EVALUATION OF COLUMN FLOTATION CIRCUITS FOR FINE COAL CLEANING

by

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(ABSTRACT)

The objective of this study was to evaluate various multi-stage circuit arrangements that may be used to improve the column flotation of micronized coal. Laboratory flotation tests were performed with two different samples of Pittsburgh No. 8 seam coal. The first coal, Coal A, was ground to two different particle sizes and subjected to both column and conventional flotation. These tests were performed to obtain an initial understanding of the operational behavior of the column process and to compare the results with those of conventional flotation. The second coal, Coal B, was used in the actual testing of three different column circuit arrangements. The experimental test results were compared to simulated results obtained using a rate-based flotation model constructed in the present work. Several hypothetical flotation circuits were also examined using the simulation model and experimental flotation rate data.

The circuit test results showed that each of the different circuit configurations possessed specific advantages in terms of throughput capacity, combustible recovery, ash rejection and sulfur rejection. However, the overall performance curves for each circuit were all found to fall on or just below the maximum separation curve predicted using the

release analysis technique. Also, the simulated results in almost all cases predicted better results than what was actually obtained. This discrepancy was attributed to the inability of the rate-based model to adequately describe restrictions associated with the carrying capacity of the column froth.

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1.0 INTRODUCTION

1.1 General

During the last several decades, froth flotation has become almost universally accepted as the primary method for the beneficiation of as-mined metalliferous ores and several industrial minerals. In 1973, approximately 1.7 million tons of ore were treated per day using froth flotation (Aplan, 1977). Along with this continuing growth in the use of the froth flotation process, there has been an increasing trend toward the complexity and size of flotation circuits and plants due to market competition, increasing production costs, and the deterioration of the higher grade reserves of easily treatable ores.

In light of these facts, it has become readily apparent that there exists a need for closer scrutiny during the design of flotation circuits as well as the need for a better overall understanding of existing flotation circuits and processes so that some means of automatic or semi-automatic control can be implemented to affect overall system improvement. The ever increasing trend toward more complex and larger flotation networks, along with the well perceived need to better control these systems, has led to greater emphasis being placed upon flotation process modelling and circuit simulation by computer. These attempts and the following results have been widely described in the literature.

Although there has been widespread acceptance of the froth flotation process in the minerals processing industries, the trend of acceptance has been somewhat less for

the coal preparation sector. In 1973, out of a total of 300 million tons of clean coal produced, only 14 million tons were produced by froth flotation (Aplan, 1977). The reasons for this slower trend of growth in implementing flotation have been due to the lower market value for coal on a unit basis, marketing and dewatering problems associated with the recovery of the finer sizes of coal, beneficiation problems associated with the high clay content of fine coal slurries, inability to date to obtain as selective a sulfur reduction than what is available with gravity separation processes, and the smaller percentage of coal preparation plant feeds existing in the size range amenable to froth flotation (i.e., 28 mesh x 0).

Since the mid-late 1970's, the growth of froth flotation circuits for use in coal processing has seen an increase due to a number of factors. First, the oil crisis of the early 1970's affected a drastic increase in the unit value of coal, which gave increased incentive to maximize coal recovery at all sizes appearing in a particular plant feed. Secondly, environmental mandates have dictated the abatement of blackwater effluents of fine coal that were previously discarded at preparation plants. An alternative to these discharges was the recovery of the inherent fine coal solids in these streams and the return of the clarified process water for reuse in the plants. Froth flotation has provided a feasible and economic alternative in many cases. Third, more environmentally restrictive regulations geared toward the unacceptable discharge of sulfur dioxides and other pollutants from coal burning facilities have necessitated the need for implementation of more intricate technology. As an alternative to the exceedingly expensive remedies

of flue gas desulfurization and scrubbing, an increased surge of interest has been directed toward the coal preparation process, particularly in the areas of fine grinding and subsequent flotation to obtain levels of deeper contaminant removal. This approach is also substantiated by the fact that remaining coal reserves are of increasingly lower quality, analogous to what has already occurred in the minerals processing industries. Finally, although current and past use of froth flotation for processing the finer sizes of coal has been largely limited to single-stage, rougher-only circuits, there has been increasing interest in applying more complex circuitry to existing process operations (Nicol, 1984). These more complicated techniques could very well involve the use of size classification and subsequent parallel flotation banks based on size as well as multiple cleaning stages to enhance performance.

One technique of froth flotation that has received increasing interest in the last several years has been column flotation. This process differs not only in the use of cylindrical tubes or columns instead of conventional rectangular tanks, but also involves the implementation of counter-current feed, where feed slurry flows into the unit in a direction opposite to the flow of collected froth product. Several accounts have confirmed that column flotation exhibits superior performance in that a more stable, quiescent environment, without turbulent mixing leads to increased recovery (Mular ., 1986). Also, the use of the counter-current feed flow concept enhances the chance of bubble-particle contact and subsequent recovery. Additionally, various methods of froth washing by auxiliary water addition through sprays, etc., have led to the production of

higher grade products.

Unfortunately, only elementary mathematical modelling approaches have been undertaken with respect to the design of column flotation circuits. Furthermore, very little if any attention has been given to the simulation of single or multiple-stage column flotation circuits for fine coal flotation.

1.2 Objectives

The primary objective of this research effort was to identify circuit arrangements and networks that may be implemented in the future for the deep cleaning of fine coal. Secondary objectives of this work were:

- to develop an improved understanding of the relationship between column performance and various process variables,
- to develop a flotation circuit simulator that can be used to evaluate the performance of single-stage and multi-stage column flotation circuits for fine coal cleaning,
- to demonstrate how conventional release analysis techniques may be used to determine the unknown modeling parameters for a particular coal feed,
- to validate the circuit simulator using test data obtained from continuous, laboratory-scale flotation tests, and
- to identify circuit arrangements and networks that may be implemented in the future for the deep cleaning of fine coal.

In addition to the objectives listed above, an overview of the observations and conclusions regarding the performance of column flotation cells will be presented with respect to the column model. These observations will be tied into the present simulator development by describing the development of various simulator methodologies that have been carried out to date. Also, a brief summary of past methods and procedures will be presented on the determination of flotation rates that can be used as input data for flotation circuit simulation.

2.0 LITERATURE REVIEW

2.1 Flotation Models

Lynch (1981) listed the variables inherent in the froth flotation process. These variables can be described by either being independent or dependent variables. The independent variables can be further subdivided into manipulated variables or disturbance variables. Manipulated variables include such items as reagent addition types and levels, pulp-froth interface height, air addition rate, and concentrate collection point within the unit. Disturbance variables include feed stream associated characteristics such as head grade, fineness of crystallization, degree of oxidation, size consist, pulp density, and overall volumetric flow rate.

The dependent variables include product stream characteristics such as recovery, grade, pulp density, and overall volumetric flow rate. Thus, it can readily be seen that the mathematical approximations to date have been directed at determining the dependent variable values from a specified set of input independent variables, commonly the determination of product recovery and grade.

There have been three different approaches to the problem of mathematically modelling the froth flotation process. These are empirical, probabilistic, and kinetic. The use or selection of a particular methodology has been determined for the most part upon the particular application or intended use of the model.

Empirical modelling is the least time consuming and most cost effective form of a flotation model to develop. This approach involves the collection of actual data around

an operating flotation cell or circuit, and the use of multiple linear or spline regression to fit a mathematical relationship to the observed values of the independent and dependent variables. Early efforts at utilizing this approach are described in the literature by Faulker, Pitt, and Smith and Lewis (Lynch ., 1981). Use of empirical flotation models as a means of controlling a flotation plant, utilizing a feedback approach, was described by Cooper (1966).

As can be quickly ascertained, the application of empirical models has been somewhat limited in scope and range of application. Although an empirical relationship can be quickly developed for an operating process, and subsequently used to control this process, the results can not be extrapolated to other systems or feed streams. Also, the range of application is limited to the particular process and collected data set on-hand.

To develop an approach with greater range and possible application to research and design problems, a second methodology of flotation modelling was formulated based on probability theory. Schuhmann (1942) envisioned that the recovery of components within a flotation feed could be related to an overall probability of recovery, which was subsequently related to the probabilities of successive events occurring within the process as they occur on an individual particle. The relationship thus developed was:

$$P = P_c \, x \, P_a \, x \, F \qquad (1)$$

where P is the probability of flotation, P_c the probability of bubble-particle collision, P_a the probability of bubble-particle adhesion after collision, and F the froth stability factor.

Tomlinson and Fleming (1965) later modified this approach to include the effects of disengagement and subsequent drainage of particles from the froth back into the pulp by:

$$F = P_b x P_f \tag{2}$$

where P_b is the probability of levitation of the base of the froth column without disengagement and P_f the probability of drainage from the froth. Further studies by these authors showed that the probability of recovery can be related to the number of bubbles available in the cell pulp and their ability to capture a specified particle and the factors needed to maintain adhesion until the discharge point of the froth overflow weir is reached.

The first attempt at applying the probabilistic approach to flotation cell banks was by Kelsall (1961) who suggested that:

$$W = W_o[1 - P]^n \tag{3}$$

where W is the weight of component remaining in the cell pulp, W_o the initial weight of component in the cell pulp, P the overall probability of recovery, and n is the number of cells in the bank. This approach was further utilized by Davis (1964) to formulate a model for a zinc flotation plant utilizing actual data to perform circuit control.

The third approach to mathematically modelling flotation processes was first postulated by Sutherland (1948) and is known as the kinetic approach. Basically, the

kinetic derivation relates the flotation recovery of a particle(s) to an analogy with chemical engineering reactions that proceed at a specified rate. Further description of the kinetic methodology as applied to flotation was given by Harris (1951), Brown (1953), and Arbiter and Harris (1962). The general equation described by these authors can be given for a semi-batch flotation cell by:

$$\frac{dC}{dt} = -kC^n \qquad (4)$$

where C is the concentration of a constituent within the cell, t the elapsed time, k the flotation rate constant, and n the order of the reaction (i.e., flotation) kinetics. According to this analysis, the rate of recovery of a particular component or species within the flotation feed is strongly related to the numerical order of the flotation process (reaction) and the retention time distribution within the cell.

Initially, the use of a non-first order rate for flotation recovery was postulated by Arbiter and Harris (1962) to explain discrepancies experienced when plotting flotation results on a semi-log scale of remaining constituents versus time. As outlined by Arbiter and Harris (1962), Loveday (1966), and Imaizumi and Inoue (1965) this deviation or nonlinearity from first-order expectation could possibly be the result of the interdependence of flotation process events of different particles upon each other as well as the variation of the flotation rate constant distribution for the same type of particles. However, Bushell (1962), Bull (1966), Imaizuni and Inoue (1965), Mika and Fuerstenau

(1969), and Aplan and Rastogi (1985) showed that for practical purposes the process of flotation could be considered to be first order over a certain range of performance. In other words, the recovery of particles into the concentrate launder are deposited at a rate which is directly proportional to the concentration of these particles as they exist in the pulp volume. As outlined by Lynch (1981), this assumption effectively reduces the complexity of the kinetic model and also facilitates the determination of rate constants for a particular set of constituents comprising a flotation feed.

Initial work with respect to retention time distributions was described by Danckwerts (1953). In these studies, retention time distribution was related to the type of mixing regime present in a continuous-flow reactor. This work was followed up by the development of a component recovery model given by Levenspiel (1972):

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

$$where \quad A = \sqrt{1 + 4k_i\tau/Pe}$$

(5)

where R_i is the fractional recovery of component i, k_i the first-order rate constant, t the retention time, and Pe is the Peclet number. Pe is a dimensionless quantity that describes the mixing intensity within the flotation cell. Techniques for determining Pe and have been described by Bull and Spottiswood (1974), Gardener (1980), and Mankosa et al. (1988).

Tomlinson and Fleming (1965) showed that the probabilistic and kinetic approaches are related to the number of bubbles available and the capture capability of these bubbles on various species within the pulp. Other investigations by Flint and Howarth (1970), Reay and Ratcliff (1973), and Luttrell and Yoon (1986), have been concerned with directly relating the kinetic aspect of flotation to the fundamental parameters of the process such as bubble quantity and bubble size distribution. Recent investigation into the behavior of the froth phase in conjunction with the kinetic modelling approach has been described by Tomlinson and Fleming (1965), Lynch (1981) and Whitten (1982). However, specific conclusions have not yet been established.

2.2 Column Flotation

During the past thirty years, the technology of column flotation has received an increasing amount of attention as a means to provide for an improved technique for treating an ever increasing amount of finer and lower grade feed material to flotation processes. The earliest plant trials with flotation columns appear to be as described by Wheeler (1966) and Wheeler and Boutin (1967). These early units were employed in a test circuit configuration for treating a copper ore. Since that time, other applications have been described for coal, copper, molybdenum, etc., for the treatment of typically-sized flotation feeds. Descriptions of these trials have been provided through the reports of Balderama and Hood (1988), Finch et al. (1988), Nicol et al. (1988), Parekh et al. (1988), and Zipperian (1988).

With respect to the mathematical modelling of flotation columns, an early theoretical treatment of counter-current feed columns was provided by Fuerstenau and Sastry (1969). This early attempt at column flotation modelling involved the use of a set of differential equations to describe the liquid and air-bubble phases. It was found that a complete solution for the general case was not possible at this early stage of technology development.

Another attempt at using the differential equation approach to column flotation was presented on a limited basis by Rice et al. (1974). In this study, emphasis was placed upon the examination of axial mixing within the columns. Again, as with the description given by Fuerstenau and Sastry (1969), the mixing profile used was based on an axially dispersed plug-flow model.

More recently, a follow-up to this early work was given by Mankosa et al. (1987). In this approach, a population balance model using differential equations was used to describe the steady-state and dynamic behavior of column flotation. It was found that the solution to this set of equations required the use of numerical techniques for solution.

In order to obtain a simpler and more useful approach, attempts have been made to describe the column flotation process on a steady-state basis. Dobby and Finch (1986) presented the use of the Levenspiel (1972) equation and the incorporation of an axially dispersed plug-flow mixing regime within flotation columns. This approach was validated and presented also by Luttrell et al. (1988) and Mankosa et al. (1988).

Although modelling on a steady-state basis does not give a complete picture of overall column performance, it lends itself very well to the theoretical computer simulation of operating columns and circuits.

As presented by Mankosa et al. (1988), for the range of length to diameter ratios normally encountered by operating columns, a Peclet number of 1 to 5 would be sufficient in describing the axially-dispersed plug-flow regime existing in a column cell operating at steady-state. Also, for this normal range of operation, it was found that a variation of the Peclet Number would affect overall recovery minimally.

More recent efforts at modelling flotation columns has been involved with describing the fundamentals of bubble capture and its relation to overall column performance. This approach has been outlined by Dobby and Finch (1987). Another presentation was given by Luttrell et al. (1988) with respect to the fundamental incorporation of bubble size and capture properties as related to the Levenspiel (1972) equation.

2.3 Flotation Circuit Simulation

The use of the mathematical principles surrounding the froth flotation process for the characterization of a typical flotation circuit was originally outlined by Gaudin (1957). In this analysis, the kinetic approach was utilized along with the idea of a separation index (SI) to characterize a flotation network. A later effort by Davis (1964) combined the probabilistic approach of Kelsall (1961) to characterize an actual zinc

flotation circuit.

Later approaches to circuit simulation fell back on the kinetic modelling approach. King, et al. (1969) presented a preliminary circuit simulation routine that was cumbersome at best, but utilized the idea of first-order flotation rates in conjunction with mass balancing to examine circuits. Smith and Bjerring (1969) employed a non-linear dynamic model incorporating the kinetic approach to flotation circuit analysis that was also preliminary and somewhat complicated.

A later program was given by Loveday and Marchant (1972) that tied in the idea of first-order kinetics to retention time and mixing regime within the cells comprising a flotation bank. In this approach a residence time was assumed for each cell, and a subsequent mass balancing routine incorporating the idea of rate constants. King (1973) and Sutherland (1976) later assumed fixed pulp densities for certain streams with mass balancing occurring outward from this assumption. Again, the idea of a first order flotation rate constant was employed.

Lynch (1981) proposed the use of the kinetic approach along with the idea of a reiteration routine for mass balancing individual cells and overall circuits. Initially input to a flotation cell bank included a characterization of the feed stream, with component rate constants, etc., and dimensional input with respect to cell and circuit volumes and arrangement. By assuming an initial value of zero concentrate flow from an individual cell, the routine could reiterate until the retention times obtained matched the cell volume with the overall feed flow rate given. From this point, the final step of the calculation

process involved the mass balancing the flows around a circuit.

The approach of Lynch (1981) could be modified to incorporate a host of variables. These included water recovery, return of floatable material from the froth back into the pulp, and the entrainment of gangue into the froth phase. Kawatra (1982) followed along these lines in trying to develop a simulation routine that would incorporate entrainment as well as primary recovery. Laurila (1983) attempted to quantify these relationships into an overall model and simulation procedure for the evaluation of coal flotation cells and circuits. One of the latest approaches along these same lines was presented by Mular et al. (1986), in which a detailed simulator for a copper-molybdenum flotation circuit was presented.

More recent research has been aimed at developing suitable methods for characterizing the froth zone with respect to drainage and entrainment. Warren (1984) attempted to separate the previously assumed idea of an overall rate constant of recovery into the two processes of true recovery and entrainment. This description was followed up by Forssberg and Subrahmanyam (1988) that presented the earlier ideas with respect to froth stability. Dobby and Falutsu (1989) presented ideas with regard to the idea of froth stability, with particular attention being paid to the observation that column flotation cells are somewhat limited with respect to the available bubble carrying capacity.

2.4 Flotation Rate Constants

To utilize the kinetic approach for flotation cell and circuit modelling, it was

envisioned early on that some means must be ascertained to facilitate the determination of flotation feed components as well as their flotation rate constants. In this regard, Morris (1952) presented a study by which it could be seen that individual flotation rate constants could be determined from semi-log plots of cumulative fraction remaining for individual components if the flotation follows a typical first-order process. Following along this same line, Brown (1953) demonstrated that it was necessary to use a modified continuous process rather than batch in order to insure first-order performance of components during coal flotation.

Jowett and Safvi (1959) showed that for practical purposes flotation proceeds at a first-order rate. Also, this presentation tied in the idea that the rates of flotation could be evaluated with respect to floatable and non-floatable fractions, and presented a methodology by which these rates could be determined easily from laboratory flotation tests.

Similarly, Kelsall (1961) showed, using the probabilistic approach, what appeared to be two primary components reporting for recovery by flotation, i.e., a fast-floating and a slow-floating component, depending upon the grades, liberation level, etc. This approach was extended to kinetic theory by Bushell (1962) by the demonstration that the individual rates of the floatable components could be determined graphically from the cumulative remaining semi-log plots by noting where the overall curve deviates from linearity.

Imaizumi and Inoue (1965) described deviation from first-order flotation as well

as the apparent distribution of flotation rate values continuously for a particular feed. Following along this same line, Loveday (1966) and Harris and Chakravarti (1969) presented various representations for rate constant distributions. Lynch (1981) gave an examination of the distribution of rate constants within a particular flotation feed, however, for practical purposes, most feed streams exhibit a fast and slow-floating recoverable component, as earlier studies had indicated. This idea was extended and confirmed by Aplan and Rastogi (1985) in work concerning the flotation rates of coal.

Another approach to the characterization of flotation feeds was presented by Dell (1959) and Dell et al. (1972). This technique, which is known as release analysis, involved the complete recovery of all the floatable fractions, and subjecting these to stage-wise recovery depending upon the level of inherent floatability. This technique provided a means by which a flotation feed could be characterized independently of the specific flotation cell.

The work by Dell et al. (1972) further demonstrated the use of release analysis to the characterization of ores for flotation with an analogy to washability or partition curves for density-based separation processes. Release analysis results were typically plotted as cumulative recovery versus cumulative elementary grade. A follow-up to this initial presentation was presented by Jowett and Sutherland (1985) with respect to the perfect separation aspect of these curves.

3.0 MATERIALS AND METHODS

3.1 Materials and Apparatus

3.1.1 Coal Samples and Preparation

The test work conducted during this study utilized two separate test coals originating from the Pittsburgh No. 8 Seam. The first coal, i.e., Coal A, was a minus 2-inch pre-sized clean coal product of an existing preparation plant. The as-received analysis of this sample is given in Table I.

Initially, this coal was allowed to air-dry to room temperature and then crushed to nominal 1/4-inch top-size using a ring-type hammermill. The coal was then mixed and split representatively into 10 lb lots and stored in a freezer to minimize oxidation. Prior to flotation testing, a single lot of the coal was removed from the freezer and allowed to come to room temperature before further size reduction. Then, a grate discharge, vane-type pulverizer was employed to reduce the nominal top-size to 60 mesh. A typical particle size distribution of the pulverized coal is given in Figure 1.

A portion of the test work with Coal A was performed on the minus 60 mesh pulverizer product. Another series of tests were conducted using a minus 200 mesh size fraction obtained by wet grinding at 30% solids in a 6-inch diameter stirred ball mill. The mill was charged with 3 mm diameter steel beads, weighing approximately 35 lbs, and rotated at a speed of 300 rpm. The mill contents were agitated for 2 minutes to obtain a minus 200 mesh particle size. Typical size distribution as measured by an Elzone 80-XY particle size analyzer is shown in Figure 2. Since the data was not mass-

Table I. Analysis of the as-received Pittsburgh No. 8 coal (Sample A).

Analysis	Description	Comment
Particle Size	2 inch x 0	Determined by screening
Ash	11.03%	Average of 3 assays
Total Sulfur	2.80%	Average of 3 assays
Heating Value	13440 Btu/lb	Average of 2 assays

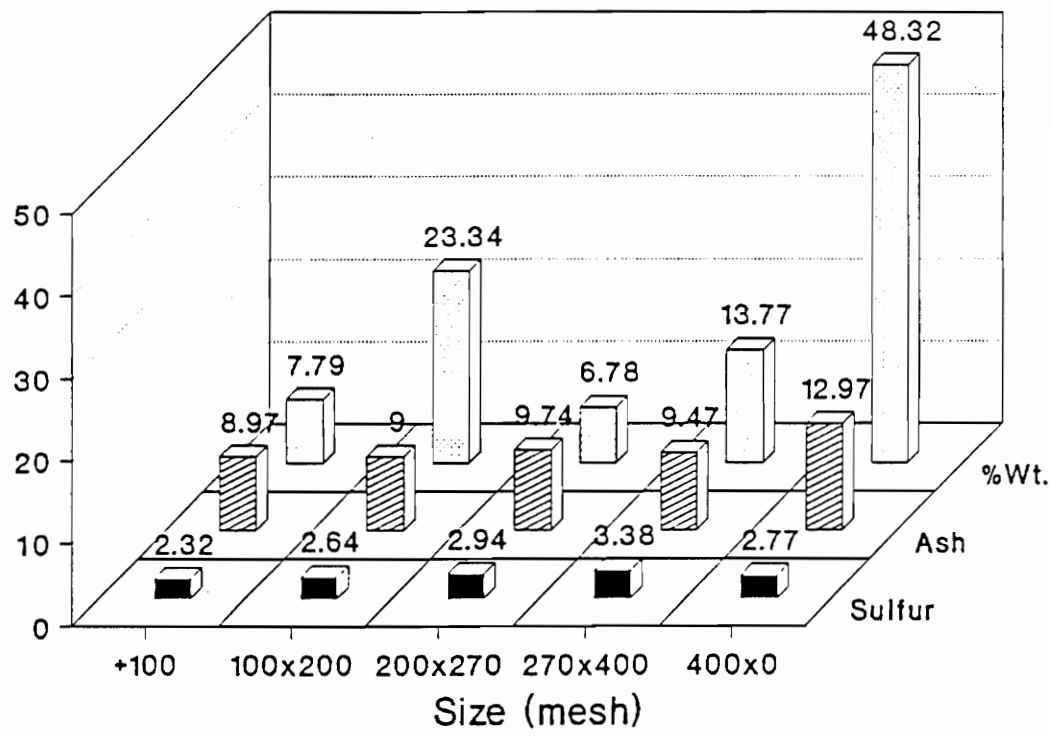


Figure 1. Raw coal analysis of Coal A after pulverization in a laboratory hammermill to 60 mesh x 0.

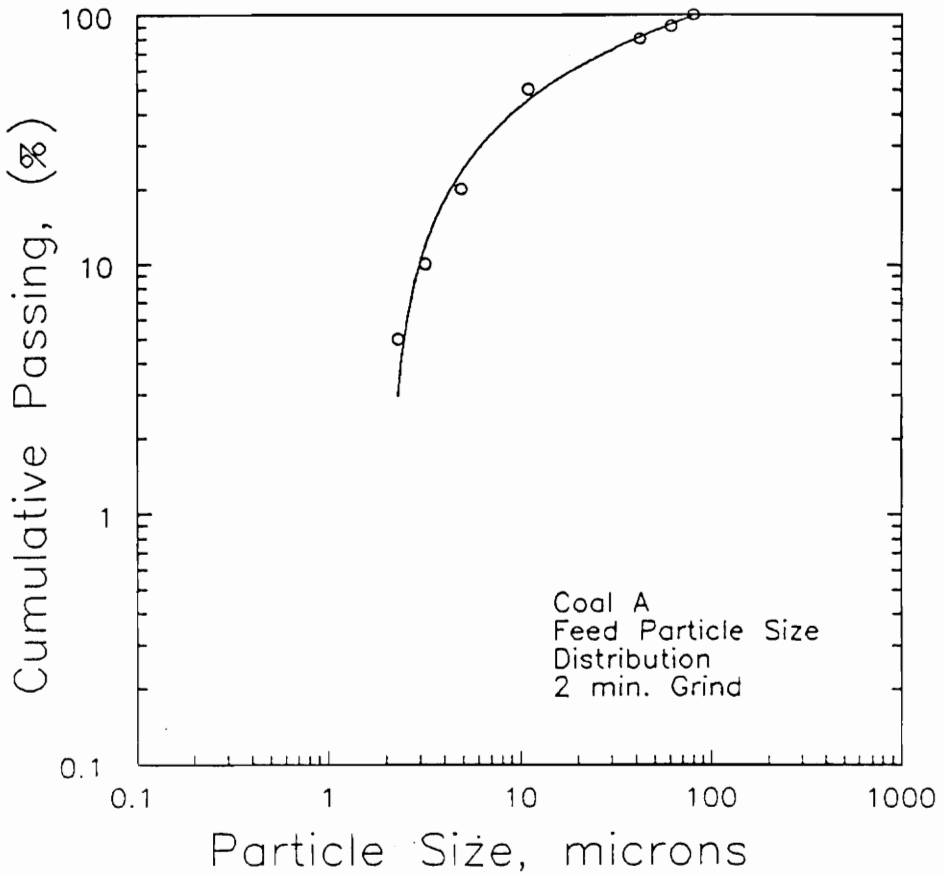


Figure 2. Particle size distribution of Coal A after micronizing in a stirred ball mill.

balanced, values below 10 μm are probably unreliable. The slurry was then further diluted and mixed with tap water to obtain the desired solids content prior to flotation.

The second coal used during the last part of the test work, i.e., Coal B, was obtained from a minus 28 mesh flotation feed stream at an operating preparation plant. The as-received analysis of the sample is shown in Table II. The slurry was shipped in 55 gallon barrels, dumped into a closed loop sump and pumping arrangement, and thoroughly mixed until completely homogenized. The slurry was then split into representative 5 gallon lots and stored until needed in the flotation tests.

Prior to flotation, a single lot of the slurry was agitated using a laboratory stirrer and placed in a 6-inch stirred ball mill for grinding. The mill and contents were rotated at a full speed setting of 300 rpm for 4 minutes to obtain a product size distribution of minus 325 mesh. A typical particle size distribution for the pulverized sample is given in Figure 3. This data was also not mass-balanced, so that values below 10 μm are probably unreliable. After the grinding was completed, the slurry was diluted with the appropriate amount of water to obtain the desired solids content prior to flotation.

All the procedures for mixing, splitting, and size reduction of coal samples A and B, along with the subsequent weighing and laboratory analysis of flotation products, were performed as close to ASTM standards as practical.

3.1.2 Column Flotation Cell

The column flotation cell(s) utilized during the course of this test work was the

Table II. Analysis of the as-received Pittsburgh No. 8 coal (Sample B).

Analysis	Description	Comment
Ash	9.49%	Average of 3 assays
Total Sulfur	3.55%	Average of 3 assays
Solids Content	24.2%	Received in slurry form
Slurry pH	7.52	Received in slurry form

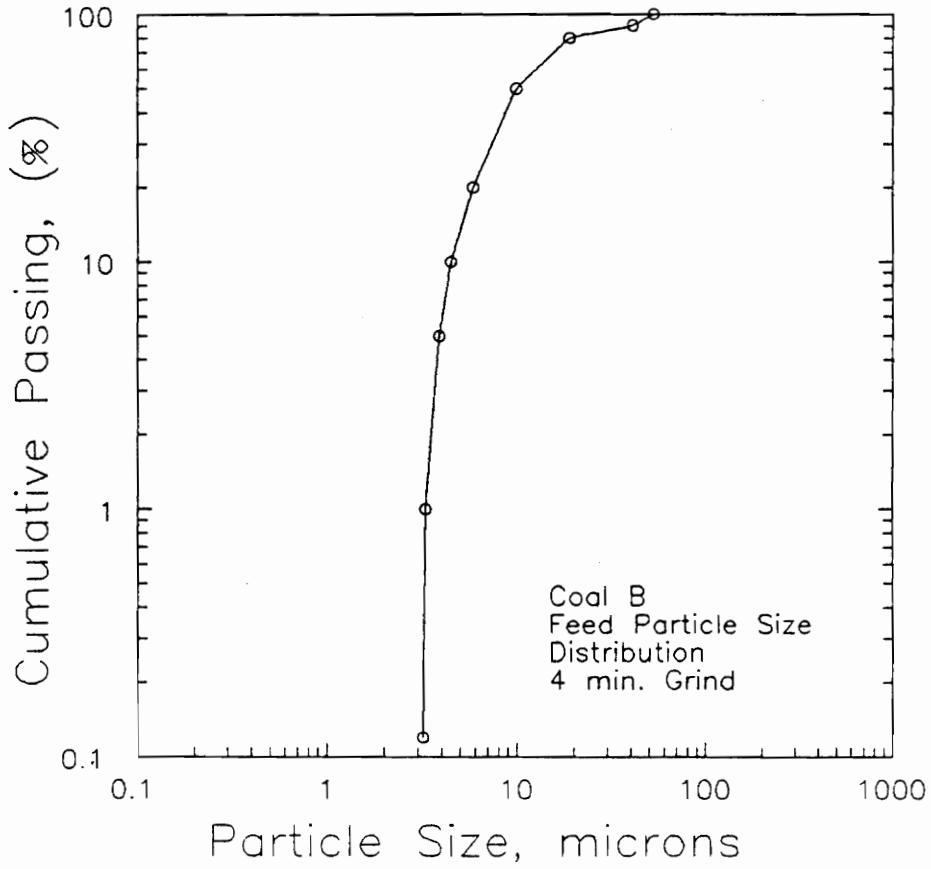


Figure 3. Particle size distribution of Coal B after micronizing in a stirred ball mill.

