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LOSS OF METHANOL AND MONOETHYLENE GLYCOL IN VLE AND LLE: PREDICTION OF PARTITION COEFFICIENT

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Abstract. *Phase equilibria of water-alcohol-hydrocarbons are important when it comes to flow assurance issues in the petroleum industry. Thermodynamic inhibitors (usually alcohols and glycols, such as methanol and monoethylene glycol) change the thermodynamic equilibrium itself, thus avoiding the hydrate zone. Accurate results for the loss of the volatile inhibitor in the gas or condensate phase (partition) are of extreme importance for the oil and gas industry. In this work, a flash algorithm using the Cubic-Plus-Association (CPA) equation of state was developed to estimate the partition of each component (water, hydrate-forming gas and inhibitor) in any phase over a wide range of temperatures and pressures in vapor-liquid equilibrium (VLE) and liquid-liquid equilibrium (LLE). Different temperature-dependent functions were optimized and evaluated for the CPA binary interaction parameters. The flash was applied to several systems with water, methane, ethane, propane, carbon dioxide, methanol and monoethylene glycol (MEG). The results were then compared with experimental data available in the literature. The loss of methanol to the gas and or condensate phases was satisfactorily predicted. Yet, the CPA underestimated the loss of MEG to the gas phase in a gas mixture containing carbon dioxide. The average absolute deviation for the predicted loss of methanol and monoethylene glycol ranged between 3-45 %.*

Keywords: *vapor-liquid equilibrium, liquid-liquid equilibrium, CPA, hydrate inhibitor, partition.*

1. INTRODUCTION

The formation of gas hydrates is one of the main problems faced by the flow assurance professionals in the oil and gas industry. Natural gas hydrates are crystalline compounds formed by water and gas molecules. Under high pressures and low temperatures, those water molecules rearrange themselves to form a solid structure that encapsulates a guest molecule, mainly light hydrocarbons, such as methane and ethane. These crystals can agglomerate and form plugs inside the pipeline, damaging equipment, impairing the production or even stopping it completely, thus causing economic and safety issues. As the world demand for energy is in constant growth, the oil industry continuously seeks new horizons. The offshore exploration in ultra-deep waters has been expanding over the last years and, in this scenario, the formation of hydrates became a major concern because production lines can often operate within the envelope of hydrate formation (Sloan and Koh, 2008).

The most common method used by the industry to avoid the hydrate formation is the injection of additives known as thermodynamic inhibitors, usually alcohols, glycols and salts. Methanol and monoethylene glycol are the most widely used in the oil and gas industry. These inhibitors shift the hydrate formation envelope to regions of higher pressures and lower temperatures, thus preventing their formation in the production lines. Methanol and MEG are often injected at rates higher than the actually necessary ones due to uncertainties in the required dosage. Methanol is relatively expensive and usually not recoverable; moreover, a large quantity of inhibitor may be required to suppress the hydrate formation temperature, thus making inhibition operations costly. MEG is used in large amounts, it is also expensive and its regeneration is required. These facts increase the operational and capital costs as well as the demand for land or deck surface area, especially on offshore installations. Thus, optimizing the amount of inhibitor injected, and keeping it to a minimum, decreases environmental and operational issues, reducing costs as well (Kontogeorgis and Folas, 2010).

Motivated by the scarcity of studies addressing this subject, a flash algorithm was developed and implemented in this work. The model utilizes the CPA equation of state (EoS) and the isofugacity condition to estimate both vapor-liquid equilibrium (VLE) and liquid-liquid equilibrium (LLE) of ternary and quaternary systems containing hydrate-

forming gases, water and thermodynamic inhibitors (methanol and MEG). In addition to the most common hydrocarbons present in natural gases (methane, ethane and propane), systems composed by carbon dioxide, an acid gas capable of forming hydrates and commonly found in natural gas reservoirs were also implemented. Additionally, different kinds of functions for the CPA binary interaction parameters in the phase equilibrium predictions were evaluated and optimized.

2. THERMODYNAMIC MODELING

For an isothermal system, it is possible to ensure the chemical equilibrium through an equality of the fugacities of each component ($\hat{f}_i$) in the present phases (Smith et al., 2004):

$$\hat{f}_i^{\alpha} = \hat{f}_i^{\beta} = \hat{f}_i^{\gamma} \tag{1}$$

The isofugacity criterion was implemented in the modeling presented in this work. The fugacity of a component can be obtained through an EoS and in this study the CPA-EoS was used. The CPA-EoS was developed by Kontogeorgis et al. (2006) with the aim of improving the prediction efficacy of cubic EoS in systems containing polar/hydrogen-bonding compounds by adding to these equations an association term based on Wertheim's first order perturbation theory. The modeling of complex multicomponent systems containing components capable of performing association is a difficult task, and often a cubic EoS cannot present satisfactory results. Complex associating systems are important in many practical cases and several of them are of interest to the oil and gas industry, especially systems composed of methanol and glycols, mostly because of their extensive use as hydrate inhibitors (Smith et al., 2004). The CPA-EoS (Kontogeorgis et al., 2006) can then be expressed in terms of pressure as:

$$P = \frac{RT}{v_m - b} - \frac{a}{v_m\left(v_m + b\right)} - \frac{1}{2}\frac{RT}{v_m}\left(1 + \frac{1}{v_m}\frac{\partial \ln\left(g\right)}{\partial\left(1/v_m\right)}\right)\sum_i x_i \sum_{A_i}\left(1 - X_{A_i}\right) \tag{2}$$

where a and b are two SRK parameters, the temperature-dependent energy parameter and the co-volume, respectively. T is the temperature, R is the universal constant of the gases, v_m is the molar volume, g is the radial distribution function, X_{Ai} is the mole fraction of molecules i that are not bonded through the active site A, and x_i is the mole fraction of the component i.

