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Catalytic hydrogenation in a packed bed bubble column reactor

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Abstract

The cocurrent downflow contactor reactor (CDCR) has been found to give low mass transfer resistances both in slurry and packed bed catalytic operation. The hydrogenation of propan-2-ol solutions of itaconic acid in the range 100–300 kPa and 20–70°C and of soyabean oil in the range 100–500 kPa and 130–160°C was studied using slurry (5% w/w Pd/C) and packed bed (3% w/w Pd/Al_2O_3 Raschig ring) catalyst. Mass transport and kinetic parameters were evaluated for both operational modes and while the slurry CDCR gave better mass transfer properties than the packed bed CDCR, the latter gave better mass transfer than conventional reactors and superior selectivity to the slurry CDCR. As has been observed with the slurry CDCR, the packed bed CDCR was found to operate under surface reaction rate control with negligible transport resistances. This was particularly evident for soyabean oil hydrogenation, which is well known to be transport controlled in conventional reactors.

Keywords: Downflow bubble column; Multiphase catalysis; Palladium catalysis; Hydrogenation; Mass transfer

1. Introduction

The wide use of bubble column reactors within the chemical industries, in particular for hydrogenation applications [1] indicates the significance of this type of reactor for heterogeneous chemical reactions. Bubble column reactors can be employed as either slurry or packed bed catalytic reactors and depending on the type of application required, can offer distinct advantages or disadvantages in each mode of operation. One of the main advantages of using slurry reactors is to utilise the superior heat and mass transfer attainable with this type of reactor compared with the packed bed equivalent [2,3]. The use of a fixed bed catalyst offers the advantage of eliminating the catalyst separation step associated with slurry reactors, enabling savings in processing time and costs. The use of a cocurrent downflow contactor reactor (CDCR) as a heterogeneous catalytic slurry reactor has allowed the development of a mass transfer efficient chemical reactor, achieving good gas–liquid and liquid–solid mass transfer with negligible resistances to diffusion [4]. The specialised entry zone of the CDCR allows a vigorously agitated gas–liquid dispersion to be formed in the upper section of the column, leading to a large degree of bubble shear and turbulence, thus generating large interfacial areas and volumetric transfer coefficients. Chemical reaction with absorption can be effected with the use of catalysts, both in slurry and fixed bed form. The presence of slurry catalysts does not affect the nature of the bubble dispersions in respect of bubble size. In the case of the fixed bed, the bubbles above the dispersion are of a size and distribution similar to that of an unpacked or slurry column. However, the bed itself does induce coalescence of the 3–4 mm H_2 bubbles giving larger 5–8 mm

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Table 1
Operating conditions for reactors used

Operating conditions	Reactor	
	Glass CDCR	Stainless steel CDCR
Reactants	Itaconic acid/H_2	Soyabean oil/H_2
Total system volume (dm^3)	10	20
Maximum reactor volume (dm^3) i.e., full gas dispersion	2[a]	1[a] and 2[b]
Pressure range		
bara	1–3	1–5
kPa	101.5–303.9	101.5–506.6
Temperature range (K)	293–323	403–433
Reactants	Itaconic acid[c]/H_2	Soyabean oil/H_2

[a] Slurry reactor.
[b] Fixed bed reactor.
[c] Itaconic acid concentration: 50 g/l.

bubbles, which rise through the bed to be broken up by the inlet stream [5]. This study reports the superior mass transfer characteristics of the CDCR using a fixed bed catalyst and, therefore removing the catalyst separation stage.

2. Experimental

To evaluate the reactor performance of the CDCR, two different types of hydrogenation reaction were used. Firstly, Itaconic acid [$HO_2CCH_2C(=CH_2)$-CO_2H] dissolved in propan-2-ol was hydrogenated as a model reaction to study the hydrogenation capabilities of the CDCR in both reactor configurations at 100–300 kPa and 20–70°C. Itaconic acid was chosen for study as it exhibits first order behaviour in hydrogen and is a fast reaction. Slurry and fixed bed reactions were undertaken in a glass CDCR, constructed from standard QVF glass. The column was 1 m in length with an i.d. of 50 mm. A glass water jacket was fitted to the column to maintain isothermal operation. The experimental set-up including an Engelhard Hydrogenation control is shown schematically in a paper by Boyes et al. [4].

