not a reflection of the natural systems encountered in the field.

This work extends the application of the new equation to multi-component light hydrocarbon systems and measures properties such as the liquid composition, density and viscosity of these mixtures. In addition, it is important for any good equation of state to be able to handle special natural gas systems known as condensate systems which exhibit a retrograde behavior of producing liquids (also known as condensates) during isothermal compression. When such behavior is manifested, most equations of state are not able to accurately predict the properties of these condensates due to inability to accurately predict liquid properties. The primary objective of this work is to evaluate the relative performance of this new Equation of State in predicting liquid density and other liquid properties, relative to the performance of two of the more popular ones in the oil and gas industry, namely: Peng-Robinson (PR) and Patel-Teja (PT). In order to evaluate the performance of the new equation, the predictions of these equations are compared to experimental data.

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The implications of this improvement is important in every facet of petroleum engineering. In the area of reservoir engineering, the application of Equation of state is important when dealing with tight pore reservoirs. This is because in such type of reservoirs, capillary pressure plays an important role in controlling the distribution of fluid and this pressure is related to the difference between pressures of the oil and water phase. Accurate liquid predictions will lead to better calculation of the pressure in the oil phase which translates to accurate predictions of capillary pressure. In addition, improvements in the prediction of liquid properties is important for the analysis of multi-phase flow (gas-condensate-water) in pipeline due to pressure continuously changing along the pipeline hence varying the amount and composition of the condensate present.

CHAPTER 2

LITERATURE REVIEW

2.1 The Ideal Gas Equation of State

The ideal gas equation was the first and simplest equation of state developed to describe the volumetric and thermodynamic properties of an ideal gas which is approximately the equivalent of the limiting behavior of a real gas as the pressure gets ever closer to zero. This equation was developed from the kinetic theory of gases and was founded on the following assumptions:

- i) Inter-molecular interactions (attractive or repulsive forces) are non-existent.
- The collisions between molecules and between the molecules and the walls of the container are elastic meaning no energy is lost during collision.
- iii) The molecules are in constant random motion in straight lines.
- iv) The volume occupied by the molecules is negligible compared to the volume of the container leading to the attractive forces between the molecules being negligible since this force weakens as distance increases between the molecules.
- v) Due to non-existent attractive forces, the kinetic energy is only a function of temperature.

After the exhaustive study of the behavior of ideal gases, two mathematical relationships were established and served as the foundation for the ideal gas equation of state:

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- For a given mass of ideal gas, volume is inversely proportional to pressure when temperature is constant. This is regarded as Boyle's Law.
- For a given mass of ideal gas, the volume is directly proportional to temperature at constant pressure. This is regarded as Charles Law.

The two mathematical relationships were then combined by Clapeyron in 1834 to give the analytical expression of the ideal gas equation of state:

$$PV = nRT \tag{2.1}$$

Where:

P represents pressure of the fluid (psia); n represents number of moles; R represents universal gas constant $(\frac{psia*ft^3}{lbmole^{o_R}})$; V represents the molar volume occupied by the fluid; T represents absolute temperature (°R).

The universal gas constant is calculated at standard conditions (P = 14.7 psia and $T = 60^{\circ}F$) knowing that the volume occupied by gases at this condition is 378.6 ft³. Below is a diagram showing the construction of a P-V diagram for an ideal gas:



Figure 2-1: P-V diagram for ideal gases (Adewumi, 2016)

As expected, the major challenge of this equation of state is its inability to represent the behavior of real fluids.

2.2 Real Gases:

The ideal EOS has no practical applications because fluids do not behave ideally however knowing the behavior of ideal gases serves as the foundation for which the behavior of real gases can then be understood. Below is the P-V behavior of a pure substance:



Figure 2-2: P-V diagram for a pure substance (Adewumi, 2016)

Comparing figures 2-1 and 2-2, it is observed that at low pressures the pure substance exhibits ideal gas behavior but generally real gas behavior deviates from that ideal gas and this deviation is more pronounced at higher pressures and lower temperatures. Some of the reasons for this deviation include:

- i) Gas molecules occupying more significant volumes
- ii) All gas molecules do not have the same shapes and sizes.
- iii) Molecular collisions are not perfectly elastic
- iv) Existence of intermolecular forces such as attraction and repulsion forces

In addition to not being able to account for these deviations, the ideal gas equation is incapable of representing certain processes that real gases go through such as the transition from one phase to another during the process of condensation at subcritical temperatures.

In order to account for this apparent discrepancy between real gas and ideal gas behavior, a correction factor popularly referred to as the compressibility factor is introduced and usually denoted as 'z'. The compressibility factor is the ratio of the real volume occupied by the gas to the volume predicted by the ideal EOS model:

$$Z = \frac{V_{real}}{V_{ideal}}$$
(2.2)

Where
$$V_{ideal} = \frac{nRT}{P}$$
 (2.3)

Put equation (2.3) in equation (2.2)

$$PV = ZnRT \tag{2.4}$$

Where equations (2.4) represents the equation of state that can be applied to real gases.

It is important to observe that at z = 1, the real gas begins to exhibit the behavior of an ideal gas. At high pressures, the number of collisions between the gas molecules and the effects of forces due to repulsion is apparent thereby making the volume occupied by the real gas

greater than the volume predicted by the ideal EOS model (Z>1). At lower pressures the effects of attractive forces are more dominant thereby making the volume occupied by the real gas less than the volume predicted by the ideal gas model (Z<1).

There are numerous methods for calculating this correction factor including empirical correlations such as the compressibility factor chart developed by Standing-Katz and the virial-expansion coefficients method. The major challenge with these procedures is that the compressibility factor varies with temperature, composition and pressure, therefore mathematical manipulations cannot be made directly. This was one of the major motivations for in depth research into equations of state.

2.3 Cubic Equations of State

As stated previously, there are different types of equations of the state but for this study the focus will be on the cubic equations of the state which falls under the semi theoretical EOS model. The reasons for this choice include (Nwankwo, 2014):

- There several properly developed analytical algorithms in the field of mathematics such as Newton Raphson that can be easily applied to solve the cubic expressions of the EOS being used.
- 2) These type of equations are extremely popular in oil and gas industry for making reservoir engineering calculations, phase equilibria calculations and predicting values of thermodynamic properties of hydrocarbon fluids.

- All cubic equations of state are originally derived for pure components and extending their applications to mixtures is relatively easy through the implementation of mixing rules.
- 4) Most Cubic Eos provide avenues for which they can be fine-tuned in order to improve their predictions of both thermodynamic and volumetric properties.

Generally, the cubic equations of state are grouped into two to four parameter equations of state. For this study, all of the equations of state explored fall in the two or three parameter EOS category.

2.3.1 Van Der Waals (VDW) Equation of State (Van der Waals, 1873)

The first attempt to successfully model the behavior of real gas was carried out in 1873 by Johannes Diderick Van der Waals. In trying to develop this EOS, Van der Waals accounted for two ideal gas assumptions: gas can have zero molecular volume and intermolecular forces (attraction and repulsion) are negligible. Accounting for these forces led to being qualitatively able to predict the coexistence of the liquid and vapor phase. The Van der Waals equation is expressed in the following form:

$$\left(P + \left(\frac{a}{(\nu)^2}\right)\right)(V - b) = RT$$
(2.5)

Where:

P represents pressure of the fluid (psia); R represents universal gas constant $(\frac{psia*ft^3}{lbmole^{o_R}})$; V represents the molar volume occupied by the fluid; T represents absolute temperature (°R).

The constants 'a' and 'b' are substance specific constants where 'a' represents an approximate measure of the attraction forces and 'b' which is also referred to as the co-volume, represents the smallest volume that one mole of molecules can occupy. Also since this equation possesses two parameters ('a' and 'b'), it is referred to as a two parameter equation of state. It is important to observe that as 'a' and 'b' approach zero, equation (2.5) collapses to the ideal Eos and in order to evaluate 'a' and 'b', criticality condition conditions are applied. Criticality conditions refer to taking the first and second derivative at the critical point. As observed from figure (2-3), the critical point is the highest point on the p-v envelope and it is also the point of inflexion (that is change of curvature occurs), therefore the first and second derivatives can be expressed mathematically as follows:

$$\left(\frac{\partial p}{\partial \overline{V}}\right)_{Pc,Tc} = \left(\frac{\partial^2 P}{\partial \overline{V^2}}\right)_{Pc,Tc} = 0$$
(2.6)



Figure 2-3: P-V isotherms for a pure substance (Adewumi, 2016)

Writing equation (2.5) in P explicit form:

$$P = \frac{RT}{V-b} - \frac{a}{V^2} \tag{2.7}$$

Applying criticality conditions to equation (2.7):

$$\left(\frac{\partial p}{\partial \overline{V}}\right)_{T_c} = \frac{-RT_c}{(V_c - b)^2} + \left(\frac{2a}{(V_c)^3}\right) = 0$$
(2.8)

$$\left(\frac{\partial^2 P}{\partial \overline{V^2}}\right)_{TC} = \frac{2RT_C}{(V_C - b)^3} - \frac{6a}{V_C^4} = 0$$
(2.9)

Divide equation (2.8) by equation (2.9)

$$\left(\frac{2a}{(V_c)^3}\right) * \frac{V_c^4}{6a} = \frac{RT_c}{(V_c - b)^2} * \frac{(V_c - b)^3}{2RT_c}$$
(2.10)

$$b = \frac{V_c}{3} \tag{2.11}$$

Substituting b in equation (2.8) or (2.9)

$$a = \frac{9}{8}RT_C V_C \tag{2.12}$$

Putting equation (2.11) and (2.12) in equation (2.7), knowing that at critical conditions P becomes P_C

$$P_C = \frac{3}{8} \left(\frac{RT_C}{V_C} \right) \tag{2.13}$$

Where:

$$Z_C = \frac{P_C V_C}{RT_C} = \frac{3}{8}$$
(2.14)

By manipulating the above equations, 'a' and 'b' can be represented in other forms such as making V_c the subject of equation (2.13) and then putting that in equation (2.11) and equation 2.12 will lead to:

$$b = \frac{RT_c}{8P_c} = \Omega_b \left(\frac{RT_c}{P_c}\right)$$
(2.15)

$$a = \frac{27R^2 T_c^2}{64P_c} = \Omega_a \left(\frac{R^2 T_c^2}{P_c}\right)$$
(2.16)

It is important to note that the critical pressure, temperature and volume are usually known. In addition to the above equations, the original van der Waals equation can be represented in cubic form hence why it is referred to as a cubic equation of state.

This can be done by multiplying equation (2.7) by (V-b) (V^2) to give:

$$V^{3} - \left(b + \frac{RT}{P}\right)V^{2} + \frac{aV}{P} - \frac{ab}{P} = 0$$
(2.17)

Since:

$$V = \frac{ZRT}{P}$$
(2.18)

Put equation (2.18) in (2.17)

$$Z^{3} - \left(1 + \frac{bP}{RT}\right)Z^{2} + \frac{aP}{(RT)^{2}}Z - \frac{abP^{2}}{(RT)^{3}} = 0$$
(2.19)

$$A = \frac{ap}{(RT)^2} \tag{2.20}$$

$$B = \frac{bP}{RT}$$
(2.21)

$$Z^3 - (1+B)Z^2 + AZ - AB = 0 (2.22)$$

There are several methods that can be used to solve the above cubic equation of state, but for this study a combination of the Newton Raphson (NR) method and a semi analytical scheme was implemented. The Newton Raphson method is an iterative procedure that provides a new estimate closer to the actual answer based on a previous guess and continues this process until the difference between the previous and present answer is negligible. This procedure was chosen due to its fast convergence rate but care must be taken when making the first guess by ensuring that the guess is very close to the ultimate solution. Below is an example of the implementation of this procedure (Ayala H., 2014):

$$f(x) = X^3 + aX^2 + bx + c = 0$$
(2.23)

Using the NR approach, the root of this equation is then:

$$X_{new} = X_{old} - \frac{f(X_{old})}{f'(X_{old})}$$
(2.24)

$$f'(X_{old}) = 3X_{old}^2 + 2aX_{old} + b$$
(2.25)

Iterations over equation 2.21 will continue until there is no significant improvement of X_{new} such that

$$|X_{new} - X_{old}| < tolerance(10^{-6})$$
(2.26)

In order for convergence to be achieved rapidly, it is important to make an educated initial guess when applying this approach to finding the z factor. It has been suggested that the initial guess when dealing with be liquid phase should be:

$$Z = \frac{bP}{RT}$$
(2.27)

The initial guess when dealing with the gas phase is:

$$Z = 1 \tag{2.28}$$

The above procedure is first implemented to find the first root of the equation, then a semi analytical approach is applied to find the other two possible roots. Below is an example of the implementation of this approach:

Factorize equation 2.20:

$$(X_1 - W)(X^2 + FX + G) = 0 (2.29)$$

Where W is the root obtained from the NR method and then other two roots are obtained below:

$$X_2 = \frac{\left(-F + \sqrt{sqrt(F^2 - 4G)}\right)}{2}$$
(2.30)

$$X_{3} = \frac{\left(-F - \sqrt{sqrt(F^{2} - 4G)}\right)}{2}$$
(2.31)

It is important to note that the behavior of the cubic equation of state is dependent on the temperature. This is shown in figure (2-4):



Figure 2-4: Multiple roots in Cubic Eos (Adewumi, 2016)

At temperatures above critical point, the cubic EOS will have one root and two imaginary complex conjugates. This is an indication of single phase conditions. For temperatures at critical point, the vapor and liquid properties are the same therefore the cubic equations predict three real and equal roots. At temperatures lower than critical temperature, the equation of state becomes non-monatomic and yields three different roots. The smallest root is taken as the specific volume of the liquid phase and the largest root is taken as the specific volume of the gas phase. The intermediate root is usually discarded because it bears no theoretical significance and it is physically meaningless.

In spite of the remarkable success of Van der Waals' EOS in comparison to the ideal gas model, it is rarely used in industrial applications due to its lack of accuracy in the vapor-liquid equilibrium calculations especially when dealing with liquid property predictions. This inaccuracy is due to the attraction or repulsion term not taking into account its dependency on temperature and the equation of state predicting a fixed and high Z factor (value of 0.38 as shown in equation (2.14)) in comparison to the typical range of 0.22 to 0.30 obtained from experimental studies such as the table shown below.

Substance	Z _c (Critical Compressibility)
Carbon Dioxide (CO ₂)	0.2744
Methane (CH ₄)	0.2862
Ethane (C ₂ H ₆₎	0.2793
N-Pentane (C ₅ H ₁₂)	0.2693
N-Hexane (C ₆ H ₁₄)	0.2659

Table 2-1 Showing the examples of critical compressibility of different fluids (Adewumi, 2016)

In order to improve the accuracy of the Van der Waals equation of state, several researchers delved into modifying the attraction and repulsion terms. The equations of state explored in this study involve the modification of only the attraction term because complex modifications of the Van der Waals repulsion have not shown any real advantage in terms of predictions compared to the original term (Adachi, et al., 1983). In addition, retaining the original Van der Waals repulsion term results in simplicity of equation and ease of solution when using analytical techniques (Nwankwo, 2014).

2.3.2 <u>Redlich-Kwong (RK) Equation of State (Redlich & Kwong, 1949)</u>:

One of the major improvements made on the Van der Waals EOS was carried out by Redlich and Kwong in 1949. They demonstrated that the attraction term is temperature dependent. They observed experimentally that the molar volume of any pure substance reduces to about 26 percent of its critical volume as the pressure approaches infinity regardless of the temperature. Their equation is expressed in the following form:

$$P = \frac{RT}{V-b} - \frac{a}{V(V+b)T^{0.5}}$$
(2.32)

$$a = \Omega_a \left(\frac{R^2 T_C^{2.5}}{P_C}\right) = 0.42747 \left(\frac{R^2 T_C^{2.5}}{P_C}\right)$$
(2.33)

$$b = \Omega_b \left(\frac{RT_c}{P_c}\right) = 0.08664 \left(\frac{RT_c}{P_c}\right)$$
(2.34)

When equation (2.32) is expressed in cubic form (in Z):

$$Z^{3} - Z^{2} + (A - B - B^{2})Z - AB = 0$$
(2.35)

$$A = \frac{ap}{R^2 T^{2.5}}$$
(2.36)

$$B = \frac{bP}{RT}$$
(2.37)

Just like the Van der Waals EOS, the RK EOS predicts one critical compressibility for all substances. In fact, the prediction of one critical compressibility of all fluids is a characteristic of two parameter equations of state. The critical compressibility predicted by this EOS is:

$$Z_C = \frac{P_C V_C}{RT_C} = \frac{1}{3}$$
(2.38)

This is an improvement on the prediction of critical compressibility made by the Van der Waals EOS but it still does not fall in the range observed from experimental studies. The RK EOS proved to be adequate for only predictions of properties for the gas phase when the reduced pressure of the substance is less than half the reduced temperature (Ahmed, 1986).

$$P_R < \frac{1}{2}T_R \tag{2.39}$$

$$P_R = \frac{P}{P_R} \tag{2.40}$$

$$T_R = \frac{T}{T_R} \tag{2.41}$$

2.3.3 <u>Soave-Redlich-Kwong (SRK) Equation of State (Soave, 1972):</u>

In 1972, Soave made modifications to the Redlich-Kwong EOS in order to improve Vapor-Liquid Equilibria (VLE) calculations. He observed that the influence of temperature on the attraction term had not been adequately represented thereby leading to poor VLE calculations. He then proposed the replacement of the $\frac{a}{T^{0.5}}$ term found in equation (2.32) with the term a[T]. The basis for his proposed improvement was founded on the assumption that an improvement in the reproduction of saturation conditions of pure substances would lead to improvement of such conditions for mixtures. The new proposed equation is expressed in the following form:

$$P = \frac{RT}{V-b} - \frac{a[T]}{V(V+b)}$$
(2.42)

$$a[T] = a\alpha(T) \tag{2.43}$$

$$a = \Omega_a \left(\frac{R^2 T_c^2}{P_C}\right) = 0.42747 \left(\frac{R^2 T_c^2}{P_C}\right)$$
(2.44)

$$\alpha(T) = [1 + m(1 - T_R^{0.5})]^2$$
(2.45)

$$m = 0.48508 + 1.55171\omega - 0.1561\omega^2 \tag{2.46}$$

$$b = \Omega_b \left(\frac{RT_C}{P_C}\right) = 0.08664 \left(\frac{RT_C}{P_C}\right)$$
(2.47)

In order to obtain the m parameter in equation (2.45), a set of $\alpha(T)$ for a number of hydrocarbons was obtained at the point where thermodynamic equilibrium is satisfied (Nwankwo, 2014). This thermodynamic equilibrium is established by the fulfillment of the following criteria:

$$Thermal Equilibrium: T^{L} = T^{V}$$
(2.48)

$$Mechanical Equilibrium: P^{L} = P^{V}$$
(2.49)

$$Chemical Equilibrium \,\mu_i^L = \mu_i^V \tag{2.50}$$

$$Diffusional Equilibrium f_i^L = f_i^V$$
(2.51)

Where L and V represent the different phases (Liquid and Vapor) present when the system is in thermodynamic equilibrium. f_i (fugacity) and μ_i (chemical potential) represent the potential for a component to move from one phase to the other. Equal fugacity of a component in the different phases present will result in zero net transfer (Nwankwo, 2014). The values of $\alpha(T)^{0.5}$ are then plotted against $T_R^{0.5}$ and straight lines are obtained for all substances with the lines passing through point $T_R = \alpha(T) = 1$. The m parameter represents the slope of these lines are it was correlated with the molecular size and shape hence equation (2.46). The ω found in equation (2.46) represents the acentric factor which was discovered by Pitzer and it represents the measure of the non-sphericity of a molecule (Pitzer, 1939). Pitzer observed that simple substances possess spherical molecules and therefore have a value of $\omega = 0$. This is not the case for more complex substances since their molecules are larger and less spherical leading to a deviation from what was observed in the case of simple substances. This deviation is given by:

$$\omega = -\log P_r^v - 1.0 \tag{2.52}$$

Similar to VDW and RK EOS, the SRK equation of state can be represented in cubic form and the expression for this equation of state is the same as equation (2.35). The only difference is the definition of the A parameter which is shown below:

$$A = \frac{a[T]P}{(RT)^2} \tag{2.53}$$

Generally, this equation proved to be an improvement over Redlich-Kwong in Vapor-Liquid Equilibria Calculations but it was still found lacking when it came to making predictions of the volumetric and thermodynamic properties of the liquid phase. The reason for this is that despite the modifications, the SRK EOS still predicts the same value for critical compressibility as the original RK EOS for all fluids and this value as stated earlier does not fall in the range that has been measured experimentally.

2.3.4 <u>Peng-Robinson (PR) Equation of State (Peng & Robinson, 1976)</u>:

Recognizing the shortcomings of SRK EOS in prediction of liquid properties, Peng and Robinson developed an equation of state to predict saturated liquid properties more accurately. This was done by modifying the denominator of the attraction term found in the
SRK model and presenting different fitting parameters to describe the temperature dependency term(a[T]). Below is the expression of this equation of state:

$$P = \frac{RT}{V-b} - \frac{a[T]}{V^2 + 2bV - b^2}$$
(2.54)

All terms in the above equation retain the same definition as the SRK model except for the following:

$$a = \Omega_a \left(\frac{R^2 T_c^2}{P_c}\right) = 0.45724 \left(\frac{R^2 T_c^2}{P_c}\right)$$
(2.55)

$$m = 0.37464 + 1.54226\omega - 0.26992\omega^2 \tag{2.56}$$

$$b = \Omega_b \left(\frac{RT_c}{P_c}\right) = 0.07780 \left(\frac{RT_c}{P_c}\right)$$
(2.57)

Expressing equation (2.54) in cubic form:

$$Z^{3} - (1 - B)Z^{2} + (A - 2B - 3B^{2})Z - (AB - B^{2} - B^{3}) = 0$$
(2.58)

Where A and B have the same definitions as the SRK model. Similar to earlier two-parameter equations of State, the Peng-Robinson EOS predicts a fixed Z_c of 0.307 for all fluids. This value is in fact closer to the experimental range observed and therefore directly translates to the improvement of calculation of volumetric properties near the critical point and the prediction of the liquid phase volumetric properties. In spite of these improvements, the equation is still limited in its performance due to its prediction of one Z_c value for all fluids which is not the case as shown in Table 2.1. This constraint made two parameter EOS a bit too rigid and therefore efforts were made in order to make these equations more dynamic.

2.3.5 <u>Three-Parameter Equations of State</u>

Cognizant of the rigidity of the two parameter EOS, it was suggested that the fixed Zc usually predicted by two parameter EOS be replaced by introducing another substancedependent adjustable Critical parameter. This will help in relaxing the constraint on the equation and therefore improve the Vapor-Liquid calculations including predictions of liquid phase properties (Valderrama, 2003). The third parameter is usually dependent on either critical compressibility or acentric factor. Several three parameter equations have been established based on this principle just discussed but the focus of this study will be on Patel-Teja and a newly developed equation of state:

a) <u>Patel-Teja(PT) Equation of State (Patel & Teja, 1982)</u>

The EOS proposed by Patel and Teja served as an extension of the works done by Soave, Peng-Robinson and other notable three parameter equations of state such as Schmidt-Wenzel EOS. The EOS is expressed as follows:

$$P = \frac{RT}{V-b} - \frac{a[T]}{V^2 + (b+c)V - bc}$$
(2.59)

Observing equation (2.59), a new parameter 'c' is incorporated into the attraction term and this requires an additional constraining condition in addition to the criticality conditions which are used to derive parameters 'a' and 'b'. This condition is shown below:

$$\zeta_c = \frac{P_c V_c}{RT_c} \tag{2.60}$$

Where ζ_c is referred to as apparent critical compressibility factor.

It was observed by Patel and Teja that ζ_c should not equal to the true critical compressibility factor of fluids because it translates to an overall loss in accuracy when making Vapor-Liquid Equilibria and volumetric calculations. For this reason, ζ_c is an empirical parameter whose value was correlated with acentric factor to yield the following for non-polar substances:

$$\zeta_c = 0.329032 - 0.0767992\omega + 0.0211947\omega^2 \tag{2.61}$$

The variables in equation (2.59) retain the same definitions similar to Peng-Robinson except for the following:

$$m = 0.452413 + 1.30982\omega - 0.295937\omega^2 \tag{2.62}$$

$$c = \Omega_c \left(\frac{RT_c}{P_c}\right) \tag{2.63}$$

$$\Omega_c = 1 - 3\zeta_c \tag{2.64}$$

 Ω_b is the smallest positive root of the following equation:

$$\Omega_b^3 + (2 - \zeta_c)\Omega_b^2 + 3\zeta_c^2\Omega_b - \zeta_c^3 = 0$$
(2.65)

Equation (2.64) is solved analytically using Newton-Raphson and the initial guess of Ω_b is presented as:

$$\Omega_b = 0.32429\zeta_c - 0.022005 \tag{2.66}$$

$$\Omega_a = 3\zeta_c^2 + 3\left(1 - 2\zeta_c\right)\Omega_b + \Omega_b^2 + \left(1 - \zeta_c\right)$$
(2.67)

In addition to the above definitions, this EOS can be expressed in cubic form as shown below:

$$Z^{3} - (C-1)Z^{2} + (A - B - C - 2BC - B^{2})Z - (BC - B^{2}C - AB) = 0$$
 (2.68)

Where A and B are the same as equation (2.53) and equation (2.37) while C is expressed as:

$$C = \frac{cP}{RT}$$
(2.69)

Below are a few note-worthy characteristics of the Patel-Teja EOS:

- 1) When the parameter 'c' is set to zero, the equation collapses to SRK EOS (equation 2.42) and when 'c' is set equal to 'b', the equation collapses to PR EOS (equation 2.54).
- 2) When the empirical parameter ζ_c is set to 0.3074 (Z_c predicted by PR EOS), the equation collapses to PR EOS. Likewise, when ζ_c is set to 0.333 (Z_c predicted by SRK EOS), the equation collapses to SRK EOS.

