REDESIGN OF INDUSTRIAL COLUMN FLOTATION CIRCUITS BASED ON A SIMPLE RESIDENCE TIME DISTRIBUTION MODEL

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ABSTRACT

The potential for improved selectivity has made column flotation cells a popular choice for upgrading fine coal. Unfortunately, recent production data from full-scale column plants indicate that many industrial installations have failed to meet original expectations in terms of clean coal recovery. Theoretical studies performed using a simple dispersion model showed that this inherent shortcoming could be largely minimized by reconfiguring the columns to operate in series as a cell-to-cell circuit. Follow-up field data showed that this low-cost modification increased flotation recovery as predicted by the dispersion model. This study presents the key findings obtained from the field investigation and provides generic guidelines for designing multi-stage column circuits.

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<u>Note</u>: Unless otherwise noted, all images are the property of the author.

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1.0 INTRODUCTION

1.1 Background

1.1.1 Preamble

Flotation, or more specifically froth flotation, is a physiochemical method of concentrating finely ground ores. The basis of this process is the utilization of the differences in the natural or created hydrophobicity of the two minerals that are to be separated. The separation involves chemical treatment of an ore pulp to create conditions favorable for the attachment of selected mineral particles to air bubble. The combined air/mineral particles have specific gravities less than the pulp and will float to the surface, forming a mineral laden froth that can be skimmed off from the flotation unit while the other minerals remain submerged in the pulp.

Flotation is the most important single mineral concentrating operation. It is also the most efficient, most complicated, most sensitive, most challenging and least understood mineral processing operation today. The art of flotation has been correctly called the most impressive wide-scale technical application of classical surface chemistry. Only in the last three decades has much progress been made in understanding its mechanisms.

Over 95% of all base metals are concentrated by flotation. In coal preparation plants using flotation, approximately 5-15% of the raw coal feed is processed through the flotation unit for recovery of fine coal particles. Virtually the entire free-world supply of copper, lead, zinc, and silver is recovered in the froth of flotation machines. As the grade (the assay of valuable constituent) of mineral deposits decreases and coal and metal prices rise, the importance of flotation increases. In turn, the necessity of improving selective recoveries becomes even more critical.

1.1.2 Coal Flotation

The basis for flotation is the creation or enhancement of the differences in hydrophobicity. By making specific mineral surfaces hydrophobic chemically, a physical separation of different particles can be achieved when air is injected into the slurry. The basic difference between ore and coal flotation is that for ores the entire tonnage is ground to flotation size, whereas for coal only the fine fraction (about 60 mesh x 0) not processed by gravity concentration is treated by flotation.

In coal flotation, the differences in the wettability of coal (hydrophobic) and rock (hydrophilic) are exploited. Hydrophobic particles strike bubbles and attach (see Figure 1), while hydrophilic particles do not attach. This attachment occurs when air is injected into the slurry containing coal and rock. The coal particles attach to the air bubbles and

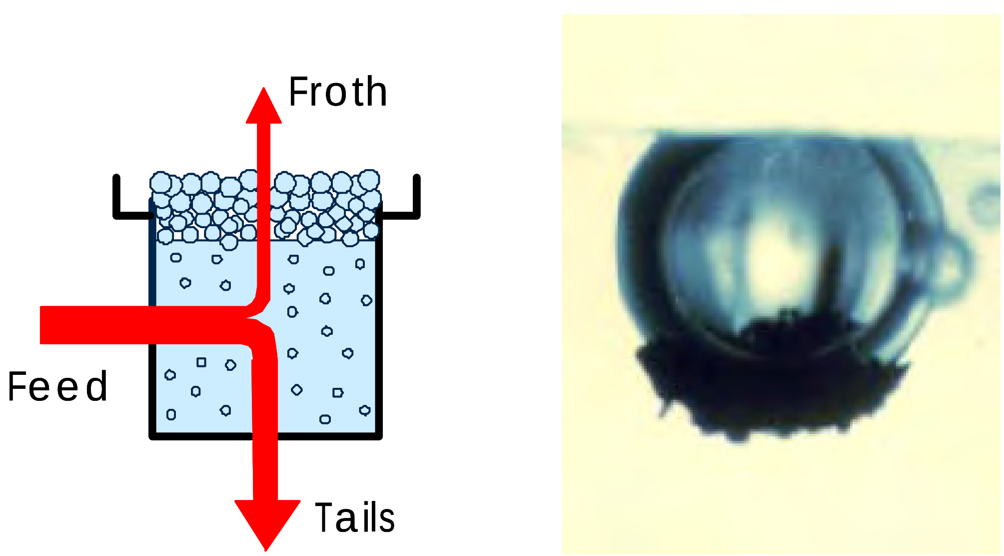

Figure 1. Attachment of coal particles to a bubble in froth flotation.

are buoyed to the pulp surface. The rock particles remain in the slurry and are rejected as tailings. The pulp phase provides an opportunity for particles to selectively attach to the air bubbles. The froth phase is used to separate bubble-particle aggregates from the pulp.

The last objective is to form at the surface of the cell a semi-stable mineral-laden froth that is fragile enough to allow entrapped rock materials to fall and stable enough to retain the mineral particles selectively. The froth must also have sufficient strength to permit mechanical or natural removal from the machine. Once the froth is out of the flotation cell, it should break down rapidly to allow pumping of the concentrate. The mineral-laden froth is also referred to as concentrate or in coal flotation "the float product".

1.1.3 Flotation Machines

Froth flotation is performed in machines specifically designed for that purpose. There are two basic types of flotation machines and adaptations of these machines. Conventional (mechanical) and column flotation cells are the two types of flotation machines used in flotation (Figure 2).

Most of the installations in the coal industry use mechanical (conventional) flotation machines. Conventional cells are open top vessels connected in rows (banks) through which the pulp flows continuously from the head (feed) end to the tail end. These machines are typically arranged in banks of 4-6 cells. The cells are individually agitated using an impeller. The impeller provides the energy needed to keep particles in suspension, suck in air (a blower may also be used) and to disperse air into small air bubbles.

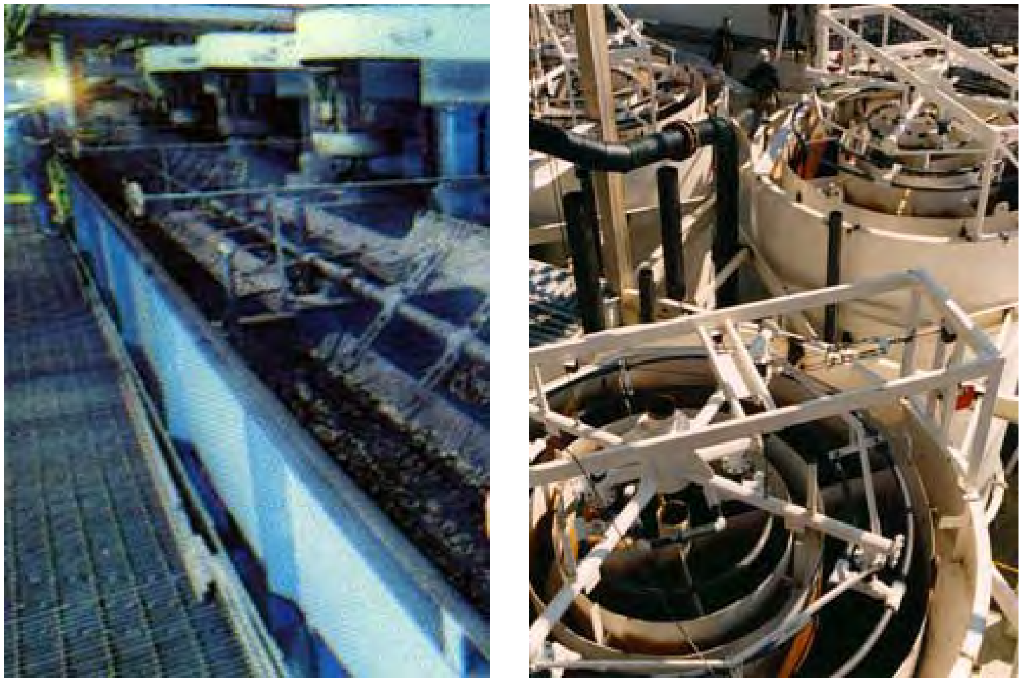

Figure 2. Conventional (left) and column (right) flotation cell installations.

A column cell consists of a vertical column into which air is added at the bottom using spargers (no impeller). Columns usually use wash water to prevent fine clays from being entrained into the clean coal froth. Gas compressors are usually required to provide the air flow and energy for air dispersion. A column cell is usually a stand-alone machine that is not connected in banks as conventional cells are.

1.1.4 Flotation Reagents

The chemistry of flotation consists of collectors, frothers, and conditioning reagents (modifiers). Coal flotation is primarily concerned with collectors and frothers. Collectors are the heart of the flotation process since they are the reagents that produce the hydrophobic film on the mineral particle. Coal, although hydrophobic, is enhanced by the collector increasing the rate at which it reports to the froth phase and helps the coal attach to the air bubble well. The collector should be applied as far back in the flotation circuit as possible because conditioning time with the flotation feed is very important. The conditioning time becomes even more important with harder to float coal such as low rank bituminous or oxidized coal. Conditioning time in a flotation circuit may range from seconds to minutes.

The most common types of collectors for coal flotation are diesel fuel, kerosene, and fuel oil. There are also environmentally friendly collectors available such as bio-diesel and tree products. The collector dosages range from 0 to 5 lb/ton. Higher rank bituminous coal may require no collector, low rank bituminous may require as much as 3 lb/ton of coal and weathered (oxidized) coal may require 5 lb/ton of coal.

The main purpose of the frothing agent is the creation of a froth capable of carrying the mineral-laden bubbles until they can be removed from the flotation machine. This objective is accomplished by imparting temporary toughness to the covering film of the bubble. The life of the individual bubble is thus prolonged until it can be further stabilized by adherence of mineral particles and joined with other bubbles at the pulp surface to form a froth. Once it is withdrawn from the flotation machine, however, the froth should break down rapidly, to prevent interference with subsequent processing operations. The frother decreases the size of the bubble in the pulp, thus increasing the surface area of the air introduced into the flotation machine. This enlarges the number of collisions of the mineral particles with the bubble and chance of the mineral to be floated to the cell surface.

The ability to lower the surface tension of water is a characteristic of all frothers. Since most organic compounds can do so to at least a limited degree, the number of frothers commercially available might be expected to be quite large. However, the following requirements on the frothing reagent have restricted the frothers to only a few chemicals. A frother must:

- be low in cost,
- be readily available,
- be effective in low concentrations,
- have no collecting properties or have collecting affinity to the valuable mineral,
- be non-toxic,
- not have a repulsive odor,
- produce a persistent froth that will last the length of the flotation bank, but that will collapse when skimmed into the launder,
- be stable during storage even at adverse atmospheric conditions (low and high temperatures),
- have strength enough to withstand the turbulence in the flotation cells and to carry the valuable mineral, and
- be brittle enough to allow the rejection of unwanted gangue material.

Frothers may consist of short-chain alcohols, derivatives, glycols, and glycol mixtures. The dosages of frother for conventional cells should typically be in the range of 7 to 15 ppm. The dosages for a column cell, which are usually higher, should typically be from 10 to 25 ppm.

1.2 Flotation Kinetics

Flotation performance is dependent on rate constant (k), retention time (τ), and cell mixing (Pe). The interrelationship between these variables is described in greater detail later in this document. Retention time is the amount of time the coal particle has to stay in the pulp. In a coal flotation circuit, 3½ to 4 minutes of residence time is needed for good recoveries. Mean retention time can be determined by using retention time distribution (RTD) testing or by using the formula:

$$\tau = V/Q \quad [1]$$

in which V is the active volume of the cell and Q is the flow rate to the cell. The number and size of the flotation cells should be adjusted to have the proper amount of time for flotation. RTD testing involves using a tracer solution, such as salt, applied to the feed end of a flotation cell and taking timed samples vs. concentration of each sample at the tails of the cell.

Cell mixing is the last but not least of the areas affecting flotation. The best flotation recoveries would come from a plug flow system rather than a well mixed system. Conventional cells are set up to try and obtain plug flow using cell to cell. Column flotation is well mixed with a greater chance of short circuiting.

The rate constant (k) indicates how fast coal floats and is dependent on coal type/size/solids content/feed rate/froth depth, reagent type/dosage, gas flow rate and sparging/agitating system. Some of the factors that can impact flotation kinetics are described below.

1.2.1 Coal Rank

Floatability is very dependent on the rank of coal. Low volatile, mid volatile, high volatile, anthracite, or oxidized (weathered) coals all float differently. Low volatile coal is the easiest to float and oxidized coal is the hardest to float. Oxidized coal causes problems with poor recoveries even when blended with higher rank coals.

1.2.2 Particle Size

Coal flotation is performed mainly on about 60 mesh x 0 coals. The size consistency within the 60 mesh x 0 is very important to coal flotation. The courser the coal the harder it is to keep attached to the air bubble. The finer the coal is will lead to poor air bubble particle collision.

1.2.3 Solids Content

The solids content of a flotation circuit should be around 3-7% depending on the classification system. At higher the solids contents, a trade-off between residence time and entrainment begins to occur.

1.2.4 Froth Depth

Froth depth changes how wet or dry the froth will be for a given reagent dosage. A deep dry froth reduces entrainment and clean coal ash. However, a very deep froth may become unstable and will lower recovery. The type of frother should be tested to determine the best product for each flotation circuit. Dosages of frother should be determined to optimize the flotation circuit. The types and dosages of frother can increase or decrease the flotation rate.

1.2.5 Aeration Rate

The gas flow rate should be sufficient to provide the bubble surface area needed to carry out the particles. Conventional cells should use about 3.5 cfm/ft^2 and columns about 5.0 cfm/ft^2. The rate the coal floats decreases with low gas flow rates.

1.2.6 Sparging System

The sparging/agitating systems should be sufficient to keep the coal particles moving for attachment to the air bubble. The energy supplied to the flotation unit should be sufficient to accomplish this task.

