Gas Dispersion and Mixing for Mineral Oxidation Reactors

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ABSTRACT

Mixers are commonly used for the transfer of oxygen in gas-liquid-solid reactions in minerals processing. The process rate is usually controlled by two steps: gas-liquid mass transfer and reaction kinetics. Some examples are described along with the factors used for analysis of the mass transfer design problem.

Typical oxygen rates and appropriate types of aeration systems or mixers are reviewed. The commonly used Rushton turbine has limitations in a number of mineral processes.

The development and characteristics of the LIGHTNIN A315 axial flow gas dispersion impeller are described. Some practical examples which highlight its advantages are briefly described.

INTRODUCTION

A number of mineral extraction processes involve oxidation reactions which are carried out in slurry form with the oxygen supplied from injected air or oxygen gas. Mixers are used to suspend the slurry and disperse the gas. Some examples are the ammoniacal leaching of nickel, the leaching of iron out of reduced ilmenite in synthetic rutile production, neutral leaching of zinc calcines, pressure oxidation and biological leaching of refractory gold, cyanide destruction and of course the well known cyanide leaching of gold.

All of these processes fall into the category of gas-liquid-solid mass transfer and while oxygen is the most common gas phase reactant this general class can include other gases. These mass transfer processes are not unique to mineral processing and similar principles apply, for example, to fermentation for antibiotic production in the pharmaceutical industry, terephthalic acid production for synthetic fibres, and flue gas de-sulphurisation in the power industry, to name a few.

This paper presents some of the background theory on mass transfer, some characteristics of various gas-liquid-solid mixing impellers and in particular the development and application of the LIGHTNIN A315 impeller.

OVERALL RATE CONSIDERATIONS

When examining a particular process it is essential to consider each of the gas-liquid-solid transfer steps as shown in Figure 1.

The first step is the dissolution of the substance from the gas phase in the liquid phase (gas-liquid mass transfer). The second step is the utilisation of the substance either as a chemical reaction with the solid, a liquid-liquid reaction followed by adsorption on the solid, or as a component in a biological reaction acting on the solid. This second step is a process driven by reaction kinetics.

In a steady state system the rate of mass or oxygen transfer in the first step, must equal that used in the reaction in the second step and of course equal to the overall rate.

$$R = R_1 = R_2$$

It may be that either R_1 or R_2 is controlling the overall rate depending on the particular processes. Some examples will serve to illustrate. In the oxidation of iron for the production of synthetic rutile from reduced ilmenite the rate of reaction of dissolved oxygen with the iron is, for practical purposes, very

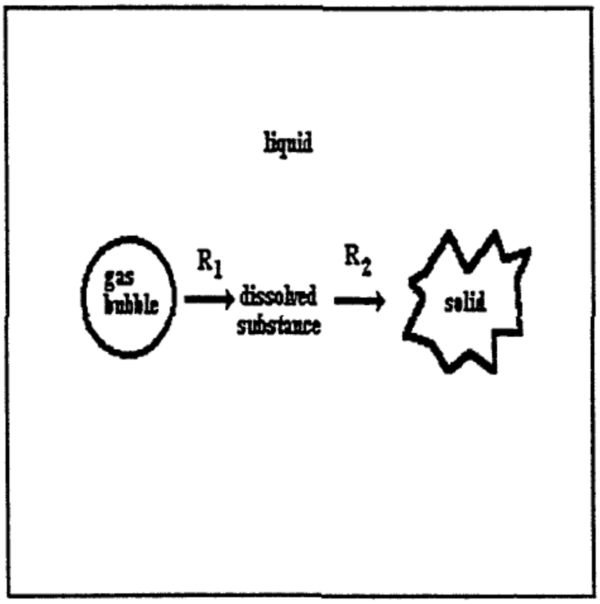

FIG 1 - Gas-liquid-solid mass transfer.

rapid ie R_2 is rapid. Thus, there is a zero or low concentration of dissolved oxygen in the liquid phase. The overall rate of the process is controlled by R_1. This gas-liquid step is largely a function of mixer power and gas rate and so the design of the mixer/gas system controls the rate of the process.

Biological reactions are generally a function of pH, temperature and other factors. They are not generally a function of dissolved oxygen concentration once a minimum level is obtained. Biological leaching of refractory gold ores, for example, does not show rate improvement above approximately 2 mg/l of oxygen. R2 then controls the overall rate. The mixer may have some effect on this rate by the maintenance of temperature and pH uniformity and by shear rates as they affect reaction rate but the reaction rate itself primarily controls.

Cyanide dissolution of gold is more complex because the reaction rate is proportional to both the cyanide and oxygen concentration.

$$R_2 \propto [CN] [O_2]$$

But the dissolved oxygen concentration, $[O_2]$, also has an effect on R_1 so that in theory $[O_2]$ adjusts automatically until $R_1 = R_2$. In practice many ores contain other oxygen consumers which may greatly exceed the oxygen required for dissolution.

The pressure oxidation of refractory gold ores is another interesting example. As Brewer and Olderstein (1989) showed, the initial stage of the process has a fast reaction rate so R_1 controls initially but later in the process the reaction slows down and so R_2 then controls.

Overall, then, it is clear that the mixer may have some effect on R_2 in terms of:

1. pH, temperature and concentration uniformity,

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2. solids suspension, and

3. shear rates.

but once these are adequately maintained there is little further direct effect of mixing.

R_1 however is more directly affected by the mixer. The gas dispersion and turbulence created by the mixing impeller has a significant effect on the gas-liquid mass transfer rate. So where R_1 controls the effect of the mixer is significant. Even if R_1 is not controlling the rate, it must at least equal R_2. In either case, it is important that the gas-liquid mass transfer be carried out as efficiently as possible, because in many cases the energy requirements are significant.

GAS-LIQUID-SOLID MASS TRANSFER

How can some of these rates be determined?

Reaction kinetics (second step)

At the conference on Gold-Mining Metallurgy and Geology in Kalgoorlie, Fraser (1984) described a method of measuring oxygen uptake rates in gold slurries aimed at pre-aeration or cyanide leaching. These measurements could be repeated over the expected leaching time to provide an estimate of the uptake rate at any stage of the process. Multiplying the uptake rate in mg/l/min by the tank volume in any stage provides a design estimate of the required oxygen input for the stage.

