The modelling and simulation of processes for the absorption of gold by activated charcoal

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SYNOPSIS
The modelling, simulation, and design of processes for the recovery of gold from solution by absorption on carbon require the combination of a suitable rate expression with a description of the flow configuration of the system. By the use of a suitable rate expression from the literature, methods are presented for the design of multi-stage countercurrent absorption systems corresponding to the commonly used carbon-in-pulp and carbon-in-column contacting systems. The sensitivity of the recovery to various design parameters is illustrated. These results have important implications in the choice and design of equipment.

SAMEVATTING
Die modellering, simulering en ontwerp van prosesse vir die herwinning van goud uit 'n oplossing deur absorpsie op koolstof vereis die kombinering van 'n geskikte tempo-uitdrukking met 'n beskrywing van die vloeikonfigurasie van die stelsel. Daar word met gebruik van 'n geskikte tempo-uitdrukking uit die literatuur, metodes voorgestel vir die ontwerp van veeltrap-teenstroomabsorpsiestelsels wat ooreenstem met die koolstof-in-pulp- en koolstof-in-kolomkontakstelsels wat algemeen gebruik word. Die gevoeligheid van die herwinning van verskillende ontwerpparameters word geïllustreer. Hierdie resultate het belangrike implikasies by die keuse en ontwerp van toerusting.

Introduction

The ability of activated carbon to absorb gold from solutions in which it is present in very low concentrations while itself loading to a fairly high concentration has made it an attractive material for the concentration of dilute gold solutions. This ability, together with the possibilities of gold recovery either by stripping or by ignition of the carbon, makes it an ideal material for use in gold recovery. In addition, the separation of the loaded carbon from the gold pulp makes possible the leaching and absorption of gold in a single stage, leading to the so-called carbon-in-pulp processes. As discussed recently[1], plants based on this principle are now being built fairly extensively. The time is therefore ripe for an examination of the design of such processes, particularly the gold-absorption units. The designer needs a rate expression for the uptake of gold on carbon as a function of the local conditions such as the gold concentration and the temperature in the unit.

Choice of Rate Expression

The rate expression chosen for the study described here was that suggested by Dixon[2] and used by Menne[1]. The latter has fitted a wide range of experimental results to this model, and has found it to give a good fit for both batch and continuous data. It is a two-parameter model and, when one of parameters (y^*) is large, it takes the same form as that used by Nicol, Fleming, and Cromberge[3] to fit their data. It can thus be stated that the model is a two-parameter one to which a wide range of data can be fitted with different types of carbon in both batch and continuous contacting equipment.

Empirical models, such as those of Fleming[4], in which time appears as an independent variable, are entirely unsuitable for the design of a continuous unit from batch data and will not be discussed further.

The rate expression used has the following form:

$$r = k(x(y^* - y) - Ky), \quad \ldots\ldots\ldots\ldots\ldots\ldots (1a)$$

which is equivalent to

$$r = k_{r}x(y^* - y) - k_{2}y, \quad \ldots\ldots\ldots\ldots\ldots\ldots (1b)$$

where r = rate of gold loading
k, k^{l}, k_{2} = rate constants
x = gold concentration in liquid phase
y = gold concentration on activated carbon
y^* = parameter with same units as y
K = equilibrium constant for gold loading.

This model could arise from a mechanism of the ion-exchange type such as

$$[Au]^{s} + [RH]^{c} \rightleftharpoons [H]^{s} + [RAu]^{c}. \quad \ldots\ldots\ldots\ldots (2)$$
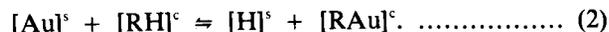
In this scheme, there is no implication about the form of the gold species in either phase, but the model emphasizes the fact that there is a site into which the material must be absorbed. It is also possible that, as with a Langmuir type of absorption, the sites are filled only with gold species, that is without anything being displaced. In any event, such a mechanism is consistent with the form of the rate expression presented.

If this mechanism is a good description of the process, then one could use it to modify the rate expression to cover other situations. For instance, when gold is stripped off the carbon, a good description of the process may include in the second term of equation (1) the concentration of the species that replaces the gold on the solid. This

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will not usually be necessary during the loading process since the high liquid-to-solid ratio in such processes means that the concentration of this species will be effectively a constant and will be incorporated in K.

The parameter y^* in the mechanism in principle represents the capacity of the carbon for gold, but in practice the values found from the fitting of the rate data are lower than the measured capacities. This could be because all the sites may not be equally available in a reasonable time scale and the parameter may represent the effective capacity.

