



Scale-up procedures for gold-adsorption systems.

Part 2: Mixing requirements

by L.C. Woollacott* and K.I. Afewu*

Synopsis

The design of in-pulp adsorption vessels in the gold industry has, for the most part, been seen as a problem involving the suspension of fine particles of ore in a slurry. The need for the vessel to be designed as a mass-transfer contactor has received insufficient attention.

The current series of papers addresses this need. The previous paper considered the question of the scale dependence of the film transfer coefficient. This paper examines the mixing considerations that are relevant. A review of the literature and an examination of a limited set of data provide some insight into the mixing conditions needed for the dispersion of the particles of adsorbent. The mixing requirements for the suspension of the ore particles are also considered, and conclusions are drawn in regard to the design of mechanically agitated vessels for gold in-pulp adsorption circuits.

* School of Process and Materials Engineering, University of the Witwatersrand, Private Bag 3, Wits, 2050 Gauteng.

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Introduction

In a previous paper¹, attention was paid to the difficult problem of the influences of scale, geometry, and mixing conditions on film mass-transfer coefficients in gold in-pulp adsorption systems. It was found that these influences can be described reasonably well by simple parametric expressions. Further, it was found that the values of all but one of the parameters in these expressions are relatively unaffected by scale, geometry, or mixing conditions over the range of conditions tested. The one parameter that is very strongly influenced by these factors is the agitation speed or, alternatively, the power input required to reach the point where the particles of adsorbent are just dispersed in the slurry. If these expressions are to be used with any confidence, a good estimate of this parameter is needed.

There is another reason for the need of such an estimate. In the design of adsorption equipment, it is important that the agitation is intense enough for the adsorbent phase to be at least just dispersed in the slurry. No particle must remain stationary at the bottom of the vessel (or floating on the surface of the slurry) for more than a short time. Operation at agitation speeds much below those that achieve this condition result in adsorption rates that are significantly lower than what is optimal (Figure 1).

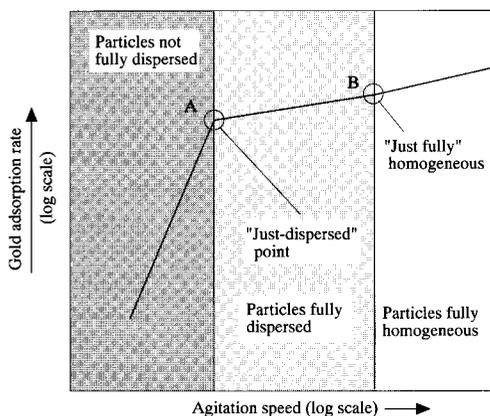


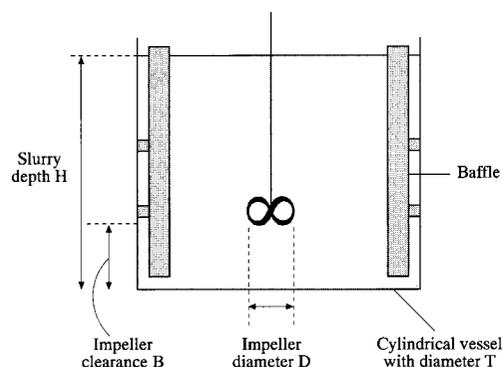
Figure 1—Influence of mixing intensity on gold-adsorption rates (after Gray and Oldshue²)

A considerable body of information exists on the suspension of solids in a liquid. This paper therefore begins with a review of the relevant literature. Thereafter, the question of the dispersion of the adsorbent phase in gold in-pulp systems is discussed, and some experimental results are examined.

Review of relevant mixing theory

The notation used is summarized at the end of the paper.

The practical aim of mixing theory is to provide the process engineer with the tools necessary to select the geometry, agitator speed, and power input that will achieve the desired process response. Various mixer configurations are available³, each with its own characteristics and areas of application. In gold-adsorption technology, the favoured configuration is the open-tank, mechanically agitated vessel (Figure 2), which is the only configuration considered in this paper.



Relevant geometric ratios:

Slurry depth,	$H = T$
Impeller clearance, B	$= 0,25 T$
Number of baffles	$= 4$
Width of baffles	$= 0,10 T$
Height of baffles	$= T$
Baffle clearance	
Off tank bottom	$= 0,075 T$
Off tank wall	$= 0,02 T$

Figure 2—Typical geometry of a carbon-in-pulp contactor

Once a particular configuration has been selected, the remaining design problem involves the selection of the rotational speed of the agitator, N , as well as the power, P , and torque, t , that must be applied to maintain that speed. These design variables are related as shown^{2, 4} in equation [1]. Other variables that are used in mixer design are the power intensity (power per unit volume), P_v , the specific power, E , the torque intensity (torque per unit volume), τ , the impeller discharge rate (or impeller pumping capacity), Q , the pumping rate per unit volume, Q_v , and the Reynolds number of the impeller, Re . Equations [2] and [3] define these variables. It should be noted that some equipment designers prefer to use the equivalent volume, V_{eq} , and the equivalent torque intensity, τ_{eq} , as defined in equation [4].

$$P = P_o \rho N^3 D^5; t = \frac{P}{2\pi N} \quad [1]$$

$$P_v = P/V; E = P/\rho v; \tau = t/V \quad [2]$$

$$Q = P_f ND^3; Q_v = Q/V; Re = \rho ND^2 / \mu \quad [3]$$

$$V_{eq} = \frac{\rho}{\rho_w} V; \tau_{eq} = t/V_{eq} \quad [4]$$

ρ and ρ_w are respectively the density of the slurry and the density of water, and D and T are the diameters of the impeller and the tank. P_o is the power number of the impeller, and reflects the difference between the theoretical and the actual power required to produce a given agitator speed. Obviously the power number is strongly influenced by the hydrodynamic conditions in the mixer, but it decreases to a constant value as the mixing moves into the fully turbulent condition (Figure 3). Similarly, the pumping number, P_f , is dependent on Re , increasing to a constant value in the fully turbulent region (Figure 3). For most mixing systems, particle suspension requires a high degree of turbulence, and the power and pumping numbers are therefore constant in the normal operating range of agitator speeds.

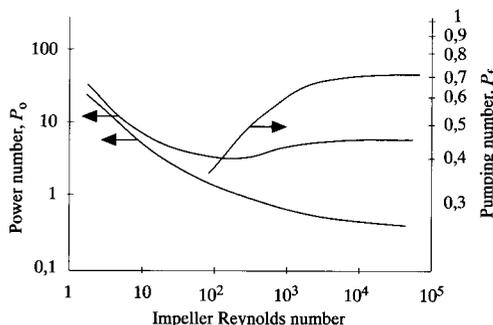


Figure 3—Dependence of the power and pumping numbers on the Reynolds number of the impeller (after Gray and Oldshue²)

Required process response in a mixer

A mixing system is designed to achieve a certain result, termed the process response of the system. It is possible that more than one process response is applicable. In gold-adsorption systems, several possibilities can be identified: suspension of the ore particles, suspension of the adsorbent phase, a more or less uniform distribution of adsorbent throughout the slurry, and a hydrodynamic condition that achieves a high rate of gold transfer from the liquid to the adsorbent phase. This paper is concerned primarily with the attainment of the dispersion of solid phase in the fluid, and peripherally with the degree to which it is distributed uniformly throughout the fluid. The relevant process responses need to be more carefully defined.

► The just-dispersed process response

If the particles in a gold-adsorption system tend to settle in the fluid, the agitation must be sufficiently vigorous for the solids to remain in suspension. Following the majority of researchers, the required process response is taken as the condition in which no particle remains stationary on the bottom of the mixer for longer than two to three seconds^{2, 4}. This condition is referred to here as off-bottom suspension.

When the solids tend to float, mixing is not a suspension but an assimilation problem. The process response in this case is defined in a way similar to that for off-bottom suspension. In gold in-pulp systems, the adsorbent phase may exhibit either type of behaviour. Because the two responses are conceptually so similar, they are referred to collectively by the same term—the just-dispersed process response. The subscript JD is applied to the various mixing variables to indicate the attainment of this condition. So, for example, N_{JD} is the agitator speed that is required for the solids to be just-dispersed in the fluid. At agitator speeds below N_{JD} , some particles will settle (or float). As the speed is increased above N_{JD} , the concentration of the solids in the fluid will move towards a more homogeneous distribution.

► The homogeneously-distributed process response

This process response has to do with the attainment of a given distribution of solids in a mixer—a matter that is relevant to gold-adsorption systems⁶. This design response is quantified in various ways based on the change in solids concentration with depth in the tank^{2, 4}.

To get even close to a homogeneous distribution requires agitator speeds—and power inputs—that are considerably greater than those needed to produce the just-dispersed condition. As Figure 1 indicates, the effort expended yields only a relatively small increase in the film coefficient, and so is invariably not economically worth while⁴. For this reason, the primary focus in this paper is on the just-dispersed process response.

Settled particles become suspended in a mixing vessel by the motion of the fluid over the bottom of the vessel. Turbulent eddies and/or viscous drag lift the particles into the body of the fluid in that region⁴. The mechanisms involved in this process are very complex and difficult to quantify. In addition, the agitator speed, and the configuration and scale of the vessel, influence the hydrodynamic conditions in ways that are not completely understood. The assimilation of particles that tend to float is even less well understood. As a consequence, it is difficult to predict the conditions needed to just disperse a solid phase in a mixer. The various approaches to this problem fall into one of two categories: predictive methods and scale-up methods.

