



TRANSFER OF O₂ FROM AIR TO MINERAL SLURRIES IN A RUSHTON TURBINE AGITATED TANK

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ABSTRACT

Oxygen transfer from dispersed air in agitated slurries represents an important unit operation in the minerals processing industry. Gold leaching, iron precipitation and especially bio-oxidation come readily to mind. However, little engineering information is available on the effect of mineral solids on such O₂ transfer. Engineering literature on the subject, i.e. information which relates slurry density and energy input to O₂ transfer, is reviewed and compared to current experimental work. The latter was carried out with minus 50 micron sand/water slurries in a 64 l laboratory tank of standard proportions, equipped with a standard Rushton turbine and variable speed drive with torque meter.

Results confirm an increasing impairment of O₂ transfer with increasing slurry density. The volumetric mass transfer coefficient, $k_L a$, in a 10 vol.% slurry shows a decrease of 30% compared to the $k_L a$ in water. At 40 vol.% sand this decrease is near 60%, depending somewhat on superficial gas velocities. Power needed for equivalent oxygen transfer increases even more steeply with slurry density. Experimental results are presented and their significance for slurry bio-oxidation discussed.

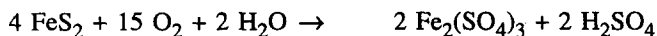
Keywords

Aeration, mineral slurries, Rushton turbine, bio-oxidation

INTRODUCTION

In many industrial processes the transfer of oxygen from the gas phase to the liquid phase plays an important role. Oxygen is required as an electron acceptor in acidic solutions. In biohydrometallurgy, aeration of slurries allows bacteria to grow. It is a major factor in both capital and operating cost [1]. Therefore, much research is done on reactor design, means of air introduction, power input to effect the aeration needed, etc. The present work aims specifically at providing design factors for agitated bio(oxidation)reactors. In mineral industry bioreactors, oxygen is usually supplied by air, and this type of bioreactors is (further) characterized by requiring the transfer of large quantities of oxygen. This results from the need to oxidize pyrite (FeS₂) or arsenopyrite (AsFeS), containing small amounts of gold. In conventional gold milling, the ores are leached with cyanide. But if gold is incorporated in a metal sulphide, then the lattice may block the way for the cyanide and the gold recovery may be as low as 20%. To free the “refractory” gold, it is necessary to destroy the lattice. This can be accomplished by oxidation of the sulphides. Several methods of oxidation are available, the oldest one involving roasting. Bio-oxidation is a recent engineering development.

In general, for pyrite and arsenopyrite the overall bio-oxidation reactions are:



As can be seen, considerable oxygen is required for the oxidation of the metal sulphides. Biotechnological treatment of gold containing ores prior to cyanide leaching has resulted in a large increase in gold recovery. Marchant *et al.* [2] reported a gold recovery of 21% on untreated concentrate and a recovery in excess of 90% after bio-oxidation. In the Bacterial Oxidation (Biox®) process, developed by Genmin, the recovery for the concentration of refractory gold ores increased from 35 to 97% in a pilot plant. Commercial Biox® plants are now in operation in several locations.

For instance, Dew and Godfrey [3] described a 580 m³ reactor with a standard Rushton turbine and ring sparger for the Sao Bento Gold mine in Brazil. The reactor was designed to handle 8.35 tonnes of pyritic sulphur per day. The motor installed has a power of 336 kW, and the oxygen demand is 755 kg/h, introduced in the vessel by 10600 Nm³/h of sparged air. Aeration tests, performed after installation of the equipment, have shown that the required oxygen mass transfer coefficient of 0.045 s⁻¹ was easily achieved. Subsequent bioreactors, installed according to the Genmin Biox® process at Ashanti, have retained the Rushton turbine agitator, despite the increase in tank size to 900 m³.

To compare engineering and cost factors on oxygen transfer to mineral slurries from air for the Delft Inclined Plate (DIP) bioreactor [4] with those for Rushton turbine agitated tanks, it was decided to measure oxygen transfer for both systems, using the same mineral (sand) for slurry make-up. The DIP bioreactor results will be presented elsewhere [5].

The effect of solids on oxygen transfer has been reported in the literature, but in most research the volume or weight fractions of solids have been low and the particles used rather large (>100µ). Usually glass beads have been used, while in industrial practice near spherical particles almost never occur. The sand particles used in this work were non-spherical and had a density of 2.61 kg/m³. Seldom has the power needed been reported.

Mills *et al* [6] did research with spherical glass beads with a mean diameter of 66µ, in the range 0–40 vol%. The objective of the current work was to follow the procedure described by Mills, but to perform the tests on a three times larger scale. The solid particles used in this work were not uniformly sized: a particle size distribution is given. The quartz particles had a mean diameter of 14µ (98% < 50µ). The volume fraction of the solids was varied between 0 and 40%.

Relation between $k_L a$ and Solids Concentration

In order to obtain a relation between oxygen transfer, power input, gas input and volume of liquid, researchers have used many different geometries and different media. Because of the variety of systems, lack of consistency (and because of different ways of data processing), a uniform correlation between the above mentioned variables in agitated two phase systems has not been found yet. Nevertheless, all equations produced for liquid/gas systems are of the following type:

$$k_L a = k \left(\frac{P}{V_l} \right)^\beta (V_s)^\gamma \quad (1)$$

where

$k_L a$	=	overall mass transfer coefficient (sec ⁻¹)
k	=	constant
P	=	power input (W)
V_l	=	volume of liquid (m ³)

V_s = superficial gas velocity (m/s)
 β, γ = exponents

Values for the exponents reported in the literature show a large variation: $0.4 < \beta < 1$ and $0 < \gamma < 0.7$. When solids are added to a gas-liquid system, a numerical approach becomes even more difficult. Very few attempts have been made to find even an empirical relation between power input, oxygen transfer coefficients, solids concentration and air input.

Only two such attempts allow comparison with the present work and are therefore described in some detail. A stirred tank with standard geometry was used by Oguz *et al* [7] to obtain information on the effect of solids on the $k_L a$ in three-phase systems. An impeller with four flat blades ($D_I = D_T/2$) was placed in a vessel of 0.145 m diameter. Air was introduced through a ring sparger with 12 holes of 0.5 mm. The slurry was deoxygenated with nitrogen. Several different solids with different sizes and densities were used to form the slurry. The viscosity was determined for all slurries with a rotational viscometer. The total power input was calculated, by taking into account the impeller power input and the isothermal gas expansion power. Figure 1 shows the influence of the gas flow rate and the total power per unit of volume.