3. RESULTS AND DISCUSSIONS

The flash algorithm presented in section 3 was applied to various systems containing hydrate-forming gases, water and thermodynamic inhibitors over a large range of temperatures and pressures.

Table 1 presents a summary of the results obtained in this work. As one can see, the systems are composed mainly by hydrocarbons (methane and ethane), and associative compounds (water, methanol and MEG). It is also possible to observe the presence of systems containing carbon dioxide, a hydrate-forming acid gas frequently present in the composition of natural gases, and which is capable of performing cross-association with the other associative compounds, thus making it an interesting gas to evaluate the modeling capacity of an associative equation of state. The average absolute deviation (AAD) and the experimental references are also shown. As it can be observed, mostly of the cases analyzed in this study fall into the vapor-liquid equilibrium region (VLE). The only exception is the carbon dioxide, water and methanol system, where also appears a liquid-liquid equilibrium region composed of a liquid phase rich in carbon dioxide and another one rich in water and methanol. In the case of a vapor-liquid equilibrium, x refers to the molar fraction in the liquid phase and thus y is the molar fraction in the vapor phase. For the liquid-liquid equilibrium, y was also used to represent the molar fraction in the condensed gas-rich phase.

Table 1. Summary of the phase equilibrium conditions and AAD for the systems investigated in this work.

System	Components	Region	T-Range [K]	P-Range [MPa]	Number of Points	% AAD		References
						x	y	
I	H_2O	VLE	280.25 - 313.45	5.14 - 13.12	9	0.03 %	24.14 %	Frost et al. (2014)
	CH_4					37.72 %	0.02 %	
	CH_3OH					0.89 %	3.42 %	
II	H_2O	VLE	278.15	5	20	0.20 %	9.27 %	Folas et al.

	CH_4		- 323.19	- 20		21.03 %	0.72 %	(2007), Kruger et al. (2018)
	$C_2H_6O_2$					0.20 %	16.52 %	
III	H_2O	VLE and LLE	243.15 - 298.15	0.16 - 9.25	22	5.91 %	31.66 %	Chang et al. (1985), Ng and Robinson (1983)
	CO_2					21.35 %	0.04 %	
	CH_3OH					6.02 %	16.19 %	
IV	H_2O	VLE	243.15 - 298.15	0.16 - 9.25	3	0.35 %	43.43 %	Ng et al. (1985)
	C_2H_6					33.03 %	5.45 %	
	CO_2					10.66 %	26.82 %	
	CH_3OH					1.85 %	19.42 %	
V	H_2O	VLE	284.15	6.89	1	0.26 %	48.88 %	Ng and Chen (1995)
	CH_4					48.54 %	1.88 %	
	CO_2					41.54 %	16.71 %	
	CH_3OH					0.31 %	11.78 %	
VI	H_2O	VLE	283.15 - 310.95	20.70 - 6.81	6	-	25.16 %	Ng and Chen (1995)
	CH_4					-	-	
	C_3H_8					-	-	
	$C_2H_6O_2$					-	14.31 %	
VII	H_2O	VLE	283.15 - 310.95	20.70 - 6.81	6	-	29.07 %	Ng and Chen (1995)
	CH_4					-	-	
	CO_2					-	-	
	$C_2H_6O_2$					-	45.02 %	
Overall					67	10.28 %	13.00 %	

Analyzing the results more carefully, it is possible to note that the highest deviations were obtained in the calculation of the gas solubility in the liquid phase (water + thermodynamic inhibitor) and in the molar fraction of water in the vapor phase. Some of the systems herein described will be presented in detail, focusing on the partition of the thermodynamic inhibitors in the vapor phase due to its importance and wide use in the oil and gas industry as a method of preventing the hydrate formation.

System I from Table 1 is composed of 0.250 mol % of methane (1), 0.463 mol % of water (2), and 0.287 mol % of methanol (3). Results obtained for this system are shown in Fig. 1, being the experimental data taken from the work of Frost et al. (2014).

