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Modelling and Control of Blackwater Treatment in Coal Preparation

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ABSTRACT

Blackwater streams emerging as effluents from coal preparation plants typically contain considerable coal values which are rejected with waste solids. Additionally, new and proposed environmental regulations severely limit the use of coal refuse ponds and encourage closed water circuitry. Optimization of flotation cells and thickening units to overcome these limitations has not been successful in most cases. Models are chosen here for simulating the two processes and examined from a theoretical and practical standpoint in application to coal preparation problems. The emphasis of this analysis is to provide a rational basis for unit process control through proper model selection and application. Strengths and weaknesses of model approaches and developmental needs are also reviewed.

INTRODUCTION

Recent developments in mining techniques, market forces and environmental preservation have focused need and opportunity for improved fine coal cleaning in both steam and metallurgical application. Wet cleaning of fine coal generates blackwater containing considerable coal values, a good portion of which may be recovered by properly designed and operated blackwater treatment/coal recovery systems.

Interest in both wet and dry fine coal recovery have increased along with attempts to utilize computer modelling (1,2,3) to carry out preliminary design of both coarse and fine coal cleaning circuits. Unfortunately, these cmputer models suffer deficiencies, particularly in their description of fine coal separations, i.e., froth flotation, refuse thickening and both coal and refuse dewatering. In order to optimize these processes in a meaningful way, adequate technological or production functions are needed. The intention here is to review modelling efforts for these processes within a systems framework and add to their capability for operation and design of blackwater treatment processes.

OVERVIEW OF COAL PREPARATION

Recent legislation including The Clean Air Act Ammendments of 1977, Resource Conservation and Recovery Act of 1976, Federal Water Pollution Control Act Ammendments of 1977, to name a few, have an impact on the coal industry (4). Expanded use of coal in the energy sector as well as increased metallurgical uses presents a challenge to remove mineral fractions (clays and pyrite), organically bound sulfur, trace metals and other undesirable elements from the carbonaceous portions. Metallurgical and steam coal generally demand different levels of treatment and final coal quality. However, fundamental cleaning requirements remain applicable to most coal users and include removal of ash contributing minerals and total sulfur. In metallurgical applications, Emerson (5) has shown that steel quality with respect to sulfur is most dependent on coal preparation i.e., total S, total ash, etc. Comparing physical coal cleaning (PCC), chemical coal cleaning (CCC) and flue gas desulfurization (FGD) for adequate removal of mineral impurities and sulfur shows that PCC will continue to function as a primary vehicle for removing extraneous material from coal (6).

Fine Coal Cleaning

Fine coal is generally defined as the minus 28 mesh fraction. One estimate suggests that the U.S. produced nearly 12 million tons on dry weight basis of slimes (a liquid slurry with very fine solids and mud-like appearance) out of 390 million raw tons processed in 1973 (7). As much as 60% of the solids in the blackwater produced may be carbonaceous and BTU losses approaching 40% have also been reported (8). Mechanized mining methods including increased use of shearing equipment may produce as much as 25% of the ROM in the 28 x 0 size (9).

Preparation designers and operators are now more eager to examine fine coal benefication for both steam and metallurgical coal for several added reasons: (1) The economic incentive for fine coal recovery now exists with delivered and washed coal ranging from $27.50 to $44.00/ton and a seven to ten percent yearly price increase (9). (2) Mechanical dewatering has become more widely practiced, particularly where large slurry ponds are not possible. (3) Large energies have already been expended to liberate the fine coal fraction and it would be desirable to recover some of this energy.

Fine coal cleaning includes hydrocyclones, froth flotation, Humphrey's spirals, concentrating tables, oil agglomeration and heavy medium separators. Those methods based on specific gravity difference such as tables, jigs, spirals, and cyclones are not likely to be extended to the finest coal sizes. Froth flotation and oil agglomeration appear most promising (9). Even so, froth flotation still suffers from inadequate ash and pyrite rejection, most likely due to the lack of sufficient research previously devoted to process optimization. Coal flotation does depend on many variables, as will be seen subsequently, and so is not so easily optimized. However, as mentioned earlier, economic incentive exists to tackle that problem now.

Thickening and Dewatering

Recovery of coal through a flotation step produces an underflow stream which must be thickened and dewatered by gravity or other applied forces. Environmental regulations generally limit the use of slurry ponds alone for blackwater treatment so mechanical thickeners are coupled into the cycle. Chemical flocculants are also used to aid solid-liquid separation. Mechanical dewatering beyond thickening has only been practiced at eastern U.S. plants where topography, etc. has limited the use of large slurry ponds (10). It should be noted that dewatering has been practiced for both the refuse (blackwater) and the product (coal) streams.

The term "blackwater" is used to describe the effluent from upstream processes which is normally discarded as tailings. One should note that blackwater streams can be of two types. One stream, for example is subject to recovery of the coal values using froth flotation. The rejected stream from flotation, when thickened, produces a final blackwater or refuse stream. The latter two terms are sometimes used interchangeably. Any stream from which fine coal can be economically beneficiated will be termed herein as a blackwater stream.

Coal dewatering technology has been reviewed in the recent edition of Coal Preparation (11). As in the case of flotation, thickening theory is not well understood and thickeners are generally designed for a limiting condition with several possible operation modes, the latter to prevent upset which may have catastrophic results.

How these processes are all integrated in a working plant flowsheet have been described in a recent paper by Zimmerman (12) where five levels of preparation complexity are compared (Table 1). Thus, fine coal cleaning is only part of the prep plant cost and optimization.

MODELLING

Computer codes exist for steady state description of coal preparation operations and include; CPSM4 (1), CPSM4 as modified (2), and COALWASH (3). The former are U.S. versions developed

Table 1
The Five Levels of Coal Preparation (after 12)