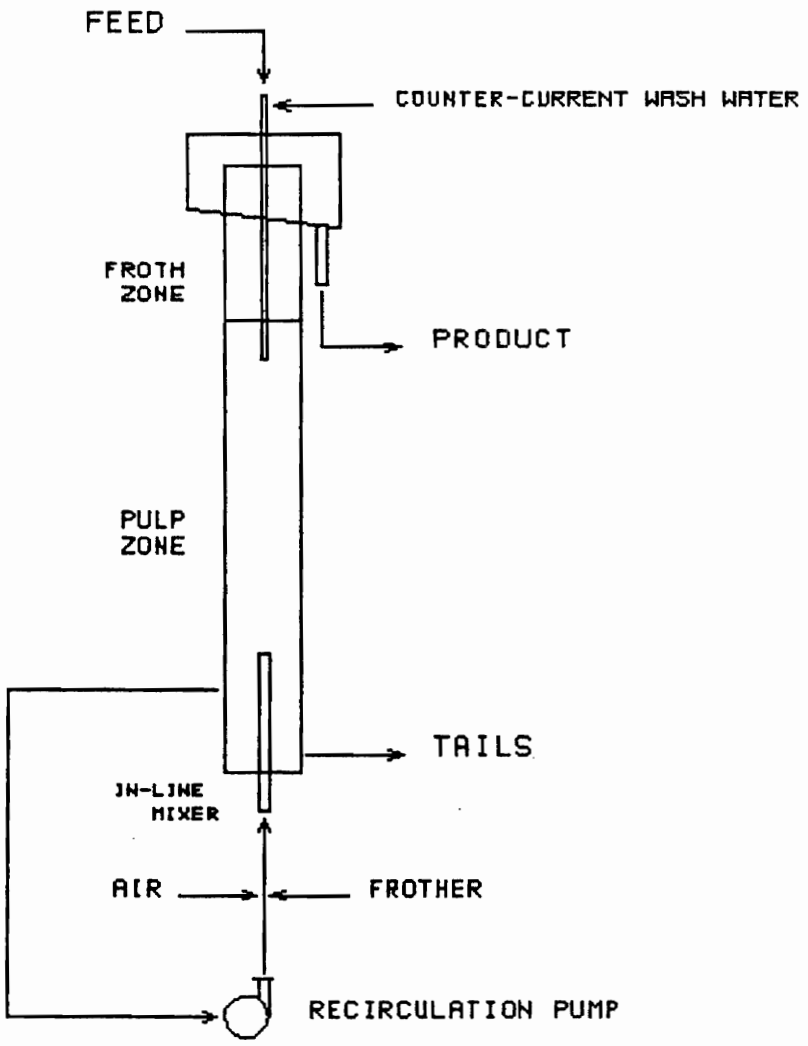


Figure 4. Schematic representation of the microbubble cell used in the evaluation of column flotation circuits.

microbubble column developed at Virginia Tech. Figure 4 provides a schematic representation of the microbubble column. Basically, the column consists of a 72-inch tall plexiglass tube with a special launder arrangement at the top of the tube for collection of the froth product. Feed slurry to the column was introduced just below the froth-pulp interface (i.e., 12 to 15 inches below the froth overflow lip). The tailings were allowed to discharge from the column near the bottom of the tube through a gravity discharge hose. The height of the pulp-froth interface was controlled by physical adjustment of the height of the discharge hose (see Figure 4).

The microbubble column flotation cells uses a novel technique for bubble generation. At the bottom of the plexiglass tube a Koflo in-line mixer was installed with its discharge going into the column. Prior to entry into the column, a special four-way fitting is used to supply compressed air to one side, and Dowfroth 1012 frother solution to the other side. A variable speed recirculation pump was used to circulate flotation pulp through the Koflo mixer. The discharge from the mixer entered the base of the column where the air bubbles continued to rise upward. Feed for the recirculation pump was obtained from pulp inside the column at a point approximately 6 inches above the bottom of the column.

Another feature inherent with the microbubble column design was the use of counter-current wash water within the froth interface zone. Specifically, the counter-current wash water assembly consisted of a hollow plexiglass cylinder (approximately 1-inch diameter), with a four-way outlet at the bottom and a single intake inlet at the top.

These inlet and outlet tubes were 1/8-inch diameter hollow plexiglass tubing. At each end of the 1-inch diameter cylinder, a special plexiglass plug was fitted to allow insertion of a 1/8-inch diameter plexiglass tube which served as the feed slurry entrance pipe (see Figure 4).

A specialized monitoring system was constructed and attached to the microbubble column(s) for monitoring fractional air hold-up within the column. Fractional hold-up is the percentage by volume of the column occupied by air. This value impacts the column performance by reducing the overall retention time. The monitoring system utilized two pressure transducers located a known distance apart connected to an overall control/power supply source. By comparing the pressure difference during actual column operation to the pressure difference for a column of water without air introduction, an idea of the fractional volume occupied by the generated air bubbles was obtained.

The pumps utilized for feeding coal feed slurry and frother solution to the microbubble column were variable speed peristaltic units, connected to the associated piping and fittings of the microbubble column via 1/4- and 1/8-inch diameter tygon tubing, respectively.

3.1.2.1 *Single-Stage Column Flotation*

A single column flotation cell was utilized for initial test work on coals A and B. For coal A, this initial column flotation testing was conducted on minus 60 mesh and minus 200 mesh particle sizes, respectively. For coal B, the testing was limited to the

minus 325 mesh particle size. The feed slurry was diluted to an appropriate solids content in a 5 gallon sump. The sump was fitted with a recirculation pump to prevent unwanted settling. Froth overflow and tailings products were allowed to discharge from the column to a common drain, with only samples being collected and saved at specified times.

3.1.2.2 *Multi-Stage Column Flotation*

To complete the simulator evaluation portion of the test work, a series of multi-stage column flotation tests were performed using three column cells arranged in three different circuits. The first type of circuit to be examined was a rougher-scavenger-cleaner (R-S-C) arrangement as shown in Figure 5. Initially, feed slurry was fed to a rougher cell for primary recovery. The froth concentrate from this cell was passed to a cleaner column for final upgrading, and the rougher tailings was sent to a scavenger column for recovery of any remaining valuables in the initial tailings. The scavenger concentrate was combined with the rougher concentrate and also sent to the cleaner cell. This type of set-up facilitated ease in operating the circuit due to limitations in the amount of equipment and space that were available. The final froth product from the cleaner cell, i.e., scavenger tailings, and the cleaner tailings were allowed to discharge to a common drain with only products collected during specified sampling periods. Sump capacities for the rougher, scavenger, and cleaner columns were 10-, 5- and 5-gallons, respectively. Each sump had its own mixer-recirculation pump assembly to

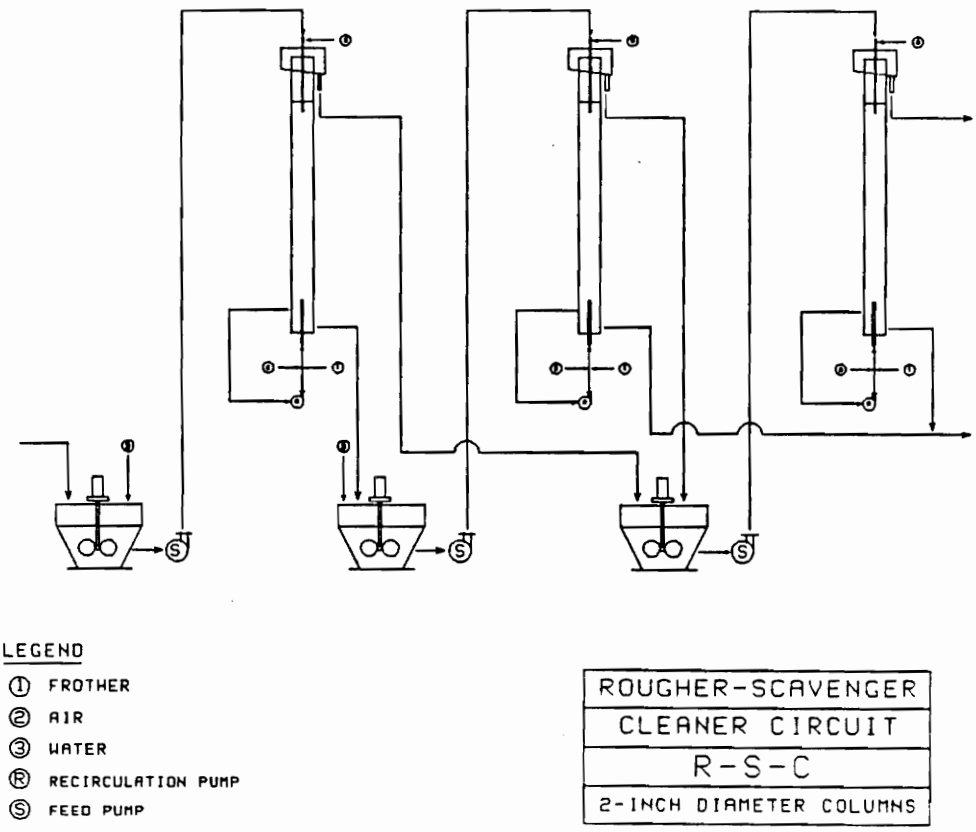


Figure 5. Schematic of the rougher-scavenger-cleaner (R-S-C) column flotation circuit.

prevent settling of solids in the feed slurry. Each column used in the circuit was fitted with its own frother solution supply and counter-current wash water assembly.

The second type of column circuit used for test work with Coal B was a modified R-S-C circuit constructed with three columns. Figure 6 provides a schematic representation of this circuit. This modified R-S-C circuit was similar to that mentioned above, however, the rougher concentrate was not recleaned in the cleaner column. Like the first R-S-C circuit, the rougher tailings were sent to a scavenger column for recovery of any remaining valuables, with the scavenger froth product being the only feed to the cleaner cell for final upgrading. The cleaner concentrate and the rougher concentrate were finally combined and discharged from the circuit to a common drain. Likewise, the scavenger tailings and cleaner tailings were combined and discharged from the circuit to a common drain. The set-up of associated sumps, pumps, etc., was identical to what was utilized for the first R-S-C circuit. The purpose of this arrangement was to determine if a higher recovery and better grade could be obtained from coal B without recleaning of the rougher concentrate.

The final circuit set-up to be examined was a rougher-cleaner-recleaner (R-C-RC) arrangement shown in Figure 7. The purpose of examining this circuit was to check if it was possible to make a high quality "superclean" coal product from coal B. The operation of this circuit was straightforward in that the froth product from each column was collected in a sump, mixed, and sent to the next succeeding stage. The tailings from each column were combined together and discharged from the circuit to a common drain.

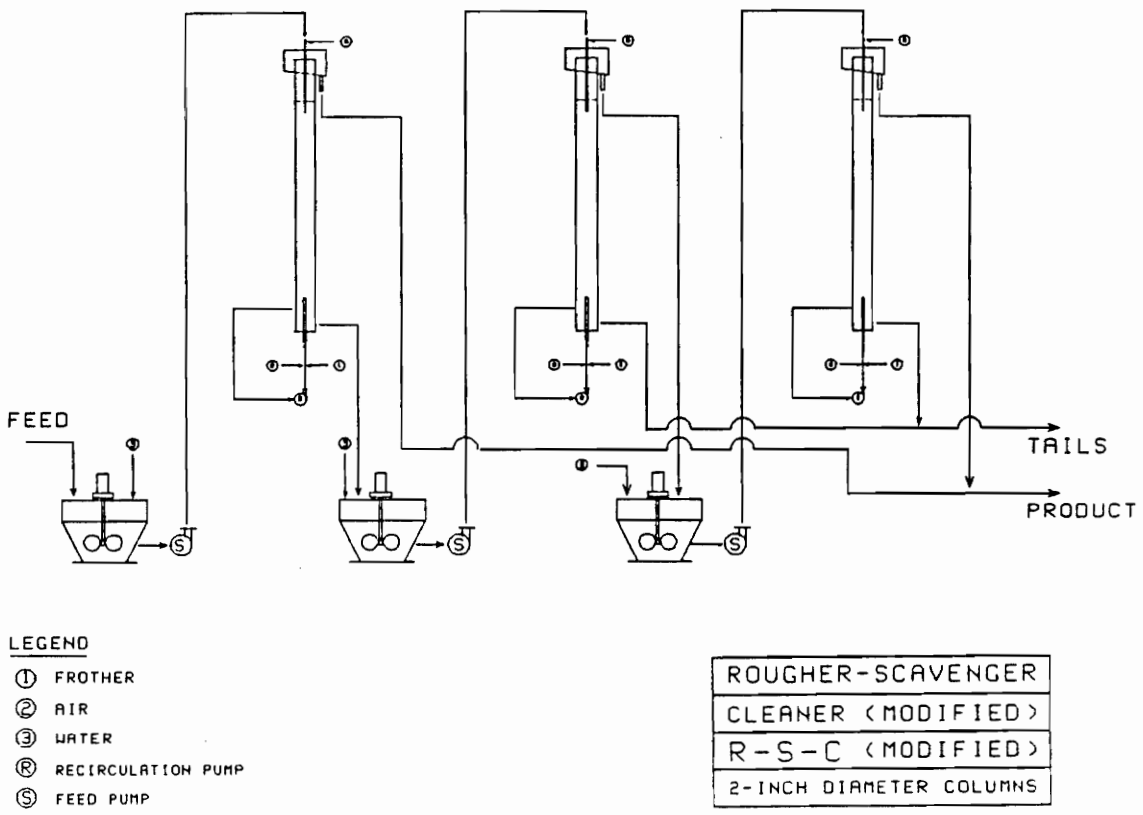


Figure 6. Schematic of the modified rougher-scavenger-cleaner (modified R-S-C) column flotation circuit.

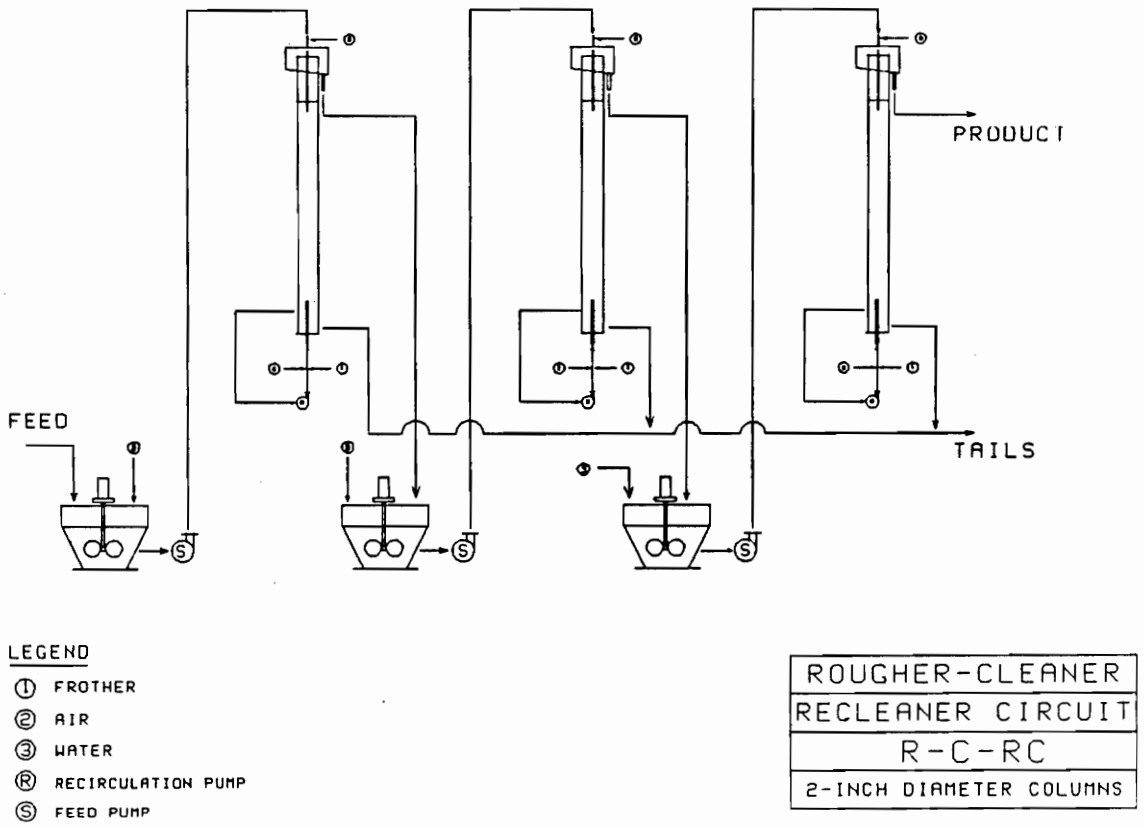


Figure 7. Schematic of the rougher-cleaner-recleaner (R-C-RC) column flotation circuit.

Likewise, the final froth product from the recleaner was allowed to discharge to a common drain. As with the previous two circuits examined, the products were only collected during specified sampling periods. The same sumps, pumps, etc., used with the R-S-C circuits were employed with this arrangement, with the larger 10 gallon sump being used for the rougher feed.

3.1.3 Batch Flotation Cell

To obtain flotation data that could be used as input to the circuit simulator, a series of release analysis tests was performed utilizing a conventional Denver Model D-25 batch flotation cell. The batch machine was equipped with a 4.5-liter capacity cell and a variable-speed impeller. Initial testing was performed on coal A at minus 200 mesh to determine the applicability of the method. Final release analysis work was performed on coal B at minus 325 mesh. As will be explained later, the results from these tests were used to obtain the actual input values for the circuit simulator.

3.1.4 Semi-Continuous Conventional Cell

Initial test work with coal A was carried out with a modified version of the batch Denver cell that was set-up to be operated in a semi-continuous fashion. Figure 8 gives a schematic representation of this arrangement. A variable-speed peristaltic pump was utilized to feed slurry into the cell tank at a preset rate just above the impeller. Another variable-speed peristaltic pump was used to remove tailings at a controlled rate from the

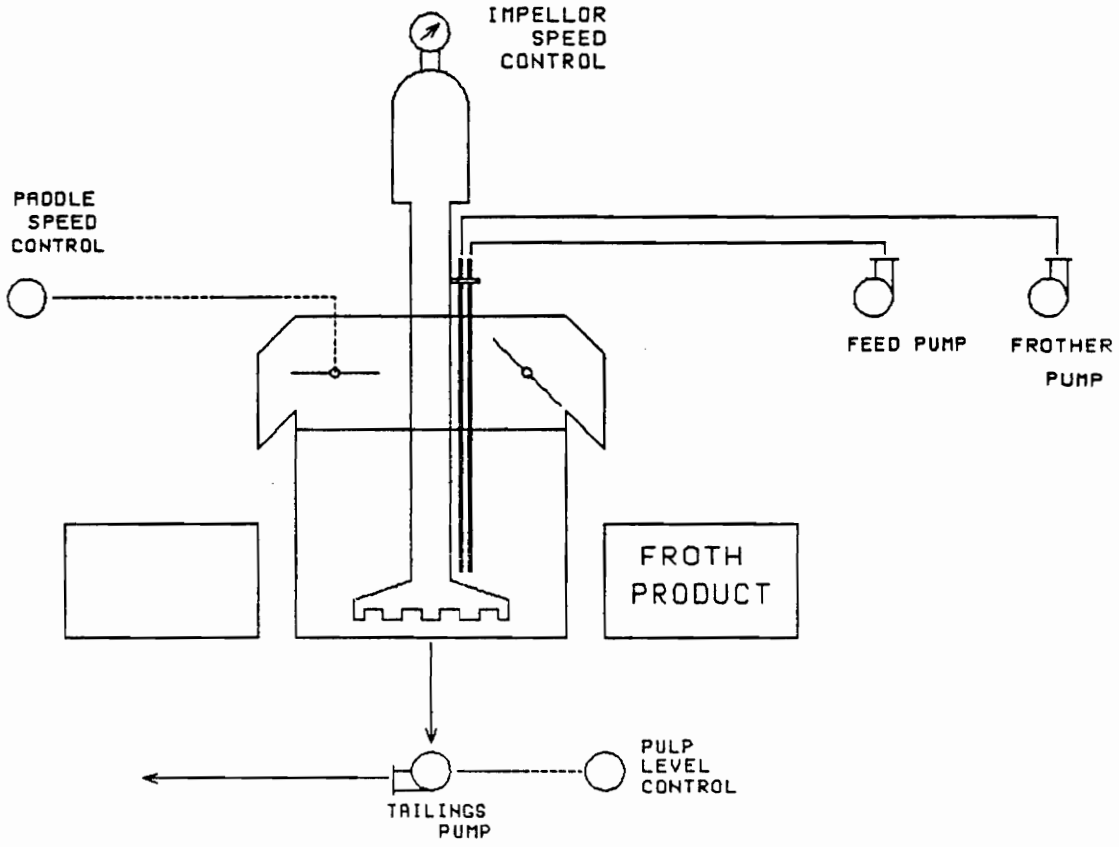


Figure 8. Schematic of the continuous conventional flotation machine.

cell tank to maintain the froth-pulp interface at a constant level in the tank. A third variable-speed peristaltic pump was utilized to add Dowfroth 1012 solution to the cell at a predetermined rate. Sufficient frother solution was added adjacent to the feed line entrance just above the impeller to maintain the desired frothing characteristics. A 3 gallon sump with attached mixer and recirculation pump was used to store the feed slurry prior to flotation.

The 4.5-liter capacity plexiglass cell was fitted with froth collecting paddles on each side that were automatically rotated using a belt connection to a variable speed motor drive. The purpose of this initial test work with coal A was to examine the differences between a semi-continuous conventional cell and the results that could be obtained using a single-stage column cell.

3.1.5 Circuit Simulator

The circuit simulator routine developed in the present work was designed with an extended amount of flexibility in mind. Although primary set-up for this program was for column flotation cells, it may also be employed for the analysis of conventional flotation circuits with only minor modifications.

The simulator routine was compiled in Turbo-Basic language to make it user friendly. This programming language also appeared to be very capable on a mathematical basis for analyzing complex circuitry. To assist ease of operation, a specialized page editor subroutine was included as part of the overall program and found

quite appropriate for the entry of input information and the retrieval of calculated results.

Input information necessary for simulator operation included three types of information. First, the feed characteristics must be entered. These include mass feed rate to the selected circuit, feed component types, component grades, weight percent for each component, component flotation rates, and solids density of the feed stream. Second, information concerning the operational characteristics of the various columns or stages within the circuit must be entered. Specific data included column diameter, length, air flow rate, air hold-up, counter-current wash water flow rate, and the Peclet number (representation of mixing profile). Finally, the information regarding the circuit arrangement must be entered. This included a description of the flow paths of concentrate and tailings streams to and from the various stages and the points of exit from the circuit. Each type of input information was contained on its own page with the simulator fully capable of altering these values on a dynamic basis prior to each simulation.

Calculated output information was also designed for presentation on a paged basis, each stage having its own page as well as a page to present overall circuit performance. Specific output values presented included mass yield, combustible recovery, concentrate and tailings grade, concentrate and tailings solids density, and cell retention time. In all of the simulation studies reported in the present work, product grade was usually represented by the product ash content. However, it should be mentioned that other parameters, such as sulfur content, heating value, etc., could have been used to represent

the product grade if so desired.

The heart of the circuit simulator was the recovery equation as given by Levenspiel (1972). A subroutine was placed in the simulator that would allow the use of this equation for the calculation of recovery for each feed component on a column-by-column basis. This recovery calculation was utilized, within the subroutine, in conjunction with an iteration loop for the mass balancing of each column or stage much the same as the procedure outlined by Lynch (1981).

Initially, for an individual column or stage, concentrate flow was assumed to be zero. Tailings flow rate was set equal to the feed flow rate plus any counter-current wash water assigned to the tailings stream by an empirical formula. From this assumption, and considering fractional air hold-up and column geometry, a retention time value was calculated. Using this retention time value, Levenspiel's equation was used to obtain a calculated concentrate flow rate. From mass balancing, a tailings flow rate was calculated and compared to the initial assumed value from above. If satisfactory convergence was not obtained, the calculated value was used as the new assumed value of tailings flow rate and the process repeated.

The calculation routine of the simulator involved the use of a mass balancing and stream routing procedure by which the selected circuit would converge. It was noted early on that the circuit balancing problem would be present when the alternatives of stream feedback and combination were selected. This problem was overcome by initializing all stream flow rates and grades to zero, going through the procedure once,

and checking for circuit flow rate convergence. If convergence was not obtained, the process was repeated.

A complete source code listing for the simulator is given in Appendix B.

3.2 Experimental Procedures

3.2.1 Single-Stage Column Tests

A single-stage column cell was used to conduct two separate series of tests with coal A. The first series of tests were performed using minus 60 mesh coal, while the second was performed using minus 200 mesh coal. For each test series, the prepared coal was placed into a 5 gallon sump, mixed, and diluted with water to the desired solids content. Mixing time in this holding sump varied from 20 to 30 minutes. The column was then brought into operation by first adjusting the air addition rate to 1.3-1.8 liters/minute. The counter-current wash water was set to 0.5 liters/minute to provide washing action for the froth zone of the column. The recirculation pump was then started and adjusted to its full speed setting of 1750 rpm.

Frother solution was prepared in a 3 gallon container by mixing Dowfroth 1012 with water to make a frother solution of 0.05% by volume. After preparing the solution, the variable-speed reagent pump was turned on and set to the desired level. This setting was arrived at by physical observation of the bubbles and their respective sizes, along with a measurement of the relative air hold-up in the column as determined by the installed pressure transducers.

Upon completion of the column start-up, the feed coal pump was started and adjusted to the desired flow rate. The operation was then allowed to come to steady-state, which was roughly 2 to 3 times the retention time of the column depending upon the selected feed rate, air hold-up, etc. As the column approached steady-state, and throughout the period of operation, water was added to the froth overflow launder to ensure that the froth product would overflow uniformly.

Sampling was accomplished for froth product and tailings samples simultaneously in order to obtain representative samples. After the collection of samples for a predetermined period, the column test conditions, in these particular tests, feed rate and relative frother dosage, were changed and the same procedure repeated until all the desired data points were obtained. Overall feed samples were collected for comparison with product analyses. These were taken from the feed sump prior to operation of the column and after the pre-mixing was completed. The samples were filtered, dried, weighed, and analyzed for ash and sulfur content as close as practical to ASTM standards. Operating conditions and settings for the two series of tests with coal A are summarized in Appendix A.

An additional series of single-stage column cell tests were performed with Coal B at a minus 325 mesh particle size. The standard operating procedures for this series of tests were very similar to what had been employed with the two series of tests with coal A. However, there was one basic difference. Preliminary testing of this coal indicated that some collector was needed in order to provide satisfactory flotation of this

particular coal. Therefore, 0.45 kg/ton of kerosene was conditioned for 10 minutes with the feed slurry following the pre-mixing stage. Again, as was practiced with coal A, a series of feed, froth product, and tailings were collected for specified periods, at each pre-selected data point, filtered, dried, weighed, and analyzed for ash and sulfur percentage as close as was possible to ASTM standards.

The single-stage column test results were used to further examine the effect of feed solids percentage on column performance, and as a part of the overall simulator evaluation program that utilized coal B. Similar to what was practiced with coal A, retention time in the column was adjusted primarily by altering feed rates during each series of tests. Operating conditions and settings for each of the three feed solids percentages examined during these test series are provided in Appendix A.

3.2.2 Semi-Continuous Conventional Cell

Figure 8 shows the standard Denver Cell which was set up to operate in a continuous mode. These tests were also conducted using a minus 60 mesh sample of coal A in order to obtain comparative data with what was obtained during single-stage column operation on the same coal.

Initially, the 4.5 liter capacity plexiglass cell was filled with water and the impeller turned on and set to a rotational speed of 900 rpm. Next, frother solution was prepared by mixing Dowfroth 1012 with water to obtain a 0.03% solution by volume in a separate 3 gallon container. Feed slurry was prepared in a 5 gallon sump by mixing

and diluting with water to obtain 5.0% solids by weight. The feed slurry conditioning time during the preparation stage was approximately 30 minutes.

After the preparation steps were completed, the reagent pump was turned on and adjusted to the desired flow rate to obtain the desired frothing action. Proper frother dosage was determined from the physical appearance of the bubbles in the froth zone of the tank. The mechanical paddles were then started and set to a top speed of 60 rpm by adjusting the speed control on the servo-controlled motor. The feed slurry pump was then started and adjusted to the desired speed depending upon the particular feed rate being investigated. The operation was then allowed to come to steady-state and normally required a period of time equal to 2-3 times the retention time.

Froth product and tailings samples were collected during specified periods for each data point examined. After collection, the samples were filtered, dried, weighed, and subjected to ash and sulfur analyses as close as practical to ASTM standards. Operating conditions and settings for these particular flotation tests are summarized in Appendix A.

3.2.3 Release Analysis

In order to characterize the various components in the feed coal, a series of tests were performed using release analysis technique. An initial evaluation of the applicability of the release analysis technique was carried out by performing a preliminary test with coal A at a 200 mesh x 0 particle size. The basic Denver Model

D-12 batch cell as described earlier was utilized to carry out this investigation.

Initially, the pulverized coal slurry was diluted and mixed with water to obtain a feed solids content of 5.0% by weight. A conditioning time of approximately 15 minutes was used for all tests. The coal slurry was transferred to the 4.5 liter cell and conditioned with 0.38 ml of Dowfroth 1012 frother for 2 minutes. The speed setting for the impeller was adjusted to 1200 rpm for all the conditioning and flotation stages.

Flotation during the first stage of this release test was allowed to proceed for a total of 14 minutes, at which time it appeared that practically all the floatable material had been recovered. During the course of this flotation stage, 0.05 ml of Dowfroth 1012 was added at the end of 9 minutes of collection to replenish any frother which had been depleted. After this initial flotation to apparent exhaustion, the collected froth product was re-slurried and floated for another 14 minutes. An additional 0.05 ml of Dowfroth 1012 was added at the end of the 10 minutes of flotation. This procedure was repeated for a third stage using the collected froth product from the second stage as feed. The non-floatable material collected as tailings from all three stages was combined into a single sample to represent the non-floatable fraction.

Upon completion of the release flotation, the final froth product was re-slurried and subjected to a kinetic analysis. Increments of the froth product were collected at the completion of 1, 3, 6, 10 and 14 minutes, respectively. Frother was replenished by adding 0.05 ml of Dowfroth 1012 at the end of 10 minutes of flotation. The froth product samples, the tailings remaining at the end of 14 minutes, and the overall non-

floatable fractions were dried, filtered, weighed, and analyzed for ash and sulfur as close as possible to ASTM standards. Operating conditions for this release study are as given in Appendix A.

