

# *Development of the Eco-Cascade-cell reactor\**

F. C. WALSH

*Wolfson Centre for Electrochemical Science, Southampton University, Southampton, UK*

N. A. GARDNER

*Ecological Engineering Limited, Macclesfield, Cheshire, UK*

D. R. GABE

*Department of Materials Engineering and Design, Loughborough University of Technology, Leicestershire, UK*

Received 29 December 1980; revised 7 August 1981

The development of multicompartment rotating cylinder electrode reactors for the removal of metal from aqueous solutions is described. Such reactors approximate to a cascade of continuously stirred tank reactors and the results illustrate that, for electrodeposition of copper powder from acid sulphate solutions, high overall conversions (about 98%) may be realised, with low exit metal concentrations (about 1 mg dm<sup>-3</sup>) and reasonable current efficiencies (65-87%).

## Nomenclature

$A$	electroactive surface area (cm <sup>2</sup> )
$C_{in}$	inlet concentration of metal (mg dm <sup>-3</sup> )
$C_{out}$	outlet concentration of metal (mg dm <sup>-3</sup> )
$C_{reactor}$	reactor concentration of metal (mg dm <sup>-3</sup> )
$f_R$	fractional conversion
$(f_R)_n$	overall fractional conversion
$F$	Faraday = 96 500 (C mol <sup>-1</sup> )
$I_L$	limiting current (A)
$K_L$	mass transfer coefficient (cm s <sup>-1</sup> )
$m$	weight of metal (g)
$M$	molecular weight of metal
$n$	number of reactor elements in the cascade
$N$	volumetric flow rate (cm <sup>3</sup> s <sup>-1</sup> )
$z$	electron change
$dm/dt$	rate of removal of metal (g s <sup>-1</sup> )

## 1. Introduction

The Eco-Cell has been developed [1] and patented [2, 3] as a compact, high rate, selective electro-

chemical reactor capable of automatic and continuous removal of metal from aqueous solutions. Typical applications for the reactor are found in hydrometallurgical production of metals, effluent treatment [4] and the removal of metal from various industrial process solutions.

The Eco-Cell concept (Fig. 1) comprises a rotating cylinder cathode inside a concentric cylindrical anode, the electrodes normally being separated by a suitably supported ion exchange membrane, typically in the form of a cloth. This membrane facilitates choice of an anode/anolyte system, somewhat independently of the cathode/catholyte, which can serve to prevent mixing of the anode and cathode reactants and products. This minimizes anode corrosion due to aggressive catholyte species. In many cases, oxygen evolution is the main anode reaction and hydrogen evolution is a secondary cathodic process; the membrane then separates a potentially hazardous and inconvenient combination of gases.

Metal is deposited on the circumferential surface of the cylinder, under mass transport control. The resultant dendritic, powdery product is continuously dislodged by a scraping mechanism

\* This paper was presented, in part, at the *Electrochemical Reaction Engineering Symposium*, Southampton University, April (1979).

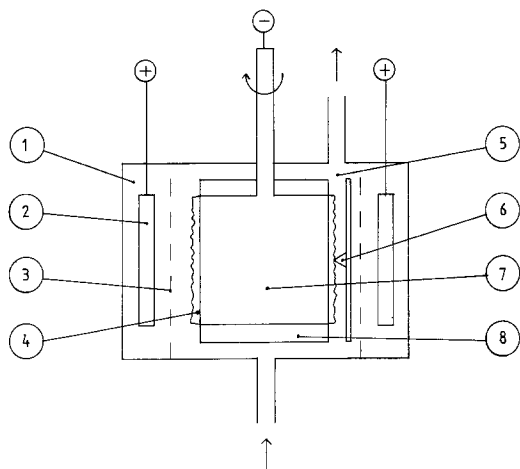


Fig. 1. Schematic diagram of an Eco-Cell. 1. Anolyte, 2. anode, 3. membrane, 4. metal powder deposit, 5. catholyte, 6. scraping mechanism, 7. rotating cylinder cathode, 8. insulating end discs. Inlet: metal-containing catholyte liquor. Outlet: metal powder fluidized in treated, metal depleted liquor.

and fluidized in the turbulent flow around the cylinder, to emerge in the metal ion depleted outlet solution.

Conventional solid-liquid separation processes, including settlement, filtration, centrifuging and hydrocycloning may then be applied to recover the metal for sale or re-use.

The high mass transport provided by turbulent flow around the rough rotating cylinder electrode [5] enables reasonably high useful currents and recovery rates of metal to be achieved, even from dilute solutions [1]. The uniform electrode potential experienced by the cylinder facilitates constant potential control resulting in a relatively high current efficiency for the selective deposition [6, 7] of a metal from solution.