For the biological oxidation of refractory gold ores Dew and Godfrey (1991) suggested that the rate of reaction of various sulphur containing minerals can be examined in a continuous pilot plant. They then estimate the corresponding oxygen demand from suggested chemical stoichiometry of reaction of the various materials.

Similarly, in the area of pressure oxidation of refractory gold ores Brewer and Olderstein (1989) showed that a batch pilot plant could be used to study the kinetics of sulphur conversion. They showed the initial stages to be gas-liquid mass transfer controlled and the later stages to be chemical reaction controlled. The batch pilot reactor gives data on both with carefully designed experiments. Provided the batch reactor is large enough to make the impeller physical dimensions large compared with the bubble size useful results on the gas liquid mass transfer versus mixer power and gas rates can also be obtained. This makes it possible to guarantee oxygen utilisation in the full scale process. Some of the scale-up precautions necessary in going from batch to continuous flow are given by Oldshue and Kubera (1992).

If of course the process under examination has rapid reaction kinetics throughout the course of the reaction then only the gas-liquid mass transfer step need be considered. The oxidation of iron can be in this category.

Gas-liquid mass transfer (first step)

The gas-liquid step is often analysed by equations based upon the two film theory of mass transfer covered by chemical engineering texts.

Sparingly soluble gases such as oxygen are liquid film controlled and so the theory relates the rate of mass transfer to a mass transfer coefficient, a transfer area, and a concentration driving force in the liquid film.

$$R_1 = k_L a\,(c^* - c)$$

The concentration driving force is normally expressed in liquid phase units. In some cases (eg pressure oxidation) it is more convenient to express it in terms of p and p*, the gas partial pressures in equilibrium with the liquid and thus related by Henry's law coefficients. This has led some (eg Oldshue and Connelly, 1977; Brewer and Olderstein, 1989) to express the mass transfer coefficient as k_Ga using pressure units though in reality it is a liquid film coefficient.

Application of this equation to a gas rising in a mixed tank can be illustrated in Figure 2. c* is the liquid concentration in equilibrium with the gas. c is normally the actual value of dissolved oxygen in the liquid. If it can be assumed that the liquid phase is well mixed then c remains constant throughout the tank. The gas, however, rises from the bottom to the top. Since c* is a function of temperature, pressure and the gas composition of the bubble it will change from bottom to top. At the bottom c* will be increased by the hydrostatic head (which is significant in large scale atmospheric tanks). When the gas is air, the bubble reaching the top will be partially stripped of oxygen and so the value of c* will be reduced. The LIGHTNIN organisation [Oldshue (1970), Fraser (1983)] generally uses the log mean of the driving force at the sparge and the surface.

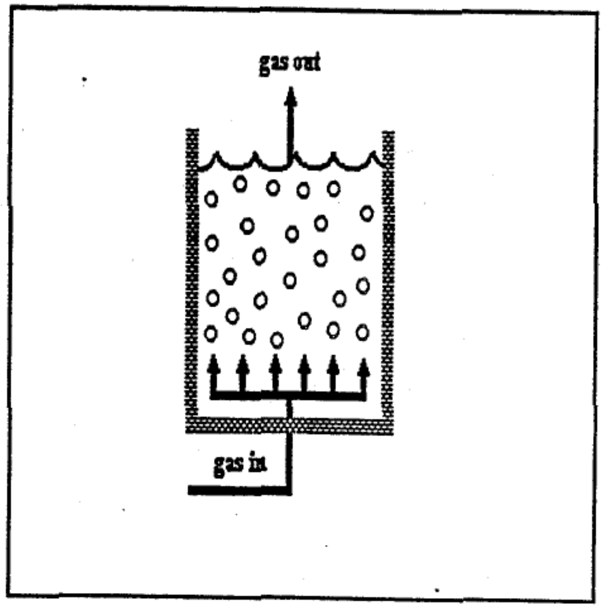


FIG 2 - Idealised model of submerged aeration.

$$R_1 = k_L a\,(c^* - c)_{LM}$$

The effect of this correction is important in typical full scale industrial tanks.

For oxygen in pure water c* can be obtained from Henry's law or handbook tabulations. In mineral slurries this is likely to be decreased by the presence of solids and salts in solution and the effect should be measured for each system under consideration. The ratio is often expressed as a beta factor.

$$\beta = \frac{c^*_{system}}{c^*_{water}}$$

La Brooy, Muir and Komosa (1991) report some data for saline solutions and pulp, and provide some corrections for dissolved oxygen meter readings in saline solutions.

It is not usually possible to separate the mass transfer coefficient, k_L and the transfer area, a. Determination of a combined k_La, even in clean water, is difficult. Results need to be evaluated with caution especially when comparisons are made using different measurement techniques. For example Oldshue and Connelly (1977) used the oxidisation of excess sulphite;

Smith, van't Riet and Middleton (1977) used de-oxygenation with nitrogen gas; Fraser (1983) used de-oxygenation with sodium sulphite catalysed by cobalt chloride; Linek, Benes and Sinkule (1991) used a dynamic pressure step method. Lally (1991) presented some precautions with the excess sulphite method.

For the dispersion of gas by a mixing impeller, k_La has been correlated (eg Oldshue, 1970) in the form:

$$k_La = x\,(P_G/V)^y\,(F)^z$$

x is a function of the type of impeller and the physical system. Base correlations are commonly developed for the air/water (coalescing) system. Even with these reproducibility is difficult. The presence of salts and solids especially where they change surface tension or viscosity will change the rate compared with clean water. These systems are sometimes referred to as non-coalescing. The ratio is commonly expressed as an alpha factor.

$$\alpha = \frac{k_La \text{ system}}{K_La \text{ water}}$$

Generally alpha must be experimentally determined for the system under consideration.

y and z depend on the impeller type, and the scale of the operations. They need to be determined experimentally. Some values could be deduced from reported data in various papers. Typically y is between 0.4 - 1.0 and z is between 0.3 - 0.6.

A consequence of the general form of the correlation is that a required k_La can be achieved by a variety of combinations of mixer powers and air rates. The air supply requires power to compress and so an optimum total power (sum of mixer and compressor) can be determined as shown in Figure 3.

