UNIVERSIDADE ESTADUAL DE CAMPINAS
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RAPHAELA GABRÍ BITENCOURT

Equilíbrio, extração e fracionamento de óleo e compostos fenólicos de café com dióxido de carbono supercrítico e etanol

Equilibrium, extraction and fractionation of oil and phenolic compounds from coffee with supercritical carbon dioxide and ethanol

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Orientador: Prof. Dr. Antonio José de Almeida Meirelles

Coorientador: Prof. Dr. Fernando Antonio Cabral

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A Ata da defesa com as respectivas assinaturas dos membros encontra-se no processo de vida acadêmica do aluno.

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RESUMO

Esse projeto objetivou obter dados de solubilidade de compostos fenólicos e óleo de café em dióxido de carbono supercrítico ($scCO_2$) na presença de cossolventes, ajustar parâmetros de modelos termodinâmicos para descrever estes dados e, posteriormente, avaliar a extração e fracionamento dessas substâncias, que se destacam por seu alto valor agregado e por sua diversidade de aplicações em diferentes ramos industriais, a partir do resíduo da produção de café solúvel. As determinações experimentais de solubilidade dos ácidos cafeico e ferúlico e do óleo de café verde foram realizadas em diferentes condições de temperatura (313 a 333 K) e pressão (20 a 40 MPa), e as quantidades de cossolvente (etanol e/ou água) adicionadas ao $scCO_2$ variaram de 0 a 10 mol%. No geral, os maiores valores de solubilidade foram obtidos nas maiores condições de pressão. Ao se adicionar aproximadamente 10 mol% de etanol ao $scCO_2$, verificou-se aumento de 120 e 30000 vezes na solubilidade dos ácidos ferúlico e cafeico, respectivamente. Os dados experimentais dos sistemas binários foram bem representados pelas equações de Peng-Robinson, Soave-Redlich-Kwong e *Cubic Plus Association* (CPA). Porém, apenas a equação CPA foi capaz de predizer satisfatoriamente os dados de equilíbrio dos sistemas ternários e/ou quaternários a partir dos parâmetros de interação binária, pois, diferentemente das equações de estado cúbicas, este modelo considera as fortes interações, como ligações de hidrogênio, que ocorrem entre as moléculas do soluto e do cossolvente. A solubilidade do óleo de café verde em $scCO_2$ mostrou-se similar à de diversos outros óleos vegetais. Ela aumentou cerca de 60% com a adição de 2,3 % (em massa) de etanol, e as amostras obtidas mostraram-se mais concentradas em ácidos graxos livres e diterpenos em relação ao óleo obtido por prensagem. Extrações com fluido supercrítico (*Supercritical Fluid Extraction* - SFE) e líquido pressurizado (*Pressurized Liquid Extraction* - PLE) em uma ou duas etapas e usando apenas dióxido de carbono e etanol como solventes foram avaliadas para obtenção de compostos fenólicos e lipídicos da borra de café. Os maiores rendimento global de extração, concentração de compostos fenólicos e taxa de extração foram verificados quando o etanol foi utilizado como solvente (extrato E). O fracionamento de E, tanto por um processo integrado de SFE + PLE ou usando o $scCO_2$ como antissolvente por meio de quatro separadores em série, forneceu frações aproximadamente quatro vezes mais concentradas que o extrato etanólico obtido em etapa única. Contudo, o uso de cossolventes, principalmente etanol, mostrou-se fundamental para extração de compostos fenólicos por consequência do aumento da sua solubilidade em $scCO_2$. O conhecimento da solubilidade de solutos em $scCO_2$ + cossolventes e de como alguns parâmetros, como temperatura, pressão e tipo e concentração de cossolvente a influenciam pode ser útil para otimizar processos de separação que utilizam o $scCO_2$ como solvente ou antissolvente.

Palavras-chaves: borra de café; solubilidade; tecnologia supercrítica; modelagem termodinâmica; equação de estado.

ABSTRACT

This work proposed to obtain solubility data of phenolic compounds and green coffee oil in supercritical carbon dioxide ($scCO_2$) with cosolvents, to describe these data through thermodynamic models, and posteriorly to evaluate the extraction and fractionation of compounds from the residue of soluble coffee production. These substances are highlighted by their high value added and its diversity of applications in different industrial sectors. Experimental measurements of solubility of caffeic and ferulic acids and green coffee oil were performed at different temperatures (313 to 333 K) and pressures (20 to 40 MPa), and the amounts of cosolvent (ethanol and/or water) added to $scCO_2$ ranged from 0 to 10 mol%. In general, the highest solubility values were obtained under the highest pressure conditions. By adding approximately 10 mol% of ethanol to $scCO_2$, it was verified an increase up to 120 and 30,000 times on the solubility of ferulic and caffeic acids, respectively. Experimental data of the binary systems were well described by the equations of Peng-Robinson, Soave-Redlich-Kwong and Cubic Plus Association (CPA). However, CPA equation was the only one suitable to predict the equilibrium data of the ternary and/or quaternary systems from the binary interaction parameters, because unlike the cubic state equations this model describes the strong interactions, such as hydrogen bonds, that occur between the solute and cosolvent molecules. Solubility data of green coffee oil in $scCO_2$ was similar to several other vegetable oils. It increased by up to 60% with the addition of 2.3% ethanol (w/w), and the samples obtained were more concentrated in free fatty acids and diterpenes in comparison with oil obtained by pressing. Supercritical fluid extraction (SFE) and pressurized liquid extraction (PLE) in one or two steps and using CO_2 and ethanol as solvents were evaluated to extract phenolic and lipid compounds from spent coffee grounds. The highest overall extraction yield, total phenolic content and extraction rate were verified when ethanol was used as a solvent (extract E). The fractionation of E, either by an integrated process with SFE + PLE or by using $scCO_2$ as antisolvent through four separators in series, favors obtaining fractions approximately four times more concentrated than the single-stage ethanolic extract. Therefore, the use of cosolvents, mainly ethanol, confirms to be crucial for the extraction of phenolic compounds, due to the increase of its solubility in $scCO_2$. In addition, knowledge on the effect of the parameters, such as temperature, pressure and type and cosolvents concentration, on the solubility of solutes in $scCO_2$ + cosolvents can be useful to optimize separation processes using $scCO_2$ as solvent or antisolvent.

Keywords: spent coffee grounds; solubility; supercritical technology; thermodynamic modeling; equation of state.

LISTA DE ABREVIATURAS, SIGLAS E SÍMBOLOS

Abreviaturas

AARD – *Average Absolute Relative Deviation* (Desvio relativo absoluto médio)

AG – *Associative Group* (Grupo associativo)

ALD – *Average Logarithmic Deviation* (Desvio logarítmico médio)

BIP – *Binary Interaction Parameters* (Parâmetros de interação binária)

CA – C*affeic Acid* (Ácido cafeico)

CE – *Cosolvent Effect* (Efeito do cossolvente)

CG – Contribuição de Grupos

CGA – *Chlorogenic acid* (Ácido clorogênico)

COSMOSAC – *Conductor-like Screening Model – Segment Activity Coefficient*

CPA – *Cubic Plus Association*

ELV – Equilíbrio Líquido-Vapor

EoS – *Equation of State* (Equação de estado)

ESV – Equilíbrio Sólido-Vapor

FA – *Ferulic Acid* (Ácido ferúlico)

FFA – *Free Fatty Acids* (Ácidos graxos livres)

GAE – *Gallic Acid Equivalent* (Equivalente em ácido gálico)

GCO – *Green coffee oil* (Óleo de café verde)

GRAS – *Generally Recognize As Safe* (Geralmente reconhecido como seguro)

MS – *Mass Spectrometry* (Detector por espectrometria de massas)

OF – *Objective Function* (Função objetivo)

PDA – *Photodiode Array Detector* (Detector de arranjo de fotodiodos)

PLE – *Pressurized Liquid Extraction* (Extração com líquido pressurizado)

PR – Peng-Robinson

RC – Regra de Combinação

SAF – *Supercritical Antisolvent Fractionation* (Fracionamento com antissolvente supercrítico)

SAFT – *Statistical Associating Fluid Theory*

scCO_2 – Dióxido de carbono supercrítico

SCG – *Spent Coffee Grounds* (Borra de café)

SFE – *Supercritical Fluid Extraction* (Extração com fluido supercrítico)

SRK – Soave-Redlich-Kwong

TAG – *Triacylglycerol* (Triacilglicerol)

TPC – *Total Phenolic Content* (Teor de compostos fenólicos totais)

VDW – van der Waals

Símbolos

a – parâmetro de energia da equação de estado cúbica (termo físico)

$A, B, u, w, \Omega_a, \Omega_b, m$ – constantes utilizadas para representar diferentes equação de estado cúbicas

a_0, c_1 – parâmetros para o cálculo de a

b – parâmetro de co-volume da equação de estado cúbica (termo físico)

C_p – capacidade térmica

E, C – porcentagem molar de etanol em base livre de soluto

f – fugacidade

g – função de distribuição radial

H – entalpia

k, a, b – parâmetros da Equação de Chrastil

k_{ij} – parâmetro de interação binária

l_{ij} – parâmetro de interação binária

M – massa molar

n – número de mols

N – número de pontos experimentais

NC – número de componentes do sistema

P – pressão

R – constante universal dos gases

S – função objetivo

T – temperatura

u – incerteza padrão

V, $\underline{V}$ – volume molar

X_{Ai} – fração molar do componente i não ligada pelo sítio A

W – porcentagem molar de água em base livre de soluto

w – fração mássica

y , x – fração molar

Y – fração molar em base livre de soluto

$\bar{Y}$ – solubilidade em g/kg solvente

Y_1 – rendimento de extração global

Y_2 – rendimento de extração em compostos fenólicos

Z – fator de compressibilidade

Letras gregas

β – parâmetro de volume (termo associativo)

γ – coeficiente de atividade

Δ – força de associação

ε – parâmetro de energia (termo associativo)

Δ – força de associação

η – densidade reduzida

ρ – densidade molar

φ , ϕ – coeficiente de fugacidade

ω – fator acêntrico

Subscritos

A , B – sítios de ligação da molécula

atm – condição atmosférica

C – propriedade crítica

cal – dado calculado

ET – etanol

exp – dado experimental

i , j – componente ou molécula

m – melting

f – fusion

mix – mistura

WA – água

Sobrescritos

assoc – contribuição do termo associativo

d – gás ideal

e – etanol

sat – saturação

vap – vapor

w – água

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CAPÍTULO 1. INTRODUÇÃO E OBJETIVOS

1.1. Introdução Geral e Justificativa

Considerando a crescente busca dos consumidores por produtos contendo ingredientes funcionais, seguros e de origem natural, é necessário o investimento em pesquisas que visem à obtenção de extratos naturais contendo compostos bioativos e com propriedades de interesse para a indústria de alimentos, cosmética e farmacêutica. Dentre os compostos com aplicações industriais, encontram-se os compostos fenólicos e lipídicos. Os compostos fenólicos, em geral, são biocompostos de alto valor agregado, pois apresentam tanto propriedades químicas como biológicas, podendo-se citar como exemplos as atividades antioxidante, quelante, anti-inflamatória, antialérgica, antimicrobiana, antiviral e anticarcinogênica (MOTA et al., 2008). Por outro lado, os óleos vegetais representam um importante produto no mercado mundial, com ampla variedade de áreas econômicas de interesse, desde a nutrição humana e animal até a produção de bioenergia (GRACIA et al., 2009).

Para a obtenção de extratos naturais, a tecnologia supercrítica é utilizada com bastante sucesso, pois vem provando ser uma tecnologia viável técnica e economicamente na extração de compostos bioativos a partir de diversas matrizes vegetais. Extrações que utilizam dióxido de carbono supercrítico ($scCO_2$) como solvente apresentam algumas vantagens em relação às técnicas tradicionais de extração, pois as propriedades do CO_2 como solvente podem ser facilmente manipuladas através do ajuste da pressão e da temperatura do processo e os extratos e resíduos produzidos são livres de solventes. Além disso, o CO_2 é barato, não inflamável, GRAS (*Generally Recognized As Safe*), disponível em alta pureza e apresenta baixos valores de temperatura e pressão críticas, sendo atrativo para extração e fracionamento de compostos termossensíveis (REVERCHON; DE MARCO, 2006).

Porém, a extração com $scCO_2$ puro é mais adequada para obtenção de compostos apolares ou de baixa polaridade, sendo necessária a utilização de cossolventes, tais como etanol e/ou água, quando se deseja extrair substâncias de maior polaridade, como os compostos fenólicos. A utilização de cossolventes, particularmente daqueles que também são considerados seguros para a indústria de alimentos, é interessante para manter esta técnica classificada como um “processo verde” e ambientalmente amigável.

O conhecimento da solubilidade de compostos de interesse em $scCO_2$ puro ou em $scCO_2$ na presença de cossolventes é de fundamental importância para o *design* e otimização de processos que empregam a tecnologia supercrítica, usando o $scCO_2$ como solvente ou antissolvente. Este conhecimento do equilíbrio de fases em condições supercríticas pode ser baseado em determinações experimentais e/ou em modelos termodinâmicos, os quais possibilitam correlacionar as informações experimentais e, com isso, usá-los como ferramenta para prever dados de equilíbrio em condições diferentes daquelas disponíveis.

No geral, as equações de estado cúbicas são os modelos termodinâmicos mais empregados para cálculo de equilíbrio de fases de sistemas em altas pressões. As equações de estado (*Equation of State* - EoS) de Peng-Robinson (PR-EoS), Soave-Redlich-Kwong (SRK-EoS) e a equação de estado por Contribuição de Grupos (CG-EoS) foram utilizadas para modelar sistemas contendo CO_2, etanol e compostos fenólicos que apresentam atividade antioxidante (BERNA et al., 2001; BERNA; CHÁFER; MONTÓN, 2001; CHAFER et al., 2004; CHÁFER et al., 2002, 2007, 2009; PAVIANI et al., 2013; ZHAN et al., 2017). No entanto, a modelagem empregando essas equações não foi avaliada visando à predição do equilíbrio de fases dos sistemas ternários a partir dos parâmetros de interação binária.

Entre as equações de estado da nova geração encontra-se a equação *Cubic Plus Association* (CPA-EoS), desenvolvida em 1996 para modelar sistemas contendo compostos fortemente polares. Ela consiste na combinação de uma equação cúbica (Soave-Redlich-Kwong) e o termo de associação proposto por Wertheim e já amplamente utilizado em equações do tipo SAFT (*Statistical Associating Fluid Theory*) (KONTOGEORGIS et al., 1996). Dessa forma, a CPA-EoS representa um modelo para sistemas contendo moléculas associativas, porém mantendo a maioria das vantagens bem conhecidas das equações de estado cúbicas. Tanto a academia como a indústria já aplicaram a CPA-EoS, embora seu uso na indústria ainda seja limitado a aplicações na área de petróleo e gás. Esta equação já foi utilizada para correlacionar satisfatoriamente a solubilidade de compostos fenólicos e fármacos em água e em solventes orgânicos (MOTA et al., 2008, 2011; QUEIMADA et al., 2009), na modelagem do equilíbrio líquido-vapor do sistema CO_2/etanol/água (PERAKIS et al., 2006), e em sistemas (contendo óleo) relevantes na produção de biodiesel (FOLLEGATTI-ROMERO et al., 2012; OLIVEIRA et al., 2008, 2009, 2010, 2011, 2014a, 2014b, OLIVEIRA; QUEIMADA; COUTINHO, 2010a, 2010b).

Outro contexto importante para a justificativa desta tese de doutorado abrange a crescente preocupação com o meio ambiente e a pressão de órgãos ambientais sobre as indústrias geradoras de resíduos orgânicos. Dessa forma, buscam-se constantemente alternativas para reutilização ou valorização desses resíduos. Apesar dos esforços recentes, a implementação de processos industriais utilizando resíduos industriais como matéria-prima ainda é um desafio a ser superado (BALLESTEROS; TEIXEIRA; MUSSATTO, 2014). Por produzir uma das bebidas mais consumidas no mundo, a indústria de café é responsável pela geração de grandes quantidades de resíduos orgânicos, dentre os quais destaca-se o material chamado de borra de café ou *spent coffee grounds* (SCG), que consiste no resíduo obtido durante a produção de café solúvel. O SCG é um material lignocelulósico rico em açúcar e em fibras alimentares, com aproximadamente 15% em massa de material lipídico (MUSSATTO et al., 2011), e com potencial antioxidante associado à presença de ácidos fenólicos (BALLESTEROS; TEIXEIRA; MUSSATTO, 2014), como os ácidos ferúlico e cafeico.

Contudo, a escassez de dados de equilíbrio de fases de sistemas contendo CO_2, cossolventes e os compostos (fenólicos e lipídicos) presentes no café motivou a realização deste trabalho. Além disso, acredita-se que a CPA-EoS é um modelo termodinâmico promissor para correlacionar e predizer satisfatoriamente esses dados experimentais de solubilidade em misturas de CO_2 e cossolventes, uma vez que interações fortes podem ocorrer entre os solutos e o cossolvente.

1.2. Objetivos

1.2.1. Objetivo Geral

Este projeto de pesquisa objetivou obter dados de equilíbrio de fases para sistemas contendo compostos fenólicos, $scCO_2$, etanol e água e óleo de café, $scCO_2$ e etanol, avaliar a equação de estado CPA para correlacionar e predizer os dados experimentais, assim como estudar o processo de extração e fracionamento do óleo de café a partir do resíduo da indústria de café.

1.2.2. Objetivos Específicos

Os objetivos específicos deste projeto são listados a seguir:

i. Obter dados de solubilidade em alta pressão do ácido ferúlico e ácido cafeico em misturas de CO_2, etanol e água;

ii. Obter dados de solubilidade em alta pressão do óleo de café em CO_2 e misturas de CO_2 e etanol;

iii. Avaliar os efeitos de temperatura, pressão e do cossolvente na solubilidade de óleo e compostos fenólicos em CO_2 + cossolventes;

iv. Obter dados do equilíbrio de fases sob baixa pressão de sistemas binários relevantes: (1) óleo de café e etanol, (2) composto fenólico e etanol e (3) composto fenólico e água;

v. Verificar a capacidade das equações de estado cúbicas e a equação CPA para:
   - Correlacionar os dados de equilíbrio dos sistemas binários;
   - Predizer dados de equilíbrio de fases dos sistemas ternários e/ou quaternários a partir dos parâmetros de interação binária.

vi. Fracionar o óleo e compostos fenólicos presentes na borra de café utilizando:
   - CO_2 em diferentes condições de temperatura e pressão;
   - CO_2 como antissolvente.

1.3. Estrutura da Tese

As etapas do desenvolvimento desta tese foram organizadas em 9 capítulos, descritos a seguir. No **Capítulo 1 – Introdução e Objetivos** é apresentado o tema deste trabalho, juntamente com justificativa para este projeto de pesquisa, além de listar os objetivos alcançados.

Capítulo 2 – Revisão Bibliográfica contempla uma revisão bibliográfica sobre os temas abordados nesta tese, como tecnologia supercrítica, equilíbrio de fases, modelagem termodinâmica, solubilidade de solutos em $scCO_2$ e compostos de interesse presentes no café.

A metodologia desenvolvida em toda esta tese está resumida de forma esquemática na **Figura 1.1**. Os resultados começam a ser apresentados no **Capítulo 3 – *Ferulic acid solubility in supercritical carbon dioxide, ethanol and water mixtures***, artigo publicado no T*he Journal of Chemical Thermodynamics*, o qual reporta dados de solubilidade do ácido ferúlico em etanol, água e em dióxido de carbono supercrítico na presença de cossolventes (etanol e uma mistura de etanol/água), avaliando como estes são influenciados pela temperatura, pressão e concentração de etanol na mistura CO_2/etanol. Além disso, verificou-se que, apesar de um bom ajuste aos dados experimentais dos sistemas binários, a modelagem com a equação de estado de Peng-Robinson não é capaz de predizer os dados de equilíbrio dos sistemas ternários e quaternários.

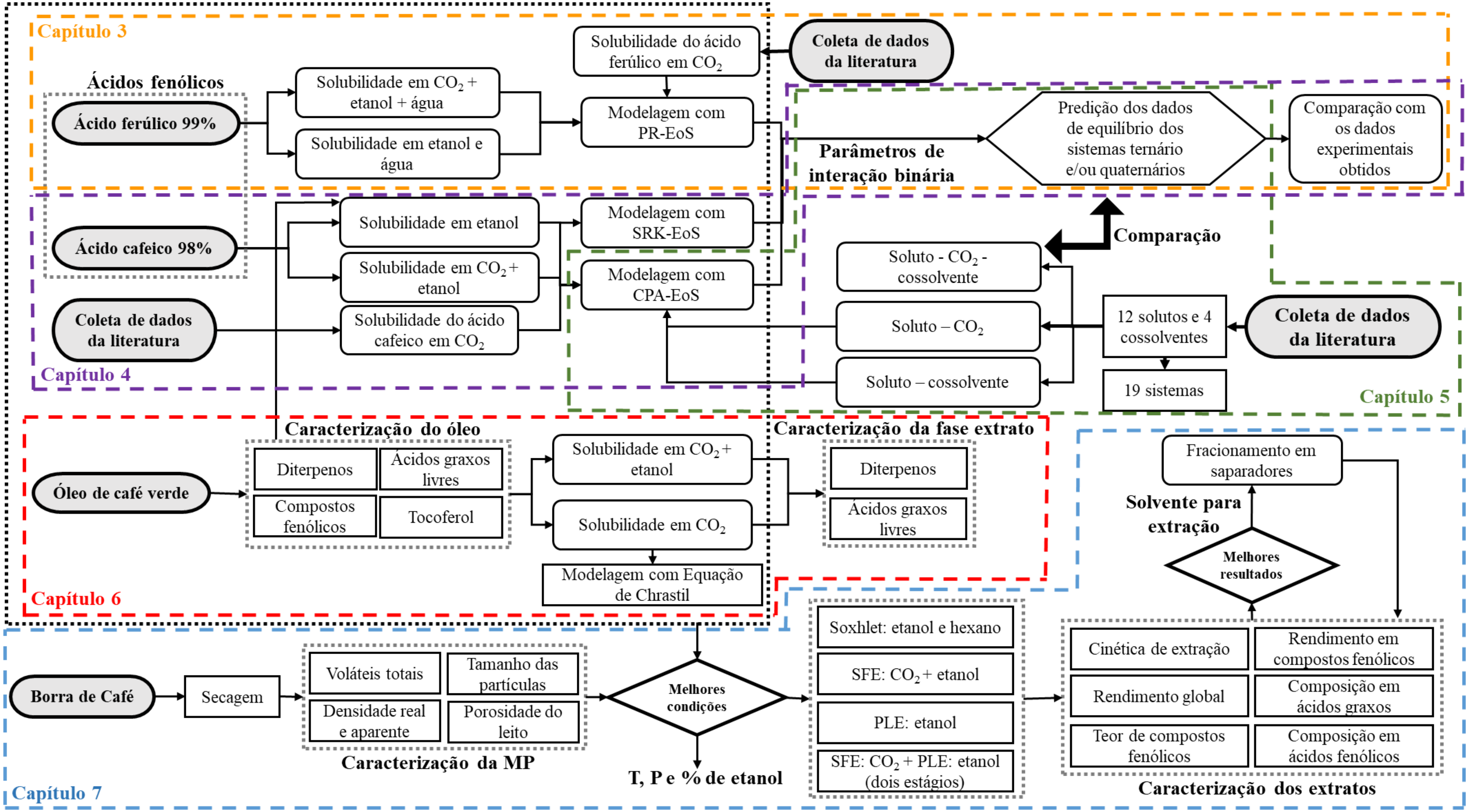

Figura 1.1 – Esquema geral das atividades realizadas nesta tese. MP: matéria-prima.

O **Capítulo 4** – ***Solubility of caffeic acid in CO_2 + ethanol: experimental and predicted data using Cubic Plus Association Equation of State*** é um artigo submetido à revista *The Journal of Supercritical Fluids*. De forma similar ao Capítulo 3, são apresentados dados experimentais da solubilidade do ácido cafeico em etanol e misturas de dióxido de carbono supercrítico e etanol. Porém, neste trabalho, comparando-se a modelagem termodinâmica usando a equação de estado Soave-Redlich-Kwong e a *Cubic Plus Association*, verificou-se que o modelo CPA prediz satisfatoriamente a solubilidade do ácido cafeico em CO_2 + etanol a partir dos parâmetros de interação binária, provalmente por considerar as interações associativas, como ligações de hidrogênio, que ocorrem entre o ácido cafeico e o etanol.

A partir dos resultados obtidos no Capítulo 4, o **Capítulo 5** – ***Prediction of solid solute solubility in supercritical CO_2 with cosolvents using CPA equation of state*** foi desenvolvido para confirmar a capacidade da CPA-EoS em predizer dados de solubilidade de diferentes solutos sólidos em dióxido de carbono supercrítico na presença de cossolventes. Neste artigo, ainda a ser submetido para o periódico *Industrial & Engineering Chemistry Research.*, um conjunto de dados foi selecionando a partir da literatura científica, avaliando-se diferentes classes de solutos e cossolventes, totalizando 19 sistemas. Bons resultados foram obtidos, principalmente comparando a predição da equação CPA com dados previamente reportados na literatura, que empregaram diferentes modelos termodinâmicos.

No **Capítulo 6** – ***High pressure phase equilibrium of the crude green coffee oil – CO_2 – ethanol system and the oil bioactive compounds*** é apresentado o artigo publicado no *The Journal of Supercritical Fluids*, no qual são apresentados dados de solubilidade do óleo de café verde comercial tanto em dióxido de carbono supercrítico puro, quanto na mistura CO_2 + etanol. Além disso, foram avaliados o teor de ácidos graxos livres (FFA) e diterpenos nas amostras obtidas. Verificou-se que o uso de CO_2 com ou sem etanol promove a extração de óleo com maior conteúdo de ácidos graxos livres e diterpenos. Em complemento, dados do sistema óleo – etanol são apresentados no Anexo A.

Enfim, o **Capítulo 7** – ***Extraction and fractionation of oil from spent coffee grounds using green solvents at high pressures*** é um artigo a ser submetido na revista *Journal of Food Engineering*, que foi idealizado com objetivo de exemplificar a importância da obtenção de dados de equilíbrio similares aos reportados nos capítulos anteriores para o *design* e otimização de processos que utilizam $scCO_2$ como solvente ou antissolvente. Foram estudados diferentes processos de extração e fracionamento do óleo de café a partir do resíduo industrial da produção de café solúvel usando apenas solventes “verdes”, com o enfoque na obtenção de frações enriquecidas em composto fenólicos.

No **Capítulo 8 – Discussão Geral** é apresentada uma discussão geral de todos os dados obtidos nesta tese, comparando os resultados reportados em diferentes capítulos. Por fim, o **Capítulo 9 – Conclusões** lista as principais conclusões obtidas a partir deste projeto de pesquisa, juntamente com sugestões para trabalhos futuros.

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CAPÍTULO 2. REVISÃO BIBLIOGRÁFICA

2.1.Tecnologia Supercrítica

A tecnologia envolvendo fluidos supercríticos ganhou destaque no final da década de 1970 e no começo da década de 1980, especialmente com a construção, na Alemanha, da primeira planta de extração que visava à remoção de cafeína de grãos de café (MCHUGH; KRUKONIS, 1986).

Um fluido supercrítico é definido como uma substância que se encontra em condições de temperatura e pressão acima dos valores críticos, ou seja, da temperatura crítica (T_c) e da pressão crítica (P_c), sendo o ponto crítico a maior condição de temperatura e pressão em que a substância pode existir como vapor e líquido em equilíbrio. Nestas condições, os fluidos apresentam densidade semelhante à de líquidos e difusividade e viscosidade como as de gases (SUN, 2002). A densidade semelhante à dos líquidos aumenta sua capacidade de solvatação (MCHUGH; KRUKONIS, 1986) e sua alta difusividade, baixa viscosidade e tensão superficial favorecem a penetração e transporte do fluido supercrítico na matriz vegetal quando comparado com processos convencionais (DUNFORD et al., 2003).

2.1.1. Extração com fluidos supercríticos

Extrações com fluido supercrítico (*Supercritical Fluid Extraction* - SFE) são processos de extração em leito fixo com possibilidades de aplicação tanto nas indústria de alimentos, quanto petroquímica e farmacêutica. Esse tipo de processo possui algumas vantagens em relação às técnicas tradicionais de extração: é um processo flexível, devido à possibilidade de alteração na propriedade de solvatação/seletividade do fluido supercrítico, por gerar extratos livres de solvente, evitando o pós-processamento (de custo elevado) para remoção dos solventes dos extratos, e por ser possível a ampliação para a escala industrial (REVERCHON; DE MARCO, 2006). Até alguns anos atrás, acreditava-se que o inconveniente deste modelo de extração era o aumento dos custos, comparando-o às técnicas tradicionais realizadas sob baixas pressões. Porém, pesquisas recentes que avaliaram o custo de manufatura de extratos obtidos por extração supercrítica reportam que o preço para aquisição da matéria-prima é um dos mais importantes para o custo do produto final (CARVALHO et al., 2015; PAULA et al., 2016).

O dióxido de carbono (CO_2) é o solvente mais comumente utilizado em SFE para aplicações alimentares, pois é barato, prontamente disponível em alta pureza, seguro de manusear, não tóxico, seguro (GRAS = *Generally recognized as safe*) e não inflamável (BRUNNER, 2005; UQUICHE; DEL VALLE; ORTIZ, 2004). Além disso, o ponto crítico do

CO_2 é relativamente baixo, temperatura de 304,1 K e pressão de 7,38 MPa (B.E. POLING, J.M. PRAUSNITZ, 2001), tornando-o adequado para extração de compostos termossensíveis.

Entretanto, o CO_2 apresenta características lipofílicas, solubilizando preferencialmente compostos apolares ou de baixa polaridade. Substâncias polares também podem ser extraídas com $scCO_2$ em altas densidades (altas pressões) e/ou com o emprego de cossolventes.

Um cossolvente apresenta a capacidade de modificar as propriedades de solvatação dos fluidos supercríticos mesmo em pequenas quantidades (PELLERIN, 1991; VASAPOLLO et al., 2004). Alguns dos cossolventes mais utilizados para extração de biocompostos são etanol e água, que apresentam polaridade maior que a do dióxido de carbono e também são solventes GRAS e sem restrições de órgãos governamentais e ambientais quanto ao seu uso.

2.1.2. Extrações sequenciais

A integração de processos é uma tendência recente para obtenção de extratos naturais contendo compostos bioativos. Esses processos maximizam os rendimentos de extração e permitem a obtenção de extratos com composições e aplicações diferenciadas, pela utilização de métodos de extração em duas ou mais etapas, nos quais a polaridade do solvente de extração aumenta ao longo de cada etapa. Podem ser combinados processos de extração com fluidos supercríticos, variando as condições de temperatura, pressão, proporção ou o tipo de cossolvente, ou ainda combinar SFE e extração com líquidos pressurizados (*Pressurized Liquid Extraction* – PLE) (VIGANÓ et al., 2016a; MORAES; ZABOT; MEIRELES, 2015).

A utilização de SFE com $scCO_2$ é interessante para a extração do material oleoso presente na matriz vegetal, para posterior obtenção, por PLE, dos compostos de maior polaridade, através de um solvente orgânico (etanol, na maioria dos casos). Neste sentido, o processo se torna mais interessante economicamente, alcançando altos rendimentos de extração para os componentes desejáveis e aproveitando de forma integral a matéria-prima (VIGANÓ; ZABOT; MARTÍNEZ, 2017).

A maioria dos processos integrados de SFE e PLE já estudados visam a obtenção de extratos enriquecidos em compostos fenólicos (PAULA et al., 2014; GARMUS et al., 2014; VIGANÓ et al. 2016b; GARCIA MENDOZA et al., 2015; MONROY et al., 2016). Porém, a extração de saponinas (BITENCOURT et al., 2014) e bixina (MORAES; ZABOT; MEIRELES, 2015) também já foram avaliadas.

2.1.3. Fracionamento com antissolvente supercrítico

Processos de separação que utilizam antissolvente supercrítico foram propostos como alternativa para processos que usam antissolvente líquido (REVERCHON, 1999). Esta técnica baseia-se no contato de uma solução contendo um ou mais compostos ativos e um solvente líquido convencional com um fluido supercrítico, como esquematizado na **Figura 2.1.** Inicialmente, a mistura é um sistema homogêneo formado pelos compostos ativos e pelo solvente. Esse sistema é então conduzido ao encontro do antissolvente à alta pressão, formando uma mistura. Como resultado da mudança de pressão, temperatura e composição, o sistema que anteriormente se situava em uma região de única fase é então deslocado para a região de duas fases. Assim, a separação é possível, induzindo a formação de partículas precipitadas do soluto (COCERO et al., 2009).

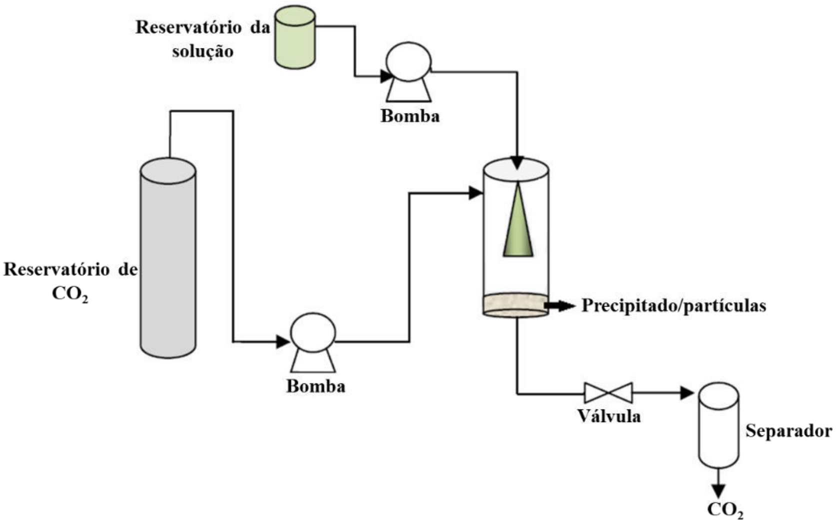

Figura 2.1 – Diagrama esquemático do processo de separação usando CO_2 como antissolvente supercrítico (Adaptado de POCHEVILLE et al., 2011).

O antissolvente supercrítico se caracteriza por apresentar difusividade que pode ser até duas ordens de magnitude maior do que a dos líquidos. Portanto, a sua difusão rápida para o solvente líquido produz a saturação do soluto e a sua precipitação em partículas com pequeno diâmetro, e que não são possíveis de obter usando antissolventes líquidos ou outras técnicas (REVERCHON, 1999).

Este comportamento é observado quando há miscibilidade completa do solvente líquido selecionado no antissolvente supercrítico, enquanto os solutos não devem ser solúveis no meio supercrítico. Além disso, é desejável que o solvente tenha maior afinidade pelo antissolvente do que pelo composto ativo, de tal maneira que, quando a solução (solvente + composto ativo) é colocada em contato com o antissolvente e ao mesmo tempo a pressão é reduzida, a solubilidade do composto ativo no solvente é drasticamente reduzida. Assim, o antissolvente causa uma queda no poder de solvatação do solvente, levando à precipitação do composto ativo na forma de partículas (FLORIS et al., 2010; KALANI; YUNUS, 2011).

Esse tipo de processo também é capaz de realizar o fracionamento (*Supercritical Antisolvent Fractionation* - SAF) de compostos de interesse pelo uso de um antissolvente supercrítico através de uma sequência de separadores (precipitadores) que operam em condições diferentes de temperatura e pressão. Isso ocorre pois, além do solvente, alguns compostos podem ser solúveis no $scCO_2$ e, por esta razão, serem solubilizados na mistura formada no precipitador e eliminados na separação seguinte. Dessa maneira, extratos de própolis (CATCHPOLE et al., 2004), alecrim-do-campo (PAULA et al., 2017) e crajirú (PAULA et al., 2018) foram fracionados.

2.2.Modelagem Termodinâmica do Equilíbrio de Fases

Com o objetivo de otimizar e até predizer resultados de um processo de separação, tal como SFE, é necessário o conhecimento do equilíbrio de fases dos sistemas envolvidos no processo. Ou seja, o conhecimento da solubilidade dos compostos de interesse em $scCO_2$ puro ou em $scCO_2$ na presença de cossolventes é de fundamental importância para o *design* de processos que empregam tecnologia supercrítica, seja com o $scCO_2$ como solvente ou antissolvente.

O conhecimento do equilíbrio de fases em condições supercríticas pode ser baseado em determinações experimentais e em modelos termodinâmicos de equilíbrio, os quais possibilitam correlacionar as informações experimentais e, com isso, serem usados como ferramenta para predição de dados de equilíbrio em condições diferentes daquelas disponíveis. Assim, o objetivo da modelagem termodinâmica do equilíbrio de fases é descrever o comportamento dos sistemas em equilíbrio, ou seja, correlacionar e predizer dados experimentais de misturas binárias, ternárias ou multicomponentes.

O estado de equilíbrio pode ser definido como aquele no qual o sistema não apresenta uma tendência espontânea a mudanças. Por exemplo, quando duas fases são colocadas em contato, ocorre uma transferência de componentes de uma fase à outra até que a composição de cada fase atinja um valor constante. Quando isto acontece, diz-se que foi atingido o estado de equilíbrio (químico). A composição de equilíbrio de cada fase pode ser muito diferente, e é esta diferença que permite, por exemplo, separar misturas por extração e destilação (PRAUSNITZ; LICHTENTHALER; DE AZEVEDO, 1998).

O equilíbrio de um sistema é estabelecido quando se atingem os equilíbrios térmico, mecânico e químico. Para um sistema fechado bifásico (heterogêneo), o equilíbrio pode ser representado pela igualdade da temperatura e pressão nas fases e do potencial químico de cada componente em ambas as fases do sistema.

$$T^{faseI} = T^{faseII} \tag{2.1}$$

$$P^{faseI} = P^{faseII} \tag{2.2}$$

$$\mu_i^{faseI} = \mu_i^{faseII} \quad i = 1,2,3...NC \tag{2.3}$$

Com o intuito de relacionar o potencial químico do componente com uma variável mensurável (concentração), funções auxiliares, tais como a fugacidade, são utilizadas. De forma geral, o equilíbrio de fases também pode ser representado pela condição de isofugacidade, isto é, as fugacidades de cada um dos componentes na mistura devem ser iguais em todas as fases.

A partir deste critério, o equilíbrio de fases pode ser calculado por duas abordagens distintas: o método γ-ϕ e o método ϕ-ϕ.

A abordagem γ-ϕ faz um tratamento distinto das duas fases, calculando o coeficiente de fugacidade (ϕ) quando há uma fase vapor envolvida, e empregando modelos baseados na energia de Gibbs em excesso para obtenção do coeficiente de atividade (γ) para a fase líquida. Estes modelos são deficientes para descrever o comportamento de fases a altas pressões, em função da compressibilidade das fases (PRAUSNITZ; LICHTENTHALER; DE AZEVEDO, 1998).

Na abordagem ϕ-ϕ, as fugacidades das fases fluidas (gases, líquidos e fluidos supercríticos) são representadas usando o coeficiente de fugacidade, cujo cálculo é realizado a partir de modelos termodinâmicos baseados nas chamadas equações de estado (EoS). Estas equações correlacionam as propriedades pressão-volume-temperatura (PVT) de fluidos, e é comumente empregado para a predição do equilíbrio líquido-vapor.

Pelo método ϕ-ϕ, as fugacidades para as fases líquida e vapor podem ser expressas da seguinte maneira:

$$f_i^{vapor} = y_i \phi_i^{vapor} P \tag{2.4}$$

$$f_i^{líquido} = x_i \phi_i^{líquido} P \tag{2.5}$$

Assim, por meio da relação termodinâmica expressa na Equação 2.6 (Prausnitz, 1998) e de uma equação de estado, é possível resolver o problema do equilíbrio entre uma fase líquida e uma vapor (ou supercrítica).

$$\ln\phi_i^{faseI} = \frac{1}{RT}\int_{V^{faseI}}^{\infty}\left[\left(\frac{\partial P}{\partial n_i}\right)_{T,V,n_{i\neq j}} - \frac{RT}{V}\right]dV - \ln\left(\frac{PV^{faseI}}{RT}\right) \quad (2.6)$$

Quando o equilíbrio de fases envolve uma fase sólida, a fugacidade desta é calculada usando uma expressão diferente. Geralmente a fase sólida é modelada como uma fase isolada, por meio da seguinte equação:

$$f_i^{sólido} = P_i^{sat}\phi_i^{sat}\exp\left(\frac{V_i^{sólido}(P - P_i^{sat})}{RT}\right) \quad (2.7)$$

onde o termo exponencial é o fator de correção de Poynting para a fugacidade do sólido puro, que leva em conta que o sólido está a uma pressão *P* diferente da pressão de saturação P_i^{sat}. Geralmente, ϕ_i^{sat}, termo para a correção de altas pressões de saturação, pode ser considerado 1,0, já que na maioria dos casos a pressão de saturação de um sólido cristalino é muito menor que 1 bar (ARCE-CASTILLO, 2002).

A Equação 2.7 é válida quando se assumem as seguintes hipóteses: o soluto permanece como um sólido puro, ou seja, não se forma uma solução sólida; o volume molar do soluto pode ser considerado constante de acordo com a pressão; e o vapor saturado do sólido se comporta como um gás ideal (CHÁFER et al., 2009).

Outra abordagem para o cálculo do equilíbrio de sistemas contendo fases sólidas, que foi aplicada nos Capítulos 4 e 5, é apresentada no Anexo B.

2.2.1. Equações de estado

As equações de estado (*Equations of State* – EoS) são expressões algébricas que relacionam a pressão e a temperatura com o volume molar para uma substância pura ou para uma mistura. A equação de estado PVT mais simples é a equação dos gases ideais. Apesar de ser amplamente utilizada em situações práticas, sua aplicabilidade se restringe a gases em condições tais que as interações intermoleculares podem ser desprezadas.

O desenvolvimento de equações de estado para fluidos reais tem seguido quatro tipo de abordagens: equações do tipo virial e correlações generalizadas, equações de estado cúbicas, equações que usam o princípio dos estados correspondentes, e equações derivadas da termodinâmica estatística (MEIRELES; PEREIRA, 2013).

2.2.1.1. Equações de estado cúbicas

Em geral, o método mais usado para predição do equilíbrio em sistema contendo fluidos supercríticos utiliza equações cúbicas derivadas da equação proposta por van der Waals (VDW-EoS), em 1873. Dentre elas, destacam-se a equação de Soave-Redlich-Kwong (SRK-EoS) (1972) e Peng-Robinson (PR-EoS) (1976). Estas equações podem ser escritas de forma genérica como:

$$P = \frac{RT}{V-b} - \frac{a}{V^2 + uVb + wb^2} \tag{2.8}$$

sendo:

$$a = \Omega_a \frac{R^2 T_c^2}{P_c} \left\{ 1 + m \left[1 - \left(\frac{T}{T_c} \right)^{0,5} \right] \right\}^2 \tag{2.9}$$

$$b = \Omega_b \frac{R\ T_c}{P_c} \tag{2.10}$$

O primeiro termo das equações de estado cúbicas representa as forças repulsivas (termo repulsivo) e o segundo termo representa as forças atrativas (termo atrativo), sendo que *a* representa o parâmetro atrativo entre as moléculas e *b* é o parâmetro de correção de volume da equação (MEIRELES; PEREIRA, 2013). Os valores de u , w , Ω_a , Ω_b e m para cada equação de estado, são apresentados na **Tabela 2.1**.

Tabela 2.1 – Valores de u , w , Ω_a , Ω_b e m para diferentes equações de estado cúbicas: VDW, SRK e PR.

EoS	u	w	Ω_a	Ω_b	m
VDW	0	0	0,421875	0,125	0
SRK	1	0	0,42748	0,08664	$0{,}480 + 1{,}574\,\omega - 0{,}176\,\omega^2$
PR	2	-1	0,45724	0,07780	$0{,}3746 + 1{,}542\,\omega - 0{,}2699\,\omega^2$

A partir das Equações 2.9 e 2.10, verifica-se que a e b são definidos a partir de parâmetros dos componentes puros: temperatura crítica (T_c), pressão crítica (P_c) e fator acêntrico (ω).