Due to its inherently high turbulence, the Eco-Cell approximates to an electrochemical, continuously stirred tank reactor (CSTR). The single pass cell shown in Fig. 2 operating under mass transport control will have catholyte terminal metal concentrations,  $C_{in}$  and  $C_{out}$  related by the fractional conversion,  $f_R$

$$f_R = \frac{C_{in} - C_{out}}{C_{in}} = 1 - \frac{1}{1 + K_L A / N} \quad (1)$$

where  $K_L$  is the mass transfer coefficient defined by

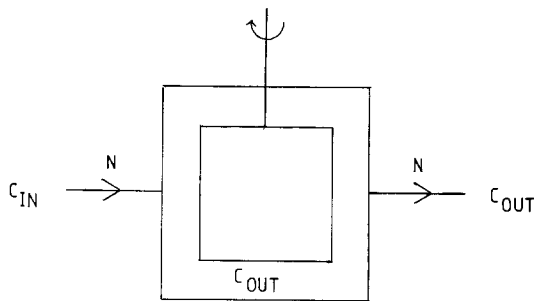


Fig. 2. The Eco-Cell as a single stage CSTR with inlet and outlet metal concentrations  $C_{in}$  and  $C_{out}$ , and a volumetric flow rate  $N$ .

$$K_L = \frac{I_L}{AzFC_{reactor}} \quad (2)$$

For the CSTR model, the assumption of perfect mixing renders the reactor concentration identical to  $C_{out}$ .  $K_L$  typically lies within the range 0.1 to  $1.0 \text{ cm s}^{-1}$  for practical Eco-Cells, and for given electrolyte conditions is dependent upon the rotational velocity and diameter of the cylinder [1].

An important characteristic of a highly turbulent, rotating cylinder electrode reactor (RCER) is that the value of the mass transfer coefficient is remarkably insensitive to axial flow through the reactor, i.e.  $K_L$  is independent of  $N$ .  $K_L A / N$  may be regarded as a scale-up factor; for a given flow rate, the reactor conversion may be enhanced by increasing  $K_L$ ,  $A$  or both. There are obvious engineering and cost limitations here, however, especially in view of the fact that the rotational power requirement for the cylinder increases approximately with the cube of the rotational velocity.

## 2. The concept of a cascade Eco-Cell

An alternative approach to the realization of a high overall conversion is to employ an array of reactors in hydraulic series (Fig. 3); such a cascade allows a stepwise reduction of the metal concentration over each reactor element. If the  $n$  elements in the cascade are identical (which could tend to minimize investment costs), and each operates under mass transport control (which results in maximum duty for given conditions), the terminal concentrations are related by the overall

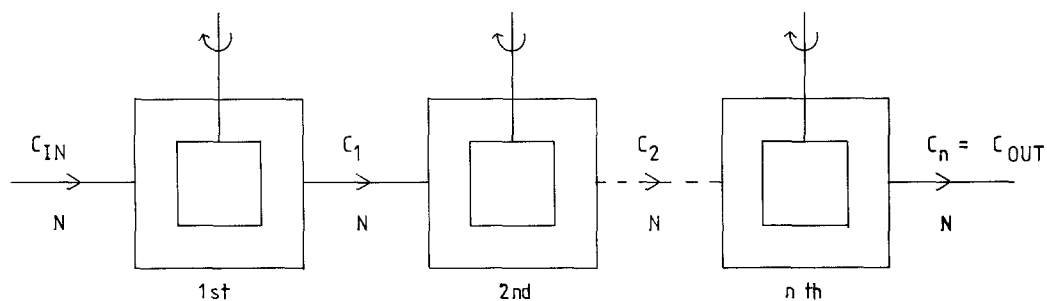


Fig. 3. A cascade of  $n$  identical Eco-Cells in hydraulic series.

fractional conversion  $(f_R)_n$

$$(f_R)_n = \frac{C_{in} - C_{out}}{C_{in}} = 1 - \frac{1}{(1 + K_L A/N)^n} \quad (3)$$

Furthermore, in the case of rotating electrode reactors, the possibility occurs of assembling all the reactor elements on a common shaft for engineering convenience, minimization of space, and to lower investment and maintenance costs. Such an arrangement has been considered for the rotor electrodes of the 'electrochemical pump cell' developed at Southampton University [8], and a multiplicity of disc electrodes on a common rotating shaft is commonly employed in the photographic industry for cathodic recovery of silver from fixing solutions [9, 10], although it must be noted that these assemblies do not function as cascades. A further possibility, in the case of

rotating cylinder reactors, is to employ a single extended cylindrical cathode, inside one reactor body, and to subdivide this into identical elements by means of regularly spaced internal baffles (Fig. 4).

Division of the cascade reactor in this manner results in a versatile system offering considerable scope for alteration of the number of compartments and the extent of each one. This construction will approximate to a series of CSTRs as long as the cylinder rotational velocity is sufficiently high (maintaining effective stirring in each element), the axial flow rate is not excessive and the baffle to cylinder aperture is sufficiently small (which discourages bypassing).