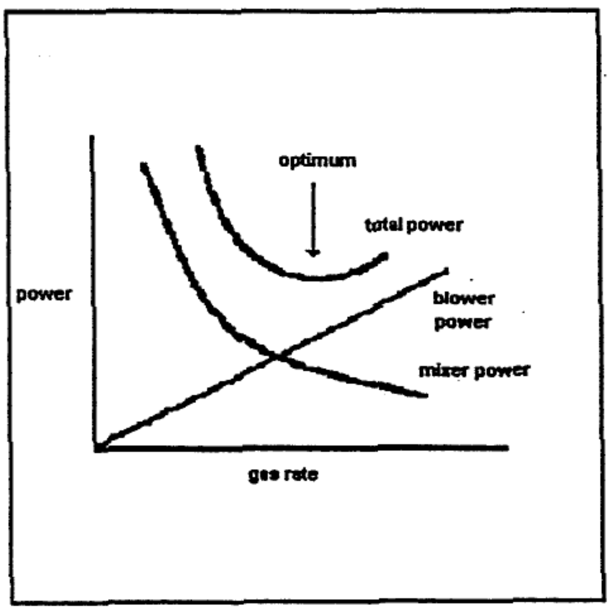


FIG 3 - Possible combinations of mixer power and gas at fixed oxygen rate.

AERATION SYSTEMS

Fraser (1984) described some of the equipment used for oxygen transfer to mineral processing slurries. These include, air injection under the LIGHTNIN A310 fluidfoil impeller, surface aeration and radial flow impellers. It is helpful to consider application of these to typical mineral processes as shown in Table 1. In addition, some values for the new process of bioleaching are shown for comparison.

TABLE 1

Application	Typical Uptake Rate mg/l/min	Typical Impeller Type	Typical Impeller Power kW/m^3
Gold Cyanide Leach	0 - 0.8	A310 Fluidflow Axial Flow	0.04
Gold Slurry Pre-Aeration	0.8 - 2	R330 Surface Aeration	0.1
Bioleach Oxidation	20	A315 Gas Dispersion Axial Flow	0.3
Reduced Ilmenite to Synthetic Rutile	100	Rushton or R100 Radial Flow	2
Gold Pressure Oxidation	500 - 1500	Rushton or R100 Radial Flow	1 - 3

The A310 fluidfoil impeller is an efficient gas dispersion device but only when operating within its range of application. With this type of device gas is driven downwards from a sparge below the impeller by the velocity of the fluid flow. Above a certain gas rate gas bubbles start to accumulate on the upper surface of the blade. Because the blade area is not large a point is soon reached where the bubbles grow sufficiently to stop the pumping of the impeller. This is described as 'flooding'. With this type of impeller the onset of flooding is sudden and at a relatively low gas rate. Thus it can only be used in low uptake rate applications.

Radial flow impellers such as the Rushton turbine are the traditional impellers for gas dispersion. Gas rising into the impeller is shed in vortices from cavities trailing behind the impeller blades as noted by van't Reit and Smith (1973) and Nienow and Wisdom (1974). A laser velocimeter vector diagram of the flow pattern from a Rushton turbine (LIGHTNIN R100) is shown in Figure 4. The high shear rates evident in the impeller outlet flow stream probably assist in the gas dispersion. Shear rates may in fact be higher than necessary for gas dispersion and are often detrimental in biological processes.

Despite the typical high power level applications shown in Table 1, the Rushton turbine can be used for effective gas

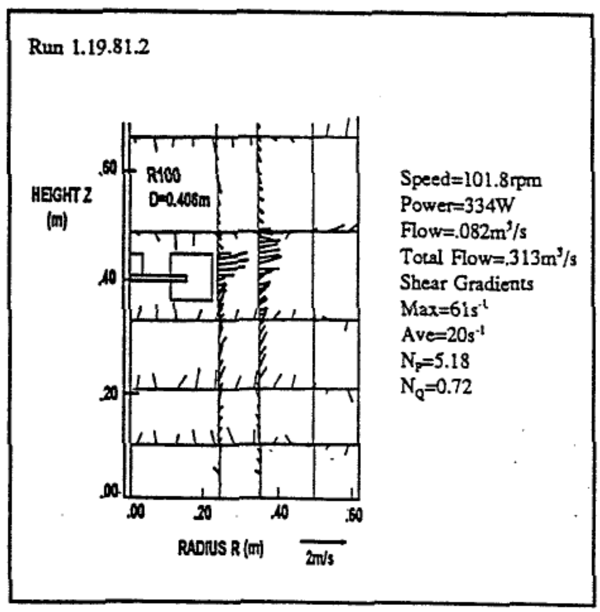


FIG 4 - R100 fluid velocity profile.

dispersion at low power levels provided there are no fast settling solids. But there can be problems with typical mineral slurries. Since much of the impeller power goes into shear, it is an inefficient flow producing device. Furthermore, the direction of flow is upward under the impeller. Both of these factors make it ineffective for the suspension of solids. A downward axial flow is preferred, as shown in Figure 5. Of course, at the high power levels typical of synthetic rutile and pressure oxidation, most solids can be suspended despite the radial flow pattern.

When multiple Rushton impellers are used (typical of the fermentation industry) the flow pattern results in zonation (even at high power levels) as shown in Figure 6. This can be detrimental, particularly in biological processes requiring good temperature and pH uniformity. The high shear rates of radial