1.3 Flotation Modeling

Despite being a complex process, flotation can often be modeled using simple kinetic expressions. For a batch flotation cell, recovery can be predicted based on:

$$R = 1\text{-}\exp(-k\tau) \qquad [2]$$

in which R is the factional recovery, k is the rate constant (1/min) and τ is the mean retention time (min). Flotation feeds are typically assumed to contain three components:

fast floating (e.g., pure coal)

slow floating (e.g., middlings)

non-floating (e.g., pure rock)

Therefore, mathematically, total recovery (R_{tot}) can be calculated from the factional recoveries of fast (R_f), slow (R_s) and non-floating (R_n) components using:

$$R_{tot} = R_f + R_s + R_n \qquad [3]$$

$$R_f = \phi_f\ (1\text{-}\exp\{-k_f\tau\}) \qquad [4]$$

$$R_s = \phi_s\ (1\text{-}\exp\{k_s\tau\}) \qquad [5]$$

$$R_n = \phi_n\ (1\text{-}\exp\{-k_n\tau\}) \qquad [6]$$

in which ϕ_f, ϕ_s and ϕ_n, are the respective mass fractions of fast, slow and non-floating solids and k_f, k_s and k_n are the corresponding rate constants of fast, slow and non-floating components. In practice, the faction of non-floating solids can usually be determined from kinetics testing and release analysis.

Kinetics testing, or timed flotation, involves collecting froth products as a function of time (see Figure 3(a)). The froth products collected from this procedure include both floatable solids as well as entrained solids. Release analysis, on the other

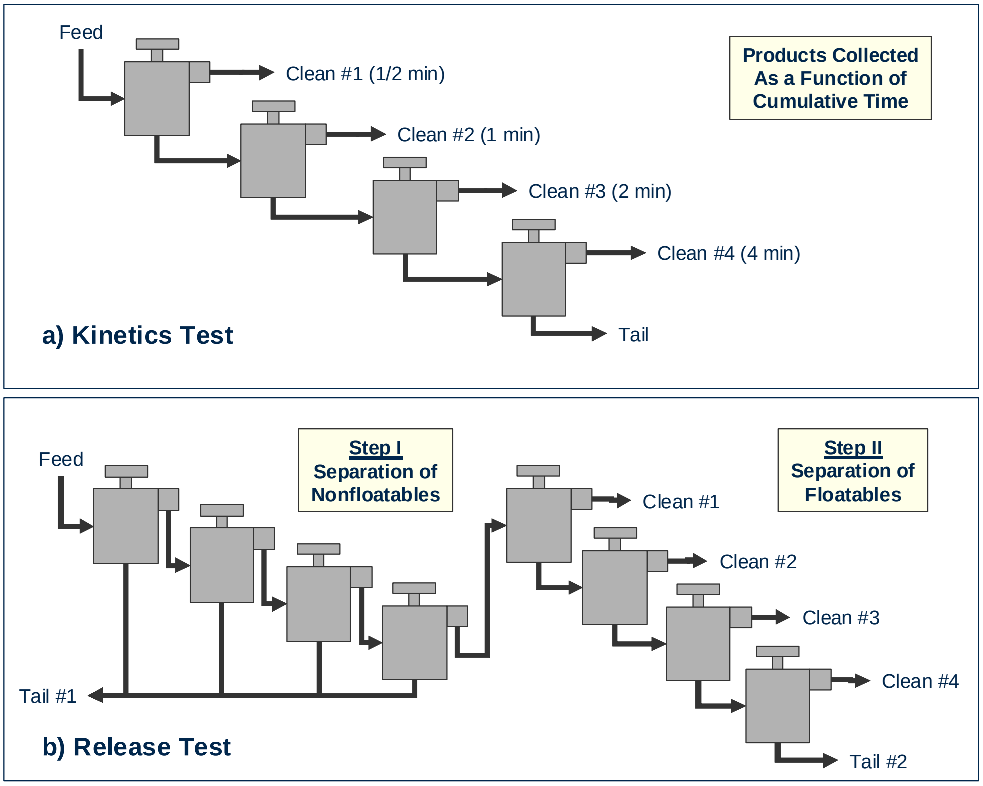

Figure 3. Schematic of laboratory flotation procedures used in (a) timed kinetics testing and (b) release analysis testing.

hand, is often used to establish the best possible separation results for flotation. It is used for plant design and for assessing flotation efficiency. As shown in Figure 3(b), the process of running a release analysis consists of two steps. The first step is to separate the floating from the non-floating coal. Step two the fast floating coal to slow floating coals are separated into grades and results are calculated to give grade recovery data and plots associated with yield and recovery. The final froth products collected from this procedure include only floatable solids since entrained solids have been removed upstream. Table 1 shows a representative set of kinetics data for a typical coal flotation system.

Table 1. Example of kinetic parameters for a typical coal flotation system.

Parameter	Fast Floating	Slow Floating	Non-Floating
Mass Fraction	0.30	0.40	0.30
Ash Content	3.5%	10.0%	85%
Rate Constant	0.80 min^{-1}	0.30 min^{-1}	0.002 min^{-1}

1.4 Objective and Approach

Froth flotation is widely considered to be the most practical and economical method for upgrading coal fines that are too small to be treated using density-based separators. Most of the installations in the coal industry use mechanical (conventional) flotation machines. The major shortcoming of conventional flotation technology is that it allows some ultrafine impurities (e.g., clay) to be hydraulically carried into the clean coal product in the process water that reports to the froth. This contamination substantially reduces the overall quality of the clean coal froth product. The entrainment problem has forced many coal operations to install column flotation cells in place of conventional flotation machines (Davis et al., 1995). Column cells are able to reduce entrainment by the addition of a countercurrent flow of wash water to the top of the froth. The water rinses ultrafine impurities such as clay from the froth back into the flotation pulp. Studies indicate that less than 1% of the feed slurry reports to the froth in a properly operated column. However, a recent field survey showed that while column technology can indeed improve clean coal quality, many industrial installations are unable to provide good coal recoveries (>80-85%) when compared to their conventional counterparts.

Since recent production data from full-scale column plants indicate that many industrial installations have failed to meet original expectations in terms of clean coal recovery, a detailed field study was undertaken at an industrial site (i) to identify the root cause of the problem responsible for the poor coal recovery in column cells and (ii) to develop low-cost methods by which this problem may be minimized. The study included experimental measurements of residence time distributions and theoretical analysis of cell mixing using a simple dispersion model. This document presents the key findings obtained from the field study and provides generic guidelines for designing multi-stage column circuits. The resultant data is used to demonstrate the large financial impacts of improving flotation performance on plant profitability.

2.0 LITERATURE REVIEW

2.1 Column Flotation

2.1.1 Overview

The major problem with conventional flotation is that it allows a fine mineral (clay) slimes to be recovered with the water reporting to froth (see Figure 4). Column cells can overcome this shortcoming by washing the clay slimes from the froth product using wash water (see Figure 5). Sufficient wash water must be added to ensure that all of the feed water that normally reports to the froth product has been replaced with fresh or clarified water. The froth depth must also be much greater than that used in conventional flotation to obtain good distribution of the wash water and to achieve the desired froth

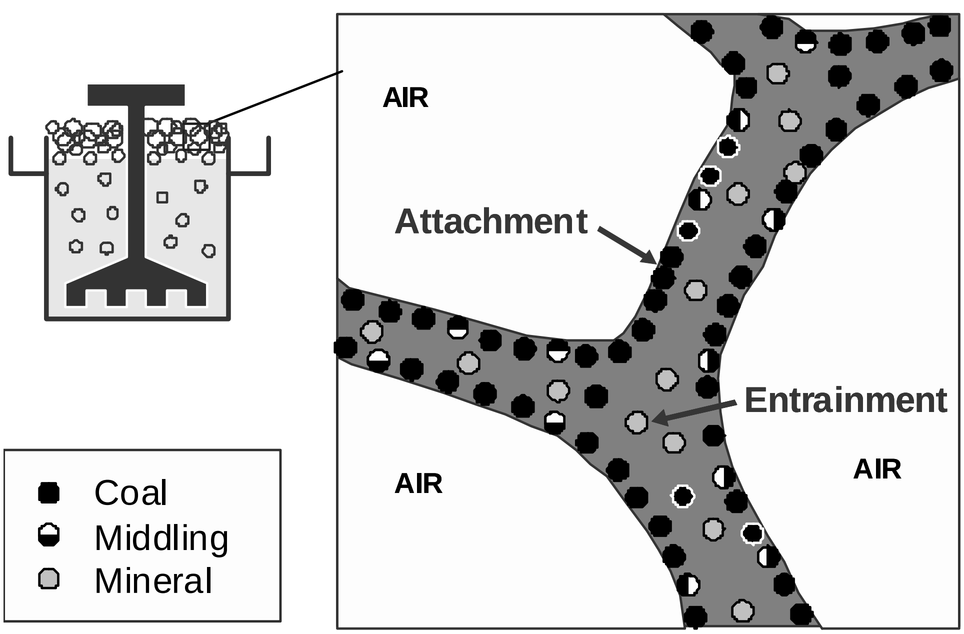

Figure 4. Entrainment of fine mineral matter with the water reporting to the froth product.

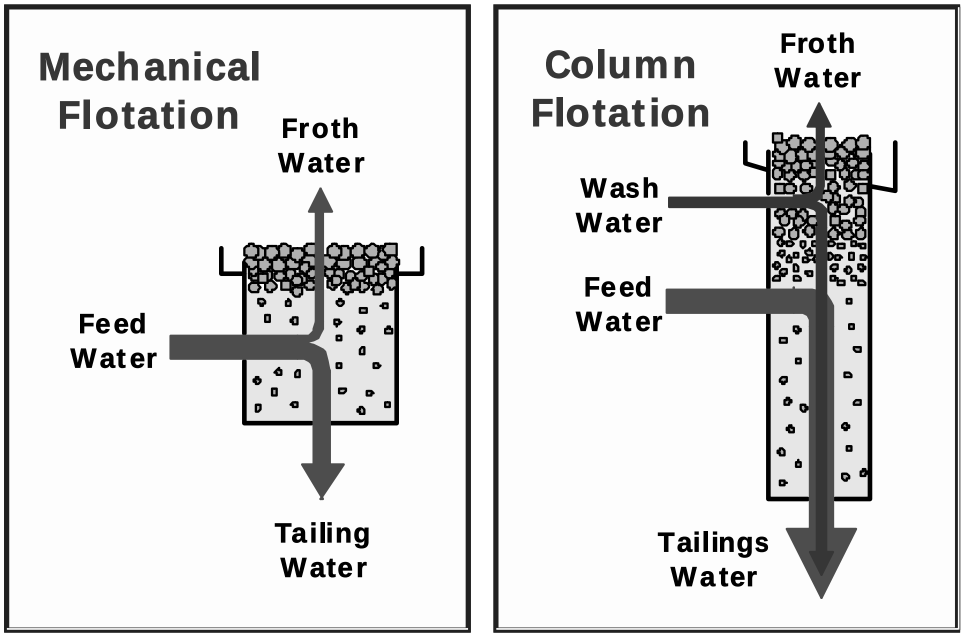

Figure 5. Comparison of how feed water reports to the froth in mechanical and column flotation cells.

washing action. This unique feature makes it possible for column cells to achieve superior ash removal when compared to conventional flotation cells. In most cases, the separation performance obtained using column cells approaches very closely to the ultimate cleanability curve predicted using flotation release analysis.

Although column cells offer superior performance in terms of ash removal, many operators in the coal industry have been reluctant to adopt the column technology. The comparatively low market value of fine coal often makes it difficult for mine management to justify the higher capital and operating costs for columns, particularly if the expenditure is for the replacement of existing conventional cells. Many operators also have the perception that columns are more difficult to operate, entail greater amounts of maintenance, and require complicated ancillary systems for compressed air and wash water. Nevertheless, several operators have identified suitable niches where the advantages of columns can be successfully exploited.

2.1.2 Commercial Installations

Table 2 provides a listing of some of the full-scale installations of columns that are currently in commercial service in the U.S. coal industry. As shown, the most popular brands of columns include Microcel, Eriez/CPT (CoalPro) and Jameson. Although the Jameson cell is not really a "column", it is included in this list since it typically uses wash water to improve ash rejection. Details related to the specific design features of the

Table 2. Examples of U.S. commercial coal column installations.

Company / Location	Brand	Number Units	Diameter (ft)	Height (ft)	Size
ANR – White Tail	Eriez/CPT	5	14	28	100 M x 0
ANR – Middle Fork	Microcel	5	10	25	100 M x 0
ANR - Holston	Microcel	1	14	30	100 M x 0
ANR - Roxanna	Microcel	2	13	28	65 M x 0
ANR – Toms Creek	Microcel	2	14	28	100 M x 0
ANR – Brooks Run	Microcel	2	15	35	100 M x 0
Carbontronics Fuel - Pawney	Jameson	1	16	12	100 M x 0
Carbontronics Fuel - Gibraltar	Jameson	1	16	12	100 M x 0
Carbontronics Fuel - Linville	Jameson	1	16	12	100 M x 0
Cline Resources – Coal Clean	Eriez/CPT	3	15	28	100 M x 0
CONSOL Energy - Keystone	Jameson	1	16	12	100 M x 0
CQ, Inc. – Ginger Hill	Jameson	1	16	12	100 M x 0
CQ, Inc. – Pleasant Ridge	Jameson	1	16	12	100 M x 0
ICG – Marrowbone	Microcel	1	8	26	150 M x 0
Massey – Lady Dunn	Microcel	3	14	30	100 M x 0
Massey– Power Mountain	Eriez/CPT	2	13	24	100x325 M
Massey– Liberty	Eriez/CPT	3	13	24	100x325 M
Massey– Bandmill	EIMCO	2	13	24	100x325 M
Massey– Marfork	Jameson	4	16	12	100 M x 0
Ohio Valley - Century	Eriez/CPT	2	14	24	100x325 M
Peabody – Sugar Camp	Jameson	1	16	12	100 M x 0
Powell Mountain – Mayflower	Ken-Flote	4	8	22	100 M x 0
Sigmon Mining – Sigmon	Eriez/CPT	2	14	28	100 M x 0
TECO– Clintwood Elkhorn	Eriez/CPT	2	10	24	100x325 M

various column technologies are available in the literature (McKay *et al*., 1988; Yoon *et al*., 1992; Davis, 1995; Manlapig *et al*., 1993; Finch and Dobby, 1990; Finch, 1995; Rubinstein, 1995; Wyslouvil, 1997). Due to economy of scale, recent trends in the coal industry have shifted toward the installation of smaller numbers of very large (>14 ft) diameter units. Although most of the installations involve the treatment of minus 100 mesh feeds, some operations have been able to justify the use of columns to treat deslimed 100x325 mesh streams. In such cases, the columns have been used to further reject high-ash slimes that bypass to the deslime underflow.

2.2 Column Operation and Design

2.2.1 Separation Performance

The primary advantage of column cells is their ability to achieve a higher level of separation performance compared to conventional flotation machines. For example, Figure 3 provides a comparison of separation results obtained using both conventional and column flotation cells at an industrial plant site. The column data were obtained using both a laboratory (2 inch diameter) test column and a full-scale (10 ft diameter) production column. For comparison, test data were also obtained from a single bank of five full-scale Denver cells (300 ft^3 each) already in operation at the preparation plant. A release analysis test was also conducted on the flotation feed to provide a baseline for comparative purposes (Dell *et al*., 1972). Release analysis is often described as the flotation equivalent of the washability procedure for density separations.