Methods for predicting the just-dispersed condition

Suspension: empirical methods

Many empirical correlations^{2, 4, 7-9} have been developed from carefully controlled tests in laboratory-scale mixing equipment with tank diameters ranging from 0,18 to 1,8 m. (Buurman *et al.*¹⁰ report work conducted in a vessel of 4,3 m diameter.) These expressions seek to predict N_{JD} given the relevant mixing configuration and the mixing conditions. They attempt to account for all the important factors that influence off-bottom suspension such as the kinematic viscosity, ν , the particle size, d_p , the density difference between solid and fluid, $\Delta\rho$, and the solids concentration, C . The most widely used is that due to Zwietering⁹:

$$N_{JD} = S \nu^{0,1} d_p^{0,2} \left(\frac{g\Delta\rho}{\rho} \right)^{0,45} C^{0,13} D^{-0,85}. \quad [5]$$

The exponents of the parameters in the Zwietering equation have been found to be independent of impeller type, vessel size, impeller clearance, and ratio of impeller to vessel diameters. The dimensionless constant S accounts for the effect of differences in contactor geometry^{8, 9}. For circular, flat-bottomed tanks with a given type of impeller, the dependence of S on the ratio of tank-to-impeller diameter, T/D , can be described by relationships of the form

$$S \propto (T/D)^a, \quad [6]$$

where a is a parameter dependent on geometry. It has a value of 0,82 for propellers and 1,3 for flat-bladed impellers.

Suspension: model-based methods

Predictive relationships based on models of the suspension mechanism have also been developed. That developed by Baldi¹¹ is particularly interesting because the expression is non-empirical yet produces an expression somewhat similar to the empirical expression of Zwietering. It is based on an energy balance and a model for the transmission of energy by microscopic eddies. A dimensionless number, Z , expresses the ratio of the measures describing the kinetic energy imparted by eddies to the energy gained by particles on suspension. In its simplest application, it assumes that, for a given mixer geometry, this number is constant, which leads to the following expression for N_{JD} :

$$N_{JD} = Z \left(\frac{g\Delta\rho}{\rho} \right)^{0,5} \frac{1}{P_0^{1/3}} \frac{d_p}{D^{2/3}} \frac{T}{D}. \quad [7]$$

Alternatively, Z can be taken to depend on both scale and geometry. Baldi's attempt to describe this dependence is based on a Reynolds number for suspension, Re^* (equation [8]). The value of Z for a particular geometry is obtained from the relationship between Z and Re^* . Baldi's approach describes suspension by disc-turbine impellers fairly well, but is less successful with other systems⁴.

$$Re^* = \frac{\rho D^3 N_{JD}}{T\mu} \quad [8]$$

Various workers have developed suspension models by focusing on the settling rate of particles and the force required to lift them¹²⁻¹⁴. Essentially, the approach involves an energy balance (which is incomplete, as Chudachek⁷ points out) so that the minimum power needed for off-bottom suspension is equal to the force required to lift the particles multiplied by their settling velocity, u_{settle} . The predictive expression takes the form shown in equation [9], where ρ_s is the density of the solid:

$$P_{JD} = gC V \rho \frac{\Delta\rho}{\rho_s} u_{settle}. \quad [9]$$

Suspension and rates of particle settling

As equation [9] suggests, the settling rate of solid particles exerts considerable influence on the intensity of agitation required in a mixer to achieve off-bottom suspension. Particles with low settling rates are easier to suspend than those with large settling rates. Gates *et al.*¹⁵, drawing on procedures employed in industry, describe a mixer design procedure that is based on the determination of particle-settling rates.

Gold-adsorption systems: Mixing requirements

Workers take account of particle-settling rates in different ways. If expressions such as equation [9] are used, the hindered-settling rate is employed^{12, 13}, although some suggest that the minimum fluidizing velocity is more appropriate¹⁴. In other approaches, the value of the particle-settling rate at low solids concentration, u_{free} , is derived either from equation [10] or from a relevant correlation^{2, 16}.

Equation [5] suggests that N_{Df} will increase with solids concentration ($N_{Df} \propto C^0$,¹³). Particles in suspension have a dampening effect on the turbulence⁴, and particle-to-particle interactions reduce the rate at which particles settle (or float). In addition, the viscosity of the slurry increases, and its rheological properties may become non-Newtonian. As Figure 4 indicates, the effective viscosity of typical gold-plant slurries is dependent on the shear rate, i.e. the slurry exhibits non-Newtonian behaviour.

Because of hindered-settling effects, some workers suggest that, above about 17 per cent solids (by volume), no further increase in agitator speed is required to achieve off-bottom suspension⁴. Others¹⁵ find that the combined effect of hindered settling, increased viscosity, and dampened turbulence require an adjustment of the agitator speed that suggests greater, not lower, particle-settling rates.

What workers usually do to account for the influence of solids concentration is first to estimate u_{free} , the particle-settling rate at low concentrations (equation [10] or equivalent), and then to modify this value to u_{slurry} , the settling rate in the slurry. One procedure^{2, 17} for doing this is presented in equation [11]:

$$u_{free} = \left[\frac{4gd_p\Delta\rho}{3\rho_L C_D} \right]^{1/2} \quad [10]$$

$$\frac{u_{slurry}}{u_{free}} = (1 - C_v)^r, \quad [11]$$

where ρ_L is the fluid density
 C_D is the drag coefficient (obtained from its relationship with the Reynolds number of the particles, Figure 5)

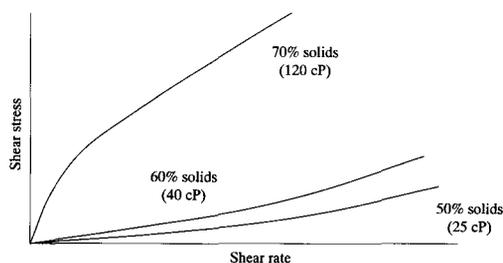


Figure 4—Non-Newtonian behaviour of typical gold-plant slurries (the effective viscosities being shown in brackets), after Bailey⁵

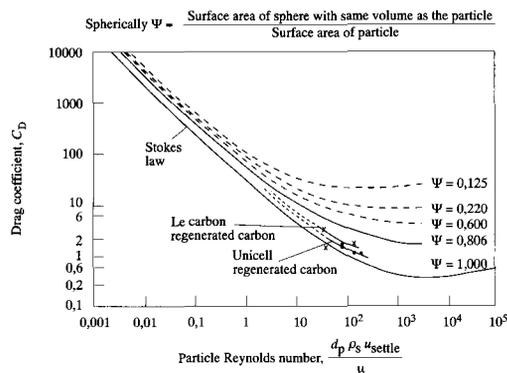


Figure 5—Drag coefficient versus particle Reynolds number (data for carbon particles settling in water are plotted on the C_D versus Re diagram due to Wadell¹⁸)

C_v is the volumetric concentration of solids

r is an exponent that varies between 4,65 and 2,33 as the flow regime moves from laminar to turbulent.

Gray and Oldshue² indicate that the nature of the suspension problem changes depending on the settling rates and concentration of the particles. They identify five categories of mixing systems, as shown in Table I. Uhl and Von Essen¹⁹ make similar distinctions by indicating that a system involving either fine particles or particles at concentrations that create hindered-settling conditions should be treated as a liquid-blending problem. They add that larger particles at moderate-to-low concentrations should be treated as a solid-suspension problem.

Table I
Types of solid-liquid mixing systems, after Table II in reference 2

	Category	Design approach
1	Low concentration of particles with settling rates < 2,5 mm/s	Treat as a single-phase liquid-blending problem
2	Low concentration of particles with settling rates from 2,5 to 100 mm/s	Suspension problem
3	Low concentration of particles with settling rates > 100 mm/s	Difficult suspension problem
4	Moderate concentration of solids with settling rates < 2,5 mm/s; slurries are moderately pseudo-plastic but show no yield stress	Treat as a liquid-blending problem
5	High concentration of solids to form slurries with high viscosities and very non-Newtonian behaviour	Care must be taken to avoid the formation of stagnant regions

Blending and process responses dependent on fluid motion

Models of fluid motion for particle suspension consider the fluid velocity across the bottom of the mixer, u_{bot} , and suggest that this velocity must exceed some critical value, u_{crit} , for the particles to be just suspended in the fluid.

Many workers^{20, 21} have shown that the fluid velocity, u , at any level in a mixer is dependent on the agitator speed, as shown in equation [12], where H is the depth of the slurry. The fluid velocity across the bottom of the vessel, u_{bot} , is governed by a similar expression. It follows that, at the just-suspended point (where $N=N_{\text{JD}}$ and $u_{\text{bot}}=u_{\text{crit}}$), the grouping on the right-hand side of equation [12] must be constant. The resulting expression, equation [13], is of limited usefulness for predictive purposes but can be used as a basis for a scale-up procedure.

$$u \propto \frac{ND^2}{(T^2H)^{1/3}} \quad [12]$$

$$u_{\text{crit}} = \text{constant} = K N_{\text{JD}} D^2 / (T^2 H)^{1/3}, \quad [13]$$

where K is a proportionality constant.