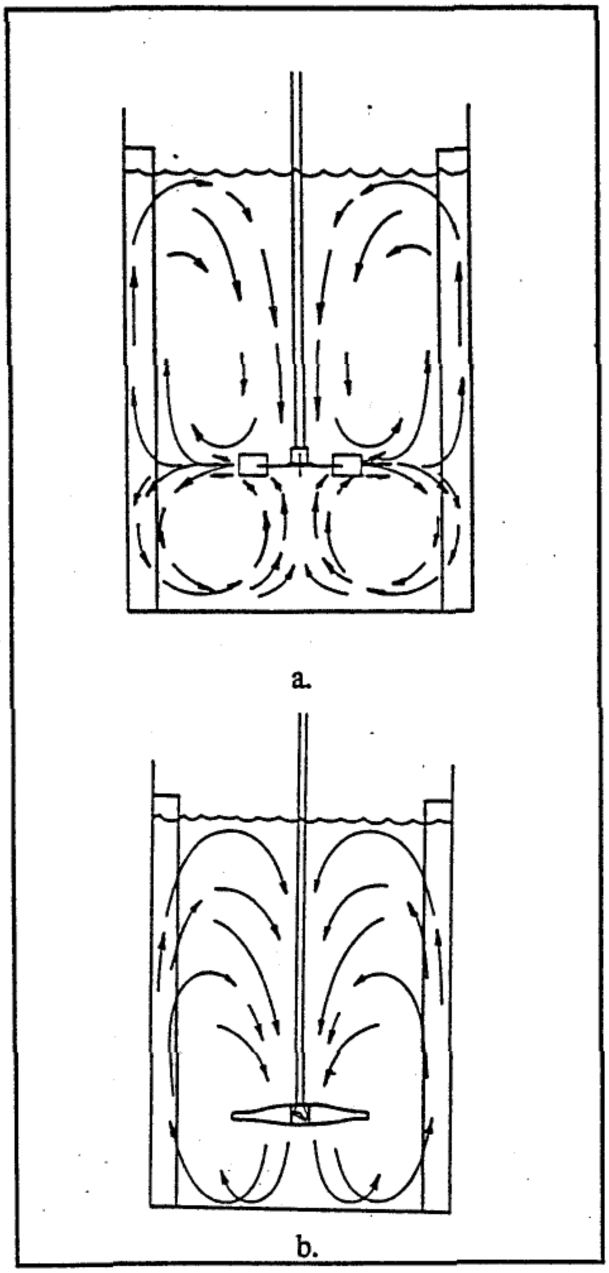


Fig 5 - Flow pattern of a Rushton turbine a) compared with the preferred axial flow for solids suspension in b).

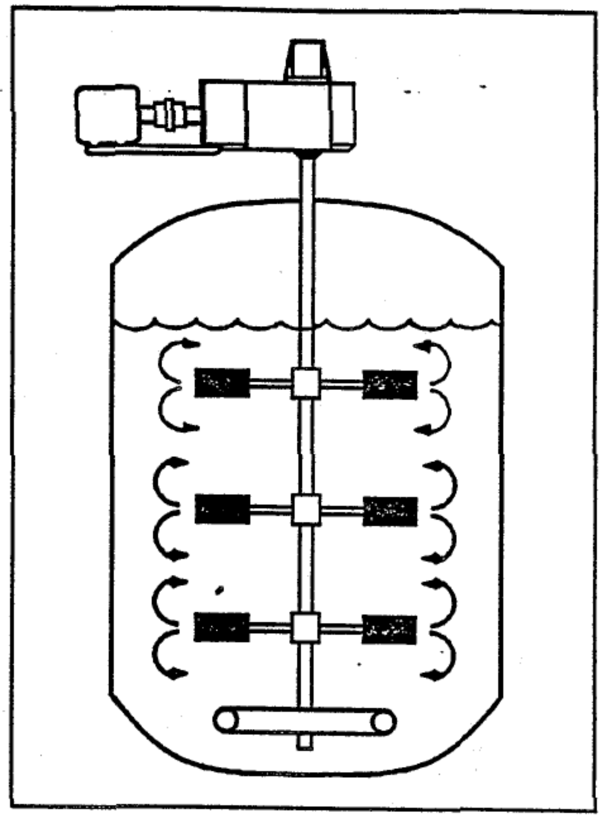


Fig 6 - Bulk fluid flow patterns with Rushton turbines.

flow impellers can be detrimental to the action of the bacteria in some biological processes.

Another disadvantage of the Rushton turbine for practical mixer design is shown in Figure 7. K is the ratio of power drawn with gas to power without gas. As gas rate increases there is a

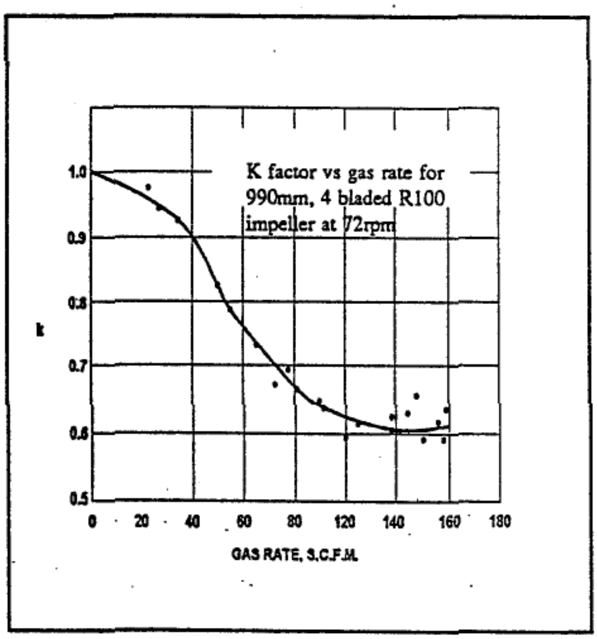


Fig 7 - Typcial curve of K factor vs gas rate at a given impeller speed.

continual reduction in power drawn. If the mixer normally operates with a high gas rate such that K = 0.6, for example, there are two possibilities:

1. Design the mixer for the power drawn at the high gas rate. If the gas flow stops or is reduced the drive then becomes overloaded.
2. Design the mixer for ungassed operation. Then with gas at normal flow it will operate lightly loaded. This means drives are perhaps 60 per cent larger than need be and costs are accordingly higher.

Attempts have been made to overcome this with a hollow blade type impeller (eg Chemineer CD6, Scaba 6SRGT and ICI Gasfoil). They do result in a more constant power draw with gas rate but one undesirable characteristic reported by Nienow (1990) is a 20 - 30 per cent jump in power on flooding. Claims of enhanced mass transfer compared with the Rushton turbine by Warmoeskerken and Smith (1989) have been disputed by Linek, Benes and Sinkule (1991). While the hollow blade turbine has a little lower power number than the Rushton both are high relative to axial flow impellers. This means for a given size and power they run at a low speed which requires a high torque. Drive size and cost are related to torque.

