COALTECH 2020

EVALUATION OF NEW AND EMERGING MINE WATER TREATMENT TECHNOLOGIES

PART C

EVALUATION OF THE ECODOSE PROCESS

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COALTECH 2020E valuation of the Ecodose Process

1 PROCESS FUNDAMENTALS

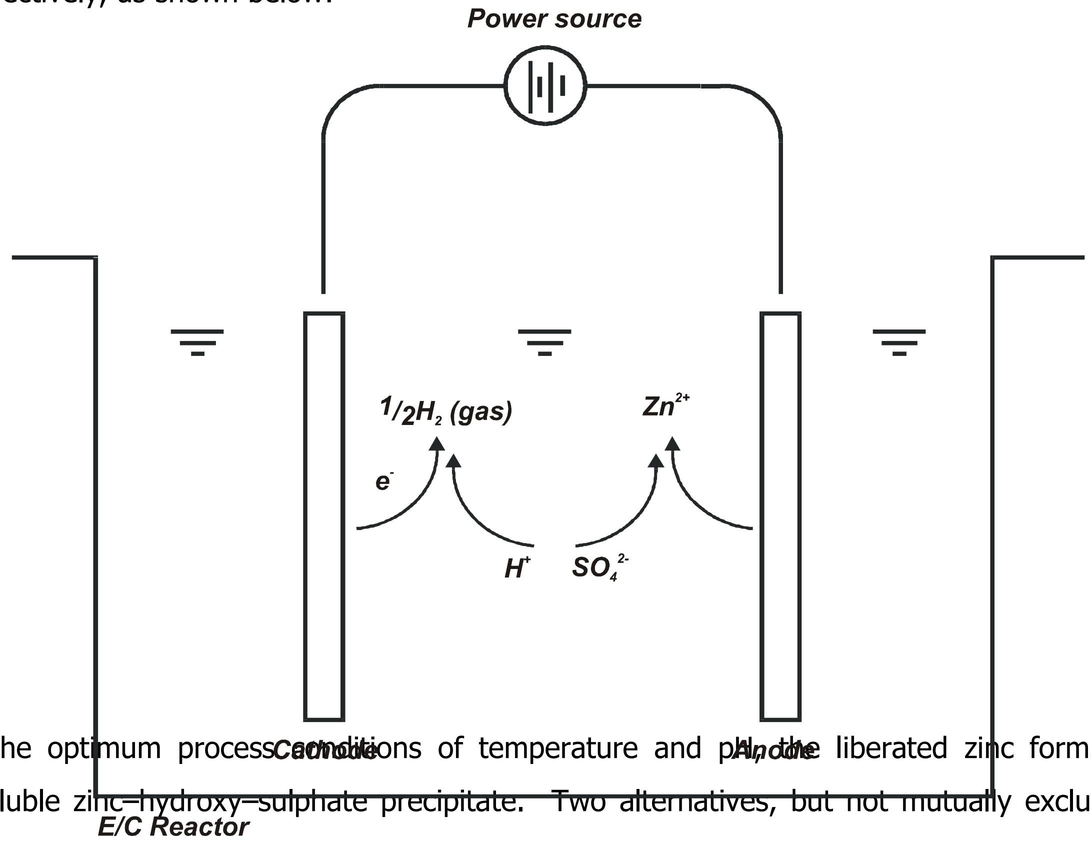
1.1 Process Description

EcoDose is an Electrochemical process which utilises the following two (2) coupled reactions:

$Zn \tau Zn^{2+} + 2e^-$

$2H^+ + 2e^- \tau H_2$ (gas)

These two (2) coupled reactions take place at the cathode electrode and anode electrode respectively, as shown below:



At the optimum process conditions of temperature and pH, the liberated zinc forms an insoluble zinc–hydroxy–sulphate precipitate. Two alternatives, but not mutually exclusive, precipitate compositions have been postulated:

- $Zn_4 (OH)_6 SO_4$ with the following mass ratio:

  $Zn : SO_4$

  2.7 : 1(kg/kg)

- $Zn_2 (OH)_2 SO_4$ polymers with the following mass ratio:

  $Zn : SO_4$

  1.36 : 1(kg/kg)

The amount of zinc required to drive the sulphate precipitation reactions can therefore vary, depending on the chemical precipitate composition.

The nett E/C treatment effect on the chemistry on acid mine water is progressive neutralisation and removal of some sulphate.

1.2 Auxiliary Reactions

The Electrochemical treatment progressively removes the free hydrogen ion (H^+) and shifts the pH of the water. As more H^+ is effectively removed, dissociation of water takes place:

$$H_2O \tau H^+ + OH^-$$

The relative abundance of OH^- increases, reflected in an increased pH.

Acid mine waters typically contain dissolved metals species, specifically aluminium, iron and manganese. As the mine water pH increases, some of these metals will start to precipitate as metal hydroxides:

$$Fe^{3+} + 3\ OH^- \tau Fe\ (OH)_3\ (s)$$

$$Fe^{2+} + 2\ OH^- \tau Fe\ (OH)_2\ (s)$$

$$A\ell^{3+} + 3\ OH^- \tau A\ell\ (OH)_3\ (s)$$

These metal hydroxides have different pH thresholds, above which the metal will start to precipitate. These pH thresholds are typically as follows:

- for ferrous iron, Fe^{2+} @ 6.5 – 7.0
- for ferric iron, Fe^{3+} @ 3.5 – 4.0
- for aluminium, $A\ell^{3+}$ @ 4.5 – 5.0

The manganese metal (Mn^{2+}) will not precipitate at the relatively low pH at which the Electrochemical reaction takes place. The threshold pH for manganese hydroxide precipitation is typically > 8.

The nett effect of the metal hydroxide precipitation is that the OH^- generated in the dissociation of water is effectively consumed. The process pH will therefore not increase above certain thresholds, until the metal hydroxides are precipitated. The plateau-effect of a temporary stable pH as the Electrochemical reaction proceeds has been well documented by a number of batch studies.

1.3 Process Constraints

1.3.1 Competing Reactions

The optimum pH for the formation of zinc-hydroxy-sulphates appears to be in the range 6.2 to 6.8. At a near-neutral pH, only low concentrations of zinc (<10 mg/ℓ) remain in solution. This implies that the bulk of the mine water acidity has to be satisfied, before the optimum process pH is achieved. There may be an imbalance in the process requirements for:

- Acidity Neutralisation
- Sulphate Precipitation

The typical situation, if any significant metal concentrations are present in the acid mine water, is that an excess of Zn^{2+} (in terms of SO_4 precipitation requirements) could be released into the water. The excess Zn would precipitate as either:

- zinc-hydroxy-sulphates
- zinc hydroxide

The nett effect of the excess Zn^{2+} is that a relatively large amount of zinc is consumed in the process of sulphate removal.

1.3.2 Partial Sulphate Removal

The Electrochemical process also has a constraint in terms of the fraction of sulphate, which can practically be removed. The sulphate ion in mine water is present in the hydrogen sulphate (HSO_4^-) and in the sulphate form (SO_4^-). The hydrogen sulphate and the sulphate anions are associated with different counter cations. In acid mine water, these counter cations may typically be H^+, $A\ell^{3+}$, Fe^{3+}, Fe^{2+}, Ca^{2+}, Mg^{2+}, Na^+ etc. The process development work done to date indicated that only the HSO_4^-/SO_4^{2-} associated with free acidity (H^+) and mineral acidity ($A\ell^{3+}$, Fe^{3+} etc) will be removed by the E/C process. The HSO_4^-/SO_4^{2-} associated with counter-cations such as Ca^{2+}, Mg^{2+}, Na^+ etc will not be removed by the E/C process.

The E/C process is therefore only applicable to acid mine water and will not be effective in neutral or pre-limed mine waters.

1.4 Electrical Power Requirements

The Electrochemical reaction is driven by electrical charge, as quantified in terms of Faraday's Law:

$M = (Z.I.t)/F$

Where $M =$ mass of metal (zinc) released over a time (gm)
$Z =$ equivalent mass of metal
$=$ (gram/mole) /valence
$I =$ electrical current (Amps)
$t =$ time (secs.)
$F =$ number of charges per equivalent mass = 96 500

In the case of a zinc electrode, the mass of zinc released by a current of 1000 A flowing for 1 second is as follows:

$$M = (65.4/2)(1000)(1)(96\ 500)$$
$$= 0.336 \text{ gram of zinc}$$

The relationship between electrical charge and electrical current:

C = $I \cdot t$

Coulomb = Ampere x Secs

The electrical power consumption in the Electrochemical reactor:

Power (W) = $V \cdot I = I^2 R$

Energy (J) = $V \cdot I \cdot T = I^2 R.t$

Where V = voltage across electrodes (volt)

I = current flow (ampere)

R = electrical resistance (ohm)

The electrical power input to an E/C reactor is typically expressed in terms of Coulomb/ℓ (or m^3) of mine water treated.