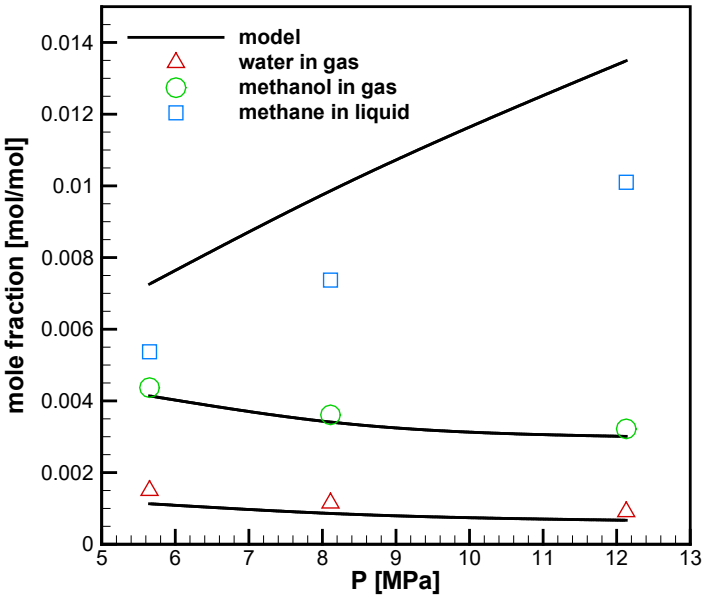

Figure 1 Experimental and predicted VLE for a water, methane and methanol system at 313.15 K (Frost et al., 2014).

Figure 1 shows the solubilities of methane in the liquid polar phase and that of the water and methanol in the gas phase at 313.15 K. All these molar fractions have the same order of magnitude, and the solubility of methane in the solution of methanol and water is greater than that presented by methanol and water in the vapor phase. These magnitudes demonstrate the difficulty in accurately modeling and measuring these values experimentally. According to Frost et al. (2014), it is not surprising that different research groups in similar conditions display distinct measured data for the water content due to the well-known challenges associated with measuring trace amounts of water in gases, and it is safe to assume that the same difficulties apply to the other compounds. As expected, water exhibited a lower gas molar fraction in comparison to methanol due to its lower volatility. In this case, methane solubility exhibited the highest AAD (34.2 %) whereas the lowest AAD was presented by methanol (5.84 %). Deviation calculated for the molar fraction of water was 25.57 %.

In order to give emphasis to the precision of the model in predicting the solubility of methanol in the gas phase, Fig. 2 demonstrates the calculated molar fraction for the three different temperatures (280.25, 298.77 and 313.15 K). One can observe that the modeling performed accurately with an average AAD of 3.41 % and, as expected, showed that the molar fraction increases with the temperature. The highest (5.84 %) and the lowest (1.83 %) deviations were presented by the highest and the lowest temperatures, respectively.

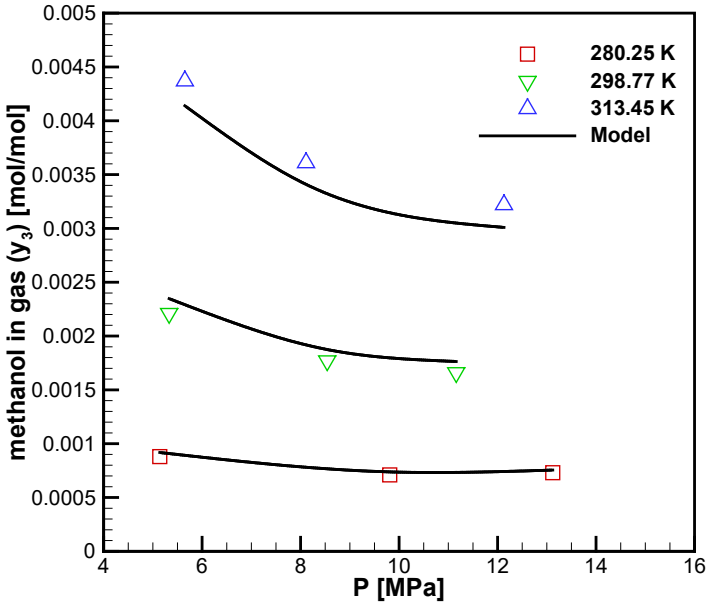

Figure 2. Experimental and predicted mole fraction of methanol in the gas phase in VLE for a water, methane and methanol system at several temperatures (Frost et al., 2014).

According to Kontogeorgis and Folas (2010), the influence of the binary interaction parameters, k_{ij}, on the prediction of the alcohol partition for methanol-water-alkane systems revealed that the correct representation of only two binary systems, alkane-methanol and water-methanol, is crucial for a satisfactory correlation of the ternary systems. In accordance with Table 1 in this work, the three binary systems were considered for system I: methane-methanol, water-methanol and water-methane.

As presented in the parameters estimation section, three different temperature-dependent functions were evaluated for the binary interaction parameters. In order to demonstrate the influence of these different approaches Fig. 3 and Fig. 4 show a comparison of the results obtained using each expression. In Figure 3, it is possible to analyze the influence of the binary interaction parameters in the solubility of methanol in the gas phase for system I. As it can be observed, the discrepancy among the results provided by the different functions is not high. The linear temperature-dependent binary interaction parameter provided better results for all temperatures with an AAD of 1.95 %. In general, the quadratic approach presented the second best performance with an AAD of 3.42 %. The only temperature for which the quadratic approach did not perform better than the inverse approach was the highest one (313.45 K). The AAD calculated for the inverse approach was 4.07 %. It is also important to emphasize that even with the linear function presenting better results for the partitioning of methanol into the vapor phase for system I, in general taking all the systems investigated in this work (I to VII) into account, the quadratic expression performed better. For this reason, the quadratic expression was selected to provide the results exhibited in this work.

