Copper electrowinning from acid mine drainage: A case study from the closed mine "Cerovo"

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A R T I C L E   I N F O

A B S T R A C T

Copper removal from acid mine drainage originating from closed copper mine "Cerovo" RTB Bor, Serbia and containing approximately $1.3\,g\,dm^{-3}$ of copper and a very small amount of Fe^{2+}/Fe^{3+} ions, has been successfully performed by the direct electrowinning method using either a porous copper sheet or carbon felt as the cathode. A cell with a fluidised bed of inert turbulent promoters, also used in this study, may be considered as unacceptable for the purpose view, having a cell voltage between 12 and 14 V.
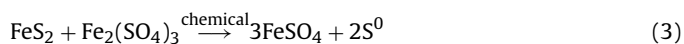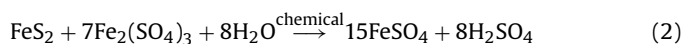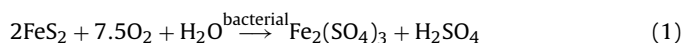
The cells used in the electrowinning experiments were compared in terms of cell voltage, pH and copper concentration. The results showed that it is possible to remove copper successfully from the mine waters with a high degree of electrowinning—higher than 92% and with a satisfactorily average current efficiency (>60%). Depending on the process time and the applied current, a final copper concentration less than $0.1\,g\,dm^{-3}$ was achieved. The specific energy consumption was approximately $7\,kWh\,kg^{-1}$ of deposited copper. A dense copper deposit was obtained when a three-dimensional electrode was used.

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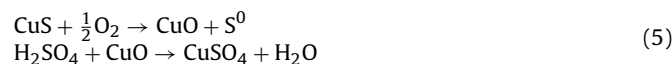
1. Introduction

1.1. Generation of mine waters

Mine waters originating from active or closed copper mines contain copper ions—sometimes in a considerable concentration ($\approx l\,g\,dm^{-3}$) usually associated with an equivalent or even a two times higher concentration of Fe^{2+}/Fe^{3+} ions as a consequence of bacterial leaching of sulphide copper and iron minerals [1]. Pyrite is the most abundant mineral in poly-metallic sulphide ore deposits and in mining waste dumps. The oxidation of pyrite and copper minerals in an aqueous environment occurs by means of two simultaneous mechanisms, *i.e.*,—chemical and biochemical involving bacteria. These can be described by the following overall stoichiometric reactions [2]:

$$2FeS_2 + 7.5O_2 + H_2O \xrightarrow{bacterial} Fe_2(SO_4)_3 + H_2SO_4 \tag{1}$$

$$FeS_2 + 7Fe_2(SO_4)_3 + 8H_2O \xrightarrow{chemical} 15FeSO_4 + 8H_2SO_4 \tag{2}$$

$$FeS_2 + Fe_2(SO_4)_3 \xrightarrow{chemical} 3FeSO_4 + 2S^0 \tag{3}$$

$$S^0 + H_2O + 1.5O_2 \xrightarrow{bacterial} H_2SO_4 \tag{4}$$
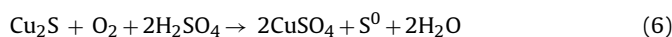
The above reactions describe how acidity is generated and maintained in dumps.

Copper minerals in mining waste dumps can also be decomposed either directly by bacterial attack or by leaching with ferric ions formed through pyrite degradation (Eqs. (1)–(4)), which can be described by the following stoichiometric reactions [3]:

Covellite:

$$CuS + \tfrac{1}{2}O_2 \rightarrow CuO + S^0$$
$$H_2SO_4 + CuO \rightarrow CuSO_4 + H_2O \tag{5}$$

Chalcocite:

$$Cu_2S + O_2 + 2H_2SO_4 \rightarrow 2CuSO_4 + S^0 + 2H_2O \tag{6}$$

Elemental sulphur will be further oxidized as described by Eq. (4).