The concept of an idealized Eco-Cascade-Cell operation is illustrated in Table 1, where an inlet metal (copper) concentration of  $100 \text{ mg dm}^{-3}$  is reduced stepwise to an outlet concentration of  $1.6 \text{ mg dm}^{-3}$ . Here, a conversion  $f_R$  of 0.5 is

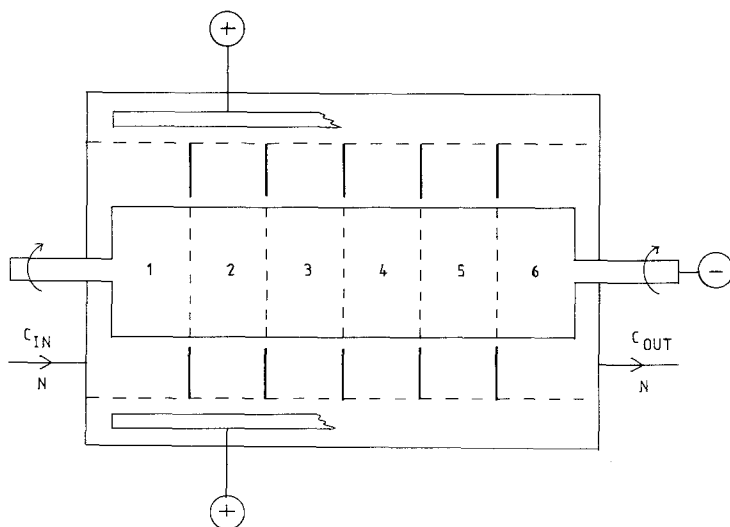


Fig. 4. Practical concept for a six-compartment cascade reactor with division of the anolyte and catholyte.

Table 1. Performance of hypothetical Eco-Cascade-Cell. Copper deposition under limiting current conditions, with a fractional conversion of 0.5 in each compartment, and a flow rate of  $1000 \text{ cm}^3 \text{ s}^{-1}$ . Cathode current efficiency 100%. Total current 300 A

Compartment No.	$C_{\text{in}}$ (mg dm $^{-3}$ )	$C_{\text{out}}$ (mg dm $^{-3}$ )	$\Delta C$ (mg dm $^{-3}$ )	$-dm/dt$ (g h $^{-1}$ )	$(f_R)_n$	$I$ (A)
1	100	50	50	180	0.500	151.88
2	50	25	25	90	0.750	75.94
3	25	12.5	12.5	45	0.875	37.97
4	12.5	6.25	6.25	22.5	0.938	18.99
5	6.25	3.125	3.125	11.25	0.969	9.49
6	3.125	1.5625	1.5625	5.625	0.984	4.75
Overall	100	1.56	98.44	354.375	0.984	300

assumed, the concentration thus halving over each element. If the electrolyte flow rate is  $1000 \text{ cm}^3 \text{ s}^{-1}$ , the individual compartment currents may be calculated by means of the following analysis.

A mass balance over each element yields the removal rate of copper,

$$-\frac{dm}{dt} = 10^{-6}N(C_{\text{in}} - C_{\text{out}}) \quad (4)$$

which will have units  $\text{g s}^{-1}$  if  $N$  is expressed as  $\text{cm}^3 \text{ s}^{-1}$  and  $C$  as  $\text{mg dm}^{-3}$ . The rate of copper removal may be related to the current,  $I$ , by application of Faraday's law and the assumption of 100% current efficiency:

$$-\frac{dm}{dt} = \frac{IM}{zF}, \quad (5)$$

For copper,  $z = 2$  and  $M = 63.54$  giving

$$-\frac{dm}{dt} = 3.292 \times 10^{-4}I. \quad (6)$$

Coupling Equations 4 and 6,

$$I = \frac{N(C_{\text{in}} - C_{\text{out}})}{329.2}. \quad (7)$$

Equation 1 shows that in order to attain a fractional conversion of 0.5 with a flow rate of  $1000 \text{ cm}^3 \text{ s}^{-1}$  (as in Table 1) the factor  $KA$  must equal  $1000 \text{ cm}^3 \text{ s}^{-1}$ , and this must be achieved by suitable choice of electrode size and rotational velocity [1].

It has already been noted that for maximum duty, each individual compartment of the Eco-Cascade-Cell should function under limiting current conditions. An additional consideration is

that electrodeposition must be maintained at or near the limiting current in order to produce metal in powder form [11], rather than as a hard, compact and adherent electroplate. The latter might present removal problems and a lowering of mass transport due to the less favourable hydrodynamics and lower electrochemically active surface area.

The most satisfactory method of supplying electrical power would undoubtedly be achieved through a potentiostatic approach, where the electrode potential in each compartment could be automatically controlled at a preselected value, regardless of metal concentration. This is rather impractical, however, as a separate control circuit would be necessary for each compartment. Moreover, each compartment would require its own independent, electrically insulated cathode and power feed brush, or separate anodes would be required in each compartment.

