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Vapor−Liquid Equilibria for Propane (R-290) + 1,1,1,2,3,3,3-Heptafluoropropane (HFC-227ea) at Various Temperatures

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Isothermal vapor−liquid equilibrium data for propane + 1,1,1,2,3,3,3-heptafluoropropane (HFC-227ea) were measured at (273.15, 283.15, 293.15, 303.15, 313.15, and 323.15) K, respectively. The experiments were carried out using a circulation-type equilibrium apparatus to measure temperature, pressure, and the compositions of the liquid and vapor phases. The experimental data were well correlated with the Peng−Robinson equation of state (PR-EOS) using the Wong−Sandler mixing rules and the Carnahan−Starling−De Santis equation of state (CSD-EOS). This system shows strong positive azeotrope.

Introduction

In recent years, the utilization of light hydrocarbons such as propane, butane, and propylene is believed to be a good solution for developing CFC-alternative refrigerants in the future because these are inexpensive, plentiful, and environmentally benign chemicals. However, they are toxic, flammable, and explosive. Other CFC alternative refrigerants are HFCs which have good performance and are more environmentally benign, but HFCs' GWP (global warming potential) is a thousand higher than that of CFCs. One of the solutions to compensate their weakness is mixing two alternatives. Mixing light hydrocarbons and HFCs can provide a good solution, making up for the weaknesses.

Vapor−liquid equilibrium (VLE) data are the most important data requested to evaluate the performance of refrigeration cycles and to determine their optimal compositions. Azeotropic mixtures have merit because their behaviors are similar to pure compounds. However, very few experimental data have been reported previously in the literature.[1−3]

In this work, isothermal VLE data for the binary mixture of propane + HFC-227ea at six equally spaced temperatures between (273.15 and 323.15) K were measured using a circulation-type equilibrium apparatus in which both phases were continuously recirculated. The experimental data were correlated with the Peng−Robinson equation of state (PR-EOS)[4] using the Wong−Sandler mixing rule[5] combined with the NRTL excess Gibbs free energy model and the Carnahan−Starling−De Santis equation of state (CSD-EOS).[6] In the range of experimental temperatures, the average absolute deviations of pressure and vapor-phase compositions between experimental and calculated values were determined, and the relevant parameters were presented. Almost all the calculated values with this model give good agreement with the experimental data, and thus the system exhibits an azeotrope.

Experimental

Chemicals. High-quality chemicals of HFC-227ea and propane were used for VLE measurement. HFC-227ea was supplied by DuPont (USA) with a guaranteed purity in mass fraction higher than 0.998, and R290 (propane) was supplied by MG Industry (UK) with a guaranteed purity in mass fraction higher than 0.995. Furthermore, we also analyzed these two pure components with a gas chromatograph. The resulted mass fraction purities of HFC-227ea and R290 were higher than 0.999 and 0.9985, respectively. So, they were used without any further purification.

Experimental Apparatus. A schematic diagram of the experimental apparatus for measuring the VLE is given in Figure 1. Because the vapor−liquid equilibrium apparatus used in this work was the same as that used in our previous work,[7−13] only a brief description is given here. It was a circulation-type in which both liquid and vapor phases were recirculated continuously. The equilibrium cell was a type-316 stainless steel vessel with an inner volume of about 85 cm^3. In its middle part, two Pyrex glass windows of 20 mm thickness were installed in the front and back so that the liquid level, mixing and circulating behaviors, and critical phenomena could be observed by a back light during operation. A stirrer, rotated at variable speeds by an external magnet, was used to accelerate the attainment of the equilibrium state and to reduce concentration gradients in both phases.

The temperature of the equilibrium cell in the water bath was maintained by a circulator from Jeio Tech, Korea. The temperature in the cell was measured with a platinum-resistance sensor and a digital temperature indicator (model F250MkII) precision thermometer from Automatic Systems Laboratories, Ltd. (United Kingdom). They were calibrated by the National Measurement Accreditation Service accredited calibration laboratory. The total uncertainty in temperature measurements is estimated to be within 0.01 K, including sensor uncertainty, 0.001 K, temperature resolution, 0.001 K, and measurement uncertainty, 0.001 K. The pressure was measured with a pressure transducer (model XPM60) and a digital pressure calibrator (model PC106), from Beamax, Finland. Pressure calibrations are traceable to National Standards (Center for Metrology and Accreditation Cert. No. M-95P077, 14.11.1995, M-M 730, 16.11.1995, and M-95P078,