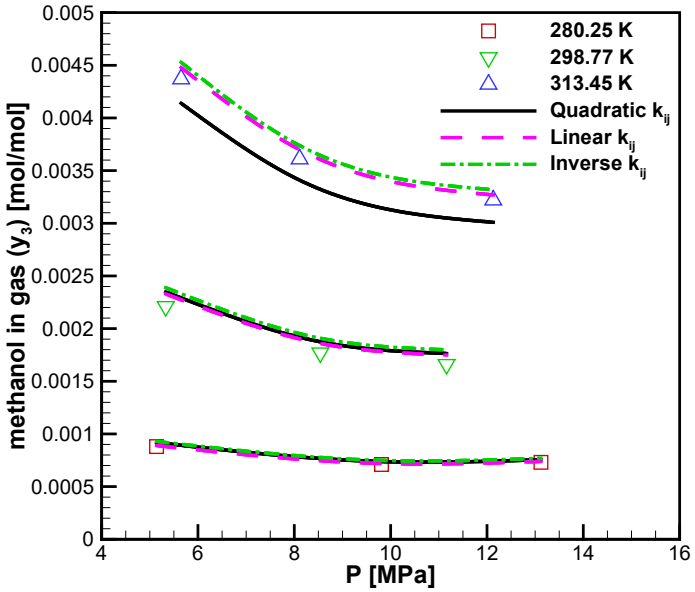

Figure 3. Experimental and predicted mole fraction of methanol in the gas phase in VLE for a water, methane and methanol system at various temperatures using different functions for binary interaction parameters (Frost et al., 2014).

Fig. 4 presents the estimated values of methane solubility for system I considering the three different types of binary interaction parameter. On analyzing this plot, the influence and importance of the binary interaction parameters in the calculation of equilibrium conditions become more evident. There is a significant deviation in the results displayed by the linear and inverse functions compared to the quadratic approach, a fact not observed in Fig. 3. The AAD exhibited by the quadratic binary interaction parameter was 34.2 %, while the linear and inverse approach presented ADD of 43.04 % and 43.74 %, respectively.

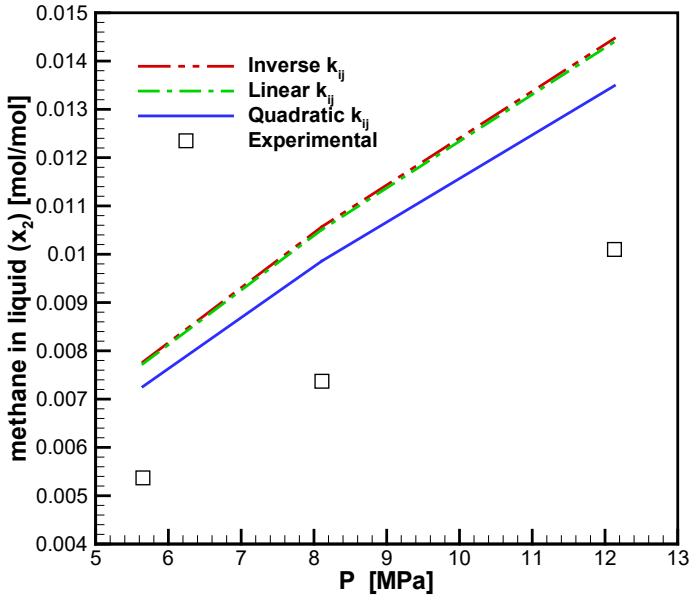

Figure 4. Experimental and predicted mole fraction of methane in the liquid phase in VLE for a water, methane and methanol system at several temperatures using different functions the for binary interaction parameters (Frost et al., 2014).

System II from Table 1 presents two different sources of experimental data. The data taken from Folas et al. (2007) is composed of 0.1925 mol % of methane (1), 0.6220 mol % of water (2) and 0.1855 mol % of MEG (3). In Fig. 5, water and MEG content in the gas phase for a temperature of 278.15 K is shown. Concentration of water in the gas phase is higher than that of MEG, in contrast to the observed in Fig. 2 for methanol. The fact that methanol is more volatile than MEG causes a greater loss of inhibitor to the gas phase, making it more difficult to recover. For that

reason, in a gas-dominant system, MEG is often preferred over methanol. AAD calculated for the molar fraction of MEG was 16.74 % and 3.49 % for water.