Several important conclusions may be drawn from the results given in Figure 6. First, the separation data obtained using the column cells fall along nearly identical

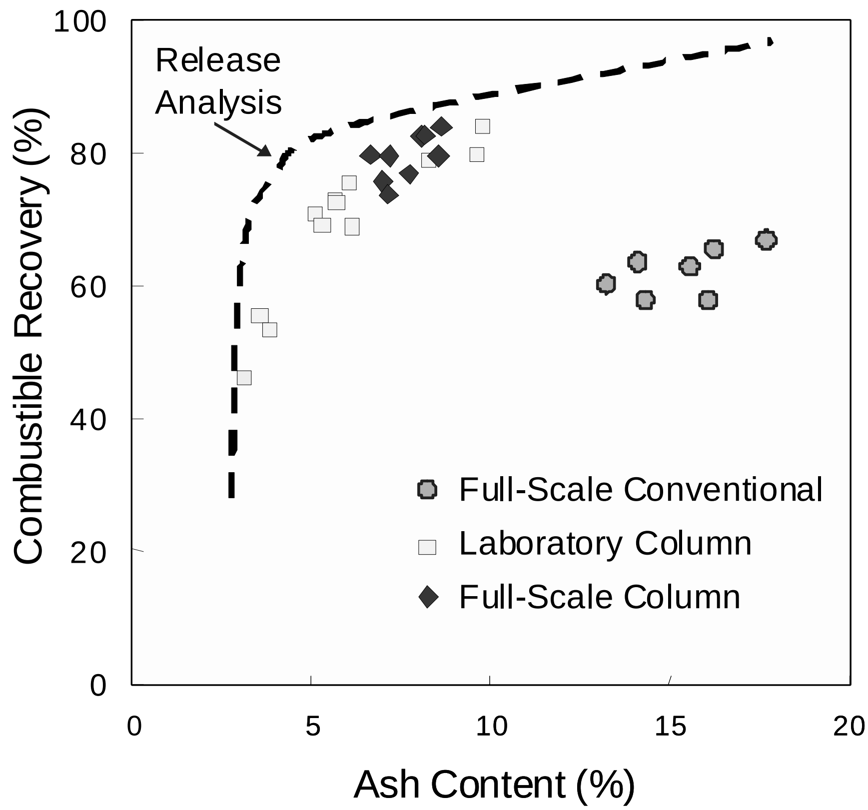

Figure 6. Comparison of column and conventional flotation cells (Davis et al., Plant Practices in Fine Coal Flotation, 1995).

recovery-grade curves regardless of column size. This finding demonstrates that the data obtained from the smaller test columns can be used to accurately predict the performance of full-scale columns.

The scale-up of separation data obtained using conventional cells is generally less reliable due to unpredictable variations in froth product quality. These variations are attributed to shifts in nonselective hydraulic entrainment created by differences in froth drainage behavior for conventional cells of different sizes (Lynch *et al.*, 1981). Second, the performance data for the column cells tend to fall just below the separation curve predicted by release analysis. This finding implies that the column cells provide a level of performance that would be difficult to achieve even after multiple stages of cleaning by conventional machines. For this particular example, the column cells reduced the concentrate ash by nearly 8 percentage points compared to the existing plant rougher bank, while increasing combustible recovery by approximately 20 percentage points.

2.2.2 Slurry Residence Time

All flotation cells must have adequate slurry residence time in order to achieve an acceptable recovery of clean coal. The mean residence time can be calculated by dividing the active volume of the flotation pulp by the volumetric flow rate of refuse. For a typical 100 mesh x 0 feed, a residence time of 3.5-4.0 minutes is often adequate for 3-5 cell bank of conventional flotation machines. The cell-to-cell arrangement used by conventional machines minimizes the bypass of feed slurry to tailings. In contrast, column cells are usually arranged in parallel, which can allow some floatable coal to bypass to tailings. As a result, columns typically require significantly more residence time to achieve the same level of coal recovery (Dobby and Finch, 1985; Mankosa *et al*., 1992). This situation is particularly serious for the relatively low profile columns that are typically employed in the coal industry This problem can lead to lower recoveries of coal if not accounted for during the column design process.

The detailed procedure for estimating the column residence time required to achieve a given recovery has been presented in the literature (Dobby and Finch, 1986a). This procedure is complicated and well beyond the scope of this article. However, as a rough rule-of-thumb, a column cell will typically require approximately twice the residence time of a 4-cell bank of conventional cells and three times the residence time of a batch laboratory flotation cell. Therefore, it would not be unusual for a column cell to require 7-8 minutes of residence time. Some industrial columns have been designed to operate with residence times of 12-15 minutes or more. It is also important to note that the active volume for columns is defined as the volume of pulp contained between the bottom of the froth and the top of the air spargers. The calculated volume is also typically reduced by 10-20% to account for the volume occupied by air bubbles. This volume is substantially less than the total volume of the column tank structure.

2.2.3 Froth Carrying Capacity

Figure 7 shows a typical relationship between combustible recovery and residence time for a coal flotation column. The volumetric feed flow rate to the column was steadily increased in this series of tests in order to reduce the mean residence time. As shown, the test data obtained at 2% solids correlated well with the theoretical performance curve predicted using Levenspiel’s (1972) equation. However, as the solids content of the feed stream was increased to 5% solids, the recovery dropped sharply and deviated substantially from the theoretical curve. This deviation was more pronounced when the solids content was increased to 10% solids. The drop in recovery with increasing solids content can be attributed to limitations associated froth carrying capacity. When this occurs, there is insufficient bubble surface area to carry all of the floatable particles through the froth. While the carrying capacity restriction can often be ignored for conventional flotation circuits, it is of great importance in the design of column cells. This is largely due to the fact that the specific surface area of the cell (ratio of the cross-sectional area to the volume) is much higher for conventional cells.

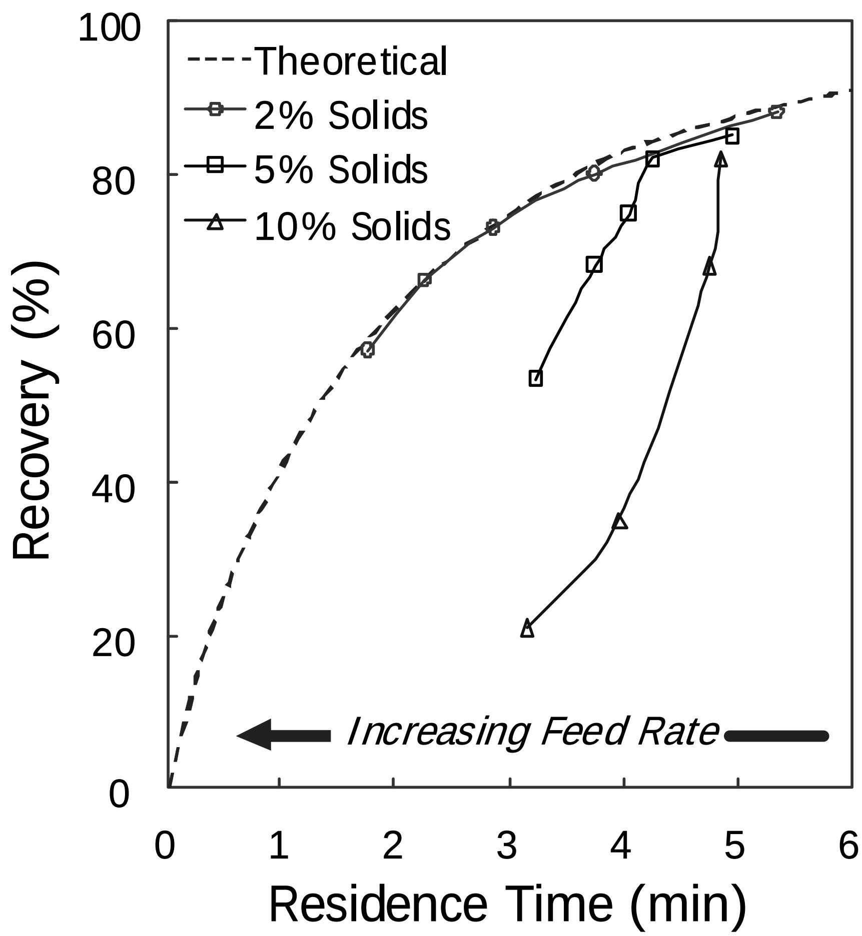

Figure 7. Effect of residence time on recovery for different feed solids contents (Luttrell et al., Proceedings of the 21st Annual International Coal, Aggregrate and Mineal Processing Exhibition and Conference, 2004).

Theoretical studies indicate that carrying capacity, which is normally reported in terms of the rate of concentrate floated per unit cross-sectional area, is linearly related to the size and density of particles in the froth (Espinosa-Gomez et al., 1988; Sastri, 1996). The froth carrying capacity may be estimated from laboratory and pilot-scale flotation tests by conducting experiments as a function of feed solids content (Finch and Dobby, 1990). Most of the full-scale columns in the coal industry operate at carrying capacities in the range of 0.06 to 0.24 tph/ft^2, with an average of about 0.12 tph/ft^2 for 100 mesh x 0 feeds. The values at the lower end of this range typically correspond to coal feeds of much finer particle size (325 mesh x 0). Once this value is known, the required column cross-sectional area can be determined by dividing the expected clean coal tonnage (tph) by the carrying capacity (tph/ft^2). However, for scale-up purposes, it is generally more convenient to calculate the clean coal tonnage for a full-scale column from a smaller test column using:

$$\frac{\text{Large Column tph}}{\text{Small Column tph}} = \left(\frac{\text{Large Diameter}}{\text{Small Diameter}}\right)^2 \qquad [7]$$

According to this relationship, a 14 ft diameter column should have a clean coal capacity that is nearly twice that of a 10 ft diameter column [i.e., $(14/10)^2 = 1.96$]. Field data suggest that this relationship holds valid for coal columns as large as 15 ft diameter. However, this suggestion appears to contradict data reported by Amelunxen (1990) stating that froth mobility problems make overflow lip length (and not cross-sectional) the appropriate scale-up parameter, i.e:

$$\frac{\text{Large Column tph}}{\text{Small Column tph}} = \left(\frac{\text{Large Diameter}}{\text{Small Diameter}}\right) \qquad [8]$$

This expression may be more appropriate for columns that operate with unstable froths created by inadequate frother additions, improper water distributors, or coarser coal feeds.

2.2.4 Froth Wash Water

The most important feature of a column is the wash water system. A froth depth of about 2-3 ft is typically required to ensure good distribution of the wash water and to prevent short-circuiting. In addition, the flow of wash water must exceed the volumetric flow of water reporting to the clean coal product to prevent entrainment of the high-ash slimes. In most cases, less than about 1% of the feed water will report to the froth product if the wash water is properly controlled. The amount of water carried by the froth can be calculated from:

$$Q = 4\,C\left(\frac{100}{P} - 1\right) \qquad [9]$$

Where Q is the water demand (gpm/ft^2), C the froth carrying capacity (tph/ft^2) and P is the froth percent solids. For example, a column cell producing 0.12 tph/ft^2 of dry clean coal at 18% solids will carry about 2.2 gpm/ft^2 of water from the pulp into the froth [i.e., 4 x 0.12 (100/18-1) = 2.2].

Entrainment should theoretically be eliminated when the number of dilution washes (defined as the froth water demand divided by the wash water addition rate) reaches a value of one. However, as shown in Figure 8, froth mixing usually requires that 1.5 dilution washes be used to fully suppress hydraulic entrainment. This constraint dictates that a wash water flow rate of at least 3.3 gpm/ft^2 [i.e., 1.5 x 2.2 = 3.3] be used in the current example to prevent the entrainment of high-ash slimes. Field data collected obtained from columns operating in the coal industry suggest that a wash water flow rate of 3.0-3.5 gpm/ft^2 is normally adequate for most commercial installations. However, higher gas and frother addition rates will typically increase the froth water demand and, as a result, the amount of wash water required. Excessive wash water flows should be avoided since the extra wash water passing downward through a column will create an undesirable reduction in the slurry retention time and, hence, a potential reduction in

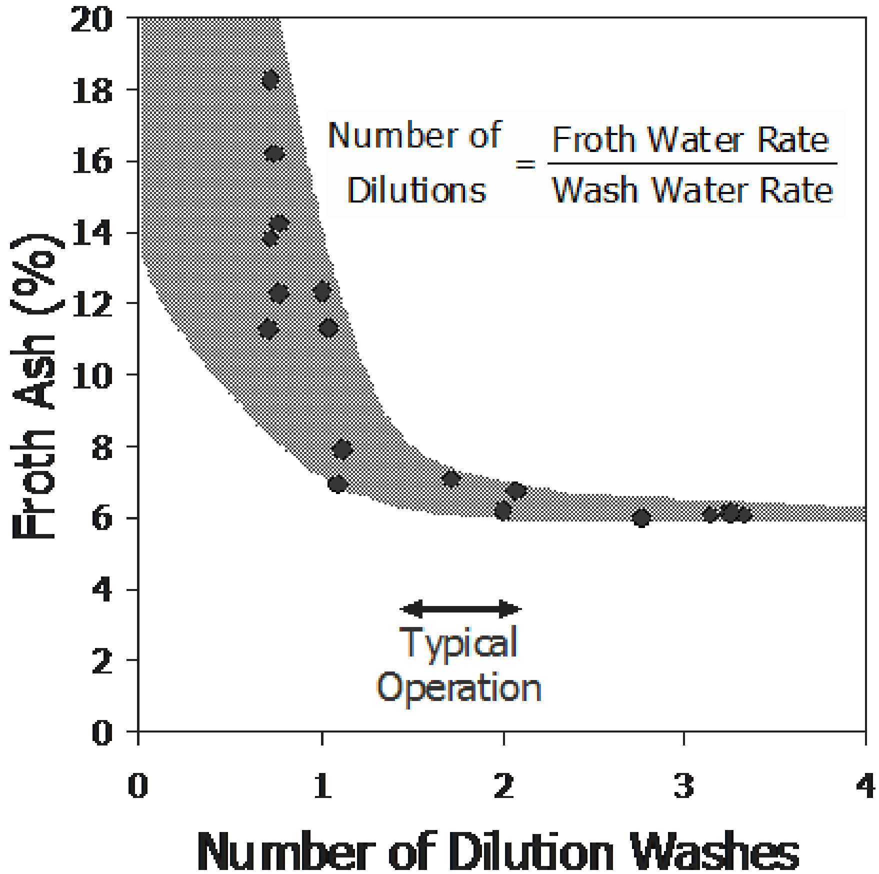

Figure 8. Effect of number of dilution washes on froth quality (Luttrell et al., Proceedings of the 21st Annual International Coal, Aggregrate and Mineal Processing Exhibition and Conference, 2004).

recovery. Very high water additions may also destabilize the froth by stripping surfactant (frother) from the bubble surfaces. High water rates may also have a detrimental impact on product grade by increasing axial froth mixing and reducing the wash water effectiveness (Yianatos et al., 1988).