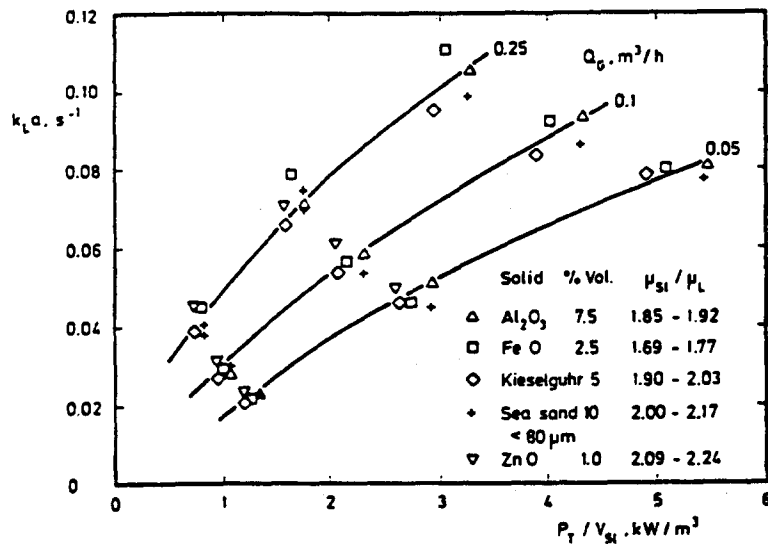


Fig.1 Influence of gas flow rate and power (from Oguz *et al* [7])

Results for six slurry systems are shown in Figure 2.

In the Oguz *et. al.* work, a relation between the different operating parameters using non-linear regression analysis was made. Instead of considering the volume fraction of the solids, Oguz put emphasis on the viscosity of the slurry. Non-linear regression analysis of the data leads to the following:

$$K_L a = 6.6 * 10^{-4} \left(\frac{\eta_{sl}}{\eta_1} \right)^{-0.39} \left(\frac{P_t}{V_{sl}} \right)^{0.75} Q_G^{0.5} \quad (2)$$

Where

- η_{sl} = dynamic viscosity of the slurry (mPas)
- η_1 = dynamic viscosity of water (mPas)
- P_t = total power input (W)
- V_{sl} = volume of the slurry (m^3)
- Q_G = gas flow rate (m^3/h)

The experiments covered the following ranges:

$$1.38 \leq \eta_{sl}/\eta_l \leq 21.3$$

$$0.05 \leq Q_G \leq 0.25 \text{ (m}^3\text{/h)}$$

$$0.78 \leq P_t/V_{sl} \leq 6.3 \text{ (kW/m}^3\text{)}$$

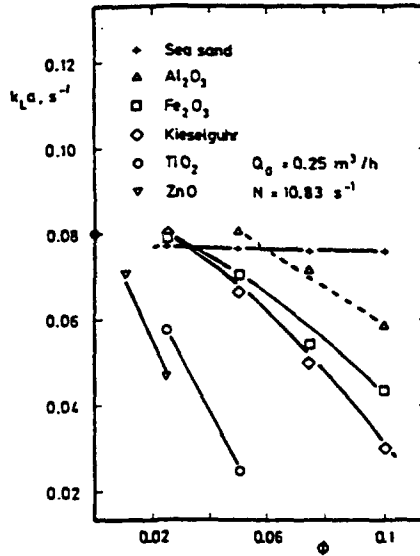


Fig.2 Effect of different solids on $k_L a$ as a function of solids fraction (from Oguz *et al* (1987))

The experiments performed by Mills *et al* [6] in a baffled cylindrical glass vessel were also aimed at obtaining a relation for the $k_L a$ on oxygen transfer from air between solids fraction, power input and superficial gas velocity. Glass beads with a mean diameter of 66 μ were suspended in the liquid in concentrations up to 40 vol. %. Mills found a general relation for the $k_L a$:

$$k_L a = (0.10 - 0.0018\phi) \left(\frac{P_t}{V_{sl}} \right)^{0.67} V_s^{0.31} \tag{3}$$

where ϕ = volume concentration of solids
 P_t = total power input (kW/m³)
 V_{sl} = slurry volume (m³)
 V_s = superficial gas velocity (cm/s)

The relation shows that when the solids concentration is increased, a significant reduction in $k_L a$ results. From Figure 3 the overall decline in $k_L a$ with increasing solids fraction is clearly seen. The research performed by Mills did not take into account solid properties such as density, size and shape factor of the particles. The relation (3) for $k_L a$ does not allow for variations in viscosity at equal percent solids.

The solids used and the scale on which Mills performed his experiments differ from this work. While Mills used almost perfectly spherical glass beads with an average diameter of 66 μ , the current research used small sand grains with a mean diameter of 14 μ , 98% < 50 μ (for a complete particle size distribution, see Figure 4). Mill also used a 0.1 M NaCl solution, which established a coalescent bubble regime. In Table 1 the differences are listed.

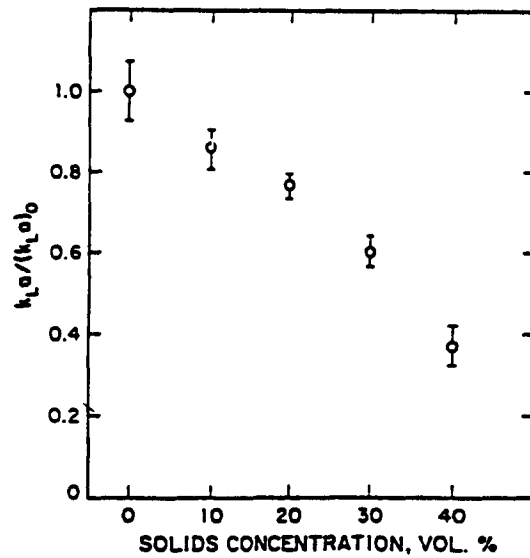


Fig.3 Overall effects of solids on k_{La} with respect to clear liquid at $N=800$ and $V_s = 1.13$ cm/s (from Mills *et al* [6]).