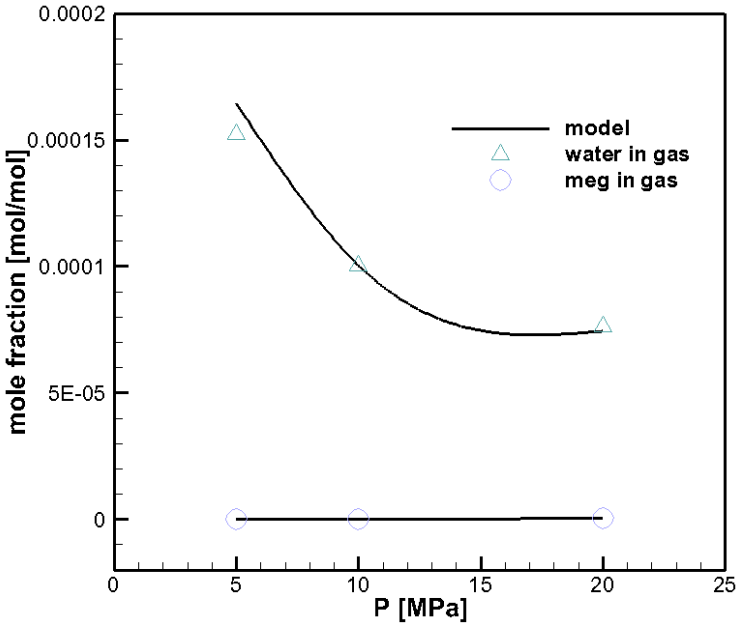

Figure 5. Experimental and predicted VLE for a water, methane and MEG system at 278.15 K (Folas et al., 2007).

In system II, the feed used by Kruger et al. (2018) ranged from 0.31-0.48 mol % for methane, 0.02-0.18 mol % for water and 0.41-0.65 mol % for MEG. Figure 6 evaluates the effect of temperature and pressure on the water solubility in the vapor phase. It can be observed that the vapor water content increases roughly exponentially with the temperature increase. A decrease of approximately 40 % in water solubility in gas phase is observed with the increase of the pressure from 6 MPa to 12.5 MPa. The CPA model provided a good description of the data with an AAD of 7.83 %.

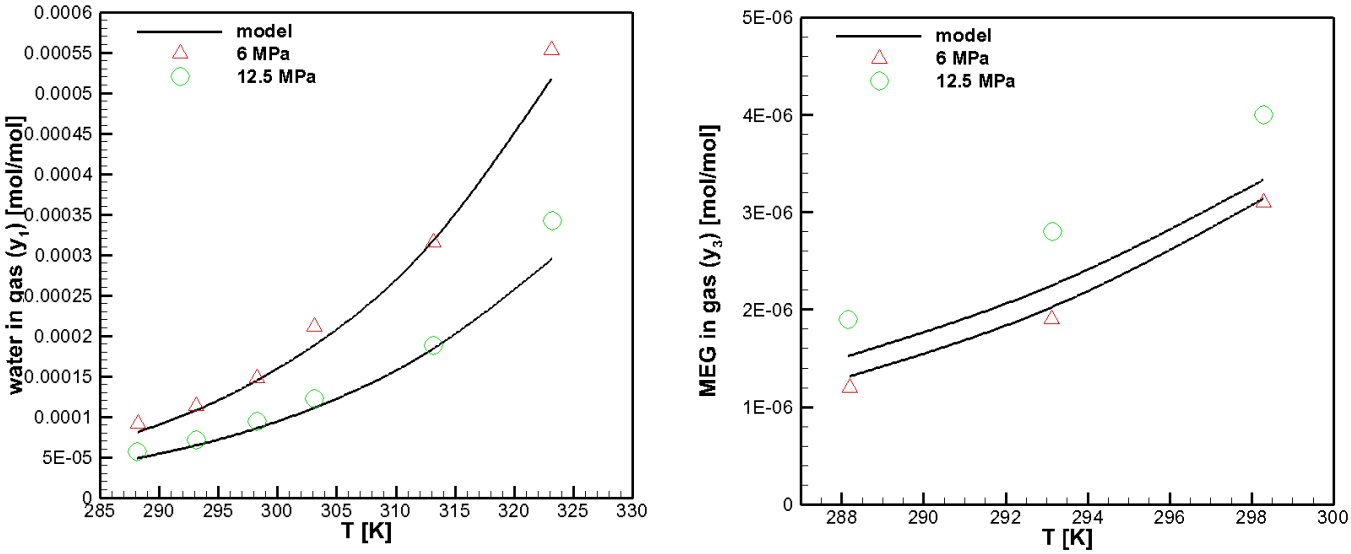

Figure 6. Experimental and predicted mole fraction of water and MEG in the gas phase in VLE for a water, methane, and MEG system at various temperatures and pressures (Kruger et al., 2018).

Once again, turning the attention to the partition of the inhibitor in the gas phase, Fig. 6 also presents the MEG content in the vapor phase for different temperatures and pressures. The model presented a less accurate performance for points with higher pressures. The CPA model presented a satisfactory qualitative prediction of the experimental data for the lower pressure (AAD = 5.83 %). The absolute average deviation calculated for the molar fraction of MEG in the higher pressure was 18.69 %.