	I	II	III	IV	V
Kind of cleaning	Crushing only (crusher or rotary breaker)	Coarse coal cleaning only	Coarse coal cleaning and simple fine coal cleaning	Coarse and flue coal cleaning plus froth flotation closed circuit	Cleaning all sizes multiple stage, two products, fine crushing, closed circuit
Scope	Crushing to a given topsize. If rotary breaker used, some impurities are removed. Used for lower ash and sulphur run-of-mine coals.	Coarse coal, usually +10 mm, washed only, with fines bypassed and added raw to product. Partial cleaning only required to meet specifications.	Coarse and fine coal cleaning down to 0.5 mm with latter size recovered raw or discarded. No closed water circuit. Total wet washing of easy coal to meet required specifications.	Standard complete plant washing all sizes including −0.5 mm by froth flotation or other means. Closed water circuit. Complete washing required for difficult coals or market conditions.	Crushing coal to increase liberation using low relative density of separation, re-treating middlings. Froth flotation. Three product separation. Closed water circuit.
Range of recovery	95 to 100 percent	80 to 95 percent	60 to 80 percent	60 to 80 percent	25 to 40 percent primary 20 to 30 percent secondary

under U.S. Department of Energy contract and the latter a U.K. model sponsored by the U.K. National Coal Board. These models basically exhibit linear-input output responses and concentrate on the coarse coal processing circuitry. Where they are deficient is in the fine coal cleaning/recovery and solid-liquid separation phases. Their process definition is essentially material balance with arbitrary partitioning for solid and liquid streams. As an example, a percentage of carbonaceous material is assumed for coal recovery in flotation and a concentration of solids in the underflow and overflow for the thickener. In order to realize true predictive capabilities, a more process oriented algorithm is needed. The intent here is to examine development of useful algorithms for flotation and thickening. Before doing so, some review of the hierarchy of modelling is worthwhile.

Modelling Hierarchy

Modelling and optimization can be realized through design models and operation models which do not have to be mutually exclusive. Focus here is on operational models which are generally not provided by equipment manufacturers.

Optimization of operation should address three elements: (1) simulation modelling of the unit process, (2) monitoring of state variables and control actions, and (3) control of physical/mechanical aspects of the system permitting operational changes necessitated by changes in state variables. The first two are subjects here and the latter deals with availability of automotion equipment, based on plant specific objectives.

System simulation is divided into three levels of scope:

a) subsystem - focus on individual mechanisms within a unit process,

b) system - attempt to simulate an entire unit process, often combining subsystems elements and employed for optimizing unit processes,

c) metasystem - combines system models to simulate conjunction of processes, ideally simulating an entire plant.

Approaches to modelling can be theoretical (phenomenological) or empirical. Theoretical models described herein are generally not first principle models while empirical models do not account for process fundamentals or mechanisms explicitly. Most models combine empirical and theoretical elements.

Models can be either dynamic or steady state. The former using time as a variable are useful in control systems whereas the latter are independent of time due to the assumption that equilibrium has been achieved. The latter are used in flowsheet evaluation of processes where time change is gradual. These generally comprise the basis for current coal preparation models. A summary of these modelling developments is included in Figure 1.

MODEL SELECTION

Herein, model selection is based on satisfying a list of required as well as certain desirable attributes described in Table 2.

The required attributes are two-fold and imply the prerequisites that models: 1) provide insight into the effects of control actions and 2) are usable by plant operation engineers. Thus, experimental and empirical models are less favorable in this context. The second attribute is somewhat subjective since the accuracy of models are largely untested by the coal preparation industry. Given this situation, it may be more important to achieve repeatability with the same degree of error under various proven conditions than obtain absolute accuracy in prediction. Models having undergone testing and use in the mineral processing industry rate higher in this regard.

Desirable attributes are somewhat self explanatory. 1) Feed conditions do vary and thus dynamic models would be most useful. 2) Control actions at the operator's disposal should be reflected.

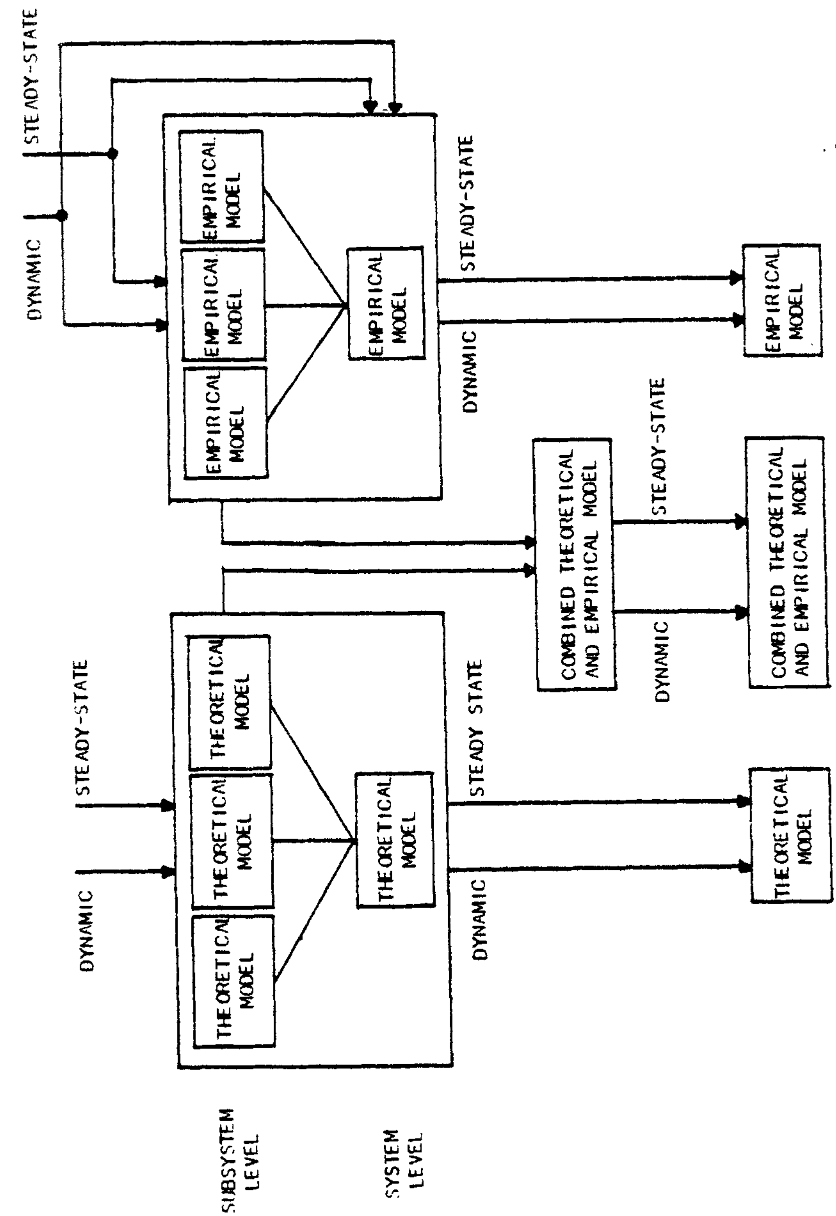

FIGURE 1 Block diagram showing the development of system level dynamic or steady-state models.

Table 2
Required and Desired Attributes of Models

REQUIRED ATTRIBUTES

1) In a general-user or readily adaptable form.
2) Accurately predictive of input/output response.