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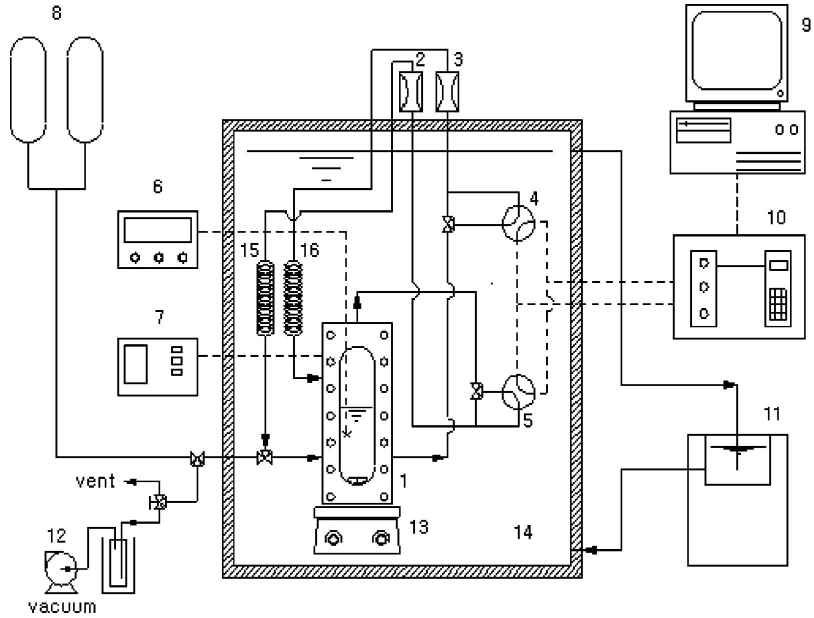

Figure 1. Schematic diagram of the experimental apparatus: 1, equilibrium cell; 2, vapor circulation pump; 3, liquid circulation pump; 4, liquid sample valve; 5, vapor sample valve; 6, temperature indicator; 7, pressure indicator; 8, sample reservoir; 9, computer; 10, gas chromatograph; 11, circulator; 12, vacuum pump; 13, magnetic stirrer; 14, constant temperature water bath; 15, vapor phase heat exchanger; 16, liquid phase heat exchanger.

16.11.1995), and the calibrator uncertainty was 0.0005 MPa; the sensor uncertainty was 0.0001 MPa; and the measurement uncertainty was 0.0001 MPa. Therefore, the total uncertainty of the pressure measurement is estimated to be within 0.001 MPa.

The vapor and liquid phases in the equilibrium cell were continuously recirculated by a dual-head circulation pump (Milton Roy Company) to reach the equilibrium state rapidly in the cell. The composition of the phases was determined by means of a gas chromatograph (Gow-Mac model 550P) connected online to the VLE cell. The response of the thermal conductivity detector (TCD) was carefully calibrated using the mixture prepared gravimetrically and the gas chromatographs (GC) with a Porapak Q column from Alltech Company. Data derived from gas chromatography were treated with a computer program (Autochro-WIN from Young-Lin Instrument Co., Ltd.). The uncertainty for the GC measurement and calibration is estimated to be within 0.001 in area fraction.

Experimental Procedures. Experiments to measure VLE data for the binary system of propane (1) + HFC-227ea (2) at various temperatures were performed by the following procedures. The system was first evacuated to remove all inert gases. A certain amount of HFC-227ea was supplied to the cell, and then the temperature of the entire system was held constant by controlling the temperature of the water bath. After the desired temperature was attained, the pressure of the pure component was measured. A certain amount of propane was introduced into the cell from a sample reservoir. The mixture in the cell was stirred continuously with the magnetic stirrer for over 1 h. Both the vapor and liquid phases were recirculated by the dual-head circulation pump until an equilibrium state was established. It is believed that 1 h or more was sufficient to obtain thermal equilibrium between the cell fluid and the thermostatic bath as well as the vapor and liquid phases. After equilibration, the pressure in the equilibrium cell was measured, and then vapor and liquid samples were withdrawn from the recycling lines by the vapor and liquid sampling valves, respectively. The compositions of the samples were measured by immediately injecting them into the GC, which was connected online to the vapor and liquid sampling valves. The GC was calibrated with pure components of known purity and with mixtures of known composition that were prepared gravimetrically. At least five analyses were performed for each phase, and the average values were considered to correspond to the equilibrium values. Considering the margin of error and the reproducibility of the GC, we generally estimated an overall uncertainty in the measurements of the composition of 0.002 in mole fraction for both the liquid and the vapor phases.