The spontaneous oxidation of other sulphide minerals existing in mine wastes, by the same or a similar mechanism as described above, is responsible for the release of other toxic elements into solution, such as: Zn, Cd, Pb, Ni, As, Mn and others. The ultimate result is a mine water discharge characterized by an elevated acidity and high concentration of metal ions associated with sulphate anions, well known in the relevant literature as acid mine drainage (AMD) [4–6]. AMDs usually have a pH value between 3 and 4 due to the elevated content of sulphuric acid as discussed earlier and described by Eqs. (1)–(4) and (6).

The flow rate of AMDs varies: from few litres to several cubic meters per minute and changes appreciably with season as well as composition. In any case, mine waters constitute a serious

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environmental problem conveying metal ions and acid to the environment, thus seriously damaging ground waters and soils. Copper has been usually removed from mine waters by cementation onto iron scrap [7,8] and sometimes by using separation/concentration technologies, such as: solvent extraction followed by electrowinning (SX/EW) [9] or column adsorption on an appropriate adsorbent or ion exchanger [10]. Other methods such as membrane techniques are still under research and development [11–13].

1.2. Electrowinning of heavy metals

The electrochemical treatment of wastewaters, containing toxic compounds (cations or anions), has attracted considerable attention of many researchers working in the area of environmental abatement for many years [15–18]. Damaging heavy metal ions can be removed by electrowinning methods from sources such as wastewaters and the metal can be recovered in a useful form. The most important advantage of electrowinning over chemical or physico-chemical processes is its environmental compatibility, due to the fact that the electron, acting as a reactant, is a "clean reagent".

Since the concentration of metal ions in wastewaters is mainly low or very low, sometimes only a few milligrams per litre, special cells with either high mass transfer features or a high specific surface electrode area must be employed for metal removal from dilute sources. Different types of electrochemical cells with three-dimensional (3D) cathodes have been developed and marketed in the past two–three decades for the efficient removal and recovery of metal from such effluents, e.g., rinse- and wastewaters [17–19]. Carbon felt, as a three-dimensional electrode material, appears to be very promising for the recovery of heavy metals from very dilute solutions due to its favourable physico-chemical properties, i.e., high specific surface area, high porosity, good chemical stability and electrical conductivity [20–26]. Other materials have also been tested as 3D cathodes, such as metal particles either as a packed or fluidised bed, metal foam [[15] and literature cited therein], porous metal plates, metallised porous plastic, etc. [27,28].

Electrochemical cells with enhanced mass transfer of ions from the bulk electrolyte to the electrode surface have also been designed to operate with effluents containing low concentrations of metals. Five- to sixfold mass transfer enhancement was achieved by means of inert turbulent promoters in the form of a packed or fluidised bed [17,19,31]. The electrochemical recovery of nickel and copper from rinse waters generated in electroplating plants is now commercially practised using a cell with mesh electrodes immersed in a fluidised bed of glass beads, known as the "Chemelec" type cell [29,30].

1.3. Acid mine drainages from Open Pit "Cerovo"

In this paper, the results of copper electrowinning from mine waters originating from the Open Pit "Cerovo" are presented. The Open Pit "Cerovo" is part of the Copper Mining and Smelting Company Bor, situated 11 km north of the city of Bor, Serbia. This open pit is now closed. The production ended in 2002 but the ex-mine still affects the quality of the surrounding ground waters.

There are two AMDs flowing from the outer lateral slope of the open pit dump (here denoted as Spring 1 and Spring 2). Both sources are characterized by a reasonably high copper concentration (>1 g dm^{-3}), indicating a high content of copper sulphide minerals in the dump, and by a very low content of iron, not exceed 10 mg dm^{-3}, as shown in Table 1, in which summarized data concerning the compositions of the AMDs are also presented. Other metal ions associated with copper in these AMDs are present in amounts much higher than the maximum allowed concentration, meaning that sulphide minerals of other heavy metals allied with

Table 1
Composition of the mine waters from the AMDs of the Open Pit "Cerovo".