To overcome these problems, alternative approaches have been adopted, utilizing simple, moderately priced, constant current transformer/rectifier power supplies. In the case of divided cell reactors (Fig. 4), it has been found possible to tailor the anode size, position and effective surface area to achieve substantially uniform electrode potential in each compartment, with a suitable current profile over the reactor length. This has been achieved by modifying the position of the anode along the reactor length, the length of anode, and by strategic masking of the anode. The number and nature of the variables involved in the resultant current density profile here, e.g. anode geometry, membrane resistance, intercompartmental current leakage via baffles, anolyte and

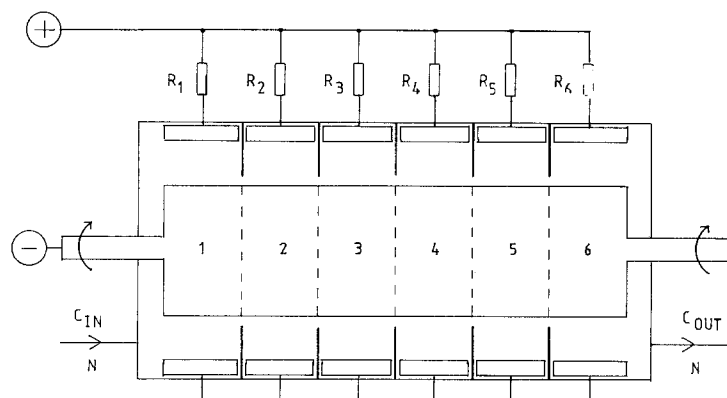


Fig. 5. Undivided cascade reactor with a separate pair of anodes in each compartment. Resistances  $R_6 > R_5 > R_4 > R_3 > R_2 > R_1$ .

catholyte resistance, and effect of gas evolution, is such that an empirical approach has been adopted, guided by experience. For undivided cell reactors (Fig. 5), individual anodes have been provided in each compartment, and a suitable resistor network has been employed to achieve the desired current density profile over the reactor. This second approach, while more costly in terms of power, anode material, additional electronic components and wiring, results in a versatile reactor, capable of being easily modified to suit different needs.

### 3. Experimental development

Table 2 briefly describes three main stages in the development of working Eco-Cascade-Cells. The first reactor constructed was a laboratory model used to examine the general aspects of the operation. Fabrication was largely in acrylic plastic for ease of assembly and optical transparency. The stainless steel cylinder rotated about a horizontal axis in a sealed catholyte compartment. Nine removable baffle plates served to divide the reactor into ten similar compartments. Provision

was made in each of the reactor elements for a separate sampling point and a reference electrode probe for electrode potential measurement. The anolyte compartments were open to the atmosphere, and utilized a static head of sodium hydroxide, with nickel expanded mesh anodes. This construction facilitated trials involving anode positioning and masking. A fluorocarbon cloth cation exchange membrane separated the catholyte and anolyte. The catholyte employed contained copper (typically  $100 \text{ mg dm}^{-3}$ ) in acid sulphate ( $1 \text{ N H}_2\text{SO}_4$ ) solutions at  $60^\circ \text{C}$  (with or without organic additions) at a flow rate of  $100 \text{ cm}^3 \text{ s}^{-1}$ . Table 3 shows four sets of results from this reactor using an inlet concentration of about  $100 \text{ mg dm}^{-3}$  copper being reduced stepwise over the ten elements to exit at about  $1.0 \text{ mg dm}^{-3}$ . This corresponds to an overall conversion of more than 0.97. It is interesting to note that the outlet concentration was identical to that in the ultimate compartment, in accordance with the CSTR model. The overall current efficiency for copper was about 68–81% i.e. some hydrogen evolution had occurred as a secondary reaction.

Table 2. Comparison of developed Eco-Cascade-Cell reactors. Current ratings are nominal

Model	Current rating (A)	n	Cylinder details			
			Diameter (cm)	Length (cm)	Rotation rate (rpm)	Compartment area (cm <sup>2</sup> )
1. Laboratory	100	10	7.6	100	2000	215
2. Commercial	1000	12	32.4	288	860	2036
3. Development/ commercial	500	6	30.6	100	730	1387

Table 3. Data for the 100 A Cascade Cell with potential monitoring

Compartment	I			II			III			IV		
	Copper concentration (mg dm <sup>-3</sup> )	<i>f<sub>R</sub></i>	Copper concentration (mg dm <sup>-3</sup> )	<i>f<sub>R</sub></i>	Copper concentration (mg dm <sup>-3</sup> )	<i>f<sub>R</sub></i>	Copper concentration (mg dm <sup>-3</sup> )	<i>f<sub>R</sub></i>	<i>-E*</i> (mV/SCE)	Copper concentration (mg dm <sup>-3</sup> )	<i>f<sub>R</sub></i>	<i>-E*</i> (mV/SCE)
Inlet	92.4	—	102.3	—	104.9	—	107.6	—	—	—	—	—
1	68.0	0.26	79.1	0.23	66.2	0.37	73.2	0.32	312 ± 43	42.4	0.42	293 ± 5
2	54.4	0.20	55.7	0.30	39.3	0.41	42.4	0.42	419 ± 90	21.2	0.50	311 ± 53
3	38.3	0.29	35.9	0.36	21.8	0.45	21.2	0.50	551 ± 91	14.2	0.33	546 ± 79
4	33.3	0.13	21.1	0.41	13.7	0.37	14.2	0.33	518 ± 109	11.2	0.21	380 ± 104
5	23.5	0.39	11.0	0.48	10.5	0.23	11.2	0.21	425 ± 100	6.5	0.42	396 ± 130
6	17.3	0.26	5.3	0.52	5.1	0.51	6.5	0.42	509 ± 106	3.9	0.40	446 ± 145
7	11.1	0.36	3.2	0.40	5.3	—	3.9	0.40	533 ± 92	2.8	0.28	511 ± 110
8	8.1	0.37	2.0	0.37	2.0	0.62	2.8	0.28	579 ± 46	2.0	0.29	604 ± 42
9	4.9	0.39	1.4	0.28	1.5	0.25	2.0	0.29	592 ± 29	1.5	0.25	624 ± 15
10 and outlet	3.1	0.37	1.15	0.20	1.2	0.20	1.5	0.25	563 ± 43	—	—	575 ± 32
<i>Conditions</i>												
Flow rate <i>N</i> (cm <sup>3</sup> s <sup>-1</sup> )	100		100		100		100				100	
Total current (A)	40		40		40		40				40	
Temperature (°C)	30		60		60		60				60	
Current efficiency (%)	68		77		79		81				81	
Overall fractional conversion ( <i>f<sub>R</sub></i> ) <sub>n</sub>	0.966		0.989		0.989		0.986				0.986	