The power input, into the E/C reactor also has the effect of heating the water, due to the energy dissipation. For example, in a perfect E/C reactor with no side-reactions and operating at an electrical current of 1000 Ampere/L with a voltage differential of 6 12, the temperature increase would be as follows:

Power = $V I$

= 12 V x 1000 A = 12 000 W

Energy = VIt, for one second

= 12 000J = 12 kJ/sec

At a mine water feed rate of 0.5 ℓ/sec, the temperature increase:

±T = E/(Sh x Q)

Where AT = temperature increase (°C)

E = energy input (kJ)

Sh = specific heat of water

= 4.22 kJ/kg/°K

Q = flow rate (kg/sec)

= 0.5 kg/sec

Thus AT = 12 kJ/(4.22 x 0.5)

= 5.7 °C

The following table gives an indication of the potential temperature increase, associated with different energy inputs:

Electrical dose (Coulomb/ℓ)	**Electrical voltage**		
	4V	**8V**	**12V**
1 000	0.9	1.9	2.8
2 000	1.9	3.8	5.7

Electrical dose (Coulomb/ℓ)	**Electrical voltage**		
	4V	**8V**	**12V**
3 000	2.8	5.7	8.5
4 000	3.8	7.6	11.4
5 000	4.7	9.5	14.2
10 000	9.5	18.9	28.4

The temperature increase can therefore, be conveniently manipulated by the voltage at which the E/C reactor operates.

2 PROCESS EVALUATION

The EcoDose process was evaluated on the basis of pilot plant operations and results at Navigation Colliery. The pilot plant configuration is shown in **Figure 2** and incorporates the following features:

- Mine water feed tank.
- Mine water feed pump.
- Mine water feed flow meter.
- Electrochemical reactor with three parallel compartments.
- Electrical power supply
- Lamella clarifier with desludging pump
- Treated mine water filter

Photograms 1 to 6 depict a number of the pilot plant components.

The pilot plant power supply had a maximum capacity of 45 kW and provided an electrical current with reversible polarity. The electrical current was supplied in cycles, which lasted 16 secs. The power supply could be operated for 25 % (4 secs out of 16 secs), 50 % (8 secs out of 16 secs), 75 % (12 secs out of 16 secs), or 100 % (16 secs out of 16 secs) of the maximum installed capacity. The polarity could be reversed at a selected frequency. The polarity reversal is important to prevent the passivation of the zinc electrodes.

The downstream clarification/filtration processes dictated the pilot plant hydraulic capacity of 48 m^3/day. The Electrochemical reactor could be operated at a much higher flow rate.

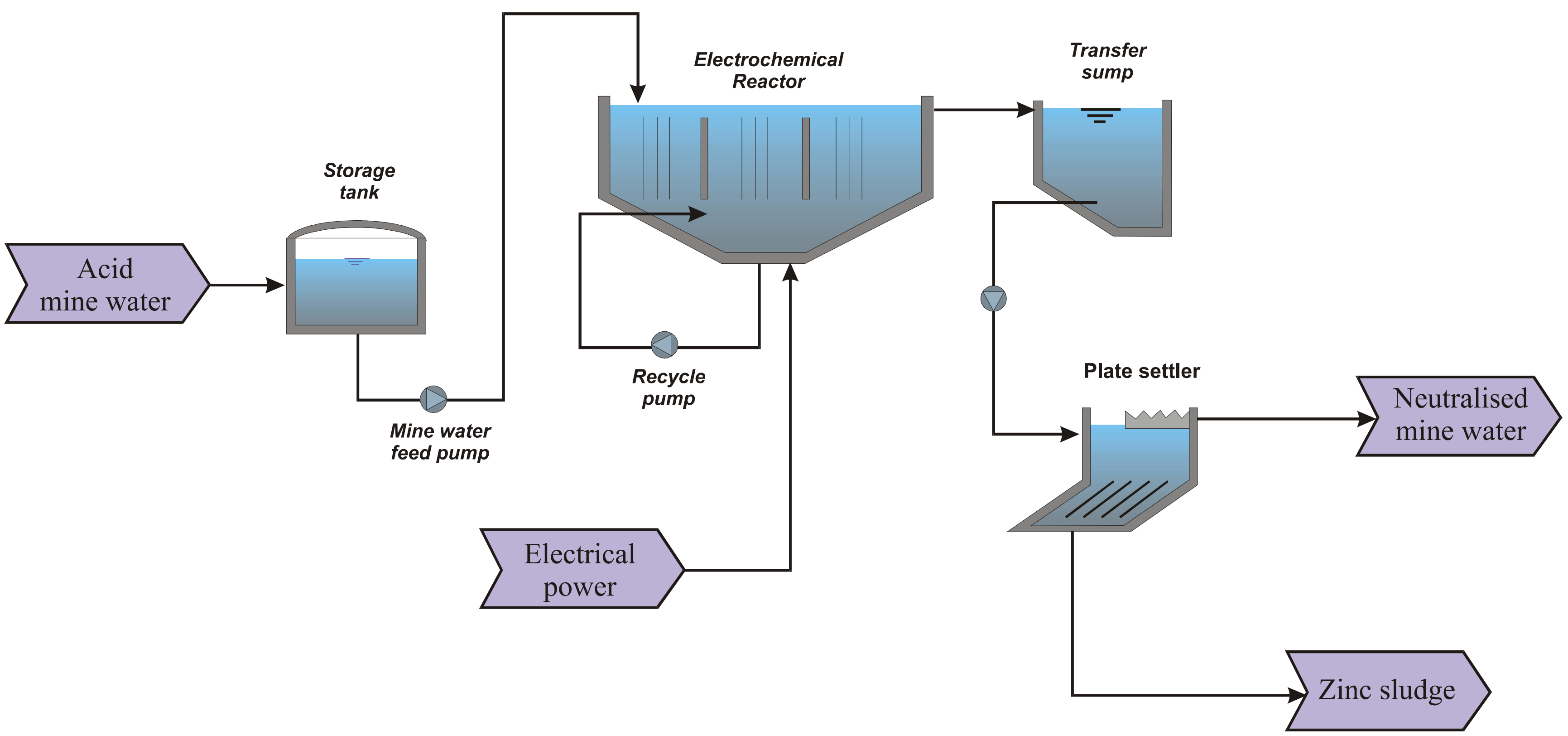
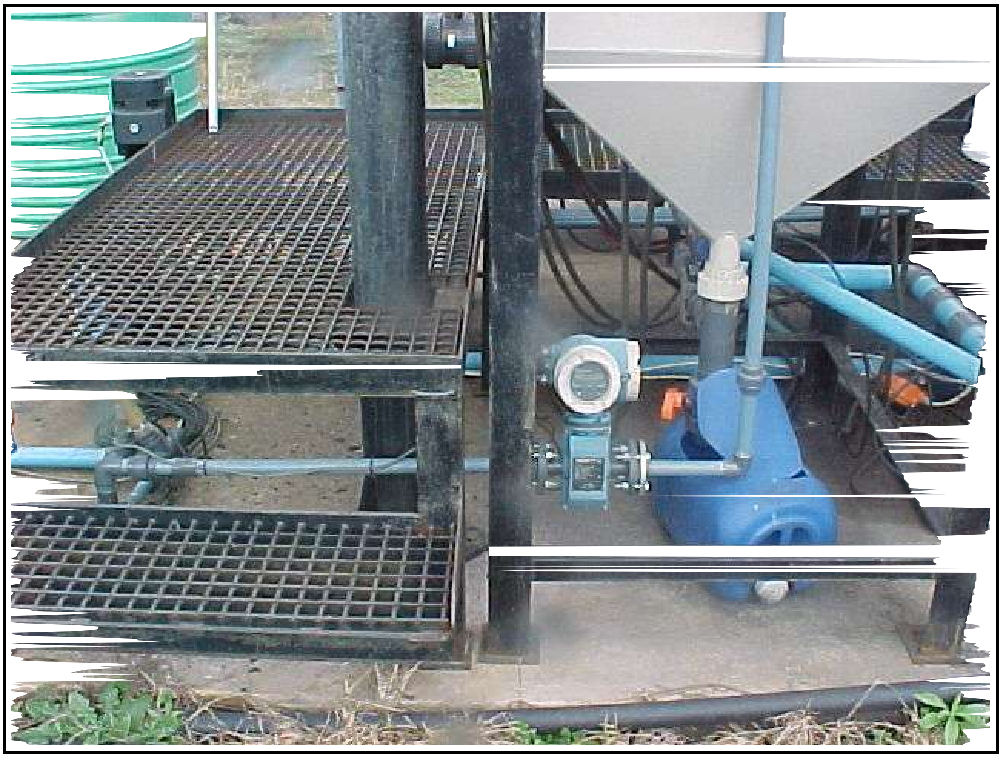