DESIRED ATTRIBUTES

1) Preferably a dynamic model to handle transients.
2) Employs sufficient level of input information capability.
3) Provides sufficient level of output information.
4) Compatible with current or near-future monitoring technology.
5) Mathematical or graphical form which lends itself to computer coding.
6) Flexibility in handling a variety of process designs.

in model input. 3) Output information must at least be supplied at the minimum to allow intelligent control by the operator. Sequential models can have decreasing but not increasing information requirements on input/output. Thus, if particle size data is required by a thickener model but not supplied by a flotation model upstream, incompatibility results. 4) If online monitoring equipment required for control is unavailable, limited model utility exists. 5) In this day of microprocessor control, models need to be in a mathematical or graphical form that can be computer coded. Physical models are impractical. 6) Sufficient flexibility to handle variation in process configuration may be more important in preliminary flowsheeting than in plant construction. Considering flotation and thickening combinations, the variables, such as arrangement of cells, feed and recycle streams and number of flotation stages may differ considerably. In metallurgical coal operation two stages of flotation is quite common, particularly when it is desirable to achieve a higher level of ash or pyrite rejection (13). Therefore, models of general use should at the very least be able to model the "simple" two stage process depicted in Figure 2 wherein the numbers associated with different streams correspond to the input for the flotation model, FLOTE, described subsequently.

Flotation Model Selection

Models able to simulate plant performance and in a general user format are relatively few and selection is based on required attributes (Table 2) in a straightforward manner. Five steady state and five dynamic simulators developed in the last ten years have been considered. Of the steady state simulators summarized by Herbst (14), only the model developed by King at the National Institute of Metallurgy, South Africa is available as a complete simulator package. Reliability and accuracy of prediction have been demonstrated in extensive testing in the metallurgical field. Dynamic simulators have been utilized in a more limited way and a general user form for large scale systems has not been reported (14). Thus dynamic modelling of flotation units does not satisfy the required attributes (Table 2) adopted in this study.

Thickening Model Selection

Options available for thickening model selection are in fact poorer than those of flotation and selection problems are of a different kind. The fundamental problem shared by all thickening models is a lack of understanding of the mechanisms of thickening. The majority of models are overly simplistic and also lack generality (15). An exception is the recent work of Lawler (16) which shows significant advance toward theoretical understanding. Unfortunately, Lawler's model is not in a general user form nor is it ready to be applied to real systems.

Most models are aimed at sizing cross sectional area of thickener units based on solids flux theory. Major differences between models are in the mathematical relations used to extrapolate from batch tests to continuous thickener design. The simple

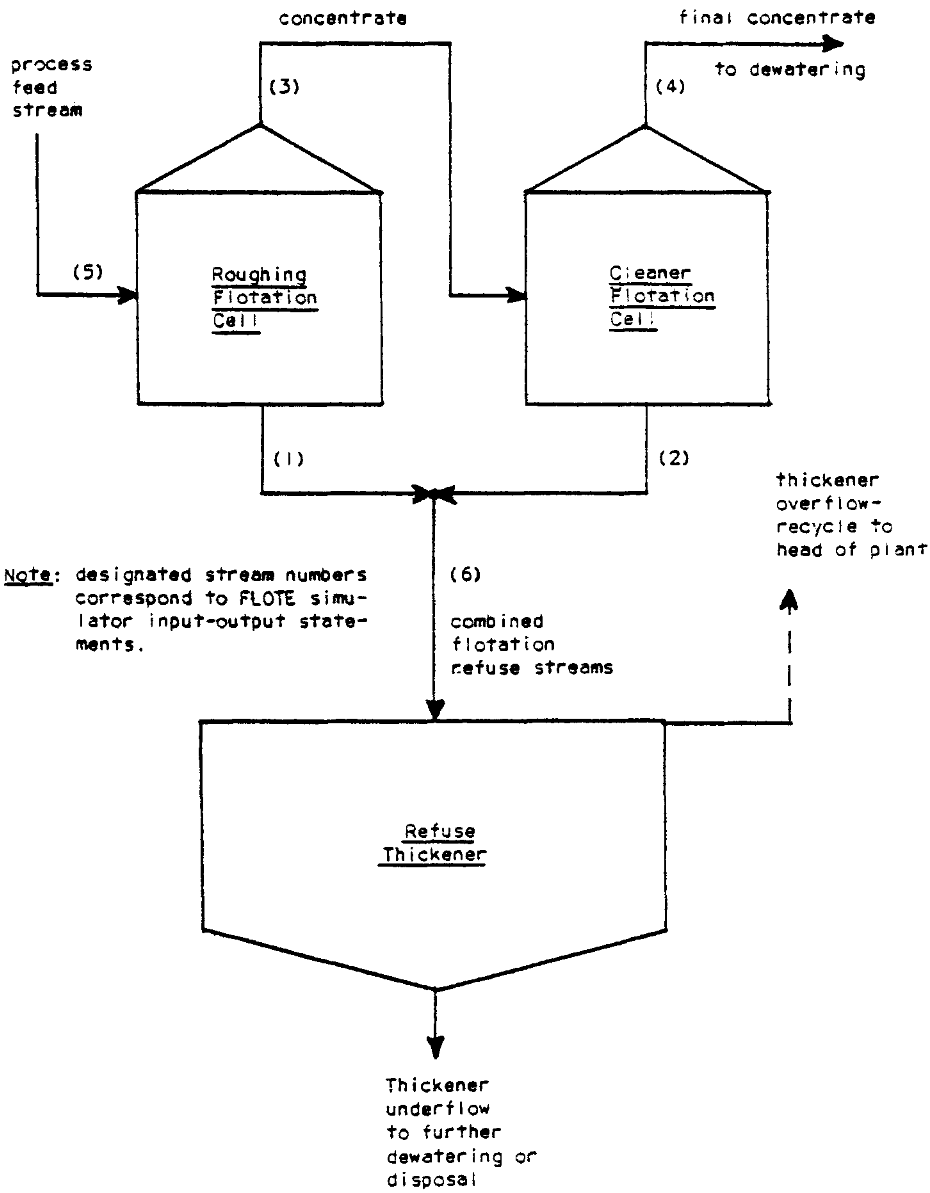

FIGURE 2 Process flow-sheet of plant configuration to be modeled in this study.

nature of the models permits classifying them in a general user category, but accuracy of the individual models is largely dependent on the particular slurry type (17). Therefore, we are constrained herein to select a model which is demonstrated reliable for thickener design for coal refuse (final blackwater) rather than a flexible model to handle any type of settling slurry. Furthermore, each model is already in a general user form and thus the accuracy criteria is also used for thickening model selection herein.

