Phase Equilibrium Engineering of Supercritical Extraction, Fractionation and Reaction Processes

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Abstract

The chemical and pharmaceutical industries have a key role in global economic development. However, the environmental impact of this industrial sector could impair its future development. Green Chemistry principles are calling for the use of renewable resources, less waste and environmentally friendly solvents (EFS). Among EFS increasing attention is given to supercritical fluids (SCF). A field that has numerous SCF applications is that related to natural products processing, which is growing driven by the fact that biomass is renewable and nature can produce many complex molecules in a highly efficient way. The design of the phase conditions in supercritical processes is directly tied to the process goal. In the case of fractionation/extraction an heterogeneous phase behavior is required over the whole range of process compositions; the opposite may be required in the case of reaction processes where homogeneous conditions are favored to avoid mass transfer limitation that hinders the reaction rate. In both cases the nature of the mixture to deal with and the process goal are critical in the selection of a SCF solvent and in the selection of the process operating variables (pressure, temperature and composition). In this work Phase Equilibrium Engineering principles are applied to the design of the phase conditions in extraction, fractionation for a wide range of natural products and reacting systems. For this purpose thermodynamic models tuned to the particular systems of interest are used to predict general diagrams of univariant lines for mixtures of key binary components. On this basis a strategy for the design of optimum operating conditions for supercritical processes is developed. This approach is illustrated with a case study of fractionation of monoglycerides obtained as byproducts of supercritical transesterification of vegetable oils.

1. High pressure processing of natural products

During the last decades, a tendency to the design and development of healthier and safer products and sustainable processes has grown in importance, based on "green chemistry" principles such as a more efficient use of energy, the replacement of traditional organic solvents by less contaminant alternatives and the use of renewable raw materials. These changes are promoted by environmental concerns and governmental regulations, as well as a higher commitment from the consumers, who are gradually modifying their habits and preferences towards more "natural" products.

In this context, high pressure and supercritical fluid technologies appear as an attractive alternative to traditional processes. Most commonly used near critical and supercritical fluids (SCF) namely carbon dioxide, ethane, propane are considered "green solvents", because they are gases at ambient conditions and therefore leave no residue in the final products after depressurization. Moreover, because of their relatively low critical temperature, thermal degradation of natural products and the subsequent generation of undesirable compounds are minimized or avoided. In general, SCF are

compatible with food and pharmaceutical products due to their very low toxicity; they are inexpensive and usually non-reactive.

From a physicochemical point of view, the main advantage of supercritical solvents is that their density (and therefore all density-related properties, like solvent power) can be tuned over a broad range, from a liquid-like to a gas-like state, by simple changes in pressure and/or temperature. Also, the phase condition can be modified by properly adjusting these variables in complex systems in which compositions also play an important role to set the system behavior.

In the case of natural products processing, new and promising opportunities arise from the use of supercritical fluids as solvents, anti-solvents, separating agents or reaction media. There is an extensive literature concerning the potential applications of high pressure and supercritical technologies in the field of food industries, including the extraction, fractionation and purification of high added value ingredients, additives and bioactive substances, precipitation and encapsulation, polymer impregnation and chemical reactions.

The supercritical fluid extraction (SFE) of solid vegetable materials is currently the main application, considering the number of published research and commercial developments. The decaffeination of coffee beans and the extraction of hops, spices and essential oils were the first processes successfully applied at industrial scale. Numerous plant species are proposed as raw material to obtain aromatic or bioactive extracts. In fact, plant extracts are now regarded as potential sources of valuable active compounds, such as monoterpenes, sesquiterpenes, diterpenes, flavonoids, carotenoids, phenols, etc. with antimicrobial, repellent, antioxidant, preservative and other properties [1]. In general, their content in the raw material is low (0.1-10%), and the co-extraction of undesired compounds should be minimized to avoid subsequent purification steps. Therefore, research efforts are focused on improving extract yield and selectivity, by the optimization of extraction operating pressure and temperature, the use of cosolvents of different polarity, as well as with a suitable raw material pre-treatment to increase extraction rate, by reducing mass transfer resistances. The extract is generally recovered in a separation vessel by simple depressurization. If several compounds or fractions are co-extracted, they can be recovered separately using a series of separation vessels operating at different pressure and/or temperature. There are in the literature extensive reviews on SFE of aromatic and bioactive compounds from plant species [1,3,4,5]. A list of selected cases is given in Table 1.

Table 1. Typical applications of supercritical fluid extraction

Plant	Type of extract	Ref.
Aniseed	Essential oil	[6]
Black pepper	Essential oil	[7],[8]
Chamomile	Sesquiterpenoid lactones	[9]
Green tea	Catechins, polyphenols	[10]
Lavender	Essential oil	[11]
Origanum	Essential oil, carvacrol, thymol	[12],[13]
Peppermint	Essential oil	[14]
Rosemary	Essential oil	[13],[15]
Sage	Essential oil, diterpenes	[13,16,17]
Tagetes minuta	Essential oil	[18]
Thyme	Essential oil, thymol	[13],[19]
Vegetable oil	Rosehip oil, Sunflower oil	[13][19]

Fractionation of liquid mixtures with supercritical fluids is another promising field of application. Continuous contact columns, similar to those used in liquid extraction and distillation columns, have been proposed and developed for the concentration or purification of valuable components from natural liquid feeds [21]. In this case, research is more oriented to the optimization of operating conditions to maximize selectivity and reduce the use of supercritical solvent, while keeping the system within the heterogeneous region. Some examples of relevance in food industry are:

- extraction of polyunsaturated fatty acids and esters (like DHA and EPA) from fish oils [22],[23];
- extraction of squalene, carotenoids, tocopherols, phytosterols, phospholipids from vegetal edible oils [24,25];
- deterpenation of citrus peel oils [26],27,28];
- concentration of oxygenated monoterpenes and other bioactive components from essential oils [29,30];
- removal of hexane and other organic contaminants from edible oils [31];
- Fractionation of oleoresins [32].

The possibility of drastically modifying solubility conditions in a supercritical fluid by changing its density has been the base for many applications in precipitation, micronization, encapsulation and impregnation processes, which have been extensively reviewed by several authors [34],35,[36].

Supercritical solvent impregnation (SSI) of polymer films with food preservative compounds has been proposed for the development of active packaging. In this case, the supercritical fluid is used not only to dissolve the active substances, but also to facilitate its diffusion into the film, which

swells and plasticizes more easily by the action of high pressure fluids. After depressurization, the solutes are "trapped" inside the polymeric matrix. Impregnation of polymer films with antimicrobials cinnamaldehyde [37] and thymol [38] are some recent examples.

High pressure and supercritical fluid technologies have been also applied as fluid media in several chemical reactions of relevance in food industry. These reactions include catalytic hydrogenation of edible oils, hydrolysis, esterification, interesterification and transesterification for the production of fatty acid esters, mono and diglycerides and shortenings. The above mentioned reactions are diffusion-controlled due to the immiscibility of the reactants. The addition of a high pressure solvent brings the system to homogenous conditions, with the subsequent increase of reaction rate and conversion [39]. Moreover, being able to control the reactants ratio also allows improving the selectivity towards the desired products. An example of this technology is the hydrogenation of sunflower oil in supercritical propane [40].
In the next section the typical binary phase diagrams between the SCF and families of natural products and their thermodynamic modeling will be discussed.

2. Thermodynamic modeling of solubilities in SCFs.

The thermodynamic modeling of the solubility of a pure solid in a SCF gives us a clear picture of the role of the fluid phase thermodynamic properties on the dramatic increase in solubility that is observed above the SCF critical conditions. If we apply the isofugacity criterion to the solubility of a solid in a supercritical phase:

$$\hat{f}_i^{\,S} = \hat{f}_i^{\,G} \tag{1}$$

$$\hat{f}_i^{\,G} = \phi_i y_i P \tag{2}$$

$$\hat{f}_i^{\,S} = P_i^0 \phi_i^0 \exp\frac{v_S(P - P_i^0)}{RT} \tag{3}$$

Equations 1 and 2 explain clearly the effects of temperature, pressure and the solid physical properties vapor pressure P^o and solid specific volume v_s on the solid solubility in the SCF phase. Under pressures well below the SCF critical pressure, the solubility y_i is directly equal to the ratio of the solid sublimation pressure and the system pressure:

$$y_i = P^o/P \tag{4}$$

This is the ideal solubility that decreases with pressure and increases with temperature. However, when the system pressure is greater than the critical pressure of the SCF a drastic increase in the solubility is observed due to the effect of the system pressure on the solute fugacity coefficient in the fluid phase ϕ_i.