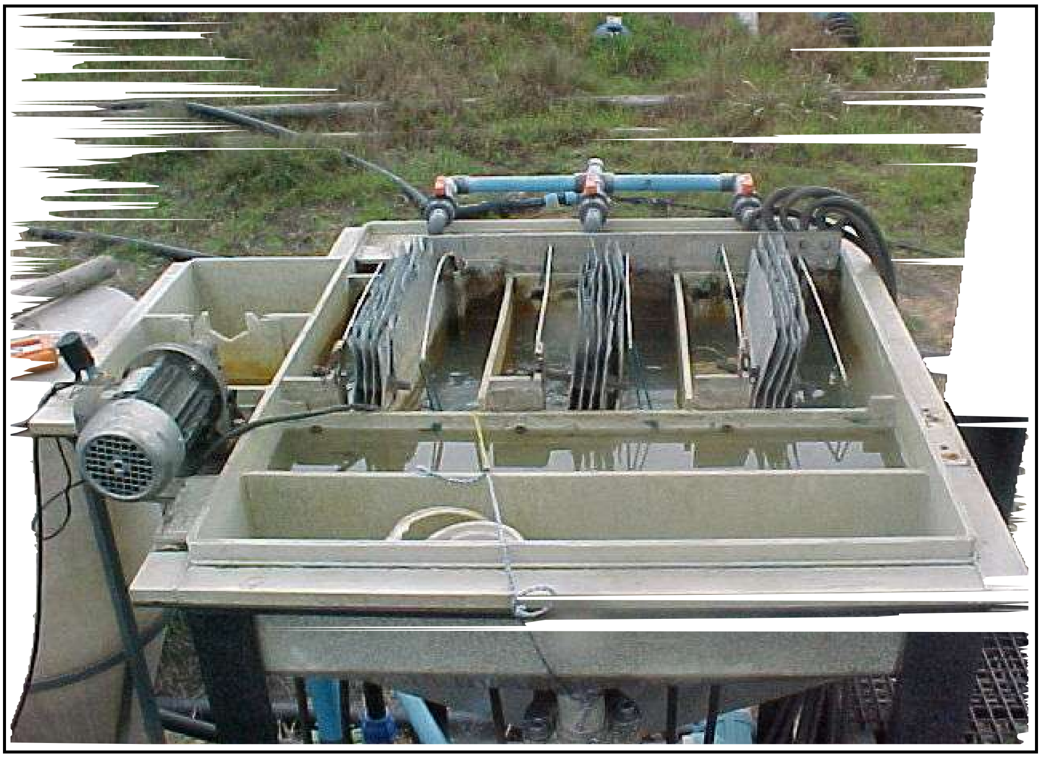
Figure 2 : Electro-Chemical Ecodose Process - Pilot Plant Flow Diagram

Photogram 1: General overview of the Ecodose pilot plant installation

Photogram 2: Mine Water Feed Tank



Photogram 3: Mine Water Feed Pump



Photogram 4: Details of the Electro-chemical reactor

Photogram 5: Sludge settler treating the Electro-chemical reactor overflow

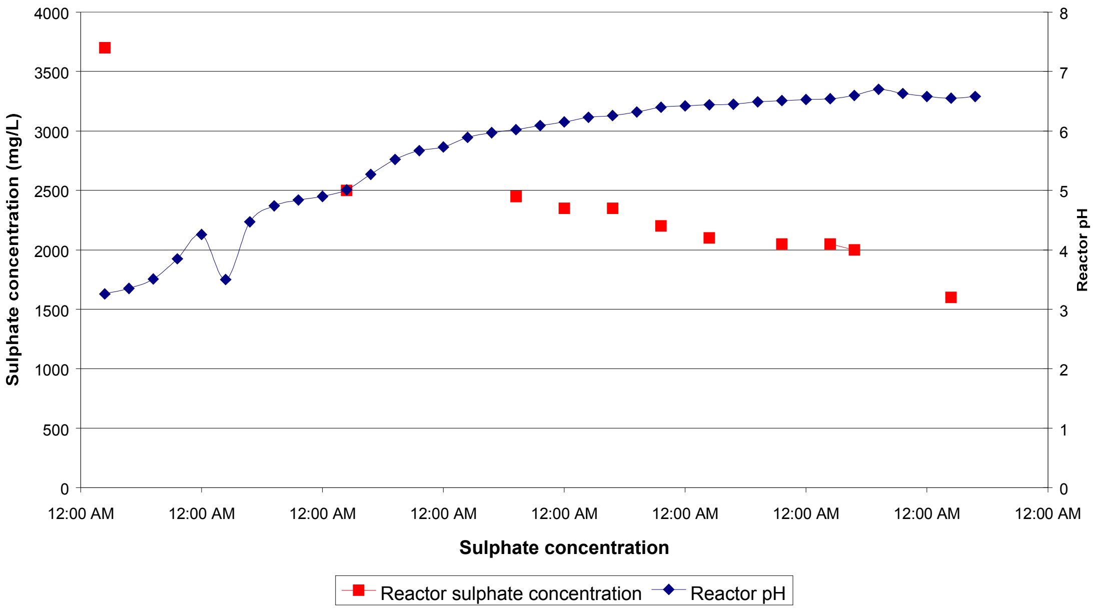
Photogram 6: Lamella settler for polishing treatment of overflow

2.1 Start-up of Process

The E/C process could be started up very rapidly. The normal start-up procedure is to first operate in a batch mode, before switching to a continuous flow mode. The batch operation was continued until the optimum pH for zinc-hydroxy-sulphate is reached, before switching to the continuous flow mode of operation.

The process could be started up within 1 to 2 hours – refer to **Figure 2.1**.

Figure 2.1 - Ecodose Process Start-up Response (11 July 2000)



2.2 Process Loading

The EcoDose Process was operated successfully in July 2000 and November 2000 on two (2) types of mine water:

- Acidic Schoongezicht mine water
- A blend of Schoongezicht mine water(90 %) and Toe Seepage (10 %)

The reactor was regularly operated at the maximum feed capacity of 48 m^3/day. The feed water sulphate concentration was typically in the range of 2 400 to 2 600 mg/ℓ for Schoongezicht mine water and in the range 3 400 to 3 600 mg/ℓ for the Schoongezicht/Toe Seepage blend water.

2.3 Process Conditions

2.3.1 Pre-treatment

The acid Mine water did not receive any form of pre-treatment.

2.3.2 Electrical Power Addition

The power input was dictated by the voltage across the electrodes, the electrical current and the fraction of time that the power was on. The electrical current was determined by the conductivity of the mine water between the electrode plates.

The charge input to the process can be expressed in terms of Coulombs per unit volume of mine water treated. The electrical charge input was varied between 2 000 to 10 000 Coulombs/ℓ for the different experiments. The electrical charge input can be related to the influent sulphate concentration, although the relationship differed for different experiments. This is illustrated below:

✜ **Pilot Test on 14 November 2000**

The acid Mine water feed was on average 42 m^3/day with a feed SO_4 concentration of 2600 mg/ℓ. The E/C reactor was operated at a pH = 6.4 to 6.6.

The electrical charge dose fluctuated between 1800 and 2200 Coulomb/ℓ. The sulphate was removed down to a concentration of 1 600 to 1 700 mg/ℓ. The calculated ratio of Zn/SO_4 was only 0.63.

✜ **Pilot Test on 16 November 2000**

The acid Mine water feed rate was 8.1 m^3/day with a feed SO_4 concentration of 3 800 to 4 000 mg/ℓ. The E/C reactor was operated at a pH = 5.4 to 5.6.

The electrical charge dose was relatively constant at 10 000 Coulomb/ℓ. The sulphate was removed down to a concentration of 1 200 to 1 600 mg/ℓ. The calculated Zn/SO_4 ratio was 1.4 to 1.6.