Although CPA can fairly represent the partition of MEG for this system, models like SRK (using van der Waals mixing rules) usually overestimate the amount of glycols in the gas phase by almost an order of magnitude. Oppositely, SRK with advanced mixing rules (Huron-Vidal types using a modified NRTL activity coefficient) can provide

reasonable prediction of the methanol partition as well. However, more interaction parameters are required in NRTL to represent the temperature dependence of phase behavior (Kontogeorgis and Folas 2010).

There are a few articles available in the literature that present a comparison between the CPA EoS and the SRK EoS with the Huron-Vidal (HV) mixing rule (Boesen et al. 2017, Folas et al. 2006 and 2007). Folas et al. (2007) concluded that for multicomponent systems both models provided satisfactory predictions of water and MEG solubility in the gas phase, however the CPA EoS demonstrated a superior predictive performance even using only one binary and temperature-independent interaction parameter.

It is known that low temperatures and high pressures are usually required for hydrate formation and in certain conditions some of the hydrate-forming gases may liquefy. The following results address exactly this scenario. Tables 2 and 3 show the liquid-liquid equilibrium (LLE) data for a system (III from Table 1) containing carbon dioxide, methanol and water. Experimental points shown in Tab. 2 allow one to observe that the molar fraction of water in the condensed carbon dioxide-rich phase is smaller than the molar fraction of methanol. The solubility of water in the condensate showed the highest deviations, followed by the molar fraction of carbon dioxide in the aqueous phase.

Table 2. Experimental and predicted composition of the aqueous solution in LLE of a water, carbon dioxide and methanol system (Ng and Robinson 1983).

	Temperature [K]	Pressure [MPa]	Carbon Dioxide in aqueous solution	% AAD	Methanol in aqueous solution	% AAD	Water in aqueous solution	% AAD
Exp.	278.50	8.72	0.0315	21.30	0.0553	1.50	0.9132	0.83
Model			0.0382		0.0561		0.9057	
Exp.	270.75	9.25	0.0360	42.42	0.1120	4.26	0.8520	2.35
Model			0.0513		0.1168		0.8320	

Table 3. Experimental and predicted composition of the condensate gas phase in LLE of a water, carbon dioxide and methanol system (Ng and Robinson 1983).

	Temperature [K]	Pressure [MPa]	Carbon Dioxide in condensate	% AAD	Methanol in condensate x 10^3	% AAD	Water in condensate x 10^3	% AAD
Exp.	278.50	8.72	0.9956	0.06	2.3900	16.41	2.0600	45.70
Model			0.9950		1.9978		3.0014	
Exp.	270.75	9.25	0.9936	0.05	4.7500	27.31	1.6200	48.49
Model			0.9941		3.4528		2.4055	

All the results presented so far were related to ternary systems. The following system (IV from Table 1) to be analyzed is composed by a quaternary mixture containing 0.0168 mol % of ethane (1), 0.0056 mol % of carbon dioxide (2), 0.8571 mol % of water (3) and 0.1205 mol % of methanol (4). Experimental data were taken from Ng et al. 1985 and the comparison with the thermodynamic modeling are shown in Tables 4 and 5.

Table 4. Experimental and predicted molar composition of the liquid phase in VLE of a water, ethane, carbon dioxide and methanol system (Ng et al. 1985).

	Temperature [K]	Pressure [MPa]	Ethane in liquid x 10^3	% AAD	Carbon Dioxide in liquid x 10^3	% AAD	Methanol in liquid	% AAD	Water in liquid	% AAD
Exp.	270.93	1.14	0.9800	36.00	1.7700	11.50	0.1299	5.41	0.8673	0.79
Model			1.3328		1.5664		0.1229		0.8742	
Exp.	275.76	2.70	1.6300	45.75	3.8100	0.88	0.1226	0.09	0.8719	0.07
Model			2.3757		3.8435		0.1225		0.8713	

Exp.	280.85	3.60	1.8500	17.33	10.2100	19.59	0.1217	0.13	0.8661	0.19
Model			2.1706		8.2102		0.1219		0.8678	

Table 4 shows molar concentrations in the liquid phase. Ethane solubility exhibited the highest deviations, while carbon dioxide solubility presented the lower deviations. The molecules of carbon dioxide can associate with the molecules of the solvent (water and methanol), making its solubility greater than that of a hydrocarbon. This ability was taken into account during the calculation. The solubility of carbon dioxide in water is, in some cases, two orders (or more) of magnitude higher than that of the ethane. This is a physical indication of the importance of solvation for such systems. Solvation implies that the carbon dioxide does not self-associate but is able to solvate with water and methanol.

Table 5. Experimental and predicted molar composition of the gas phase in VLE of a water, ethane, carbon dioxide, methanol system (Ng et al. 1985).