Patel-Teja EOS showed comparable performance to Peng-Robinson when making Vapor-Liquid Equilibria calculations but was shown to be superior when predicting liquid volumetric properties. Another notable feature of this EOS is its superior performance when handling mixtures constituting of heavy hydrocarbons and polar substances.

b) New Equation of State (Nwankwo, 2014)

This equation was developed based on the observation made by (Yun, et al., 1998) that the attraction parameter of the Van der Waals equations of state (i.e. EOS that retain the Van der Waals repulsion term and modify the attraction term) is influenced by volume and this influence can be captured a $\pi(v)$ function. Yun et al (1998) also observed that the accuracy of EOS is dependent on both the number of terms in the $\pi(v)$ function and the relationship between the numerators of the various terms in the function. It is important to know that the $\pi(v)$ function is derived noting that all equations derived from the original Van der Waals EOS can be represented as follows:

$$P = \frac{RT}{V-b} - \frac{a}{\pi(v)V^2}$$
(2.70)

Where a and b retain their expressions as defined by their specific equation of state but the expression of $\pi(v)$ is equal to 1 for Van der Waals EOS and the expression changes for different equations. Below is a derivation of the $\pi(v)$ function for the three equations of state that are pertinent to this study and an additional table showing the $\pi(v)$ functions of the other EOS derived from the Original Vdw EOS:

<u>SRK EOS</u>

Equating the attraction terms from equation (2.5) and (2.42)

$$\frac{a_w}{(v)^2} = \frac{a_{srk}}{V(V+b)}$$
(2.71)

Multiply both sides by V²

$$a_{w} = \frac{a_{srk}}{1 + \frac{b}{v}} = \frac{a_{srk}}{\pi_{srk}(v)}$$
(2.72)

<u>PR EOS</u>

Equating the attraction terms from equation (2.5) and (2.54)

$$\frac{a_w}{(v)^2} = \frac{a_{pr}}{v^2 + 2bv - b^2}$$
(2.73)

Multiply both sides by V²

$$a_w = \frac{a_{pr}}{1 + \frac{2b}{v} - \frac{b^2}{v^2}} = \frac{a_{pr}}{\pi_{pr}(v)}$$
(2.74)

<u>PT EOS</u>

Equating the attraction terms from equation (2.5) and (2.59)

$$\frac{a_w}{(v)^2} = \frac{a_{PT}}{V^2 + (b+c)V - bc}$$
(2.75)

Multiply both sides by V²

$$a_{w} = \frac{a_{PT}}{1 + \frac{(b+c)}{v} - \frac{cb}{v^{2}}} = \frac{a_{PT}}{\pi_{PT}(v)}$$
(2.76)

Table 2-2 Showing a summary	ν of π(v)	functions	for several eq	uations of st	ate (Nwankwo, 201	4)
		3	, , ,			

Equation of State	$\pi(v)$ function
Soave-Redlich-Kwong (SRK) (2-parameter)	$1 + \frac{b}{v}$
	2
Peng-Robinson (PR)	$1 + \frac{2b}{2} - \frac{b^2}{2}$
(2-parameter)	$v v^2$
Patel-Teja(PT)	$1 + \frac{(b+c)}{cb} - \frac{cb}{cb}$
(3-parameter)	$v v^2$
Martin	$2c c^2$
(3-parameter)	$1 + \frac{1}{v} + \frac{1}{v^2}$
Adachi-Lu-Sugie (ALS)	$1 + \frac{b3 - b2}{b2b3} + \frac{b2b3}{b2b3}$
(3-parameter)	$v v^2$

Given the above expressions of $\pi(v)$ functions in table 2.2, we can therefore surmise that the general form of this function is as follows:

$$\pi(v) = 1 + \frac{k}{v} + \frac{l}{v^2} + \dots$$
(2.77)

Given the above developments, (Yun, et al., 1998) concluded that the different forms of $\pi(v)$ functions translate to the differences in the ability of the EOS to make predictions accurately (Nwankwo, 2014). For example, comparing the attraction term of both the SRK EOS (equation 2.42) and PR EOS (equation 2.54), it is noted that they both have the same definition of the temperature dependency term(a[T]) and yet PR EOS seems to significantly outperform SRK especially when the liquid properties are concerned. The explanation for this according to Yun et. al., 1998 is that the PR EOS has a greater number of terms that have volume as its denominator in its $\pi(v)$ function as shown in Table 2.2.

We also notice from the table above that the $\pi(v)$ function of martin EOS is very similar to that of PR EOS so it can be inferred that they both have comparable degrees of accuracy (Nwankwo, 2014). Further observing Table 2.2, it is noted that all EOS listed except SRK have the same number of terms but some of these equations such PT and ALS have shown to perform better than the others. According to Yun et. al., 1998 the reason for this is because there is a weak relationship between the numerators of the $\pi(v)$ function for PT and ALS in comparison to PR and Martin EOS. This weak relationship translates to the equation being more elastic and powerful thereby making the equation more competent for regression of data and therefore leading to improvement in its predictive capability.

In line with these observations, Yun et. al.(1998) proposes two options that will help improve subsequent equations that are developed from the original equations of state. These are:

- 1) Increasing the number of terms in the $\pi(v)$ function.
- 2) Maintaining the number of terms already present in the $\pi(v)$ function but decreasing the strength of relationship between the numerators of the terms.

Based on these suggestions, Nwankwo 2014 utilizes the second option to develop an EOS that builds on the strength of PT EOS and stretches its $\pi(v)$ function as proposed. The first option was discarded in this case because the resulting development would lead to a polynomial higher than 3 making the equation cumbersome to implement since it will be impossible to represent in cubic form. In order to achieve this New EOS, 153 structural forms were developed which satisfy the criteria of minimizing the error in the prediction of vapor pressure and liquid density data. These structural forms were then tested with empirical data to determine which form provided predictions that minimized error the most. Below is the resulting equation that proved to have the best performance:

$$P = \frac{RT}{V-b} - \frac{a[T]}{V(V+b) + c(V-b) + c(c-b)}$$
(2.78)

$$\pi_{new}(v) = 1 + \frac{(b+c)}{v} - \frac{(2bc-c^2)}{v^2}$$
(2.79)

Although this equation showed an improvement in prediction of liquid densities over PT and PR, VLE calculations were not explored. All terms in equation (2.78) have similar definitions to PT EOS except the following:

$$m = 0.359 + 0.288\omega + 1.846\omega^2 \tag{2.80}$$

Also, similar to the other EOS looked at in this study, this equation can be represented in cubic form as follows:

$$Z^{3} + (C-1)Z^{2} + (A-B-C-3BC-B^{2}+C^{2})Z + (2BC+2B^{2}C-BC^{2}-C^{2}-AB) = 0$$
 (2.81)

2.4 Extension to Mixtures

All equations developed in the previous section can only be applied to pure substances and since the fluids handled in the Oil and Gas industry are multi-component mixtures, it is of importance that the applications of these equations is extended to mixtures (Sadus, 1992a). Several mixing rules have been developed for this purpose but for this study the general mixing rules developed by Van der Waals is implemented:

$$(a\alpha)_m = \sum_i \sum_j \theta_i \theta_j (a\alpha_i a\alpha_j)^{\circ} 0.5$$
(2.82)

$$b_m = \sum \theta_i b_i \tag{2.83}$$

$$C_m = \sum \theta_i C_i \tag{2.84}$$

$$a\alpha_{i} = \Omega_{ai} \left(\frac{R^{2} T_{Ci}^{2}}{P_{Ci}}\right) \left[1 + m_{i} \left(1 - T_{Ri}^{0.5}\right)\right]^{2}$$
(2.85)

$$b_i = \Omega_{bi} \left(\frac{RT_{Ci}}{P_{Ci}}\right) \tag{2.86}$$

$$C_i = \Omega_{ci} \left(\frac{RT_{Ci}}{P_{Ci}}\right) \tag{2.87}$$

Where $(a\alpha)_m$, b_m and C_m represents mixture constants while θ_i and θ_j represent the mole fraction of the components of a particular phase (Vapor or Liquid). In many cases for mixtures, the molecules of the components are dissimilar both chemically and physically and the interactions between the molecules have an influence on the volumetric and thermodynamic properties. In order to account for these interactions, a correction factor known as binary interaction parameter(k_{ij}) is introduced into equation (2.81):

$$(a\alpha)_m = \sum_i \sum_j \theta_i \theta_j \sqrt{a\alpha_i a\alpha_j} (1 - k_{ij})$$
(2.88)

Where k_{ij} represents the interaction between i and j.

In addition to the above definitions, the following changes are made when solving the cubic equations of state:

$$A = \frac{(a\alpha)_m P}{(RT)^2}$$
(2.89)

$$B = \frac{b_m P}{RT}$$
(2.90)

$$C = \frac{C_m P}{RT}$$
(2.91)

CHAPTER 3

METHODOLOGY

In this chapter, procedures for the determination of the number of phases and the numerical schemes required to carry out Vapor-Liquid Equilibria calculations will be described. In addition, procedures required to calculate thermos-physical properties of hydrocarbon systems will be explained.

3.1 Vapor-Liquid Equilibrium (VLE) Flash

VLE calculations also known as flash calculations are extremely essential in ascertaining whether a given mixture at a given pressure and temperature will exist as a single phase or split into two phases (gas and liquid in this case). In addition to detecting phase split, VLE flash enables the calculations of the amount of phase present and the composition of each phase at equilibrium. An important parameter essential to this procedure is the equilibrium constant which is defined as the ratio of the mole fraction of a component in the gas phase (yi) to the mole fraction of the same component in the liquid phase (xi). This parameter is dependent on pressure, temperature and composition of each phase present and is represented below as:

$$K_i = \frac{y_i}{x_i} \tag{3.1}$$

Using equation 3.1 in addition to the material balance associated with two phase mixtures will yield an objective function developed by Rachford H. H. and Rice J. D. in 1952 known as the Rachford-Rice function which is used in calculation of the amount of phases present (Rachford & Rice, 1952). Below is a derivation of this function:

The total amount of phases can be expressed as:

$$F = V + L \tag{3.2}$$

Where V represents the number of moles of vapor phase and L is the number of moles of liquid phase.

Divide equation 3.2 by F to give:

$$1 = \left(\frac{V}{F}\right) + \left(\frac{L}{F}\right) = \alpha_g + \alpha_L \tag{3.3}$$

Where α_q is the fraction of gas phase and α_L is the fraction of liquid phase

Apply equation 3.3 to a multicomponent system:

$$y_i \alpha_g + x_i \alpha_L = z_i \tag{3.4}$$

Equation 3.4 was rewritten as

$$y_i \alpha_g + x_i (1 - \alpha_g) = z_i \tag{3.5}$$

Applying equation 3.1 to equation 3.5:

$$y_i \alpha_g + \left(\frac{y_i}{K_i}\right) \left(1 - \alpha_g\right) = z_i \tag{3.6}$$

Making y_i the subject of equation 3.6

$$y_i = \frac{z_i K_i}{1 + \alpha_g (K_i - 1)}$$
(3.7)

Expressing equation 3.6 in x_i and then making it the subject of equation:

$$x_i = \frac{z_i}{1 + \alpha_g(K_i - 1)}$$
(3.8)

Given the following the constraints:

$$\sum y_i = 1 \tag{3.9}$$

$$\sum x_i = 1 \tag{3.10}$$

With the above constraints, equation 3.6 and equation 3.7 are represented as:

$$\sum \frac{z_i K_i}{1 + \alpha_g(K_i - 1)} = 1 \tag{3.11}$$

$$\sum \frac{z_i}{1 + \alpha_g(K_i - 1)} = 1 \tag{3.12}$$

Equation 3.11 and 3.12 can be rewritten as:

$$\sum \frac{z_i K_i}{1 + \alpha_g(K_i - 1)} - 1 = 0 \tag{3.11}$$

$$\sum \frac{z_i}{1 + \alpha_g(K_i - 1)} - 1 = 0 \tag{3.12}$$

The objective function can be obtained by taking the difference between equation 3.11 and

3.12

$$f(\alpha_g) = \sum_{1}^{nc} \frac{z_i(K_i - 1)}{1 + \alpha_g(K_i - 1)} = 0$$
(3.13)

Equation 3.13 is a non-linear function with one variable and can be solved using the Newton-Raphson iterative scheme as shown below

$$\alpha_g^{new} = \alpha_g^{old} - \left(\frac{f(\alpha_g^{old})}{f'(\alpha_g^{old})}\right)$$
(3.14)

Where $f'(\alpha_q^{old})$ represents the first derivative of equation 3.13 and is shown below as:

$$f'(\alpha_g^{old}) = \sum_{1}^{nc} \frac{z_i(K_i-1)^2}{\left[1 + \alpha_g(K_i-1)\right]^2}$$
(3.15)

For this iterative scheme, convergence is achieved when:

$$\left|\alpha_{g}^{new} - \alpha_{g}^{old}\right| < 10^{-10} \tag{3.16}$$

It is important to note that this iterative procedure has its setbacks since it is dependent on how close the initial guess of α_g is to the solution. Also the procedure fails if $f'(\alpha_g^{old})$ has value close to 0 because during the update of $f(\alpha_g^{old})$, the new value falls outside the physically acceptable range and therefore yields negative concentrations. In order to remedy such occurrence, the bisection strategy is implemented.

3.1.1 <u>Bisection Strategy</u>

This strategy was selected because it is simple, fail-proof and robust. The only disadvantage of this procedure is that it is relatively slow and this can be alleviated by using the bisection strategy to obtain a rough approximation (Initial guess) to the solution and then Implementing Newton-Raphson since it converges quadratically when the initial guess is close to the solution. Below is a detailed step-step process of implementing this procedure:

1) While carrying out the Newton-Raphson procedure, the following condition should be satisfied at each iteration for the procedure to be implemented:

$$\frac{1}{1-Ki_{max}} < \alpha_g < \frac{1}{1-Ki_{\min}} \tag{3.17}$$

If equation 3.17 is not fulfilled, the iterative scheme should switch to the following:

2) State the upper ($lpha_g U$) and lower ($lpha_g L$) initial values for $lpha_g$

$$\alpha_g U = 0 \left[f \left(\alpha_g = 0 \right) \text{ should be positive} \right]$$
(3.18)

$$\alpha_g L = 1 \left[f(\alpha_g = 1) \text{ should be negative} \right]$$
 (3.19)

3) Update the α_g value using the following scheme

$$\alpha_g^{new} = 0.5 * (\alpha_g U + \alpha_g L) \tag{3.20}$$

4) Evaluate $f(\alpha_g)$ and update the values of $\alpha_g U$ and $\alpha_g L$ as follows:

If
$$f(\alpha_g^{new}) > 0$$
, $\alpha_g U = \alpha_g^{new}$ (3.21)

If
$$f(\alpha_g^{new}) < 0$$
, $\alpha_g L = \alpha_g^{new}$ (3.22)

5) Steps 1 – 4 will continue until equation 3.16 is satisfied

3.1.2 Initial guess for K_i's

Since the composition of possible phases is unknown prior to implementing the Newton-Raphson routine, the value of Ki cannot be calculated and therefore an initial guess has to be provided. Numerous correlations have been developed to give an estimate of Ki values but for this study, Wilson's empirical correlation is used to initialize the values as shown below:

$$K_{i} = \left(\frac{1}{P_{ri}}\right) Exp\left[5.37(1+\omega_{i})\left(1-\left(\frac{1}{T_{ri}}\right)\right)\right]$$
(3.23)

Given that equation 3.23 provides an estimate, rigorous thermodynamic considerations have to be applied in order to yield more accurate K_i values (Ayala H., 2014). These considerations can be represented in a form known as the Iso-fugacity Criteria.

3.1.3 <u>Iso-Fugacity Criteria</u>

Fugacity is the measurement of the potential for a component in a mixture to transfer from one phase to another. For a system in a state of equilibrium, the net transfer of heat, momentum and mass must be zero (Ayala H., 2014). In order to satisfy these conditions for equilibrium, temperature, pressure and chemical potential have to be equal in all present phases. Due to the one to one relationship between Gibbs energy to chemical potential, fugacity is directly related to chemical potential and therefore the fugacity of all components in each phase must be equal at equilibrium. This form of equilibrium is expressed as:

$$f_{li} = f_{gi} \tag{3.24}$$

In order to calculate the values of fugacity, the fugacity coefficients have to be evaluated. The fugacity coefficient is defined as the ratio of the fugacity of a substance to its partial pressure. For a two phase system containing N_c components, the fugacity coefficient of components for both vapor and liquid phase are expressed below as:

$$ln\phi_{gi} = ln\left(\frac{f_{gi}}{y_i * P}\right) = \left(\frac{1}{RT}\right) \int_{\infty}^{V} \left[\left(\frac{dP}{dn_i}\right) - \left(\frac{RT}{V}\right)\right] dv - lnZ_g$$
(3.25)

$$ln\phi_{li} = ln\left(\frac{f_{li}}{x_i^*P}\right) = \left(\frac{1}{RT}\right) \int_{\infty}^{V} \left[\left(\frac{dP}{dn_i}\right) - \left(\frac{RT}{V}\right)\right] dv - lnZ_l$$
(3.26)

The $\left(\frac{dP}{dn_i}\right)$ parameter is obtained by deriving the P-explicit form of any cubic equation of state. For this study, the formula of the fugacity coefficient is written in a generalized form that applicable to any cubic equation of state.

$$ln\phi_{i} = -\ln(Z-B) + \left(\frac{A}{(m_{1}-m_{2})B}\right) \left[\frac{2\sum_{j=1}^{n_{c}}A_{ij}C_{j}}{A} - \left(\frac{B_{i}}{B}\right)\right] \ln\left[\frac{Z+m_{2}B}{Z+m_{2}B}\right] + \left(\frac{B_{i}}{B}\right)(Z-1)$$
(3.27)

The parameter m1 and m2 are definitions specific to different equations. Below is a table representing the definitions of m1 and m2 for the equations used for VLE calculations.

EOS	m1	m2
Peng-Robinson	$1 + \sqrt{2}$	$1 - \sqrt{2}$
Patel-Teja	$\left(\frac{b+c}{2b}\right) + \frac{1}{b}\sqrt{bc + \left(\frac{b+c}{2b}\right)^2}$	$\left(\frac{b+c}{2b}\right) - \frac{1}{b}\sqrt{bc + \left(\frac{b+c}{2b}\right)^2}$
New	$\left(\frac{b+c}{2b}\right) + \frac{1}{b}\sqrt{2bc - c^2 + \left(\frac{b+c}{2b}\right)^2}$	$\left(\frac{b+c}{2b}\right) - \frac{1}{b}\sqrt{2bc - c^2 + \left(\frac{b+c}{2b}\right)^2}$

Table 3-1 Showing the m1 and m2 definitions for three different equations of state

Equation 3.27 can be applied to both Liquid and Vapor phase provided the z factor of the chosen phase is already calculated by solving the cubic expression for any EOS that is being implemented. Just like the equation 3.27, a generalized cubic equation state can be utilized to obtain the z factor. The generalized cubic equation utilized in this work was developed by coats in 1985 and is expressed as follows

$$Z^{3} + [(m_{1} + m_{2} - 1)B - 1]Z^{2} + [A + m_{1}m_{2}B^{2} - (m_{1} + m_{2})B(B + 1)]Z - [AB + m_{1}m_{2}B^{2}(B + 1)] = 0$$
(3.28)

Where:

$$A = \sum_{i} \sum_{j} c_{i} c_{j} A_{ij}$$
(3.29)

$$B = \sum x_i b_i \tag{3.30}$$

$$C = \sum x_i C_i \tag{3.31}$$

$$A_{ij} = (1 - k_{ij}) (a\alpha_i a\alpha_j)^{0.5}$$
(3.32)

$$a\alpha_{i} = \Omega_{ai} \left(\frac{R^{2} T_{Ci}^{2}}{P_{Ci}}\right) \left[1 + m_{i} \left(1 - T_{Ri}^{0.5}\right)\right]^{2}$$
(3.33)

$$b_i = \Omega_{bi} \left(\frac{RT_{Ci}}{P_{Ci}}\right) \tag{3.34}$$

$$C_i = \Omega_{ci} \left(\frac{RT_{Ci}}{P_{Ci}}\right) \tag{3.35}$$

It is important note that x_i represents the composition of whatever phase is being evaluated and parameters m1 and m2 are obtained from Table 3.1.

Based on the equilibrium criterion that fugacities of all present phases must be equal, an iterative scheme known as Successive substitution method (SSM) is developed. The purpose of SSM is to update previous K_i values using fugacities that have been calculated using equation 3.27. Below is development of SSM procedure when updating the K_i values:

 K_i is related to the fugacity coefficient of the gas and liquid phase as follows:

$$K_{i} = \frac{\phi_{li}}{\phi_{gi}} = \frac{\left(\frac{f_{li}}{x_{i}*P}\right)}{\left(\frac{f_{gi}}{y_{i}*P}\right)} = \frac{y_{i}}{xi} \left(\frac{f_{li}}{f_{gi}}\right)$$
(3.36)

Implementing SSM, equation 3.36 is then represented in the correct step-formulation as:

$$K_i = \left(\frac{y_i}{x_i}\right)^n \left(\frac{f_{li}}{f_{gi}}\right)^n = (K_i)^n \left(\frac{f_{li}}{f_{gi}}\right)^n$$
(3.37)

Equation 3.37 will be continually implemented until the fugacity ratios of all components in the system are close to unity. This criterion is represented by the following inequality:

$$\sum_{i}^{n} \left(\left(\frac{f_{li}}{f_{gi}} \right) - 1 \right)^{2} < 10^{-14} \tag{3.38}$$

3.2 Phase Stability

One of the drawbacks of flash calculations is that it is computationally intensive and therefore knowing if a phase split occurs at a particular temperature and pressure before carrying out flash calculations will help in reducing computational time. There have been several approaches that have been developed for this purpose such as the bring back technique developed by Risnes et al (1981) but for this work, the phase stability criteria developed by Michelsen (1982) is implemented because though it is similar to Risnes et al in terms of procedure, it also provides interpretation for the cases where trivial solutions are found. Michelsen proposed the creation of a second phase within the mixture and then a test is carried out to verify if the system is stable or not. Since this work deals with vapor-liquid equilibrium only, the test carried out for two separate cases:

1) Assuming the second phase is gas

2) Assuming the second phase is liquid

Below is a step by step procedure for two cases:

Vapor Phase:

- 1) Calculate the fugacity of the mixture using overall composition (Z_i)
- 2) Create a Vapor-like second phase
 - a) Used Wilson's correlation to obtain initial k-values
 - b) Calculate second-phase mole numbers, Y_i

$$Y_i = Z_i K_i \tag{3.39}$$

c) Obtain the sum of the mole numbers,

$$S_{\nu} = \sum_{i}^{n} Y_{i} \tag{3.40}$$

d) Normalize the second-phase mole numbers to get mole fractions:

$$y_i = \frac{Y_i}{S_v} \tag{3.41}$$

- e) Calculate the second-phase fugacity (f_{yi}) using the corresponding EOS and the composition of the second phase
- f) Update the K_i values as follows:

$$R_{i} = \frac{f_{zi}}{f_{yi}} * \frac{1}{s_{v}}$$
(3.42)

$$K_i^{n+1} = K_i^n * R_i (3.43)$$

- g) Check for the following conditions upon iteration:
 - I. Convergence is achieved when:

$$\sum_{i}^{n} (R_{i} - 1)^{2} < 1 * 10^{-10}$$
(3.43)

II. Trivial solution is approached:

$$\sum_{i}^{n} (lnK_{i})^{2} < 1 * 10^{-4}$$
(3.44)

If the convergence criterion has not been attained, go back to step b with the newly calculated K_i values.

Liquid Phase:

The same steps used for the vapor phase are applicable to the liquid phase except for the following changes:

$$Y_i = \frac{Z_i}{K_i} \tag{3.45}$$

$$S_L = \sum_i^n Y_i \tag{3.46}$$

$$x_i = \frac{Y_i}{S_L} \tag{3.47}$$

$$R_i = \frac{f_{xi}}{f_{zi}} * S_L \tag{3.48}$$

After carrying out the above procedures for both the liquid and vapor phase, below are conditions required to decide if the hydrocarbon system is stable and therefore will not split

- $S_L < 1$ and $S_v < 1$
- The test for both phases converge to trivial solution
- The test for one phase converges to trivial solution and while for the other phase S < 1

If any of these conditions are not fulfilled, then it can be deduced that phase splitting will occur and therefore flash calculations can be carried.

3.3 Calculation of Phase properties

Once flash calculations has been carried out, the compositions of gas and liquid phase can be computed and then several thermo-physical properties are calculated as follows:

1) Molecular weight

The molecular weight of each phase is a function of the mole fraction and molecular weight of the individual components found in the mixture. It is calculated as follows:

$$Mw_g = \sum_{i=1}^n y_i Mw_i \tag{3.49a}$$

$$Mw_L = \sum_{i=1}^n x_i Mw_i \tag{3.49b}$$

2) Density

The density of a phase is calculated using the compressibility of the phase which is predicted from a chosen equation of state. From the real gas law, the density of each phase is calculated

$$\rho_L = \frac{P}{RT} \left(\frac{M w_L}{Z_L} \right) \tag{3.50a}$$

$$\rho_g = \frac{P}{RT} \left(\frac{M w_g}{Z_g} \right) \tag{3.50b}$$

Due to the poor liquid predictions generally made by two equations state, a volume shift technique was developed to alleviate this issue. The idea for this technique was first introduced by martin in 1979 but was thoroughly applied by Peneloux et al in 1982 with the goal of improving volumetric capabilities of the SRK EOS. In 1988, Jhaveri and Youngren went ahead to show that this technique could be applied to any two-constant EOS, hence the reason why this procedure is applied in this study to the Peng-Robinson equation. Below is how this volume translation concept is applied:

$$V_L = V_L^{EOS} - \sum_{I=1}^n x_i c_i$$
 (3.50c)

$$V_g = V_g^{EOS} - \sum_{l=1}^n y_l c_l$$
 (3.50d)

Where: V_L^{EOS} and V_g^{EOS} represent the liquid and vapor molar volumes calculated by EOS

C_i represents component dependent volume shift parameters.

