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A Continuous Mixture Computational Algorithm for Reservoir Fluids Phase Behavior

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ABSTRACT

Development of compositional reservoir simulators is presently hampered due to the complexity of the existing computational algorithms of reservoir fluid phase behavior calculations. Through the application of the theory of polydisperse fluid mixtures we have developed a new algorithm for reservoir fluid phase behavior calculations. In this algorithm a reservoir fluid is considered to consist of a continuous mixture with a defined molecular weight/composition distribution. The proposed algorithm requires less computational time than the time required by the existing pseudocomponent VLE calculation methods. It is also shown that the new algorithm is applicable to varieties of equations of state and mixing rules. Results of the proposed technique are compared with results of the other existing techniques, and with the simulated multi-pseudocomponent reservoir fluid mixture data.

INTRODUCTION

In complex mixtures, such as petroleum reservoir fluids consisting of a large number of components, it is very difficult to identify components and their concentrations. Thus description of such mixtures by mole fractions of components is practically impossible, and a simplified technique which is called "pseudocomponent method" has been generally used. By this method the lighter components (C_1 to C_6) are identified and the heavier components (C_{7+}) are splitted into a number of fractions which are characterized by their average boiling points, liquid densities, and molecular weights. Then phase equilibrium calculations are performed by ordinary thermodynamic techniques in which each fraction is considered as a pure component.[1-3] When applying the

References and illustrations at end of paper.

pseudocomponent method to a system containing various families of compounds, such as paraffins, naphthenes, and aromatics, it is necessary to extend the pseudocomponent description of the mixtures up to C_{20+} or more so that the experimental data can be matched by the prediction. Phase equilibrium calculation for the latter case is intricate, and in addition to the difficulty mentioned above, a great deal of computational time will be necessary. As noted by Mathias and Benson,[4] for a typical system encountered in engineering calculations the computational time depends primarily on the terms involving pairs of components in the equation of state mixing rule summations.

An alternative method which can significantly reduce the difficulties mentioned above stems from the idea of polydisperse (or continuous) thermodynamics. That is, instead of mole fractions, a density distribution function which is a function of some measureable property such as normal boiling point, liquid density, or molecular weight is introduced to describe composition of many-component mixtures such as petroleum reservoir fluids.

Initially, the concept of continuous mixtures was applied in polymer solution theories and prediction of liquid-liquid separation in such solutions.[5-7] Aris and Gavalas[8] introduced functional analysis in the thermodynamic description of polydisperse system and kinetics of continuous reactions such as polymerization or cracking. Concerning petroleum distillation, flash calculation schemes which are based on simple models, such as Raoult's law, have been reported for continuous mixtures.[9-11] For example, Hoffman[10] presented a numerical integration method. In this method, he adopted integrated form of Clausius-Clapeyron equation and Raoult's law to compute vapor pressures of petroleum fluids. He used Gaussian distribution with normal boiling temperatures considered as the independent variable to describe the vapor and liquid phases. Similar procedures as Hoffman's were presented by Rätzsch and Kehlen,[12] whose treatment was based on

Raoult's law with compositions described by normal density function to derive thermodynamic properties of continuous mixtures. Edmister[13] developed a graphical procedure by which flash vaporization calculations of continuous mixtures were performed for petroleum distillation.

An approach using equation of state for solving phase equilibrium problems in continuous mixtures was developed by Gualtieri et al.[14] They used the van der Waals equation of state (vdW EOS) to solve the fractionation of a polydisperse impurity dissolved in a solvent, the shift of the critical point due to the presence of a polydisperse impurity, and calculation of the cloud-point surface and critical point of a completely polydisperse system. Similar results for flash calculation of continuous mixtures which were based on the Redlich-Kwong equation of state were reported by Briano.[15] Cotterman and Prausnitz[16] introduced two techniques for phase equilibrium calculation of reservoir fluids, one of which is a continuous model and the other is a semi-continuous model. In both of their techniques they used the Soave equation of state, with two interation parameters a and b (considered as linear expressions with respect to molecular weight), to perform phase equilibria calculation.

Due to the lack of a general algorithm for continuous mixture phase equilibrium calculation in all of the previous attempts reported above a number of crude assumptions has been made. Such approximations have hampered the effective use of continuous mixture theory in practical phase equilibrium calculations and development of accurate prediction techniques. In the present report we introduce a new computational algorithm by which we can perform phase equilibrium calculation of continuous mixtures consisting of species with a wide range of molecular weight distribution such as a petroleum reservoir fluid. In part II we introduce the basic theory behind this new technique. In part III we apply this technique, together with the continuous vdW EOS, to a specific continuous mixture which resembles a gas-condensate reservoir fluid. In part IV we demostrate VLE flash calculations using the proposed technique with the continuous Peng-Robinson equation of state (PR EOS) and compare results of the calculation with simulated gas-condensate VLE flash data.