The design of the wash water distributor can also significantly affect column performance. In some cases, the distribution piping is intentionally submerged below the cell dip so that a drained froth can form above the distributor. This arrangement allows the depth of the drained froth and the extent of froth drainage to be varied by raising or lowering the distributor. Changes to the vertical position of the water distributor can be used to somewhat control the split of water between the clean coal and refuse streams. In some cases, multi-level concentric distribution rings may also be used to overcome problems associated with poor froth mobility. The inner rings are typically located above the outer rings to reduce drainage and improve the fluidity of the froth in the center of the column. In other cases, the water distributor may be located just above the top of the froth. This arrangement does not allow the froth mobility to be controlled by adjusting the distributor location, but it does make if easier to identify and correct plugging problems that may severely impact the performance of the distribution network.

2.2.5 Gas Sparging

The air sparging system is the most important, and perhaps most controversial, component in the design of a coal flotation column. Ideally, the spargers should produce small, uniformly sized bubbles at a desired aeration rate. The spargers should also be non-plugging, wear-resistant and allow for easy, on-line servicing. Three primary types of column sparging systems are used:

- porous bubblers fabricated from filter cloth, punctured rubber tubes or sintered metals/plastics (e.g., Ken-Flote and many "homemade" columns),
- high-velocity air or air-water injectors (e.g. EIMCO, CPT/Eriez and Minnovex columns), and
- circulating systems incorporating inline or external agitators such as static mixers (e.g., Microcel column).

Details related to these systems have been discussed extensively in the technical literature (Dobby and Finch, 1986b; Xu and Finch, 1989; Huls et al., 1991; Yoon et al., 1992; Groppo and Parekh, 1992; Finch, 1995).

Figure 9 shows test results obtained from a laboratory comparison of three popular column spargers. In general, the data for all three systems tended to fall along nearly identical separation curves. This suggests that the selectivity of the separation is largely determined by the bubble-particle attachment process and is relatively independent of the method used to produce the air bubbles. Similar data have been

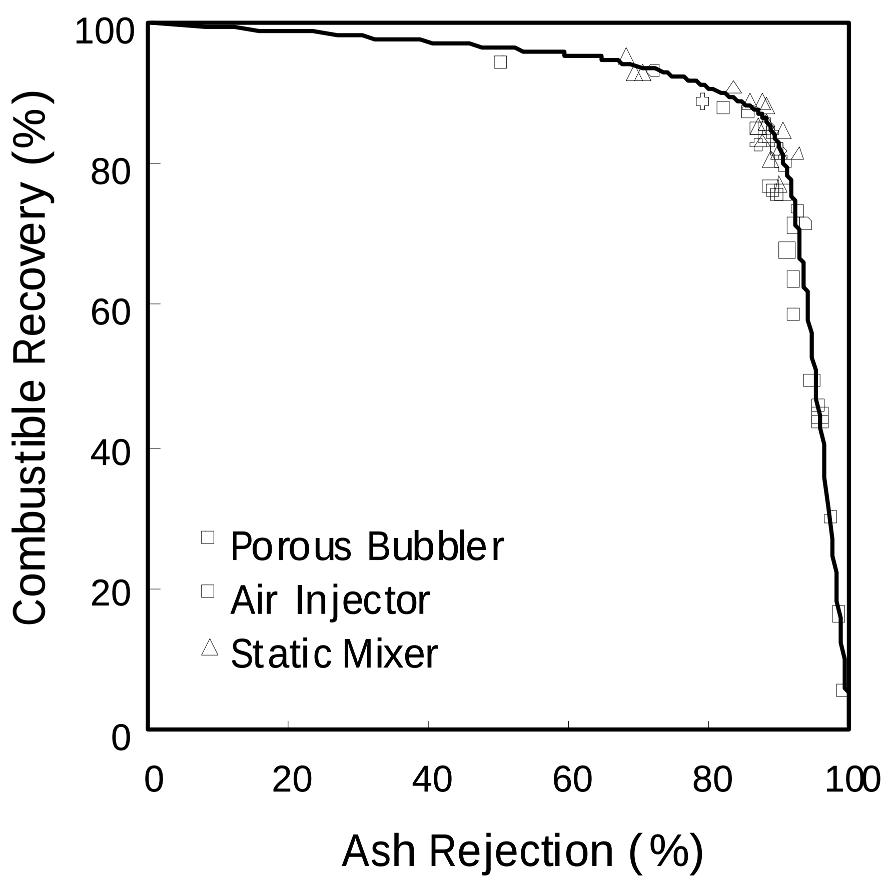

Figure 9. Laboratory comparison of three column sparging systems (Kohmuench, Precombustion of Hazardous Air Pollutant Precursors, Ph.D. Dissertation, Virginia Tech, 1998).

reported by Honaker et al. (1995). This statement does not, however, imply that sparger selection is unimportant. The test data indicate that the porous bubbler and static mixer generally produced higher overall recoveries than the high-velocity air injector. In addition, the static mixer was generally less sensitive to changes in operating conditions and tended to produce data points higher on the separation curve. On the other hand, systems such as the porous bubbler have been shown to plug easily and require frequent maintenance in industrial operations. Capital costs for the static mixer system are generally higher due to the purchase of a slurry pump and wear can become an issue when very coarse feeds (>100 mesh) are treated. All these factors need to be carefully considered prior to selecting a sparging system.

2.2.6 Aeration Rate

Differences in spargers can normally be attributed to variations in the amount of bubble surface area generated by each device. This capability is commonly reported in terms of the superficial bubble surface area rate (S_b), which is defined as the total bubble surface area per unit of time passing through a given column cross-sectional area. This value can be calculated by:

$$S_b = 30.5\, V_g / D_b \qquad [4]$$

in which V_g is the superficial gas rate (cfm/ft^2) and D_b is the bubble diameter (mm). The superficial air rate may be calculated by dividing the volumetric gas flow rate (cfm) by the column cross sectional area (ft^2). The impact of S_b on flotation recovery is illustrated by the test results given in Figure 10. These data show that recovery increases sharply as S_b increases above 50 sec^{-1} and eventually reaches a plateau at 100-150 sec^{-1}. These

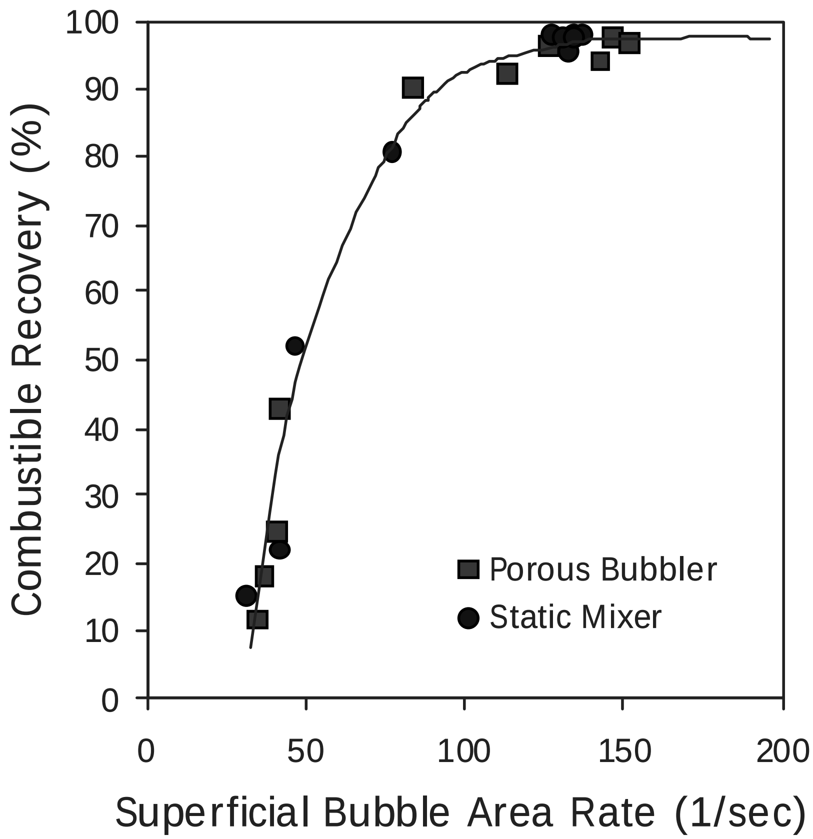

Figure 10. Effect of S_b on flotation recovery for two different air sparging systems (Kohmuench, Personal Communication, 2000).

values indicate that nearly 1 million square feet bubble surface area passes through a 14 ft diameter column every minute! In addition, the data indicate that the relationship between recovery and S_b is generally independent of sparger type. This finding implies that most commercial sparging systems can achieve a target recovery provided that a proper balance is established between gas rate and bubble size.

It should also be noted that operation at S_b values above 150 sec^{-1} often produces poor separations because of “runaway” froths that cannot be effectively washed. In most industrial cases, a superficial gas rate in the range of 3.5-5.0 cfm/ft^2 would be suitable for coal applications. These values correspond to total aeration rates of 540-770 scfm of air for a full-scale 14 ft diameter column. The gas rates at the lower end of the range would generally be used for spargers that generate smaller bubbles, while the higher gas rates are typically needed for less efficient spargers. A proper combination of gas rate and bubble size will generally provide a gas holdup in the flotation pulp in the range of 15-18%. The holdup can be determined by installing pressure transducers at two different levels along the height of the flotation column. A low air holdup indicates that the production capabilities of the column are being underutilized.

Caution should also be used during the metering of gas flow rates. A properly designed system should be equipped with a flow meter that is calibrated to read correctly at a specified operating pressure. The operating pressure should be held constant by placing a pressure regulator ahead of the flow meter. By placing the air flow control valve after the flow meter, the flow meter will always operate at its design pressure. If the flow meter is placed after the control valve, then the operating pressure and true gas flow rate are both unknown. Improper metering of the gas flow rate can be a particularly serious problem when laboratory and pilot-scale tests are conducted for the purpose of collecting scale-up information.

A great deal of confusion also exists regarding the specification of compressors for column applications. In fact, several installations of columns in the coal industry have required the purchase of additional compressors to reach the original design flow rate. Much of this confusion is related to improper use of gas flow terminology (Sullair Bulletin, 1992). For example, column manufacturers normally report gas flow rates as a "standard" volumetric flow per time. This value is only valid at 1 atm (14.7 psia) of pressure and 20^{o}C (68^{o}F) of dry air. The “actual” flow rate specified by compressor manufacturers is typically reported in terms of “inlet” conditions or “free air.” Although this amount of air enters the compressor, it is not necessarily the amount of air delivered to the column due to compressor seal leakage. As a result, the “actual” flow may be only 95% of the “inlet” flow. Furthermore, corrections to the gas flow rate must be made to account for differences in elevation (atmospheric pressure) and humidity. Air temperature generally has little impact on the capacity of an oil-flooded screw compressor, but may affect the performance of an air-cooled compressor. These complications generally require that professionals be consulted to ensure that the compressor is properly sized for the specified air requirements. This item is not an issue with the Jameson cell that requires no compressed air.

2.2.7 Froth Handling

Froth handling is a major problem associated with coal flotation columns. Concentrates containing large amounts of ultrafine (325 mesh x 0) coal generally become excessively stable, creating serious problems related to backup in launders and downstream handling. Attempts to overcome this problem by selecting weaker frothers or reducing frother dosage have not been successful and generally lead to lower column recoveries. Therefore, several circuit modifications have been adopted by the coal industry to deal with the froth stability problem. For example, column launders need to be considerably oversized with steep slopes to reduce backup. Horizontal froth travel distances must be kept as short as possible, while adequate vertical head must be provided between downstream operations and the column launder. Most of the newer column installations include a deaeration tank to permit time for the froth to collapse (see

Figure 11). Special provisions are also normally required to ensure that downstream dewatering units can accept the large froth volumes. Standard screen-bowl centrifuges equipped with 4 inch inlets may need to be retrofitted with 6 inch or larger inlets to minimize flow restrictions. Some installations may require defoaming agents to deal with the froth stability problem. These reagents can represent a large operating cost and must be included in any detailed economic analysis.

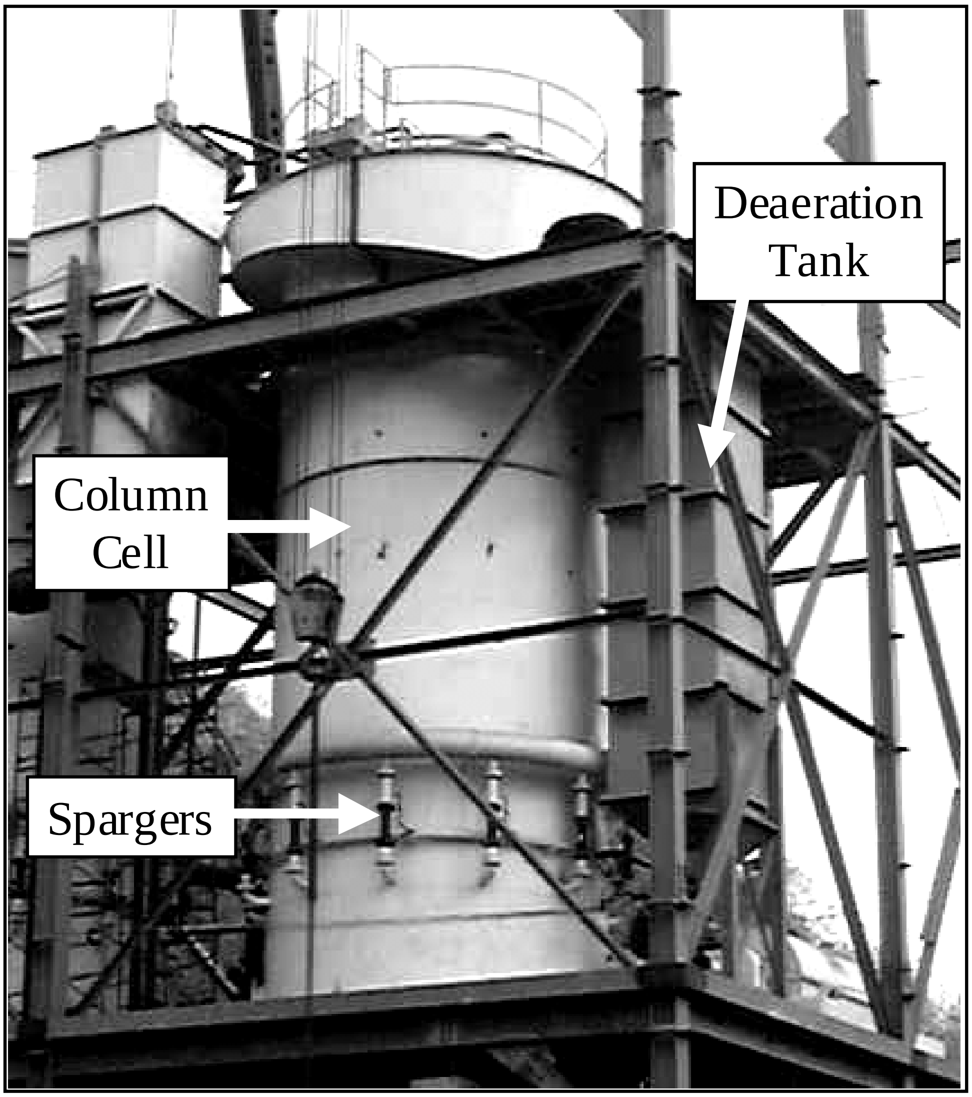

Figure 11. Column with deaeration tank used to control froth handling problems.