Although C_i was proposed by Peneloux et al, the calculation of this parameter was generalized by Jhaveri and Youngren who proposed that the C_i is a function of the pure substance co-volume (b_i) and is therefore expressed as equation 3.50e.

$$c_i = s_i b_i \tag{3.50e}$$

 S_l represents proportionality constants that are estimated by matching saturated liquid density at $T_r = 0.7$ for each pure component. These proportionality constants are EOS dependent and the values applicable to the Peng-Robinson equation are shown below for several components

Compound	Si Values
N2	-0.1927
CO ₂	-0.0817
H ₂ S	-0.1288
C1	-0.1595
C2	-0.1134
C3	-0.0863
i-C4	-0.0844
n-C4	-0.0675
i-C5	-0.0608

n-C5	-0.039
n-C6	-0.008
n-C7	0.0033
n-C8	0.0314
n-C9	0.0408
n-C10	0.0655

Table 3-2 Showing shift parameters for Peng-Robinson EOS (Curtis & Michael, 2000)

Once volume translation procedure has been applied as shown in equation 3.50c and 3.50d, the density can then be calculated as follows:

$$\rho_L = \left(\frac{Mw_L}{V_L}\right) \tag{3.50f}$$

$$\rho_g = \left(\frac{Mw_g}{V_g}\right) \tag{3.50g}$$

- 3) Viscosity
 - a) Gas Phase (Lee, et al., 1966):

Numerous empirical correlations have been proposed for estimating the viscosity of natural gases but for this study, the empirical correlations developed by Lee, Gonzalez and Eakin was implemented and is expressed as follows:

$$\mu_g = 1 * 10^{-4} K_v EXP\left(x_v \left(\frac{\rho_g}{62.4}\right)^{y_v}\right)$$
(3.51a)

$$K_{\nu} = \frac{(9.4 + 0.02MW_g)T^{1.5}}{209 + 19MW_g + T}$$
(3.51b)

$$y_{\nu} = 2.4 - 0.2X_{\nu} \tag{3.51c}$$

$$x_{\nu} = 3.5 + \frac{986}{T} + 0.01 M W_g \tag{3.51d}$$

b) Liquid Phase (Lohrenz, et al., 1964):

In order to calculate the viscosity of the liquid phase, the empirical correlation proposed by Lohrenz, Bray and clark (1964) is implemented as shown below:

$$\mu_L = \mu^* + \xi_m^{-1} [(a_0 + a_1\rho_r + a_2\rho_r^2 + a_3\rho_r^3 + a_4\rho_r^4)^4 - 1 * 10^{-4}]$$
(3.52a)

Where:

$$\mu_L$$
 = Liquid viscosity (cp)

 μ^* = viscosity at atmospheric pressure (cp)

 ξ_m = Mixture viscosity parameter (cp⁻¹)

 ρ_r = reduced liquid density

The coefficients of reduced liquid density in equation 3.52a are constants proposed by Jossi, stiel and Thodos where:

$$a_0 = 0.1023, a_1 = 0.023364, a_2 = 0.058533, a_3 = -0.040758, a_4 = 0.0093324$$

$$\mu^* = \frac{\sum_i x_i \mu_i^* \sqrt{MW_i}}{\sum_i x_i \sqrt{MW_i}}$$
(3.52b)

$$\mu_i^* = \frac{34*10^{-5}*T_{ri}^{0.94}}{\xi_i} \quad for \ T_{ri} \le 1.5$$
(3.52c)

$$\mu_i^* = \frac{17.78 \times 10^{-5} (4.58T_{ri} - 1.67)^{0.625}}{\xi_i} \quad for \ T_{ri} > 1.5 \tag{3.52d}$$

$$\xi_{i} = \frac{5.4402T_{pci}^{\frac{1}{6}}}{\sqrt{MW_{i}}P_{pci}^{\frac{2}{3}}}$$
(3.52e)

$$\xi_m = \frac{5.4402T_{pc}^{\frac{1}{6}}}{\sqrt{MW_L} P_{pc}^{\frac{2}{3}}}$$
(3.52f)

Where

$$T_{pc}$$
 = pseudocritical temperature (°R) = $\sum x_i T_{ci}$ (3.52g)

 P_{PC} = pseudocritical pressure (psia)= $\sum x_i P_{ci}$ (3.52h)

$$\rho_r = \left(\frac{\rho_L}{MW_L}\right) V_{pc} \tag{3.53i}$$

$$V_{pc}$$
 = pseudocritical VOLUME = $\sum x_i V_{ci}$ (3.52J)

3.4Estimation of critical properties for Heptane plus fractions

When evaluating certain gas compositions especially gas condensates, it important to make sure that heavy fractions of hydrocarbon systems are properly characterized because they affect the accuracy of the predictions made by the equations of state. Characterization of these heavy fractions can be extremely challenging due to the fractions containing different components within itself. In order to solve such problem several empirical correlations have been established which have provided relatively accurate predictions on the critical pressure and temperature and can therefore make it possible for the any particular EOS to carry out calculations easily. For this study, the correlation established by Riazi and Daubert is being used for charactering C₇⁺ due to its simplicity and its accuracy across a wider range of specific gravities in comparison to other well-known correlations. Below is expression of this correlation:

$$\vartheta = a * MW^b * SG^c * EXP[d * MW + e * SG + f * (MW * SG)]$$
(3.53)

Where $\vartheta = physical property$

a-f represents constants that change depending on the property being evaluated. Below is a table listing the constants used to calculate Critical pressure and Critical Temperature.

θ	Тс	Рс
а	544.4	45203
b	0.2998	-0.8063
С	1.0555	1.6015
d	-0.00013	-0.0018078
е	-0.61641	-0.3084
f	0	0

Table 3-3 Showing Riazi and Daubert Coefficients

CHAPTER 4

RESULTS AND DISCUSSION OF RESULTS

The main goal of this chapter is to measure and evaluate the performance of the new equation of state in comparison to the Patel-Teja and Peng-Robinson's equation of state after undergoing flash calculations. The experimentally measured thermo-physical properties is used as reference when determining which equation provides the prediction with the least deviation from the experimental value. The following statistical methods are used in order to carry out error analysis and therefore measure the performance the different equations:

Absolute Average Deviation (AAD) =
$$\frac{1}{n}\sum_{i=1}^{n} \left| \frac{N_{experimental} - N_{predicted}}{N_{predicted}} \right|$$
 (4.1)

Root Mean Square Deviation (RMSD) =
$$\sqrt{\frac{1}{n}\sum_{i=1}^{n} \left(\frac{N_{experimental}-N_{predicted}}{N_{predicted}}\right)^2}$$
 (4.2)

Using the above equations, the following properties are evaluated: Composition of the Liquid phase, Viscosity of the Liquid phase and Density of the Liquid phase. These properties are evaluated for both single and multi-component hydrocarbon systems. The multi-component systems in these studies not only varied in the number of components present, but also in the amount of each component in the system. Below are the results of the predictions of the different properties by the different equations of state.

4.1 Prediction of Liquid Phase Compositions

In order to accurately predict the density or viscosity of any phase, the composition of that phase is of significant importance because with this information the molecular weight of the

phase can be calculated and this parameter is essential in calculating density and viscosity of the phase. Below is a list of mixtures for which the composition of the liquid phase was calculated:

Mixture	System	NDP	T range	P range
			(deg R)	(psia)
1	C1-C3	19	344 – 385	27.5 - 300
2	C1-C2	30	285 - 360.2	25.8 - 350
3	C2-nC4	12	546 - 654.45	164 -773
4	C1-iC4	11	366 - 493	383 - 1209
5	C1-nC4	10	420 – 493	190 - 1052

Table 4-1: Showing the Mixtures for which the Liquid Phase Compositions is evaluated

1) Mixture 1 (Wichterle & Riki, 1972):

This mixture is a binary mixture constituting of methane and propane whose experimental data and predicted values for the three different equations of state are shown in Appendix A. Upon carrying out error analysis for the different EOS as shown in Table 4.2, the least AAD and RMSD values were registered by the new equation of state with values of **3.53% and 4.33%** while Peng-Robinson recorded the largest deviations of **527% and 870%**. The reason for the extremely high values for Peng-Robinson is because at certain pressures and temperatures, the model assumes that a single phase is present instead of two phases. The new equation and Patel-Teja are robust enough to not only detect the presence of two phases but make good predictions with the new equation performing slightly better.

Table 4-2: Error Analysis of Liquid composition for Methane-Propane System

EOS	AAD	RMSD
PR	5.269819086	8.696163
PT	0.141253862	0.141852

NEW	0.035336517	0.043278
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Below are 45-degree plots pictorially showing the deviations of the data predicted by the different EOS models from the experimental values.



Figure 4-1: Plots showing Liquid composition predictions for a Methane-Propane system

Looking at figure 4-1, the New EOS does the best job of predicting the experimental data hence while it lies closest to the 45-degree line. Observing the figure on the top, there are certain data points predicted by the Peng- Robison model that are extremely far from the 45-degree points, these are points where the model was unable to detect the presence of two phase and therefore assumed the overall composition of the mixture as the composition of the Liquid phase.

2) Mixture 2 (Wichterle & Kobayashi, 1972):

This binary mixture constitutes of methane and Ethane whose experimental data and predicted values for the three different equations of state are shown in Appendix A. The experimental data for this mixture is obtained at Pressure range of 26 – 350 psia and temperature range of 285 – 360.2 °R. In terms of prediction, the new equation performs best with the New EOS having an AAD of **9.8%** and Patel having an AAD of **13.57%**. Similar to mixture 1, the PR-EOS had problems making accurate predictions at certain pressures and temperatures because it was unable to detect the presence of two phase and therefore assumed the overall mixture as the mixture of the liquid phase. Below is a table showing the results of the error analysis of the predictions of state.

EOS	AAD	RMSD
PR	2.928524	3.946912
PT	0.135671	0.140795
NEW	0.097697	0.111984

Table 4-3: Error Analysis of Liquid composition for Methane-Ethane System

In addition to having the table above, below is a 45-degree plot showing the deviations of the predicted values from experimental data



Figure 4-2: Plots showing Liquid composition predictions for a methane-ethane system

Similar to Mixture 1, the large deviations of data points of the PR EOS model are representative of the temperature and pressure conditions where the presence of two phase is not detected.

3) Mixture 3 (Lhotak & Wichterle, 1981):

This binary mixture constitutes of Ethane and n-butane whose experimental data and predicted values for the three different equations of state are shown in Appendix A. The experimental

data for this mixture is obtained at Pressure range of 164 -773 psia and temperature range of 546 – 654.45 °R. In terms of prediction, the Patel-Teja and New equation of state make nearly identical predictions with the new equation having a slight edge in accuracy. The New equation has an AAD value of **4.016%** and Patel-Teja has an AAD value of **4.2%**.

Table 4-4: Error Analysis of Liquid composition for Ethane-nButane System

EOS	AAD	RMSD
PR	0.416143	0.644833
РТ	0.042462	0.050757
NEW	0.040157	0.055095

4) Mixture 4 (May, et al., 2015):

This binary mixture constitutes of Methane and i-butane whose experimental data and predicted values for the three different equations of state are shown in Appendix A. The experimental data for this mixture is obtained at Pressure range of 383 – 1209 psia and temperature range of 366 – 493 °R. Similar to the previous three mixtures, the new equation performs the best based on the results of the error analysis in table 4.5 but unlike the previous mixtures, the Peng-Robinson performs much better than Patel-Teja. This is because Peng Robinson was designed to handle dry gas and wet gas mixtures which are hydrocarbon systems that constitute an extremely high amount of methane.

EOS	AAD	RMSD
PR	0.036152	0.040026
РТ	0.144756	0.146602
NEW	0.03342	0.037141

Table 4-5: Error Analysis of Liquid composition for Methane-ibutane System

5) Mixture 5 (May, et al., 2015):

This binary mixture constitutes of methane and n-butane whose experimental data and predicted values for the three different equations of state are shown in Appendix A. Similar to mixture 4, the methane constitutes over 90% of the overall composition hence the reason why Peng-Robinson outperforms Patel-Teja. For Mixture 5, the new equation slightly outperforms Peng-Robinson as shown in the results of the error analysis presented in table 4.6

EOS	AAD	RMSD
PR	0.041667	0.044528
РТ	0.171986	0.172373
NEW	0.041245	0.041782

Table 4-6: Error Analysis of Liquid composition for Methane-nbutane System

In addition, an overall error analysis was carried out after collating the predictions made for all hydrocarbon systems tested. The results of this analysis are shown in Table 4-7 and Figure 4-3.

EOS	AAD	RMSD
PR	2.363297	4.825254
PT	0.128972	0.137107
NEW 0.05932		0.076577

Table 4-7: Error Analysis of Liquid composition for all 5 mixtures



Figure 4-3: Plots Showing Liquid Composition Predictions for all hydrocarbon systems explored

We observe from Table 4-7 that Peng-Robinson records the highest percentage error of over 100% due to its inability to detect two phase splits at certain pressures and temperature conditions. Patel does a better job compared to Peng Robinson with an average absolute

deviation of 12.9% while the new equation performs even better than the Patel-Teja with an AAD value of 6%.

4.2Prediction of liquid Densities

Density of the liquid phase is extremely important in several engineering applications such as pipeline simulation and separator design and so it is therefore important that this property is predicted accurately. Similar to the results of the predictions of liquid phase Compositions, the new equation proved to be more accurate in the prediction of Liquid phase densities when compared to the predictions made by Peng-Robinson and Patel-Teja. Below is a list of mixtures for which this property was predicted at different pressures and temperatures.

r	1		1
System	NDP	T range (°F)	P range (psia)
C3 -nC4	101	-29.4 – 224.3	67 – 1026.1
C3 -nC4	96	-27.67 – 224.33	69 – 1032.15
C3 -nC4	90	-27.67 – 224.33	71 - 1022
C3 -iC4	100	-27.67 – 224.33	57 - 1025
C3 -iC4	80	-9.67 – 224.33	67 - 1022
C3 -iC4	85	-27.67 – 188.33	36 - 1028
nC4-iC4	102	-27.67 – 224.33	69 - 1022
nC4-iC4	105	-27.67 – 224.33	45 - 1029
nC4-iC4	105	-27.67 – 224.33	71 – 1026
C3-nC4-iC4	101	-27.67 – 224.33	72 - 1025
C3-nC4-iC4	100	-27.67 – 224.33	70 - 1027
C3-nC4-iC4	94	-27.67 – 224.33	22 - 1028
Total	1159		•

Table 4-8: Mixtures for which Liquid Density is evaluated (Kayukawa, et al., 2005)

It is important to point out that some of the mixtures have similar components but the

compositions of these components vary.
Propane-nButane System

For this study, three mixtures containing these components were investigated and below are the results

a) Mixture 1 (25% Propane):

The density of the liquid phase of the binary mixture was measured experimentally at temperatures ranging from -29 to 225 °F and pressures ranging from 65 to 1030 psia. The density was then calculated and an error analysis was then carried out. As seen in the table 4.2B, Patel-Teja performs better than Peng-Robinson with an AAD value of 1.77% in comparison to 4.44% recorded by Peng-Robinson. The new EOS performs slightly better than Patel-Teja recording an AAD value of 1.53% and the same pattern is observed in the RMSD values recorded by the different models.

EOS	AAD	RMSD
PR	0.044423	0.047613
PT	0.017788	0.021389
NEW	0.015258	0.016499

Table 4-9: Error Analysis of Liquid densities for 25% Propane-nButane system



Figure 4-4: Column Chart of AAD errors for a 25% Propane-nButane system

b) Mixture 2 (50% Propane):

The density of the liquid phase of this binary mixture was measured experimentally at temperatures ranging from -27 to 225 °F and pressures ranging from 69 to 1030 psia. The calculated densities at various combinations of temperatures and pressures is recorded in Appendix A. Similar to mixture 1, the new equation performs better than Peng-Robinson and Patel-Teja, recording the least AAD and RMSD values as shown in table 4-10

Table 4-10: Error Analysis of Liquid densities for a 50% Propane-nButane system

EOS	AAD	RMSD
PR	0.045672	0.049158
PT	0.02005	0.02626
NEW	0.015812	0.017098

c) Mixture 3 (75% Propane):

The densities for this mixture were measured at temperatures ranging from -27 to 225 °F and pressures ranging from 70 and 1025 psia. Below is a summary of the Error analysis of the predictions made by the different EOS models.

Table 4-11: Error Analysis of Liquid densities for a 75% Propane-nButane system

EOS	AAD	RMSD
PR	0.046572	0.050793
PT	0.022692	0.030804
NEW	0.015315	0.01686



Figure 4-5: Column Chart of AAD errors for a 75% Propane-nButane system

As shown in table 4-11 and figure 4-5, the new equation of state performed the best with AAD value of **1.53%** and RMSD value **1.69%**.

Propane-iButane System

Three mixtures with different compositions of propane and i-butane were explored. Below are the results:

a) Mixture 4 (25% Propane):

The densities for this mixture were measured at temperatures ranging from -27 to 225 °F and pressures ranging from 57 and 1025 psia. Below is a summary of the Error analysis of the predictions made by the different EOS models.

EOS	AAD	RMSD
PR	0.048229	0.052024
PT	0.022259	0.028688
NEW	0.01842	0.020072

Table 4-12: Error Analysis of Liquid densities for a 25% Propane-iButane system

As shown in table 4-12, the new equation of state performed the best with AAD value of 1.84% and RMSD value 2%.

b) Mixture 5 (50% Propane):

The densities for this mixture were measured at temperatures ranging from -9 to 225 °F and pressures ranging from 67 and 1025 psia. Below is a summary of the Error analysis of the predictions made by the different EOS models.

EOS	AAD	RMSD
PR	0.047427	0.051606
PT	0.025182	0.036104
NEW	0.018133	0.019995

Table 4-13: Error Analysis of Liquid densities for a 50% Propane-iButane system



Figure 4-6: Column Chart of AAD errors for a 50% Propane-iButane system

As shown in Table 4-13 and figure 4-6, the predictions of the new equation of state is more accurate as highlighted by the low values of AAD and RMSD in comparison to Peng-Robinson and Patel-Teja.

c) Mixture 6 (75% Propane):

The densities for this mixture were measured at temperatures ranging from -27 to 190 °F and pressures ranging from 36 and 1028 psia. Below is a summary of the Error analysis of the predictions made by the different EOS models.

EOS	AAD	RMSD
PR	0.048582	0.052544
РТ	0.021083	0.028635
NEW	0.017064	0.018636

Table 4-14: Error Analysis of Liquid densities for a 75% Propane-iButane system

As shown in Table 4-14, the predictions of the new equation of state is more accurate as highlighted by the low values of AAD and RMSD in comparison to Peng-Robinson and Patel-Teja.

nButane-iButane System

For this hydrocarbon system, three mixtures were explored with varying compositions of the components. Below are the results of the error analysis carried out on the predictions made by the different equations of state:

a) Mixture 7 (25% nButane)

The densities for this mixture were measured at temperatures ranging from -27 to 225 °F and pressures ranging from 69 and 1025 psia. Below is a summary of the Error analysis of the predictions made by the different EOS models.

EOS	AAD	RMSD
PR	0.04955	0.052834
РТ	0.020073	0.022538
NEW	0.019764	0.021015

Table 4-15: Error Analysis of Liquid densities for a 25% nButane-iButane system

As shown in Table 4-15, the new equation of state performs better than Peng-Robinson and slightly better Patel-Teja

b) Mixture 8 (50% nButane)

The experimental data for this mixture were obtained at temperatures ranging from -27 to 225 ^oF and pressures ranging from 45 and 1030 psia. A summary of the results of the Error Analysis is shown in Table 4-16.

EOS	AAD	RMSD
PR	0.048007	0.050968
PT	0.018721	0.020869
NEW	0.018179	0.01939

Table 4-16: Error Analysis of Liquid densities for a 50% nButane-iButane system

Similar to Mixture 7, the new equation performs better than Peng-Robinson but just slightly better than Patel-Teja.

c) Mixture 9 (75% nButane)

The experimental data for this mixture were obtained at temperatures ranging from -27 to 225 °F and pressures ranging from 70 and 1026 psia. A summary of the results of the Error Analysis is shown in Table 4-17.

EOS	AAD	RMSD
PR	0.049139	0.053336
PT	0.020472	0.024836
NEW	0.019399	0.023245

Table 4-17: Error Analysis of Liquid densities for a 75% nButane-iButane system

The new equation and Patel-Teja performed very well in their predictions as highlighted by the minute difference in their AAD and RMSD values.

Propane-nButane-iButane System

Similar to the previous hydrocarbon system, three mixtures were explored with varying compositions of the components. Below are the results of the error analysis carried out on the predictions made by the different equations of state:

a) Mixture 10 (20% Propane & 60% n-butane)

The experimental data for this mixture were obtained at temperatures ranging from -27 to 225 °F and pressures ranging from 72 and 1025 psia. A summary of the results of the Error Analysis is shown in Table 4-18.

Table 4-18: Error Analysis of Liquid densities for Mixture 10 (20% Propane & 60% n-butane)

EOS	AAD	RMSD
PR	0.044374	0.047637
PT	0.017841	0.021549
NEW	0.01523	0.016503

b) Mixture 11 (34% Propane & 33% n-butane)

The experimental data for this mixture were obtained at temperatures ranging from -27 to 225 ^oF and pressures ranging from 70 and 1027 psia. A summary of the results of the Error Analysis is shown in Table 4-19.

Table 4-19: Error Analysis of Liquid densities for Mixture 11 (34% Propane & 33% n-butane)

EOS	AAD	RMSD
PR	0.046769	0.050461
PT	0.019939	0.024905
NEW	0.016815	0.018273



Figure 4-7: Column Chart of RMSD errors for Mixture 11 (34% Propane & 33% n-butane)

c) Mixture 12 (60% Propane & 20% n-butane): The experimental data for this mixture were obtained at temperatures ranging from -27 to 225 °F and pressures ranging from 22 and 1028 psia. A summary of the results of the Error Analysis is shown in Table 4-20.

Table 4-20: Error Analysis of Liquid densities for Mixture 12 (60% Propane & 20% n-butane)

EOS	AAD	RMSD
PR	0.047701	0.051378
РТ	0.021401	0.028754
NEW	0.016178	0.017488

Observing Tables 4-18, 4-19 and 4.20, the new equation of state excels in the three mixtures given that it reports the least AAD and RMSD values in all cases.

In addition to the separate error analysis carried out for every mixture explored, the data for all

12 mixtures were compiled and then error analysis was carried out. Observing figure 4-8 and 4-

9, the plot of the predictions of the new EOS model is closer to the 45-degree Line in

comparison to the other two EOS models thereby showing that the new EOS improves

predictions of liquid density.

EOS	AAD	RMS
PR	0.046866	0.05042
PT	0.020256	0.025988
NEW	0.016874	0.018325

Table 4-21: Summary of Error Analysis of Liquid densities for all Hydrocarbon systemsconsidered



Figure 4-8: Plot Showing Density measurements of PR and New EOS for all mixtures



Figure 4-9: Plot Showing Density measurements of PT and New EOS for all mixtures

In spite of this improvement, further analysis was carried out to investigate the pressure and temperature region where the NEW EOS performs the best and where it fails. In order to do this, predictions of density at same pressures or temperatures were collated and then error analysis was carried out. Below is a summary of the results obtained:

4.2.1 Error Analysis for pressure

		AAD		
Pressure	PR	PT	NEW	No. of Data Points
(psia)				
0-100	0.058811157	0.016115	0.017788	42
100-200	0.05362975	0.013671	0.015504	55
200-300	0.047074924	0.01624	0.014726	146
300-400	0.044528371	0.021158	0.016175	85

 Table 4-22: Summary of AAD calculations at different pressure ranges

400-500	0.044789924	0.022774	0.01653	91
500-600	0.045441279	0.02434	0.017178	184
600-700	0.045945743	0.023409	0.017149	94
700-800	0.045795827	0.022103	0.017488	163
800-900	0.04794769	0.01993	0.018099	117
900-1000	0.047954235	0.020202	0.018983	96
>1000	0.048907717	0.018598	0.019674	94

Table 4-23: Summary of RMSD calculations at different pressure ranges

		RMSD		
Pressure	PR	РТ	NEW	No. of Data Points
(psia) 0-100	0.058937271	0.016558	0.017898	42
100-200	0.054503959	0.01506	0.016303	55
200-300	0.050135625	0.018984	0.016014	146
300-400	0.048260901	0.027125	0.017747	85
400-500	0.04863076	0.030253	0.018265	91
500-600	0.049561966	0.033165	0.018741	184
600-700	0.050198189	0.030869	0.018973	94
700-800	0.05035919	0.028027	0.019821	163
800-900	0.051533975	0.024175	0.019805	117
900-1000	0.052133822	0.024259	0.021224	96
>1000	0.052597637	0.021584	0.021454	94

Figure 4-10: Plot Showing deviations of AAD values at different pressure ranges





Figure 4-11: Plot Showing deviations of RMSD values at different pressure ranges

Observing figure 4-10 and 4.11, the new EOS shows to have the least AAD and RMSD values for most of the pressure ranges tested. At extremely low pressures (0-100 psia and 100-200 psia) and pressures over 1000psia, Patel-Teja provides slightly better predictions but in all pressure ranges observed, the new equation performs better than Peng-Robinson.

4.2.2 Error Analysis for Temperature

		AAD		
Temp (deg F)	PR	РТ	NEW	No. of data points
-27.67	0.062264	0.020188	0.019975	144
-9.67	0.067855	0.021894	0.024226	14
8.33	0.062553	0.020267	0.021703	154
44.33	0.060947	0.018094	0.022139	167
80.33	0.055609	0.013044	0.020111	161
116.33	0.045454	0.008716	0.015098	150
152.33	0.031195	0.013446	0.010258	140
188.33	0.018129	0.030997	0.010067	125
224.33	0.022119	0.052793	0.013732	98

Table 4-24: Summary of AAD calculations at different temperatures

		RMS		
Temp (deg F)	PR	РТ	NEW	No. of data points
-27.67	0.052598	0.021584	0.021454	144
-9.67	0.067865	0.021939	0.024276	14
8.33	0.062732	0.020911	0.022181	154
44.33	0.061157	0.019197	0.02277	167
80.33	0.055997	0.015434	0.0212	161
116.33	0.046058	0.010512	0.016359	150
152.33	0.033271	0.017758	0.012023	140
188.33	0.021382	0.037415	0.012633	125
224.33	0.029336	0.058455	0.017458	98

Table 4-25: Summary of RMSD calculations different temperatures



Figure 4-12: Plot Showing deviations of AAD values at different temperatures



Figure 4-13: Plot Showing deviations of RMSD values at different temperatures

Observing figure 4-12 and 4-13, the new equation performs better than Peng Robinson for all temperatures explored but this is not the case with Patel-Teja. For temperatures below 130 °F, Patel-Teja performs slightly better than the new equation with the largest disparity in predictions occurring between 70 and 120 °F (about 0.6% difference in error). Above 130 °F, the new equation is superior in its predictions for hydrocarbon mixtures with the largest disparity occurring at about 225 °F (about 5% difference in error). Patel-Teja EOS performs better than Peng-Robinson at temperatures lesser than 170°F but at higher temperatures, Peng-Robinson makes more accurate predictions.

