

# A Unified Model for Representing Densities and Viscosities of Hydrocarbon Liquids and Gases Based on Peng-Robinson Equation of State

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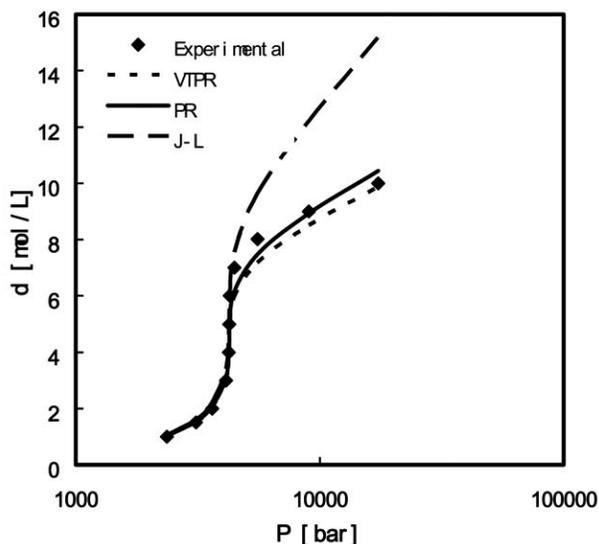
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**Abstract:** This work presents a volume translation for the Peng-Robinson (PR) cubic equation of state (EOS) for a better description of densities. A viscosity model has been developed based on the similarity between  $PVT$  and  $T\mu P$  relationship. The viscosity model can also describe the relation of the saturated vapor pressure with temperature. The volume translated PR EOS and the EOS based viscosity model are applicable to both liquid and gas phase and both sub and supercritical regions for pure hydrocarbons.

**Key Words:** Equation of state, volume translation, viscosity model, hydrocarbons.

## INTRODUCTION

The van der Waals type cubic equation of state, such as the Soave-Redlich-Kwong (SRK) and Peng-Robinson (PR) equations of state [1, 2], is widely used for representing vapor-liquid equilibrium data in the oil and natural gas industry. However, it is well known that the SRK EOS usually produces poor results for saturated liquid and supercritical densities and excess enthalpies. To improve the volumetric property prediction ability of the SRK EOS, Peneloux and Rauzy [3] proposed a constant volume translation method for liquid densities. For further improving the capability of the SRK EOS representing volumetric properties, especially in the near critical region, Chou and Prausnitz [4], Ji and Lempe [5], Wang and Gmehling [6] proposed a few temperature-dependent volume translation methods. Later, in order to obtain a better description of saturated liquid densities for PR EOS, Ahlers and Gmehling [7] adopted the idea of a constant volume translation and proposed a volume translated PR EOS (VTPR). In the VTPR model, the temperature dependent volume translation method of Ji and Lempe (J-L model) was also tested for the PR EOS. In Ahlers' study, the weakness of the J-L model was highlighted. Indeed, a temperature dependent volume translation in the sub critical region would lead to over-crossing isotherms in the supercritical region Fig. (1) shows the comparison of the calculated molar densities based on different volume translation models with experimental data and original PR EOS at the critical isotherm of carbon dioxide; Pfohl [8] also observed similar problems for a temperature dependent volume translation based on Soave-Redlich-Kwong equation of state). Up to now, the method of a temperature dependent volume translation is limited to the sub critical region, and the over-crossing problem in the supercritical region is unsolved.



**Fig. (1).** Comparison of the calculated molar densities based on different volume translation models with experimental data and original PR EOS at the critical isotherm of carbon dioxide.

Numerous viscosity graph and correlation for hydrocarbon liquids and gases are available in the literatures; there are main drawbacks in their utilization:

1. Application range and accuracy are limited.
2. Smooth transition in the near-critical region cannot be achieved since the viscosity of liquid phase and gas phase are calculated by using different graphs/correlations.
3. Separate density correlation is required since density is general involved in evaluating the fluid viscosity.

The similarity between the  $PVT$  and  $T\mu$  (viscosity) $P$  relationship was first pointed out by Philips in 1912 [9]. Little and Kennedy [10] developed the first EOS-based viscosity model using the van der Waals EOS. Lawal [11]

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proposed a viscosity model based on the four-parameter Lawal-Lake-Silberberg EOS that was applicable to pure hydrocarbons and their mixtures. In 1997, Guo *et al.* proposed two viscosity models, one of them is based on the Petal-Teja EOS and another is based on the PR EOS [12]. In 1999, Guo revised the viscosity model based on the PR EOS and the prediction accuracy was improved (see Appendix A, hereafter named as PR viscosity model) [13]. The major advantages of developing a viscosity model based on an equation of state are:

1. The viscosity of both gas and liquid phases can be described by a single model, achieving smooth transition of liquid/gas viscosity in the near-critical region.
2. Both high pressure and low pressure data can be correlated, and density is not involved in evaluating the fluid viscosity.