THEORY OF PHASE EQUILIBRIUM OF CONTINUOUS MIXTURES

For a mixture with a large number of components, composition may be replaced with a composition distribution function $F(I, I_0, \eta)$ whose independent variable I is some measureable property such as molecular weight, boiling point, or the likes,[15] with mean value of I_0 and variance of η. The composition distribution function $F(I, I_0, \eta)$ is normalized such that

$$\int_I F(I) dI = 1$$

over the entire range of I.

Several single-variable distribution functions have been proposed for the composition distribution functions of continuous mixtures by different investigators as it is reported in Table 1. For example, normal distribution function was adopted by Kehlen and Rätzsch[12] and also by Hoffman[10] for polymers and reservoir fluids; the Schulz[17] distribution function was used by Gualtieri et al.[14] for their fundamental studies on continuous mixtures; the gamma distribution function was selected by Cotterman and Prausnitz[16] for petroleum reservoir fluids. To choose a certain continuous distribution function, some general knowledge about composition of the many-component mixtures under consideration is necessary. For example, the exponential-decay distribution function[18] can be applied for the gas condensate reservoir fluids; while the so called "gamma" distribution function[18] may be applied for heavy petroleum fluid mixtures, asphaltenes, and resins.

In the case of the continuous mixtures an extensive thermodynamic property, such as the compressibility factor Z, may be considered as a function of temperature T, pressure P and the extensive distribution function F(I):

$$Z = Z[T, P, F(I)] \tag{1}$$

where I is an independent variable such as molecular weight or boiling point.

For such a mixture from the expression of the compressibility factor one can derive expressions for other thermodynamic properties. Here, we present expression for the chemical potential of a fraction in a continuous mixtures which is needed for the phase equilibrium calculation, that is

$$\mu(I) = \mu^o(I) + \int_v^\infty [\frac{\partial P}{\partial F(I)} - \frac{RT}{v}] dv - RT \ln \frac{v}{RTF(I)} + RT \tag{2}$$

or

$$\mu(I) = \mu^o(I) + \delta(v, T, I_0, \eta : \{I\}) - RT \ln \frac{v}{F(I, I_0, \eta)} \tag{3}$$

where $\mu^o(I)$ is the chemical potential of the continuous reference state at temperature T. With regard to the above equation we have assumed that there is only one family of continuous mixtures in the system. Extension of the above formulas to the case of a multi-family continuous mixtures is straightforward.

According to classical thermodynamics for phase equilibrium calculation between several phases in equilibrium, it is necessary to equate temperatures, pressures, and chemical potentials of components in every phase. For vapor-liquid equilibrium of a continuous mixture, this means that the following conditions must be satisfied:

$$T = T^L = T^V \tag{4}$$

$$P = P^L[v_L, T, F_L(I)] = P^V[v_V, T, F_V(I)] \tag{5}$$

$$\mu^L(I) = \mu^V(I) \tag{6}$$

Considering the fact that distribution varialbe I changes from some initial value of I_1 to very large values, eqn. (6) is representative of very many equations.

To illustrate the difference between discrete and continuous vapor-liquid equilibrium concepts, Fig.(1) which represents a comparative flash VLE calculation scheme is reported. Considering the fact that distribution functions are not additive $F_f(I)$, $F_L(I)$, and $F_V(I)$ can not possess the same functional forms in a specific flash calculation scheme. Considering eqn.(3), we can rewrite the phase equilibrium conditions for a one-family continuous mixture as the followings:

$$P = P^L(T, v_L, Io_L, \eta_L) = P^V(T, v_V, Io_V, \eta_V) \qquad (7)$$

$$\delta^L(T, v_L, Io_L, \eta_L:\{I\}) - RT\ln\frac{v_L}{F(Io_L, \eta_L:\{I\})}$$

$$= \delta^V(T, v_V, Io_V, \eta_V:\{I\}) - RT\ln\frac{v_V}{F(Io_V, \eta_V:\{I\})} \qquad (8)$$

In order to perform phase equilibrium calculation, it is necessary to use an equation of state to calculate pressure and chemical potential in the liquid and vapor phases in equilibrium.