4.3 Prediction of liquid Viscosities

Similar to analysis carried out when calculating the density of the liquid phase, the viscosity of the liquid phase was calculated for different hydrocarbon systems and the predictions made by the different EOS models were compared. Below is a list of mixtures that were explored for this analysis:

System	NDP	T range (°R)	P range (psia)
C1 -C2 (Diller, 1984)	7	360 - 451	589 - 3120
C1 -C2 (Diller, 1984)	8	306 - 414	396 -2597
C1 -C2 (Diller, 1984)	4	414	859 – 2079
C1-nC7	16	582 - 852	5012 - 17505
N2, Co2, C1, C2, C3, iC4, nC4, ic5, nC5, nC6, nC7, nC8, nC9, nC10, nC11, C12+	4	243	2814 - 5015
C3	12	500 - 860	1016 - 4981
nC4	16	560 - 680	200 - 3000
Total	67		

Table 4-26: Showing the Mixtures for which the Liquid Viscosity is evaluated

1) Methane-Ethane Mixture:

For this hydrocarbon system, three mixtures of varying compositions of both components are explored for different pressures and temperatures. The experimental values and predictions made by the three different EOS are summarized in Appendix A. Below are the summary of the error calculations made for the different mixtures: a) Mixture 1 (50% methane):

Table 4-27: Error Analysis of liquid viscosity for 50% Methane-Ethane Mixture

EOS	AAD	RMSD
PR	0.157087	0.190881
PT	0.065532	0.067723
NEW	0.017346	0.020361



Figure 4-14: Column chart of RMSD analysis for 50% Methane-Ethane Mixture

b) Mixture 2 (68% methane):

Table 4-28: Error Analysis of liquid viscosity for 68% Methane-Ethane Mixture

EOS	AAD	RMS
PR	0.140512	0.164604
РТ	0.065816	0.068143
NEW	0.018544	0.023344



Figure 4-15: Column chart of AAD analysis for 68% Methane-Ethane Mixture

c) Mixture 3 (34.53% methane)

EOS	AAD	RMS
PR	0.212122	0.219679
PT	0.07541	0.07922
NEW	0.025108	0.030066

Table 4-29: Error Analysis of liquid viscosity for 34.53% Methane-Ethane Mixture



Figure 4-16: Plot Showing Viscosity measurements for 34.53% Methane-Ethane Mixture

Based on tables 4-27, 4-28 and 4-29, the new equation of state has the least AAD and RMSD values therefore it makes better predictions in comparison to Peng-Robinson and Patel-Teja. It was also observed that at higher compositions of the Lighter Component (Methane in this case), the AAD and RMSD of the different EOS models are lower hence predictions are better. The reason for this is because the higher the mole fraction of the lighter component in a binary system, the lesser the impact of the disparity in sizes of the components hence the Binary interaction parameters are of less importance.

2) Mixture 4:

This mixture constitutes of methane and n-heptane and experimental values were provided between temperatures of 582-852 °F and pressures of 5000 – 18000 psia. It is important to note that for this hydrocarbon system, the conditions explored are in the single phase region so the mixture does not split into two phases. Below are the results of the error analysis of the predictions made by the different EOS model.

EOS	AAD	RMS
PR	0.266353	0.293955
PT	0.237758	0.264746
NEW	0.204611	0.232466

Table 4-30: Error Analysis of liquid viscosities for Methane- nHeptane Mixture



Figure 4-17: Column chart of RMSD analysis for Methane- nHeptane Mixture

From figure 4-17, the new equation shows a slight improvement over Peng-Robinson and Patel-Teja EOS. Compared to the methane-ethane system, it is observed that the errors recorded in terms of AAD and RMS value for this hydrocarbon system is much higher and this is because the size of n-C7 molecule is much large than methane. Although this was not done in this study, the inclusion of Binary-interaction parameters will lead to a tremendous improvement in the predictions. 3) Mixture 5: This is a gas condensate system consisting of hydrocarbon components ranging from methane to n-C12+. Also included in this system, are acid gases such as carbon dioxide. Viscosities were measured experimentally at various pressures with a temperature of 243°F. Below are the predictions made by the different EOS models at these conditions:

EOS	AAD	RMS
PR	0.248448	0.308386
PT	0.336466	0.429524
NEW	0.216662	0.270366

Table 4-31: Error Analysis of liquid viscosities for gas condensate system

As observed from Table 4-31, the new equation of state shows an improvement in Viscosity predictions over Peng-Robinson and Patel-Teja. Another observation made, is that Peng-Robinson performs better than Patel-Teja which is usually not the case because Patel has an additional parameter in its equation that helps to improve its accuracy. The reason for this can be attributed to the behavior of the different EOS models in figure 4-12 and 4-13 where it is observed that above 170°F, the Patel-Teja EOS performs poorly in comparison to Peng-Robinson and the New equation when dealing with hydrocarbon mixtures.

4) Pure Substances:

In addition to numerous mixtures that have been explored, two single component systems were explored between temperatures of 40 – 400°F and pressures 200 – 5000 psia. Below are details on the results of error analysis carried out on the two pure substances explored:

a) Propane:

EOS	AAD	RMS
PR	0.078576	0.101968
РТ	0.105305	0.112203
NEW	0.033188	0.043224

Table 4-32: Error Analysis of liquid viscosities for Propane



Figure 4-18: Column chart of AAD analysis for Propane

b) N-butane:

EOS	AAD	RMS
PR	0.126353	0.1311
PT	0.059929	0.063038
NEW	0.036323	0.0429

Table 4-33: Error Analysis of liquid viscosities for n-Butane



Figure 4-19: Column chart of AAD analysis for n-Butane

It can be observed in both cases that the new equation improved the accuracy of predictions by having the least AAD values in comparison to Patel-Teja and Peng-Robinson EOS. Another observation is that Peng-Robinson performs better than Patel-Teja for propane as shown in figure 4-18 while the reverse is the case for the N-butane. The reason for this can be attributed to the observation made in figure 4-12 and 4-13. For propane, the temperatures for which calculations were carried out were way above 170 °F, which is the region where Peng-Robinson performs better than Patel-Teja EOS. For butane, most of the temperatures investigated are found below 170 °F where Peng-Robinson's performance is not as good as Patel-Teja.

In addition, the data for the 7 hydrocarbon systems were compiled and error analysis was carried out. Below is a summary of the results of the analysis:

EOS	AAD	RMS
PR	0.205665	0.28046
РТ	0.129636	0.177353
NEW	0.085501	0.135771

Figure 4-20: Summarized Results for all Hydrocarbon systems considered Table 4.20 reveals that the new equation of state performs the best due to having the least Average absolute errors (8.6%) and Root mean square error (13.6%). it is also observed that the errors recorded in this analysis is relatively higher in comparison to the results obtained from analyzing liquid densities (Table 4-21). The reason for this is that the components of the mixtures analyzed for liquid viscosities are more dissimilar in shape and molecular weight compared to the mixtures analyzed for liquid densities.

4.4 Limitations of new equation of state

Despite the superiority of the new equation in the prediction of Liquid properties when dealing with mixtures containing lighter components, the equations has shown not have the best predictions under certain conditions that will be elaborated below.

4.4.1 <u>Mixtures with heavy hydrocarbons</u>

The new equation shows its limitations when dealing with mixtures containing heavy hydrocarbons (components with molecular weight over 90 lb/lbmol). Below is list of various heavy hydrocarbon mixtures for which liquid viscosities are calculated:

Mixture	Components	NDP	T range	P range
			(deg F)	(psia)
1	nC4, nC6,Co2, nC10	5	188-252	39 - 60
2	nC4, nC6, nC10, Co2	5	188 - 252	365
3	nC5, nC6. nC7, nC10, Co2	4	188 - 252	715
4	nC5-nC6-nC7-nC10-Co2	4	177 - 263	34 - 75
5	nC4,nC6,nC10	5	123 - 252	44 - 95
6	nC5,nC6, nC7,nC10	4	123 - 252	365
7	nC5, nC10	5	123 - 252	715

Table 4-34: List of Heavy Hydrocarbon Mixtures (Barrufet & El Sayed Salem, 1996)

1) Mixtures with less than 60% of heaviest component

For this category, 4 mixtures were explored at various pressures and temperatures. All EOS models were able to make reasonable predictions but Patel-Teja performance is superior in comparison to New and Peng-Robinson EOS. This is in line with the observation made in Patel & Teja (1982) where it is stated that the most important

feature of the Patel-Teja equation is its applicability to mixtures containing heavy

hydrocarbons. Below is a summary of the error analysis of the calculations made:

a) Mixture 1:

EOS	AAD	RMS
PR	0.299912	0.302235
РТ	0.112053	0.131736
NEW	0.354617	0.35795

Table 4-35: Error Analysis of liquid viscosities for Mixture 1 (nC4-nC6-nC10-Co2)



Figure 4-21: Plot of Viscosity predictions for Mixture 1 (nC4-nC6-nC10-Co2)

b) Mixture 2:

EOS	AAD	RMS
PR	0.20369	0.205543
РТ	0.047837	0.054051
NEW	0.269995	0.271983

Table 4-36: Error Analysis of liquid viscosities for Mixture 2 (nC4-nC6-nC10-Co2)



Figure 4-22: Plot of Viscosity predictions for Mixture 2 (nC4-nC6-nC10-Co2)

c) Mixture 3:

EOS	AAD	RMS
PR	0.29034082	0.29063
РТ	0.132705228	0.134141
NEW	0.359837363	0.360088

Table 4-37: Error Analysis of liquid viscosities for Mixture 3 (nC5-nC6-nC7-nC10-Co2)



Figure 4-23: Plot of Viscosity predictions for Mixture 3 (nC5-nC6-nC7-nC10-Co2)

d) Mixture 4:

EOS	AAD	RMS
PR	0.20006	0.20173
РТ	0.052481	0.058591
NEW	0.27773	0.278603

Table 4-38: Error Analysis of liquid viscosities for Mixture 4 (nC5-nC6-nC7-nC10-Co2)



Figure 4-24: Plot of Viscosity predictions for Mixture 4 (nC5-nC6-nC7-nC10-Co2)

Observing Table 4-37, 4-38, 4-39 and 4-40, Patel-Teja makes the most accurate predictions due to it having the least AAD and RMS values in all cases. It is also observed that for this set of mixtures, Peng-Robinson performs better than the new equation of state and this due to the improvement made by Peng-Robinson on their original EOS with the goal of yielding more accurate predictions of vapor pressure for heavy hydrocarbons (Robinson & Peng, 1978). 2) Mixtures with over 60% of heaviest component:

For this category, the liquid viscosities of a set of three mixtures were explored at various pressures and temperatures. The Patel-Teja EOS performed well in comparison to the new and Peng-Robinson EOS. It was also observed that Peng-Robinson estimated unreasonable liquid Viscosities at the various pressures and temperatures for these set of mixtures hence high percentage errors were recorded during error analysis. Below is a summary of the error calculations that was carried out from the predictions.

a) Mixture 5 (61.4% n-Decane)

EOS	AAD	RMS
PR	0.587239	0.669012334
РТ	0.140953	0.141541604
NEW	0.387007	0.387140312

 Table 4-39: Error Analysis of liquid viscosities for Mixture 5(nC4-nC6-nC10)



Figure 4-25: Plot showing Viscosity predictions for Mixture 5

b) Mixture 6 (70% n-Decane)

EOS	AAD	RMS
PR	0.969925	0.969947
РТ	0.157133	0.158666
NEW	0.41553	0.415797

Table 4-40: Error Analysis of liquid viscosities for Mixture 6 (nC5-nC10)



Figure 4-26: Plot showing Viscosity predictions for Mixture 6 (nC5-nC10)

c) Mixture 7 (51.1% n-Decane & 10.3% n – Hexane)

EOS	AAD	RMS
PR	0.966823	0.96683
РТ	0.122662	0.127641
NEW	0.366396	0.367215

Table 4-41: Error Analysis of liquid viscosities for Mixture 7 (nC5-nC6-nC7-nC10)



Figure 4-27: Plot showing Viscosity predictions for Mixture 7 (nC5-nC6-nC7-nC10)

It can be observed from Table 4-41, 4-42 and 4-43 that Patel-Teja performs the best and Peng-Robinson is unable to make reasonable predictions hence the extremely high AAD and RMS values in all cases.

4.4.2 <u>Evaluation of Gas Property (Z factor)</u>

For this section, several hydrocarbon mixtures including gas condensates were evaluated for gas Z factor and then error analysis was carried out on the predictions to evaluate which EOS performs the best. Below is a list of the hydrocarbon systems that were explored for this property.

Table 4-42: List of Hydrocarbon systems used for gas Z factor calculations (Nwankwo, 2014)

System	NDP	T range (deg F)	P range (psia)
H2S,Co2, N2, C1, C2,C3,iC4,nC4,iC5,nC5,nC6,C7+	7	219	700 - 4825
Co2, N2, C1, C2,C3,iC4,nC4,iC5,nC5,nC6,C7+	7	209	700 - 4786
H2S,Co2, N2, C1, C2,C3,iC4,nC4,iC5,nC5,nC6,C7+	7	250	700 - 4190
C1,C2,C3,nC4,N2,Co2	33	77 - 122	26 - 2176

C1,C2,C3, nC4	35	77 - 1225	19 - 2332
Total	89		

Composition							
H2S	0.003	0.003	0.003	0.003	0.003	0.003	0.004
Co2	0.6352	0.6395	0.6514	0.6579	0.6639	0.6706	0.6716
N2	0.0386	0.0399	0.041	0.0417	0.0421	0.0411	0.0388
C1	0.1937	0.1988	0.2008	0.207	0.2084	0.2037	0.1994
C2	0.0303	0.0307	0.0308	0.0309	0.0313	0.0315	0.0318
C3	0.0174	0.0172	0.017	0.0169	0.017	0.0145	0.0184
i-C4	0.0033	0.0032	0.0031	0.003	0.003	0.0032	0.0035
n-C4	0.0093	0.0088	0.0085	0.0082	0.0082	0.0088	0.0097
i-C5	0.0039	0.0036	0.0033	0.0031	0.003	0.0033	0.0039
n-C5	0.0047	0.0042	0.0038	0.0036	0.0035	0.0038	0.0046
n-C6	0.0051	0.0049	0.0046	0.0042	0.0036	0.003	0.0034
C7+	0.0551	0.0458	0.0324	0.0202	0.0127	0.0101	0.0113
P (psia)	4825	4100	3300	2600	1900	1200	700
T (deg. F)	219	219	219	219	219	219	219
Expt Z	0.951	0.8276	0.7522	0.7483	0.7882	0.8437	0.8167
Z (PR)	0.804066	0.730493	0.684045	0.690295	0.737325	0.816939	0.882717
Z (PT)	0.815727	0.745295	0.699514	0.704793	0.74981	0.826103	0.888676
Z (New)	0.791333	0.714782	0.661824	0.664282	0.71682	0.807015	0.878603

1) Mixture 1 (Carbon Dioxide-Rich and Sour Gas Condensate):

Table 4-43: Composition of Carbon Dioxide-Rich and Sour Gas Condensate Mixture

Eos	AAD	RMS
PR	0.088152	0.095292
РТ	0.075367	0.083652
NEW	0.10664	0.113169

Table 4-44: Error Analysis of Gas Compressibility for Carbon Dioxide-Rich and Sour Gas Condensate Mixture



Figure 4-28: Column chart of AAD analysis for Carbon Dioxide-Rich and Sour Gas Condensate Mixture

Composition							
H2S	0	0	0	0	0	0	0
Co2	0.0033	0.0033	0.0034	0.0035	0.0035	0.0036	0.0038
N2	0.0032	0.0033	0.0033	0.0033	0.0033	0.0033	0.0033
C1	0.942	0.9438	0.9451	0.9461	0.9468	0.9473	0.9467
C2	0.231	0.023	0.231	0.0231	0.0232	0.0233	0.0236
C3	0.0082	0.0082	0.0082	0.0082	0.0082	0.0082	0.0083
i-C4	0.0023	0.0023	0.0023	0.0023	0.0023	0.0023	0.0023
n-C4	0.0025	0.0025	0.0025	0.0025	0.0025	0.0025	0.0026
i-C5	0.0012	0.0012	0.0012	0.0012	0.0012	0.0012	0.0012
n-C5	0.0008	0.0008	0.0008	0.0008	0.0008	0.0008	0.0009
n-C6	0.0014	0.0013	0.0013	0.0013	0.0013	0.0012	0.0013
C7+	0.0012	0.0103	0.0089	0.0077	0.0069	0.0063	0.006
P (psia)	4786	4000	3300	2600	1900	1300	700
T (deg. F)	209	209	209	209	209	209	209
Expt Z	1.019	0.984	0.895	0.933	0.933	0.947	0.969
Z (PR)	0.923624	0.9327	0.80435	0.899014	0.902049	0.918584	0.948322
Z (PT)	0.981765	0.975892	0.852872	0.93283	0.930046	0.94015	0.961422
Z (New)	0.936692	0.940967	0.805447	0.904519	0.907684	0.924301	0.95284

2) Mixture 2 (Sweet Gas Condensate System):

Table 4-45: Composition of Sweet Gas Condensate System and Gas Z-factor predictions

Eos	AAD	RMS
PR	0.052566	0.06038
РТ	0.015751	0.023122
NEW	0.046125	0.054732

Table 4-46: Error Analysis of Sweet Gas Condensate System



Figure 4-29: Column chart of AAD analysis for Sweet Gas Condensate System

Composition							
H2S	0.282	0.277	0.272	0.27	0.273	0.278	0.318
Co2	0.0608	0.0644	0.0669	0.0685	0.0694	0.0699	0.0679
N2	0.0383	0.0455	0.0476	0.0473	0.0461	0.0434	0.0394
C1	0.4033	0.4382	0.4641	0.4807	0.4844	0.4688	0.4331
C2	0.0448	0.0471	0.0481	0.0487	0.0493	0.0496	0.0494
C3	0.0248	0.0243	0.0239	0.0237	0.0239	0.0252	0.0277
i-C4	0.006	0.0055	0.0051	0.0049	0.0049	0.0055	0.0067
n-C4	0.0132	0.012	0.0111	0.0106	0.0106	0.0114	0.014
i-C5	0.0079	0.0068	0.006	0.0055	0.0053	0.0058	0.0074
n-C5	0.0081	0.0069	0.006	0.0054	0.0052	0.0057	0.0071
n-C6	0.0121	0.0069	0.0078	0.0066	0.006	0.0063	0.0077
C7+	0.0991	0.063	0.0412	0.0286	0.0217	0.0192	0.0214
P (psia)	4190	3600	3000	2400	1800	1200	700
T (deg. F)	250	250	250	250	250	250	250
Expt Z	0.883	0.806	0.799	0.809	0.842	0.888	0.935
Z (PR)	0.773871	0.722339	0.721336	0.742898	0.780642	0.83633	0.882337
Z (PT)	0.795175	0.75323	0.751054	0.770082	0.804151	0.854058	0.894108
Z (New)	0.772407	0.72026	0.716062	0.738191	0.779657	0.839119	0.886148

3) Mixture 3 (Highly Sour Gas Condensate System):

Table 4-47: Composition of Highly Sour Gas Condensate System and Gas Z-factor predictions

Eos	AAD	RMS
PR	0.084811	0.087864
PT	0.057137	0.060343
NEW	0.086327	0.089985

Table 4-48: Error Analysis of Highly Sour Gas Condensate System



Figure 4-30: Column chart of AAD analysis for Highly Sour Gas Condensate System

4) Mixture 4 (C1-C2-C3-nC4-N2-Co2):

EOS	AAD	RMS
PR	0.011326	0.015113
РТ	0.002164	0.00368
NEW	0.007748	0.010815

Table 4-49: Error Analysis of Mixture 4 (C1-C2-C3-nC4-N2-Co2)



Figure 4-31: Column chart of AAD analysis for Mixture 4 (C1-C2-C3-nC4-N2-Co2)

5) Mixture 5 (C1-C2-C3-nC4):

EOS	AAD	RMS
PR	0.012701	0.016678
PT	0.003857	0.007238
NEW	0.008509	0.011714

Table 4-50: Error Analysis of Mixture 5 (C1-C2-C3-nC4)



Figure 4-32: Column chart of AAD analysis for Mixture 5 (C1-C2-C3-nC4)

Observing the above hydrocarbon systems, Patel-Teja outperforms both the new and Peng-Robinson equation in the prediction of gas compressibility for all 5 hydrocarbon systems. It is also important to note that Peng-Robinson out performs the new equation in sour condensate systems (presence of H₂s,) and systems where CO₂ constitutes over 2% of the mixture as shown in Table 4-46 and Table 4-50. Generally, the new equation performs better than Peng-Robinson for sweet condensate mixtures that have less than 2% CO₂ as shown in mixture 2 (Table 4-48), mixture 4 (Table 4-51) and mixture 5 (Table 4-52).

CHAPTER 5

CONCLUSIONS AND RECOMMENDATIONS

Based on the Vapor-Liquid equilibria calculations carried out using the New equation of state, the following inferences can be made:

5.1 <u>CONCLUSIONS</u>

- The new equation of state was successfully applied to multi-component systems and showed better performance_compared to Peng-Robinson and Patel-Teja when making predictions for properties of the liquid phase (composition, density and viscosity). This is in spite of implementing the Peneloux volume shift for the Peng-Robinson equation
- The new equation makes better predictions compared to Peng-Robinson EOS when dealing with hydrocarbon systems having a high composition of heavy components.
 Peng Robinson is unable to handle mixtures that constitute over 60% of heavy components.
- The new equation of state makes comparable predictions to Patel-Teja when calculating Liquid densities at temperatures below 130°F but at higher temperatures, the new equation is superior.
- The new equation makes comparable predictions to Peng-Robinson for gas phase properties (gas Z factor in this case) when dealing with sweet condensate mixtures (containing no H₂S) containing little to no carbon dioxide. It under performs when dealing with sour condensate mixtures and condensate mixtures with high CO₂ content.

- Patel-Teja is superior in performance for gas Z factor predictions when dealing with both sweet and sour condensate mixtures. It is also makes the best predictions when dealing with mixtures that have high CO₂ content.
- The Patel-Teja EOS performs better than the new equation of state when predicting liquid densities for hydrocarbon mixtures that have high compositions of heavy components.

5.2 <u>Recommendations</u>

- Improve the 'm' parameter correlation to fine tune the predictions made by the new equation so as to yield better predictions of vapor phase property and more accurate predictions for heavier components
- Develop suitable Binary interactions parameters that will help improve the predictions the new equation of state
- Further studies can be carried out by using the new equation of state in three phase flash calculations

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APPENDIX A: Physical Properties for Pure Components

Compound	Formula	PC	TC	Acentric	$MW\left(\frac{lb}{lbmol}\right)$	$VC\left(\frac{ft^3}{n}\right)$
		(psia)	(°F)	Factor		(lbm)
Methane	CH ₄	667.8	-116.63	0.0104	16.043	0.098797
Ethane	C_2H_6	707.8	90.09	0.0979	30.07	0.078284
Propane	C₃H ₈	616.3	206.01	0.1524	44.097	0.072703
Isobutane	C ₄ H ₁₀	529.1	274.98	0.1848	58.124	0.071399
n-Butane	C ₄ H ₁₀	550.7	305.65	0.201	58.124	0.070298
I-Pentane	C ₅ H ₁₂	490.4	369.1	0.2223	72.151	0.067899
Neopentane	C_5H_{12}	464	321.13	0.1969	72.151	0.0673
n-Pentane	C ₅ H ₁₂	488.6	385.7	0.2539	72.151	0.067497
n-Hexane	C ₆ H ₁₄	436.9	453.7	0.3007	86.178	0.068799
n-Heptane	C ₇ H ₁₆	396.8	512.8	0.3498	100.205	0.069098
n-decane	$C_{10}H_{22}$	304	652.1	0.4885	142.266	0.067908
Hydrogen Sulphide	H₂S	1300	212.45	0.0948	34.08	0.046097
Carbon Dioxide	CO ₂	1071	87.9	0.2667	44.01	0.034401
Nitrogen	N ₂	493.1	-232.51	0.0372	28.0134	0.051011
Water	H ₂ O	3198.8	705.16	0.3443	18.0153	0.049
Oxygen	02	731.4	-181.43	0.0216	31.9988	0.0367

Table 5-1: Critical Properties

APPENDIX B: Results for New EOS

Liquid Composition Predictions

1) Mixture 1(80% CH₄- C₃H₈):

Components	Mol. Frac.	PC (psia)	TC (°F)	Acentric Factor	$MW\left(\frac{lb}{lbmol}\right)$	$\operatorname{VC}\left(\frac{ft^{3}}{lbm}\right)$
CH_4	0.8	667.8	-116.63	0.0104	16.043	0.098797
C_3H_8	0.2	616.3	206.01	0.1524	44.097	0.072703

Table 5-2: Properties of Components for Mixture 1

Table 5-3: Comparison of Liquid Compositions for Mixture 1

Т	Р	Expt. CH₄	PR	РТ	NEW
(°F)	(psia)	Composition			
-75	51.5	0.0443	0.80000002	0.037368	0.039956
-75	100	0.0899	0.80000004	0.077216	0.085603
-75	150	0.1358	0.146810679	0.11791	0.132093
-75	300	0.2709	0.291953962	0.238061	0.268685
-75	400	0.3656	0.384887057	0.317052	0.358037
-75	500	0.458	0.475493532	0.395731	0.44683
-108.4	30.5	0.0377	0.8	0.032284	0.035305
-108.4	52.5	0.0667	0.80000006	0.057305	0.064152
-108.4	75	0.0958	0.80000002	0.082834	0.093532
-108.4	99	0.1263	0.141128279	0.110011	0.124753
-108.4	200	0.2545	0.283300838	0.224193	0.255337
-108.4	300	0.3969	0.420683176	0.338273	0.384966
-113.6	30	0.0409	0.8	0.034042	0.037496
-113.6	50	0.0692	0.80000005	0.05815	0.0653
-113.6	75	0.1052	0.80000003	0.088232	0.09993
-113.6	100	0.1379	0.152307647	0.11828	0.134449
-113.6	200	0.2737	0.302419416	0.238668	0.272083
-113.6	300	0.4207	0.449722813	0.361207	0.411193
-116	300	0.4379	0.464351326	0.372826	0.424425

2) Mixture 2:

Table 5-4: Properties of Components for Mixture 2

Components	Mol. Frac.	PC (psia)	TC (°F)	Acentric Factor	$MW\left(\frac{lb}{lbmol}\right)$	$\operatorname{VC}\left(\frac{ft^{3}}{lbm}\right)$
CH ₄	0.65	667.8	-116.63	0.0104	16.043	0.098797
C_2H_6	0.35	707.8	90.09	0.0979	30.07	0.078284

Table 5-5: Comparison of Liquid Compositions for Mixture 2

Т	Р	Expt. CH ₄	PR	РТ	NEW
(°F)	(psia)	Composition			
-99.8	160	0.1875	0.65	0.162817	0.168754
-99.8	250	0.31	0.314682	0.279133	0.294392
-99.8	350	0.4526	0.454972	0.411253	0.435038
-113.4	61.5	0.0681	0.65	0.060174	0.078327
-113.4	100	0.1364	0.65	0.114172	0.116256
-113.4	160	0.2299	0.65	0.203865	0.213748
-113.4	250	0.3854	0.385088	0.341888	0.361838
-116.1	60.5	0.0714	0.65	0.060088	0.074398
-116.1	100	0.1386	0.65	0.120282	0.123024
-116.1	160	0.2384	0.65	0.213107	0.22383
-116.1	250	0.3955	0.401231	0.356426	0.377306
-118.3	60	0.0752	0.65	0.062802	0.072729
-118.3	95.5	0.138	0.65	0.118348	0.120995
-118.3	140	0.2138	0.65	0.188883	0.197694
-118.3	200	0.3202	0.65	0.285938	0.302234
-118.3	300	0.4975	0.50398	0.454106	0.479956
-124.7	61.5	0.0897	0.65	0.075929	0.074856
-124.7	104	0.1707	0.65	0.148275	0.153836
-124.7	180	0.31	0.65	0.281212	0.297286
-124.7	275	0.5019	0.507691	0.455995	0.481662
-150	30.8	0.0685	0.65	0.055825	0.053941
-150	45.5	0.1087	0.65	0.091158	0.092562
-150	81	0.205	0.65	0.178561	0.187361
-150	120	0.3164	0.65	0.278649	0.294448
-175	25.8	0.109	0.65	0.083737	0.085424
-175	28.8	0.123	0.65	0.094923	0.097536
-175	40	0.164	0.65	0.137331	0.143254
-175	50	0.2186	0.65	0.176125	0.184791
-175	70	0.2953	0.65	0.256732	0.270127
-175	100	0.4382	0.65	0.386907	0.404685