Coupled with unit design models should be thickener control strategies to improve performance and minimize cost. These are discussed subsequently. The upshot is that a construction suggested by Wilhelm and Naide (18) is chosen for design of a thickener. For coal slurries the model also satisfies some of the desirable attributes in Table 2.

FLOTATION MODEL

One can go into great lengths describing the theory behind the flotation and thickening models but space does not permit that here. Suffice it to say that in normal flotation, coal can be floated from a system of inhomogeneous suspended solids because of its hydrophobicity which may be enhanced by chemical addition. The gangue preferentially adheres to the water and remains in the pulp and is withdrawn with excess water from the lower part of the cell. Flotation is a kinetic process dependent on a number of factors with an overall rate constant depending on several mechanisms or subprocesses. Figure 3 shows bubble and particle rate described as a series of subprocesses, each with a certain probability. Relating these probabilities to various measured parameters of the system constitutes the major difference between models.

Basics of FLOTE

FLOTE, the computer coded program developed at NIM (19), and an accessory program for the estimation of kinetic parameters

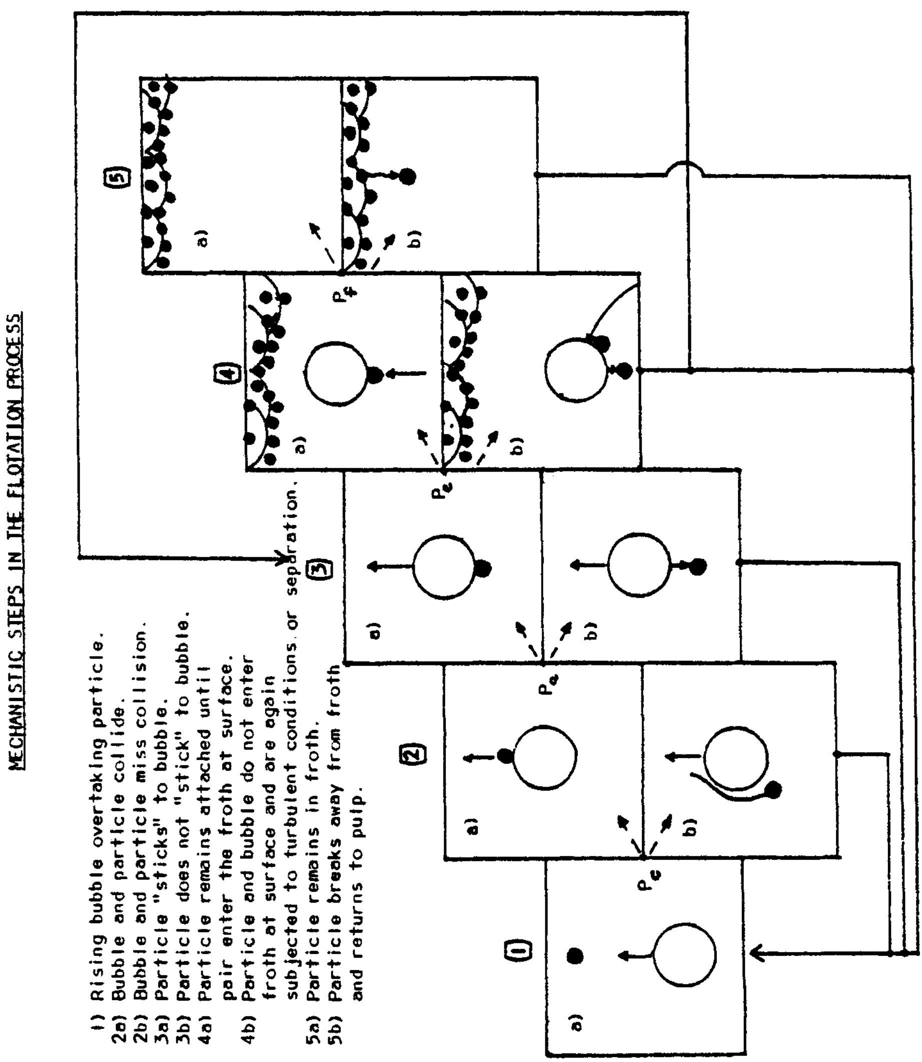

FIGURE 3 The mechanistic steps in flotation.

(20), provide a general purpose simulator written in FORTRAN and available in tape form. FLOTE can be applied to flotation plants of any configuration except those which use regrinders in the process. The NIM program is based on the distributed constant flotation model described in a paper by King (21), and has been subjected to revision over the years.

The formulation of the model as described by King (21) involves certain assumptions: flotation cell pulp undergoes perfect mixing through which rises a cloud of bubbles. Hydrophobic particles then attach to these rising bubbles and are transported upward through the pulp into the froth. A fraction, $1 - \gamma$, of the solids that enter into the froth returns to the pulp. The rate of flotation depends primarily on:

(1) the surface activity, k, which is assumed to be independent of particle diameter,

(2) the mineral composition, G, which is given by a vector of values (g_1, g_2, $g_3 \ldots g_5$) representing the fraction of each mineral type in the particle (gangue excluded), and

(3) the particle size, D.

The surface activity, k, essentially represents a measure of the particles' ability to adhere to the bubble (hydrophobicity). The G-vector accounts for the non-liberated condition of solids typically expected reporting to the flotation cell and reconciles the differences in flotation properties resulting from different mineral proportions by attributing specific k values to each G-class.

Two conditional distributions: f(G/D), f(k/G) and one size distribution, f(D) need to be specified. The k-vector is usually represented by an average value and consequently, there remains one specific k-value for each G-class.

The aggregate of particles is assumed to be divided into a finite number of classes, each characterized by values for k, G, and D. $f_T(k,G,D)$ represents the total fraction of solids in each class. The model is based on the hypothesis that the flotation of each class of solids can be assumed to obey a first-order rate law which is independent of the concentration of particles in other classes (i.e., no interclass effects). The equation takes the form:

$$r(k,G,D) = \gamma k\emptyset(D)ASWf_T(k,G,D) \qquad (1)$$

where r(k,G,D) is the rate of recovery in the concentrate of particles belonging to the particular class specified by k, G, and D; γ is the froth transmission coefficient and indicates the fraction of solids that remain in the froth after entering the froth; k is the surface activity described earlier; $\emptyset(D)$ is a function that models the effect of particle size on flotation rate; A is the total bubble surface area per unit volume of pulp; S is the fraction of the bubble surface area that is not covered by particles and is thus available for flotation; W is the mass of solids in the cell; and $f_T(k,G,D)$ is defined above.