DEVELOPMENT OF THE LIGHTNIN A315 IMPELLER

Some experiments in fermenters where the upper radial flow impellers were replaced with axial flow (see Figure 8) showed promise of improved process performance. But the upper impellers readily flooded. Work was undertaken to develop an impeller of this type but with improved gas handling. Approximately 32 different types of impellers were tested over a range of powers and gas flow rates to optimise the design. Some

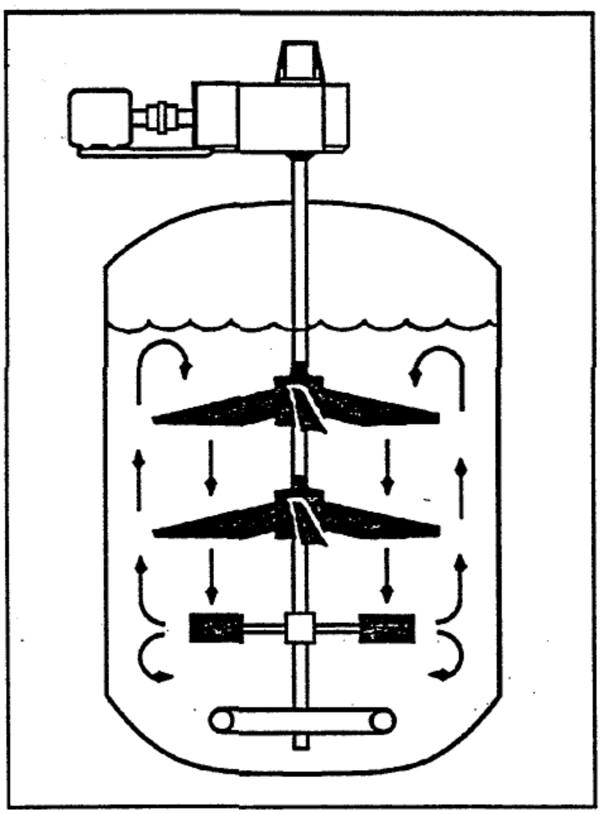


FIG 8 - Bulk fluid flow patterns with upper axial flow turbines.

were rejected on the basis of sensitivity of power draw to impeller proximity to the tank bottom and poor pumping performance.

The selected design is designated the LIGHTNIN A315 which is patented in USA, Australia and other countries. It is shown in Figure 10. Note that the blade has a wide section which allows much more gas accumulation than the A310 impeller. Furthermore the area of the blades relative to the swept area is 87 per cent compared to 22 per cent for an A310. This increased 'solidity' improves its ability to pump against the head developed by rising gas bubbles and helps to prevent gas by-pass through the impeller in the lower velocity region near the hub. The blade is given a camber and essentially some twist to provide a fluidfoil flow enhancement.

Pumping performance

LIGHTNIN uses its laser doppler velocimeter [Weetman and Salzman (1981)] to quantify impeller flow direction and magnitude. A velocity vector plot of the A315 flow pattern is shown in Figure 9. The plot shows a strong downward axial flow with maximum velocities at about 75 per cent of the impeller diameter in the discharge stream. This pattern is similar to other high efficiency axial flow impellers but quite different from the radial flow of the Rushton turbine of Figure 4.

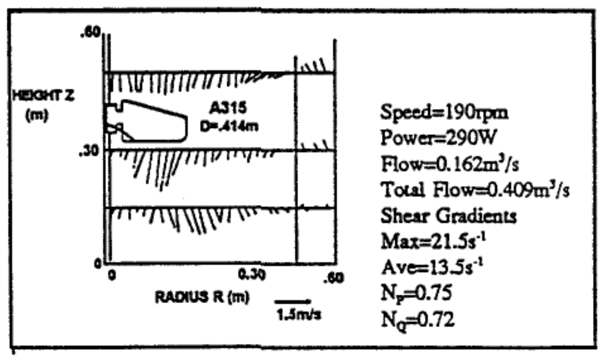


FIG 9 - A315 fluid velocity profile.

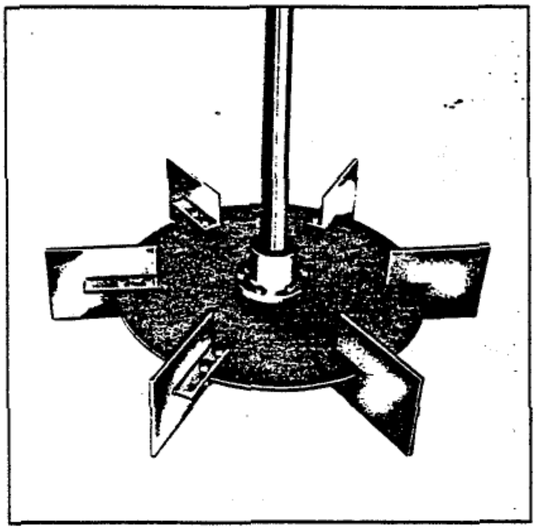
Integration over the impeller area permits accurate determination of the primary pumping capacity and impeller flow number. Integration over a larger area gives the total circulating flow. A comparison of primary flow to power of some impellers relative to the pitched blade turbine is given in Table 2.

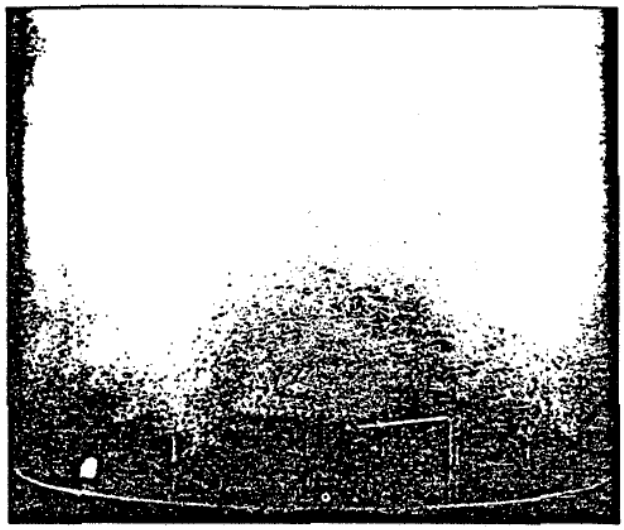
TABLE 2

IMPELLER TYPE		$(Q/P)_R$*
A200 -	45o Pitched Blade Turbine	1.00
A315 -	Fluidfoil Gas Dispersion Impeller	1.34
A310 -	Fluidfoil Impeller Axial Flow	1.51
R100 -	Rushton Turbine Radial Flow	0.18
*$(Q/P)_R$ is ratio (Q/P) / $(Q/P)_{A200}$ at constant Q and D		

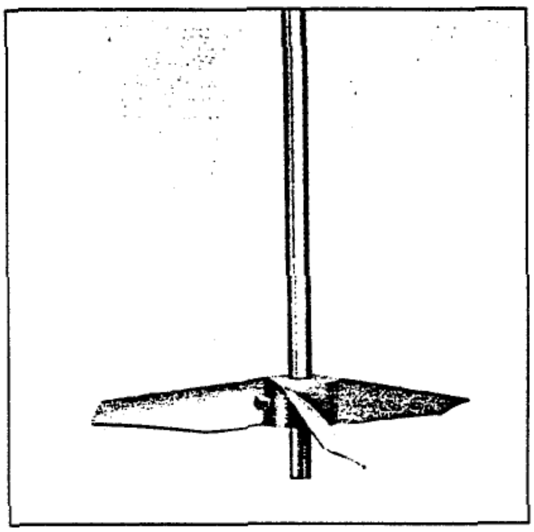
While the A315 sacrifices some flow efficiency compared to the A310 it is better than a pitched blade turbine and much better than the Rushton turbine.