3) Mixture 3:

Components	Mol. Frac.	PC (psia)	TC (°F)	Acentric Factor	$MW\left(\frac{lb}{lbmol}\right)$	$\operatorname{VC}\left(\frac{ft^3}{lbm}\right)$
C_2H_6	0.6	707.8	90.09	0.0979	30.07	0.078284
$n-C_4H_{10}$	0.4	550.7	305.65	0.201	58.124	0.070298

Table 5-6: Properties of Components for Mixture 3

Table 5-7: Comparison of Liquid Compositions for Mixture 3

T (05)	P	Expt. C ₂ H ₆	PR	РТ	NEW
(°F)	(psia)	Composition			
86	164.038	0.257	0.6	0.233597	0.228238
86	179.412	0.288	0.6	0.261704	0.257974
86	238.4425	0.397	0.6	0.367474	0.369557
86	247.2898	0.403	0.6	0.383039	0.385932
122	250.9157	0.269	0.6	0.258443	0.272823
122	320.8241	0.358	0.382643	0.353137	0.359904
122	393.4881	0.46	0.480559	0.448011	0.460442
122	410.6026	0.486	0.502761	0.469837	0.483476
122	496.6101	0.6	0.609422	0.576511	0.595453
194.45	537.2208	0.346	0.360377	0.339276	0.352125
194.45	715.7625	0.5	0.525738	0.498496	0.517939
194.45	772.4724	0.533	0.580094	0.548742	0.568921

4) Mixture 4

Table 5-8: Properties of Components for Mixture 4

Components	Mol. Frac.	PC (psia)	TC (°F)	Acentric Factor	$MW\left(\frac{lb}{lbmol}\right)$	$\operatorname{VC}\left(\frac{ft^3}{lbm}\right)$
CH ₄	0.9144	667.8	-116.63	0.0104	16.043	0.098797
i- C ₄ H ₁₀	0.0856	529.1	274.98	0.1848	58.124	0.071399

Table 5-9: Comparison of Liquid Compositions for Mixture 4

T P Expt. CH4 PR PT NEW						
	Т	Р	Expt. CH ₄	PR	РТ	NEW

(°F)	(psia)	Composition			
32.576	1208.022	0.516	0.537281	0.456184	0.545168
14.738	1134.052	0.5237	0.544009	0.459033	0.512387
-3.19	1059.648	0.5371	0.555555	0.465319	0.523181
-21.172	984.0828	0.5561	0.572368	0.475434	0.538353
-21.19	976.8309	0.5523	0.568835	0.472473	0.534996
-21.19	876.1746	0.4971	0.518425	0.429964	0.486821
-21.19	640.7779	0.3734	0.394827	0.324946	0.368122
-21.19	383.1904	0.23	0.247032	0.200819	0.227539
-57.388	828.7471	0.6151	0.628091	0.512083	0.587884
-75.568	742.3045	0.6565	0.66752	0.538803	0.622676
-93.82	653.1061	0.7133	0.721764	0.576898	0.671181

5) Mixture 5:

Table 5-10: Properties of Components for Mixture 5

Components	Mol. Frac.	PC (psia)	TC (°F)	Acentric Factor	$MW\left(\frac{lb}{lbmol}\right)$	$\operatorname{VC}\left(\frac{ft^3}{lbm}\right)$
C1	0.9354	667.8	-116.63	0.0104	16.043	0.098797
n-C4	0.0646	550.7	305.65	0.201	58.124	0.070298

Table 5-11: Comparison of Liquid Compositions for Mixture 5

Т	Р	Expt. CH ₄	PR	РТ	NEW
(°F)	(psia)	Composition			
32.486	1051.38	0.4246	0.439971	0.360817	0.409997
14.594	992.35	0.4331	0.449525	0.36491	0.417968
-3.334	932.7394	0.448	0.462878	0.371614	0.42899
-19.534	190.1448	0.1071	0.116202	0.089889	0.104113
-19.534	481.2361	0.2669	0.281142	0.220587	0.25656
-19.534	733.6022	0.395	0.409909	0.32456	0.377279
-19.534	870.9532	0.4601	0.475476	0.377612	0.438933
-19.534	956.2355	0.5008	0.514879	0.409268	0.475811
-21.334	871.6784	0.4646	0.480707	0.381335	0.443599
-39.388	812.793	0.4882	0.506283	0.396484	0.464721

Liquid Density Predictions

Propane-nButane System

a) Mixture 1 (25% Propane):

Т	Р	Exp. Liquid	PR	РТ	NEW
(°F)	(psia)	Den. $\left(\frac{Lb}{ft^3}\right)$			
-29.3998	1005.2	39.28344	41.52375	40.02048	39.96504
-29.3998	945.3577	39.25847	41.49579	39.99201	39.93509
-29.3998	874.4921	39.22913	41.46236	39.95795	39.89923
-29.3998	801.2334	39.19854	41.4274	39.92233	39.86169
-29.3998	717.9381	39.16296	41.38715	39.88132	39.81843
-29.3998	654.7886	39.13611	41.35628	39.84984	39.7852
-29.3998	579.2818	39.10427	41.31896	39.81178	39.74498
-29.3998	503.8185	39.07244	41.2812	39.77325	39.70424
-29.3998	443.5697	39.04622	41.25071	39.74214	39.6713
-29.3998	360.6225	39.01063	41.20823	39.69878	39.62535
-29.3998	284.6226	38.97692	41.16879	39.65849	39.58263
-29.3998	218.6738	38.94883	41.13415	39.6231	39.54505
-29.3998	136.8724	38.912	41.09063	39.57863	39.49777
-29.3998	67.25412	38.88141	41.05311	39.54026	39.45694
8.33	1016.861	37.99118	40.25335	38.78266	38.78475
8.33	943.2981	37.95498	40.20871	38.7374	38.7377
8.33	875.1738	37.92064	40.16683	38.69494	38.6935
8.33	799.5075	37.88256	40.11969	38.64712	38.6437
8.33	727.7572	37.84635	40.07436	38.60113	38.59575
8.33	654.3679	37.80827	40.02736	38.55342	38.54596
8.33	581.1673	37.76956	39.97981	38.50514	38.49553
8.33	500.3231	37.72774	39.9265	38.45099	38.4389
8.33	435.6942	37.6934	39.88326	38.40705	38.39291
8.33	365.6843	37.65657	39.83579	38.3588	38.34234
8.33	288.6111	37.61537	39.78274	38.30486	38.28574
8.33	215.0188	37.57604	39.73129	38.25252	38.23077
8.33	145.7487	37.53796	39.68214	38.2025	38.17816
8.33	76.01442	37.49988	39.63192	38.15137	38.12433
44.33	1016.615	36.7008	38.90294	37.4706	37.54295
44.33	945.2271	36.65835	38.84699	37.41415	37.48494
44.33	871.0837	36.6134	38.78795	37.35457	37.42365
44.33	798.1006	36.56907	38.72887	37.29493	37.36224
44.33	728.0472	36.526	38.67125	37.23674	37.30226
44.33	653.4397	36.4798	38.60885	37.1737	37.23722
44.33	582.3276	36.43485	38.54836	37.11256	37.17407
44.33	507.9666	36.38803	38.484	37.0475	37.10679
44.33	435.1575	36.34121	38.41986	36.98263	37.03963
44.33	363.1606	36.29439	38.3553	36.91731	36.97191
44.33	292.3241	36.24757	38.29064	36.85185	36.90398
44.33	217.0784	36.19763	38.22066	36.78099	36.83033

Table 5-12: Comparison of Liquid Densities for Mixture 1

44.00	4 4 4 4 9 9 9	26.4.4002	20 45456	26 74 202	00 75747
44.33	144.1823	36.14893	38.15156	36./1098	36./5/4/
44.33	76.01442	30.10211	38.08572	36.64424	30.08/91
80.33	1026.1	35.30484	37.41028	36.02536	36.18617
80.33	941.1226	35.30179	37.32349	35.93827	36.09777
80.33	860.0463	35.24061	37.23885	35.85329	36.01141
80.33	790.6166	35.18754	37.16486	35.77898	35.93582
80.33	729.6862	35.14072	37.09874	35./1256	35.86819
80.33	654.3099	35.08266	37.01535	35.62875	35.78277
80.33	579.9635	35.02336	36.9313	35.54425	35.69653
80.33	508.7788	34.9678	36.84908	35.46156	35.61203
80.33	436.927	34.90974	36.76428	35.37622	35.52473
80.33	362.2904	34.84793	36.67417	35.28551	35.43179
80.33	290.4386	34.788	36.58539	35.19608	35.34005
80.33	218.1952	34.72745	36.49401	35.104	35.24544
80.33	145.8357	34.66377	36.40025	35.00947	35.14816
116.33	1015.817	33.92899	35.71647	34.39029	34.66103
116.33	946.3149	33.8622	35.62252	34.29655	34.56716
116.33	872.244	33.79602	35.51997	34.1942	34.46456
116.33	797.854	33.72298	35.41434	34.08874	34.35873
116.33	723.6961	33.64869	35.30625	33.98078	34.25028
116.33	648.6099	33.57378	35.19383	33.86843	34.13731
116.33	581.907	33.50698	35.09129	33.76592	34.0341
116.33	510.0551	33.43519	34.97786	33.65249	33.91976
116.33	436.9415	33.35778	34.8591	33.53365	33.79981
116.33	363.0011	33.27787	34.73535	33.40975	33.67458
116.33	289.5974	33.19671	34.60863	33.2828	33.54606
116.33	214.9753	33.11244	34.47558	33.14942	33.41082
116.33	144.6609	33.0319	34.34603	33.01946	33.27881
152.33	1014.077	32.37953	33.8094	32.55656	32.9644
152.33	939.5272	32.28901	33.67301	32.42134	32.83104
152.33	864.2089	32.19662	33.53068	32.28019	32.69172
152.33	794.6922	32.10859	33.39499	32.14557	32.55875
152.33	722.5503	32.01558	33.24947	32.00114	32.41597
152.33	654.8176	31.92505	33.10816	31.86083	32.27714
152.33	577.9039	31.82018	32.94177	31.69553	32.11343
152.33	505.8925	31.71904	32.7798	31.53452	31.9538
152.33	434.7659	31.61604	32.61342	31.36904	31.78955
152.33	355.9668	31.49742	32.42098	31.17748	31.59919
152.33	288.7562	31.39317	32.24944	31.0066	31.42913
152.33	215.4249	31.27518	32.0537	30.81142	31.23462
188.33	1010.465	30.67088	31.61819	30.45773	31.03772
188.33	938.7149	30.55476	31.43489	30.2773	30.86303
188.33	870.2135	30.43864	31.25256	30.09775	30.68918
188.33	796.5922	30.31067	31.04782	29.89607	30.49387
188.33	727.8152	30.18519	30.84752	29.69866	30.30269
188.33	651.9023	30.04223	30.61515	29.46954	30.08077

188.33	580.8192	29.90176	30.38555	29.243	29.86136
188.33	504.6597	29.74382	30.12496	28.98569	29.61214
188.33	439.5377	29.60148	29.88852	28.75202	29.38583
188.33	360.1584	29.41857	29.58069	28.44746	29.09091
188.33	285.4783	29.23503	29.26804	28.13766	28.79101
224.33	1008.42	28.71626	29.06149	28.01884	28.81852
224.33	943.0951	28.55831	28.81759	27.78062	28.59331
224.33	869.5753	28.3729	28.52633	27.49608	28.32472
224.33	797.3464	28.17875	28.22033	27.19703	28.04298
224.33	725.0015	27.97274	27.89077	26.87478	27.74011
224.33	652.8305	27.75049	27.53465	26.52635	27.41363
224.33	579.9635	27.51015	27.14151	26.14135	27.05428
224.33	508.8948	27.24857	26.71768	25.72577	26.66839
224.33	436.0422	26.95391	26.22993	25.24667	26.22668
224.33	363.7263	26.6218	25.67391	24.69897	25.72705

b) Mixture 2 (50% Propane):

Table 5-13: Com	parison of L	iquid Densities	for Mixture 2
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Т	Р	Exp. Liquid	PR	РТ	NEW
(°F)	(psia)	Den. $\left(\frac{Lb}{ft^3}\right)$			
-27.67	1011.364	38.31456072	40.67292	39.05273	39.06473
-27.67	943.1386	38.283971	40.63758	39.01675	39.02745
-27.67	870.2425	38.25213272	40.59937	38.97784	38.98711
-27.67	793.6624	38.21779732	40.55873	38.93644	38.94415
-27.67	721.1434	38.18471048	40.51977	38.89673	38.90291
-27.67	651.6702	38.15224792	40.48198	38.85821	38.86287
-27.67	573.2627	38.11728824	40.43879	38.81416	38.81705
-27.67	507.8361	38.08669852	40.4023	38.77693	38.77829
-27.67	428.7323	38.05049028	40.35761	38.73132	38.73076
-27.67	361.3767	38.018652	40.31906	38.69196	38.6897
-27.67	284.6226	37.98244376	40.27456	38.6465	38.64224
-27.67	214.5112	37.94873264	40.23336	38.6044	38.59825
-27.67	144.5594	37.91751864	40.19172	38.56184	38.55371
-27.67	69.40068	37.88006184	40.14637	38.51546	38.50514
8.33	1013.322	37.05039372	39.3579	37.77619	37.85531
8.33	944.502	37.0135612	39.31166	37.72935	37.80732
8.33	872.4616	36.97548012	39.26257	37.67961	37.75634
8.33	799.696	36.93615048	39.21228	37.62863	37.70403
8.33	727.7137	36.89682084	39.16179	37.57744	37.65147
8.33	656.4275	36.85686692	39.11106	37.52599	37.59857
8.33	580.8047	36.81566444	39.05642	37.47054	37.54153

8.33	508.5032	36.7763348	39.00336	37.41668	37.48606
8.33	437.2751	36.7357566	38.95028	37.36277	37.4305
8.33	365.0897	36.69392984	38.89565	37.30727	37.37322
8.33	290.7432	36.6514788	38.83847	37.24915	37.31319
8.33	218.4852	36.60902776	38.78198	37.19171	37.25378
8.33	145.1975	36.56657672	38.72374	37.13245	37.19243
8.33	71.90984	36.52162856	38.6645	37.07216	37.12993
44.33	1011.248	35.69133616	37.89193	36.35795	36.5214
44.33	943.8203	35.64701228	37.83265	36.29824	36.46096
44.33	871.5914	35.599567	37.76809	36.23319	36.39506
44.33	798.6227	35.55087316	37.70171	36.16629	36.32722
44.33	725.0015	35.50217932	37.63352	36.09753	36.25743
44.33	656.2389	35.45473404	37.56869	36.03214	36.19099
44.33	582.5741	35.40354308	37.49796	35.96076	36.11841
44.33	508.2422	35.35172784	37.4252	35.88732	36.04364
44.33	436.0422	35.30053688	37.35315	35.81454	35.96947
44.33	362.8706	35.24809736	37.27868	35.73929	35.89269
44.33	290.6707	35.19628212	37.20372	35.6635	35.81527
44.33	218.0211	35.14259404	37.12673	35.58563	35.73563
44.33	144.4724	35.0876574	37.04714	35.50507	35.65313
44.33	77.33426	35.03709072	36.97297	35.42996	35.5761
80.33	1009.45	34.25299504	36.24931	34.77481	35.04331
80.33	937.8302	34.19306416	36.1658	34.69123	34.95976
80.33	863.4112	34.13126044	36.07714	34.60245	34.87095
80.33	798.9853	34.07694808	35.99875	34.52393	34.79234
80.33	718.4022	34.00827728	35.89848	34.42345	34.69165
80.33	654.8031	33.95334064	35.81752	34.34228	34.61023
80.33	582.4581	33.8890398	35.72337	34.24786	34.51542
80.33	508.4452	33.82286612	35.62468	34.14882	34.41588
80.33	427.311	33.7485768	35.51359	34.03728	34.30364
80.33	356.9965	33.68365168	35.41472	33.93796	34.20355
80.33	282.9546	33.61373232	35.30785	33.83053	34.09518
80.33	223.5616	33.55629856	35.21998	33.74215	34.0059
80.33	143.5876	33.47763928	35.09846	33.61985	33.88219
116.33	1032.148	32.74473456	34.42943	33.02924	33.42766
116.33	939.7737	32.64859544	34.28458	32.88525	33.28582
116.33	868.7486	32.57305756	34.16971	32.77102	33.1732
116.33	796.3456	32.49439828	34.04928	32.65122	33.055
116.33	723.5366	32.41324188	33.92457	32.5271	32.93244
116.33	652.2939	32.3314612	33.79883	32.40188	32.80871
116.33	582.6321	32.2503048	33.67208	32.27561	32.68382
116.33	508.9673	32.16228132	33.53368	32.13765	32.54725
116.33	435.085	32.07176072	33.39005	31.99438	32.40528
116.33	364.9881	31.98311296	33.24896	31.85355	32.26557
116.33	291.5699	31.88759812	33.09577	31.70052	32.11359
116.33	217.0204	31.7883376	32.93406	31.53884	31.95281

152.33	1017.355	31.06791848	32.28826	30.98211	31.54346
152.33	947.7363	30.96990652	32.1364	30.8323	31.39837
152.33	874.7532	30.8644032	31.97132	30.66939	31.24055
152.33	799.3914	30.7520328	31.79402	30.49435	31.07093
152.33	726.9595	30.6396624	31.61647	30.31896	30.90095
152.33	652.2069	30.52042492	31.42519	30.12991	30.71767
152.33	582.5016	30.40493312	31.23873	29.9455	30.53885
152.33	509.8811	30.28007712	31.0353	29.74416	30.34357
152.33	437.0865	30.15022688	30.82093	29.53182	30.13757
152.33	363.8713	30.012261	30.5935	29.3063	29.91873
152.33	292.0485	29.871798	30.35737	29.07186	29.69121
188.33	1011.306	29.23877808	29.80419	28.61833	29.38656
188.33	947.1852	29.09893936	29.60071	28.41929	29.19791
188.33	872.8097	28.93662656	29.35244	28.17634	28.9679
188.33	801.335	28.7761866	29.09992	27.92913	28.73413
188.33	728.816	28.60575816	28.82783	27.6626	28.48248
188.33	656.9061	28.42783836	28.53983	27.38029	28.21642
188.33	583.6039	28.23368728	28.22452	27.07093	27.92553
188.33	512.0857	28.030172	27.89182	26.74413	27.61913
188.33	437.2025	27.8023098	27.51133	26.3698	27.26951
188.33	366.6561	27.56695624	27.11561	25.97968	26.90697
224.33	1002.039	26.9251964	26.80692	25.77875	26.81956
224.33	939.9333	26.72480252	26.49702	25.47845	26.54349
224.33	864.7891	26.46260492	26.09055	25.08445	26.18299
224.33	796.9113	26.20290444	25.6874	24.69351	25.82751
224.33	727.9457	25.91261424	25.23397	24.25354	25.43056
224.33	649.2626	25.53492484	24.64609	23.68247	24.92137
224.33	580.3115	25.15036836	24.04562	23.09812	24.4092
224.33	510.9834	24.68777688	23.32189	22.39165	23.80589

c) Mixture 3 (75% Propane):

Т	Р	Exp. Liquid	PR	РТ	NEW
(°F)	(psia)	Den. $\left(\frac{Lb}{ft^3}\right)$			
-27.67	1021.88	37.28824	39.7737	38.02791	38.11288
-27.67	948.4905	37.25516	39.73168	37.98518	38.06928
-27.67	876.7257	37.21957	39.69007	37.94284	38.02605
-27.67	800.8418	37.18461	39.6455	37.89746	37.97968
-27.67	717.4885	37.14404	39.59583	37.84688	37.92795
-27.67	656.558	37.11407	39.55904	37.80941	37.8896
-27.67	584.6772	37.08036	39.51511	37.76464	37.84375

Table 5-14: Comparison of Liquid Densities for Mixture 3

-27.67	510.4467	37.04415	39.46913	37.71775	37.79569
-27.67	438.4064	37.00732	39.42388	37.6716	37.74834
-27.67	365.4232	36.97049	39.3774	37.62417	37.69964
-27.67	293.5859	36.93428	39.331	37.5768	37.65096
-27.67	218.3402	36.89682	39.28169	37.52645	37.59916
-27.67	145.9517	36.85999	39.23356	37.47726	37.54852
-27.67	74.68007	36.82191	39.18548	37.42811	37.49785
8.33	1006.999	35.93356	38.33559	36.6382	36.80433
8.33	942.573	35.8986	38.28705	36.58913	36.75483
8.33	877.3639	35.85927	38.23724	36.53875	36.70398
8.33	798.8838	35.81432	38.17636	36.47716	36.64176
8.33	728.4969	35.77187	38.12087	36.42099	36.58498
8.33	651.8733	35.72567	38.05947	36.35882	36.52207
8.33	581.0512	35.68509	38.00176	36.30036	36.46287
8.33	506.9803	35.6389	37.9404	36.23818	36.39983
8.33	437.3766	35.59582	37.88177	36.17873	36.33952
8.33	363.8133	35.54962	37.81874	36.11479	36.2746
8.33	290.5981	35.50343	37.7549	36.04999	36.20873
8.33	218.1662	35.45723	37.69059	35.98469	36.14229
8.33	144.5884	35.40979	37.62407	35.91711	36.07345
8.33	71.0106	35.36172	37.5563	35.84821	36.00319
44.33	1017.035	34.52518	36.74417	35.10761	35.37494
44.33	947.1562	34.47399	36.67437	35.03752	35.3051
44.33	876.5662	34.42217	36.60255	34.96538	35.23317
44.33	798.9853	34.36661	36.52204	34.88447	35.15244
44.33	722.9999	34.3098	36.44151	34.80351	35.07159
44.33	653.8603	34.25737	36.36675	34.72831	34.99643
44.33	582.3276	34.2018	36.28783	34.6489	34.917
44.33	509.62	34.14562	36.20593	34.56644	34.83444
44.33	437.7827	34.08943	36.12325	34.48316	34.75098
44.33	363.9874	34.03013	36.03642	34.39564	34.66319
44.33	291.947	33.97207	35.94971	34.3082	34.57538
44.33	218.4707	33.91027	35.8592	34.21687	34.48356
80.33	1017.471	33.02941	34.93889	33.37871	33.77134
80.33	938.918	32.95699	34.83272	33.27291	33.66734
80.33	862.9761	32.88582	34.72739	33.16789	33.56405
80.33	794.0395	32.8209	34.62936	33.07011	33.46782
80.33	727.6121	32.75535	34.53262	32.97357	33.37275
80.33	654.6725	32.68418	34.42367	32.8648	33.26556
80.33	582.6902	32.61176	34.31321	32.75446	33.15674
80.33	509.3009	32.53623	34.1974	32.6387	33.04249
80.33	436.4483	32.45944	34.07904	32.52034	32.92556
80.33	363.7118	32.38078	33.9573	32.3985	32.80509
80.33	286.0294	32.29588	33.82306	32.26405	32.67202
80.33	218.2387	32.21847	33.7021	32.1428	32.5519
80.33	145.3716	32.13481	33.56784	32.00812	32.41832

116.33	1018.268	31.38755	32.87681	31.41338	31.9618
116.33	946.0684	31.29953	32.74159	31.27974	31.83253
116.33	870.4746	31.20464	32.59527	31.13507	31.69256
116.33	800.6678	31.11599	32.45554	30.99685	31.5588
116.33	727.9602	31.01985	32.30494	30.84781	31.41452
116.33	654.513	30.91996	32.14716	30.69156	31.26323
116.33	582.139	30.81946	31.98565	30.53152	31.10822
116.33	501.759	30.70334	31.79863	30.34606	30.92853
116.33	436.7964	30.60595	31.64106	30.18968	30.77697
116.33	362.1744	30.49046	31.45226	30.00214	30.59515
116.33	293.0783	30.37996	31.26929	29.82019	30.41871
116.33	218.9784	30.25635	31.06337	29.61518	30.21984
152.33	1008.478	29.51908	30.45346	29.11451	29.86075
152.33	942.0218	29.40421	30.27277	28.93757	29.69303
152.33	868.8066	29.27311	30.06471	28.73375	29.49997
152.33	800.1746	29.14576	29.86023	28.53333	29.31029
152.33	723.9137	28.99656	29.62103	28.29876	29.08852
152.33	648.1893	28.84236	29.36947	28.05187	28.8554
152.33	583.2848	28.70502	29.14119	27.82765	28.64401
152.33	513.8116	28.54645	28.88209	27.5729	28.40423
152.33	437.7247	28.36541	28.57809	27.27362	28.12316
152.33	366.54	28.1825	28.27112	26.97091	27.83969
152.33	289.3363	27.97149	27.90844	26.61249	27.5053
188.33	1009.493	27.23546	27.58795	26.41125	27.41358
188.33	941.8768	27.05005	27.30356	26.13565	27.15956
188.33	865.2677	26.82843	26.95745	25.80012	26.85145
188.33	794.4311	26.60244	26.61119	25.46426	26.5445
188.33	719.3015	26.35086	26.20881	25.07369	26.18967
188.33	652.5985	26.0999	25.81574	24.69178	25.84531
188.33	585.1123	25.8196	25.37411	24.26209	25.46155
188.33	508.2277	25.45439	24.7996	23.70184	24.96839
188.33	432.3728	25.02801	24.12613	23.04243	24.40151
224.33	1000.414	24.39374	23.96433	23.00874	24.36652
224.33	943.3126	24.11718	23.54429	22.60647	24.01134
224.33	868.2265	23.69018	22.92009	22.00869	23.49034
224.33	796.4327	23.197	22.21797	21.33617	22.91615
224.33	728.8595	22.61766	21.41446	20.56606	22.27871
224.33	658.2361	21.79174	20.31737	19.51281	21.45493

Propane-iButane System

d) Mixture 4 (25% Propane):