Table 1. Characteristic Properties of the Chemicals (Critical Temperature T_c, Critical Pressure P_c, Acentric Factor w) from Database REFPROP 6.01[18]

component	chemical formula	molecular weight	T_c/K	P_c/MPa	ω
propane (1)	C_3H_8	44.10	369.85	4.248	0.1524
HFC-227ea (2)	CF_3CHCF_3	170.03	375.95	2.980	0.3632

Correlation

In this work, the experimental VLE data were correlated with the Peng−Robinson equation of state (PR-EOS)[4] and the Carnahan−Starling−De Santis equation of state (CSD-EOS).[6]

PR-EOS.

$$P = \frac{RT}{v_M - b} - \frac{a(T)}{v_M(v_M + b) + b(v_M - b)} \tag{1}$$

with

$$a(T) = \left(0.457235\frac{R^2T_c^{\,2}}{P_c}\right)\alpha(T) \tag{2}$$

$$b = 0.077796\frac{RT_c}{P_c} \tag{3}$$

$$\alpha(T) = [1 + k(1 - \sqrt{T/T_c})]^2 \tag{4}$$

$$k = 0.37464 + 1.54226\omega - 0.26992\omega^2 \tag{5}$$

Table 2. Coefficients (Temperature Dependent α and β) of Pure Components for the CSD-EOS[17]

	α_0	$\alpha_1 \cdot 10^3$	$\alpha_2 \cdot 10^6$	β_0	$\beta_1 \cdot 10^4$	$\beta_2 \cdot 10^7$
component	$kJ \cdot m^3 \cdot kg^{-1} \cdot mol^{-2}$	$kJ \cdot m^3 \cdot kg^{-1} \cdot mol^{-2} \cdot K^{-1}$	$kJ \cdot m^3 \cdot kg^{-1} \cdot mol^{-2} \cdot K^{-2}$	$m^3 \cdot kg^{-1} \cdot mol^{-1}$	$m^3 \cdot kg^{-1} \cdot mol^{-1} \cdot K^{-1}$	$m^3 \cdot kg^{-1} \cdot mol^{-1} \cdot K^{-2}$
propane	$2.988 \cdot 10^3$	−2.629	−1.097	0.1430	−1.765	−0.5785
HFC-227ea	$5.988 \cdot 10^3$	−3.438	−0.995	0.2190	−2.940	−0.5619

Table 3. Comparison of the Measured Pure Component Vapor Pressures *P*, with Reference Vapor Pressures P_{REF} From the Database REFPROP 6.01[18]

T/K	*P*/MPa	P_{REF}/MPa	ΔP^a/MPa	$\lvert \Delta P/P \rvert^b$
		Propane		
273.15	0.476	0.474	0.002	0.004
283.15	0.632	0.636	−0.004	0.006
293.15	0.833	0.836	−0.003	0.004
303.15	1.079	1.079	0.000	0.000
313.15	1.366	1.369	−0.003	0.002
323.15	1.709	1.713	−0.004	0.002
average value				0.003
		HFC-227ea		
273.15	0.195	0.193	0.002	0.010
283.15	0.280	0.278	0.002	0.007
293.15	0.389	0.389	0.000	0.000
303.15	0.527	0.530	−0.003	0.006
313.15	0.704	0.706	−0.002	0.003
323.15	0.920	0.922	−0.002	0.002
average value				0.005

[a] $\Delta P = P - P_{REF}$. [b] $|\Delta P/P| = (|P - P_{calcd}|)/P$.

where the parameter *a* is a function of temperature; *b* is constant; *k* is a constant characteristic of each substance; *ω* is the acentric factor; *P* (MPa) is the pressure; P_c (MPa) is the critical pressure; *T* (K) is the absolute temperature; T_c (K) is the critical temperature; T_r is the reduced temperature; and υ_M is the molar volume.

The Wong−Sandler mixing rules[5] were used in this work to obtain EOS parameters for a mixture from those of the pure components. These mixing rules for a cubic equation of state can be written as

$$b_m = \frac{\sum_i \sum_j x_i x_j (b - a/RT)_{ij}}{(1 - A_\infty^E/CRT - \sum_i x_i a_i/RTb_i)} \tag{6}$$

with

$$(b - a/RT)_{ij} = \frac{1}{2}[(b - a/RT)_i + (b - a/RT)_j](1 - k_{ij}) \tag{7}$$

and

$$\frac{a_m}{b_m} = \sum_i x_i \frac{a_i}{b_i} + \frac{A_\infty^E}{C} \tag{8}$$

where *C* is a numerical constant equal to $\ln(\sqrt{2} - 1)/\sqrt{2}$ for the PR-EOS used in this work. The single adjustable parameter (k_{ij}) for each binary pair is referred to as the Wong−Sandler mixing rule parameter. Also, A_∞^E is an excess Helmholtz free-energy model at infinite pressure that can be equated to a low-pressure excess Gibbs energy model.[14] In this study, we used the nonrandom two-liquid (NRTL) model[15] given by

$$\frac{A_\infty^E}{RT} = \sum_i x_i \frac{\sum_j x_j G_{ji} \tau_{ji}}{\sum_r x_r G_{ri}} \tag{9}$$