Element	Spring 1 (mg dm^{-3})	Spring 2 (mg dm^{-3})	Allowed values (mg dm^{-3})
Mn	40.1	65.7	/
Cu	1050	1550	0.1
Fe	<10	<10	
Ni	0.5	1.1	0.05
Co	2.1	5.6	0.2
Y	0.9	3.3	/
Al	>74	>884	/
Cd	0.2	0.4	0.005
Zn	>20	>27	0.2
Mg	>670	>947	/
Be	0.003	0.074	0.0002

copper are also present in the dump. The very low concentration of iron in the AMDs could mean a lack of pyrite and chalcopyrite in the dump.

There are no other anions present beside sulphate ions. The concentration of SO_4^{2-} is also high, >1200 mg dm^{-3}, according to the mechanism of the oxidation of minerals. The pH value is 3.4 for Spring 1 and 3.6 for Spring 2, while the corresponding specific conductivities are 10.73 and 7.1 mS cm^{-1}, respectively. The flow rate varies between 1.5 and 5 dm^3 min^{-1}.

Mine waters from the AMDs, flow directly into the "Cerova" River, which has, according to the local laws, a water quality between 1st and 2nd category before the inflow of the AMDs which make it fourth category water downstream of these springs.

Hitherto, no attempt has been made to treat copper bearing mine waters by the direct electrowinning method. The main reason for not considering direct electrowinning for copper or some other valuable metal removal from mine waters is the reasonably high concentration of Fe^{2+}/Fe^{3+} ions, usually present in these natural sources. The Fe^{2+}/Fe^{3+} couple greatly influences the electrowinning of copper, acting as a simultaneous redox reaction, thus decreasing significantly the current efficiency and disabling the efficient removal of copper to a desired final concentration. Unexpectedly, the very low concentration of Fe-ions in the AMDs (see Table 1) allows a pioneering attempt to be made of direct electrowinning of copper from such waters before the further processing of sulphuric acid neutralization and removal of other toxic ions. With the exception of Fe-ions, the other ionic species present in the AMDs do not affect the copper electrowinning process.

2. Experimental

Samples of AMD waters were taken from Spring 1 and Spring 2. Except for spontaneous clarification, no any other pre-treatment of the mine waters was performed prior to electrowinning. The samples were mixed to obtain an average composition of water for the copper electrowinning study. The initial copper concentration of the mixture used in the experiments was ≈1.3 g dm^{-3}. The initial pH was ≈3.52 and the initial specific conductivity ≈9 mS cm^{-1}.

In all experiments, the water volume was kept constant at 3.5 dm^3.

2.1. Experimental set-up

A hydrodynamic circuit was formed using a reservoir (1), centrifugal pump (2), valves (3), a flow-meter (4) and a cell (5), all connected with plastic hose-pipes, as is schematically presented in Fig. 1.

The electrical circuit consisted of a P/Gstat (7) (Amel 555) as a power supply and a voltmeter (6) (MASTECH MY – 64), all connected by wires to the electrodes.
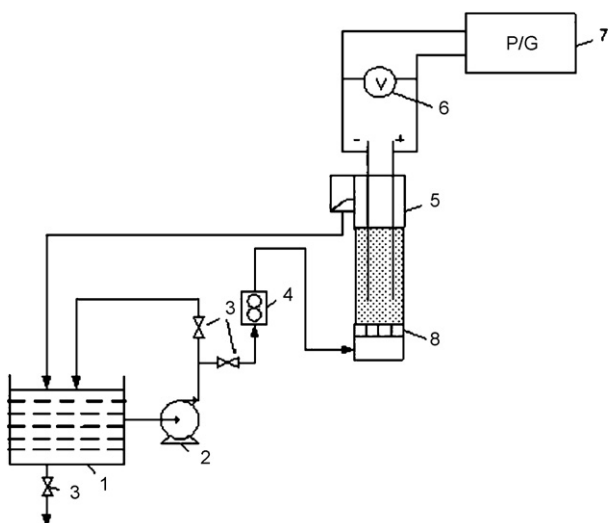
Fig. 1. Schematic representation of the experimental set-up: (1) reservoir, (2) centrifugal pump, (3) valves, (4) flow-meter, (5) electrolytic cell, (6) voltmeter, (7) P/Gstat and (8) flow-distributor.