	Temperature [K]	Pressure [MPa]	Ethane in gas	% AAD	Carbon Dioxide in gas	% AAD	Methanol in gas x 10^3	% AAD	Water in gas x 10^3	% AAD
Exp.	270.93	1.14	0.8748	4.14	0.1239	29.06	0.6200	14.70	0.7200	39.63
Model			0.9109		0.0879		0.7111		0.4346	
Exp.	275.76	2.70	0.8265	7.39	0.1729	35.41	0.5700	17.39	0.4100	35.29
Model			0.8876		0.1116		0.4709		0.2653	
Exp.	280.85	3.60	0.7708	4.84	0.2267	15.99	1.5600	26.17	0.7800	55.35
Model			0.8081		0.1904		1.1517		0.3482	

On the other hand, Table 5 shows the gas phase. Once again, the highest deviations are related to the mole fraction of water in the vapor phase, which is even smaller than the solubility of ethane in the liquid phase. Results obtained for methanol in the gas phase presented a better agreement with experimental data.

System VI is composed of 0.3007 mol % of methane (1), 0.0334 mol % of propane (2), 0.5161 mol % of water (3) and 0.1498 mol % of MEG (4). Experimental data were obtained from Ng and Chen (1995). Figure 7 shows the water content in the gas phase for 283.15 K and 310.95 K, and the MEG content in the gas phase for 310.95 K. AAD calculated for the molar fraction of water was 25.16 %. AAD calculated for molar fraction of MEG was 14.31 %.

Figure 8 shows the system VII composed of 0.3007 mol % of methane (1), 0.0334 mol % of carbon dioxide (2), 0.5161 mol % of water (3) and 0.1498 mol % of MEG (4). AAD calculated for molar fraction of water was 29.07 %. AAD calculated for molar fraction of MEG was 45.02 %. MEG concentration in system VII is bigger than in system VI, in a likely demonstration that the existence of carbon dioxide increases the loss of MEG to gas phase. The cross-association between the carbon dioxide and MEG might be the reason for this increase of the solubility of MEG in the gas phase. However, it is worthwhile to point out that the CPA underpredicts the MEG concentration in the system VII.

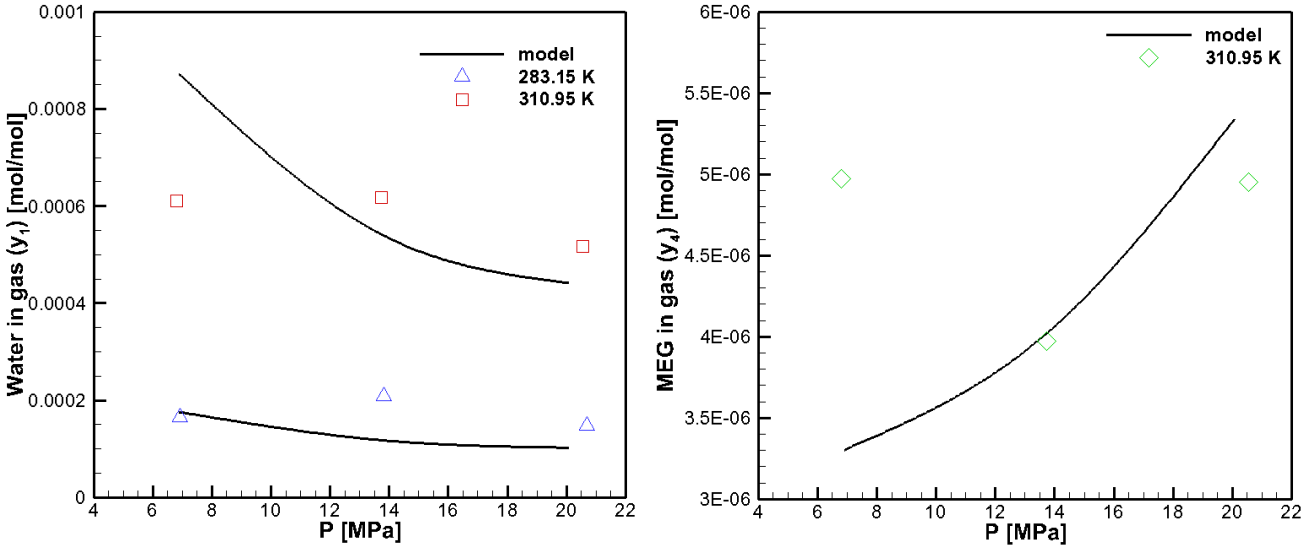

Figure 7. Experimental and predicted mole fraction of water and MEG in the gas phase in VLE for a water, methane, propane and MEG system at 283.15 K and 310.95 K (Ng and Chen 1995).