The strong dependence of solubility and density is the basis of the Chrastil [41] correlation of solute solubilities in supercritical fluids. This correlation gives the solubility (y_i) as a function of the solvent density and usually is written as follows:

$$ln(y_i) = k_i \, ln(\rho) + A_i \tag{5}$$

where A_i is a temperature dependent term and ρ is the SCF density under the system conditions.
The solubility of liquid solutes in supercritical fluids is also computed applying the isofugacity criterion for phase equilibrium at a given pressure and temperature:

$$f_i^L = f_i^G \qquad i=1,2..NC \tag{6}$$

The same equation of state is applied to both phases for the computation of the component fugacities. The application of cubic equations of state with classical mixing rules has been applied for the correlation of solutes solubilities in supercritical extraction and fractionation processes when the mixtures are moderately non ideal. Recently Fornari et al. [42] presented a review of applications of the van der Waals family of equation of state to the correlation of solubilities in SCFs. However when there are strong polar or associating interactions, or highly asymmetric systems, the application of cubic equations with classical mixing rules is severely limited. A typical situation arises when we need to predict LL and VLL equilibria in highly asymmetric systems, like ethane or propane with triglycerides. This capability of the equation of state is highly needed in SCF applications for tuning the solvent pressure and temperature to different process needs. These systems present low energetic interaction; however, a cubic equation of state is not able to correlate the multiphase equilibria with a single set of parameters [43]. An equation of state with a repulsive term based on hard spheres like the Carnahan – Starling is more adequate for this purpose. The GC-EOS equation of state [44] is derived from two contributions to the mixture Helmholtz function (A):

$$A = A^{rep} + A^{att} \tag{7}$$

This model has a repulsive hard spheres term, and a group contribution van der Waals type attractive energy term. The addition of a group-based associating contribution derived from the statistical association fluid theory (SAFT) was proposed by Gros et al. [45], to derive the GCA-EOS model.

$$A = A^{rep} + A^{att} + A^{assoc} \tag{8}$$

When dealing with natural products, a group contribution approach is very fruitful because with a few groups a large variety of compounds can be represented and provides the model with predicting capability. For example a complex molecule like linoleic acid methyl ester can be simply described by:

$$(CH3)_2(CH2)_{11}(CH=CH)_2(CH2COO)_1$$
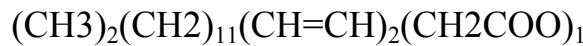
The application of equations of state to large molecules is hindered by the lack of information on critical properties of these compounds and most authors resort to empirical estimation methods. In the GCA-EOS model a property that is derived from critical properties or vapor pressure data, is the molecule critical diameter. Bottini et al [46] proposed to obtain the molecule critical diameter from experimental values of infinite dilution activity coefficients of a non polar molecule, like hexane, in a high molecular weight molecule. Later, based on this date, Espinosa et al [47] proposed a correlation of critical diameters as a function of the molecule van der Waals volume (r_{vdW}):

$$log(dc) = 0.4152+0.4128 \, log(r_{vdW}) \tag{9}$$

In this way the critical diameters of homologous families are also obtained by a group contribution approach.

3. Phase diagrams of binary systems of solutes with SCFs

The fluid phase behavior of binary systems is useful to understand the types of phase equilibria that SCFs can exhibit with different solutes. Van Konynenburg and Scott [48] found five types of phase behavior that cover most of experimentally studied systems. Also they have shown that all these behaviors can be qualitatively predicted by the van der Waals equation of state. Type I is typical of binary mixtures of similar size molecules and low non ideality, where complete liquid miscibility is observed up to the binary mixture critical point. Therefore, we have a continuous locus of the mixture critical points from the one of the light components to the heavy component. When the non ideality of the binary system increases we run into a region of liquid-liquid (LL) equilibria at low temperatures, but still there is a continuous locus of critical points between the two pure component critical points. When the binary system is highly asymmetric in molecular size, even with systems with moderate non ideality a different phenomena is observed. This condition is typical of binary mixtures of SCFs like ethane, propane or CO_2 with a large variety of natural products. In this type of binary mixtures, due to the appearance of liquid immiscibility near the light component critical point, there is a discontinuity in the locus of critical points of the binary mixtures with a transition from LV binary critical points starting from the heavier component to LL critical points that ends in a LLV line, this is Type V behavior. The later condition normally arises at high molar concentrations of the lighter component and it looks like a decrease of its solvent power near its critical point, creates a segregation of a second liquid phase. The LLV line starts in the lower critical end point and ends in the upper critical end point (UCEP). From the UCEP starts a LV critical line that ends in the lighter component critical point. Luks [49] gives a lucid description of how molecular interactions and size asymmetry determine the different types of binary phase equilibria (Fig. 1). In this figure univariant lines are plotted in which there is only one degree of freedom, like vapor pressure lines, LLV lines and LV and LL critical lines.
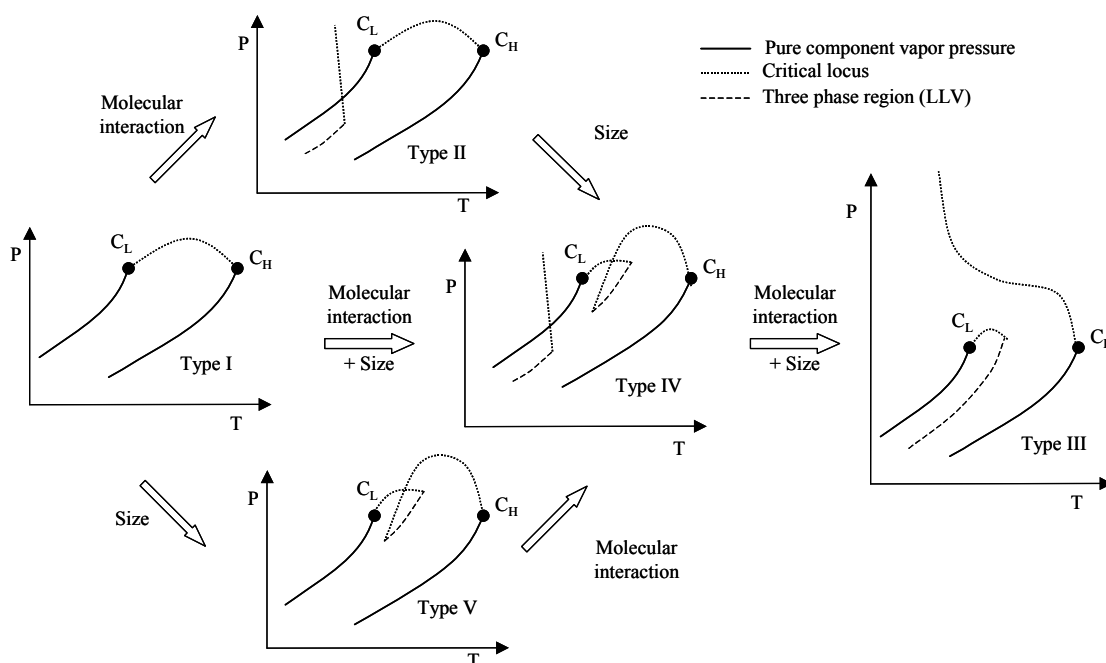

Fig. 1 Transition of types of phase equilibria with size asymmetry and molecular interactions.

Type IV phase behavior combines both the effects of size-asymmetry and high non ideality. Finally, when the low and high temperature liquid LLV line merge, the LL critical line diverges towards pressures above the critical point of the light component and the Type IV behavior switches to Type III. The latter has a continuous divergent locus of critical points that starts as a LV critical line and ends as a LL critical line at high pressure.