As equações de estado podem ser escritas, de forma completamente equivalente, em termos do fator de compressibilidade ($Z = PV/RT$), as quais são explicitamente cúbicas. Uma forma geral das equações de estado cúbicas é apresentada a seguir:

$$Z^3 - (1 + B - uB)Z^2 + (A + wB^2 - uB - uB^2)Z - AB - wB^2 - wB^3 = 0 \tag{2.11}$$

onde:

$$A = \frac{aP}{R^2T^2} \tag{2.12}$$

$$B = \frac{aP}{R\ T} \tag{2.13}$$

Para estender as equações de estado para o cálculo de misturas binárias ou multicomponentes utilizam-se regras de mistura, que consistem em equações que expressam a dependência dos parâmetros da equação de estado com a composição da mistura. A mais utilizada é a regra de mistura clássica proposta por van der Waals, com dois parâmetros de interação binária, k_{ij} e l_{ij}.

$$a = \sum_i \sum_j a_{ij} y_i y_j \tag{2.14}$$

$$b = \sum_i \sum_j b_{ij} y_i y_j \tag{2.15}$$

Sendo:

$$a_{ij} = \sqrt{a_i a_j}\left(1 - k_{ij}\right) \tag{2.16}$$

$$b_{ij} = \frac{(b_i + b_j)}{2}(1 - l_{ij}) \tag{2.17}$$

2.2.1.2. *Cubic Plus Association*

A equação *Cubic Plus Association* (CPA-EoS), desenvolvida por Kontogeorgis e colaboradores (1996), consiste fundamentalmente na junção da equação SRK a um termo que aplica a teoria de perturbação de Wertheim para associação molecular, amplamente utilizada em equações do tipo SAFT (*Statistical Associating Fluid Theory*). Seu desenvolvimento foi motivado pela necessidade de modelar sistemas contendo compostos fortemente polares, pois estas espécies formam ligações de hidrogênio e, muitas vezes, apresentam um comportamento termodinâmico incomum por haver forte interação, seja entre moléculas da mesma espécie (auto-associação), ou entre moléculas de diferentes espécies (associação cruzada). Estas interações podem afetar fortemente as propriedades termodinâmicas dos fluidos. Assim, os equilíbrios químicos entre *clusters* devem ser levados em consideração a fim de desenvolver um bom modelo termodinâmico (HARTONO; KIM, 2004).

No modelo CPA-EoS, cada molécula de composto associativo é idealizada como contendo um ou mais sítios, por meio dos quais a mesma pode se ligar com outras moléculas presentes. Em termos do fator de compressibilidade, a equação CPA pode ser escrita como:

$$Z^{CPA} = Z^{SRK} + Z^{assoc} \tag{2.18}$$

A contribuição do fator de compressibilidade da equação de estado de Soave-Redlich-Kwong (SRK) é representada por:

$$Z^{SRK} = \frac{V}{V-b} - \frac{a}{RT(V+b)} \tag{2.19}$$

e a contribuição do termo de associação é dada por:

$$Z^{assoc} = -\frac{1}{2}(1+\rho\frac{\partial(\ln g)}{\partial\rho})\sum_i x_i \sum_{A_i}(1-X_{A_i}) \tag{2.20}$$

As letras *i* e *j* são usadas para indexar as moléculas, e as letras maiúsculas *A* e *B* são usados para indexar os sítios de ligação em uma determinada molécula *i*, x_i é a fração molar do componente *i*, ρ é a densidade molar do componente *i*, g é a função de distribuição radial e X_{Ai} é a fração molar do componente *i* não ligada ao sítio *A*, ou seja, a fração molar de monômeros. X_{Ai} é calculada a partir da resolução do sistema de equações formado por:

$$X_{Ai} = \frac{1}{1+\rho\sum x_j \sum X_{Bj}\Delta^{AiBj}} \tag{2.21}$$

onde $\Delta^{A_iB_j}$ é a força de associação entre dois sítios associativos de duas moléculas diferentes, calculado por:

$$\Delta^{A_iB_j} = g(\rho)\left[\exp\left(\frac{\varepsilon^{A_iB_j}}{RT}\right)-1\right]b_{ij}\beta^{A_iB_j} \tag{2.22}$$

Assim, dois parâmetros adicionais, referentes às contribuições associativas do composto, são necessários: a energia de associação entre os sítios de sua molécula (ε) e o parâmetro de volume de associação (β). Comparando esta equação com SRK-EoS e PR-EoS, a CPA-EoS apresenta complexidade matemática consideravelmente maior, uma vez que não é cúbica em volume molar e requer a obtenção de cinco parâmetros para os componentes puros: três provenientes do termo físico ($a = f(a_0 = \Omega_a \frac{R^2 T_C^2}{P_C}, m)$ e b) e dois para o termo associativo (ε e β).

Dentre outras funções já propostas, neste trabalho foi utilizada o modelo simplificado de esfera rígida para a função de distribuição radial g (KONTOGEORGIS et al., 1999):

$$g(\rho) = \frac{1}{1 - 1{,}9\eta} \tag{2.23}$$

$$\eta = \frac{1}{4} b\rho \tag{2.24}$$

Uma forma geral da CPA-EoS, para uma mistura de componentes, é apresentada na Equação 2.25. Os parâmetros a e b já foram definidos nas Equações 2.9 e 2.10 (usando as constantes para SRK-EoS apresentadas na **Tabela 2.1**).

$$P = \frac{RT}{V-b} - \frac{a}{V(V+b)} + \frac{RT}{2V}\left(1 + \frac{1}{V}\frac{\partial \ln(g)}{\partial \rho_i}\right)\sum_i x_i \sum_{A_j}(1 - X_{A_i}) \tag{2.25}$$

Além da regra de mistura aplicada ao termo cúbico da equação, uma regra de combinação é necessária para calcular os parâmetros do termo associativo quando a equação CPA é aplicada a misturas. Diversas regras de combinação (RC) já foram estudadas, sendo que suas nomenclaturas e/ou numerações variam de acordo com o autor. Destacam-se as apresentadas na **Tabela 2.2,** as quais foram empregadas e testadas para algumas misturas de compostos que apresentam associação. Porém, verificou-se que não há uma regra de combinação universal que represente de forma satisfatória todas as misturas (FOLAS et al., 2005; KONTOGEORGIS et al., 2006; VOUTSAS; YAKOUMIS; TASSIOS, 1999).

Como a caracterização de um sistema com a equação CPA é totalmente dependente de suposições quanto aos sítios de ligação presentes na molécula, é necessário o conhecimento das teorias que dizem respeito aos esquemas de associação, ou seja, que definem a quantidade e os tipos de sítios associativos. Usa-se normalmente uma classificação simplificada das moléculas de acordo com o número e tipos de sítios presentes, seguindo a terminologia de Huang e Radosz (1991). Neste sistema considera-se que uma molécula pode ser não associada (interagindo com as demais somente por forças de dispersão) ou pode apresentar um ou mais sítios de associação, e os sítios de associação podem ser de dois tipos: doador ou receptor de elétrons.

Tabela 2.2 – Regras de combinação para os parâmetros de energia (ε) e volume (β) de associação (VOUTSAS; YAKOUMIS; TASSIOS, 1999).

Regra de Combinação (RC)	Equações
RC-1	$\varepsilon^{A_iB_j} = \frac{1}{2}\left(\varepsilon^{A_i} + \varepsilon^{B_j}\right)$, $\beta^{A_iB_j} = \frac{1}{2}\left(\beta^{A_i} + \beta^{B_j}\right)$
RC-2	$\varepsilon^{A_iB_j} = \frac{1}{2}\left(\varepsilon^{A_i} + \varepsilon^{B_j}\right)$, $\beta^{A_iB_j} = \sqrt{\beta^{A_i}\,\beta^{B_j}}$
RC-3	$\varepsilon^{A_iB_j} = \sqrt{\varepsilon^{A_i}\,\varepsilon^{B_j}}$, $\beta^{A_iB_j} = \sqrt{\beta^{A_i}\,\beta^{B_j}}$
RC-4 (Regra de Elliot)	$\Delta^{A_iB_j} = \sqrt{\Delta^{A_i}\,\Delta^{B_j}}$

Boa capacidade descritiva da CPA-EoS foi verificada em diversos trabalhos visando à modelagem do equilíbrio de fases de sistemas relevantes na produção de biodiesel (FOLLEGATTI-ROMERO et al., 2012; OLIVEIRA et al., 2008, 2009, 2010, 2011, 2014a, 2014b, OLIVEIRA; QUEIMADA; COUTINHO, 2010a, 2010b).

Alguns modelos da CPA-EoS já foram ajustados para sistemas contendo compostos fenólicos. Mota et al. (2008) e Queimada et al. (2009) mediram a solubilidade de compostos fenólicos em água e ajustaram satisfatoriamente o modelo da CPA-EoS aos dados obtidos. A solubilidade de alguns fármacos também foi analisada em diferentes solventes orgânicos e, em geral, boa correlação dos dados foi obtida pela CPA-EoS (MOTA et al., 2011).

A modelagem do equilíbrio líquido-vapor do sistema CO_2/etanol/água foi realizada por Perakis et al. (2006) visando à sua predição. Resultados satisfatórios foram obtidos ao utilizar uma modificação da equação CPA original, na qual a PR-EoS foi utilizada como substituta da SKR-EoS.

2.3.Borra de Café

Por ser uma das bebidas mais populares do mundo e famosa por seu efeito estimulante, o café é uma das principais *commodities* brasileiras. Além disso, o Brasil é o maior produtor e exportador mundial de café, sendo o Estado de Minas Gerais responsável por mais de 50% da produção brasileira (CECAFE, 2017).

Das várias espécies no gênero *Coffea* (família *Rubiaceae*) somente duas são produzidas comercialmente: a *Coffea arabica* L. (café arábica) e a *Coffea canephora* (café robusta ou conilon) (ZUORRO; LAVECCHIA, 2013).

Como consequência do grande mercado mundial de café, as indústrias deste ramo são responsáveis pela geração de grandes quantidades de resíduos orgânicos, entre os quais destaca-se o material chamado de borra de café ou *spent coffee grounds* (SCG), que consiste no resíduo obtido durante o tratamento de grãos torrados e triturados de café com água quente ou vapor, para a preparação de café solúvel (BALLESTEROS; TEIXEIRA; MUSSATTO, 2014).

Este tipo de resíduo demanda alta quantidade de oxigênio durante a decomposição devido à presença de cafeína residual, taninos e polifenóis (VARDON et al., 2013). Com a crescente preocupação com o meio ambiente, diversas pesquisas têm sido necessárias no sentido de buscar alternativas para reutilização ou valorização desse resíduo de café.

Além de ser incorporado na ração animal e usado como adubo em plantações, o SCG pode ser usado para a produção de combustível em caldeiras industriais (SILVA et al., 1998), como substrato para o cultivo de microrganismos (MACHADO et al., 2012), fonte de compostos de alto valor (PANUSA et al., 2013), e como matéria-prima para a produção de álcool combustível (MUSSATTO et al., 2012), de carvão ativo (KANTE et al., 2012) e biodiesel (CAETANO et al., 2014; KONDAMUDI; MOHAPATRA; MISRA, 2008; VARDON et al., 2013). Além disso, há relatos dos extratos de SCG protegerem a pele contra o fotoenvelhecimento induzido por UVB (CHOI et al., 2016).

O SCG é um material lignocelulósico rico em açúcar, fibras alimentares e material lipídico, e com potencial antioxidante, associado à presença de compostos fenólicos (BALLESTEROS; TEIXEIRA; MUSSATTO, 2014). Portanto, este resíduo também pode ser explorado como matéria-prima de baixo custo para a obtenção de novos ingredientes funcionais de valor elevado para a indústria de alimentos e farmacêutica.

2.3.1. Compostos fenólicos do café

Os compostos fenólicos, também conhecidos como polifenóis, representam uma das principais classes de metabólitos secundários de plantas. Estes compostos têm sido muito estudados devido à sua influência na qualidade dos alimentos. Estruturalmente, os compostos fenólicos podem ser definidos como substâncias que possuem um anel aromático como um ou mais grupos hidroxila ligados diretamente a uma estrutura cíclica. Dependendo da sua estrutura, podem ser classificados em fenóis simples, ácidos fenólicos, flavonoides e taninos.

Os ácidos fenólicos caracterizam-se por terem um anel benzênico, um grupamento carboxílico e um ou mais grupamentos de hidroxila e/ou metoxila na molécula, conferindo propriedades antioxidantes tanto para os alimentos como para o organismo. Os dois principais grupos dos ácidos fenólicos são os ácidos benzóicos, que possuem sete átomos de carbono (ácidos salicílico, gentísico; p-hidroxibenzóico, protocatequínico, vanílico, gálico e siríngico) e os ácidos cinâmicos, que possuem nove átomos de carbono (ácidos cinâmico, o-cumárico, m-cumárico, p-cumárico, caféico, ferúlico e sinápico). As estruturas químicas básicas das duas principais classes de ácidos fenólicos são apresentadas na **Figura 2.2**. O tipo, número e posição das cadeias laterais ligadas ao anel aromático que diferenciam os ácidos fenólicos (SOARES, 2002).

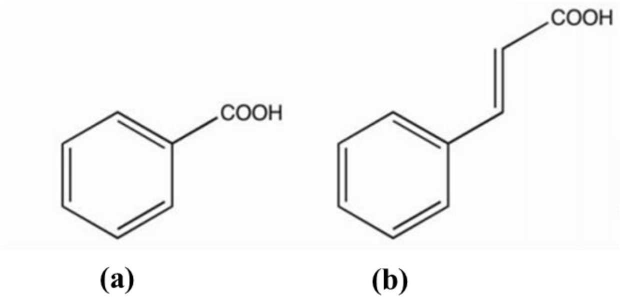

Figura 2.2 – Estrutura química básica: (a) ácidos benzoicos e (b) ácidos cinâmicos (KHODDAMI; WILKES; ROBERTS, 2013).

De acordo com Pérez-Hernández et al. (2012), o teor de compostos fenólicos totais em grãos de café verde, obtido em equivalente de ácido clorogênico, é 55,84 mg/g. Os ácidos clorogênicos (CGA) são os principais compostos fenólicos presentes em café. Eles são ésteres de ácidos trans-cinâmicos, tais como os ácidos cafeico, ferúlico e p-cumárico, com o ácido quínico. As principais classes de CGA em café são ácidos cafeoilquínicos (CQA), dicafeoilquínicos e feruloilquínicos, sendo os CQA os encontrados em maior quantidade (CLIFFORD, 1985).

O processo de torrefação influencia o teor de ácidos clorogênicos em grãos de café, assim como as condições de plantio e a espécie em questão (Arábica ou Robusta). De acordo com Clifford (1999), durante a torrefação, há uma destruição e transformação progressiva dos CGA, sendo perdidos de 8-10% para cada 1% de perda de matéria seca.

Em amostras de SCG, o teor de compostos fenólicos totais, expresso em equivalente de ácido gálico, varia de 2,65 a 35,5 mg/g, dependendo dos solventes utilizados para extração e da espécie do café (BRAVO et al., 2013; PANUSA et al., 2013; ZUORRO; LAVECCHIA, 2012). Panusa et al. (2013) verificaram que o café Robusta apresenta maior teor de CGA comparado ao café Arábica.

2.3.2. Compostos lipídicos do café

Os compostos lipídicos são substâncias caracterizadas pela sua baixa solubilidade em água, reflexo da natureza hidrofóbica das suas estruturas químicas. Em geral, os óleos vegetais representam um importante produto no mercado mundial, com ampla variedade de áreas econômicas de interesse, desde a nutrição humana e animal até a produção de bioenergia (GRACIA et al., 2009).

O teor de lipídios em grãos de café é pouco influenciado pelo processo de torrefação. Segundo Mussato et al. (2011), o teor de óleo em grãos crus pode variar de 7% a 17%, enquanto para grãos torrados o teor varia de 11% a 17% (ESQUIVEL; JIMÉNEZ, 2012).

A borra de café apresenta alto teor de lipídios, pois estes compostos não são eficientemente extraídos em meio aquoso durante a produção de café solúvel. Porém, os valores podem ser bastante diferentes de acordo com o *blend* formado. Segundo Kondamudi et al. (2008), a borra de café apresenta ~15% de óleo, porém há trabalhos que relatam 2,3% (b.s) (BALLESTEROS; TEIXEIRA; MUSSATTO, 2014), 6,0% (b.s) (CAETANO et al., 2014) e 12,5% (b.s) (CRUZ et al., 2012). Esses lipídios incluem principalmente triacilgliceróis (75%), esteróis (5%) e diterpenos (19%), além de pequena fração de tocoferóis (CRUZ et al., 2012).

A composição do óleo de café em ácidos graxos é bastante similar em diversos trabalhos, não havendo grandes diferenças entre a composição majoritária do óleo extraído dos grãos e o óleo extraído da borra de café. O principal ácido graxo presente em óleo extraído de borra de café é o ácido linoleico (C18:2), representando ~44%, seguido pelo ácido palmítico (C16:0) com ~32%, o ácido oléico (C18:1) com ~10%, e o ácido esteárico (C18:0) com ~7% (CRUZ et al., 2012). Esta alta homogeneidade na fração lipídica é relevante para a otimização de processos para a reutilização deste resíduo, como na produção de biodiesel.

De acordo com Segall et al. (2005), os triacilgliceróis majoritários presentes em óleo de café são dilinoleil-palmitoil-glicerol (PLL, 25,90%), dipalmitoil-linoleil-glicerol (PPL, 13,74%), trilinoleína (LLL, 11,76%), palmitoil-oleil-linoleil-glicerol (POL, 8,76%), palmitoil-linoleil-estearil-glicerol (PLS, 8,73%) e dilinoleil-estearil-glicerol (SLL, 8,28%).

O óleo de café não é comumente utilizado como óleo comestível, mesmo quando extraído de grãos crus ou torrados de boa qualidade. Este fato ocorre provavelmente devido à presença de grande quantidade de material insaponificável, que oferece problemas na etapa de refino (HARTMAN et al., 1968).

Caveol e cafestol são os dois principais ésteres terpênicos presentes nos lipídios de café, cujas aplicações são relatadas por Tsukui et al. (2014). Tocoferóis e tocotrienóis, constituintes da vitamina E, também estão presentes no café (GONZÁLEZ et al., 2001). Quanto aos esteróis, estão presentes em borra de café: colesterol (~0,5%), campesterol (~16%), estigmasterol (~25%), sitosterol (~49%) e Δ5-avenasterol (~7%) (LAGO; ANTONIASSI, 2001).

2.3.3. Tecnologia supercrítica aplicada à borra de café

Alguns estudos envolvendo SFE de borra de café (SCG) são encontrados na literatura. Além da extração do óleo de SCG (AHANGARI; SARGOLZAEI, 2013; AKGÜN et al., 2014; COUTO et al., 2009; DE MELO et al., 2014), também há estudos visando à co-extração de óleo e diterpenos (ACEVEDO et al., 2013; BARBOSA et al., 2014), produção de polihidroxialcanoatos (poliéster) a partir do óleo extraído por SFE (CRUZ et al., 2014) e obtenção de extratos com atividade antioxidante (ANDRADE et al., 2012). Além disso, Andrade e colaboradores (2012) identificaram presença dos ácidos protocatecuico, clorogênico, hidroxibenzóico e caféico nos extratos de SCG obtidos com CO_2 supercrítico, enquanto no extrato obtido com CO_2 + etanol apenas ácido clorogênico foi detectado.

O óleo obtido por SFE apresenta algumas variações na composição em ácidos graxos, quando comparada com o óleo de SCG extraído com hexano, porém nos dois tipos de extrato verificou-se que os ácidos graxos majoritários foram os ácidos linoleico, palmítico, oleico e esteárico (AHANGARI; SARGOLZAEI, 2013).

Contudo, verificou-se que não há na literatura científica estudos utilizando a tecnologia de fracionamento dos compostos presentes no café utilizando o dióxido de carbono como antissolvente supercrítico ou por extrações sequenciais em dois ou mais estágios.

2.4. Solubilidade de Solutos em Dióxido de Carbono Supercrítico

Por ser relevante para o design de processos de separação que aplicam a tecnologia supercrítica, a solubilidade de uma extensa gama de compostos em CO_2 supercrítico já foi estudada. Em particular, desde que em condições próximas ao ponto crítico, a densidade do solvente pode mudar por um fator de dez com apenas pequenas mudanças na temperatura e na pressão. Dessa forma, fica claro que a solubilidade do soluto também pode variar por um grande fator (ALVAREZ et al., 2005).

Dados de solubilidade de solutos em $scCO_2$ apresentam um comportamento típico, que pode ser verificado pela relação entre o logaritmo da solubilidade e o logaritmo da densidade do $scCO_2$ (exemplo na **Figura 2.3a**). Em muito casos, essa relação apresenta um comportamento linear, sendo bem descrito pela equação proposta por Chrastil (CHRASTIL, 1982). Além disso, é comum verificar que a solubilidade aumenta com o aumento da pressão à temperatura constante, e o cruzamento das isotermas de solubilidade, conhecido como *Crossover Pressure,* que representa o valor da pressão em que a solubilidade se torna independente da temperatura (exemplo na **Figura 2.3b**). O conhecimento deste ponto de cruzamento pode ser de grande utilidade quando são necessárias pequenas diferenças na seletividade, como na separação de isômeros (SHI; KASSAMA; KAKUDA, 2007).

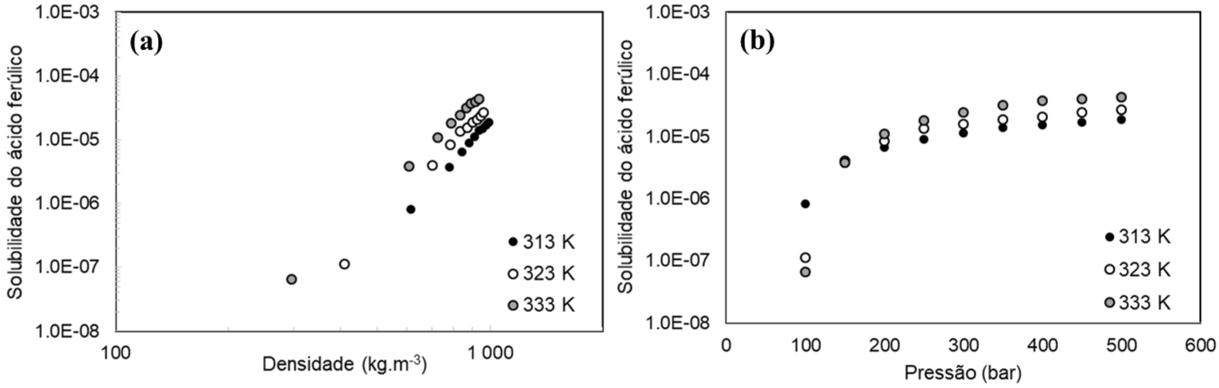

Figura 2.3 – Comportamento da solubilidade do ácido ferúlico em $scCO_2$: (a) relação linear entre logaritmo da solubilidade e logaritmo da densidade do CO_2; (b) *crossover pressure* (Murga et al., 2003).

No geral, as equações de estado cúbicas são os modelos termodinâmicos mais utilizados para descrever sistemas em altas pressões, como as equações de estado de Peng-Robinson e Soave-Redlich-Kwong. Além disso, uma extensa lista de modelos empíricos ou semi-empíricos é encontrada para correlacionar dados de solubilidade em $scCO_2$.

2.4.1. Solubilidade de compostos fenólicos

O estudo da solubilidade de compostos fenólicos em $scCO_2$ foi realizado por diversos autores. Para os ácidos fenólicos, por exemplo, os valores de solubilidade são muito baixos, da ordem de 10^{-7} e 10^{-9}, em fração molar (MURGA et al., 2003, 2004; SUN; LI; QUAN, 2005). Portanto, verifica-se que a utilização de cossolventes, como etanol e/ou água, para manipular a polaridade da mistura $scCO_2$/etanol/água é necessária para a extração destes compostos.

Diante da diversidade de compostos de interesse para serem avaliados, o estudo da solubilidade de compostos fenólicos em misturas de CO_2 e cossolvente são escassos. Dentre os compostos fenólicos que se destacam por suas atividade antioxidante, catequina (BERNA et al., 2001), resveratrol (BERNA; CHÁFER; MONTÓN, 2001), epicatequina (CHÁFER et al., 2002), quercetina (CHÁFER et al., 2004), ácido gálico (CHÁFER et al., 2007), ácido cinâmico (CHÁFER et al., 2009) e curcumina (PAVIANI et al., 2013; ZHAN et al., 2017) já foram avaliados. As equações de estado utilizadas para modelagem do equilíbrio de fases nestes trabalhos foram: Peng-Robinson, Soave-Redlich-Kwong e a Equação de Estado por Contribuição de Grupos (CG-EoS).

A solubilidade do ácido ferúlico em misturas de CO_2 e etanol também foi avaliada, porém a modelagem dos dados experimentais não foi feita por equações de estado cúbicas. Neste caso apenas um modelo empírico foi obtido (SOVOVÁ, 2001).

2.4.2. Solubilidade de óleo e compostos lipídicos

Dados de solubilidade de óleos ou compostos lipídicos em $scCO_2$ são mais frequentes na literatura, comparados com os dados de compostos fenólicos, principalmente porque compostos lipofílicos apresentam alta solubilidade em $scCO_2$. Porém, a modelagem termodinâmica desses dados é menos frequente. A dificuldade de modelagem desses sistemas está no fato de que óleos vegetais são misturas complexas, contendo diversos compostos. Portanto, abordagens de pseudo-componentes que consideram um óleo formado apenas pelos componentes majoritários devem ser adotadas.

O equilíbrio de fases do óleo de girassol + etanol + CO_2 foi investigado por Herna et al. (2008) e as composições experimentais obtidas foram correlacionadas com um modelo baseado na equação CG-EoS, considerando o óleo de girassol como um sistema ternário. Além disso, estudos mostram que a solubilidade de compostos lipídicos em $scCO_2$ aumenta com a adição de etanol (CATCHPOLE et al., 1998; COCERO; CALVO, 1996; SALDAÑA et al., 2010). Quanto aos ácidos graxos, a solubilidade em misturas de $scCO_2$ e cossolventes dos ácidos láurico, mirístico, palmítico e esteárico foram reportadas (GARLAPATI; MADRAS, 2008, 2008).

Especificamente, a solubilidade do óleo de café em dióxido de carbono supercrítico foi avaliada por De Azevedo et al. (2008), Sandi et al. (2012) e Cornelio-Santiago et al. (2017), porém os valores relatados são muito diferentes entre si.

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CAPÍTULO 3. FERULIC ACID SOLUBILITY IN SUPERCRITICAL CARBON DIOXIDE, ETHANOL AND WATER MIXTURES

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Ferulic acid solubility in supercritical carbon dioxide, ethanol and water mixtures

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Abstract

Ferulic acid (FA) is a phenolic compound present in many natural extracts. To obtain high FA extraction yields using supercritical carbon dioxide ($scCO_2$), the extraction should be performed in the presence of co-solvents. This work reports FA solubility data in $scCO_2$/ethanol/water mixtures and its thermodynamic modeling using the Peng-Robinson equation of state. FA solubility in $scCO_2$/ethanol was measured by a dynamic method at different temperatures (313, 323 and 333 K), pressures (20, 30 and 40 MPa) and molar ratios of the $scCO_2$/ethanol mixture (95:5 and 90:10). FA solubility in $scCO_2$ with ethanol and water was evaluated at 323 K and 30 MPa, at three different molar ratios of the $scCO_2$/ethanol/water mixture (97.4:2:0.6, 93.5:5:1.5 and 88:9.3:2.7). Also measured was the solubility data of FA in water and in ethanol at ambient pressure. By using 10% ethanol, the FA solubility in $scCO_2$ + ethanol attained values 120 times higher than FA solubility in pure $scCO_2$ and about 50 times lower than solubility in ethanol. The lowest value was obtained at 313 K and 20 MPa and the highest at 333 K and 40 MPa (1.9×10^{-3}); however, the co-solvent effect decreased when increasing pressure. Finally, the presence of the co-solvent is very important to obtain natural extracts rich in ferulic acid. The prediction of FA solubility in ternary and quaternary systems using the binary interaction parameters did not produce good results. Calculated values were smaller than the experimental values, probably due to the strong interactions between acid molecules and alcohol.

Keywords: Ferulic acid, solubility, supercritical carbon dioxide, co-solvents, Peng-Robinson equation of state.

1. Introduction

Research on natural products with therapeutic benefits to the consumer has gained prominence in recent years. These products are usually plant extracts rich in bioactive compounds such as phenolic compounds. Phenolic compounds, also known as polyphenols, are secondary metabolites of different vegetable species. Depending on their structure, they can be subdivided in different classes of compounds, including simple phenols, phenolic acids, flavonoids and tannins [1]. Phenolic acids are characterized by having a benzene ring, a carboxylic acid group and one or more hydroxyl and/or methoxyl groups in the molecule. The structural characteristics of these acids confer to them antioxidant properties, enhancing the health benefits and contributing to human body well-being [2].

Among the phenolic compounds is ferulic acid (FA), whose structure is shown in Figure 1. FA is one of the most abundant phenolic acids in nature and it is present in grains (oats, barley, rye, corn and wheat), fruits (grape, orange, banana, tomatoes), vegetables (beets, spinach, radish) and drinks (coffee, yerba mate tea, orange juice) [3–6]. This compound exhibits a variety of biomedical, pharmaceutical and industrial applications [7], and its presence is associated with anti-inflammatory, antimicrobial, anti-allergic and anti-cancer activities in addition to the prevention of cardiovascular diseases [7–9].

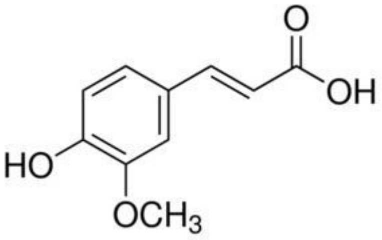

FIGURE 1. Chemical structure of ferulic acid.

Supercritical technology is an interesting alternative widely used to obtain natural extracts. Extractions using supercritical carbon dioxide ($scCO_2$) have some advantages over traditional extraction techniques, since they generate extracts free of toxic residues. Furthermore, carbon dioxide is cheap, inert, non-flammable and presents low critical temperature and pressure values, making attractive for extraction and fractionation of thermosensitive compounds. However, extraction with $scCO_2$ is appropriate to obtain nonpolar or low polarity compounds, while the use of co-solvents is necessary to extract compounds of higher polarity.

Knowledge on solubility of compounds in pure $scCO_2$ or in $scCO_2$ with the presence of co-solvents is important for optimizing the processes employing $scCO_2$ as a solvent or antisolvent. The behavior of ferulic acid in $scCO_2$ was studied in prior works [10–12] and solubility values reported are relatively low, between 10^{-6} and 10^{-5} in mole fraction. Therefore, the use of co-solvents is required when extracts enriched in this compound are desired. Ethanol and water are the major co-solvents used in supercritical extraction because they are also “generally recognized as safe” (GRAS) and “green”.

Equilibrium data of systems containing phenolic compounds, CO_2 and ethanol are rare. Studies with catechin [13], resveratrol [14], epicatechin [15], quercetin [16] and trans-cinnamic acid [17] are reported in literature. Moreover, thermodynamic models commonly used to describe systems containing supercritical fluids are derived from cubic equation proposed by van der Waals in 1873, such as the Soave-Redlich-Kwong [18] and Peng-Robinson [19] equations.

In this context, the present paper provides a phase equilibrium study of the $scCO_2$ (1) - ethanol (2) - water (3) - ferulic acid (4) system and evaluates thermodynamic modeling using the Peng-Robinson equation by fit of the binary systems and the ability of this equation to predict multicomponent system data.

2. Experimental

2.1.Materials

Suppliers and purity of the materials used in this work are listed in Table 1. None of them was subjected to further purification. Ferulic acid purchased from Sigma-Aldrich is the trans-4-hydroxy-3-methoxycinnamic acid, which has CAS number 537-98-4.

TABLE 1

Materials used in this work, its suppliers and mass fraction purity.

Compound	Supplier/Source	Mass fraction purity[a]
Ferulic acid	Sigma-Aldrich	0.99
Carbon Dioxide	White Martins Gases Industriais	0.995
Ethanol	Synth	0.995[b]
Water	Milli-Q® Direct–Q 3 UV	> 0.9999[c]

[a]As reported by the supplier;

[b]Volumetric fraction purity.

[c]Based on supplier's specification: total organic carbon < 10 ppb.

2.2.Experimental procedure

2.2.1. Measurements of solubility in ethanol and water

FA solubility in ethanol and water was evaluated at atmospheric pressure (94.3 kPa) using a 50 mL glass cell surrounded by a jacket with water circulation from a thermostatic bath to maintain the temperature. The system was prepared by adding a sufficient content of solute to a fixed volume of solvent, thus securing the presence of excess solute. The mixture was stirred with a magnetic stirrer (1 h) and allowed to settle until complete separation of the phases (24 h). The times of agitation and rest were defined in preliminary tests. Supernatant samples were collected using a syringe and solubility was calculated as the ratio between the mass of the solute and the mass of the solution collected.

2.2.2. Measurements of solubility in $scCO_2$/ethanol/water mixtures

The experiments were performed using a dynamic method similar to that described by Lopes et al. [20] in an extraction unit as schematized in Figure 2. The unit consists of a CO_2 cylinder, a thermostatic bath, pressure gauges (Bourdon), a heat exchanger, a high pressure pump, a supply tank, an extractor, a flask for collecting samples, a gas flow meter, a volume totalizer and a co-solvent pump, as well as needle and retention valves.

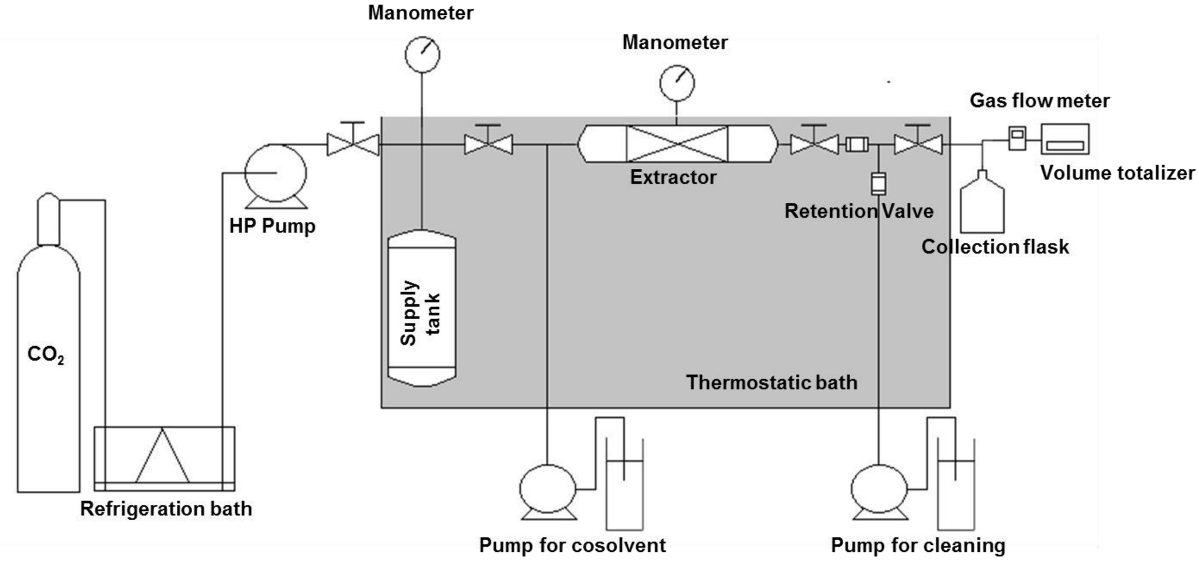

FIGURE 2. Layout of the experimental apparatus for determination FA solubility in $scCO_2$/ethanol/water mixtures.

The extractor, whose capacity is about 100 mL (~ 30 cm length), was filled with FA mixed with glass spheres (~ 5mm diameter). The extractor was pressurized with CO_2 to achieve the select pressure (20, 30 or 40 MPa), while the thermostatic bath kept the temperature controlled (313, 323 or 333 K). The system was stabilized in static mode overnight. After about 1.5 h of extraction under desired conditions of temperature, pressure and CO_2/ethanol proportion, the samples were collected every 16 g of spent CO_2, measured by volume totalizer. Such prior extraction time is required to create the proportion of ethanol and CO_2 within the extractor. The CO_2 flow rate was fixed around 1.1×10^{-5} kg s^{-1} (0.4 L min^{-1} measured at ambient conditions), which it was considered by preliminary tests sufficient to ensure equilibrium during the solubility measurements.

In fact, the preliminary tests included different extractor's volume and length, and different solvent flow rates. Considering that the initial equilibrium time in the static mode overnight lasted at least 12 h and the solubility values measured in the first samples and in subsequent samples were the same within the experimental error, equilibrium conditions were obtained in all experiments.

Different $scCO_2$/ethanol/water mixtures were generated by varying the co-solvent flow rate (0.1, 0.05 or 0.02 mL min^{-1}). Ethanol or a previously prepared mixture of ethanol and water (90:10 w/w) was used as the co-solvent.

At the end of the experiments, the samples were concentrated in rotary evaporator (Marconi, MA-120 model, Brazil) and kept in an oven (Marconi, MA030/12 model, Brazil) under vacuum at 323 K until constant weight. Finally, the FA solubility was calculated as the ratio between the FA mass collected and the mass of all solvents consumed during extraction.

2.2.3. FA quantification

The gravimetric method was used for quantification of FA in the samples; however, for comparison some samples were analyzed by spectrophotometry (UV-VIS Spectrophotometer Aquamate Orion 8000, Thermo Scientific, USA). Analyses performed at 322 nm, after obtaining a calibration curve. The difference between measurements obtained by the two methods was around 2.9%.

2.3.Thermodynamic modeling

FA solubility (y_{FA}) in all solvents was calculated by Equation 1, which represents the equality of solute fugacity in the solid phase and solute fugacity in solution (light phase). Note that Equation 1 requires the following hypotheses: ferulic acid in the solid phase remains as a pure solid, not forming any solid solution; the molar volume of the solute can be considered constant within the pressure range of interest; and pure ferulic acid in the vapor phase behaves as an ideal gas [17].

$$y_{FA} = \frac{P_{FA}^{sat} \exp\left(\frac{\underline{V}_{FA}\left(P - P_{FA}^{sat}\right)}{RT} \right)}{\phi_{FA}^{light\ phase} P} \tag{1}$$

P_{FA}^{sat} is the saturation pressure of FA obtained by Saldaña et al. [21] at different temperatures, $\underline{V}_{FA}$ is the molar volume of FA and $\phi_{FA}^{light\ phase}$ represents the fugacity coefficient of FA in the light phase which is calculated from the Peng-Robinson equation of state (PR-EoS) (Equation 2).

$$P = \frac{RT}{\underline{V}_{mix} - b_{mix}} - \frac{a_{mix}}{\underline{V}_{mix}\left(\underline{V}_{mix} + b_{mix}\right) + b_{mix}\left(\underline{V}_{mix} - b_{mix}\right)} \tag{2}$$

Van der Waals classical mixing rule was used:

$$a_{mix} = \sum_i \sum_j a_{ij} y_i y_j \quad , \quad a_{ij} = \sqrt{a_i a_j}\left(1 - k_{ij}\right) \tag{3}$$

$$b_{mix} = \sum_i \sum_j b_{ij} y_i y_j \quad , \quad b_{ij} = \frac{\left(b_i + b_j\right)}{2}\left(1 - l_{ij}\right) \tag{4}$$

k_{ij} and l_{ij} are the binary interaction parameters of energy and size, respectively. These parameters were calculated by minimizing an objective function (S) (Equation 5) which evaluates the difference between the experimental and calculated values of solubility. This procedure was performed in Matlab® software using the Simplex method [22], along with a routine for calculating the phase equilibrium.

$$S(k_{ij}, l_{ij}) = \sum_{j=1}^{N} \sum_{i=1}^{2} \left[\left(\frac{y_{cal}(i) - y_{\exp}(i)}{y_{\exp}(i)} \right)^2 \right]_j \tag{5}$$

The parameters a_i and b_i are dependent on the critical properties and acentric factor of pure substances. The values used in this work are presented in Table 2.

TABLE 2

Pure component physical properties.

Number	Compound	M/g mol^{-1}	T_C/K	P_C/MPa	ω	$\underline{V}$/cm^3 mol^{-1}
1	CO_2	44.01	304.1^a	7.38^a	0.225^a	-
2	Ethanol	46.07	513.9^a	6.14^a	0.644^a	-
3	Water	18.02	647.3^a	22.12^a	0.344^a	-
4	Ferulic acid	194.18	854.6^b	3.64^b	1.194^c	176.64^d

^{a}Poling et al. [23];

^{b}Calculated by Mota et al. [24] using the Marrero-Gani method [25];

^{c}Calculated by Moncada et al. [26] using the Clapeyron Equation;

^{d}Calculated by Moncada et al. [26].

3. Results

3.1. FA solubility in ethanol and water

FA solubilities in ethanol and water were measured at nine different temperatures (293 to 333 K) and the results are shown in Table 3. The values of solubility in ethanol ranged from 0.023 to 0.058 (mole fraction), therefore an increase of approximately 150% was observed when increasing the temperature from 293 to 333 K. FA solubility data in ethanol is scarce in the literature. Buranov and Mazza [27] reported FA solubility in ethanol at room temperature of 65 mg mL^{-1} (≈ 0.0195 mole fraction). Zhuo et al. [28] evaluated FA solubility in ethanol within the temperature range of 288 to 318 K, and under the same conditions the data of this work was 11% higher on average than that obtained by Zhuo et al..

The FA solubility in water presented lower values than FA solubility in ethanol by an order of magnitude of 10^{-5} on a molar fraction. The values obtained are 38% smaller on average than those of Mota et al. [24], and very different from those obtained by Noubigh et al. [29].

TABLE 3

FA solubility in ethanol and water at atmospheric pressure (P = 94.3 kPa) [a].

T/K	$100y^{e}_{FA}$ [b]	$u(100y^{e}_{FA})$	$100000y^{w}_{FA}$ [c]	$u(100000y^{w}_{FA})$
293	2.34	0.02	4.6	0.2
298	2.60	0.01	4.9	0.1
303	2.92	0.01	6.2	0.3
308	3.07	0.03	7.9	0.5
313	3.51	0.06	11.3	0.4
318	4.05	0.03	13.5	0.2
323	4.5	0.1	17.1	0.3
328	4.9	0.1	22.0	0.4
333	5.8	0.1	27	1

[a] Standard uncertainties *u* are *u*(T/K) = 0.05 and *u*(P/kPa) = 0.2;

[b] Mole fraction of FA in ethanol;

[c] Mole fraction of FA in water.

Figure 3 presents the comparison between FA solubility data in ethanol and water obtained in this work and those already reported in the literature. The data from Mota et al. [24] were transformed from g L^{-1} to mole fraction considering a dilute solution of ferulic acid, so water density values were used at different temperatures. It is probable that the deviation between the Noubigh et al. data [29] occurs due to differences in the polymorphic form of ferulic acid. According Sohn and Oh [30], ferulic acid has two polymorphic forms, which have different solubility in water. The polymorphic forms exhibit differences in DSC thermograms: form I has an endotherm peak corresponding to the melting point at 448.15 K and form II has three endothermic peaks at 378.15, 421.15 and 455.15 K.

The ferulic acid used in this work was analyzed by DSC with a heating rate of 10 $K.min^{-1}$ within a range from 293.15 to 493.15 K. Its DSC thermogram showed one endothermic peak (form I) with melting temperature at 448.43 K. Mota et al. [24] also confirmed the form I of their ferulic acid crystal.