Working equations for a perfectly mixed cell can be developed using equation 1. A mass balance on any particular class of solids yields

$$M_i f_i(k,G,D) - M_t f_t(k,G,D) = M_f f_f(k,G,D) \qquad (2)$$

$$= \gamma k\emptyset(D)AS_{AV}Wf_T(k,G,D), \qquad (3)$$

where M is the total mass flowrate of solids and i, t, and f denote feed, tailing, and froth streams respectively.

Equation 3 turns out to be a non-linear integral equation, and an explicit solution for $f_T(k,G,D)$ is not available. A convergent iterative solution solves the equation and is implemented in the computer program (19).

Model Assumptions and Restriction

The most important assumption in the construction of the model is that the solid particles reporting to the flotation cell can be classified according to their individual flotation rates. Specification of a particle class is based on particle size, degree of liberation (defined as the relative proportions of minerals in each particle), and flotation properties as reflected by surface activity. The first two properties can be measured directly while the latter must be inferred from observations of the ore in a flotation cell subject to continuous or batch testing.

The plant water balance and holding times in the flotation stages are computed from a priori knowledge of the percent solids in the concentrate stream, aeration rate, bubble size, and other input. To allow for the presence of heavier minerals, the average specific gravity of the solids in the flotation feed is regarded as a variable. How this aspect of the model affects the validity of the model is discussed later.

Computer Input and Output

A summary of the data which must be provided as input to FLOTE is given in Table 3. Some of the parameters, such as bubble size and average bubble residence time, are independent of the state variable observations. Other parameters in the model will have frequency distributions which are characteristic of a sub-set of the state variables and the control actions. Thus, certain members of the parameter set will vary in accordance with changes in these variables (22).

The program algorithm as well as the output format are well suited for handling complicated plant configurations, although this is not of great importance in coal flotation plants since they are typically simple in design. Stream by stream analysis provides a detailed record of the following for each point in the system circuit:

(1) the recovery of each mineral,
(2) the grade of each mineral,
(3) the grade of each mineral in the various D-classes,
(4) the recovery of each mineral in the various D-classes,
(5) the particle-size distribution of the total solids,
(6) the recovery of gangue in the various D-classes,
(7) the recovery of total solids in the various D-classes,
(8) the solid and water flowrates.

Table 3
A summary of the data which must be provided as input to FLOTE (modified after, 19)

Plant Specifications:

1) Number of stages, stream routes and identification, location of nodes (a point where two or more streams coverage and leave as one).
2) Flotation cell (stage) volumes.

State Variables:

1) Number of k-classes, number of D-classes, number of G-classes, number of minerals.
2) Vector and name for mineral fractions in each G-class.
3) Value of D for each D-class.
4) Fractional distribution of particles among G-classes in each D-class.
5) Fractional distribution of particles among D-classes.
6) Specific gravity of solids (average).
7) Solid feed rate.
8) Percent solids in feed.

Model Parameters:

1) Value of k for each k-class.
2) Fractional distribution of particles among k-classes in each G-class.
3) Fraction of froth produced in each state that does not return to the pulp but passes over the froth lip (froth transmission coefficient, γ).
4) Δ in the $\emptyset(D)$ function - the largest particle size that will float in each G-class (a vector).
5) ϵ in the $\emptyset(D)$ function - the turbulence intensity parameter for each G-class (a vector).
6) Bubble size, d_b, and average bubble residence time, $\bar{\tau}$.

Control Actions:

1) Percent solids in the flotation cell.
2) Percent solids in the concentrate.
3) Aeration rate for each stage.

Adaptability for coal flotation

FLOTE is a versatile model as already discussed. How the model can be adapted to the particular problems associated with coal flotation encompasses the body of this section.

The major inputing variable enabling characterization of a coal slurry, is the G-class, indicating the minerals present in the system as well as their relative proportions. In the simple case of a slurry with coal and gangue as constituents (sulfur is assumed to be unimportant), the particles can be classified into one of five categories (G-classes) depending on the mineral composition of each particle. These types may include particles which are composed of (1) 100% coal, (2) 100% gangue, (3) 75% coal and 25% gangue, (4) 50% coal and 50% gangue, and (5) 25% coal and 75% gangue, then the following G-class notation can be used:

G-Class	Mineral Content	Mineral Fraction Vector
1	100% coal, 0% gangue	(1.0)
2	0% coal, 100% gangue	(0.0)
3	75% coal, 25% gangue	(.75)
4	50% coal, 50% gangue	(.50)
5	25% coal, 75% gangue	(.25)

Gangue values are not included in the vector since they are not considered to represent a "mineral" component.

A more comprehensive case might include descriptions of the oxidation characteristics of the individual coal particles as well as provide a mass balance for pyrite in the system. The following represents G-class notation for such a case:

G-Class	Mineral Content	Mineral Fraction Vector
1	100% unoxycoal	(1.0, 0.0, 0.0, 0.0)
2	100% modoxycoal	(0.0, 1.0, 0.0, 0.0)
3	100% oxycoal	(0.0, 0.0, 1.0, 0.0)
4	100% pyrite	(0.0, 0.0, 0.0, 1.0)
5	100% gangue	(0.0, 0.0, 0.0, 0.0)
6	75% unoxycoal, 10% pyrite, 15% Gangue	(.75, 0.0, 0.0, .10)
7	75% modoxycoal,10% pyrite, 15% Gangue	(0.0, .75, 0.0, .10)
8	75% oxycoal, 10% pyrite, 15% Gangue	(0.0, 0.0, .75, .10)

where unoxycoal is the unoxidized coal, modoxycoal is the moderately oxidized coal, and oxycoal represents the oxidized coal fraction. In either case, the oxidation characteristics of the coal may abruptly change due to intrinsic coal properties or mining or storage strategy.

THICKENING MODEL

Herein "thickening" corresponds to the range of particle concentration and interparticle cohesiveness which observes line settling behavior. Line settling may include both zone settling and compression depending on characteristics of the solids. These are often dealt with separately in thickener design. Zone settling properties of the solids generally dictate thickener cross-sectional area while compression properties often dictate depth requirements.

There are currently three physical testing methods available for the development of thickener size and performance specifications: continuous piloting, semi-continuous testing, and predictions from batch settling tests (17). Of these, batch settling tests are the most frequently used method. They are generally inexpensive requiring only ordinary laboratory equipment and small amounts of sample.