2.3 Residence Time Modeling

2.3.1 Mean Residence Time

A considerable period of contacting time is necessary in flotation systems to ensure that all coal particles have the opportunity to collide and adhere to air bubbles. This contacting time is commonly referred to as the flotation time or residence time. As a rule-of-thumb, a mean residence time of at least 3½-4 minutes is typically required in conventional flotation machines to achieve good recoveries of well-floating bituminous coals. The required residence time may be even longer for difficult-to-float coals that are very fine or superficially oxidized. The mean residence time (τ) can be estimated by dividing the active cell volume (V) by the volumetric flow rate (Q) of slurry passing through the cells, i.e.:

$$\tau = V / Q \qquad [10]$$

Note that the active volume of the cell discounts the volume of the impeller mechanism and any hold-up of gas due to rising air bubbles. For example, a bank of four 14.2 m^3 machines (85% active volume) fed 817.6 m^3/hr of slurry would have $\tau \approx 3.5$ minutes (4 cells x 14.2 m^3/cell x 0.85 / 817.6 m^3/hr x 1 hr/60 min = 3.5 minutes). While this rule-of-thumb is useful for a first-pass evaluation, the specification of a minimum residence time can be very misleading in some cases.

The mean residence time represents the average length of time that it takes an element of fluid to travel through the flotation cell. In a plug-flow or batch system, all fluid elements are exposed to the same residence time (Figure 12). Likewise, in a perfectly-mixed system, some elements begin to exit immediately after being introduced due to the perfect dispersion. However, in a continuous bank containing partially mixed or more than one agitated tank, a wide distribution of residence times may exist for each element of fluid entering the system. The differences in this dispersed-flow system are caused by fluid channeling, short-circuiting, stagnant regions, or other conditions that allow some of the flotation pulp to exit the machine at a time different from the mean residence time.

Because of differences in fluid dispersion, a wide range of flotation recoveries may be obtained for the same mean residence time (Finch and Dobby, 1990). This important interaction can be mathematically described by an equation originally derived by Wehner and Wilhelm (1956), which states that the total recovery (R) for any first-order process is given by:

$$R = 1 - \frac{4A\exp\{Pe/2\}}{(1+A)^2\exp\{(A/2)Pe\} - (1-A)^2\exp\{(-A/2)Pe\}} \qquad [11]$$

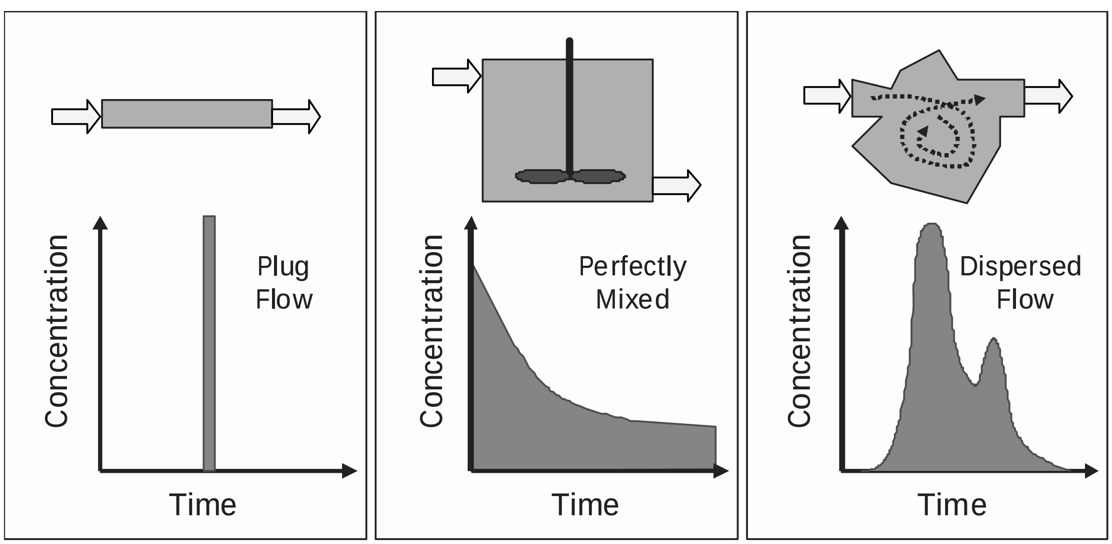

Figure 12. Passage of a fluid through plug-flow, perfectly-mixed and dispersed-flow systems.

$$A = \sqrt{1 + 4k\tau / Pe} \qquad [12]$$

where τ is the mean residence time, k is the flotation rate constant, and Pe is the Peclet number. For flotation systems, the terms τ and k are usually expressed in units of minutes and 1/minutes, respectively. As a result, Equation [11] can be plotted as a function of two dimensionless terms (kτ and Pe) as shown in Figure 13.

The product kτ represents the degree of completion of the kinetic process, while parameter Pe represents the intensity of dispersive mixing. The operating band shown in Figure 13 is bounded by the upper plug-flow (Pe=∞) curve and the lower perfectly-mixed (Pe=0) curve. While a plug-flow system is obviously more desirable since it provides a much higher recovery than a perfectly-mixed system at the same kτ, plug-flow conditions

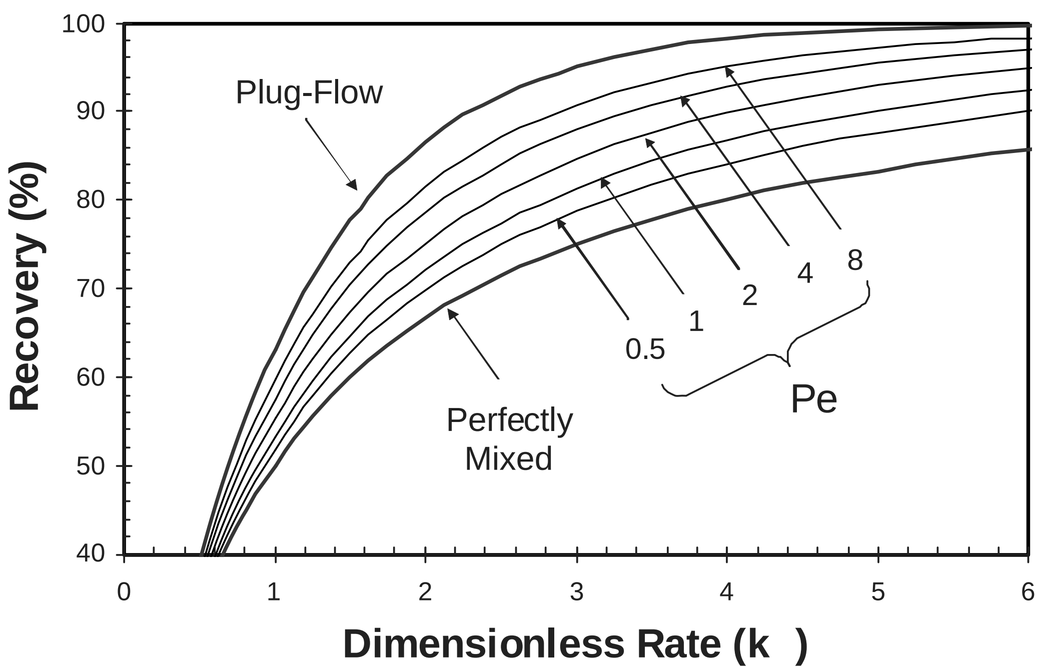

Figure 13. Recovery as a function of rate (k), residence time (τ) and Peclet number (Pe).

are difficult to duplicate in industrial practice. As a result, most commercial flotation machines are designed with multiple agitated tanks arranged in a cell-to-cell bank in an attempt to approach plug-flow behavior. A simple calculation shows that the total recovery (R_N) for a bank of N tanks in series is:

$$R_N = R_i + R_i(1-R_i) + R_i(1-R_i)^2 + R_i(1-R_i)^3 + ... + R_i(1-R_i)^N = 1-(1-R_i)^N \quad [6]$$

where R_i is the fraction recovery defined at $\tau_i = \tau/N$. For a bank of perfectly-mixed cells, Equations (2)-(4) can be combined and mathematically simplified to yield:

$$R_N = 1 - \left(\frac{N}{N + k\tau}\right)^N \quad [13]$$

As expected, the performance curve provided by this equation approaches that of a plug-flow system as N approaches infinity. However, in commercial practice, banks are typically limited to four or five tanks in series since additional cells provide little incremental improvement in recovery compared to the increased cost of purchasing additional cells within the bank.

2.3.2 Residence Time Distributions

Unfortunately, field studies indicate that the performance curve described by Equation [11] is often inadequate to describe the kinetic response of individual flotation cells. Instead, most industrial machines operate with $0<Pe<\infty$. Therefore, Pe must be determined to establish the appropriate recovery versus $k\tau$ curve for industrial flotation systems. Pe can be estimated from an experimental residence time distribution (RTD) curve. The RTD is obtained by:

(i) adding a tracer substance to tag an incoming element of fluid to a flotation machine and

(ii) monitoring the concentration of the tracer in the tailings as a function of time.

An example of RTD data for a four cell bank of conventional cells is shown in Table 2. The mean residence time (τ) can be calculated from these data using (Levenspiel, 1972):

$$\tau = \frac{\sum tC\Delta t}{\sum C\Delta t} \qquad [14]$$

where C is the tracer concentration at any time t and Δt is the time increment between samples [$\Delta t = (t_{i+1}+t_{i-1})/2$]. This procedure gives τ=3.47 minutes for the data provided in Table 3. Pe is related to the spread of the RTD, which can be quantified by the

Table 3. Example of RTD data for a four-cell bank of conventional machines.

Time (min)	Δt (min)	C (ppm)	tCΔt (ppm-min^2)	CΔt (ppm-min)	t^2CΔt (ppm-min^3)
0	--	0	0	0	0
1	1	480	480	480	480
2	1	780	1560	780	3120
3	1	740	2220	740	6660
4	1	560	2240	560	8960
5	1.5	370	2775	555	13875
7	2.5	130	2275	325	15925
10	4	15	600	60	6000
15	---	0	0	0	0
			∑=12,150	∑=3,500	∑=55,020

$$\tau = \frac{\sum tC\Delta t}{\sum C\Delta t} = \frac{12{,}150}{3{,}500} = 3.47 \text{ min} \qquad \sigma_\theta^2 = \frac{\sum t^2 C\Delta t}{\tau^2 \sum C\Delta t} - 1 = \frac{55{,}020}{3.47^2(3{,}500)} - 1 = 0.304$$

dimensionless variance (σ_θ^2) given by (Levenspiel, 1972):

$$\sigma_\theta^2 = \frac{\sum t^2 C\Delta t}{\tau^2 \sum C\Delta t} - 1 = 2\left(\frac{1}{Pe} - \frac{1-e^{-Pe}}{Pe^2}\right) \qquad [15]$$

Due to the complexity of this equation, it must be solved either graphically or by numerical iteration. For convenience, the graphical solution is plotted in Figure 14 for the range of Pe values typically observed in flotation. By inspection, Pe=5.4 for σ_θ^2=0.304 determined from the RTD data in Table 3. Since Pe is known, Equation [11] can now be used to define the recovery curve for this flotation system.

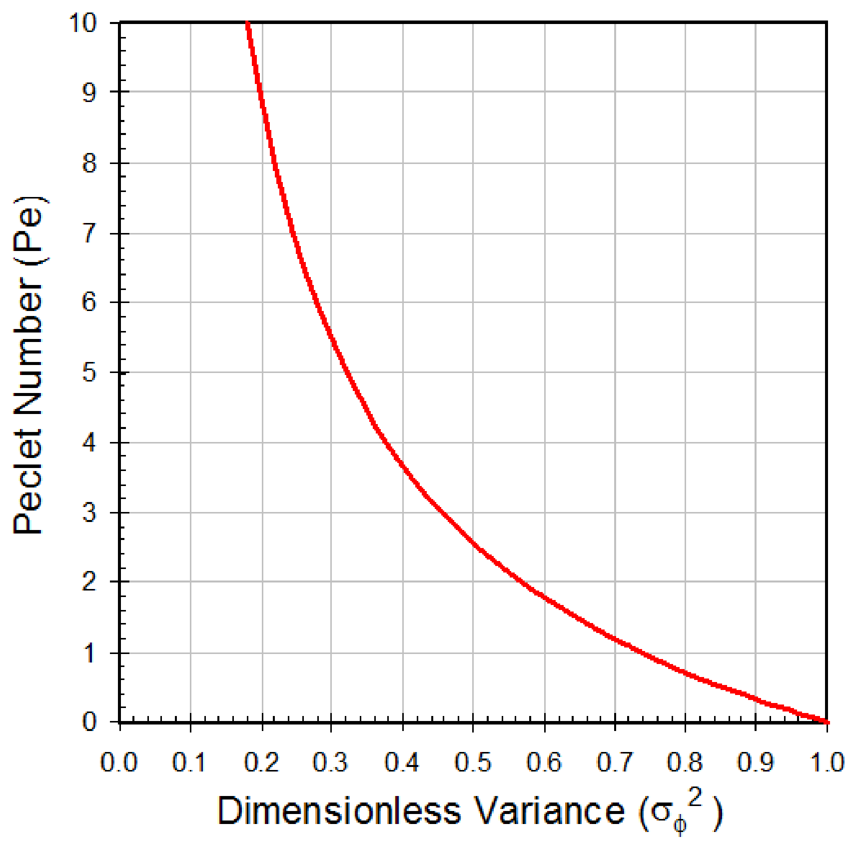

Figure 14. Graphical solution used to find the Peclet Number from RTD data.