\* Referring to the potential between a SCE reference electrode probe and the rotating cylinder shaft near Compartment 1.

Table 4. Data for the 1 kA commercial cascade reactor

Compartment No.	Copper concentration (mg dm <sup>-3</sup> )	Fractional conversion, $f_R$
Inlet	88.8	—
1	73.8	0.169
2	52.5	0.289
3	40.0	0.238
4	29.5	0.263
5	21.0	0.288
6	14.2	0.324
7	9.75	0.313
8	6.75	0.308
9	4.75	0.296
10	3.6	0.242
11	2.8	0.222
12	2.25	0.196
<i>Conditions</i>		
Catholyte flow rate $N$ (cm <sup>3</sup> s <sup>-1</sup> )		2236
Cell voltage (V)		6.4
Cell current (A)		900
Temperature (°C)		55.5
Current efficiency (%)		65
Overall fractional conversion ( $f_R$ ) <sub>n</sub>		0.975

The cell was fitted with potential probes in each compartment and during some processing full potential monitoring was achieved (Table 3).

The above results were considered promising and were used in order to scale-up to a commercial reactor. A cascade cell having twelve compartments has been installed by a Danish company, in order to remove copper from an acid effluent stream [4]. In this case, the cascade reactor is used as a second-stage treatment, following first-stage reduction of the copper concentration by a single compartment Eco-Cell, the combined plant treating about 8 m<sup>3</sup> h<sup>-1</sup> of electrolyte. Typically, the cascade reactor receives a copper concentration of 80–100 mg dm<sup>-3</sup>, and reduces this to about 2 mg dm<sup>-3</sup>. Table 4 shows concentration profiles for the reactor, an inlet of about 89 mg dm<sup>-3</sup> being reduced to an outlet of about 2 mg dm<sup>-3</sup> with an overall current efficiency of about 60%. The concentration profiles are shown in Fig. 6.

While both of the above reactors operated in a satisfactory manner, it was considered desirable to develop an improved 'second generation' cascade reactor with a view to:

(a) lowering capital costs;

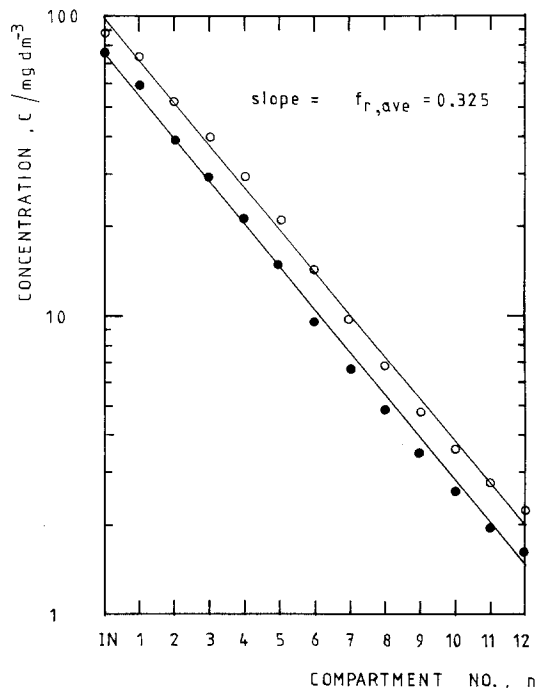


Fig. 6. Concentration profiles for the IKA cascade reactor.

- (b) improving engineering and minimizing maintenance;
- (c) facilitating routine maintenance;
- (d) attaining enhanced electrochemical performance;
- (e) improving the ease and efficiency of the metal product removal process.

A major change has been to use a cylindrical cathode rotated vertically about its axis, as opposed to the previous horizontal design. This has several advantages including:

1. minimization of floor space;
2. improved mechanical operation;
3. greater ease of product recovery.