A rate expression that has such a capacity term has the advantage that, if there are other materials that compete with the gold for the absorption sites, one can readily take their effect into account by writing an equivalent rate expression for them while allowing for all species in the mass balance in the absorbed material. Thus, for instance, the influence of copper complexes competing with gold for sites, as discussed by Fleming[4], could be allowed for.

Fleming has also mentioned the influence of a mass-transfer limitation on the rate of loading. There is evidence from his work for limitation both by film diffusion and particle diffusion. This is probably best handled by the use of an effectiveness factor[5]. As the rate expression is linear with respect to the concentration of the liquid phase, there will be a single value of the effectiveness factor for a given particle size of carbon and mixing conditions. Thus, the effectiveness factor will be a correction factor for the rate expression that either can be estimated separately or will be already incorporated if the measurement of the kinetic constants is done under the same conditions.

Using this rate expression as a basis, we shall now proceed to discuss the design of gold-absorption units. If equation (1) is not an entirely adequate description of the rate, the methods outlined in this paper can still be applied with some small modifications using other rate expressions.

The values of the constants used for the rate expression and for other parameters are given in Table I. They are reasonable values for these processes and are consistent with those reported by Menne[1].

Previous Modelling and Design Methods

Fleming[4] used his empirical rate expression to model pilot-plant performance, but, as his rate includes time as a variable, he had to assume that the mean residence time in his vessel has the same meaning as time in a batch system. This assumption is valid only under a very limited set of circumstances. Menne[1] plotted his results on a McCabe–Thiele diagram, and it is clear from his graph that there is not, indeed, a very close approach to equilibrium. When stage efficiencies become too low, all the effective error lies in the prediction of these efficiencies, and the system must be designed on a rate basis. As pointed out by Fleming[4], with the slow approach to equilibrium that occurs during gold loading, the McCabe–Thiele method is not a satisfactory design procedure.

Modelling of Existing Processes

It is convenient for us to look at the various types of unit in use, and to discuss the modelling of such systems. The two types most commonly used are a continuous countercurrent mixed-stage system called carbon-in-pulp (CIP), and a method known as carbon-in-column (CIC). The latter system is based on fluidized stages in which there is periodically a short reverse-flow time that serves effectively to move the solid countercurrent to the liquid. The models described here can be used in the calculation of the 'steady-state' conditions for the process units. While continuous CIP is a true steady-state operation, the other process, in which the solid is moved periodically, is never at true steady state. However, it does reach cyclic steady states, in which each cycle is the same as the previous one.

In all these transient situations, we make the assumption that the residence time of the liquid in any stage is much less than that of the solid. This is obviously true in that we are working with a dilute feed solution that is being used to load the solid to a high concentration. It follows that the rates at which concentration changes occur in the liquid phase are very much higher than in the solid phase. The concentration of solids can therefore be assumed to remain constant during the passage of a typical element of liquid through a stage of the system. This assumption of pseudo-steady-state greatly simplifies the subsequent analysis.

Model for CIC Process

Single Stage

The single stage in the CIC tower consists of a fluidized bed of solid carbon particles through which the liquid flows. It is known[6] that a reasonable model for such a bed is as follows. The solid fluidized particles are assumed to be well mixed; that is, all the particles in the bed are assumed to visit all parts of the bed equally. If all the solid particles enter and leave the bed together, it follows that they have all been exposed to the same conditions, and hence at any time they will all have absorbed the same amount of solute. The liquid is assumed to pass through the bed in plug flow; that is, each portion of liquid passes straight through the bed in the same time, undergoing the same concentration-time history while doing so.

The differential mass balance on the liquid is thus

$$Q \frac{dx}{dz} = -\frac{M}{H} r, \quad \dots\dots\dots\dots\dots\dots\dots\dots (3a)$$

where Q = liquid flow (t/h)
  x = liquid concentration (g/t)
  M = amount of solid in the stage (t)
  H = height of the stage (m)
  z = distance coordinate up the stage (m)
  r = rate of absorption of solute into unit mass of carbon (g/t per hour).

Substituting the rate expression from equation (1), we thus obtain

$$Q \frac{dx}{dz} = -\frac{M}{H} \{k_1 x(y^* - y) - k_2 y\}. \quad \dots\dots\dots\dots (3b)$$

As discussed above, the fact that the residence time of the liquid in the bed is of the order 10^{-3} that of the solid enables us to neglect the change in solid concentration, y, during the passage of a particular portion of fluid through the bed.