The partial hydrogenation of soyabean oil is of particular interest as the reaction in conventional slurry reactors is usually carried out under diffusion controlled conditions and packed bed operation is often avoided due to the fouling of the catalysts caused by the oil [6,7]. The reaction is also well established and poses fairly demanding selectivity requirements. For these reasons the hydrogenation of soyabean oil was chosen for study as a reaction of commercial significance. Triglyceride oils were hydrogenated in a stainless steel column of 0.5 m length and i.d. of 50 mm. A stainless steel heating jacket was fitted around the column using diesel oil as the heating medium. The reaction conditions used for both itaconic acid and soyabean oil hydrogenation are given in Tables 1–3.

For a direct comparison, these two reactions were studied in the packed bed CDCR under the same reaction conditions as for the slurry CDCR, using a

Table 2
Catalyst data

Slurry catalyst – 5% Pd/C – Type 37 (Johnson Matthey)	
Metal loading	5.1 (%w/w)
Metal area	13.9[a] m^2/g
Mean particle size	19 μm
Density	2.15×10^3 kg/m^3
Packed bed catalyst	
Support	α-Al_2O_3
Metal loading	3% (w/w)[b]
Size	19 mm
N_2 surface area	1.34 m^2/g
Pore volume	0.4 cm^3/g
Mean pore diameter	1.2 μm
Packing density	622 g/l

[a] Metal distributed on exterior surface – no pore diffusion effects.
[b] Metal was assumed to be available to reactants at surface concentration of hydrogen since pore diameter is in the macropore range and surface area is very low – typical of α-Al_2O_3.

Table 3
Average initial composition of soyabean oil

Fatty acid	Palimitic, C16 (0)	Stearate, C18 (0)	Oleate, C 18 (1)	Linoleate, C18 (2)	Linolenate, C18 (3)
Percentage	11.0	3.5	25.0	52.5	8.0

5% Pd/C, Type 37 slurry catalyst as supplied by Johnson Matthey and a commercially prepared 3 wt% Pd/Al_2O_3 Raschig ring catalyst. For successful operation of the CDCR, it is essential to enable the vigorously agitated bubble dispersion to be formed and therefore it is necessary to leave a catalyst free space of at least 0.3 m at the top of the reactor; the packed bed reactor length was 1 m.

3. Reaction studies

3.1. Palladium catalysed hydrogenation of itaconic acid

3.1.1. Slurry reactors

The reaction kinetics were evaluated from the classical film model for first order reactions [2,7].

$$\frac{C^*}{R_A} = \frac{1}{k_L a} + \frac{\rho_p d_p}{6W}\left(\frac{1}{k_s} + \frac{1}{k_r}\right) \tag{1}$$

and the liquid–solid mass transfer coefficient (k_s) was calculated from the Frössling equation [8].

$$Sh = 2.0 + 0.76 Re^{1/2} Sc^{1/3}. \tag{2}$$

For the CDCR, the Frössling equation was applied using the main column velocity to determine the Reynolds number (*Re*), which gives conservative (low) values of *Re*, *Sh* and k_s. Consideration of the entry region will yield much larger values of *Re*, *Sh* and k_s.

The volumetric gas–liquid mass transfer coefficient $k_L a$ is obtained by extrapolation and, depending on the experimental scatter can be subject to errors from as little as ±2% to 40–50% inaccuracy especially if $k_L a$ is very large. In this work most of the errors lay in the range 2–10%. The hydrogen solubility in propan-2-ol was determined from cited reference values [9] and those for soyabean oil [10] determined from a well known literature correlation; the diffusivities of hydrogen in propan-2-ol [11] and soyabean oil [10] were calculated using accepted literature sources.