However, in all of these proposed models, the saturated vapor pressure was not considered in the correlation of experimental data of the  $T\mu P$  relation. Therefore, vapor-liquid equilibrium computation and viscosity calculations cannot be performed using a single model. Thus, thermodynamic consistency in process and reservoir simulation should be improved.

In this paper, a modified temperature-dependent volume translation is proposed for the PR EOS. The new model will be both suitable in the near and far from the critical point, as well as in the sub- and supercritical regions for the correlation of saturation density and  $PVT$  behavior. A viscosity model will be proposed based on the PR EOS and the similarity between the  $PVT$  and  $T\mu P$  relationship. The viscosity model will be applied for both the saturation vapor pressure and viscosity prediction of hydrocarbon liquids and gases. Based on the idea of volume translation for the PR EOS, the new viscosity model will be extended to supercritical region using a viscosity translation method.

## MODEL DEVELOPMENT

### Volume Translation for PR EOS

The equation of state used in this work is the Peng-Robinson EOS:

$$P = \frac{RT}{v-b} - \frac{a}{v^2 + 2bv - b^2} \quad (1)$$

Where  $a$  and  $b$  are energy and size parameters, respectively. The pure component parameters  $a$  and  $b$  can be obtained from the generalized correlation using critical constant  $T_c$  and  $P_c$

$$a = (0.45724R^2T_c^2 / P_c)\alpha(T_r) \quad (2)$$

$$b = 0.07780RT_c / P_c \quad (3)$$

Where the  $\alpha$ -function proposed by Twu *et al.* [14] is used:

$$\alpha(T_r) = T_r^{N(M-1)} \exp \left[ L \left( 1 - T_r^{NM} \right) \right] \quad (4)$$

The Twu's  $\alpha$ -function is continuous at the critical point when different parameters in the sub- and supercritical region are chosen. Moreover, it extrapolates very well at supercritical conditions without exhibiting abnormal charac-

teristics, and this function can approach zero, when there is no interaction between the molecules exists. Based on this  $\alpha$ -function, the Twu's parameters  $L$ ,  $M$ , and  $N$  were fitted for more than 65 compounds based on vapor pressure data. Part of these parameters was published in literature [15] and will be used in this paper.

With these new parameters, the accuracy for saturated vapor pressures has been significantly improved compared to the original PR EOS.

In this work, the VTPR model and the Chou-Prausnitz near-critical volume translation term will be modified as follows (hereafter named as T-VTPR):

$$v = v^{PR} + c - \Delta v_c \beta(T) \quad (5)$$

$$\Delta v_c = v_c^{PR} - v_c^{exp} \quad (6)$$

Where  $v^{PR}$  refers to the molar volume calculated from PR EOS without volume translation. The constant  $c$  can be calculated from following generalized relation [7]:

$$c = 0.252 \left( \frac{RT_c}{P_c} \right) (1.5448Z_c - 0.4024) \quad (7)$$

In which  $Z_c$  is the experimental critical compressibility. The following  $\beta$ -function is proposed:

$$\beta(T) = \frac{0.35}{0.35 + 0.5|d_r \alpha(T_r) - T_r|} \quad (8)$$

In Eq. (8) the Twu  $\alpha$ -function is used and the original parameter  $N$  and  $M$  will be retained, while the parameter  $L$  is replaced by parameter  $c_1$

$$\alpha(T_r) = T_r^{N(M-1)} \exp \left[ c_1 \left( 1 - T_r^{NM} \right) \right] \quad (9)$$

Because the  $\alpha$ -function is continuous at the critical point, the parameter  $c_1$  can be determined from the regression of saturated liquid densities in the sub critical region and from the  $PVT$  data in the supercritical region separately. The values of parameter  $c_1$  are vary from negative to positive for different substances, so that the  $\alpha$ -function will vary between zero and large positive values. For the correlation of  $PVT$  data in sub critical region, the calculated saturation liquid density and the critical density data from PR EOS are used to calculate  $d_r$  in Eq. (8)

$$d_r = d_{sat}^{PR} / d_c^{PR} \quad (10)$$

In Equation (10),  $d_r$  is a reduced density along the saturated liquid line,  $d_{sat}^{PR}$  is a calculated saturated liquid density (mol/L), and  $d_c^{PR}$  is a calculated critical density, both  $d_{sat}^{PR}$  and  $d_c^{PR}$  are calculated from PE EOS. In this case,  $d_r$  is only temperature dependent, the characteristics of Eq. (5) can be summarized as