In this regard, Peters and co-workers [50,51,52] have done a great deal of experimental measurements to elucidate the expected phase behavior of supercritical fluids with the homologous series of many substrates like hydrocarbons (paraffinic and aromatics), alcohols, acids, among others. The evolution of the phase behavior of CO2, ethane and propane with molecular size, for alkanes, aromatics, carboxylic acids and alcohols are shown in Table 2. On the other hand Fig. 2 shows LCEP and UCEP for propane with high molecular weight alkanes and triglycerides. As it was explained, the temperature range between the LCEP and UCEP limits the appearance of the liquid immiscibility; therefore the plot shown in Fig.2 is a simple way to visualized binary phase behavior as the hydrocarbon chain increases.

Table 2. Evolution of the phase behavior of the homologous series of organic compounds with supercritical fluids.

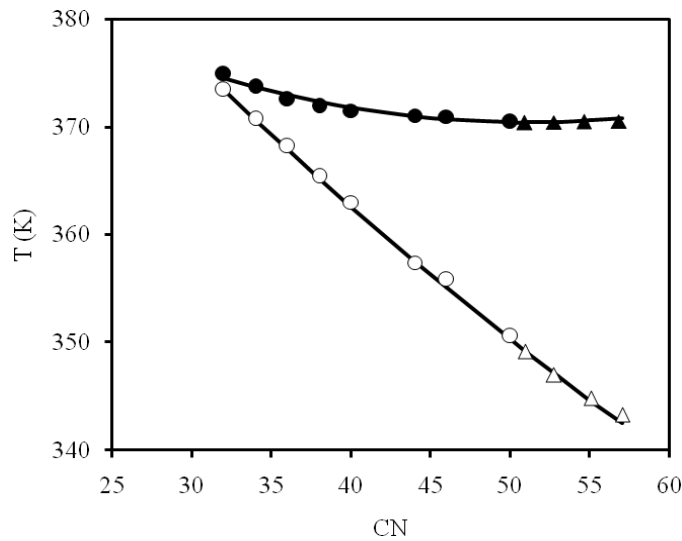| Solvent	Type I	Type II	Type IV	Type V	Type III
n-alkanes					
CO_2	C_1-C_6	C_7-C_{12}	C_{13}		C_{14}-C_{32}
Ethane	C_1-C_{17}			C_{18}-C_{23}	C_{24}-C_{28}
Propane	C_1-C_{29}			C_{30}-C_{50}	C_{60}
n-alcohols					
CO_2	C_1-C_2	C_3-C_4	C_5		C_6-C_{14}
Ethane				C_1-C_{10}	C_{10}-C_{16}
Propane	C_1-C_{16}			C_{17}-C_{26}	
n-alkanoic acids					
Propane	C_1-C_{14}			C_{15}-C_{22}	
alkyl-benzene					
CO_2	C_1-C_8	C_9-C_{12}			C_{13}-C_{22}
Ethane	C_1-C_{15}			C_{16}-C_{23}	

Fig. 2 Propane with paraffins (●○) and triglycerides (▲△). ○△: LCEP and ●▲:UCEP.

4. Phase scenarios in supercritical fluid processing of natural products

Among the supercritical solvents, low and high critical temperature (Tc) fluids depict different properties with regard to solvent power and selectivity. Low Tc solvents have critical temperatures close to ambient temperature. They have moderate solvent power and selectivities for high molecular weight or polar material. On the other hand, high Tc SCFs have high solvent power at supercritical conditions and low selectivity. CO_2, ethane and propane are typical examples of low Tc fluids and methanol, toluene and water of high Tc solvents. CO_2 and water are the most common solvent choice as a low and high temperature SCF. Both are readily available, cheap, non-flammable and non-toxic. These properties make them ideal solvents for sustainable processes. They have very different solvent properties, but we have to keep in mind that supercritical fluids can be tuned to meet the required solvent conditions. Water plays a significant role in the conversion of biomass and CO_2 in the extraction, and fractionation of high value thermally labile natural products.

Different phase scenarios are required in extraction or fractionation of natural products. It is known that even though the potential solvents are few: CO_2, ethane or propane, their solvent properties are highly sensitive to temperature, pressure and composition (in the case of mixed-solvents). The fact that the solvent power of SCFs is density dependent, and density can be changed dramatically with moderate changes in pressure or temperature, makes it possible to tune the SCF to the process needs. As an example, Fig. 3 shows the effect of temperature and pressure on the solubility of *Salvia officinalis* essential oil in supercritical CO_2[30]. These data when plotted against pressure show a temperature inversion point (known as crossover) that disappears when plotted against the SCF density.
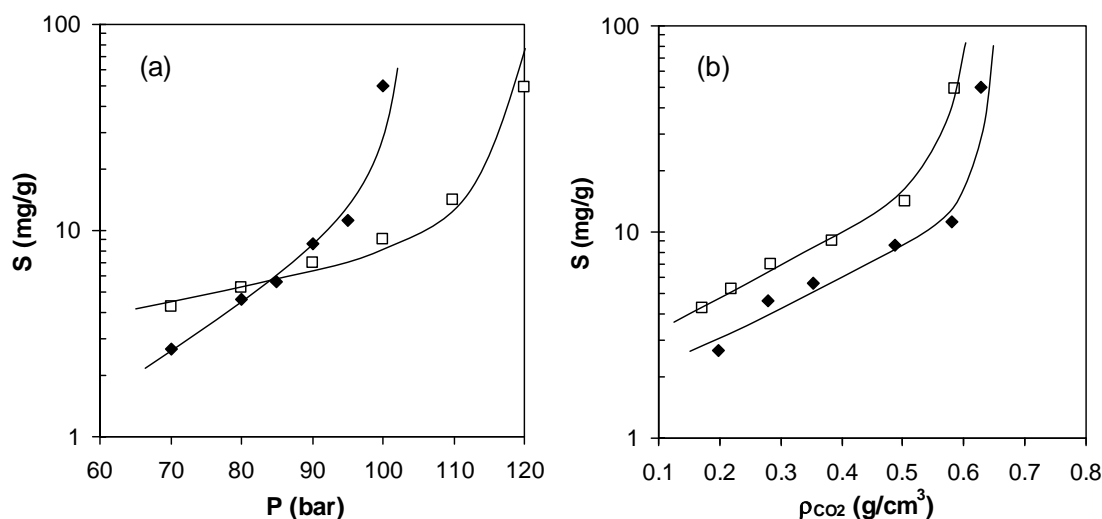
Fig. 3 Solubility of *S. officinalis* essential oil in CO_2 : (a) as a function of pressure, (b) as a function of pure CO2 density. ($\blacklozenge$) T=313 K, ($\square$) T=323 K,

The design of the phase conditions is based on the solubility and selectivity needs of the process under consideration. In extraction processes high solubility, or even complete miscibility of the solutes with the SCF is required. There are two alternatives, extraction from a liquid phase or from a solid matrix (seeds or plant material). In contrast, fractionation processes requires a careful design of the phase scenario to guarantee a two fluid phase (heterogeneous) conditions over the whole composition range of the separation process. As discussed in the previous section, these properties are dependent of the molecular size and the chemical nature of the solutes to be extracted or fractionated and their interactions with the SCF.

The phase scenarios for micronization can vary depending on the phase equilibria between the SCF and the solute to be micronized. In micronization process as rapid expansion of supercritical solutions (RESS) or gas anti-solvent, in which a cosolvent dissolves both the SCF and the solute, there are several micronization variants [40,41,42] determined by the specific phase equilibria between the SCF, the solute and the cosolvent. Different phase scenarios can be designed to achieve the desired micronization goal.

In the case of supercritical reaction processes, by the addition of a SCF, it is possible to carry out an otherwise heterogeneous reaction under homogeneous supercritical conditions over the whole reaction pathway. This is particularly important when we are dealing with solid catalyzed reactions of heavy substrates with gases [39].

4.1 Design of phase scenarios for green processes

The first step in the design of a phase scenario is the solvent selection. For this purpose we considered mainly three solvents: CO_2, ethane and propane, because they are low temperature solvents and inherently safe to be applied to nutraceuticals, bioactive and edible products. For each solvent we will discuss the design of the phase scenario for typical supercritical separation processes. The binary phase behavior of the SCF with the main components of natural products will guide us in the selection of operating conditions. The evolution of the phase behavior of the homologous series of certain organic families with supercritical fluids is given in Table 2.