Table 4. Vapor−Liquid Equilibrium Measurements for the Propane (1) + HFC-227ea (2) System at (273.15 to 323.15) K

Experimental Data					
P/MPa	x_1	y_1	*P*/MPa	x_1	y_1
T = 273.15 K			*T* = 283.15 K		
0.195	0.000	0.000	0.280	0.000	0.000
0.320	0.115	0.406	0.396	0.081	0.315
0.384	0.205	0.523	0.504	0.199	0.460
0.428	0.306	0.599	0.572	0.305	0.552
0.459	0.406	0.644	0.622	0.423	0.611
0.479	0.498	0.698	0.648	0.512	0.662
0.496	0.598	0.731	0.664	0.608	0.703
0.505	0.664	0.752	0.673	0.662	0.727
0.511	0.731	0.779	0.684	0.733	0.766
0.514	0.813	0.820	0.687	0.763	0.782
0.510	0.866	0.846	0.689	0.805	0.806
0.509	0.907	0.877	0.688	0.851	0.836
0.496	0.966	0.943	0.673	0.942	0.916
0.476	1.000	1.000	0.632	1.000	1.000
T = 293.15 K			*T* = 303.15 K		
0.389	0.000	0.000	0.527	0.000	0.000
0.564	0.109	0.332	0.696	0.081	0.244
0.689	0.225	0.488	0.855	0.195	0.423
0.776	0.344	0.579	0.983	0.327	0.541
0.835	0.469	0.646	1.074	0.479	0.631
0.862	0.562	0.688	1.117	0.591	0.687
0.885	0.665	0.731	1.145	0.725	0.754
0.892	0.752	0.775	1.150	0.797	0.791
0.895	0.776	0.789	1.151	0.841	0.828
0.893	0.815	0.812	1.136	0.900	0.877
0.892	0.858	0.841	1.079	1.000	1.000
0.875	0.929	0.904			
0.833	1.000	1.000			
T = 313.15 K			*T* = 323.15 K		
0.704	0.000	0.000	0.920	0.000	0.000
0.823	0.046	0.140	1.197	0.109	0.249
0.908	0.086	0.231	1.351	0.187	0.366
1.004	0.142	0.324	1.495	0.276	0.440
1.105	0.202	0.399	1.607	0.368	0.504
1.240	0.315	0.501	1.687	0.449	0.572
1.364	0.477	0.607	1.785	0.611	0.654
1.412	0.532	0.637	1.800	0.629	0.680
1.446	0.579	0.663	1.830	0.757	0.762
1.478	0.706	0.761	1.837	0.800	0.773
1.481	0.796	0.801	1.826	0.807	0.807
1.459	0.899	0.872	1.816	0.913	0.884
1.424	0.970	0.954	1.709	1.000	1.000
1.366	1.000	1.000			

with

$$G_{ji} = \exp(-\alpha_{ji}\tau_{ji}) \text{ and } \tau_{ji} = (g_{ji} - g_{ii})/(RT) \tag{10}$$

where

$$A_{ij} = (g_{ij} - g_{jj}) \tag{11}$$

The critical temperature (T_c), critical pressure (P_c), and acentric factor (*ω*) for both HFC-227ea and propane that were used to calculate the parameters for the PR-EOS are provided in Table 1. We have set the nonrandomness parameter, α_{ij}, equal to a fixed value of 0.3 for all of the binary mixtures studied here.