Evaluating the influence of temperature on FA solubility, the results indicate that FA solubility in water is more temperature dependent than its solubility in ethanol, reaching values up to 6 times higher by increasing the temperature from 293 K to 333 K (4.6×10^{-5} to

27.1×10^{-5}). This FA behavior is similar to vanillin [31] and the phenolic compounds of grape [32].

Comparing these results with FA solubility data in pure $scCO_2$, it was observed that FA solubility in water is about 10 times higher than FA solubility in $scCO_2$, while FA solubility in ethanol is up to 1000 times higher [12]. This result confirms the importance of using co-solvents, such as ethanol and water, to obtain extracts with high FA content. Furthermore, it is important to cite that Buranov and Mazza [27] evaluated FA solubility in mixtures of ethanol and water, and observed a high value in the ethanol/water 80:20 (v/v) mixture. These results suggest the use of an ethanol/water mixture as a co-solvent.

3.2.FA solubility in the $scCO_2$/ethanol mixture

FA solubility in $scCO_2$ + ethanol was measured at different temperatures (313, 323 and 333 K), pressures (20, 30 and 40 MPa) and molar ratios of the $scCO_2$/ethanol mixture (approximately 95:5 and 90:10), as shown in Table 4. The lowest solubility values were obtained at 313 K and 20 MPa, independent of ethanol concentration. The highest solubility values were obtained at 323 K and 30 MPa using 5% ethanol (4.3×10^{-4}) and at 333 K and 40 MPa with 10% ethanol (1.9×10^{-3}). Sovová [10] reported some values of FA solubility in $scCO_2$ at lower pressures with up to 3.3% ethanol and the corresponding results are consistent with those of the present work.

FA solubility in the mixture of $scCO_2$/ethanol 90:10 reached values 120 times higher than solubility in pure $scCO_2$ (333 K and 20 MPa) and 50 times lower than FA solubility in ethanol. Additionally, when using 10% ethanol as a co-solvent, FA solubility presented values on average 3.5 times higher than FA solubility in water at the same temperature.

There are few approaches that attempt to explain the significant increase in solubility of this type of compound in $scCO_2$ + co-solvent, as a consequence of the increase in solvent density caused by introduction of the co-solvent in the system. According to Huang et al. [33], dipole-dipole interactions also promote a considerable co-solvent effect. However, the most significant increase in solubility is due to strong hydrogen bonds formed between the solute and the alcohol molecules [33–36].

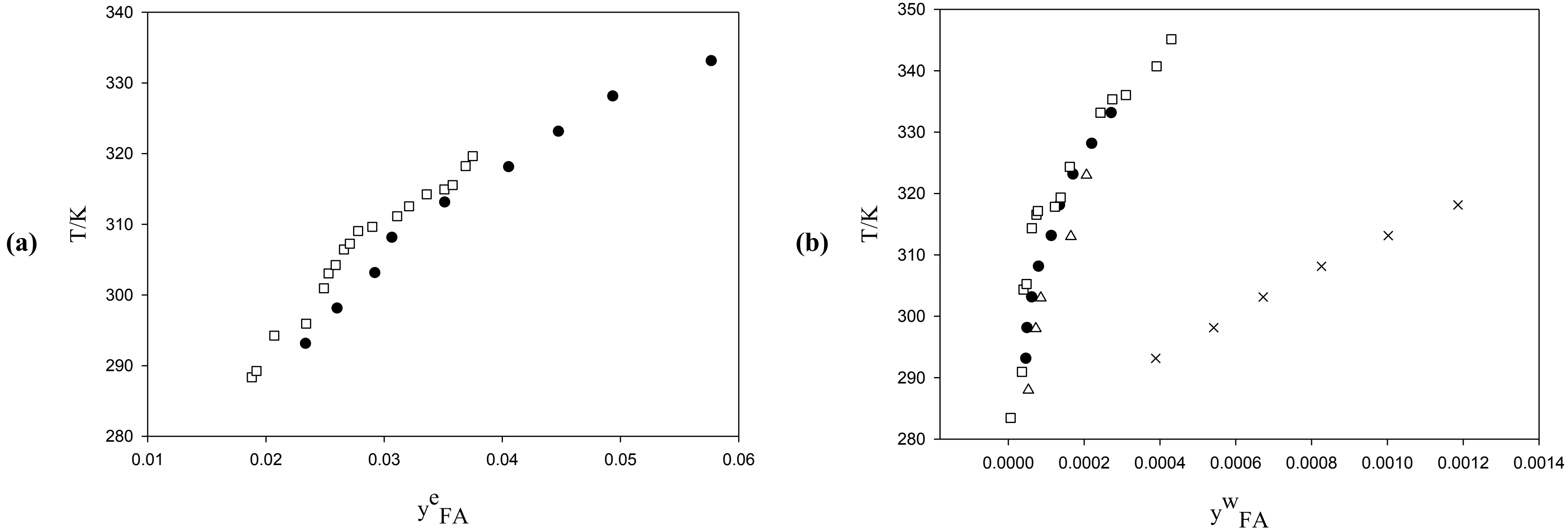

FIGURE 3. FA solubility in (a) ethanol and (b) water: ● this work; □ Zhuo et al. [28]; △ Mota et al. [24]; × Noubigh et al. [29]. y_{FA}: mole fraction of ferulic acid.

TABLE 4

FA solubility in $scCO_2$/ethanol mixtures[a].

T/K	P/MPa	E/% [b]	$10000y_{exp}$ [c]	$u(10000y_{exp})$	$10000y_{cal}$ [d]
313	20	5.78	1.89	0.02	1.41
		10.92	4.9	0.8	3.36
	30	5.37	2.8	0.4	1.99
		10.58	5.6	0.1	4.95
	40	5.45	2.92	0.08	3.29
		10.62	7	1	7.84
323	20	5.45	2.60	0.05	1.74
		10.91	8	1	5.03
	30	5.43	4.30	0.07	3.24
		10.52	10.4	0.9	7.60
	40	5.48	3.9	0.2	5.39
		10.34	8.4	0.7	11.31
333	20	5.48	3.36	0.08	1.13
		11.70	13	1	9.06
	30	5.46	4.0	0.1	2.44
		10.59	16.5	0.8	11.37
	40	5.51	3.87	0.05	4.41
		10.43	19	2	21.11

[a] Standard uncertainties u are $u(T/K) = 0.05$, $u(P/MPa) = 0.1$ and $u(E/\%) = 0.18$;

[b] Molar percentage of ethanol on a solute-free basis;

[c] Experimental mole fraction of FA in $scCO_2$ + ethanol;

[d] Calculated mole fraction of FA in $scCO_2$ + ethanol using the new PR-EoS binary interaction parameter (k_{ij}) for ethanol – FA. (Table 7).

Figure 4a shows the experimental data at 323 K to facilitate visualization of the effect of pressure on solubility values. By increasing the ethanol fraction in the solvent mixture, the effect of pressure becomes less significant, as indicated by the similar solubility values obtained at $Y_{ET} = 0.10$ for different pressures. This behavior was also found by Catchpole et al. [36]. On the other hand, the effect of temperature on solubility data remains relevant when adding ethanol as a co-solvent (Figure 4b).

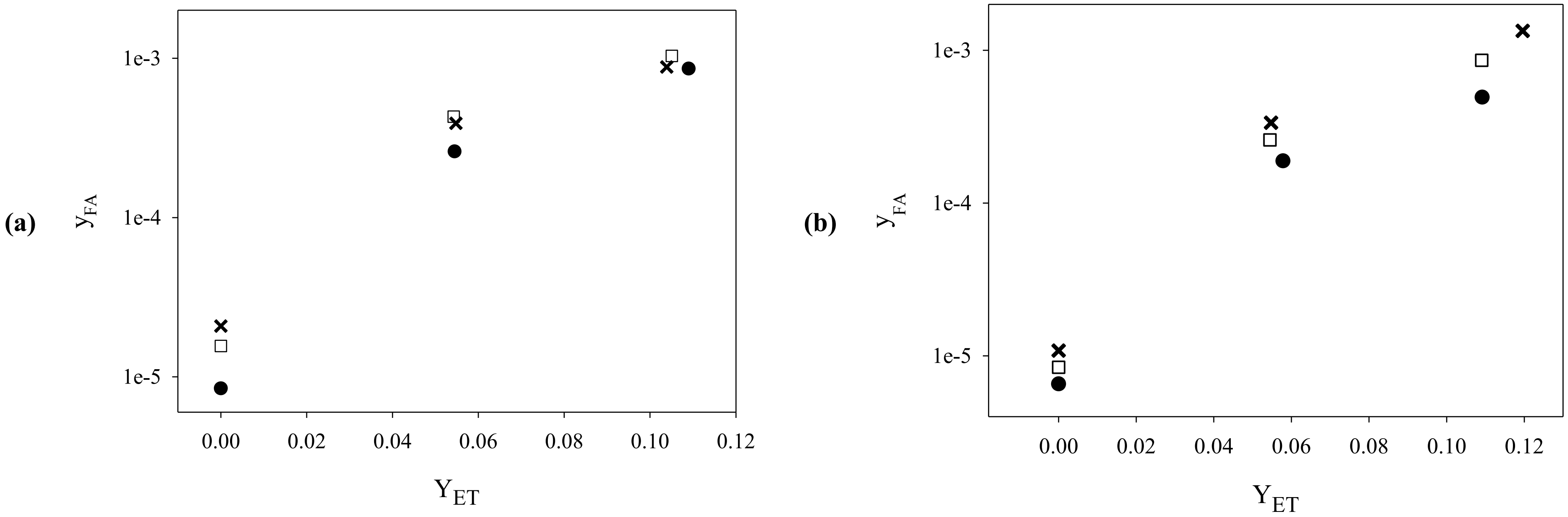

FIGURE 4. FA solubility (y_{FA}) in scCO$_2$/ethanol mixtures at (a) 323 K: ● 20 MPa; □ 30 MPa; ×40 MPa; and (b) 20 MPa: ● 313 K; □ 323 K; × 333 K. FA solubility data in pure scCO$_2$ was obtained from Murga et al. [12]. Y_{ET}: mole fraction of ethanol on a solute-free basis.

Also examined was the effect of the co-solvent (CE), which was calculated by the ratio of solubility in the ternary system (in $scCO_2$/ethanol) and solubility in a binary system (in $scCO_2$) under the same conditions of temperature and pressure (Equation 6).

$$CE = \frac{y_{FA}^{CO2+ETHANOL}}{y_{FA}^{CO2}} \tag{6}$$

The results for the mixture containing 5% ethanol are shown in Figure 5. It was observed that addition of 5% of the co-solvent increased FA solubility by up to 30 times compared with FA solubility in pure $scCO_2$. At lower pressure the highest co-solvent effect was observed at higher temperatures. However, at higher pressures the higher co-solvent effect was obtained at the lowest temperature (313 K). This behavior was also observed when using 10% ethanol (chart not shown). The inversion of the co-solvent effect was also verified in the case of aspirin solubility data in $scCO_2$ + ethanol and $scCO_2$ + methanol [33].

Furthermore, Figure 5 shows the effect of pressure on FA solubility using ethanol as a co-solvent. As noted earlier, when increasing the pressure there was a decrease in the co-solvent effect. This decrease occurs due to variation in the co-solvent local composition around the solute molecules. In fact, the local composition seems to approach the bulk concentration and the effect of the co-solvent-solute interactions are less pronounced at higher pressures [33].

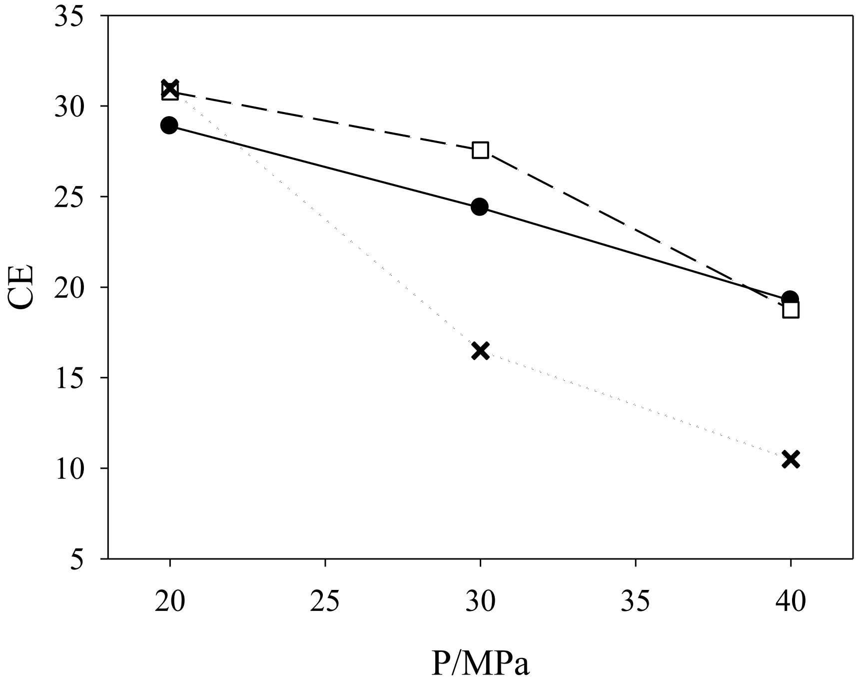

FIGURE 5. Co-solvent effect in $scCO_2$/ethanol 95:5 mixture: ● 313 K; □ 323 K; ×333 K.

3.3. FA solubility in the $scCO_2$/ethanol/water mixture

The FA solubility in $scCO_2$/ethanol/water mixtures was evaluated at 323 K, 30 MPa and three different molar ratios of the $scCO_2$/ethanol/water mixture (97.4:2:0.6, 93.5:5:1.5 and 88:9.3:2.7) (Table 5). Although results in literature indicate higher FA solubility in the ethanol/water 80:20 (v/v) mixture [27], the ratio of ethanol and water was set based on vapor-liquid equilibrium data of the CO_2-ethanol-water system [37]. Therefore, to guarantee the formation of a single fluid phase, the ethanol/water ratio of 90:10 (w/w) was used for all experiments.

TABLE 5

FA solubility in $scCO_2$ + ethanol + water at T = 323 K and P = 30 MPa[a].

E/% [b]	W/% [c]	$10000y_{exp}$ [d]	$u(10000y_{exp})$	$10000y_{cal}$ [e]
2.07	0.59	0.77	0.03	0.280
5.04	1.43	3.5	0.1	0.653
9.37	2.66	10.6	0.2	2.089

[a] Standard uncertainties *u* are *u*(T/K) = 0.05, *u*(P/MPa) = 0.1, *u*(E/%) = 0.05 and *u*(W/%) = 0.02;

[b] Molar percentage of ethanol on a solute-free basis;

[c] Molar percentage of water on a solute-free basis;

[d] Experimental mole fraction of FA in $scCO_2$ + ethanol + water;

[e] Calculated mole fraction of FA in $scCO_2$ + ethanol + water using the PR-EoS binary interaction parameters shown in Table 6.

For the same content of co-solvents (ethanol and/or water), FA solubility in $scCO_2$ in presence of ethanol and water is slightly lower than the solubility when using only ethanol as a co-solvent (Figure 6). Under the conditions tested, this result indicates that it is more interesting to use ethanol as a co-solvent to obtain extracts with ferulic acid. There are a few studies when use a mixture of ethanol and water as a co-solvent to extract phenolic compounds. For example, Paes et al. [38] and Zulkafli et al. [39] identified an ethanol/water mixture as the best co-solvent for the extraction of anthocyanins from blueberry and phenolic compounds from bamboo leaves, respectively. These results confirm the relevance of the information on solubility data of phenolic compounds in $scCO_2$ with co-solvents for developing the corresponding extraction processes; however, they are still rare in the literature.

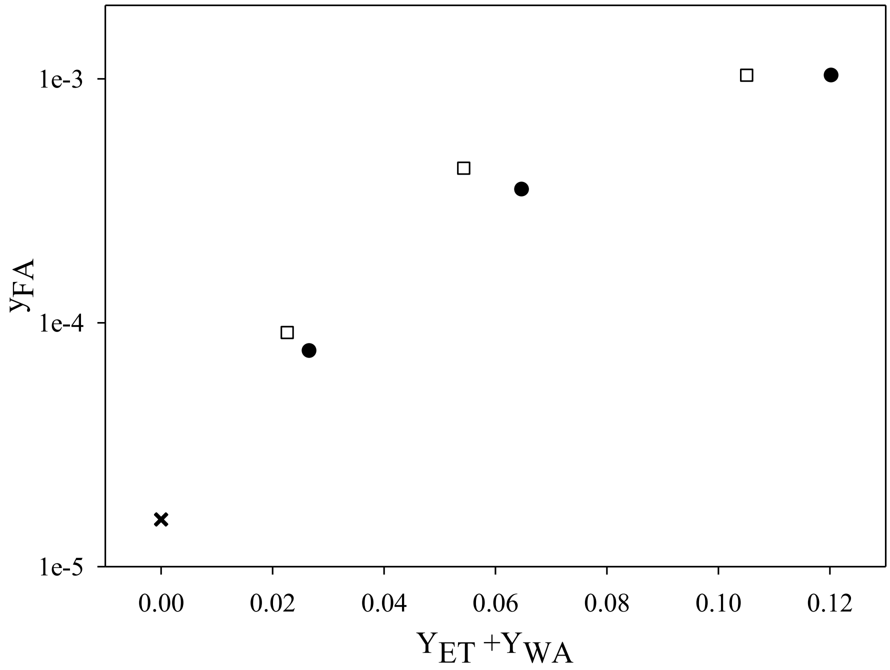

FIGURE 6. Comparison of FA solubility (y_{FA}) in $scCO_2$/ethanol and $scCO_2$/ethanol/water at 323 K and 30 MPa: ● ethanol and water as co-solvents; □ ethanol as a co-solvent; × without co-solvent. Y_{ET} and Y_{WA}: mole fraction of ethanol and water on a solute-free basis, respectively.

3.4. Thermodynamic modeling

Firstly, the binary interaction parameters (k_{ij} and l_{ij}) involving all the system components were obtained in order to predict FA solubility in $scCO_2$/ethanol/water mixtures. The binary interaction parameter k_{ij} (with $l_{ij} = 0$) was fitted for ethanol (2) - ferulic acid (4) and water (3) - ferulic acid (4) systems. A linear dependence of this parameter was obtained with temperature as follows: $k_{24} = -4.8034\times10^{-4}*T/K + 0.3894$ ($r^2 = 0.998$) and $k_{34} = -3.3779\times10^{-4}*T/K + 0.1395$ ($r^2 = 0.976$). The other binary interaction parameters used are shown in Table 6.

TABLE 6

PR-EoS binary interaction parameters for the CO_2 (1) – ethanol (2) – water (3) – FA (4) system.

Pair i-j	Parameter	T/K			Reference
		313	323	333	
12	k	0.0923	0.0877	0.0830	[40]
	l	-0.00505	-0.01317	-0.02128	
13	k	0.2611	0.1910	0.1209	[40]
	l	0.2175	0.1702	0.1229	
14	k	0.5511	0.5277	0.5404	Data from [12][a]
	l	0.3314	0.2941	0.3721	
23	k	0.0282	0.0231	0.0181	[40]
	l	0.1318	0.1257	0.1195	
24	k	0.2389	0.2341	0.2293	This work
	l	0	0	0	
34	k	0.0337	0.0303	0.0269	This work
	l	0	0	0	

[a] Values readjusted using data ranged from 20 to 40 MPa.

The prediction of FA solubility data in $scCO_2$/ethanol using the PR-EOS thermodynamic model did not produce good results. The calculated values varied within a range of 3 and 14 times lower than the experimental data. For this reason, these calculated solubility values were not shown. Catchpole et al. [36] also obtained values 20 times lower when predicting the urea solubility data in mixtures of $scCO_2$/ethanol using the PR-EoS.

According to Ting et al. [34], the addition of polar co-solvents may form specific interactions such as hydrogen bonding between the solute and the co-solvent molecules, resulting in a greater increase in solubility because there are not only non-specific interactions. Therefore, other interactions occur in this type of system in addition to the attraction and repulsion forces that a cubic equation of state such as PR-EoS takes into account.

The prediction of FA solubility in $scCO_2$/ethanol/water showed values up to 5.4 times lower than the experimental data (mean absolute deviation about 75%) (Table 5). The inclusion of water in the system adds further complexity to the molecular interactions and justifies the deviation observed in the predication.

An alternative for better data prediction using the PR-EoS would consider the complex formed between solute-ethanol as an additional component in the system, as performed by some authors [34,36]. However, for this procedure it is necessary to obtain the equilibrium constant of the reaction for formation of this complex. In this work, a new binary interaction parameter between ferulic acid and ethanol was adjusted based on FA solubility data in the $scCO_2$/ethanol mixture (fixing the remaining parameters), similar to the procedure conducted by Huang et al. [33]. There was an important dependence of the parameter with concentration of ethanol in the solvent mixture and the system temperature (Table 7). The average deviation between experimental and calculated data using these new parameters was 28.5% (Table 4).

TABLE 7

New PR-EoS binary interaction parameters for the ethanol – FA system.

C/% [a]	Parameter	T/K		
		313	323	333
5	k_{24}	-0.1648	-0.2744	-0.1119
10	k_{24}	0.0639	-0.0028	-0.0170

[a]Molar percentage of ethanol on a solute-free basis.

It was emphasized in this work that like in other studies cited [16,33], the binary interaction parameters were fitted from experimental data of the ternary system. However, it is very important to obtain models able to predict the behavior of multicomponent systems based on the binary interaction parameters. The SAFT equation was satisfactory to predict the systems containing aromatic compounds, $scCO_2$ and ethanol [41]. Another promising alternative is the Cubic-Plus-Association equation [42] which consists of combining the cubic equation (Soave-Redlich-Kwong) and SAFT equation.

4. Conclusion

Experimental data of ferulic acid solubility in $scCO_2$ in the presence of ethanol and water are reported in this work. Small contents of ethanol or the ethanol/water mixture as a co-solvent (≈ 10%) produced a considerable increase in FA solubility compared to pure $scCO_2$. Furthermore, FA solubility presented higher values when pure ethanol was used as a co-solvent.

Although the fitting of the binary interaction parameter present low deviations between experimental and calculated data, prediction of the ternary and quaternary systems did not produce good results, indicating the need for using other equation of state more appropriate for systems with co-solvents and strong polar interactions.

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CAPÍTULO 4. SOLUBILITY OF CAFFEIC ACID IN CO_2 + ETHANOL: EXPERIMENTAL AND PREDICTED DATA USING CUBIC PLUS ASSOCIATION EQUATION OF STATE

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Solubility of caffeic acid in CO_2 + ethanol: experimental and predicted data using Cubic Plus Association Equation of State

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Abstract

This work evaluated the solubility of caffeic acid (CA) in mixtures of supercritical carbon dioxide ($scCO_2$) and ethanol at different temperatures (313, 323 and 333 K), pressures (20, 30 and 40 MPa) and concentrations of ethanol (2.2, 5.4 and 10.2 mol%). The Soave-Redlich-Kwong and Cubic Plus Association (CPA) equations of state were used to correlate the binary mixture data and to predict the ternary system data. CA solubility in ethanol is approximately 10^6 times higher than its solubility in pure $scCO_2$. By using 10.2 mol% ethanol in $scCO_2$, CA solubility increased 30,000 times at 313 K and 20 MPa. Both models provided reasonable descriptions of the experimental data for the binary systems. However, CPA-EoS can better describe the strong interactions between acid molecules and ethanol, and can predict that the addition of small amounts of ethanol to $scCO_2$ provides a large increase in CA solubility.

Keywords: 3,4-dihydroxycinnamic acid; supercritical carbon dioxide; cosolvent; phase equilibrium; thermodynamic modeling; Soave-Redlich-Kwong.

1. Introduction

Caffeic acid, or 3,4-dihydroxycinnamic acid (CA), is an abundant phenolic acid that represents between 75% and 100% of the total content of hydroxycinnamic acids in many fruits [1]. CA is rarely found in its free form. For instance, the ester formed by caffeic and quinic acids is one of the chlorogenic acids found in fruits (blueberries, kiwis, plums, cherries, apple, ciders), vegetables, wine, olive oil, and, in high concentrations, in coffee [1,2]. Thus, this compound is present in significant amounts in the human diet. The CA structure (**Fig. 1**) is characterized by having a benzene ring, a carboxylic acid group and two hydroxyl groups in the molecule, conferring it antioxidant properties [3]. Besides its antioxidant activity, CA presents as well anti-ischemia, anti-thrombosis, anti-hypertension, anti-fibrosis, anti-virus and anti-tumor properties [4-6].

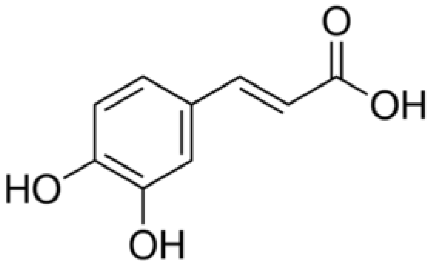

Fig. 1. Caffeic acid chemical structure.

The production of natural extracts rich in phenolic compounds, such as CA, is interesting for food, chemical and pharmaceutical industries. Extractions using supercritical carbon dioxide ($scCO_2$) as a solvent have some advantages over traditional extraction techniques. Mainly, it is a flexible and selective process that provides clean extracts, i.e. free of toxic residues. Furthermore, carbon dioxide is cheap, available in high purity, non-toxic, non-flammable and "Generally Recognized as Safe" (GRAS) for food applications.

Knowledge on the solubility of phenolic compounds in $scCO_2$ is very important for design and optimization of processes aiming to obtain natural extracts rich in phenolic compounds using supercritical technology. However, CA presents low solubility in $scCO_2$ (approximately 10^{-8} and 10^{-9} in mole fraction [7]) and consequently low extraction yields are obtained using this solvent. The low solubility of CA in $scCO_2$ is associated with the high polarity of this compound. Thus, its extraction from vegetable sources using $scCO_2$ should be performed in the presence of polar co-extractants to increase the solvent polarity. Ethanol is the most used cosolvent because it is also considered a "green solvent". It is thus important to know the CA solubility in CO_2/ethanol mixtures. Other phenolic compounds with antioxidant properties had their solubility previously reported in CO_2 + ethanol [8-17].

The thermodynamic modeling of these systems makes possible the prediction of data for multicomponent systems and at different conditions of temperature and pressure from those experimentally studied. Cubic state equations, such as Peng-Robinson (PR-EoS) and Soave-Redlich-Kwong (SRK-EoS), are the most commonly used to describe systems under high pressures. However, the prediction of the phase equilibrium of systems containing phenolic compounds, $scCO_2$ and ethanol using these equations does not provide good results [16], because the presence of polar cosolvents may induce specific interactions, such as hydrogen bonding between the solute and the cosolvent, which a cubic equation of state cannot properly take into account [18]. Most of the prior works on solubility of biomolecules in mixtures of CO_2 and cosolvent correlated the data by fitting binary interaction parameters to the experimental data of the ternary system [19,20]. Only Ting et al. [21] and Yang and Zhong [22] attempted the prediction of solubility data of ternary systems.

One promising model for polar and associating solvents is the Cubic-Plus-Association Equation of State (CPA-EoS), which consists in the combination of a cubic equation (Soave-Redlich-Kwong) and the association term proposed by Wertheim and widely used in SAFT-type (Statistical Associating Fluid Theory) equations of state [23]. Compared with SAFT, CPA-EoS is a simpler yet accurate model for associating mixtures, while keeping the well-known advantages of cubic equations of state. The CPA-EoS has already been successfully used to describe the CA solubility in water and other biomolecules in organic solvents [24,25] and in the liquid-vapor equilibrium modeling of the CO_2/ethanol/water system [26].

This work addresses the measurement of the solubility data of caffeic acid in mixtures of supercritical carbon dioxide and ethanol and evaluates the ability of Cubic-Plus-Association Equation of State (CPA-EoS) to predict the behavior of these systems.

2. Experimental

2.1.Chemicals

Caffeic acid with purity ≥ 98% (CAS Number 331-39-5, Sigma-Aldrich, USA), carbon dioxide 99.5% w/w (White Martins Gases Industriais, Brazil) and ethanol 99.5% (Synth, Brazil) were used.

2.2.Solubility measurements

2.2.1. Caffeic acid - ethanol system

Solubility measurements of caffeic acid in ethanol were performed using a jacketed glass cell coupled to a thermostatic bath (Polystat, Cole Parmer, USA). The system was assembled adding excess solute to a fixed volume of solvent. The mixture was stirred for 1 h with a magnetic stirrer and then allowed to equilibrate and phase separate for 24 h. A syringe was used to collect the sample of the upper phase which was then weighted (Accurate Analytical Balance with precision +/- 0.0001 g, model XT 220A, Sweden). Finally, the sample was concentrated in a rotary vacuum evaporator (Marconi rotary evaporator, model MA-120, Brazil) and kept in an oven (Marconi, model MA030/12, Brazil) under vacuum (Marconi vacuum pump model MA057/1, Brazil) at 323 K until constant mass. Solubility in mass fraction was calculated by the ratio of the solute mass and the solution mass collected by syringe.

2.2.2. Caffeic acid - CO_2 - ethanol system

The solubility of caffeic acid in CO_2 - ethanol mixtures was measured by a dynamic method in a supercritical extraction unit, as detailed in studies published recently [27,28]. The investigation was conducted at different conditions of temperature (313, 323 and 333 K), pressure (20, 30 and 40 MPa), and amounts of ethanol (2 mol%, 5 mol% and 10 mol% on a solute free-basis).

An extractor with 100 mL capacity (32 cm length) was manually packaged with caffeic acid mixed with glass beads (5 cm diameter). Measured in the end of the process (atmospheric conditions), the CO_2 flow rate was around 0.4 L/min and the ethanol flow rate varied according to the desired mixture at 0.02, 0.05 and 0.1 mL/min. In previous tests, the CO_2 flow rate 0.4 L/min was found to be low enough to ensure that phase equilibrium was achieved during solubility measurements. Considering the local atmospheric conditions as 298 K and 0.095 MPa, the values for the density of ethanol and carbon dioxide were assumed as 0.78 g/mL and 1.65 g/L, respectively.

The CO_2 amount used for each measurement depended on the proportion CO_2/ethanol mixture evaluated to guarantee sufficient sample mass to be well quantified: 5, 10 and 20 L of CO_2 were used for the solubility measurements with ethanol concentrations of 10, 5 and 2.5 mol% ethanol, respectively.

The samples were concentrated in a rotary vacuum evaporator (Marconi, model MA-120, Brazil) and kept in an oven (Marconi, model MA030/12, Brazil) under vacuum (Marconi, model MA057/Brazil) at 323 K until constant weight. The samples were stored at 247 K until quantification.

2.3. Quantification of caffeic acid

The quantification of caffeic acid was performed by spectrophotometry (UV-VIS spectrophotometer Aquamate Orion 8000, Thermo Scientific, USA) at 217 nm. For that purpose, calibration curves with solutions of known concentrations of caffeic acid varying from 0.001 to 0.01 mg/mL were prepared ($R^2 > 0.99$).

3. Thermodynamic modeling

The CA solubilities in CO_2, ethanol, and CO_2 + ethanol were calculated using the equation of equilibrium based on the equality of solute fugacity in the solid and solution (light) phases. Assuming a pure solid phase and no significant temperature dependence on the difference between the heat capacity of the pure liquid and solid (ΔC_p), **Equation 1** [29] is obtained.

$$y_{CA} = \frac{\varphi^{liquid}_{pure\,CA}}{\varphi^{liquid}_{CA}} \exp\left[-\frac{\Delta H}{R}\left(\frac{1}{T} - \frac{1}{T_m}\right) + \frac{\Delta C_p}{R}\left[\frac{T_m}{T} - 1 - \ln\left(\frac{T_m}{T}\right)\right] - \left[\frac{(P - P_{atm})\Delta V_{CA}}{RT}\right]\right] \tag{1}$$

where y_{CA} is the solute mole fraction solubility; *R* is the gas constant; *T* is the absolute temperature; *T*T_m is the melting temperature; φ is the coefficient of fugacity; ΔH is the phase transition enthalpy; $\Delta C_{p,CA} = C^{liquid}_{p,CA} - C^{solid}_{p.CA}$ and $\Delta V_{CA} = V^{liquid}_{CA} - V^{solid}_{CA}$.

Fugacity coefficients were calculated from the Soave-Redlich-Kwong or the Cubic Plus Association equations of state. In both cases, the binary interaction parameters were estimated by minimizing an objective function (*OF*) (**Equation 2**) which evaluates the difference between the experimental and calculated values of solubility.

$$OF(k_{ij}) = AARD\,y\,(\%) = \frac{100}{N}\sum_{j=1}^{N}\left[\frac{|y_{cal} - y_{\exp}|}{y_{\exp}}\right]_j \tag{2}$$

3.1.Soave-Redlich-Kwong Equation of State (SRK-EoS)

The SRK-EoS has the following form:

$$P = \frac{RT}{V - b} - \frac{a}{V(V + b)} \tag{3}$$

where *V* is molar volume; *a* and *b* are the pure component parameters of energy and co-volume, respectively. A Soave-type temperature dependency of the energy parameter is used:

$$\mathrm{a}(T) = a_0[1 + c_1(1 - \sqrt{T/T_c})]^2 \tag{4}$$

The *a* and *b* parameters are calculated by using the critical properties and the acentric factor (ω) of the pure components, as presented in **Table 1**. For mixtures, the parameters were calculated using the classical van der Waals one-fluid mixing rules, with one binary interaction parameter k_{ij}.

$$a = \sum_i \sum_j a_{ij} y_i y_j \qquad a_{ij} = \sqrt{a_i a_j}\,(1 - k_{ij}) \tag{5}$$

$$b = \sum_i b_i y_i \tag{6}$$

3.2.Cubic Plus Association Equation of State (CPA-EoS)

As mentioned previously, the CPA equation of state results from the combination of the SRK equation and the Wertheim association term [23]. In CPA, each associative compound is conceptualized as having one or more sites through which it may bond to other molecules, and the EoS can thus be described as follows:

$$P=\frac{RT}{V-b}-\frac{a}{V(V+b)}+\frac{RT}{2V}\left(1+\frac{1}{V}\frac{\partial \ln(g)}{\partial \rho_i}\right)\sum_i x_i \sum_{A_j}(1-X_{A_i}) \tag{7}$$

where i and j are used to index the molecules, and the capital letters A and B are used to index the bonding sites on a particular molecule i; x is the mole fraction; ρ is the molar density; g is a radial distribution function and X_{Ai} is the mole fraction of component i not bonded at site A;

The X_{Ai} is calculated by solving the following set of equations:

$$X_{Ai}=\frac{1}{1+\rho\sum x_j \sum X_{Bj}\Delta^{AiBj}} \tag{8}$$

where $\Delta^{A_iB_j}$ is the association strength between two sites belonging to two different molecules.

$$\Delta^{A_iB_j}=g(\rho)\left[\exp\left(\frac{\varepsilon^{A_iB_j}}{RT}\right)-1\right]b_{ij}\beta^{A_iB_j} \tag{9}$$

The simplified hard-sphere radial distribution function (g) was used [30]:

$$g(\rho)=\frac{1}{1-1{,}9\eta} \tag{10}$$

$$\eta=\frac{1}{4}b\rho \tag{11}$$

Apart from the three pure component parameters (a_0, c_1 and b) for non-associating components, two more parameters, related to the associative contribution of the compound, are required: the energy of association between the sites of molecule (ε) and the parameter of association volume (β). These five parameters are obtained simultaneously from fitting the equation to vapor pressure and liquid density data of pure compounds.

For mixtures, the classical van der Waals mixing rule are used (**Equations 5 and 6**) for the *a* and *b* parameters. For the association parameters several combining rules have been suggested [31]. The Elliot combining rule and CR-2 [32] are the most used. For many mixtures, both of these combining rules provide accurate descriptions of the phase equilibria. In this work the CR-2 (**Equations 12 and 13**) was used for the mixtures studied.

$$\beta^{A_i B_j} = \sqrt{\beta^{A_i}\, \beta^{B_j}} \tag{12}$$

$$\varepsilon^{A_i B_j} = \frac{1}{2}\left(\varepsilon^{A_i} + \varepsilon^{B_j}\right) \tag{13}$$

Table 1

Pure component physical properties.

Compound	M (g/mol)	T_C (K)	P_C (MPa)	ω	T_m (K)	ΔH_f (J/mol)
CO_2	44.01	304.13[a]	7.377[a]	0.223[a]		
Ethanol	46.07	513.92[a]	6.137[a]	0.643[a]		
Caffeic acid	180.16	993.05[b]	5.809[b]	1.056[c]	469[c]	13638.78[c]

[a] Multiflash™ database [33]; [b] Murga et al. [7]; [c] Saldaña et al. [34].

It is not rare to find **Equation 1** in a simplified form, in which only the term with ΔH is considered, since the other terms may be neglected when compared to that one. In this way, for comparison purposes and in order to evaluate the influence of ΔC_p and ΔV on the modeling performance, two different approaches for SRK (SRK-1 and SRK-2) and four approaches based on the CPA model (CPA-1 to CPA-4) were tested. Details on the differences will be presented and discussed in section 4.3.

4. Results and discussion

4.1. Caffeic acid - ethanol system

One of the objectives of this work is to predict CA solubility in CO_2 + ethanol. For this analysis, it is important to obtain experimental phase equilibrium data for all the binary systems involved in the ternary mixture. For this reason, phase equilibrium measurements were conducted for the system caffeic acid – ethanol.

CA solubility in ethanol was measured at nine temperatures in the range from 293 K to 333 K, as shown in **Table 2** and **Fig. 2.** CA solubility doubles by increasing the temperature from 293 K to 333 K. Zhang et al. [35] evaluated CA solubility in ethanol/water mixtures, and the solubility at 298 K and 90.05% ethanol was 0.0621 in mass fraction (0.016 in mole fraction). This value is consistent with those obtained in this work, since Zhang et al. [35] shows that CA solubility should not increase when adding more ethanol in the ethanol/water mixture. Recently, Ji et al. [36] also evaluated the solubility of caffeic acid in ethanol. The average absolute deviation between their data and the results presented in this work is approximately 25%. In addition, CA solubility in water present lower values than the solubility in ethanol, varying from 5×10^{-5} to 3×10^{-4} in mole fraction for temperatures within the range 288 K to 323 K [25], and CA solubility in ethanol is approximately 10^6 times higher than CA solubility in $scCO_2$ [7]. All these results attest the importance of using cosolvents, such as ethanol, to obtain extracts with high CA content.

Table 2

CA solubility in ethanol (mole fraction) at atmospheric pressure.

T/K	$x_{CA} \times 10^2$
293.15	1.41 ± 0.05
298.15	1.51 ± 0.08
303.15	1.58 ± 0.03
308.15	1.73 ± 0.05
313.15	1.86 ± 0.05
318.15	2.13 ± 0.05
323.15	2.3 ± 0.1
328.15	2.5 ± 0.2
333.15	2.8 ± 0.1

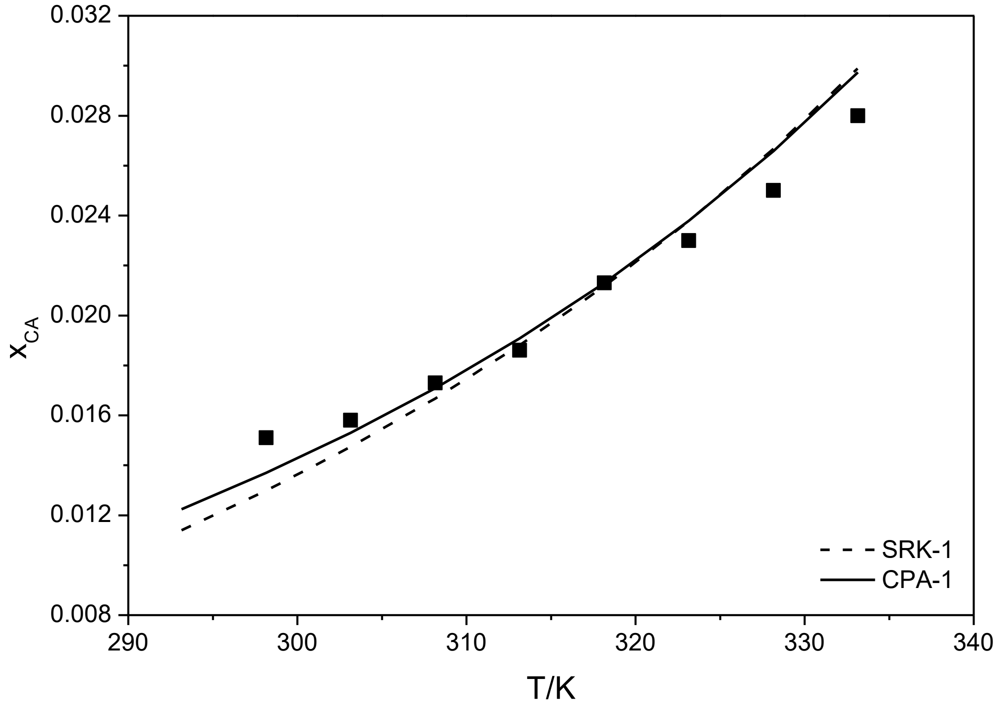

Fig. 2. Thermodynamic modeling of the CA solubility in ethanol (x_{CA}) at atmospheric pressure using SRK-1 and CPA-1 models.

4.2. Caffeic acid - CO_2 - ethanol system

The CA solubility in CO_2 and ethanol mixtures was evaluated at three values for temperatures (313, 323 and 333 K), pressures (20, 30 and 40 MPa) and concentrations of ethanol (2.2, 5.4 and 10.2 mol%), as shown in **Table 3**. There are no available data in the literature for this mixture. The lowest CA solubility values were obtained at 313 K and 20 MPa, for all the proportions of ethanol studied. However, the conditions of temperature and pressure that presented the greatest solubility varied according to the amount of ethanol in the CO_2/ethanol mixture. For example, using 2.2 mol% ethanol the highest CA solubility was found at 333 K and 30 MPa. On the other hand, with 10.2 mol% ethanol in the CO_2/ethanol mixture the greatest values were found at higher pressures (40 MPa).

Although the CA solubility in the mixture CO_2 + ethanol was still much lower than in pure ethanol, the solubility of CA in $scCO_2$ + 10.2% ethanol reached values 10^5 times higher than its solubility in pure $scCO_2$ (313 K and 20 MPa) [7]. As mentioned in previous works, the increase in solubility using cosolvent occurs mainly due to the strong hydrogen bonds formed between the solute and the alcohol molecules [18,37-39]. However, the increase in the density of the solvent mixture with the addition of cosolvent in the system and the dipole-dipole type interactions can also be responsible for this behavior [38].

Other phenolic compounds for which the solubility in CO_2 + ethanol was previously reported did not present such a huge increase in solubility when adding ethanol. This is probably due to the fact that CA solubility in $scCO_2$ is rather small and caffeic acid is a fairly polar compound.

Although the highest solubility values were observed at the higher temperature and pressure conditions, the influence of temperature and pressure decreases when the quantity of cosolvent in the mixture is increased. The dependencies of the solubility with temperature and pressure are shown in **Fig. 3** and **Fig. 4**, in which the proximity of the solubility data is noteworthy when using 10.2 mol% ethanol. Similar behavior was also found for solubility of urea [39] and ferulic acid [16].

The experimental data suggests a logarithmic dependency of the solubility data with ethanol concentration. The increase in CA solubility due to the increase in the concentration of ethanol from 5.4 mol% to 10.2 mol% is not as important as the increase from 0 to 2.2 mol% ethanol. Then, an increase on ethanol concentration above 10.2 mol% should not promote a significant further increase on the CA solubility. Thus, the use of greater amounts of ethanol should be avoided, since this ethanol addition would increase the extract production cost, because of the purification step associated with the necessity of cosolvent evaporation after extraction.