Integration of (3b) regarding y as constant yields

$$\frac{k_1 x_{out}(y^* - y) - k_2 y}{k_1 x_{in}(y^* - y) - k_2 y} = \exp(-M k_1 (y^* - y)/Q), \quad (4)$$

where x_{in} and x_{out} are the concentrations of liquid entering and leaving the stage at any time. Equation (4) can be solved for the exit concentration from the bed:

$$x_{out} = \beta x_{in} + (1 - \beta) k_2 y / k_1 (y^* - y), \quad\quad (5)$$

where $\beta = \exp(- M k_1 (y^* - y)/Q$.

A mass balance over the whole bed, neglecting the hold-up in the liquid phase, gives

$$\frac{dy}{dt} = \frac{Q}{M}(x_{in} - x_{out})$$

$$= \frac{Q}{M}(1 - \beta) \{x_{in} - k_2 y/[k_1 (y^* - y)]\}. \quad\quad (6)$$

If $x_{in}(t)$ is known, this differential equation can be integrated numerically provided a suitable boundary condition is available for $y(t)$. The values of $x_{out}(t)$ can then be obtained from equation (5).

Calculation for Tower

Consider the tower shown in Fig. 1. The diagram shows three stages, but the extension to any other number is obvious. The superscripts o and f on the y's represent values at the beginning and end of each operating cycle. The solid arrows represent the continuous flow of liquid, and the dashed arrows represent the solid transfers that are carried out at the end of each cycle. The concentration of the liquid feed, x_0, is constant. If we take a value for y_1^0, the concentration of the solid phase that enters stage 1 at the beginning of a cycle, equations (5) and (6) can be integrated numerically to give the concentration on the solid, y_1^f, at the end of the cycle. We also obtain $x_1(t)$, the time-varying concentration of the stream flowing out of stage 1 into stage 2. Thus, in exactly the same manner, we can use equations (5) and (6) to solve for y_2 from a given y_2^0. Similar calculations are applied to every stage of the tower for one time cycle. (In practice, the differential equations for all the stages are integrated simultaneously to avoid the need for storage of the intermediate values of liquid concentration.)

After the calculations for one time cycle have been completed, the operation of moving the solid down the column is represented by putting

$$y_1^o = y_2^f$$

$$y_2^o = y_3^f \text{ etc.,} \quad\quad\quad\quad\quad\quad\quad\quad\quad\quad\quad\quad (7)$$

and y_3^o = the concentration of fresh feed = 0 (or any other fixed value).

The calculation of the behaviour during the next time cycle is then carried out exactly as before. This procedure is repeated until the similarity of the values obtained during two successive cycles indicates that the values representing the steady-state operation have been obtained.
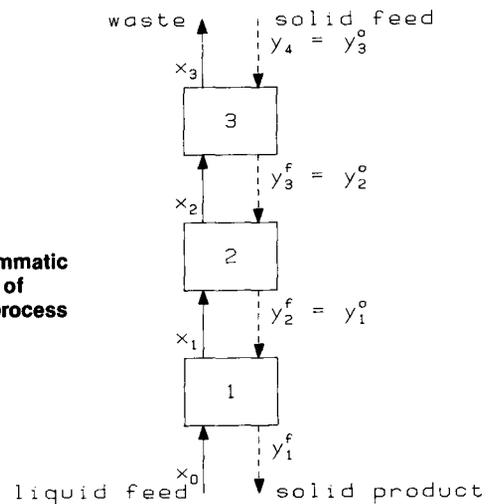

Fig. 1—Diagrammatic representation of flows in a CIC process

Bypassing

The model for the single stage may be improved if allowance is made for the bypassing of the liquid stream—that is by the assumption that some of the liquid flows through the fluidized bed without making contact with the solid. This is known to occur to a considerable extent in fluidized beds[6] owing to spouting and 'bubble' formation. The stage can then be represented as in Fig. 2. The reactor section still obeys equations (5) and (6), except that the flow through the reactor will only be $(1 - b)Q$. The exit concentration, x_{out}, is given by (5), but the feed concentration to the next stage is given by

$$x_i = b x_{i-1} + (1 - b) x_{out}. \quad\quad\quad\quad\quad\quad (8)$$

Allowance for Partial Movement of Solid

Suppose now that not all of the solid in a stage is moved to the next stage at the end of each time interval. Thus, the concentration of the solid in a stage at any time is represented by a distribution function. Thus, let $m(y)dy$ be the fraction of particles having a concentration between y and $y + dy$.