The electrical charge input has to be modified to pace the feed mine water flow and sulphate load. The Zn/SO_4 ratio differs for various mine water and operating conditions (specifically the reactor pH). This latter factor makes the prediction of the electrical charge dose requirements difficult.

2.3.3 Temperature

The process was not sensitive to the feed mine water temperature. The reactor temperature did increase, depending on the electrical power input to the process. This is again demonstrated by comparing two (2) different pilot plant tests.

✜ **Pilot Test on 14 November 2000**

The average energy input to the process was 5.85 kW, which equated to 12.1 kJ/ℓ treated. The feed mine water temperature was 25 °C and the reactor temperature operated at steady state was 29 °C.

✜ **Pilot Test on 16 November 2000**

The average energy input to the steady state process was 9.9 kW, which equated to 106 kJ/ℓ treated. The feed water temperature was 25 °C and the reactor temperature during the last part of the test was as high as 49 °C.

The E/C reaction could therefore increase the mine water temperature, depending on the energy input.

2.3.4 Toxicity/inhibition

No form of toxicity or inhibition was recorded in any of the pilot plant tests.

The only malfunction was recorded during a pilot test performed on a highly polluted mine water (Toe Seepage). The Toe Seepage conductivity was so high (> 1 200 mS/m), that the electrical control system interpreted the high electrical currents as a short circuit and the plant was automatically shut down.

2.4 Process Control

The E/C process control is based on the synchronisation of two aspects:

- The feed mine water flow and the associated sulphate load.
- The electrical charge dose as dictated by the electrical current.

It is also known that the zinc-hydroxy-sulphate precipitation is optimised at a pH in the range 6.4 to 6.8. The process control is designed to achieve the target optimum pH.

The process control strategy is, therefore, based on the following approach:

- Select an average mine water feed flow rate and associated sulphate load.
- Calculate the electrical charge requirements to achieve the optimum sulphate removal.
- Set the potential differential between the electrodes
- The electrical current (and by implication the electrical charge dose, Coulomb/ℓ) will stabilise at a certain level, mainly dictated by the reactor liquor conductivity and temperature.
- The Mine water feed rate is controlled from a feedback signal originated by the reactor pH meter, to maintain the target optimum pH level.

The use of process control instrumentation in the E/C reactor must also be approached with caution. The electrical fields set up in the E/C reactor will interfere with most process instrumentation, such as pH meters. The pH meter must therefore be installed on an isolated re-circulating side-stream.

2.5 Process Performance

The E/C process ability to neutralise mine water and precipitate sulphate is well proven. The process can, however, only remove the fraction of the total sulphate associated with free and mineral acidity.

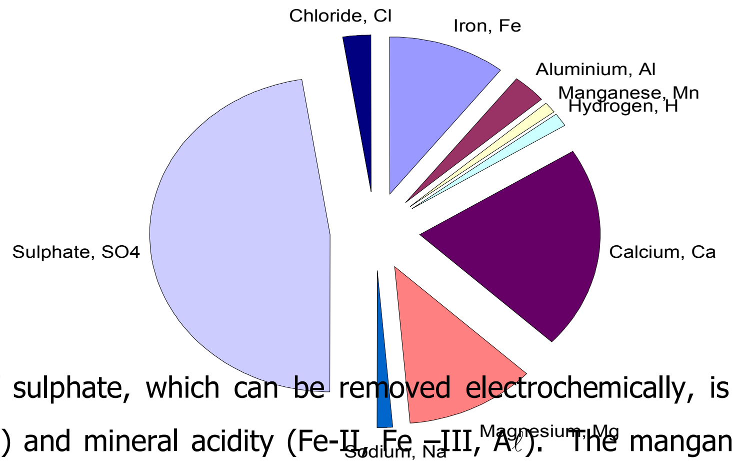
The ionic composition of a typical mine water used in the pilot plant testing is reflected below in terms of the major ionic species:

Table 2.5 (a) High calcium/magnesium mine water composition

Cations			Anions		
Species	**mg/ℓ**	**mequi/ℓ**	**Species**	**mg/ℓ**	**mequi/ℓ**
Calcium	470	23.5	Sulphate	2550	53.1
Magnesium	158	13.0	Chloride	103	2.9
Sodium	38	1.7	Bicarbonate	0	0
Iron -II	33	1.2			
Iron -III	200	10.8			
Aluminium	31	3.4			
Manganese	28	1.0			
Hydrogen	pH = 2.75	1.8			
Total		**56.4**			**56.0**

The ionic composition is also shown graphically in **Figure 2.5 (a)** below:

Figure 2.5(a) - Chemical composition of the Acid Mine Water



The fraction of sulphate, which can be removed electrochemically, is associated with the free acidity (H^+) and mineral acidity (Fe-II, Fe -III, Aℓ). The manganese will typically not precipitate under the E/C reactor operating conditions, and is excluded from the estimate of the fraction of sulphate which could be precipitated. The estimated fraction of sulphate, which could precipitate, is therefore:

$$SO_4 \text{ removal fraction} = (1.2 + 10.8 + 3.4 + 1.8)/56.4$$

$$= 30.5\ \%$$

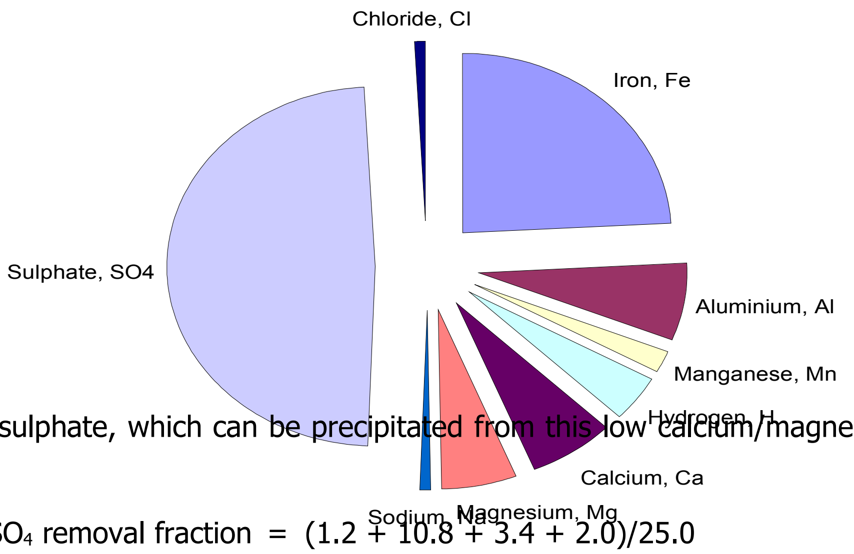
By contrast, if the acid mine water contains a relatively low calcium/magnesium content, a much higher sulphate fraction may be removed electrochemically. The ionic composition a low calcium/magnesium mine water is reflected below:

Table 2.5 (b) Low calcium/magnesium mine water composition

Cations			Anions		
Species	**mg/ℓ**	**mequi/ℓ**	**Species**	**mg/ℓ**	**mequi/ℓ**
Calcium	65	3.2	Sulphate	1 150	24.0
Magnesium	35	2.9	Chloride	15	0.4
Sodium	12	0.5	Bicarbonate	0	0.0
Iron -II	33	1.2			
Iron -III	200	10.8			
Aluminium	31	3.4			
Manganese	28	1.0			
Hydrogen	pH = 2.7	2.0			
Total		**25.0**			**24.4**

The ionic species composition is shown in **Figure 2.5 (b)**.

Figure 2.5(b) - Chemical composition of the Acid Mine Water



The fraction of sulphate, which can be precipitated from this low calcium/magnesium water is:

$$SO_4 \text{ removal fraction} = (1.2 + 10.8 + 3.4 + 2.0)/25.0$$
$$= 69\ \%$$

The performance of the E/C process is therefore very dependent on the ionic species composition of the feed Mine water.

The fraction of sulphate removed in the pilot plant tests are summarised below:

Date	Sulphate Concentration		Sulphate Removal (%)
	Feed (mg/ℓ)	Product (mg/ℓ)	
11 July 2000	3 700	2 000	46
13 July 2000	2 650	1 650	38
14 November 2000	2 650	1 650	38
15 November 2000	2 650	1 850	30
16 November 2000	3 460	1 200	65

The percentage sulphate removal is therefore quite variable, depending on the feed water composition.

2.6 Residue Generation

The E/C process generates a well-flocculated zinc-hydroxy-sulphate and metal hydroxide sludge. The composition of the sludge is dependent on two aspects:

- The metal composition of the feed acid mine water, specifically in terms of the iron and aluminium concentrations.
- The zinc : sulphate ratio in the zinc-hydroxy-sulphate precipitate.