Table 3

Mole fraction CA solubility in CO_2 and ethanol mixtures (y_{CA}). Y_{et}: mole percentage of ethanol on a solute free-basis;

Y_{et} (%)	T/K	P/MPa	$y_{CA}\times10^6$
2.23 ± 0.01	313.15	20	0.58 ± 0.08
		30	0.8 ± 0.2
		40	1.7 ± 0.4
	323.15	20	1.9 ± 0.2
		30	0.86 ± 0.06
		40	0.92 ± 0.09
	333.15	20	1.2 ± 0.2
		30	4.8 ± 0.9
		40	2.6 ± 0.3
5.37 ± 0.03	313.15	20	6 ± 1
		30	14 ± 1
		40	37 ± 5
	323.15	20	22 ± 2
		30	30 ± 3
		40	36 ± 2
	333.15	20	11 ± 2
		30	59 ± 3
		40	54 ± 3
10.2 ± 0.1	313.15	20	60 ± 2
		30	69 ± 5
		40	77 ± 4
	323.15	20	66 ± 5
		30	80 ± 1
		40	90 ± 5
	333.15	20	44 ± 3
		30	84 ± 4
		40	91 ± 4

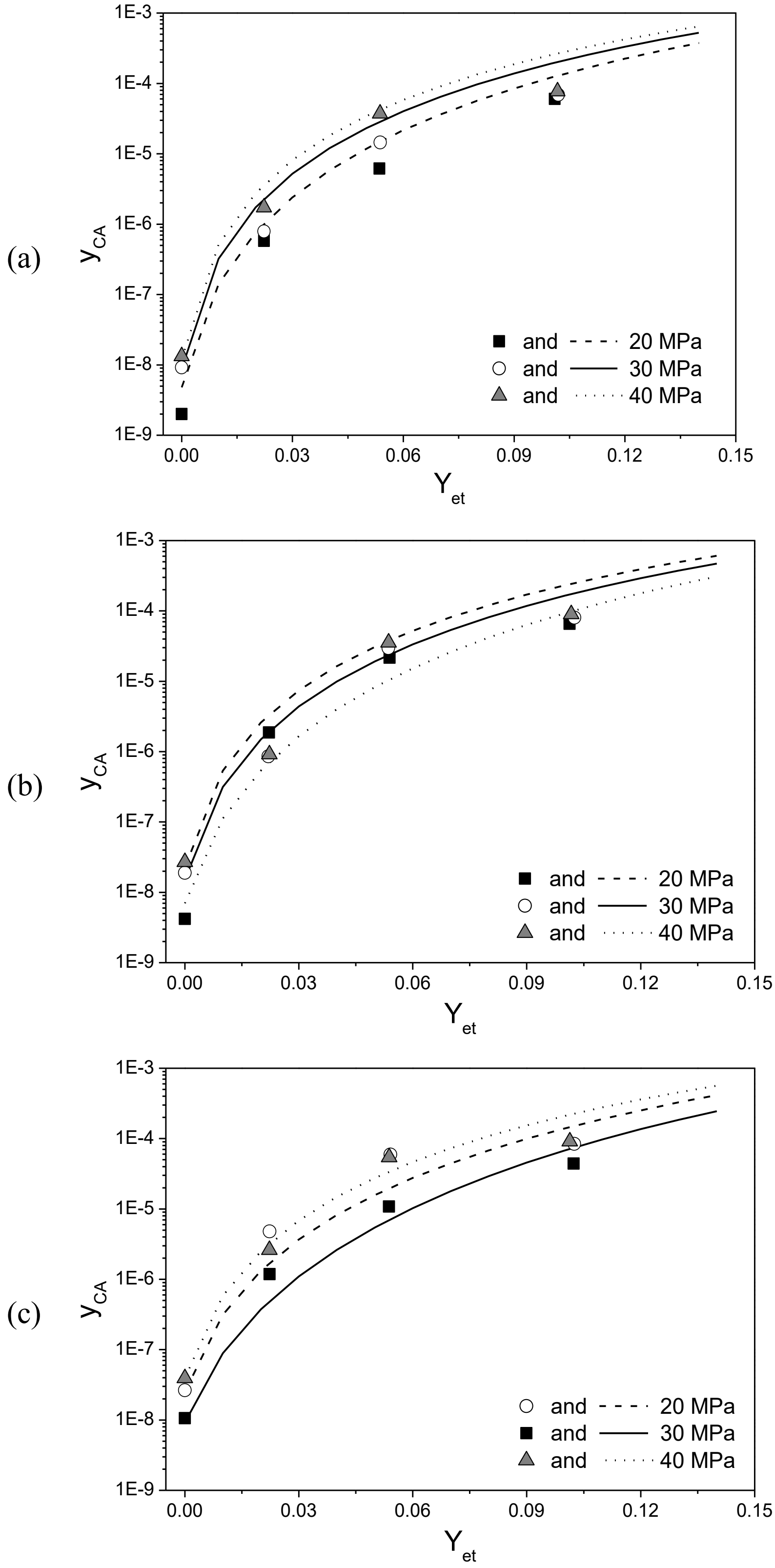

Fig. 3. Prediction of CA solubility in CO_2/ethanol mixtures (y_{CA}) from CPA-1: (a) 313 K; (b) 323 K; (c) 333 K. Experimental data of CA solubility in pure $scCO_2$ were obtained from Murga et al. [7]. Y_{et}: ethanol mole fraction on a solute free-basis; Symbols: experimental data; Lines: predicted data.

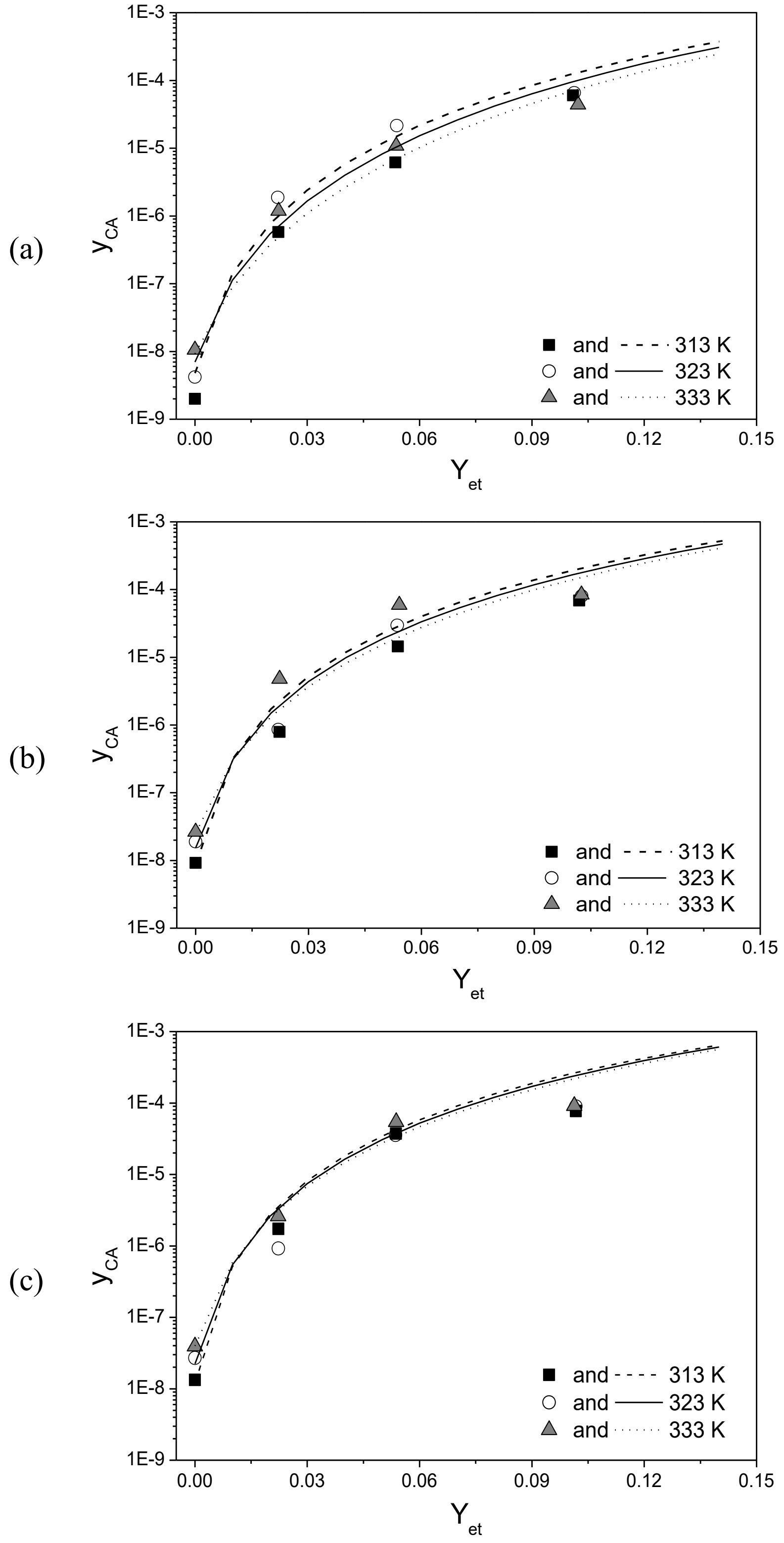

Fig. 4. Prediction of CA solubility in CO_2/ethanol mixtures (y_{CA}) from CPA-1: (a) 20 MPa; (b) 30 MPa; (c) 40 MPa. Experimental data of CA solubility in pure $scCO_2$ were obtained from Murga et al. [7]. Y_{et}: ethanol mole fraction on a solute free-basis; Symbols: experimental data; Lines: predicted data.

The ratio of the solubility in the ternary system (CO_2/ethanol mixture) and the solubility in a binary system ($scCO_2$) under the same conditions of temperature and pressure provides the cosolvent effect (**Equation 14**).

$$co-solvent \quad effect = \frac{y_{CA}^{CO_2+Ethanol}}{y_{CA}^{CO_2}} \tag{14}$$

As a consequence of the results aforementioned, i.e. solubility values less dependent on temperature and pressure at larger concentrations of ethanol, the cosolvent effect is lower for high pressure and temperature values when approximately 10 mol% of ethanol is used as cosolvent (**Fig. 5**). In addition, a significant increase of up to 30,000 times in the solubility of caffeic acid in $scCO_2$ was observed by adding 10.2 mol% ethanol at 313 K and 20 MPa (cosolvent effect = 30,000).

For systems containing 2.2 and 5.4 mol% ethanol (**Fig. A.1** of the Appendix A. Supplementary data), the highest cosolvent effect were 290 and 3075, respectively. Although the highest cosolvent effect was also observed at 323 K and 20 MPa by using 2.2 mol%, there was no clear trend for the influence of temperature and pressure on the solubility data.

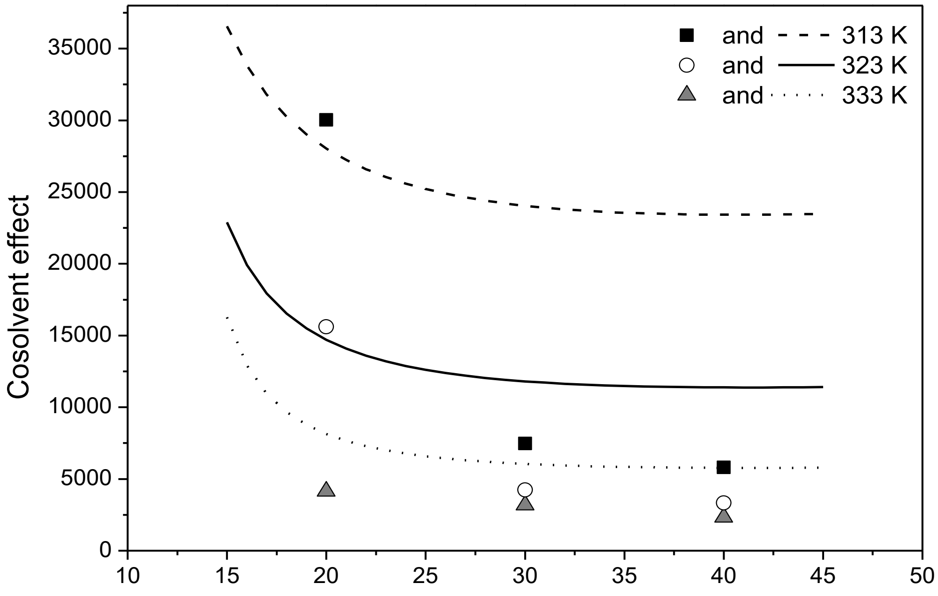

Fig. 5. Cosolvent effect on CA solubility in CO_2 with 10.2 mol% ethanol. The lines represent cosolvent effect calculated according to CPA-1.

4.3. Thermodynamic modeling

The melting temperature (T_m) and enthalpy of fusion (ΔH_f) required by **Equation 1**, are presented in **Table 1**. While for SRK equation, a_0, c_1 and b are calculated from the critical properties, CPA parameters for pure components (a_0, c_1, b , ε and β) are obtained by fitting vapor pressure and/or liquid density data of the pure compounds. For the solvents CO_2 and ethanol, these parameters were obtained from literature, as indicated in **Table 4**. CO_2 was here considered as a non-associating compound. For caffeic acid, the parameters were fitted to vapor pressure data [42] using as initial estimation the values reported by Mota el al. [24]. It was not possible to use the parameters from Mota et al. [24] because in our work the parameters ε and β were not fitted for each type of associative group of the molecule.

The values for k_{ij}, the binary interaction parameters (BIP) **(Equation 5),** were estimated by fitting the experimental data with the SRK and CPA EoS. Their values are reported in **Table 5**. Experimental data from Murga et al. [7] and Lim et al. [43] were used to estimate the BIP between CO_2 and caffeic acid and CO_2 and ethanol, respectively.

A good description of the binary systems was achieved with low average absolute relative deviations (AARD). The largest AARD was obtained for caffeic acid - CO_2 system (28.7%), however **Fig. A.2** of the Appendix A shows that the deviation between these experimental and calculated points are high at low pressures, near to the critical point.

As previously mentioned, different approaches for SRK (SRK-1 and SRK-2) and CPA (CPA-1 to CPA-4) models were evaluated, in which different values of ΔC_p and ΔV were considered, as shown in **Table 6.** The ΔV value adopted was estimated assuming that the liquid molar volume is about 20% higher than the solid molar volume reported by Moncada et al. [44], as suggested by Goodman et al. [45]. The ΔC_p adopted value was of 50 J/mol K. However, no significant influence of this term on the results was observed.

No fitting was performed for the ternary mixtures. These were predicted using the BIP's from the binary mixtures. Although the SRK model was as good at fitting the binary systems as the CPA-EoS, the prediction of ternary system was not successful. The comparison between the results for the prediction of the ternary system (caffeic acid – CO_2 – ethanol) using CPA-EoS and SRK-EoS are shown in **Fig. 6**. It was observed that SRK-EoS greatly underestimates the solubility values independently of ΔC_p and ΔV values adopted.

Table 4

CPA parameters of pure components.

Compound	a_0 (J m^3/mol)	$b \times 10^5$ (m^3/mol)	c_1	Associative Group (AG)	Number of AG	ε (J/mol)	$\beta \times 10^{-3}$	*AARD P* (%)
CO_2	0.35[a]	2.72[a]	0.76[a]					
Ethanol	0.86716[b]	4.91[b]	0.7369[b]	OH	1	21532[b]	8.0[b]	
Caffeic acid	3.8896[c]	15.6[c]	1.9534[c]	OH and COOH	3	19090.37[d]	27.88[d]	7.8

[a] Oliveira et al [40]; [b] Folas et al. [41]; [c] Mota et al. [25]; [d] Fitted from vapor pressure data of caffeic acid reported by Chen et al. [42].

Table 5

Binary interaction parameters from SRK-EoS and CPA-EoS for CA (1) – CO_2 (2) – ethanol (3) system.

Pair ij	$k_{ij} = A + B \times (T/K)$				*AARD* (%)		Data reference
	SRK-1		CPA-1		SRK	CPA	
	A	B × 10^4	A	B × 10^4			
12	0.167	– 0.170	0.084	-2.12	20.0[a]	28.7[a]	Murga et al. [7]
13	0.0149	-1.45	-0.0424	3.70	6.9[a]	5.0[a]	This work
23	0.2199	4.13	0.0407	3.01	2.9[b]	14.8[b]	Lim et al. [43]

[a] *AARD y* ; [b] *AARD P* .

Table 6

Different model approaches evaluated, based in **Equation 1**.

Model	$\Delta C_{p,CA}$ (J/mol K)	ΔV_{CA} (m^3/mol)	*ALD* [a]
SRK-1 and CPA-1	0	0	0.288[b]
SRK-2 and CPA-2	50	2.549×10^{-5}	0.321[b]
CPA-3	50	0	0.314[b]
CPA-4	0	2.549×10^{-5}	0.297[b]

[a] $ALD = \sum_{j=1}^{N} \frac{1}{N} \left| \log \frac{y_i^{calc}}{y_i^{\exp t}} \right|$; [b] *ALD* for prediction of phase equilibrium data in ternary system CO_2 – ethanol – caffeic acid using CPA model.

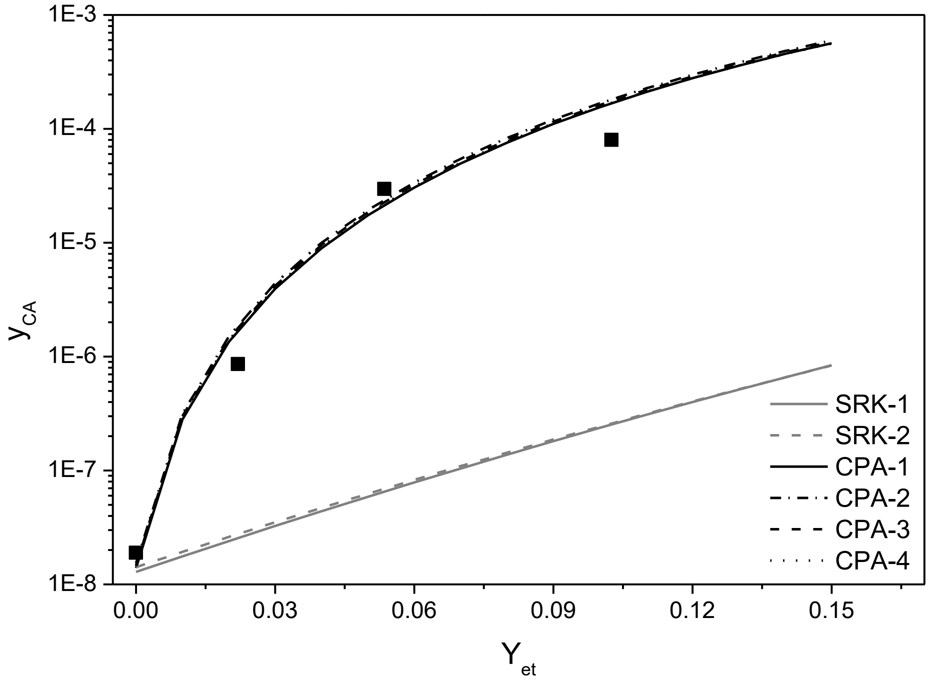

Fig. 6. Comparison between SRK and CPA model for prediction of ternary system caffeic acid - CO_2 - ethanol at 323.15 K and 30 MPa.

As expected, the predicted CA solubility using CPA-EoS was much closer to solubility experimental data than that of the SRK-EoS prediction. Unlike SRK, CPA takes into account the associative interactions such as hydrogen bonding, which take place between the caffeic acid and ethanol.

Fig. 6 also shows that similar results were obtained using the various approaches with CPA-EoS. Comparing the average logarithmic deviation (*ALD*) between the experimental and calculated data (**Table 6**), the best results were obtained with CPA-1 (*ALD* = 0.28). In addition to providing the lowest ALD value, CPA-1 is more suitable mainly because it allows to use the simplified form of **Equation 1**.

Comparing the predictions from CPA with those from SAFT and PR + COSMOSAC, which were previously reported [21,22], all the three presented reasonable results. Ting et al. [21] evaluated PR + COSMOSAC for the prediction of solubility of different solutes in CO_2 + cosolvents and 0.48 was the minimum *ALD* value obtained. However CPA, besides a slightly better description of the experimental data, also provides most of the advantages of cubic equations of state, such as simplicity of application, low computational overhead and good correlation for phase equilibria data at high pressures. The results by PR + COSMOSAC are also of interest because the prediction of the ternary system requires only the melting temperature and enthalpy of fusion of the solute.

In general, the CPA-EoS predictions provide a good description of the dependence of caffeic acid solubility with the concentration of ethanol, for the temperature and pressure ranges here evaluated, as shown in **Fig. 3** and **Fig. 4**. Both experimental and calculated data for the ternary system showed that the effect of pressure becomes less significant with the increase of ethanol content in the CO_2/ethanol mixture, as shown in **Fig. 3** where both experimental and predicted data become closer to each other at higher concentrations of ethanol.

However, mainly at higher pressures, according to the CPA predictions, the effect of temperature on CA solubility is not relevant. This behavior is shown in **Fig. 4b** and **Fig. 4c**, by the overlapping lines.

As observed for the experimental data, CPA predicted the higher cosolvent effect at lower temperatures and pressure when CO_2/ethanol (90:10 mol/mol) was used as solvent (**Fig. 5)**. However, for other proportions of CO_2 and ethanol, the prediction of the cosolvent effect indicated the opposite behavior (**Fig. A.1** of the Appendix A).

Finally, although accurate trends for prediction of the CA solubility in CO_2/ethanol were obtained, there was some divergences between the behavior of calculated and experimental data when the influence of temperature and pressure on cosolvent effect were analyzed.

5. Conclusions

Experimental data on the solubility of caffeic acid in ethanol and mixtures of supercritical carbon dioxide and ethanol were reported in this work for the first time in the open literature. As expected, the importance of the use of cosolvent to obtain natural extracts by supercritical technology was confirmed, with an increase of up to 30,000 times observed in the solubility of caffeic acid when 10 mol% of ethanol was added to $scCO_2$.

The use of concentrations higher than 10 mol% of ethanol should be analyzed, in terms of cost benefit, due to the small increase in the solubility of CA on these conditions. When compared with SRK-EoS, the CPA model showed an excellent prediction of the solubility of caffeic acid in supercritical carbon dioxide in the presence of ethanol.

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Abbreviations

Nomenclatures

AARD - Average Absolute Relative Deviation

$$AARD\ P(\%) = \frac{100}{N}\sum_{j=1}^{N}\left[\frac{\left|P_{cal} - P_{\exp}\right|}{P_{\exp}}\right]_j$$

$$AARD\, y\,(\%) = \frac{100}{N}\sum_{j=1}^{N}\left[\frac{\left|y_{cal} - y_{\exp}\right|}{y_{\exp}}\right]_j$$

ALD - Average Logarithmic Deviation - $ALD = \sum_{j=1}^{N}\frac{1}{N}\left|\log\frac{y_i^{calc}}{y_i^{\exp t}}\right|$

AG - association group

CA - caffeic acid

CPA - Cubic Plus Association

COSMO-SAC - COnductor-like Screening Model - Segment Activity Coefficient

EoS - Equation of State

OF - objective function

PR - Peng-Robinson

SAFT - Statistical Associating Fluid Theory

SRK - Soave-Redlich-Kwong

$scCO_2$ - supercritical carbon dioxide

List of symbols

a - energy parameter in the physical term

a_0, c_1 - parameters for calculating a

b - co-volume parameter in the physical term

C_p - heat capacity

g - radial distribution function

H - enthalpy

k_{ij} - binary interaction parameter

N - number of experimental data

P - pressure

R - gas constant

T - temperature

V - molar volume

X_{Ai} - fraction of molecule not bonded at site A

y , x - mole fractions

Greek symbols

β - association volume

ε - association energy

Δ - association strength

ρ - molar density

η - reduced fluid density

φ - coefficient of fugacity

ω - acentric factor

Subscripts

A , B - molecules sites

atm - atmospheric condition

C - critical properties

cal - calculated

$\exp$ - experimental

i , j - pure component indexes

m - melting

f - fusion

Superscript

vap - vapor

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Appendix A. Supplementary data

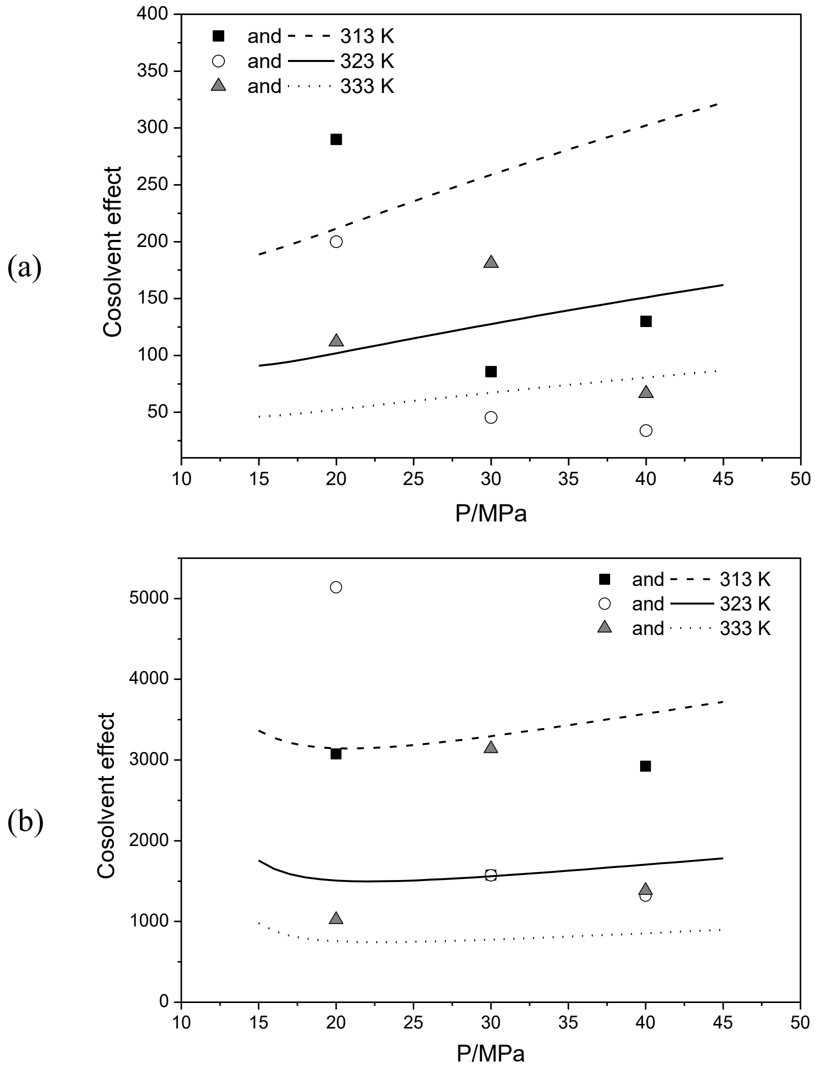

Fig. A.1. Cosolvent effect on CA solubility in CO_2 with (a) 2.2 mol% ethanol and (b) 5.4 mol% ethanol. The lines represent cosolvent effect calculated from data predicted with CPA-1.

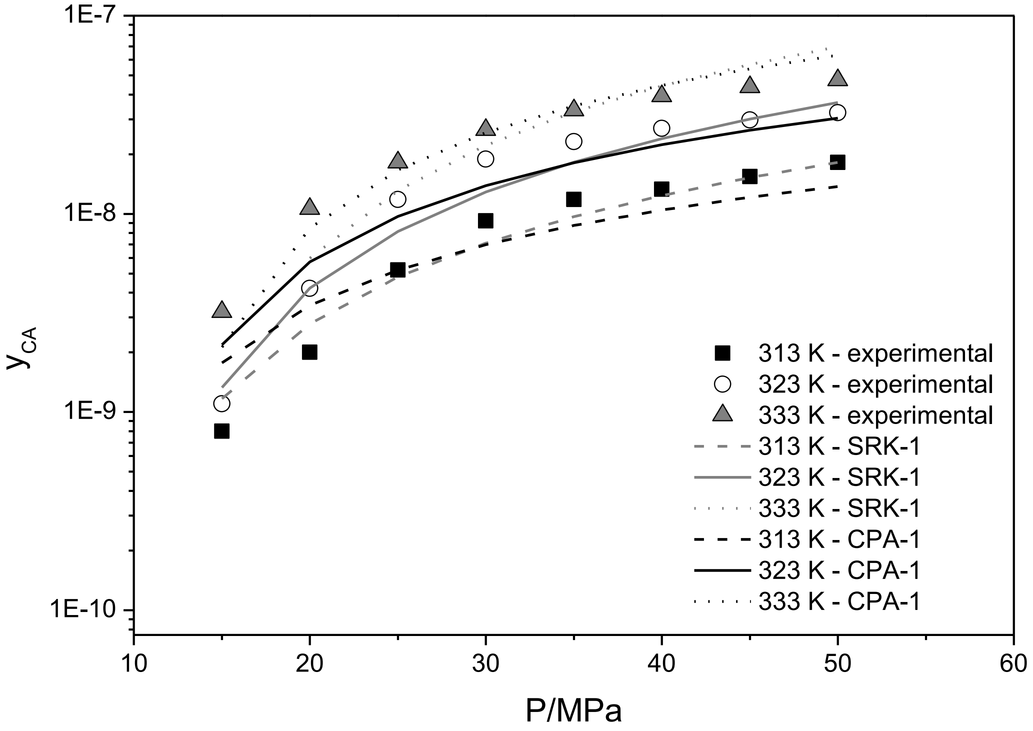

Fig. A.2. Thermodynamic modeling from SRK-1 and CPA-1 models of the CA solubility in $scCO_2$ (y_{CA}).

CAPÍTULO 5. PREDICTION OF SOLID SOLUTE SOLUBILITY IN SUPERCRITICAL CO_2 WITH COSOLVENTS USING CPA EQUATION OF STATE

Manuscript to be submitted to Industrial & Engineering Chemistry Research

Prediction of solid solute solubility in supercritical CO_2 with cosolvents using CPA equation of state

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Abstract

Supercritical carbon dioxide ($scCO_2$) is widely used in separation processes applied to food, chemistry, pharmaceutical and material industries. Therefore, knowledge on solubility of solid solutes in $scCO_2$, with or without cosolvents, is essential for the design and optimization of extraction, fractionation and purification processes. This work evaluated the accuracy of Cubic Plus Association equation of state (CPA-EoS) to predict the solubility of solutes in ternary and/or quaternary systems from the binary interaction parameters. The solubility of 12 solid solutes in $scCO_2$ in the presence of different organic cosolvents was investigated, totalizing 19 systems at pressures between 8 and 40 MPa, temperatures ranging from 308 K to 353 K and concentrations of cosolvent varying from 0.73 to 10 mol%. The overall average logarithmic deviation (ALD) between experimental and calculated data was 0.43. The accuracy of CPA-EoS is higher than that obtained from PR+COSMOSAC, which was previously reported in the literature. Furthermore, the effects of the temperature, pressure, type and concentration of the cosolvent on the solubility were well estimated through CPA model. However, it is important to assume a suitable number of association sites for solute and cosolvent molecules, and have a reliable data source for vapor pressure and liquid density of the pure components, as well as for phase equilibria of binary systems.

Keywords: thermodynamic modeling; Cubic Plus Association; phase equilibrium; supercritical carbon dioxide; biocompounds.

1. Introduction

In recent years, fast progress has been occurred toward using the supercritical technology in industrial separation processes. Apart from “green” extractions to replace the organic solvents, applications involving supercritical fluids include supercritical fluid fractionation[1–3], supercritical fluid chromatography[4–6], uses as antisolvent for micro- and nanoparticle formation[7,8] and applications where supercritical fluids are utilized as a reaction medium [9–11]. Thus, chemical, petroleum, pharmaceutical, food, biotechnology, biomedical, nuclear, material and polymer processing industries are examples of application areas for separation processes using this technology[12].

Carbon dioxide is the most common gas applied for supercritical fluid extraction due to its low cost, high purity, safety and low critical properties. Since pure supercritical carbon dioxide ($scCO_2$) is a lipophilic solvent, mixtures with organic solvents, especially alcohols, are used to increase the polarity of the mixture fluid.

Solubility data of solutes in $scCO_2$ in the presence of cosolvents are essential information for design and optimization of separation processes using $scCO_2$. Besides the experimental measurements, thermodynamic models are useful and low cost alternatives to obtain phase equilibrium information. For that purpose, the use of reliable thermodynamic models which are able to accurately predict the phase equilibria as a function of temperature, pressure, and composition of a specific system, is crucial.

Several semi-empirical equations have also been proposed to describe the solubility of solutes in $scCO_2$ with[13–15] or without[16–18] entrainer. Due to its easy handling and good results, the cubic equations of state such as Peng-Robinson (PR-EoS) and Soave-Redlich-Kwong (SRK-EoS) are the models most used to correlate the solubility data under high pressures. Despite these thermodynamic models being able to describe the binary system data with satisfactory results, in most cases the fitted binary interaction parameters are not suitable to estimate the phase equilibrium in multicomponent systems (with cosolvents), since the strong interactions that can occur between the molecules of solute and the cosolvent are not taken into account in the cubic equations of state[19–21].

The solubility of solutes in $scCO_2$ and cosolvents mixtures were predicted using the Statistical Associating Fluid Theory (SAFT)[22] and PR+COSMOSAC[23] models. In addition, Bitencourt et al.[24] have recently reported good prediction for caffeic acid solubility in CO_2/ethanol mixtures using the Cubic Plus Association Equation of State (CPA-EoS) from binary interaction parameters. CPA-EoS consists of the combination of physical term derived from SRK-EoS and an association term as proposed by Wertheim perturbation theory[25] and widely used in SAFT-type equations. The need to accurately describe systems containing strongly polar compounds, which form hydrogen bonds and often present an unusual thermodynamic behavior because the strong interactions, either between molecules of the same species (self-association) or between molecules of different species (cross-association), was one of the main motivations behind the development of this EoS.

Compared to SAFT, CPA-EoS is a simpler model for associating mixtures, which keep the well-known advantages of cubic equations of state, while still presenting a high performance for associating compounds. Although CPA has been widely investigated for several application in the academic community, its use in industry is far from being widespread. Nevertheless, diverse applications in the oil and gas industry are benefitting from the use of this EoS, as well as, diverse companies having in-house versions of the model. Therefore, this work will evaluate the accuracy of the CPA equation of state to predict solubility for various solid solutes in supercritical carbon dioxide in the presence of different cosolvents.

2. Thermodynamic modeling

2.1.Solid-fluid equilibrium

Solid solute solubility was described through the equality of fugacity for the solute *i* in its pure solid phase and in the fluid phase. Considering a pure solid phase with no solid-solid phase transitions and neglecting the Poynting term and the effect of pressure in melting temperature (T_m) and enthalpy ($\Delta_{fus}H$), the fugacity of the solute *i* in its pure solid phase at a given temperature *T* and pressure *P* is obtained from the fugacity of its hypothetical liquid under the same conditions, only needing T_m and $\Delta_{fus}H$ of the solute [26]. Thus, when an equation of state is selected, the following expression for the mole fraction solubility of the solute (y_i) is used:

$$y_i = \frac{\varphi_i^{pure\,liquid}}{\varphi_i^{liquid}} \exp\left[-\frac{\Delta_{fus}H}{R}\left(\frac{1}{T}-\frac{1}{T_m}\right)\right] \tag{1}$$

where *T* is the absolute temperature, *R* is the gas constant and φ is the fugacity coefficient, which was calculated by the Cubic Plus Association equation of state (CPA-EoS) in this work.

2.2.Cubic Plus Association

CPA-EoS consists of the combination of a physical (cubic) term with an association contribution term. Although there are variants, the term which consider physical interactions, such as attraction and repulsion forces, is the one from the SRK-EoS. The association contribution, originally proposed by Wertheim, accounts for intermolecular hydrogen bonding and solvation effects. Thus, the CPA-EoS is expressed in terms of the compressibility factor as:

$$Z^{CPA} = Z^{SRK} + Z^{assoc} = \frac{1}{1-b\rho} - \frac{a\rho}{RT(1+b\rho)} - \frac{1}{2}(1+\rho\frac{\partial(\ln g)}{\partial\rho})\sum_i x_i \sum_{A_i}(1-X_{A_i}) \tag{2}$$

where *i* is the component index, A indexes the bonding sites on molecule *i*, a is the pure component energy parameter, b is the co-volume pure component parameter, ρ is the molar density, x_i is the molar fraction of component i, g is the function of radial distribution and

X_{Ai} is the molar fraction of component i not bound to site A, i.e represents the mole fraction of monomers.

The pure component energy parameter (a) has a Soave-type temperature dependency:

$$a(T) = a_0[1 + c_1(1 - \sqrt{T/T_c})]^2 \tag{3}$$

X_{Ai} is calculated from the association strength between two sites belonging to two different molecules ($\Delta^{A_iB_j}$) through the resolution the following set of equations:

$$X_{Ai} = \frac{1}{1 + \rho \sum x_j \sum X_{Bj} \Delta^{AiBj}} \tag{4}$$

where:

$$\Delta^{A_iB_j} = g(\rho)\left[\exp\left(\frac{\varepsilon^{A_iB_j}}{RT}\right) - 1\right] b_{ij} \beta^{A_iB_j} \tag{5}$$

The simplified hard-sphere radial distribution function (g) was used [27]:

$$g(\rho) = \frac{1}{1 - 1{,}9\eta} \tag{26}$$

$$\eta = \frac{1}{4} b\rho \tag{27}$$

Therefore, apart from the three pure component parameters (a_0, c_1, and b) for non-associating components, two more parameters related to the associative contribution of the compound are required: the energy of association between the sites of molecule (ε) and the parameter of association volume (β).

When dealing with mixtures, the energy and co-volume parameters of the physical term were calculated employing the conventional van der Waals one-fluid mixing rules (**Equations 8, 9 and 10**), fitting one binary interaction parameter (k_{ij}) for the energy parameter. For the parameters from the associative contribution several combining rules have been suggested [28]. The Elliot combining rule and CR-2 [29] are the most used, because both provide accurate descriptions of the phase equilibria for many mixtures. In this work the CR-2 (**Equations 11 and 12**) was used.

$$a = \sum_i \sum_j a_{ij} y_i y_j \quad , \tag{8}$$

$$a_{ij} = \sqrt{a_i a_j}\,(1 - k_{ij}) \tag{9}$$

$$b = \sum_i b_i y_i \tag{10}$$

$$\beta^{A_i B_j} = \sqrt{\beta^{A_i}\, \beta^{B_j}} \tag{11}$$

$$\varepsilon^{A_i B_j} = \frac{1}{2}\left(\varepsilon^{A_i} + \varepsilon^{B_j}\right) \tag{12}$$

3. Selection of data

The solutes investigated in this work were selected based on the dataset availability needed to evaluate the CPA-EoS for prediction of its solubility in CO_2 + cosolvent, as shown in **Figure 1**. Critical temperature, melting temperature and enthalpy of fusion of the solutes are required, as shown in **Equations 1 and 3**. Experimental data for vapor pressure and liquid density at different temperatures are used to fit CPA pure component parameters (a_0, c_1, b, ε and β) and phase equilibrium data of the binary systems of the components present in the ternary mixtures are relevant to obtain the binary interaction parameters (k_{ij}). Finally, ternary system data were only used to compare experimental and predicted data.

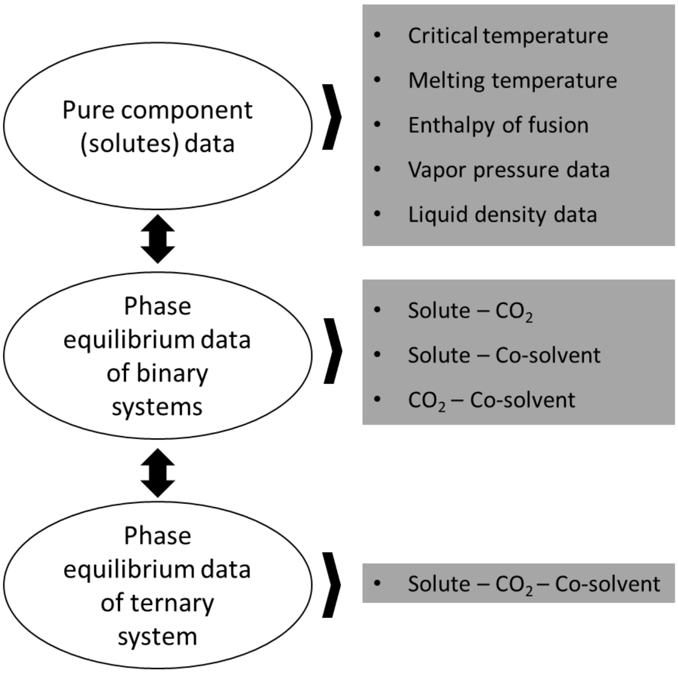

Figure 1. Dataset required to evaluate the prediction of ternary systems equilibrium data using CPA model.

4. Results

Several works have reported solubility data for solutes in mixtures of CO_2 plus cosolvent. However, mainly due to the lack of experimental data for vapor pressure and for the binary equilibrium cosolvent – solute, 12 solutes were selected, as shown in **Figure 2**. Different classes of compounds were investigated, including aromatic, aliphatic molecules, alcohol and acid substances. Thus, the solubility of 12 solid solutes in $scCO_2$ in the presence of different organic cosolvents was investigated, at pressures between 8 and 40 MPa, temperatures ranging from 308 K to 353 K and concentration of cosolvent varying from 0.73 and 10 mol% (**Table 1)**. Polar and non-polar cosolvents were also analyzed, totalizing 19 systems. Among them, only one is a quaternary system, containing trans-ferulic acid, CO_2, ethanol and water.

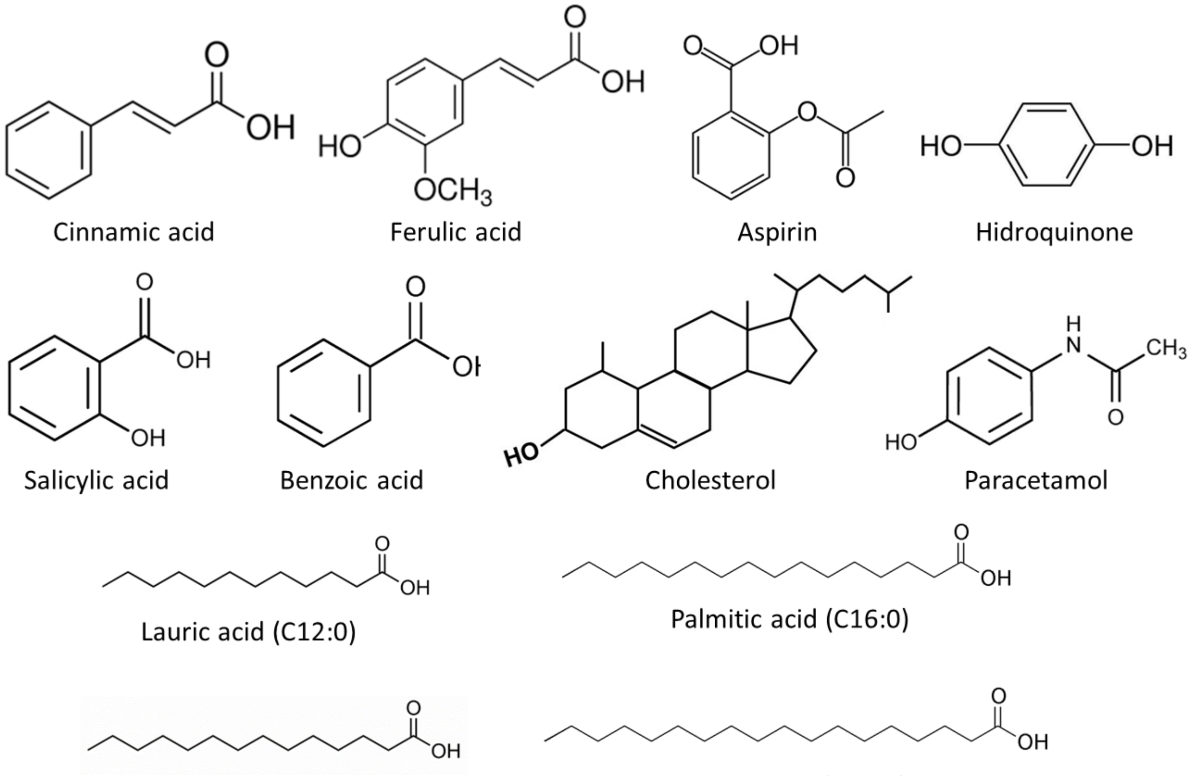

Figure 2. Chemical structure of the solutes investigated in this work.