The equivalent of equation (3a) is now

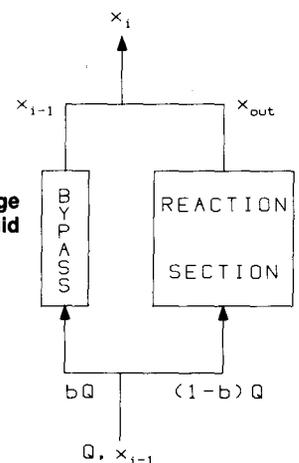$$Q\frac{dx}{dz} = -\frac{M}{H}\int_0^\infty m(y)\, r(y,x)\, dy, \quad\quad\quad (9)$$

Fig. 2—Flows through a stage allowing for the bypassing of liquid

where the integral term represents the total effective rate of absorption by the solid allowing for the different rate on solid of each concentration. By substitution of the form of the rate equation,

$$Q\frac{dx}{dz} = -\frac{M}{H}\int_0^\infty m(y)\,\{k_1 x(y^* - y) - k_2 y\}\,dy$$

$$= -\frac{M}{H}k_1 x y^* \int_0^\infty m(y)dy\ -$$

$$(k_1 x + k_2)\int_0^\infty y m(y)dy. \qquad (10)$$

The first integral term is equal to unity; the second gives the average solid concentration $\bar{y}$. Thus,

$$Q\frac{dx}{dz} = -\frac{M}{H}\{k_1 x(y^* - \bar{y})\}. \qquad (11)$$

the rest of the derivation for the single stage follows, with y replaced by its average value $\bar{y}$. Thus, the only change necessary in the calculation is the trivial one of replacing equations (7) representing the mixing process at the end of a cycle, which now become

$$y_i^o = F y_{i+1}^f + (1 - F)y_i^f, \qquad (12)$$

where F is the fraction of solid moved down at the end of each cycle time. Note that equation (12) reduces the equation (7) when $F = 1$.

Model for CIP Process
Fully-mixed-flow Stage
Consider the stage (tank) shown in Fig. 3. Both the flows of solid and liquid are continuous, and the stage is well mixed, i.e. flow as in a continuous stirred tank reactor (CSTR) in both phases. Thus, the concentration of the liquid and the concentration on the solid are uniform and equal to their exit values throughout the stage. Because of the CSTR flow, there is a spread of residence times on the solid particles, which therefore have a distribution of concentrations. Let $m_i(y)\,dy$ be the mass fraction of material in (and leaving) the i th tank having a concentration between y and $y + dy$. A balance on material of this concentration in the i th tank gives
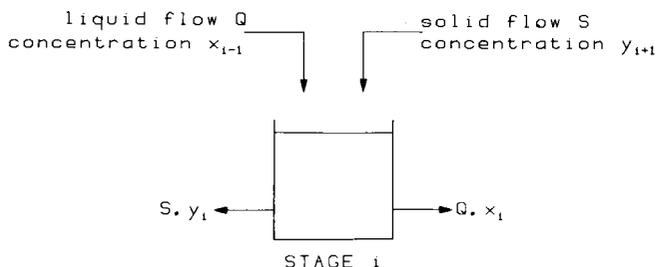

liquid flow Q
concentration x_{i-1} solid flow S
                                 concentration y_{i+1}

S. y_i Q. x_i

STAGE i

Fig. 3—One stage of the CIP process

$$Sm_{i+1}(y)\,dy = Sm_i(y)\,dy + Mm_i(y)\,r(x,y) -$$
$$Mm_i(y - dy)\,r(x, y - dy), \qquad (13)$$

where S is the flowrate of the solid and M is the holdup of solids in the tank. Thus,

$$S[m_{i+1}(y) - m_i(y)] = M\frac{dm_i(y)\,r(x,y)}{dy}. \qquad (14)$$

Multiplying across by y and integrating between the limits zero and infinity gives

$$S\int_0^\infty [m_{i+1}(y) - m_i(y)]y\,dy = M\int_0^\infty y\frac{dm_i(y)\,r(x,y)}{dy}dy,$$

i.e.

$$S(\bar{y}_{i+1} + \bar{y}) = M[ym_i(y)\,r(x,y)]\Big|_0^\infty -$$
$$M\int_0^\infty m_i(y)r(x,y)dy. \qquad (15)$$

The term in square brackets is zero at both limits. The integral term becomes equal to the rate at the average concentration in the tank (since $r(x,y)$ is linear in y). Thus, putting

$M/S = \bar{t}$, the average residence time,

we obtain

$$\bar{y}_i = \frac{\bar{y}_{i+1} + k_1 x_i y^* \bar{t}}{k_1 x_i \bar{t} + k_2 \bar{t} + 1}. \qquad (16)$$

An overall balance on the stage yields the relationship between the liquid concentrations:

$$S(\bar{y}_i - \bar{y}_{i+1}) = Q(x_{i-1} - x_i). \qquad (17)$$

An alternative derivation of (16) is outlined in the Addendum. That derivation does not depend directly on the linearity of the rate expression in terms of the solid concentration, y, and so could be used with other non-linear rate expressions.