In the horizontal designs, metal was removed from the reactor after a prolonged period of operation, the electrolytic current and normal electrolyte flow were turned off, and the reactor was backwashed with a minimum volume of nitric acid, to chemically dissolve the copper. This proved to be a rapid and successful method of product removal, and in the case of the commercial reactor, the resultant copper nitrate concentrate could be recycled to the first-stage Eco-Cell, producing copper powder. One of the inherent advantages of the Eco-Cell, that is, product

recovery as metal powder, was lost, however. An additional consideration is that certain of the nobler metals, e.g. the platinum group, are difficult to dissolve chemically, and so the method has a more limited range of application.

In the improved vertical design, a scraper mechanism has been included to remove powder from the cathode at suitable intervals, say once every eight hours. At such times, a backwash cycle would be initiated, such that the normal flow (bottom-to-top) could be reversed to facilitate powder removal, aided by gravity. Product removal is also encouraged by having apertures in the baffle plates. In normal operation, the apertures are covered by a close-fitting scraper blade arrangement, but during the metal recovery operation the blade is moved out of the baffle space. The provision of multiple blades (one per compartment) results in a much smaller displacement and decreased blade wear compared with a single blade.

A 500 A Cascade Cell was constructed for commercial use having important differences (see Table 2); in particular the catholyte flow rate was reasonably high at  $580$  to  $1000 \text{ cm}^3 \text{ s}^{-1}$  and only six compartments were provided. Comparable conversion data was obtained for this reactor but in addition polarization and electrical power measurements were made for varying conditions of usage. The reactor was equipped with three pairs of anodes, each pair extending over two compartments. Each anode pair was separately controlled and could be used at will. The basic polarization characteristics were defined using galvanostatic control, a reference electrode probe near the cylinder in each compartment and a lower brush engaging on the rotating cylinder shaft below compartment No. 1 and  $1 \text{ mol dm}^{-3}$  sulphuric acid at  $60^\circ \text{C}$  generating hydrogen only at the cathode (Fig. 7). For a given cell current the measured potential generally increased with higher compartment number, a particularly rapid increase occurring for potentials more negative than  $-450 \text{ mV}$  (SCE). The total cell voltage was also measured as a function of current. This showed an ohmic relationship and yielded an effective reactor resistance of  $0.015 \Omega$  when only anode pair No. 1 was in use.

Copper sulphate additions were made to the electrolyte and the polarization characteristics

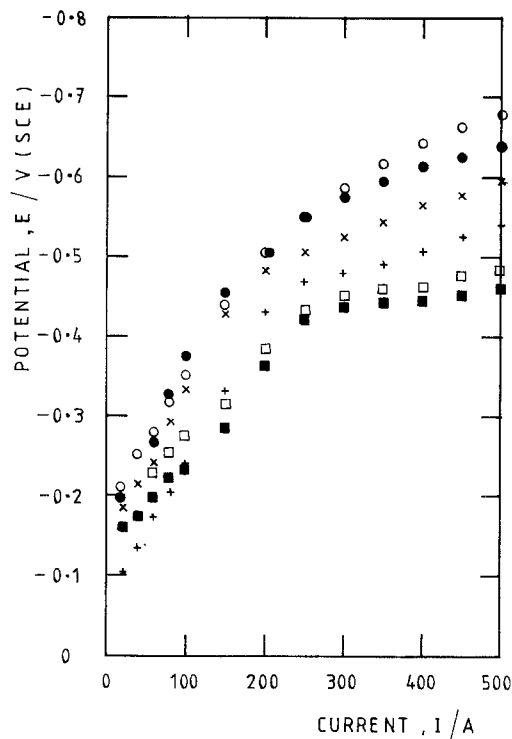


Fig. 7. Polarization behaviour for hydrogen evolution in the 500 A cascade reactor. Compartment number:  $\circ$ , 1;  $\bullet$ , 2;  $\times$ , 3;  $+$ , 4;  $\blacksquare$ , 5;  $\square$ , 6.

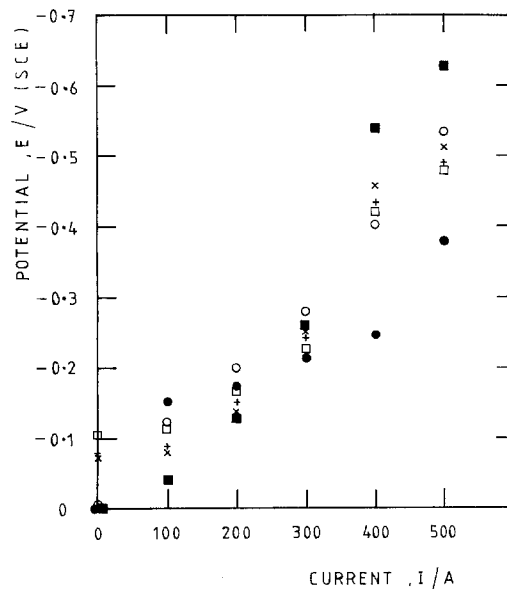


Fig. 8. Polarization behaviour for copper deposition in the 500 A cascade reactor. Compartment number:  $\bullet$ , 1;  $\circ$ , 2;  $+$ , 3;  $\times$ , 4;  $\square$ , 5;  $\blacksquare$ , 5.



measured again. With an input concentration of  $152 \text{ mg dm}^{-3} \text{ Cu}^{2+}$  behaviour was rather complex (Fig. 8) but a tendency to exhibit a limiting current density in the potential range  $-350$  to  $-450 \text{ mV}$  suggests an approach to mass transport control. When copper dosing to the batch circulation reservoir was stopped, the concentration depleted rapidly as copper was stripped out of solution. Under these circumstances the concentration decayed exponentially and Fig. 9 shows the results recorded after 30, 37 and 47 min of continuous deposition at  $300 \text{ A}$  and  $2.7 \text{ V}$  cell voltage.

