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Performance of bipolar trickle reactors

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The performance of the bipolar trickle reactor has been studied using the electrochemical tracer technique. The theoretical equations for a semi-infinite dispersion model have been fitted to the experimental responses for the reactor with and without electrochemical reaction. Hydrodynamic parameters and reaction rate constants for copper deposition as functions of both the film Reynolds number and the dimensions of the bipolar trickle reactor have been derived and are interpreted in this paper.

List of Symbols

(Bo)	Bodenstein number (uL_p/D)
C	amplitude of the response curve (dimensionless)
C^0	area under the response curve (mol cm^{-3} s)
D	dispersion coefficient (cm^2 s^{-1})
h	film thickness (cm)
k/h	first order reaction rate constant (s^{-1})
L	length of the reactor (cm)
L_p	length of the ring (cm)
n_r	number of rings in a single layer
(Pe)	Peclét number (uL/D)
$(Re)_f$	film Reynolds number $\dfrac{v}{2n_r(r_o + r_i)\nu}$
r_i, r_o	inner and outer radii of the ring (cm)
t	time (s)
u	mean liquid velocity (cm s^{-1})
v	volumetric liquid velocity (cm^3 s^{-1})
τ	residence time (s)
ν	kinematic viscosity (cm^2 s^{-1})

1. Introduction

The application to the bipolar trickle tower of an electrochemical tracer technique has been described in the previous papers [2, 3]. It has been shown that the behaviour in the absence of reaction is best interpreted in terms of a model of a fast and slow-moving phase with dispersion in the fast phase and exchange of material between both phases. In the presence of a sufficiently high field, reaction takes place predominantly in the fast phase. The bipolar character of the reactors is adequately described by a zoned reactor model. The effect of changes in the boundary conditions is small, that is, the behaviour of the reactor is dominated by its internal characteristics.

It has been shown that the reactor can nevertheless be adequately described in terms of a semi-infinite dispersion model, that is in terms of an average Peclét number, residence time and reaction rate constant. The variation of these parameters (determined by this model) as a function of the experimental variables (flow rate, reactor length and diameter, ring size and presence or absence of cocurrent gas flow) is analysed in this paper.

2. Experimental

The experimental procedure and data analysis have been described in the previous two papers [1, 2]. Experiments on the effects of bed length were performed in reactors 4 cm in diameter. The lengths studied were 10.3, 15.5 and 19.8 cm, namely 14, 21 and 26 layers, each containing 30 graphite Raschig rings of $\frac{1}{4}$ inch diameter. Experiments with cocurrent nitrogen flow (140 cm^3 min^{-1}) were performed on inactive beds of length 10.3, 15.5 and 25 cm (33 layers).

The effect of the bed diameter was studied on columns of 10.3 cm length packed with 14 layers of $\frac{1}{4}$ inch Raschig rings. The diameters of the beds were 4 and 4.88 cm, and these contained 30 and 43 rings respectively in each layer. The range of liquid flow rates investigated was 200–600 cm^3

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min^{-1}; measurements were performed for increments in volumetric liquid velocity of $50\,\text{cm}^3\,\text{min}^{-1}$.

Experiments with different particle sizes were made on inactive beds, having a diameter of 6.2 cm. Each bed contained 8 layers with either 71 $\frac{1}{4}$ inch rings or 17 $\frac{1}{2}$ inch rings; the respective lengths were 5.5 and 9.5 cm. The range of liquid flow rate investigated was $400–750\,\text{cm}^3\,\text{min}^{-1}$, with increments of $50\,\text{cm}^3\,\text{min}^{-1}$.