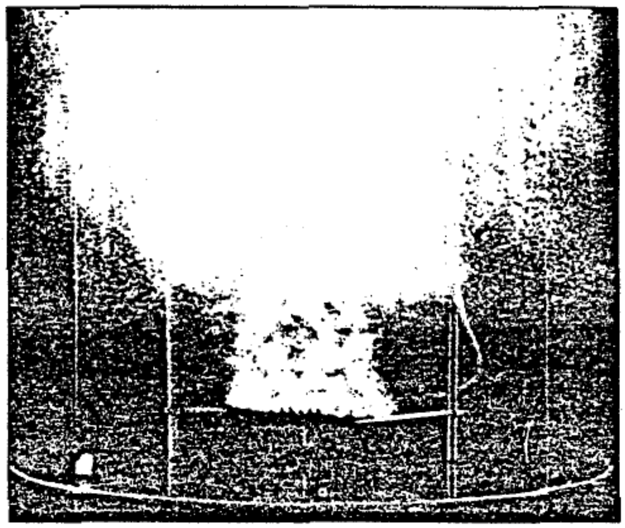
Shear rates or velocity gradients are also shown on the vector diagrams. Because of the different flow pattern the rates for the A315 may not be directly comparable with the Rushton turbine but are probably of the order of ⅓. This is significant for biological processes which may be shear sensitive.



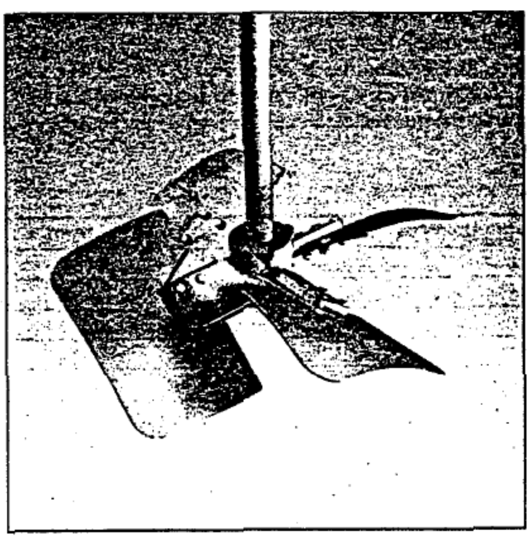


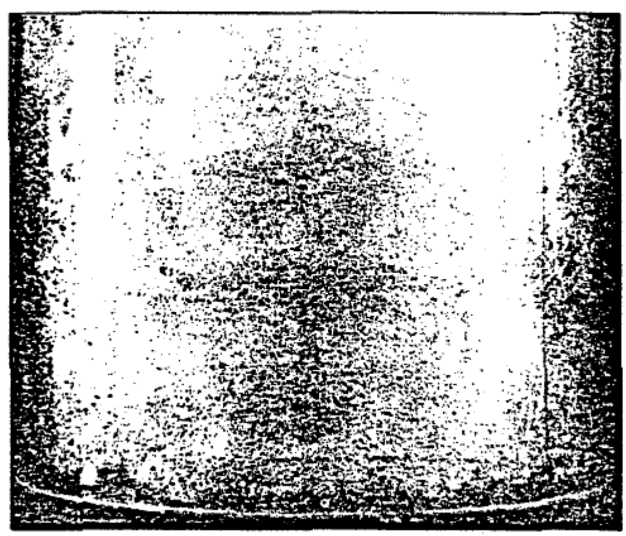
Rushton Turbine (LIGHTNIN R100)





The LIGHTNIN A310





The LIGHTNIN A315

FIG 10 - Comparison of gas dispersion with various impellers.

The high flow provides effective solid suspension and heat transfer - important requirements in, for example, bioleaching of mineral slurries.

Physical gas dispersion

This is best demonstrated in a series of photographs in Figure 10. These have been taken at $P_G/V = 0.2$ kW/m^3, F = 0.010 m/sec in a 1.22 m diameter tank using air and water. The sparge is set well below the impeller for clarity. The gas rate was set to approach the limit of gas handling for the Rushton turbine.

The Rushton turbine is giving a fine bubble dispersion, but it is unable at this gas rate, to drive the bubbles to the bottom of the tank. The A310 impeller axial flow pattern is completely overwhelmed by the gas flow. The A315 is still able to maintain most of the axial flow (though there is some radial component) which drives the bubbles to the bottom and gives greater gas hold-up. While the bubbles are not quite as fine, the holdup together with better blending result in improved mass transfer in a number of configurations.

Solids suspension

Comparative solids suspension capability is illustrated in the photographs in Figure 11 using beads for clarity. With the Rushton turbine most of the beads remain on the bottom whereas they are all suspended with the A315 impeller.

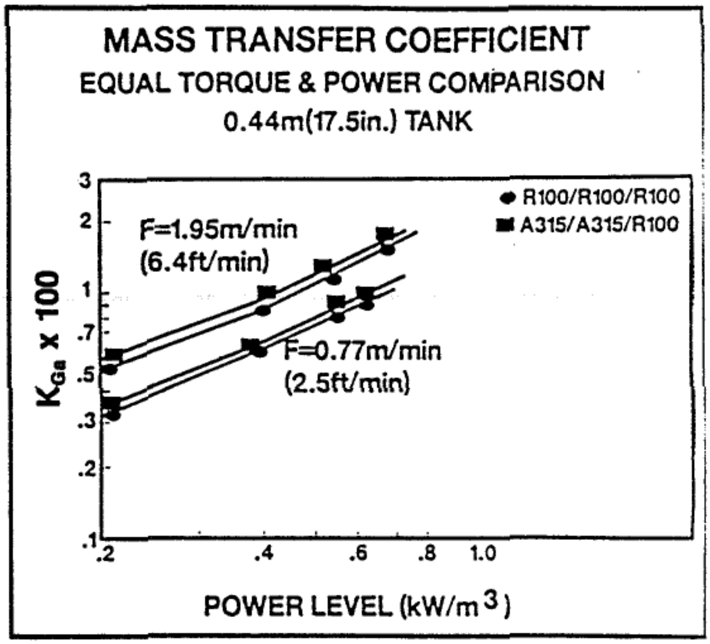


FIG 12 - Mass transfer coefficient vs power level.