$$v = v^{PR} + c(T) \quad (T < T_c) \quad (11)$$

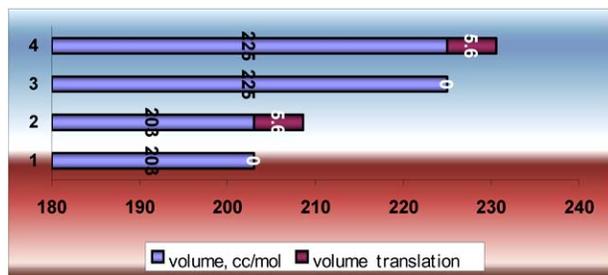
In the supercritical region:

$$d_r = d^{PR}(T_c, P) / d_c^{PR} \quad (12)$$

In Equation (12),  $d_r$  is the reduced density calculated from the PR EOS along the critical isotherm line at a given pressure  $P$  and it is used as reference for the volume translation. In this case, the  $\beta$ -function is both temperature and pressure dependent:

$$v = v^{PR} + c(T, P) \quad (T > T_c) \quad (13)$$

By the above definitions we can summary that the reference state is, in the sub critical region, along the saturated liquid density line up to critical point; in the supercritical region, along the critical isotherm density line. Therefore, the volume translation will be continuous. It is well known that a temperature dependent volume translation will not influence the calculation of equilibrium condition, as mentioned by many authors [6, 15]. The introduction of  $d_r$  as reference state can eliminate the over-crossing problem. In the supercritical region, especially along the critical isotherm line and far from the critical point, the value of  $d_r$  is positive, and its value will increase significantly with increasing pressure. This leads to the  $\beta$ -function approaching a small value at high pressures. Fitting the parameter  $c_1$  to the experimental data hereafter does not influence the results. The introduction of the reduced density in the  $\beta$ -function can significantly eliminate the over-crossing problem. At the critical point, predicted critical volume by the T-VTPR model will lead to  $v = v_c^{exp} + c$ , which is closest to the experimental critical volume. The critical volumes of propane calculated from these different volume translation models are shown in Fig. (2). From Fig. (2) it can be seen that the T-VTPR shows the smallest departure from the experimental values, while VTPR shows the largest departure.



**Fig. (2).** Comparison of different volume translations with experimental data and original PR EOS at the critical point of propane; (1) experimental; (2) T-VTPR; (3) original PR; (4) VTPR;

Following equation can be obtained at critical isotherm from the T-VTPR model:

$$v = v^{PR} + c - \Delta v_c \frac{0.35}{0.35 + 0.5(d_r - 1)} \quad (14)$$

### Viscosity Model Based on PR EOS

Based on the similarity of  $PVT$  and  $T\mu P$  relationship, the positions of  $T$  and  $P$  in the PR EOS are interchanged;  $v$  is

replaced by  $\mu$ , and gas constant  $R$  is replaced by  $r$  (defined subsequently). Eq. (1) is then transformed to the following PR viscosity equation:

$$T = \frac{rP}{\mu - b} - \frac{a}{\mu^2 + 2b\mu - b^2} \quad (15)$$

$$a = 0.45724r_c^2 P_c^2 / T_c \quad (16)$$

$$b = 0.07780r_c P_c / T_c \quad (17)$$

It is well known that the critical compressibility factor  $Z_c$  can be used to represent the gas constant  $R$  for the  $PVT$  properties with

$$R = \frac{P_c v_c}{Z_c T_c} \quad (18)$$

Therefore,  $r_c$  can be calculated from following expression:

$$r_c = \frac{T_c \mu_c}{Z_c P_c} \quad (19)$$

In this paper, the critical viscosity  $\mu_c$  is calculated from the empirical correlation proposed by Uyehara and Watson [16]:

$$\mu_c = 7.7T_c^{-1/6} M_w^{1/2} P_c^{2/3} \quad (20)$$

The temperature dependence of  $r$  can be correlated from the experimental saturated vapor pressure data at specified temperature and saturated liquid viscosities [17, 18]. In the regression, the following expression was used to represent the relationship of the saturated liquid density vs. temperature:

$$\log \mu = -A + B/T \quad (T/K) \quad (21)$$

For each substance, specific parameters  $A$  and  $B$  are fitted according to a defined temperature range, and the accuracy is usually within 1% (average relative deviation). Therefore, the saturated liquid viscosity is dependent with temperature; with the temperature known, the saturated liquid viscosity will be specified from Eq. (21). Finally, a relationship for the temperature dependence of  $r$  was obtained:

$$r = r_c [1 + m(T_r - 1)]^{-4} \quad (22)$$

The parameters  $m$  of alkanes was generalized into following expression of acentric factor  $\omega$

$$m = 1.538 + 1.40\omega - 0.554\omega^2 \quad (23)$$

Eq. (15)~(17) and (19)~(22) can be used to predict the saturated vapor pressure at specified temperatures and viscosities, or predict the viscosities of saturated liquid and gas phase at specified temperature and pressure. In solving Eq. (15), three viscosity roots were obtained at the specified temperature and pressure at the sub critical region. The correct viscosity root is chosen as follows:

- (1) In the sub critical gas region (when the pressure is lower than the saturated vapor pressure at prevailing temperature), choose the smallest real root greater than  $b$ , as  $b$  representing the asymptotic value of ideal gas viscosity at infinite temperature.
- (2) In the liquid region (when the pressure is greater than the saturated vapor pressure at prevailing temperature), choose the maximum real root. The saturated vapor pressure ( $P^s$ ) were calculated using the Antoine equation for  $P^s < 1500\text{mmHg}$ , and from the Lee-Kesler equation for  $P^s > 1500\text{mmHg}$  in the root selection procedure.
- (3) In the supercritical region (when  $T > T_c$ , as described subsequently), only a real root is obtained.

**Extension of the Viscosity Model to Supercritical Region by the Idea of Temperature Dependent Volume Translation of PR EOS**

For the extension of above viscosity model to the supercritical region, the prediction results of the proposed PR viscosity model is tested at first to predict the viscosity of supercritical fluids. The deviations of the predicted viscosities to the experimental data ( $\Delta\mu = \mu^{PR1} - \mu^{exp}$ ) of n-butane were plotted in Fig. (3). From Fig. (3) it can be seen that  $\Delta\mu$  shows a linear behavior with the increase of pressure. Therefore, a linear term with form  $c_o + c_1 P_r$  is chosen for the correction of  $\Delta\mu$ :

$$\mu = \mu^{PR1} - \Delta\mu_c (Q + s(1 + Q)P_r) \tag{24}$$

where

$$\Delta\mu_c = \mu_c^{PR1} - \mu_c^{exp} \tag{25}$$

Where  $\mu_c^{exp}$  is calculated from Eq. (20). To avoid a over-crossing problem for the viscosity in the supercritical region, a reference state is chosen as following:

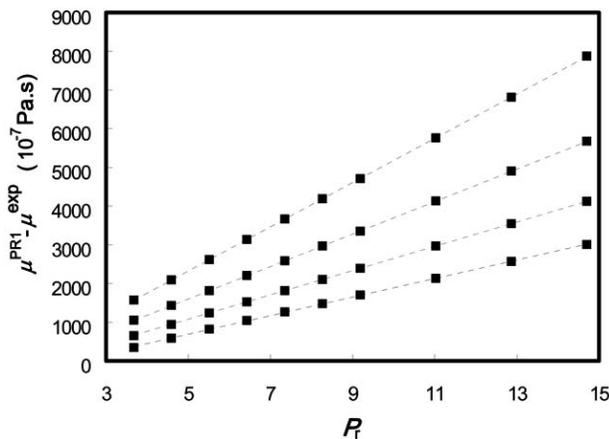


Fig. (3). Deviations of the predicted viscosities to the experimental data of n-butane.

$$s = \delta \exp\left(\sum_{i=1}^3 Q_i (\mu_r^{-1} - 1)^i\right) \tag{26}$$

Where for methane the parameter  $\delta$  equals to  $-1$ , and for other substances  $\delta$  equals to  $1$ . The reduced viscosity  $\mu_r$  is a reference state chosen for the viscosity translation. It is defined as

$$\mu_r = \mu^{PR1}(T, P_c) / \mu_c^{PR1} \tag{27}$$

The above proposed PR viscosity model and its extension to supercritical region are hereafter named as PR1 model.

**RESULTS AND DISCUSSIONS**

Table 1 and Table 2 list the calculated results for the saturated liquid densities and supercritical PVT based on the PR, VTPR and the proposed T-VTPR models and the comparison to the experimental data [19] for selected substances. From Table 1 it can be seen that the deviations of the calculated saturation liquid densities from experimental data are remarkable reduced by the new model than the VTPR and PR EOS. Fig. (4) shows the detailed calculation results for the different models and a comparison with experimental data for the saturated liquid densities of propane and hexane. From Fig. (4) is obvious that both the VTPR and PR EOS do not provide satisfactory prediction in the near critical region, while the T-VTPR model can provide a better description of the experimental data. Table 2 shows the deviations of the calculated results to the experimental supercritical PVT data [19] of carbon dioxide, methane, ethane and propane are 2.28% by the T-VTPR model, 2.73% by the VTPR and 2.81% by the PR EOS. Considering these data cover a large temperature and pressure ranges and are taken from many different sources [19], the new model provides results that are close to the average experimental error distribution. Fig. (5) show the detailed calculation results of supercritical PVT behavior of propane based on these three models. From these Figures, it is obvious that the original PR EOS gives over-predictions for the fluid density at the supercritical region, and the VTPR model gives better prediction results. The deviations of the VTPR model to the experimental data can be minimized using the T-VTPR model, and the over-crossing problem can be avoided with the application of the proposed volume translation method.