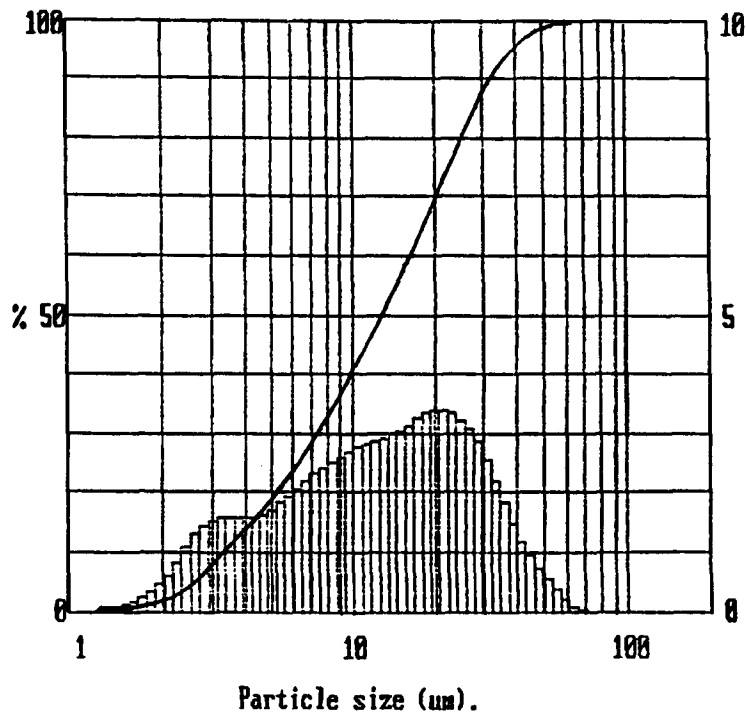


Fig.4 Particle size distribution of the solids

The shape of the particles is shown in Figure 5. The effects of shape and size of solids on oxygen transfer and power draw in air sparged tanks is still undocumented. This makes comparison of the present data with earlier results difficult, especially when viscosity of slurries may be a critical parameter.

TABLE 1 Comparison of variables between the current work and the work done by Mill *et al* [6]

	Current research	Research by Mills <i>et al</i>
Type of solids	SiO ₂ (ground sand)	Glass beads
Size of solids	< 50 μ	66 μ
Shape of solids	Irregular, with sharp edges	Spherical
Volume of tank	63.8 l	16.8 l
Density of solids	2.61 g/cm ³	2.46 g/cm ³
Impeller rotation	500–700 min ⁻¹	600–800 min ⁻¹
Superficial gas velocity	0.34/0.68/1.02/1.36 cm/s	0.36/0.65/1.13 cm/s
Bubble regime	Non-coalescing	Coalescing
Data processing	First order (vd Aart [8])	By Dang <i>et al</i> [9]

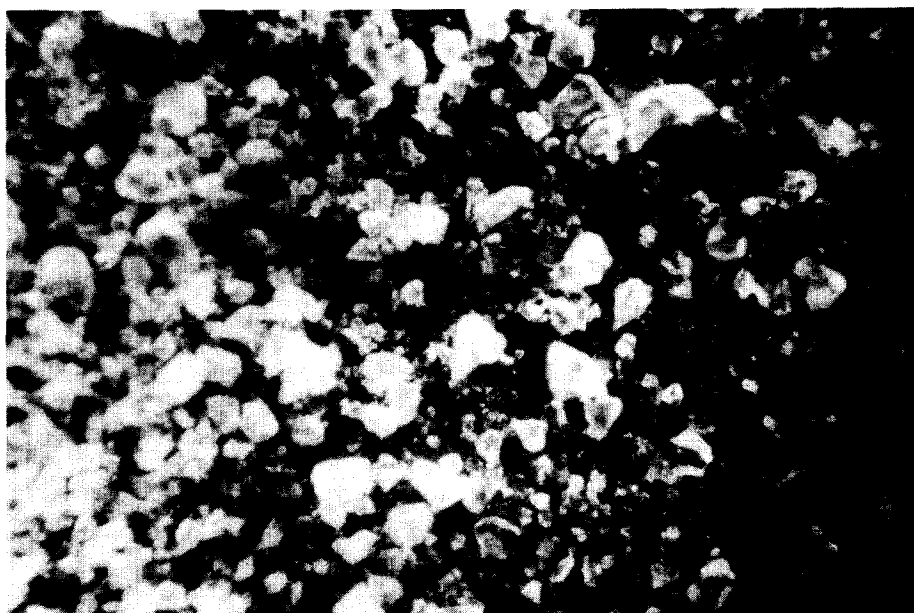


Fig.5 Solid particles used in this work (scale 110:1)

Description of Equipment and Measuring Techniques

All experiments were performed in a vessel with standard geometry, see Figure 6.

As can be seen from Figure 6, the vessel had a flat bottom. Four baffles were placed in the vessel to prevent the formation of a vortex. The spacing between the baffles and the wall of the vessel was 44 mm (0.01 D_T). The experimental set up is shown in Figure 7. A mechanically variable speed drive, with a range of 149–1050 rpm, was used to change the impeller rotation. A light sensor and frequency meter were used to measure the impeller rotation.