Calculation for Process
Consider a set of N such stages. If we know the concentration of the solid feed, y_{N+1}, and the concentration of the liquid product x_N, that is the solid stream into and the liquid stream out of the Nth tank, straight-forward substitution into equations (16) and (17) yields the corresponding values for the $(N-1)$th tank, that is y_N and x_{N-1}.

Repetition of this procedure the requisite number of times yields the values for the solid product (y_1) and the liquid feed (x_o). Determination of the concentrations of the two products from given feed conditions is simply a matter of using some slightly sophisticated form of guesswork (such as *regula falsi*) to determine the value of x_N that gives the known x_0.

Periodic Process

We can also consider the alternative mode of operation in which the solid flow is not continuous. Instead, a fraction of the solid in each tank is moved to the next tank after a given time interval. This process is similar to the CIC process except that we replace the plug flow assumption for the liquid flow by the CSTR model. We again use the pseudo-steady-state assumption, and neglect the change in solid concentration during the passage of a portion of liquid through the tank. Equating the right-hand sides of (15) and (17), we obtain

$$Q(x_{i-1} - x_i) = M \, r(x_i, \bar{y}_i), \quad \ldots\ldots\ldots\ldots\ldots (18)$$

from which we can obtain the exit concentration at any time:

$$x_i = \frac{(M/Q)k_2 \bar{y}_i + x_{i-1}}{1 + (M/Q)k_1(y^* - y_i)}. \quad \ldots\ldots\ldots\ldots (19)$$

The unsteady-state balance of the solid is then

$$\frac{d\bar{y}_i}{dt} = r(x_i, \bar{y}_i) = \frac{Q}{M}(x_{i-1} - x_i) . \quad \ldots\ldots\ldots\ldots (20)$$

The set of differential equations (20) can be integrated over a time cycle as described for the CIC process. Equation (12) again represents the result of the periodic movement of solids. The calculation is repeated until the cyclic steady-state values are obtained.

Results

Table I shows recoveries calculated for different numbers of stages by use of the models given above. Case 1 is the base case for the comparisons that follow, and the other cases have the same parameters except as noted.

Case 2 shows the recoveries in a continuous stirred-tank (CIP) process (as described above) in which the carbon holdup per stage and the flow of liquid and (effective) flow of solids are the same as in case 1. We see that the single-stage recovery is significantly lower in case 2, but that the effect decreases as more stages are used.

Cases 3 to 5 show the effect of the bypassing parameter, as discussed earlier. Consideration of these and the other cases shows that the bypassing is probably the single most important parameter in the design of this sort of process; it is probably more important that this should be modelled correctly than that the correct kinetics should be used for the absorption process.

Cases 6 to 8 demonstrate the reductions in recovery that result in a CIC process when there is incomplete transfer of solids from stage to stage at the end of each cycle. While the losses are not so large as to be unacceptable when the complete transfer of solids is difficult to accomplish, it is obvious that this is not a preferred mode of operation.

Cases 9 to 11 show the reduction in recovery to be expected as the result of a reduction in the stage height, and hence the holdup of solids in the stage. In practice, too small a stage height would result in even lower recoveries because of the increased bypassing that would

TABLE I
RECOVERY OF GOLD IN VARIOUS CIC AND CIP PROCESSES AS CALCULATED FROM THE MODELS

Conditions for case 1:

Liquid feed rate	100 t/h
Liquid feed concn	0,05 g/t
Carbon holdup per stage	0,59 t
Cycle time (forward-flow period)	24 h
Fraction of solid moved after each cycle	1,0
Ratio of liquid-to-solid flowrates	4000
Fraction bypassing	$b = 0$