2.2. Cell and electrodes

Electrowinning of copper was performed in an electrochemical cell (30 cm × 5 cm × 5 cm) made from Plexiglas. A perforated electrolyte distributor was built in, dividing the cell body into two parts and serving to produce a uniform electrolyte velocity over the whole cross-section of the cell. It also served as a glass bead bed support when the cell was used with a fluidised bed of glass particles as mass transfer promoters. The lower part of the cell served as a calming section into which the electrolyte was introduced through an opening at the bottom. At the top of the cell, there was an overflow from which the electrolyte flowed gravitationally back to the reservoir.

The part above the electrolyte distributor served as the reacting compartment into which the electrodes were placed in a side-by-side configuration. A stainless steel grid was used as the anode, while three different cathodes were tested in the experiments:

A. A flat copper plate (5 cm × 10 cm × 0.05 cm), used in the first case, was immersed in a fluidised bed of glass beads (d_p = 1 mm), in the inter-electrode space and serving as inert turbulence promoters to enhance mass transfer of copper ions from the bulk to the cathode [17,30]. The fixed bed height was 10 cm while, in its fluidised state, the height was 13 cm. The fluidised bed of glass particles was only employed with the flat copper electrode.
B. Porous copper sheet (5 cm × 10 cm × 0.2 cm; RETEC, Porous Metal, USA), having a pore density 100 ppi, a porosity $\varepsilon \approx 0.64$ and a specific surface area 1670 m^{-1} was used as a three-dimensional cathode.
C. Carbon felt (5 cm × 10 cm × 1.5 cm) was tested as a three-dimensional electrode in the third case. Its porosity was $\varepsilon \approx 0.95$ and specific surface area 22,000 m^{-1}. The carbon felt was supplied by Sigry GmbH, Germany.

2.3. Other experimental details

All the electrowinning experiments were performed in the galvanostatic mode of operation at different currents and under different hydrodynamic conditions. The cell voltage was monitored by a voltmeter (MASTECH MY – 64) enabling data concerning the energy consumed in the process to be derived. The pH value of the processed mine water was also periodically measured using a

WTW inoLab-720 pH-meter. Samples of the mine water were periodically assayed and the concentration of copper was determined by atomic absorption spectrophotometer using a PerkinElmer-403 instrument. The flow rate through the cell was maintained constant throughout the experiment by adjusting with valves (3) and monitoring by a flow-meter (METRIC 24 P) (4). The experiments were performed at ambient temperature (approximately 20 °C).

3. Results and discussion

3.1. Process stoichiometry

During the electrowinning of copper, two cathodic reactions usually occur consecutively or even simultaneously, the stoichiometry of which is given by the following equations:

- Reduction of copper ions:

$$Cu^{2+} + 2e^- = Cu \tag{7}$$

- As a competitive reaction, hydrogen evolution occurs simultaneously when the operating current density used for copper deposition becomes higher than the value of the limiting current density, for a given concentration and hydrodynamic conditions [32,33]:

$$2H^+ + 2e^- = H_2 \tag{7a}$$

- Oxygen evolution occurs on the anode according to the overall stoichiometric reaction:

$$2H_2O = O_2 + 4H^+ + 4e^- \tag{8}$$

3.2. Cell voltage–time relationship

Cell voltage U can be expressed by the following equation:

$$U = U_0 + |\eta_a| + |\eta_C| + IR + |\eta_{H_2}|_* + \Delta E \tag{9}$$

where U_0, denotes the decomposition potential (V); η_a, η_c, η_{H_2}, the anodic, cathodic and hydrogen evolution overvoltage (*the last term has to be taken into account only when hydrogen evolution occurs simultaneously with metal deposition), respectively (V); I, current (A); R, ohmic resistance (Ω), ΔE, potential drop inside the three-dimensional cathode (V) [32].

