SUPERCRITICAL FLUID EXTRACTION
OF EVENING PRIMROSE OIL
KINETIC AND MASS TRANSFER EFFECTS

ESTRAZIONE CON FLUIDI SUPERCRITICI DELL'OLIO DI ENAGRO
STUDIO DELLA CINETICA E DEL TRASPORTO DI MATERIA

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ABSTRACT

For processing utilization, supercritical fluid extraction requires a thorough understanding of the relevant phase equilibria, mass balance, and kinetic factors that impact on the successful recovery of extracts. In this study, we have determined the factors contributing to the kinetics and mass transfer of evening primrose oil (EPO) from its ground seed matrix, to supplement previously determined solu-

RIASSUNTO

L'utilizzazione a livello industriale del processo di estrazione con fluidi allo stato supercritico richiede una approfondita conoscenza dei pertinenti equilibri di fase, dei bilanci di materia e delle cinetiche che influenzano il recupero degli estratti. Al fine di integrare i dati di solubilità ed alcuni dati analitici precedentemente determinati sull'olio di enagro estratto con CO₂ supercritica, in questo lavoro sono stati

- Key words: evening primrose, extraction, kinetics, mass transfer, oilseed, supercritical fluid.
INTRODUCTION

Supercritical fluid extraction (SFE) has matured considerably as a processing technique over the last two decades due to fundamental thermodynamic, phase equilibria, and mass transfer studies, which have provided a basis for design of industrial scale extraction processes. For the extraction of oils from seeds and other natural products, SFE provides the processor with the ability to alter the solvent power of the fluid, and hence its selectivity, by varying the fluid's density as a function of extraction pressure and temperature. This factor, coupled with the superior mass transport properties exhibited by solutes dissolved in supercritical media, assures rapid removal of solutes from natural product matrices.

Studies on the solubility of seed oils in our laboratory (FRIEDRICH et al., 1982; SNYDER et al., 1984) as well as others (STAHL et al., 1980; 1983) have provided a basis on which to design processes for the rapid and efficient removal of oil from a variety of seed and natural product matrices (KING and LIST, 1996). Other investigators in the 1980's (FATTORI, et...
al., 1988; EGGERS et al., 1985; EGGERS and SIEVERS, 1989a) provided preliminary information on the kinetics of oil extraction, which, along with basic mass transfer data for solutes in supercritical fluids (KING et al., 1983; LAHIERE and FAIR, 1987; BARTOLOMEO et al., 1990; GOTO et al., 1990; LIM et al., 1990; ZEHNDER and TREPP, 1993), are required for the successful design of a supercritical fluid extraction process (RIZVI et al., 1986).

In the absence of the latter data, modelling studies of the SFE process become important since they provide the basis of optimizing the extraction process as well as a guide to process scaleup. Early research by BRUNNER (1984) explored the fundamental variables that influenced the mass transfer of solutes from solid substrates and provided the basis for several other modelling studies involving oilseed substrates (HONG et al., 1990; SAKAKI et al., 1990; FAVATI and FLORENTINI, 1993; SOVOVA, 1994; COELHO et al., 1995; STASTOVA et al., 1996) which were to follow. More recently, SFE kinetics have been interpreted in terms of a more fundamental physicochemical approach (BARTLE et al., 1990; GOTO et al., 1996) or have employed chromatographic models to describe oil depletion from packed beds of seed (GOODRUM et al., 1996). REVERCHON (REVERCHON, 1992; REVERCHON and OSSEO, 1993; POLETTO and REVERCHON, 1996), in particular, has been very active in testing alternative models for the SFE of oilseeds and other natural product matrices.

Previously, we have reported on the solubility of evening primrose oil (EPO) in supercritical carbon dioxide (SC-CO<sub>2</sub>) and characterized its extraction from the seed (FAVATI et al., 1991a,b). Others have also employed SC-CO<sub>2</sub> for the extraction of EPO (LEE et al., 1994) and further characterized the resultant oil with respect to its thermal stability (ANDRICH et al., 1994). The importance and commercial value of EPO is related to its gamma-linolenic acid content (HUANG and MILLS, 1996) and use as a nutraceutical and curative agent (FAVATI et al., 1991a,b; FANTOZZI et al., 1993); hence it is a seed oil of high value whose economic value is well matched with the attendant costs of conducting SFE.