Т	Р	Exp. Liquid	PR	PT	NEW
(°F)	(psia)	Den. $\left(\frac{Lb}{ft^3}\right)$			
-27.67	1013.482	38.42319	40.93156	39.30113	39.3135
-27.67	948.7661	38.39322	40.89622	39.26518	39.27626
-27.67	875.0143	38.35888	40.85546	39.22369	39.23327
-27.67	802.5533	38.32517	40.8149	39.1824	39.19044
-27.67	729.5992	38.29146	40.77354	39.14027	39.1467
-27.67	655.4412	38.2565	40.73093	39.09686	39.1016
-27.67	584.1696	38.22342	40.68944	39.05457	39.05762
-27.67	513.1299	38.18908	40.64755	39.01185	39.01315
-27.67	439.2331	38.15287	40.60338	38.96679	38.96621
-27.67	367.3813	38.11916	40.55983	38.92236	38.91987
-27.67	290.7287	38.08233	40.51272	38.87426	38.86966
-27.67	218.4127	38.04674	40.46763	38.82821	38.82154
-27.67	145.0525	38.01054	40.42124	38.7808	38.77194
-27.67	68.90755	37.97308	40.37237	38.73084	38.71961
-27.67	57.75413	37.96809	40.36514	38.72346	38.71187
8.33	1016.165	37.15402	39.62172	38.03031	38.11027
8.33	939.0195	37.11157	39.56713	37.97505	38.05369
8.33	877.7555	37.07973	39.52316	37.93052	38.00807
8.33	805.0769	37.03853	39.47026	37.87694	37.95313
8.33	721.7526	36.99046	39.4086	37.81446	37.889
8.33	649.2336	36.94926	39.35404	37.75915	37.83216
8.33	578.223	36.90743	39.29976	37.70411	37.77555
8.33	510.1277	36.8731	39.24691	37.65049	37.72036
8.33	434.3888	36.83127	39.18718	37.58986	37.65787
8.33	363.2912	36.79007	39.13015	37.53196	37.59813
8.33	290.7142	36.74699	39.07096	37.47183	37.53603
8.33	214.8883	36.70017	39.00804	37.40787	37.4699
8.33	143.6166	36.65772	38.94784	37.34666	37.40652
8.33	75.5793	36.61714	38.88938	37.28719	37.34488
44.33	1010.277	35.82243	38.1597	36.61655	36.78126
44.33	1010.625	35.81994	38.16002	36.61687	36.78159
44.33	947.4607	35.77936	38.10173	36.55821	36.72225
44.33	868.6906	35.7263	38.02775	36.48372	36.64683
44.33	795.5334	35.67573	37.95771	36.41318	36.57534
44.33	722.1587	35.62516	37.88612	36.34104	36.50217
44.33	655.1076	35.57834	37.81948	36.27387	36.43396
44.33	579.0497	35.5234	37.74243	36.19617	36.35499
44.33	506.0086	35.47097	37.66691	36.11999	36.27747
44.33	434.9835	35.41915	37.592	36.04437	36.20044

Table 5-15: Comparison of Liquid Densities for Mixture 4

44.33	364.5675	35.36609	37.51621	35.96784	36.12239
44.33	291.2073	35.31053	37.43559	35.88639	36.03922
44.33	212.1616	35.25122	37.34674	35.79657	35.94738
44.33	145.7197	35.20065	37.27039	35.71935	35.8683
44.33	72.56251	35.14322	37.18448	35.6324	35.77915
80.33	1009.029	34.41219	36.52839	35.04528	35.31539
80.33	945.0676	34.361	36.45043	34.96732	35.23754
80.33	873.8685	34.30231	36.36184	34.87868	35.14895
80.33	799.4204	34.23427	36.26705	34.78382	35.05405
80.33	727.5686	34.17184	36.17339	34.69004	34.96015
80.33	650.2634	34.1013	36.0701	34.58658	34.85646
80.33	581.8925	34.04074	35.97647	34.49275	34.76231
80.33	506.0086	33.97082	35.86991	34.38591	34.65499
80.33	437.2316	33.90652	35.7708	34.28649	34.55501
80.33	364.1324	33.8366	35.66268	34.17796	34.44575
80.33	291.0333	33.76543	35.55153	34.06633	34.3332
80.33	218.0936	33.69302	35.43741	33.95165	34.21742
80.33	145.2701	33.61935	35.32008	33.83366	34.09812
116.33	1010.059	32.87458	34.69247	33.28431	33.68487
116.33	939.8172	32.79905	34.57733	33.16996	33.57231
116.33	870.4891	32.72413	34.46053	33.05392	33.45801
116.33	800.2182	32.64547	34.33875	32.93289	33.3387
116.33	723.7541	32.55807	34.20211	32.79703	33.20467
116.33	655.8763	32.47879	34.07696	32.67254	33.08175
116.33	583.4734	32.39202	33.93923	32.53546	32.94629
116.33	508.8078	32.30025	33.79225	32.3891	32.80152
116.33	436.2163	32.20973	33.64418	32.24156	32.65543
116.33	365.8148	32.11858	33.49532	32.09313	32.5083
116.33	293.4554	32.02182	33.33649	31.93464	32.35102
116.33	217.557	31.91819	33.16297	31.76134	32.17881
116.33	148.0113	31.81893	32.99718	31.59561	32.01389
152.33	1009.9	31.26082	32.59822	31.28427	31.84683
152.33	946.7355	31.17092	32.45559	31.14371	31.71079
152.33	873.4623	31.06355	32.28413	30.97469	31.54716
152.33	800.9433	30.9543	32.10758	30.80058	31.37857
152.33	720.0992	30.82757	31.90195	30.5977	31.18206
152.33	653.1061	30.71957	31.72382	30.42186	31.01171
152.33	582.4291	30.60283	31.52754	30.22799	30.82384
152.33	512.4048	30.4811	31.32373	30.02654	30.6286
152.33	439.059	30.34812	31.09914	29.8044	30.41324
152.33	366.0469	30.21016	30.8629	29.57049	30.18644
152.33	294.1371	30.06532	30.6162	29.32596	29.94928
152.33	219.6165	29.90988	30.34369	29.05547	29.68692
188.33	1008.942	29.43355	30.17192	28.97761	29.74375
188.33	934.7119	29.28435	29.93035	28.74158	29.52006
188.33	867.7479	29.14514	29.70075	28.51719	29.30762

188.33	799.9136	28.99468	29.45549	28.2774	29.08085
188.33	722.8549	28.81552	29.15927	27.98767	28.80725
188.33	655.6008	28.65133	28.88318	27.71746	28.55253
188.33	583.3863	28.46405	28.56546	27.40629	28.25982
188.33	510.5048	28.26178	28.21843	27.06609	27.94069
188.33	438.3483	28.0464	27.84328	26.69785	27.59651
188.33	367.3668	27.81542	27.43629	26.29766	27.22423
188.33	293.4554	27.55197	26.96125	25.82937	26.7915
224.33	1024.084	27.33909	27.37726	26.33464	27.35607
224.33	938.6569	27.07752	26.95711	25.92805	26.9812
224.33	866.1814	26.83717	26.56736	25.55083	26.63505
224.33	796.0411	26.58434	26.15445	25.15112	26.27028
224.33	728.0327	26.31465	25.71248	24.72314	25.88242
224.33	645.9847	25.95132	25.10843	24.13782	25.35746
224.33	575.7283	25.59486	24.50713	23.5545	24.84222
224.33	506.9803	25.18533	23.80863	22.87555	24.25557
224.33	431.8071	24.63409	22.83999	21.93001	23.47154

e) Mixture 5 (50% Propane):

Tab	ole 5	-16:	Com	parison	of L	liquid	Densities	for	Mixture	5
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Т	Р	Exp. Liquid	PR	РТ	NEW
(°F)	(psia)	Den. $\left(\frac{Lb}{ft^3}\right)$			
-9.67	988.9851	37.00982	39.58261	37.89737	37.99073
-9.67	944.1249	36.98734	39.55309	37.86744	37.96019
-9.67	869.7204	36.95113	39.50357	37.8172	37.9089
-9.67	795.4754	36.91305	39.45343	37.7663	37.8569
-9.67	731.3106	36.87872	39.4095	37.72169	37.81128
-9.67	657.3847	36.84001	39.35817	37.66955	37.75793
-9.67	579.5573	36.79943	39.30328	37.61378	37.7008
-9.67	506.5452	36.7601	39.25098	37.5606	37.64628
-9.67	436.985	36.72202	39.20039	37.50915	37.59348
-9.67	361.1156	36.67957	39.14435	37.45213	37.53491
-9.67	294.0645	36.64211	39.09405	37.40093	37.48226
-9.67	219.1669	36.59904	39.03698	37.3428	37.42244
-9.67	147.1265	36.55846	38.98117	37.28594	37.36384
-9.67	74.76709	36.51663	38.92419	37.22785	37.30392
44.33	1006.404	35.02086	37.3337	35.72861	35.95769
44.33	946.5035	34.97903	37.2744	35.66902	35.89799
44.33	874.144	34.92597	37.20151	35.59576	35.82453
44.33	795.2143	34.86854	37.12038	35.51419	35.74267
44.33	727.2205	34.81734	37.04909	35.44247	35.67064

44.33	655.3687	34.76428	36.97228	35.36517	35.59293
44.33	578.9917	34.70685	36.88891	35.28123	35.50846
44.33	510.4757	34.65378	36.81254	35.20429	35.43097
44.33	438.4354	34.59822	36.73055	35.12167	35.34765
44.33	361.0141	34.53642	36.64043	35.03079	35.25593
44.33	291.2073	34.48148	36.55777	34.94739	35.17164
44.33	219.5875	34.42217	36.4701	34.85888	35.08209
44.33	138.1922	34.35413	36.36849	34.75624	34.9781
44.33	67.76175	34.29545	36.27833	34.66509	34.88563
80.33	1013.047	33.56691	35.60124	34.06646	34.41335
80.33	945.2562	33.50636	35.5115	33.97694	34.32485
80.33	866.7036	33.43394	35.40499	33.87064	34.21968
80.33	791.7044	33.36527	35.30063	33.76644	34.1165
80.33	719.2289	33.29535	35.19717	33.66309	34.01408
80.33	653.1931	33.23292	35.10054	33.56652	33.9183
80.33	579.4993	33.15926	34.98991	33.45591	33.80851
80.33	505.7185	33.08372	34.87604	33.34198	33.69531
80.33	441.4086	33.02004	34.77408	33.23993	33.59381
80.33	367.0042	32.94326	34.6528	33.11846	33.47288
80.33	293.8615	32.86522	34.52989	32.99526	33.3501
80.33	220.3997	32.78531	34.40251	32.8675	33.2226
80.33	144.5739	32.70103	34.2666	32.73108	33.08628
116.33	1005.157	31.95377	33.61216	32.16598	32.6592
116.33	946.3584	31.88448	33.50552	32.06042	32.55641
116.33	869.8364	31.79083	33.36263	31.91894	32.41858
116.33	804.3517	31.70718	33.23644	31.79392	32.29675
116.33	718.5473	31.59918	33.0652	31.62421	32.13126
116.33	657.0802	31.5199	32.93813	31.4982	32.00833
116.33	585.5764	31.42501	32.78532	31.34659	31.86033
116.33	514.8849	31.32887	32.62856	31.19096	31.70833
116.33	438.5804	31.22149	32.45246	31.016	31.53735
116.33	360.9561	31.1085	32.26528	30.82987	31.35531
116.33	296.0951	31.01111	32.10206	30.66742	31.19631
116.33	211.9295	30.87751	31.88	30.44617	30.97959
152.33	1021.865	30.13774	31.37402	30.03958	30.71387
152.33	944.6325	30.01226	31.17651	29.84579	30.52864
152.33	867.6173	29.88304	30.97039	29.64345	30.33531
152.33	790.5296	29.74694	30.7539	29.43085	30.13224
152.33	723.8266	29.62209	30.55749	29.23786	29.94799
152.33	653.8893	29.48724	30.34148	29.02549	29.74535
152.33	581.965	29.34366	30.10732	28.79513	29.52568
152.33	505.559	29.18322	29.84347	28.53532	29.27813
152.33	432.4453	29.01966	29.57429	28.26997	29.02558
152.33	356.2133	28.84111	29.2734	27.97297	28,74328
152.33	290.4531	28.67755	28,99441	27.69713	28.48154
188.33	1011.93	28.2262	28.67616	27.48682	28.39458

940.9195	28.05577	28.40411	27.22246	28.14798
869.1112	27.8766	28.11113	26.93767	27.88296
788.5136	27.65935	27.75731	26.59361	27.56376
725.19	27.47893	27.45755	26.30195	27.29414
651.0176	27.24857	27.07719	25.93162	26.95327
578.5276	27.0051	26.66834	25.53313	26.58866
507.0964	26.74041	26.22016	25.09565	26.19149
434.5774	26.44076	25.70486	24.59156	25.73883
361.9278	26.09678	25.10428	24.00194	25.2181
998.7027	25.75654	25.36427	24.36796	25.58761
931.5066	25.48124	24.93936	23.95955	25.21987
861.9753	25.16847	24.45177	23.4909	24.80144
798.6808	24.85009	23.95366	23.01209	24.37858
727.5106	24.43619	23.31033	22.39349	23.84075
652.9466	23.91554	22.49825	21.61199	23.17885
587.0993	23.33309	21.58803	20.73438	22.4675
513.536	22.39043	20.11486	19.30536	21.42238
	940.9195 869.1112 788.5136 725.19 651.0176 578.5276 507.0964 434.5774 361.9278 998.7027 931.5066 861.9753 798.6808 727.5106 652.9466 587.0993 513.536	940.919528.05577869.111227.8766788.513627.65935725.1927.47893651.017627.24857578.527627.0051507.096426.74041434.577426.44076361.927826.09678998.702725.75654931.506625.48124861.975325.16847798.680824.85009727.510623.91554587.099323.33309513.53622.39043	940.919528.0557728.40411869.111227.876628.11113788.513627.6593527.75731725.1927.4789327.45755651.017627.2485727.07719578.527627.005126.66834507.096426.7404126.22016434.577426.4407625.70486361.927826.0967825.10428998.702725.7565425.36427931.506625.4812424.93936861.975325.1684724.45177798.680824.8500923.95366727.510623.9155422.49825587.099323.3330921.58803513.53622.3904320.11486	940.919528.0557728.4041127.22246869.111227.876628.1111326.93767788.513627.6593527.7573126.59361725.1927.4789327.4575526.30195651.017627.2485727.0771925.93162578.527627.005126.6683425.53313507.096426.7404126.2201625.09565434.577426.4407625.7048624.59156361.927826.0967825.1042824.00194998.702725.7565425.3642724.36796931.506625.4812424.9393623.95955861.975325.1684724.4517723.4909727.510624.4361923.3103322.39349652.946623.9155422.4982521.61199587.099323.3330921.5880320.73438513.53622.3904320.1148619.30536

f) Mixture 6 (75% Propane):

Т	Р	Exp. Liquid	PR	РТ	NEW
(°F)	(psia)	Den. $\left(\frac{Lb}{ft^3}\right)$			
-27.67	1010.842	36.96362	39.53941	37.75038	37.85927
-27.67	948.6936	36.93178	39.50172	37.71207	37.82036
-27.67	871.5333	36.8937	39.45432	37.66388	37.77138
-27.67	797.1869	36.85562	39.408	37.61678	37.72346
-27.67	724.7259	36.81941	39.36224	37.57021	37.67606
-27.67	654.8901	36.7832	39.31753	37.5247	37.62971
-27.67	582.023	36.74699	39.27024	37.47654	37.58061
-27.67	514.2177	36.71141	39.22562	37.43109	37.53424
-27.67	436.0422	36.67021	39.17344	37.3779	37.47993
-27.67	367.2072	36.63525	39.12681	37.33035	37.43134
-27.67	290.1775	36.59467	39.07385	37.27632	37.37607
-27.67	213.699	36.55347	39.02043	37.22179	37.32024
-27.67	148.5334	36.51913	38.97423	37.17462	37.2719
-27.67	76.73961	36.4798	38.92259	37.12186	37.21778
8.33	36.53507	34.99526	37.19917	35.4519	35.63896
8.33	1004.084	35.61954	38.06402	36.32777	36.52307
8.33	944.7195	35.5902	38.01649	36.27977	36.47489
8.33	803.0899	35.50405	37.90052	36.16261	36.35714
8.33	720.6358	35.46347	37.83125	36.09259	36.28669
8.33	656.616	35.42789	37.77655	36.03726	36.23099

8.33	585.939	35.38481	37.71518	35.97517	36.16842
8.33	506.7773	35.33425	37.6452	35.90433	36.09697
8.33	440.2628	35.29117	37.58534	35.8437	36.03577
8.33	363.8713	35.24185	37.51535	35.77278	35.96411
8.33	294.5142	35.19691	37.45061	35.70715	35.89773
8.33	221.0524	35.14946	37.38076	35.6363	35.82601
8.33	148.8235	35.09827	37.31076	35.56525	35.75399
8.33	76.43503	35.04958	37.23921	35.49259	35.68028
44.33	69.42969	33.42083	35.2688	33.59248	33.89906
44.33	1005.273	34.18932	36.41393	34.744	35.04667
44.33	941.0356	34.14125	36.34524	34.67515	34.97841
44.33	872.099	34.08881	36.27017	34.59986	34.90373
44.33	802.1907	34.03512	36.19252	34.52197	34.8264
44.33	720.1427	33.97144	36.09938	34.42848	34.73353
44.33	659.1107	33.92213	36.02861	34.35743	34.66288
44.33	583.9665	33.86282	35.93968	34.26809	34.57399
44.33	510.8673	33.80289	35.85117	34.17913	34.48539
44.33	437.9277	33.74233	35.76079	34.08824	34.39479
44.33	364.0889	33.67991	35.6671	33.99396	34.30073
44.33	294.3691	33.61998	35.57651	33.90274	34.20963
44.33	220.0517	33.55567	35.47754	33.80302	34.10994
44.33	143.6746	33.48763	35.37311	33.69773	34.00454
80.33	119.5548	31.67285	33.01711	31.43427	31.89825
80.33	1008.681	32.64922	34.54864	32.96235	33.3979
80.33	938.831	32.58242	34.44703	32.86131	33.29913
80.33	864.557	32.50876	34.33617	32.75103	33.19127
80.33	791.7479	32.43572	34.22452	32.63992	33.08253
80.33	724.3923	32.36642	34.11847	32.53432	32.97914
80.33	647.8992	32.28589	33.99461	32.41093	32.85825
80.33	579.6734	32.21098	33.88088	32.29758	32.74711
80.33	510.4177	32.13481	33.7621	32.17911	32.63089
80.33	439.2766	32.05491	33.63637	32.05363	32.50771
80.33	361.6087	31.96563	33.49448	31.91193	32.36848
80.33	294.1806	31.88635	33.36712	31.78464	32.24331
80.33	222.4883	31.79957	33.22712	31.64459	32.10548
116.33	192.2189	29.68576	30.33436	28.87681	29.55724
116.33	1021.749	30.96616	32.431	30.95065	31.55076
116.33	942.0218	30.86315	32.26911	30.79112	31.3974
116.33	870.7501	30.76764	32.11906	30.64318	31.25518
116.33	801.654	30.67275	31.96841	30.49459	31.11232
116.33	720.5488	30.55851	31.78452	30.31312	30.93784
116.33	653.7878	30.45987	31.62693	30.15751	30.78823
116.33	579.3108	30.3475	31.44389	29.97664	30.61433
116.33	508.5612	30.237	31.26226	29.79704	30.44165
116.33	438.827	30.12338	31.07511	29.61182	30.26359
116.33	365.1622	29.99915	30.8677	29.40634	30.06606

116.33	293.1363	29.87242	30.65413	29.19451	29.86248
116.33	218.3257	29.73508	30.41958	28.96155	29.63864
152.33	1027.029	29.10706	29.95364	28.60929	29.41705
152.33	943.6317	28.95348	29.70628	28.36796	29.18995
152.33	864.4845	28.79991	29.45799	28.12561	28.96219
152.33	799.3189	28.66819	29.24241	27.91509	28.76464
152.33	719.8381	28.49901	28.96401	27.64306	28.50981
152.33	652.4679	28.34793	28.71289	27.39751	28.28026
152.33	575.6558	28.16564	28.40694	27.09808	28.00104
152.33	509.0979	27.99646	28.12211	26.819	27.74161
152.33	439.088	27.80481	27.79909	26.50206	27.44806
152.33	361.1156	27.57694	27.40506	26.11469	27.09113
188.33	1005.418	26.90959	26.87779	25.71525	26.80132
188.33	944.1684	26.74478	26.58891	25.43656	26.54693
188.33	860.3654	26.49007	26.16011	25.02278	26.17117
188.33	791.3708	26.2491	25.77182	24.64793	25.83312
188.33	714.1236	25.93384	25.28933	24.18185	25.41647
188.33	650.4664	25.6629	24.84319	23.75046	25.0352
188.33	576.0039	25.30519	24.24619	23.1723	24.53234
188.33	574.3505	25.29645	24.23179	23.15834	24.52033
188.33	510.2727	24.94186	23.62383	22.56796	24.01942

nButane-iButane System

g) Mixture 7 (25% nButane)

Т	Р	Exp. Liquid	PR	РТ	NEW
(°F)	(psia)	Den. $\left(\frac{Lb}{ft^3}\right)$			
-27.67	1011.93	39.23974	41.70159	40.18291	40.13421
-27.67	945.2707	39.21103	41.66847	40.1492	40.09882
-27.67	880.0181	3.91049	41.6357	40.11584	40.06377
-27.67	800.4937	39.14922	41.59528	40.07469	40.02048
-27.67	728.8014	39.11614	41.55838	40.0371	39.98091
-27.67	650.5825	39.0818	41.51761	39.99555	39.93714
-27.67	575.8154	39.04871	41.47813	39.9553	39.89468
-27.67	504.9933	39.01688	41.44022	39.91665	39.85387
-27.67	433.272	38.98441	41.40142	39.87707	39.81204
-27.67	361.6958	38.95195	41.36217	39.83702	39.76968
-27.67	290.3371	38.91949	41.32253	39.79656	39.72683
-27.67	218.0066	38.8864	41.28184	39.755	39.68278
-27.67	144.9655	38.85269	41.24019	39.71246	39.63762

Table 5-18: Comparison of Liquid Densities for Mixture 7

-27.67	69.8503	38.81711	41.19677	39.66808	39.59048
8.33	1015.31	38.00679	40.48669	38.99981	39.00717
8.33	945.5317	37.97183	40.44224	38.95477	38.96042
8.33	874.3616	37.93437	40.39629	38.9082	38.91203
8.33	798.6372	37.89504	40.34669	38.85791	38.85974
8.33	728.5839	37.85884	40.30015	38.8107	38.8106
8.33	655.7168	37.82013	40.25066	38.7605	38.75829
8.33	582.197	37.7808	40.20075	38.70984	38.70544
8.33	508.3872	37.74147	40.14949	38.6578	38.65111
8.33	435.0705	37.70152	40.0978	38.6053	38.59622
8.33	364.2774	37.66219	40.04711	38.5538	38.54233
8.33	290.5256	37.62161	39.99349	38.4993	38.48522
8.33	216.2227	37.57978	39.9386	38.44347	38.42665
8.33	142.0357	37.53733	39.88288	38.38679	38.36711
8.33	73.05564	37.498	39.83024	38.33321	38.31076
44.33	1012.96	36.7189	39.14011	37.69229	37.77098
44.33	944.3134	36.67645	39.08372	37.63543	37.71265
44.33	873.8104	36.6315	39.02488	37.57608	37.6517
44.33	801.1609	36.58531	38.96323	37.51388	37.58777
44.33	726.1037	36.53724	38.89841	37.44846	37.52046
44.33	653.3092	36.49104	38.83443	37.38386	37.45392
44.33	581.3123	36.44172	38.77002	37.3188	37.38683
44.33	508.6338	36.39428	38.70382	37.25191	37.31777
44.33	438.4499	36.34745	38.63873	37.18611	37.24976
44.33	364.437	36.29689	38.5688	37.11539	37.17658
44.33	291.4103	36.24757	38.49846	37.04423	37.10283
44.33	216.9768	36.19513	38.42533	36.97021	37.02603
44.33	144.6754	36.14519	38.35287	36.89683	36.94976
44.33	71.27167	36.09212	38.2778	36.82076	36.8706
80.33	1015.498	35.38169	37.64855	36.24891	36.41712
80.33	942.4569	35.32801	37.57052	36.17066	36.33785
80.33	875.5364	35.27494	37.49764	36.09755	36.2637
80.33	799.8266	35.21563	37.4135	36.01312	36.17799
80.33	727.5541	35.15758	37.33145	35.93074	36.09428
80.33	654.687	35.10014	37.24693	35.84586	36.00793
80.33	581.2688	35.03959	37.15987	35.7584	35.91884
80.33	509.1704	34.98028	37.07244	35.67051	35.82922
80.33	437.1445	34.91973	36.98308	35.58065	35.73746
80.33	362.3919	34.85543	36.8881	35.48511	35.63976
80.33	289.6989	34.7905	36.79346	35.38984	35.5422
80.33	217.1364	34.7262	36.69662	35.29231	35.44217
80.33	144.951	34.6619	36.59781	35.19275	35.33989
80.33	72.93961	34.59572	36.49665	35.09075	35.23493
116.33	1020.328	33.9652	35.9853	34.64523	34.92458
116.33	944.9081	33.89279	35.87926	34.53952	34.81894
116.33	866.0074	33.81537	35.76533	34.4259	34.70528

116.33	798.6227	33.74858	35.66548	34.32629	34 60554
116 33					51.00551
110.55	724.9144	33.67304	35.5534	34.21444	34.49342
116.33	650.5389	33.59688	35.43713	34.09836	34.37694
116.33	581.0802	33.52384	35.32549	33.98685	34.26492
116.33	508.2422	33.44518	35.20505	33.8665	34.14387
116.33	433.5621	33.36152	35.07777	33.73924	34.01569
116.33	364.1904	33.28349	34.95585	33.61728	33.89269
116.33	292.0485	33.20046	34.82507	33.48638	33.76047
116.33	217.0059	33.11181	34.68438	33.34546	33.6179
116.33	146.5319	33.02691	34.54762	33.2084	33.47898
152.33	1018.152	32.43447	34.1005	32.83449	33.25153
152.33	944.0233	32.34207	33.96039	32.69573	33.11497
152.33	870.0975	32.24906	33.81601	32.55268	32.97411
152.33	797.7525	32.15541	33.66988	32.40786	32.83139
152.33	722.4778	32.05491	33.51239	32.25171	32.67737
152.33	653.9183	31.96001	33.36373	32.10426	32.53182
152.33	580.8482	31.85763	33.19937	31.94116	32.37067
152.33	509.504	31.75525	33.03251	31.77548	32.20682
152.33	436.5354	31.64538	32.85474	31.59887	32.03197
152.33	361.9858	31.53113	32.66498	31.41022	31.84498
152.33	290.5691	31.41627	32.4747	31.22089	31.65707
152.33	219.0074	31.30077	32.27479	31.02181	31.45922
188.33	1021.213	30.77263	31.96273	30.78925	31.37738
188.33	942.3409	30.64154	31.7568	30.58678	31.18174
188.33	870.5906	30.51855	31.5609	30.3941	30.99556
188.33	798.2311	30.3887	31.35423	30.19076	30.79906
188.33	726.6694	30.25635	31.13988	29.97979	30.59519
188.33	651.6557	30.11215	30.90335	29.74688	30.37012
188.33	581.2108	29.96731	30.66884	29.51584	30.14686
188.33	508.7498	29.81436	30.41347	29.26407	29.90361
188.33	436.5789	29.65268	30.14284	28.99705	29.64567
188.33	360.0713	29.47226	29.83538	28.69337	29.35241
188.33	286.5516	29.2881	29.51643	28.37793	29.04794
188.33	220.9799	29.11267	29.20884	28.07324	28.75404
224.33	1015.73	28.85922	29.46225	28.40622	29.21066
224.33	939.3821	28.67693	29.17423	28.12529	28.94537
224.33	868.299	28.49838	28.88873	27.84676	28.68278
224.33	794.9823	28.30173	28.57402	27.53967	28.39381
224.33	725.8572	28.10446	28.2554	27.22865	28.10182
224.33	647.3336	27.86349	27.86274	26.84516	27.74291
224.33	580.152	27.64062	27.49547	26.48624	27.40827
224.33	507.3719	27.37655	27.0571	26.05744	27.01046
224.33	436.2453	27.09063	26.57686	25.58702	26.57704
224.33	362.0874	26.7529	26.00285	25.02353	26.06314

h) Mixture 8 (50% nButane)