Generally, it is impossible to solve eqn's.(7) and (8) simultaneously for the liquid and vapor continuous mixtures in equilibrium. That has been the reason for the crude simplifying assumptions proposed by the previous investigators.[14-16]

At this stage we assume that there exist the first n derivatives of the chemical potential. As a result eqn's.(7) and (8) can be replaced by the following set of equations:

$$P = P^L(T, v_L, Io_L, \eta_L) = P^V(T, v_V, Io_V, \eta_V) \qquad (9)$$

$$\mu^L(T, v_L, Io_L, \eta_L:\{I\}) = \mu^V(T, v_V, Io_V, \eta_V:\{I\}) \qquad (10)$$

$$\frac{\partial\mu^L(T, v_L, Io_L, \eta_L:\{I\})}{\partial I} = \frac{\partial\mu^V(T, v_V, Io_V, \eta_V:\{I\})}{\partial I} \qquad (11)$$

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$$\cdots\cdots\cdots\cdots\cdots\cdots\cdots\cdots\cdots\cdots$$

$$\frac{\partial^n\mu^L(T, v_L, Io_L, \eta_L:\{I\})}{\partial I^n}$$

$$= \frac{\partial^n\mu^V(T, v_V, Io_V, \eta_V:\{I\})}{\partial I^n} \qquad (12)$$

To perform vapor-liquid equilibrium calculation for a one-family continuous mixture, according to the phase rule of continuous mixtures (to be published later) we need at most 5 equations. We now assume that there exists a field distribution function

$$F_x(I) = F_x(I, \eta_x, I_{ox}) \qquad (13)$$

such that the following four equations can effectively represent the n equations given by eqn's.(10)-(12).

$$\int_I \mu^L(I) F_x(I) dI = \int_I \mu^V(I) F_x(I) dI \qquad (14)$$

$$\int_I \frac{\partial\mu^L(I)}{\partial I} F_x(I) dI = \int_I \frac{\partial\mu^V(I)}{\partial I} F_x(I) dI \qquad (15)$$

$$\int_I \frac{\partial^2\mu^L(I)}{\partial I^2} F_x(I) dI = \int_I \frac{\partial^2\mu^V(I)}{\partial I^2} F_x(I) dI \qquad (16)$$

$$\int_I \frac{\partial^3\mu^L(I)}{\partial I^3} F_x(I) dI = \int_I \frac{\partial^3\mu^V(I)}{\partial I^3} F_x(I) dI \qquad (17)$$

The equality of pressures, eqn.(9), together with eqn's.(14)-(17) will constitute 5 equations which are needed for VLE calculation of a one-family continuous mixture. Functional form and parameters of the field distribution function $F_x(I, \eta_x, I_{ox})$ will depend on the nature of the system under consideration. In some instances, it may be chosen to be identical with the distribution function of the feed.

CONTINUOUS vdW EOS FOR GAS-CONDENSATES

To illustrate utility of the proposed continuous mixture phase equilibrium algorithm we present here a simple example in which for the equation of state the vdW EOS of mixtures,

$$P = RT/(v-b) - a/v^2 \qquad (18)$$

$$a = \sum_i \sum_j x_i x_j (a_i a_j)^{1/2} \qquad (19)$$

$$b = \sum_i x_i b_i \qquad (20)$$

and for the distribution function the exponential-decay distribution function

$$F(I) = (1/\eta)\exp[-(I-Io)/\eta] \qquad (21)$$

are used.

In order to extend the vdW EOS to continuous mixtures eqn's.(19) and (20) for a and b must be replaced with the following two expressions, respectively.

$$a = \{\int_I F(I)[a(I)]^{1/2} dI\}^2 \qquad (22)$$

$$b = \int_I F(I) b(I) dI \qquad (23)$$

When we apply the above equations for a homologous series of hydrocarbon compounds, $[a(I)]^{1/2}$, and $b(I)$ will be polynomials relating to the molecular weight of hydrocarbons compound. For the family of paraffins, Fig's.(2) and (3) indicate that $[a(I)]^{1/2}$ and $b(I)$ can be accurately correlated to the following 3rd order polynomials, respectively.

$$[a(I)]^{1/2} = a_0 + a_1 I + a_2 I^2 + a_3 I^3 \qquad (24)$$

$$b(I)=b_0+b_1 I+b_2 I^2+b_3 I^3 \qquad (25)$$

By introducing eqn's.(24), (25) and (21) into eqn's.(22) and (23), the following expressions for a and b will be obtained

$$a = [a_0+a_1(I_0+\eta)+a_2(I_0^2+2I_0\eta^2)$$
$$+a_3(I_0^3+3I_0^2\eta+6I_0\eta^2+6\eta^3)]^2 \qquad (26)$$

$$b = b_0+b_1(I_0+\eta)+b_2(I_0^2+2I_0\eta^2)$$
$$+b_3(I_0^3+3I_0^2\eta+6I_0\eta^2+6\eta^3) \qquad (27)$$