The proposed generalized PR1 viscosity equation, Eqs. (15)~(17) and (19)~(22), have been applied to calculate the saturated liquid viscosities of pure hydrocarbons. The results calculated by the PR1 model and the original PR viscosity model are summarized in Table 3. Fig. (6) shows a comparison of the predicted saturated viscosities based on the PR1 and original PR viscosity model for undecane. From Table 3 and Fig. (6) we can see that the new model has achieved a significant improvement to the original PR viscosity model.

The predicted saturated vapor pressure for 23 pure substances based on the PR1 model at specified temperatures and viscosities were compared with the experimental data [17, 18], the results are summarized in Table 4. Satisfactory results have been achieved with an average relative deviation (AAD%) equals to 3.01%. The original PR viscosity model was not used to do parallel calculation because the saturated vapor pressure data was not employed for its parameter fitting.

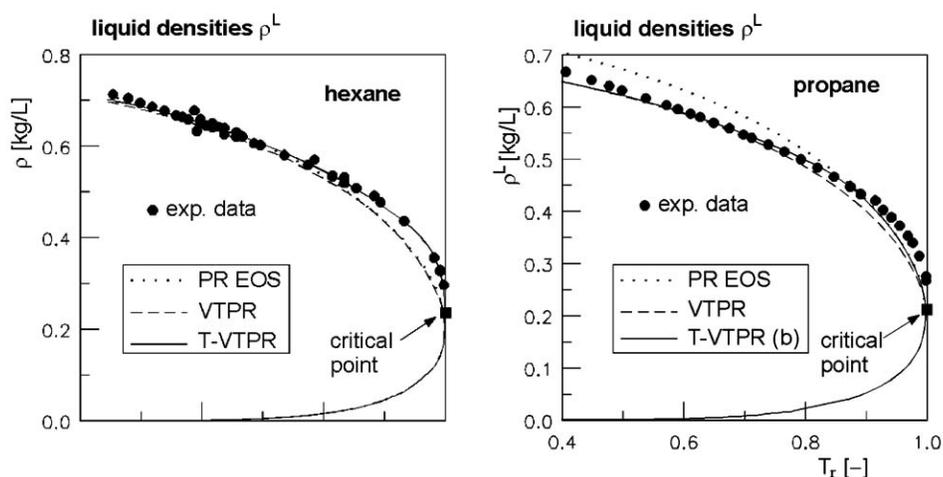
The predicted results for the viscosities of supercritical fluids based the viscosity translated PR1 model and the original PR viscosity model of Guo *et al.* are summarized in

**Table 1. Derivations in Calculated Saturation Densities Based on Different Volume Translation Models for Peng-Robinson Equation of State**

Component	Data Points	T-Range (K)	$v_c$ (Cm <sup>3</sup> /mol)	$c_1$	AAD%		
					T-VTPR	VTPR	PR
Carbon dioxide	34	217.00-304.10	94.0	5.50	2.64	7.07	6.00
Methane	103	91.00-190.53	99.0	17.4	1.74	1.98	9.65
Ethane	55	100.00-305.33	148.0	2.08	1.51	2.46	6.86
Propane	92	100.08-598.37	203.0	8.20	1.63	3.49	4.12
n-Butane	49	135.08-410.86	255.0	4.71	0.62	0.89	4.06
n-Pentane	126	153.15-439.76	304.0	14.8	1.72	1.81	2.42
n-Hexane	323	183.15-506.15	370.0	2.71	0.68	1.72	1.30
n-Heptane	305	183.15-523.15	432.0	4.00	0.64	2.34	1.41
n-Octane	244	223.15-568.77	492.0	3.60	1.08	3.93	4.27
n-Nonane	112	223.15-423.15	548.0	5.50	0.72	1.95	5.11
n-Decane	167	243.15-598.15	603.0	8.66	0.45	0.77	3.49
Dodecane	149	263.15-643.15	713.0	9.63	0.64	0.89	2.53
Cyclohexane	282	279.91-473.15	308.0	7.00	0.29	0.30	5.22
Water	92	288.15-593.15	56.0	3.26	0.49	3.97	17.98
Overall	2133				1.06	2.40	5.32

**Table 2. Comparison of Calculated PVT Properties of Supercritical Fluids Based on Different Volume Translation Models for Peng-Robinson Equation of State**

Component	Data Points	T-Range (K)	P-Range (MPa)	$c_1$	AAD%		
					T-VTPR	VTPR	PR
Carbon dioxide	993	304.65-513.15	0.0990-250.0	-12.0	3.04	3.40	3.50
Methane	2490	193.00-673.15	0.0850-579.5	-10.0	1.39	1.44	2.50
Ethane	353	305.63-623.15	0.1013-48.00	65.1	2.32	3.58	2.49
Propane	505	369.85-548.15	0.1013-68.95	-51.6	2.39	2.50	2.75
Overall	4341				2.28	2.73	2.81

**Fig. (4).** Calculated results from PR, VTPR and T-VTPR models for saturated liquid densities and the comparison with experimental data of propane and hexane.