CO_2 has been the solvent of choice, for well known reasons, in most supercritical processes. The solvent power of CO_2 is good for moderately polar and medium to low molecular weight compounds. For instance, important families in this range are the essential oils which are complex mixtures of terpenes and sequiterpenes, oxygenated terpernic derivatives, waxes and a non volatile

fraction. The extraction and fractionation of essential oils with supercritical fluids is attractive due to the use of safe solvents that leave no residues and the processes work at moderate temperatures that are suitable for thermally labile products. In Table 2 we can see that CO_2 with paraffinic components in the range C7-C12 have Type II phase behavior with LL immiscibility only at low temperatures. Outside this LL region there is complete miscibility and a continuous locus of the VL critical line between the critical point of the mono terpenic component and the critical point of CO2. The selection of proper values of operating temperature and pressure can be obtained from the phase envelope curve of a mixture of CO_2 with the main terpenic component. The phase envelope curve is obtained at a given global composition of the mixture to be extracted or fractionated. The main coordinates in a phase envelope diagram are the mixture critical point, the temperature and pressure at the maximum pressure (crioconderbar). For example in a mixture of CO_2 with a molar fraction of limonene of 0.02, the critical point is Tc=318.1 and Pc=91.3 bar and the criocondenbar is Pmax=147.2 at T=376.4 K. This is a clear indication that at temperatures lower than 370 K the operating pressure should be well below 150 bar to avoid entering the single phase condition. For example operating at 333K in orange peel oil deterpenation the selected operating pressure is 100 bar [33]. In a similar fractionation process the separation of β-ocimene (terpene) from the oxygenated fraction of *Tagetes minuta* oil at 323 K, the operating pressure for fractionation is 90 bar [30]. At lower temperatures lower pressures can be selected for the oil fractionation. If the goal of the process is the extraction and not the fractionation, the critical point of the mixture gives the minimum pressure required at Tc for complete miscibility, looking at the phase envelope diagram at higher temperatures the pressures of complete miscibility are readily obtained.

From Table 2 we can see that for binaries of CO_2 with alkanes of carbon number higher than 14 (for instance sesquiterpenes) a Type III behavior are observed.

Very often in supercritical processes we have to deal with size-asymmetric mixtures and, therefore, we should expect Types III, IV or V phase behavior, i.e. mixtures that show liquid phase split at temperatures near the critical point of the solvent. Fig. 4 illustrates these types of phase diagrams. If we are looking for high solvent power we should avoid diagrams with a rapidly divergent L1=L2 critical curve (cases a and b in Fig. 4) since in the near critical region, where we should operate our process, the system will show liquid immiscibility. Of course the more favorable diagrams are of Type I or II like the ones that we have already discussed for essential oil components, also a type III behavior with no rapidly divergent L1=L2 critical curve, may be feasible for extraction or fractionation because by increasing the pressure high enough the liquid immiscibility is avoided. An intermediate situation is that given by the case "c " of Fig. 4, a divergent critical line with a low negative, or near null, slope in the critical region; we still can achieve homogenous liquid phase but at pressures somewhat higher than the required for case "d" (Type V). An interesting example of this type of behavior is the concentration of docohexanoic acid methyl ester (DHA) and eicopentanoic acid methyl ester (EPA) by fractionation of a fish oil methyl esters mixture. The fractionation temperature has to be low (333 K) to avoid thermal degradation of the mixture. On the other hand the fractionation pressure is determined by looking at the type III binary diagram of CO2 with methyl oleate (a representative component of the lighter fraction) to select a pressure in the VL region of Fig. 5. At 333K a pressure of 145.0 bar is chosen for the fractionation process [53,54].

It is interesting to compare ethane with CO2 as a SCF solvent for orange oil desterpenation and fractionation of fish oil methyl esters. Using a similar approach the operating pressures for desterpenation with ethane is 63 bar. If we compare the pressures for both solvents ethane and CO2 in reduced coordinates they are very similar and close to Pr=1.5 [55]. The reduced pressure for the concentration of DEA and EPA are again similar however the values are higher (Pr=2). Higher pressures are required to increase the solubilities of heavier compounds in the supercritical phase. The concentration of α−tocopherol from deodorizer distillate FAME follows a similar approach and a pressure of 150 bar was used by Fang et al [56] to remove the methyl esters and finally a pressure of 200 bar was used to recover overhead the enriched tocopherol fraction.
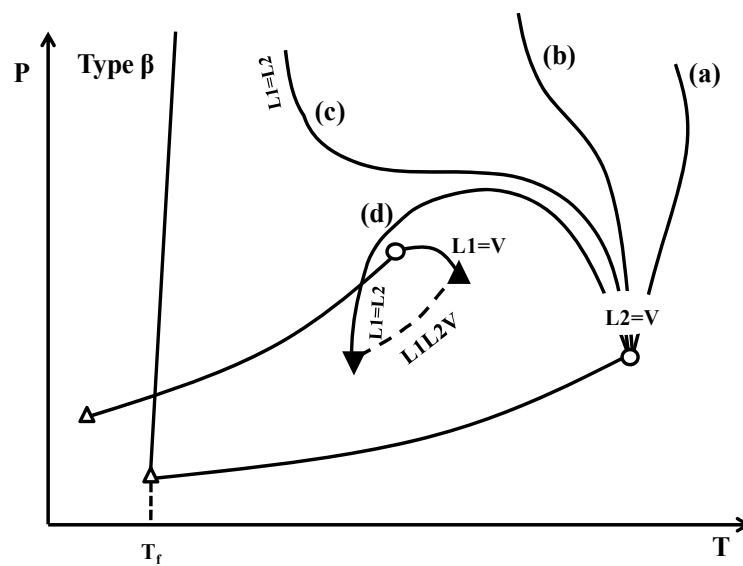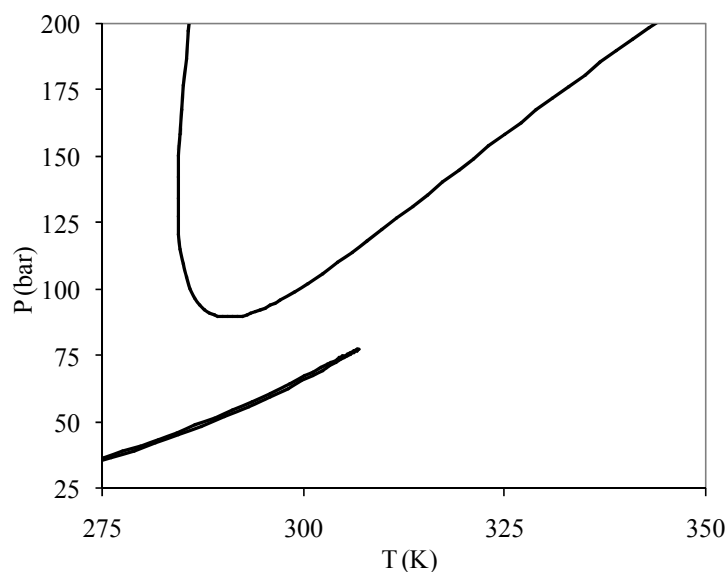
Fig. 4 Binary phase behavior of types III, IV and V

Fig. 5. General phase equilibrium diagram of CO_2 − Methyl oleate

The solubilities of triglycerides with more than 50 carbons in CO_2 are very low and the system presents a type III behavior with LL immiscibility up to very high pressures. This behavior can be advantageous to recover valuable components or eliminate contaminants from oil by operating at pressures in the 150-200 bar range. Examples of this behavior are the removal of hexane, squalene or oleic acid from olive oil [57].

Propane is a better SCF solvent for high molecular weight material. For instance it is completely miscible with alcohols up to C16 and with carboxylic acids up to C14. With triglycerides of carbon number less of C30 has Type I or II phase behavior and above it is Type V up to C56 (Fig 2). The Type V behavior of propane was the basis of the Solexol process for oil refining using propane

under selected conditions of pressure and temperature. At temperatures below the oil-propane LCEP, propane and the oil are completely miscible and only colored material and highly polar material are not dissolved. However, if the temperature is increased a liquid phase is segregated where vitamins and carboxylic acids are extracted. At higher temperatures the solubility of the oil in the propane phase is very low and propane can be used to remove contaminants or valuable products from the oil [31]. Another interesting approach is the use of mixed-solvents near critical fluids, for instance of CO_2 and propane, to meet requirements of non-flammability and high solvent power solvents for fixed oils [20]. In what follows we analyze a case study of supercritical reaction and fractionation process.