Rate expression $r = 0,12 \, x \, (3600 - y) - 0,022 \, y$

Case	Parameter	Recovery in number of stages, %				
		1	2	3	4	5
1	See above	75,6	93,5	98,6	99,7	99,9
2	CIP as case 1	53,8	76,0	87,3	93,5	96,8
3	Bypass $b = 0,1$	71,5	90,6	97,1	99,2	99,8
4	0,2	66,5	86,9	94,9	98,1	99,3
5	0,3	60,7	82,3	91,9	96,3	98,4
6	Solid moved $F = 0,9$	74,6	92,8	98,3	99,6	99,9
7	0,75	73,0	91,7	97,8	99,5	99,8
8	0,5	70,2	89,7	96,8	99,1	99,7
9	Height $H = 0,75$	69,7	89,3	96,5	99,0	99,7
10	0,5	59,6	80,7	90,7	95,7	98,1
11	0,25	40,9	61,4	73,8	82,0	87,6
12	As case 1 but $Q = 10$	85,8	98,4	99,87	99,99	99,99
13	As case 2 but $Q = 10$	71,3	90,5	97,2	99,3	99,8

occur in such a bed.

Cases 12 and 13 are the same as cases 1 and 2, except that the flow of liquid has been drastically reduced. The holdup of solids remains unchanged. The very long contact time between the solid and the liquid phases means that the two phases effectively come to equilibrium. The continuous CIP process of case 13 corresponds to a true equilibrium stage, as can be seen from the McCabe-Thiele diagram in Fig. 4, which also shows case 2 for comparison. While the McCabe-Thiele is a good method of design in case 13, the running of the process under these conditions is not likely to be economic since it is desirable that the flowrate of the liquid should be as high as possible (or equivalently, that a given flow should be treated in as small a system as possible). In case 2, which is more realistic, it can be seen that we would need to assume a stage efficiency of about 50 per cent in order to get a reasonable design. As there is no real basis for the estimation of such low efficiencies, the use of a rate model is much more reasonable. Case 12 is a CIC process in which the liquid leaving a stage is always in equilibrium with the concentration of the solid in the stage at that time. In this case, the concept of equilibrium stage (McCabe-Thiele) is not an adequate design technique, as can be seen by a comparison of the recoveries for low numbers of stages between cases 12 and 13. Even under these 'equilibrium' conditions, the different assumptions about the pattern of liquid flow and liquid residence-time distribution result in different recoveries in the two cases. This also implies that, if best use is to be made of the feed streams, it is important that equipment with appropriate flow patterns should be designed. If, however, there is a good practical reason for the use of the equivalent of mixed tanks, then one would have to accept lower recoveries or compensate by using more or larger stages.
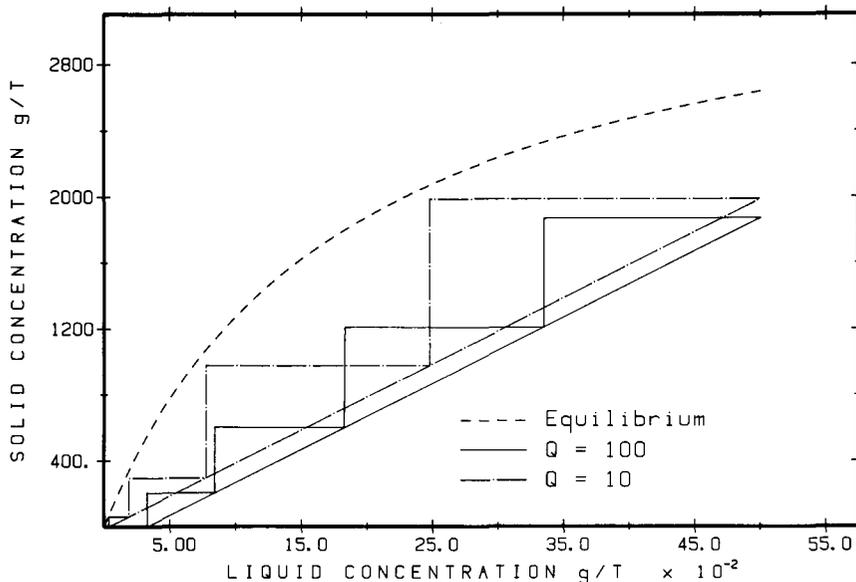
Fig. 4—McCabe–Thiele diagrams for cases 2 and 13

Conclusions

The discussion and results presented in this paper indicate that it is important that the process should be modelled accurately if the results are to enable one to design and cost the required equipment adequately. This applies both to the modelling associated with the rate equation and to that associated with the flow characteristics of the apparatus. Even though one may be using fairly simple and idealized models, the difference between the various assumptions can lead to significant differences in the size of the equipment for a given duty.