CSD-EOS. The CSD-EOS is given by

$$\frac{P\upsilon_M}{RT} = \frac{1 + y + y^2 - y^3}{(1 - y)^3} - \frac{a}{RT(\upsilon_M + b)} \tag{12}$$

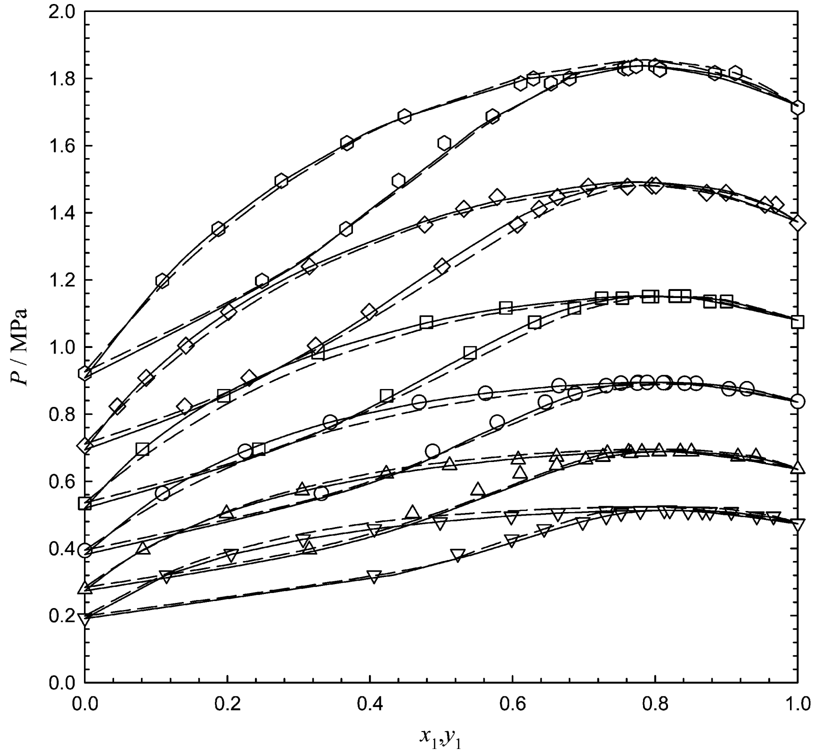

Figure 2. $P-x-y$ diagram for the propane (1) + HFC-227ea (2) system. Experimental data at: ▽, 273.15 K; △, 283.15 K; ○, 293.15 K; □, 303.15 K; ◇, 313.15 K; and ⬡, 323.15 K. −, Calculated with the PR-EOS using W−S mixing; ---, calculated with the CSD-EOS.

Table 5. Optimal Values of the Binary Parameters k_{12} and f_{12}, Adjustable Parameters for the NRTL Model τ_{21} and τ_{12}, and Absolute Average Relative Deviations of P and Absolute Average Deviation of y

	PR-EOS					CSD-EOS		
T/K	k_{12}[a]	t_{21}[b]	τ_{12}[b]	AARD-P[c]	AAD-y[d]	f_{12}[a]	AARD-P[c]	AAD-y[d]
273.15	−0.0349	2.4929	5.4426	0.003	0.008	0.131	0.023	0.013
283.15	0.0017	2.2917	5.0456	0.004	0.014	0.126	0.012	0.016
293.15	0.2258	1.5390	2.5811	0.003	0.007	0.111	0.015	0.005
303.15	0.1141	2.2813	3.0849	0.005	0.005	0.113	0.013	0.005
313.15	−0.0160	2.5326	5.0622	0.005	0.008	0.141	0.009	0.006
323.15	0.2036	1.8516	2.1418	0.004	0.008	0.124	0.008	0.009
average values				0.004	0.008		0.013	0.009

[a] k_{12} and f_{12} are interaction parameters. [b] $\tau_{ij} = A_{ij}/RT$. The unit of A_{ij} and A_{ji} is kJ·mol^{-1}. [c] AARD-$P = 1/N \sum|(P_{\text{exptl}} - P_{\text{calcd}})/P_{\text{exptl}}|$. [d] AAD-$y = 1/N \sum|y_{\text{exptl}} - y_{\text{calcd}}|$.

Where

$$y = \frac{b}{4v_{\text{M}}} \tag{13}$$

and v_{M} is the molar volume. For a pure component, the temperature dependence of a and b is represented by the following forms

$$a = \alpha_0 \exp(\alpha_1 T + \alpha_2 T^2) \tag{14}$$

$$b = \beta_0 + \beta_1 T + \beta_2 T^2 \tag{15}$$

The CSD-EOS using the approach of Morrison and McLinden[16] was adopted to correlate the VLE data. The coefficients of α_0, α_1, and α_2 in eq 14 and β_0, β_1, and β_2 in eq 15 were cited from REFPROP 5.0[17] and listed in Table 2.