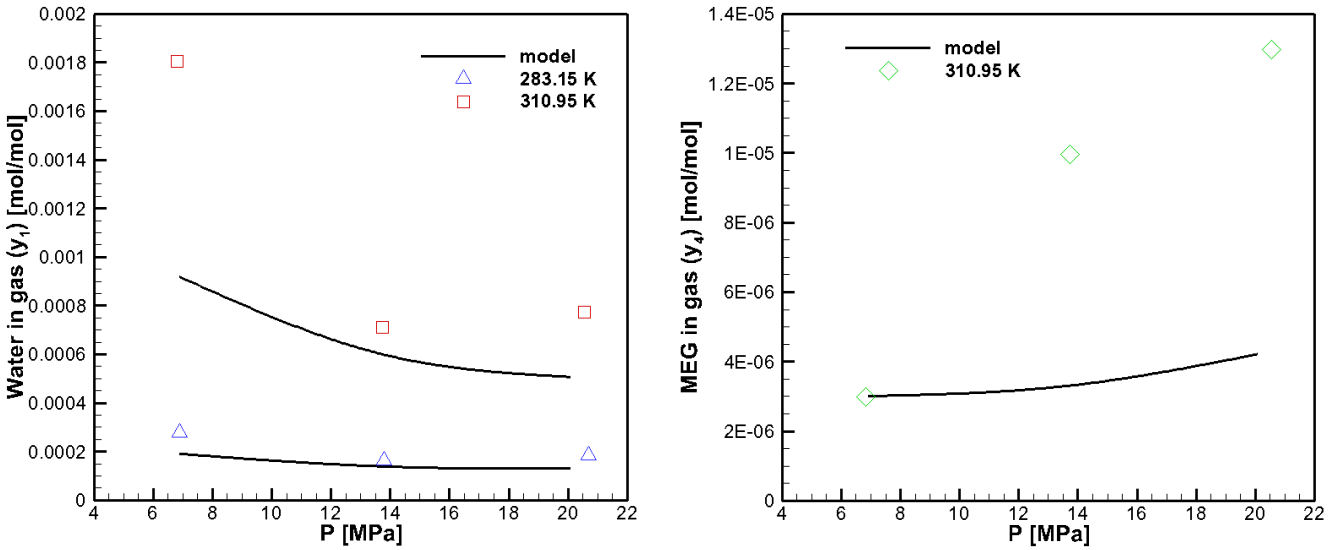

Figure 8. Experimental and predicted mole fraction of water and MEG in the gas phase in VLE for a water, methane, carbon dioxide and MEG system at 283.15 K and 310.95 K (Ng and Chen 1995).

It is important to point out that the presence of systems containing carbon dioxide and hydrogen sulfide (hydrate-forming acid gases frequently present in the composition of natural gases), which can perform cross-association with other associative compounds (hydrate inhibitors), should be investigated in greater detail to evaluate some possible limitations of CPA.

Despite the satisfactory performance exhibited by the CPA EoS in this work, association theories present theoretical and practical limitations and there are many unanswered questions. It is possible to see a detailed discussion regarding this subject in Kontogeorgis (2013). Only a few of these problems will be highlighted in this paper. One topic that requires special attention is the modeling of systems containing acid gases, such as CO_2. Carbon dioxide is in reality a quadrupolar molecule (not a self-associating molecule), and the model does not take this type of interaction into account. However, these quadrupolar interactions can be implicitly accounted via binary interaction parameters. Due to this fact, for highly polar and quadrupolar systems (e.g. CO_2 with hydrocarbons), non-zero binary interaction parameters must be always implemented in order to obtain satisfactory results (Kontogeorgis 2013). During the modeling of the vapor-liquid equilibrium of CO_2-alkanes systems, Tsivintzelis et al. (2015) reported better predictive results (without binary interaction parameters) when considering CO_2 as a self-associative compound. Although there is no experimental evidence for strong self-associating interactions between CO_2 molecules, the addition of self-association effects in the modeling may indirectly and effectively account for quadrupolar interactions. Regardless of the modeling approach, very similar results were obtained when interaction parameters were used. According to Tsivintzelis et al. (2015), higher values of binary interaction parameters are necessary when CO_2 is treated as a non-associating fluid.

In general, the model described in this work showed a good performance, given the inherent difficulty in predicting traces of water content and thermodynamic inhibitors in the gas or condensate phase. Results showed a good agreement with experimental data, especially regarding the partition of methanol and MEG into the gas phase (systems I and II), which is one of the most critical and important datum for the oil and gas industry due to its wide use in large amounts as a tool in prevention of hydrates formation. However, it was found that CPA underpredicted the loss of MEG to gas phase in a gas mixture containing carbon dioxide. In most cases, the model tends to overestimate the solubility of the gas in the liquid phase.

4. CONCLUSIONS

The objective of this work was to implement a thermodynamic model capable of predicting hydrate inhibitor partition. CPA EoS was utilized to calculate vapor-liquid equilibrium (VLE) and liquid-liquid equilibrium (LLE) of ternary and quaternary systems containing hydrate-forming gases (methane, ethane, propane and carbon dioxide), water and thermodynamic inhibitors (methanol and monoethylene glycol). Different approaches for the EoS binary interaction parameter were also implemented and tested. Results were compared with experimental data available in the literature and proved to be satisfactory for the loss of methanol and MEG to gas and/or condensate phases, especially due to the difficulty associated with predicting traces of water content and thermodynamic inhibitors in the gas and condensate phases. Yet, CPA underpredicted the loss of MEG to gas phase in a gas mixture containing carbon dioxide. The average absolute deviation for the prediction of loss of methanol and mono-ethylene glycol ranged between 3-45 %. The authors also realized that more experimental data are needed to validate the thermodynamic models.

5. ACKNOWLEDGEMENTS

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