The majority of modeling efforts are aimed at: 1) simulating the behavior of continuous thickeners, and 2) modeling the settling of solids in a batch cell. Coe and Clevenger (23) directed their efforts towards the former and Kynch (24) the latter. A comprehensive overview of thickening models (17, 25) provides a description of recent efforts aimed at enlarging and modifying the approaches of Coe and Clevenger (23) and Kynch (24), but the scope of this work does not permit space for that discussion.

Wilhelm and Naide Design Model

a) continuous thickening theory

In a continuous thickener, the downward velocity of the solids with respect to the thickener wall is the result of two components which give rise to the equation of solids flux in a continuous thickener at any layer i:

$$G_i = C_i V_i + C_i U, \tag{4}$$

where G_i is the solids flux at any layer i, C_i is the local solids concentration at layer i, V_i is the settling velocity of solids due to gravity, and U is the velocity of the solids caused by the underflow withdrawal of solids. The product $C_i V_i$ is often called the "settling (batch) flux" while the product $C_i U$ is called the "transport (underflow) flux" where the bracketed terms are those of Wilhelm and Naide (18).

Figure 4A shows a typical flux curve at one underflow pumping rate. Note the location of the minimum in the flux curve. If the solids' settling velocity is a function only of the local solids concentration, then the thickener is designed by and must be operated at the limiting solids flux, G_L, for this particular underflow pumping rate, viz:

$$G_L = C_L V_L = C_L U \tag{5}$$

where G_L is the limiting solids flux, C_L is the corresponding limiting solids concentration and V_L is the limiting solids settling velocity. C_u, the underflow concentration, can be obtained from Figure 4A by drawing a tangent to the minimum in the flux curve and intersecting the G_L and the underflow flux line. C_u is read off of the underflow flux line.

Wilhelm and Naide have shown that the underflow withdrawal rate can be expressed in terms of concentrations and settling velocities at the limiting conditions by taking the derivative of Equation 4 at the minimum in Figure 4A giving:

$$U = \left(\frac{-d(C_i V_i)}{dC_i} \right)_{i = L} \tag{6}$$

Equation 6 cannot be differentiated until the relationship between settling velocity and concentration is determined. At present there are no models which allow prediction of this relationship based on fundamental parameters of the system, although attempts to

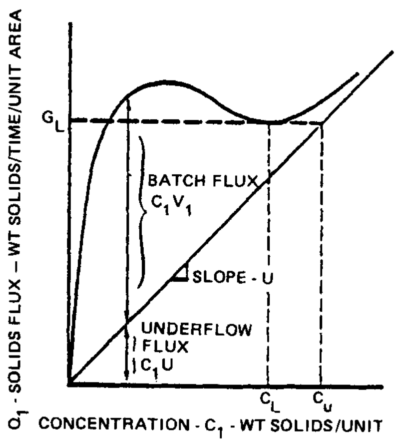

Figure 4A. (After, 18)

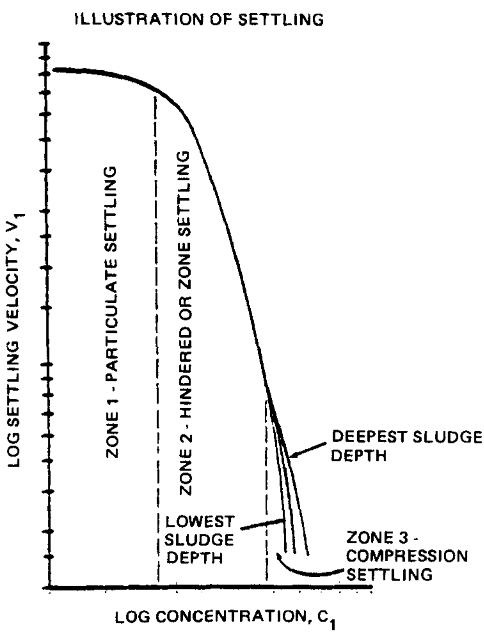

FIGURE 4 Batch Flux and Settling Zones (after 18).

describe a relationship based on particle-size distribution, zeta potential, and other elements describing particle behavior for specific experimental systems have recently been reported (15,16). Thus, batch testing is required.

Figure 4B is a representative settling curve determined from multiple batch testing in the typical Coe and Clevenger fashion. The section of the settling curve corresponding to zone settling can be approximated by a straight line whose equation is:

$$V_i = aC_i^{-b} \tag{7}$$

and "b" is calculated from the slope of the line in Figure 4B.

If Equation 7 is substituted into Equation 6 and rearranged, this leads to:

$$G_L = ab \left(\frac{b-1}{b} C_u\right)^{-b+1} \tag{8}$$

which is the design expression since unit area = G_L^{-1}.

b) batch thickening theory

Multiple batch tests can be used to determine the settling curve. However, this method is cumbersome and difficult if flocculation of the slurry is required (18). As the initial concentration (solids) is increased, flocculation generally becomes less efficient so that measurements of the initial settling rates can only be made for a narrow range of dilute concentrations.

An alternative to measuring the relationship between settling velocity and concentration is to use a single batch settling curve (24). Details are presented elsewhere (26) but results give values which can be used in Equation 8 for thickener design.

Thus, the original contributions to a thickening design model presented by Wilhelm and Naide can be summarized by realizing that Kynch did not treat steady-state continuous thickening explicitly. Wilhelm and Naide (18) extended Kynch's analysis for batch settling to continuous thickening and point out what they perceive as an error in Talmadge and Fitch's (27) approach. They conclude that the underflow concentration in the continuous thickener corresponds to the average concentration of the slurry bed in the batch tests, and the limiting concentration in the continuous thickener corresponds to the concentration just below the interface in the batch tests.

Fitch subsequently (25) claimed that Wilhelm and Naide's method has a misconception of existing art even though it is able to predict thickener unit areas. In fact he claims that the Talmadge and Fitch construction is simpler and more general. Nevertheless, Wilhelm and Naide's contribution is significant according to Fitch (25) because it is the first to recognize and utilize the deep column principle for Kynch interpretation of a batch settling curve and is one of the few that compare experimentally the predictions of a batch settling model with actual continuous thickening of the same material.

PARAMETER STUDIES WITH FLOTE

At the outset, it was hoped that the predictive capabilities of FLOTE for coal preparation could be examined based on the results of reported flotation tests (7,28) but this was impossible because of required model input characteristics. Nevertheless, it is believed to be worthwhile to estimate certain model parameters, propose a hypothetical input situation and plant configuration, and examine the sensitivity of changes in certain control actions on process performance. For example, the rate of aeration must be maintained at a critical limit to ensure sufficient interaction between valuable coal particles and air bubbles (29). Conversely, over-aeration can actually lead to a decrease in the grade of coal in the concentrate and aeration rate control has been suggested (29) as more important than control of the froth depth.