A second model relates the fluid velocities in a mixer to the speed of the impeller tip, ϕ (also known as the peripheral speed). The model⁷ focuses on the vertical flows of fluid, both downwards through an area A_i swept by the impeller and upwards through an area A_w in the outer region adjacent to the vessel walls. In geometrically similar tanks, these two areas remain in the same ratio and

$$A_w \propto A_i = \frac{\pi D^2}{4}. \quad [14]$$

The mean downward velocity through the impeller region, u_{vert} , is determined by Q , the pumping capacity of the impeller, and the pumping number, P_f (equation [3]). If, in geometrically similar systems, P_f can be assumed to have the same value, then u_{vert} is proportional to the tip speed, ϕ (equation [15]). What this implies is that, in two geometrically similar systems having the same tip speed, the downward velocity of the fluid will be the same. Provided the fluid-flow patterns are also the same, it follows from equations [14] and [15] that, for equal tip speed, the fluid motion will be equal in the outer region of the tank and also, by implication, across the bottom of the tank. If this is true, then any process response that depends primarily on fluid motion should occur at the same tip speed provided the fluid-flow patterns in the two systems are similar:

$$u_{\text{vert}} = \frac{Q}{A_i} = \frac{4P_f ND^3}{\pi D^2} \propto ND = \phi / \pi. \quad [15]$$

A similar approach can be developed with a different characteristic velocity, u_s , known as the apparent superficial velocity¹⁹ or the bulk velocity¹⁶. The term is defined in equation [16], where A is the cross-sectional area of the tank. Proponents of this approach suggest that it is more characteristic of fluid velocities within a mixer than is the tip speed¹⁶. By this view, the same process response should be achieved at the same superficial velocity in vessels of different scale—again, provided the patterns of fluid flow are similar. It should be noted that, in systems having the same T/D ratio and the same pumping number, equations [15] and [16] correspond exactly and $u_{\text{vert}} \propto u_s \propto \phi$:

$$u_s = Q / A = \frac{4P_f ND^3}{\pi T^2}. \quad [16]$$

Assimilation of floating particles

The assimilation of floating particles has received much less attention than has the suspension of settled solids⁴. Whether the particles of adsorbent tend to settle or float has a strong bearing on N_{JD} because greater mixer speeds and power inputs are required to assimilate floating particles than to suspend settled particles⁴.

Joosten *et al.*²² report a significant relationship for estimating the speed to just assimilate floating particles—equation [17], which is far less well established than Zwietering's equation. It was derived for a system involving a four-bladed impeller with the blades pitched at 45 degrees and a vessel of 0,27 to 1,8 m diameter. To promote the assimilation of the particles, a single baffle was installed to create an off-centre vortex. The solids were coarse, in the size range 2 to 10 mm.

$$N_{\text{JD}} = [0,036 (T/D)^{3,65} \left(\frac{\Delta\rho}{\rho}\right)^{0,42} \left(\frac{g}{D}\right)^{0,5}]^{1/7} \quad [17]$$

Scale-up procedures

The predictive expressions given earlier were tested for vessels with diameters of up to 1,8 m or so. For mixers considerably larger than this, the validity of these expressions is uncertain. In addition, considerable care must be exercised to ensure that a given expression has been derived from tests on systems that are similar and relevant to those under consideration. For example, Bohnet and Niesmak²³ found that nine expressions from the literature gave predictions that varied between -56 and 250 per cent of the experimental values they had obtained. They attributed these large variations to basic differences in the mixing systems from which the correlations had been derived.

Where the available predictive expressions are not applicable, or where they are of uncertain validity, the required mixing information must be determined experimentally. The scale dependence of measured parameters makes this problematic in that it is frequently unpractical to conduct the measurements at the scale of interest. Measurements are therefore conducted on a suitable scale based on the same solid-liquid system that will be mixed on a larger scale. The measurements obtained are then interpolated, i.e. scaled up, to yield the information required for the large-scale system.

The key to this approach is that the hydrodynamic conditions in the small-scale vessel are either similar to those that will occur in the larger vessel or differ in a way that can be accounted for reliably. Three types of similarity are recognized¹⁹: geometric, kinetic, and dynamic. Two systems are geometrically similar if the ratios of all the geometric dimensions (for both the vessel and the impeller) in one system are the same as in the other. The systems are kinetically similar if the patterns of fluid flow are geometrically similar. Dynamic similarity exists when the ratios of the relevant forces (e.g. Reynolds numbers, shear and viscous forces) are the same at geometrically equivalent points in the two systems. It is usually assumed that kinematically similar systems are first geometrically similar and that dynamic similarity assumes kinematic similarity. Deviations from strict geometric similarity are sometimes necessary to give kinematic similarity²⁴.

Two approaches to scale-up are followed: that employing scale-up rules and that employing scale-up equations. The first method postulates that the desired process response is achieved at the same value of a relevant mixing parameter on different scales. The second attempts to describe the scale dependence of a relevant mixing parameter.

The identification of the mixing parameter to be used as a basis for the scale-up procedure depends on the mixing system and the type of process response desired. This choice, as well as the 'correct' application of the scale-up procedure, requires considerable experience in that the process is fraught with risk. Uhl¹⁹ has suggested that scale-up is 'almost entirely an art, supported by limited concepts and a kit of imperfect tools'. However, for many mixing situations, it is the only practical approach that is available. A number of excellent reviews and articles are available on the subject^{2, 21, 24-28}.

Scale-up rules

Behind the use of scale-up rules is the idea that some key mixing parameter is very closely linked to the hydrodynamic condition that controls the process response of interest. Consequently, it is assumed that systems of different scale are dynamically similar (and so will have the same process response) when the values of that parameter are the same in both systems. So, for example, the 'rules' of constant tip speed and constant power intensity reflect the belief that the same process response will be obtained at different scales when the tip speeds or power intensities are the same. Usually, the scale-up rules are applicable only to geometrically similar systems.

The rationale behind such rules comes from an understanding of the physical factors that control the process response. The understanding may be the result of experience, a reliable correlation, or a meaningful suspension model. So, for example, equation [15] and the suspension model on which it is based provide a justification for the constant-tip-speed rule, while equation [9] lends support to the constant-power-intensity rule.

These two rules, especially the latter, have been used very extensively^{2, 24, 26, 29}. More recently, the rules of constant torque intensity and constant superficial velocity have received favour²⁴ for systems where shear and fluid motion strongly influence the process response. For turbulent flow conditions in geometrically similar systems, the constant-tip-speed rule corresponds exactly to these two rules²⁴ provided P_o and P_f do not change with the scale.

The choice of which rule to use in a given application comes from an understanding of which aspect of the hydrodynamic conditions in the system (such as shear rate, fluid motion, and energy input) most affects the desired process response. For blending and the suspension of fine solids, fluid motion is the physical factor that is important, and the scale-up rules that are recommended are constant torque intensity or constant superficial velocity^{2, 19}.

The widespread use of scale-up rules, particularly the constant-power-intensity rule, have led in some quarters to the mistaken impression that they are universally applicable and significant. Many workers warn against such error^{19, 25}. Although the approach has been used successfully many times, there have also been many instances of design failures.

Scale-up equations

In many scale-up situations, the process response is scale-dependent in a way that cannot be described by a neat and simple scale-up rule. In these situations, the scale-up procedure must be based on correlations between mixing and geometric variables^{19, 25, 26}. The correlations are termed scale-up equations or translational equations.

Geometric similarity between the two scales of interest is assumed for these correlations, so that any dimension on one scale is related to the equivalent dimension on the other scale by a single scale ratio, R . This ratio involves linear dimensions and is defined as in equation [18]. The subscripts 1 and 2 indicate the two different scales, scale 2 being the larger.

$$R = \frac{T_2}{T_1} = \frac{D_2}{D_1} = \left(\frac{V_2}{V_1}\right)^{1/3} = \text{etc.} \quad [18]$$

The scale dependence of mixing variables can be described by a simple power relationship. The scale-up equation for agitator speed, for example, is shown in equation [19]. Again, the subscripts refer to the relevant scale. Any mixing parameter, including those used in scale-up rules, can be scaled using similar expressions.

$$N_{JD2} = N_{JD1}R^n = N_{JD1}\left(\frac{T_2}{T_1}\right)^n \quad [19]$$

The value of the scale-up exponent n is obviously of crucial importance, and can be obtained from the literature or from equipment suppliers. It can also be inferred from the predictive expressions given earlier, or it can be determined experimentally for the specific mixing system under consideration.

For the suspension of particles that settle at appreciable rates in a fluid, the recommended scale-up equation is based on the power intensity¹⁹—equation [20]. The recommended scale-up exponent is $-0,25$, although this value may vary somewhat with the nature of the mixing system.

$$E_{JD,2} \propto E_{JD,1} (T_2/T_1)^{-0,25} \quad [20]$$

Provided P_o and P_f are constant, a scale-up equation based on one mixing parameter can be converted to an equivalent equation based on a different mixing parameter. This suggests that the choice of mixing parameter is not important. Here, experience is needed. Different equipment manufacturers prefer one mixing parameter over another for a variety of reasons. Clearly, if P_o or P_f is not constant with scale, the choice of the scale-up equation can be very important:

$$X_{JD} \propto (TD)^{tD} D^d \quad [21]$$

It is possible to extend the scale-up approach to systems that diverge from geometric similarity. Rautzen²⁵ does not recommend this practice, but suggests that the effect of variation in geometry be examined before the scale-up procedure is applied. In the present paper, an expression of the form of equation [21] is considered where X_{JD} represents any mixing variable at the just-dispersed point (e.g. N_{JD} or E_{JD}).