Release analysis testing was also performed with coal B at a 325 mesh x 0 particle size. The test procedure was nearly identical to what was used with coal A except for the variation of flotation times and the addition of collector to better enhance the flotation response as compared to coal A. After the pre-mixing step was completed, 0.45 kg/ton of kerosene was conditioned with the feed slurry prior to addition of frother. In this case, it appeared that three stages of cleaner flotation were sufficient, followed by a kinetic recovery test to complete the analysis. Total collection for the three stages of release flotation were 26, 30 and 21 minutes, respectively. No frother was added during these three stages as had been practiced with coal A as it appeared the replenishment of frother was not necessary.

The final floatable portion of the feed material was collected and re-slurried following the third stage of release flotation. The floated product was subjected to kinetic evaluation as was practiced with coal A. The actual collection times of froth product during this analysis were at 2, 4, 8, 12 and 20 minutes, respectively. These froth products, the final tailings, and the overall sample of non-floatable material remaining after the release flotation tests, were filtered, dried, weighed, and assayed for ash and sulfur as close to ASTM standards as was possible. Operating conditions and flotation times for these release analysis tests are as shown in Appendix A.

3.2.4 Semi-Continuous Column Circuits

The three separate column circuits were operated utilizing Coal B at a 325 mesh x 0 particle size, These tests were conducted to obtain actual data for comparison with the developed simulator and to observe the performance of the columns in a three-stage circuit set-up. In these tests, the feed slurry was conditioned with 0.45 kg/ton kerosene for 10 minutes prior to feeding into the rougher stage of the circuits. The operation of each column, within each circuit, during the course of these tests was identical to the procedure for single-stage operation. For each circuit tested, i.e., R-S-C, R-S-C (Modified) and R-C-RC, three identical feed rates were examined. The operating conditions and settings with each circuit evaluation test are summarized in Appendix A.

Before taking any samples during the circuit operation, enough time had to be allowed for the particular column being analyzed to come to steady-state. This required that the preceding column stage(s) feeding this column would be at steady-state. Also, enough time had to be allowed to ensure that ample feed slurry was available in the column's feed sump before it could be put into operation. Considering these operational factors, it was decided to devote the entire course of one test to a single feed rate.

At pre-specified times, the circuit column stages were sampled, simultaneously where practical, with froth product tailings samples cut out at the same time for each particular column. Feed samples were taken directly from the sump feeding the particular column being sampled. Also, an overall feed sample was taken from the circuit feed sump that supplied the raw feed in each of the first columns. All samples

were filtered, dried, weighed, and analyzed for ash and sulfur percentage as close to ASTM standards as practical. The operating conditions, feed rates examined, and overall settings for all circuits are included in Appendix A.

4.0 EXPERIMENTAL RESULTS

4.1 Single-Stage Column Flotation

Initial experimental work was performed on Coal A (60 mesh x 0), Coal A (200 mesh x 0) and Coal B (325 mesh x 0) using single-stage column flotation. The evaluation of Coal A, 60 mesh x 0 and 200 mesh x 0 particle sizes, was to obtain preliminary data on the column flotation response of this particular coal. The single-stage evaluation of Coal B, 325 mesh x 0 particle size, was to assist in the determination of necessary input data for the simulator, and to further the comparative study of simulated results versus actual three-stage circuit results.

4.1.1 Coal A, 60 mesh x 0 Particle Size

Figures 9, 10, and 11 show the percent yield versus retention time, percent yield versus froth product ash content, and percent yield versus total froth product sulfur percentage, respectively, for the single-stage flotation results of Coal A, 60 mesh X 0 particle size. As can be seen, the analysis of the flotation product samples was broken down into a size-by-size analysis, with the product samples screened at size classes of +100 mesh, 100 x 200 mesh, 200 x 270 mesh, 270 x 400 mesh, and -400 mesh.

During this series of tests, the column was set at an air flow rate of 1200 ml/minute and counter-current wash water flow rate at 500 ml/minute. A post-test evaluation of frother dosage, based on product sample weights, revealed an average of 1.26 kg/ton. The retention time within the column was varied by primarily adjusting the

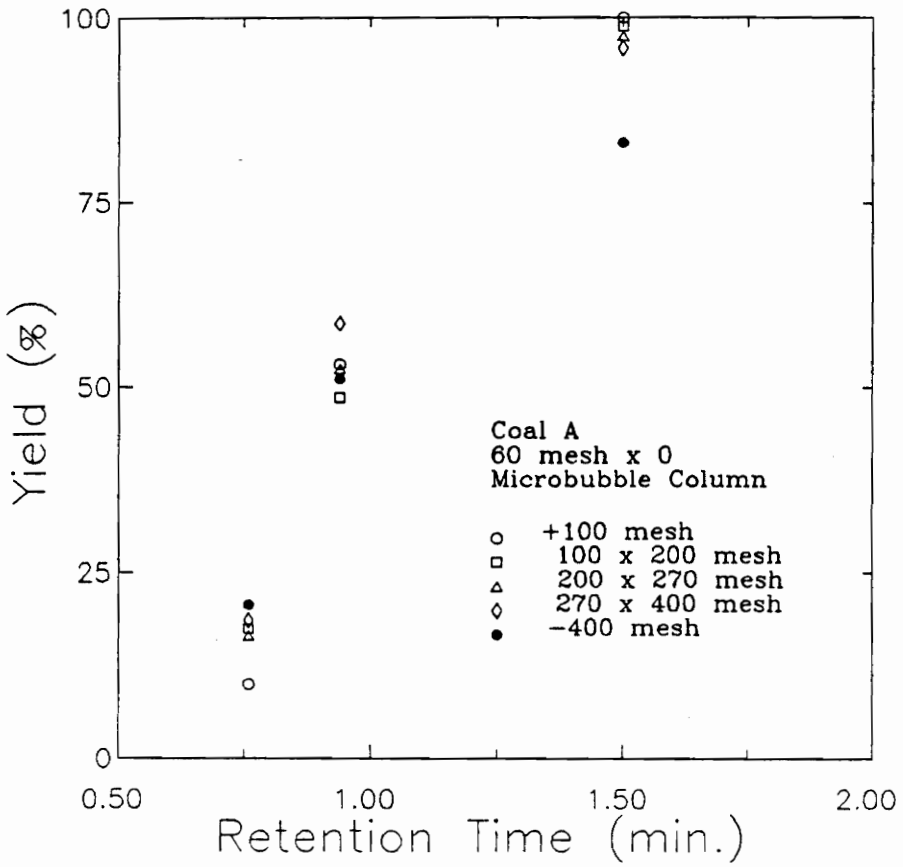


Figure 9. Clean coal yield versus mean retention time for the column flotation of Coal A (60 mesh x 0).

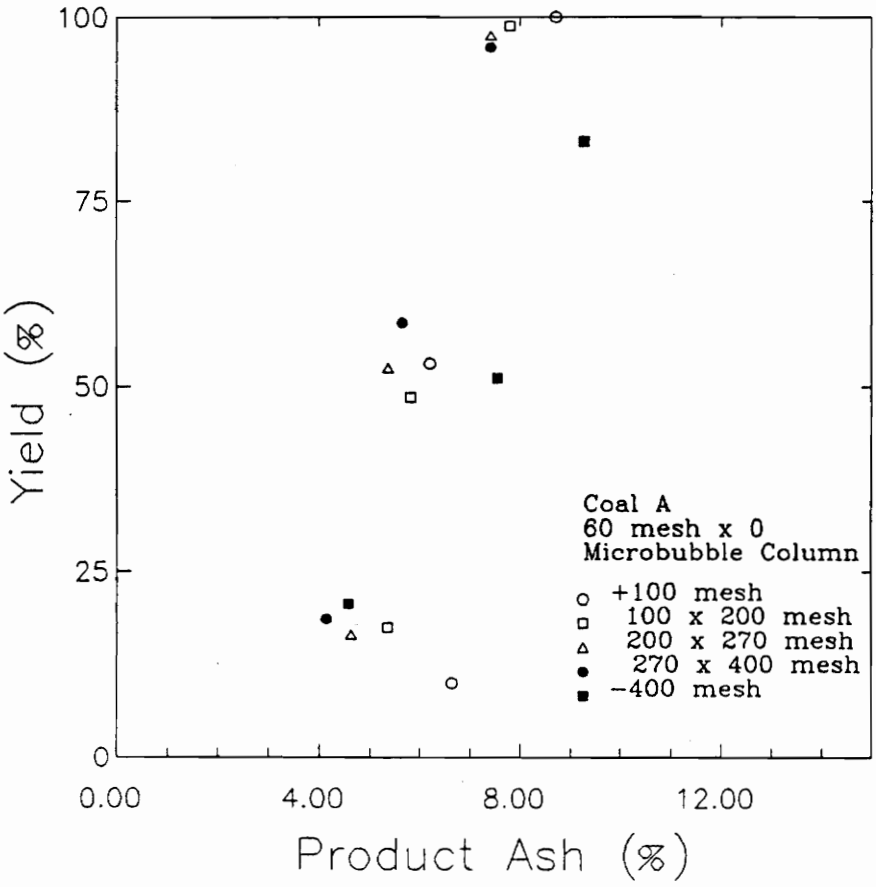


Figure 10. Clean coal yield versus clean coal ash content for the column flotation of Coal A (60 mesh x 0).

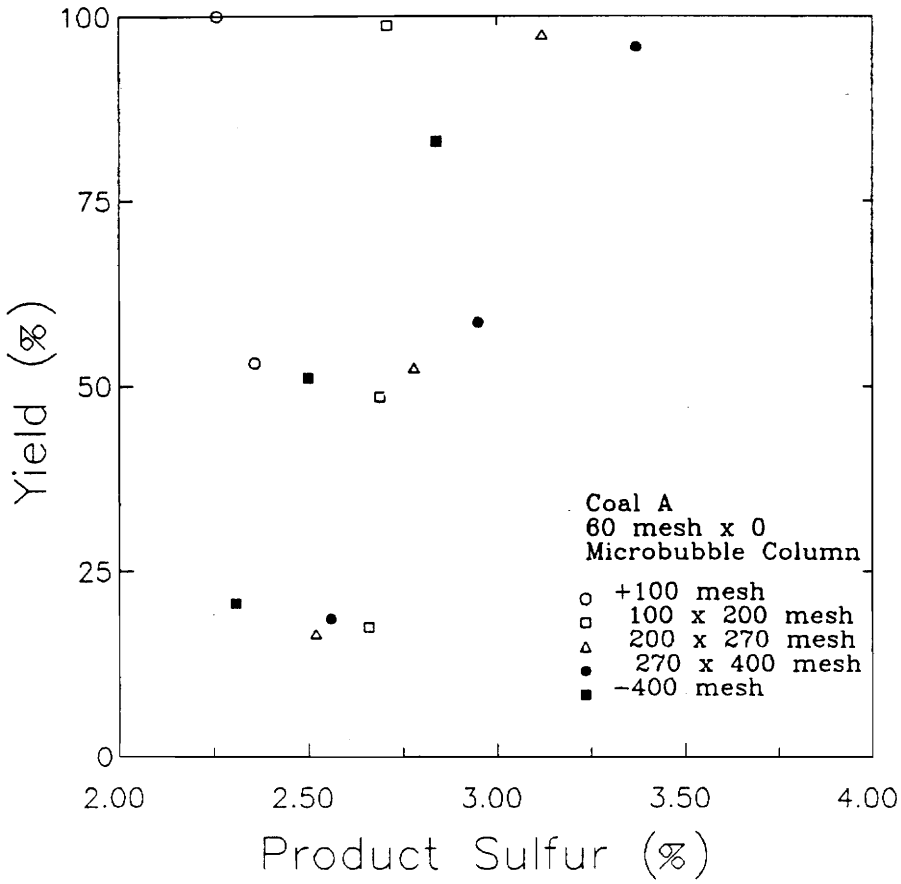


Figure 11. Clean coal yield versus clean coal total sulfur content for the column flotation of Coal A (60 mesh x 0).

feed flow rate from 106-1500 ml/minute. Feed slurry was introduced into the column at a point 16-inch below the froth overflow lip during the course of these tests. Froth height and counter-current wash water addition were maintained at 12-inch and 6-inch below the overflow lip, respectively. Feed solids percentage for all retention times investigated was around 5.0%. A detailed tabulation of test conditions is as shown in Appendix A.

As can be seen from Figure 9, the different particle size classes exhibited different flotation responses. It appeared that the fastest recoverable species existing in the feed was the 270 X 400 mesh fraction. This class was followed by the 200 x 270 mesh class, 100 x 200 mesh class, +100 mesh class, and -400 mesh class, respectively. Furthermore, from Figure 10 it seemed that this same trend was exhibited with lowest to highest froth product ash content.

Thus, it appeared that the coarser +100 mesh and 100 x 200 mesh size fractions were not as well liberated upon pulverizing to 60 mesh top-size as the following fractions were. Another factor present could be that these two upper-size class fractions contained larger particles, and thus, were more difficult to recover due to the larger size.

The -400 mesh size class was the slowest recoverable size class of the fractions examined. Also, it appeared to be the most impure with regard to froth product ash content. This result was probably due to the presence of slimes and coatings, much of which could actually have been created during the size reduction process from the particle size classes existing above this class.

Overall, the single-stage flotation response of the 60 mesh x 0 sized Coal A did not give a typical flotation retention time response. As can be seen from Figure 9, the initial tail of the curve dies off abruptly before reaching 0 minute retention time for all the size classes analyzed. From this, it appeared that their may be an inherent problem with some recovery mechanism taking place within the column.

4.1.2 Coal A, 200 mesh x 0 Particle Size

A second series of single-stage column flotation tests were performed with Coal A, this time, the nominal feed particle size being 200 mesh x 0 as prepared using the stirred ball mill to grind the pulverized 60 mesh x 0 feed coal. Again, as with the 60 mesh x 0 column flotation results, it appeared that some effect of bubble loading, available air, etc., was influencing the final flotation response. With this in mind, it was decided to perform two sets of retention time tests, the first at 5.0% feed solids, and the second at 2.0% feed solids to check if the problem might be alleviated.

During the course of these two sets of tests, retention time was varied by primarily adjusting the feed slurry flow rate from 150-600 ml/minute Froth height was maintained at a constant 12-inch level below the froth overflow lip. Post-test evaluation of frother dosage indicated average usage of 1.41 kg/ton for the 5.0% feed solids set, and 0.77 kg/ton for the 2.0% feed solids set, respectively. A detailed tabulation of test conditions is as shown in Appendix A.

Figure 12 gives the flotation response of percent yield versus retention time for

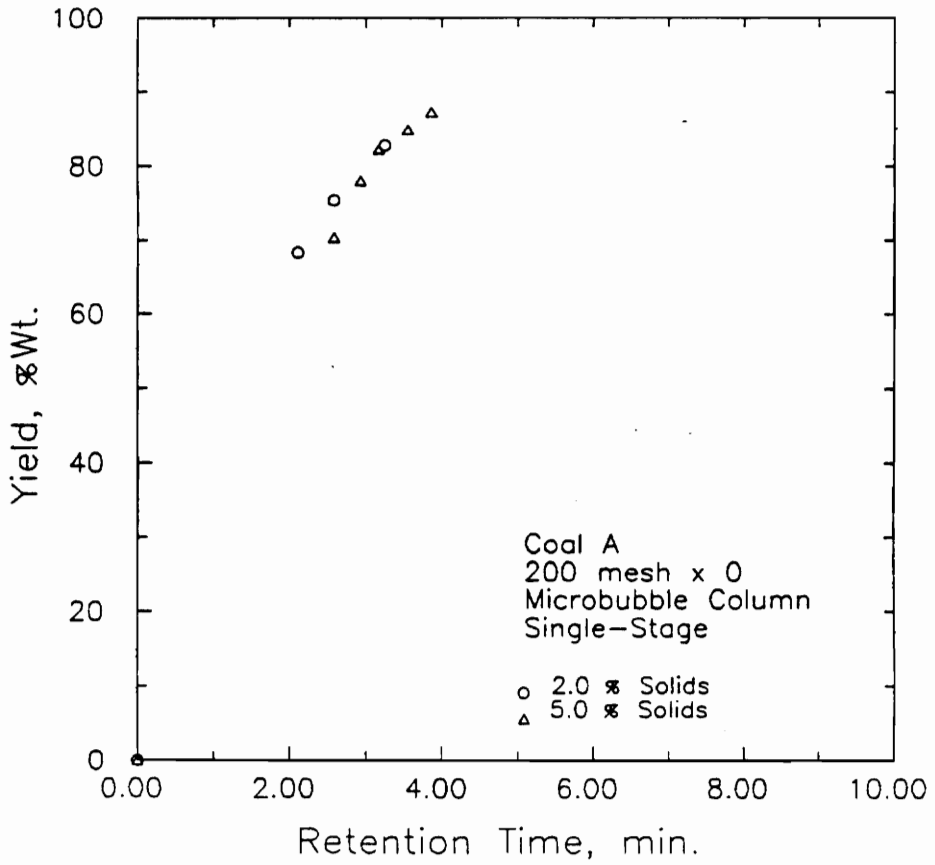


Figure 12. Clean coal yield versus mean retention time for the column flotation of Coal A (200 mesh x 0).

both feed solids percentages studied. As can be seen from Figure 12, the 5.0% feed solids set gave a result very similar to what had been observed with the 60 mesh x 0 single-stage flotation. Again, the initial tail of the curve dipped down prematurely before reaching 0% yield at 0 minute retention time. With this in mind, it was decided to increase air flow rate from 1300-1500 ml/minute for a second set of flotation tests, and to lower feed solids to 2.0% to prevent any possible froth overloading problems.

As can be seen from Figure 12, the flotation performance, with regard to retention time, is somewhat improved with 2.0% feed solids as compared to 5.0% feed solids. The initial tail of the curve approaches 0% yield at 0 min retention time as traditional results have shown. Figures 13 and 14 give percent yield versus froth product ash content for the 5.0% feed solids and 2.0% feed solids tests, respectively. Also, it appeared that a lower froth product ash was obtainable at 2.0% feed solids. Thus, there appeared to definitely be a limit to bubble loading or carrying capacity inherent in the recovery mechanisms taking place in a column.

Observing Figure 13, it appeared that a possible liberation limit, or wall, had been encountered at the lower levels of percent yield for the 5.0% feed solids material, at a froth product ash content of around 4.00%. However, at 2.0% feed solids, this limit was reduced to the neighborhood of 3.50% froth product ash, with a more gradual slope, indicating less probability of liberation problems.

Even though there did seem to be an inherent problem with the column recovery mechanism(s) at the higher feed solids percentage, it would appear that there would be

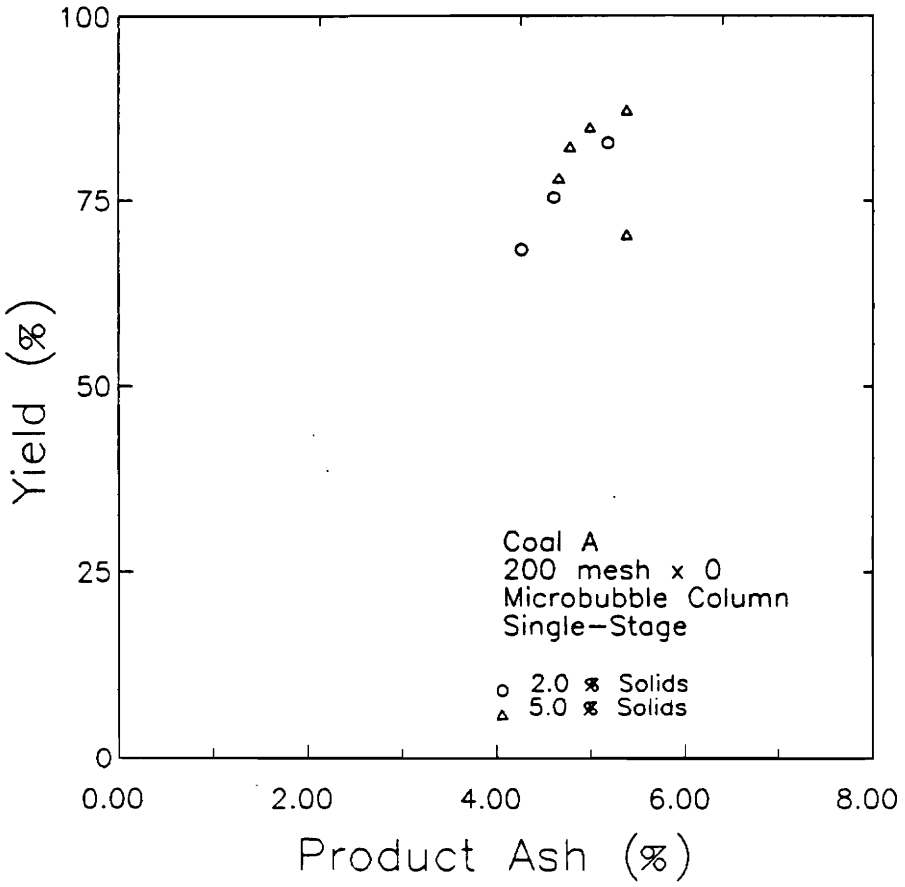


Figure 13. Clean coal yield versus clean coal ash for the column flotation of Coal A (200 mesh x 0).

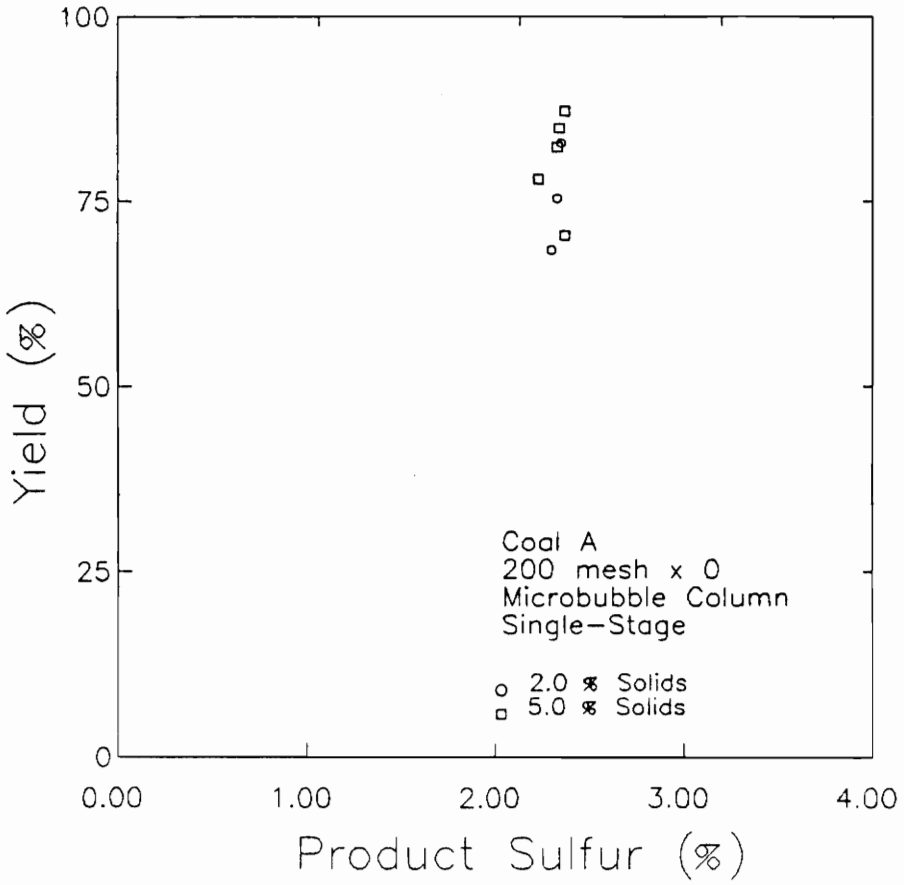


Figure 14. Clean coal yield versus clean coal total sulfur content for the column flotation of Coal A (200 mesh x 0).

trade-off point where recovery and grade would be balanced with overall throughput.

4.1.3 Coal B, 325 mesh x 0 Particle Size

The final series of single-stage column flotation tests was carried out using Coal B, 325 mesh x 0 particle size, prepared by grinding the 60 mesh x 0 pulverized feed coal in the stirred ball mill. This series of tests was another evaluation of froth product yield versus retention time and ash content at three separate feed solids percentages, 2.0%, 5.0%, and 10.0%. Retention time was varied by primarily adjusting feed slurry flow rate from 100-600 ml/minute. Air flow rate and counter-current wash water flow rate were 1500 ml/minute and 400 ml/minute, respectively. Kerosene collector was pre-conditioned with the feed slurry to a level of 0.45 kg/ton throughout these three sets of tests. Post-test evaluation of frother dosage showed consumption in the range of 0.29-2.19 kg/ton. Froth height relative to the product overflow lip was 12- to 15-inches during the investigation. A detailed listing of the operating conditions is as shown in Appendix A.

As can be seen from Figure 15, the lower feed solids, 2.0%, gave flotation performance nearest to the traditional expectation, with the 10.0% feed solids set giving the most erratic response with respect to percent yield and retention time. Definitely, there did appear to be a problem with carrying capacity. From Table III, it appeared that the best overall quality was obtained at 10.0% feed solids, however, percent yield was substantially lower. From this series of tests involving Coal B, 325 mesh x 0 particle

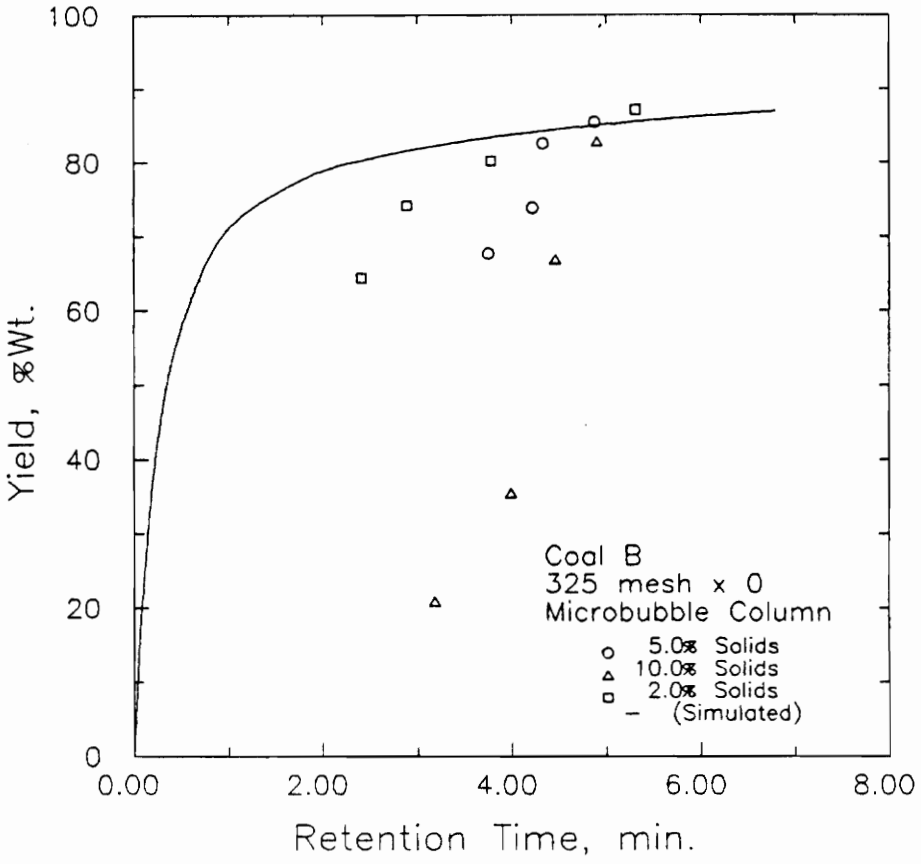


Figure 15. Clean coal yield versus mean retention time as a function of feed solids content for the column flotation of Coal B (325 mesh x 0).

Table III. Column flotation results obtained for Coal B after pulverization to a topsize of 325 mesh.

Feed Solids Content	Time (min.)	Yield (%)	Ash (%)	Sulfur (%)
2.0%	2.41	69.3	2.91	3.16
	2.89	77.6	3.09	3.17
	3.78	82.6	3.26	3.26
	5.31	87.1	3.10	3.13
5.0%	3.75	67.7	3.56	3.29
	4.22	74.6	3.37	3.25
	4.33	82.6	3.57	3.29
	4.87	85.5	3.58	3.28
10.0%	3.18	20.9	2.39	3.15
	3.99	34.9	2.53	3.16
	4.46	66.9	3.22	3.29
	4.90	82.9	3.31	3.33

Notes:

- Percentage ash and sulfur reported as cumulative values.
- Sulfur content measured as total sulfur.

size, it was decided to proceed with column circuit testing at a general feed solids level of 5.0% to check how actual values compare to simulated results. This would further stress the importance of making trade-offs between yield and overall throughput at a level of feed solids that appeared most practical.

4.2 Semi-Continuous Conventional Cell

A series of tests was performed with Coal A, 60 mesh x 0 particle size, using the semi-continuous Denver cell as previously described in Section 3.1.4. The purpose of this analysis was to obtain a comparative correlation between conventional single-stage flotation and single-stage flotation results that were being gathered. The test results obtained with the semi-continuous Denver cell were as shown in Figures 16 and 17, which give percent yield versus retention time and percent yield versus froth product ash content, respectively. Retention time was varied during these particular tests by altering the feed slurry flow rate to the cell in the range 500-2000 ml/minute The percent solids of the feed slurry was maintained constant at 5.0%. A rotational impeller speed of 1200 rpm was maintained during the course of these tests. Frother consumption was adjusted to keep an average dosage of 1.50 kg/ton.