5. Case study. Production and separation of fatty acid methyl esters and acylglycerides by supercritical technology

Alternative technologies have been proposed in the last decade [58] to improve the conventional process of biodiesel production from unrefined oils and fats. One process of high interest is the non-catalytic continuous supercritical methanolysis, which can process raw materials with high free fatty acids (FFA) and water contents. The vegetable oil is treated with an excess of alcohol (40:1 methanol to oil molar ratio) at 593 K and 150 bar to achieve complete conversion. However, working under milder conditions, acylglycerides could also be obtained (543 K to 573 K). Acylglycerols are employed in the food and pharmaceutical industries as emulsifiers. In this example, a method for the fractionation of FAMEs + acylglycerides mixtures with CO_2 as a green solvent is studied.

5.1 Continuous supercritical alcohol transesterification

Fig. 6 shows a schematic diagram of a continuous supercritical alcoholisys process [59]. In this process the alcohol and raw oil are pumped in a molar ratio of 40 to 1 and mixed at a medium temperature (403 K) - low pressure ($\approx$ 8.4 bar) to obtain a single liquid phase. The liquid alcohol + oil solution is then pumped at 150 bar through a series of heat exchangers to increase the temperature of the system to the reaction conditions (553 K to 593 K). The vegetable oil transesterification to fatty acid esters takes place through three consecutive reactions in a tubular reactor. According to the reaction time (10 min. to 30 min.) and operating conditions, mono and diglycerides (MG and DG) are also produced as intermediate products. After heat exchange with the feed to the reactor the reaction products and excess alcohol are expanded to atmospheric pressure and the alcohol is recycled. In this way a glycerol and biodiesel/oily phases are separated by decantation. Fatty esters, mono and diglycerides remain in the biodiesel or oily phase.
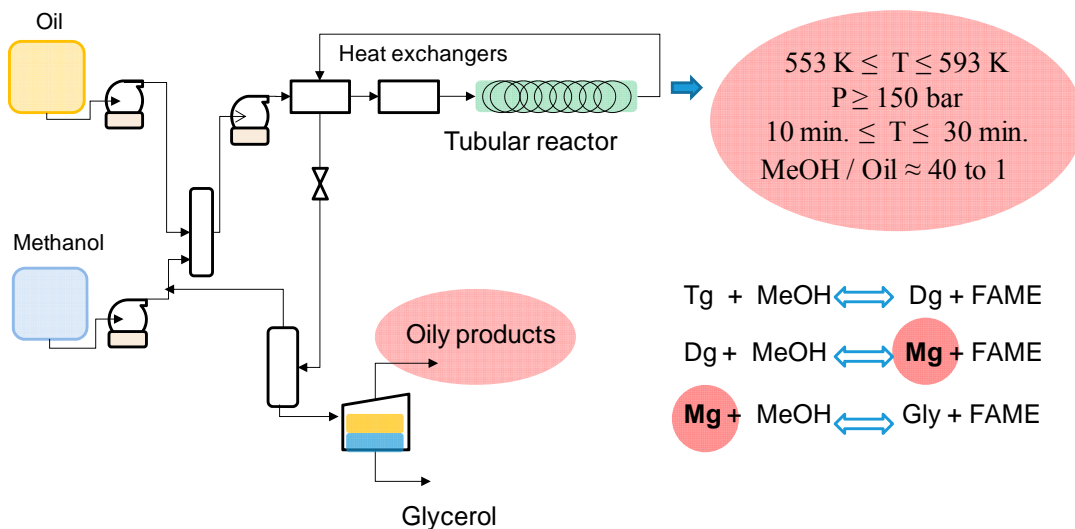

Fig. 6. Schematic diagram of the supercritical alcohol transesterification process

The operating conditions should be properly selected to have a single phase in the system, to avoid mass transfer limitations, and to have proper reaction kinetics. Velez et al. [59] study the supercritical methanol transesterification of high oleic sunflower oil. The operation at a pressure of ≈150 bar ensures a single phase reaction system working with molar ratios in the order of 40 to 1 alcohol to oil and in a range of temperature of 553 K to 608 K. In the vegetable oils + methanol binary system, the liquid partial miscibility persists up to temperatures in the order of 403 K to 423 K where the phase transition from liquid-liquid-vapor to liquid-vapor equilibria takes place [60]. Further temperature increments (553 K to 593 K) turn liquid-vapor equilibria in a single phase if the final operating pressure is in the order of ≈150 bar. Also, the advance of the reaction aids to the phase transition because the presence of fatty esters in the system reduce the critical temperature of the reacting system [61].
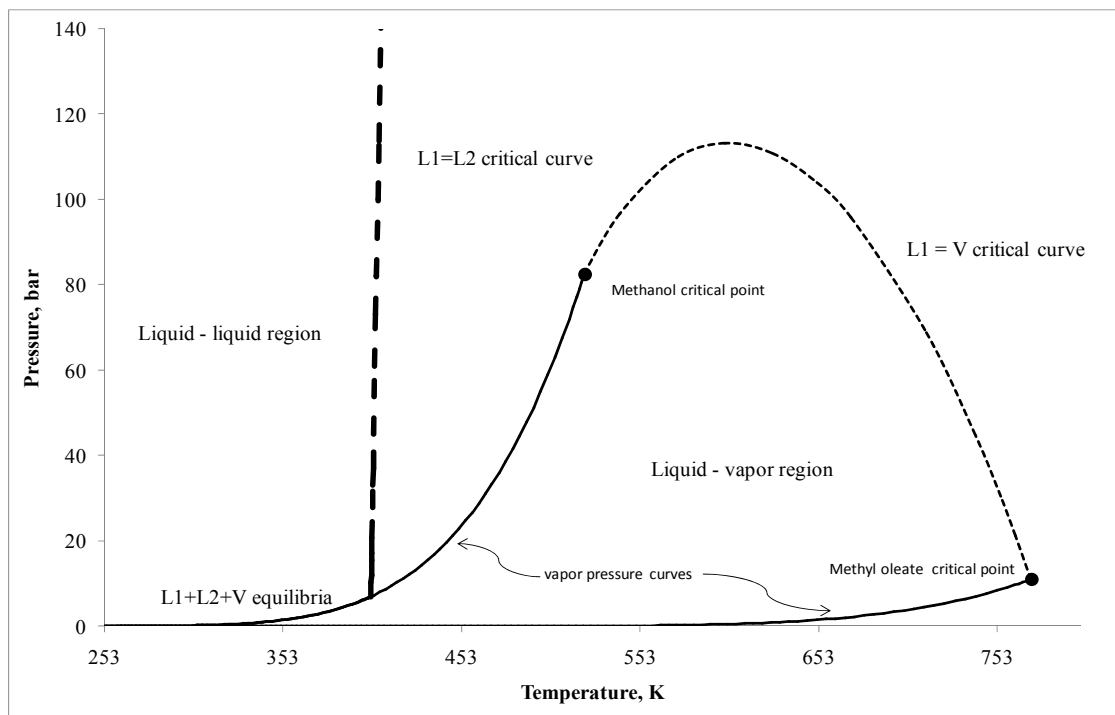

Fig. 7. GCA-EoS prediction of the general phase behavior of the methanol + methyl oleate binary system. Diagram calculated with GPEC [62].