The Reynolds number (*Re*) was calculated using the liquid properties since it has been suggested by Doraiswamy and Sharma [7] that for very turbulent systems and small particles, k_s is independent of particle size in relation to the calculation of *Re*. In this work, use of the limiting form of the Frössling equation

$$Sh = \frac{k_s d_p}{D_A} = 2 \tag{3}$$

gave negative values of $1/k_r$ when applied to Eq. (1). Further attempts to use relationships involving slip velocities [12] gave similar results where $1/k_r$ is determined from the slope of the C^*/R_A versus $1/W$ plot. The value of k_s was calculated using the full form of the Frössling equation (Eq. (2)). It is worth noting that a paper by van der Zon et al. [13] also showed that carbon particles agglomerate at the bubble–liquid interface, causing enhancement of k_s, a result that is relevant to this work, especially with polar solvents.

The apparent energies of activation (E_A) were obtained from the Arrhenius relationship, allowing the identification of the controlling reaction regime [8].

$$k_r = A \exp\left(\frac{-E_A}{RT}\right). \tag{4}$$

3.1.2. Fixed bed reactors

For successful operation of the fixed bed CDCR approximately 300 mm of packing free space at the top of the column is required to enable the jet from the orifice to be created fully and the bubble dispersion to become stabilised.

Itaconic acid when hydrogenated in the slurry and fixed bed CDCR showed first order behaviour in hydrogen and zero order behaviour with respect to itaconic acid and the reaction rate was of the form

$$R_A = k_r P_{H_2}^{1.0} C_{IA}^{0}. \tag{5}$$

The fixed bed CDCR was therefore analysed as follows:

3.1.2.1. Mass transfer coefficients and reaction kinetics The mass transfer coefficients k_La and k_s were evaluated from the Reiss [14] and Hirose [15] correlations (see below).

(i) Gas–liquid mass transfer coefficient k_La: the Reiss [14] correlation employed c.g.s. units, which can be applied to packed columns and was employed to evaluate k_La from the dimensional equation

$$k_La = 5.48 \times 10^{-3} e_l^{0.5} \tag{6}$$

for which

$$e_l = U_l \delta_l, \tag{7}$$

where e_l is the liquid energy dissipation term (g/cm^2 s), U_l is the superficial liquid velocity (cm/s), and δ_l is the liquid phase pressure drop (dyne/cm^3).

The liquid phase pressure drop δ_l can be obtained from the Ergun equation [16] for fluid flow through packed bed columns.

$$\delta_l = \frac{150(1-\varepsilon_B)^2}{\varepsilon_B^3}\frac{U_l\mu}{d_p^2} + \frac{1.75(1-\varepsilon_B)}{\varepsilon_B^3}\frac{U_l^2\rho}{d_p}. \tag{8}$$

(ii) The liquid–solid mass transfer coefficient k_s was calculated using the Hirose equation [15] for liquid to particle mass transfer in a fixed bed reactor with cocurrent gas–liquid downflow, which is a specific variation of the Frössling equation

$$\frac{k_s d_p}{D_A} = 0.8\frac{(Re_L)^{0.5}}{\varepsilon_L}\left(\frac{\mu_L}{\rho_L D_A}\right)^{0.33} \quad \text{for } 20 \le Re \le 200 \tag{9}$$

or

$$\frac{k_s d_p}{D_A} = 0.53\frac{(Re_L)^{0.58}}{\varepsilon_L}\left(\frac{\mu_L}{\rho_L D_A}\right)^{0.33} \quad \text{for } 200 \le Re \le 5000. \tag{10}$$

For the CDCR, Eq. (10) was used as Re>200.

(iii) The surface reaction rate constant k_r for fixed bed reactors was evaluated from the manipulation of simple rate equations as follows,

From gas phase to bulk liquid

$$R_A = k_La(C^* - C_L). \tag{11}$$

From bulk liquid to catalyst surface

$$R_A = k_s a_p(C_L - C_S). \tag{12}$$

Reaction at the catalyst surface

$$R_A = k_r a_p \eta C_S^n, \tag{13}$$

where n=1, for a first order reaction, η=1, for catalyst used.