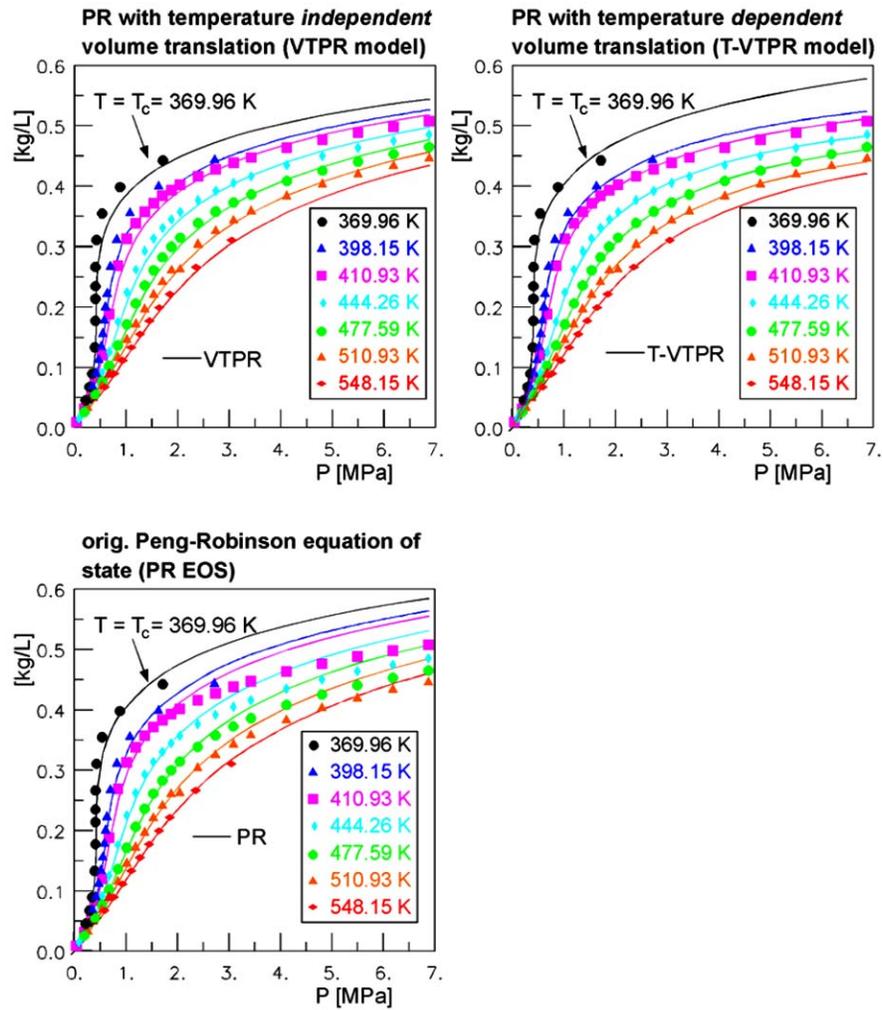


Fig. (5). Comparison of calculated results with experimental supercritical volumetric data of propane based on PR EOS, VTPR and T-VTPR models. Experimental: (+) 369.96K (critical temperature); (□) 398.15K; (□) 410.93K; (Δ) 444.26K; (O) 477.59K; (×) 510.93K; (◇) 548.15K.

Table 3. Prediction Results for the Saturated Liquid Viscosities of Pure Hydrocarbons Based on PR and PR1 Viscosity Models

Substance	T-range (K)	Data Points	PR AAD%	PR1 AAD%
Methane	133.4~190.6	5	81.80	5.79
Ethane	213.8~305.4	18	78.62	4.31
Propane	258.9~369.8	20	72.20	4.37
n-Butane	297.6~425.2	26	69.63	5.65
n-Pentane	328.8~469.8	14	73.27	5.33
n-Hexane	355.2~507.4	14	78.21	2.86
n-Heptane	378.2~540.5	13	16.25	10.62
n-Octane	398.2~568.8	16	18.50	1.10
n-Nonane	416.2~594.8	36	14.32	5.13
n-Decane	432.6~617.9	37	16.21	4.78
Undecane	446.8~638.2	35	17.60	3.05
Dodecane	461.2~658.9	37	19.50	4.24

Table 3. contd....