In the application of the CSD-EOS for a mixture, there exist the effective molecular parameters a_{m} and b_{m} (the subscript m refers to the mixture) defined by using the following mixing rules

$$a_{\text{m}} = \sum_{i=1}^{n}\sum_{j=1}^{n} x_i x_j a_{ij} \tag{16}$$

$$b_{\text{m}} = \sum_{i=1}^{n}\sum_{j=1}^{n} x_i x_j b_{ij} \tag{17}$$

where n is the number of components in the mixture.

When $i = j$, the values of a_{ii} and b_{ii} are those of a and b for the pure components which are determined by eqs 14 and 15. The values of a_{ij} and b_{ij} can be obtained if nearly any experimental property of the mixture is known. For the binary systems, the values of a_{12} and b_{12} can be expressed in the following forms

$$a_{12} = (1 - f_{12})(a_1 a_2)^{1/2} \tag{18}$$

$$b_{12} = \frac{1}{8}(b_1^{1/3} + b_2^{1/3})^3 \tag{19}$$

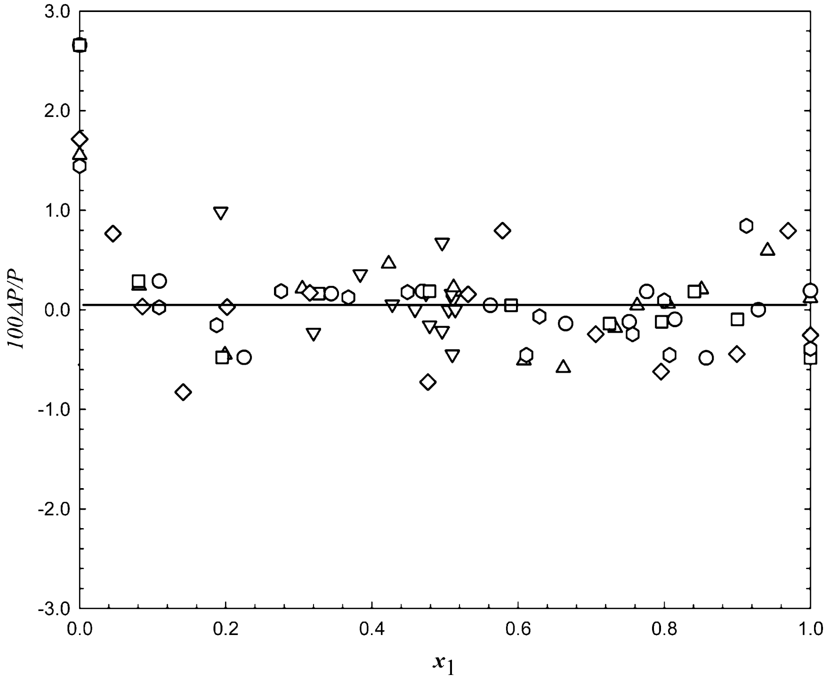

Figure 3. Deviation of pressure for the system propane (1) + HFC-227ea (2) from the PR-EOS using the W−S mixing rule at: ▽, 273.15 K; △, 283.15 K; ○, 293.15 K; □, 303.15 K; ◇, 313.15 K; and ⬡, 323.15 K.

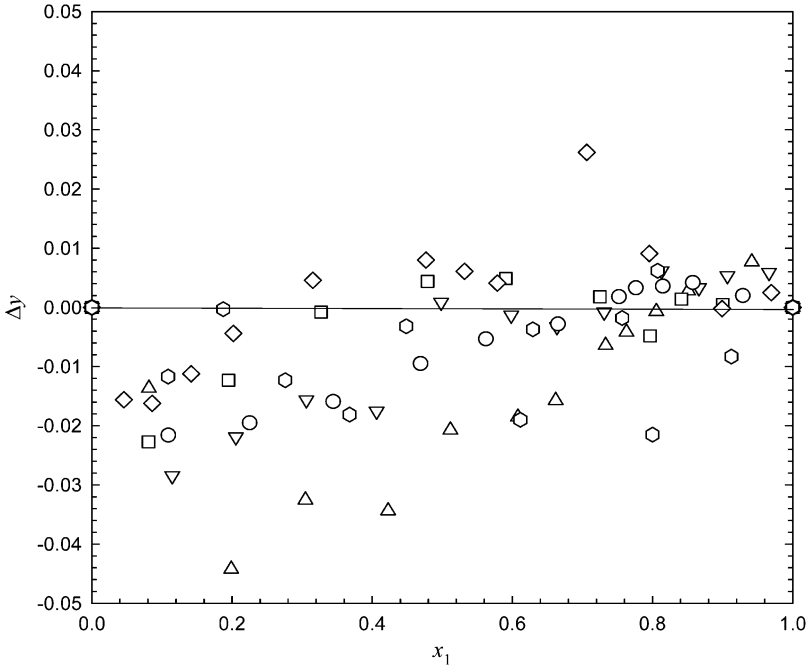

Figure 4. Deviation of vapor composition for the system propane (1) + HFC-227ea (2) from the PR-EOS using the W−S mixing rule at: ▽, 273.15 K; △, 283.15 K; ○, 293.15 K; □, 303.15 K; ◇, 313.15 K; and ⬡, 323.15 K.

where f_{12} is the interaction parameter and a_i and b_i are properties of the pure components.