The typical metal composition of the E/C treatment process sludge is reflected below in the sample/analysis done on the 14 November 2000:

Zinc = 256 g/kg solids

Iron = 82 g/kg solids

Aluminium = 33 g/kg solids

Manganese = 1.4 g/kg solids

The E/C sludge composition can also be calculated on the basis of a theoretical model, based on the chemical composition of the feed mine water. The sludge production calculation is based on the following mine water composition.

Table 2.6 Typical Acid Mine Feed Water Composition

Cations			**Anions**		
Species	**mg/ℓ**	**mequi/ℓ**	**Species**	**mg/ℓ**	**mequi/ℓ**
Calcium	470	23.5	Sulphate	2550	53.1
Magnesium	158	13.0	Chloride	103	2.9
Sodium	38	1.7	Bicarbonate	0	0
Iron -II	33	1.2			
Iron -III	200	10.8			
Aluminium	31	3.4			
Manganese	28	1.0			
Hydrogen	pH = 2.75	1.8			
Total		**56.4**			**56.0**

Assuming that all the iron and aluminium is precipitated and that the sulphate is precipitated as a $Zn_4 (OH)_6 SO_4$ precipitate, the sludge composition is as follows:

- $Fe (OH)_2$ = 53 mg/ℓ (1.3%)
- $Fe (OH)_3$ = 383 mg/ℓ (9.0%)
- $A\ell (OH)_3$ = 89 mg/ℓ (2.1%)
- $Zn_4(OH)_6 SO_4$ = 3723 mg/ℓ (87.6%)

4247 mg/ℓ

The sludge flocculates well and also settles readily in conventional gravity solids separation devices. Sludge settling rates of 1.2 – 1.3 m/hour have been reported.

The composition of the sludge is very important from a zinc recovery point of view. It is desirable to limit other contaminants, such as the other metal hydroxides, in the sludge. This will reduce the amount of low value constituents, which must be thickened, dewatered, dried and transported to a zinc refinery.

The observed zinc : sulphate mass ratio of the sludge varied substantially, as reflected below:

Date	**Zn : SO_4 mass ratio in sludge**
11 July 2000	1.73
12 July 2000	3.53
13 July 2000	2.50
14 November 2000	0.61

15 November 2000	0.66
16 (a) November 2000	0.70
16 (b) November 2000	1.47

The Zn : SO_4 mass ratio was substantially lower during the November 2000 pilot testing, compared to the July 2000 pilot testing campaign. The lower zinc requirement is encouraging, but the mechanism to consistently achieve this was not established.

Note that the theoretical minimum zinc : sulphate mass ratio, assuming a $ZnSO_4$ precipitate, is 0.72.

Earlier observations also indicated that the zinc-hydroxy-sulphate might not be stable during prolonged sludge storage. It is therefore essential to rapidly dewater and dry the sludge for further zinc recovery.

2.7 By-product Generation

A potentially valuable by-product is generated in the form of hydrogen gas, H_2. The amount of hydrogen gas formed is directly proportional to the electrical charge dose. For every 1 000 A flowing for 1 second, the mass of H_2 released is calculated as follows:

$$M = (^1/_1)\ (1000)(1)\ /\ 96500$$
$$= 10.4\ \text{mg},\ H_2$$

3 FULL-SCALE TREATMENT PLANT CONCEPTUAL DESIGN AND COSTING

The potential application of the treatment technology to a full-scale installation was investigated, based on the results of the pilot plant work.

The potential application was taken to be one of the three regional mine water treatment facilities proposed by one of the previous Coaltech projects (Task 6.8.2). The specific regional facility is for one of Acid Mine Water in the Brugspruit Catchment, which is projected to have a modular size of 5 Mℓ/day at the following projected average mine water quality:

Conductivity = 560 mS/m
pH = 2.8
TDS = 4 210 mg/ℓ
Calcium = 430 mg/ℓ
Magnesium = 50 mg/ℓ
Sodium = 50 mg/ℓ
Sulphate = 2 530 mg/ℓ
Chloride = 55 mg/ℓ
Iron ≤ 200 mg/ℓ

Aluminium ≤ 75 mg/ℓ
Manganese = 27 mg/ℓ

The conceptual integrated sulphate removal process is shown in **Figure 3-C**.

The conceptual treatment process design was based on achieving the following product water quality targets:

- Sulphate, SO_4 ≤ 1 600 mg/ℓ (37 % removal)
- pH range 6.5 – 7
- Iron ≤ 1 mg/ℓ
- Aluminium ≤ 1 mg/ℓ
- Manganese ≤ 10 mg/ℓ

-

3.1 Process Design Criteria

3.1.1 Iron Oxidation Reactor

The iron oxidation reactor size was based on an iron oxidation rate, using air as oxidant of:

$R = 0.2$ kg Fe/m^3/hour

The oxygen supply was based on the stoichiometric amount of 0.15 kg O_2/kg Fe, with and oxygen transfer efficiency of 5 %.

3.1.2 Primary Electrochemical reactor

The primary Electrochemical reactor was designed to supply 80 % of the total electrical dose of 3 610 Coulomb/ℓ.

Each individual Electrochemical cell was configured to have 6 zinc plates, each with dimension 1 300 mm long and 1 040 mm deep. The effective retention time was established to be 15 – 20 minutes. A total of 40 Electrochemical cells were included in the primary reactor.

3.1.3 Primary Thickener

The primary Electrochemical thickener was designed to accept the following hydraulic and solids loading rates:

Upflow velocity = 0.8 m/hour
Solids loading = 9 – 10 kg SS/m^2/day

3.1.4 Secondary Electrochemical Reactor

The secondary Electrochemical reactor was designed to supply 20 % of the total required electrical dose of 360 Coulomb/ℓ.

The individual Electrochemical reactor cells were identical to the primary Electrochemical reactor cells. The effective residence time was set to be 4 – 6 minutes. The secondary Electrochemical reactor contained 20 cells.

3.1.5 Secondary Thickener

The secondary Electrochemical thickener was designed to accept the following hydraulic and solids loading rates

Upflow velocity = 0.8 m/hour
Solids loading = 350 kg SS/m^2/day

3.2 Treatment Process Description

3.2.1 General Process Description

The practical implementation of the Ecodose Electrochemical Process requires a number of treatment steps for partial sulphate removal from acid mine water. These include:

- Iron oxidation

- pH adjustments to allow selective precipitation of iron and aluminium metals
- Sulphate precipitation as a zinc – hydroxy – sulphate
- Dewatering and drying of the zinc containing slurry

The overall process flow diagram is shown in **Figure 3.2.1(a)-C** and the plant layout is shown in **Figure 3.2.1 (b)-C.**

3.2.2 Iron Oxidation

The first process step involves iron oxidation in a packed bed biological reactor. The ferrous iron must be oxidised to ferric iron to allow effective removal in the first Electrochemical reactor. If the iron remains in the ferrous form, it may not be effectively removed with the other metals. This may result in contamination of the zinc sludge, which will decrease the value of the zinc sludge for the recovery of zinc.