3.0 EXPERIMENTAL RESULTS AND DISCUSSION

3.1 Dye Mixing Studies

Conventional flotation machines are typically arranged in series with the tailings from each cell providing feed to the next. This multi-stage arrangement makes it possible for conventional machines to maintain good coal recoveries even at relatively short residence times. Column cells, on the other hand, are usually fed in parallel and operate independently from each other. This layout dictates substantially more residence time be provided for columns to maintain good recoveries of floatable particles (Nicol and Bensley, 1988). However, even at long residence times, the lack of downstream scavenging creates a potential for feed to bypass to tailings in a relatively short time.

In order to qualitatively assess the severity of mixing and short-circuiting in a column flotation cell, a series of tests were conducted using colored dye (dark red) to visually observe the mixing behavior of fluid passing through a clear Plexiglas flotation cell. The 30-inch diameter cell was filled with fluid to a height of 60 inches to provide a length-to-diameter of 2:1, which is geometrically similar to many industrial column cells. The flow rate of gas to the cell, which was equipped with a Microcel™ sparging, was held constant at 2.0 cm/sec for all tests. A feed flow rate equivalent to a superficial velocity of 0.25 cm/sec was maintained by injecting fresh water through a 4-way feed distributor placed near the top of the liquid level. This feed flow rate provided a total volumetric residence time of approximately 10 minutes. However, when corrected for air hold-up, the mean residence was probably in the range of 7-8 minutes. The transport of the dye through the cell was observed at one-second intervals using a digital color camera equipped with a 16-frame timed-sequence image grabber.

Figure 15 shows the results obtained from the first test conducted using the tracer dye. The first image (at 1 sec) is located in the top-right corner of the composite photograph, while the last image (at 16 secs) is located at the bottom-left corner. The

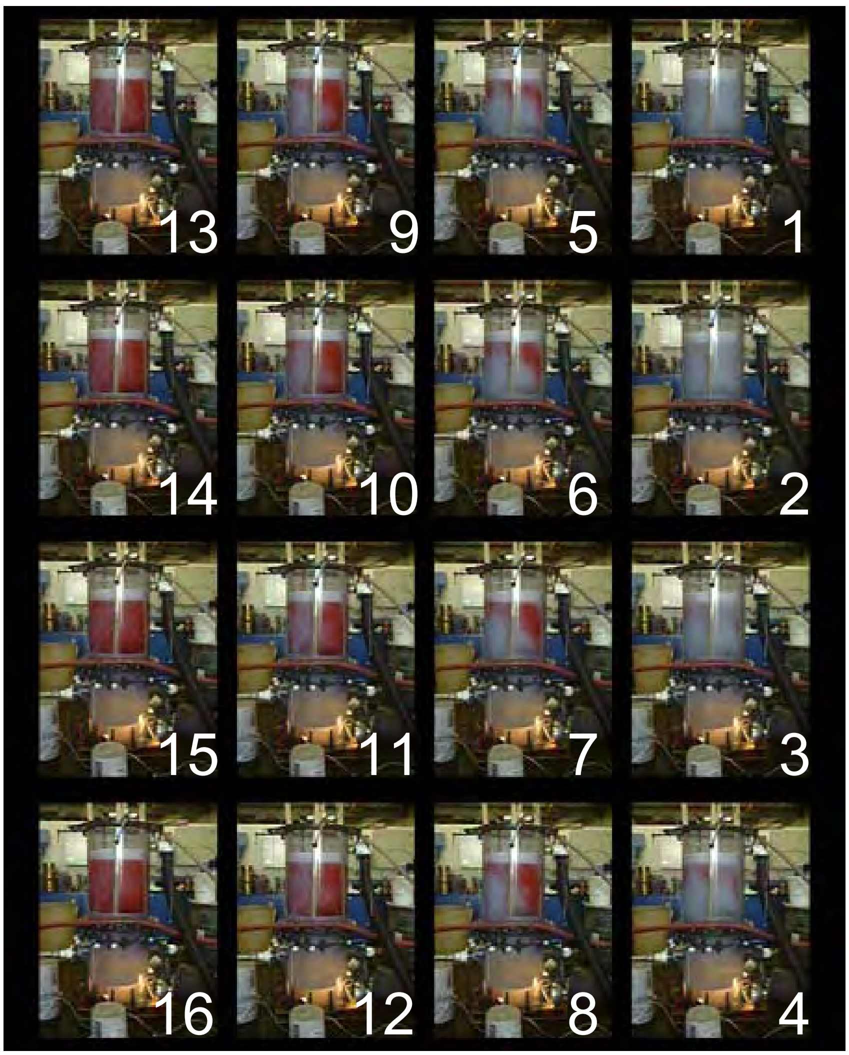

Figure 15. Visualization of column mixing using a red dye and timed-sequence images at one-second intervals.

timed sequence of images shows that the dye is first visible through the column wall at about 4 seconds. Surprisingly, the dye appeared to be thoroughly mixed throughout the entire cell volume after only about 13-14 seconds. This finding suggests that column cells operated under similar conditions are likely to be well-mixed as opposed to plug-flow. Contributors to the high degree of internal mixing are likely to include the small length-to-diameter ratio and the agitation and circulation provided by the sparger pump.

Another series of tests were conducted using a column cell equipped with baffles designed to minimize mixing. The baffles consisted of four horizontal partitions that split the cross-sectional of the column into four "pie" slices of equal area. The baffles passed from just below the top of the liquid level, down pass the the sparger inlets, and terminated just above the central tailings discharge. Unfortunately, the test data showed that the dye appeared to be mixed throughout the cell volume after only about 15-16 seconds (Figure 16). In addition, the data showed that the partitions tended to provide unequal distribution of the feed in each quadrant, which would be an undesirable result for industrial operations. Therefore, the qualitative data obtained in this study suggests that this type of baffling cannot effectively minimize large internal mixing.

3.2 RTD Measurements

In order to quantitatively establish the extent of mixing in an industrial column, an RTD measurement was conducted at an industrial column flotation plant. The resultant RTD curve is plotted in Figure 17. The RTD calculations showed that the single column operated with a relatively long residence time of τ=12.0 minutes. However, the calculations showed that the column was relatively well mixed, as indicated by the low

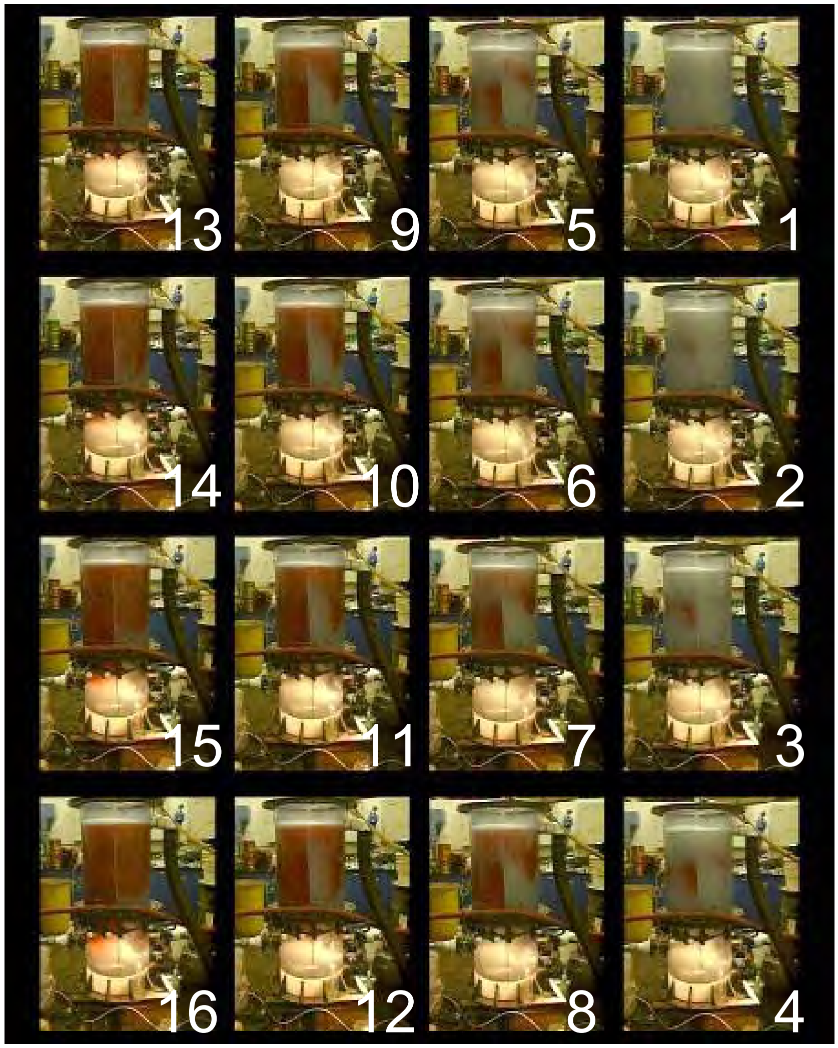

Figure 16. Visualization of column mixing using a red dye and timed-sequence images (column equipped with baffles) at one-second intervals.

Pe=1.58. In fact, the tracer distribution suggests that nearly 10% of the feed slurry exited the cell after just one minute. Therefore, despite the long residence time, the columns achieved a recovery of only 73.7% (two-run average) due to the large dispersion.

One method of improving the performance of the column installation would be to reconfigure the cells to operate in series as opposed to the parallel feeding arrangement currently in use. As shown in Figure 18, the operating point defined by Pe=1.58 and

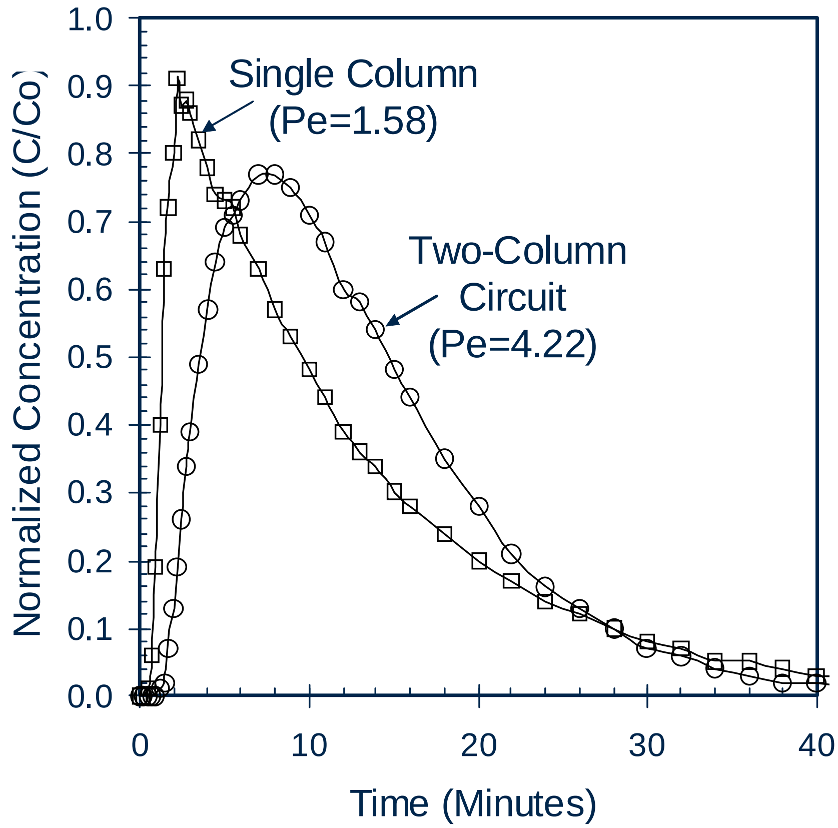

Figure 17. Comparison of RTD data for single- and two-stage column circuits.

R_i=73.7% corresponds to a dimensionless rate of $k\tau_i$=1.97 for the current single-cell configuration. Two cells connected in series would reduce the dimensionless rate to $k\tau_i$=1.97/2=0.99, which corresponds to R_i=54.5% according to Equation [11] and Figure 17. The combined recovery (R_2) for the two cells in series can be estimated using Equation [13] as:

$$R_2 = 1-(1-R_i)^2 = 1-(1-0.545)^2 = 79.3\% \quad [16]$$

Likewise, $k\tau_i$=1.97/3=0.66 for three columns in series corresponds to R_i=42.7%. The projected total recovery for three columns (R_3) is:

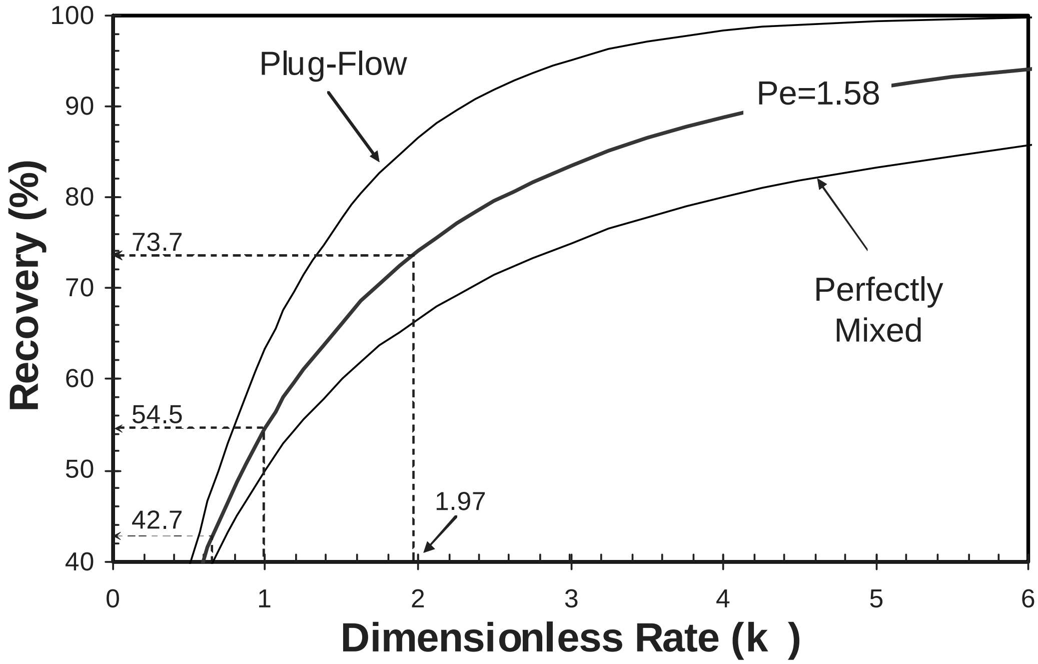

Figure 18. Recovery curve for projecting two- and three-stage column circuit performance...