GCA-EoS prediction of the system methanol+methyl oleate shows a type II diagram of the classification of van Konynenburg and Scott (Fig. 7), with liquid-liquid equilibria at temperatures lower than the critical end point (403 K). According to the model predictions the system methyl oleate + methanol exhibit a single phase at pressures greater than 120 bar in the range of temperature normally used in the supercritical methanolisys.
When the reaction is carried out at 593 K, 150 bar and reaction times higher than 30 min., a biodiesel phase with a concentration of fatty acid methyl esters (FAMEs) greater than 95 wt.% [59] is obtained. However, when the reactor operates under milder reaction conditions a sizeable concentration of mono a diglycerides are obtained. Velez et al. [59] and more recently Soto et al [63] achieved a biodiesel with ≈80 wt.% FAME, ≈16 wt.% monoglycerides and ≈4 wt.% diglycerides when the reactor operates at 573 K, 150 bar and ≈15 min. of reaction time. Fig. 8 shows the reaction products obtained in the supercritical methanol transesterification of high oleic sunflower oil with a molar ratio of methanol to oil of 40 to 1 and different reaction temperatures, reported previously elsewhere [63].
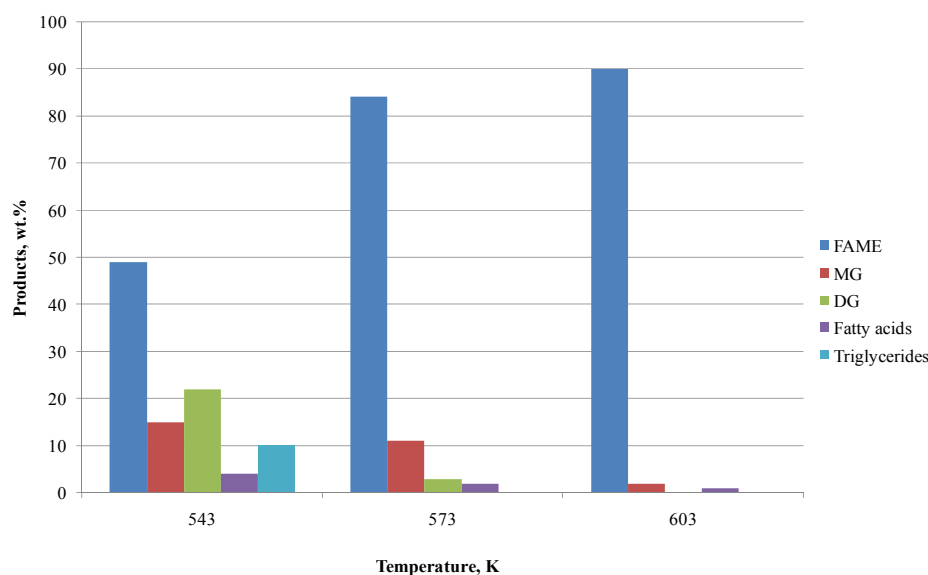
Fig. 8. Reaction products obtained in supercritical methanol transesterification of high oleic sunflower oil at 543 K, 573 K and 603 K with a methanol to oil molar ratio of 40 to 1. Data reported by Soto et al. [63].

The separation of acylglycerides and fatty acid esters is interesting because MG/DG have applications as emulsifiers in the food industry and acylglycerides are relevant in the elaboration of personal care products due to its excellent skin compatibility. Therefore, the separation of the reaction products to obtain FAME and MG/DG is an attractive alternative. In order to design the operating conditions of the separation process with CO_2, it is necessary to study the thermodynamic phase behavior of the reaction mixtures with this solvent as a function of pressure, temperature and product concentrations.

5.2 Binary systems

Fig. 9 shows the phase behavior of the binaries CO_2+methyl oleate (Fig 9A) and CO_2+monoolein (Fig 9B) in a pressure – composition diagram at different temperatures (298, 313, 323 and 333 K). GCA-EoS predictions of the systems are in agreement with experimental data reported previously by Inomata et al. [64] for methyl oleate and Nilsson [65] for monoolein.

The binaries of CO_2 + fatty acid methyl esters (C16 to C20) present a type III diagram in the classification of van Konynenburg and Scott (see GCA-EoS predictions in Fig. 10). The fatty acid methyl esters + CO_2 binary systems have a minimum temperature and pressure in the L1=V critical curve near the CO_2 critical point. Figure 9A shows that the binary CO_2+methyl oleate presents liquid-liquid-vapour equilibria at room temperature (298 K) and a pressure close to the CO_2 vapour pressure. The partial miscibility in these systems at this temperature ends at pressures higher than 100 bar. As expected, higher operating temperatures require higher pressures to obtain a complete miscibility between the substrate and solvent. For example, the binary methyl oleate +CO_2 exhibit at 313 K liquid-vapor equilibria at 100 bar and only above 125 bar a single fluid phase is reached. Fig. 9.B shows the solubility of monoolein in liquid and supercritical CO_2 predicted by the GCA-EOS, together with experimental data reported in the literature [65] for this system.
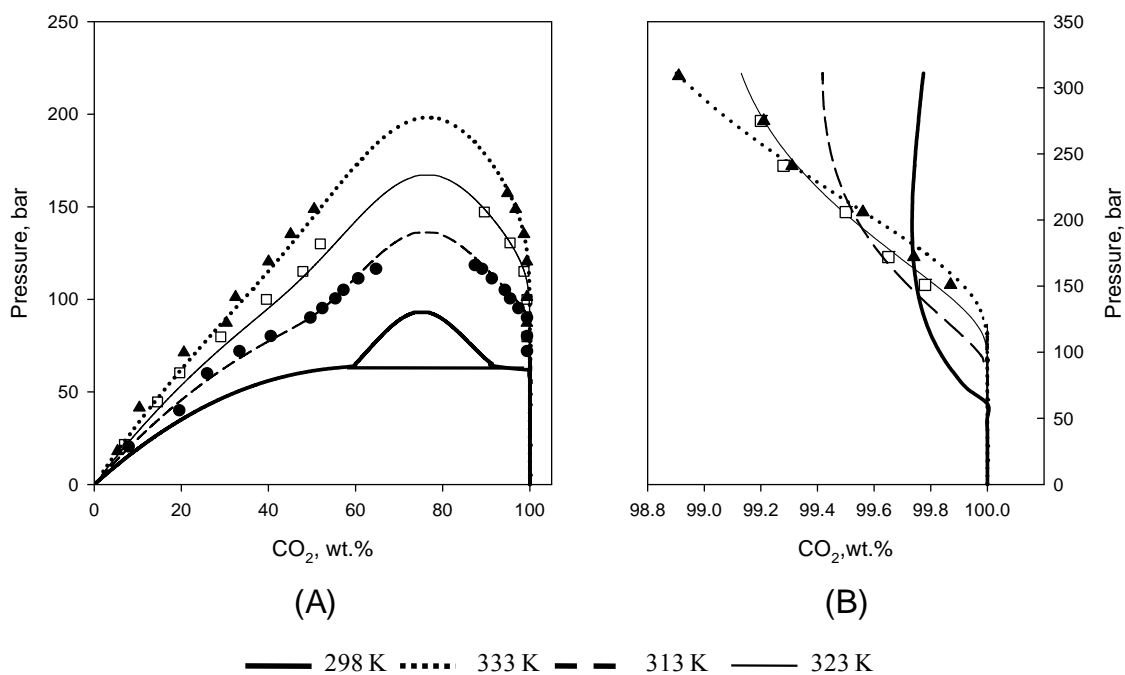
Fig. 9. Pressure concentration diagram at different temperatures for the binary systems of CO_2 with: A) methyl oleate; B) monoolein. Lines are GCA-EoS predictions. Symbols are experimental data at (●) 313 K, (□) 323 K and (▲) 333 K for (A) methyl oleate [64] and (B) monoolein [65].
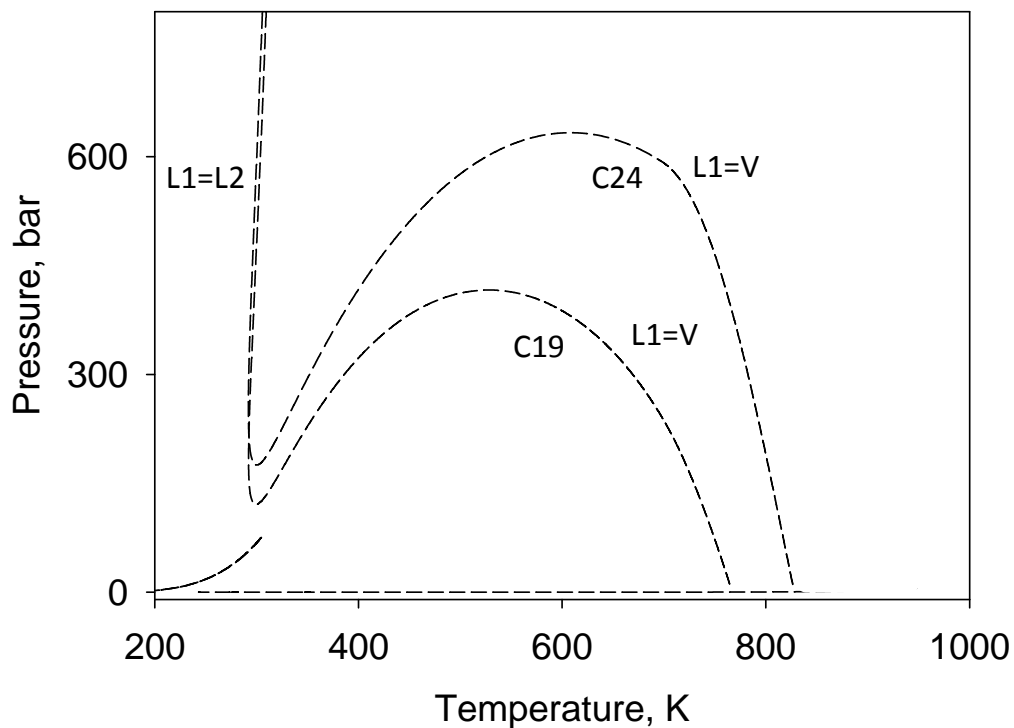