The parameters of both PR-EOS and CSD-EOS were obtained by minimizing the following objective function

$$\text{objective function} = \frac{1}{N}\sum_{j}^{N}\left[\left(\frac{P_{j,\text{exptl}} - P_{j,\text{calcd}}}{P_{j,\text{exptl}}}\right)\right]^2 \qquad (20)$$

Table 6. Azeotropic Pressure P_{az} and Azeotropic Liquid Mole Fraction x_{az} for Propane (1) + HFC-227ea (2) as a Function of Temperature

T/K	azeotropic pressure P_{az}/MPa	azeotropic composition x_{az} (propane mole fraction)
273.15	0.514	0.841
283.15	0.689	0.806
293.15	0.895	0.802
303.15	1.150	0.784
313.15	1.488	0.798
323.15	1.838	0.770

Results and Discussion

Table 3 shows the comparison of measured vapor pressures of pure propane and HFC-227ea with those calculated from the database REFPROP 6.01,[18] which is considered to be reliable for the pure compounds considered and consistent with other literature data. As can be seen in this table, the deviations of

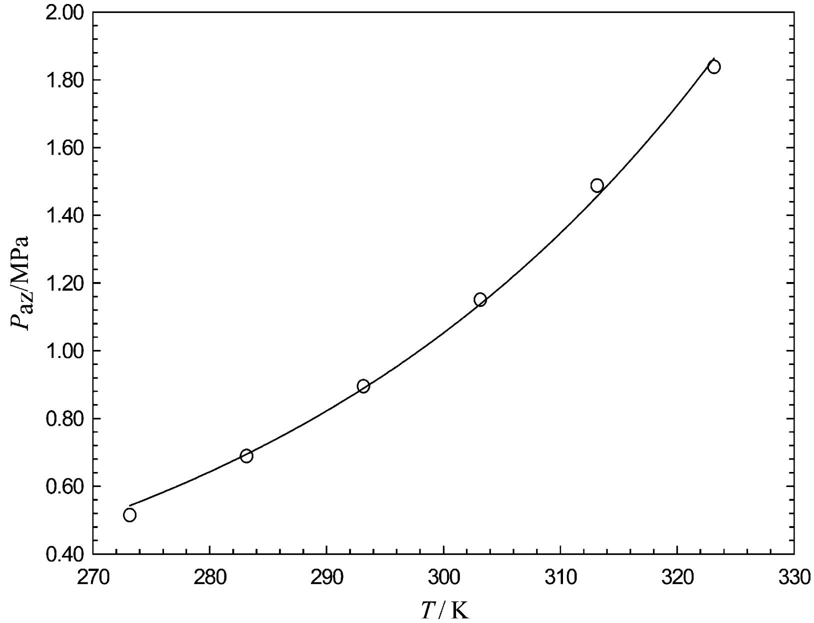

Figure 5. Variation of the azeotropic pressure with temperature for propane (1) + HFC-227ea (2). The equation of the fitting line is $P_{\mathrm{az}} = 2.506 \cdot 10^{-4}(T/\mathrm{K})^2 - 0.123(T/\mathrm{K}) + 15.405$ (323.15 K ≥ T ≥ 273.15 K).