In a first step for the thermodynamic modeling with CPA, pure component parameters were obtained by fitting vapor pressure and liquid density data. However, liquid density data were not easily accessible on the scientific literature and databases for the trans-ferulic acid and paracetamol. On other hand, the values available for cholesterol, salicylic acid, benzoic acid were not used because the model was unable to fit these data. Therefore, pure component parameters for these solutes were calculated only from vapor pressure data.

For this purpose, it is also important to analyze the number of associating sites in the molecule. The properties and approaches used for each pure component are presented in **Table 2**. For each compound, ε and β parameters were assumed as a unique value, i.e they were not fitted for each type of associative group of the molecule. In addition, the solvents CO_2, hexane and acetone were assumed as a non-associating compound.

Then, the binary interaction parameters were calculated for all the binary combinations involved in the ternary or quaternary systems. References for experimental data, the values of k_{ij} fitted and the average absolute relative deviation (AARD) are shown in **Table 3**. Mainly for cosolvent – solute systems good description of the experimental data was verified.

Finally, prediction of the solubility for each solute in the ternary or quaternary system using CPA-EoS was performed, and the average logarithmic deviation (ALD) was used to evaluate the accuracy of the prediction, as shown below:

$$ALD = \sum_{j=1}^{N} \frac{1}{N} \left| \log \frac{y_i^{calc}}{y_i^{\exp t}} \right| \tag{13}$$

where N is the number of experimental points and y_i is the mole fraction solubility of the solute in mixtures of $scCO_2$ and cosolvent.

In general, accurate results for prediction of the solubility of solutes in $scCO_2$ + cosolvents using CPA model were verified. ALD value for each solute (**Table 1)** varied from 0.04 to 1.5. Few solutes such as lauric acid, myristic acid, hydroquinone and salicylic acid presented the highest ALD values. According to Ting et al. [23], higher deviations for the carboxylic acids with short chain (lauric and myristic acids) can be associated to their lower melting temperature (316.6 K and 326.5 K, respectively), which are very close to the temperatures of solubility measurements. In this way, excluding the myristic acid solubility data at 318 K the ALD value decreased from 1.32 to 0.75.

The ALD value obtained by evaluating all the system simultaneously was 0.43, and excluding lauric and myristic acids the ALD reached 0.30. Both values are quite lower than those reported by Ting et al. [23] (ALD = 0.64), when PR+COSMOSAC equation was used for prediction of solubility for 23 solutes in $scCO_2$ and cosolvents. Excluding trans-ferulic acid, all the solutes here evaluated were also investigated by Ting et al. Applying the SAFT model to predict solubility of naphthalene, phenanthrene, anthracene, pyrene and benzoic acid in $scCO_2$ with cosolvents, Yang et al. [22] reported AARD values (between 4.7% and 48.5%) lower than those obtained with cubic EoS (ranging from 15.0% to 70.8%). This difference was not so relevant probably because these authors have mainly evaluated systems containing non-associating compounds. For instance, to predict the solubility of caffeic acid, which has three association sites, the values estimated using a cubic EoS were more than 100 times lower than experimental data [24]. In addition, the CPA EoS presented good accuracy when predicting the solubility of caffeic acid in $scCO_2$ + ethanol, with ALD = 0.29.

The dispersion between the experimental and calculated data can be visualized in **Figure 3**, in which the points overlap the black solid line in an ideal situation. Underestimated prediction of the solubility data was noted in most of the cases. Hydroquinone solubility was the only one significantly overestimated. **Figure 3** categorizes the compounds in aromatic acids (●), aromatic alcohols (■), aliphatic acids (▲) and aliphatic alcohols (×). No trend for better results was verified when analyzing different classes of compounds, since the highest differences between experimental and calculated data were observed for an aromatic acid (salicylic acid), aliphatic acid (lauric acid) and aromatic alcohol (hydroquinone).

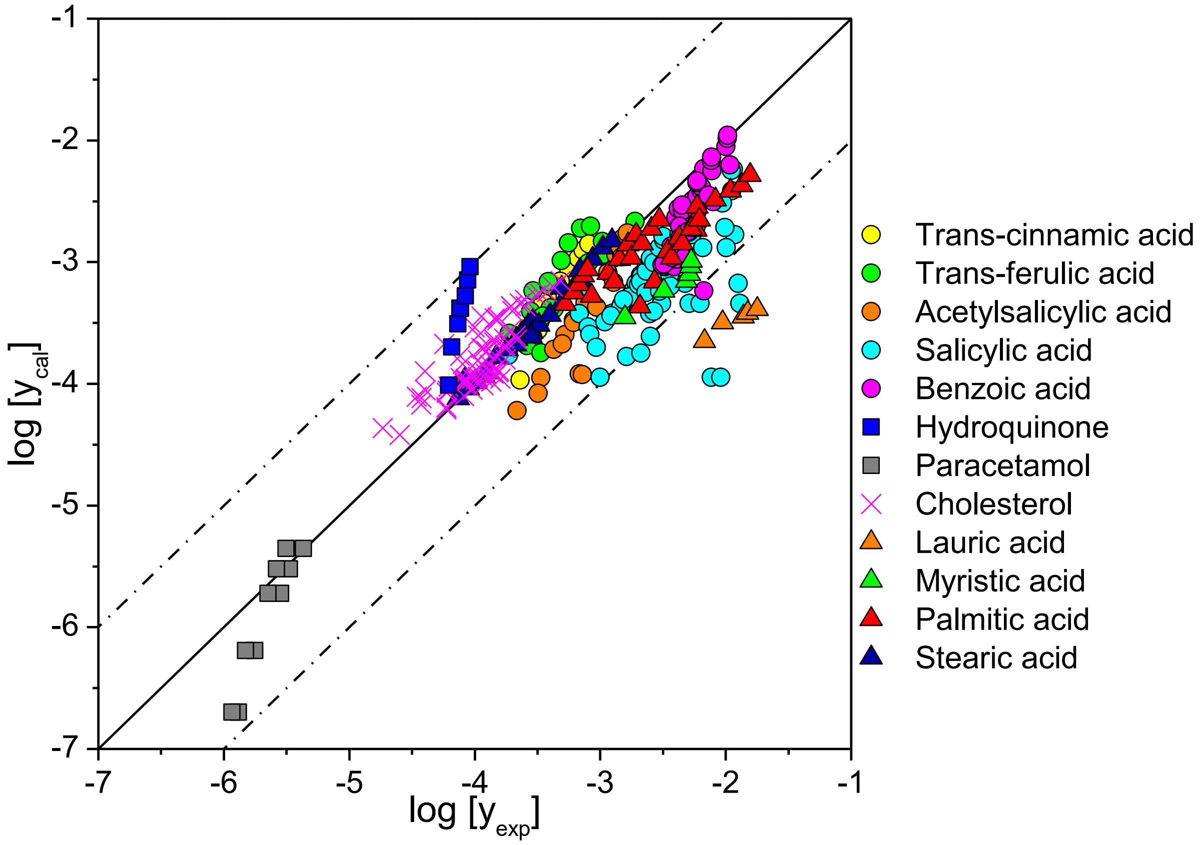

Figure 3. Comparison between experimental and predicted values of the solubility of 12 solutes in $scCO_2$ with cosolvent.

Table 1. Ternary and quaternary systems investigated in this work and the average logarithmic deviation between experimental and predicted data using CPA model.

Solute[a]	Cosolvent	Cosolvent (mol%)	Pressure (MPa)	Temperature (K)	ALD
Trans-cinnamic acid[30]	Ethanol	2-4	10-40	313-353	0.126
Trans-ferulic acid[19]	Ethanol	5-10	20-40	313-333	0.189
	Ethanol/water (90:10 w/w)	2-10	30	323	0.054
Aspirin (acetylsalicylic acid)[31,32]	Ethanol	3	10-20	318-328	0.282
	Acetone	3	10-20	318-328	0.380
Salicylic acid[33,34]	Ethanol	1.47-6.91	8.7-15.7	308-318	0.706
	Methanol	3.5	10.1-20.1	318-328	0.666
	Acetone	3.5	8.6-20.1	318-328	0.461
Benzoic acid[35]	Methanol	3-3.5	10.1-30.4	308-318	0.227
	Hexane	2.2-7	15.2-35.4	308-318	0.105
Hydroquinone[36]	Methanol	2.8	10.3-32.7	308	0.684
Paracetamol[37]	Ethanol	0.85	10-25	313	0.289[b]
Cholesterol[38,39]	Methanol	3	10-21	318-328	0.075[c]
	Acetone	3-6	10-22	318-338	0.205[c]
	Hexane	3.5-6	9-22	318-338	0.227[c]
Lauric acid[40]	Ethanol	0.5	9.9-22.7	308	1.55
Myristic acid[40]	Ethanol	0.73-4.16	9.9-22.7	308-318	1.32 (0.754)
Palmitic acid[41,42]	Ethanol	0.73-4.16	8.21-22.8	308-318	0.297
Stearic acid[41]	Ethanol	0.73-1.98	12.8-22.6	308-318	0.043

[a]Reference for experimental data of ternary or quaternary systems; [b]Assuming paracetamol with one associating site; [c]Assuming cholesterol as a non-associating molecule.

Table 2. Properties and parameters for pure component.

Compound	T_C (K)	T_m (K)	$\Delta_{fus}H$ (kJ/mol)	Associating Group (AG)	Number of AG	CPA parameters					AARD (%)	
						a_0 (J m^3/mol)	$b \times 10^5$ (m^3/mol)	c_1	ε (kJ/mol)	$\beta \times 10^{-3}$	P	ρ
Solutes[a]												
Trans-cinnamic acid[30,43]	797.3[30]	406.1[44]	22.2[44]	COOH	1	4.45	12.9	1.32	19.1	1.62	3.6	1.1
Trans-ferulic acid[45]	947.97[46]	443[47]	12.3[47]	OH and COOH	2	5.12	17.8	1.82	21.4	3.03	1.3	
Acetylsalicylic acid[43,48]	762.9[32]	408[49]	33.8[48]	COOH	1	4.99	12.6	0.93	19.4	27.3	2.0	13.0
Salicylic acid[50]	739[43]	431.3[44]	24.5[44]	COOH and OH	2	4.25	17.7	1.51	10.0	60.3	0.9	
Benzoic acid[51,52]	755[53]	395.4[23]	17.1[23]	COOH	1	4.07	20.9	1.73	19.1	60.5	2.0	
Hydroquinone[43,54]	766.9[55]	445.1[23]	27.1[23]	OH	2	2.19	10.6	1.17	19.2	40.6	3.3	0.5
Paracetamol[56]	828[55]	441[57]	27.7[57]	OH or NH	1	5.00	13.9	1.84	18.0	16.2	3.3	
				OH and NH	2	2.77	16.7	3.08	18.0	65.6	3.2	
Cholesterol[38]	1168.2[38]	420.2[23]	27.4[23]	non-associating	0	3.83	19.5	1.96			3.5	
				OH	1	3.84	18.1	1.46	21.3	10.1	5.2	
Lauric acid[58,59]	743[60]	316.6[23]	36.1[23]	COOH	1	6.79	22.4	1.82	17.9	0.226	1.8	1.2
Myristic acid[58,59]	763[60]	326.5[23]	45.0[23]	COOH	1	8.54	25.7	1.58	25.0	0.568	0.4	1.9
Palmitic acid[58,59]	785[60]	334.7[23]	53.0[23]	COOH	1	9.08	29.4	2.02	15.2	4.71	0.4	1.4
Stearic acid[58,59]	803[60]	343[49]	61.2[49]	COOH	1	11.88	34.8	1.76	17.9	9.4	4.1	4.7
Solvents[b]												
CO_2[61]				non-associating	0	0.35	2.72	0.76				
Ethanol[62]				OH	1	0.86716	4.91	0.7369	21.532	8.0		
Methanol[62]				OH	1	0.40531	3.0978	0.43102	24.591	16.1		
Acetone[49]				non-associating	0	1.39	6.14	0.788				
Hexane[63]				non-associating	0	2.3681	10.79	0.8313				
Water[27]				OH	2	0.1228	1.452	0.6736	16.655	69.2		

[a] Reference for vapor pressure and/or liquid density experimental data; [b] Reference for the CPA pure component parameters.

Table 3. Binary interaction parameters as a function of the temperature: $k_{ij} = A + B \times T / K$.

Solute	Cosolvent	CO_2 – solute				Cosolvent – solute			
		Reference[a]	k_{ij}		AARD (%)	Reference[a]	k_{ij}		AARD (%)
			A	B			A	B	
Trans-cinnamic acid	Ethanol	30,64	-0.3845	0.00147	20.4	65	-0.145	0.000429	2.5
Trans-ferulic acid	Ethanol	66	-0.007645	0.000215	20.3	19,67	-0.0675	0.000436	4.8
	Water					19,44,67	0.000792	-0.00000845	29.3
Acetylsalicylic acid	Ethanol	32	0.1795	-0.000689	19.6	68	-0.0559	0.0000107	6.9
	Acetone					68	-0.2225	0.000410	0.4
Salicylic acid	Ethanol	33,69,70	-0.0318	0.000247	8.2	71,72	-0.0043	-0.000356	25.0
	Methanol					73	-0.0671	-0.0000660	16.6
	Acetone					73	-0.1224	-0.0000961	0.9
Benzoic acid	Methanol	35,70,74	-0.0203	0.000286	36.3	75	-0.0294	0.0000321	21.7
	Hexane					75	0.0167	-0.0000226	37.2
Hydroquinone	Methanol	76,77	-0.6003	0.00163	22.2	78	-0.4912	0.00131	0.9
Paracetamol[b]	Ethanol	37,79	0.0505	-0.0000209	39.2	80,81	-0.2419	0.000552	6.7
Cholesterol[c]	Methanol	82	0.3709	-0.000529	34.6	83	0.01185	-0.0000447	3.0
	Acetone					83	0.01139	0.000136	6.8
	Hexane					84	-0.02930	0.000231	2.6
Lauric acid	Ethanol	40	0.1800	0.000000	66.2	85	-0.000209	0.00000650	10.1
Myristic acid	Ethanol	40,86	-0.0929	0.000640	18.1	87	-0.2642	0.000952	13.0
Palmitic acid	Ethanol	41,42	-0.0389	0.000432	13.3	87,88	-0.0965	0.000325	17.9
Stearic acid	Ethanol	89	-0.0209	0.000348	9.6	90	-0.0892	0.000276	18.4

[a]Reference for experimental data; [b]Assuming paracetamol with one associating site; [c]Assuming cholesterol as a non-associating molecule.

Considering that ternary solubilities of solids in systems containing cosolvents are complex functions of temperature, pressure and cosolvent/cosolute composition[13], it is essential to observe the influence of different parameters such as temperature, pressure, concentration and type of the cosolvent, and number of associating sites of the solute, in the results.

4.1.Temperature and pressure effects

Solubility data of solutes in $scCO_2$ exhibit a typical behavior, as verified by the straight line when plotting the logarithm of solubility versus the logarithm of CO_2 density (which is a function of temperature and pressure). Commonly, the solubility increases when increasing pressure at given temperature, and the crossing of solubility isotherms, known as "crossover pressure”, represent the pressure at which solubility is independent of temperature. Knowledge on this crossing point can be useful when small differences in selectivity are required, such as in the separation of isomers[91]. By adding a cosolvent to the system, the crossover pressure point increases because it promotes an increase in the mixture critical point of the supercritical mixture[92].

Figure 4a presents the effects of temperature and pressure on the solubility of stearic and palmitic acids in $scCO_2$ + 0.73% ethanol. The highest values of solubility were obtained at higher pressure. However, the influence of the pressure is more relevant at lower pressures ($P \leq 15$ MPa). On the other hand, temperature increases the solubility more effectively at higher pressures. These solubility behaviors were also verified for the predicted data obtained using CPA model, indicating the accuracy of this equation to describe the influence of temperature and pressure on solubility data of solutes in CO_2 plus cosolvents. No crossover pressure was observed in the range of temperature and pressure here studied. However, **Figure 4a** indicates that crossover pressures are probably close to 12 MPa for these carboxylic acids, which were confirmed through the prediction with the CPA equation.

Temperature and pressure effects on the solubility can also be influenced by the concentration of cosolvent. **Figure 4b** shows the pressure effect on the solubility of trans-cinnamic acid when different amounts of ethanol were added to $scCO_2$ (0, 2 and 4 mol%). Experimental data indicates that the pressure effect becomes less important when higher concentration of cosolvent are used, as also mentioned in other studies [19,20,24]. Although CPA has shown a good accuracy of the CPA to describe this mixture, the decrease on pressure effect when cosolvent was added in $scCO_2$ was not well estimated. It is worth mentioning that CPA correctly predicted the effect of pressure on solubility data for the other systems in which pressure effect remained relevant at higher concentrations of cosolvent.

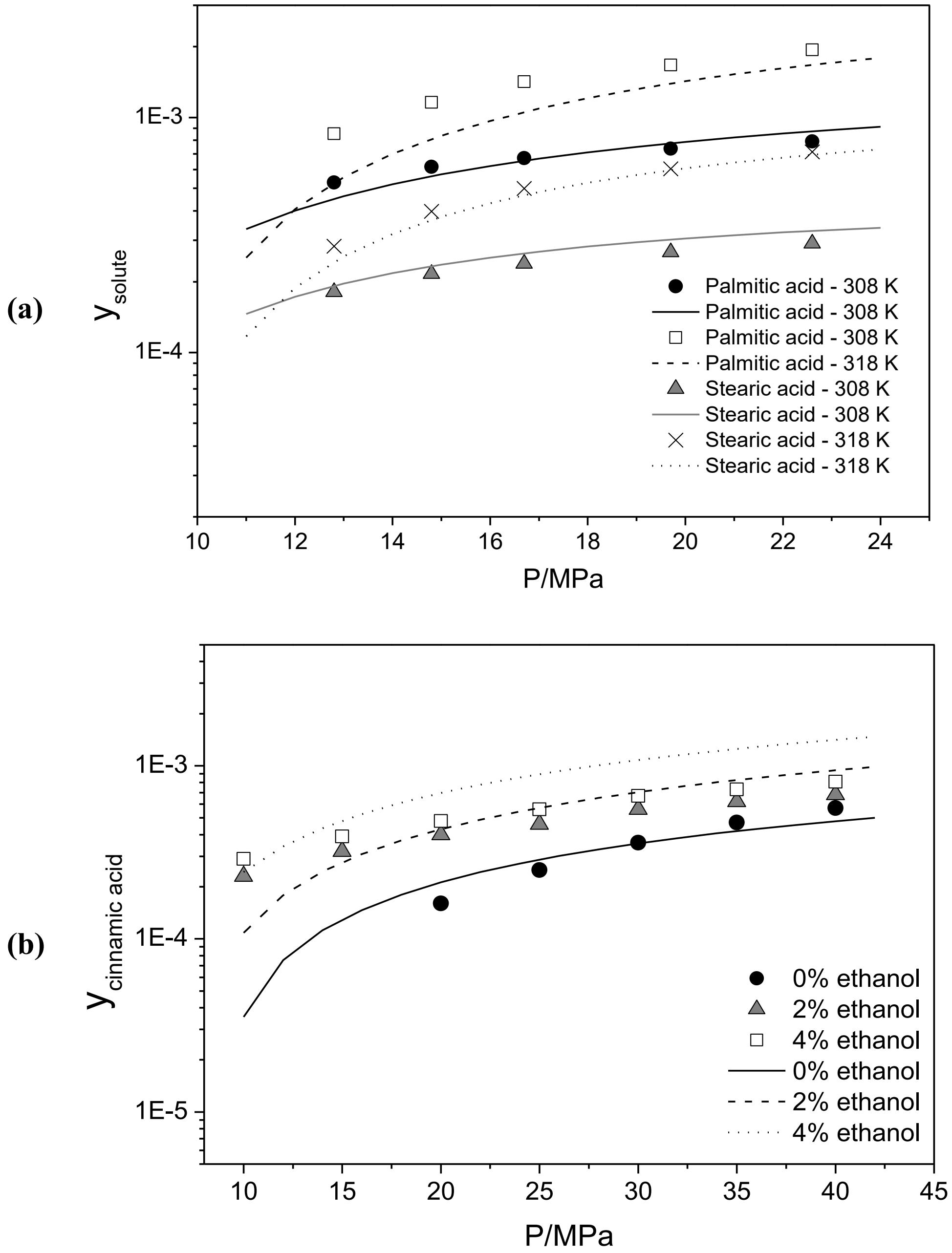

Figure 4. Effect of the temperature and pressure on: (a) Solubility of stearic and palmitic acids in CO_2 + 0.73% ethanol; (b) solubility of trans-cinnamic acid in CO_2 + ethanol at 313 K. The lines correspond to prediction data through CPA modeling.

4.2.Effect of the cosolvent type

It is also relevant to evaluate the effect of the cosolvent on the ternary solubilities of solids and on the prediction of phase equilibria data. Although there are few studies using non-polar organic solvent as the entrainer, the polar organic solvents such as methanol, ethanol and acetone are the most commonly used. The solubility of solid solutes in $scCO_2$ in the presence of a cosolvent, and consequently the choice of a cosolvent, is influenced by the ability of the solute to form hydrogen bond with the cosolvent and by the structure (size) and polarity of the molecule[93]. Therefore, there is not a specific cosolvent that should be used for all the solutes because the increase in the solubility of solids in ternary systems (CO_2 + cosolvent) is attributed to the increase in solvent density, dipole-dipole interactions and strong hydrogen bonds formed between the solute and the molecules of cosolvent.[31]

In this section, two examples are presented aiming to show the good ability of CPA-EoS to predict solubility data in $scCO_2$ when adding different cosolvent molecules. **Figure 5a** presents the solubility of the aspirin in $scCO_2$ + ethanol and $scCO_2$ + acetone. For both experimental and predicted data, the use of ethanol as cosolvent further increases aspirin solubility when compared to acetone. This result is probably associated with ethanol's polarity and with the fact of ethanol being an associating compound, in contrast to acetone.

The enhancement of the cholesterol solubility in $scCO_2$ with the addition of methanol and acetone as cosolvents was very similar (**Figure 5b**). At 328 K the acetone promotes a slightly higher increase in solubility of cholesterol compared to when methanol is used as a cosolvent. However, there was no visible difference at 318 K. In contrast to aspirin solubility behavior (**Figure 5a**), temperature showed to be most important than the cosolvent type to improve the solubility of cholesterol in $scCO_2$ + cosolvent. Even presenting different behaviors, the estimation of solubility data from CPA modeling was suitable for both systems presented in **Figure 5**, mainly at higher pressures, away from the critical point.

It is worth mentioning that the prediction of the behavior of solubility in the presence of cosolvents using thermodynamic models is also important to select a cosolvent, such as in the initial steps of design and optimization of processes.

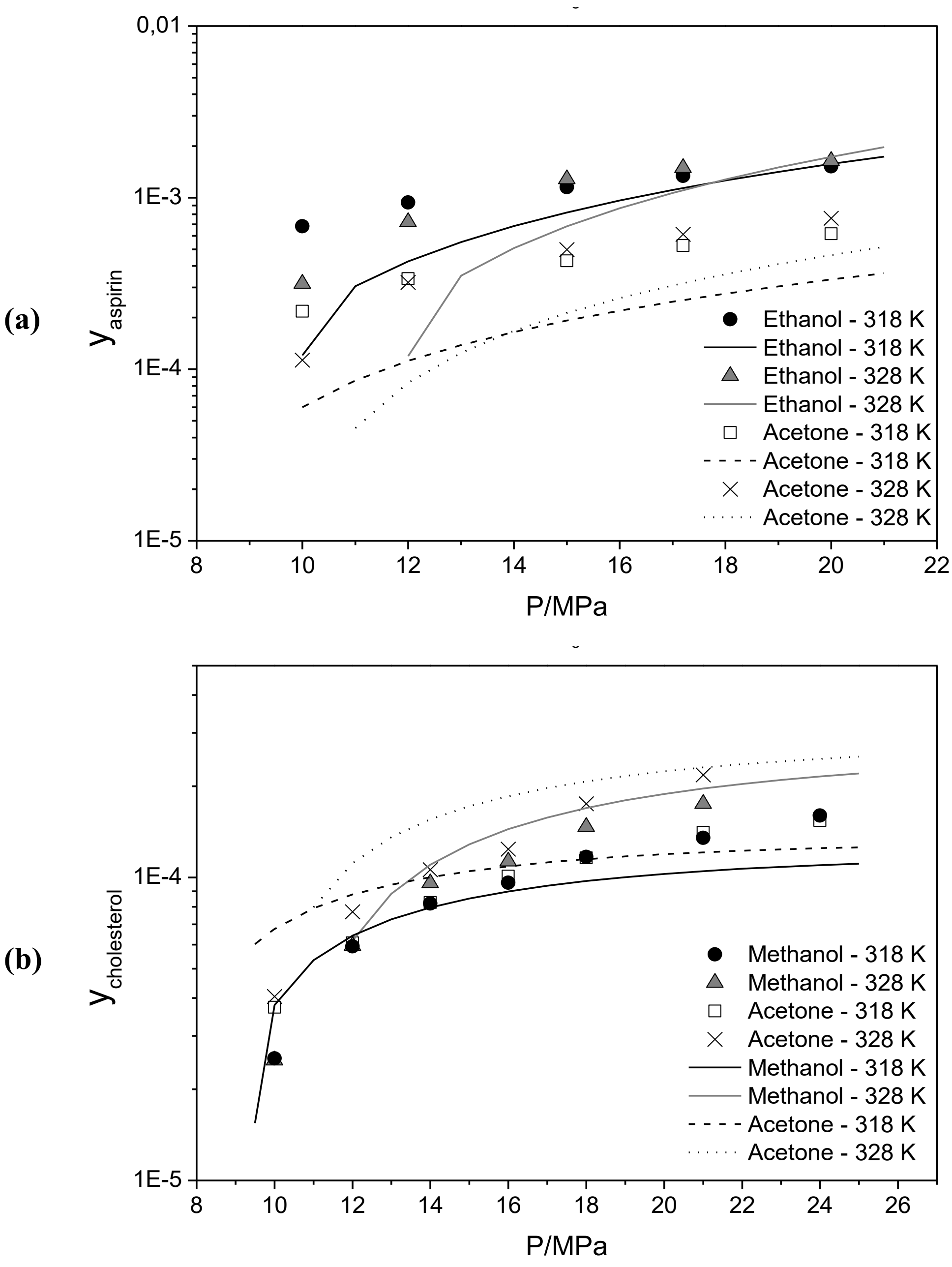

Figure 5. Effect of cosolvent type on: (a) aspirin solubility in CO_2 + 3% cosolvent; (b) cholesterol solubility in CO_2 + 3% cosolvent. The lines correspond to prediction data through CPA modeling.

4.3. Effect of the cosolvent concentration

To verify the quality of predictions using CPA when analyzing the effect of cosolvent concentration on solubility data, systems with diverse cosolvents were selected, which presented quite different influence of the cosolvent concentration on the phase equilibrium.

In general, the solute solubility increases at higher concentrations of cosolvent. However, while the solubility of trans-ferulic acid increases following an exponential behavior, with values up to 100 times higher than its solubility in pure $scCO_2$ when using 10 mol% ethanol (**Figure 6a**), the increase on benzoic acid solubility by increasing the hexane concentration is slightly relevant (**Figure 6b**). In both cases, predicted data followed the behavior of the experimental one.

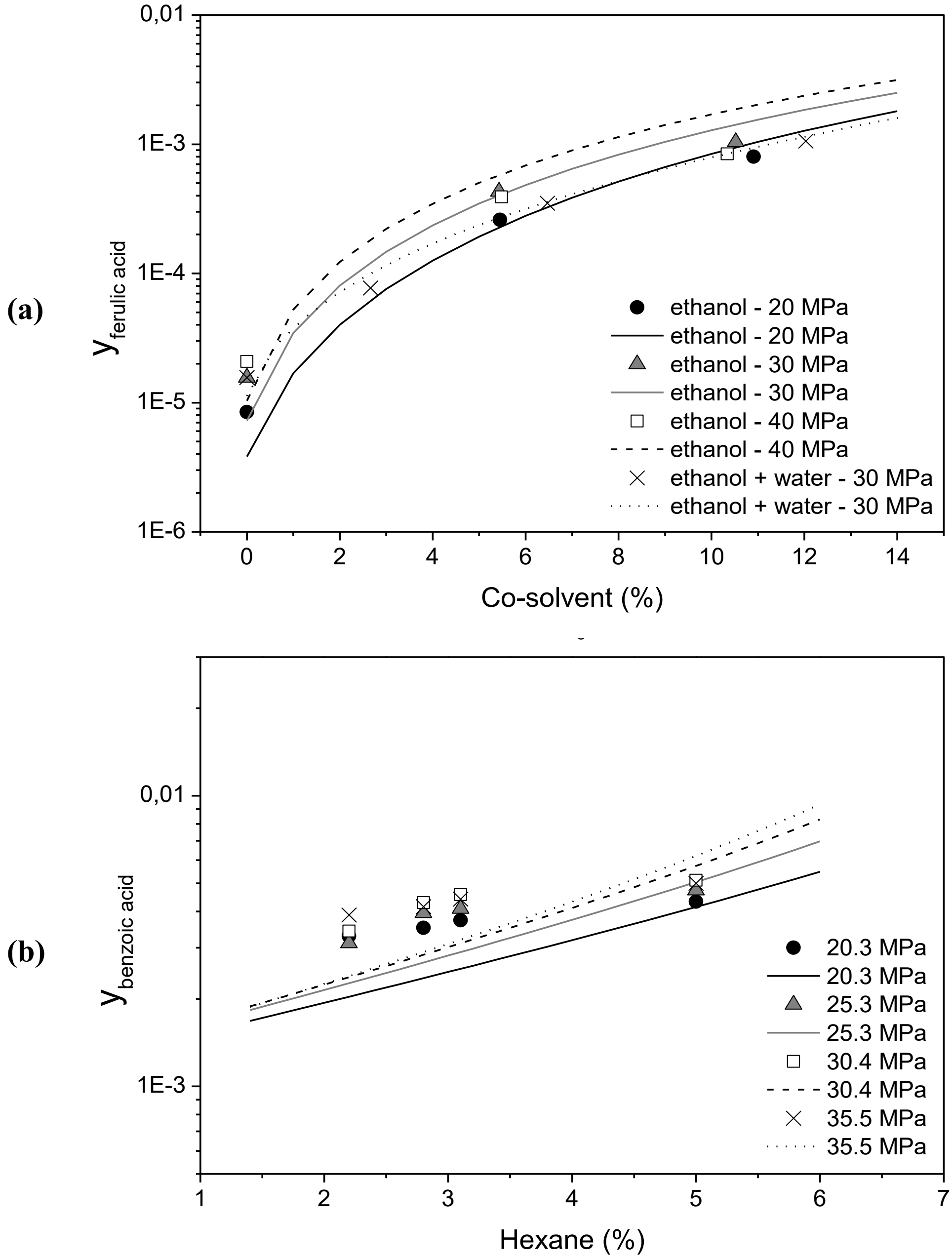

Figure 6. Effect of cosolvent concentration (molar basis) on: (a) trans-ferulic acid solubility in CO_2 + cosolvent at 323 K; (b) benzoic acid solubility in CO_2 + hexane at 308 K. The lines correspond to prediction data through CPA modeling.

The predictions for the only quaternary system investigated in this work are presented in **Figure 6a**. The excellent accuracy obtained with CPA indicates the possibility of applying this model to estimate the phase equilibrium in multicomponent systems. Furthermore, since there was good prediction in the quaternary system, it emphasizes and confirms that some faulty predictions for ternary systems can occur due to the reliability of both the solubility data in CO_2 + cosolvent and/or the pure components and the binary phase equilibrium data.

4.4.Effect of the association schemes

From the first step of thermodynamic modeling, which consist on fitting the pure component parameters of CPA-EoS, it is important to assume the type and number of association sites for each compound of the system. This section exemplified how these considerations can change the results for prediction. **Figure 7a** shows that the solubility of cholesterol in CO_2 + methanol was overestimated when cholesterol was assumed as an associating molecule due to the presence of one hydroxyl group. On other hand, CPA predictions presented good accuracy when non-association was considered. Thus, this result indicated that cholesterol cannot bond through this site, probably because it is a compound of large molecular chain, and the group will be too obstructed to promote hydrogen-bonding.

In the scientific literature, paracetamol has already been assumed both as a molecule with one [49] or two[57] association sites. Comparing the estimation of paracetamol solubility in CO_2 + ethanol from these different approaches (**Figure 7b**), a relevant difference in the prediction was also noted. Although both results were not that good, it is more suitable predict its solubility in $scCO_2$ + ethanol by assuming paracetamol as a molecule with one association site.

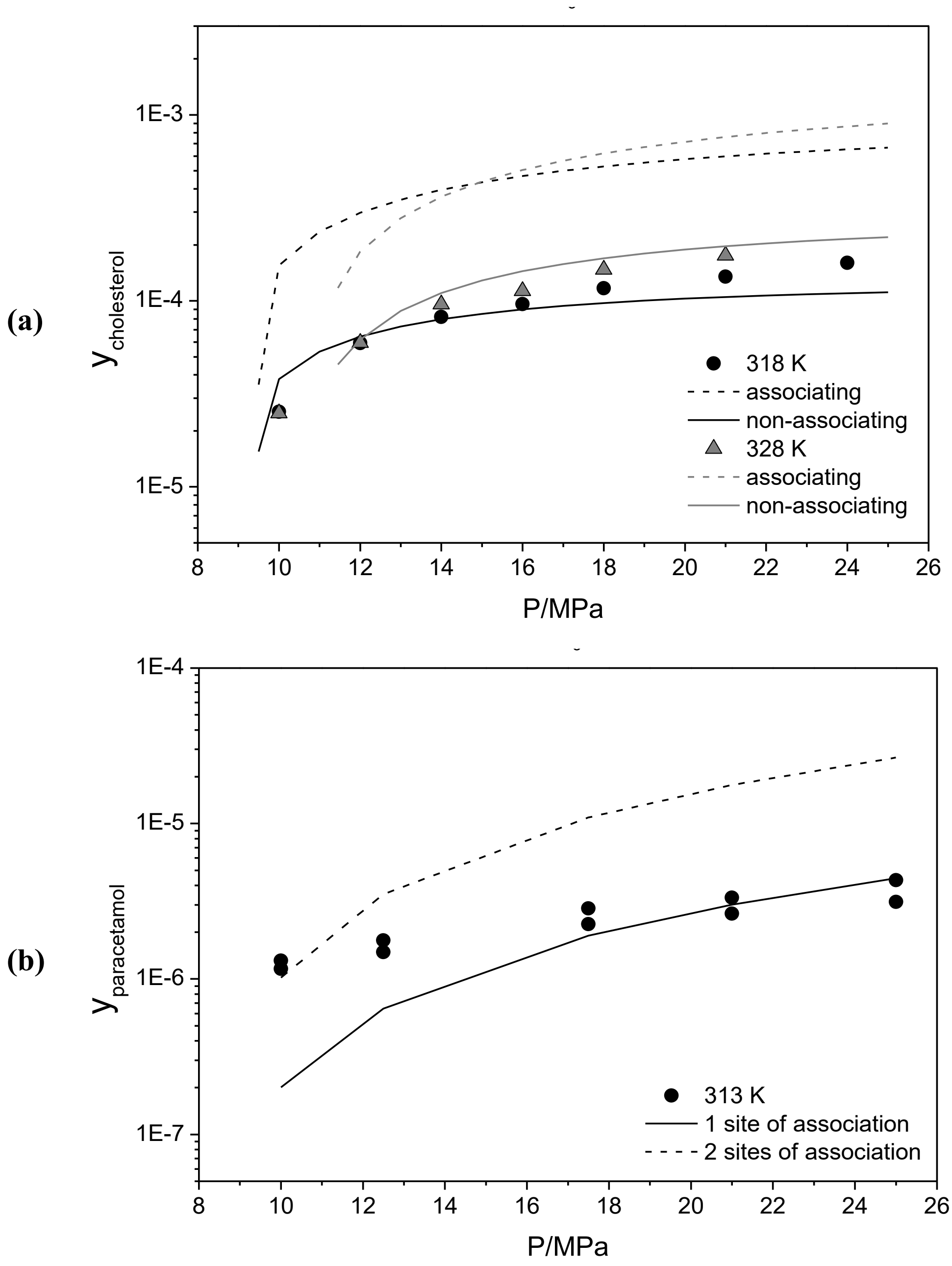

Figure 7. Effect of the number of association sites on CPA prediction of: (a) cholesterol solubility in CO_2 + 3% methanol; (b) paracetamol solubility in CO_2 + 0.85% ethanol at 313 K. The lines correspond to prediction data through CPA modeling.

5. Conclusions

The CPA equation presents good capacities in the description and prediction of solubility for solutes in supercritical carbon dioxide in the presence of organic cosolvents. These results were verified in systems containing different classes of compounds, both for solute and cosolvent. Furthermore, the effects of temperature, pressure, type and concentration of the cosolvent on the solubility of solids in carbon dioxide were calculated with reasonable accuracy.

However, it is needed to select suitable association squemes for solute and cosolvent molecules, and have a reliable source for experimental data of the main parameterization properties.

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Nomenclature

Abbreviations

$AARD$ - Average Absolute Relative Deviation - $AARD\,y\,(\%) = \frac{100}{N}\sum_{j=1}^{N}\left[\frac{\left|y_{cal} - y_{\exp}\right|}{y_{\exp}}\right]_j$

AG - association group

ALD - Average Logarithmic Deviation - $ALD = \sum_{j=1}^{N}\frac{1}{N}\left|\log\frac{y_i^{calc}}{y_i^{\exp t}}\right|$

CPA - Cubic Plus Association

COSMO-SAC - COnductor-like Screening Model - Segment Activity Coefficient

EoS - Equation of State

PR – Peng-Robinson

SAFT - Statistical Associating Fluid Theory

SRK - Soave-Redlich-Kwong

$scCO_2$ - supercritical carbon dioxide

List of symbols

a - energy parameter in the physical term

a_0, c_1 - parameters for calculating a

b - co-volume parameter in the physical term

g - radial distribution function

H - enthalpy

k_{ij} - binary interaction parameter

N - number of experimental data

P - pressure

R - gas constant

T - temperature

X_{Ai} - fraction of molecule not bonded at site A

x, y - mole fractions

Greek symbols

β - association volume

ε - association energy

Δ - association strength

ρ - molar density

η - reduced fluid density

φ - fugacity coefficient

Sub and superscripts

A, B - molecules sites indexes

C - critical property

cal - calculated

exp - experimental

i, j - pure component indexes

m - melting

fus – fusion

vap – vapor

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CAPÍTULO 6. HIGH PRESSURE PHASE EQUILIBRIUM OF THE CRUDE GREEN COFFEE OIL – CO_2 – ETHANOL SYSTEM AND THE OIL BIOACTIVE COMPOUNDS

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High pressure phase equilibrium of the crude green coffee oil - CO_2 – ethanol system and the oil bioactive compounds

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Abstract

Green coffee oil (GCO) is highlighted for applications in cosmetics industry, however GCO is usually obtained by cold pressing. In order to develop its extraction by supercritical technology experimental equilibrium data are required. This work presents phase equilibrium data of systems containing GCO, supercritical carbon dioxide ($scCO_2$) and ethanol, and evaluates the content of free fatty acids (FFA) and diterpenes in the samples. GCO solubility was evaluated at different temperatures (40-60 °C), pressures (200-400 bar) and $scCO_2$/ethanol ratios (0-5.7% w/w of ethanol). GCO solubility in $scCO_2$ was similar to other vegetable oils and Chrastil semi-empirical model fit well to the experimental data. By using 2.9% ethanol, the GCO solubility increased 63%. The samples presented FFA, kahweol and cafestol contents up to 3.4, 4.4 and 4.0 times higher than commercial GCO, respectively. The use of $scCO_2$ with or without ethanol promotes the extraction of GCO with higher FFA and diterpene contents.

Keywords: green coffee oil, supercritical carbon dioxide, ethanol, solubility, diterpenes and free fatty acids.

1. Introduction

In general, vegetable oils represent an important product on the world market, with a wide variety of economic areas of interest, from human and animal nutrition to the production of bioenergy [1]. The oil extracted by cold pressing from unroasted coffee beans, known as green coffee oil (GCO), has been indicated for applications in the pharmaceutical and cosmetic industry. Formulations containing GCO in cosmetics appear to be safe for topical applications and show good skin compatibility [2]. GCO is more expensive than most vegetable oils and it is composed of triacylglycerols (75%), steroids (5%) and diterpenes (19%), as well as small fractions of tocopherols and phenolic compounds [3]. Its composition is dependent on the coffee species, being the oil obtained from *Coffee arabica* the most suitable for cosmetic applications, since it offers the highest sun protection factor among the cultivated and commercialized coffee species [4].

Emollient properties associated with the presence of fatty acids, mostly linoleic and palmitic acids, are attributed to GCO, as well as evidence that this oil is capable of absorbing UVB radiation [5]. Furthermore, this product is a potent antioxidant due to the presence of chlorogenic acids and tocopherols [6,7]. There are reports of biological properties, including anticancer and anti-inflammatory properties, mainly because of its high content of unsaponifiable components, such as diterpenes, cafestol and kahweol, which are exclusively present in this oil [8,9].

GCO is usually obtained by cold pressing since the use of solvent, such as hexane, may leave residues in the final product. Although the lipid content in coffee beans varies from 7% to 17% [10], this extraction option makes the product expensive, since it generates about 30% less oil than the process with chemical solvents [11]. It is noteworthy that, in addition to green coffee beans, by-products of the coffee processing industry, such as spent coffee grounds and defective coffee beans, also have potential for application due to their high lipid content [12]. Thus, studies of methods that could provide higher GCO extraction yields, while maintaining its properties of interest, are necessary.

In this context, processes using supercritical carbon dioxide ($scCO_2$) stand out. In fact, this is an extraction technique whose advantages over alternative extraction methodologies have already been extensively described in the scientific literature.

The use of supercritical technology for extraction of coffee oil was evaluated in a few works. Araújo and Sandi [13] performed extractions from green and roasted coffee beans at different temperatures (60-90 °C) and pressures (235-380 bar), obtaining supercritical extracts at 70 °C/253 bar with a concentration of diterpenes 48% lower than in the oil extracted with hexane using a Soxhlet system. De Azevedo et al. [14] studied the extraction of oil and caffeine at 50-70 °C and pressures ranging between 15.2 and 35.2 MPa. In addition, the use of ethanol and isopropanol (5% w/w) provided higher extraction yields of caffeine, oil and chlorogenic acids [15]. More recently, Oliveira et al. [16] evaluated the extraction of oil, phenolic compounds and diterpenes from green coffee beans in a pressure range from 174.5 to 323.5 bar and temperatures from 65.9 to 94.1 °C, obtaining extracts enriched in diterpenes compared with oil obtained by pressing.

The optimization of $scCO_2$ processes can be based on experimental data of solubility in $scCO_2$ or on mathematical models that describe this behavior. Coffee oil solubility in $scCO_2$ was evaluated by Cornelio-Santiago et al. [17], de Azevedo et al. [14] and Sandi et al. [18], but the reported values are very different from each other.

The extraction of green coffee oil together with its minor compounds is a research subject still little investigated in the literature, mainly in the case of using ethanol as a cosolvent for $scCO_2$. The use of ethanol as cosolvent becomes particularly interesting for extracting GCO or for obtaining oil fractions, both enriched in its bioactive compounds. Therefore, this work provides equilibrium experimental data for systems containing green coffee oil, $scCO_2$ and ethanol, and also evaluates the content of diterpenes and free fatty acids (FFA) in the light equilibrium phase.