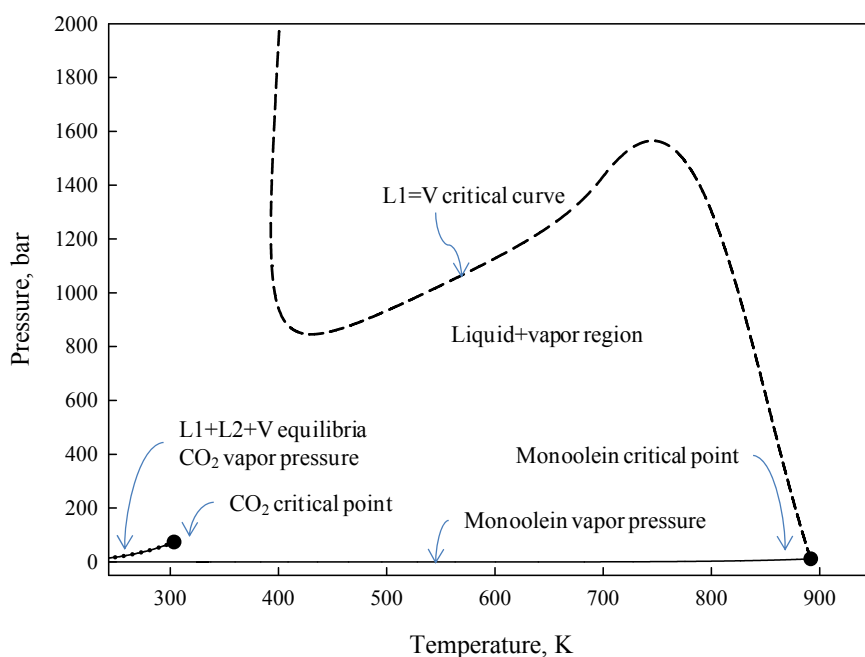**Fig. 10.** General phase equilibrium diagram of CO_2 – Methyl oleate

Fig. 11. General phase equilibrium diagram of CO_2 – Monoolein

The binary CO_2 – monoolein is also type III (Fig. 11), but with a divergent supercritical line, that extends the LL region at temperatures much higher than the critical temperature of CO_2, due to the polar nature and higher molar mass of monoglycerides. GCA-EOS predictions indicate the solubility of MG in the liquid solvent phase (L2) at 298 K and 75 bar is 0.013 wt.%. Moreover, it increases with pressure (0.15 wt.% at 200 bar). At temperatures higher than the CO_2 critical point, the system presents liquid-vapour equilibria. Nilsson et al. [65] measured monoolein solubility in supercritical CO_2 at 323 K and 333 K. The authors found a MG solubility of 0.22 wt.% at 323 K and 151 bar. Furthermore, it increases with pressure up to 0.7 wt.% at 241 bar. GCA-EOS predictions are in good agreement with the experimental data (see Fig. 9.B). Diolein solubility, according to these authors, is 0.16 wt.% at 323 K and 151 bar and 0.8 wt.% at 241 bar, showing a similar behaviour to MG. Therefore, it can be assumed that under the studied operating conditions, MG/DG distribute equally between the liquid-vapor, and liquid-liquid phases in equilibrium with CO_2. Thus, all the acylglycerides present in the reaction products are modeled as MG.

5.3 CO_2+FAME+MG ternary system
Experimental data of the ternary CO_2 + fatty acid methyl esters + MG/DG was reported by Soto et al.[63]. These authors showed that GCA-EoS predictions of this system, agree well with the experimental data under conditions of liquid-liquid-vapor, liquid-liquid and liquid-vapor equilibria. Fig. 12 shows the phase behavior of the system CO_2 + methyl oleate + monoolein predicted by the GCA-EoS at different operating temperatures and pressures. At room temperatures (298 K) the system exhibit liquid-liquid-vapor equilibria at pressures nearly the vapor pressure of CO_2. Both binaries of CO_2 with monoglycerides and fatty acid esters show liquid-liquid-vapor equilibria at 298 K. Therefore, reaction products of mixtures of esters and MG/DG present partial liquid miscibility in all range of relative product concentrations at temperatures lower than the critical temperature of CO_2, with small variations of pressure because the reaction products has negligible vapor pressure respect to CO_2. In the example shown in Fig. 12 at 69 bar and 298 K , when the oily or heavy liquid phase has a composition of 40 wt.% CO_2, 38 % FAME and 22% of MG/DG. The CO_2 liquid phase,

on the other hand, has a concentration of 5 wt.% esters and a negligible fraction of MG/DG components (tie line shown with red dots in Fig 12).
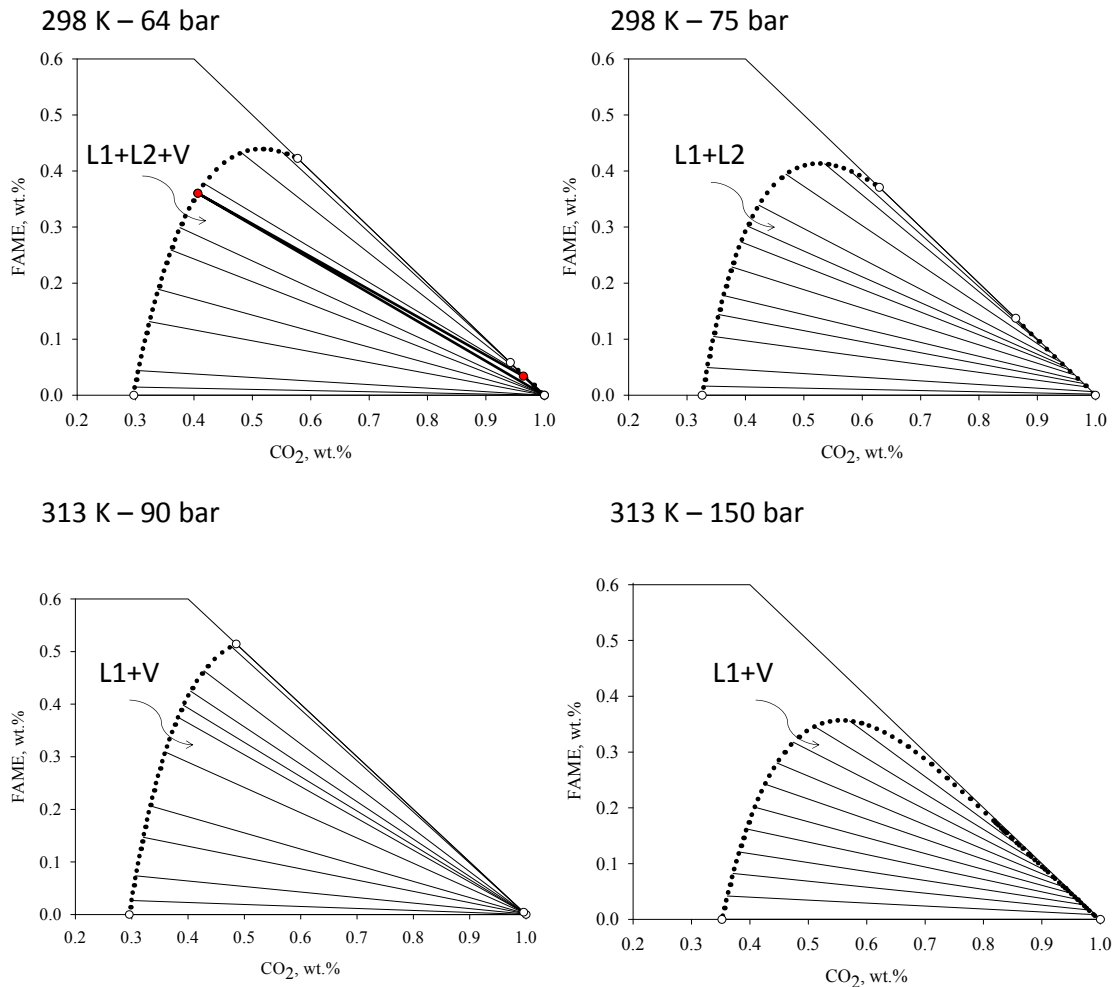