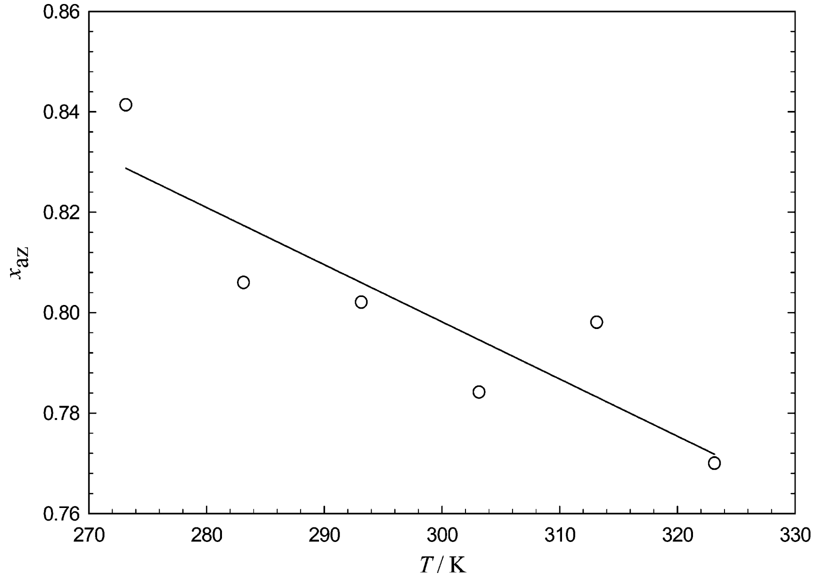

Figure 6. Variation of the azeotropic composition with temperature for propane (1) + HFC-227ea (2). The equation of the fitting line is $x_{\mathrm{az}} = -1.114 \cdot 10^{-4}\ T/\mathrm{K} + 1.140$ (323.15 K ≥ T ≥ 273.15 K).

vapor pressure (ΔP) between measured and calculated values from REFPROP 6.01[18] are within 0.005 MPa. The average values of absolute relative deviations of P ($|\Delta P/P|$) between measured and calculated values were 0.001 for propane and 0.003 for HFC-227ea. The experimental data of isothermal VLE for the binary system of propane (1) + HFC-227ea (2) are listed in Table 4. These tables list the measured mole fraction of the liquid and vapor phases, pressures, and temperatures in equilibrium and the deviations between measured and calculated pressures (ΔP) and vapor compositions (Δy), point by point.

Figure 2 shows the comparison of measured and calculated values with the PR-EOS and CSD-EOS for the binary system of propane (1) + HFC-227ea (2) at various temperatures. The experimental VLE data at (273.15, 283.15, 293.15, 303.15, 313.15, and 323.15) K are shown as reverse triangles, triangles, circles, squares, diamonds, and hexagons, respectively. The black solid lines represent the calculated data by the PR-EOS, and the black dashed lines represent the calculated data by the CSD-EOS.

The interaction parameters of the binary mixtures for each isotherm, the binary parameters of the NRTL model with the Wong−Sandler mixing rules, and the average absolute relative deviations of pressure (AARD-P) and average absolute deviations of vapor-phase composition (AAD-y) between measured and calculated values are reported in Table 5. The average values of AARD-P and AAD-y for all the six temperatures were 0.004 and 0.008 for the PR-EOS and 0.013 and 0.009 for the CSD-EOS, respectively. Both the average absolute deviations of pressure and vapor-phase compositions of PR-EOS are less than those of CSD-EOS. So, only the results of error estimations with the PR-EOS for P and y were shown in Figures 3 and 4, respectively. From these figures and the low average deviations of P and y, we conclude that both the PR-EOS and the CSD-EOS could be used for predicting VLE data for propane (1) +

HFC-227ea, and the PR-EOS is more or less better than the CSD-EOS for these systems.

This mixture exhibited an azeotrope. Azeotropic data of this mixture have been determined at the compositions between 0.77 and 0.85 in mole fraction and at pressures between (0.51 and 1.84) MPa in the temperature range from (273.15 to 323.15) K. Azeotropic compositions and pressures are shown in Table 6 and Figures 5 and 6. As can be seen in these figures, azeotropic compositions and pressures are dependent on the temperature. In the range of experimental temperatures, the azeotropic compositions were correlated by the empirical equation, $x_{az} = -1.139 \cdot 10^{-3}\ T/\mathrm{K} + 1.140$, and the azeotropic pressure was correlated by the empirical equation, $P_{az}\ /\mathrm{MPa} = 1.541 \cdot 10^{-0.123}\ T/\mathrm{K} + 2.506 \cdot 10^{-4}\ T^2/\mathrm{K}^2$. These equations were fitted within the temperature range $T \geq 273.15$ K.

Conclusions

VLE data for binary systems of propane (1) + HFC-227ea (2) were measured at five equally spaced temperatures between (273.15 and 323.15) K using a circulation-type equilibrium apparatus. The experimental data were correlated with the PR-EOS using the Wong−Sandler mixing rules and the CSD-EOS. Calculated results with these equations have given satisfactory results in comparison with the experimental data. This system shows strong positive azeotropes for the full temperature range studied here. According to the results of this work, the PR-EOS gave better correlation results than the CSD-EOS for propane (1) + HFC-227ea (2).

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