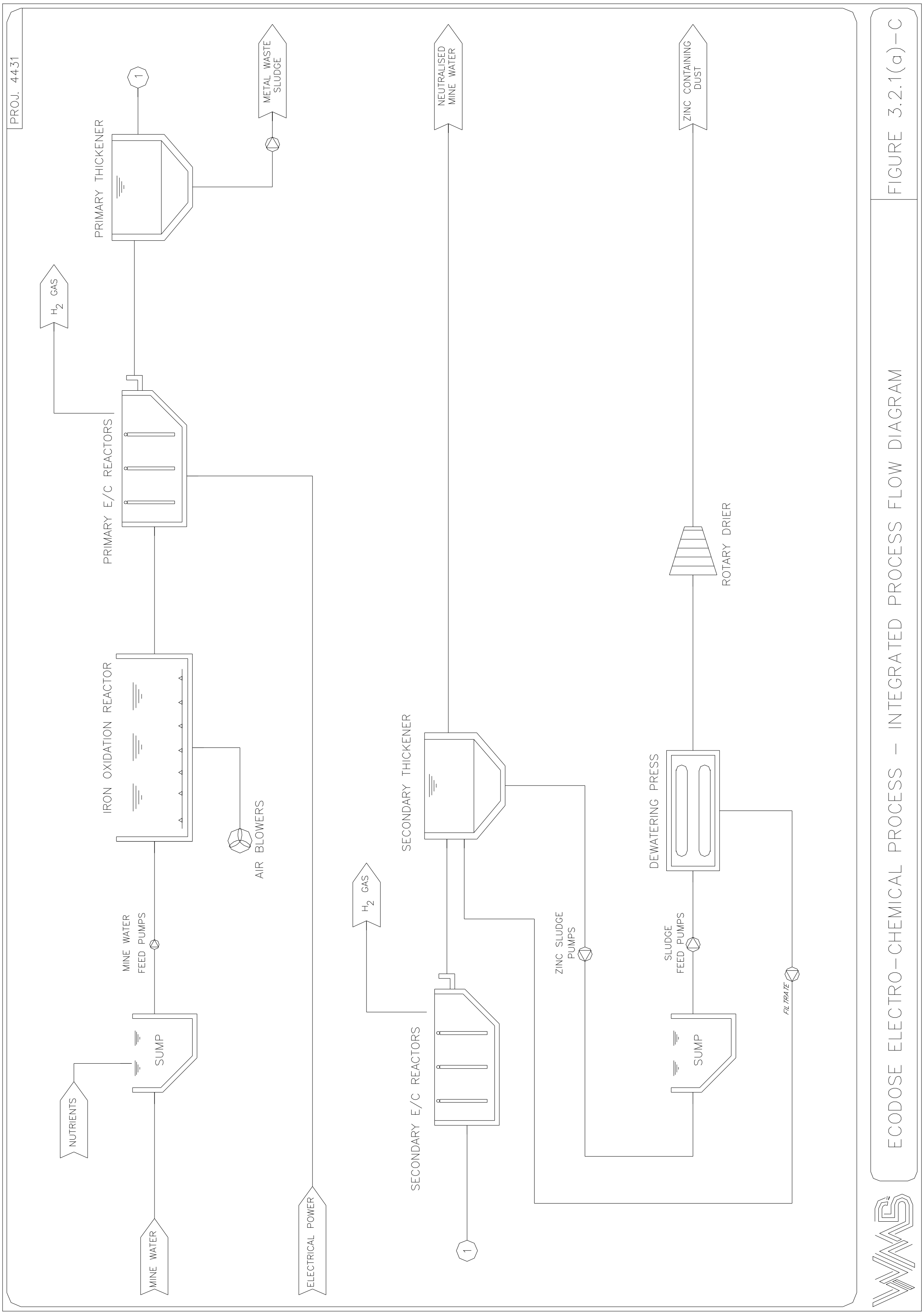
The iron oxidation reactor is an aerated packed bed reactor. The reactor is packed with a synthetic media, to allow attached biological growth. The media must have a high specific surface area (m^2/m^3) to maximise the opportunity for microbiological growth. Nutrients in the form of urea (nitrogen source) and phosphoric acid (phosphorus source) are added to further stimulate microbiological growth.

The iron oxidation reactor is aerated using medium bubble diffusers, which distribute air over the entire floor to allow even aeration.

Figure 3.2.2 – C shows the process flow diagram for the iron oxidation step.

3.2.3 Primary Electrochemical Reactor

The first Electrochemical reactor is designed to neutralise acidity and elevate the pH to a point where the ferric iron and aluminium precipitate as hydroxides. The target pH is 5.5. The objective is to preferentially remove the metals from the acid mine water, to minimize the downstream contamination of the zinc-hydroxy-sulphate with other metal hydroxides.



ECODOSE ELECTRO-CHEMICAL PROCESS – INTEGRATED PROCESS FLOW DIAGRAM

FIGURE 3.2.1(a)–C

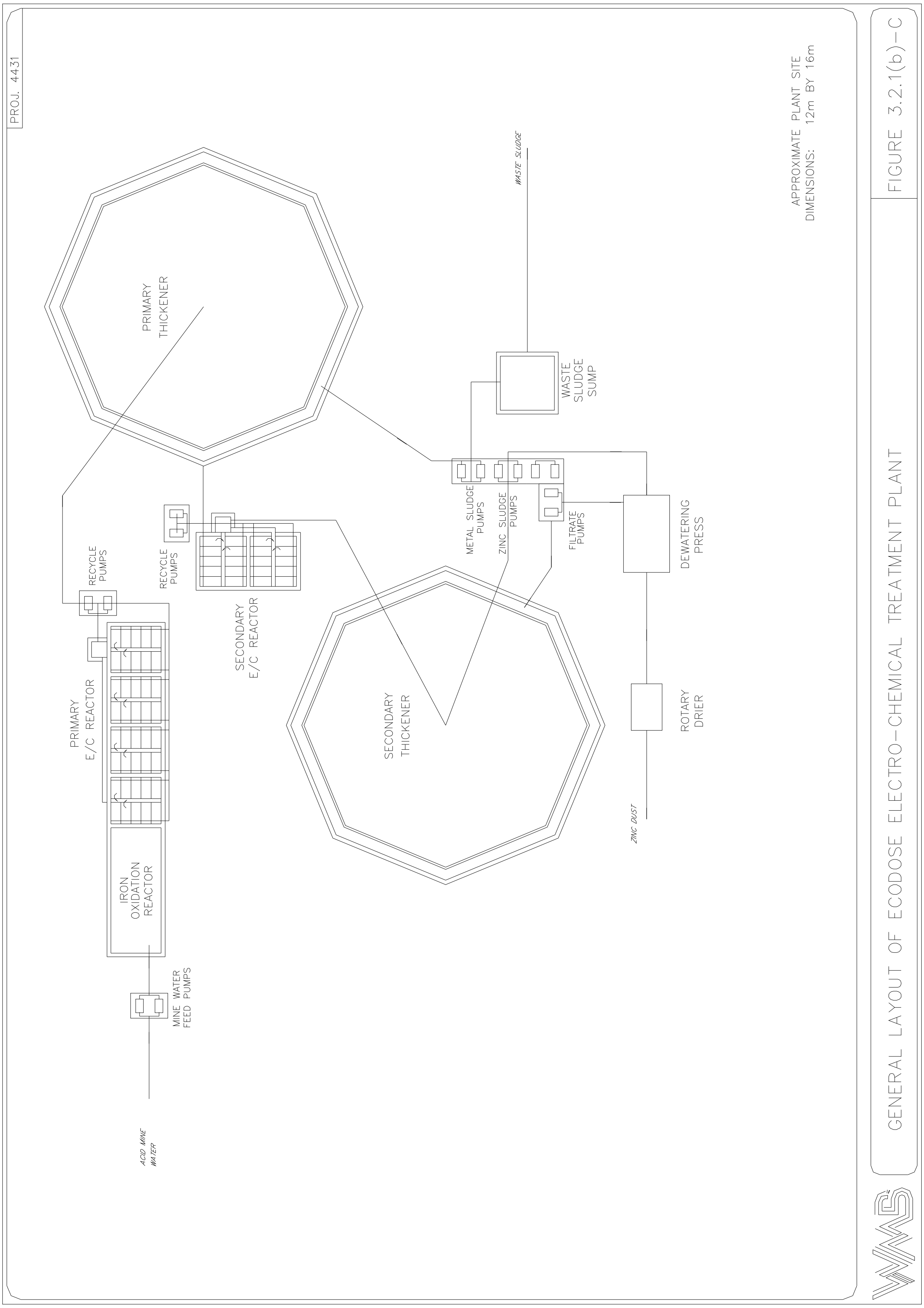
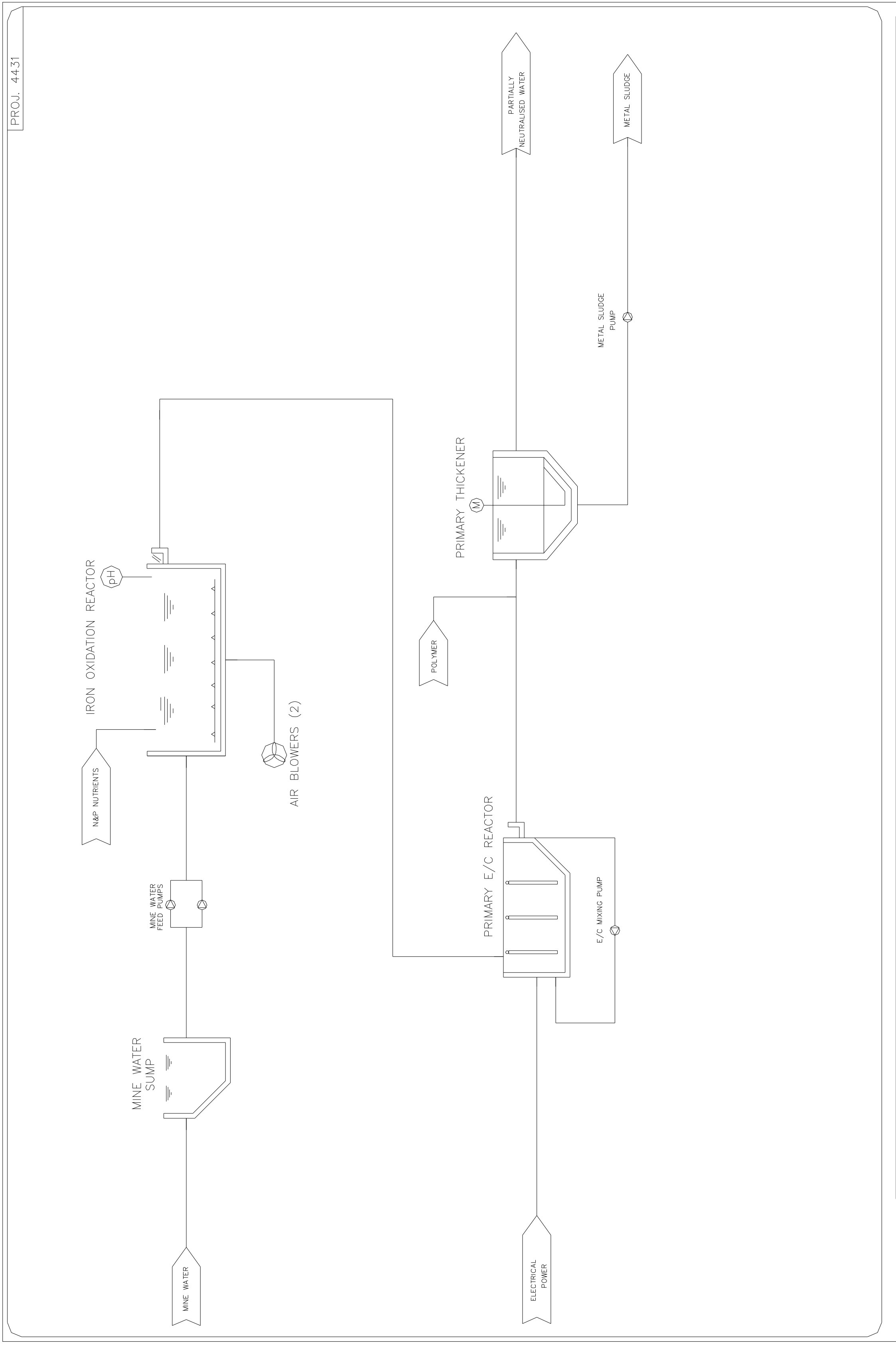