2. Experimental

2.1. Materials

Raw green coffee oil, extracted by cold pressing from unroasted *Coffea arabica* beans in February 2016, was supplied by Cooxupé - Regional Cooperative of Coffee Growers in Guaxupé (Guaxupé, Brazil). Carbon dioxide 99.5% w/w (White Martins Gases Industriais, Brazil) and absolute ethanol 99.5% (Synth, Brazil) were used as solvents.

Additional reagents were also needed: acetic acid 100% anhydrous (Merck, Germany), deionized water (Millipore, Milli-Q, USA), ethanol 95% (Synth, Brazil), ethyl ether 98% (Synth, Brazil), helium 99.995% (Linde Gases Industriais, Brazil), hexane >99% (Êxodo Científica, Brazil), methanol 99.8% (Sigma-Aldrich, USA), nitrogen 99.9% (Linde Gases Industriais, Brazil), potassium hydroxide (Êxodo Científica, Brazil), sodium hydroxide (Merck, Germany), tert-Butyl methyl ether ≥99.0% (Sigma-Aldrich, USA) and toluene 99.9% (Sigma-Aldrich, USA).

2.2.Oil characterization

The first step in GCO characterization was based on its content of minor compounds: free fatty acids, phenolic compounds, tocopherols and diterpenes. **Table 1** summarizes the analysis with the respective methodologies used.

Oil acidity, evaluated by titration with NaOH, was 2.4%, higher than values reported for *Coffee arabica* [19], which varied between 1.0 and 1.5% depending on the region of culture. The content of total phenolic compounds in GCO was within the range presented in the literature by Gutfinger [20] and Fuentes et al. [21] for different olive oil samples. Additionally, the content of tocopherols was similar to that reported for coffee oil in previous studies [22]. Finally, diterpene contents of GCO will be discussed in posterior sections (item 3.3).

Table 1

GCO characterization with regards to the minor compounds.

Analysis	Methodology	Reference	Result
Free fatty acids	Tritation with NaOH	[23]	2.4%
Diterpenes	GC/MS	[16,24]	
Kahweol			3.4 g/kg
Cafestol			11.6 g/kg
Total Phenolic	Spectrophotometry	[25]	0.088 mg GAE[a]/g
Tocols	UPLC/MS	[26]	
α - tocopherol			0.133 mg/g
β/γ - tocopherol			0.347 mg/g

[a] GAE: gallic acid equivalent

2.3.Measurements of solubility

The GCO - CO_2 - ethanol system was only evaluated via analysis of the supercritical phase (extract phase). Thus, the measurements represent the solubility of GCO in CO_2/ethanol mixtures. **Fig. 1** shows the supercritical extraction unit used in this work, located at the EXTRAE laboratory (UNICAMP, Brazil), and which was already employed in prior works developed for the same research group [27,28]. The experimental runs were carried out at different temperatures (40, 50 and 60 °C), pressures (200, 300 and 400 bar) and ethanol quantities in the CO_2/ethanol mixture (0.0, 2.9 and 5.7% w/w).

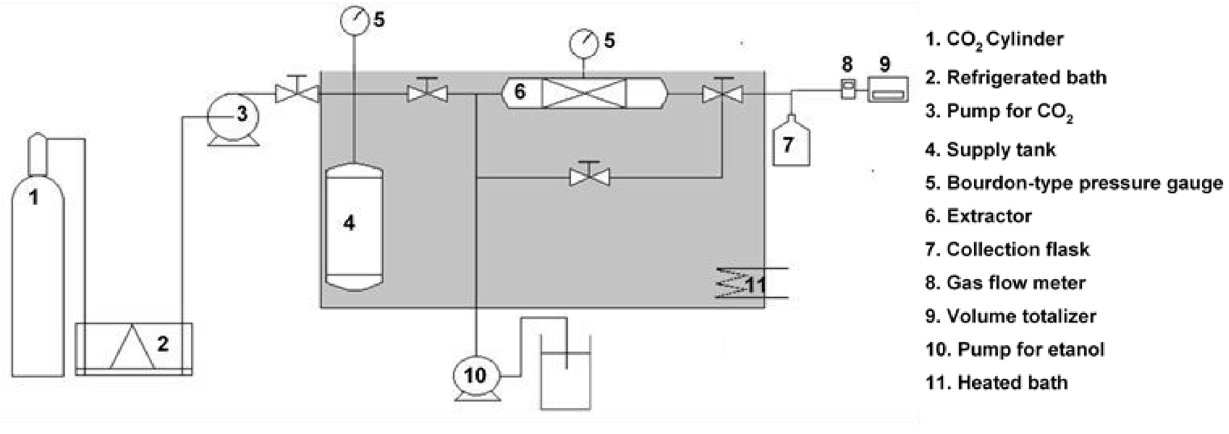


Fig. 1. Experimental unit for determination of the green coffee oil (GCO) solubility in $scCO_2$ and $scCO_2$ + ethanol.

First, paper soaked with approximately 15 g of GCO was packaged in a stainless steel extractor (100 mL - 32 cm length). The extractor was pressurized with pure CO_2 until reaching the desired pressure condition and the temperature was adjusted by a thermostatic bath in which the extractor was submerged. After 30 minutes for stabilization of the system, the supercritical extraction process started using a CO_2 flow rate of approximately 0.8 L min^{-1} (1.32 g min^{-1}) and ethanol flow rate of 0.05 mL min^{-1} (0.039 g min^{-1}) or 0.1 mL min^{-1} (0.079 g min^{-1}). One sample was collected for every 10 L of spent CO_2. Finally, the samples were concentrated in a rotary evaporator (Marconi rotary evaporator, model MA-120, Brazil) and maintained in an oven (Marconi, model MA030 / 12, Brazil) under vacuum at 50 °C until reaching constant weight (Precisa, model XT220A, Sweden).

Extraction curves of oil mass versus solvent mass were constructed and good linear adjustments indicated the GCO solubility in pure $scCO_2$ or $scCO_2$ + ethanol. Thus, the solubility ($\overline{Y}$) was expressed by the ratio between mass of oil and solvent (CO_2 or CO_2 + ethanol). GCO solubility in $scCO_2$ was modeled by the equation proposed by Chrastil [29], which correlates the solubility of solids and liquids in supercritical fluids as a function of density and temperature. This equation has been widely used to correlate and extrapolate solubility experimental data of several compounds in supercritical fluids due to its ability to reproduce experimental data and its relative simplicity compared with state equations. The Chrastil equation was rewritten, in terms of mass fraction, as presented below [30].

$$\ln \overline{Y} = (k-1).\ln \rho + \left(\frac{a}{T} + b \right) \tag{1}$$

$\overline{Y}$ is the solubility (kg/kg CO_2), ρ is the solvent density ($kg.m^{-3}$), T is the temperature (K) and k, a and b are parameters to be adjusted to the experimental data. CO_2 density values at the different temperatures and pressures were obtained from Angus et al. [31] and adjustment of the parameters (k, a and b) was performed by the Nelder and Mead method [32], minimizing the objective function given below (S) (**Equation 2**).

$$S(k,a,b) = \sum_{j=1}^{N} \sum_{i=1}^{2} \left[\left(\frac{\overline{Y}_{cal}(i) - \overline{Y}_{\exp}(i)}{\overline{Y}_{\exp}(i)} \right)^2 \right]_j \tag{2}$$

2.4.Analytical methods

2.4.1. Quantification of free fatty acids (FFA)

The FFA contents were evaluated by high-performance liquid chromatography (HPLC) according to published methods [33,34]. The HPLC system was equipped with a 100 Å Phenogel™ size exclusion column (300 mm × 7.8 mm ID, 5-µm) (Phenomenex, Torrance, CA, USA) and a refractive index detector (Shimadzu, model RID-10A, Japan). The column temperature was set at 40 °C and the injection volume (Shimadzu, model SIL-20A, Japan) was 20 µL. Elution was carried out in isocratic mode using a mobile phase composed of a toluene solution with 0.25% acetic acid (v/v) at a flow rate of 1 mL min^{-1}. FFA contents were quantified by a calibration curve (external calibration), constructed with GCO and linoleic acid (60-74% of purity, Sigma-Aldrich, USA) at six levels of known concentrations. The oil content of the samples to be analyzed by HPLC was 10 mg mL^{-1}, prepared by dilution in toluene.

2.4.2. Quantification of diterpenes

First the samples were hot saponified, according to the methodology adapted from Chartier et al [24]. About 200 mg of the sample and 2 mL of a 1.25 M KOH solution in methanol were mixed and maintained in a thermostatic bath (Marconi MA-120, Brazil) at 70 °C for 1 h. Then this solution was washed, supplemented with 2 mL of tert-butyl methyl ether (MTBE) and 2 mL of Milli-Q water, and centrifuged (Eppendorf 5430 R, USA) at 3000 rpm and 25 °C for 5 min. The supernatant was collected and the heavy phase was washed two more times for extraction of the unsaponifiable compounds present in the sample. The total supernatant collected was dried in a concentrator with N_2 flow (Tecnal TE-019, Brazil). Then, 3mL of MTBE was added to the dry sample and the final solution was filtered through Sartorius PTFE 0.20 µm syringe filters.

The identification and quantification of kahweol and cafestol was performed by gas chromatography (Shimadzu CG-2010, Japan) coupled to mass spectrometry (Shimadzu GCMS-QP2010 Plus, Japan), using the external calibration method with standards of kahweol and cafestol (Santa Cruz Biotechnology, USA) diluted in MTBE and at concentrations ranging from 90 to 4900 mg L^{-1}.

The chromatographic conditions were adapted from Oliveira et al. [16]. 1.0 µl of the solution was injected into a capillary column (30 m x 0.25 mm x 0.25 µm composed of 5% diphenyl/95% dimethyl polysiloxane) (Restek RTx®-5MS, USA) in split mode (ratio 1:21.5).

Helium was used as the carrier gas at 1.2 mL min^{-1} and the temperature of the injector and the source of ionization were 270 °C and 300 °C, respectively. The oven temperature started at 50 °C for 1.0 min, increased at 20 °C min^{-1} up to 250 °C, followed by an increase of 5 °C min^{-1} up to 300 °C, a temperature that was maintained for additional 4 min, totaling 25.0 min. The mass scan range was 40–650 m/z.

2.4.3. Oil composition in fatty acids and triacylglycerols

Determination of the fatty acid composition was performed by gas chromatography of the fatty acid methyl esters according to the official procedure Ce 1f-96 of the American Oil Chemists' Society [23]. Thus, the oil was first derivatized to form fatty acid methyl esters according to the method of Hartman and Lago [35].

Oil composition in triacylglycerols (TAG) was evaluated by the methodology of Antoniosi Filho et al. [36], which consists of a statistical algorithm for estimating the distribution of fatty acids in the TAG molecules. This method results in a large number of TAGs and in order to reduce the number of components, all structural isomers have been divided into a set of components with the same number of carbon and double bonds. The sets of isomers were named according to the major TAG and groups with a total TAG concentration lower than 0.5% (w/w) were ignored.

3. Results and discussion

3.1.GCO - CO_2 system

The GCO - CO_2 system was investigated at three different temperatures (40, 50 and 60 °C) and pressures (200, 300 and 400 bar). Extraction curves of oil mass versus solvent mass were constructed and linear adjustments of all curves were obtained with $R^2 \geq 0.99$, as shown in **Fig. 2.** Thus, it was possible to obtain the solubility data of the GCO in pure $scCO_2$ and $scCO_2$/ethanol (97.1: 2.9 w/w). It was observed that the slope of each extraction curve was altered after some extraction time, an effect observed when ethanol starts to be altogether collected together with the extract, indicating that the CO_2/ethanol solvent mixture was obtained at the selected mass ratio.

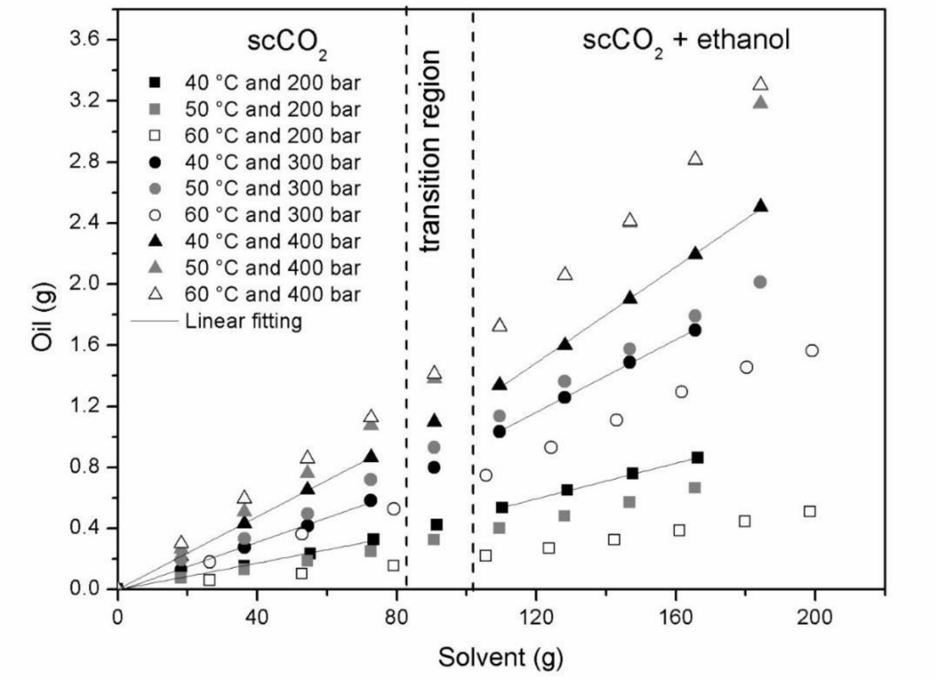


Fig. 2. Extraction curves of GCO at different pressures and temperatures.

The average values of GCO solubility in scCO$_2$ varied between 2.0 and 14.8 g/kg CO$_2$, being the lowest value obtained at 60 °C and 200 bar and the highest solubility found at 60 °C and 400 bar (**Table 2**). Data presented by Sandi et al. [18] and Cornelio-Santiago et al. [17] are very different from those obtained in this work. However, our solubility data in scCO$_2$ were closer to those reported by Azevedo et al. [14], although it is not possible to assign deviation values to both data sets since the temperature and pressure conditions were not equivalent.

Comparing the GCO solubility with the solubility of other vegetable oils, a small deviation was verified under the same conditions of temperature and pressure [37,38]. The divergence may be attributed to differences in the oil compositions, not only that part of the composition based on the fatty acids distribution, but also to the presence of other components. As previously mentioned, GCO has a higher content of unsaponifiable matter [3].

Table 2

GCO solubility in CO_2 and CO_2 + ethanol at different pressures and temperatures.

Ethanol 100 W[a]	T °C	P bar	ρ kg.m^{-3}	$\bar{Y}^b$ experimental g/kg solvente	$\bar{Y}^b$ calculated g/kg solvent		
					Chrastil Equation	Equation by Azevedo et al. [14][c]	Equation by del Valle and Aguilera [38][d]
0.0	40	200	841	4.0 ± 0.6	3.77	4.85	3.42
		300	910	7.5 ± 0.3	7.76	8.36	7.36
		400	956	11.8 ± 0.3	12.2	11.7	11.9
	50	200	785	3.1 ± 0.4	2.98	4.01	2.85
		300	871	7.6 ± 0.3	7.69	8.21	7.84
		400	923	13.2 ± 0.7	13.1	12.2	13.8
	60	200	725	2.0 ± 0.3	2.08	3.02	2.17
		300	830	6.8 ± 0.2	7.16	7.69	8.08
		400	890	14.8 ± 0.7	13.5	12.4	15.9
2.9 ± 0.1	40	200		6.0 ± 0.1			
		300		12.3 ± 0.2			
		400		16.2 ± 0.5			
	50	200		4.7 ± 0.3			
		300		11.7 ± 0.3			
		400		18.9 ± 0.8			
	60	200		3.3 ± 0.1			
		300		9.8 ± 0.1			
		400		19 ± 1			
5.7 ± 0.1	50	300		18 ± 1			

[a] W: mass fraction of ethanol on a solute-free basis;

[b] $\bar{Y}$: green coffee oil solubility in CO_2 or CO_2 + ethanol;

[c] Equation reported by Azevedo et al. [14] for GCO solubility in $scCO_2$;

[d] Equation reported by Del Valle and Aguilera [38] for solubility of soybean, sunflower, cottonseed and corn oils in $scCO_2$.

In general, GCO solubility was higher when increasing the pressure; however the effect of temperature on GCO solubility was not so simple. At pressures bellow 310 bar, lower temperatures showed the higher solubility values, and at pressures above 350 bar the behavior was reversed, as shown in Fig. 3.

The pressure at which there is a cross between the isotherms is known as the crossover point and it is common for oil solubility data in $scCO_2$. GCO solubility data present a region within the pressure range of 310 and 350 bar in which the transition of temperature dependence occurred. This transition region is probably due to the fact that GCO represents a mixture of components, as has also been reported in literature, including for GCO [14,39].

The inversion of temperature dependence observed in the solubility data was explained by King and Bott (1993), apud Ozkal et al. [40], and occurs due to the concurrent effects of reducing the density of $scCO_2$ and increasing the volatility of fatty acids, which accompany the increase in temperature. Thus, at lower pressures the variation of the solvent density is more significant than the change in vapor pressure of the solute, and consequently the solubility increases with a decrease in temperature.

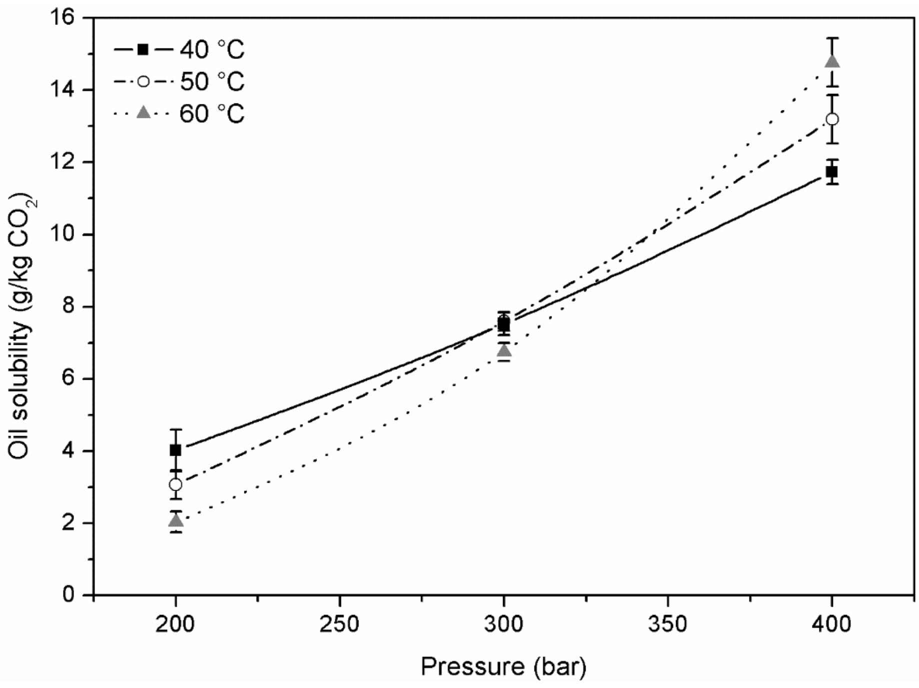


Fig. 3. GCO solubility in CO_2 as a function of pressure.

The Chrastil semi-empirical equation was applied to correlate the experimental data. **Fig. 4** shows the linear behavior obtained between the natural logarithm of the GCO solubility in $scCO_2$ and the natural logarithm of CO_2 density, under the same conditions of temperature and pressure. The parameters of the Chrastil equation that correlate the experimental data of GCO solubility in $scCO_2$ are also presented in **Fig. 4**. The average absolute relative deviation (AARD), calculated according to Equation 3, was 3.85%.

$$AARD\ (\%) = \frac{100}{N}\sum_{N=1}^{N}\frac{\left|\overline{Y}_{\exp} - \overline{Y}_{calc}\right|}{\overline{Y}_{\exp}} \quad (3)$$

, N = number of experimental data

Two other correlations derived from the Chrastil equation were used to evaluate the experimental data, as shown in **Table 2**. Del Valle and Aguilera [38] adjusted an equation for solubility data of soybean, sunflower seed, cottonseed and corn oils, and Azevedo et al. [14] reported an equation for the solubility of green coffee oil in $scCO_2$. All calculated values were similar to the experimental data reported in this work, indicating that green coffee oil presents solubility comparable to most vegetable oils. The deviations between our data and the corresponding values calculated using the equations proposed by Del Valle and Aguilera and Azevedo et al. were 7.5% and 17.3%, respectively.

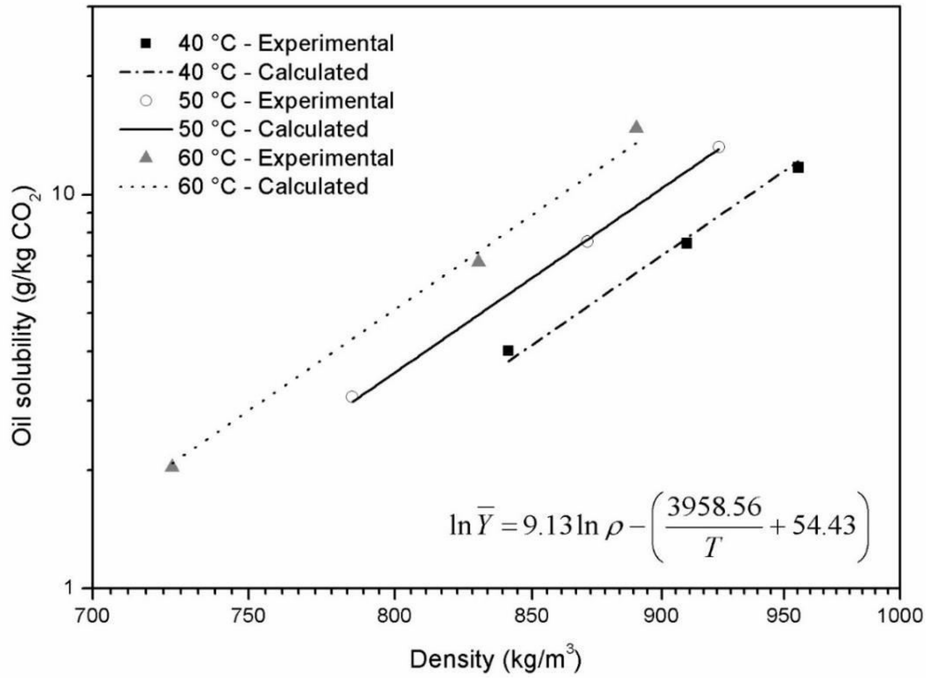


Fig. 4. GCO solubility in CO_2 as a function of CO_2 density.

3.2.GCO - CO_2 - ethanol system

The system GCO - CO_2 - ethanol was evaluated with respect to the effects of temperature (40, 50 and 60 °C) and pressure (200, 300 and 400 bar) using one of the selected ratios of the CO_2/ethanol mixture (97.1:2.9 w/w). At 50 °C and 300 bar, the system was also measured when adding 5.7% of ethanol by mass, on a solute-free basis. **Table 2** showed that

GCO solubility increases with the addition of ethanol. This behavior was expected, since it has been reported for other types of oils and lipid compounds [41–43] and also for compounds present in coffee oil [44]. When adding ethanol and increasing the pressure, the density of the solvent mixture and the physical interactions between solute and solvent are higher, contributing to increase the solubility [45].

The effects of temperature and pressure were similar to those observed in the GCO – CO_2 system, including the crossover point between the isotherms. However, it should be noted that the pressure values at which the inversion of behavior occurs changed was into the range from 330 to 370 bar (**Fig. 5**). The higher crossover point compared to that with pure $scCO_2$ (310 to 350 bar) can occur because ethanol addition increases the critical point of the supercritical mixture [45].

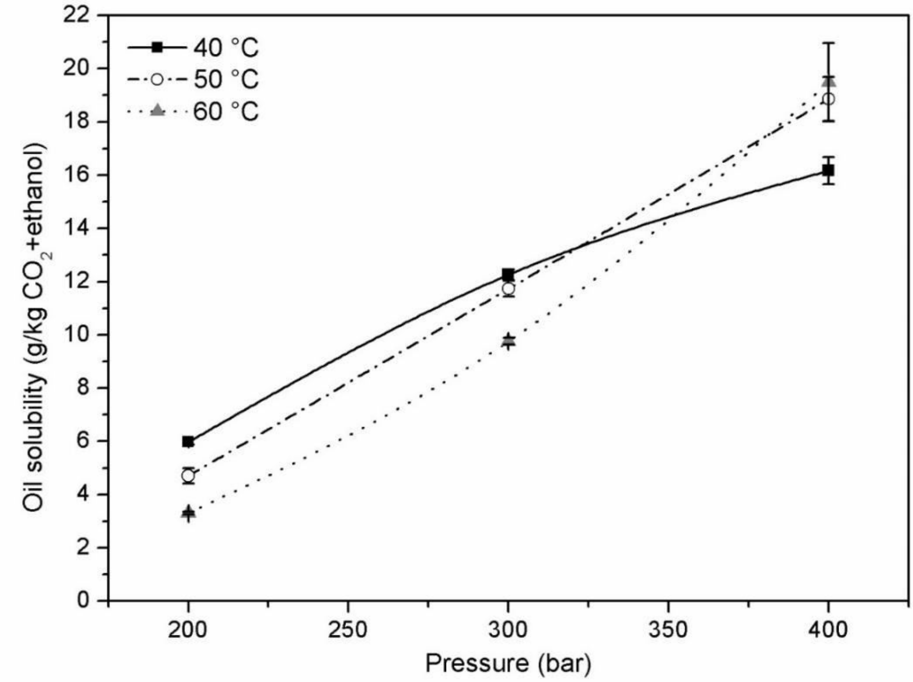


Fig. 5. GCO solubility in CO_2 + ethanol as a function of pressure.

At 50 °C and 300 bar, when the amount of ethanol in the system is increased from 0 to 2.9% and then to 5.7%, the solubility of GCO increases by 54% (**Table 2**), indicating that the use of an even higher proportion of ethanol may be of interest.

In studies on the solubility of biocompounds in $scCO_2$ with the addition of a cosolvent, it is common to investigate the influence of temperature and pressure upon the cosolvent effect by evaluating the ratio between the solubility in the CO_2 + ethanol mixture and the solubility in pure CO_2. Values of the cosolvent effect obtained in this work ranged from 1.32 to 1.63, indicating that the addition of 2.9% of ethanol to $scCO_2$ increased GCO solubility by up to 63%. For comparison purposes, the increase in fish oil solubility was 89% at 60 °C and 300 bar with the addition of 4.6% of ethanol [43].

In general, the influence of the cosolvent decreased when the system pressure is increased, an effect that can be attributed to the reduction of differences between the local densities and the bulk solution density as it moves away from the critical point [46]. The highest average values of the cosolvent effect were obtained at 60 °C/200 bar and 40 °C/300 bar. Nevertheless, due to the propagation of errors associated with the calculation of this property and, consequently, to the corresponding high standard deviations, the differences of values observed for this effect are not statically significant, so that the influence of temperature and pressure on the cosolvent effect could not be clearly evaluated.

Experimental data of the GCO - CO_2 – ethanol system were not described using the Chrastil equation due to the scarcity of density data for the solvent mixture (CO_2 + ethanol). Thermodynamic modelling using equations of state are also complicated, because GCO is composed by different classes of compounds in a proportion more significant than in the case of refined oils. Although Cornelio-Santiago et al. [17] considered coffee oil as a mixture of 6 triacylglycerols (TAG) to model their solubility data on pure $scCO_2$, the Peng-Robinson state equation failed to predict the solubility of the binary (TAG and $scCO_2$) and multicomponent (TAGs and $scCO_2$) systems. In the present case, the corresponding modelling difficulties are higher since the samples obtained in this work contain significant amounts of FFAs and diterpenes.

3.3.Free fatty acid and diterpene contents

In order to evaluate the distribution of FFA and diterpenes present in the extract phase of the systems GCO - CO_2 and GCO - CO_2 - ethanol, three extraction curves (400 bar: 40, 50 and 60 °C) were selected and their contents of FFA and diterpenes (kahweol and cafestol) were measured.

Fig. 6a shows the mass of FFA extracted per solvent mass. The curves are straight lines with $R^2 \geq 0.99$ and the data behavior as a function of temperature was similar to that observed for oil extraction: at higher temperatures greater quantities of free fatty acids are extracted. At 40 °C and 60 °C, **Fig. 6a** shows an important difference in the inclination of the straight lines obtained for the "CO_2" and "CO_2 + ethanol" regions. This result indicates an 11% increase in FFA extraction when ethanol is added to the system.

Fig. 6b and **6c** present the extraction curves for kahweol and cafestol at 400 bar as a function of the solvent mass, respectively. Straight lines with R^2>0.97 were adjusted and higher amounts of diterpenes, both kahweol and cafestol, were obtained at lower temperatures.

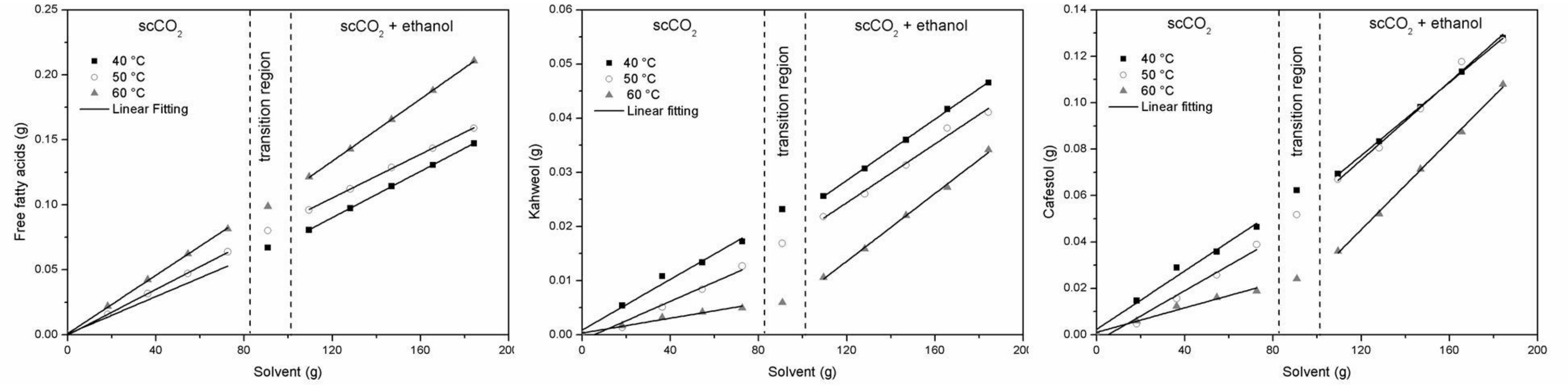
Furthermore, the slope differences between the straight lines obtained for the "CO_2" and "CO_2 + ethanol" regions were more significant at 60 °C.

Because the linear fittings presented good results, the concentrations of FFA and diterpenes in the samples obtained for the GCO - CO_2 and GCO - CO_2 - ethanol systems were calculated by the average of the points shown in **Fig. 6**. The results are presented in **Table 3**.

The samples of the extract phase obtained with $scCO_2$ and $scCO_2$ + ethanol (referred to as SF and SEF, respectively) had a fatty acid content up to 3.4 times greater than the commercial green coffee oil, whose acidity was 2.1%. Although the addition of ethanol improves the FFA extraction, this increase is less important than the increase observed in the extraction of other oil components, since there was a reduction in the FFA concentration of the samples compared with the samples obtained using pure CO_2. In other words, because the mass of extract (oil) obtained with CO_2 + ethanol was much higher than the mass with pure CO_2, the corresponding acid content in the SEF samples are lower than the values obtained in the case of the SF samples.

Note that the acid content of GCO presented in **Table 3** is different from that previously reported in **Table 1.** This fact occurs because the methods of analysis were different: the HPLC method analyzes only the free fatty acids, while the acidity determination by titration quantifies any acid compound present in the samples.

In this work, both kahweol and cafestol represent 1.5% of GCO (15 g/kg), but the samples from phase equilibrium of the GCO – CO_2 – ethanol system presented diterpene contents up to 60 g/kg, which corresponds to an increase of 4.4 and 4.0 times in the contents of kahweol and cafestol, respectively. Oliveira et al. [16] reported green coffee oil obtained by pressing with 12.1 and 20.3 g {kahweol + cafestol}/kg, depending on the storage time.



(a) (b) (c)

Fig. 6. Extraction curves of (a) free fatty acids (FFA) (b) kahweol and (c) cafestol at 400 bar.

Table 3

Contents of FFA and diterpenes in samples of the GCO - CO_2 - ethanol system.

Sample	FFA %	Kahweol g/kg oil	Cafestol g/kg oil	Diterpenes {Kahweol + Cafestol} g/kg oil
GCO[a]	2.1	3.4	11.6	15.0
SF[b]				
400 bar / 40 °C	6.1 ± 0.4	15 ± 3	41 ± 9	56 ± 10
400 bar / 50 °C	5.9 ± 0.4	12 ± 4	37 ± 10	49 ± 10
400 bar / 60 °C	7.2 ± 0.2	4 ± 1	17 ± 4	21 ± 4
SEF[c]				
400 bar / 40 °C	5.7 ± 0.4	13 ± 2	38 ± 5	51 ± 5
400 bar / 50 °C	4.4 ± 0.5	15 ± 2	46 ± 4	61 ± 4
400 bar / 60 °C	5.8 ± 0.7	15 ± 2	46 ± 6	61 ± 6

[a] *GCO*: crude green coffee oil (pressing);
[b] *SF*: supercritical fraction;
[c] *SEF*: supercritical + ethanol fraction.

Araújo and Sandi [13] obtained supercritical extracts with diterpene contents ranging from 4.13 to 8.6 g/kg. Chartier et al. [24] reported contents that ranged from 57.6 to 73.3 g/kg as a result of a transesterification reaction and liquid-liquid extraction. Tsukui et al. [47] showed a diterpenes yield ranging from 75.8 to 131.0 g/kg by microwave assisted extraction. The comparison between these methods becomes complicated because variations in the diterpene content may be also due to factors related to the cultivation of the coffee bean and due to part of the terpenes and sterols that are in the form of esters bounded to free fatty acids.

Furthermore, the main effect of the addition of ethanol to the system was verified at 60 °C, the temperature representing the worst condition for extraction of diterpenes with pure CO_2 (sample SF). At this condition, the addition of ethanol (SEF sample) caused an increase of 3.6 and 2.7 times in the concentration of kahweol and cafestol, respectively.

In addition, the results of **Table 3** indicate that triacylglycerols from GCO present lower solubility in CO_2 and CO_2 + ethanol than FFA, kahweol and cafestol, since the contents of FFA, kahweol and cafestol for both samples (SF and SEF) are higher than the corresponding content in green coffee oil obtained by pressing (GCO).

3.4.Fatty acid and triacylglycerol composition

Three coffee oil samples had their fatty acid composition analyzed: commercial green coffee oil (GCO), one supercritical fraction (SF) and one of the fractions obtained with supercritical carbon dioxide and ethanol (SEF). **Table 4** indicates that there was no difference between the major fatty acids in the samples and that the mass percentage of each acid also remained similar. In all samples palmitic and linoleic acids represent 78 to 80% of the fatty acids. The composition obtained in this work is similar to the ones reported in prior works [48–51].

Table 4

Fatty acid composition of the green coffee oil samples.

Fatty acids	Symbol	C*x*:*y*^{b}	M^a g mol^{-1}	*GCO*^{d} 100 w^c	*SF*^{e} 100 w^c	*SEF*^{f} 100 w^c
Palmitic acid	P	C16:0	256.4	33.78	36.07	35.01
Stearic acid	S	C18:0	284.5	7.80	6.87	7.35
Oleic acid	O	C18:1	282.5	9.31	9.03	9.21
Linoleic acid	Li	C18:2	280.4	44.26	44.04	43.99
Linolenic acid	Ln	C18:3	278.4	2.62	2.03	2.31
Arachidic acid	A	C20:0	312.5	1.39	1.37	1.39
Gondoic acid	Ga	C20:1	310.5	0.25	0.17	0.24
Behenic acid	Be	C22:0	340.6	0.60	0.42	0.50

^{a}*M:* molar mass;

^{b}C*x*:*y* where x = number of carbons and y = number of double bonds;

^{c}*w*: mass fraction;

^{d}*GCO*: crude green coffee oil (pressing);

^{e}*SF*: supercritical fraction;

^{f}*SEF*: supercritical + ethanol fraction.

From the fatty acid composition, it was possible to determine the probable triacylglycerol (TAG) composition of the green coffee oil, as shown in **Table 5**. The main TAG represents the component with the greatest mass fraction of the isomer set defined by the number *x* of carbon atoms in the fatty acid moieties and by the number y of double bonds. PLiP, LiLiP, PLiS and POLi were the major TAGs found, representing more than 65% of the total. This composition is similar to that calculated by Cornelio-Santiago et al. [17].

Table 5

Hypothetical triacylglycerol composition of the green coffee oil[a].

main TAG[b]	*x*:*y*[c]	*M*[d] g mol^{-1}	100 *w*[e]
PPP	48:0	807.3	0.51
POP	50:1	833.4	4.80
POA	54:1	889.5	0.93
PLiP	50:2	831.4	22.93
PLiS	52:2	859.4	10.84
PLiA	54:2	887.5	4.43
PLiBe	56:2	915.5	1.42
PLnP	50:3	829.3	0.72
POLi	52:3	857.4	9.65
SOLi	54:3	885.4	2.42
AOLi	56:3	913.5	0.73
LiLiP	52:4	855.4	22.59
LiLiS	54:4	883.4	5.58
LiLiA	56:4	911.5	1.57
PLnLi	52:5	853.4	1.41
LiLiO	54:5	881.4	3.76
LiLiLi	54:6	879.4	5.70

[a] Calculated from fatty acid composition of commercial green coffee oil;

[b] Major component in the triacylglycerol (TAG) set with the same values of x and y;

[c] *x*:*y*, where *x* = number of carbons (except carbons of glycerol) and *y* = number of double bonds;

[d] *M*: molar mass;

[e] *w*: mass fraction.

4. Conclusion

Phase equilibrium experimental data of systems containing green coffee oil, supercritical carbon dioxide and ethanol were reported in this work. The effect of temperature and pressure on equilibrium data was similar in systems with $scCO_2$ or $scCO_2$ + ethanol: at higher pressures greater solubility values are obtained. GCO solubility in $scCO_2$ presented values similar to those available in literature for other vegetable oils. By using 2.9% of ethanol, GCO solubility increased 63% over the solubility in pure $scCO_2$ and the crossover pressure point increased about 20 bar.

Samples of the extract phase were more enriched in free fatty acids, kahweol and cafestol than the green coffee oil obtained by pressing. In contrast to the behavior observed for FFA extraction, the samples most concentrated in diterpenes were obtained at lower temperatures. On the other hand, the effect of ethanol on the solubility of diterpenes was greater at higher temperatures.

The solvents tested in this work proved to be effective for obtaining green coffee oil with its bioactive compounds, therefore being an alternative to extraction by pressing. In addition, the use of supercritical technology improves the extraction of free fatty acids and diterpenes, especially when adding ethanol as a cosolvent to the extraction system.

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CAPÍTULO 7. EXTRACTION AND FRACTIONATION OF OIL FROM SPENT COFFEE GROUNDS USING GREEN SOLVENTS AT HIGH PRESSURES

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Extraction and fractionation of oil from spent coffee grounds using green solvents at high pressures

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Abstract

The coffee industry generates large amounts of organic waste, such as the residue obtained during instant coffee production, called spent coffee grounds (SCG). In this work, different methods for extraction and fractionation of oil from SCG, focused on phenolic compounds content (TPC), were investigated at 333 K and 40 MPa, in one or two steps, with different solvents: pure supercritical CO_2 ($scCO_2$), ethanol and $scCO_2$/ethanol mixture (90:10 w/w). The fractionation was also evaluated in four separators in series at 323 K and different pressures and concentrations of ethanol. Ethanolic extract (E) presented the highest overall yield (25 wt%). Samples from the first separator in the fractionation of E showed TPC up to four times higher than the original ethanolic extract. Similar enrichment in TPC was observed by two-steps extraction. The extraction methods proposed in this work are efficient to obtain different extracts with potential of applications in pharmaceutical and food industries.

Keywords: coffee oil; supercritical carbon dioxide; ethanol; cosolvent; antisolvent; phenolic compounds.

1. Introduction

Due to its stimulating effect, coffee is one of the most popular beverages of the world and the second largest traded commodity after petroleum (Murthy and Madhava Naidu, 2012). As a consequence, the coffee industries are responsible for the production of large amounts of organic waste, such as the material called spent coffee grounds (SCG), which consists of the residue obtained during the treatment of roasted coffee beans with hot water or steam for the production of instant coffee, also called soluble coffee (Ballesteros et al., 2014).

SCG is a lignocellulosic material rich in sugar and dietary fiber, with approximately 15% of lipid material (Mussatto et al., 2011), and antioxidant potential associated to the presence of phenolic acids (Ballesteros et al., 2014). The total phenolic compounds content in SCG extracts, expressed in caffeic acid equivalent, reached 587.7 mg/g extract (Andrade et al., 2012). However, the values reported in the literature depend on the coffee species and the extraction solvents and methods. Although the roasting process influences the content of phenolic compounds in coffee beans, the oil content ranged from 7 to 17% (Mussatto et al., 2011) and 11 to 17% (Esquivel and Jiménez, 2012) for green and roasted beans, respectively.

A similar value is reported for oil content in SGC (Kondamudi et al., 2008), since the lipid fraction is not efficiently extracted with water, during the production of soluble coffee.

In addition to being incorporated in animal feed and used as fertilizer in plantations, SCG can be applied for the production of fuels in industrial boilers (Silva et al., 1998), as a substrate for the cultivation of microorganisms (Machado et al., 2012), as source of phenolic compounds (Panusa et al., 2013; Zuorro and Lavecchia, 2012), and as a raw material for the production of fuel alcohol (Mussatto et al., 2012), active carbon (Kante et al., 2012) and biodiesel (Caetano et al., 2014; Kondamudi et al., 2008; Vardon et al., 2013).

In this context, researches to find alternatives for reuse or recovery of this type of residue are important. Furthermore, there is a trend for works following the principles of green chemistry. Green technology applied to food and herbal plants extraction must develop and encourage the utilization of procedures that reduce and/or eliminate the use or production of hazardous substances. Thus, it is necessary to reduce the use of non-eco-friendly organic solvents, and to foster the use of novel environmentally friendly extraction techniques (Mustafa and Turner, 2011), mainly, applying the solvents that follow this philosophy, called as “green solvents”.

Among the techniques that obtain high-purity natural extracts with high extraction yields are the processes carried out under high pressure, such as pressurized liquid extraction (PLE) and supercritical fluid extraction (SFE). The Generally Recognized as Safe (GRAS) solvents, such as CO_2, ethanol and water are the most used by these techniques.

Extractions using supercritical carbon dioxide ($scCO_2$) provide solvent-free extracts and CO_2 is cheap, inert, non-toxic and suitable for extraction of thermosensitive compounds. By applying SFE, mainly with $scCO_2$, the use of cosolvent may be important to improve the range of extracted compounds, since they can modify the solvation properties of the supercritical fluids even at low concentrations. Following the green philosophy, the use of ethanol and/or water as cosolvent is recommended. In addition, combinations of SFE and PLE in two or more extraction steps, in which the polarity of the solvent increases in each step, represent an alternative to maximize extraction of the bioactive compounds, as verified in prior works (Bitencourt et al., 2014; Garmus et al., 2014; Paula et al., 2014).