To conclude the review of scale-up procedures it is worth stating that a scale-up rule is really a scale-up equation with a particular value for the scale-up exponent n . Obviously, this value will be zero if the scale-up equation and rule are based on the same mixing parameter. However, if the scale-up equation is based on a different parameter, the value may not be zero. So, for example, the constant-power-intensity rule is equivalent to an agitator-speed scale-up equation (equation [19]) having a value of $-0,67$ for n . If the value of this exponent were -1 , the scale-up equation would be equivalent to a constant-torque-intensity scale-up rule.

Design of contactors for gold in-pulp adsorption

Gold slurries are three-phase systems consisting of fine particles of ore at concentrations of about 37 per cent by volume (50 per cent by mass) in an aqueous solution along with coarse particles of adsorbent at concentrations of about 2 per cent. The rheology of gold slurries is non-Newtonian but, as Figure 4 indicates, only moderately so at the solids concentrations commonly found in adsorption systems.

In the development of carbon-in-pulp (CIP) technology, the adsorption vessels were designed on the assumption that adequate suspension (or assimilation) of the adsorbent phase should occur once the ore particles had been adequately suspended. Essentially, the vessels were designed as mixers. Investigations showed that, with this approach, reasonable distributions of carbon in the slurry could be maintained under normal operating conditions⁶. However, as Figure 1 suggests, uniform distribution of the adsorbent may not be as important a factor from a mass-transfer perspective as ensuring that the adsorbent is at least just dispersed in the slurry.

If contactors for gold in-pulp adsorption are to be designed optimally, among the many factors that must be considered, those affecting the rate of gold adsorption should be given special prominence. The contactors need to be designed as mass-transfer units, rather than as mixers. To do this, attention must be given to three primary factors:

- suspension of the fine particles of ore (to ensure that the system remains fluid)

- ▶ dispersion of the adsorbent phase (to ensure that the particles are at least just dispersed)
- ▶ influence of the design on film-transfer coefficients (to give as high a gold-adsorption rate as economically possible).

The last aspect was addressed in the previous paper¹. The scale-up procedure developed there required a reasonable estimate of E_{jd} , the specific power at the point where the adsorbent is just dispersed. The information is the same as that required for an understanding of the dispersion needs of the adsorbent phase.

As the suspension of the ore particles fits categories 1 and 4 in Table I, this aspect of the design is treated as a liquid-blending problem controlled by fluid motion. Scaling of the just-suspended point for the ore particles should therefore be based on the constant-torque-intensity or constant-superficial-velocity rule.

Experiments can be conducted to show the torque intensity (or superficial velocity) at which the particles of ore will be adequately suspended in a small-scale system. The same torque intensity (or superficial velocity) must then be achieved on the larger scale.

Alternatively, values of these variables can be selected based on previous experience. Some typical values are reported in the literature. Hicks *et al.*¹⁶ describe a procedure used industrially in which each type of blending or suspension problem is associated with a given range of superficial velocities. They recommend superficial velocities in the range 0,09 to 0,18 m/s for the suspension of low concentrations of particles with settling rates of less than 2 mm/s, or the blending of moderately low-viscosity liquids (Table I in reference 16). Lyons² based his design procedure on a tip speed of 3,5 m/s. Two South African equipment suppliers indicate that the design of CIP adsorption vessels begins with typical values of either torque intensity or superficial velocity. The values cited were between 6 and 7 Nm/m³ for the equivalent torque intensity, τ_{eq} , and around 18 mm/s for the superficial velocity.

Little attention has been given in CIP technology to the prediction of the just-dispersed point for the adsorbent phase. In the design of South African CIP contactors, it is usually assumed that dispersion of the carbon phase needs to be addressed only when the slurry density is very much lower than the carbon density. In such situations, the contactors are designed to the same scale-up rules as are employed for the suspension of the ore particles.

The adsorbent-gold slurry system is more complex than the two-phase systems for which most scale-up procedures have been developed and tested. It is therefore not certain whether these procedures are fully applicable to the dispersion of the adsorbent in gold-extraction systems. To throw some light on this problem, the specific problem of the dispersal of carbon in CIP contactors is examined more closely.

Dispersion of carbon in CIP processes

Theory

It is unlikely that successful estimation of the point at which carbon is just dispersed in gold slurries can be achieved by use of the predictive expressions discussed previously. The systems are too complex. Scale-up procedures must therefore be employed. However, the most appropriate procedure is not immediately apparent.

The first step must obviously be to assume that the slurry acts as a single-phase liquid. This appears to be a safe assumption provided that the particles of ore are small and are reasonably well suspended and distributed in the slurry. Figure 4 suggests that such a slurry will have an effective viscosity in the region of 25 cP, although the values will be higher for ores that are very finely ground. The design problem is then seen to be one in which a solid at low concentrations must be dispersed in a liquid of near-Newtonian rheology. The complication introduced by the possibility that the adsorbent may tend to float is considered later.

From Table I it is clear that the design procedure for the suspension of particles of adsorbent depends on their effective settling rates. The table refers to a critical settling rate of 2,5 mm/s. If the particles settle at a rate below this critical value, the suspension problem should be treated as a liquid-blending problem, and a scale-up rule should be used for the design. If they settle at rates greater than this, the design should be based on a scale-up equation, equation [20] being the one most widely employed.

To ascertain which design approach is appropriate, information on the settling rates of carbon particles in the slurry is needed. Settling rates are difficult to measure. However, rough estimates can be made by use of equation [10] and the density data given in Table II. For a value of the applicable drag coefficient, C_D , reference can be made to the settling-rate data reported by Baguley³⁰. The relevant information is presented in Table III and is plotted on curves of C_D versus particle Reynolds number¹⁸ in Figure 5. What this figure reveals is, firstly, that the particles of regenerated carbon behaved almost like spheres, sphericities between 0,93

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Table II

Densities of the phases in CIP systems

Phase	Typical density, kg/m ³
Gold slurry	1450
Pure carbon (graphite)	2100 to 2300
Dry activated carbon ³⁰	600 to 960
Wet activated carbon: After Bailey ⁶	1600 to 1760
After Baguley ³⁰	1300 to 1400

Table III

Settling rates and drag coefficients of regenerated carbon in water

Size class, mm	-2,38 +1,7	-1,7 +1,4	-1,4 +1,18	-1,18 +1,1
Representative size, mm	2,01	1,54	1,29	1,09
Unicell *Density: 1400 kg/m ³ *Settling rate, mm/s Drag coefficient †Reynolds number, Re _p	95,2 1,2 191	79,4 1,3 122	68,5 1,4 88	56,5 1,8 62
Le Carbon *Density: 1230 kg/m ³ *Settling rate, mm/s Drag coefficient †Reynolds number, Re _p	57,8 1,8 116	53,2 1,6 82	50,8 1,5 65	37,0 2,4 40

* Data obtained from references 6 and 30

† Re_p = ρ_sd_pμ_{free}/μ

and 0,97 being indicated; and, secondly, that the maximum drag coefficient for carbon particles is in the region of 1,2 to 2. Because the slurry has a much greater viscosity than that of water, settling is likely to occur under more laminar conditions at much lower settling rates. Figure 6 shows the limits of the settling rates for the largest and smallest particles typically found in CIP systems. The lower limits were calculated

from Stokes Law (equation [22]) on an assumed viscosity of 25 cP, and are indicated by the dark lines in the diagrams. Because the particles have sphericities so close to unity, an adjustment for particle shape is not needed. The upper limits were determined from equation [10] with a drag coefficient of 1,2.

$$u_{\text{settle}} = \frac{g d_p^2 \Delta \rho}{18 \mu} \quad [22]$$

Figure 6 shows how the difference between the densities of the carbon and the slurry might affect the settling rates of the carbon particles. The effective settling rates in CIP operations will be somewhere between these limits—probably closer to the lower values associated with laminar settling conditions. (It should be noted that, in the evaluation of these settling rates, attention is focused, not on the degree of turbulence in the contactor, but on the flow conditions that arise because of the way the particle settles.)

In Figure 6, a band of settling rates has been superimposed to indicate the region where dispersion of the particles can be treated as a liquid-blending problem. What is indicated is that, if all the carbon particles are small, their settling rates will be below the limit of 2,5 mm/s provided the carbon-slurry density difference is no more than about 180 kg/m³. However, the situation changes significantly as the size of the carbon increases. With large particles, carbon-slurry density differences greater than 20 kg/m³ (0,02 g/ml) result in settling rates greater than the critical value of 2,5 mm/s. At greater density differences still, the settling rates are significant. For example, with particles of 2,5 mm, a density difference of 50 kg/m³ will result in a settling rate of 6,8 mm/s or more, while a density difference of 100 kg/m³ will result in a settling rate of 13,6 mm/s or more.