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$$R_3 = 1-(1-R_i)^3 = 1-(1-0.427)^3 = 81.2\% \qquad [17]$$

The use of more than three columns may not be practical since bubbles may not be able to rise against the increased downward flow rate passing through each column. It should be noted that, in practice, the projected improvements are likely to slightly different from these values since the reconfiguration may result in higher flow rates that could change the dispersion in each cell.

3.3 Circuit Reconfiguration

In light of the potential improvements afforded by a cell-to-cell layout, the piping in the column plant was reconfigured so that (i) the feed slurry to columns A and B was entirely diverted to column A and (ii) the tailings stream from column A was fed to column B. The columns were sampled before and after the modification. RTD measurements were also taken for the modified two-stage circuit. The performance calculations and experimental results for these tests are summarized in Table 4. As expected, the modification reduced axial dispersion by increasing Pe from 1.58 to 4.22 (see Figure 17). The modification did not substantially change the mean residence time since twice the feed flow passed through twice the cell volume after the modification. However, the change in feeding configuration shifted the performance for the overall circuit from a recovery of 73.7% (two-run average) for the single column to 79.8% (two-run average) for the two-column bank. This improvement agrees very well with the 79.3% recovery projected from the series flow calculation given in Equation [16].

The results obtained from the trial run were promising enough to move ahead with the reconfiguration of the entire five column circuit to a cell-to-cell system. In order

Table 4. Example of RTD data for single- and two-stage column circuits.

Circuit Type	Feed Ash (%)	Clean Ash (%)	Tails Ash (%)	Yield (%)	Recovery (%)
Single-Stage Circuit (τ=11.9 min, Pe= 1.58)					
Run #1	54.83	15.02	79.84	38.58	72.59
Run #2	52.45	13.86	79.59	41.29	74.80
Average	**53.64**	**14.44**	**79.72**	**39.94**	**73.69**
Two-Stage Circuit (τ=12.9 min, Pe= 4.22)					
Run 1 - Column 1	56.55	15.94	83.21	39.63	76.67
Run 1 - Column 2	83.21	17.87	84.25	1.57	7.66
Run 1 - Column 1&2	56.55	15.98	84.25	40.58	78.46
Run 2 - Column 1	51.39	18.28	77.27	43.87	73.75
Run 2 - Column 2	77.27	18.78	82.23	7.82	27.93
Run 2 - Column 1&2	51.39	18.33	82.23	48.26	81.09
Average (Column 1&2)	**53.97**	**17.16**	**83.24**	**44.42**	**79.77**

to avoid excessive downward velocities in the columns, the layout was modified so that the feed was split between the two end columns in the five cell bank and directed inward to the center cell (Figure 19). Note that, due to piping constraints, a small amount of feed also had to be bypassed directly to columns #2 and #4.

Daily production data obtained before and after the circuit modifications are plotted in Figure 20. Performance indicators such as average recovery and ash values are summarized in Table 5. The daily values show substantial data scatter due to the large variations in the feed quality from the pond reclaim operation that feeds the columns. For this reason, weekly averages have been included on the plot so that the production trends

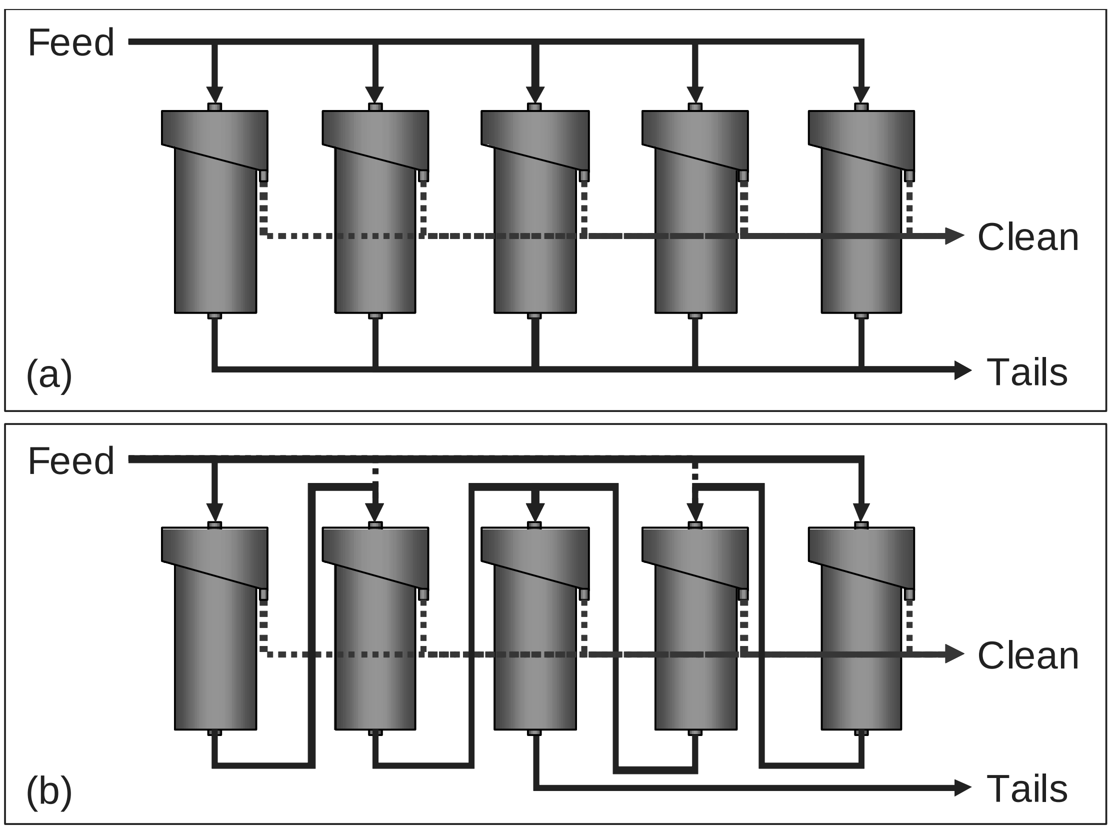

Figure 19. Column circuit layout (a) before and (b) after cell-to-cell reconfiguration.

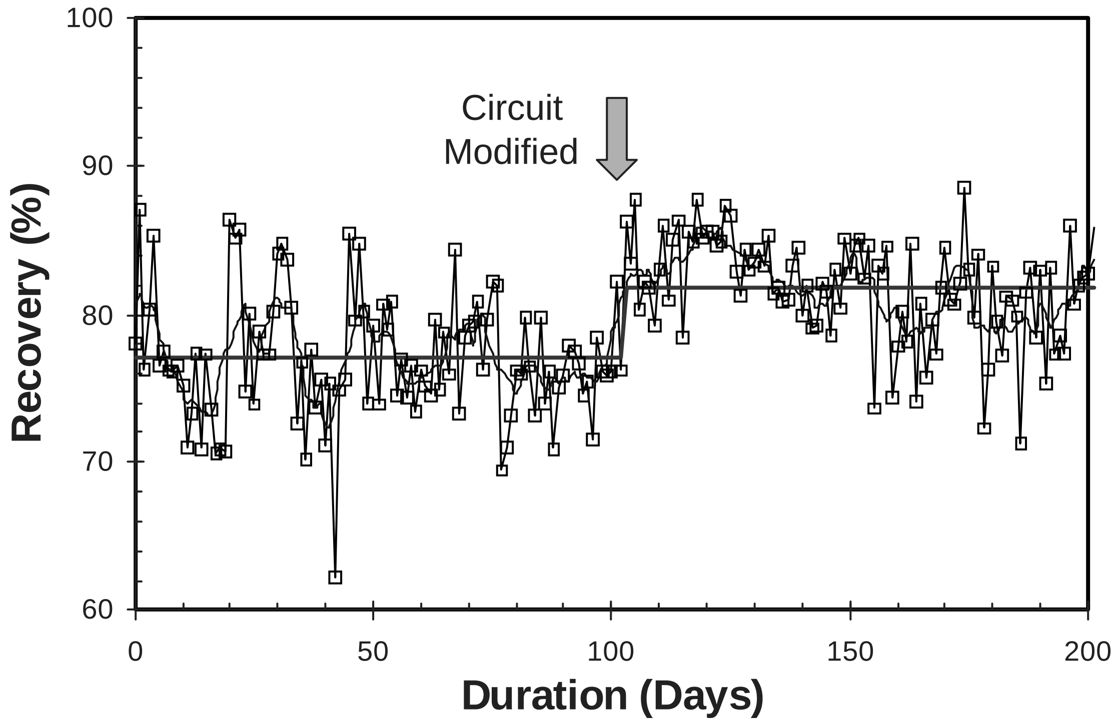

Figure 20. Production statistics before and after the cell-to-cell column reconfiguration.

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can be more clearly visualized (thick solid line). Prior to the modification, the columns produced an average recovery of 77.0% with a standard deviation of 4.3%. After the change, the recovery improved to an average value of 81.9%, with a slightly smaller standard deviation of 3.5%. As such, recovery improved by 4.9 percentage points. This corresponds to a production increase valued at >$300,000 US annually for this particular plant site.

Table 5. Production statistics before and after the full cell-to-cell column reconfiguration.

	Feed Ash (%)	Clean Ash (%)	Tailing Ash (%)	Yield (%)	Recovery (%)
BEFORE					
Average	48.76	16.31	77.78	47.39	**77.00**
Maximum	55.11	37.38	83.25	67.95	87.10
Minimum	41.13	9.31	68.88	37.13	62.25
Std Dev	3.02	3.95	2.24	5.99	4.26
AFTER					
Average	49.72	15.26	82.37	48.82	**81.85**
Maximum	57.07	29.71	85.28	63.41	88.65
Minimum	42.80	9.54	78.42	34.51	71.26
Std Dev	3.13	4.46	1.23	5.56	3.48

4.0 SUMMARY AND CONCLUSIONS

1. Column flotation cells can produce a higher overall separation performance than conventional flotation machines when processing high-ash coal feeds. If properly operated, the separation performance achieved by columns approaches the theoretical maximum cleanability predicted by release analysis.

2. The improved separation performance of flotation columns can be largely attributed to froth washing which minimizes the nonselective hydraulic entrainment of ultrafine slimes into the froth product. Wash water rates in the range of 3.0-3.5 gpm/ft^2 are typically required to maintain the proper number of dilution washes needed to prevent entrainment.

3. The capacity of industrial coal flotation columns is normally controlled by limitations associated with froth carrying capacity. This constraint, which is very sensitive to variations in particle size, establishes the total column cross-sectional area (column number and diameter) required for a given application. Carrying capacities commonly fall in the range of range of 0.06 to 0.24 tph/ft^2, with an average of about 0.12 tph/ft^2 for 100 mesh x 0 feed coals. The lower carrying capacities typically correspond to finer feeds (325 mesh x 0).

4. Data obtained to date suggest that the selectivity of coal columns is largely independent of column size and sparger design. The spargers must be capable of dispersing air into small bubbles at gas rates typically exceeding 3.5 scfm/ft^2 in order

to maintain acceptable coal recoveries and production rates. Sparger performance can be measured by the parameter S_b, which represents the amount of bubble surface area generated per unit time for a given column cross-sectional area. For optimum performance, S_b values greater than approximately 80-100 sec^{-1} should be maintained.

5. Column cells can have both positive and negative impacts on downstream operations in coal plants. For example, the ability of columns to removal of clay slimes and reduce total ash content can lead to a significant increase in the filtration rate and a lowering of the equilibrium moisture content of the clean coal filter cake. On the other hand, difficulties associated with the increased stability of column froths may create serious material handling problems and require the installation of deaeration tanks or possibly the use of expensive defoaming agents. These factors must be fully addressed prior to completing the design of any industrial circuit.

6. Residence time distribution (RTD) measurements conducted at a coal flotation plant indicates that the poor recovery performance observed in some column flotation installations may be partly due to the inadvertent short-circuiting of feed to tailings cause by strong axial mixing. Mathematical calculations using a first-order kinetic model indicate that this inherent shortcoming can be minimized by reconfiguring banks of column cells to operate as a cell-to-cell circuit as opposed to the parallel feeding arrangement commonly used by column manufacturers.

7. The feed piping at a coal column flotation plant was modified from a parallel to a series layout. The modification improved the recovery by nearly 5 percentage points,

as predicted by the mathematical projections based on residence time distributions. This improvement increased the profitability of the column circuit at this particular site by approximately $300,000 US annually. This practice is now recommended for all column flotation circuits treating fine coal.