The term ΔE refers here only to the carbon cloth and the porous copper sheet acting as three-dimensional electrodes. For the galvanostatic mode of operation, used in these experiments, and for the employed electrolyte, which has a rather low specific conductivity, an elevated cell voltage during the electrowinning process is to be expected. The generation of protons in the anode reaction (8) leads to a decrease in cell voltage due to increased electrolyte conductivity, as shown in Fig. 2.

At the very beginning of the electrowinning process, the cell voltage had its highest value for all three cells and it then started to
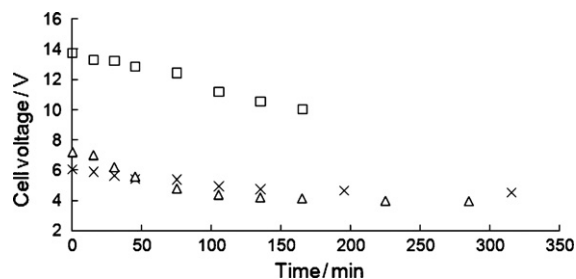

Fig. 2. Plot of cell voltage versus time for different cathodes: (□) cell with fluidised bed of inert particles (I = 0.7 A); (×) cell with porous copper sheet (I = 2 A); (△) cell with carbon felt (I = 0.7 A).

decrease with time, reaching a constant value after a sufficiently long period of time. In the case of the cell with a fluidised bed of glass beads, the cell voltage was almost two times higher than that in the cells with the three-dimensional cathodes, due to the increased ohmic resistance caused by the presence of the inert particles in the inter-electrode space [14]. The presence of a non-conductive dispersed phase in an electrolyte (gas bubbles or solid particles) decreases the apparent conductivity of such a two-phase system by an amount depending on the concentration of dispersed phase in the electrolyte. For a bed of glass beads, fluidised by an electrolyte, the apparent conductivity can be described by the well known Bruggeman equation [14]:

$$\kappa = \kappa_0 \varepsilon^{1.5} \tag{10}$$

where κ and κ_0 represent the conductivity in the fluidised bed and in the electrolyte alone, respectively; ε, porosity of the fluidised bed.

This explains the more than two times higher cell voltage in the fluidised bed cell in comparison to the cell with carbon felt as cathode. The significantly higher cell voltage is the main drawback for the application of a fluidised bed cell with inert turbulence promoters for the purpose of metal removal from less conductive electrolytes. The higher cell voltage causes a proportionally higher specific energy consumption, which may significantly reduce the positive effect of the mass transfer enhancement, particularly when going towards higher percentage of metal removal from solution.

A comparison between the carbon felt and the porous copper sheet cathodes showed that the cell with the porous copper sheet had the lowest ohmic resistance. The smallest cell voltage was achieved in the cell with the porous copper sheet cathode. As can be seen from Fig. 2, this cell voltage was almost the same as that with the carbon felt cathode even though the operating current was three times smaller in the second case. This is due to the higher deposition overvoltage of copper onto carbon than the deposition potential of copper onto copper and the much higher potential drop within the three-dimensional cathode, ΔE, in the case of the carbon felt cathode, having a seven times higher thickness than the cathode made from the porous copper sheet. Thus, the porous copper cathode is the most suitable electrode for the purpose considered here—the treatment of mine waters.

During the process, the cell voltage decreased progressively for each of the cathodes. This can be attributed to an increase in the conductivity of the electrolyte during the electrowinning process, due to the generation of H+ ions according to Eq. (8), which is also connected with the change in electrolyte acidity during the process. In the case of the carbon felt and porous copper sheet electrodes, the cell voltage reached a constant value after a certain period. The lack of a further decrease in cell voltage is associated with simultaneous hydrogen evolution on the cathode, which consumes the produced protons and results in a tendency for the cell voltage to increase because the hydrogen overvoltage has now to be taken into account in Eq. (9). Hydrogen bubbles increase the overall concentration of the gas phase in the inter-electrode space, which increases the ohmic drop in the same way as the fluidised inert particles. Leaving the electrowinning process for a longer period, the cell voltage will increase.

3.3. pH–time relationship

It is well known that during the electrowinning of metals, the electrolyte acidity increases when the anodic reaction is oxygen evolution, due to the splitting of water according to Eq. (8). This effect causes a progressive decrease in the pH value of the processed mine water with time, as presented in Fig. 3.