The converse of this argument is also true: it is important that the equipment with the best characteristics should be chosen. In this case, countercurrent plug flow would obviously be the best from a size point of view, but practical difficulties militate against the building of such equipment. Hence, one might choose the CIC process (as long as the stages are designed to minimize bypassing), rather than the CIP process. Once again, practical implications might indicate that the CIP process should be adopted. At least, using the design equations given here, one can quantitatively evaluate the loss in process efficiency relative to the gain in equipment simplicity. .

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References

1. MENNE, D. Predicting and assessing carbon-in-pulp circuit performance. CIM XIV International Mineral Processing Conference, Toronto, 1982.
2. DIXON, S., CHO, E.H., and PITT, C.H. The interaction between gold cyanide, silver cyanide and high surface area charcoal. *AIChE Symposium Series,* vol. 74, no. 173. 1978.
3. NICOL, M.J., FLEMING, C.A., and CROMBERGE, G. The absorption of gold cyanide onto activated carbon. I. The kinetics of absorption from pulps. *J. S. Afr. Inst. Min. Metall.,* vol 84, no. 2. Feb. 1984. pp. 50–54.
4. FLEMING, C.A. Recent developments in carbon-in-pulp technology in South Africa. Osseo-Asare, K.R., and Miller, J.D. (eds.). Warrendale (U.S.A.), Metallurgical Society of AIME, 1982.

5. SATTERFIELD, C.N., and SHERWOOD, T.K. *The role of diffusion in catalysis.* Reading (U.S.A.), Addison-Wesley, 1963. Chap. 3.
6. GLASSER, D., KATZ, S., and SHINNAR, R. The measurement and interpretation of contact time distributions for catalytic reactor characteristics. *I&EC Fundamentals,* vol. 12. 1973. pp. 165–183.

Addendum: Alternative Derivation of Equations Describing the CIP Process

Consider the series of fully mixed stages (which behave as CSTRs) shown in Fig. 5. (Note that the numbering of the stages has been reversed for the sake of easier notation below.) Suppose that the concentration of the solid feed, y_0, is uniform. Various solid particles will spend varying amounts of time in tank 1 before leaving, and therefore the stream marked y_1 will consist of particles of various concentrations.
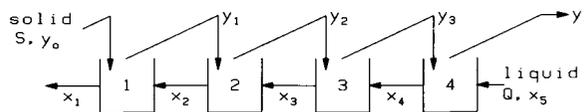

Fig. 5—Series of fully mixed stages making up a CIP process

Let us determine the concentration, y_1, of a particle of initial concentration y_0 that spends time t immersed in liquid of (constant) concentration x_1. Under these conditions,

$$\frac{dy}{dt} = - k_f x(y^* - y) - k_2 y, \quad\quad\quad (A1)$$

with the initial condition $y(t = 0) = y_0$.
The solution to this differential equation is

$$y_1(x,t) = A(x,t) + B(x,t)/y_0 , \quad\quad\quad (A2)$$

where $A(x,t) = \dfrac{k_f x y^*}{k_f x + k_2} (1 - B(x,t))$

and $B(x,t) = \exp(-(k_f x + k_2)t)$.

The average value of the concentration on the solid leaving the first tank will be given by

$$\bar{y}_1 = \int_0^\infty y_1(x,t_1) E(t_1) \, dt_1, \qquad \qquad \text{(A3)}$$

where $E(t)$ is the residence-time distribution in the tank, i.e. $E(t)dt$ is the fraction of the solids flow spending a time between t and $(t + dt)$ in the tank. For CSTR behaviour,

$$E(t) = \frac{1}{T} e^{-t/T}, \qquad \qquad \text{(A4)}$$

where $T = M/S$ is the mean residence time. Thus,

$$\bar{y}_1 = \int_0^\infty (A(x,t) + B(x,t)y_0)e^{-t/T}(1/T) \, dt. \qquad \text{(A5)}$$

Now consider the exit stream from the second tank. A particle of solid that has spent time t_1 in tank 1 and time t_2 in tank 2 will have a concentration

$$y_2(t_1,t_2) = A(x_2,t_2) + B(x_2,t_2)y_1,$$

where $y_1 = A(x_1,t_1) + B(x_1,t_1)y_0$.