From Figure 16, it appeared that a somewhat different result was obtained with the single-stage conventional cell as compared to the results of the single-stage column operating on the same type of feed. In these tests, the flotation retention time response exhibited a more classical shape, in that, the initial tail seemed to approach 0% yield at

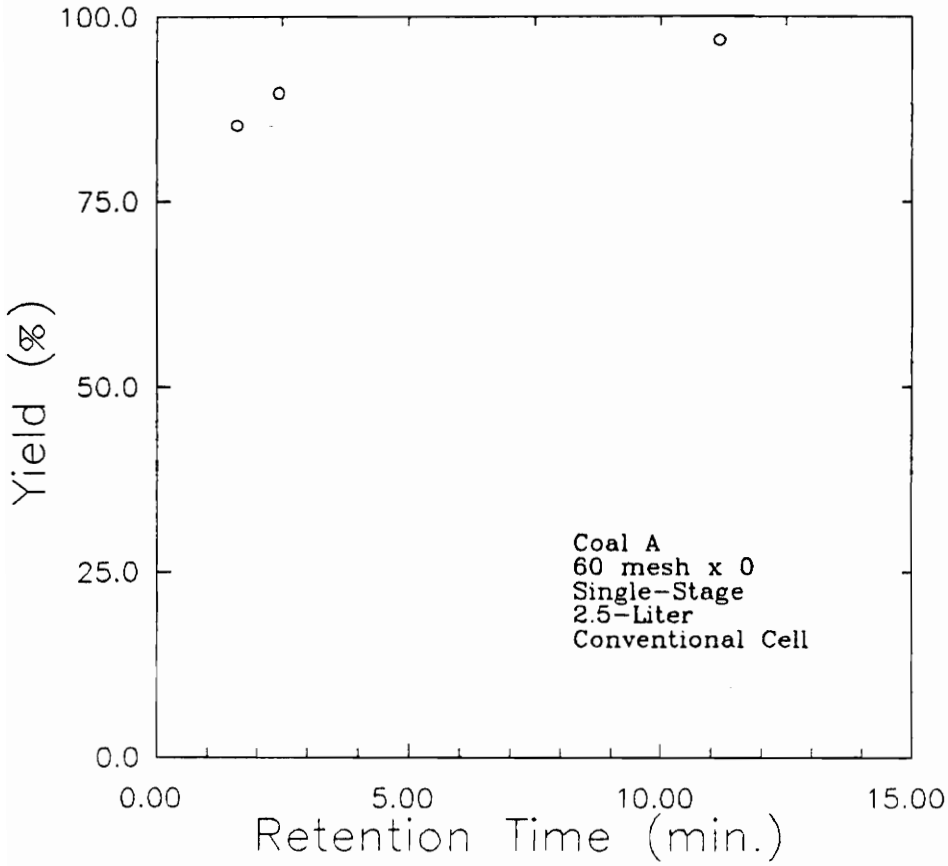


Figure 16. Clean coal yield versus mean retention time for the conventional flotation of Coal A (60 mesh x 0).

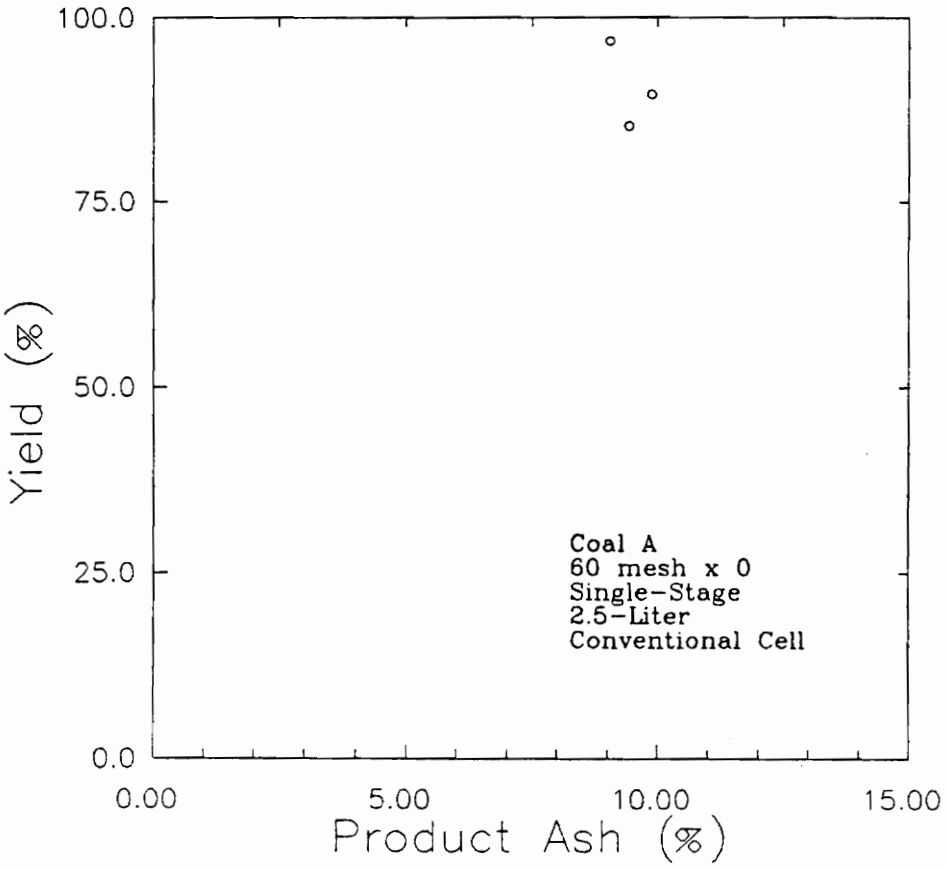


Figure 17. Clean coal yield versus clean coal ash content for the conventional flotation of Coal A (60 mesh x 0).

0 minute retention time. This observation appeared to lend support to the premise that a carrying capacity problem in fact does exist with the recovery mechanism(s) that exist in column flotation.

However, from Figure 17, the semi-continuous conventional flotation of this particular coal resulted in a lower quality froth product as compared to what was obtained during column flotation. This result would seem to confirm the fact that there is an additional secondary cleaning of the froth in a column (see Section 3.1.2).

4.2.1 Release Analysis

The technique of release analysis and kinetic flotation as previously described in Section 3.1.3 was employed to evaluate Coal A, 200 mesh x 0 particle size, and Coal B, 325 mesh x 0 particle size. The release work with Coal A was carried out in order to investigate the applicability of this approach to determine representative flotation components, component grades, and their respective rates of flotation for a particular coal. The later testing with Coal B was to identify feed components to incorporate them into the circuit simulation and evaluation program to be carried out later during the circuit analysis portion of this study as will be discussed later.

4.2.2 Coal A

Figure 18 shows a typical release curve for Coal A, 200 mesh x 0 particle size. This curve was constructed by plotting the cumulative yield of froth product versus

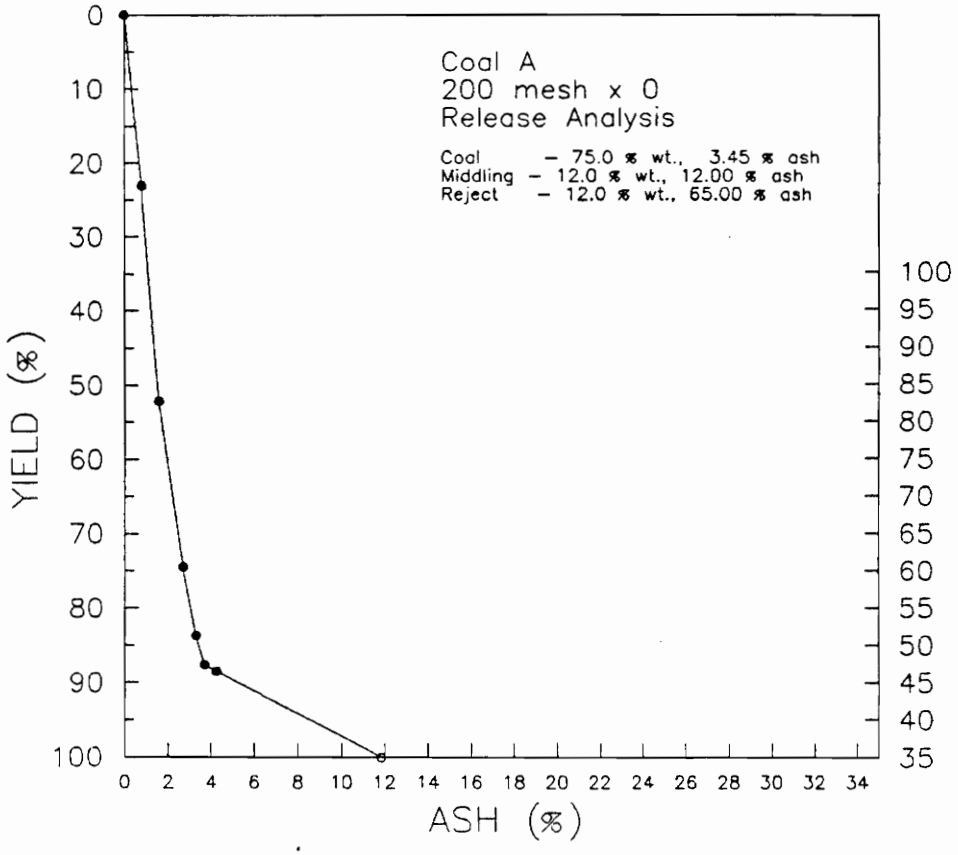


Figure 18. M-curve representation of flotation release analysis data obtained for Coal A (200 mesh x 0).

cumulative froth product ash content. Table IV gives the cumulative flotation data for the kinetic portion of the release analysis curve along with the assay of nonfloatable material (i.e., flotation tails). Experimental conditions are given in Appendix A.

It can be seen from Figure 18, that certain points of transition along the release curve could perhaps be taken to identify components of related ash content and that these could be singled out. A straight line was drawn between these apparent transition points to identify the breakdown of the overall feed into various components by relative weight percentage, which was determined readily by computing the difference in cumulative yields at these points. For this particular coal, it appeared that the feed may be represented by three components, "coal", 76.0% by weight, "middlings", 12.0 percent by weight, and "reject", 12.0% by weight.

Also, it was found by constructing another line for each component, initiating from 0% yield, and running parallel through to the defining lines of the component classes, the relative grades of the three components could be determined from the point at which this line intersects the 100% yield axis. These relative grades for the coal, middlings, and reject components were found to be 3.45%, 12.00%, and 65.00% ash, respectively.

These component values were taken and used in the simulator to identify values for flotation rate constants that might be appropriate for general simulation of the flotation behavior of this coal. It was found that the simulated results closely matched the single-stage results reported in Section 4.1.2 (i.e., Coal A, 200 mesh x 0 particle

Table IV. Release analysis results and kinetic flotation data obtained for Coal A at a 200 mesh topsize.

Product	Time (min.)	Yield (%)	Ash (%)
Float #1	1	23.02	3.12
Float #2	3	52.27	3.35
Float #3	6	74.66	3.74
Float #4	10	83.74	4.03
Float #5	14	87.63	4.28
Tail #5	14	(1.00)	(50.90)
Non-Float	--	(11.38)	(66.16)

Notes:

- Feed Solids = 5.0%
- Impeller Speed = 1200 rpm
- Release Stages = 3
- Float #1-#5 reported as cumulative values.
- Tail #5 and Non-Float reported as individual values.

size, 5.0% solids) when rate constant values of 2.84, 0.75, and 0.01 min^{-1} were used to represent the coal, middlings, and reject components, respectively (see Table V). Thus, these values were believed to be acceptable for this particular coal.

Thus, it appeared that a suitable technique had been found by which components, associated grades, and respective flotation rate constants may be found using release analysis and utilized in the simulator and circuit analysis work that was to finalize this program.

4.2.3 Coal B

Another release analysis was carried out on Coal B, 325 mesh x 0 particle size, using the procedure as described in Section 3.2.3. The set-points and operating conditions are as given in tabular form in Appendix A. The feed slurry was pre-conditioned in the Denver cell prior to actual flotation with 0.45 kg/ton kerosene collector to enhance flotation performance as indicted from preliminary test work.

Table VI gives the cumulative yield percentage versus cumulative product ash content at the respective flotation times used in the kinetic portion of this analysis. Figure 19 gives the release curve obtained from this information. As can be seen, as with the work mentioned under Section 4.3.1 with Coal A, there appeared to be three components present, these similarly named "coal", "middlings", and "reject". By using a similar construction of tangent and parallel lines as had been done with Coal A, it was determined that the weight percentages and ash contents of these components were

Table V. Feed component characteristics for Coal A at a 200 mesh topsize.

Component	Weight (%)	Ash (%)	k (min^{-1})
Coal	76.0	3.45	2.84
Middlings	12.0	12.00	0.75
Reject	12.0	65.00	0.01

Table VI. Release analysis results and kinetic flotation data obtained for Coal B at a 325 mesh topsize.

Product	Time (min.)	Yield (%)	Ash (%)
Float #1	2	10.00	2.25
Float #2	4	21.11	2.32
Float #3	8	37.64	2.40
Float #4	12	51.86	2.55
Float #5	20	66.47	2.80
Tail #5	20	(5.97)	(7.08)
Non-Float	--	(27.56)	(26.58)

Notes:

- Feed Solids = 5.0%
- Impeller Speed = 1200 rpm
- Release Stages = 3
- Float #1-#5 reported as cumulative values.
- Tail #5 and Non-Float reported as individual values.

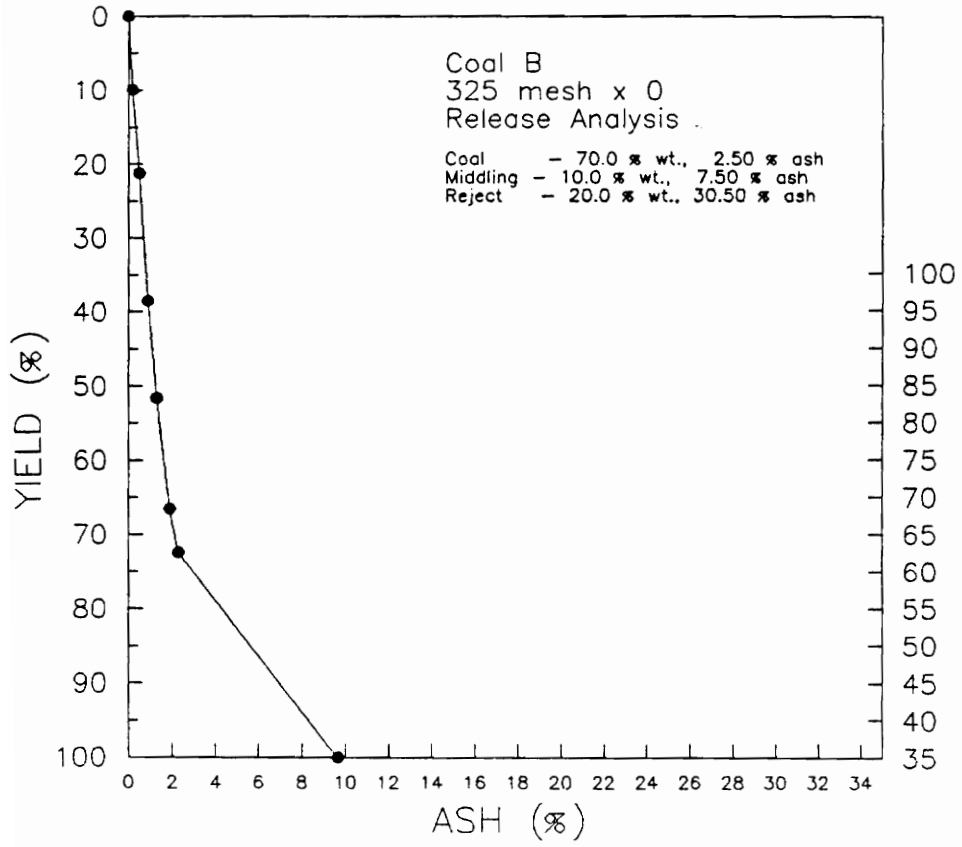


Figure 19. M-curve representation of flotation release analysis data obtained for Coal B (325 mesh x 0).

70.0%, 10.0%, and 20.0% by weight, and 2.50%, 7.50%, and 30.50% ash, respectively.

In comparison to Coal A, 200 mesh x 0 particle size, there seemed to be nearly the same amount of the first species, "coal", however, it had a lower ash content, probably owing to the greater degree of liberation obtained by grinding from 200 mesh to 325 mesh top-size. In contrast, the analysis of the non-floatable portion indicated an increased weight percentage in this component class, 10.0% greater, and a lower ash level of 30.50%. Allowing for source difference between the two samples of Pittsburgh No. 8, this effect may have been due to the fact that the recovery mechanism(s) present in the Denver cell was somewhat limited in recovery ability of the finer sizes of coal even though a greater degree of liberation was obtained.

Continuing further, the developed simulator was utilized to attempt a fit of flotation rate constant values to the as determined component classes. The best fit, when comparing simulated results with actual single-stage results with this coal, came at rate constant values of 4.50, 0.75, and 0.07 $minute^{-1}$. These values are as shown in Table VII, and as plotted alongside the actual single-stage results presented in Figure 15.

4.3 Column Circuit Tests

In order to provide comparative actual data for the computed values that were to be obtained with the simulator, three series of tests were performed with Coal B, 325 mesh x 0 particle size. These tests involved the operation of three separate circuits made

Table VII. Feed component characteristics for Coal B at a 325 mesh topsize.

Component	Weight (%)	Ash (%)	k (min^{-1})
Coal	70.0	2.50	4.50
Middlings	10.0	7.50	0.75
Reject	20.0	30.50	0.07

up of identical column flotation cells as outlined in Section 3.1.2.2. The practicality of these tests was insured, and the volume of work held to a minimum, by employing a uniform feed solids of 5.0%, by weight. In all these tests, kerosene collector, in the amount of 0.45 kg/ton, was pre-conditioned with the feed slurry prior to circuit operation in order to enhance flotation recovery.

4.3.1 Rougher-Scavenger-Cleaner

The first circuit to be investigated was the rougher-scavenger-cleaner arrangement (R-S-C), as described in Section 3.1.2.2, and shown in Figure 5. The analysis of this circuit involved operation at four separate feed slurry flow rates, 50 ml/minute, 100 ml/minute, 200 ml/minute, and 300 ml/minute These rates were obtained by adjusting the speed of the slurry feed pump going into the rougher cell for each separate feed rate that was examined. The feed flow rates to the succeeding column stages were arrived at by physically measuring the streams feeding that stage and adjusting the slurry feed pump speed accordingly.

During the course of this series of tests, air flow rate was maintained constant at 1500 ml/minute. Counter-current wash water was adjusted to a constant level of 400 ml/minute. Froth height and feed entry point were maintained at 15-inch and 15.5-inches, respectively, for all the column stages and for each feed rate examined. Retention times were calculated by obtaining tailings slurry flow rates and measuring air hold-up values. A detailed summary of this test series operating conditions is as given

in Appendix A.

Figures 20 through 23 give the analysis results for each of the respective feed rates examined in the format of simple flowsheets. As can be seen, the best overall circuit yield and recovery, respectively, were obtained at the lowest feed rate examined, 50 ml/minute This result would be expected in that the greatest amount of retention time is available at the lowest feed rate. Likewise, the lowest yield and recovery were at the highest feed flow rate, 300 ml/minute, these being 69.4% and 74.8%, respectively.

However, upon examining Figures 20 through 23, it can be seen that the overall froth product ash from the circuit remained relatively unchanged, at 2.81%, 2.91%, 3.03%, and 2.08%, for the feed slurry flow rates of 50 ml/minute, 100 ml/minute, 200 ml/minute, and 300 ml/minute, respectively. Total froth product sulfur actually increased from 2.95% to 3.26% during the increase in feed flow rate from 50 ml/minute to 300 ml/minute Overall tailings ash decreased from 32.36% to 25.69% as circuit feed flow rate was increased from 50 ml/minute to 300 ml/minute. No distinct trend could be obtained for the corresponding total sulfur content of the circuit tailings.

On an individual basis, the best rougher stage performance, with respect to yield, was at a slurry feed flow rate of 50 ml/minute, 86.2%. Froth product ash was 3.35% at this feed rate. As feed flow rate was increased from 50 ml/minute to 300 ml/minute, the rougher stage yield dropped down to 68.1%, and froth product ash decreased to 3.02%. As with the results for overall circuit performance, this seems to be what would be expected with the decrease in rougher retention time. Total froth product sulfur

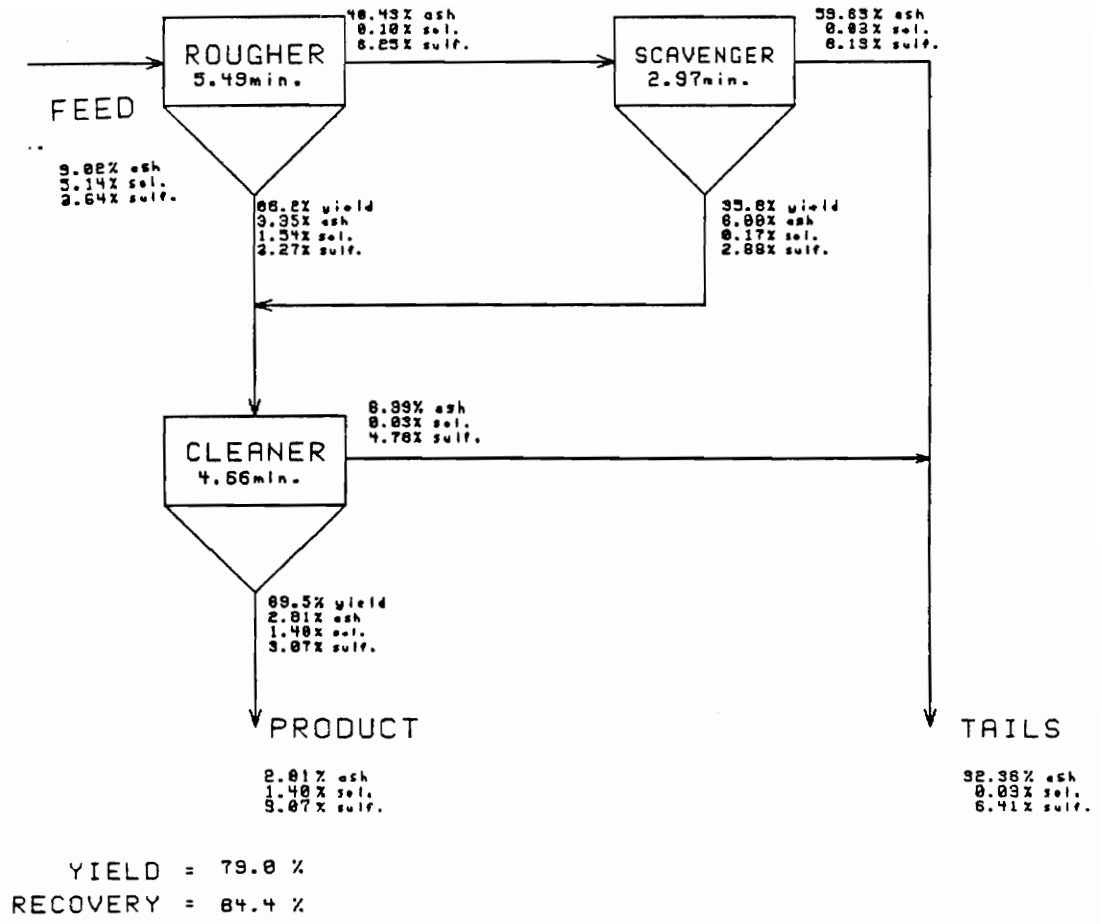
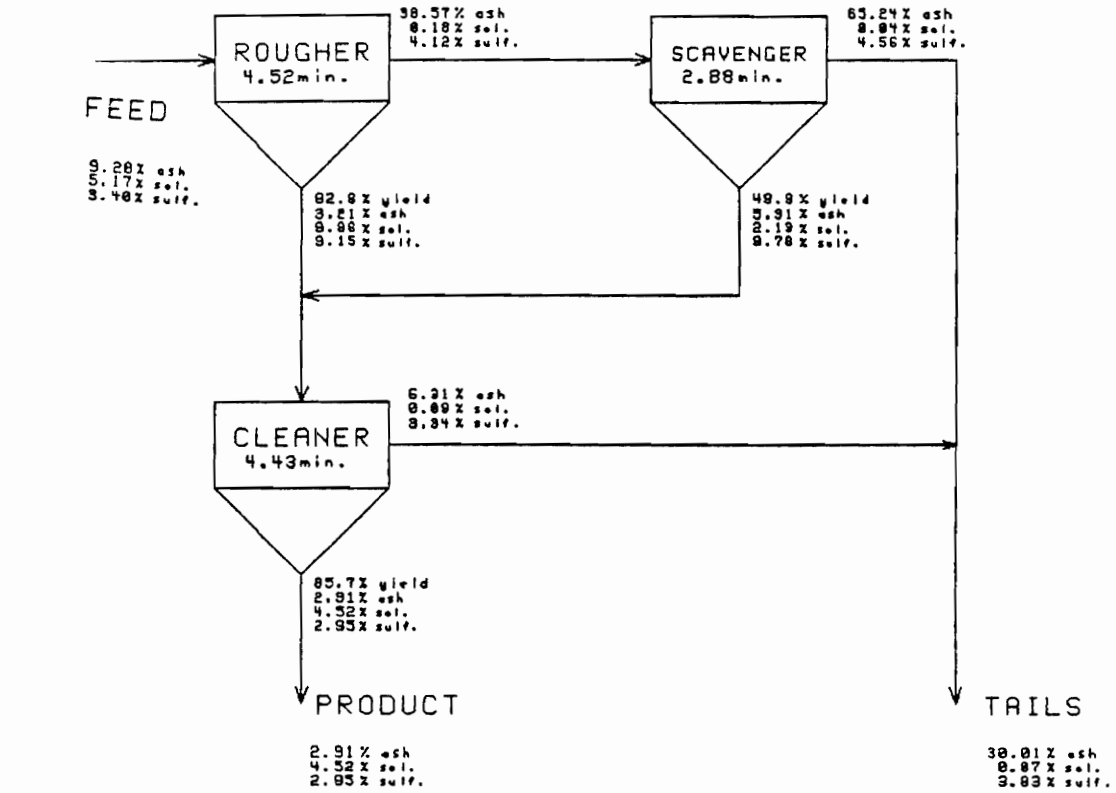


Figure 20. Results obtained for the R-S-C circuit at a 50 ml/min feed flowrate.



YIELD = 76.5 %
RECOVERY = 81.9 %

Figure 21. Results obtained for the R-S-C circuit at a 100 ml/min feed flowrate.

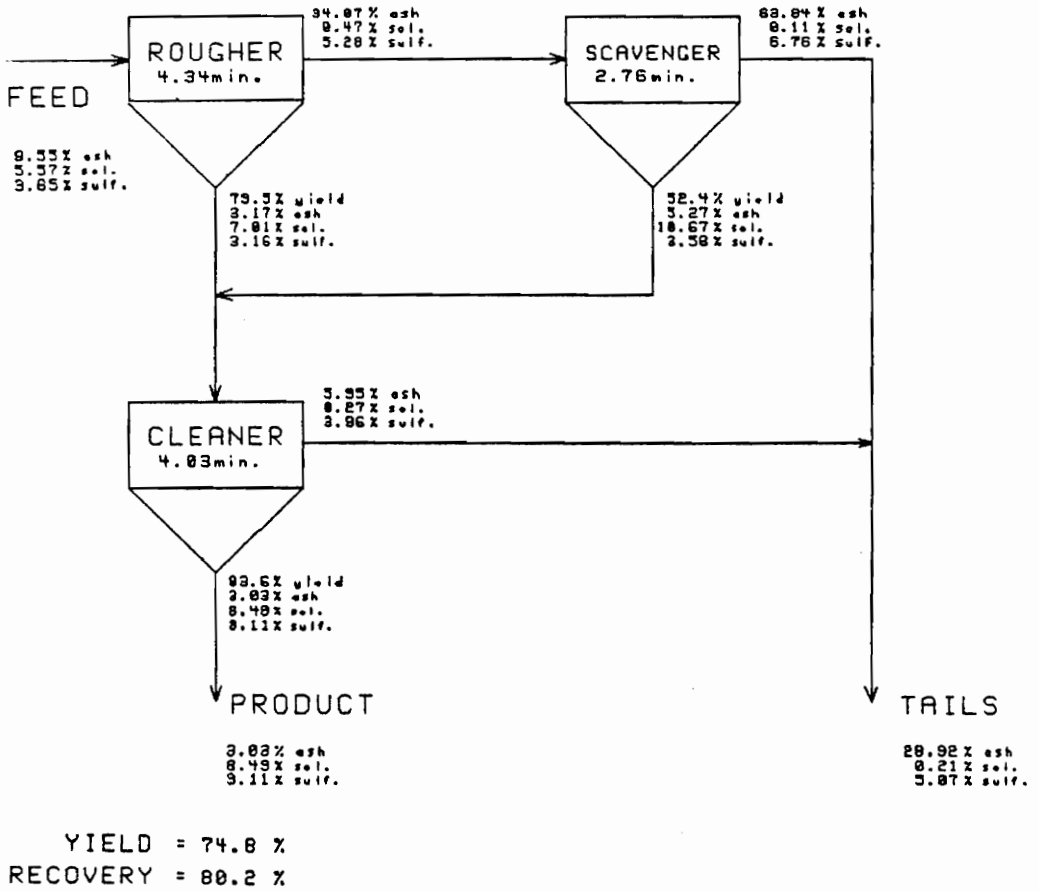


Figure 22. Results obtained for the R-S-C circuit at a 200 ml/min feed flowrate.

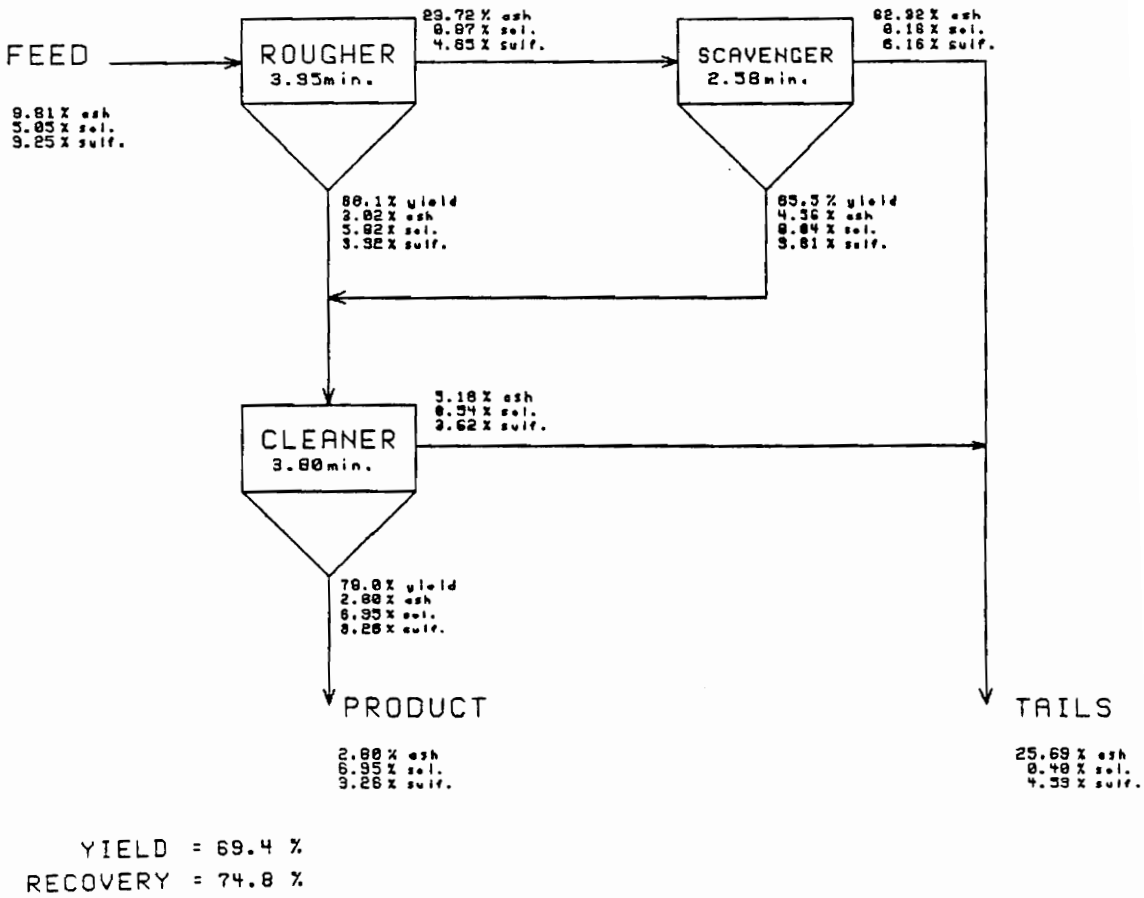


Figure 23. Results obtained for the R-S-C circuit at a 300 ml/min feed flowrate.

remained relatively the same at an average value of 3.21%.