The fastest change in pH of the mine water was at the very beginning of the process. With the passing of time, the pH decrease became slower and tended to asymptote to a constant value close to
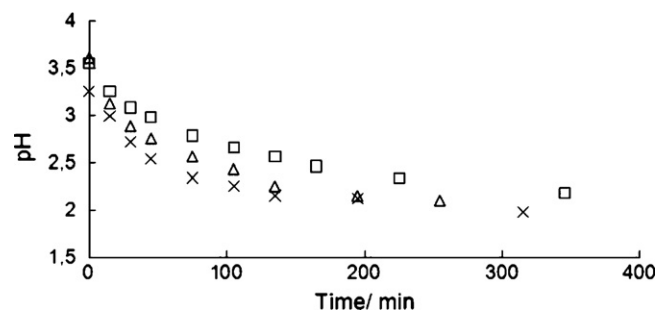

Fig. 3. Plot of pH versus time in the cell with the porous copper cathode for different operating currents: (×) 2 A; (△) 0.9 A; (□) 0.5 A.

pH ≈ 2. The decrease in pH slowed down significantly when simultaneously with copper deposition, hydrogen evolution commenced on the cathode (reaction (7a)), consuming the protons produced in the anode reaction (8), thus achieving a steady state condition when the pH no longer changed with time. The increased acidity of the mine water after completion of the electrowinning of copper could be beneficial if there is an organized dump leaching process of residual sulphide minerals in the mine waste material or at the edges of the ore body. This is usually associated with the closure procedure of copper mines with the aim of speeding up the leaching process, keeping the AMDs under control and reducing their life time. Otherwise, self-leaching could occur spontaneously for decades and damage the environment. In the case of organized leaching, the mine waters are usually recycled in the leaching process after depletion of their copper content, to be reused as leaching solutions. If there is no promoted leaching of dumps, the consumption of chemical is increased because of the necessity to neutralize the acid produced in the electrowinning process, which would increase the overall costs of the treatment of AMDs.

3.4. Change of copper ion concentration with time for different cells

The effect of current intensity, I, or current density, i, in the electrowinning of metals is well known and is defined *via* the electrochemical reaction rate $r = i/zF$. This means, the higher the current density, the faster is the depletion of reacting ions. For a batch recirculating system, as was used in this study, the concentration change of reacting ions may be described by the following relationship [33]:

$$C(t) = C_i \exp\left\{ -\frac{tQ}{V} \left[1 - \exp\left(-\frac{k_L S}{Q} \right) \right] \right\} \tag{11}$$

where $C(t)$ and C_i (g dm^{-3}) are the actual and initial concentration of copper in the mine water, respectively; t is the process time (s); Q is the flow rate through the cell (m^3 s^{-1}); V is the volume of the mine water (m^3); k_L is the mass transfer coefficient (m s^{-1}) and S is the wetted electrode surface (m^2).

The change in copper ion concentration with time is presented in Fig. 4 for the three cathodes and for different operating currents, allowing the process rate and the quality of the deposit to be compared. As shown in Fig. 4, the depletion rate of copper ions using either the carbon felt or the copper plate as the cathode is very similar for the same current intensity. Coming into the second third of the process, copper powder appeared on the copper plate; hence electrowinning with this electrode was terminated.