The average value of y_2 will be given by

$$\bar{y}_2 = \int_0^\infty \int_0^\infty y_2(t_1,t_2) \, E(t_1) \, E(t_2) \, dt_1 \, dt_2,$$

which, abbreviating the notation somewhat, can be expressed as

$$\bar{y}_2 = \iint (A_2 + A_1B_2 + B_1B_2y_0)E_1E_2 dt_2 t_1$$

$$= \int E_1 \int A_2E_2 + \int E_1A_1 \int B_2E_2 + y_0 \int B_1E_1 \int B_2E_2.$$

Now $\int_0^\infty E(t)dt = 1$. Thus,

$$\bar{y}_2 = \int A_2E_2 + \int B_2E_2 \left\{ y_0 \int B_1E_1 + \int E_1A_1 \right\}$$

$$= \int A_2E_2 + \int B_2E_2 \bar{y}_1$$

$$= \int (A_2 + B_2\bar{y}_1)E_2, \qquad \qquad \text{(A6)}$$

which is the same relationship obtained in (A5) between $\bar{y}_1$ and y_0. Either integral can be evaluated to yield the same result as that given in equation (16), viz

$$\bar{y}_i = \frac{(\bar{y}_{i-1} + k_1xy^*\bar{t})}{(1 + (k_1x_i + k_2)\bar{t})} \cdot \qquad \qquad \text{(A7)}$$

INFACON 86

In 1974, Johannesburg hosted the 1st International Ferro-alloys Congress (INFACON 74), which was organized jointly by Mintek (then the National Institute for Metallurgy), The South African Institute for Mining and Metallurgy, and the Ferro Alloy Producers' Association. Stemming from the success of the Congress, the idea of an ongoing series of congresses in various countries was proposed by the organizers, and an international committee was formed to ensure that the proposal was implemented. The second congress was held in Lausanne (Switzerland) in 1980, and the third in Tokyo in 1983. South Africa was very well represented at both meetings,

having the largest of the visiting delegations. The fourth congress will be held in Rio de Janeiro in 1986, and it is hoped that, once again, South Africa will be well represented.

The dates for INFACON 86 are 31st August to 3rd September, and the venue will be the Inter-Continental Hotel in Rio. The organizers, ABRAFE, have already mailed leaflets to people who are on their mailing list. Anyone else who is interested in receiving information should write to Mr Douglas Melhem Junior, Secretary General, ABRAFE, C.x. Postal 1031, CEP 01051 Sao Paulo, Brazil.

South African safety record

A delegation of 12 occupational safety and health experts from Canada and the U.S.A. recently completed a three-week visit to South Africa with high praise for the standard of safety in local industries. The delegation leader, Hugh Douglas, said that the visit had brought home the fact that South Africa is a world leader in activities designed to ensure that all work places are healthy and safe.

Douglas, who draws on 33 years experience as senior loss control co-ordinator and safety official with Exxon Corporation of America, said that the delegation had visited a broad spectrum of firms across the country, and had met industrialists, trades union officials, and government representatives.
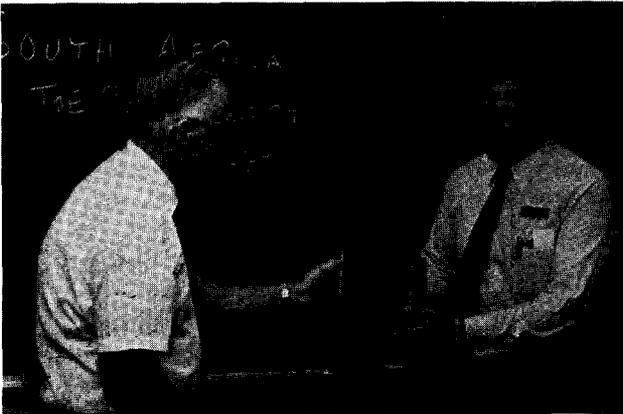
The leader of the safety delegation, Hugh Douglas, presents the General Manager of the National Occupational Safety Association (NOSA), Bunny Matthysen, with a plaque

He reported that delegation members had made the following observations based on this experience.

- The 5-star system to audit safety performance worldwide had been developed in South Africa.
- Every management team with whom they came into contact had demonstrated a deep commitment to occupational health and safety.
- The leaders of the labour movement had displayed a strong interest and a keen desire to be involved.
- The leading-edge thinking of industrial physicians, academics, and researchers in the field of health and safety was impressive.
- The decline in national injury-frequency rates and in Worker's Compensation payments was unique. In other Western World countries, the trends were up.
- The world records set by Foskor Mines and Mobil Oil, for working more than 20 million and 9 million manhours respectively without a disabling injury, were striking evidence that the South African approach is effective when fully applied at the work contact level. The visits to these and other award-winning sites had convinced the members of the delegation that the award winners merited every recognition that they had received.

The delegation to South Africa was sponsored by People to People International. Started in 1956 by President Dwight Eisenhower, it promotes international friendship through personal contact and is a non-profit, non-political organization dedicated to promoting effective and meaningful communication between members of the world community.