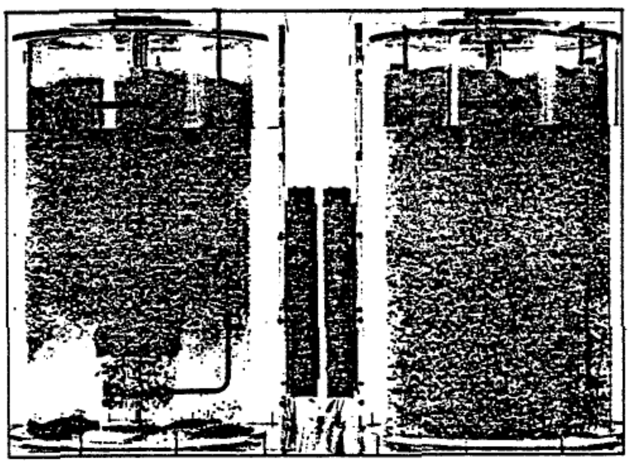


FIG 11 - Comparative solids suspension of Rushton turbine (R100) and A315 impellers.

Mass transfer

Care is necessary in the determination of mass transfer performance as it can be affected significantly by what seem to be minor factors. For example, the studies of Smith, van't Riet and Middleton (1977) on Rushton turbines showed a doubling of the k_La when going from a 'coalescing' system (eg air-water) to a 'non-coalescing' system (eg air-electrolyte). Comparison of impellers may give different results, depending upon the technique used in measuring k_La, purity of solutions, pH control, presence of certain metals (principally copper) and other factors.

Comparisons made by LIGHTNIN (Kubera and Oldshue, 1991) in a typical tall, slim biological fermenter configuration using an excess sulphite method, with very good reproducibility, showed that the best results were obtained using a lower Rushton turbine for initial gas dispersion and two upper A315 impellers. Comparison with three Rushton turbines is shown in Figure 12. Despite the lower shear rates, some improvement in mass transfer is achieved with the A315 impellers. Blending to uniformity in the same configuration is shown in Figure 13, with improvements evident with the A315 impellers.

Roustan *et al* (1991) compared an A315 impeller versus a Rushton turbine in a single impeller configuration at low power levels and gas rates. They used a cobalt catalysed unsteady state re-aeration technique. Measured k_La was 30 - 40 per cent higher with the A315.

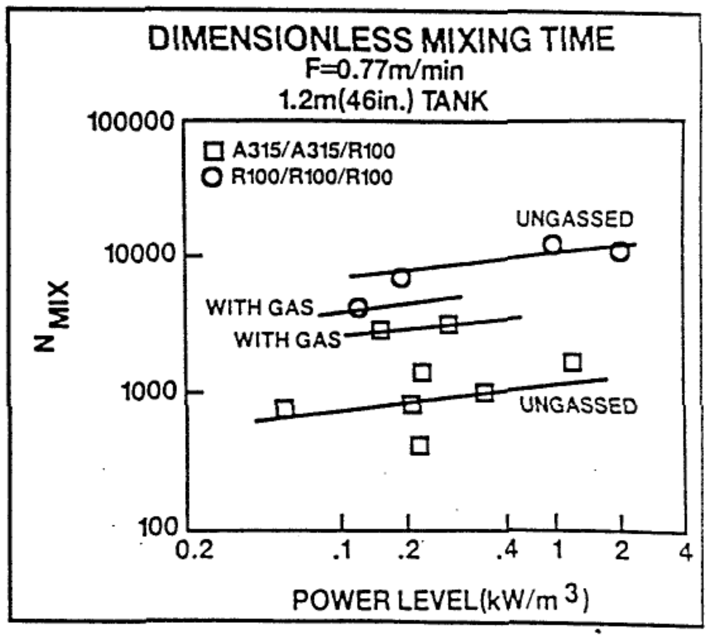


FIG 13 - Dimensionless mixing time vs power level.

At the same conference Nienow (1991) indicated k_La might be independent of impeller type at the same power/volume and superficial gas velocity in coalescing fluids. Mackon, McFarlane and Nienow (1991), however, showed that gas holdup could be significantly enhanced with axial flow impellers, where coalescence is severely inhibited. This may explain significant mass transfer improvements reported for axial flow impellers in some systems but not in others.

These results highlight the need for plant design purposes to obtain at least some indicative data of the effect of the process slurry concerned. In gold bioleaching for example, van Aswegen *et al* (1991) report adequate oxygen transfer at significantly lower power input with axial flow impellers. It is possible gas coalescence is inhibited and gas holdup is enhanced by the particular surface tension and viscosity of these systems.

Power draw

At high power/volume levels, the power draw of all gas dispersion impellers is affected by the presence of the gas. This is an important factor in practical mixer design.

The A315 shows improved characteristics particularly at lower power levels. Figure 14 provides some data for a number of impellers at 0.2kW/m^3 (typical for bioleaching). As the gas rate increases, there is a slow increase in power with the A315 before finally dropping quickly. This drop is usually associated with flooding but the A315 continues to disperse gas in this condition albeit with a more radial flow pattern. The improvement compared with other impellers is evident.

Normal operation will be unflooded with less than ±10 per cent variation in power draw. This means that drives do not need to be oversized for variable gas loading as is common with Rushton turbines.

The relatively low power number of the A315 reduces drive torque so smaller and less expensive drives can be used.

Air injection

Best location for the air injection sparge or pipes is in the high velocity and turbulent region under the impeller and evident in Figure 9. High turbulence will maximise the mass transfer coefficient. Locating it in the maximum velocity will maximise gas handling without flooding.