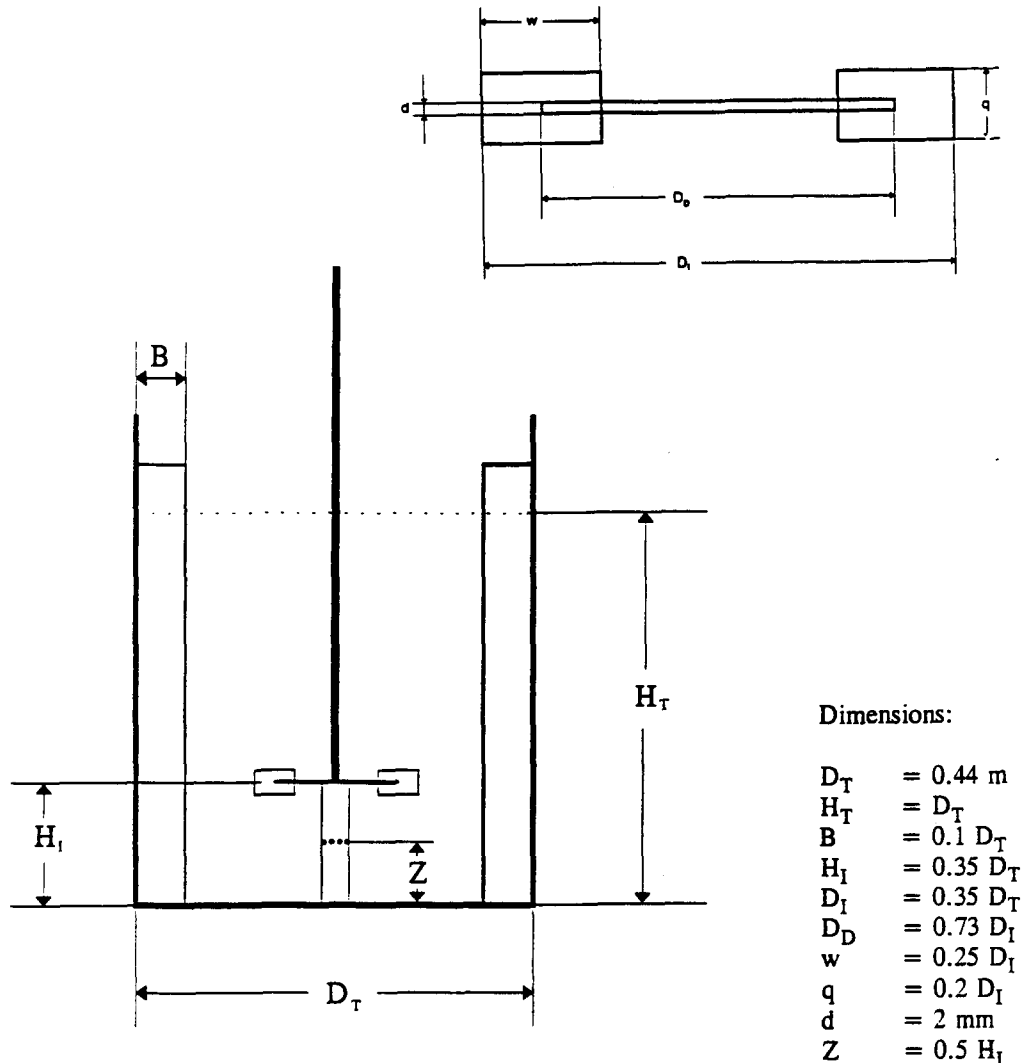


Fig.6 Impeller and vessel geometry

Vessel and impeller were made of stainless steel. The vessel had a small glass window near the bottom. A 15 cm diameter standard six-blade flat disk (Rushton) turbine was used. To prevent the impeller shaft from deflecting, a bearing was installed at the bottom of the vessel. Especially at high impeller rotation, the bearing turned out to be valuable because of the rather rough conditions. Gas was introduced through 8 holes in the bearing housing, situated halfway between the bottom of the vessel and the impeller. The diameter of the holes was 2 mm. To prevent the holes from being blocked by the particles, gas was passed through the holes continuously. At the beginning of the test, the slurry was deoxygenated with nitrogen gas. For the $k_L a$ measurement, the nitrogen was switched to air.

The impeller rotation required to keep solids in ungasged suspension was calculated according to the Zwietering correlation [9]. The rotational speeds used in this work were approximately a factor 2 higher than the "just suspension rotation", a large enough factor to allow assumption of complete suspension of the particles in the present work [10].

A strain-gauge torque transducer was used to determine the impeller power input. The torque transducer used is made by Vibro-meter AG, the type being TT 106, with a maximum torque of 20 Nm. The torque transducer is an extension of the impeller shaft and consists of a square shaft containing a circuit with strain gauges. These gauges respond to shear stresses in the elastic shaft under torsion.

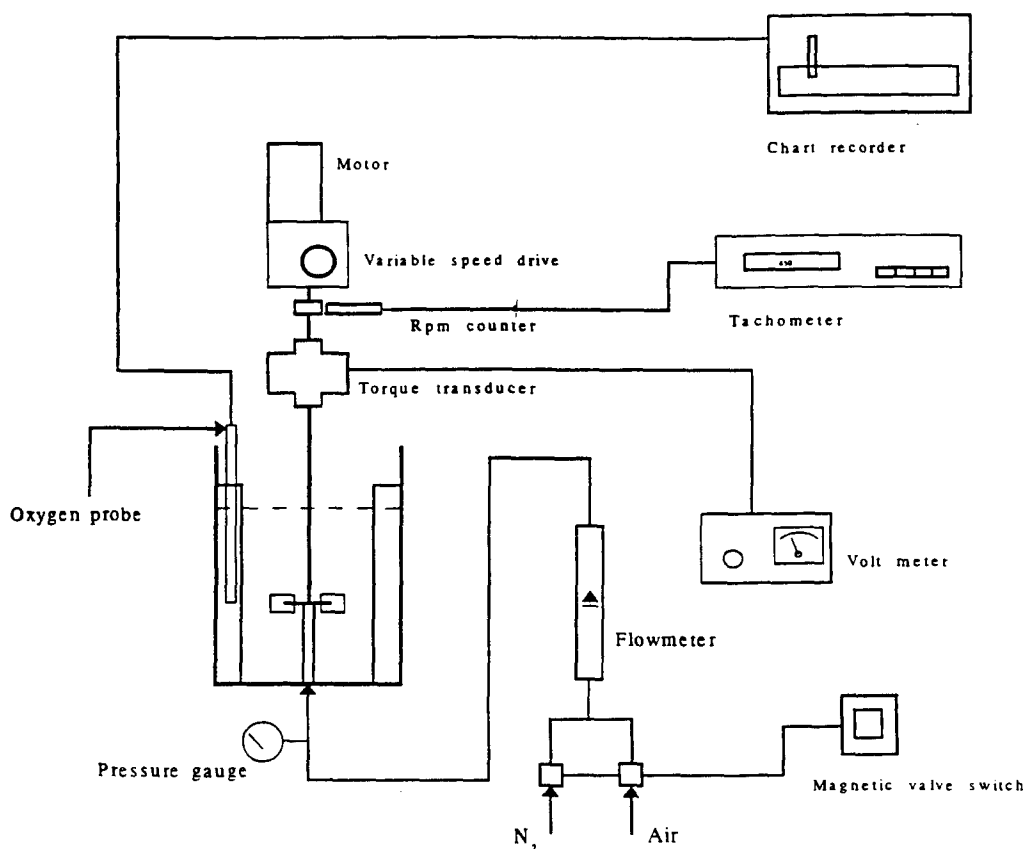


Fig.7 Experimental set up

Measurement of the Volumetric Mass Transfer Coefficient (k_1a)

The k_1a was determined by measuring the increase in oxygen concentration in the liquid with an oxygen probe. Because of the turbulent conditions and the solids present in the liquid, it was necessary to place the oxygen probe directly behind a baffle to protect its Teflon membrane. The membrane was positioned at impeller height. Although the membrane did not break, solids got behind the plastic protector placed around the membrane and eventually reached the electrolyte, at which point signals from the electrode system became less clear. At high solids fractions, placing the probe behind a baffle extended the life of the probe from + 8 tests to + 20 tests. The total electrode time constant for the probe was determined according to Dang *et al* [11] and was found not to change substantially with changing conditions, principally slurry density. This signified that the membrane resistance dominated, which is in agreement with Mills' work. The average total time constant for the probe used in this work was 7.5 seconds.