FIGURE 3.2.1(b)-C

GENERAL LAYOUT OF ECODOSE ELECTRO-CHEMICAL TREATMENT PLANT



ECODOSE– PRIMARY STEP NEUTRALISATION IN E/C TREATMENT PROCESS

FIGURE 3.2.2–C

The primary Electrochemical reactor incorporates 80 individual cells. Each cell is equipped with six (6) zinc plates hanging from the electrical contacts. The electrical current flow in alternating the mode to the zinc plates to drive the Electrochemical reactions.

Each bank of 10 cells is equipped with a recycle pump to maintain good contact between the mine water and the zinc electrodes. The contents of the cells is also kept mixed to prevent the local deposition of metal hydroxide solids.

Hydrogen gas must be vented from the reactor to prevent the risk of explosion. The gas could potentially be recovered and sold as a valuable byproduct.

The Electrochemical reactor overflows to a primary thickener. The predominantly metal hydroxide solids are separated from the partially neutralised mine water. The metal hydroxide sludge is pumped to disposal.

Figure 3.2.2-C shows the process configuration of the primary Electrochemical process with associated reactor and thickener.

3.2.4 Secondary Electrochemical Reactor

The partially neutralised mine water is further treated in a second stage Electrochemical reactor. The objective of the second stage Electrochemical reactor is the formation and removal of a relatively pure form of zinc-hydroxy-sulphate. This will allow efficient further dewatering and drying of the zinc containing sludge.

The secondary Electrochemical reactor incorporates 20 individual cells. Each cell is equipped with six (6) zinc plates hanging from electrical contacts. The electrical current flows in alternating mode to the zinc plates to drive the Electrochemical reaction.

Each bank of 10 cells is equipped with a recycle pump to maintain good contact between the mine water and the zinc electrodes. The contents of the cells is also kept mixed to prevent the local deposition of solids.

Hydrogen gas must be vented from the reactor to prevent the risk of explosion. The gas could potentially be recovered and sold as a valuable byproduct.

The Electrochemical reactor overflows to a secondary thickener. The predominantly zinc-hydroxide-sulphate sludge is separated from the neutralised mine water. The zinc containing sludge is pumped to the solids processing facility.

The process configuration for the secondary Electrochemical reactor and thickener is shown in **Figure 3.2.4-C**.