In this paper, we have extended our previous studies to determine the kinetic and mass transfer dependency of the extraction of EPO with SC-CO<sub>2</sub>, and modelled the extraction rates using the methods of HONG, et al. (1990) and BRUNNER (1984). Interpretation of the SFE rate of EPO from the seed matrix can be facilitated with the aid of concepts advanced by EGGERS (EGGERS et al., 1985; EGGERS and SIEVERS, 1989b), and these have been utilized to generalize the extraction rate dependency on CO<sub>2</sub> pressure and flow rate. Extraction models such as those advocated here, have been employed by others for the SFE of fungal lipids (CYNAROWICZ-PROVOST et al., 1995; CYNAROWICZ-PROVOST, 1996), algae lipids (MENDES et al., 1996), and commodity seed oils (YOO and HONG, 1996). As we shall subsequently show, they work equally well for modelling the SFE of EPO.

MATERIALS AND METHODS

Experimental

The experimental apparatus used in these extraction studies has been previously described (FAVATI et al., 1991a,b) and will not be repeated here. The extraction vessel consisted of a 316 SS tube, pressure rated to 76 MPa, having dimensions of 1.75 cm i.d. x 30 cm. in length. Tests were conducted on 50 grams of evening primrose seeds that had been ground to pass a 0.355 mm sieve. The evening primrose seeds were obtained from Efamol Ltd. (Guilford, England). Their oil content was 26.1% by weight on a dry basis; gamma-linolenic acid represented 10% by weight of the total fatty...
acid content. Both oil and moisture content of the seeds were determined using standard AOCS methods (AOCS, 1989).

Extractions were performed at 20, 30, 50 and 70 MPa, over a temperature range of 40°-60°C. A fixed quantity of CO₂ (2.5 kg) was used for each experiment with extracts being collected after 125, 250, 375, 500, 1,000, 1,500, 2,500 g of CO₂ had passed through the seed bed. Different CO₂ flow rates were also employed during this study: 9, 18, 27 g/min. The CO₂ flow rate was determined with the aid of a Micro Motion Model D6 (Boulder, CO, USA) mass flow meter placed immediately before the gas booster pump. These CO₂ mass values were also checked by measuring the gas volume utilized with a dry test meter (Singer Model DTM-115, American Meter Division, Philadelphia, PA, USA), placed after the receiver vessel, and then converting the differential volume of CO₂ that was measured to a mass basis.

Since a small amount of water was coextracted with the EPO, the collected extracts were diluted with anhydrous ethyl ether (J.T. Baker, Phillipsburg, NJ, USA), and dried over sodium sulfate (EM-Science, Gibbstown, NJ, USA). The extraction yields were then determined by gravimetry after removing the solvent under vacuum.

Calculations

The calculation of the theoretical extraction curves utilized the models developed by HONG et al. (1990) and BRUNNER (1984), in which the SFE is partitioned into two sectors: one in which the rate of oil extraction is constant, followed by a second sector where the oil extraction is governed by diffusion out of the seed matrix. The first part of the model assumes that steady state mass transfer is applicable and that it can be used to represent values obtained during the initial period of the SFE, where there is a linear dependence of the amount of oil extracted with the mass of the supercritical fluid. The second part of the model applies to the SFE at longer extraction times, where the unsteady state predominates, and mass transfer is limited by diffusion.

Using the convention of BRUNNER (1984) and HONG et al. (1990), the amount of oil extracted per unit time during the constant rate period may be expressed as:

\[
m = k_g A_s \Delta C_m \quad (1)
\]

and

\[
A_s = \frac{6 (1 - \Psi) V_t}{d} \quad (2)
\]

where \(k_g\) is the external mass transfer coefficient (m/s), \(A_s\) is the specific mass transfer area (m²), \(\Psi\) is the void volume fraction, \(V_t\) is the total volume in the bed (m³), \(d\) is the particle diameter (m), and \(\Delta C_m\) is the concentration difference between the oil at the mass transfer interface and in the bulk phase in g/m³.