The k_1a was measured as a function of solids fraction, the impeller rotation (500 and 700 min^{-1}) and superficial gas velocity (0.34 to 1.68 cm/s). The volume of the slurry (solids plus liquids) was kept constant at 63.8 litre. The parameters investigated were the volumetric mass transfer coefficient and the impeller power input as a function of slurry density in volume %.

Viscosity of the Slurries

The viscosity of a slurry appears an important parameter in the mechanisms controlling oxygen transfer. Equations exist that give a relation between viscosity and solids fraction, but usually they do not take into account the shape factor or the presence of a particle size distribution (instead of a mono-sized material).

The following procedure was followed to measure the viscosity. A suspension with a certain solids concentration was prepared and the slurry placed in a viscometer as quick as possible, in order to measure the viscosity before solids settled on the bottom of the container. The values obtained will be slightly inaccurate because of the above mentioned problem:

Sand vol %	0	10	20	30	40
Sand wt %	0	22.5	39.5	52.8	63.5
Viscosity (mPas)	1	4	6	11.5	62

RESULTS

Approach

The objective was to perform all $k_{\text{L}}a$ experiments twice, but this appeared to be rather inconvenient. At low solids concentrations few problems occurred, but the amount of time required to obtain results increased with increasing solids fraction in the slurry. Firstly, the Teflon membrane of the oxygen probe had to be replaced often at high solids tests. With 40 vol % solids, the Teflon membrane lasted for only 7 or 8 experiments. The second problem was with the bearing of the impeller shaft. The shaft was positioned in a Teflon bearing, mounted in a stainless steel cylinder. The SiO_2 particles used in this work had sharp edges, as a consequence of which the Teflon deteriorated quickly, especially when running at high impeller rotations. The bearing had to be replaced five times. As the slurries contained more solids, an increasing amount of nitrogen was required to deplete the slurry of oxygen (as was to be expected). Because of these difficulties, double $k_{\text{L}}a$ experiments were performed for most of the 10, 20 and 30% slurries, but only for some of the 0% and 40% slurries. Where two experiments were done, the average of the two values is given. Deviation of the individual measurements from the average value was below 5 %.

Since the influence of power input on the $k_{\text{L}}a$ is more interesting from a process economical point of view, results of the relation between impeller rotation and the volumetric mass transfer coefficient are not presented here. The interested reader can obtain this relationship, and others, from the experimental data summary [10].

Influence of Impeller Power Input on $k_{\text{L}}a$

It was the objective to determine the influence of total power input (impeller power and gas expansion power) on the volumetric mass transfer coefficient. No difficulties were encountered with measuring the impeller power. To estimate the gas expansion power, the pressure at the sparger holes had to be identified. A pressure gauge was installed towards the end of the experimental work, but this device turned out to be very inaccurate. The pressure gauge gave different values for identical conditions, when read at different times. As a result, values for the gas expansion power have not been used in this paper. From the different superficial gas velocities investigated, the correlation between $k_{\text{L}}a$ and impeller power input is presented for two, see figures 8 and 9. The power input is calculated per unit volume (kW/m^3). The reference volume is the volume of the slurry (liquids plus solids). The $k_{\text{L}}a$ values for pure water (solid squares in the graphs) are higher than for all suspensions, at all superficial gas velocities. The $k_{\text{L}}a$ values for the slurries lie all in a rather narrow range, and differences between the respective slurries are less discernible than the difference between clear water and the sand slurries. The relation between impeller power input and $k_{\text{L}}a$ was also discussed in the work of Mills *et al.* [6]. He used the total power input in his research; the contribution of the gas expansion power did not exceed 15%. Mills' $k_{\text{L}}a$ values correlated linearly with the total power input when plotted in a double log graph. The $k_{\text{L}}a$ values obtained in the present study show the same linearity when plotted that way.