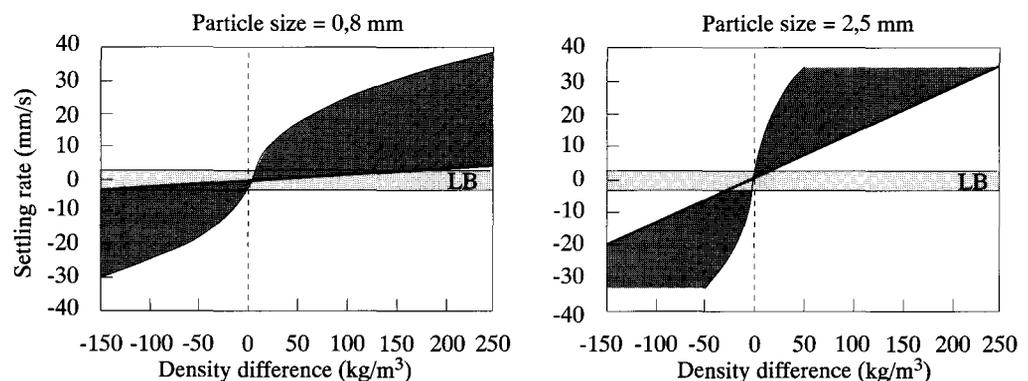


Figure 6—Estimated settling rates of carbon particles in a slurry having a viscosity of 25 cP. (The dark line indicates the limiting settling velocity under laminar-flow conditions. The heavily shaded region indicates the range of settling rates when the particles settle under more turbulent conditions. The range of settling velocities over which a liquid-blending design approach applies is indicated by the lightly shaded region LB)

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Figure 7 shows the effect of particle size more clearly. The density difference—the carbon-slurry density difference that will result in a settling velocity of 2,5 mm/s—is plotted as a function of particle size. (Only the minimum settling rates are considered in the figure.) If the density difference is less than this value, the settling rates are low, and a liquid-blending approach with a scale-up rule is indicated. If they are greater, the design should be based on a solids-suspension approach using a scale-up equation.

Figure 7 is not intended as a design tool: in practice, the designer finds a range of particle sizes, and the slurry density fluctuates to some extent. In addition, Figure 7 applies only to a situation in which the slurry viscosity is 25 cP. Rather, what it indicates is that the design of contactors on a liquid-blending basis may not be optimal for a particular case. Even rather small fluctuations in slurry density (or relatively small slurry-carbon density differences) will mean that, at the least, the larger particles will have significant settling rates—a situation not adequately catered for by the liquid-blending approach. Consequently, the designer cannot be sure that the contactor has been adequately designed to keep the carbon dispersion above the just-dispersed point.

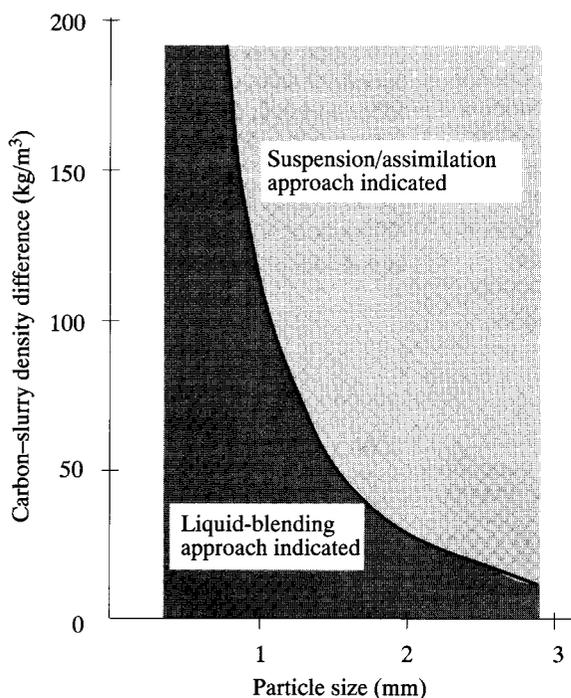


Figure 7—Regions indicating the appropriate approach for the design of CIP contactors. (The graph shows how the design approach depends on the particle size of the carbon and on the carbon-slurry density difference when the slurry has a viscosity of 25 cP. The boundary is defined by a settling rate of 2,5 mm/s. The settling rate is calculated by use of equation [22])

To account for these situations, the designer should consider using suspension principles and a scale-up equation. However, a problem arises in that it is not certain how the scale-up equations developed for simple two-phase mixing systems will apply to more complex three-phase systems in which the particles settle in a slurry. The problem is further aggravated by the general lack of information about systems in which the particles have a tendency to float. To address these various problems here, some relevant experimental data are examined.

Experimental data

The previous paper¹ described a series of mass-transfer tests conducted on CIP systems involving an adsorption suite of synthetic gold slurry and virgin carbon that was fairly representative of South African conditions. The mixer configuration used was that given in Figure 2. The diameters of the tanks involved in the test programme ranged from 0,185 to 1,2 m. The impellers were all 20-degree three-bladed hydrofoils. The particle densities were low so that the particles tended to float at low levels of agitation intensity.

The just-dispersed point was identified from data on both the agitator and the power input. The method employed was to identify the break-point in the relationship between film coefficient and either agitator or power input, point A in Figure 1. The method has been shown to give similar results to those obtained by other methods^{8, 3}, although some discrepancies have been reported⁸. Nevertheless, for the current purpose, this technique is the most appropriate because mass-transfer effects are of primary concern.

The power input was determined by measurement of the restraining torque necessary to prevent rotation of the drive arrangement. (The details are presented elsewhere^{1, 32}.) Some workers account for system friction when measuring power^{33, 34}, but most do not (or do not report doing so). For the data considered here, both types of measurements were made, and two sets of power data were therefore available. In the previous paper¹, the correlation of film coefficients with power or specific power was reported as being essentially the same for both sets of power data. However, it was found that the CIP dispersion data were very sensitive to the power data, and both sets of data gave poor correlations. The data set giving the better correlations was used in the work reported here.

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Table IV
Data on the dispersion of activated carbon in a typical gold slurry

T mm	D mm	N_{jD} r/min	P_{jD} W
185	110	288	2,86
305	110	472	7,86
	130	365	7,22
	142	329	5,95
330	110	570	14,60
	130	404	12,74
	142	371	10,72
690	275	255	90,37
1200	480	135	333,8

It is important to appreciate that the CIP data presented here (Table IV) are a byproduct of a study on mass-transfer coefficients. Although the information is limited in extent, it is still interesting in that no body of data relevant to the just-dispersed condition of an adsorbent phase in a gold slurry has been reported previously.

Results

As an examination of the applicability of available scale-up procedures for the prediction of the just-dispersed point of the adsorbent in gold in-pulp systems, five questions are addressed. Thereafter, relevant conclusions are drawn and appropriate recommendations are made.

Question 1: How did the power number vary with the agitator speed?

In the design of a mixing system, it is necessary to select both the agitator speed and the power needed to maintain that speed. For large-scale systems, one of these terms is estimated by use of an appropriate scale-up procedure, while the other is inferred by use of the power number, P_o , and equation [1]. Errors in the value of the power number employed will result in errors in the selection of one of these important design variables. (Other things being equal, such errors are minimized if the agitator speed is inferred from the power because P_o is raised to the third power in the calculation.) Clearly, the scale dependence of P_o should be understood.

Figure 3 shows that P_o is constant in the fully developed turbulent regime, where solid dispersion normally occurs. In fully turbulent, fully baffled vessels, it is usually assumed that the power number is dependent only on impeller type and geometry¹⁰. Curves relating the power number to the Reynolds number are available³⁵. It is sometimes assumed that the power number is the same for impellers of any diameter⁸. However, it is very evident that this is far from being universally true. Bates³⁵ reports a variation of 20 per cent in P_o with impeller diameter. For a disc impeller, Chapman⁸ found that the power number increased from 3,15 to 5,95 when the impeller diameter changed from 97 to 280 mm.

He also reported that P_o increased with D/T , and found it to be very sensitive to minor differences in impeller geometry. A recent careful study³⁶ detailed such effects for a 45-degree pitched-blade impeller (downflow) in the form of equation [23]. That correlation suggests that a tenfold increase in scale will increase the power number by a factor of 1,8, even though the systems are geometrically similar. Clearly, variations of such magnitude, if ignored, will lead to serious error. If P_o can vary significantly in the simple two-phase systems from which the above conclusions have been drawn, the potential for variation in complex gold-slurry systems is significant:

$$P_o = 0,653 T^{0,26} \left(\frac{T}{D}\right)^{0,11} \left(\frac{B}{T}\right)^{-0,23} n_b^{0,68} \theta^{1,82}, \quad [23]$$

where B is the clearance of the impeller above the tank bottom
 n_b is the number of blades in the impeller
 θ is the angle of the blades in radians (rd).

Figure 8 shows the power number as a function of agitator speed below and above the just-dispersed point in the gold-adsorption systems tested. P_o is not plotted against the Reynolds number of the impeller because of uncertainty about the viscosity. Although the three vessels were almost geometrically similar, very different power numbers were obtained. The following points emerge from an examination of Figure 8.