In order to derive expression for the chemical potential in the continuous mixtures eqn.(18) will be replaced in eqn.(3) with the following result

$$\mu(I)=D_0 +D_1 I+D_2 I^2+D_3 I^3 \qquad (28)$$

where

$$D_0 =-\ln(v-b)+\frac{b_0}{v-b}-\frac{2a_0Q(I,\eta)}{RTv}-\ln\eta+I_0/\eta \qquad (29)$$

$$D_1 = \frac{b_1}{v-b} - \frac{2a_1Q(I,\eta)}{RTv} -1/\eta \qquad (30)$$

$$D_2 = \frac{b_2}{v-b} - \frac{2a_2Q(I,\eta)}{RTv} \qquad (31)$$

$$D_3 = \frac{b_3}{v-b} - \frac{2a_3Q(I,\eta)}{RTv} \qquad (32)$$

$$Q(I,\eta)=2[a(I)]^{1/2}[a_0+a_1(I_0+\eta)+a_2(I_0^2+2I_0\eta$$
$$+2\eta^2)+a_3(I_0^3+3I_0^2\eta+6I_0\eta^2+6\eta^3)]$$

According to eqn's.(29)-(32) D_0, D_1, D_2 and D_3 are all functions of temperature, volume, and the distribution variance η.

To perform VLE flash calculation we need to replace eqn.(28) in eqn's.(14)-(17). For the present example n=3, but since, $I_0= I_{0_L}= I_{0_V}= I_{0_f}=$ molecular weight of methane, we will only need to consider up to the first derivatives of the chemical potential in the criteria of equilibrium. As a result, we will need to solve the following set of 4 equations:

$$P = \frac{RT}{v_L-b_L} - \frac{a_L}{v_L^2} = \frac{RT}{v_V-b_V} - \frac{a_V}{v_V^2} \qquad (33)$$

$$D_0^L-D_0^V+(D_1^L-D_1^V)(I_{0_x}+\eta_x)$$
$$+(D_2^L-D_2^V)(I_{0_x}^2+2I_{0_x}\eta_x+2\eta_x^2)$$
$$+(D_{3L}-D_{3V})(I_{0_x}^3+3I_{0_x}^2\eta_x+6I_{0_x}\eta_x^2+6\eta_x^3)=0 \qquad (34)$$

$$D_1^L-D_1^V+2(D_2^L-D_2^V)(I_{0_x}+\eta_x)$$
$$+3(D_3^L-D_3_V)(I_{0_x}^2+2I_{0_x}\eta_x+2\eta_x^2) = 0 \qquad (35)$$

On Fig.(4) the P-T diagrams resulting from the simultaneous solution of the above 4 equations for $\eta_f=15.0$ and for various values of η_x are

reported. Also reported on this figure is the P-T diagram based on a previous continuous mixture VLE computational scheme proposed by Gualtieri et al.[14] According to Fig.(4) the present technique is more versatile in representing VLE data of different reservoir fluids.

Application of the proposed algorithm for the vdW EOS is a simple example of continuous mixture phase equilibrium calculation. In what follows we introduce the proposed algorithm for a sophisticated equation of state which is extensively used for reservoir fluid phase behavior calculation.

CONTINUOUS PR EOS FOR GAS-CONDENSATES

The PR EOS for mixtures,

$$P = \frac{RT}{v-b} - \frac{a(T)}{v(v+b)+b(v-b)} \qquad (36)$$

$$a(T)=\sum_i \sum_j x_i x_j (a_{ii}a_{jj})^{1/2} = [\sum_i x_i a_{ii}^{1/2}]^2 \qquad (37)$$

$$b=\sum_i x_i b_i \qquad (38)$$

where

$$a_{ii}(T)=a(T_{ci})[1+k_i(1-T_{ri}^{1/2})]^2 \qquad (39)$$

$$a(T_{ci})=0.45724R^2T_{ci}^2/P_{ci} \qquad (40)$$

$$b_i=0.0778RT_{ci}/P_{ci} \qquad (41)$$

$$k_i=0.37464+1.54226w-0.26992w^2 \qquad (42)$$

has received a wide spread acceptance in phase behavior calculation of reservoir fluids.[1] In order to extend this equation of state to continuous mixtures we introduce eqn.(37) in the following form:

$$[a(T)]^{1/2} =a_1-a_2T^{1/2} \qquad (43)$$

where
$$a_1=\sum_i x_i a_{i1} , a_2=\sum_i x_i a_{i2} \qquad (44)$$

and

$$a_{i1}=[a(T_{ic})]^{1/2}(1+k_i) \qquad (45)$$

$$a_{i2}=[a(T_{ci})/T_{ci}]^{1/2} k_i \qquad (46)$$

On Fig's(5)-(7) graphical representation of a_{i1}, a_{i2}, and b_i for paraffin hydrocarbons versus molecular weight are reported. We have been able to represent $a_1(I)$, $a_2(I)$, and $b(I)$ of paraffins, by the following third order polynomials with respect to molecular weight I:

$$a_1(I)=a_{10}+a_{11}I+a_{12}I^2+a_{13}I^3 \qquad (47)$$

$$a_2(I)=a_{20}+a_{21}I+a_{22}I^2+a_{23}I^3 \qquad (48)$$

$$b(I)=b_0+b_1I+b_2I^2+b_3I^3 \qquad (49)$$

Now, by using the exponential decay-distribution

function and a similar procedure as in the case of the vdW EOS we can derive the continuous mixture expressions for parameters a_1, a_2, and b of the PR EOS as the following

$$a_1 = a_{10} + a_{11}(I_0 + \eta) + a_{12}(I_0^2 + 2I_0\eta + 2\eta^2)$$
$$+ a_{13}(I_0^3 + 3I_0^2\eta + 6I_0\eta^2 + 6\eta^3) \tag{50}$$

$$a_2 = a_{20} + a_{21}(I_0 + \eta) + a_{22}(I_0^2 + 2I_0\eta + 2\eta^2)$$
$$+ a_{23}(I_0^3 + 3I_0^2\eta + 6I_0\eta^2 + 6\eta^3) \tag{51}$$

$$b = b_0 + b_1(I_0 + \eta) + b_2(I_0^2 + 2I_0\eta + 2\eta^2)$$
$$+ b_3(I_0^3 + 3I_0^2\eta + 6I_0\eta^2 + 6\eta^3) \tag{52}$$

Knowing the continuous PR EOS chemical potential of components of the mixture can be derived in the following form:

$$\mu = D_0 + D_1 I + D_2 I^2 + D_3 I^3 \tag{53}$$

where

$$D_0 = -RT\ln(v-b) + C_1 b_0$$
$$+ C_2(a_{10} + a_{20}T^{\frac{1}{2}}) - RT\ln\eta + RTI_0/\eta \tag{54}$$

$$D_1 = C_1 b_1 + C_2(a_{11} + a_{21}T^{\frac{1}{2}}) - RT/\eta \tag{55}$$

$$D_2 = C_1 b_2 + C_2(a_{12} + a_{22}T^{\frac{1}{2}}) \tag{56}$$

$$D_3 = C_1 b_3 + C_2(a_{13} + a_{23}T^{\frac{1}{2}}) \tag{57}$$

and

$$C_1 = \frac{RT}{v-b} - \frac{a(v+b)}{b(v^2+2vb-b^2)}$$
$$- \frac{a}{2.828b^2} \ln\frac{v-0.414b}{v+2.414b} \tag{58}$$

$$C_2 = \frac{Q(\eta)}{1.414b} \ln\frac{v-0.414b}{v+2.414b} \tag{59}$$

$$Q(\eta) = a_{10} + a_{11}(I_0+\eta) + a_{12}(I_0^2 + 2I_0\eta + 2\eta^2)$$
$$+ a_{13}(I_0^3 + 3I_0^2\eta + 6I_0\eta^2 + 6\eta^3)$$
$$- [a_{20} + a_{21}(I_0+\eta) + a_{22}(I_0^2 + 2I_0\eta + 2\eta^2)$$
$$+ a_{23}(I_0^3 + 3I_0^2\eta + 6I_0\eta^2 + 6\eta^3)]T^{\frac{1}{2}} \tag{60}$$

To perform VLE flash calculation we need to replace eqn.(36) in eqn's.(14)-(17). In this example n=3, but since $I_{ox} = I_{0L} = I_{0v} = I_{0f} =$ molecular weight of methane, we will only need to consider up to the first derivative of the chemical potential in the criteria of equilibrium. As a result, we need to solve the following set of 4 equations.

$$P = \frac{RT}{v_L - b_L} - \frac{(a_{1L} - a_{2L}T^{\frac{1}{2}})^2}{v_L(v_L+b_L) + b_L(v_L-b_L)}$$

$$= \frac{RT}{v_V - b_V} - \frac{(a_{1V} - a_{2V}T^{\frac{1}{2}})^2}{v_V(v_V+b_V) + b_V(v_V-b_V)} \tag{61}$$

$$(D_0^L - D_0^V) + (D_1^L - D_1^V)(I_0 + \eta_x) + (D_2^L - D_2^V)$$
$$(I_0^2 + 2I_0\eta_x + 2\eta_x^2) + (D_3^L - D_3^V)$$
$$(I_0^3 + 3I_0^2\eta_x + 6I_0\eta_x^2 + 6\eta_x^3) = 0 \tag{62}$$

$$(D_1^L - D_1^V) + 2(D_2^L - D_2^V)(I_0 + \eta_x)$$
$$+ 3(D_3^L - D_3^V)(I_0^2 + 2I_0\eta_x + 2\eta_x^2) = 0 \tag{63}$$