Similarly, cleaner performance with respect to yield appeared to be at highest yield at the lowest feed flow rate of 50 ml/minute, being 89.5%. Yields for this stage were 89.5%, 85.7%, 83.6%, and 78.5%, for feed flow rates of 50 ml/minute, 100 ml/minute, 200 ml/minute, and 300 ml/minute, respectively. As mentioned previously for the overall circuit results, froth product sulfur content from the cleaner stage remained relatively unchanged.

However, the scavenger stage performance was somewhat unexpected when compared to the rougher, cleaner, and overall circuit results. The lowest scavenger yield, 35.8%, occurred at the lowest feed flow rate, 50 ml/minute Likewise, the highest yield, 65.5%, occurred at the highest circuit feed flow rate, 300 ml/minute Also, scavenger froth product ash decreased from 6.00% to 4.56% as feed flow rate to the circuit was increased from 50 ml/minute to 300 ml/minute.

It appeared that these performance results of the scavenger stage would indicate that a substantially greater amount of floatable material became available in the scavenger feed as circuit feed flow rate was increased because of the decrease in retention time in the rougher. In addition, it can be seen that the feed material being sent to the scavenger stage, as the feed flow rate was increased, was consistently higher in solids and lower in ash content. Analysis of the rougher stage tailings indicated an increase in solids from 0.10% to 0.87%, and a decrease in ash from 40.43% to 23.72%, respectively, as the feed flow rate to the circuit was increased from 50 ml/minute to 300 ml/minute. Thus,

it is readily apparent that seemingly more floatable, lower ash material was being sent to the scavenger stage as feed flow rate was increased to the circuit.

In conjunction with the observation that retention time to the rougher stage was consistently decreased as circuit feed flow rate was increased, the presence of more floatable material in the scavenger stage feed may be also due to the inherent carrying capacity problem previously mentioned with respect to the rougher stage as an ever increasing amount of more floatable material was trying to be recovered.

It appeared that the performance of the overall circuit was distinctly similar to trends being observed in the rougher stage. The performance of the cleaner stage also showed similar trends to overall circuit results, however, the range in yield variation was not as large as with the rougher stage and the overall circuit. This observation could have possibly been affected due to an increasing amount of the more floatable, lower ash material reporting to the cleaner feed from the scavenger stage as feed flow rate to the circuit was increased. The seemingly constant level of froth product ash from the cleaner stage could have been the result of a liberation limit with this size and type of coal, or resulted due to the negating effects of rougher and scavenger stage froth products. The increasing trend of total froth product sulfur from the circuit as feed flow rate was increased could have been the result of the inherent sulfur content of the floatable species or the increased floatability of liberated pyrite.

Comparing these results to the single-stage column flotation tests reported in Section 4.1.3 and shown in Figure 16, it appeared that yield was not substantially

improved using this three-stage configuration, however, froth product ash content was indeed lowered. Similar to yield, total froth product sulfur content remained relatively unchanged.

4.3.2 Rougher-Scavenger-Cleaner (Modified)

In order to check if any advantages might be gained with respect to yield and quality, a rougher-scavenger-cleaner (modified), R-S-C (modified), circuit was utilized with three column flotation cells as described in Section 3.1.2.2 and as shown in Figure 6. From the results obtained with the R-S-C circuit outlined in the previous section, it became apparent that overall circuit performance seemed to be somewhat controlled by performance in the rougher stage. Therefore, it was decided to remove the rougher leg from the cleaner path of the circuit, with only the scavenger froth product being upgraded in the cleaner column.

Figures 24 through 27 give the results from operating the R-S-C (modified) circuit on Coal B, 325 mesh x 0 particle size, presented in the format of simple flowsheets. As feed flow rate was increased from 50 ml/minute to 300 ml/minute, it can be seen that overall circuit yield and recovery remained relatively unchanged at average values of 88.8% and 95.0%, respectively. Likewise, overall circuit froth product ash and total sulfur remained much the same, at average values of 3.32% and 3.06%, respectively. Overall circuit tailings ash was 66.90%, 54.39%, 59.43%, and 58.77%, for circuit feed flow rates of 50 ml/minute, 100 ml/minute, 200 ml/minute, and 300 ml/minute,

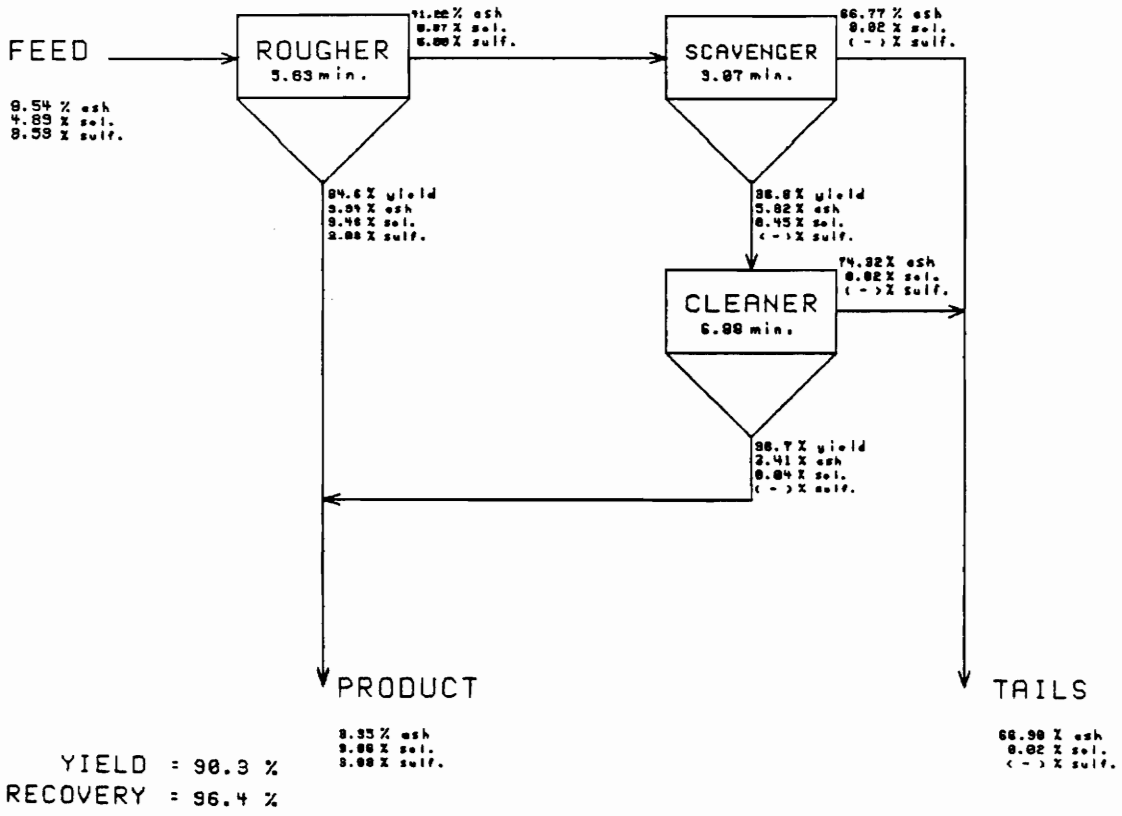
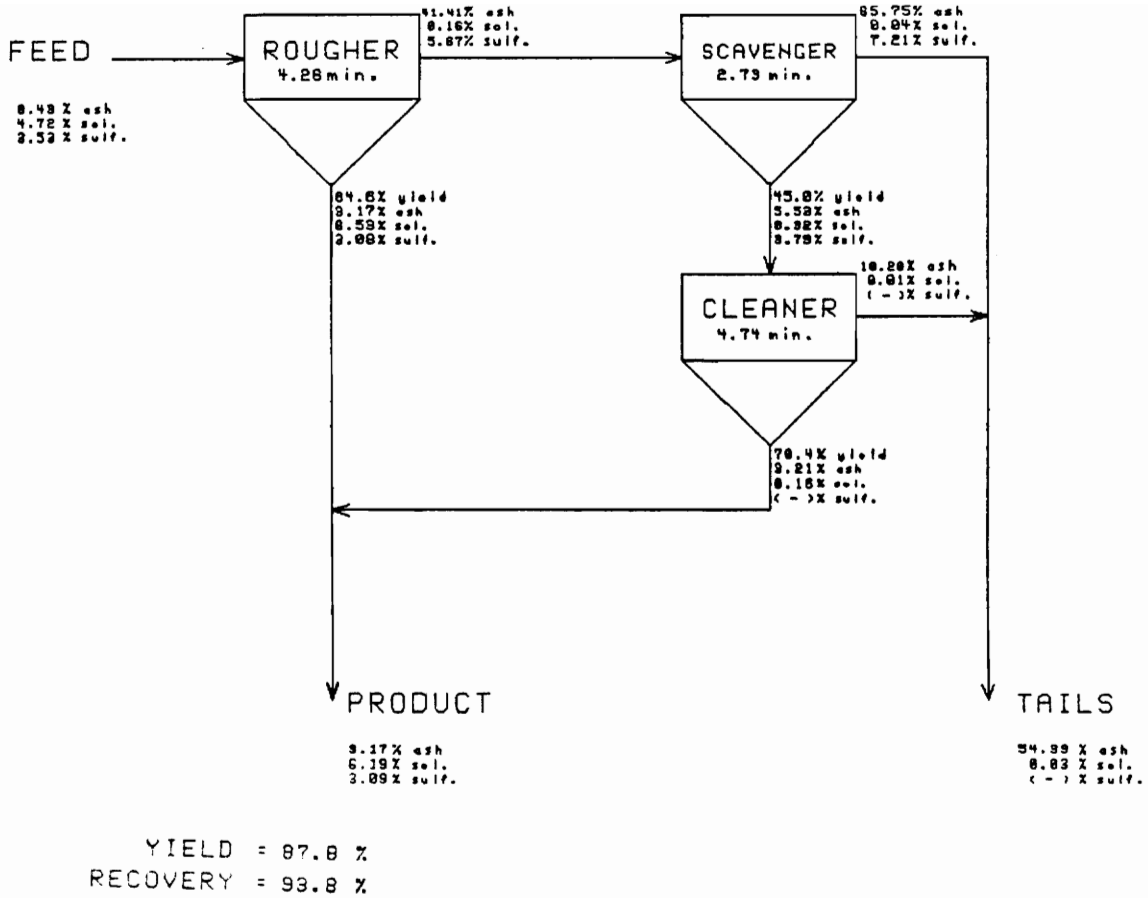


Figure 24. Results obtained for the modified R-S-C circuit at a 50 ml/min feed flowrate.



YIELD = 87.8 %
RECOVERY = 93.8 %

Figure 25. Results obtained for the modified R-S-C circuit at a 100 ml/min feed flowrate.

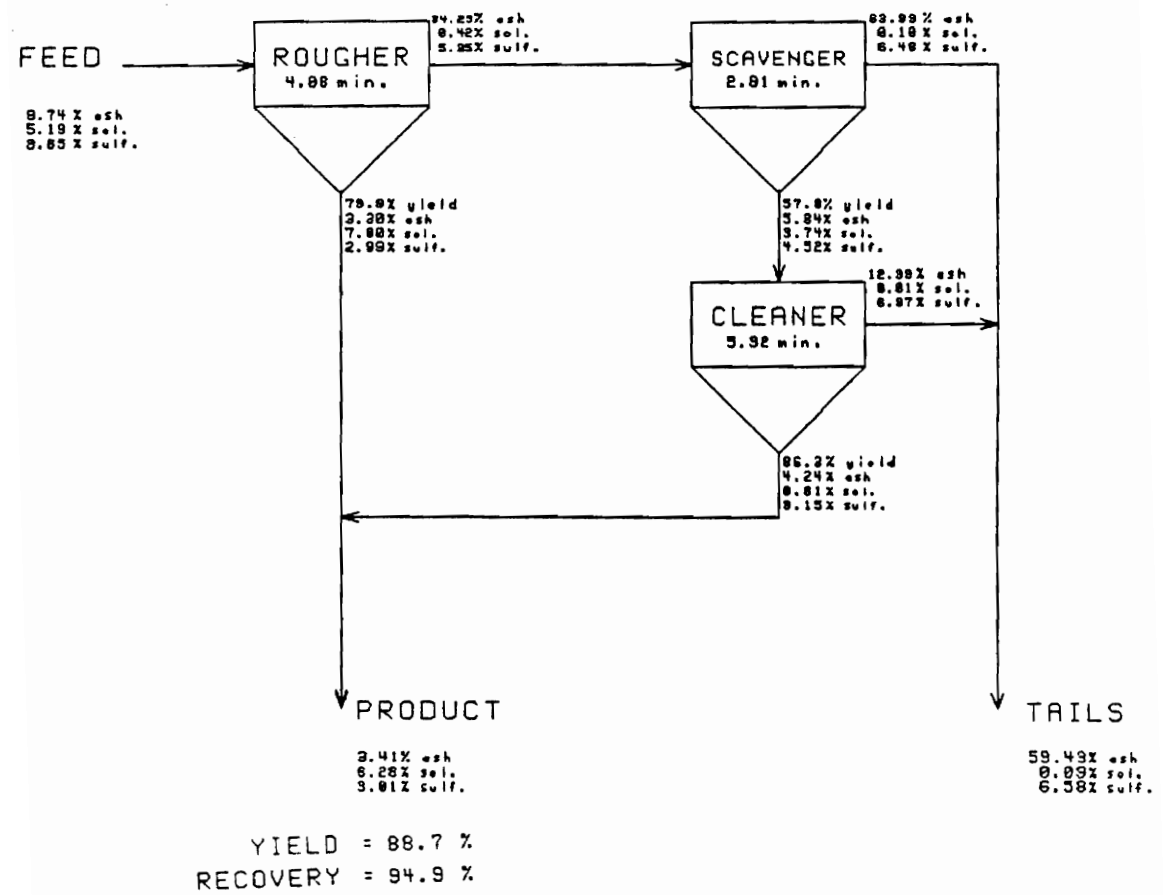


Figure 26. Results obtained for the modified R-S-C circuit at a 200 ml/min feed flowrate.

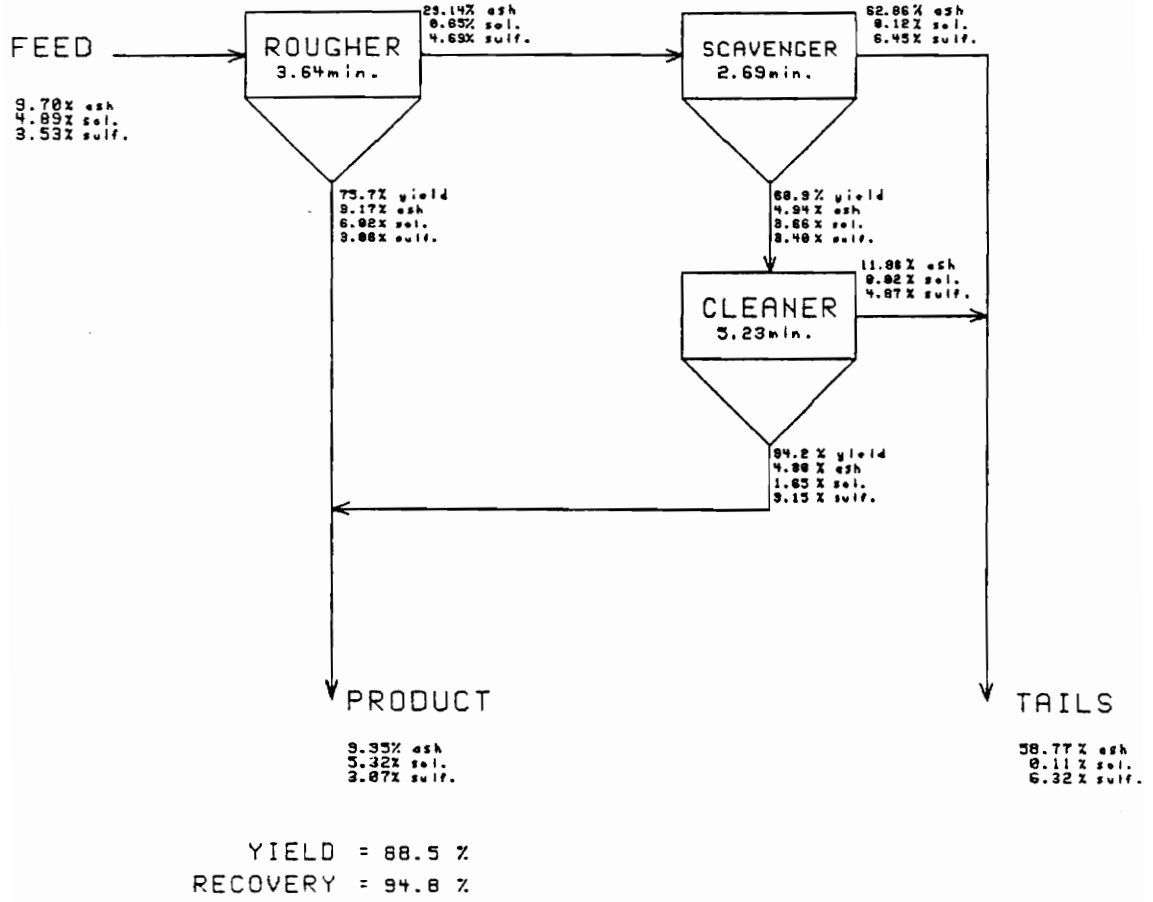


Figure 27. Results obtained for the modified R-S-C circuit at a 300 ml/min feed flowrate.

respectively, which is roughly the same value when considering the sampling and experimental analysis accuracy that was available.

With respect to individual performance, it can be seen that as the circuit feed flow rate was increased from 50 ml/minute to 300 ml/minute, the rougher stage retention time decreased from 5.63 minute to 3.64 minute Yield from this stage decreased during these feed flow rate changes from 84.6% to 75.7%. However, rougher froth product ash and total sulfur remained relatively constant at average values of 3.25% and 3.05%, respectively. Although rougher yield seemed to follow rougher retention time, rougher concentrate quality remained relatively unchanged and traced overall circuit performance very closely.

Again, as had been the case with the R-S-C circuit, scavenger performance was inherently opposite to the rougher stage results. As circuit feed flow rate was increased from 50 ml/minute to 300 ml/minute, scavenger stage yield increased from 36.8% to 60.9%. Scavenger stage concentrate ash trended downward with results of 5.82%, 5.53%, 5.84%, and 4.94%, for the circuit feed flow rates of 50 ml/minute, 100 ml/minute, 200 ml/minute, and 300 ml/minute, respectively.

Similar to the previous work with the R-S-C circuit, it appeared that scavenger stage performance was linked to an increased amount of more floatable, lower ash species in the rougher tailings as the feed flow rate to the circuit was increased. This was also exhibited by noting the increase in solids concentration and decrease in ash content of the rougher tailings from 0.07% to 0.65%, and 41.22% to 29.14%,

respectively, as the circuit feed flow rate was increased. Thus, there was again confirmation that an ever increasing amount of more floatable, lower ash species became available to the scavenger stage as overall feed rate was increased.

The highest yield with the cleaner stage, 96.7%, was obtained at the lowest circuit feed flow rate, 50 ml/minute As circuit feed flow rate was increased from 100 ml/min to 300 ml/minute, the cleaner stage yield varied from 70.4% to 94.2%, while cleaner concentrate ash content increased from 3.21% to 4.30%.

It can be observed that, at the lowest circuit feed flow rate of 50 ml/minute, a very small amount of material was apparently available to be recovered in the cleaner feed. This fact is exhibited by the fact that the scavenger concentrate was 0.45% solids. Also, cleaner stage retention time at this lowest feed rate was 6.88 minute Thus, almost complete recovery of all available floatable material was being achieved. As the circuit feed flow rate was increased, more and more floatable species appeared in the cleaner stage feed, with a somewhat different trend taking place. Initially, at the lowest feed flow rate of 50 ml/minute it appeared that possibly water recovery in addition to solids recovery by flotation may have been the ruling mechanisms, whereas, with increasing the feed flow rate from 100 ml/minute to 300 ml/minute, perhaps the mechanisms of retention time and the inherent carrying capacity problem were the ruling mechanisms at work.

In comparison with the previous R-S-C circuit and the single stage results (see Section 4.1.3), it can be seen that overall yield and froth product ash content are

increased. Unlike in the R-S-C circuit results, the cleaner column of the R-S-C (modified) circuit seemed to be more influenced by the performance of the scavenger stage rather than the rougher stage. Furthermore, although overall circuit concentrate yield was increased by removing the cleaner leg from the rougher side of the circuit, overall circuit quality with respect to ash deteriorated, as it appeared that a cleaning stage is indeed necessary for the rougher concentrate. Overall circuit quality seemed to be influenced primarily by rougher concentrate quality. However, limitations due to time and equipment availability prohibited the use of an additional cleaner stage to make a four-column circuit.

4.3.3 Rougher-Cleaner-Recleaner

The third type of column circuit to be evaluated using Coal B, 325 mesh x 0 particle size, was the rougher-cleaner-recleaner, (R-C-RC), arrangement as described in Section 3.1.2.2 and as shown in Figure 7. The purpose of this series of tests was to determine if this particular coal could be up-graded with respect to quality by succeeding stages of recleaning, or that a liberation limit had been reached. Also, the results obtained from this particular configuration, could be used in the comparative analysis of simulated results for this particular feed coal.

Figures 28 through 31 give the results for circuit feed flow rates of 50 ml/minute, 100 ml/minute, 200 ml/minute, and 300 ml/minute, respectively, in the form of simple flowsheets. Observing Figures 28 through 31, it can be seen that as feed flow rate was

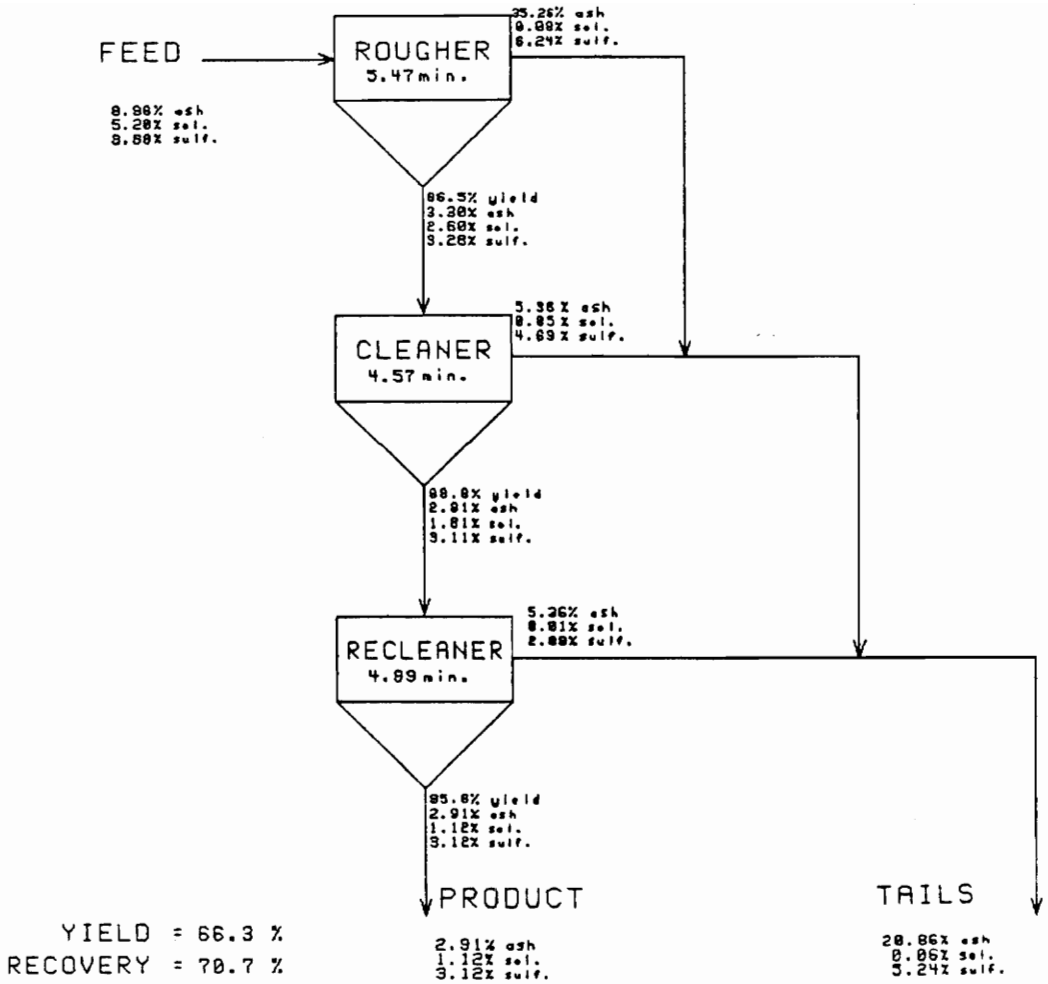


Figure 28. Results obtained for the R-C-RC circuit at a 50 ml/min feed flowrate.

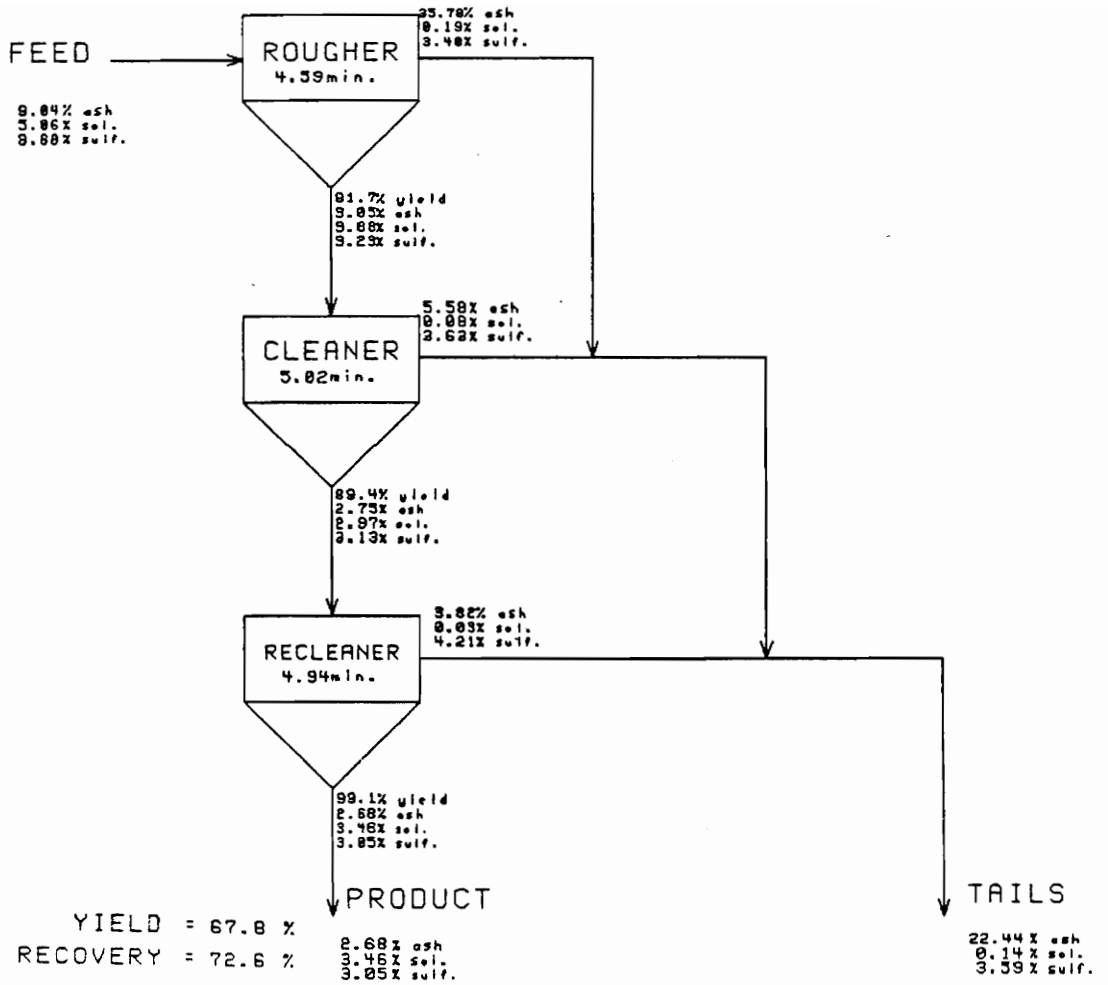


Figure 29. Results obtained for the R-C-RC circuit at a 100 ml/min feed flowrate.