3. Results and discussion

The form of typical transient responses at the outlet of the reactor to a delta function perturbation at the inlet has been illustrated in Fig. 4 of the preceeding paper [2]. All the data discussed in this paper have been derived by fitting in the time domain to the semi-infinite dispersion model, i.e. to:

$$C = \frac{C^0}{2\tau}\left[\frac{(Pe)}{\pi(t/\tau)^3}\right]^{1/2} \exp\left\{\left[\frac{-(Pe)(1-t/\tau)^2}{4(t/\tau)}\right] - \frac{k}{h}t\right\} \tag{1}$$

or when

$$k = 0$$

to

$$C = \frac{C^0}{2\tau}\left[\frac{(Pe)}{\pi(t/\tau)^3}\right]^{1/2} \exp\left[\frac{-(Pe)(1-t/\tau)^2}{4(t/\tau)}\right]. \tag{2}$$

One of the most interesting sets of results is the relationship between the film Reynolds number $(Re)_f$ and the Bodenstein number (Bo) (Fig. 1). This number

$$(Bo) = \frac{(Pe)}{L}L_p = \frac{uL_p}{D} \tag{3}$$

represents mixing over the length of a single Raschig ring. Since there is no effect of bed length and, bearing in mind that the tower has a uniform geometry (layers of rings packed between the polyester nets), it appears that mixing is initiated by each mesh and dies out in the film. (Bo) increases with $(Re)_f$ but the slope is small showing that although dispersion increases less rapidly than convection, there is almost complete compensation of the two opposing effects (Equation 3).

Fig. 2 shows the relation of (Bo) with $(Re)_f$ for different column lengths ($\frac{1}{4}$ inch Raschig rings) measured with cocurrent nitrogen flow (flow rate

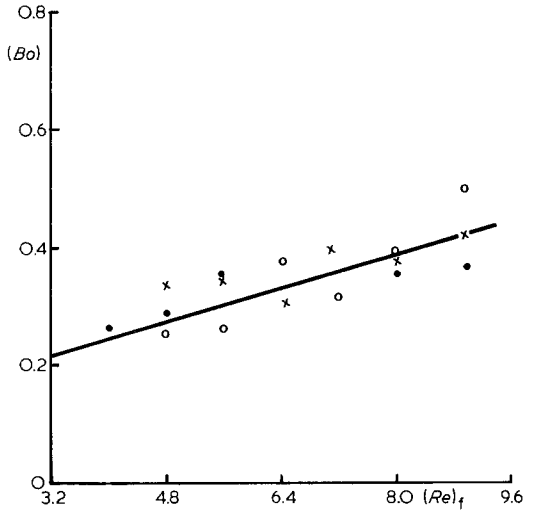


Fig. 1. (Bo)–$(Re)_f$ relationship. Ring size, $\frac{1}{4}$ inch; diameter of the reactor, 4 cm. Lengths of the reactors: ○ 10.3 cm, × 15.5 cm, • 19.8 cm.

$140\,\text{cm}^3\,\text{min}^{-1}$). The Bodenstein number now remains independent of $(Re)_f$ for the flow region investigated. The effect of the gas flow evidently is to enhance dispersion so that it now increases as rapidly as convection. This gas flow therefore increases the mixing initiated by a mesh.

Fig. 3 shows the variation of the residence time with $(Re)_f$. It is noteworthy that the residence time does not show a linear relationship with column length so that the mean liquid velocity $u = L/\tau$ must increase with column length. There-

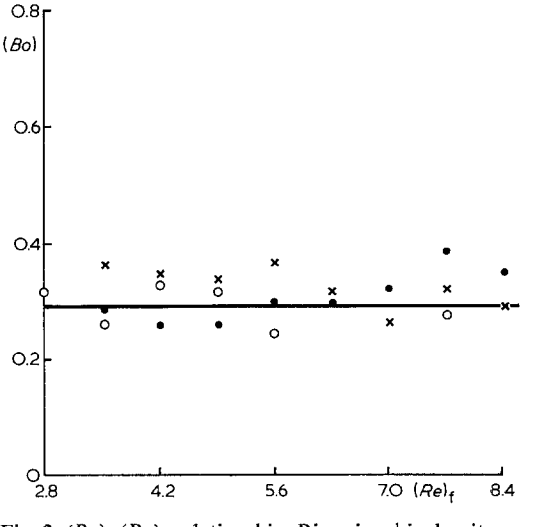


Fig. 2. (Bo)–$(Re)_f$ relationship. Ring size, $\frac{1}{4}$ inch; nitrogen flow rate, 140 ml min^{-1}. Lengths of the reactors: ○ 10.3 cm, × 15.5 cm, • 19.8 cm.