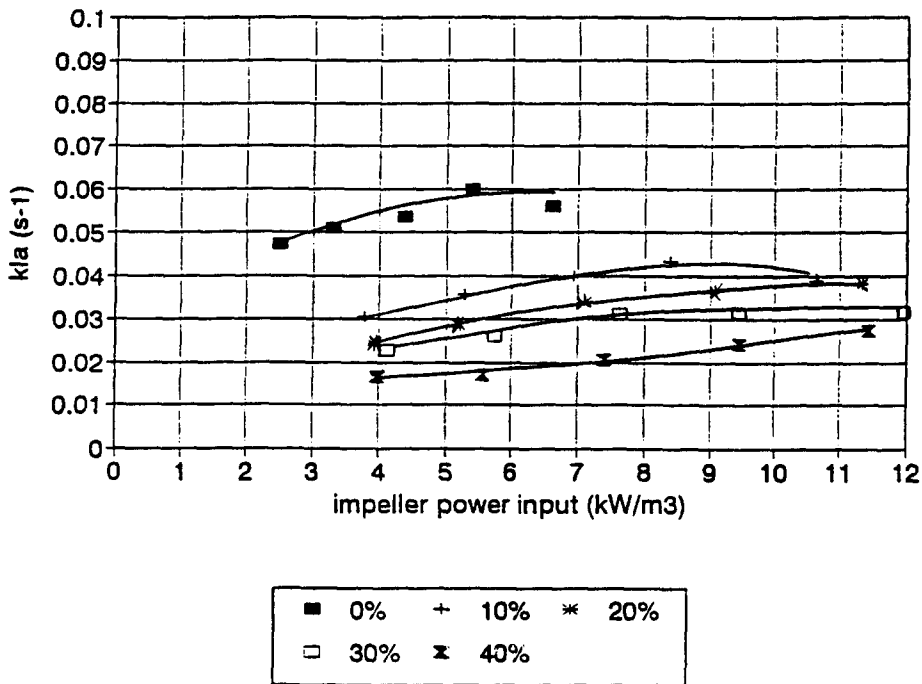


Fig.8 Relation between impeller power input and k_{La} at $V_s=0.34$ cm/s

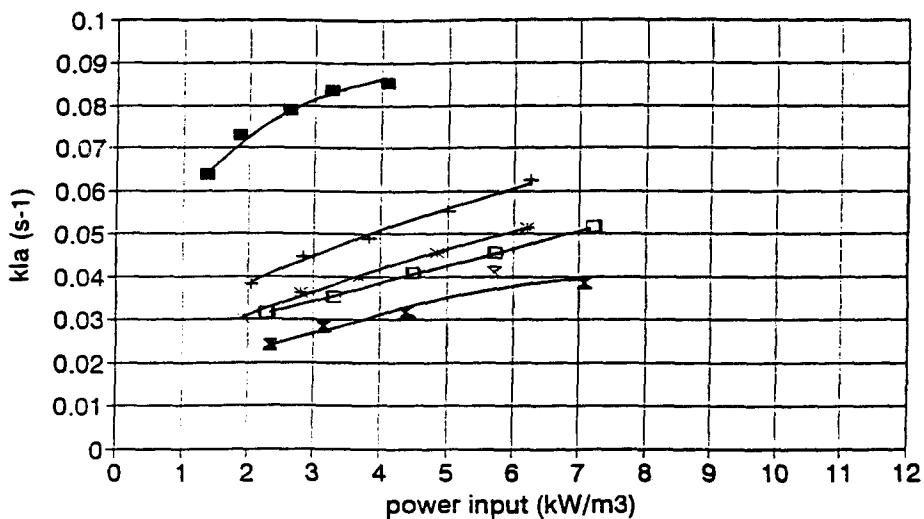


Fig.9 Relation between impeller power input and k_{La} at $V_s = 1.36$ cm/s

Overall Effects of Solids on k_{La}

The overall effect of the solids concentration on the volumetric mass transfer coefficient is shown in Figures 10 and 11. The first one, Figure 10, shows the solids concentration at a superficial gas velocity of 0.34 cm/s, while the values for Figure 11 were obtained at the highest superficial gas velocity investigated, i.e. 1.36 cm/s. On the vertical axis is put the ratio of k_{La} to the k_{La} of 0% solids. The ratio is obtained by dividing the k_{La} by the k_{La} at 0 vol. % solids for that specific impeller rotation. As can be seen, adding solids to the liquid causes a significant decrease in k_{La} at the same impeller speed. Solids fractions between 0 and 10 vol. % were not tested, so it is actually not known whether the linear decline from the ordinate is appropriate

as depicted. There have been reports in the literature indicating small amounts of solids may increase the k_{Ia} [12]. In this work emphasis was put on the effect of high concentrations of solids on the k_{Ia} . The decrease in k_{Ia} from 0 to 10 vol. % sand seems to be independent of superficial gas flow and averages about 30% for all four velocities tested. Between 10 and 20 vol. % solids a steady decline in k_{Ia} value is detected, although less than between 0 and 10 vol. %. Again the decrease in k_{Ia} appears to be independent of superficial gas velocity. At all superficial gas velocities tested the k_{Ia} does not decrease significantly, or at all, when the solids fraction is increased from 20 to 30 %. The impeller rotation does not seem to have an influence on the magnitude of the increase or reduction in the volumetric mass transfer coefficient. The last part of the graphs show a clear decline in k_{Ia} value between 30 and 40 vol. % solids. The total drop in k_{Ia} lies between 55 and 65%, with an average of 60%. At each particular superficial gas velocity, the decrease in relative k_{Ia} values is the least for the highest impeller rotations. The viscosity of 40 vol % sand slurries is such, however, that these observations have little significance for bio-oxidation practice. Double log plotting of viscosity impellor power input and superficial gas velocity resulted in this work's exponents for a Oguz type correlation:

$$k_{\text{Ia}} = 0.036 \left(\frac{\eta_{\text{sl}}}{\eta_{\text{l}}} \right)^{-0.22} \left(\frac{P_{\text{l}}}{V_{\text{sl}}} \right)^{0.39} V_{\text{s}}^{0.24} \quad (4)$$

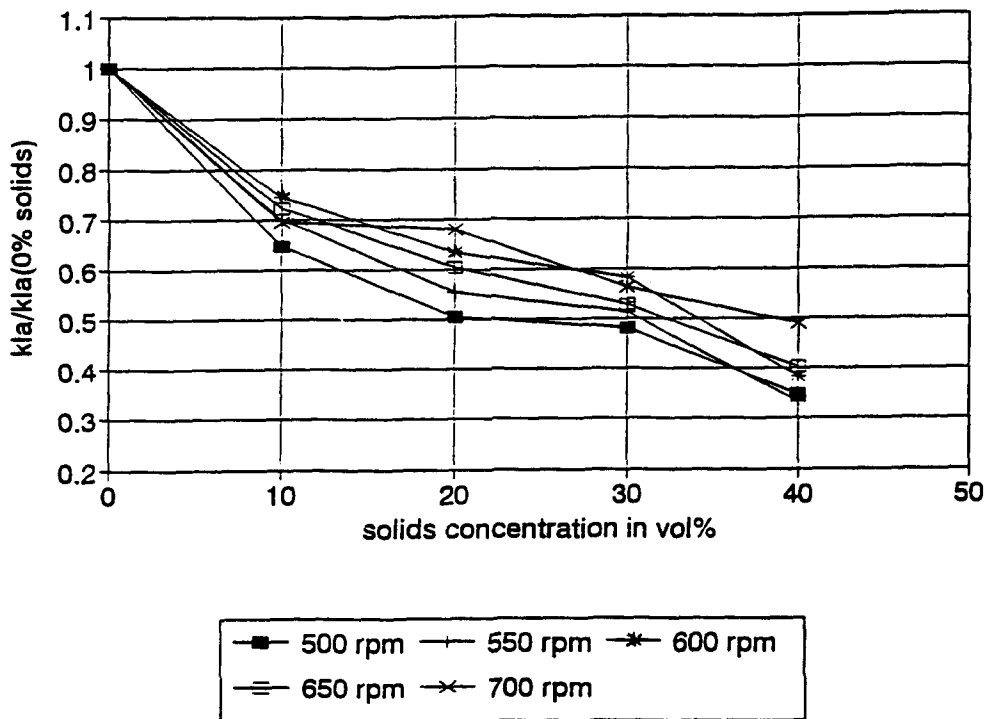


Fig.10 Overall effect of solids on k_{Ia} at $V_{\text{s}} = 0.34 \text{ cm/s}$

DISCUSSION OF THE RESULTS

In Figure 12 the overall effect of the solids is given for the experimental values and for the values as calculated with the equations of Oguz and Mills for the test conditions at $N=700 \text{ min}^{-1}$ and a superficial gas velocity of 1.02 cm/s .