- ▶ **The power number is not constant in the region of the just-dispersed point.** In the smallest vessel well above the just-dispersed point, the power number is essentially constant with agitator speed. (Its value of 0,36 agrees well with the value of 0,35 given by the manufacturer.) However, in the region of the just-dispersed point, its value is about 0,41. In the two larger vessels, the power number decreases markedly with increased speed.
- ▶ **Large-scale agitation of gold slurries by hydrofoil impellers probably does not occur under fully turbulent conditions.** In a given mixing system, a variation in the power number with agitator speed indicates that the system is not fully turbulent. It can be seen that, in all cases, the particles of adsorbent were just dispersed before the slurry was in the fully turbulent regime. This is particularly evident in the larger vessels. These observations agree with those of Gray and Oldshue², who report that, in the suspension of slurries, fully turbulent conditions can occur only near the impeller.

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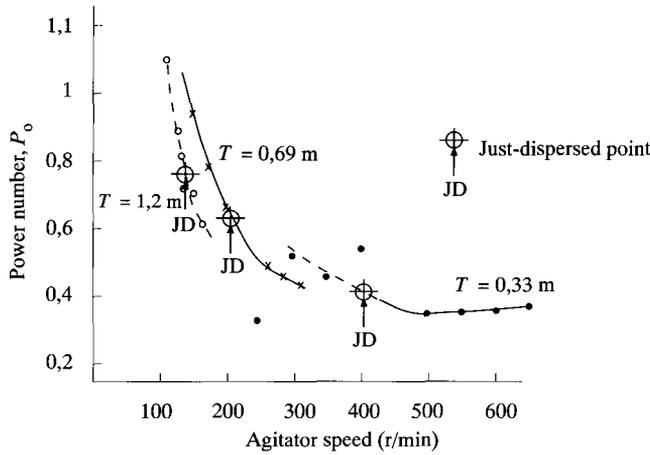


Figure 8—Power numbers for the CIP systems tested

Table V
Power numbers in gold CIP adsorption vessels

	Power number, P_0			
	T (m) = T/D =	0,33 2,538	0,69 2,509	1,2* 2,5
For $N=N_{JD}$ in Figure 8 Minimum P_0 when $N>N_{JD}$		0,41 0,35	0,64 0,42	0,77 0,62
Values of P_0 scaled by use of equation [18]		[0,35]†	0,42	0,48

* The large impeller was lined with rubber, and the others were not
† This value was taken as the reference point for the scale-up calculation

Table VI
Comparison of various correlations for N_{JD}

Correlation	$d1$	$t1$	Impeller type
Equation [17]: Zwietering ⁹	-0,85	0,82	45° pitched blade Flat blade Hydrofoil
Gold CIP data	-0,85	1,3	
Gold CIP data	-0,77	1,09	
	$t1$	$d1$	p
Equation [20]: Baldi ¹¹	1,0	-0,667	-0,333
Gold CIP data	0,99	-0,766	-0,118
Equation 17: Joosten ²²	1,91	-0,707	
Gold CIP data	1,09	-0,77	

Table VII
Values of key dispersion variables in geometrically similar CIP systems

Design variable	T , mm D , mm T/D	330 130 2,538	690 275 2,509	12000 480 2,500
N_{JD} , r/min		404	255	135
θ_{JD} , m/s		2,75	3,67	3,39
E_{JD} , W/kg		0,31	0,24	0,17
τ_{JD} , Nm/m ³		10,7	13,2	17,4
u_{sJD} , m/s		0,104	0,142	0,132

* The values of u_s were calculated on the assumption that the pumping number was constant at 0,6

► The power number at the just-dispersed point is scale-dependent. It is not clear from Figure 8 whether the power number in the fully turbulent region is likely to be scale-independent in geometrically similar vessels. What is evident is that, in the region of interest—the just-dispersed point, the power number is scale-dependent. Table V presents the relevant data.

Question 2: How well do the suspension correlations given in the literature describe gold-adsorption systems?

The scale-up equations developed from the Zwietering and Joosten expressions take the form of equation [21], while the Baldi equation reduces to the form of equation [24]. Table VI compares the values of the exponents in these equations with those obtained for the CIP data:

$$N_{JD} \propto (T/D)^{t1} D^d P_0^p \quad [24]$$

It can be seen that the published values of the exponents in the Zwietering expression are close to the values obtained from the CIP data. This is a little surprising in view of the differences in the associated mixing systems. In the Joosten expression, the CIP data give exponents that are similar for $d1$, but not for $t1$. In the Baldi expression, the agreement is good for the value of $t1$, not very good for the value of $d1$, and poor for the value of p .

Question 3: Are any of the key dispersion variables independent of scale?

To determine whether or not it is appropriate to apply a scale-up rule for the dispersion of the adsorbent phase in gold slurries, the scale dependence of the mixing parameters at the just-dispersed point is examined. Relevant data for the three geometrically similar systems tested are presented in Table VII.

As expected, N_{JD} —the agitator speed to just disperse the carbon—decreases with an increase in the size of the vessel, while the specific power and the specific torque decrease and increase respectively. The variation in the tip speed to just disperse the adsorbent, θ_{JD} , varies with scale, but does not show any specific trend. It is interesting to note that the apparent superficial velocity at the just-dispersed point is the variable that is closest to being scale-independent, although this observation may arise from the assumption that the pumping number is the same at the three scales.

Question 4: What scale-up equations are appropriate for the dispersion of the adsorbent phase?

In Table VIII, the scale dependence of various mixing parameters in the CIP systems tested is demonstrated, all the available data being considered. Table IX demonstrates the same information but only for those systems that were geometrically similar.

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Table VIII

Scale dependence of key dispersion variables

Correlation	Correlation coefficient
$N_{JD} \propto (T/D)^{1.09} D^{-0.775}$	0,987
$[\theta_{JD} \propto (T/D)^{1.09} D^{0.23}]$	[0,956]
$u_{sJD} \propto (T/D)^{-0.91} D^{0.23}$	0,937
$E_{JD} \propto (T/D)^{-0.15} D^{-0.351}$	0,624
$\tau_{JD} \propto (T/D)^{-1.237} D^{0.417}$	0,619

NB: The correlation involving tip speed is identical to that involving agitator speed after modification of the exponent on D to account for the fact that $\theta = ND$

Table IX

Scale dependence of key dispersion variables in geometrically similar systems

Correlation	Correlation coefficient
$N_{KD} \propto T^{-0.837}$	0,98
$E_{JD} \propto T^{-0.462}$	0,98
$\tau_{JD} \propto T^{0.373}$	0,97

Table X

Comparison of scale-up exponents for particle dispersion

Source	Exponent on N_{JD}	Exponent on E_{JD}	Exponent on τ_{JD}
Zwietering ⁹	-0,85	-0,55	
Baldi ¹¹	-0,67	0*	
Chapman ⁸	-0,76	-0,28	
Rautzen ²⁵	-0,75	-0,25	
Uhl ¹⁹		-0,25†	
Gold CIP data:			0,5
From Table IX	[-0,84]	[-0,46]	(0 - 0,67)‡
From Table VIII	-0,77	-0,35	0,37

* Baldi held that the exponent for a disc impeller should be zero but that, for other impellers, it would be a function of geometry

† The value of the exponent varies with the mixing system

‡ The value of this exponent depends on the concentration and settling rate of the solids. When particles settle in the fluid at rates greater than about 0,5 cm/s, the exponent decreases from 0,5 to zero as the concentration of the solids increases from zero to that of a dense slurry. If the settling rates are very high, the exponent may be as high as 0,67 for a dilute slurry¹⁹

As can be seen from Table VIII, N_{JD} correlates with the geometric variables significantly better than does the specific power, E_{JD} , or the torque intensity, τ_{JD} . Chapman *et al.*⁸ make a similar observation, finding that the minor scatter in the N_{JD} data is magnified in the E_{JD} data. Their explanation for this was that there were minor geometric differences in their experiments between the vessels of different sizes: some vessels had a sparge pipe or a centre support bearing at the base of the vessel, while others did not. Their power data appear to be much more sensitive to these slight differences in geometry than are their data for agitator speed.

Something similar may have affected the CIP data reported here. The impellers of 142 and 480 mm diameter were lined with rubber, while the others were not. Further, in the large vessel, the support structure for a stabilizing bearing near the top of the tank was just submerged in the slurry.

The observed sensitivity does not invalidate the usefulness of correlations based on power (or variables derived from them), but it does indicate that caution is needed in their use.

Table X compares values for relevant scale-up exponents from the literature with values obtained from the CIP data. Two values for each exponent are given for the CIP data. The first, from Table IX, indicates the value obtained for geometrically similar vessels. The second, from Table VIII, which is derived from the broader correlations involving all the available data, should be the more reliable.

The values of all the exponents on N_{JD} are very similar, with the exception of the value reported by Baldi. The values are in the region of -0,75. This value indicates that the scale-up equation is closer to a constant-power-intensity rule (when it would have a value of -0,67) than to a constant-torque-intensity rule (when its value would be -1).

The agreement between the various exponents on E_{JD} is not good. Nienow⁴, following Chapman⁸, recommends that the more conservative exponent of -0,28 should be used. This value is in good agreement with the value of -0,25 recommended by Rautzen and others^{2, 25}, who present various scale-up procedures based on industrial practice. The more reliable exponent from the CIP data is -0,35.