Т	Р	Exp. Liquid	PR	PT	NEW
(°F)	(psia)	Den. $\left(\frac{Lb}{ft^3}\right)$			
-27.67	1012.51	39.5163	41.86586	40.38907	40.31979
-27.67	940.731	39.48571	41.83177	40.35438	40.28317
-27.67	867.5013	39.45387	41.79659	40.31856	40.24533
-27.67	795.7365	39.42328	41.76171	40.28304	40.20778
-27.67	722.8114	39.39144	41.72584	40.24651	40.16912
-27.67	647.9138	39.35898	41.68857	40.20852	40.12888
-27.67	577.6428	39.32839	41.65317	40.17244	40.09063
-27.67	499.8009	39.29406	41.61347	40.13196	40.04767
-27.67	431.6911	39.26409	41.5783	40.0961	40.00957
-27.67	358.0698	39.23163	41.53983	40.05684	39.96782
-27.67	284.6516	39.19979	41.50097	40.01718	39.9256
-27.67	213.5685	39.16795	41.46287	39.97828	39.88415
-27.67	143.5151	39.13611	41.42486	39.93945	39.84273
-27.67	67.89229	39.10178	41.38328	39.89697	39.79735
8.33	1013.424	38.28085	40.67648	39.22931	39.21214
8.33	947.2577	38.24776	40.63629	39.18858	39.16964
8.33	873.5639	38.21093	40.59094	39.14261	39.12163
8.33	800.3052	38.17285	40.54524	39.09626	39.07318
8.33	729.019	38.13664	40.50016	39.05052	39.02533
8.33	653.9618	38.09794	40.45201	39.00167	38.97415
8.33	578.8612	38.05985	40.40313	38.95205	38.92212
8.33	509.6635	38.02365	40.35745	38.90566	38.87343
8.33	434.1858	37.98307	40.30689	38.8543	38.81946
8.33	363.6973	37.94624	40.25897	38.80561	38.76823
8.33	291.6569	37.90753	40.20928	38.75509	38.71502
8.33	218.8478	37.8682	40.15829	38.70323	38.66033
8.33	146.6624	37.82887	40.10696	38.65101	38.60519
8.33	73.08465	37.78829	40.05381	38.59691	38.548
44.33	1008.58	37.01294	39.359	37.94815	37.99781
44.33	946.8806	36.9761	39.3108	37.89953	37.94766
44.33	869.2562	36.92928	39.24922	37.83739	37.88351
44.33	794.7067	36.88371	39.18907	37.77666	37.82076
44.33	728.9175	36.84188	39.13513	37.7222	37.76442
44.33	653.8748	36.79506	39.07261	37.65904	37.69902
44.33	582.0085	36.74949	39.01169	37.59748	37.6352
44.33	512.3612	36.70579	38.95165	37.53679	37.57221

Table 5-19: Comparison of Liquid Densities for Mixture 8

44.33	437.4636	36.65772	38.88595	37.47034	37.50316
44.33	364.1324	36.60965	38.82043	37.40406	37.4342
44.33	288.7417	36.55909	38.75181	37.33461	37.36183
44.33	219.8196	36.51351	38.68791	37.26991	37.29433
44.33	144.6609	36.46357	38.61691	37.19799	37.21918
44.33	74.56404	36.4155	38.54941	37.12958	37.14759
80.33	1015.788	35.69945	37.90763	36.54157	36.67572
80.33	944.4149	35.64701	37.8354	36.46908	36.60189
80.33	864.4265	35.58646	37.75282	36.38616	36.51733
80.33	793.7495	35.53339	37.67834	36.31136	36.44098
80.33	728.9755	35.48408	37.6088	36.24148	36.36957
80.33	645.9122	35.42165	37.51773	36.14996	36.27594
80.33	578.4696	35.36796	37.44217	36.07398	36.19811
80.33	504.1666	35.30865	37.35717	35.98848	36.11042
80.33	436.7674	35.25434	37.27841	35.90922	36.02903
80.33	364.611	35.19566	37.19226	35.82249	35.93983
80.33	289.0317	35.13261	37.09989	35.72946	35.84403
80.33	218.9784	35.07392	37.01225	35.64114	35.75293
80.33	142.7029	35.00837	36.91448	35.54256	35.65108
80.33	73.99839	34.94844	36.82422	35.45151	35.55685
80.33	45.25186	34.92222	36.78582	35.41275	35.51668
116.33	1028.871	34.31168	36.29838	34.98754	35.22689
116.33	937.3371	34.22802	36.17696	34.86635	35.10509
116.33	866.1379	34.16185	36.07995	34.76949	35.00764
116.33	792.6762	34.09318	35.97738	34.66705	34.90446
116.33	724.0877	34.02763	35.87923	34.56898	34.80559
116.33	651.5252	33.95709	35.77276	34.46256	34.69817
116.33	576.207	33.88092	35.65921	34.34902	34.58343
116.33	502.6147	33.80663	35.54511	34.23488	34.46792
116.33	438.1308	33.74046	35.44243	34.1321	34.36377
116.33	364.5965	33.66243	35.32205	34.01157	34.24146
116.33	291.2218	33.58314	35.19824	33.88753	34.11538
116.33	231.2051	33.51697	35.09406	33.78309	34.00907
116.33	141.9777	33.41646	34.93398	33.62251	33.8453
152.33	1027.551	32.81965	34.47148	33.22904	33.59865
152.33	939.3821	32.71664	34.31524	33.07406	33.44523
152.33	872.6356	32.63736	34.19306	32.95283	33.32511
152.33	797.9991	32.54621	34.05219	32.81301	33.18645
152.33	723.5221	32.45257	33.90684	32.66868	33.04319
152.33	656.471	32.36642	33.77159	32.53433	32.9097
152.33	586.2871	32.2734	33.62525	32.3889	32.76506
152.33	510.8093	32.17227	33.462	32.22657	32.60344
152.33	435.3171	32.06614	33.29212	32.05756	32.43492
152.33	366.6416	31.96688	33.13133	31.89749	32.27512
152.33	291.3088	31.85514	32.94747	31.71434	32.092
152.33	226.4478	31.75525	32.78233	31.54969	31.92713

152.33	144.8349	31.6254	32.56457	31.33239	31.70913
188.33	1018.471	31.18091	32.37527	31.21851	31.75082
188.33	944.4004	31.06168	32.19506	31.04098	31.57814
188.33	873.0417	30.94868	32.01429	30.86283	31.40483
188.33	798.9563	30.82507	31.81848	30.6698	31.21697
188.33	727.9312	30.70209	31.62222	30.47626	31.02854
188.33	654.1504	30.57162	31.40857	30.26547	30.82326
188.33	582.9222	30.43802	31.19181	30.05151	30.6148
188.33	509.2574	30.29506	30.9555	29.8181	30.38732
188.33	435.9697	30.14711	30.70656	29.57204	30.14742
188.33	366.221	29.99665	30.45507	29.32323	29.90474
188.33	292.7592	29.8306	30.17248	29.04337	29.63167
188.33	220.0662	29.65767	29.87202	28.7454	29.34082
224.33	1012.568	29.33741	29.95466	28.90618	29.64389
224.33	943.4722	29.18634	29.71438	28.67124	29.42036
224.33	871.7654	29.02153	29.45087	28.41354	29.17542
224.33	799.3914	28.84673	29.16831	28.13713	28.91301
224.33	725.3785	28.65695	28.85949	27.83493	28.62655
224.33	655.8183	28.46717	28.54793	27.5299	28.33795
224.33	582.3711	28.25242	28.19241	27.18162	28.00919
224.33	510.3887	28.02518	27.81209	26.80874	27.65828
224.33	435.8392	27.76922	27.37693	26.38157	27.25794
224.33	363.5523	27.49392	26.90361	25.91614	26.82425

i) Mixture 9 (75% nButane)

Т	Р	Exp. Liquid	PR	PT	NEW
(°F)	(psia)	Den. $\left(\frac{Lb}{ft^3}\right)$			
-27.67	1014.033	39.81346	42.02601	40.59175	40.50191
-27.67	941.6447	39.78474	41.99314	40.5583	40.46642
-27.67	873.6364	39.7554	41.96193	40.52653	40.43267
-27.67	800.0006	39.72356	41.92775	40.49173	40.39569
-27.67	726.8289	39.69484	41.89341	40.45675	40.35848
-27.67	654.0634	39.66363	41.85885	40.42154	40.32099
-27.67	582.3711	39.63679	41.82441	40.38644	40.28358
-27.67	508.6773	39.60307	41.78859	40.34993	40.24463
-27.67	435.6796	39.57373	41.75269	40.31331	40.20552
-27.67	363.3347	39.54377	41.71668	40.27658	40.16625
-27.67	291.7294	39.50943	41.6806	40.23976	40.12686
-27.67	217.586	39.47822	41.64279	40.20117	40.08551
-27.67	145.183	39.44513	41.6054	40.16299	40.04456

Table 5-20: Comparison of Liquid Densities for Mixture 9

27.67	72.27200	20 44267	44 56700	40.4244	40.00004
-27.67	/2.3/396	39.41267	41.56/33	40.1241	40.00281
-27.67	946.547	38.53805	41.99538	40.56058	40.46884
-27.67	/2/.829/	38.43255	41.89388	40.45723	40.35899
8.33	1018.79	38.57301	40.86509	39.45847	39.41704
8.33	870.8662	38.50497	40.77882	39.371	39.32524
8.33	799.4785	38.46876	40.73635	39.32793	39.27997
8.33	655.4557	38.40258	40.64896	39.23925	39.18663
8.33	581.5589	38.36201	40.6032	39.19279	39.13765
8.33	506.9513	38.32642	40.55634	39.1452	39.08743
8.33	437.5071	38.29334	40.51211	39.10027	39.03996
8.33	364.0599	38.25401	40.46468	39.05207	38.98898
8.33	290.4386	38.21717	40.41644	39.00303	38.93705
8.33	216.0341	38.17597	40.36696	38.95271	38.88369
8.33	144.0517	38.13851	40.31837	38.90328	38.83121
8.33	71.98236	38.10355	40.269	38.85303	38.77779
8.33	861.9318	37.23019	40.77353	39.36564	39.31961
8.33	652.9756	37.10346	40.64743	39.2377	39.185
44.33	1020.98	37.31946	39.584	38.2112	38.2324
44.33	939.1936	37.27513	39.52323	38.14987	38.16881
44.33	801.4075	37.19273	39.41843	38.04405	38.05892
44.33	724.3488	37.14591	39.35844	37.98345	37.99589
44.33	581.6749	37.05976	39.24462	37.86841	37.87603
44.33	509.2719	37.01793	39.18543	37.80856	37.81357
44.33	435.3896	36.97236	39.124	37.74641	37.74863
44.33	362.74	36.92679	39.06254	37.68422	37.68355
44.33	288.8867	36.87997	38.99896	37.61984	37.61611
44.33	217.528	36.83439	38.93642	37.55651	37.54966
44.33	144.7334	36.78507	38.87148	37.4907	37.48052
44.33	71.72129	36.73825	38.80512	37.42345	37.40975
44.33	1016.064	35.99973	39.58038	38.20755	38.22861
44.33	793.822	35.84054	39.41257	38.03813	38.05277
80.33	938.8455	35.94479	38.0845	36.75289	36.85145
80.33	863.0486	35.89111	38.01027	36.67831	36.77497
80.33	724 8709	35 7906	37 8711	36 53841	36 63128
80.33	647 8847	35 73254	37 7913	36 45815	36 5487
80.33	582 11	35 68509	37 72177	36 38819	36 47663
80.33	505 25//	35 626/1	37 63888	36 30/76	36 39058
80.33	/3/ 5193	35 57085	37.56096	36 2263	36 3095/
80.33	264 0744	25 51520	27 / 917/	26 1/65	26 227
80.33	201 1622	35.31323	37 30706	36 06208	36 13056
<u>80.33</u>	291.1030	25 20017	27 21077	35.00200	36 01702
80.33 80.22	1/15 2061	25 2/112	27 22/2/	25 0021500	25 05015
00.33	14J.3001 72 24572	25.24112 25.20101	27 12600	25.00/3/	22.22012
00.33	13.345/2	33.28181 24.56701	30,13009	35.19192	25.80504
00.33	341.0040	24.20/UI	2000000	36 53044	26 62120
80.33	1024.8/09	34.37280	37.8/11	30.53841	30.03128
116.33	1024.795	34.63818	36.58486	35.30479	35.50451

116.33	872.5341	34.50645	36.39197	35.11204	35.30958
116.33	798.8983	34.4409	36.29528	35.01537	35.21165
116.33	654.8611	34.30793	36.09917	34.81921	35.0126
116.33	578.223	34.23614	35.99081	34.71075	34.90233
116.33	509.1704	34.16684	35.89062	34.61042	34.80019
116.33	433.7797	34.09443	35.77829	34.4979	34.68547
116.33	363.1461	34.02513	35.67012	34.38949	34.57477
116.33	290.9462	33.94959	35.55644	34.2755	34.45819
116.33	218.4272	33.87593	35.4389	34.15757	34.33737
116.33	146.2563	33.79852	35.31838	34.03658	34.21318
152.33	1018.065	33.16987	34.79741	33.58074	33.90389
152.33	942.2104	33.08622	34.67092	33.45509	33.77874
152.33	866.399	33.00007	34.54063	33.32563	33.64968
152.33	800.0151	32.92265	34.42318	33.20889	33.53317
152.33	723.8266	32.82901	34.28424	33.07074	33.39516
152.33	652.1489	32.74536	34.14922	32.93643	33.26085
152.33	580.8482	32.65671	34.01048	32.79836	33.1226
152.33	507.575	32.56307	33.86295	32.65148	32.97535
152.33	435.9407	32.46693	33.71351	32.50262	32.82589
152.33	362.7255	32.37142	33.555	32.34463	32.66704
152.33	293.0493	32.2734	33.39829	32.18835	32.50964
152.33	218.4127	32.16603	33.22352	32.01392	32.33366
152.33	146.2708	32.0599	33.04716	31.83776	32.15558
188.33	1025.897	31.56859	32.79607	31.65815	32.13498
188.33	940.4844	31.44686	32.60228	31.46689	31.94773
188.33	865.0356	31.33324	32.42376	31.29065	31.77508
188.33	797.68	31.22524	32.25804	31.12699	31.61467
188.33	717.7641	31.09664	32.05293	30.92435	31.41594
188.33	644.4909	30.97428	31.85598	30.72969	31.22489
188.33	576.8451	30.85566	31.66582	30.54164	31.04021
188.33	506.0811	30.72956	31.45742	30.33545	30.83757
188.33	429.501	30.58785	31.21972	30.10011	30.60608
188.33	362.0148	30.4555	30.99847	29.88087	30.39025
188.33	286.6386	30.30005	30.7365	29.62106	30.13421
188.33	220.0371	30.15585	30.49016	29.37647	29.89291
224.33	1010.552	29.74569	30.43254	29.39464	30.06661
224.33	939.1791	29.60086	30.2026	29.1693	29.85061
224.33	869.9959	29.45353	29.96801	28.93936	29.63032
224.33	798.3617	29.29371	29.71147	28.68784	29.3895
224.33	724.3198	29.11892	29.42975	28.41153	29.12517
224.33	649.3061	28.93163	29.12456	28.11205	28.83898
224.33	583.3428	28.75683	28.83707	27.82979	28.56958
224.33	505.1383	28.53521	28.46878	27.46792	28.22478
224.33	437.2171	28.32795	28.11993	27.1248	27.89855
224.33	365.3942	28.08885	27.71485	26.72583	27.52032
224.33	288.9592	27.8073	27.23172	26.24904	27.07023

Propane-nButane-iButane System

j) Mixture 10 (20% Propane & 60% n-butane)

Т	Р	Exp. Liquid	PR	PT	NEW
(°F)	(psia)	Den. $\left(\frac{Lb}{ft^3}\right)$			
-27.67	1012.829	39.23725	41.48397	39.96995	39.92214
-27.67	946.1409	39.20853	41.45182	39.93723	39.88778
-27.67	874.6807	39.17794	41.417	39.90177	39.85051
-27.67	798.7243	39.14548	41.37955	39.86362	39.81038
-27.67	727.206	39.11489	41.34386	39.82725	39.7721
-27.67	656.4565	39.08367	41.30814	39.79085	39.73374
-27.67	582.8642	39.05246	41.27054	39.75251	39.69331
-27.67	512.7383	39.02062	41.23428	39.71553	39.65427
-27.67	438.4644	38.98816	41.1954	39.67586	39.61236
-27.67	364.321	38.95445	41.1561	39.63575	39.56993
-27.67	287.6974	38.92074	41.11495	39.59374	39.52544
-27.67	218.8768	38.89015	41.07752	39.5555	39.48492
-27.67	145.8792	38.85706	41.03731	39.51443	39.44133
-27.67	72.1274	38.82335	40.99616	39.47236	39.39665
8.33	1020.241	37.98369	40.26263	38.78043	38.78877
8.33	944.8355	37.94686	40.21595	38.73312	38.73965
8.33	873.3318	37.91128	40.17108	38.68763	38.69237
8.33	799.9861	37.87444	40.12442	38.6403	38.64315
8.33	730.1503	37.83824	40.07938	38.59461	38.59558
8.33	655.0206	37.80015	40.03024	38.54475	38.54361
8.33	583.2703	37.7627	39.98264	38.49641	38.4932
8.33	509.3299	37.72462	39.93286	38.44586	38.44041
8.33	438.0583	37.68654	39.88417	38.3964	38.3887
8.33	362.9141	37.64658	39.83207	38.34344	38.33328
8.33	289.2638	37.60725	39.7802	38.29071	38.27803
8.33	218.0501	37.56792	39.72929	38.23892	38.2237
8.33	146.5754	37.52859	39.6774	38.18613	38.16825
8.33	73.62129	37.48801	39.62361	38.13136	38.11066
44.33	1016.223	36.69705	38.90407	37.46095	37.54062
44.33	944.4149	36.65335	38.84662	37.403	37.48115
44.33	870.5616	36.6084	38.78656	37.34241	37.41892
44.33	800.2907	36.5647	38.72848	37.28379	37.35865
44.33	725.7121	36.51851	38.66581	37.22051	37.29352

Table 5-21: Comparison of Liquid Densities for Mixture 10

44.33	654.8031	36.47356	38.6052	37.1593	37.23045
44.33	577.3238	36.42424	38.5378	37.0912	37.1602
44.33	507.3574	36.37867	38.47583	37.02857	37.09553
44.33	434.8094	36.3306	38.41045	36.96245	37.02717
44.33	359.8393	36.28128	38.34163	36.89283	36.9551
44.33	288.4516	36.23321	38.27486	36.82526	36.88506
44.33	217.9776	36.18577	38.20772	36.75728	36.81451
44.33	143.9937	36.1352	38.13588	36.68451	36.73888
44.33	74.37549	36.08651	38.06695	36.61465	36.66617
80.33	1023.62	35.34736	37.403	36.00823	36.17764
80.33	941.2531	35.29117	37.3171	35.92205	36.0903
80.33	869.2127	35.22999	37.24038	35.84506	36.01219
80.33	799.0143	35.17818	37.16414	35.76852	35.93445
80.33	726.4373	35.12074	37.0837	35.68774	35.85233
80.33	655.3977	35.0708	37.00333	35.60699	35.77014
80.33	583.0528	35.01212	36.91973	35.52297	35.68453
80.33	509.475	34.94657	36.83281	35.43558	35.59538
80.33	433.9102	34.88539	36.74148	35.3437	35.50152
80.33	363.6828	34.82484	36.65461	35.25628	35.41209
80.33	290.4821	34.7699	36.56195	35.16298	35.31652
80.33	218.4417	34.70123	36.46853	35.06887	35.21997
80.33	146.3869	34.6413	36.37277	34.97234	35.12079
116.33	1016.571	33.91526	35.70556	34.37069	34.65159
116.33	942.718	33.84596	35.60357	34.26898	34.54988
116.33	873.912	33.78042	35.50623	34.17187	34.45269
116.33	801.9151	33.7105	35.40185	34.06769	34.34833
116.33	728.7434	33.6387	35.29297	33.959	34.23933
116.33	654.8611	33.56441	35.18003	33.8462	34.12609
116.33	579.2528	33.48638	35.06113	33.72739	34.00668
116.33	506.8788	33.41022	34.94398	33.61027	33.88881
116.33	433.9102	33.33218	34.82236	33.48862	33.76623
116.33	363.3637	33.25477	34.7012	33.36737	33.64387
116.33	289.7424	33.17237	34.57078	33.23676	33.51188
116.33	217.644	33.08934	34.43883	33.10454	33.37805
116.33	145.4731	33.00444	34.30226	32.96759	33.23918
152.33	1024.302	32.38702	33.80854	32.54826	32.96767
152.33	943.7913	32.28901	33.65857	32.39967	32.82138
152.33	871.0837	32.19849	33.51852	32.26086	32.68461
152.33	799.6815	32.10734	33.3764	32.11994	32.54567
152.33	725.2335	32.00933	33.22302	31.96781	32.39555
152.33	654.3534	31.91382	33.07169	31.81764	32.24723
152.33	579.7749	31.81081	32.90643	31.65355	32.08503
152.33	506.6903	31.70593	32.73795	31.48618	31.91941
152.33	431.8071	31.59481	32.55802	31.30731	31.74222
152.33	362.6385	31.48931	32.38462	31.13481	31.57114
152.33	290.5256	31.37506	32.19575	30.94677	31.38442

152.33	218.8913	31.25895	31.99912	30.75082	31.18956
188.33	1015.397	30.66151	31.60291	30.43675	31.0301
188.33	942.4279	30.5404	31.41301	30.24994	30.84956
188.33	872.2875	30.42054	31.22252	30.0625	30.6684
188.33	796.2731	30.2857	31.0064	29.84976	30.46278
188.33	726.9305	30.15772	30.79953	29.64603	30.26587
188.33	653.2076	30.01538	30.56826	29.41815	30.04562
188.33	579.1947	29.86556	30.32287	29.17622	29.81181
188.33	508.4162	29.7151	30.07418	28.93085	29.57472
188.33	434.5193	29.55029	29.79779	28.6579	29.31104
188.33	363.1171	29.38174	29.51193	28.37528	29.03811
188.33	290.4966	29.19882	29.19891	28.06537	28.73899
224.33	1020.705	28.68941	29.07104	28.02395	28.8369
224.33	939.0775	28.48652	28.76095	27.72135	28.55132
224.33	865.4562	28.30049	28.46181	27.42935	28.27627
224.33	796.2876	28.11008	28.16122	27.13585	28.00036
224.33	722.0572	27.89782	27.81392	26.79658	27.68227
224.33	657.4282	27.69556	27.48687	26.47691	27.3835
224.33	582.9512	27.4446	27.07544	26.07442	27.00885
224.33	509.6925	27.17616	26.62577	25.63399	26.60125
224.33	437.7247	26.87588	26.1276	25.14519	26.15245
224.33	364.2339	26.5319	25.54001	24.56703	25.62784

k) Mixture 11 (34% Propane & 33% n-butane):

Т	Р	Exp. Liquid	PR	РТ	NEW
(°F)	(psia)	Den. $\left(\frac{Lb}{ft^3}\right)$			
-27.67	1026.216	38.53743	40.95986	39.35886	39.35797
-27.67	941.7027	38.49935	40.91589	39.31411	39.31149
-27.67	869.6914	38.45752	40.87794	39.27547	39.27132
-27.67	805.7151	38.42818	40.84384	39.24073	39.23518
-27.67	719.635	38.38947	40.79737	39.19338	39.18586
-27.67	650.5534	38.35826	40.75957	39.15485	39.1457
-27.67	572.7696	38.32268	40.71646	39.1109	39.09985
-27.67	504.9063	38.29021	40.67837	39.07204	39.05927
-27.67	437.6667	38.259	40.64016	39.03306	39.01853
-27.67	364.2339	38.22591	40.59791	38.98993	38.97341
-27.67	290.2791	38.19158	40.55478	38.94589	38.92729
-27.67	222.9814	38.15912	40.51502	38.90527	38.88472
-27.67	148.0258	38.12228	40.47015	38.85941	38.8366
-27.67	70.83656	38.0842	40.42326	38.81148	38.78624