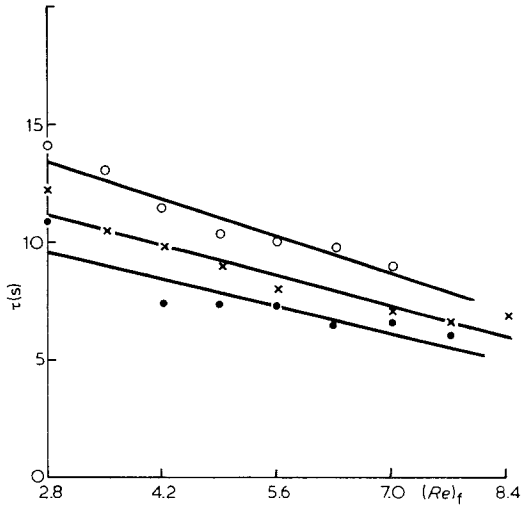


Fig. 3. τ–$(Re)_f$ relationship: size of the rings, $\frac{1}{4}$ inch; diameter of the reactor, 4 cm; nitrogen flow rate, 140 ml min^{-1}. Lengths of the reactors: ○ 10.3 cm, × 15.5 cm, • 19.8 cm.

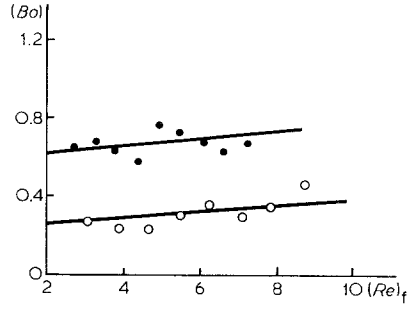


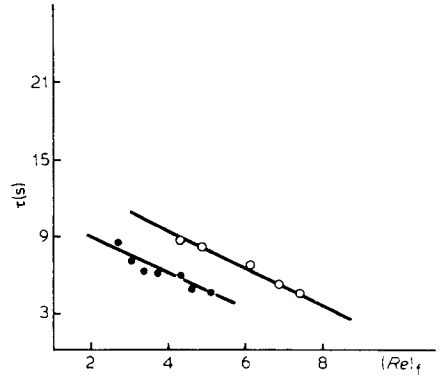
Fig. 4. (Bo)–$(Re)_f$ relationship. Size of the rings, $\frac{1}{4}$ inch. Diameter of the reactors: • 4.88 cm, ○ 4 cm; length of the reactors, 10.3 cm.

fore the dispersion coefficient deduced from (Pe) and u must increase with column length and this is consistent with the observation [5–10] that D is not constant along the length of short reactors such as those used in this study. However, the effect may be due in part to changes in the proportion of the stagnant zone which have been shown to be present in the reactor [2] and it must be borne in mind that the dispersion model is simply a representation of several mixing processes.

Fig. 4 indicates that better mixing occurs in reactors of smaller diameter. This result is in line with previous parametric measurements [4], which have shown that bipolar reactors with smaller diameters give better current efficiencies for the formation of HBrO from sodium bromide. This reaction is dependent on anolyte and catholyte mixing.

Fig. 5 illustrates the effect of particle size on the (Bo)–$(Re)_f$ and τ–$(Re)_f$ relationships with a constant number of layers and hence different lengths of column. The reactor with longer rings and longer column lengths shows higher (Bo) values, but at the same time a higher dispersion coefficient, implying that while the source of mixing is the nets, dispersion nevertheless still occurs in the film falling down the rings.