The required external mass transfer coefficient, \(k_g\), is computed using the following correlation:

\[
Sh[1 + 1.5(1 - \Psi)] = m Re^{1/2}Sc^{1/3} \quad (3)
\]

where \(Sh = \) Sherwood number = \(k_g d/D_g\), \(Re = \) Reynolds number = \(\mu V_t/\rho\), \(Sc = \) Schmidt number = \(\mu/\rho D_g\), \(D_g = \) Binary diffusion coefficient (m²/s).

The required Sherwood, Reynolds and Schmidt numbers were computed for the experimental data points and are presented in Table 1 for various pressures, temperatures and fluid velocities, \(u\). Utilizing the computed values for \(Re\), \(Sh\), and \(Sc\) and estimates for \(V_t\), \(\Psi\), and \(d\) (see Table 1), the mass transfer coefficient, \(k_g\), can be determined using eq. 3. Density (\(\rho\)) values for SC-CO₂ were obtained from KENNEDY (1954), while the relevant fluid viscosities (\(\mu\)) utilized were taken from VUKALOVICH and ALTUNIN (1968). The binary diffusion coefficient...
Table 1 - Mass transfer coefficients for the constant rate period of extraction.

<table>
<thead>
<tr>
<th>Temp.</th>
<th>Pres.</th>
<th>u(m/s)</th>
<th>Re</th>
<th>Sc</th>
<th>Sh</th>
<th>$K_e$(m/s)</th>
</tr>
</thead>
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<tr>
<td>40</td>
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<td>0.50793</td>
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<td>0.504045</td>
<td>96.78308</td>
<td>0.012554</td>
<td>3.55E-07</td>
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<tr>
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<td>90.90668</td>
<td>0.011773</td>
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</tr>
<tr>
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<tr>
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<tr>
<td>60</td>
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<td>0.685789</td>
<td>88.65987</td>
<td>0.010744</td>
<td>4.28E-07</td>
</tr>
</tbody>
</table>

The following parameters were used for calculation of mass transfer coefficients: $\Psi = 0.04$, $d$ (particle diameter = 0.0002828 m, $D_o = 8.0 \times 10^{-9}$ m$^2$/s, $V_i = 7.31 \times 10^{-5}$ m$^3$. Temp. (°C), Pres. (MPa) was estimated to be 8.0E-9 m$^2$/s, based on an average of values given by HONG (1990) for soybean oil in CO$_2$. Such an approximation is reasonable since the composition of EPO is similar to that of soybean oil, a triglyceride-based seed oil.

Implicit in the derivation of eq. 1 is that concentration of extracted oil at the mass transfer interface is constant. By analogy with HONG et al. (1990), $\Delta C_m$ was assumed to be 95% of the equilibrium value. The experimental values for Re, Sc and Sh were plotted (Fig. 1), and linear regression used to determine the constant m in eq. 3. HONG et al. (1990) reported a value for m equal to 0.035; in this study, m was found to be 0.2548. As noted in Fig. 1, the calculated values are somewhat scattered, yielding a regression coefficient of only 0.59. Despite this correlation, the computed mass transfer coefficients were found to be more than adequate to model the steady state sector of the extraction curves for EPO in SC-CO$_2$.

Inspection of the computed values for $K_e$ in Table 1 indicate that external mass transfer coefficient increases with increasing fluid velocity. The values of $K_e$
computed for the EPO extractions are an order of magnitude higher than those reported for soybean oil at the same particle diameter. This can be attributed to the higher flow rates utilized in this study.

For the unsteady state mass transfer sector, we have used the diffusion equation given by HONG et al. (1990), where:

\[
m(t)/m_0 = A \exp(-Kt) \quad (4)
\]

and

\[
K = F_0/t \\
F_0 = D_s t/d^2
\]

Here, \(m(t)\) is the amount of oil that remains in the seeds at time \(t\), \(m_0\) is the initial amount of oil in the seed bed at time zero, \(F_0\) is the Fourier number that results from a series solution to the original differential equation, and \(D_s\) is the diffusivity in the solid phase. The parameters \(A\) and \(K\) can be determined by linear regression of the experimental data in the diffusion-controlled sector. The computed diffusivities are approximately the same for all of the experimental runs, and do not exhibit the sensitivity to temperature noted by HONG et al. (1990).