A case is chosen to examine systems response to changes in the control actions listed in Table 4. Percent coal solids in the head grade to the flotation unit, aeration rate for each cell, and rate constant for the coal fractions have been independently varied. Where possible, model parameters were estimated from

Table 4
Date to be Supplied on Continuous Basis

STATE VARIABLES:

1) Particle size distribution.
2) Fractional distribution of particles among G-classes in each D-class.
3) Percent solid in feed.
4) Solid feed rate.

CONTROL ACTIONS:

1) Percent solids in the flotation cell.
2) Percent solids in the concentrate.
3) Aeration rate for each stage.
*4) Froth depth.
*5) Reagent addition rates
*6) pH.

RESPONSES (Concentrate and Tailings Streams):

1) Particle size distributions.
2) Relative proportions of "minerals" in the solids.
3) Solid flowrate.
4) Percent solids.

*These additional control actions require monitoring beyond the immediate requirements of FLOTE (29).

typical operation values found in analogous systems, i.e., flotation in the mineral field. Results of the calculations may suggest methods of optimization as well as needed experimental research based on input requirements of FLOTE.

For the study, the plant configuration described in Figure 2 was chosen and the numbered streams in this diagram correspond to the numbers of the streams in the output statement from FLOTE. Performance results are represented in terms of the production function:

$$C \leq \text{FLOTE} \quad (A_1, A_2, K_c, PC) \tag{9}$$

where C = total recovery calculated by FLOTE for fixed value of aeration rate in Stage 1 (A_1) and Stage 2 (A_2) and for some un-

Table 5
Coal Recovery as function of aeration rate and coal content

Aeration Rate m^3/sec	Coal Recovery (%)			
	Stage 1		Stage 2+	
0.05	32.0	(35.9)‡	8.8	(15.7)
0.10	46.5	(48.2)	40.9	(52.2)
0.25	60.0	(60.7)	55.0	(60.0)
0.50	65.9	(67.0)	62.1	(64.3)
0.75	68.0	(69.5)	64.1	(65.6)
1.0	69.2	(70.9)	64.7	(66.0)
1.75	70.8	(72.9)	65.0	(66.3)
2.5	71.4	(73.7)	65.4	(66.5)

‡ bracketed values at 30% coal content
unbracketed values at 50% coal content

+ since Stage 2 results are dependent on the performance of Stage 1, an arbitrarily chosen aeration rate for Stage 1, corresponding to a marginal recovery of 10% recovery/m^3/sec aeration, has been adopted in this instance. This translates into approximately 0.5 m^3/sec. aeration rate for Stage 1.

specified chemical conditioning such that a specific flotation rate constant for coal, (K_c) is established. PC is the percent of coal in the solids reporting to the flotation unit.

Results are given for the first case in Table 5 and show the effects of varying aeration rates on coal recovery for Stage 1 and then Stage 2 corresponding to:

$$C \leq \text{FLOTE } (A_1, A_2^*, K_c = 8.5 \times 10^{-4}, 30 \ \& \ 50) \qquad (9a)$$

where A_1 is independent and A_2^* is determined at $MR_1 \leq 0.1$ where MR_1 is marginal coal recovery from Stage 1 and is expressed as a decimal fraction recovery/m^3/sec incremental aeration. This is somewhat arbitrary since cost difference for variation in aeration is probably negligible due to equipment characteristics (30).

Table 6 gives results for variation in pyrite flotation rate constant, K_p, and aeration rates corresponding to:

$$C \leq \text{FLOTE}\ (A_1,\ A_2,\ K_p = 1.0 \times 10^{-4}\ \&\ 3.0 \times 10^{-5},\ 30) \qquad (9b)$$

where A_1 and A_2 are independently varied for values of rate constant operating under normal conditions ($K_p = 1.0 \times 10^{-4}$) and suppressed pyrite flotation in both stages ($K_p = 3.0 \times 10^{-5}$).

Results of Table 5 suggest that variation in percent coal in flotation feed from 30 to 50% shows little effect on recovery for the aeration ranges examined except at extremely low rates ($< 0.1\ m^3$/sec). These results also suggest that the marginal recovery "optimum" criterion (10%) is met at 0.5 m^3/sec for Stage 1 and 0.25 m^3/sec for Stage 2 for either 30 or 50% coal in the feed.

Table 6 shows the overall recovery corresponding to Equation 9b. If aeration costs are independent of aeration rate then one would maximize Equation 9b subject to a given final pyrite grade, Gp, viz.

$$Gp \leq \text{FLOTE}\ (\ A_1,\ A_2,\ K_p,\ PC\ ) \qquad (10)$$

Practical ranges of aeration in flotation recovery of fine coal are 2 - 4 $ft^3/min/ft^2$ cell surface (30). For normal cell sizes and flow conditions here, this would translate into 0.35 m^3/sec - 0.70 m^3/sec for Stage 1 (70 m^3 volume) and 0.63 m^3/sec - 1.25 m^3/sec for Stage 2 (40 m^3). Under the conditions of Equation 9b, FLOTE predicts that coal recovery monotonically increases whereas in practice over-aeration can actually decrease yield. Furthermore, for metallurgical coal, pyrite rejection might require

Table 6
Coal Recovery and Pyrite Grade as Function of Aeration Rate and Pyrite Flotation Rate Constant (K_p)

Aeration Rate m^3/sec Stage 1 / Stage 2	Coal Recovery .01	Coal Recovery .05	(Pyrite Grade) .01	(Pyrite Grade) .05
.01	ǂ 10.0 (1.1)[a] + 10.0 (0.9)	16.4 (1.0) 16.4 (0.9)	16.0 (1.0) 16.0 (0.9)	15.7 (1.2) 15.8 (1.0)
.05	10.9 (1.2) 10.9 (1.0)	33.3 (1.5) 33.4 (1.2)	42.6 (1.6) 42.6 (1.3)	52.2 (2.0) 52.3 (1.5)
0.1	11.0 (1.2) 11.0 (1.0)	34.7 (1.5) 34.7 (1.2)	45.6 (1.8) 45.7 (1.4)	60.0 (2.3) 60.1 (1.7)
0.5	11.1 (1.2) 11.1 (1.0)	35.7 (1.6) 35.7 (1.3)	47.7 (2.0) 47.7 (1.5)	65.6 (2.8) 65.6 (2.1)

Specific rate constants (sec^{-1}) in force are as footnotes indicate:

K_c 100% Coal	K_p 100% Pyrite	100% Gangue	85% Coal 10% Gangue 5% Pyrite
ǂ .001	.00010	.000001	.00085
+ .001	.00003	.000001	.00085

a) Coal recovery and final pyrite grade both in percent are respectively first and second, (bracketed) values in matrix.

b) Values in matrix correspond to aeration rates in columns and rows.

c) Values in matrix correspond to 30% coal in solids reporting to flotation cell (PC).