With the carbon felt cathode, a good dense deposit of copper was obtained, firmly attached to carbon fibres making them copper coloured. The dense deposit was obtained throughout the electrowinning period due to the very low true current density (based on the real electrode surface), although the geometrical current density was reasonably high. Here the edge effect was
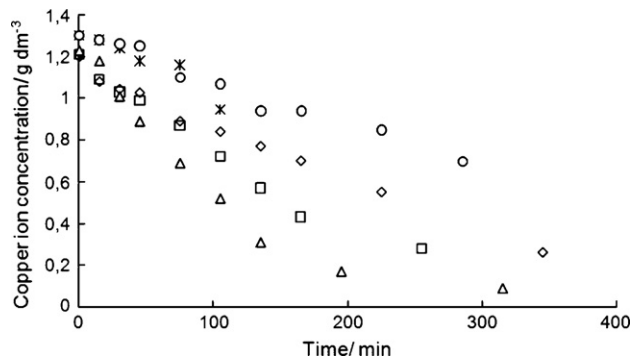
Fig. 4. Plot of copper ion versus concentration with time in the cell with different cathodes: porous copper sheet (($\Diamond$) $I = 0.5$ A; ($\square$) $I = 0.9$ A; ($\triangle$) $I = 2$ A); ($\times$) fluidised bed of inert particles ($I = 0.7$ A); ($\bigcirc$) carbon felt ($I = 0.7$ A).

operative since the edges of the electrode were the first place to be covered by copper, while the interior surface remained virgin. After a longer process time, a profile of copper colour across the felt thickness appeared as a consequence of the potential profile, which is stronger for thicker three-dimensional electrodes. Due to the potential profile across a three-dimensional electrode, the part closer to the anode is more active while the opposite part is almost inactive.

Using the porous copper cathode leads to faster depletion even though it works at a lower current than the previous two cathodes. This means that the current passing through the cell is spent to the greatest extent on the deposition of copper. For this reason, further experiments were performed with this cathode. On increasing the operating current, the concentration of copper in the mine water decreased more rapidly. However, a linear concentration–time relationship is lost after a certain time, indicating that the operating current becomes higher than the instantaneous limiting current for given experimental conditions [33]. A curvilinear concentration–time relationship appeared when hydrogen evolution started (reaction (7a)), occurring simultaneously with copper deposition. Here part of the operating current is being spent on the parasitic reaction, thus significantly affecting the current efficiency.

A dense deposit is obtained using porous copper as a cathode even after 4 h of electrowinning, which corresponds to a copper concentration of 0.2 g dm^{-3}, or even less. Only the edge effect was registered as a slightly darkly coloured deposit of copper, indicating that the local current density was higher in those areas than the limiting current, causing the appearance of copper powder at the electrode edges.

3.5. Change of current efficiency with copper concentration

The current efficiency is defined as the ratio between the real mass of deposited copper and that computed accordingly to Faradays Law. Since the mass of deposited copper can be expressed as: $[C_i - C(t)]V = m$, the current efficiency may be defined by the following equation:

$$\eta_e = \frac{[C_i - C(t)]VzF}{ItM} \tag{12}$$

where z is the number of exchanged electrons in the reaction; F is the Faraday constant (A s mol^{-1}); I is the current intensity (A); M is the molecular weight (g mol^{-1}).

Fig. 5 shows that the current efficiency has a maximum value between 80 and 100%, at the beginning of the electrowinning process. When the concentration of copper falls below 0.3 g dm^{-3}, the current efficiency rapidly decreases, due to simultaneous hydrogen evolution.
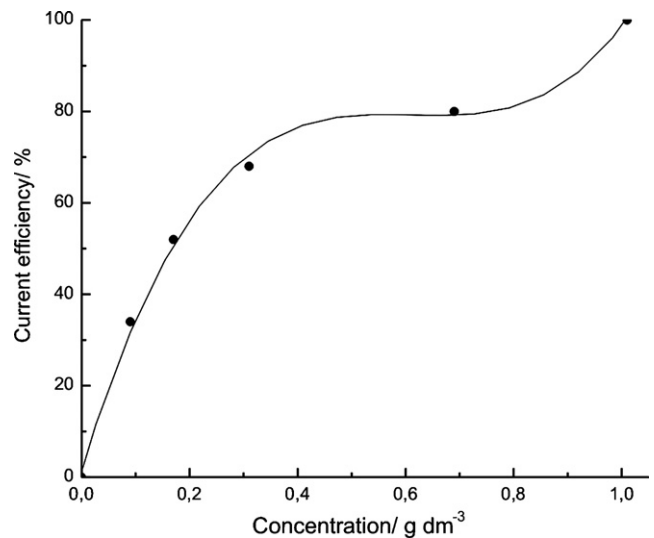

Fig. 5. Plot of the current efficiency versus the actual concentration of copper ions in the cell with the porous copper sheet cathode ($I = 2$ A).