Analysis of the data for the  $500 \text{ A}$  cascade reactor (Figs 9 and 10) showed that a reasonable conversion took place in compartments 1–4 but in the last two compartments conversion was very poor. This could be attributed to internal channelling or bypassing of high concentration electrolyte from high concentration (lower) to low concentration (higher) compartments, thereby effectively destroying the concept of cascade flow.

Upon disassembly and examination, a relatively wide gap ( $2 \text{ mm}$ ) was evident between the reactor body and the catholyte baffle plates facilitating

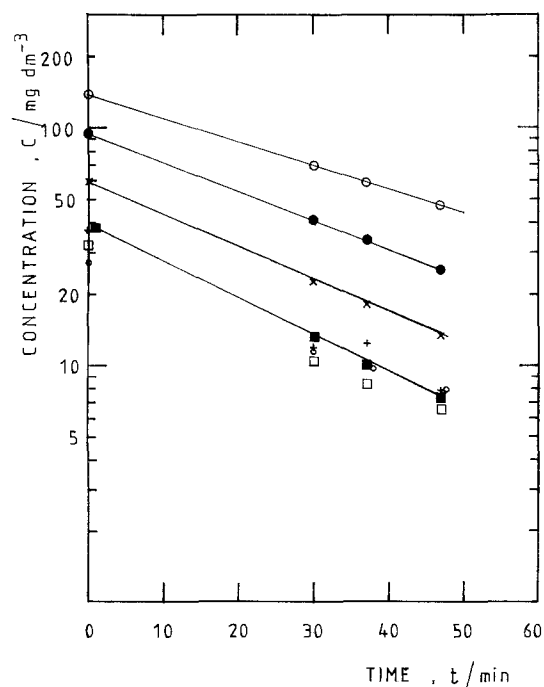


Fig. 9. Concentration decay in the  $500 \text{ A}$  cascade reactor. Compartment number:  $\circ$ , inlet;  $\bullet$ , 1;  $+$ , 2;  $\times$ , 3;  $\square$ , 4;  $\blacksquare$ , 5;  $\diamond$ , 6.

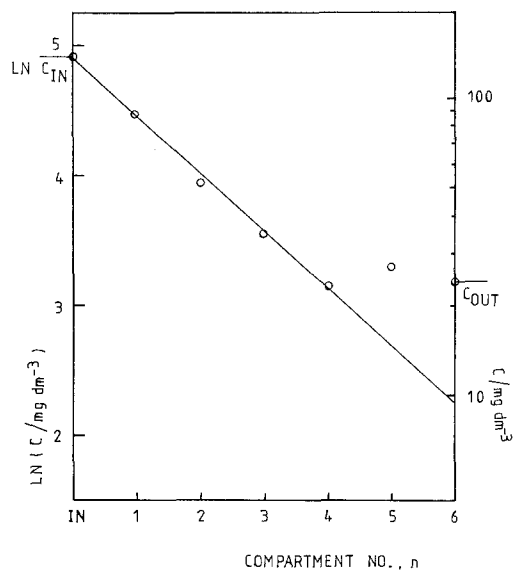


Fig. 10. Concentration profiles for the early  $500 \text{ A}$  cascade reactor showing destruction of cascade action in higher compartments due to a bypass fault condition.  $55^\circ \text{ C}$ ,  $1000 \text{ cm}^3 \text{ s}^{-1}$ ,  $400 \text{ A}$ . Overall cathode current efficiency  $87\%$ . Solid line shows predicted performance, according to Equation 3.

such adverse channelling of flow. The reactor performance may be readily restored by redesign of critical construction features, or by resealing. The results serve to show the importance of correct design and assembly of high conversion cascade reactors.

#### 4. Discussion

A cascade of CSTRs is a powerful reactor assembly, capable of high overall conversions and low exit concentrations of metal suitable for effluent discharge. A turbulent rotating cylinder electrode geometry is a particularly advantageous combination of flow pattern and geometry. A uniform electrode surface is provided enabling a near equipotential electrode to be used thereby allowing potentiostatic control to be exercised, while high mass transfer rates may be realized giving rise to high rates of conversion. In addition, the mass transfer to the electrode is largely independent of the actual flow rate, being governed primarily by the rotational velocity of the cylinder, and its surface roughness and area. This is in marked contrast to plug flow reactors such as packed beds and unstirred filter press cells where the mass transfer

is strongly dependent on the flow rate. Mass transport controlled operation ensures maximum reactor duty.