To complete the examination of the scale dependence of key variables, the effect of changes in impeller diameter in a given vessel is considered. Chapman⁸ found that the exponent dI in equation [25] was -1,5. He claimed reasonable agreement with the value that would arise from Zwietering's equation, i.e. -1,67. From the CIP data, the value of the exponent was found to be -1,45 for the tank of 305 mm diameter and -1,73 for the tank of 330 mm diameter. A more reliable value was obtained from the regression on the full data set given in Table VIII. This value is -1,67, which agrees exactly with Zwietering:

$$N_{JD} \propto D^{dI} \quad [25]$$

It emerges from this discussion that, for the dispersion of carbon in the CIP system tested, the most reliable scale-up equation involves the agitator speed. This equation is not sensitive to minor geometrical differences between the scales, and is in good agreement with the correlations found in the literature.

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References

- WOOLLACOTT, L.C., and AFEWU, K.I. Scale-up procedures for gold-adsorption systems. Part 1: Adsorption kinetics. *J. S. Afr. Inst. Min. Metall.*, vol. 95, no. 4. Jul./Aug. 1995. pp. 167-177.
- GRAY, J.B., and OLDSHUE, J.Y. *Mixing: theory and practice*. Uhl, V.W., and Gray, J.B. (eds.). New York, Academic Press, 1986. vol. 3, chap. 12.
- SWEENEY, E.T. *An introduction and literature guide to mixing*. Fluid Engineering Series: vol. 5. Cranfield, (UK), BHRA, 1978.
- NIENOW, A.W. *Mixing of liquids by mechanical agitation*. Ulbrecht, J.J., and Patterson, G.K., (eds.). Gordon and Breach Science Publishers, 1985. chap. 8.
- BAILEY, D. Private communication.
- Bailey, P.R. *The extractive metallurgy of gold in South Africa*. Stanley, G.G. (ed.). Johannesburg, South African Institute of Mining and Metallurgy, 1987. Chap. 9.
- CHUDACHEK, M.W. *Ind. Eng. Chem. Fundam.*, vol. 25. 1986. pp. 391-401.
- CHAPMAN, C.M., NIENOW, A.W., COOKE, M, and MIDDLETON, J.C. *Chem. Eng. Res. Des.*, vol. 61. 1983. pp. 71-81.
- ZWIETERING, T.N. *Chem. Engng Sci.*, vol. 8. 1958. pp. 244-253.

Question 5: Does the design procedure for the suspension of ore particles result in insufficient agitation for adequate dispersion of the carbon?

This question can be answered only for the specific adsorption systems for which data are available. To address the question to the system tested, a comparison must be made between the data in Table VII and the scale-up rules used for the design of CIP contactors (as described earlier).

The superficial velocities recommended by Hicks¹⁶ are in the range 0,09 to 0,18 m/s, while South African CIP plants are designed with values around 0,18 m/s. The latter value would be adequate for the dispersion of carbon in the systems tested—Table VII shows values for u_{SD} ranging from 0,1 to 0,14 m/s.

The equivalent torque-intensity values used in the design of CIP contactors are typically in the range 6 to 7 Nm/m³. For the systems tested, this translates into torque intensities in the range 8,7 to 10,2 Nm/m³. Table VII indicates that this range would be inadequate for carbon dispersion in the systems tested—the values of τ_{TD} increasing from 10,7 Nm/m³ as the scale is increased.

Table XI presents the torque intensities and superficial velocities that would result from scaling up of the agitator speed as recommended in the previous discussion. As can be seen, the values are all greater than those indicated by the scale-up rules that are usually applied to the design of CIP contactors. The tip speeds are also greater than the range of 200 to 300 m/min that is thought to be safe for minimizing the attrition rate of carbon. These observations, as well as the scale-up equations that have been recommended, must remain tentative until a stronger set of data is available than that presented here.

Discussion

The dispersion data examined here originated from systems in which the carbon particles tended to float. As such, the conclusions that can be drawn provide some insight into an area of mixing that has received very little attention. The following points have emerged.

- An examination of the power numbers obtained suggests that the dispersion of the carbon occurred before the slurry was fully turbulent. In addition, it was found that the power number at the just-dispersed point was significantly dependent on scale, even in geometrically similar systems. One consequence of this is that the calculation of power from scaled-up agitator speeds by use of a constant power number would under-estimate the power requirements. In addition, scale-up equations based on different mixing parameters are not likely to correspond to one another in the way that they would if the power and pumping numbers were constant and scale-independent. Therefore, for gold in-pulp systems, one scale-up equation may prove to be more reliable than another.
- No appropriate scale-up rule was found for the dispersion of carbon particles. The agitator speed and specific power decreased, and the torque intensity increased, with increasing scale in geometrically similar vessels. The indication was not as clear for the apparent superficial velocity, u_{SD} . In that case, there was some variation with a change in scale, but the variation was reasonably small. It will require a better set of data to reveal whether the u_{SD} for the dispersion of adsorbent particles remains essentially constant with scale for gold-adsorption systems.
- It is apparent from (2) that the scale-up of the system tested requires an appropriate scale-up equation. This suggests that the mixing in the specific CIP system tested was a suspension-assimilation problem and could not be treated as a liquid-blending problem. Power and torque intensities appear to be rather sensitive to minor variations in system geometry, and therefore do not offer as reliable a basis for scale-up as equations based on agitator speed. The scale-up equation recommended by Rautzen²⁵, but with a scale-up exponent of -0,76, seems to be the most appropriate—equation [26]. It should be noted that Rautzen recommends that strict geometric similarity should be maintained when this scale-up equation is used.

$$N_{D,2} = N_{D,1} (T_2/T_1)^{-0,76} \quad [26]$$

Table XI

Tentative scale-up data for the CIP systems tested
($\rho = 1450 \text{ kg/m}^3$, $F_0 = 0,35$, and $F_1 = 0,6$)

Parameter	Measured	Scaled	
		6	12
Tank diameter, m	1,2	2,4	4,8
Impeller diameter, m	0,48	0,96	1,92
$N_{D,1}$, r/min	135	40	23,5
Tip speed, m/min		302	354
Torque intensity, Nm/m ³		16,8	23,3
Superficial velocity, m/s		0,20	0,23

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10. BUURMAN, C., RESOORT, G., and PLASCHKES, A. *Fifth European Conference on Mixing, Wurzburg, Germany, 10-12 June, 1985*. Cranfield (UK), BHRA Fluid Engineering, paper 5.
 11. BALDI, G., CONTI, R., and ALARIA, E. *Chem. Engng. Sci.*, vol. 33, 1978, p. 21.
 12. KNEULE, F. *Chem. Ing. Tech.*, vol. 55, 1983, p. 275.
 13. KOLA, V. *Coll. Czech. Chem. Comm.*, vol. 26, 1961, p. 613.
 14. CLOETE, F.L.D., and COETZEE, M.C. *Powder Technology*, vol. 46, 1986, pp. 239-243.
 15. GATES, L.E., MORTON, J.R., and FONDY, P.L. *Chem. Eng.*, vol. 83, 24th May, 1976, pp. 144-150.
 16. HICKS, R.W., MORTON, J.R., and FENIC, J.G. *Chem. Eng.*, vol. 83, 26th April, 1976, pp. 102-110.
 17. MAUDE, A.D., and WHITMORE, R.L. *Br. J. Appl. Phys.*, vol. 9, 1958, p. 477.
 18. WADELL, H. *J. Franklin Inst.*, vol. 217, 1934, p. 459.
 19. UHL, V.W., and VON ESSEN, J.A. *Mixing: theory and practice*. Uhl, V.W., and Gray, J.B. (eds.). New York, Academic Press, 1986, vol. 3, chap. 15.
 20. NIENOW, A.W., and BARTLETT, R. *Proceedings First European Conference on Mixing and Centrifugal Separation*. Cranfield (UK), BHRA, 1975, pp. B1-B15.
 21. LEVINS, D.M., and GLASTONBURY, J.R. *Trans. Inst. Chem. Eng.*, vol. 50, no. 32, 1972, p. 132.
 22. JOOSTEN, G.E.H., SCHILDER, J.G.M., and BROORE, A.M. *Trans. Inst. Chem. Eng.*, vol. 55, 1977, p. 220.
 23. BOHNET, M., and NIESMAK, G. *Ger. Chem. Eng.*, vol. 3, 1980, p. 57.
 24. OLDSHUE, J.Y. *Mixing of liquids by mechanical agitation*. Ulbrecht, J.J., and Patterson, G.K. (eds.). Gordon and Breach Science Publishers, 1985, chap. 9.
 25. RAUTZEN, R.R., CORPSTEIN, R.R., and DICKEY, D.S. *Chem. Eng.*, 25th Oct., 1976, pp. 119-126.
 26. RUSHTON, J.H. *Chem. Eng. Progress*, vol. 47, no. 9, 1951, pp. 485-488.
 27. PENNY, W.R. *Chem. Eng.* 22 Mar., 1971, p. 86.
- (4) It is apparent from the scale dependence of the power number that there will be some difficulty with the estimation of the power and torque required on a large scale. For the scaling up of film coefficients, which requires a value of E_{JD} on the larger scale, there will also be difficulty. In the absence of correlations of power number against Reynolds number for gold-slurry systems, it is recommended that both the agitator speed and the specific power should be scaled, rather than the application of the power number to a scaled-up value of the agitator speed. The recommended scale-up equation is as follows:
- $$E_{JD,2} \propto E_{JD,1} (T_2/T_1)^{-0.25} \quad [27]$$
- A conservative value for the scale-up exponent $t1$ is suggested because of the sensitivity of the CIP data to power measurements. Chapman's value of -0.28 (Table VIII) is recommended, but the impact of values as low as -0.35 or -0.4 should also be tested.
- Scaling up of the specific power, rather than the specific torque, is recommended here because this appears to be the more common practice in suspension problems where a scale-up equation (rather than a scale-up rule) is employed². Table VI suggests that both variables correlated with the CIP data equally poorly.