$G_p \leq 2.0\%$. Thus, with this constraint, operation of Stages 1 & 2 should be at aeration rates of 0.5 m^3/sec each for pyrite suppression and maximum coal recovery. Specific values show coal recovery to be 65.6% and pyrite grade to be 2.1%.

It should be noted that the parameter set remained constant for each process run despite the fact that the parameters of FLOTE may vary as a result of varying the aeration rate. It was not known what parameters would be affected and to what extents. It is believed that the general shape of the response curves would be similar however.

The surface activity, K is probably the overriding factor in determining flotation cell performance and although it is specified as a model parameter, we know from experience that the surface activity can be manipulated through reagent addition. Thus, for typical ranges of control actions, probably the most important area requiring optimization is the type and quantity of reagents added to alter the surface activities of selected mineral constituents. This single control action will have the most significant effect on causing gangue and pyrite rejection and increasing the total quantity of coal recovered through flotation.

WILHELM AND NAIDE DESIGN MODEL AND OPERATION STRATEGIES

As a design tool, the model proposed by Wilhelm and Naide (18) appears to offer specific advantages for handling coal refuse slurries having been subjected to testing for this type of slurry. If the design model is flawed (25), it is because it does not account for the compression of the solids at the bottom of the thickener explicitly. Nevertheless, a method is proposed based on their model for determining the depth requirements of a thickener under conditions of low compression. However, a method for choosing a depth in the case of solids at high solids inventory which undergoes compression is not given. The reader is referred to Fitch's review (25) for such approaches.

There has been some discussion (18) on whether or not a thickener can be operated at more than one underflow concentration for a particular unit area. One side of the discussion reasons that at a particular feed rate, there is only one underflow concentration and underflow pumping rate that will give a stable bed level. If low underflow concentrations are expected, then this is approximately correct. However, if compression is considered to be significant, e.g., high underflow concentrations, the settling rate of a particular concentration layer may be affected by the presence of solids above it. Different underflow densities can be obtained from the same unit area, and different operating strategies can be employed.

Most thickeners are operated in one of three ways, with:

1) constant interface height,
2) constant underflow concentration,
3) constant underflow pumping rate.

All three methods take advantage of the increased settling velocities and higher underflow concentrations made possible by compression to maintain a stable sludge blanket in the wake of varying solids feed rates (18).

The most practical operating method for coal refuse thickeners is the first method, constant interface height. Since in many cases, the thickener underflow will be reporting to a slurry pond for increased dewatering, the underflow concentration is not as important as in other applications. If final dewatering by mechanical pressing or vacuum filtration is to be done, strategy #2 might be preferred.

The real weakness in the design model and operation strategy, is that there is no provision for determining the effect of varying flocculant dose (when applicable) on the performance characteristics of the thickener. The design model handles this implicitly in that the batch tests reflect performance under varying flocculant doses. However, flocculant dose rates can be used as part of an operation strategy as well and may be particularly important in high-capacity thickeners for design as well as operation (31). Lawler's (16) approach to developing a particle-mechanism thickening model may eventually adequately account for flocculant addition.

Much of the same monitoring equipment is required by the thickening unit as required by the flotation unit. Two additional factors which must be monitored in thickeners are:

1) Sludge level height and
2) Torque on the mechanical rake

Wells (29) describes how this can be accomplished satisfactorily.

FLOTE AND WILHELM AND NAIDE MODEL ATTRIBUTES

Required and desired attributes were given in Table 2. Desirable attributes are reviewed here:

1) Neither model is dynamic. Technical limitations for this for each model have already been discussed. Steady state simulators will have to suffice for the near term.

2) Level of input information for FLOTE is high. Thus we have a versatile tool for optimizing. Some of the shortcomings of FLOTE input include: a) poor handling of the froth characterization and no provisions for inputing the froth height and b) no provisions for differentiating between mineral specific gravities. The level of input demanded by the Wilhelm and Naide (18) design model is relatively simple requiring only two constants derived from batch settling tests and knowledge of the solids feed rate.

3) The level of output from FLOTE is also extensive, yielding a stream analysis of particle size data as well as recovery-of-mineral information. The output of the thickening design model is overly simplistic, even in this simple application, inasmuch as it does not predict depth values; only the cross-sectional area of the thickener is specified. Furthermore, no particle characteristics of overflow or underflow are reported. This would be useful for obtaining final dewatering information.

4) The monitoring requirements of state variable and control actions, as well as other pertinent information, can apparently be satisfied with the current level of instrumentation technology (29,32).

5) While FLOTE is already computer coded, the simple nature of Wilhelm and Naide's design model does not necessitate a computer solution.

6) FLOTE is equipped to simulate any flotation plant configuration one is likely to encounter in the flotation of coal. The sixth desirable attribute is not particularly relevant to the thickening design model.

CONCLUSIONS AND RECOMMENDATIONS

The optimization of unit processes in coal preparation requires a rational basis. Systems modeling has been an accepted means for supplying this rationale in other fields including coarse coal cleaning and should be applied here as well.

FLOTE, a distributed-rate-constant simulator which accounts for most of the control actions applied to coal flotation units, should prove quite useful in optimizing the design and operation of the same. Two limitations of the model are: 1) oversimplification of the froth zone and the control actions affecting its performance, and 2) possible errors introduced into the mass balance as the result of using an average specific gravity to characterize solids.

Process performance depends mostly on surface activity of the particles and the control action affecting this most is the type and quantity of reagent addition(s). Relations between surface activity and reagent addition are badly needed.

This work has surely pointed out the need for a more comprehensive, theoretically-based, thickening model. The Lawler (16) model is a step in the right direction. Present model emphasis is directed towards design. Dynamic models are needed if total process optimization is to be realized.

Of the models available for design, the approach proposed by Wilhelm and Naide (18) appears to correlate well to the settling properties of coal refuse slurries. The effects of compression and depth requirements need to be determined in the design equation. Thickening design and operation should not be optimized independent of one another as Wilhelm and Naide tend to suggest. Furthermore, in view of the effectiveness of flocculant addition in increasing settling rates and thus reducing the size requirements of thickeners, theoretical efforts should accelerate an understanding of flocculant addition effects, although fruition appears some years off.

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