Table 5-22: Comparison of Liquid Densities for Mixture 11

8.33	1018.08	37.25266	39.66202	38.09803	38.16174
8.33	937.6562	37.20959	39.60784	38.04316	38.10538
8.33	866.8776	37.17213	39.55944	37.99412	38.05497
8.33	797.2159	37.13592	39.51113	37.94515	38.00459
8.33	722.8549	37.09534	39.4588	37.8921	37.94995
8.33	656.1519	37.05789	39.41118	37.84379	37.90016
8.33	579.5718	37.01606	39.35568	37.78748	37.84206
8.33	505.3994	36.97423	39.30106	37.73204	37.78479
8.33	438.2033	36.93677	39.25082	37.68102	37.73205
8.33	366.4385	36.89557	39.19635	37.62569	37.67478
8.33	289.9745	36.85125	39.13735	37.56572	37.61265
8.33	222.9959	36.81129	39.08483	37.51232	37.55726
8.33	145.4006	36.76572	39.02297	37.4494	37.49191
8.33	72.19992	36.72327	38.96359	37.38896	37.42906
44.33	1014.338	35.92357	38.22222	36.70413	36.84939
44.33	939.6577	35.87487	38.15665	36.63808	36.78235
44.33	865.4708	35.82618	38.09035	36.57127	36.71448
44.33	800.2762	35.78248	38.03109	36.51154	36.65374
44.33	722.6663	35.73067	37.95929	36.43914	36.58005
44.33	649.9153	35.68072	37.89071	36.36996	36.50955
44.33	580.6741	35.63328	37.82424	36.30287	36.44114
44.33	509.0544	35.58334	37.75421	36.23217	36.36895
44.33	433.098	35.52965	37.67847	36.15567	36.29075
44.33	379.3034	35.49094	37.62388	36.10051	36.23431
44.33	288.7271	35.42602	37.53012	36.00572	36.13719
44.33	219.5295	35.37545	37.45686	35.93161	36.06116
44.33	147.112	35.32176	37.37861	35.85242	35.9798
44.33	73.11366	35.2662	37.29691	35.76968	35.89469
80.33	1026.361	34.54141	36.62981	35.16879	35.41573
80.33	944.7775	34.47524	36.53562	35.07449	35.32123
80.33	865.3547	34.40969	36.44172	34.98045	35.2269
80.33	796.1716	34.35101	36.35808	34.89664	35.14275
80.33	717.6045	34.28483	36.2609	34.79924	35.04486
80.33	641.097	34.21866	36.16393	34.702	34.94702
80.33	572.9436	34.15873	36.0755	34.61328	34.85766
80.33	505.0803	34.10005	35.98543	34.52288	34.76651
80.33	438.4789	34.04012	35.895	34.43207	34.67485
80.33	354.3713	33.96208	35.77774	34.31426	34.55579
80.33	283.1867	33.89528	35.6757	34.21168	34.45196
80.33	218.2097	33.83285	35.58019	34.1156	34.35458
80.33	147.9098	33.76418	35.47416	34.00887	34.24626
80.33	78.08846	33.69551	35.36592	33.89986	34.13545
116.33	1015.368	33.03627	34.80802	33.41818	33.79114
116.33	947.2577	32.96573	34.70241	33.31316	33.68737
116.33	861.5112	32.87521	34.56557	33.17702	33.55274
116.33	788.354	32.79655	34.44518	33.05719	33.43415

116.33	714.5297	32.71477	34.32006	32.93261	33.31074
116.33	647.3046	32.63923	34.20277	32.81577	33.19489
116.33	575.8299	32.55683	34.07431	32.68775	33.06784
116.33	507.546	32.47629	33.94776	32.56155	32.94248
116.33	431.227	32.3839	33.80155	32.41567	32.79741
116.33	354.1393	32.28839	33.64839	32.26274	32.64514
116.33	286.102	32.20161	33.50827	32.12273	32.50556
116.33	217.9196	32.11234	33.36284	31.97729	32.36038
116.33	146.1838	32.01558	33.20395	31.81827	32.2014
152.33	1026.956	31.41502	32.77127	31.47055	31.99962
152.33	937.5546	31.29204	32.58054	31.28228	31.81664
152.33	863.0051	31.18653	32.41489	31.11869	31.6576
152.33	790.4861	31.0804	32.24746	30.95329	31.49672
152.33	723.9137	30.97865	32.08785	30.79553	31.34323
152.33	651.1626	30.86565	31.90638	30.61608	31.16856
152.33	577.9909	30.74766	31.71574	30.42745	30.98487
152.33	497.7994	30.61344	31.49647	30.21033	30.77335
152.33	435.1285	30.50482	31.31677	30.03225	30.59977
152.33	363.8858	30.37559	31.10259	29.81983	30.39263
152.33	291.1783	30.23887	30.87188	29.59076	30.16914
152.33	214.8883	30.0878	30.61482	29.3352	29.91967
188.33	1014.642	29.57464	30.3608	29.17394	29.90105
188.33	944.2119	29.43792	30.14545	28.96311	29.70033
188.33	872.4761	29.29559	29.91495	28.73736	29.48557
188.33	799.7105	29.14264	29.66824	28.49565	29.25582
188.33	727.1045	28.98157	29.40745	28.24	29.01308
188.33	654.7305	28.81239	29.13087	27.96873	28.75584
188.33	583.8505	28.63697	28.84146	27.68463	28.48687
188.33	508.9383	28.43845	28.51219	27.36111	28.18119
188.33	437.2025	28.23618	28.17004	27.02449	27.86397
188.33	364.5385	28.01332	27.79084	26.6508	27.51302
224.33	1016.339	27.38592	27.5489	26.5081	27.48824
224.33	943.8783	27.16679	27.21232	26.18157	27.18581
224.33	873.9265	26.93893	26.86164	25.8413	26.87181
224.33	798.5647	26.67673	26.44979	25.44154	26.50467
224.33	729.4396	26.41391	26.03392	25.03768	26.13602
224.33	652.4389	26.08741	25.51597	24.53431	25.68035
224.33	583.6764	25.75405	24.98933	24.02185	25.2218
224.33	515.117	25.37761	24.38059	23.42836	24.69932
224.33	441.2781	24.88193	23.58393	22.64882	24.03184

I) Mixture 12 (60% Propane & 20% n-butane):

Т	Р	Exp. Liquid	PR	PT	NEW
(°F)	(psia)	Den. $\left(\frac{Lb}{ft^3}\right)$			
-27.67	1024.142	37.70838	40.16486	38.46041	38.5195
-27.67	944.8791	37.67093	40.12034	38.41512	38.47304
-27.67	864.7601	37.63534	40.07471	38.36868	38.42537
-27.67	786.5846	37.59914	40.02956	38.32271	38.37814
-27.67	722.1877	37.5648	39.99189	38.28434	38.33869
-27.67	654.0199	37.53546	39.95154	38.24322	38.29639
-27.67	577.5993	37.49675	39.9057	38.1965	38.24827
-27.67	510.4902	37.46179	39.86491	38.15491	38.20541
-27.67	433.0109	37.42559	39.81717	38.10621	38.15517
-27.67	358.1858	37.38501	39.77041	38.05848	38.10589
-27.67	288.5676	37.35317	39.72628	38.01343	38.05934
-27.67	218.0356	37.31509	39.68097	37.96715	38.01146
-27.67	140.6869	37.27576	39.63055	37.91562	37.9581
-27.67	70.03885	37.24454	39.58381	37.86783	37.90856
-27.67	22.35036	37.21833	39.55188	37.83517	37.87468
8.33	1028	36.36119	38.78148	37.12199	37.25737
8.33	936.2203	36.30937	38.71408	37.05381	37.18823
8.33	869.4883	36.27129	38.66425	37.00339	37.13706
8.33	794.8953	36.22822	38.60771	36.94615	37.07893
8.33	716.1976	36.18327	38.54707	36.88474	37.0165
8.33	651.2641	36.14456	38.49624	36.83325	36.96411
8.33	584.3436	36.10523	38.44309	36.77937	36.90925
8.33	503.2383	36.05654	38.37758	36.71295	36.84156
8.33	435.027	36.01534	38.32154	36.6561	36.78355
8.33	358.882	35.96852	38.25792	36.59153	36.71761
8.33	287.9004	35.8961	38.19757	36.53025	36.65497
8.33	214.2356	35.87987	38.13384	36.4655	36.5887
8.33	147.8227	35.83679	38.07538	36.40609	36.52783
8.33	72.519	35.78872	38.00792	36.33748	36.45747
44.33	1024.78	34.98028	37.23167	35.62818	35.85862
44.33	932.9569	34.91536	37.14261	35.53866	35.76894
44.33	868.7196	34.86978	37.07905	35.47475	35.70486
44.33	794.0685	34.81672	37.00383	35.39909	35.62894
44.33	725.2045	34.76678	36.9331	35.32791	35.55745
44.33	655.0496	34.71496	36.85967	35.25399	35.48315
44.33	582.8497	34.66127	36.78259	35.17636	35.40505
44.33	510.4757	34.60696	36.70373	35.0969	35.32503
44.33	436.7384	34.55078	36.62167	35.01417	35.24163
44.33	361.3332	34.49209	36.53589	34.92764	35.15431

Table 5-23: Comparison of Liquid Densities for Mixture 12

44.33	293.1218	34.43841	36.45658	34.8476	35.07345
44.33	219.7616	34.37972	36.3694	34.75957	34.9844
44.33	145.1975	34.31917	36.27869	34.66791	34.89158
44.33	73.86785	34.25986	36.18984	34.57806	34.80047
80.33	1013.54	33.50136	35.47642	33.94309	34.29151
80.33	944.502	33.44018	35.38668	33.85354	34.20297
80.33	872.2585	33.37463	35.2906	33.75762	34.10808
80.33	796.5342	33.30471	35.1874	33.65454	34.00603
80.33	727.7717	33.24041	35.09135	33.55856	33.91094
80.33	654.0199	33.17049	34.98573	33.45297	33.80624
80.33	581.994	33.09995	34.87984	33.34706	33.70112
80.33	509.591	33.02878	34.77052	33.23766	33.59245
80.33	435.3751	32.95324	34.65529	33.12227	33.47771
80.33	364.2339	32.8802	34.54163	33.00839	33.36435
80.33	289.0462	32.80092	34.41789	32.88431	33.24071
80.33	218.9059	32.72538	34.29888	32.7649	33.12157
80.33	145.7342	32.64485	34.1708	32.63628	32.99308
116.33	1012.38	31.90383	33.49046	32.04566	32.54055
116.33	942.631	31.82267	33.36601	31.92243	32.42057
116.33	872.7662	31.74027	33.2376	31.79523	32.29669
116.33	799.0724	31.65037	33.09782	31.65671	32.1617
116.33	722.0572	31.55423	32.94664	31.50681	32.01556
116.33	652.9756	31.46496	32.80627	31.36754	31.87972
116.33	582.9947	31.37257	32.65911	31.22147	31.73716
116.33	509.0689	31.27143	32.49781	31.06126	31.5807
116.33	435.9552	31.16905	32.33186	30.89631	31.41953
116.33	363.9439	31.0648	32.16162	30.72694	31.25393
116.33	289.8004	30.9543	31.97861	30.54471	31.07562
116.33	220.7043	30.8463	31.8003	30.36696	30.90156
152.33	1017.804	30.14211	31.21363	29.88123	30.55913
152.33	940.8905	30.02038	31.01876	29.68994	30.37636
152.33	873.6074	29.90863	30.84092	29.5153	30.20955
152.33	798.9853	29.77878	30.63485	29.31284	30.01624
152.33	728.9175	29.6533	30.43203	29.11347	29.82596
152.33	655.0496	29.51471	30.2073	28.89242	29.61512
152.33	580.8192	29.36988	29.96884	28.65769	29.39139
152.33	502.9773	29.20944	29.70332	28.39607	29.14225
152.33	432.9674	29.05711	29.44904	28.14525	28.90366
152.33	363.5523	28.89917	29.1802	27.87971	28.65141
152.33	286.0875	28.71064	28.85715	27.56009	28.34831
188.33	1004.345	28.02268	28.47037	27.2846	28.20015
188.33	931.405	27.8354	28.1901	27.01212	27.9463
188.33	871.3448	27.67621	27.94529	26.77402	27.725
188.33	715.5595	27.2236	27.23748	26.08512	27.08789
188.33	646.8405	27.00011	26.88326	25.73995	26.77084
188.33	569.4047	26.72543	26.44352	25.31093	26.3793
188.33	501.9475	26.46011	26.01619	24.89331	26.00138
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188.33	431.6041	26.15046	25.51282	24.40016	25.56018
224.33	1017.5	25.53368	25.24481	24.2494	25.47182
224.33	940.5569	25.20468	24.76277	23.78586	25.05501
224.33	871.5043	24.87381	24.27969	23.3213	24.64113
224.33	800.3632	24.48738	23.71654	22.77962	24.16439
224.33	726.6404	24.01418	23.03634	22.12498	23.59866
224.33	653.9618	23.43922	22.22105	21.33937	22.93938
224.33	583.0673	22.68883	21.18663	20.33974	22.14403

Liquid Viscosity Predictions for Light Component Systems

Methane-Ethane System

a) Mixture 1 (50% methane):

Table 5-24: Comparison of Liquid Viscosities for Mixture 1

Т	Р	Exp. Visc.	PR	РТ	NEW
(°F)	(psia)	(cp)			
-9.67	3119.999	0.05573	0.065873467	0.053894	0.057411
-9.67	2604.897	0.05239	0.059210925	0.049503	0.0527
-9.67	2108.374	0.04833	0.052704175	0.045036	0.047931
-9.67	1615.564	0.0434	0.045914674	0.04013	0.042745
-9.67	1105.973	0.03691	0.037694441	0.033749	0.036161
-99.67	1090.178	0.07913	0.106190085	0.074515	0.078571
-99.67	589.536	0.07408	0.094506157	0.067914	0.071514

b) Mixture 2 (68% methane):

Т	Р	Exp. Visc.	PR	PT	NEW
(°F)	(psia)	(cp)			
-45.67	2596.093177	0.0487	0.055864	0.046305	0.048968
-45.67	2107 63/201	0.0448	0.0/9686	0.04215	0.044564

Гable 5-254: Comparison (f Liquid Viscosities f	or Mixture 2
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\ /	(F)	1-1-1			
-45.67	2596.093177	0.0487	0.055864	0.046305	0.048968901
-45.67	2107.634201	0.0448	0.049686	0.04215	0.044564778
-45.67	1792.843726	0.04205	0.045564	0.039254	0.041515954
-45.67	1479.938744	0.03904	0.041204	0.036058	0.038183805
-45.67	1193.865793	0.03508	0.03671	0.032599	0.034631986
-99.67	2097.815128	0.06504	0.085895	0.062717	0.066121426
-99.67	1247.529853	0.05775	0.070335	0.053723	0.05655305

-99.67	741.5502864	0.05318	0.060832	0.047905	0.05037089
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c) Mixture 3 (34.53% methane)

Т	Р	Exp. Visc.	PR	РТ	NEW
(°F)	(psia)	(cp)			
-45.67	2078.801	0.077	0.099012248	0.073557	0.078506
-45.67	1688.373	0.07338	0.09129948	0.068962	0.073568
-45.67	1268.792	0.07024	0.082980081	0.063897	0.068119
-45.67	859.4082	0.06574	0.074748634	0.058744	0.062577

Table 5-26: Comparison of Liquid Viscosities for Mixture 3

Methane- nHeptane System

Т	Р	Exp. Visc.	PR	PT	NEW
(°F)	(psia)	(cp)			
122	5013	0.219	0.258032	0.251037	0.2348
122	7515	0.254	0.32038	0.312028	0.295682
122	10003	0.286	0.379681	0.370041	0.353841
122	12497	0.323	0.436431	0.425554	0.40967
122	15001	0.356	0.490721	0.478655	0.463195
122	17505	0.393	0.542383	0.52918	0.514207
212	7503	0.181	0.225695	0.21999	0.212441
212	10003	0.209	0.277205	0.270344	0.262562
212	12507	0.235	0.327802	0.319813	0.311947
212	15010	0.262	0.37708	0.367998	0.360152
302	5013	0.115	0.122549	0.119674	0.117601
302	7504	0.142	0.164124	0.160271	0.157978
302	10011	0.162	0.206974	0.202129	0.199694
302	12505	0.184	0.250037	0.244211	0.241694
392	5012	0.092	0.092358	0.090532	0.090818
392	7504	0.112	0.124941	0.122346	0.122775

Table 5-27: Comparison of Liquid Viscosities for Methane-nHeptane System

Gas Condensate System

Table 5-28: Properties of Components for Gas Condensate System

Comp.	Mol.	PC	тс	Acentric	$MW\left(\frac{lb}{lbmol}\right)$	$\operatorname{VC}\left(\frac{ft^3}{lbm}\right)$
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	Frac.	(psia)	(°F)	Factor		
N2	0.0708	493.1	-232.51	0.0372	28.0134	0.051011
Co2	0.0062	1071	87.9	0.2667	44.01	0.034401
H2S	0	1300	212.45	0.0948	34.08	0.046097
C1	0.7104	667.8	-116.63	0.0104	16.043	0.098797
C2	0.0757	707.8	90.09	0.0979	30.07	0.078284
C3	0.0348	616.3	206.01	0.1524	44.097	0.072703
i-C4	0.0064	529.1	274.98	0.1848	58.124	0.071399
n-C4	0.0143	550.7	305.65	0.201	58.124	0.070298
i-C5	0.005	490.4	369.1	0.2223	72.151	0.067899
n-C5	0.0056	488.6	385.7	0.2539	72.151	0.067497
n-C6	0.0075	436.9	453.7	0.3007	86.178	0.065985
n-C7	0.0107	396.8	512.8	0.3498	100.205	0.062475
n-C8	0.0136	360.6	564.22	0.4018	114.232	0.06197
n-C9	0.0086	332	610.66	0.4455	128.259	0.061787
n-C10	0.0061	304	652.1	0.4885	142.286	0.062144
n-C11	0.0041	288.7	690.44	0.535	156.313	0.063057
C12+	0.0202	477.8944	891	0.6	232	0.064572

Table 5-29: Comparison of Liquid Viscosity for Gas Condensate System

Т	Р	Exp. Visc.	PR	РТ	NEW
(°F)	(psia)	(cp)			
243	5014.7	0.264	0.225295	0.215629	0.21919
243	4214.7	0.277	0.290121	0.291686	0.281104
243	3514.7	0.292	0.369154	0.393473	0.355391
243	2814.7	0.312	0.479103	0.54978	0.457083

<u>Propane</u>

Table 5-30: Comparison of Liquid Viscosity for Propane

Т	Р	Exp. Visc.	PR	РТ	NEW
(°F)	(psia)	(cp)			
40	1113.3	0.133745	0.160272	0.122382	0.129141
40	1117.6	0.133572	0.160394	0.122464	0.129228
220	1016	0.043986	0.03857	0.036545	0.040893
220	1016.8	0.043809	0.038591	0.036564	0.040911
220	2010.4	0.061134	0.056531	0.051983	0.056723
220	2011.9	0.061108	0.056553	0.052001	0.056743
220	4001.9	0.081097	0.084347	0.073709	0.080359
220	4010.5	0.081169	0.084466	0.073799	0.080457
400	3974.9	0.047122	0.046292	0.044094	0.047754
400	3985	0.047185	0.046373	0.044166	0.047831
400	4969.8	0.05485	0.054024	0.050837	0.055053
400	4981.1	0.055276	0.05411	0.050911	0.055133

n-butane:

T(F)	Р	Exp. Visc	PR	РТ	NEW
	(psia)	(cp)			
100	200	0.143	0.158029	0.130736	0.132356
100	300	0.1445	0.161299	0.133213	0.134974
100	400	0.146	0.164556	0.135676	0.137575
100	500	0.1472	0.167802	0.138126	0.14016
100	600	0.1485	0.171037	0.140564	0.142729
100	700	0.15	0.174263	0.142991	0.145284
100	800	0.1512	0.177479	0.145408	0.147826
100	1000	0.154	0.183886	0.150212	0.152872
130	700	0.1293	0.139772	0.117779	0.120914
130	800	0.1305	0.142722	0.120039	0.123307
130	1000	0.1335	0.148597	0.124527	0.128049
130	1500	0.14	0.16316	0.135587	0.139704
160	1500	0.1225	0.133651	0.113841	0.118452
160	2000	0.129	0.146802	0.123936	0.129116
190	2000	0.114	0.122022	0.105453	0.110838
220	3000	0.1125	0.123671	0.107252	0.113708

Liquid Viscosity Prediction for Mixtures with heavy hydrocarbons

1) Mixture 1:

Comp.	Mol. Frac.	PC (psia)	TC (°F)	Acentric Factor	$MW\left(\frac{lb}{lbmol}\right)$	$\operatorname{VC}\left(\frac{ft^3}{lbm}\right)$
n-C4	0.177	550.7	305.65	0.201	58.124	0.070298
n-C6	0.177	436.9	453.7	0.3007	86.178	0.068799
n-C10	0.56	304	652.1	0.4885	142.266	0.067908
Co2	0.086	1071	87.9	0.2667	44.01	0.034401

Table 5-32: Properties of Components for Mixture 1

Table 5-33: Comparison of Liquid Viscosity for Mixture 1

T(F)	P(psia)	Exp. Visc.	PR	РТ	NEW
		(cp)			
123.998	364.6255	0.35	0.257391	0.33763	0.245744
156.002	364.6255	0.307	0.225214	0.290701	0.210598
188.006	364.6255	0.275	0.196258	0.248888	0.180165
219.992	364.6255	0.25	0.170433	0.212022	0.154018
251.996	364.6255	0.232	0.147563	0.179808	0.131679

2) Mixture 2:

Table 5-34: Prop	oerties of Co	mponents for	[.] Mixture 2
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Comp.	Mol. Frac.	PC (psia)	TC (°F)	Acentric Factor	$MW\left(\frac{lb}{lbmol}\right)$	$\operatorname{VC}\left(\frac{ft^3}{lbm}\right)$
n-C4	0.15	550.7	305.65	0.201	58.124	0.070298
n-C6	0.15	436.9	453.7	0.3007	86.178	0.068799
n- C10	0.476	304	652.1	0.4885	142.266	0.067908
Co2	0.224	1071	87.9	0.2667	44.01	0.034401

Table 5-35: Comparison of Liquid Viscosity for Mixture 2

T(F)	P(psia)	Exp. Visc.	PR	РТ	NEW
		(cp)			
123.998	714.7473	0.304	0.253498	0.323133	0.238756
156.002	714.7473	0.279	0.219846	0.275285	0.203594
188.006	714.7473	0.244	0.190077	0.23345	0.173598
219.992	714.7473	0.216	0.163989	0.197285	0.148206
251.996	714.7473	0.172	0.141296	0.166315	0.126817

3) Mixture 3:

Comp.	Mol. Frac.	PC (psia)	TC (°F)	Acentric Factor	$MW\left(\frac{lb}{lbmol}\right)$	$\operatorname{VC}\left(\frac{ft^3}{lbm}\right)$
n-C5	0.088	488.6	385.7	0.2539	72.151	0.067497
n-C6	0.264	436.9	453.7	0.3007	86.178	0.068799
n-C7	0.094	396.8	512.8	0.3498	100.205	0.069098
n-C10	0.467	304	652.1	0.4885	142.266	0.067908
Co2	0.087	1071	87.9	0.2667	44.01	0.034401

Table 5-36: Properties of Components for Mixture 3

Table 5-37: Comparison of Liquid Viscosity for Mixture 3

T(F)	P(psia)	Exp. Visc.	PR	РТ	NEW
		(cp)			
188.006	364.6255	0.278	0.199817	0.248916	0.182795
219.992	364.6255	0.252	0.173196	0.212027	0.156257
230	364.6255	0.231	0.165512	0.20146	0.148766
251.996	364.6255	0.209	0.149661	0.179803	0.13356

4) Mixture 4:

Table 5-38: F	Properties	of	Components	for	Mixture 4
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Comp.	Mol. Frac.	PC (psia)	TC (°F)	Acentric Factor	$MW\left(\frac{lb}{lbmol}\right)$	$\operatorname{VC}\left(\frac{ft^3}{lbm}\right)$
n-C5	0.075	488.6	385.7	0.2539	72.151	0.067497
n-C6	0.225	436.9	453.7	0.3007	86.178	0.068799
n-C7	0.08	396.8	512.8	0.3498	100.205	0.069098
n-C10	0.397	304	652.1	0.4885	142.266	0.067908
Co2	0.223	1071	87.9	0.2667	44.01	0.034401

Table 5-39: Comparison of Liquid Viscosity for Mixture 4

T(F)	P(psia)	Exp. Visc.	PR	РТ	NEW
		(cp)			
188.006	714.7473	0.243	0.193678	0.234201	0.176391
219.992	714.7473	0.219	0.166765	0.197913	0.150531
230	714.7473	0.197	0.159089	0.187656	0.143302
251.996	714.7473	0.172	0.143395	0.166841	0.128727

5) Mixture 5:

Comp.	Mol. Frac.	PC (psia)	TC (°F)	Acentric Factor	$MW\left(\frac{lb}{lbmol}\right)$	$\operatorname{VC}\left(\frac{ft^3}{lbm}\right)$
n-C4	0.193	550.7	305.65	0.201	58.124	0.070298
n-C6	0.193	436.9	453.7	0.3007	86.178	0.068799
n-C10	0.614	304	652.1	0.4885	142.266	0.067908

Table 5-40: Properties of Components for Mixture 5

Table 5-41: Comparison of Liquid Viscosity for Mixture 5

T(F)	P(psia)	Exp. Visc.	PR	РТ	NEW
		(cp)			
123.998	44.6717	0.381	0.00698	0.3332938	0.23960419
156.002	54.67933	0.336	0.00744	0.288550286	0.20587115
188.006	64.68695	0.287	0.19251	0.248340939	0.17647465
219.992	74.69457	0.247	0.16784	0.212569645	0.15105864
251.996	94.70981	0.217	0.14606	0.181360489	0.12952827

6) Mixture 6:

Table 5-42: Properties of Components for Mixture 6

Comp.	Mol. Frac.	PC (psia)	TC (°F)	Acentric Factor	$ \begin{pmatrix} MW \\ \frac{lb}{lbmol} \end{pmatrix} $	$VC\left(\frac{ft^3}{lbm}\right)$
n-C5	0.097	488.6	385.7	0.2539	72.151	0.067497
n-C6	0.289	436.9	453.7	0.3007	86.178	0.068799
n-C7	0.103	396.8	512.8	0.3498	100.205	0.069098
n-C10	0.511	304	652.1	0.4885	142.266	0.067908

Table 5-43: Comparison of Liquid Viscosity for Mixture 6

T(F)	P(psia)	Exp. Visc.	PR	PT	NEW
		(cp)			
188.006	39.59537	0.266	0.007441	0.246641	0.177891
219.992	44.6717	0.241	0.007779	0.210818	0.152041
230	49.74803	0.228	0.007911	0.200612	0.144827

7) Mixture 7:

Comp.	Mol. Frac.	PC (psia)	TC (°F)	Acentric Factor	$MW\left(\frac{lb}{lbmol}\right)$	$\operatorname{VC}\left(\frac{ft^3}{lbm}\right)$
n-C5	0.288	488.6	385.7	0.2539	72.151	0.067497
n-C10	0.712	304	652.1	0.4885	142.266	0.067908

Table 5-44: Properties of Components for Mixture 7

Table 5-45: Comparison of Liquid Viscosity for Mixture 7

T(F)	P(psia)	Exp. Visc.	PR	PT	NEW
		(cp)			
177.998	34.66408	0.334	0.007163	0.269992	0.18764
199.004	44.6717	0.297	0.007447	0.245766	0.170303
219.992	54.67933	0.259	0.007743	0.223228	0.154502
242.006	64.68695	0.237	0.008061	0.201327	0.139446
262.994	74.69457	0.21	0.008389	0.182095	0.126469