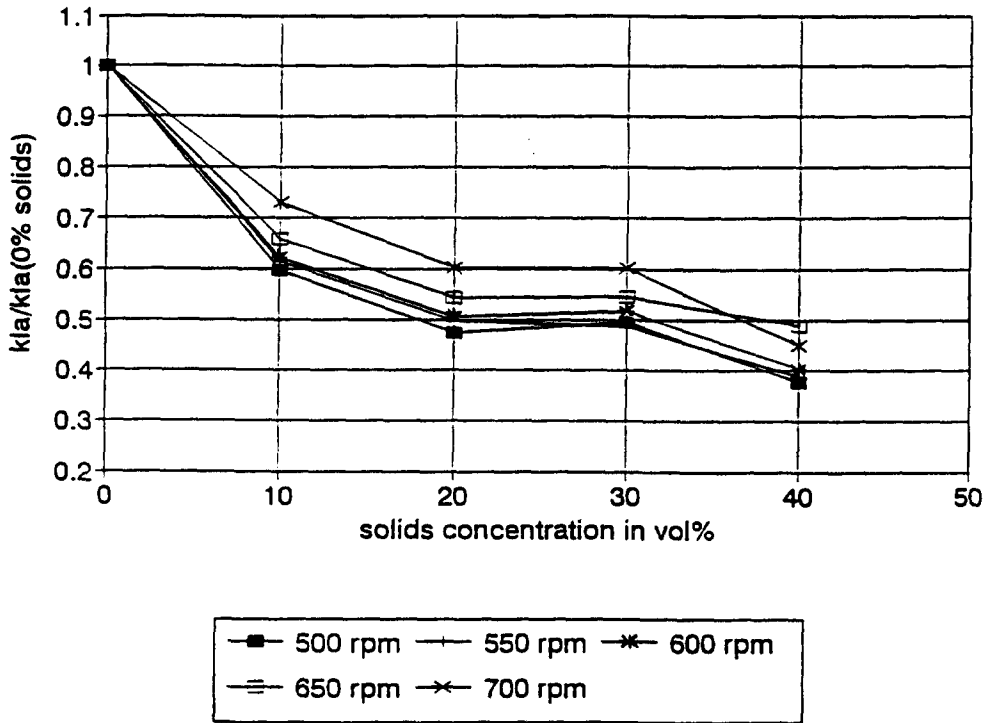


Fig.11 Overall effect of solids on $k_L a$ at $V_g = 1.36$ cm/s.

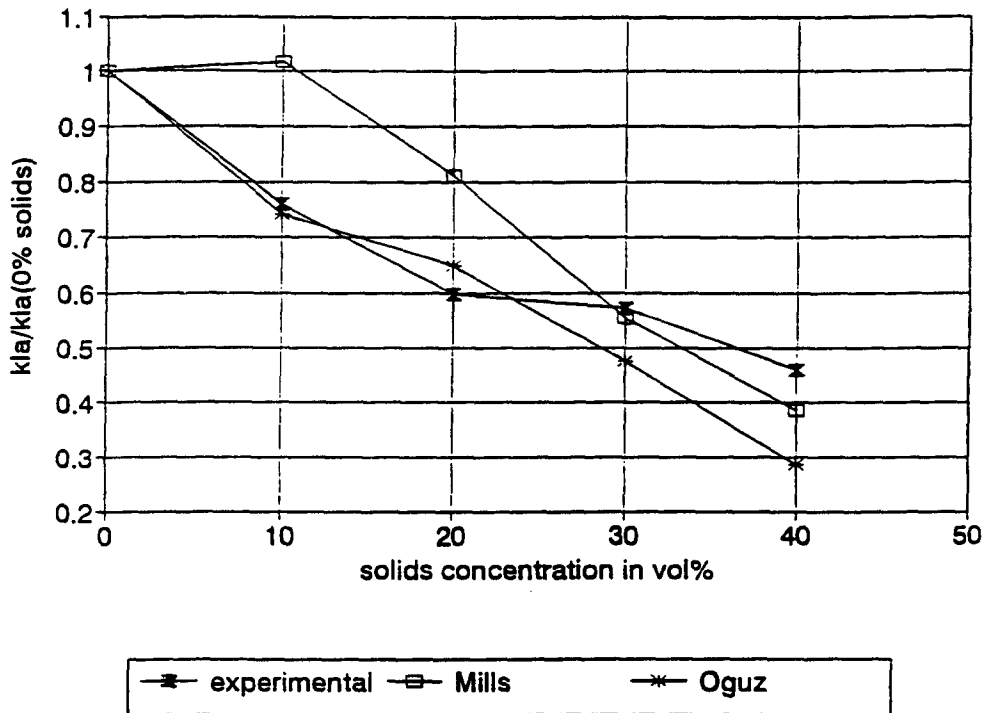


Fig.12 Experimental values and values calculated with the Mills and Oguz equations at 700 rpm and $V_g = 1.02$ cm/s.

The difference between the experimental work and the predicted values with Mills' correlation is obvious. As described by Mills, the decrease in k_{ja} is most detectable in the slurries containing more than 10 vol.% solids. One explanation could be the following: the increase in energy input as required for the slurry in the current work is probably higher than the increase in power input in Mills' work would have been, due to a difference in viscosity. The increase in viscosity when irregular quartz particles are added is larger than with spherical glass beads. The line representing the equation of Oguz shows more resemblance with the current work. The first parts of both lines coincide almost completely. Going beyond 20 vol. % shows a deviation. The overall trend presented by the relation of Oguz corresponds better with the experimental values than the values calculated with the relation of Mills. This suggests a better correlation with k_{ja} and slurry viscosity than with k_{ja} and solids fraction. One remark has to be made: the viscosity of the 40 vol. % slurry and the gas flow rate used in this work all exceed the range in which Oguz performed his experiments. Because of the (still undefined) influence of particle size, particle size distribution, shape factor and density on the viscosity, a relation between k_{ja} and slurry viscosity is expected to be preferred to a relation between k_{ja} and solids fraction. To exemplify, an examination of Oguz's results in Figures 1 and 2 is useful. It still needs to be explained why the effect of 1 vol. % ZnO on the k_{ja} is equivalent to that of 10 vol. % sea sand, and indeed why beach sand has no apparent effect on the k_{ja} up to 10 vol.%.