Rearranging Eqs. (11) and (12), and substituting for C_L,

$$C_S = C^* - R_A\left(\frac{1}{k_La} + \frac{1}{k_s a_p}\right). \tag{14}$$

Rearranging Eq. (13), for first order reactions and catalyst used;

$$k_r = \frac{R_A}{a_p C_S}. \tag{15}$$

From Eqs. (14) and (15) the value of k_r can be obtained; k_La and k_s were evaluated using Eqs. (6) and (10) as already described.

The various kinetic parameters for the catalytic hydrogenation of Itaconic acid in the different types of reactor are shown in Table 4, following calculation from the above equations.

Table 4 shows that relatively high values of k_La were obtained and such values are comparable with an intensively mixed 120 cm^3 stirred tank reactor; as such the results indicate negligible gas–liquid mass transfer resistance. Furthermore, liquid–solid mass transfer resistance (X_{L-S}) was small (≤3%) and the reaction, in both reactors is largely controlled by intrinsic or surface reaction controlled kinetics. The values of the apparent energy of activation confirms this although in the fixed bed CDCR the E_A values at the higher temperature end of the range indicates the contribution of liquid–solid mass transfer resistance, which is perhaps not surprising.

Table 4
Mass transfer and kinetic parameters for Itaconic acid hydrogenation

CDC reactor	Temperature (K)	Pressure (kPa)	k_La (s^{-1})	$k_s \times 10^2$ (m/s)	$k_r \times 10^4$ (m/s)	X_{L-S} (%)	E_A (kJ/mol)
Slurry	293–343	130–270	0.58–3.3	3.6–6.6	11.7–75.9	3.1	41–43
Fixed bed	303–333	130–270	0.21±0.01	0.027	0.04	1.4	33–45

For itaconic acid hydrogenation, the packed bed CDCR was operated with a catalyst free space at the top of the reactor and the bubble dispersion initially was maintained above the packed catalyst particles (H_2 absorption only) and then extended down the full length of the column and maintained within the catalyst particles (H_2 absorption with simultaneous chemical reaction) to allow characterisation of two different operating zones. Low rates of hydrogen uptake were observed, when the dispersion was maintained above the packing, with rates increasing significantly once the dispersion was extended into packing and maintained below the bed of catalyst. The increase in rates of reaction indicates the presence of two separate zones within the reactor. To evaluate further the effect of the catalyst in the packed bed CDCR, calculations were undertaken to obtain the rate of hydrogen uptake using completely mixed (STR) and plugflow models of mass transfer for a full column length bubble dispersion in the absence of catalyst. Calculations were also undertaken to take into account the effect of inert Raschig ring supports in the reactor, with no deposited catalyst. The results are shown in graphical form in Fig. 1 and indicate that the rates of hydrogen uptake do not vary significantly from the rates determined in the absorption only zone and are lower than reaction rates when catalyst is present. The significant effect of the presence of catalyst on reaction rate further emphasises the two distinct zones available within the packed bed CDCR and the expected advantage of extending the dispersion into the bed.

The evaluation of the mass transfer performance of the packed bed CDCR, from transport parameters and reaction kinetics indicated that calculated values of the volumetric gas–liquid mass transfer coefficient (k_La) were found to be in the range 0.21–0.22 s^{-1} and were similar but smaller in value than those obtained in the slurry CDCR for the same reaction. This is possibly due to the fact that the majority of the gas–liquid mass