It should be noted that Eqn's.(62) and (63) are similar to eqn's.(34) and (35) which were derived for the vdW EOS. By using eqn's.(61)-(63) the saturation pressure, P-T diagrams, of three different hypothetical gas-condensate reservoir fluids are calculated and they are reported in Fig's.(8)-(10). Also reported in these figures are the P-T diagrams of the same reservoir fluids assuming to contain 6, 10, and 20 pseudocomponents, respectively. According to these figures, the proposed continuous mixture model can represent phase behavior of a many-component mixture effectively.

CONCLUSIONS

We have proposed a continuous mixture phase equilibrium algorithm which can be used for vapor-liquid equilibrium calculations involving reservoir fluids. The proposed algorithm is applied to the case of gas-condensate systems with analyticl results. Accurate prediction of reservoir fluid phase behavior by using a pseudocomponent method requires the assumption of a large number of pseudocomponents and as a result excessive computer time is needed. The proposed continuous mixture technique reduces the required computer time significantly while it retains accuracy of the predictions. The computer time needed for the proposed scheme is roughly equivalent to the time needed for a two-components mixture VLE calculation.

The proposed continuous mixture technique is applicable to varieties of reservoir fluids, equations of state, and mixing rules. In the present report its application for some hypothetical gas condensate systems using two representative equations of state, the van der Waals and the Peng-Robinson equations, is demonstrated.

In order to apply the proposed algorithm for phase behavior prediction of realistic reservoir fluids we need to include the unlike-interaction parameters in the equation of state mixing rules. In addition, knowledge about the nature of the molecular weight distribution function of species of reservoir fluids will be necessary. In our ongoing research activities we are addressing these and other pertinent questions for the development of an accurate and simple reservoir fluid phase behavior prediction package.

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Table 1: Several Single Variable Distribution Function