Substance	T-range (K)	Data Points	PR AAD%	PR1 AAD%
Tridecane	473.2~676.0	35	21.30	4.85
Tetradecane	484.2~691.2	38	22.30	4.14
Pentadecane	494.9~707.1	36	19.60	6.68
Hexadecane	505.6~722.4	5	8.88	4.98
Heptadecane	515.2~736.0	36	19.56	10.23
Octadecane	523.4~747.8	34	16.26	2.24
Overall		455	17.52	5.02

Data sources (Lu, 1982; Ma, 1993).

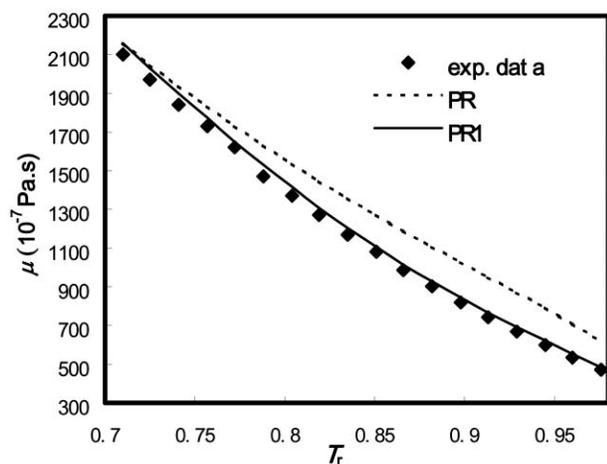


Fig. (6). Comparison of the predicted saturated viscosities based on the PR1 and original PR viscosity model for undecane.

Table 5. The specified parameters obtained from the viscosity translation are listed in Table 6. Fig. (7) shows the

comparisons of the predicted supercritical viscosities based on the PR1 and original PR viscosity model for n-butane. From Table 5 it can be seen that although the viscosity data of API [20] cover a large range of temperature and pressure, the calculated results from the new model and the original PR viscosity model are both satisfactory, better results have been achieved from the new model.

## CONCLUSIONS

A temperature-dependent volume translation model is proposed for Peng-Robinson equation of state, which is applicable both in the near and far from the critical point, as well as in the sub- and supercritical regions. Satisfactory correlation results for saturation density and *PVT* behavior of light hydrocarbons and carbon dioxide are obtained. A new viscosity model is also proposed in this paper based on the Peng-Robinson equation of state and the similarity between the *PVT* and *TμP* relationship. Using a viscosity translation method based on the idea of volume translation for the PR EOS, the viscosity model has been applied to calculate the saturation vapor pressures and viscosities of hydrocarbon liquids and gases.

Table 4. Prediction Results for the Saturated Vapor Pressure of Pure Substances Based on the PR1 Viscosity Model\*

Substance	T-range (K)	Data Points	AAD%
Carbon dioxide	212.9~304.2	33	1.88
Nitrogen	88.4~126.2	38	0.59
Water	453.2~647.4	29	1.58
Methane	133.4~190.6	57	3.16
Ethane	213.8~305.4	33	2.50
Propane	258.9~369.8	42	1.64
n-Butane	297.6~425.2	27	2.08
n-Pentane	328.8~469.8	30	1.99
n-Hexane	355.2~507.4	42	2.00
n-Heptane	378.2~540.5	28	1.40
n-Nonane	416.2~594.8	19	2.88
n-Decane	432.6~617.9	44	3.46
Undecane	446.8~638.2	38	4.21
Dodecane	461.2~658.9	37	2.87

Table 4. contd....

Substance	T-range (K)	Data Points	AAD%
Tridecane	473.2~676.0	20	3.50
Tetradecane	484.2~691.2	18	4.21
Pentadecane	494.9~707.1	18	5.00
Hexadecane	505.6~722.4	20	5.22
Heptadecane	515.2~736.0	20	6.40
Octadecane	523.4~747.8	20	6.98
i-Butane	286.2~408.8	37	1.09
i-Petane	322.3~460.4	18	1.46
Overall		693	3.01

\*Data Sources (Lu, 1982; Ma, 1993)

Table 5. The Predicted Results for the Viscosities of Supercritical Fluids Based the Viscosity Translated PR1 and the Original PR Viscosity Models\*

Substance T-range (K)	P-range (10 <sup>5</sup> Pa)	Data Points	PR AAD%	PR1 AAD%
Methane 344~477	137.8~689.1	60	5.55	4.44
Ethane 310~427	206.7~689.1	51	2.64	3.04
Propane 377~510	206.7~689.1	58	4.26	2.93
n-Butane 427~510	137.8~689.1	49	3.74	2.70
n-Pentane 477~510	68.9~551.3	36	3.57	2.67

\*Data source (Lee, 1965).