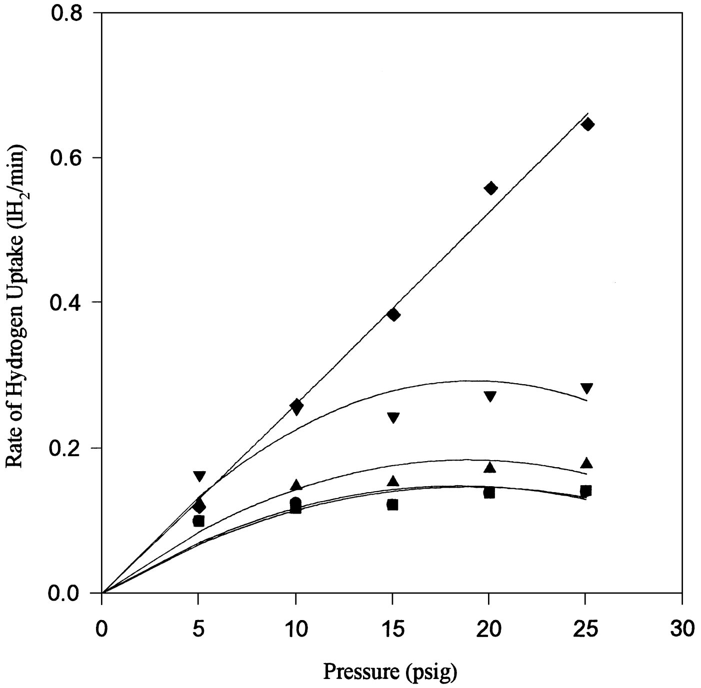

Fig. 1. The effect of catalyst on hydrogen uptake in a packed bed CDCR: (●) experimental hydrogen uptake, dispersion above packing, *no catalyst*; (■) hydrogen uptake calculated using *plugflow model*, dispersion above packing, *no catalyst*; (▲) hydrogen uptake calculated using *CSTR model*, dispersion above packing, *no catalyst*; (▼) experimental hydrogen uptake, dispersion into 0.2 m inert packing with *no catalyst* on surface; (◆) experimental hydrogen uptake, dispersion into 0.2 m packing 3% (w/w) Pd *catalyst* on surface.

Table 5
Mass transport and kinetic parameters for soyabean oil hydrogenation

CDC reactor	Temperature (K)	Pressure (kPa)	k_La (s^{-1})	$k_s \times 10^2$ (m/s)	$k_r \times 10^4$ (m/s)	X_{L-S} (%)	E_A (kJ/mol)
Slurry	403–433	100–400	1.11–3.33	12.3–15	2.1–4.8	0.2–0.3	47–58
Fixed bed	408	100–400	1.0–1.2	0.025	0.009	0.4	40–50

transfer occurs in the upper most turbulent region of the reactor and in the case of the packed bed reactor no catalyst is present in this section, whereas for the slurry reactor, catalyst is present, therefore increasing the degree of gas–liquid mass transfer by enhancement for the slurry CDCR compared to the packed bed CDCR. Resistances to liquid–solid mass transfer were found to be less than 5%, indicating that the reaction was not limited by resistances to liquid–solid mass transfer and largely controlled by the reaction kinetics. Although the packed bed catalyst was of uniform metal dispersion, the pore diameter is large and energy of activation values indicate that pore diffusion is negligible otherwise E_A values would have been only 20–25 kJ/mol [17].

3.2. *Palladium catalysed hydrogenation of triglycerides*

Following the successful study of the hydrogenation of itaconic acid in the slurry and packed bed CDCR, the more complex triglyceride oil hydrogenation was undertaken. Triglycerides (vegetable oils) require partial hydrogenation when used as feed stocks for edible oil manufacture due to the high content of unstable fatty acid chains with three or more carbon–carbon double bonds (polyunsaturates). Polyunsaturates are prone to oxidation under ambient conditions leading to off flavours and polymerisation, resulting in short shelf lives.

An ideal hydrogenation of triglycerides is represented by the sequence [6]

$$\underset{\text{C18:3}}{\text{Linolenic}} \xrightarrow{H_2} \underset{\text{C18:2}}{\text{Linoleic}} \xrightarrow{H_2} \underset{\text{C18:1}}{\text{Oleic}} \xrightarrow{H_2} \underset{\text{C18:0}}{\text{Stearic}}$$

Partial hydrogenation reduces the content of polyunsaturates providing a stable product at room temperature, with an ideal hydrogenation removing linolenic acid (C18:3) and retaining linoleic acid (C18:2), an essential fatty acid for human diet. Minimum formation of stearic acid (C18:0) and trans isomers in the final product are also required. High contents of stearic acid and trans isomers lead to high melting point fats which are both unsuitable for frying oil or soft margarine applications.