Conclusions

There are two mixing issues in gold in-pulp adsorption systems. The first involves the suspension of fine particles of ore so that settlement does not occur during operation and re-suspension can be achieved readily after a shutdown. The second involves the dispersion of the adsorbent phase in the slurry. Mass-transfer considerations dictate that, at the very least, the adsorbent must be just dispersed in the slurry so that the full surface area of these particles is available for gold adsorption. In addition, if film-transfer coefficients are to be scaled successfully, it is necessary to scale up the value of E_{JD} —the specific power at the point where the adsorbent is just dispersed. This is required irrespective of whether it is needed for the design of the mixing conditions in the contactor.

In regard to the first mixing issue, the suspension of fine particles of ore, no specific data were examined, and current design practices were therefore accepted without comment. The procedures involve scale-up rules based on equal fluid motion. Constant-equivalent-torque-intensity or constant-superficial-velocity is the rule commonly applied. Typical values that are used are 6 to 7 Nm/m³ for the equivalent torque intensity, and around 0.18 m/min for the apparent superficial velocity. The constant-power-intensity or constant-tip-speed rule is no longer favoured for the scaling up of mixing requirements. However, tip speeds need to be in the range 200 to 300 m/min if high levels of adsorbent attrition are to be avoided.

In regard to the second mixing issue, the dispersion of the adsorbent phase, two factors need to be assessed. The first concerns the likely settling rates of the particles. The second is whether the adsorbent phase has a tendency to settle or float at the slurry densities envisaged in the contactor.

If the settling rates of the particles are low—below about 2.5 mm/s—the recommended design procedure is the same as that employed for the suspension of the ore particles, whether the particles tend to float or settle. Settling rates are a function of the particle size, the slurry viscosity, and the difference between the adsorbent and slurry densities. An idea of the minimum settling rates can be obtained from equation [22] by the assumption of particle sphericity, Newtonian rheology, and laminar settling conditions (Figures 6 and 7).

If the settling rates of a good proportion of the adsorbent particles are significantly greater than 2.5 mm/s, the use of an appropriate scale-up equation is recommended. Whether the particles have a tendency to settle or float has, at least in principle, a bearing on which equation to use. Normal practice for systems of particles that tend to settle is to scale-up on power intensity by use of equation [27], although some workers have found this approach to be rather sensitive to minor differences in the geometry of the system. In systems where the particles tend to float, little guidance is given in the literature. The work reported here indicates that the relevant scale-up equations are very similar to those used when particles tend to settle. Again, the scale-up equation based on a power variable is prone to error, and scale-up based on agitator speed is recommended, i.e. equation [26].

For the scaling up of film-transfer coefficients, a scale-up procedure is required for E_{JD} , the specific power needed to just disperse the adsorbent phase. Equation [27] is the appropriate vehicle but, as already mentioned, caution should be exercised because of its sensitivity to minor deviations in the geometry of the system. In addition, the scale-up exponent (usually -0.25) may be closer to -0.4 in systems involving particles of adsorbent that tend to float.

The primary concern in the first two papers in this series (this and reference 1) has been to provide an understanding of those aspects of contactor design which will most affect the rates of gold adsorption. The next paper will examine how these aspects can impact on the performance of operating plants.

Nomenclature

A	Cross-sectional area of the vessel
A_i	Area swept by the impeller
A_w	Area adjacent to the vessel walls, where the fluid flow is upwards
a	Parameter (equation [6])
B	Impeller clearance
C	Solids concentration (by mass)

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28. VON ESSEN, J.A. Lecture notes, Philadelphia Mixers Division, Philadelphia Gear Corporation, 1980.	C_D	Drag coefficient (or friction factor)	t	Torque applied to maintain the speed of the agitator
29. JOHNSTONE, R.E., and THRING, M.W. <i>Pilot plants, models and scale-up methods</i> . New York, McGraw-Hill, 1957.	C_V	Volumetric concentration of solids	tI	Scale-up exponent
30. BAGULEY, W. Carbon school lecture notes. Randburg, Mintek, 14-18th October, 1985.	E	Specific power or power per unit mass	V	Volume of fluid being mixed
31. COETZEE, M.C., and CLOETE, F.L.D. <i>J. S. Afr. Inst. Min. Metall.</i> , vol. 89, no. 4. 1989. pp. 99-109.	D	Impeller diameter	V_{eq}	Equivalent volume (equation [4])
32. AFEWU, K.I. Scale-up in carbon-in-pulp adsorption systems. Johannesburg (South Africa), University of the Witwatersrand, M.Sc. thesis.	dI	Scale-up exponent	u	Fluid velocity
33. NAGATA, S. <i>Mixing—principles and applications</i> . Kondansha Ltd, Japan, and Halstead Press USA, 1975. pp. 12-20.	d_p	Particle nominal diameter	u_{bot}	Fluid velocity across the bottom of a vessel
34. CHUDACHEK, M.W. <i>Chem. Engn Sci.</i> , vol. 40, no. 3. 1985. pp. 385-392.	g	Gravitational constant	u_{crit}	Critical velocity for suspension
35. BATES, R.L., FONDY, P.L. and CORFSTEIN, R.R. <i>Ind. Eng. Proc. Des. Dev.</i> , vol. 2. 1963. p. 310.	H	Depth of the fluid	u_{free}	Settling rate at low solids concentration
36. REWATKAR, V.B., RAGHAVA RAO, K.S.M.S., and JOSHI, J.B. <i>Chem. Eng. Comm.</i> , vol. 88. 1990. pp. 69-90.	JD	Subscript referring to the just-dispersed point	u_s	Apparent superficial velocity
	K	Proportionality constant (equation [13])	u_{settle}	Particle settling velocity
	N	Agitator speed	u_{slurry}	Settling rate at high solids concentration
	n	Scale-up exponent	u_{vert}	Fluid velocity past the impeller
	n_b	Number of blades on the impeller	X	Generic mixing parameter
	P	Power applied to maintain agitator speed N	Z	Baldi number (equation [8])
	P_f	Pumping number (or flow number)	$\Delta\rho$	Density difference between suspended solids and suspending fluid
	P_o	Power number	θ	Blade angle
	P_V	Power intensity or power per unit volume	μ	Viscosity of the fluid
	p	Scale-up exponent	ν	Kinematic viscosity
	Q	Pumping capacity of the impeller	ρ	Density of the solid-fluid system being mixed
	Q_V	Pumping capacity per unit volume	ρ_s	Density of solids
	r	Parameter	ρ_w	Density of water
	R	Scale-up ratio	τ	Torque intensity or torque per unit volume
	Re	Reynolds number of the impeller	τ_{eq}	Equivalent torque (equation [4])
	Re^*	Reynolds number for suspension	Φ	Speed of impeller tip or peripheral velocity
	Re_p	Reynolds number of the particle		
	S	Constant in the Zwietering equation		
	T	Vessel diameter		

Sasol Coal develops safer ventilation*

Sasol Coal, in conjunction with the CSIR's air-research department Aerotek, has developed a safe, efficient, and economical jet fan for the ventilation of continuous-miner headings.

Tests with a prototype unit proved that the fan conforms to all requirements. Sascoal Engineering has now begun manufacturing the fans commercially.

Mr Conrad Kahts, a senior project engineer in Sasol Coal's Project and Research Services Department and a member of the development team, points out that the elimination of ducting for incoming air in a heading is a major advantage of the new system. Ducting for incoming air in conventional ventilation systems is frequently damaged, leading to ineffective ventilation. This gave rise to a decision by Sasol Coal in 1989 to develop a new ventilation method.

The greatest advantage of the jet fan is the turbulence it generates to dilute gases and reduce the risk of a build-up of methane in a heading. The system provides for one jet fan per heading.

The jet fan is light, compact, and manoeuvrable. It can be re-located by two people, and can be mounted out of the way of moving equipment without risk of damage and without obstructing shuttle cars and other mining vehicles.

The jet fan does not require ducting since it directs a powerful jet of air at a speed of 40 to 60 metres per second onto the face from a distance of up to 35 metres.

The jet fan is being manufactured in three sizes: 4, 5.5, and 7.5 kilowatts, the volume of air delivered varying from 1.8 to 3.0 cubic metres per second.

A further feature of the newly-designed jet fan, according to Mr Kahts, is that it achieves a high level of entrainment—air in the vicinity of the fan is drawn into the air stream that it creates. Tests have established that the entrainment is 1.0 to 1.5 times the volume of air produced by the fan itself. 'This is a very high level of optimization, with between 3.6 and 7.5 cubic metres per second being delivered to the face', he added. ♦

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