**RESULTS AND CONCLUSIONS**

The rate of extraction of EPO from the respective seeds using SC-CO\(_2\) can best be understood by examining its dependence on several processing variables. In Fig. 2, the weight fraction of oil removed from the seeds per total mass of oil collected and CO\(_2\) used is plotted as a func-

<table>
<thead>
<tr>
<th>(u (\text{m/s}))</th>
<th>Temp. (°C)</th>
<th>Pres. (MPa)</th>
<th>(D_s (\text{m}^2/\text{s}))</th>
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<td>50</td>
<td>4.64E-12</td>
</tr>
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</tr>
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<td>4.16E-12</td>
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<td>70</td>
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</tr>
<tr>
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<td>60</td>
<td>50</td>
<td>7.28E-12</td>
</tr>
<tr>
<td>0.00062</td>
<td>60</td>
<td>70</td>
<td>7.6E-12</td>
</tr>
<tr>
<td>0.00125</td>
<td>60</td>
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<td>4.96E-12</td>
</tr>
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<td>0.00188</td>
<td>60</td>
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<td>6.8E-12</td>
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<td>50</td>
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**Fig. 1** - Dimensionless number correlation for obtaining mass transfer coefficient for constant rate sector of extraction curve.
tion of the mass of CO<sub>2</sub> passed through the seed bed for four different extraction pressures at 50°C. It is apparent from the differential extraction yields in Fig. 2, that a CO<sub>2</sub> pressure of 20 MPa will require a considerable larger mass of CO<sub>2</sub> to complete the extraction. Increasing the extraction pressure diminishes the total CO<sub>2</sub> mass requirements for extraction, and as shown for practical purposes, only 0.5 kg of CO<sub>2</sub> is needed to remove most of the EPO from the seed bed. The results shown in Fig. 2 are a direct consequence of the influence of the CO<sub>2</sub> density on EPO solubility in the fluid phase.

As shown in Fig. 3, the oil solubility increases with temperature in °K for each of the fitted lines representing approximating a constant fluid density, similar to the trends recorded by FAT-TORI et al. (1988). The data trend in Fig. 3 clearly shows that fluid density is the controlling variable with respect to oil solubility and that by selecting different pressures and temperatures, the processor can enhance or diminish oil solubility in the SC-CO<sub>2</sub> phase.

It has previously been shown (FAVATI et al., 1991a,b) that the flow rate dependence of seed oil extraction can be expressed as a function of the mass of extract(oil)/mass of extracted seed as a function of extraction time. Such plots clearly indicate the quantity of extract that one can expect as a function of extraction time and the advantages attendant with using higher CO<sub>2</sub> flow rates. The dependence of extraction yield for a given mass of substrate can be normalized by dividing by the mass of seeds in the extraction vessel, i.e., to eliminate the time dependence of oil yield by plotting mass of extract/mass of seeds versus mass of CO<sub>2</sub> used/mass of seed after EGGERS (1989a). This is shown in Fig. 4 for the extraction of EPO at 20 MPa and 313°K for three different CO<sub>2</sub> flow rates (9, 18, 27 g/min). The superimposition of each curve is consistent with previous results noted at 70 MPa and 333°K (FAVATI et al., 1991a,b). Fig. 4 also indicates that the EPO yield is not appreciably effected by the quoted three-fold difference in flow rate, the m<sub>ex</sub>/m<sub>seeds</sub> being equal to 16.9 at approximately m<sub>CO2</sub>/m<sub>seeds</sub> equal to 53.

It is interesting to use the above approach to contrast the difference in extraction efficiency of EPO conducted on
an experimental extractor versus studies performed on systems defined as a "pilot plant" or production plant (EGGERS and SIEVERS, 1989b). This comparison has been made by EGGERS and SIEVERS (1989a,b) and CLAVIER et al. (1995) and we have used the former's data on EPO extraction for comparison purposes. Figure 5 shows our extraction results at 30 MPa and 40°C (circles) with those found by EGGERS and SIEVERS (1989a) for pilot and production scale plant runs. The plots of $m_{ex}/m_{seeds}$ versus $m_{CO_2}/m_{seeds}$ show that there is good agreement between production plant run conducted under the same extraction conditions and our data, up to a $m_{CO_2}/m_{seeds}$ of approximately 30. The data for EGGERS pilot plant run tends to be somewhat lower than our experimental values over most of the encompassed range of $m_{CO_2}/m_{seeds}$ values. We have no rationale for this disagreement, however it is encouraging to note that overall, the results from experimental, pilot, and production plant runs are similar in magnitude, thereby allowing extrapolation of experimental data as a guide for optimizing plant scaleup. However, EGGERS and SIEVERS (1989 b) have cautioned against scaling up at constant specific mass flows ($m_{CO_2}/m_{seeds}$) to rationalize the differences they have observed in going from pilot to production plant operation.