3.6. Specific energy consumption

For the galvanostatic mode of operation, the specific energy consumption, E_e, defined as the energy spent in the process per unit mass of deposited copper, may be expressed by the equation [19,34]:

$$E_e = \frac{zF \int U(t)\,dt}{M \int \eta_e(t)\,dt} \tag{13}$$

The specific energy consumption depends on the cell voltage and the current efficiency—both functions of time. Thus, the current efficiency depends indirectly on time as a function of the concentration which is a variable of the process time, given by Eqs. (11) and (12).

Fig. 6 shows that the consumption of energy is relatively high at the beginning of the process for both cathodes because of the highest cell voltage value in the initial period of the process. During the process time, as the concentration of copper ions falls, the energy consumption also decreases following the change in cell voltage. As the concentration of copper ions approaches zero the specific energy consumption increases and tends to infinity because the current efficiency approaches zero. In the case of the porous copper sheet, this is noticeable. In case of the carbon felt electrode operated at a lower current, it should be expected to effect further depletion.
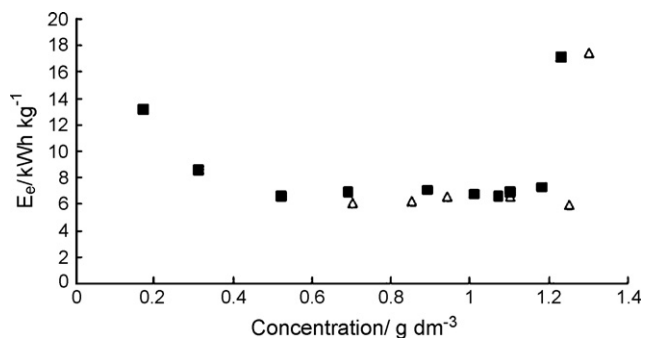

Fig. 6. Specific energy consumption versus the concentration of the reacting ions: ($\blacksquare$) porous copper sheet ($I = 2$ A); ($\triangle$) carbon felt ($I = 0.7$ A).

4. Conclusions

Copper removal from AMDs containing a small amount of Fe^{2+}/Fe^{3+} ions has been successfully performed by the direct electrowinning method using either a porous copper sheet or carbon felt as the cathode. A high degree of electrowinning, a satisfactorily good current efficiency and a good, dense metal deposit was obtained with both these cathodes. However, the cell with the porous copper cathode had better features than that with the carbon felt cathode in terms of current efficiency and specific energy consumption. Depending on the process time as well as the applied current, a final copper concentration, less than $0.1\,g\,dm^{-3}$, was achieved in the cell with the porous copper cathode, dense metal being produced.

A cell with either carbon felt or a porous copper sheet cathode can operate at higher geometrical current densities than a cell with inert turbulence promoters because of the high internal surface of these cathodes, allowing higher reaction rates than those with the cell with inert turbulence promoters. A cell with a fluidised bed of inert particles is not suitable to be for this purpose.

The cell voltage decreases with time due to decreasing pH as a consequence of oxygen evolution as an anode reaction which increases the acid content. Ranking the cells in view of the achieved cell voltage shows that the cell with the porous copper sheet cathode is the most suitable, having the lowest cell voltage. When using the carbon felt cathode, high cell voltage is registered due to the increased potential drop within the cell as well as the higher overvoltage of copper deposition onto carbon. The cell with the fluidised bed of inert turbulent promoters may be considered as unacceptable, having a cell voltage between 12 and 14 V.

The decrease in pH of the treated mine water with time due to the anodic oxygen evolution causes an increase in conductivity and acidity of the mine water. This fact may have a beneficial effect if the treated water is being recycled in a leaching stage, if present, leading to a decrease in the consumption of sulphuric acid. The opposite effect is an elevated consumption of chemicals needed to neutralize sulphuric acid formed during the electrowinning process prior to its release into a receiving water course.

Acknowledgement

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