Fig. 6 shows the variation of the rate constant for copper deposition with $(Re)_f$ for the various column lengths. This relationship is reminiscent of the (Bo)–$(Re)_f$ relationship and confirms that the reaction is essentially mass transfer controlled over an 'active' part of the rings. The slope of the rate constant–$(Re)_f$ plot is higher for measurements with cocurrent gas flow and this is again parallel to



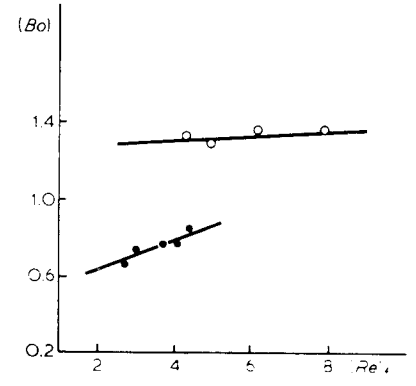


Fig. 5. Variation of τ and (Bo) with the ring size. Number of layers, 8; diameter of the reactors, 6.2 cm; ○ 9.5 cm long reactor packed with 17 $\frac{1}{2}$ inch Raschig rings in a single layer; • 5.5 cm long reactor packed with 71 $\frac{1}{4}$ inch Raschig rings in a single layer.

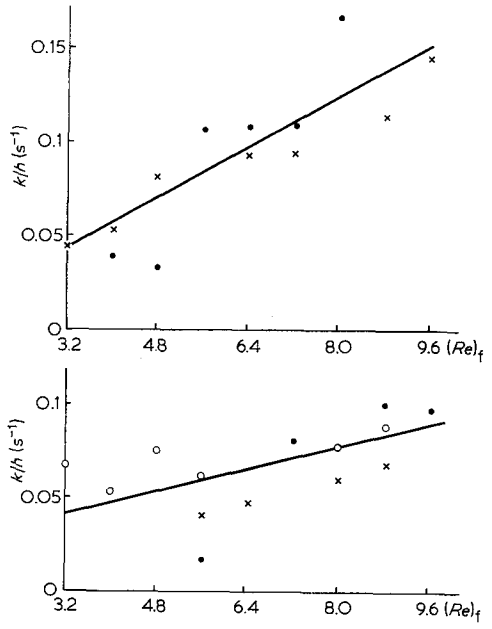


Fig. 6. Variation of reaction rate constant with $(Re)_f$. Size of the rings, $\frac{1}{4}$ inch; diameter of the reactors, 4 cm. Lengths of the reactors: • 10.3 cm, × 15.5 cm, ○ 19.8 cm. The data in the upper figure were determined with a cocurrent nitrogen flow at a rate 140 $cm^3 min^{-1}$.

the increase of the dispersion in the reactor. Both phenomena must therefore have a common cause, namely the increased flow fluctuations (rippling) in the film.

Despite the fact that there are a number of reviews and papers [5, 8–10] which describe the flow conditions in packed beds and in packed beds with trickle flow [4, 6, 11–24], the comparison with published data is not straightforward. Most trickle towers applied in chemical engineering contain randomly packed beds of Raschig rings whereas the system investigated here contains an ordered packing between net separators. The tailing reported by most investigators has the effect of producing relatively high scatter in the derived data. The value of (Pe) or (Bo) has been reported to show a slight linear increase with Reynolds number* and to lie in the range of the experimental results found in this study [6, 7, 11, 17, 20, 22].

4. Conclusions

The experiments have shown that in the flow range investigated here mixing originates at the mesh separators and persists over the lengths of the Raschig rings. Higher Reynolds numbers, smaller column diameters, longer columns and cocurrent gas flow favour mixing in the bipolar trickle reactor. Although the bipolar reactor is characterized by a non-linear distribution of voltage and reaction (including mass transfer controlled regions, activation controlled regions and regions without reaction), the average rate constant k/h is linearly dependent on $(Re)_f$ and this suggests that mass transfer dominates the measured performance for reactions such as copper deposition. The simple semi-infinite dispersion model gives a reasonable explanation for the performance of the reactor in the presence of a first order reaction.

* The Reynolds number has often been expressed in terms of the mass flow rate across the cross-section of the empty reactor.

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