To compare the calculated extraction rate curves, derived from the theory and calculations discussed in the Materials and Methods section with experimental data versus predicted extraction curve.
data, we plotted the mass of oil extracted versus time for extractions at several temperatures and \( \text{CO}_2 \) flow rates. As shown in Fig. 6, the incremental experimental data points have been plotted for three different pressures and the extraction curve calculated from theory superimposed for each set of data corresponding to the different extraction pressures. As indicated in Fig. 6, the agreement between experiment and theory is quite encouraging at all three extraction pressures.

To further test the agreement between theory and our experimental results, we have plotted mass of EPO recovered versus time at both 50° and 60°C for three different extraction pressures as shown in Figs. 7 and 8. Once again, the agreement between the experimental data points, regardless of pressure, and the curve from the extraction model is excellent, lending credence to the concepts and calculations incorporated for describing both the mass transfer and diffusion-controlled sectors of the extraction rate curves. Only in the 30 MPa case in Fig. 7 does there appear to be a small disagreement between theory and experiment, most likely due to experimental error. Results for extractions at 20 MPa were not plotted since at this low extraction pressure the curves are linear and are of limited interest in terms of confirming the agreement between the employed model and experimental data.

Similarly, plots were also constructed for the three different flow rates utilized in extracting the EPO for several combinations of extraction pressure and temperature (Figs. 9-11). For the SFE conducted at a \( \text{CO}_2 \) flow of 9g/min, the agreement between theory and the experimental data is excellent as shown in Fig. 9. Excellent agreement between the extraction data and model curve is also obtained for the extractions conducted at a flow rate of 18g/min of \( \text{CO}_2 \) as noted in Fig. 10. For the final case examined, extraction of EPO using a \( \text{CO}_2 \) flow of 27 g/min, there appears to be an error in the derived experimental data in the diffusion-controlled regime of the extraction conducted at 50 MPa and 60°C, since the calculated extraction curve from the model parallels the results found in the previous figures (i.e., the diffusion controlled sector of the extraction curve for 50 MPa is below that calculated for 70 MPa). Again, the data and model curve

![Fig. 7 - Extraction rate curves for EPO at 50°C as a function of extraction pressure - Experimental data vs. predicted extraction curve.](image1)

![Fig. 8 - Extraction rate curves for EPO at 60°C as a function of extraction pressure - Experimental data vs. predicted extraction curve.](image2)
are in good agreement for the extraction conducted at 70 MPa and 60°C, suggesting that the data determined at 50 MPa and 60°C are too high. This appears to be a good example of how theory can be used to diagnose errors in experimentally-derived data.

In conclusion we feel that the results presented in this study and their interpretation in terms of theory and the parameters that control the SFE of EPO adequately describe the extraction kinetics and its dependance on mass transfer and kinetic factors, such as diffusion. We have verified that CO₂ density plays a predominant role in controlling EPO solubility in SC-CO₂, and that for the range covered by our experimental data, a three-fold increase in CO₂ flow rate does not substantially reduce the EPO concentration solubility or mass recovered in the steady state region of extraction.

Our laboratory extraction results are in general agreement with extraction trends determined under the same pressure and temperature conditions on a pilot or production plant noted by other investigators. However subtle differences do exist as one transcends from the steady state to non-steady state regime of extraction, hence the caution advocated by POLETTO and REVERCHON (1995) should be heeded with respect to scaleup of the extraction.

Utilization of the binary mass transfer model of HONG et al. (1990) has been found to produce excellent agreement with our experimentally-derived extraction curves at different pressures, temperatures and flow rates.
Names are necessary to report factually on available data; however, the USDA neither guarantees nor warrants the standard of the product, and the use of the name by USDA implies no approval of the products to the exclusion of others that may also be suitable.

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