Another separation process, called as Supercritical Antisolvent Fractionation (SAF), applies the supercritical fluid as an antisolvent. This technique bases on the continuous contact between the $scCO_2$ and a polar liquid mixture in a pressurized precipitation vessel

(Reverchon and De Marco, 2006). The polar mixture is formed by a solvent and the compounds that have been extracted from the natural matrix. In the vessel the supercritical fluid dissolves the polar solvent and the compounds soluble in the mixture $scCO_2$ + solvent, selectively precipitating heavier and more polar compounds (Gonzalez-Coloma et al., 2012). This process can also occur using precipitators or separators in series, in which the soluble portion in the first separator is send to the next one, where different conditions of temperature and pressure can promote a new precipitation. SAF has already been reported in prior works to improve the obtainment of phenolic compounds and anthocyanins from plant extracts (Catchpole et al., 2004; Paula et al., 2018; Sánchez-Camargo et al., 2016; Visentín et al., 2011).

Some previously studies using green technologies at high pressures for obtaining extracts from SCG were found in the literature, as summarized in **Table 1**. Thus, aiming to fill some gaps this work evaluated different methods for obtainment of extracts from SCG in single or two steps using $scCO_2$ and ethanol as solvents, and the fractionation of SCG extracts using $scCO_2$ as antisolvent through separations in series.

2. Material and methods

2.1. Raw material and characterization

Industrial waste from the soluble coffee production was kindly donated by *Cocam Cia. de Café Solúvel e Derivados* (Catanduva, Brazil). This material (SCG), formed by a blend of *Coffea arabica* L. and *Coffea canephora*, was dried in an air circulation oven (model MA 035 Marconi, Brazil) at 333 K for approximately 24 h, packed in plastic bags and stored at 247 K. The final moisture + volatile content was (2.0 ± 0.2)%, measured by gravimetric method AOAC 930.04 (1997).

The size distribution of the particles was determined through a set of sieves ranged from 8 to 100 Mesh (Model 1868, Bertel, Brazil). The mean particle diameter (0.87 mm) was calculated by method ASAE (1997) after measurement of the mass of the retained fractions in each sieve (Semi-analytical balance model AS5500C, Marte, Brazil). The following size distribution was obtained: 18.1% (2.66 mm), 22.9% (1.82 mm), 25.1% (1.00 mm), 8.1% (0.60 mm), 8.0% (0.39 mm), 7.5% (0.21 mm), and 10.5% (0.12 mm).

Table 1

Prior studies using green solvents at high pressure for obtainment of extracts from SCG.

T (K)	P (MPa)	Extraction method	Solvent	Compound of interest	Reference
313-473	3.4-17.2	PLE	ethanol/water	phenolic compounds	Shang et al. (2017)
313-328	19	SFE	CO_2	oil	de Melo et al. (2014)
313-343	14-19	SFE	CO_2 + ethanol	oil and diterpernes	Barbosa et al. (2014)
306-340	11.5-28.4	SFE		oil, sterols and tocopherols	Akgun et al. (2014)
313-353	9.8-37.9	SFE	CO_2	oil and diterpenes	Acevedo et al. (2013)
313-342	20-30	SFE	CO_2 + ethanol	oil	Ahangari and Sargolzaei (2013)
313-333	10-30	SFE	CO_2 + ethanol	phenolic compounds	Andrade et al. (2012)
313-333	15-30	SFE	CO_2 + ethanol	oil	Couto et al. (2009)

PLE: pressurized liquid extraction; SFE: supercritical fluid extraction.

Real density of SCG particles was evaluated by helium gas pycnometry (Quantachrome pycnometer Auto Ultrapyc 1200e) and the apparent density was calculated from the SCG mass required to fill a known volume (Uquiche et al., 2004). The results were (1.23 ± 0.01) g/cm³ and (0.30 ± 0.01) g/cm³, respectively. Finally, the bed porosity (0.75 ± 0.01) was calculated from apparent and real densities (Rahman et al., 1996).

2.2.Chemicals

Carbon dioxide 99.5% w/w (White Martins Gases Industriais, Brazil), ethanol 99.5% (Dinâmica, Brazil) and hexane ≥ 98.5% (Synth, Brazil) were used as solvents in the extraction and fractionation processes.

Other chemicals needed in this work are listed as follows: acetonitrile (TEDIA, USA), acetic acid anhydrous (Merck, Germany), ammonium chloride (Êxodo Científica, Brazil), Folin-Ciocalteu reagent (Dinâmica, Brazil), Milli-Q water (Milli-Q® Direct–Q 3 UV), methanol (TEDIA, USA), sodium carbonate (Dinâmica, Brazil), sodium chloride (Dinâmica, Brazil), sodium hydroxide (Merck, Germany) and sulfuric acid (Êxodo Científica, Brazil).

Commercial green coffee oil obtained by cold pressing was donated by Cooxupé (Brazil). Gallic, protocatechuic, caffeic, p-cumaric, ferulic, sinapic, and trans-cinamic acids with purity ≥ 98% were purchased from Sigma-Aldrich (USA).

2.3.Extraction procedures

2.3.1. Low pressure extractions

In order to obtain comparable samples from SCG and to verify the influence of pressure and the solvent type, low pressure extractions were also evaluated. 5 g of dry SCG were packed in a filter paper cartridge into a Soxhlet extractor. About 150 mL of hexane or ethanol was heated until the boiling temperature and maintained under reflux for 6 hours.

2.3.2. High pressure extractions

The experimental procedures at high pressures were performed in an experimental unit as shown in **Fig. 1.** The unit consists in one extractor (Ex), four separators (S1 to S4) whose pressures were controlled by backpressure valves, a CO_2 pumping system containing a diaphragm metering pump P1 (Model Milroyal MD93(112)S(F) 4M500/Z, Milton Roy Europe, France) and a pneumatic pump P2 (Maq'agua, Brazil), a pumping system for liquid solvents at atmospheric conditions P3 (Model Optos 2SM, Eldex Laboratories Inc., USA), a coriolis mass flow meter (MF) (Model RHMO15NT+CMM01, Metroval, Brazil) and a thermostatic bath (TB) (Model polystat 2C15, Cole Parmer Instrument Company, USA) to control CO_2 temperature for the pumping. The temperature of the extractors and separators are controlled by an electric heating system using electric resistances.

Aiming to maximize the extraction of coffee oil and phenolic compounds, all extractions were performed at 40 MPa and 333 K, in one or two stages. Single-step extractions were evaluated using three different solvents: CO_2, ethanol and CO_2 + ethanol (90.6:9.4 w/w). Sequential extraction was performed as follows: first, the coffee oil was extracted using CO_2 as solvent, and then, ethanol was used to obtain an extract from the residue of the first step (step 2). **Fig. 2** summarizes the extraction methods and solvents evaluated in this work.

Basically the extractor Ex was filled with approximately 25 g of dry SCG and pressurized until 40 MPa with the extraction solvent (Pump P1 was used to pump CO_2 and P3 for ethanol). After pressurization, the solvent was drained through the extractor and the extract was collected before the separator 1 (S1) at set time intervals, aiming to evaluate the kinetic of the extraction. When the CO_2/ethanol mixture was used as solvent, CO_2 and ethanol were pumped separately, but mixed before the extractor.

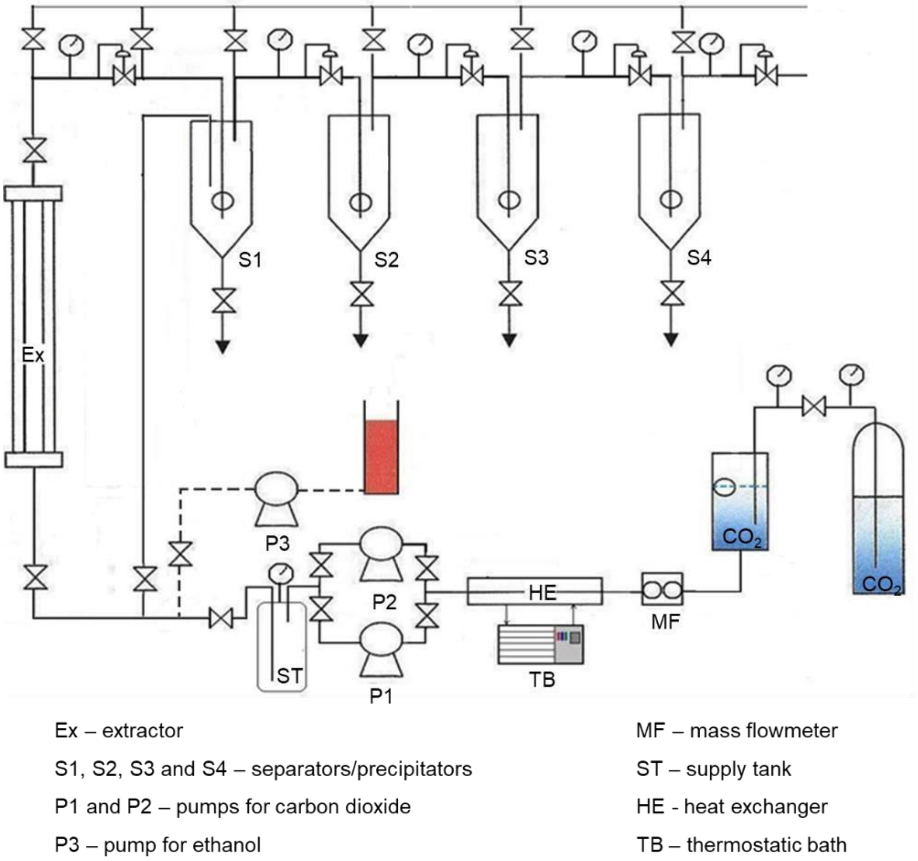

Fig. 1 – Schematic diagram of the experimental extraction and fractionation unit.

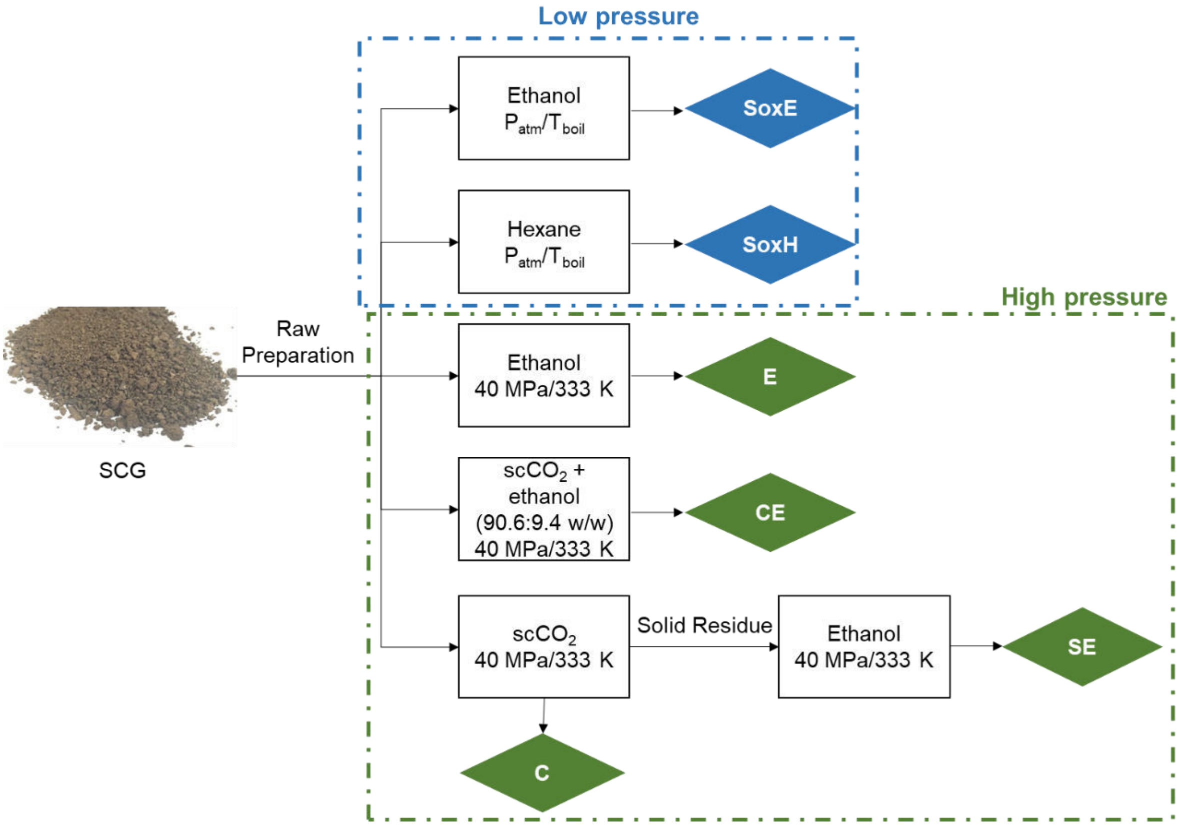

Fig. 2 – Flowchart of the different solvents and methods evaluated for obtaining extracts from SCG.

The mass flow rates of the solvents were 0.9 kg/h and 0.47 kg/h for extractions with pure CO_2 and pure ethanol, respectively. For SFE with CO_2 + ethanol, the mass flow rates were about 0.9 kg CO_2/h and 0.1 kg ethanol/h.

2.4.Fractionation procedures

The fractionations were performed in the same experimental unit (**Fig. 1**) from two different extracts: the extract obtained with ethanol under high pressure (E) and the extract obtained by SFE with the mixture CO_2 + ethanol as solvent (CE).

The experimental procedure was similar to the extraction methodology, mainly for packaging and initial pressurizing of the extractor. However, after the extraction, the extract was directly sent to the separators S1, S2, S3 and S4 in sequence. The separators were previously pressurized at different pressures, so that S1 was at the higher pressure and S4 was at the lower one. In each separator, the precipitation of the compounds occurred according to their solubility under the conditions of temperature and pressure, whereby four fractions were collected at the end of the process. Apart from four precipitated fractions, one additional sample, named “cleaning” (CL), was obtained, which corresponds to the sample generated by washing the pipe between the separators with ethanol. Dashed lines in **Fig. 3** represent the pipe excerpt washed.

For fractionation of CE, about 0.9 kg/h of CO_2 and 0.1 kg/h of ethanol were pumped directly into the extractor, while the separators were previously pressurized. For fractionation of E the mass flow rate of CO_2 for the separators was about 2 kg/h and ethanol drained by the extractor in different mass flow rates: 0.22 kg/h or 0.11 kg/h.

Five different fractionation conditions were evaluated, as shown in **Fig. 3**. The first three conditions correspond to the fractionation of the ethanolic extract (E). Conditions 1 and 3 were carried out at same pressures of 30 MPa/20 MPa/10 MPa /P_{atm} (S1/S2/S3/S4) and differed by the concentration of ethanol in the CO_2/ethanol mixture within the separators: 9.6% for condition 1 and 5.0% for condition 3. Condition 2 was also performed with about 10% ethanol in separators, but under the follow pressures: 20 MPa/10 MPa/8 MPa/P_{atm}. In addition, the two conditions for fractionation of the extract CE (Conditions 4 and 5) differ only in the concentration of ethanol in the separator, setting the pressures at 30 MPa/20 MPa/10 MPa /$P_{atm.}$

As previously mentioned, during the fractionation experiments, the extract was sent to the separators, which were already pressurized. Thus, prior pressurizing of the separators was required, and the solvent for pressurizing the separators varied according to the condition evaluated. For the conditions 1, 2, 3 and 5 the pressurization was performed with pure CO_2, and the same proportion of the CO_2/ethanol mixture of the extraction was used in pressurization for condition 4.

All extracts and fractions were concentrated in a rotary vacuum evaporator (Rotary evaporator Marconi, MA-120, Brazil), and dried in an oven (Marconi, MA030/12, Brazil) under vacuum (Precision Vacuum Pump model D25 with Marathon Electric motor 223711) at 323 K until constant weight (Analytical balance Precisa XT 220A). The samples were stored at 247 K for further analysis. First, the extraction and fractionation methods were evaluated based on the overall extraction yields, which was calculated by the ratio of the total mass of extract and the initial mass of the raw material.

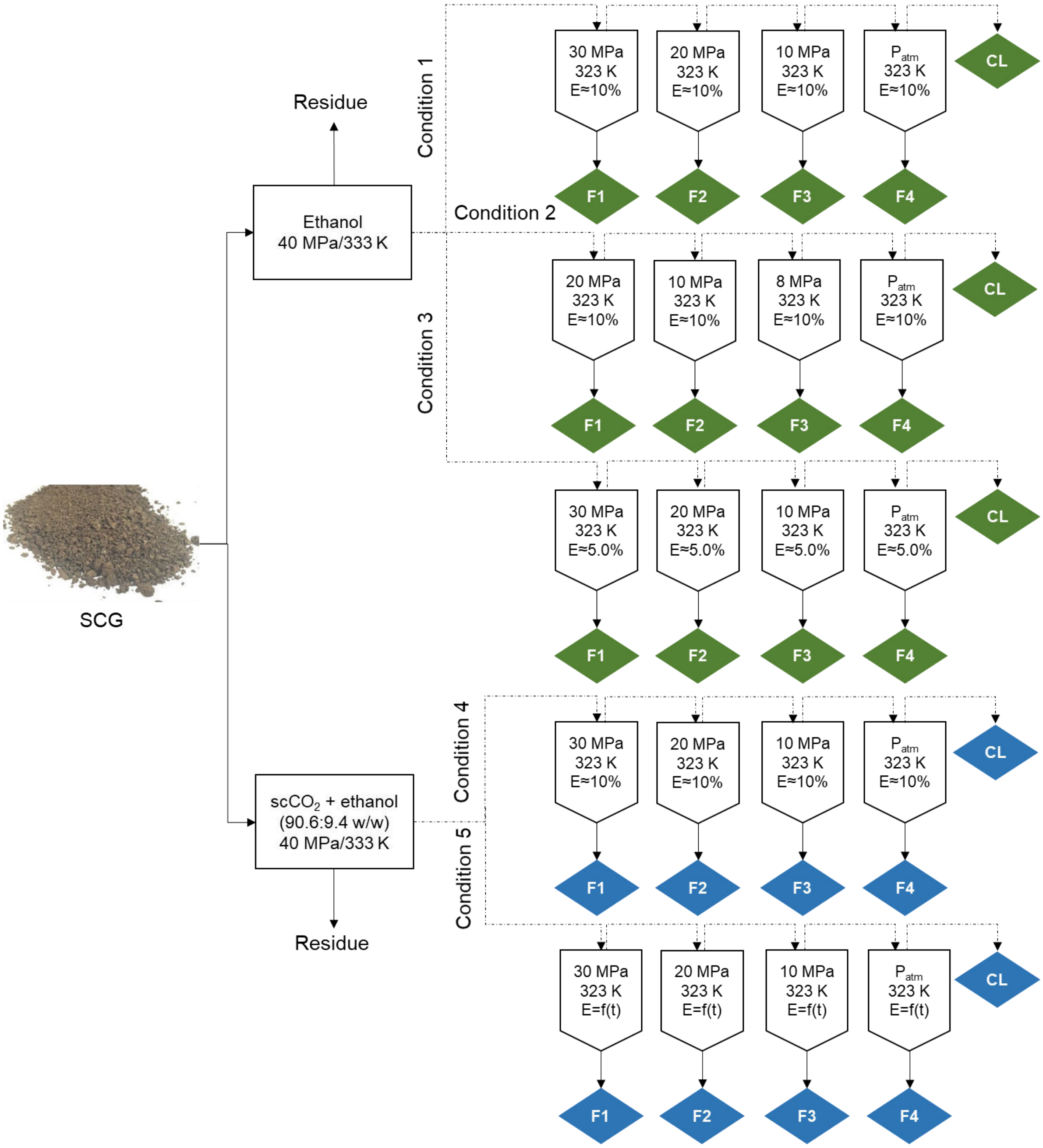

Fig. 3 – Flowchart of the different fractionation conditions evaluated in this work.

2.5. Analytical methods

2.5.1. Total phenolic content (TPC)

Quantitative measurements of total phenolic compounds were performed by spectrophotometry using Folin-Ciocalteu reagent (Singleton and Rossi, 1965) by an adapted methodology from Fuentes et al. (2012). Up to 0.5 g of sample was weighted into a 15 mL falcon tube in which 1 mL of hexane and 1 mL of methanol/water solution (60:40 v/v) were added. Then, the tube was shaken (Vortex-Genie 2 model, Scientific Industries, USA) for 2 minutes and centrifuged (Centrifuge model BR4i, Jouan, France) at 3500 rpm, 10 min and 278 K.

From 50 to 400 μL of the methanolic (lower) phase was used, depending on the content of total phenolic compounds in the sample. The volume of the solution was completed until 2 mL with deionized water and 0.2 mL of Folin-Ciocalteu reagent was added. After 3 minutes, 0.4 mL of sodium carbonate solution 35% and 1.4 mL of deionized water were added. Finally, after 2 hours, absorbance at 725 nm was measured (UV-VIS Spectrophotometer Aquamate Orion 8000, Thermo Scientific, USA).

A standard curve was plotted using 2 mL of known concentration solutions of gallic acid, between 2.5 and 15 mg/L, reacted with the Folin-Ciocalteu reagent following the same methodology described above.

2.5.2. Phenolic acids composition

The composition of the samples in phenolic acids was performed by ultra-high performance liquid chromatography (UHPLC). 5 μL of the methanolic phase (section 2.5.1) were directly injected into the SQD/UPLC (Waters Acquity, USA) at the elution and Photodiode Array Detector conditions as reported by Schwarz et al. (2009). However, since it generally provides lower detection limits, the detection of the components was also evaluated by a single-quadrupole mass spectrometry with electrospray ionization interface as described by Gruz et al. (2008).

2.5.3. Fatty acids composition

The determination of fatty acid composition was performed in duplicate by gas chromatography of the fatty acid methyl esters according to the official procedure Ce 1f-96 of the American Oil Chemists' Society (AOCS, 2009). Thus, firstly, the oil was derivatized to form fatty acid methyl esters according to Hartman and Lago (1973).

2.6.Statistical analysis

All procedures for obtaining extracts or fractions as well the quantitative analysis were performed in triplicate, and the results were submitted to analysis of variance to compare means with one-way ANOVA and post hoc Tukey’s test ($p < 0.05$), using the statistical software Minitab version 16 (Minitab, State College, PA, USA).

3. Results and discussion

3.1.Kinetic study

For a better understanding of the extraction process it is important to evaluate the extraction kinetics through curves obtained by plotting the extraction yield versus S/F, which corresponds to the ratio of the solvent mass and the raw material mass. **Fig. 4** shows that all the extraction methods were performed until depleting the vegetable source (SCG), making it possible to optimize the spent mass of solvent and, consequently, the extraction time.

In general, all the curves presented a typical behavior. According to Sovová (1994), a typical curve for supercritical extraction can be described by three steps: (1) Period of constant rate of extraction, (2) Step with decreasing extraction rate, and (3) Diffusion-controlled stage. In the first step, the extraction of the most easily accessible solute occurs and, generally, corresponds to about 50 to 90% of the total extractable mass. For SFE with pure CO_2, after approximately 20 minutes of extraction (S/F ≈ 15) more than 80% of total extraction yield has already been obtained, and the diffusion-controlled stage was achieved at S/F = 25 (corresponding to 600 g of CO_2 spent).

The single-step extractions employing ethanol (PLE: ethanol and SFE: CO_2 + ethanol) presented the highest extraction rates. In these cases, the diffusion-controlled stage was achieved when approximately 300 g of solvent had been spent (S/F ≈ 12). In addition, at S/F = 6.5 (t = 21 minutes), PLE extraction reached 89% of the total extraction yield. For SFE with CO_2 + ethanol, only 12 minutes (S/F ≈ 8.6) were necessary to obtain the same percentage.

The kinetic study can be also necessary to determinate the extraction time for posterior fractionation procedures, since the fractions precipitated were only collected in the end of the fractionation, i.e. it was not possible to visualize the fractions during the process. For fractionation procedures, S/F for extraction stage with ethanol and CO_2 + ethanol was settled in S/F = 12 (300 g of ethanol and t = 80 min) and S/F = 33 (825 g of solvent and t ≈ 45min), respectively. Although **Fig. 4** shows these S/F values for extraction is probably not economically viable, this condition was set aiming to compare, through global mass balance, the extraction yield and total phenolic content (TPC) for extraction and fractionation experiments.

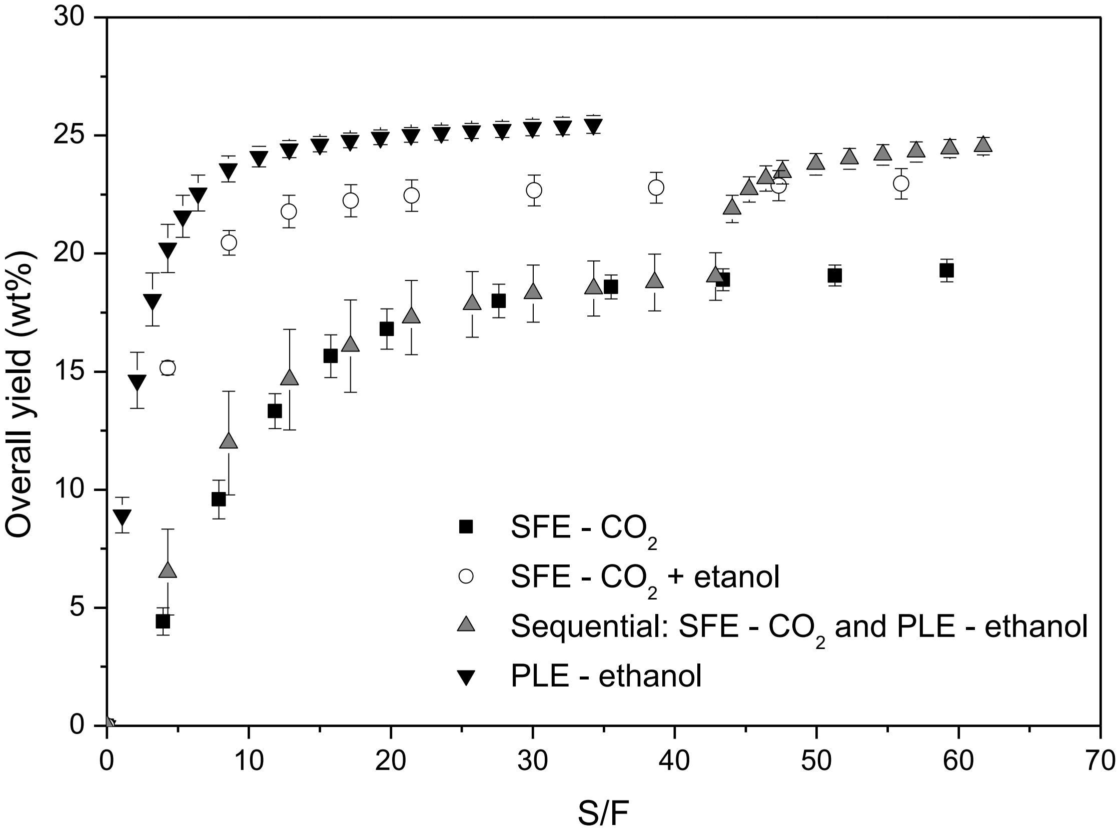

Fig. 4 – Kinetic study of coffee oil extraction from SCG at 333 K and 40 MPa.

3.2.Single-Step extractions

Besides the extractions by Soxhlet with ethanol and hexane, which are important for comparison with the extractions using green solvents at high pressures, the single-step extractions were performed with three different solvents: pure CO_2, pure ethanol and a CO_2/ethanol mixture (90.6:9.4 w/w). **Table 2** summarizes the extraction yields and TPC, so that one can compare the performance of the different methods and solvents investigated.

Extractions using ethanol as solvent presented the highest overall extraction yields, both for extractions under high (PLE) or low (Soxhlet) pressure. Although coffee beans were previously submitted to extraction with water, this result confirms that an important content of compounds of higher polarity is still found in SCG.

Comparing the use of ethanol by Soxhlet with PLE, there are no significant differences between overall extraction yields; both processes have obtained about 25 g of extract/100 g SCG. Recently, data on extraction from SCG using PLE with ethanol and water mixtures were reported, however no extraction yield was informed (Shang et al., 2017). Andrade et al. (2012) reported extraction yields using Soxhlet extractor of 12% and 15% with hexane and ethanol as solvents, respectively.

Table 2

Yields of extraction and total phenolic content of coffee oil from SCG [a].

	Sample	Method	Solvent	Pressure (MPa)	Temperature (K)	Y_1	TPC	Y_2
Single-Step	SoxE	Soxhlet	Ethanol	P_{atm}	T_{boil}	26.0 ± 0.5^A	4.7 ± 0.3^B	1.23 ± 0.07^A
	SoxH		Hexane	P_{atm}	T_{boil}	20.0 ± 0.5^C	0.06 ± 0.01^C	0.012 ± 0.003^C
	C	SFE	CO_2	40	333	19.3 ± 0.6^C	0.20 ± 0.03^C	0.04 ± 0.01^C
	CE		CO_2 + ethanol (90.6:9.4 w/w)	40	333	23.0 ± 0.8^B	0.8 ± 0.1^{BC}	0.19 ± 0.02^C
	E	PLE	Ethanol	40	333	25.5 ± 0.5^A	4.9 ± 0.2^B	1.25 ± 0.05^A
Two-Steps	SC	SFE	CO_2	40	333	19 ± 1^C	0.24 ± 0.04^C	0.04 ± 0.01^C
	SE	PLE	Ethanol	40	333	5 ± 1^D	17 ± 4^A	0.90 ± 0.06^B
		Total		-	-	24.6 ± 0.5^{AB}		

[a] Means that do not share a capital letter are significantly different at $p \leq 0.05$; P_{atm}: atmospheric pressure in Campinas, Brazil; T_{boil}: boiling temperature of the solvent at P_{atm}; TPC: total phenolic content (mg GAE/g extract); Y_1: overall yield of extraction (g extract/100 g SCG); Y_2: total phenolic compounds yield (mg GAE/g SCG); GAE: gallic acid equivalent; SFE: supercritical fluid extraction; PLE: pressurized liquid extraction.

In the same way, solvents with similar polarity can be compared, such as CO_2 and hexane. It was also not verified a significant difference between the results, so that extracts obtained by using SoxH and C did not present statistical difference for overall extraction yields at $p \leq 0.05$ (≈ 20g extract/100g SCG). This value is higher than those reported by Ahangari and Sargolaei (2013) and Cruz et al. (2014), who informed yields of 16.7% and 14%, respectively.

In the scientific literature there are some studies reporting the extraction of coffee oil from SCG using $scCO_2$ at different temperatures and pressures, but there is no information at 40 MPa. The extractions performed by Andrade et al. (2012) at 10, 20 and 30 MPa presented yields of 0.43, 9.1 and 9.8%, respectively. Cruz et al. (2014) extracted 11.9 g of oil per 100 grams of SCG at 323 K and 25 MPa. The extractions carried out by Ribeiro et al. (2013) at 323 K and 25 MPa produced an average yield of 12.1% and a viscous oil solution. The SFE with pure $scCO_2$ reported in this work presented overall yield of 19%. Although pressure is a relevant parameter to increase the coffee oil extraction, the differences in overall yield can also be caused by climatic factors, coffee blend and raw material pretreatment.

The overall yield for SFE with CO_2 + ethanol (CE) followed the sequence SoxE = E > CE > SoxH = C. Couto et al. (2009) obtained an extraction yield of 19.4% at 323 K and 20 MPa using the mixture CO_2/ethanol (93.5:6.5 w/w) against 13.1% for pure $scCO_2$. In general, the use of ethanol as cosolvent promotes the increase of oil extraction yields, as confirmed by the green coffee oil solubility reported by Bitencourt et al. (2017). This increase can be attributed to the higher solvent density and changes in chemical and physical intermolecular forces in the system.

Regarding the total phenolic content (TPC) in the extracts, **Table 2** shows similar conclusions to the previous ones when analyzing the overall extraction yield, as follows: SoxE = E > C = SoxH. On the other hand, extract CE presented TPC statistically equal ($p \leq 0.05$) to all other extracts obtained in a single step. The higher TPC was obtained with ethanol as solvent, which was about 4.9 mg GAE/g extract. TPC of the extracts obtained in this work were lower than most of those shown in literature. Prior works reported values up to 500 mg/g extract by ultrasound extraction (Andrade et al., 2012). However, these values are greatly dependent on the coffee blends, and also on the solvent and method of extraction.

For extracts obtained by Soxhlet system with ethanol as solvent, the total phenolic content was 151 mg/g, expressed in caffeic acid equivalent (CAE) (Andrade et al., 2012). The maximum TPC obtained by PLE with water/ethanol mixture (75:25) was 12.4 mg/g at 433 K and 17.2 MPa (Shang et al., 2017). TPC reported for supercritical extracts obtained with pure

$scCO_2$ (C) varied from 24 to 57 mg CAE/g extract (Andrade et al., 2012). In accordance with the results reported in the present work, Andrade et al. (2012) did not observe any statistical difference between TPC in the extracts obtained by supercritical extraction using pure CO_2 or CO_2 + ethanol.

The results presented above suggest that extraction methods containing ethanol as solvent or cosolvent guarantee a better performance, since they provide the highest overall extraction yields and TPC. However, the appearance of the extract obtained with $scCO_2$ (C) is quite different from the one obtained with ethanol (E): while the first extract is a clear oil, the second one is a dark oily extract (**Fig. 5a**). Thus, it seems that the best solution would be to obtain a sequence of fractions with different industrial applications along a unique process.

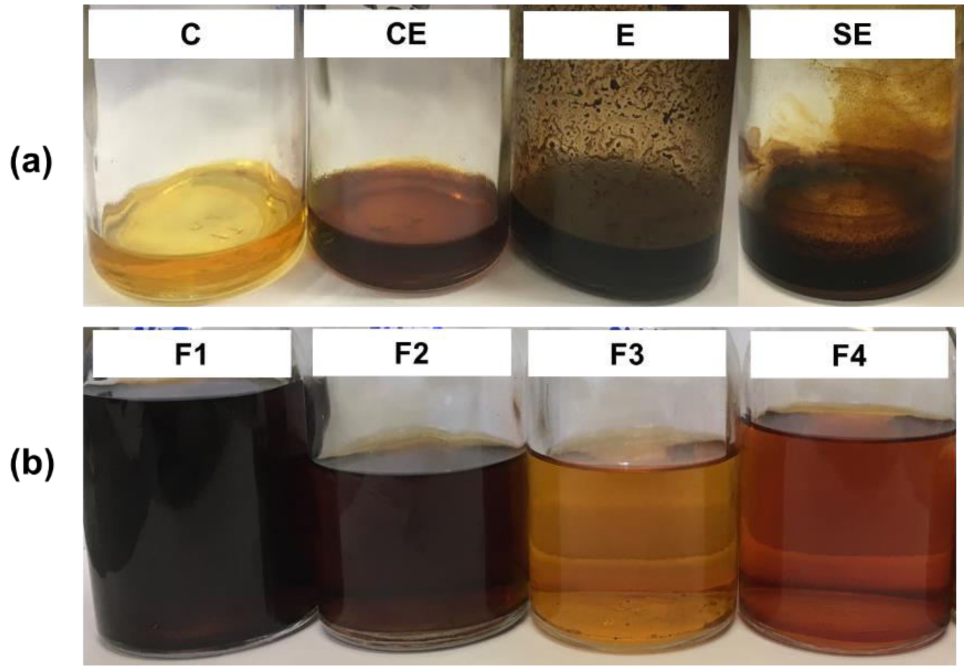

Fig. 5 – Pictures of: (a) extracts from SCG obtained using different solvents: single-step extractions with CO_2 (C), CO_2 + ethanol (CE), ethanol (E); two-step extraction with ethanol as second stage (SE), and (b) the four fractions obtained by fractionation of E (condition 2).

3.3.Two-Steps extractions

Integration of processes is a recent trend towards obtaining natural extracts containing bioactive compounds. These processes maximize the extraction yields and obtain extracts with different compositions and applications by using extraction methods in two or more steps, in which the extraction solvent polarity increases throughout each stage (Viganó et al., 2016; Moraes et al., 2015). Thus, the process becomes more economically interesting, achieving high extraction yields for the desired components and taking full advantage of the raw material (Viganó et al., 2017).

In this work, the sequential extractions in two steps were performed aiming at to fractionate the compounds obtained when ethanol is used as solvent in a single-step extraction. In this way, the low polar solvent CO_2 was used in the first stage to extract the lipid portion from SCG.

Kinect extraction curves for the two-step extraction are also shown in **Fig. 4**, where the jump of the points represents the beginning of the extraction with ethanol. Besides, approximately 19% of the extraction yield for SFE (first step), the second step of extraction presented overall yield of (5 ± 1) g extract/g SCG, totalizing 24.6 ± 0.5 g extract/g SCG (**Table 2**). This total is similar ($p \leq 0.05$) to the extraction yield by PLE in a single-step. However, this two-step method allowed obtaining two extracts with different industrial applications.

The main advantage of this kind of extraction is confirmed by TPC of the extract obtained during the second stage of extraction (SE), which presented about 17 mg GAE/g extract. This value is more than three times higher than TPC for extract E (4.9 ± 0.2 mg/g). Garmus et al. (2014) and Martinez-Correa et al. (2017a, 2017b) have also obtained enriched fractions in phenolic compounds through integrated process combining SFE followed by extraction with ethanol as solvent.

Visually, the performance of the process to separate the oily fraction from the rich fraction in phenolic compounds is shown in **Fig. 5a**, in which the sample obtained in the second extraction step (SE) is seen as a dry extract adhered to the walls of the vial, i.e. the extract SE is solid at environmental conditions, unlike the others.

3.4.Fractionations

By varying pressure, temperature and concentration of ethanol in the four separators in series, several experiments were performed to fractionate the oil extracted from the SCG, aiming to determine the best operating conditions to obtain fractions enriched in phenolic compounds.

Tables 4 and 5 present the yields and contents of total phenolic compounds in each separator for the fractionation of the ethanolic extract obtained by PLE (E) and the extract obtained using $scCO_2$ + ethanol as solvent (CE), respectively. All the results were analyzed through calculations of overall mass balance and component mass balance, comparing the sum of the yields for the four fractions with the overall extraction yield for the equivalent single-stage extraction. The deviation values observed were lower than 11% and 35%, for global and component balances, respectively.

Table 4

Fractionation of ethanolic extract (E) from SCG: yield and total phenolic content [a].

	Sample	T (K)	P (MPa)	E (%)	Y_1	TPC	Y_2
	E	333	40	100	25.5 ± 0.5^{A}	4.9 ± 0.2^{D}	1.25 ± 0.05^{A}
Condition 1	F1	323	30	9.6 ± 0.1	0.6 ± 0.2^{F}	20 ± 5^{AB}	0.12 ± 0.07^{E}
	F2	323	20	9.6 ± 0.1	0.52 ± 0.06^{F}	11 ± 3^{BCD}	0.055 ± 0.006^{E}
	F3	323	10	9.6 ± 0.1	3.3 ± 0.4^{D}	2.3 ± 0.3^{D}	0.075 ± 0.002^{E}
	F4	323	P_{atm}	9.6 ± 0.1	19.3 ± 0.3^{BC}	3.2 ± 0.2^{D}	0.62 ± 0.04^{CD}
	CL	-	-		0.4 ± 0.3^{F}	19 ± 4^{AB}	0.07 ± 0.05^{E}
	Total				24.1 ± 0.4^{A}	3.9 ± 0.4^{D}	0.9 ± 0.1^{B}
Condition 2	F1	323	20	9.8 ± 0.1	0.8 ± 0.1^{EF}	26 ± 10^{A}	0.21 ± 0.09^{E}
	F2	323	10	9.8 ± 0.1	1.2 ± 0.5^{EF}	5 ± 2^{CD}	0.05 ± 0.01^{E}
	F3	323	8	9.8 ± 0.1	2.2 ± 0.3^{DE}	2.5 ± 0.5^{D}	0.06 ± 0.02^{E}
	F4	323	P_{atm}	9.8 ± 0.1	19.9 ± 0.9^{B}	3.3 ± 0.5^{D}	0.66 ± 0.08^{C}
	CL	-	-		0.20 ± 0.03^{F}	27 ± 5^{A}	0.05 ± 0.02^{E}
	Total				24.2 ± 0.3^{A}	4.3 ± 0.2^{D}	1.03 ± 0.04^{B}
Condition 3	F1	323	30	5.0 ± 0.1	1.11 ± 0.03^{EF}	19 ± 12^{ABC}	0.2 ± 0.1^{E}
	F2	323	20	5.0 ± 0.1	0.7 ± 0.1^{EF}	14 ± 3^{ABCD}	0.10 ± 0.03^{E}
	F3	323	10	5.0 ± 0.1	3.5 ± 0.9^{D}	2.4 ± 0.5^{D}	0.08 ± 0.01^{E}
	F4	323	P_{atm}	5.0 ± 0.1	18 ± 1^{C}	2.6 ± 0.5^{D}	0.47 ± 0.06^{D}
	CL	-	-		0.5 ± 0.2^{F}	7 ± 3^{BCD}	0.042 ± 0.009^{E}
	Total				24 ± 1^{A}	3.7 ± 0.1^{D}	0.89 ± 0.05^{B}

[a] Means that do not share a capital letter are significantly different at $p \leq 0.05$; P_{atm}: atmospheric pressure in Campinas, Brazil; TPC: total phenolic content (mg GAE/g extract); Y_1: overall yield of extraction (g extract/100 g SCG); Y_2: total phenolic compounds yield (mg GAE/g SCG); GAE: gallic acid equivalent; E: mass percentage of ethanol.

Almost the totality of the extract has precipitated in separators 3 and 4, named F3 and F4, respectively (**Table 4**). This behavior was partially expected, since extracts from SCG are composed mainly by a lipid portion and, in general, oils present high solubility in pure $scCO_2$ at any temperature and pressure. Consequently, during the fractionation process, these compounds continue dissolved in the mixtures of CO_2 and ethanol created inside the separator, being dragged to the next one.

Although the highest overall yield was obtained in F3, this fraction presented one of the lowest TPC. For all conditions evaluated, the fractions precipitated in separator 1 (F1) presented the highest TPC, followed by F2 > F4 > F3. Total phenolic content in F1 reached values up to 6 times higher than TPC in the extract E. The difference of TPCs for the fractions can be visualized in **Fig. 5b**, in which it was verified a much darker coloration associated to the presence of phenolic compounds in fractions 1 and 2.

The standard deviation in the experimental data did not allow concluding the best condition for the fractionation of E, since the contents and the extraction yields of phenolic compounds for the same fractions in different conditions of fractionation were not different ($p \leq 0.05$). However, condition 3 showed a slightly different behavior. Although condition 3 stands out for the lowest TPC in the fraction collected after the cleaning of the pipe (CL), it has the disadvantage because TPC for fractions from separator 1 and 2 did not presented statistically significant differences.

Similar to the results of the fractionation of E indicated, the proportion of CO_2/ethanol into the separator is relevant. For this reason, the fractionation of the CE extract was evaluated at two conditions, which only differed in the concentration of ethanol in CO_2/ethanol mixture inside the separator (**Table 5**). As previous mentioned, experimentally this difference was obtained by pressurizing the separators early in the process with a mixture of CO_2/ethanol similar to the one used for the extraction step (Condition 4) or with pure $scCO_2$ (Condition 5). Condition 5 was designed in order to promote higher precipitation of the phenolic compounds in first separator, since an abrupt drop in the concentration of ethanol within the separators is obtained. In this way, for the Condition 5 the concentration of ethanol inside the separator decreases from S1 to S4 and it is close to zero at $t = 0$ min, increasing over time. Thus, mass percentage of ethanol inside the separator (E) must be represented as a function of the time (**Table 5**)

Regarding the overall yield, while condition 4 presented the largest precipitated fraction in separator 4 (F4), F3 is the major fraction obtained by condition 5. This result is probably due the lower concentration of ethanol in the separators in condition 5 compared with condition 4, favoring early precipitation. In a similar way to the previous results (**Table 4**), F1 presents the lowest overall yield and the highest content of phenolic compounds. In addition, F1 enriched about four times compared to TPC of the extract CE.