Distribution		Mean	Variance	Application
Normal	$1/(2\pi\eta)^{\frac{1}{2}}\exp[-(I-I_0)^2/2\eta]$	I_0	η	Petroleum, Polymer
Gamma	$1/[\eta^\alpha\Gamma(\eta)](I-I_0)^{\alpha-1}\exp[-(I-I_0)/\eta]$	$\alpha\eta$	I_0^2/η	Heavy Petroleum, Polymer
Schulz	$1/[I\Gamma(\eta)](I\eta/I_0)^\eta\exp(-I\eta/I_0)$	I_0	$I_0^2/\eta^{\frac{1}{2}}$	Heavy Petroleum, Polymer
Exponential	$1/\eta\exp[-(I-I_0)/\eta]$	$I_0+\eta$	η	Gas Condensate
Weibull	$\alpha\eta(I-I_0)^{\eta-1}\exp[-\alpha(I-I_0)^\eta]$	$\alpha^{-1/\eta}\Gamma(1+1/\alpha)$	$\alpha^{-2/\eta}\{\Gamma(1+2/\eta)-[\Gamma(1+1/\eta)]^2\}$	Petroleum, polymer
Tung	$\eta/I_0(I/I_0)^{\eta-1}\exp[-(I/I_0)^\eta]$	$I_0\Gamma(1+1/\eta)$	$I_0^2[\Gamma(1+2/\eta)-\Gamma(1+1/\eta)]^2$	Polymer, Colloids
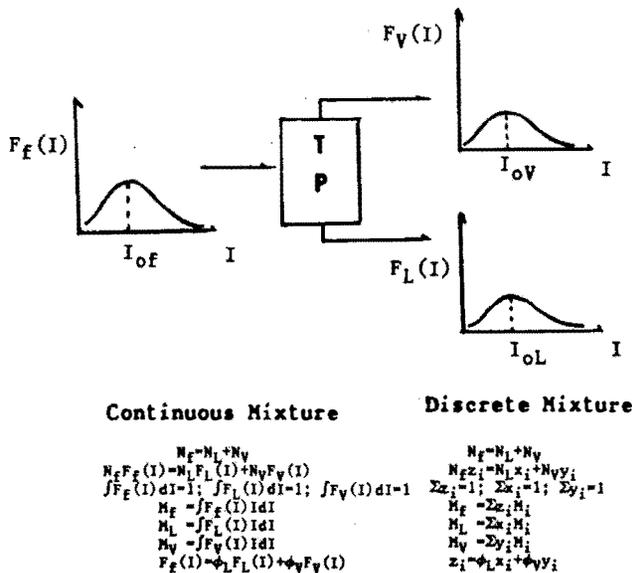
Fig. 1—Flash VLE material balance comparison of continuous and discrete mixtures.
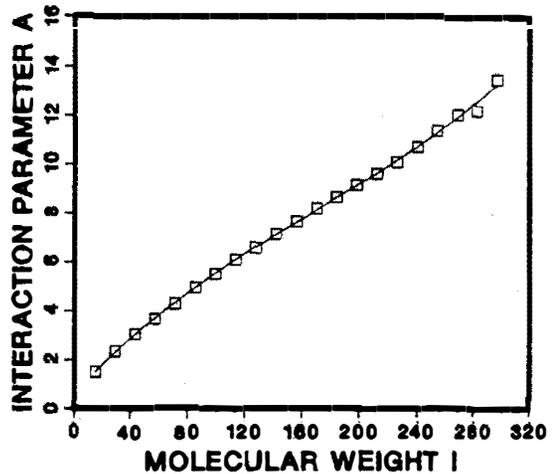
Fig. 2—Interaction parameter ai of the vdW EOS vs. molecular weight of paraffins.
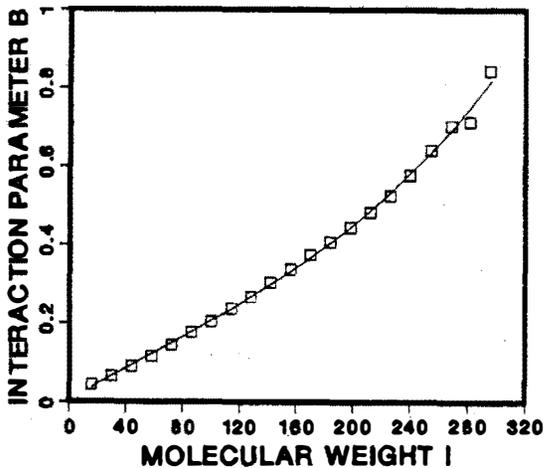
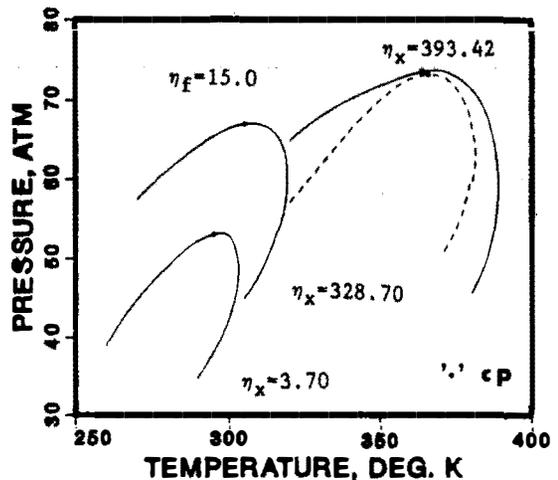Fig. 3—Interaction parameter bi of the vdW EOS vs. molecular weight of paraffins.

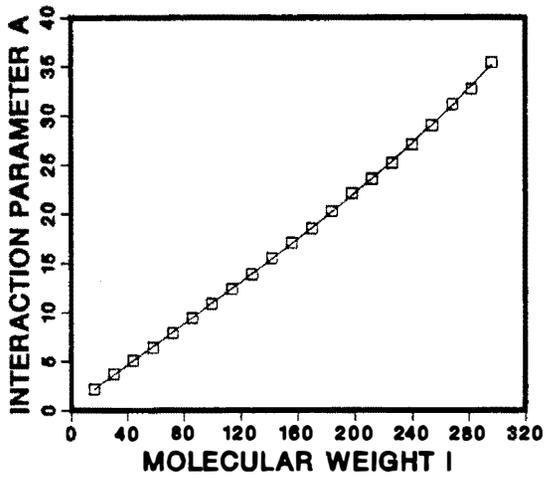Fig. 4—P-T diagram of a continuous mixture based on the vdW EOS.