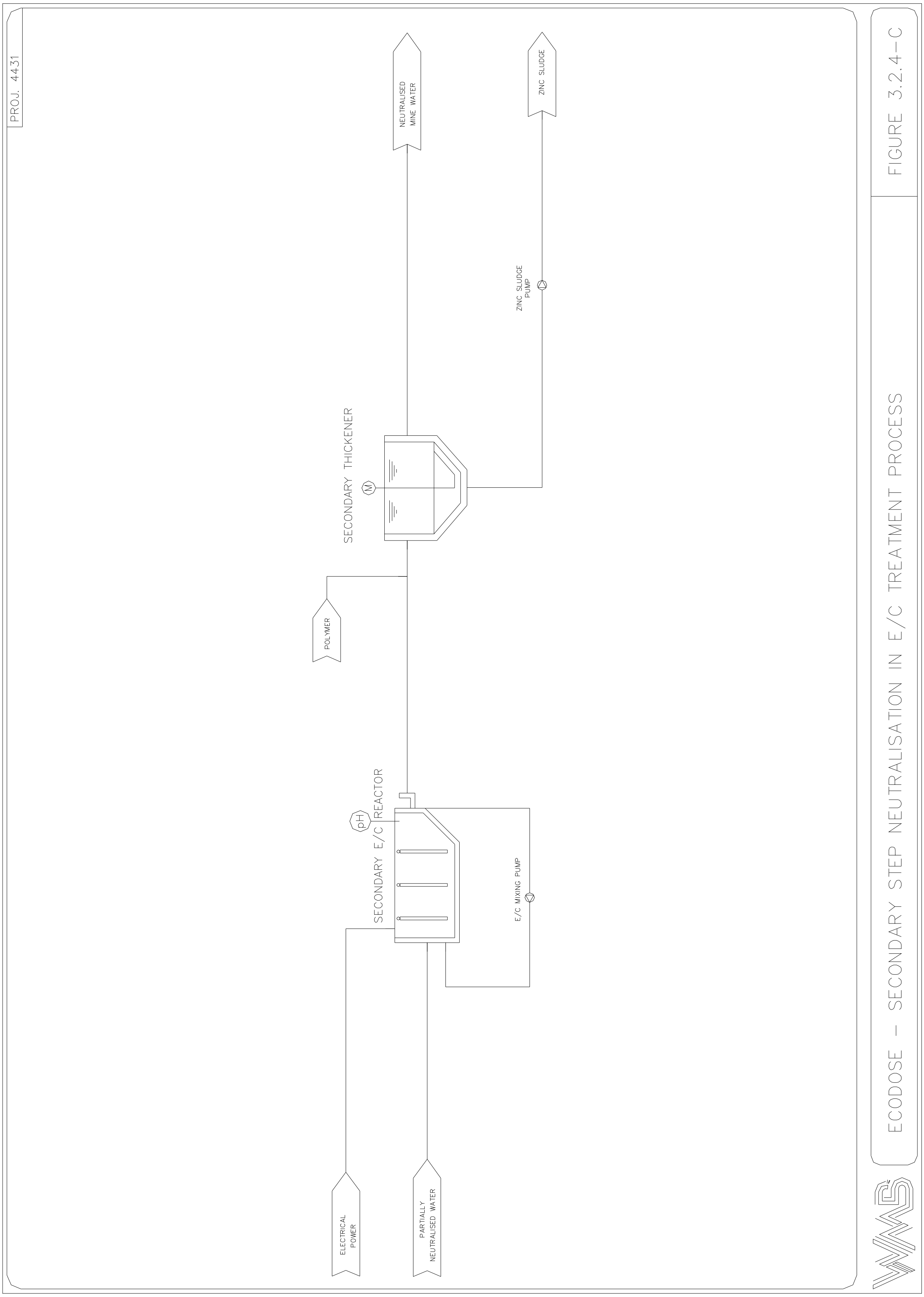


FIGURE 3.2.4-C

ECODOSE – SECONDARY STEP NEUTRALISATION IN E/C TREATMENT PROCESS

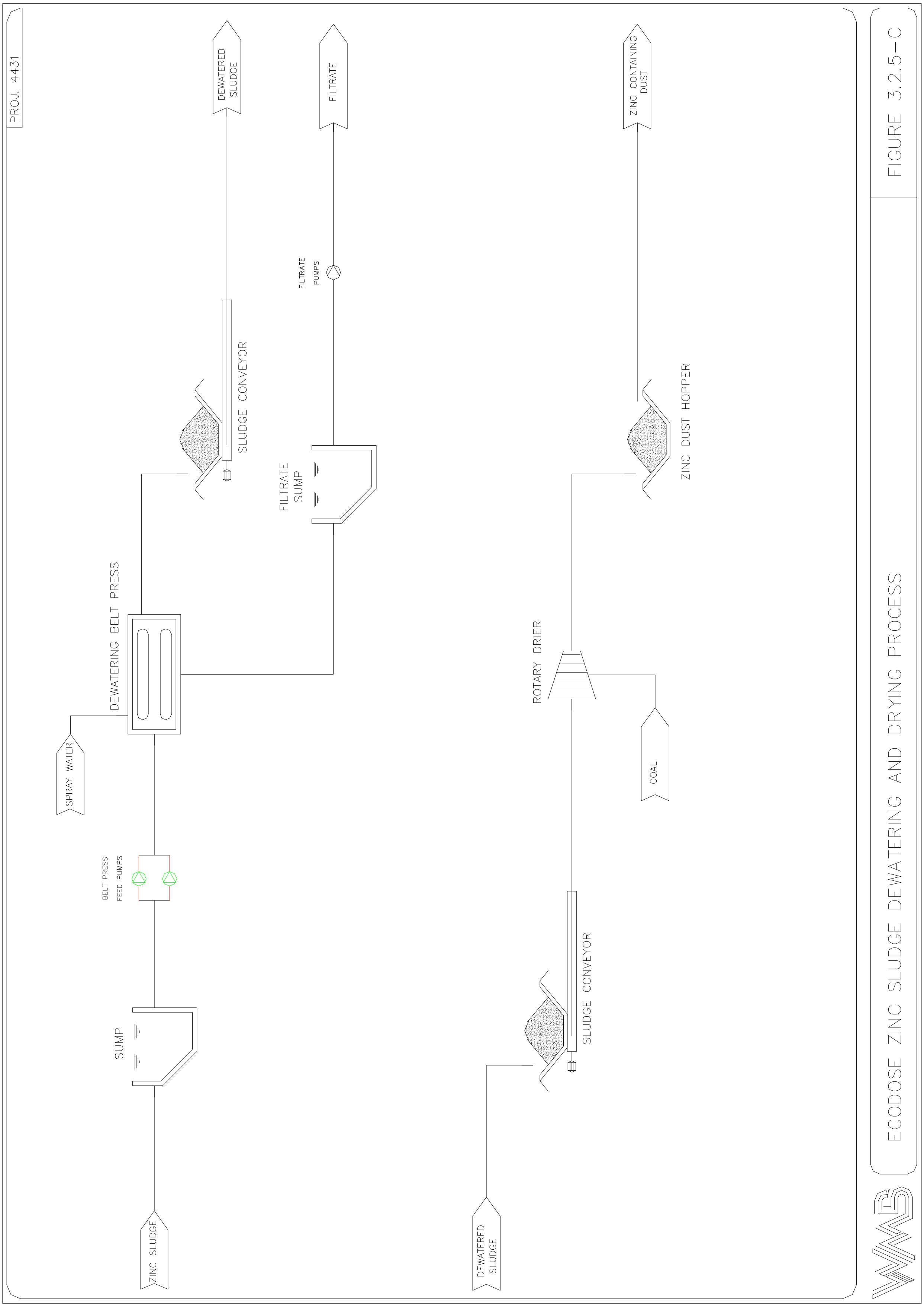


FIGURE 3.2.5–C

ECODOSE ZINC SLUDGE DEWATERING AND DRYING PROCESS

3.5 Operating and Maintenance Cost

The operating and maintenance costs were calculated to include the major components of:

- Chemicals
- Electrical power
- Operating personnel and labour
- Maintenance and repair

The operating and maintenance costs reflect the price indices of January 2002.

3.5.1 Chemical Usage

The chemical usage is estimated on the basis of the average mine water flow and load conditions.

The estimated chemical usage is summarised below:

Chemical type	Consumption (kg/day as pure)	Unit cost (R/kg as chemic)	Daily cost R/day
Primary thickener - polymer	25	35	875
Secondary thickener – Polymer	25	35	875
Dewatering press - polymer	156	35	5 471
Zinc metal	1 213	7.6	9 221
Hydrogen gas	185	0	0
Total			**16 442**

It was assumed that 80 % of the zinc is recovered from the zinc-hydroxy-sulphate sludge. No credit was given for the sale of hydrogen gas.

3.5.2 Electrical power

The installed power and the power drawn from each major individual mechanical equipment item were estimated. The electrical power cost was estimated using a unit rate of :

0.12 R/kWhr

The estimated installed power and the daily energy cost are summarised below:

Equipment item	Number installed	Installed power kW/unit	Operating hours/day	Daily power cost R/day
Mine water feed pumps	2	17	24	39.17
Aeration blowers	2	7	24	17.28
Primary E/C reactor power supply	1	667	24	1 920.00
Primary E/C reactor mixer pumps	2	5	24	12.67
Primary thickener bridge	2	1	24	6.91
Metal sludge pumps	2	5	24	12.67

Equipment item	Number installed	Installed power kW/unit	Operating hours/day	Daily power cost R/day
Secondary E/C reactor power supply	1	167	24	480.00
Secondary E/C reactor mixer pumps	2	5	24	12.67
Secondary thickener bridge	1	1	24	3.46
Zinc sludge pumps	2	7	2	1.44
Sludge dewatering press	1	37	24	85.25
Sludge drier	1	45	24	103.68
Sludge conveyor belts	2	3	24	17.05
Filtrate/centrate pumps	2	7	24	16.13
Total				**2 728**

3.5.3 Operating personnel and labour

The cost associated with operating personnel and labour was based on a reasonable assessment of the staffing requirements for a treatment plant.

The proposed categories of operating personnel and labour and associated monthly cost are summarised below:

Personnel Category	Number	Unit Cost (R/Month)	Monthly Cost (R/Month)
Plant superintendent	1	9 000	9 000
Process operators	2	6 500	13 000
Process assistants	1	4 500	4 500
Labourers	3	2 000	6 000
Total cost			**R32 500**

3.5.4 Maintenance cost

Planned and preventative maintenance and associated repair work are costed as a fraction of the capital investment into different components of the plant. The estimated annual maintenance cost is summarised below:

Infrastructure component	Capital expenditure R	Annual allowance for maintenance %	Annual maintenance cost R/year
Civil & building works	4.924	0.5	24 620
Mechanical equipment	6.646	2.0	132 920
Electrical & instrumentation	2.460	3.0	73 800
Total	**R14.030**		**R231 340**

3.5.5 Coal

Coal would be utilised as heat source for the drying of the zinc sludge. The coal cost is estimated as follows:

- Coal consumption = 3.8 ton/day
- Coal unit cost= R100/ton pebble coal
- Daily cost = R380/day

3.5.6 Transport

The dried zinc-hydroxy-sulphate sludge is transported to a zinc refinery located in Gauteng. It is assumed that the transport cost would be as follows

- Zinc-containing dry sludge = 31 ton/day
- Transport rate = R0.37/km-ton (100 km one way)
- Daily cost = R1 156/day

3.5.7 Operating and Maintenance summary

The different operations and maintenance cost component are summarised as follows:

Cost component	Daily cost (R/day)	Unit cost (R/m^3)
1. Chemicals	16 442	3.29
2. Electrical power	2 728	0.55
4. Operating personnel	1 068	0.21
5. Maintenance	633	0.13
6. Coal	380	0.07
7. Transport	1 156	0.23
Total		**4.48**