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6.0 APPENDIX

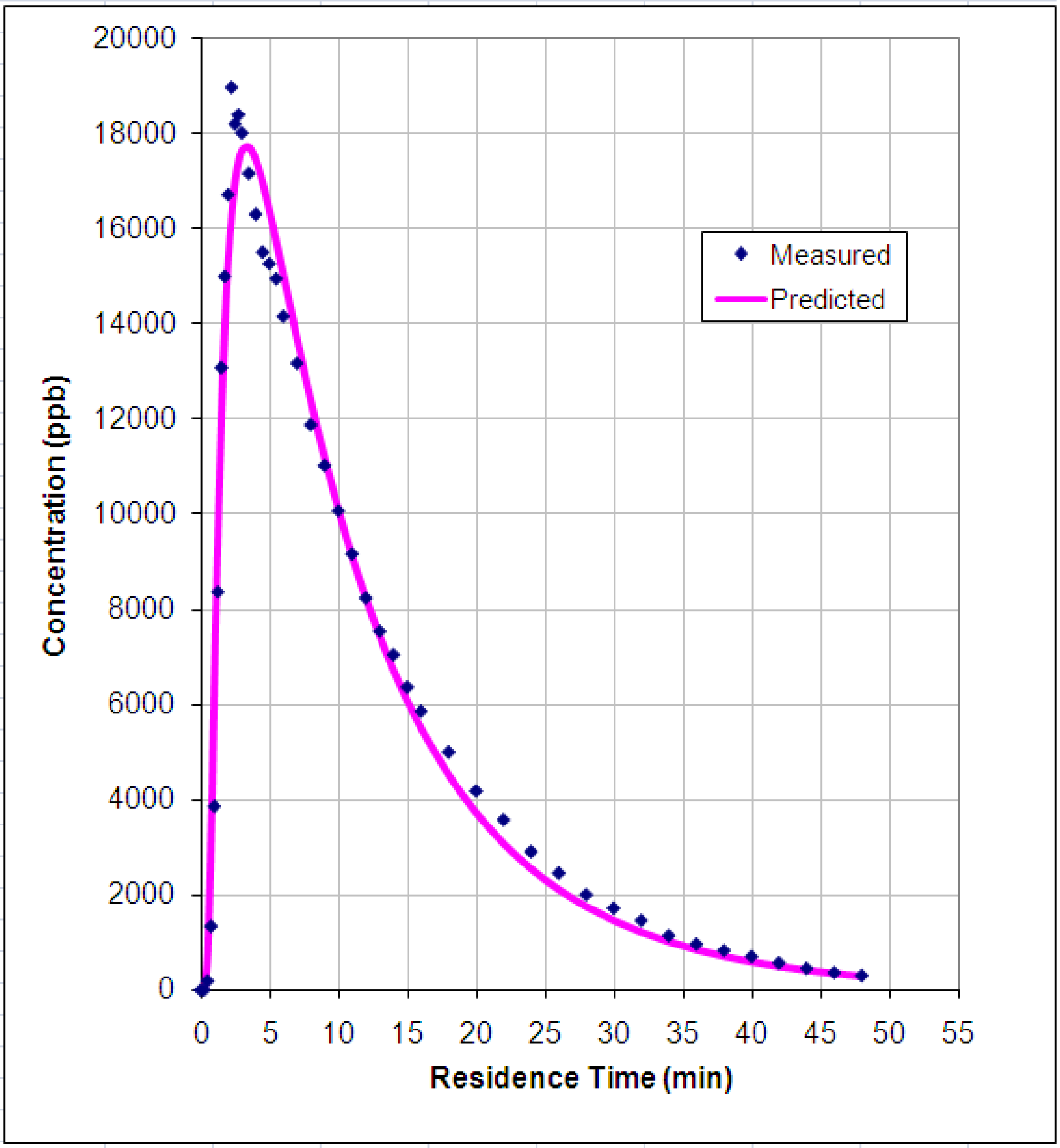
Appendix

RTD Data for Single Column

Dispersion Calculations	Measured	Fitted	Units
Baseline Concentration	0.00	0.00	ppb
Mean Retention Time	**12.00**	11.45	min
Variance (Required)	90.57	85.24	min^2
Variance (Calculated)	90.57	85.24	min^2
Peclet (Reset)			(Delete to run!)
Peclet (Calculated)	**1.59**	1.46	---

Curve Fit Parameters	
Parameter #1	5.69
Parameter #2	1.76
Parameter #3	41282

Performance Calculations		
Active Volume	11529	gal
Estimated Flow Rate	**960**	gpm



		Measured Data				Curve Fit Data			
Time	Conc'n	Conc'n	Calculated	Calculated	Calculated	Conc'n	Calculated	Calculated	Calculated
(min)	(ppb)	(ppb)	t(C)(dt)	(C)(dt)	t^2(C)(dt)	(ppb)	t(C)(dt)	(C)(dt)	t^2(C)(dt)
0.00	0	0	0	0	0	0.0	0	0	0
0.25	0	0	0	0	0	7.7	0	1	0
0.50	193	193	9	24	3	804.8	38	102	14
0.75	1343	1343	120	192	75	3427.9	331	529	207
1.00	3868	3868	570	651	499	6715.0	1109	1268	971
1.25	8379	8379	1722	1531	1938	9725.8	2312	2055	2601
1.50	13100	13100	3692	2685	5076	12164.3	3762	2736	5173
1.75	15020	15020	5712	3515	9282	14024.7	5320	3274	8644
2.00	16740	16740	7444	3970	13957	15390.7	6894	3677	12927
2.25	19000	19000	9493	4468	20174	16359.2	8434	3969	17921
2.50	18230	18230	11053	4654	26250	17016.3	9908	4172	23532
2.75	18420	18420	12026	4581	31568	17432.2	11303	4306	29671
3.00	18040	18040	13103	4558	37671	17662.2	12612	4387	36260
3.50	17190	17190	28624	8808	93029	17725.9	28753	8847	93447
4.00	16330	16330	31425	8380	117844	17445.6	32973	8793	123650
4.50	15530	15530	33851	7965	143868	16964.6	36561	8603	155384
5.00	15290	15290	36599	7705	173844	16370.1	39585	8334	188029
5.50	14970	14970	39716	7565	208510	15716.0	42113	8022	221093
6.00	14180	14180	41903	7288	240943	15036.1	44206	7688	254186
7.00	13190	13190	88953	13685	578191	13676.3	93315	14356	606551
8.00	11900	11900	94088	12545	705656	12382.3	97720	13029	732898
9.00	11040	11040	97495	11470	828708	11187.4	100171	11785	851456
10.00	10090	10090	100368	10565	953491	10100.3	101117	10644	960609
11.00	9178	9178	101157	9634	1062149	9118.5	100898	9609	1059434
12.00	8250	8250	100211	8714	1152427	8234.9	99782	8677	1147493
13.00	7551	7551	98756	7901	1234453	7441.2	97976	7838	1224700
14.00	7059	7059	98618	7305	1331336	6728.5	95646	7085	1291217
15.00	6374	6374	97389	6717	1412144	6088.5	92923	6408	1347385
16.00	5865	5865	94852	6120	1470210	5513.4	89914	5801	1393669
18.00	5005	5005	184790	10870	3141430	4530.8	170750	10044	2902751
20.00	4190	4190	174705	9195	3319395	3733.5	157021	8264	2983391
22.00	3584	3584	163254	7774	3428334	3084.2	143170	6818	3006578
24.00	2913	2913	149431	6497	3436913	2553.5	129667	5638	2982333
26.00	2459	2459	134300	5372	3357500	2118.5	116799	4672	2919983
28.00	2009	2009	120636	4468	3257172	1760.7	104738	3879	2827916
30.00	1723	1723	108228	3732	3138612	1465.8	93569	3227	2713494
32.00	1465	1465	98828	3188	3063668	1222.1	83324	2688	2583039
34.00	1145	1145	86130	2610	2842290	1020.2	73997	2242	2441892
36.00	965	965	73850	2110	2584750	852.8	65557	1873	2294482
38.00	833	833	66526	1798	2461462	713.6	57958	1566	2144429
40.00	706	706	60021	1539	2340819	597.8	51144	1311	1994630
42.00	569	569	52275	1275	2143275	501.2	45058	1099	1847365
44.00	456	456	44075	1025	1895225	420.6	39637	922	1704380
46.00	366	366	36990	822	1664550	353.2	34822	774	1566972
48.00	305	305	31537	671	1482239	296.9	30555	650	1436064
	Totals	345013	2834524	236139	#######	345686	2653441	231660	#######
					SSQ	3.6E+07			

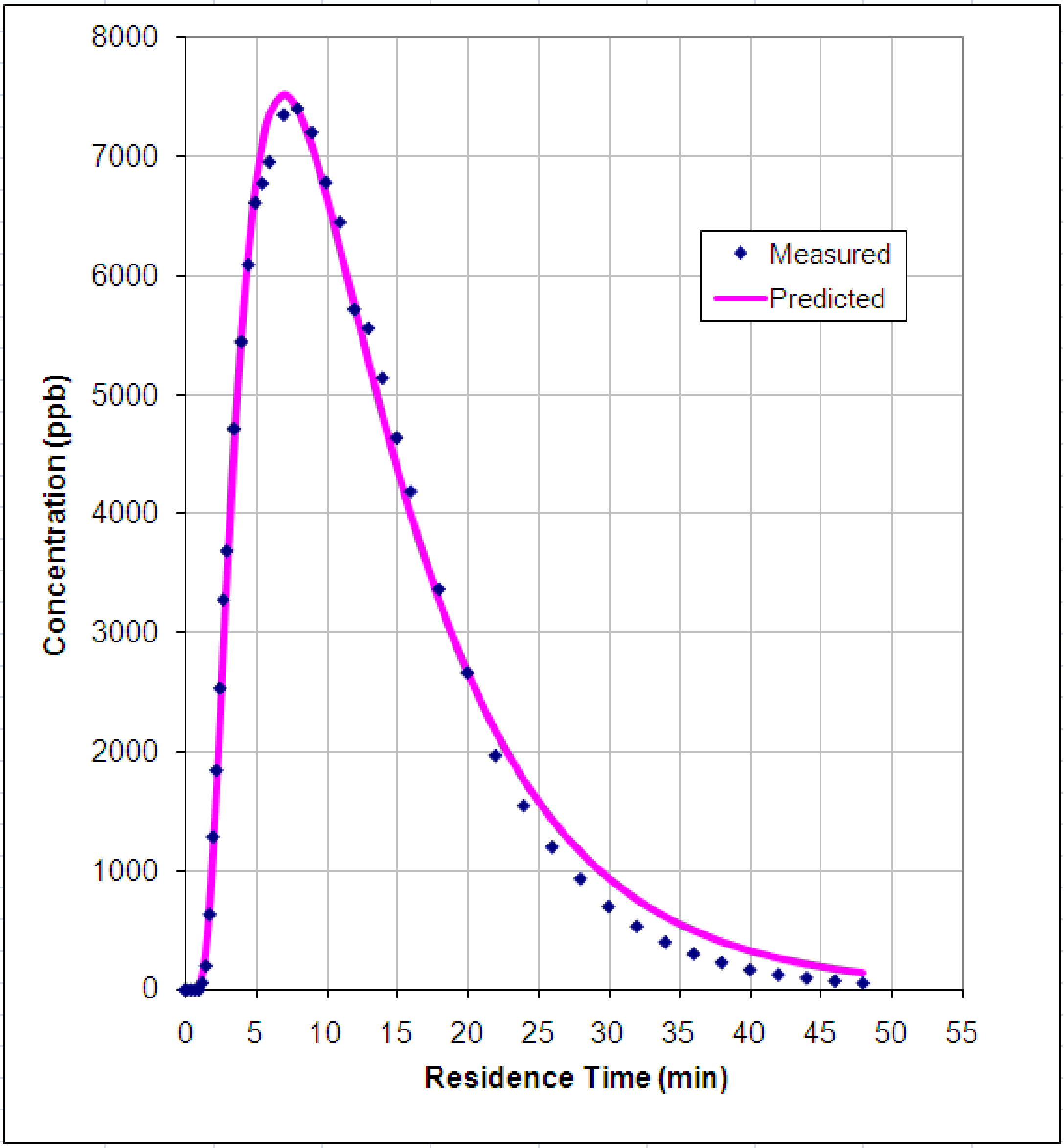
Appendix

RTD Data for Two Columns in Series

Dispersion Calculations	Measured	Fitted	Units
Baseline Concentration	0.00	0.00	ppb
Mean Retention Time	**13.03**	13.79	min
Variance (Required)	61.68	77.10	min^2
Variance (Calculated)	61.68	77.10	min^2
Peclet (Reset)			(Delete to run!)
Peclet (Calculated)	**4.22**	3.60	---

Curve Fit Parameters	
Parameter #1	9.17
Parameter #2	3.61
Parameter #3	13098

Performance Calculations		
Active Volume	23058	gal
Estimated Flow Rate	**1769**	gpm



		Measured Data				Curve Fit Data			
Time (min)	Conc'n (ppb)	Conc'n (ppb)	Calculated t(C)(dt)	Calculated (C)(dt)	Calculated t^2(C)(dt)	Conc'n (ppb)	Calculated t(C)(dt)	Calculated (C)(dt)	Calculated t^2(C)(dt)
0.00	0	0	0	0	0	0.0	0	0	0
0.25	0	0	0	0	0	0.0	0	0	0
0.50	0	0	0	0	0	0.0	0	0	0
0.75	0	0	0	0	0	2.2	0	0	0
1.00	0	0	0	0	0	29.5	3	4	3
1.25	61	61	9	8	10	135.1	23	21	26
1.50	202	202	45	33	62	363.5	86	62	118
1.75	634	634	170	105	276	723.1	221	136	359
2.00	1287	1287	450	240	844	1193.1	449	240	842
2.25	1845	1845	832	392	1768	1739.5	779	367	1655
2.50	2534	2534	1300	547	3088	2327.3	1207	508	2867
2.75	3280	3280	1908	727	5008	2926.7	1724	657	4525
3.00	3691	3691	2505	871	7202	3514.7	2315	805	6655
3.50	4716	4716	6831	2102	22200	4596.8	6591	2028	21419
4.00	5448	5448	9529	2541	35733	5503.6	9469	2525	35509
4.50	6095	6095	12264	2886	52124	6218.5	12455	2931	52933
5.00	6617	6617	15096	3178	71704	6751.8	15402	3243	73160
5.50	6780	6780	17584	3349	92314	7125.0	18213	3469	95620
6.00	6960	6960	19751	3435	113570	7362.9	20826	3622	119752
7.00	7356	7356	46527	7158	302426	7525.0	48386	7444	314508
8.00	7407	7407	55361	7382	415209	7396.3	55955	7461	419663
9.00	7209	7209	62118	7308	528003	7090.1	61567	7243	523323
10.00	6789	6789	66491	6999	631660	6683.3	65424	6887	621527
11.00	6453	6453	69521	6621	729965	6226.4	67776	6455	711648
12.00	5719	5719	69989	6086	804874	5752.1	68877	5989	792081
13.00	5564	5564	70519	5642	881484	5281.3	68959	5517	861986
14.00	5144	5144	72279	5354	975767	4826.6	68228	5054	921080
15.00	4642	4642	70949	4893	1028753	4395.6	66861	4611	969481
16.00	4190	4190	68448	4416	1060944	3992.2	65005	4194	1007584
18.00	3369	3369	128503	7559	2184551	3273.9	123524	7266	2099907
20.00	2667	2667	114684	6036	2178996	2670.6	112946	5945	2145968
22.00	1970	1970	97377	4637	2044917	2171.2	101679	4842	2135252
24.00	1548	1548	80914	3518	1861022	1761.5	90452	3933	2080388
26.00	1199	1199	68675	2747	1716875	1427.1	79714	3189	1992848
28.00	933	933	57564	2132	1554228	1155.3	69724	2582	1882536
30.00	701	701	47386	1634	1374194	934.7	60610	2090	1757687
32.00	533	533	38254	1234	1185874	756.1	52417	1691	1624940
34.00	401	401	30822	934	1017126	611.6	45136	1368	1489490
36.00	302	302	24605	703	861175	494.7	38723	1106	1355288
38.00	229	229	19647	531	726939	400.2	33114	895	1225230
40.00	169	169	15522	398	605358	323.9	28240	724	1101349
42.00	129	129	12218	298	500938	262.1	24024	586	984981
44.00	101	101	9890	230	425270	212.2	20393	474	876918
46.00	77	77	8010	178	360450	171.8	17278	384	777527
48.00	59	59	6392	136	300424	139.1	14614	311	686858
	Totals	125010	1500936	115176	26663322	126448	1639389	118856	31775495
					SSQ	1459330			