Table 6. Specified Parameters of PR1 Viscosity Model Obtained from the Viscosity Translation in the Supercritical Region

Substance	Q <sub>1</sub>	Q <sub>2</sub>	Q <sub>3</sub>	Q
Methane	3.10	0.26	-0.48	-4.44
Ethane	-0.76	1.27	-0.42	-18.83
Propane	-0.60	0.83	-0.28	-14.69
n-Butane	0.086	0.21	-0.14	-12.26

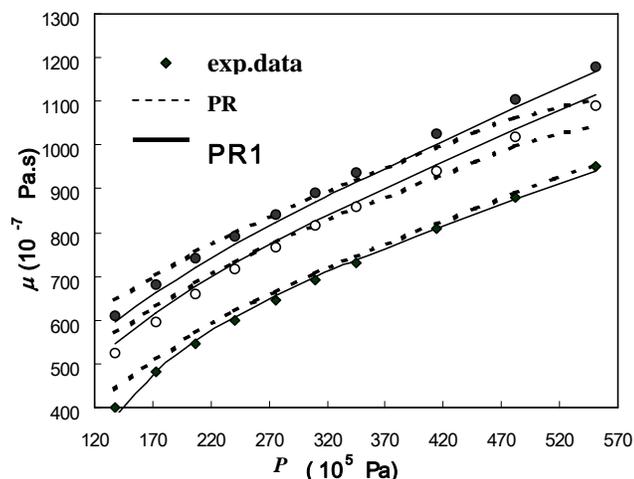


Fig. (7). Comparison of the predicted supercritical viscosities based on the PR1 and original PR viscosity model for n-butane. (•) 444.26K; (o) 460.93K; (■) 494.26K.

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**APPENDIX A**

PR viscosity equation:

$$T = \frac{rP}{\mu - b_1} - \frac{a}{\mu(\mu + b) + b(\mu - b)} \quad (\text{A1})$$

In which  $a$  and  $b$  are calculated from Eq. (16) and (17), and

$$r = r_c \cdot \left[ 1 + k_1 \left( \sqrt{T_r P_r} - 1 \right) \right]^{-2} \quad (\text{A2})$$

$$b_1 = b \cdot \left( \exp \left[ k_2 \left( \sqrt{T_r} - 1 \right) \right] + k_3 \left( \sqrt{P_r} - 1 \right)^2 \right) \quad (\text{A3})$$

In which  $r_c$  is calculated from Eq. (19), and parameters  $k_1, k_2, k_3$  are calculated from the following generalized expressions:

$\omega < 0.3$

$$k_1 = 0.829599 + 0.350857\omega - 0.747680\omega^2$$

$$k_2 = 1.94546 - 3.19777\omega + 2.80193\omega^2$$

$$k_3 = 0.299757 + 2.20855\omega - 6.64959\omega^2$$

$\omega \geq 0.3$

$$k_1 = 0.956763 + 0.192829\omega - 0.303189\omega^2$$

$$k_2 = -0.258789 - 37.1071\omega + 20.5510\omega^2$$

$$k_3 = 5.16307 - 12.8207\omega + 11.0109\omega^2$$

**NOMENCLATURE**

$\alpha$	=	Energy parameter of the PR equation of state
$A, B$	=	Specified parameters in Eq. (21)
$b$	=	Volumetric parameter of the PR equation of state
$c$	=	Temperature independent volume correction for the PR equation of state
$c_1$	=	Volume translation parameter of the T-VTPR model
$d$	=	Density (mol/liter)
$L, M, N$	=	Two-Bluck-Cunningham-Coon alpha function parameters
$m$	=	Parameter in Eq. (22)
$M_w$	=	Molar mass in Eq. (20)
$P$	=	Pressure (bar)
$Q$	=	Parameter in Eq. (24)
$r$	=	Generalized viscosity constant in Eq. (15)
$R$	=	Generalized gas constant
$s$	=	Parameter in Eq. (24)
$T$	=	Absolute temperature (K)
$v$	=	Molar volume (liter/mol)
$Z$	=	Compressibility factor

**Greek letter** $\alpha$  = Temperature dependent function $\beta$  = Temperature dependent function $\delta$  = Parameter of viscosity translation $\mu$  = Viscosity ( $10^{-7}$ Pa.s) $\omega$  = Acentric factor $\rho$  = Density (kg/liter)**Subscript** $c$  = Critical property $r$  = Reduced property

sat = Saturation state

**Superscript**

s = Saturated state

exp = Experimental

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