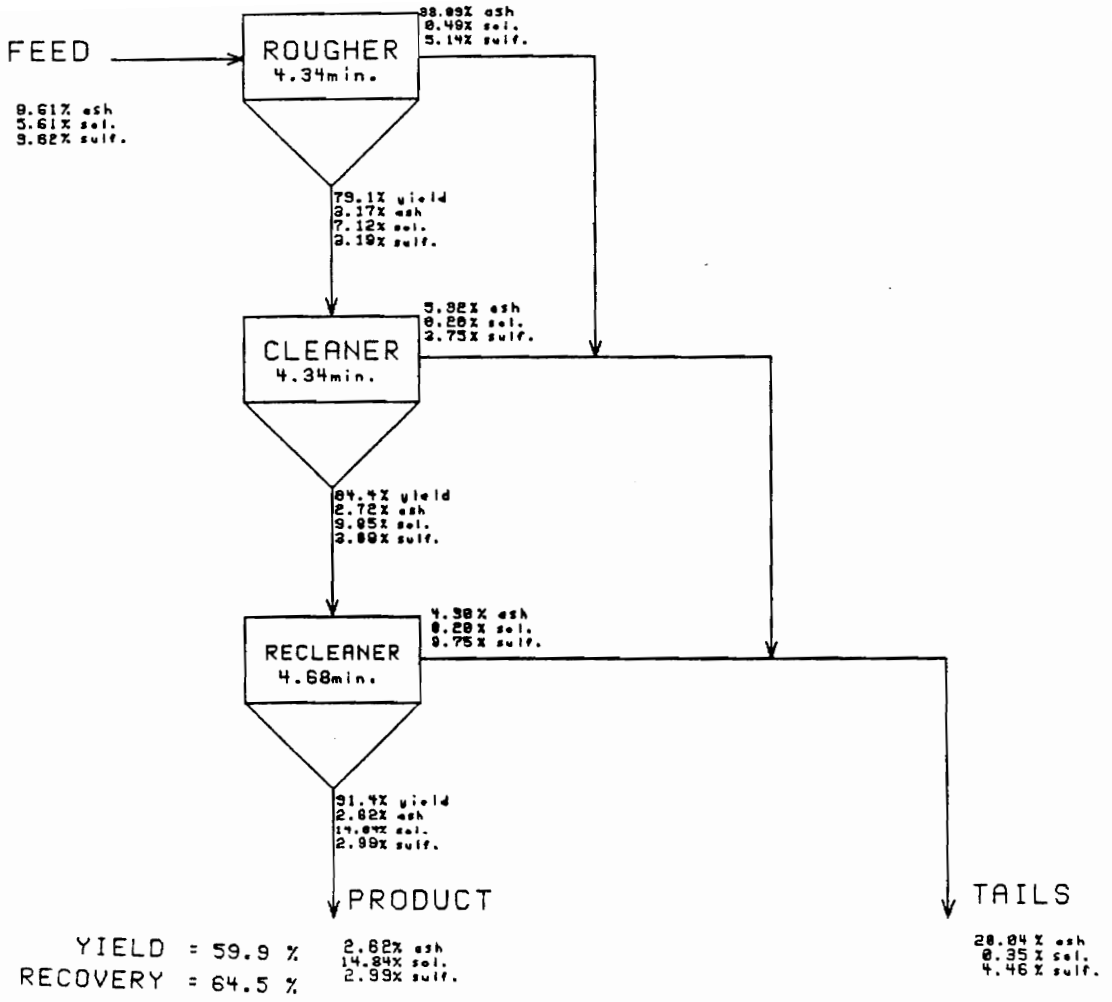


Figure 30. Results obtained for the R-C-RC circuit at a 200 ml/min feed flowrate.

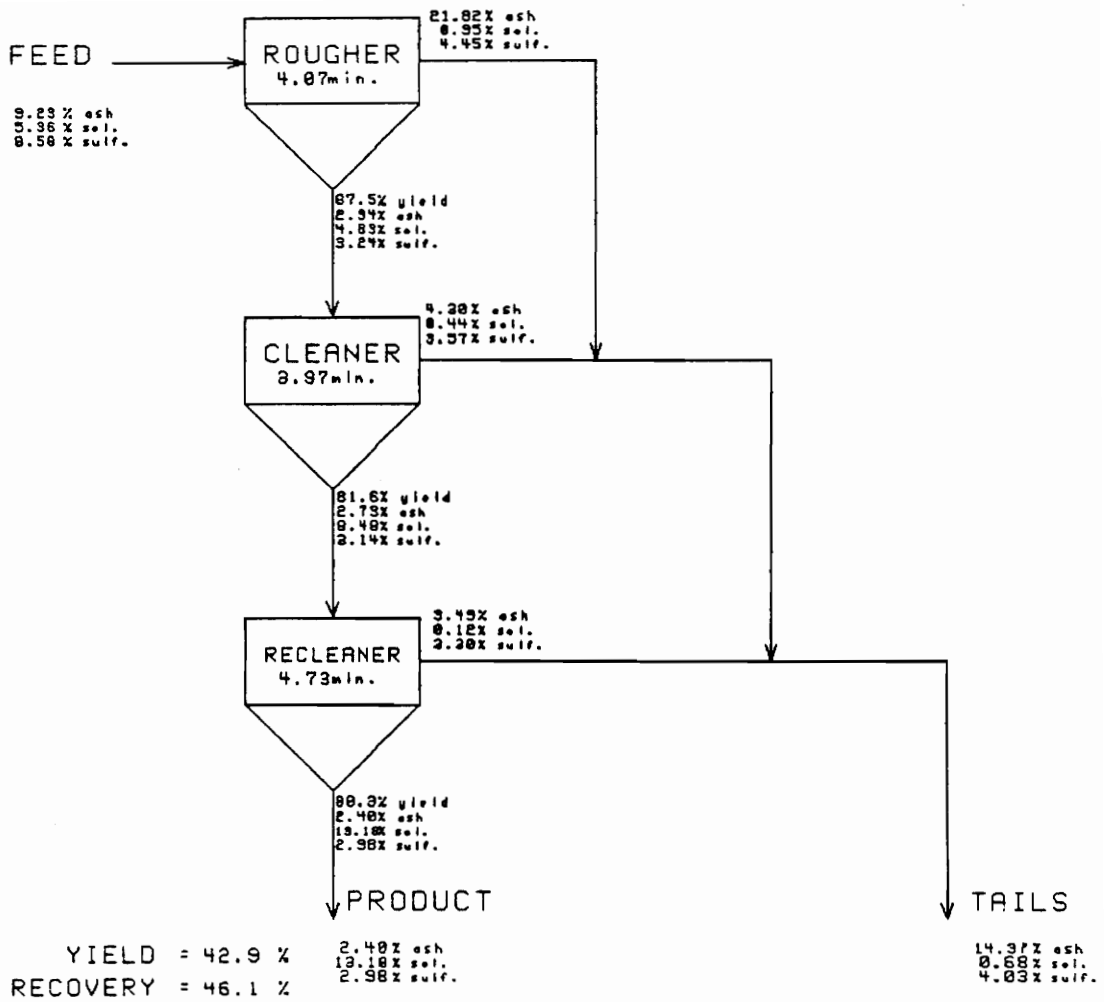


Figure 31. Results obtained for the R-C-RC circuit at a 300 ml/min feed flowrate.

increased from 50 ml/minute to 300 ml/minute, overall circuit yield and recovery decreased from 66.3%, 70.7% to 42.9%, 46.1%, respectively. An exception to this trend occurred at the feed rate of 100 ml/minute, with yield and recovery at 67.8% and 72.6%. Total circuit froth product ash content decreased from 3.30% to 2.40% as the feed flow rate was increased from 50 ml/minute to 300 ml/minute Tailings ash for the overall circuit was 20.86%, 22.44%, 20.04%, and 14.37%, for the circuit feed rates of 50 ml/minute, 100 ml/minute, 200 ml/minute, and 300 ml/minute, respectively.

As the circuit feed flow rate was increased from 50 ml/minute to 300 ml/minute, the retention time in the rougher stage decreased from 5.47 minute to 4.07 minute Correspondingly, the rougher stage yield dropped from 86.5% to 67.5%. During this increase in circuit feed flow rate, the rougher concentrate ash decreased from 3.30% to 2.94%, while concentrate total sulfur content remained relatively unchanged at an average value of 3.23%.

The cleaner stage retention time increased from 4.57 minute to 5.02 minute as circuit feed flow rate was increased from 50 ml/minute to 100 ml/minute, and then decreased from 5.02 minute to 3.97 minute as feed flow rate was increased from 100 ml/minute to 300 ml/minute Concurrently, cleaner product yield increased from 88.8% to 89.4%, and then decreased to 81.6%, for these same feed flow rates. Cleaner stage concentrate ash and total sulfur remained relatively unchanged at average values of 2.75% and 3.12%, respectively, as the feed flow rate to the circuit was increased.

It appeared that perhaps a similar trend was observed in this circuit arrangement

between the rougher and cleaner, as had been seen with the scavenger and cleaner in the R-S-C (modified) circuit results detailed in Section 4.4.2. As the feed flow rate was increased from 50 ml/minute to 100 ml/minute, there was less recovery of floatable material in the rougher, due to decrease in retention time or the presence of a carrying capacity associated factor, which resulted in a succeeding increase in the retention time of the cleaner stage. This fact was further confirmed by the increase in feed solids to the cleaner stage from 2.60% to 9.88% as the feed rate was increased from 50 ml/minute to 100 ml/minute After 100 ml/minute feed flow rate was reached, a further increase in feed flow rate brought about trends in the cleaner stage comparable to the previous performance in the rougher, where retention time, carrying capacity, or froth overloading became the mechanisms controlling flotation recovery.

The recleaner column exhibited a trend similar to rougher stage and overall circuit performance. Froth product yield decreased from 95.6% to 80.3% as the circuit feed flow rate was increased from 50 ml/minute to 300 ml/minute Froth product ash content dropped off correspondingly from 2.91% to 2.40%, while total froth product sulfur remained relatively unchanged at an averaged value of 3.04%. Although cleaner stage yield showed a trend quite similar to the rougher stage yield, there was a 7.5% greater drop in yield for a smaller drop in retention time, 4.99 minute to 4.73 minute, when the feed flow rate was increased in the circuit.

Comparing these results to what had been seen with the two previous circuits, as well as the single-stage flotation results presented in Section 4.1.3 and shown in Figure

15, it can be seen that with a polishing stage of cleaning and recleaning, this particular coal could be significantly upgraded with respect to ash. However, froth product total sulfur content could not be lowered any appreciable amount, indicating a locked condition of pyrite with floatable material, high susceptibility of liberated pyrite to flotation recovery, or a high level of organic sulfur content in the lower ash species of the floatable material.

There was also a significant improvement in froth product quality with respect to ash content. The use of one or more stages of recleaning might not be desirable from the standpoint of recovery due to the frequently appearing problem of froth overloading, carrying capacity, inherent with the columns, especially apparent with the increase in concentration of the more floatable, lower ash components as would be present during repeated stages of cleaning and recleaning.

Overall, considering all three circuit analysis results, it appeared that rougher stage column performance showed a distinct effect upon overall circuit performance. Circuit yield was increased at higher feed rates by the use of a scavenger stage, and overall circuit quality was improved through the use of cleaning and recleaning. However, the results to date indicate that circuit throughput might be limited using cleaner stages on the rougher side of the circuit due to an inherent problem identified with the recovery mechanism(s) present in the column cells.

4.4 Circuit Simulation

In order to determine the validity and capability of the developed simulator, a series of circuit simulations were performed. These theoretical tests incorporated the three circuit types outlined in Section 3.1.2.2, R-S-C, R-S-C (modified), and R-C-RC, used in the actual testing of Coal B, 325 mesh x 0 particle size. A deliberate effort was made to insure that identical conditions were incorporated into the simulation of each circuit, at each slurry feed flow rate that was actually tested.

Feed coal characteristic input data for the simulations was obtained by taking the feed component data derived from the release analysis testing of Coal B described in Section 4.3.2. Flotation rate constant values were arrived at by taking these results and fitting to the single-stage column flotation results presented in Section 4.1.3 and shown in Figure 16. Using this approach, it was found that three components of the feed coal could be classified. These components were "coal," percent by weight, percent ash, "middlings," percent by weight, percent ash, and "reject," percent by weight, percent ash, with respective rate constants of 4.50, 0.75, and 0.07 $minute^{-1}$.

Upon further examination of these feed component characteristics, it was decided that the component percentages derived directly from release analysis did not fit as well for the Coal B sample as had been the case for Coal A in earlier test work (see Section 4.3.1). Thus, some further adjustment was necessary on the Coal B components to better duplicate single-stage column flotation data. This was effected by the addition of an fourth component, "non-floatable," 6.0% by weight, 100% ash with flotation rate

constant of 0 minute^{-1}. As can be seen from Figure 15, this gave single-stage simulated results much closer to the actual test work at 5.0% solids. The derived feed component breakdown and values are as shown in Table VIII, these being used exclusively throughout the simulations.

When the circuit simulations were performed, it was found that there was some difficulty in arriving at retention times in the program that would correspond identically to the actual circuit test work using the same values of feed rate and air hold-up. This apparent problem probably stemmed from the fact that a suitable and comprehensive froth recovery model had not been developed, to date, and incorporated into the simulator. However, it was found that adjustment of the counter-current wash water flow rate during the simulations would give retention times that would be the same as actual values. Although the level of counter-current wash water would have no significant effect upon simulated results, with the program as was developed, it should be remembered that there would be substantial contribution in actual tests, especially with respect to product grade. Thus, an area of the simulator that appeared to be lacking was in the need for development of a substantial model incorporating the counter-current wash water effect.

4.4.1 Rougher-Scavenger-Cleaner

Figures 32 through 35 give the results for the rougher-scavenger-cleaner, (R-S-C), simulations using the selected input feed components of Coal B, 325 mesh x 0 particle

Table VIII. Adjusted feed component characteristics for Coal B at a 325 mesh topsize.

Component	Weight (%)	Ash (%)	k (min^{-1})
Coal	70.0	2.50	4.50
Middlings	10.0	7.50	0.75
Reject	14.0	10.0	0.07
Non-Float	6.0	100.0	0.00

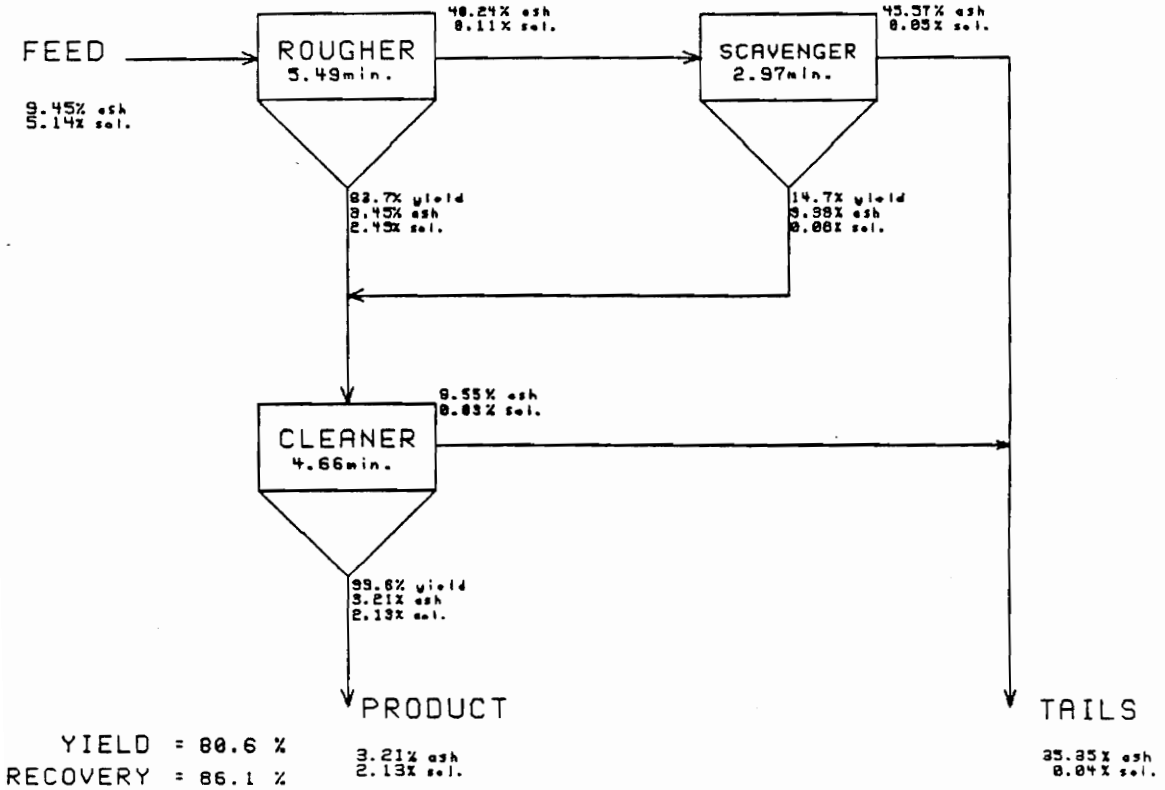


Figure 32. Simulated results for the R-S-C circuit at a 50 ml/min feed flowrate.

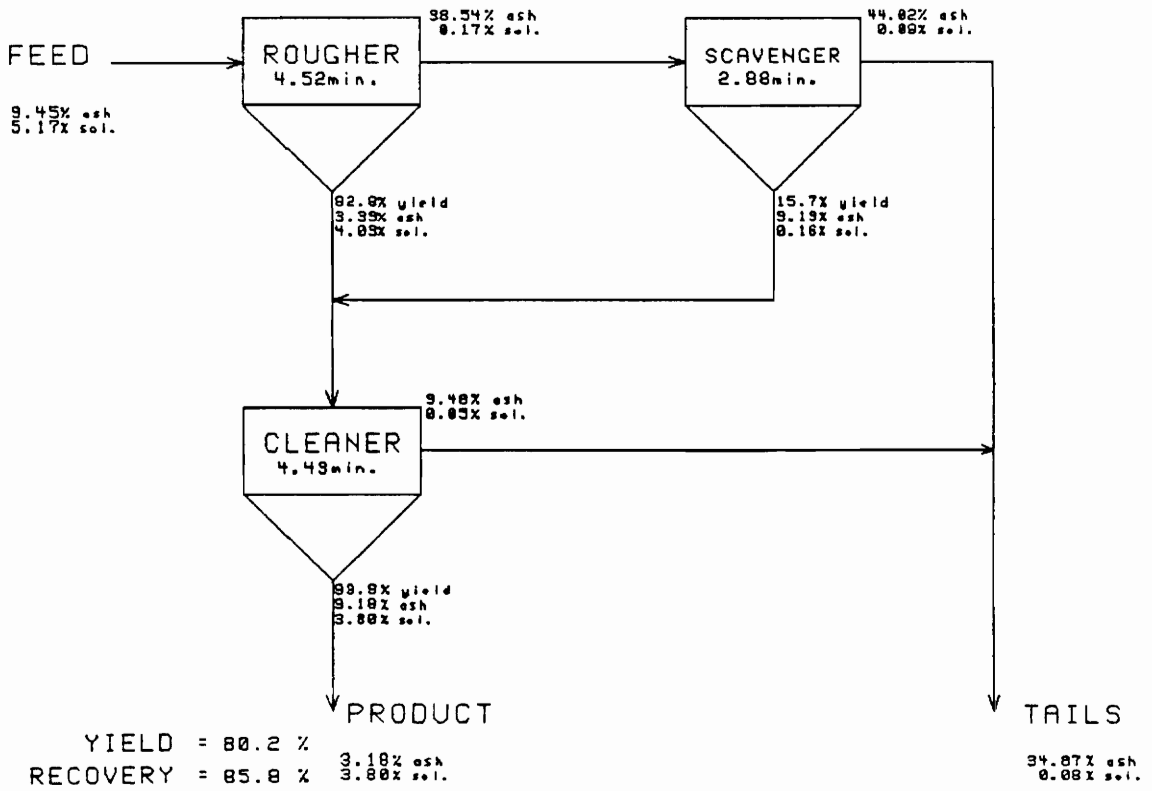


Figure 33. Simulated results for the R-S-C circuit at a 100 ml/min feed flowrate.

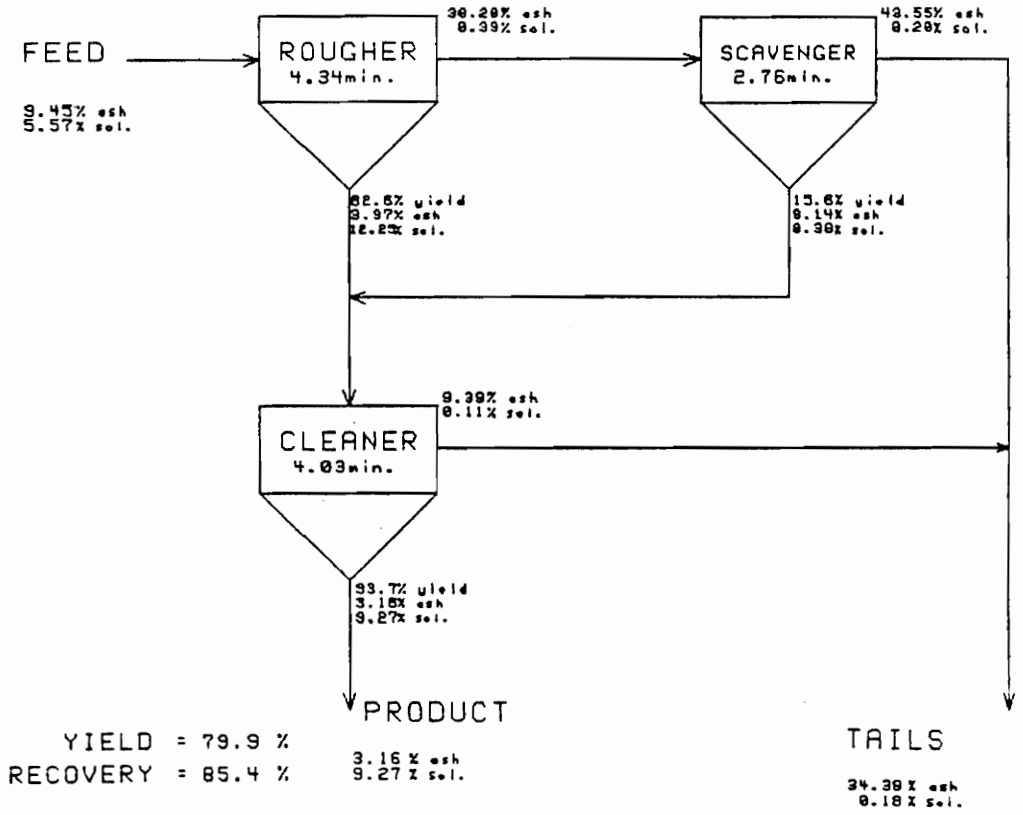


Figure 34. Simulated results for the R-S-C circuit at a 200 ml/min feed flowrate.

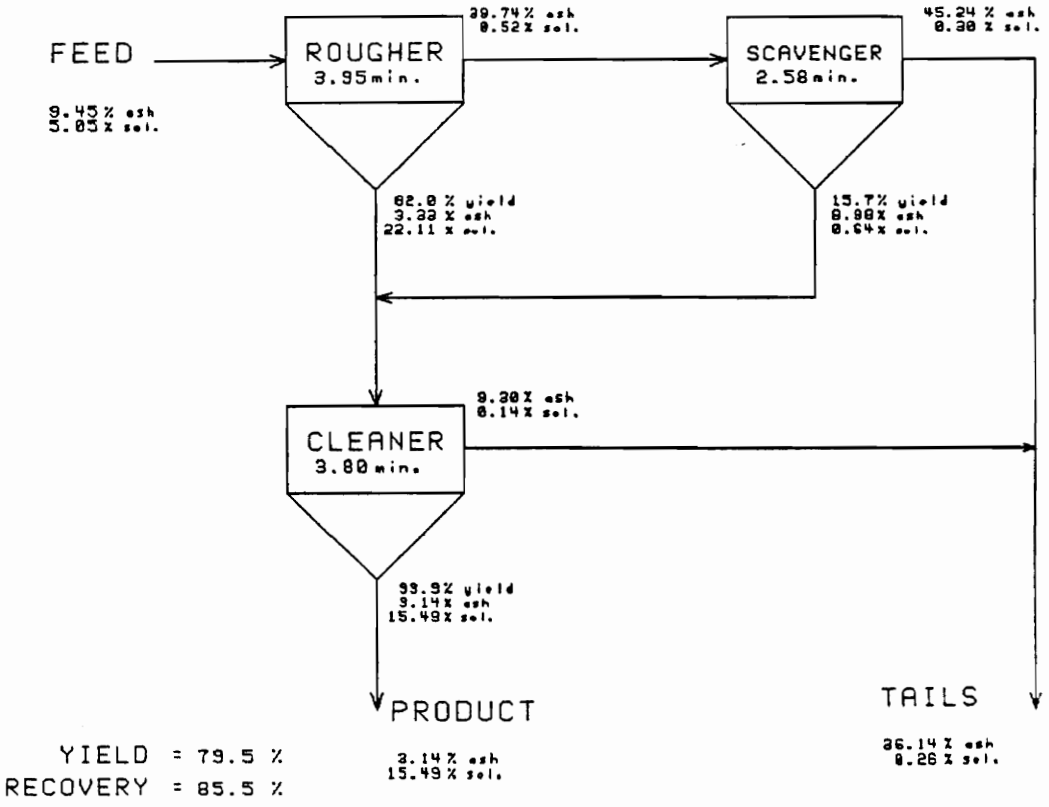


Figure 35. Simulated results for the R-S-C circuit at a 300 ml/min feed flowrate.

size, reflecting the circuit feed slurry flow rates of 50 ml/minute, 100 ml/minute, 200 ml/minute, and 300 ml/minute As can be seen, the highest overall circuit yield and recovery occurred at the lowest feed flow rate of 50 ml/minute, 80.6% and 86.1%, respectively. This result goes along with the fact that the greatest amount of available circuit retention time would be available at the lowest feed flow rate. Likewise, the lowest overall circuit yield and recovery were at the highest feed flow rate of 300 ml/minute, 79.5% and 85.5%, respectively.

As circuit feed flow rate was increased from 50 ml/minute to 300 ml/minute, total circuit concentrate ash decreased from 3.21% to 3.14%. Circuit froth product solids concentration increased from 2.13% to 15.49% for this same increase in feed rate. The trend for froth product ash content falls along the lines of what would be expected due to a consistent lowering of available retention time for the less floatable, higher ash components. However, the increase in circuit concentrate solids percentage followed the simple solids recovery model previously incorporated into the simulator (see Section 3.1.5). The total circuit tailings ash content increased from 35.35% to 36.14% as the feed flow rate was varied from 50 ml/minute to 300 ml/minute, which would not be expected from the theoretical developments surrounding retention time and recovery inherent in the program as designed.

Similarly, rougher stage performance followed overall circuit performance closely, as feed flow rate was increased from 50 ml/minute to 300 ml/minute, froth product yield decreased from 83.7% to 82.0%. Likewise, froth product ash content

decreased from 3.45% to 3.33% for the same increase in feed flow rate. These results can be easily explained, in that, a lower amount of retention time was available at the higher rates, and a greater percentage of the more floatable, lower ash species of the feed coal showed up as froth product.

However, observing Figures 32 through 35, it can be seen that a somewhat unexpected set of results were obtained from the scavenger column. As the feed rate was increased from 50 ml/minute to 100 ml/minute, froth product yield increased from 14.7% to 15.7%. As feed flow rate was increased further from 100 ml/minute to 300 ml/minute, the yield remained relatively unchanged. Likewise, upon increasing the feed flow rate from 50 ml/minute to 100 ml/minute, froth product ash decreased from 9.39% to 9.19%, and then remained fairly constant with further increase in feed flow rate from 100 ml/minute to 300 ml/minute Apparently, this initial decreasing trend followed by level performance probably resulted due to the presence of an inversion point on the recovery curve for one of the feed component species that was being input to the simulation, although, overall scavenger performance changed very little during the increase in feed flow rate to the circuit.

Also, observing froth product yield and ash values for the cleaner stage, it can be seen that only froth product ash content dropped slightly from 3.21% to 3.14%, while froth product yield remained relatively constant, as feed flow rate was increased from 50 ml/minute to 100 ml/minute. Possibly, the previously mentioned effects of different component recovery rates and the balancing effect of the rougher and scavenger stages

upon cleaner feed component composition negated any apparent trends. Overall, cleaner performance changed very little upon the increase in feed flow rate to the circuit.

As was the case with the actual R-S-C circuit test results outlined in Section 4.4.1 and shown in Figures 20 through 23, it seemed that the overall controlling column during these simulations was the rougher stage. Comparing simulated results further with actual test work, it can be seen that at the lower feed rates of 50 ml/minute and 100 ml/minute, circuit concentrate yield and ash contents were very close. However, as feed flow rate is increased to 300 ml/minute, although a similar trend was followed with respect to product yield, the spread in yield values was greater with the actual circuit performance. Additionally, overall concentrate ash was somewhat lower in the actual circuit operation.

The simulated performance of the rougher column stage showed a similar comparative trend with actual test work. Again, at the lower feed rates, froth product yield and ash were nearly the same, showing the same trend with an increase in feed rate, however, the trend in actual yield values showed a wider spread, with the ash contents consistently lower than the simulated values.

Upon comparing the simulated scavenger stage results with actual values presented in Section 4.4.1, it can be seen that drastically different results were obtained during actual circuit operation, where scavenger performance was consistently better. With this difference in mind, along with note taken of the apparent problems associated with recovery capacity inherent in the actual operation of columns, a visualization of what may have occurred can be formed.

During the simulations, no allowance was made for this inherent problem, however, in actual operation, the more floatable, lower ash fractions of the feed reported to the scavenger in the rougher stage tailings, which was also reflected in the lower concentrate ash content of the actual rougher stage as the feed flow rate was increased.

Furthermore, actual operation of the cleaner stage in the R-S-C arrangement produced a much lower performance with respect to recovery than was exhibited during the simulations. The effect of the inherent froth recovery or bubble carrying capacity problem seems to have been magnified in the cleaner stage during the actual operation due to the increased percentage of more floatable, lower ash components that were amenable to recovery to the concentrate launder. Theoretically, the cleaner stage of the circuit should show the best overall performance as was the case during the simulations.

4.4.2 Rougher-Scavenger-Cleaner (Modified)

Figures 36 through 39 give the simulated results for the modified rougher-scavenger-cleaner (modified R-S-C) circuit described in Sections 3.1.2.2 and 4.4.2, in the format of simple flowsheets. Again, as had been practiced with the previous R-S-C circuit, the input feed characteristics and operating conditions were entered and set so as to duplicate actual circuit test work during this simulation as near to identical as was possible. Feed components and associated characteristics were those as derived for Coal B, 325 mesh x 0 particle size, as was presented in Table VIII.