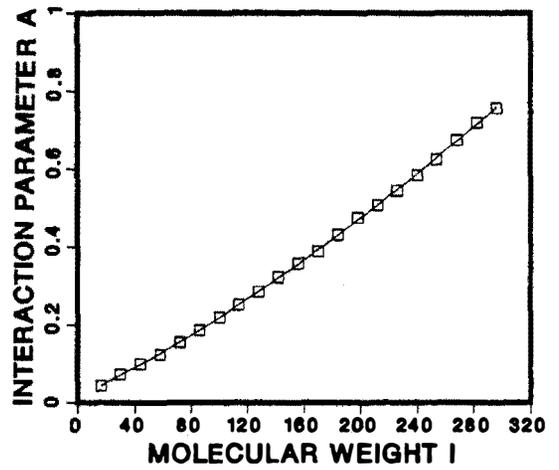Fig. 5—Parameter ai1 of the PR EOS vs. molecular
weight of paraffins.

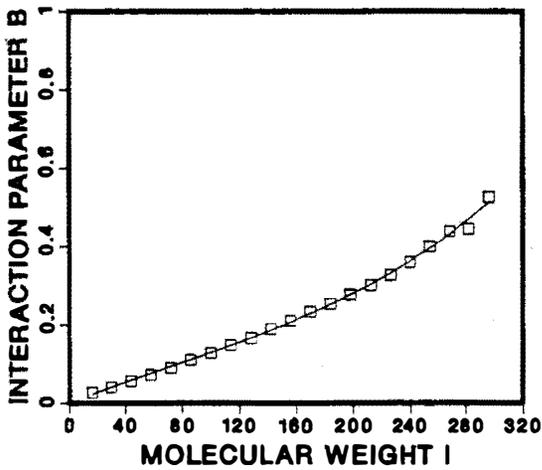Fig. 6—Parameter ai2 of the PR EOS vs. molecular
weight of paraffins.

Fig. 7—Parameter bi of the PR EOS vs. molecular
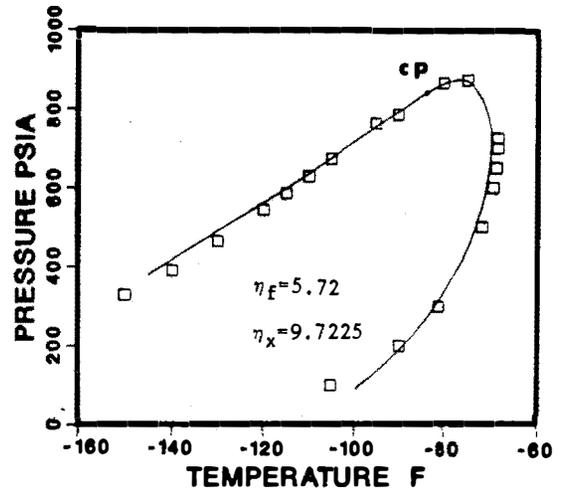weight of paraffins.

Fig. 8—P-T diagram of a continuous mixture with
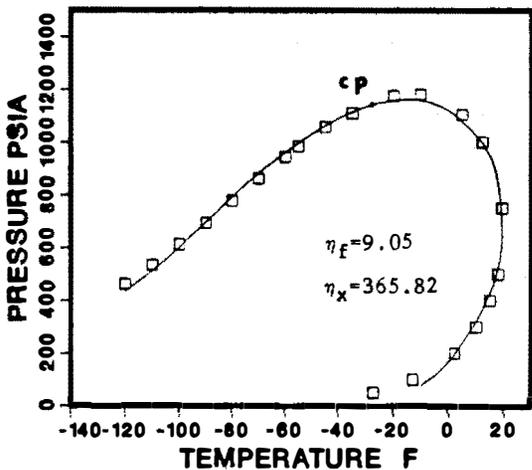$\eta_f = 5.72$ based on the PR EOS.

Fig. 9—P-T diagram of a continuous mixture with
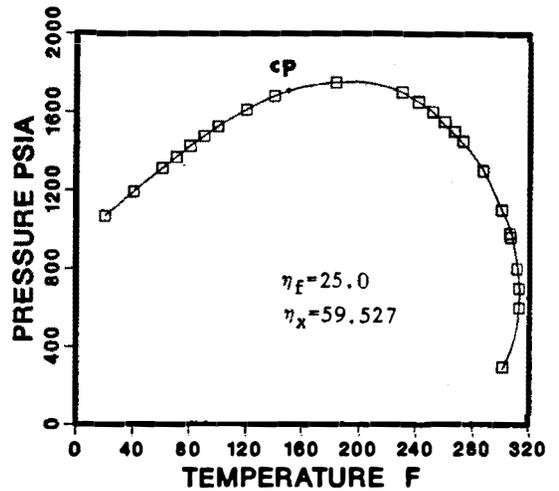$\eta_f = 9.05$ based on the PR EOS.

Fig. 10—P-T diagram of a continuous mixture with
$\eta_f = 25.0$ based on the PR EOS.