Table 6
Reaction selectivities for soyabean oil hydrogenation

CDC reactor	SL_0	SL_n	*trans*-Acids	Iodine value
Slurry	5–84	2.5–4.6	25–29%	73–110
Fixed bed	4–17	2.6–5.6	27–32%	73–110

A palladium catalyst was used to carry out soyabean oil hydrogenation at 1–4 bara and 403–433 K. Reaction orders, reaction kinetics, mass transfer parameters, energies of activation and selectivities were determined for both packed bed and slurry operation and are given in Tables 5 and 6. It should be noted that for packed bed operation, the usual problem of catalyst fouling by the oil was not encountered in this study, however, it is thought that the thin layer of oil covering the catalyst actually protected the catalyst from deactivation by oxidation during shutdown and allowed rapid reactivation each time a new reaction was undertaken by the fresh batch of oil being used. There are no available conversion versus time data but the same batch of catalyst was employed over a 2–3 month period without significant activity change.

The order of reaction in hydrogen was determined by varying hydrogen pressure using the following rate equation; it is assumed that the reaction order with respect to triglyceride is zero, which is generally accepted in the literature. In any case the triglyceride concentration is effectively constant for the conversion range studied (<20%).

$$R_A = k_r p H_2^*, \tag{16}$$

where pH_2^* is the hydrogen pressure.

The reaction order with respect to hydrogen was found to be 1.3 for soyabean oil. The results suggest that the measured rate may be a combination of two simultaneous steps, one being first order in hydrogen and the other being second order in hydrogen. Palladium catalysts were used and it had been reported that palladium and platinum can directly reduce conjugated trienes to monoenes [18,19]. This would account for second step, with a parallel reaction converting the linolenate chains directly to oleate chains.

$$\underset{\mathrm{A}}{\text{Trienes}} \xrightarrow{\mathrm{k_1}} \underset{\mathrm{B}}{\text{Dienes}} \xrightarrow{\mathrm{k_2}} \underset{\mathrm{C}}{\text{Monoenes}} \xrightarrow{\mathrm{k_3}} \underset{\mathrm{D}}{\text{Saturate}}$$

The $k_{\mathrm{L}}a$ values in the slurry and fixed bed CDCR were very comparable, slightly more so than those for itaconic acid (Table 4). However, in the latter case the possible enhancement effect due to carbon agglomeration around the bubbles [13] might be greater than for soyabean oil which is largely hydrocarbon in nature, with a smaller liquid–gas interfacial effect. The magnitude of the $k_{\mathrm{L}}a$ values suggests good gas–liquid contacting with a low gas absorption resistance and the values obtained are greater than those attained in large conventional stirred reactors simply because the CDCR has no significant scale-up problems, while stirred reactors on a large scale cannot employ high stirring speeds and operate with a much lower gas holdup. The liquid–solid transport resistances ($X_{\mathrm{L-S}}$) are small (<1%) and again suggest that the reaction rate is controlled by the intrinsic reaction kinetics. Further confirmation is provided by the apparent energy of activation values, which indicate, very strongly surface reaction rate control and are in agreement with that for olefinic double bond hydrogenation.

The selectivities in comparison with the slurry CDCR were similar but slightly better in the packed bed CDCR for linolenate removal, linoleate retention and lower stearate production. The selectivities of the individual fatty acids were calculated in conjunction with the known product compositions using Eqs. (17) and (18);

$$\underset{\mathrm{A}}{\text{Trienes}} \xrightarrow{k_1} \underset{\mathrm{B}}{\text{Dienes}} \xrightarrow{k_2} \underset{\mathrm{C}}{\text{Monoenes}} \xrightarrow{k_3} \underset{\mathrm{D}}{\text{Saturate}}$$

$$C_{\mathrm{B}} = \frac{C_{\mathrm{A}}}{(K_1 - 1)} + \left(C_{\mathrm{B}}^0 + \frac{C_{\mathrm{A}}^0}{(1 - K_1)}\right)\left(\frac{C_{\mathrm{A}}}{C_{\mathrm{A}}^0}\right) \tag{17}$$