Observing Figures 36 through 39, it can be seen that as the circuit feed flow rate

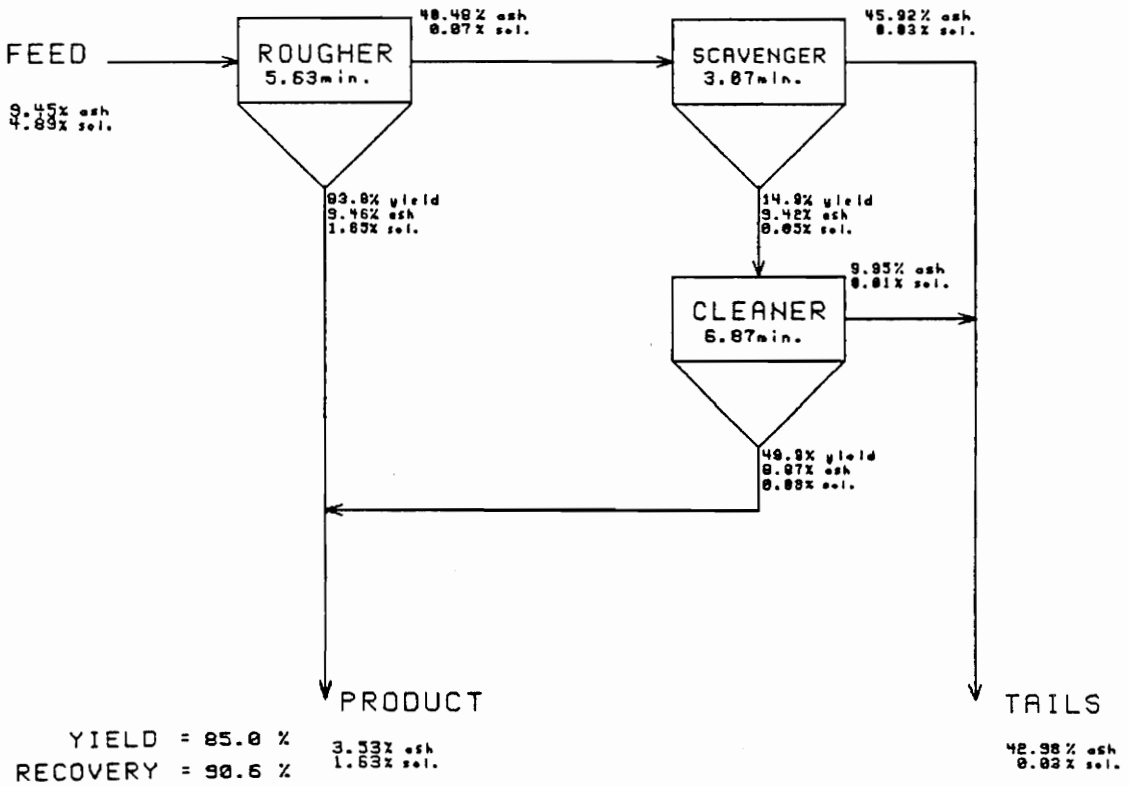


Figure 36. Simulated results for the modified R-S-C circuit at a 50 ml/min feed flowrate.

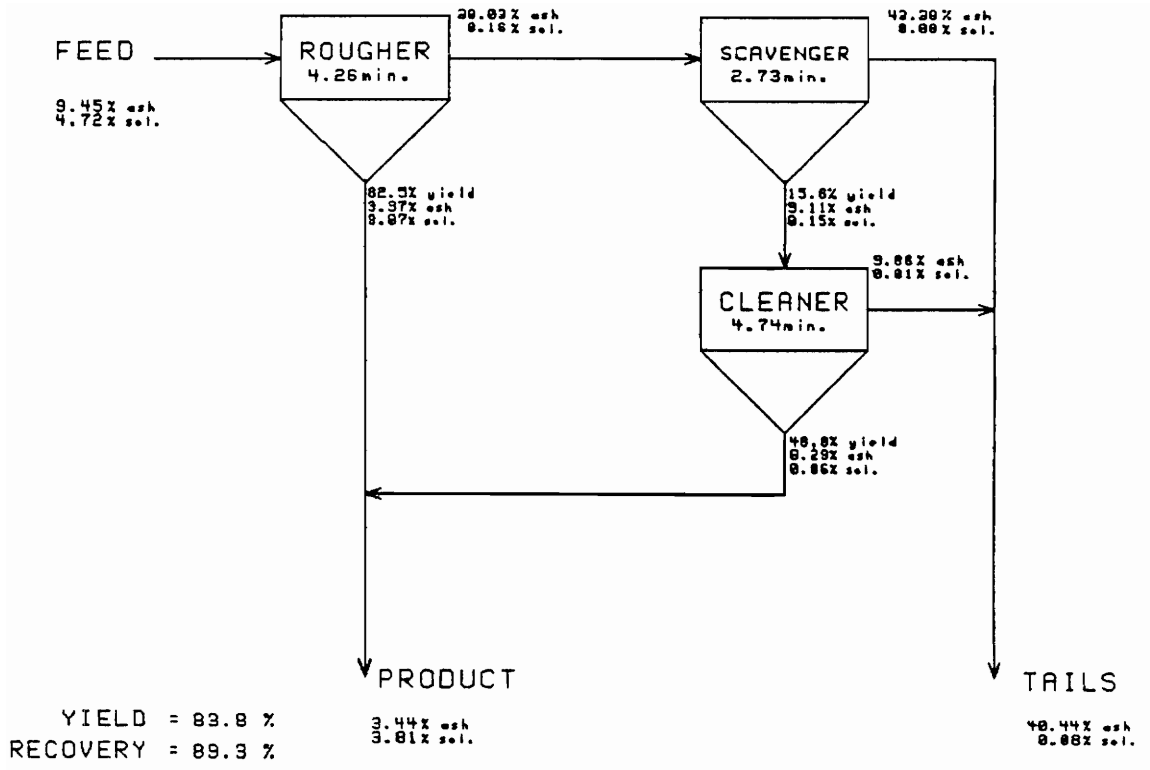


Figure 37. Simulated results for the modified R-S-C circuit at a 100 ml/min feed flowrate.

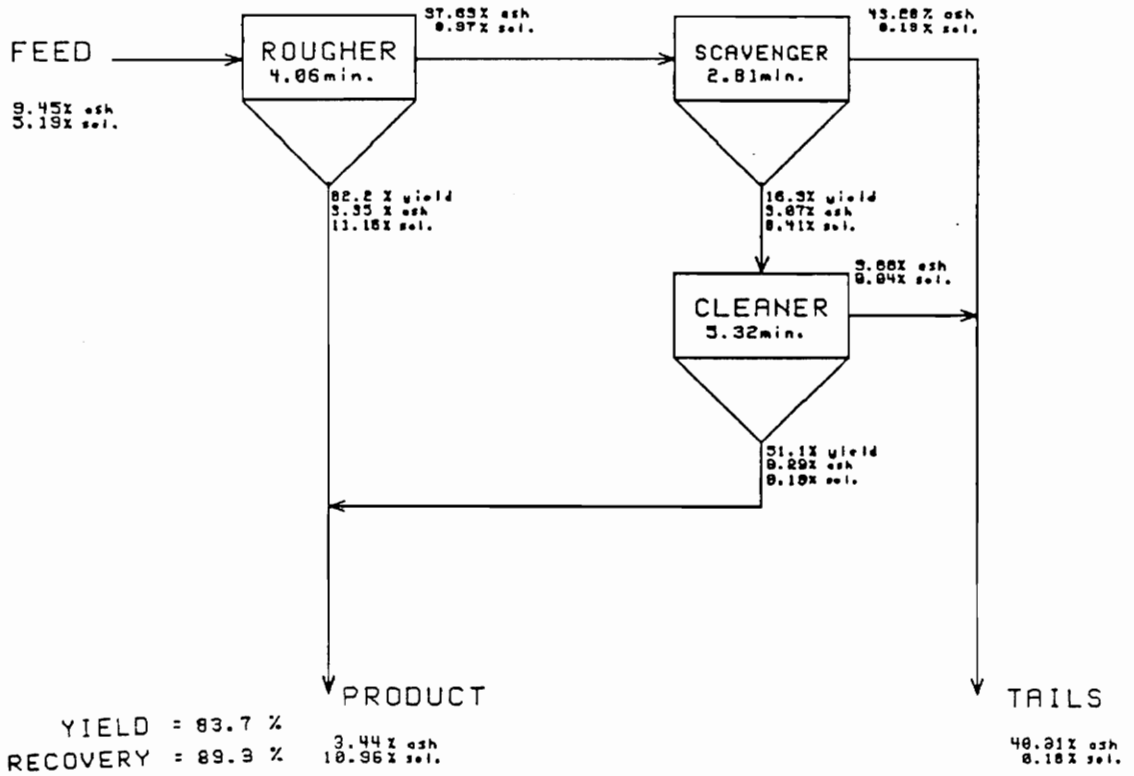


Figure 38. Simulated results for the modified R-S-C circuit at a 200 ml/min feed flowrate.

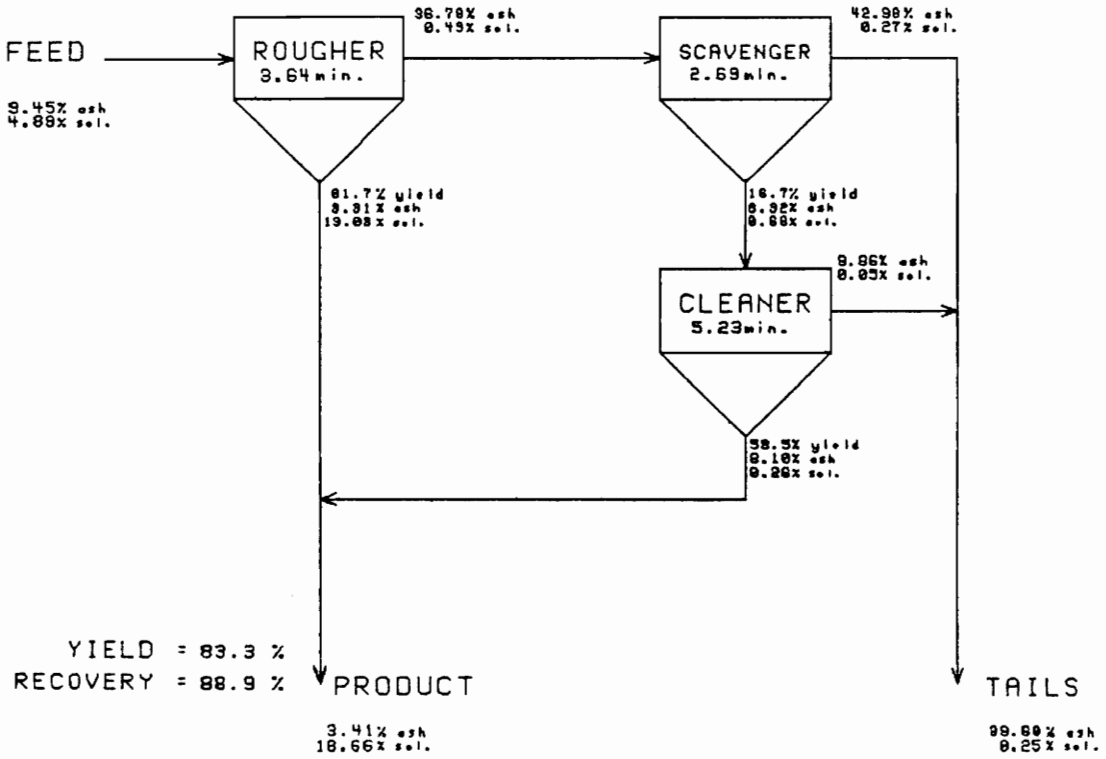


Figure 39. Simulated results for the modified R-S-C circuit at a 300 ml/min feed flowrate.

was increased from 50 ml/minute to 300 ml/minute, overall circuit yield and recovery decreased from 85.0%, 90.6% to 83.3%, 88.9%, respectively. Overall circuit concentrate ash decreased from 3.53% to 3.44% as feed flow rate was increased from 50 ml/minute, and then remained relatively the same as feed flow rate was subsequently increased. Overall circuit tailings ash decreased from 42.98% to 39.60% as the feed flow rate was varied from 50 ml/minute to 300 ml/minute.

The observed drop in circuit recovery can be explained by the fact that the overall retention time of the circuit was reduced as the feed flow rate was increased. This observation agrees with conventional theory and the expectation of the simulator as developed. Also, the decrease in overall tailings ash was expected due to the fact that, as the feed rate was increased, there was less and less retention time available, thus, steadily allowing an increasing percentage of the more floatable, lower ash species present in the feed to report to the tailings.

The initial drop, followed by the relatively flat behavior of the circuit concentrate ash was apparently the result of action and interaction of the individual component recovery rates and the balancing effect of all the column stages as laid out in this circuit arrangement. However, on an individual basis, rougher froth product yield followed the same trend as was seen for the overall circuit. As circuit feed flow rate was increased from 50 ml/minute to 300 ml/minute, rougher stage concentrate yield decreased from 83.8% to 81.7%. Unlike overall circuit froth product ash, the rougher concentrate ash dropped from 3.46% to 3.31% as the feed rate was varied. These results seem to follow

directly from the theoretical concepts of recovery and retention time.

As circuit feed flow rate was increased from 50 ml/minute to 300 ml/minute, the scavenger stage yield increased slightly from 14.9% to 16.7%. During this same feed rate change, the froth product ash decreased from 9.42% to 8.92%. Apparently, as less and less retention time was available in the rougher stage, the more floatable, lower ash components became available in a greater percentage in the scavenger feed, thus, improving performance as the feed flow rate was increased. This explanation is also justified by considering the rougher concentrate ash drop observed during the feed flow rate increase. However, considering the range of improvement in froth product yield and ash, overall performance was virtually constant for the scavenger column.

The range of performance variation for the cleaner stage was somewhat wider than what was seen for the scavenger. Observing Figures 36 through 39, it can be seen that as the feed flow rate was increased from 50 ml/minute to 300 ml/minute, concentrate yield increased from 49.3% to 53.5%, except for the 100 ml/minute feed flow rate, where froth product yield was 48.0%. As can be seen from Figures 36 through 39, this deviation from the expected trend can be explained by the fact that retention time in the cleaner cell was lower at 100 ml/minute circuit feed flow rate than at any of the other feed rates examined. The reason for this lowering of cleaner retention time was to make simulation identical with actual circuit operation. Likewise, froth product ash decreased from 8.87% to 8.10% as feed flow rate was increased except for the feed rate of 100 ml/minute, which had the lowest retention time. Overall, cleaner performance was as

would be expected considering flotation recovery rates and retention time.

In comparison with the previous simulations of the R-S-C circuit described in Section 4.4.1, it can be seen that overall circuit performance was improved with respect to recovery, and deteriorated with respect to overall froth product ash, when using this modified configuration. Again, as was the case before, rougher stage performance seemed to have a priority in controlling total circuit results. In that there was no cleaner stage implemented in polishing the rougher concentrate, overall quality with respect to final concentrate was sacrificed. Considering only the difference in retention times between the simulations, rougher column results were nearly identical using this R-S-C (modified) circuit as the results obtained using the R-S-C circuit previously examined.

Although there was slightly more variation in scavenger stage performance when using the R-S-C (modified), the overall range was negligible and approximately the same results were obtained as with the R-S-C configuration. As was discussed in Section 4.4.1, the primary factors controlling scavenger performance during these simulations appeared to be related to the individual and cumulative recovery rates, the interaction of recovery rates, and feed composition that appeared in the rougher stage tailings.

The variation and trends appeared to be similar for cleaner stage performance in both the R-S-C and R-S-C (modified) simulations, however, both recovery and froth product ash were much higher and lower, respectively, for the previous circuit (see Figures 32 through 35). Thus it seemed that cleaner performance was strongly controlled by the nature of the preceding stage feeding it. In that the purpose of a cleaner stage,

in any leg of a circuit, would be the polishing of a previous concentrate, it appeared that optimum operation would be where recovery would be as high as possible, as was the case with the previous R-S-C configuration.

In comparison with the actual R-S-C (modified) circuit results discussed in Section 4.4.2 and presented in Figures 24 through 27, it can be seen that overall yield and quality were better for the actual operation. The higher recoveries of froth product during actual test work as compared to the simulations may be explained by the fact that actual scavenger stage performance was drastically different than what resulted during the simulations owing to the fact that a substantially greater percentage of the more floatable, lower ash components were available in the scavenger leg of the actual circuit, effectively allowing this leg to operate more efficiently. This was further confirmed by the fact that rougher yield was somewhat lower in the actual case, especially for the higher feed flow rates of 200 ml/minute and 300 ml/minute, with substantially lower concentrate ash content. In other words, the scavenger leg of the circuit was provided higher quality material with respect to ash and floatability in the actual tests. In addition to better operating conditions for the scavenger in the actual circuit run, the polishing cleaner stage was provided also with a much more desirable feed with respect to ash and floatability. Even though rougher stage performance dipped below the simulated values, the overall better actual performance for the scavenging side of the circuit counter-balanced rougher output allowing for better overall circuit performance for the feed flow rates examined.

Similar to what was experienced in the simulation of the R-S-C circuit, there was substantially more spread in froth product yield values, especially for the scavenger stage during actual testing. This fact could also confirm that there is indeed other mechanisms at work during coal recovery in a column besides the singular effect of flotation recovery rates and retention times. Again, as previously noted, it appeared from the simulations performed with this circuit, that best performance could be obtained on Coal B, 325 mesh x 0 particle size, by separating the roughing and scavenging legs of the circuit, allowing for an additional stage of cleaning and possibly, recleaning for each separate leg, and insuring that the cleaner(s) stage(s) are receiving only the lowest ash material.

4.4.3 Rougher-Cleaner-Recleaner

Figures 40 through 43 give the simulated results for the rougher-cleaner-recleaner, (R-C-RC), circuit that was described in Section 3.1.2.2, using the input feed characteristics of Coal B, 325 mesh x 0, that was derived from the release analysis (see Section 4.3.2) and as presented in Table VIII. Again, as in the previous two simulations, operating conditions were set up in these simulations to reflect identical parameters as to what occurred during the actual tests.

Observing Figures 40 through 43, it can be seen that as the circuit feed flow rate was increased from 50 ml/minute to 100 ml/minute, overall circuit yield and recovery remained relatively unchanged at average values of 77.6% and 83.1%, respectively. During this feed rate increase, overall concentrate ash content from the circuit dropped

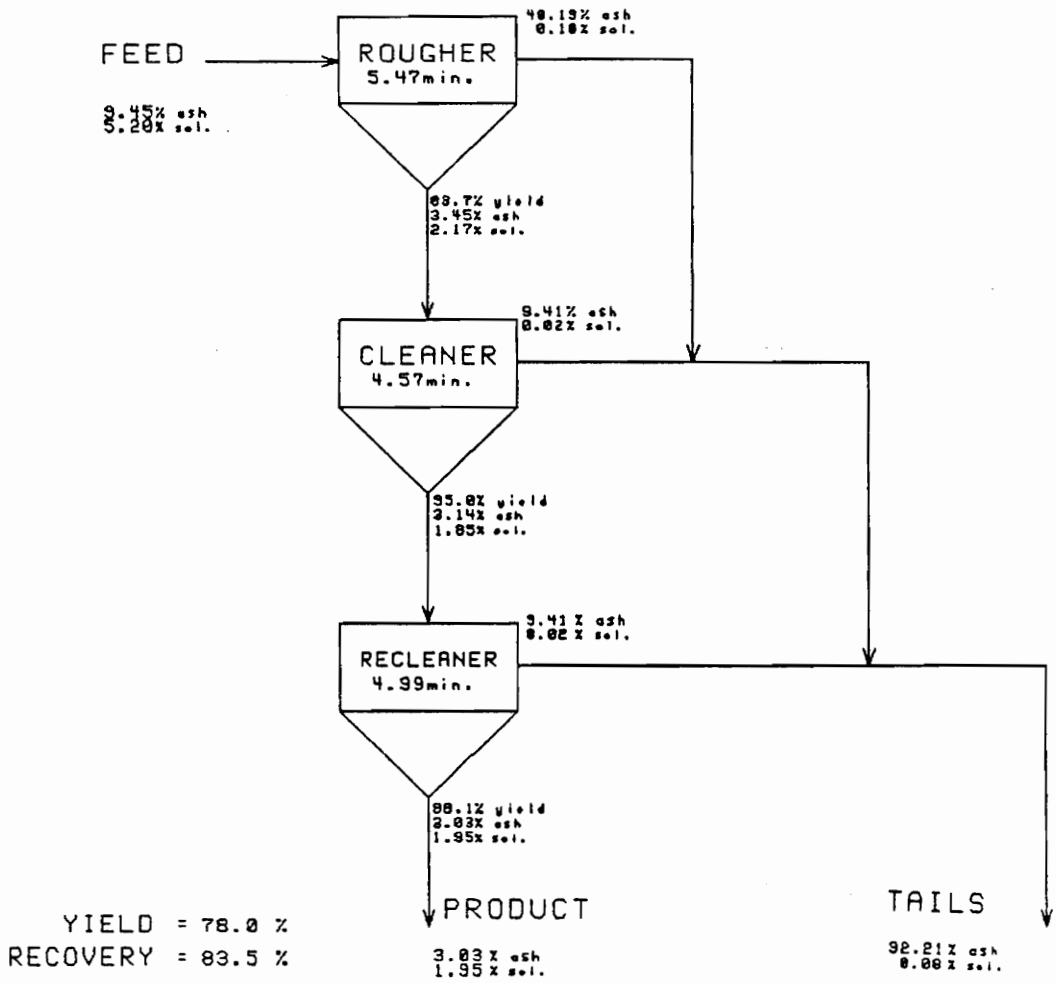


Figure 40. Simulated results for the R-C-RC circuit at a 50 ml/min feed flowrate.

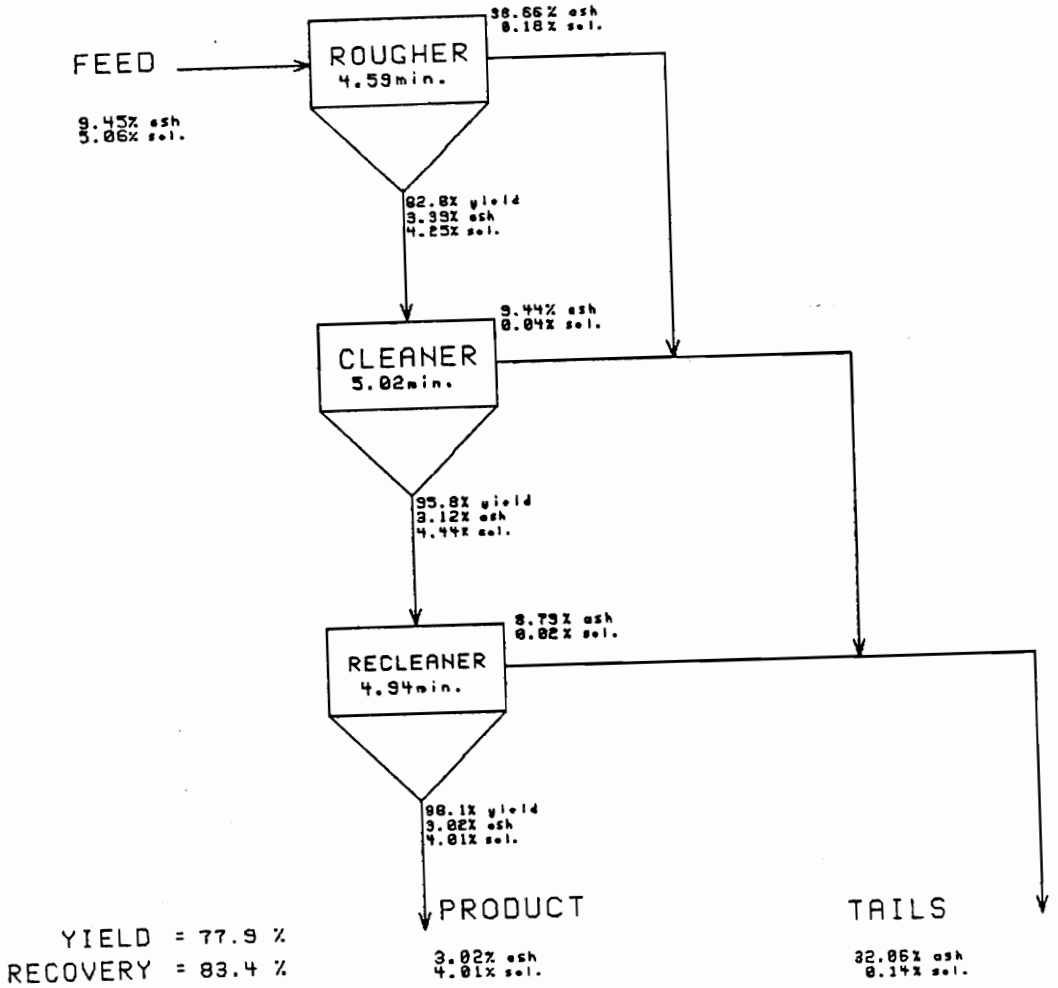


Figure 41. Simulated results for the R-C-RC circuit at a 100 ml/min feed flowrate.

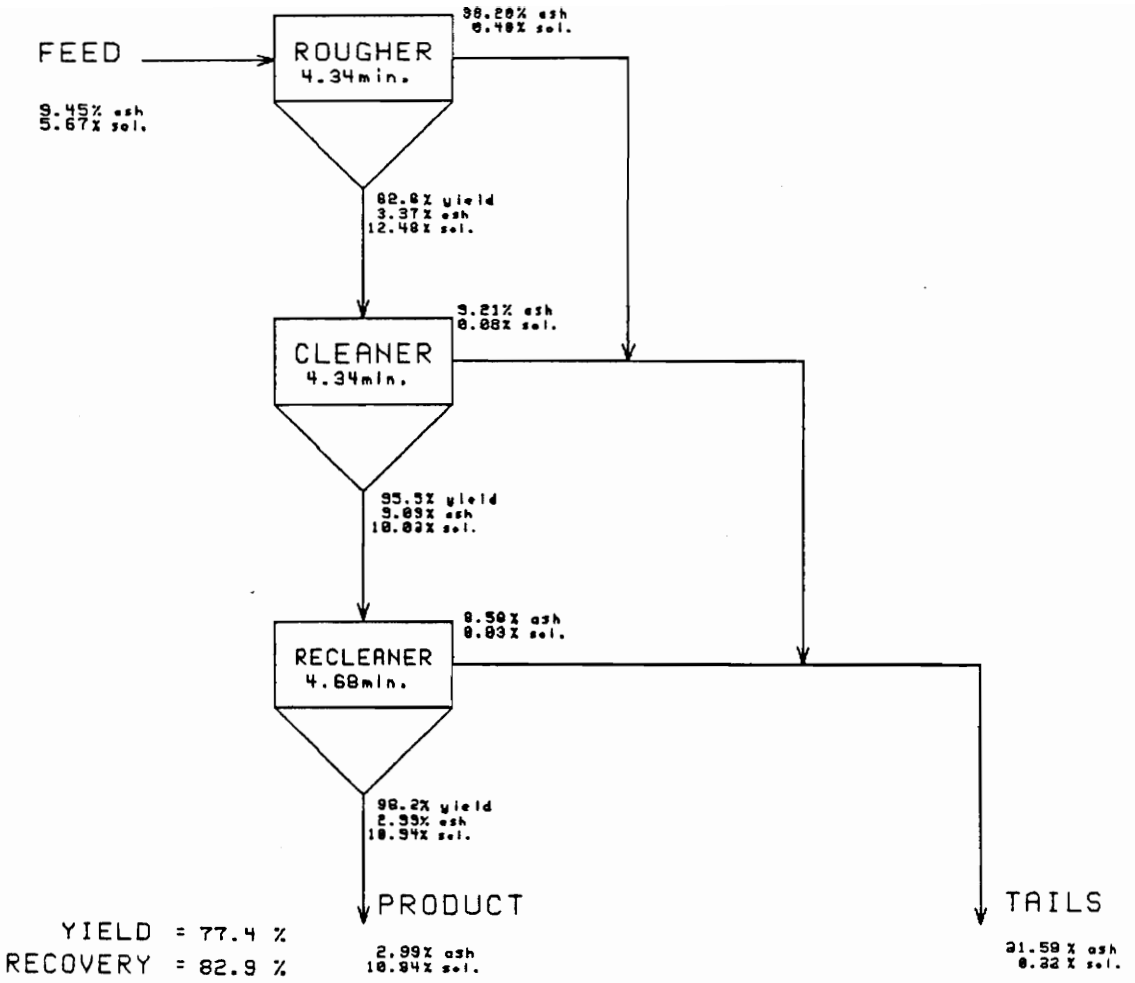


Figure 42. Simulated results for the R-C-RC circuit at a 200 ml/min feed flowrate.

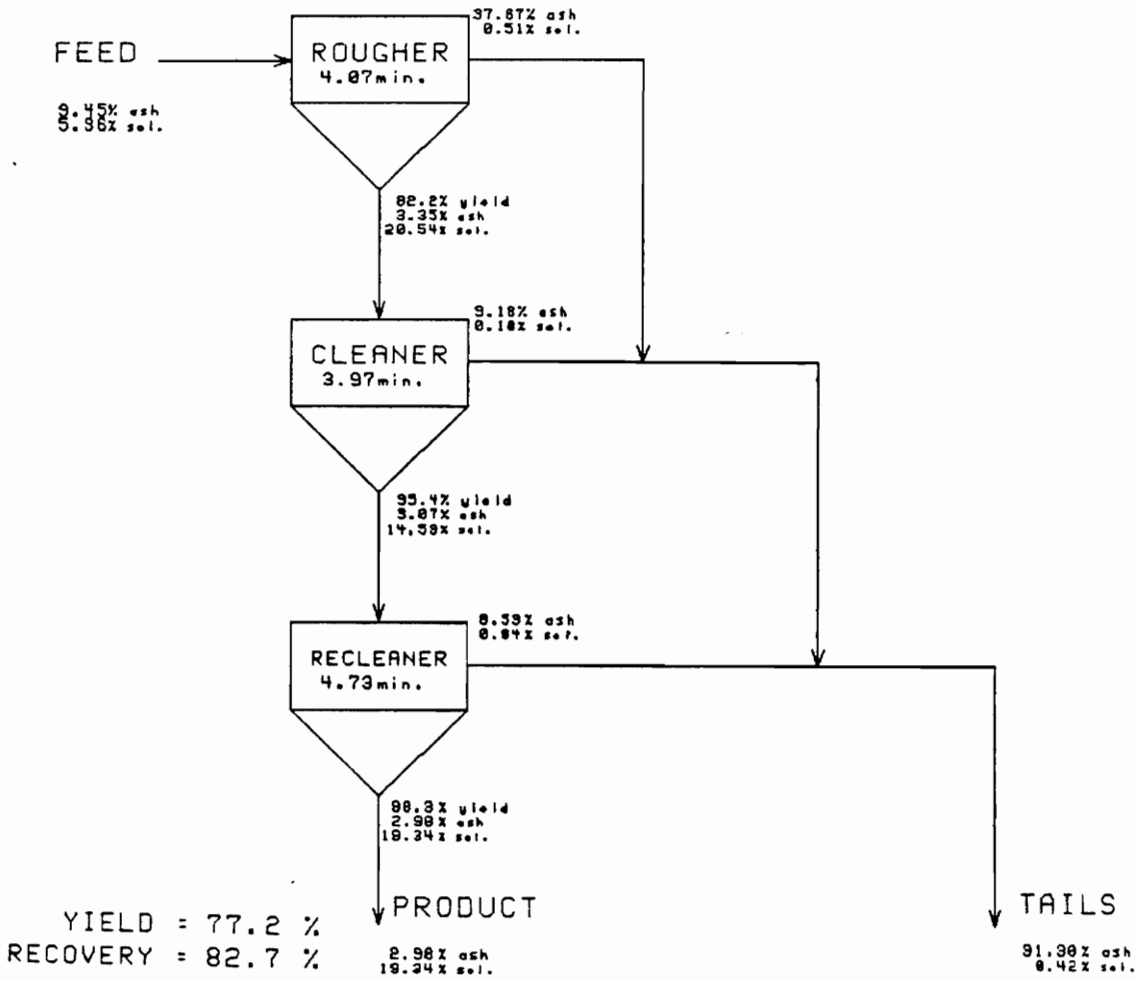


Figure 43. Simulated results for the R-C-RC circuit at a 300 ml/min feed flowrate.

slightly from 3.03% to 2.98%. Likewise, overall circuit tailings ash decreased from 32.21% to 31.30%, almost an insignificant drop. On an individual basis, it can be seen that rougher stage performance exhibited a slightly more noticeable trend, however, overall circuit performance was seemingly related more closely to rougher results than to the other two stages of the circuit. As flow rate was increased from 50 ml/minute to 300 ml/minute, rougher froth product yield decreased from 83.7% to 82.2%, while froth product ash content decreased from 3.45% to 3.35%. These results were expected in light of the traditional theories of recovery rates of various components and retention times.