As expected, due to their low solubility in $scCO_2$, the phenolic compounds precipitated in the first separators. However, this precipitation does not occur entirely in the first separator, since phenolic compounds were still collected in the last separator and during the cleaning of the pipe (fraction CL).

Table 5

Fractionation of the extract obtained with $scCO_2$ + ethanol (CE) from SCG: yield and total phenolic content [a].

	Sample	T (K)	P (MPa)	E (%)	Y_1	TPC	Y_2
	CE	333	40	9.4	23.0 ± 0.8^{A}	0.8 ± 0.1^{BC}	0.19 ± 0.02^{A}
Condition 4	F1	323	30	11 ± 1	0.61 ± 0.08^{F}	3.2 ± 0.6^{A}	0.020 ± 0.004^{E}
	F2	323	20	11 ± 1	1.8 ± 0.2^{F}	1.6 ± 0.3^{B}	0.030 ± 0.006^{DE}
	F3	323	10	11 ± 1	8.6 ± 0.9^{D}	0.5 ± 0.1^{C}	0.04 ± 0.01^{CDE}
	F4	323	P_{atm}	11 ± 1	11 ± 2^{C}	1.0 ± 0.2^{BC}	0.11 ± 0.03^{B}
	CL				0.5 ± 0.1^{F}	1.7 ± 0.5^{B}	0.008 ± 0.003^{E}
	Total				22.7 ± 0.8^{A}	0.9 ± 0.2^{BC}	0.21 ± 0.04^{A}
Condition 5	F1	323	30	f(t)	0.7 ± 0.2^{F}	3.4 ± 0.9^{A}	0.023 ± 0.002^{DE}
	F2	323	20	f(t)	2.6 ± 0.6^{F}	1.3 ± 0.3^{BC}	0.033 ± 0.002^{DE}
	F3	323	10	f(t)	13.6 ± 0.9^{B}	0.58 ± 0.07^{BC}	0.079 ± 0.004^{BCD}
	F4	323	P_{atm}	f(t)	6.0 ± 0.6^{E}	1.6 ± 0.2^{BC}	0.09 ± 0.02^{BC}
	CL				0.26 ± 0.02^{F}	1.2 ± 0.2^{BC}	0.0029 ± 0.0005^{E}
	Total				23.2 ± 0.3^{A}	1.0 ± 0.1^{BC}	0.23 ± 0.03^{A}

[a] Means that do not share a capital letter are significantly different at $p \leq 0.05$; P_{atm}: atmospheric pressure in Campinas, Brazil; TPC: total phenolic content (mg GAE/g extract); Y_1: overall yield of extraction (g extract/100 g SCG); Y_2: total phenolic compounds yield (mg GAE/g SCG); GAE: gallic acid equivalent; E: mass percentage of ethanol.

Fig. 6 summarized the results reported in **Tables 4 and 5**, allowing the comparison between the fractionation of E (**Fig. 6a**) and CE (**Fig 6b**). While the major part of the extract was collected at S4 for fractionation of E, **Fig.6b** shows a better mass distribution of the precipitated samples in each separator during the fractionation of CE. In addition, by comparing **Fig. 6a** with **6c** and **Fig. 6b** with **6d** it can be concluded that, as a consequence of TPC, the distribution of the overall yield (Y_1) is different from the distribution of the total phenolic yield (Y_2) in separators. For any condition of fractionation the higher Y_2 was verified for F4.

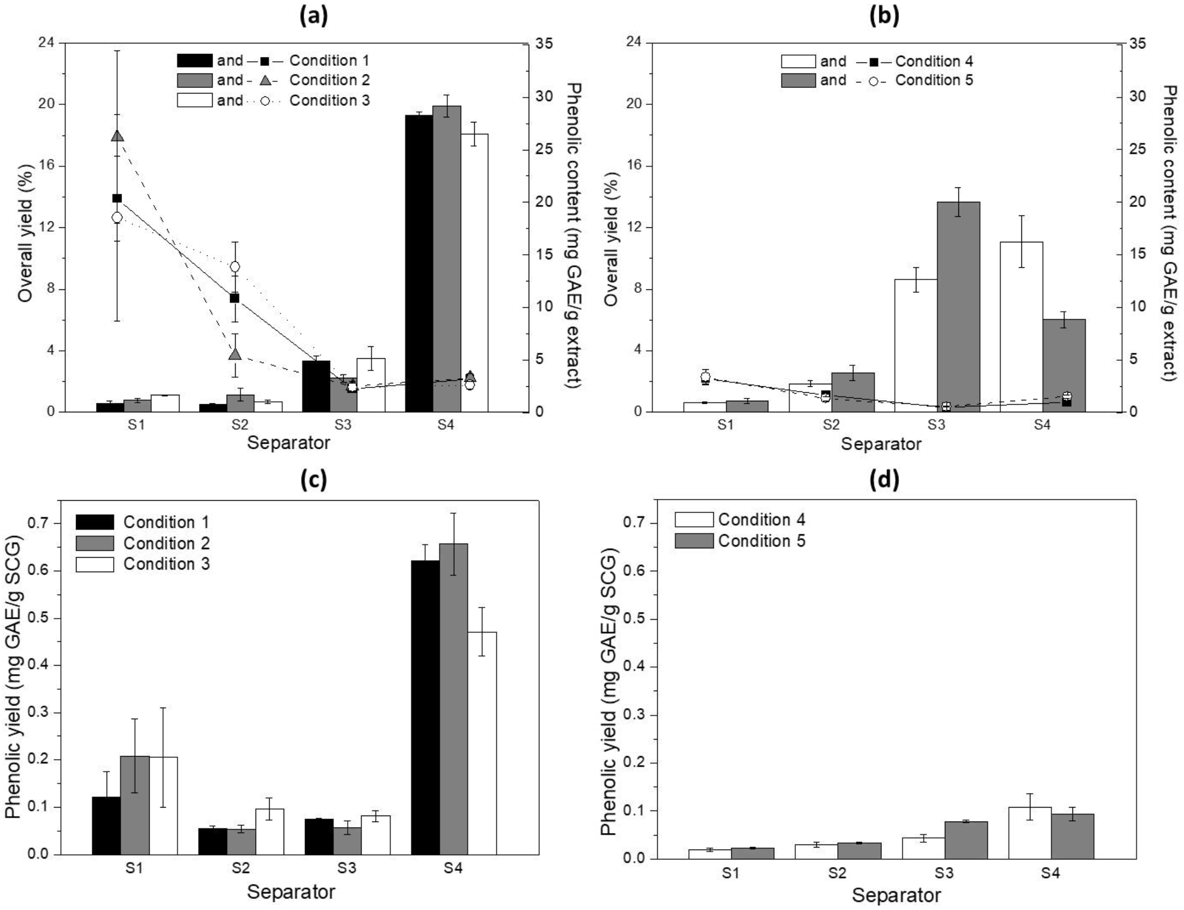

Fig. 6 – Overall extraction yields and TPC for fractionation of (a) E and (b) CE, and phenolic extraction yields for fractionation of (c) E and (d) CE. The lines represent total phenolic contents (TPC) and the columns report the yields.

Finally, the fractionation procedure generated fractions as enriched in phenolic compounds as the samples from two-step extraction, as discussed in section 3.3. However, the total phenolic yield for the fraction with the highest TPC (F1- condition 2) was 0.21 ± 0.09, against 0.9 ± 0.06 for the extract SE. Thus, the two-step extraction method has the advantage of providing a sample with higher extraction yield for phenolic compounds, while the fractionation using the series of separators is faster, since the extraction and fractionation processes occur simultaneously.

3.5.Phenolic acids composition

Samples were analyzed for the composition in phenolic acids by ultra-high performance liquid chromatography (UHPLC) using two detection systems: Photodiode Array Detector (PDA) and (b) Mass Spectrometry (MS). The sequence of elution for the phenolic acids is visualized by the chromatograms in **Fig. 7**. Calibration curves for all phenolic acids in both detection methods presented $R^2 \geq 0.99$.

Due to the scarcity of sample mass, the fractions collected during the fractionation procedures were not evaluated. **Table 5** shows that only ferulic, protocatechuic and caffeic acids could be measured. Contrary to expectations, the major phenolic acid in our samples was protocatechuic acid. Besides chlorogenic acids, Andrade et al. (2012) detected the presence of epicatechin and tannic, p-hydroxybenzoic, caffeic and gallic acids in samples from SCG.

An important difference between the concentrations was verified by comparing the results using PDA and MS. For all samples, quantifications performed by PDA obtained the highest values. These results are consistent, since co-eluted compounds that absorb light at a given wavelength are jointly quantified by PDA. On the other hand, MS detector only quantified the phenolic acids in their free form. This difference is relevant for coffee samples with high content of chlorogenic acids, since they represent the family of esters formed between certain trans-cinnamic acids, mainly caffeic and ferulic acid, with quinic acid (Clifford, 1999).

Table 5

Content of phenolic acids (mg/kg extract) in extracts from SCG measured by UHPLC.

Sample	Ferulic acid		Protocatechuic acid		Caffeic acid	
	PDA	MS	PDA	MS	PDA	MS
GCO	2.34 ± 0.1	2.17 ± 0.1	-	1.07 ± 0.1	-	-
SoxE	11.67 ± 0.1	6.95 ± 0.1	43.00 ± 0.1	24.48 ± 0.1	11.68 ± 0.1	1.31 ± 0.1
C	3.76 ± 0.1	0.29 ± 0.1	-	0.70 ± 0.1	-	-
CE	4.69 ± 0.1	3.53 ± 0.1	13.01 ± 0.1	6.04 ± 0.1	9.01 ± 0.1	-
E	22.83 ± 0.1	4.37 ± 0.1	39.42 ± 0.1	28.35 ± 0.1	22.83 ± 0.1	3.93 ± 0.1
SE	18.14 ± 0.1	17.14± 0.1	111.19 ± 0.1	92.55 ± 0.1	9.55 ± 0.1	9.46 ± 0.1

PDA: Photodiode Array Detector; MS: Mass Spectrometry.

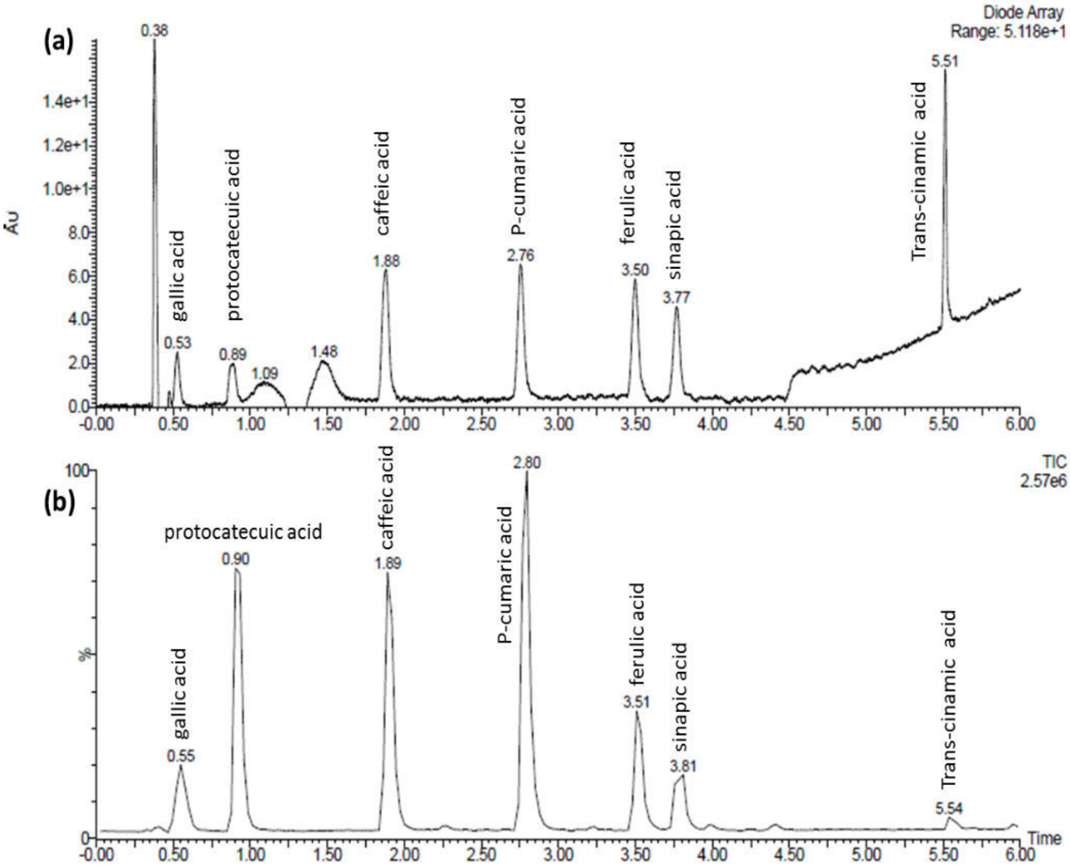

Fig. 7 – UHPLC chromatogram of phenolic acids obtained by detector (a) Photodiode Array Detector - PDA and (b) Mass Spectrometry - MS.

3.6.Fatty acids composition

The composition of the extracts and fractions was also evaluated based on the fatty acids distribution. For the samples from fractionation procedures, only fractions with visual appearance similar to oil were measured, which were mainly the fractions collected in separator 3 and 4 (**Table 6)**. In general, all the extracts obtained from single-step extractions have presented about 41% and 59% for saturated and unsaturated fatty acids, respectively, while the commercial green coffee oil obtained by cold pressing (GCO) presented 56% of unsaturated fatty acids. However, for all samples, the three major fatty acids were linoleic, palmitic and oleic acids, summing more than 85% of the total.

Although there was no relevant difference for the fatty acid composition of the oil fractions, all conditions for fractionation indicated the increase of unsaturated fatty acids content in F3. The first oily fractions obtained during the fractionation of E presented composition of 38% and 62% for saturated and unsaturated fatty acids, respectively. In addition, the largest difference between the compositions of the fractions was approximately 4%, obtained for the fractionation of CE (Condition 5).

Table 6

Fatty acid composition of extracts and fractions from SCG in mass percentage [a].

Cx:y[b]	GCO[n]	Extracts					Fractions from E							Fractions from CE			
							Condition 1		Condition 2			Condition 3		Condition 4		Condition 5	
		C	CE	E	SoE	SoH	F3	F4	F2	F3	F4	F3	F4	F3	F4	F3	F4
C16:0[c]	33.8	30.6	30.6	30.5	30.7	30.7	27.1 ± 0.1	31.1 ± 0.1	27.0 ± 0.1	29.3 ± 0.1	30.9 ± 0.1	26.3 ± 0.2	30.8 ± 0.6	28.7 ± 0.0	32.0 ± 0.0	29.0 ± 0.1	33.8 ± 0.3
C18:0[d]	7.9	8.5	8.6	8.5	8.5	8.6	9.1 ± 0.0	8.4 ± 0.0	9.1 ± 0.1	8.4 ± 0.0	8.5 ± 0.0	9.1 ± 0.1	8.2 ± 0.2	8.6 ± 0.0	8.4 ± 0.1	8.5 ± 0.0	8.4 ± 0.1
C18:1[e]	9.3	10.5	10.6	10.4	10.5	10.5	10.6 ± 0,0	10.4 ± 0.1	11.0 ± 0.3	10.5 ± 0.0	10.5 ± 0.1	11.1 ± 0.1	11.4 ± 0.9	10.6 ± 0.0	10.4 ± 0.0	10.6 ± 0.2	11 ± 1
C18:2[f]	43.9	44.2	44.1	44.1	44.1	43.7	45.9 ± 0.2	44.0 ± 0.2	45.4 ± 0.0	45.6 ± 0.2	43.9 ± 0.2	46.0 ± 0.1	43.4 ± 0.4	45.5 ± 0.1	43.2 ± 0.2	45.6 ± 0.2	40.5 ± 0.6
C20:0[g]	1.4	0.9	0.9	1.0	0.9	0.9	1.1 ± 0.0	1.0 ± 0.0	1.05 ± 0.0	1.0 ± 0.0	0.9 ± 0.0	1.0 ± 0.0	1.3 ± 0.3	1.0 ± 0.0	0.9 ± 0.0	1.0 ± 0.0	0.8 ± 0.0
C18:3[h]	2.8	3.6	3.7	3.8	3.7	3.8	4.2 ± 0.0	3.6 ± 0.0	4.4 ± 0.0	3.6 ± 0.0	3.6 ± 0.0	4.3 ± 0.2	3.4 ± 0.1	3.8 ± 0.0	3.5 ± 0.1	3.7 ± 0.1	3.6 ± 0.1
C20:1[i]	0.3	0.7	0.6	0.7	0.6	0.7	0.8 ± 0.0	0.6 ± 0.0	0.8 ± 0.0	0.6 ± 0.0	0.7 ± 0.0	0.8 ± 0.0	0.6 ± 0.0	0.7 ± 0.0	0.6 ± 0.0	0.7 ± 0.0	0.6 ± 0.0
C22:0[j]	0.7	1.0	1.1	1.1	1.1	1.1	1.3 ± 0.0	1.0 ± 0.0	1.4 ± 0.0	1.0 ± 0.0	1.0 ± 0.0	1.4 ± 0.0	1.0 ± 0.0	1.2 ± 0.0	1.0 ± 0.0	1.1 ± 0.0	1.0 ± 0.0
Sat[k]	43.7	41.0	41.1	41.1	41.2	41.3	38.5 ± 0.1	41.4 ± 0.1	38.4 ± 0.2	39.7 ± 0.1	41.3 ± 0.1	37.8 ± 0.0	41.2 ± 0.5	39.5 ± 0.1	42.3 ± 0.1	39.5 ± 0.1	44.0 ± 0.5
Unsat[l]	56.3	59.0	58.9	58.9	58.8	58.7	61.5 ± 0.1	59.6 ± 0.1	61.6 ± 0.2	60.3 ± 0.1	58.7 ± 0.1	62.2 ± 0.0	58.8 ± 0.5	60.5 ± 0.1	57.8 ± 0.1	60.5 ± 0.1	56.0 ± 0.5
R[m]	0.78	0.70	0.70	0.70	0.70	0.70	0.65 ± 0.0	0.73 ± 0.0	0.65 ± 0.0	0.78 ± 0.0	0.63 ± 0.0	0.71 ± 0.0	0.62 ± 0.0	0.66 ± 0.0	0.70 ± 0.0	0.61 ± 0.0	0.70 ± 0.0

[a] Standard deviation < 0.05 were represented as 0.0;

[b] Cx:y where x = number of carbons and y = number of double bonds;

[c] Palmitic acid; [d] stearic acid; [e] oleic acid; [f] linoleic acid; [g] arachidic acid; [h] linolenic acid; [i] gadoleic acid; [j] behenic acid;

[k] Total percentage of saturated fatty acids;

[l] Total percentage of unsaturated fatty acids;

[m] R=Ratio between the percentages of saturated and unsaturated fatty acids;

[n] Green coffee oil obtained by cold pressing.

The fractionation methodology using $scCO_2$ as an antisolvent proposed in this work is a promising and efficient one, since it promotes fractionation of phenolic compounds and fatty acids. Besides the enriched fraction in phenolic compounds from the first separator, samples obtained from S3 and S4 are oily fractions with slightly different composition in fatty acids. However, the process should be further investigated in order to optimize the operation conditions.

4. Conclusions

This work evaluated different methods for obtaining coffee oil from spent coffee grounds (SCG) under high pressure and using only the "green" solvents CO_2 and ethanol. The use of high pressure, approximately 40 MPa, was important to increase the overall extraction yields. Extractions performed with ethanol as a solvent presented the highest overall extractions yield (25%) and TPC (4.56 mg GAE/g extract).

Due to the low solubility of phenolic compounds in $scCO_2$, these compounds precipitated in the first separators. However, adjustments are required in conditions of fractionation since phenolic compounds were still collected in the last separator S4 and during the cleaning of the pipe between the separators.

Fractions enriched in total phenolic compounds (TPCs approximately four times higher than the extract obtained by single-step extraction) were obtained by two methods: 1) two-step extraction with SFE with pure $scCO_2$ followed by PLE with pure ethanol and 2) fractionation of SCG extract using $scCO_2$ as antisolvent through four separators in series.

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CAPÍTULO 8. DISCUSSÃO GERAL

Este capítulo tem como objetivo apresentar, de forma geral, uma discussão dos resultados reportados nos Capítulos 3 a 7, visando sintetizar as informações obtidas neste trabalho. O resultado primordial e comum a todos os artigos que compõem esta tese refere-se ao aumento da solubilidade de compostos fenólicos e lipídicos e, consequentemente, dos seus rendimentos de extração quando se utilizam cossolventes alcoólicos juntamente com $scCO_2$.

Inicialmente, neste trabalho fez-se o estudo da solubilidade dos ácidos ferúlico e cafeico e do óleo de café em misturas de CO_2 + etanol em diferentes condições de temperatura, pressão e concentração de etanol. Ao utilizar aproximadamente 10 mol% de etanol adicionados ao $scCO_2$, verificou-se aumento de 120 e 30000 vezes na solubilidade dos ácidos ferúlico e cafeico, respectivamente. Para ambos os ácidos fenólicos, não é recomendável a adição de mais do que 10 mol% de etanol ao CO_2, uma vez que os resultados indicaram que, em concentrações maiores de etanol, o aumento na solubilidade pode não justificar o aumento no custo do processo, principalmente relacionado à remoção do cossolvente (líquido) da amostra final.

Apesar de a solubilidade do ácido cafeico em $scCO_2$ + etanol apresentar valores menores que a solubilidade do ácido ferúlico nas mesmas condições de pressão, temperatura e concentração de etanol, este ácido fenólico apresentou maior aumento da sua solubilidade com a adição de etanol, quando comparada à sua solubilidade em $scCO_2$ puro. O aumento na solubilidade com a adição de cossolventes é consequência do aumento da densidade da mistura solvente, das interações dipolo-dipolo e principalmente das fortes interações químicas, como ligações de hidrogênio, que podem ocorrer entre as moléculas do soluto e do cossolvente (CATCHPOLE et al., 2005; HUANG et al., 2005; SAUCEAU et al., 2004; TING et al., 1993). Dessa forma, o comportamento observado para o ácido cafeico provavelmente está associado a sua maior polaridade e ao maior número de sítios associativos presentes em sua molécula.

Para o óleo de café, o uso do etanol também promoveu o aumento da solubilidade em CO_2 supercrítico, porém de forma menos significativa: valores 2,4 vezes maiores foram obtidos ao adicionar 5,7% (em massa) de etanol ao $scCO_2$. Além disso, na extração deste óleo com $scCO_2$, verificou-se a extração preferencial de alguns compostos presentes no óleo. Dessa forma, as amostras obtidas mostraram-se mais enriquecidas em diterpenos e ácidos graxos livres que o óleo de café verde comercial, produzido por prensagem. Todas essas informações indicam a complexidade do processo de extração em questão e justificam a dificuldade de modelagem termodinâmica desse sistema. Assim, os dados foram correlacionados apenas pela equação semi-empírica proposta por Chrastil, e apresentaram dados de solubilidade bastante similares a diversos outros óleos vegetais (DEL VALLE; AGUILERA, 1988).

No geral, os maiores valores de solubilidade do ácido cafeico, ferúlico e óleo de café verde em misturas de CO_2 e etanol foram obtidos em maiores pressões, assim como verificado sem a presença do cossolvente. Por outro lado, também é comum para estes sistemas a redução do efeito cossolvente (calculado pela razão entre a solubilidade na mistura $scCO_2$/cossolvente pela solubilidade em $scCO_2$, nas mesmas condições de temperatura e pressão) em maiores condições de pressão. Este comportamento ocorre devido à redução da diferença entre composição local do cossolvente ao redor das moléculas do soluto e a sua composição *bulk*[1]. Ou seja, a composição local se aproxima da concentração *bulk* e o efeito das interações cossolvente-soluto são menos pronunciadas em pressões mais altas (HUANG et al., 2005).

Para os dados de solubilidade do óleo de café em $scCO_2$ e $scCO_2$ + etanol foi possível detectar um ponto de cruzamento entre as isotermas da solubilidade, indicando que, nesta condição de pressão em que o cruzamento ocorre, há inversão no comportamento da solubilidade com a temperatura. Por outro lado, a solubilidade dos ácidos cafeico e ferúlico em $scCO_2$ na presença de etanol não apresentaram clara tendência com variações na temperatura.

Para posterior obtenção de todos os parâmetros de interação binária envolvidos nos sistemas ternários ou quaternários avaliados, dados de equilíbrio dos sistemas ácido ferúlico-etanol, ácido ferúlico-água e ácido cafeico-etanol foram medidos. Verificou-se que a solubilidade do ácido cafeico em etanol é ligeiramente menor que a solubilidade do ácido ferúlico, além de sofrer menor influência com a variação da temperatura (ZHOU et al., 2007). Para ambos os ácidos, a solubilidade em etanol é maior que a solubilidade em água, indicando que a utilização do etanol como cossolvente é mais recomendada.

Dados de equilíbrio de fases do sistema óleo de café verde-etanol também foram medidos (Anexo A). Foram obtidos dados de composição das fases alcoólicas e oleosas, considerando o óleo como um pseudo-componente. As composições dessas fases quanto ao teor de diterpenos (caveol e cafestol) não puderam ser quantificadas adequadamente e, por esse motivo, os resultados não foram incluídos no Capítulo 6.

Quanto à modelagem termodinâmica dos dados de equilíbrio obtidos nesta tese, verificou-se que a equação de estado *Cubic Plus Association* estimou adequadamente, a partir dos parâmetros de interação binária, os dados experimentais de solubilidade dos ácidos ferúlico e cafeico em $scCO_2$ na presença de cossolventes. A capacidade preditiva das equações de estado cúbicas (Peng-Robinson e Soave-Redlich-Kwong) na modelagem termodinâmica do equilíbrio

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[1] Concentração *Bulk*: refere-se à concentração do cossolvente em pontos distantes da superfície do soluto, uma definição adaptada a partir da sugestão proposta pela IUPAC (2014).

foi muito inferior que a das equações que aplicam a teoria de perturbação de Wertheim e consideram possíveis associações entre os componentes do sistema (autoassociação ou associação cruzada). Dessa forma, a equação CPA assume as interações químicas que ocorrem entre as moléculas, além das forças de atração e repulsão que as equações de estado cúbicas levam em conta.

A solubilidade dos ácidos ferúlico e cafeico nos sistemas ternário e quaternário foram fortemente subestimadas pela modelagem termodinâmica que emprega as equações de Peng-Robinson e Soave-Redlich-Kwong a partir dos parâmetros de interação binária. Comparando os resultados apresentados nos Capítulos 3 e 4, nota-se que a predição dos dados de solubilidade do ácido cafeico foi ainda pior que a obtida para o ácido ferúlico: para ácido cafeico a predição mostrou-se mais de 100 vezes menor que os dados experimentais, enquanto para o ácido ferúlico valores até 14 vezes menores foram calculados. Essas diferenças provavelmente não estão associadas às diferenças no modelo termodinâmico, e sim ao fato de o ácido cafeico poder se associar com o etanol por meio de mais sítios associativos presentes em sua estrutura, fazendo com que mais interações ocorram e não estejam sendo contabilizadas pela equação de estado cúbica. Dessa forma, também vale mencionar que, quando essas interações químicas entre os componentes do sistema não ocorrem, a predição utilizando as equações de estado cúbicas pode fornecer bons resultados.

A partir desses resultados, sistemas similares e reportados na literatura foram selecionados para confirmar a boa capacidade preditiva da equação CPA para dados de solubilidade de solutos em misturas de $scCO_2$ e cossolventes. Dentre os solutos e cossolventes investigou-se tanto moléculas contendo sítio ativos de associação, quanto moléculas não associativas. A equação conseguiu reproduzir adequadamente o comportamento dos dados experimentais quando os efeitos da temperatura, pressão, tipo e concentração de cossolventes foram avaliados. Porém, um fator limitante para a utilização deste modelo termodinâmico pode ser a necessidade do conhecimento de um grande conjunto de dados iniciais, incluindo tanto propriedades dos componentes puros, quanto informações sobre o equilíbrio de fases dos sistemas binários relevantes.

O Capítulo 7 apresenta exemplos de processos de extração e fracionamento a partir da borra de café, para os quais o conhecimento dos dados de solubilidade de alguns ácidos fenólicos presentes no café e do óleo de café verde pode ser importante para orientar, inicialmente, a seleção das melhores condições de temperatura, pressão e concentração de etanol para o processo. Como esperado, as extrações com etanol fornecerem os maiores rendimentos globais e maiores teores de compostos fenólicos. Porém, o uso de etanol como solvente ou cossolvente também se destacou por apresentar as maiores taxas de extração, e consequentemente, por atingir o período da extração controlado pela difusão mais rapidamente.

Pela solubilidade do óleo de café mostrar comportamentos diferentes com a variação da temperatura dependendo da condição de pressão, a temperatura intermediária de 323 K foi fixada em todos os quatro separadores. Além disso, como desejava-se primeiramente maximizar a extração dos compostos fenólicos e, posteriormente, promover sua precipitação nos primeiros separadores, o processo inicial de extração foi realizado apenas com solventes contendo etanol, seja puro ou em mistura com $scCO_2$. E, para a máxima precipitação dos compostos fenólicos no primeiro separador era necessária uma redução significativa na solubilidade desses compostos, feita pela formação de diferentes misturas de CO_2/etanol dentro dos separadores. Por limitações na bomba de alta pressão, as concentrações de etanol (em base livre de soluto) criadas dentro dos separadores foram em torno de 5% ou 10% (em mol). Dessa forma, resumidamente, a precipitação dos compostos foi avaliada apenas em função da pressão e da concentração de etanol dentro dos separadores, que mostraram ser os principais parâmetros que influenciam no valor da solubilidade de compostos fenólicos e lipídicos presentes no café em $scCO_2$ na presença de cossolventes, como apresentado nos Capítulos 3, 4 e 6.

Apesar do teor de compostos fenólicos para as amostras deste trabalho ser menor que os da maioria dos extratos de borra de café já reportados em trabalhos científicos anteriores, pôde-se afirmar que o processo de fracionamento de extratos usando o $scCO_2$ como antissolvente forneceu frações aproximadamente quatro vezes mais concentradas em compostos fenólicos, comparadas ao extrato obtido em etapa única de extração. Porém, as demais frações coletadas ainda indicavam a presença de significativa quantidade desses compostos. O uso de concentrações de etanol menores que 5% dentro dos separadores pode fornecer melhores resultados, pois irá promover uma queda mais acentuada na solubilidade dos ácidos fenólicos.

O processo de fracionamento por extração em dois estágios, no qual inicialmente procedeu-se a extração da borra de café com $scCO_2$ puro (SFE) seguida da extração com etanol puro (PLE), forneceu frações enriquecidas em compostos fenólicos de forma similar ao observado no processo de fracionamento que empregou o $scCO_2$ como antissolvente. Porém, as extrações sequenciais apresentaram amostras com maior rendimento de extração para compostos fenólicos e o fracionamento por separadores em série tem a vantagem de ser um processo mais rápido, uma vez que a extração e o fracionamento ocorrem simultaneamente.

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CAPÍTULO 9.

CONCLUSÕES

9.1.Considerações Finais

- O uso de etanol como cossolvente é de fundamental importância para aumentar a solubilidade de compostos fenólicos em $scCO_2$. Ao se utilizar aproximadamente 10 mol% de etanol, verificou-se um aumento de até 120 e 30000 vezes na solubilidade dos ácidos ferúlico e cafeico, respectivamente.
- Para ambos os ácidos fenólicos estudados, não é recomendada a adição de mais de 10 mol% de etanol como cossolvente ao $scCO_2$.
- Para o óleo de café, o uso do etanol também promove o aumento da solubilidade em $scCO_2$, porém de forma bem menos significativa. Para este sistema, a modelagem termodinâmica é um desafio por se tratar de um sistema multicomponente e por serem obtidas amostras mais concentradas em diterpenos e ácidos graxos livres que o óleo obtido por prensagem.
- No geral, assim como em $scCO_2$ puro, os maiores valores de solubilidade dos ácidos cafeico e ferúlico e do óleo de café em misturas de $scCO_2$ e etanol foram obtidos em maiores pressões. Porém, verificou-se que o efeito cossolvente tende a decrescer com o aumento da pressão.
- Enquanto para os dados do sistema contendo óleo de café foi possível detectar um ponto de cruzamento das isotermas, que representa o valor da pressão em que há a inversão no comportamento da solubilidade com a temperatura, as solubilidades dos ácidos cafeico e ferúlico em $scCO_2$ + etanol não apresentaram uma clara tendência com as variações na temperatura.
- As equações de estado cúbicas de Peng-Robinson e de Soave-Redlich-Kwong não foram capazes de predizer adequadamente os dados de solubilidade do ácido ferúlico e ácido cafeico em misturas de dióxido de carbono supercrítico e cossolventes. Para ambos os compostos, os dados de solubilidade foram subestimados por esses modelos termodinâmicos.
- A equação de estado *Cubic Plus Association* mostrou ser um modelo termodinâmico eficaz para predizer dados de solubilidade de solutos sólidos em $scCO_2$ na presença de cossolventes. Seu uso é mais recomendado quando o soluto e cossolventes são moléculas capazes de se associarem fortemente através de pontes de hidrogênio.
- O conhecimento de dados de equilíbrio de fases ou solubilidade de compostos de interesse em misturas de $scCO_2$ e cossolvente é importante fonte de informação para a otimização e *design* de processos de separação que utilizam a tecnologia supercrítica.

- O resíduo da produção de café solúvel (ou instantâneo) confirmou ser importante fonte de óleo e compostos fenólicos. Além disso, o uso de solventes "verdes" e processos sob altas pressões representam técnicas de extração alternativas que se destacam por produzir extratos naturais livres de solventes tóxicos.
- O fracionamento de extratos usando o $scCO_2$ como antissolvente forneceu frações aproximadamente quatro vezes mais concentradas em compostos fenólicos comparadas ao extrato obtido em etapa única de extração.
- Processos de fracionamento, baseados tanto no uso do dióxido de carbono supercrítico como solvente ou antissolvente ou em processos integrados com a utilização de líquidos pressurizados, são interessantes para recuperar compostos de matrizes vegetais, cuja composição apresenta compostos de diferentes polaridades. No geral, esses processos representam uma alternativa para o aproveitamento integral das matérias-primas.

9.2.Sugestões para Trabalhos Futuros

- Avaliação econômica do processo de fracionamento dos compostos presentes na borra de café utilizando quatro separadores em série e por extrações sequenciais em duas ou mais etapas.
- Análise mais detalhada de composição das frações obtidas para verificar o possível fracionamento e enriquecimento das amostras em diferentes compostos, como em relação aos insaponificáveis totais, diterpenos e tocoferóis.
- Avaliação da atividade antioxidante das amostras obtidas a partir da borra de café.
- Estudo do processo de fracionamento de compostos presentes em outros resíduos da indústria de café, como a torta remanescente da extração do óleo de café verde.

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ANEXOS

A. Dados do equilíbrio de fases do sistema óleo de café verde-etanol

A solubilidade mútua do sistema óleo de café verde - etanol foi avaliada em 4 temperaturas (293, 303, 313 e 323 K) e os resultados de cada componente em cada fase, em fração mássica, são apresentados na **Tabela B.1**. Como esperado, um aumento da solubilidade foi observado com o aumento da temperatura, porém, as duas fases apresentaram dependência distinta com a temperatura. Por exemplo, a fração mássica do óleo na fase rica em etanol variou de 0,13 a 0,29 (aumento de 123%) ao aumentar a temperatura de 293 para 323 K.

Tabela B.1 – Dados de solubilidade mútua do sistema composto por óleo café verde (1) + etanol (2). w_1 e w_2: fração mássica de óleo e etanol, respectivamente.

Temperatura (K)	Composição global		Fase rica em etanol		Fase rica em óleo	
	w_1	w_2	w_1	w_2	w_1	w_2
293	0,369	0,631	0,135	0,865	0,807	0,193
303	0,318	0,682	0,143	0,857	0,781	0,219
313	0,314	0,686	0,180	0,820	0,715	0,285
323	0,503	0,497	0,287	0,713	0,625	0,375

A **Figura B.1** apresenta os dados em função da fração mássica de etanol. Uma inclinação mais acentuada na fase rica em óleo foi observada, indicando que a temperatura tem mais influência na miscibilidade dos dois compostos nesta fase. Este comportamento também foi verificado para outros óleos vegetais como óleo de semente de algodão, girassol, milho, canola, palma, macaúba e farelo de arroz (DA SILVA et al., 2010; FOLLEGATTI-ROMERO et al., 2010).

Em relação aos ácidos graxos, a composição do óleo de café verde apresenta similaridades com o óleo de semente de algodão, pois este também apresenta os seguintes ácidos graxos como majoritários: ácido linoleico (56,4%), palmítico (22,8%) e oleico (16,0%). Comparando os dados de solubilidade mútua obtidos neste trabalho com dados similares para o óleo de semente de algodão, verificou-se que a fração mássica do etanol na fase rica em óleo de café apresentou–se 22% maior que para o óleo de semente de algodão. Já para a fase rica em etanol, o desvio da fração mássica de etanol foi de 8,6% (FOLLEGATTI-ROMERO et al., 2010).

Os trabalhos que apresentam dados de equilíbrio de fases para sistemas contendo óleos vegetais + etanol presentes na literatura visam o conhecimento sobre a solubilidade mútua desses componentes para uma melhor compreensão do processo de produção de biodiesel e melhoria das taxas da reação de transesterificação, que deve ser realizada com excesso de etanol e, portanto depende da solubilidade do etanol na fase oleosa. Além disso, os dados de equilíbrio de fases para sistemas compostos de óleos vegetais + etanol aquoso ou etanol anidro são necessários para a concepção de processos de extração líquido-líquido no refino de óleos vegetais (DA SILVA et al., 2010).

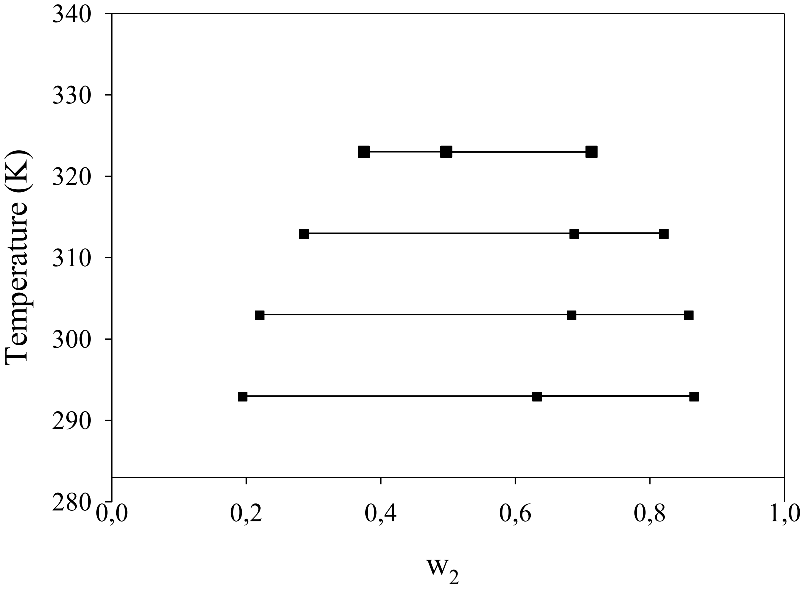

Figura B.1 – Solubilidade mútua do sistema composto por óleo de café verde (1) + etanol (2). w_2: fração mássica de etanol.

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B. Equilíbrio sólido-líquido: Equação aplicada nos Capítulos 4 e 5

Considerando o equilíbrio sólido-líquido entre um soluto, designado como componente i, e um solvente, o equilíbrio de fases pode ser representado por (PRAUSNITZ; LICHTENTHALER; DE AZEVEDO, 1998):

$$f_i(sólido\,puro) = f_i(soluto\,em\,solução\,líquida) \tag{A.1}$$

$$f_i(sólido\,puro) = \gamma_i x_i f_i^0 \tag{A.2}$$

onde x_i é a solubilidade em fração molar do soluto no solvente, γ_i é o coeficiente de atividade do soluto *i* fase líquida e f_i^0 é a fugacidade do componente *i* no estado de referência. Para o coeficiente de atividade da fase líquida, define-se o estado de referência como um líquido puro subresfriado. Desconsiderando a solubilidade do solvente na fase sólida, a equação de equilíbrio pode ser escrita como:

$$x_2 = \frac{f_2(sólido\,puro)}{\gamma_2 f_2(líquido\,subresfriado\,puro)} \tag{A.3}$$

Essas duas fugacidades dependem apenas das propriedades do soluto, ou seja, são independentes da natureza do solvente. Dessa forma, a razão dessas duas fugacidades pode ser calculada pelo ciclo termodinâmico apresentado na **Figura A.1**. A variação na energia livre de Gibbs do componente i ao ir do estado termodinâmico **a** ao **d** está relacionada às fugacidades do sólido e líquido subresfriado por:

$$\Delta \underset{a\to d}{\underline{G}} = RT \ln \frac{f_i(líquido\,subresfriado\,puro)}{f_i(sólido\,puro)} \tag{A.4}$$

sendo:

$$\Delta \underset{a\to d}{\underline{G}} = \Delta \underset{a\to d}{\underline{H}} - T\Delta \underset{a\to d}{\underline{S}} \tag{A.5}$$

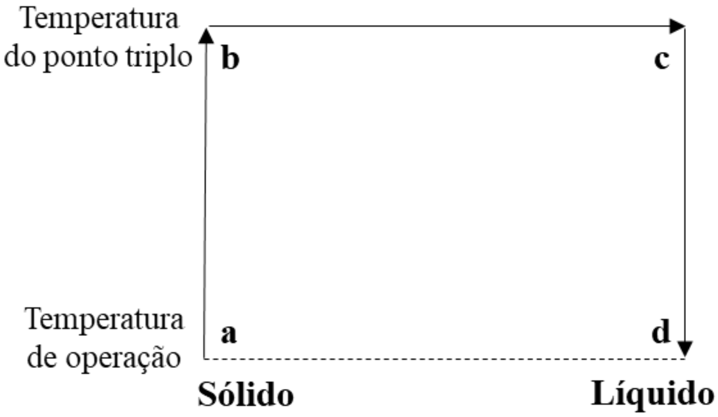

Figura A.1 – Ciclo termodinâmico para o cálculo da fugacidade do líquido puro subresfriado (Adpatado de (PRAUSNITZ; LICHTENTHALER; DE AZEVEDO, 1998).

Sabendo-se que entalpia e entropia são funções de estado, ou seja, independentes do caminho termodinâmico, é possível calculá-las com o caminho **a → b → c → d** (**Figura A.1**). Assim as variações de entalpia e entropia de **a → d** podem ser calculadas pelas Equações A.6 e A.7, respectivamente.

$$\Delta \underset{a\to d}{\underline{H}} = \Delta_{fus}\underline{H} + \int_{T_t}^{T} \Delta c_p dT \tag{A.6}$$

$$\Delta \underset{a\to d}{\underline{S}} = \Delta_{fus}\underline{S} + \int_{T_t}^{T} \frac{\Delta c_p}{T} dT \tag{A.7}$$

No ponto triplo pode-se afirmar que:

$$\Delta_{fus}\underline{S} = \frac{\Delta_{fus}\underline{H}}{T_t} \tag{A.8}$$

Finalmente, substituindo as Equações A.5, A.6, A.7 e A.8 na Equação A.4, e assumindo que Δc_p é constante no intervalo de temperatura de T até T_t , obtém-se:

$$\ln\frac{f_2^L}{f_2^S} = \frac{\Delta_{fus}\underline{H}}{RT_t}\left(\frac{T_t}{T}-1\right) - \frac{\Delta c_p}{R}\left(\frac{T_t}{T}-1\right) + \frac{\Delta c_p}{R}\ln\frac{T_t}{T} \tag{A.9}$$

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