and

$$\begin{aligned} C_{\mathrm{C}} = {} & \frac{K_1 C_{\mathrm{A}}}{(1 - K_1)(1 - K_2)} + \left(\frac{K_1}{K_1 - K_2}\right) \\ & \times \left(\frac{C_{\mathrm{A}}}{(K_1 - 1)} - C_{\mathrm{B}}^0\right)\left(\frac{C_{\mathrm{A}}}{C_{\mathrm{A}}^0}\right)^{K_1} \\ & + \left(C_{\mathrm{C}}^0 + \frac{K_1 C_{\mathrm{A}}^0}{(1-K_2)(K_1-K_2)} + \frac{K_1 C_{\mathrm{B}}^0}{(K_1-K_2)}\right)\left(\frac{C_{\mathrm{A}}}{C_{\mathrm{A}}^0}\right)^{K_2}, \end{aligned} \tag{18}$$

where C_{A}^0, C_{B}^0, C_{C}^0 are initial concentrations and C_{A}, C_{B}, C_{C} are final concentrations of triene, diene and monoenes, respectively. The above equations were solved using the half integral method [20]. Selectivities SL_n and SL_0 were given by the following:

$$\mathrm{SL}_n = \frac{k_1}{k_2} \quad \text{and} \quad \mathrm{SL}_0 = \frac{k_3}{k_2}, \quad \text{where } K_1 = \frac{k_2}{k_1} \quad \text{and } K_2 = \frac{k_3}{k_1}.$$

It can be concluded that the CDCR is a device of very high mass transfer efficiency and when used as a chemical reactor gives reaction rates which are almost entirely reaction rate controlled. The application of mass transfer advantages achieved with the slurry CDCR to a packed bed reactor has been achieved with the packed bed CDCR also able to achieve effective gas–liquid and liquid–solid mass transfer and reaction occurring mainly under surface reaction rate controlled conditions with negligible resistances to mass transfer. This is particularly true for the hydrogenation of soyabean oil, a reaction which is normally carried out under diffusion controlled conditions in conventional reactors, but under almost diffusion free conditions in the CDCR.

4. Nomenclature

A	Arrhenius constant (dimensionless)
a_{p}	catalyst interfacial area ($\mathrm{m}^2/\mathrm{m}^3$)
a	gas–liquid interfacial area per unit volume dispersion ($\mathrm{m}^2/\mathrm{m}^3$)

C^*	equilibrium concentration of gas at the gas–liquid interface (mol/m^3)
C_L	concentration of hydrogen in the bulk liquid (mol/m^3)
C_s	concentration of hydrogen at the catalyst surface (mol/m^3)
C_{Cat}	concentration of catalyst (mol/m^3)
C_{IA}	concentration of itaconic acid (mol/m^3)
d_p	diameter of catalyst particle (m)
D_A	diffusivity of hydrogen (m$^{2/}$s)
e_l	liquid energy dissipation (g/cm^{2}s)
E_A	energy of activation (kcal/mol) or (kJ/mol)
k_La	volumetric gas–liquid mass transfer coefficient (s^{-1})
k_s	liquid–solid mass transfer coefficient (m/s)
k_r	surface reaction rate constant (m/s)
n	reaction order (dimensionless)
P	reactor pressure (kPa)
P_{H_2}	hydrogen pressure (kPa)
R_A	reaction rate per unit volume of dispersion (mol/m^3 s)
R	gas constant (J/mol K)
Re	Reynolds number (dimensionless)
Sc	Schmidt number (dimensionless)
Sh	Sherwood number (dimensionless)
T	temperature (K or °C)
U_l	superficial liquid velocity (m/s)
W	catalyst loading (kg/m^3)
X_{L-S}	% liquid–solid mass transfer resistances (%)
ε_B	bed voidage (dimensionless)
ε_L	liquid voidage (dimensionless)
δ_L	pressure drop per unit bed height in a fixed bed reactor (dyne/cm^3)
ρ_L	density of the liquid (kg/m^3)
ρ_p	density of the catalyst particle (kg/m^3)
μ_L	viscosity of the liquid (kg/m s^2)
η	catalyst effectiveness factor (dimensionless)

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