Location of the sparge above the impeller, while attractive in reducing the head (and hence energy) requirements of the blower, reduces gas handling because of lower velocities. It is also analogous to the use of a larger diameter sparge ring with Rushton turbines which Oldshue and Connelly (1977) showed to be quite detrimental to the mass transfer coefficient.

CASE HISTORIES

Case one - zinc leaching

Zinc concentrates are roasted to a calcine which is largely zinc oxide but contains significant amounts of iron, lead, cadmium, copper and cobalt. Leaching is carried out in stages with increasing acid concentration. The neutral leach is the first stage and contacts spent electrolyte with calcine. The reaction requires the oxidation of ferrous ion to ferric with the oxygen obtained from air sparged under the impeller. While solids concentration is typically less than ten per cent, the solids have a high specific gravity at 5.2 and settle rapidly.

A large A315 impeller has been constructed from stainless steel to provide oxygen transfer and solids suspension in this process in an Australian operation.

Case two- cyanide destruction

Residue slurry or tailings from gold extraction plants contain significant quantities of cyanide. While many Australian plants in warm climates achieve cyanide degradation naturally by exposure to sun light and air in tailings ponds, other plants in cold climates or having more stringent effluent controls need to remove cyanide. The Inco process described by Devuyst *et al* (1989) uses sulphur dioxide and air which react to oxidise the cyanide.

Recent use of the A315 impeller instead of the Rushton turbine is providing effective oxygen transfer and good suspension of the tailings solids in this process. The ability to design to a required quantity of oxygen provides the right mixer selection for the process.

Case three- biological leaching of refractory gold ores

Significant oxygen transfer enhancement was noted for A315 impellers in the only commercial bioleach plant in South Africa

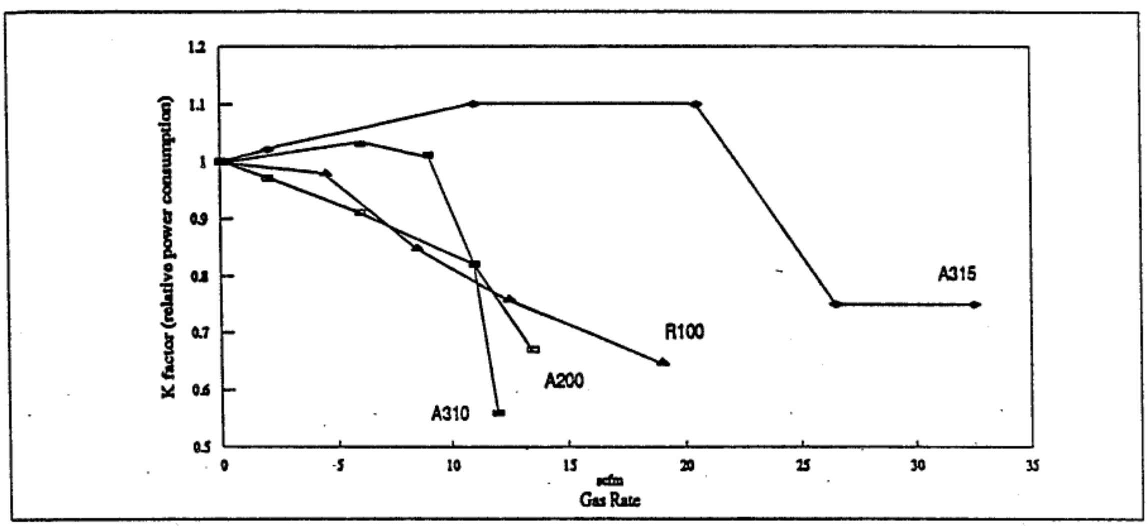


FIG 14 - Change in power consumption vs gas rate for axial and radial flow impellers.

which uses the Genmin bioleaching process described by van Aswegen *et al* (1991). Additional installations are in operation or manufacture for the first two commercial Australian bioleach projects. Sulphide concentrates are acted upon by specially acclimatised bacteria which oxidise the sulphide minerals. In so doing the solid particles are broken down to liberate the gold for CIP recovery. Oxygen is transferred from air dispersed by the mixer and sulphuric acid is produced. This is neutralised prior to CIP extraction. Close temperature control is required but this is complicated by the highly exothermic reaction.

This process has an intermediate oxygen uptake rate where the A315 is very effective for oxygen transfer. Its high flow capacity achieves good solids suspension and heat transfer from the exothermic process. The A315 also has a good power draw characteristic at the typical power levels of this process so plant operation is simplified.

CONCLUSIONS

1. Typical gas-liquid-solid mineral processes are controlled by either the gas-liquid mass transfer step or the subsequent liquid-solid reaction, adsorption, or biological reaction.
2. Techniques are available to analyse and correlate both the mixing impeller gas-liquid mass transfer performance and the kinetics of the subsequent reaction.
3. Selection of the appropriate type of impeller can be determined by the oxygen uptake rate and other process requirements such as solids suspension, heat transfer and fluid shear rates.
4. The LIGHTNIN A315 impeller is particularly useful in the medium power range between the capabilities of the A310 fluidfoil and the Rushton turbine. It has equivalent or better oxygen transfer, high axial flow for solids suspension, blending and heat transfer, and a relatively constant power draw for economy of drive selection.
5. The A315 is not necessarily the best solution for every gas-liquid mixing process but it is proving beneficial in a number of applications in the minerals, energy, pharmaceutical, food and chemical industries.
6. The A315 advantages have been demonstrated in practical mineral processes such as zinc leaching, cyanide destruction and bioleaching of refractory gold ores.

NOMENCLATURE

a	Specific transfer area	m^2/m^3
c	Concentration of material in liquidkg/m^3	
c*	Saturation concentration in equilibrium with the gas	kg/m^3
F	Superficial gas velocity	m^3/m^2/sec
K	Ratio of gassed power/ungassed power P_G/P_{UG}	-
k_L	Volumetric mass transfer coefficient	m/sec
k_Ga	Mass transfer (liquid phase) coefficient expressed in pressure units	g.mole/bar m^3 sec
p	Gas partial pressure	bar
P	Power	kiloWatts
Q	Impeller primary pumping rate	m^3/sec
R	Volumetric mass transfer rate	kg/sec m^3

Subscripts

G	gas phase or gassed operation
L	liquid phase
UG	ungassed operation
[ ]	concentration

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