Fig. 12. Phase behavior of the ternary system CO_2+FAME+MG/DG at 298 K and 313 K. Methyl oleate and monoolein were selected in the thermodynamic model as representative components of FAME and MG, respectively. (○) binary phase equilibria predictions. GCA-EoS parameters are reported elsewhere [63].

The system shows liquid-liquid equilibria at pressures barely higher than the vapor pressure of CO_2 and temperatures lower than the CO2 critical temperature. Fig. 12 shows the phase equilibria of the ternary system at 298 K and 75 bar. The behavior of the liquid phases is similar at the one described for the liquid-liquid-vapor. In this particular case the solvent phase has a higher concentration of esters. As mentioned earlier, the ester exhibits a greater solubility in liquid CO_2 with the increase of pressure. When the temperature is increased to 313 K the system shows liquid-vapor equilibria. The temperature increment reduces the solvent power of CO_2 to dissolve esters and it is necessary to increase the pressure in order to get higher fatty ester concentration in the solvent phase. For example, according to GCA-EoS predictions the ester solubility is lower than 0.5 wt.%. at 313 K and 90 bar, when the oily liquid phases has a CO_2 concentration between 30 to 45 wt.%. A pressure increment to 150 bar turns the binary CO_2+fatty ester completely miscible at 313 K, and as expected, in the ternary system increased the ester solubility up to 15 wt.% fatty esters in the solvent phase.

5.4 CO$_2$ fractionation of MG and FAME.

An analysis of the phase behavior of the binary and ternary systems: FAME, MG and CO$_2$ indicates that the separation process should be carried out at low temperature (298 K to 313 K) in order to avoid high operating pressures. If the separation process operates at pressures greater than 150 bar the solubility of MG/DG in CO$_2$ increases drastically. At this pressure the FAME + CO$_2$ binary at exhibits complete miscibility at temperatures lower than 313 K. Moreover the phase behavior of the ternary indicates that a higher concentration of fatty ester in the solvent CO$_2$ phase induces   the dissolution of MG/DG compounds. The esters are a sort of co-solvent in the system that turns MG/DG more soluble in CO$_2$. Therefore, different phase design conditions should be analyzed when operating under liquid or supercritical conditions.

In order to compare operating conditions (temperature, pressure and solvent load) of a counter-current column, a multi-component extraction was simulated based on the GCA-EOS predictions of phase equilibria. The simulation routine is the GCEXTRACT program developed by Andersen [66] upgraded to use the GCA-EOS model. A seven-stage isothermal extractor was assumed for the fractionation of a reaction product containing 80 wt.% of FAME and 20 wt% of MG. At a given pressure, the solvent load was calculated specifying a FAME concentration of 0.2 wt.% in the MG/DG refined product.
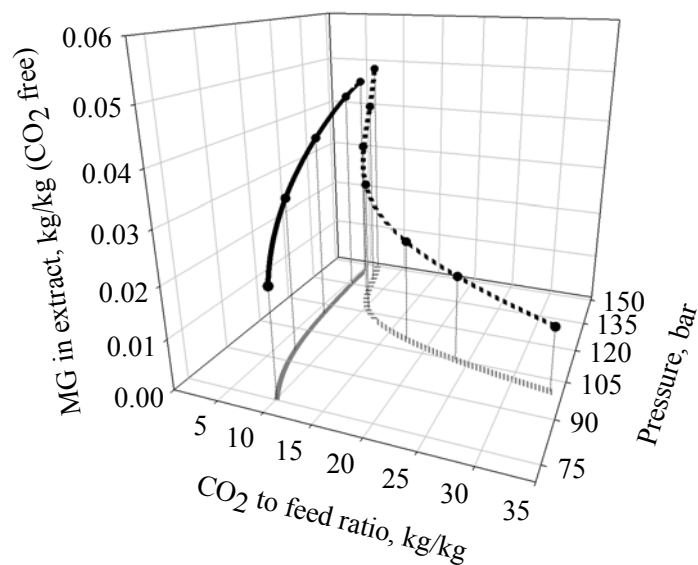
(A)
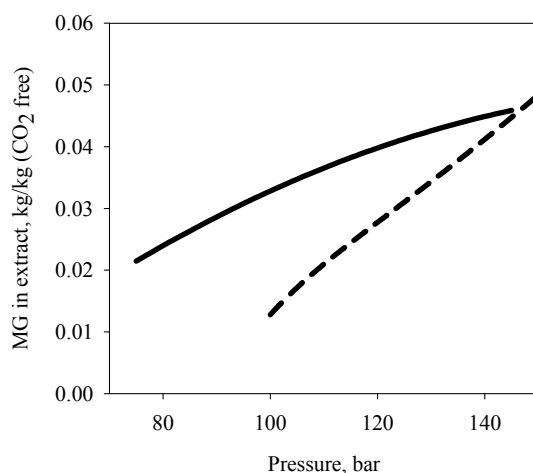
(B)

Fig. 13. Seven-stage countercurrent liquid and supercritical fractionation processes for FAME/MG separation using CO_2 at 298 K (solid lines) and 313 K (dashed lines), respectively. (A) MG content in the extract phase (on a CO_2-free basis) as a function of the operating pressure and the CO_2/feed mass ratio. (B) Projection of the MG content in the extract phase as a function of pressure.

The operation at 313 K requires a higher solvent load respect to the lower temperature, at a given pressure. On the other hand, the supercritical operation has a higher selectivity that allows a better MG recovery also at a given pressure. On the other hand, Fig. 13B shows that the MG in the extract at 298 K increases almost linearly with pressure from about 2.1 wt% to 4.5 wt.% at 75 bar and 145

bar, respectively. The operation of the extractor at 313 K shows also a linear increment of the MG content in the extract. However, the supercritical operation is more sensitive and MG in the extract increases from 1.3 wt.% to 4.8 wt.% at 100 bar and 150 bar, respectively.

In conclusion, if less than 2 wt.% of MG in the extract (CO_2 free) is desired, , the extractor should be operated at 313 K and 110 bar (or lower) using a solvent to feed mass ratio of 17 kg CO_2 / kg of feed. On the other hand, if a MG content in the extract of 3 wt.% is acceptable, the operation at 298 K is more attractive because it is possible to operate the extractor at 90 bar using a solvent to feed mass ratio of 7.5 kg CO_2 / kg of feed.

Summary

In this work we wanted to highlight the importance of applying phase equilibrium engineering principles when dealing with pressure-intensified processes, which have already proved to be of great interest in the food processing industry. In order to ensure efficient operation we must be able to design and control the system phase behavior. The road to the phase design is guided by the mixture to deal with and the process goals. The mixture components molecular interaction with the SCFs determine the binary mixtures phase behavior. The use of thermodynamic models makes it possible to explore different phase scenarios to carry out our process, identify the limits of the operating variables and select by computer simulation optimum process conditions.

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