Relevance of Oxygen Transfer Rate and Power Demand to Bio-Oxidation

In designing bio-oxidation systems for sulphidic minerals, large amounts of oxygen need to be transferred from air to slurries. Whereas the rate constant k_{ja} is of prime importance to the design engineer, the costs of oxygen are ultimately going to decide the feasibility of the project [13, 14]. To assess such costs directly, one might prefer to express the oxygen transfer rate (OTR) in units related to equipment size ($\text{kg O}_2/\text{m}^3/\text{h}$) and to operating costs ($\text{kg O}_2/\text{kWh}$). The OTR's have been calculated with the formula for a representative driving force of $C = 5 \text{ g O}_2/\text{m}^3$ slurry:

$$OTR = k_{La} * C \quad (5)$$

The values are given in Table 2 up to 20 vol. % sand. Logically the same trends as with k_{ja} values are detectable, since C is constant. The increase in OTR with increasing impeller rotation is larger when the superficial gas velocity is larger.

Power demand values for oxygen transfer, expressed as $\text{kg O}_2/\text{kWh}$, are also listed in Table 2. The values are based on the OTR's in Table 2 and the measured power input for the listed testwork conditions. For a certain superficial gas velocity, the power per kg O_2 transferred almost doubles when the impeller rotation is increased from 500 to 700 min^{-1} . As a consequence, when an increase in oxygen transfer is needed, much attention must be paid to the cost of power. Let's look specifically at the 10 vol % or approximately 20 wt % solids. Increasing the rpm of the impeller improves the oxygen transfer, but at substantial costs. By moving from 500 to 700 rpm, the O_2 transferred per kWh nearly halves. Increasing the air flow rate appears more economic. Tripling the flow doubles the $\text{kg O}_2/\text{kWh}$. However, this does not include gas expansion power. A comparison must, therefore, be made between the costs of air compressor power and the costs of increasing impeller power. The former depends, of course, on the depth of the tank in which the slurry is contained (blower vs compressor). In that regard it is interesting that the bio-oxidation industry has elected to scale up to very deep, single agitator tanks, instead of shallow tanks with multiple, short shaft agitators and possibly a sloping bottom as proposed for the DIP bioreactor [4]. Space is normally not a problem at a mineral processing site. It is still an open question whether the improved oxygen transfer due to the high hydrostatic head and longer bubble residence times compensates for the extra compressor power and maintenance.

Further insight might be gained from the work of Fugie *et al.* [15], given in graph form in the excellent book by Fan [16], relating the $\text{kg BOD}/\text{kWh}$ to the BOD removal rate/ m^2 floor area/day for various biological waste water treatment processes. The standard, rock media trickling filter is still the most energy efficient unit at 2–3 $\text{kg BOD}/\text{kWh}$. It also requires the largest land area.

TABLE 2 Oxygen Transfer Rates and Power Demand

Gas Velocity	rpm	kgO ₂ /m ³ /hour			kgO ₂ /kWh		
		0 vol. %	10 vol. %	20 vol. %	0 vol. %	10 vol. %	20 vol. %
0.34 cm/s	500	0.84	0.55	0.43	0.34	0.17	0.11
	550	0.92	0.64	0.51	0.28	0.12	0.10
	600	0.95	0.72	0.60	0.22	0.10	0.08
	650	1.08	0.78	0.65	0.20	0.09	0.07
	700	1.00	0.70	0.68	0.15	0.07	0.06
0.68 cm/s	500	0.99	0.60	0.46	1.58	0.25	0.18
	550	1.13	0.75	0.52	0.47	0.23	0.15
	600	1.29	0.78	0.62	0.41	0.17	0.13
	650	1.40	0.86	0.73	0.36	0.14	0.12
	700	1.35	0.95	0.81	0.27	0.12	0.10
1.02 cm/s	500	1.02	0.64	0.51	0.68	0.29	0.22
	550	1.16	0.75	0.59	0.53	0.25	0.19
	600	1.22	0.85	0.64	0.43	0.21	0.16
	650	1.38	0.94	0.76	0.37	0.18	0.14
	700	1.49	1.12	0.89	0.32	0.17	0.13
1.36 cm/s	500	1.14	0.68	0.54	0.82	0.32	0.28
	550	1.30	0.81	0.65	0.69	0.29	0.23
	600	1.42	0.87	0.72	0.54	0.23	0.19
	650	1.51	0.98	0.81	0.46	0.20	0.16
	700	1.54	1.12	0.92	0.38	0.18	0.15

Oxygen transfer in agitated tanks decreases sharply with increasing solids content, and even more so for the power demand than for the rate. This has been confirmed in bacterial oxidation systems for auriferous pyrites [17]. Remains the problem of divergence of the present small scale test results from the values obtained in industry and specifically those reported for the Sao Bento bio-oxidation plant. The published figures [3] suggest 2.2 kgO₂/kWh agitator energy and a 1.3 kgO₂/m³/h transfer rate. Especially the first figure is so different from results reported in Table 2 that it raises serious questions about the importance of gas expansion energy in bio-oxidation systems.

The earlier comparison made for the 10 vol. % slurry appears to favour high superficial gas velocities. This raises another question. Why have Fluid Bed Bioreactors not been used? Here the major energy comes from gas expansion (some from pumping). Fan [16] in his review of FBB's comes to the conclusion that all the various types appear capable of reaching 0.6 kg BOD/m³/h, which is half the oxygen uptake rate reported for the Sao Bento bio-oxidation plant. Put another way, by adapting already developed bioreactors from the waste water industry, such as the external circulating three-phase fluidized bed bioreactor, the bio-oxidation industry could possibly move away from creating mass transfer surface (air bubbles) by means of rotating impellers, and the attendant need to carry out extensive and lengthy research, such as herein presented. Three phase agitated bio-systems are so complex that their mathematical definition and resultant scientific optimization are still decades away. Opportunism is recommended.

SUMMARY

Many interrelationships and aspects of oxygen transfer from air in mineral slurries still need to be worked out, such as effects of particle size, particle shape, particle density, slurry viscosity, agitator type, tank design, sparger design and depth, effects of bio-decomposition products on surface tension of the liquid, etc. Power consumption and scale-up factors are also important to bio-oxidation engineering. Present results suggest that larger benefits could be gained from increasing sparge air volume than from an increase in agitator power. This suggests further that serious consideration should be given to using more shallow, even agitator free bioreactors which would then be based on three-phase fluid bed or surface aerator design.

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