In comparison with other high conversion reactors, the Eco-Cascade-Cell does not suffer the limitations of high pressure drops which may occur as product builds up in packed beds or filter press cells. Nor does it have the problems of severely uneven potential distribution. Unlike the majority of other electrochemical reactors for metal removal, the metal product may be recovered directly for re-use or sale; this is particularly important in the case of precious metals.

Despite the importance of cascade electrochemical reactors, there has been a surprising paucity of published work in this field. Sudall [12] and Pickett [13] have considered the theoretical aspects of a cascade arrangement with respect to electrolyte flow and electrical connections. The first author has attempted to utilize experimental data from a single impeller-stirred CSTR in order to forecast, by computer, the behaviour of a cascade of similar elements.

Conventional filter press place cells are often designed on the basis of a number of CSTR compartments in hydraulic series; the CSTR model being a more reasonable approximation in the case of low throughput of electrolyte and thorough gas stirring.

A reactor often employed in larger scale photographic silver recovery, the 'Hickman Cell' [14] employs a number of carbon plate anodes and interleaved stainless steel cathodes, with a paddle agitator moving between the plates. While this system has been considered as a cascade cell, in practice, considerable bypassing and non-uniform electrode potential result in a non-idealized performance. In addition, mechanical disassembly, followed by scraping is necessary to remove silver.

Returning to the Eco-Cascade-Cell Reactor, it is interesting to note that if an infinite number of elements were provided, the system might be expected to approximate to a plug flow reactor (PFR), with an enhanced overall conversion. If each compartment behaves according to the plug flow model, the conversion over each element would be given by:

$$f_{\text{R}} = 1 - \exp\left(\frac{-K_{\text{L}}A}{N}\right). \quad (8)$$

Assuming values of  $K_{\text{L}}$ ,  $A$  and  $N$  identical to those in the CSTR example, a value of  $f_{\text{R}} = 0.632$  is obtained, which is somewhat greater than the 0.5 in the comparable CSTR case. In the case of a cascade of  $n$  identical plug flow elements, the overall conversion is given by:

$$(f_{\text{R}})_n = 1 - \exp\left(\frac{-nK_{\text{L}}A}{N}\right). \quad (9)$$

Comparing CSTR and PFR conversions for a hypothetical cascade, it may be seen that as  $n$  increases, the difference between the two models diminishes and asymptotic behaviour occurs.

Another interesting comparison may be made by removing the baffles from a cascade reactor. In this case, the reactor behaves as one compartment, with an increased surface area of  $nA$ . Examining this case for a CSTR, the fractional conversion is given by:

$$f_{\text{R}} = 1 - \frac{1}{nK_{\text{L}}A/N} \quad (10)$$

and this result is equivalent to operating the  $n$  elements in series electrical connection and parallel flow. A comparison of the data reveals that an overall conversion of 0.875 may be attained by a three-compartment reactor, in contrast to the seven projected compartments of an unbaffled reactor.

Clearly the practical cascade reactors only approach the hypothetical or theoretical models and despite careful control of flow, rotation rates, temperature and concentration, etc., a variable conversion factor was found from compartment to compartment (see Tables 3 and 4). Although this must partly be attributed to imperfect construction of the cascade cell compartment dividers, thereby allowing bypassing to occur, it is also attributable to a degree of randomness in the power electrodeposit.

The great virtue of potentiostatic control on the single-compartment Eco-Cell, which may ensure nearly 100% efficiency, becomes much more difficult for the multi-compartment cascade cell and efficiency is consequently lower. Notwithstanding this deficiency it is quite clear that the cascade cell acts as a very effective 'stripping' or 'polishing' stage in the removal of metal from process effluent as the data of Tables 3 and 4 bear witness.

### Acknowledgements

The authors gratefully acknowledge the assistance of the staff of Ecological Engineering Limited in the preparation of this paper.

### References

- [1] F. S. Holland, *Chem. Ind.* (1978) 453.
- [2] F. S. Holland, British Patent 1505736 (1978).
- [3] F. S. Holland, US Patent 4028199 (1977).
- [4] F. S. Holland and H. Rolskov, Effluent Water Treatment Convention, Birmingham (1978).
- [5] D. R. Gabe, *J. Appl. Electrochem.* **4** (1974) 91.
- [6] D. R. Gabe and F. C. Walsh, *Proc. Interfinish 80*, Kyoto, Japan (1980) p. 486.
- [7] F. C. Walsh and D. R. Gabe, *Surface Tech.* **12** (1981) 25.
- [8] M. Fleischmann, R. E. W. Jansson, G. A. Ashworth and P. J. Ayre, Provisional British Patent Application 18305 (1974).
- [9] C. E. Duisenberg and R. H. Bedford, US Patent 2791555 (1957).
- [10] W. M. Adams, US Patent 3342718 (1967).
- [11] N. Ibl, *Adv. Electrochem. Electrochem. Eng.* **2** (1966) 49.
- [12] S. J. Sudall, MSc thesis, UMIST (1970), PhD thesis, UMIST (1975).
- [13] D. J. Pickett, 'Electrochemical Reactor Design', Elsevier, Oxford (1977) p. 383.
- [14] K. Hickman, C. Sandford and W. Weyerts, *J. Soc. Motion Picture Engineers* **17** (1931) 568.