Further examination of the results of Figures 40 through 43 shows that an increase in feed flow rate from 50 ml/minute to 300 ml/minute slightly increased the cleaner concentrate yield from 95.0% to 95.4%. At the same this, this action decreased the froth product ash content from 3.14% to 3.07%. This very modest increase in yield, with corresponding drop in froth product ash appeared to be the result of the fact that as the feed rate was increased, the yield of rougher product dropped, with the concurrent appearance of a larger percentage of more floatable, lower ash material in the feed to the cleaner.

Likewise, with the increase in feed flow rate from 50 ml/minute to 300 ml/minute, the yield of the recleaner increased only slightly from 98.1% to 98.3%. The corresponding drop in froth product ash was 3.03% to 2.98%. Again, the best material with respect to ash and floatability was increasingly greater in concentration in the feed

to the recleaner as the cleaner froth product ash was reduced. Also, it can be seen that there seemed to be a step-wise increase in performance going from the first cleaner stage to the second, however, the variation in yield was not as great from the recleaner as it was for the cleaner.

Comparing these simulated results with the R-S-C and modified R-S-C simulated results (see Figures 32 through 39), it can be ascertained that the best overall circuit performance with respect to yield was obtained with the R-S-C (modified) circuit, the lowest overall yield performance obtained with this circuit, the R-C-RC. On the other hand, this circuit arrangement provided for the lowest overall froth product ash from the circuit even though rougher stage results were very much the same with all three circuits examined during the simulations. It can be seen that the best overall cleaner stage performance was be obtained when the quality entering within the feed of this stage is limited to only the highest quality material, as was the case with the present circuit that was examined.

In comparison with the actual R-C-RC circuit results presented in Figures 28 through 31, it can be noted that the actual performance with respect to overall circuit concentrate yield and ash was exceptionally lower than simulated results for any of the circuit configurations here evaluated. Also, the range of froth product ash and yield values was much greater during actual operations.

With further comparison, it can also be seen that the range of variation with respect to concentrate yield and ash content seems to grow wider with each successive

stage of cleaning and recleaning in the actual R-C-RC test work, which is exactly opposite to what happened during the simulations. Furthermore, instead of a step-wise increase in performance when preceding to the next successive stage of cleaning, as was the case with the simulated results, there seemed to be a step-down in actual performance between the cleaner and recleaner as the 300 ml/minute feed flow rate was reached.

These drastic differences between actual and simulated results serve to further the concept that there are distinct problems with bubble carrying capacity, froth overloading, etc. inherent during the actual operation of columns that was not allowed for in the simulator. This problem seems to be compounded or magnified when there is an increasingly greater percentage of the more floatable, lower ash components appearing in a column stage feed, as is the case with one or more successive stages of cleaning and recleaning.

4.5 Simulation of Other Circuits

As was mentioned previously (see Section 4.4.1), it appeared that some additional advantage might be obtained in terms of recovery and grade if one or more stages of column flotation could be added to the simple three-stage R-S-C (modified) circuit case. In that it proved impractical to run actual circuit tests, due to the absolute requirement for additional process units and necessary support equipment, along with the physical problems in controlling such a large system, additional circuits employing more than three columns as well as feedback streams were examined utilizing the developed

simulator.

Figures 44 through 46 give the simulated results for a four-stage column circuit, noted as circuit AA, for three separate feed flow rates in the form of simple flowsheets. In this arrangement, the features of the R-S-C and R-S-C (Modified) circuits are further extended by the use of an additional cleaner column. The feed input characteristics input to the simulator were the same as previously used in Section 4.4.3 results for the three-stage circuit simulations with arbitrary counter-current wash water amounts maintained to give comparable retention times. The three feed slurry flow rates examined were 50 ml/min, 150 ml/minute, and 300 ml/minute. Feed slurry solids density was held constant at 5.00%.

Observing Figures 44 through 46, it can be seen that overall circuit recovery and yield dropped from 86.4%, 80.8% to 84.8%, 79.3%, respectively, as feed flow rate was increased from 50 ml/minute to 300 ml/minute. Overall circuit concentrate and tailings ash dropped from 3.23%, 35.70% to 3.43%, 33.65%, respectively, as the feed flow rate was increased. The rougher stage yield decreased from 83.5% to 80.8%, with an associated froth product ash drop of 3.43% to 3.26% as the feed rate was increased. The rougher leg cleaner stage and the scavenger stage maintained relatively constant yield values for this increase in feed flow rate, at average values of 95.6% and 15.64%, respectively. The scavenger leg cleaner had the most notable change in yield, increasing from 50.3% to 63.7% as feed flow rate was increased from 50 ml/minute to 300 ml/minute. Froth product ash values dropped slightly during this increase in feed flow

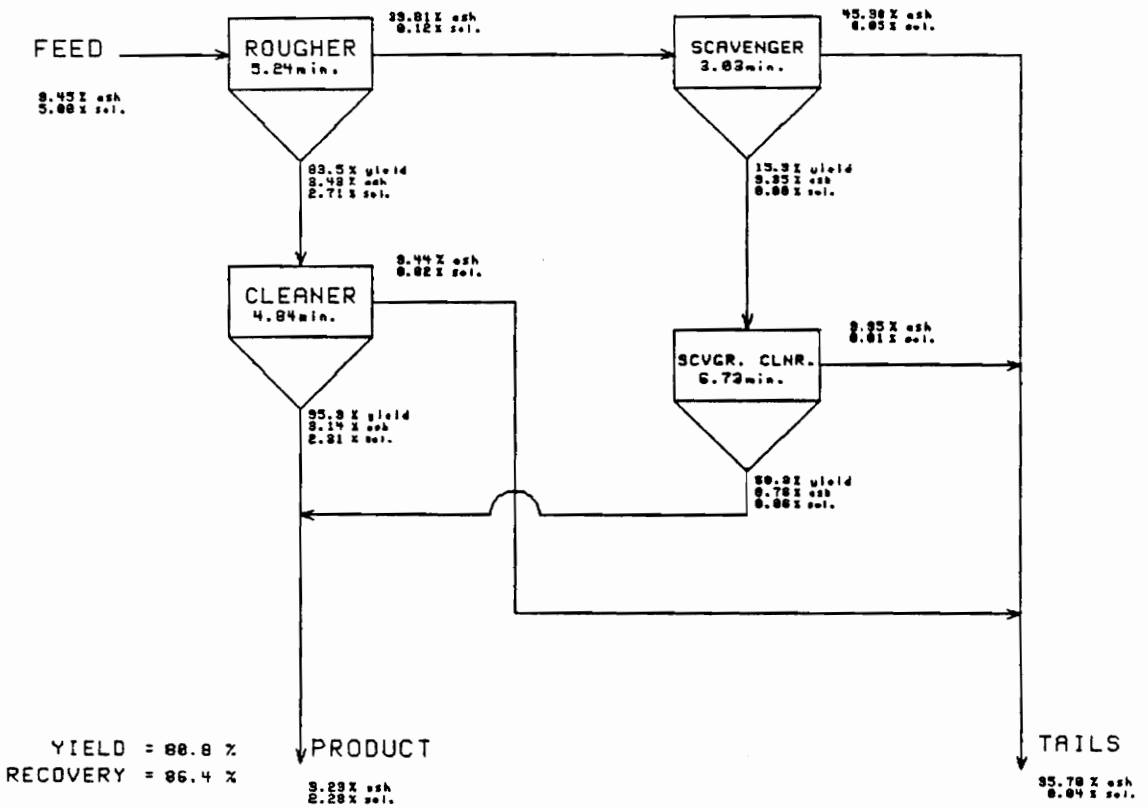
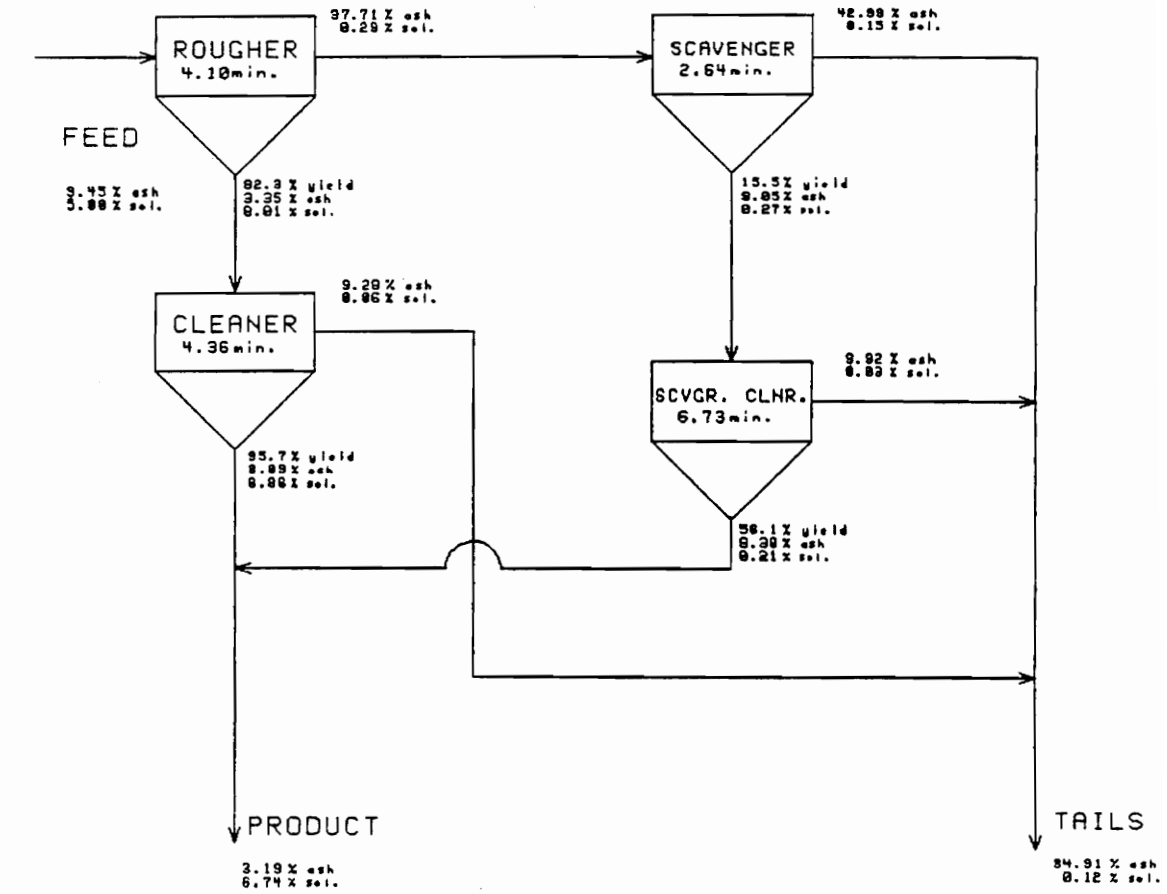


Figure 44. Simulated results for the 4-column "AA" circuit at a 50 ml/min feed flowrate.



YIELD = 80.3 %
RECOVERY = 85.2 %

Figure 45. Simulated results for the 4-column "AA" circuit at a 150 ml/min feed flowrate.

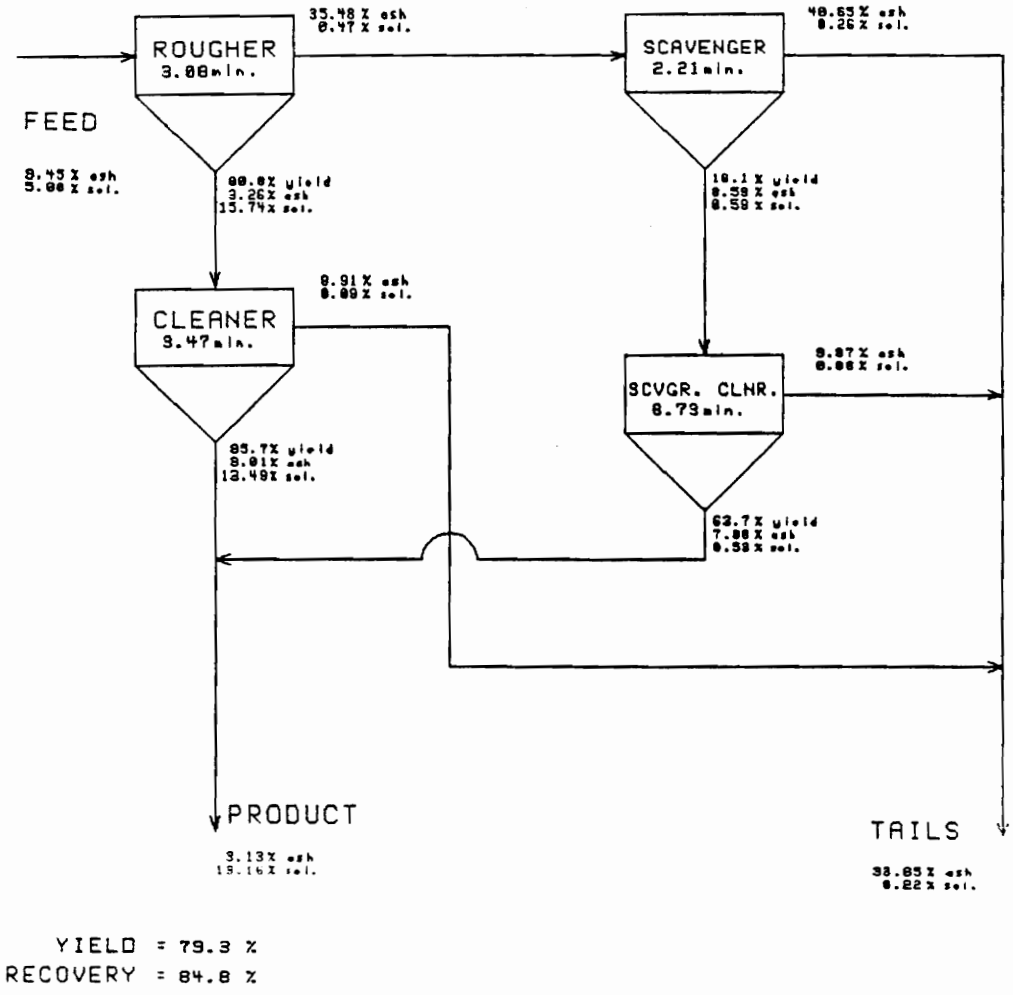


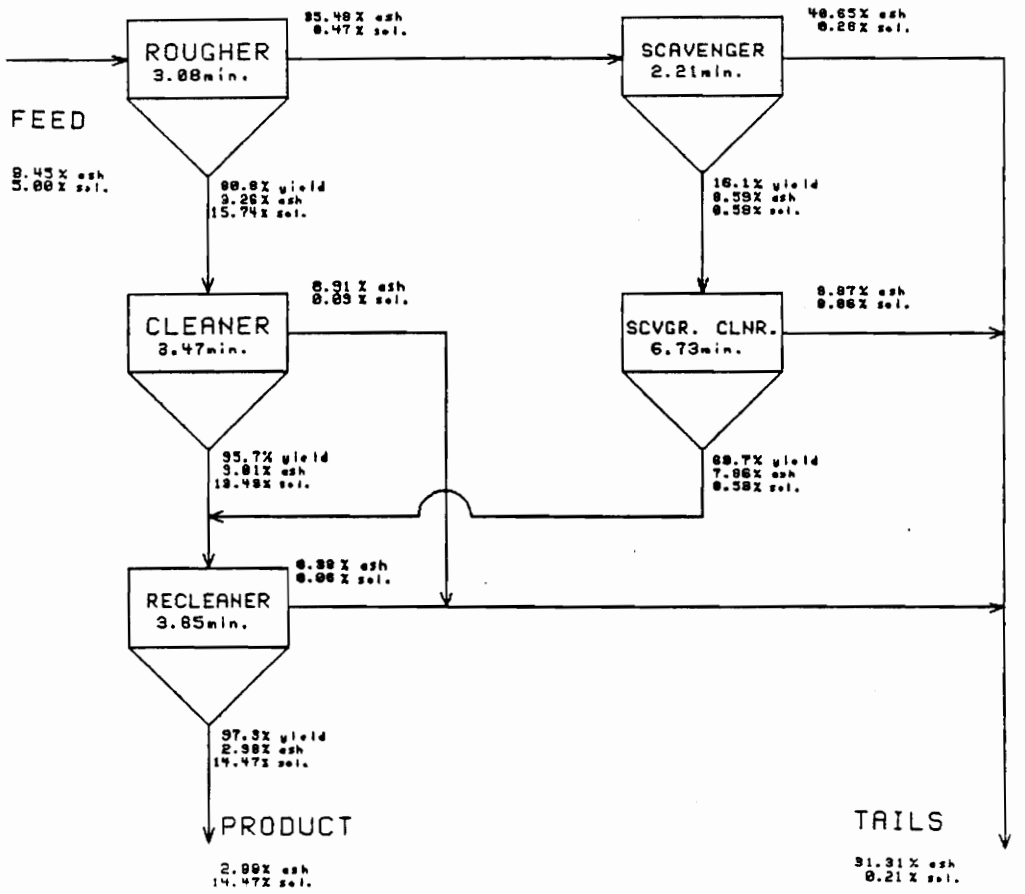
Figure 46. Simulated results for the 4-column "AA" circuit at a 300 ml/min feed flowrate.

rate.

Comparing these results with those of the R-S-C (Modified) circuit given in Figures 36 through 39, it can be seen that overall circuit yield and recovery was slightly improved by adding a fourth column to the circuit. Although a lower circuit ash was obtained with the three-stage circuit at the lowest feed flow rate of 50 ml/minute, it was slightly lower with the four-stage circuit at the highest feed flow rate of 300 ml/minute, giving a distinct advantage to this arrangement with respect to throughput.

Continuing along the same lines, Figure 47 gives the results for a five-stage circuit, circuit AB, simulated with the same input feed characteristics at a circuit feed flow rate of 300 ml/minute In this arrangement, the fifth column serves as a recleaner for both the rougher and scavenger product legs of the circuit. Comparing these results with those of the circuit AA at the same feed flow rate (see Figure 46), it can be seen that overall circuit recovery and yield were decreased from 84.8%, 79.3% to 82.5%, 77.0%, respectively. At the same time, circuit concentrate ash was decreased from 3.13% to 2.97%, which is a significant drop with respect to the changes that were observed thus far with the simulations. Thus, with this arrangement, it appeared that product quality was improved at the expense of recovery, much the same as the case of the R-C-RC circuit previously analyzed.

Early on in the test program, it was found desirable to study the effect of feedback streams upon overall circuit performance. However, due to the lack of a circuit control system as well as physical and equipment limitations, actual test work employing



YIELD = 77.2 %
RECOVERY = 82.7 %

Figure 47. Simulated results for the 5-column "AB" circuit at a 300 ml/min feed flowrate.

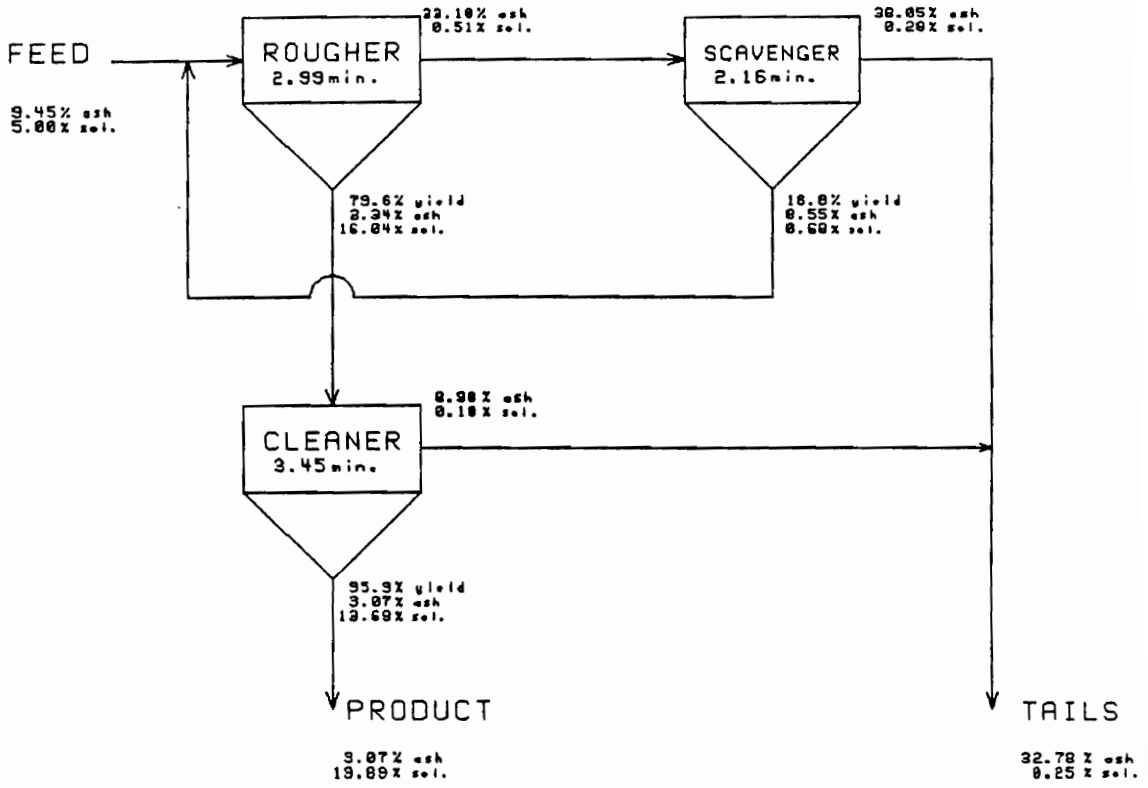
this methodology was not carried out. Thus, feedback arrangements were only examined in the context of simulations, utilizing the same feed input characteristics as above.

Figures 48 through 50 give the simulated results for three separate cases of feedback utilizing the simple R-S-C circuit in the form of simple flowsheets. Circuits AC, AD, and AE use feedback of the scavenger concentrate to the rougher feed, the combined scavenger concentrate and cleaner tails to the rougher feed, and the cleaner tails to the rougher feed, respectively. All these simulations employed a feed flow rate of 300 ml/minute.

As can be seen from the results presented by Figures 48 through 50, overall circuit recovery and yield ranged from 84.0%, 78.5% to 86.0%, 80.5%, respectively, going from circuit AC to circuit AE. With respect to recovery and yield, circuit AD was intermediate between the two cases. Similarly, circuit concentrate ash ranged from 3.07% for circuit AC to 3.18% for circuit AE.

Thus, for these three cases of feedback examined, it appeared that the best product recovery was obtained with only the feedback of the cleaner tails for reprocessing. It was interesting to note that the ash content of the cleaner tails stream of circuit AE, 9.17%, is very close to the head ash of the feed stream, 9.45%. Although best quality was obtained with circuit AC, recovery was apparently sacrificed due to the flow rate and low solids density of the scavenger concentrate stream that may have reduced the retention time of the rougher.

Comparing these results with the three-stage simulations of the R-S-C (Modified)



YIELD = 78.5 %
RECOVERY = 84.0 %

Figure 48. Simulated results for the 3-column "AC" circuit at a 300 ml/min feed flowrate.

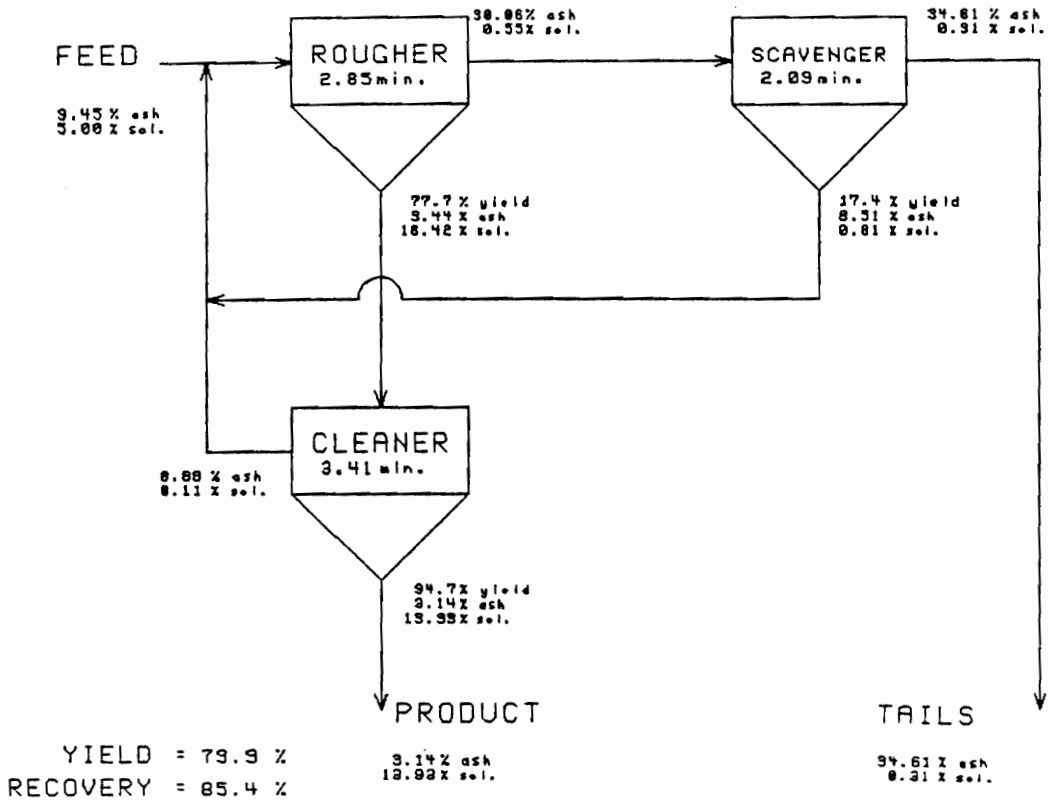


Figure 49. Simulated results for the 3-column "AD" circuit at a 300 ml/min feed flowrate.

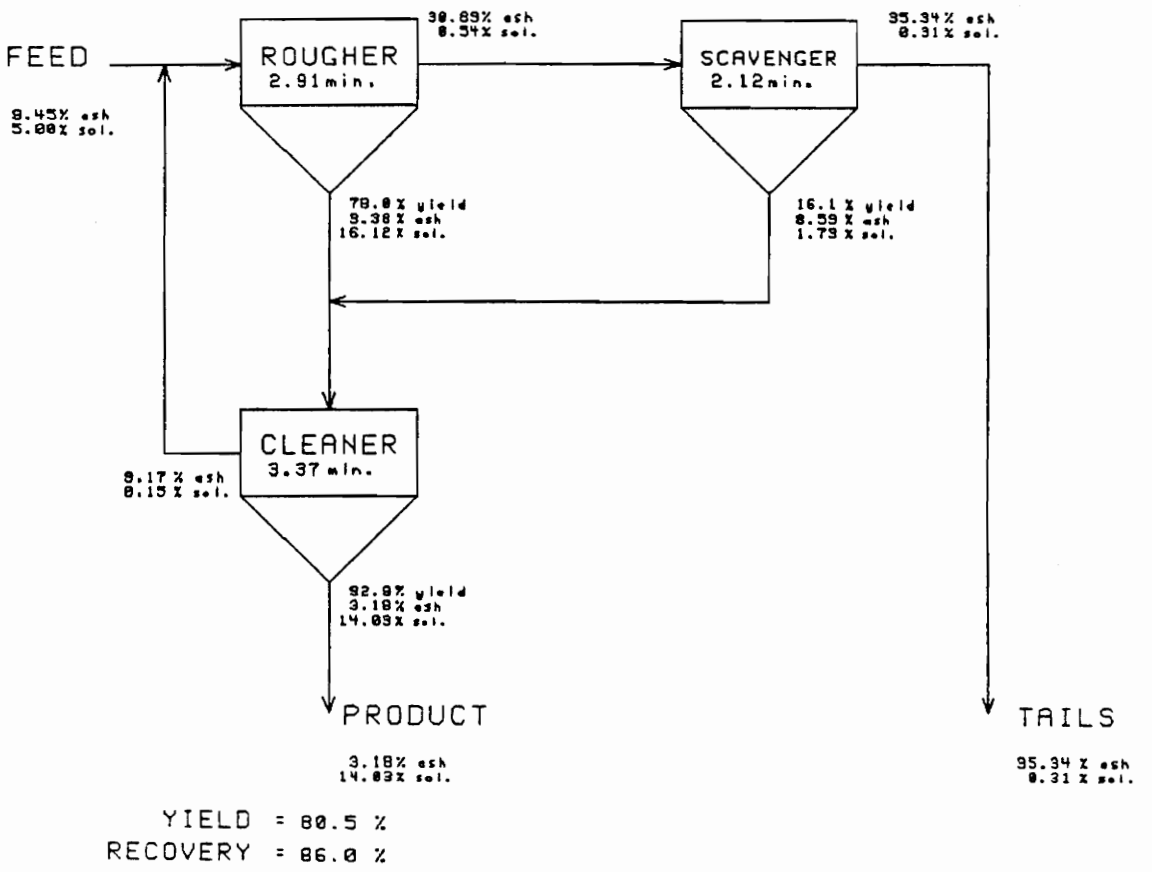


Figure 50. Simulated results for the 3-column "AE" circuit at a 300 ml/min feed flowrate.

circuit shown in Figure 39, it can be seen that the obtained recovery and product grade values were very close to what was obtained in that case. Thus, it can be concluded that the effective reduction in retention time by the use of feedback streams subdued any advantages to secondary recovery or reprocessing.

The final simulations performed as a conclusion to this study were to check if any further advantages might be gained from adding even more columns with the employment of some form of stream feedback. Figures 51 through 53 give simulated results for six-stage column circuits that employ feedback in one form ar another to make the comparison. Similar to what was utilized previously, all simulations were at a feed flow rate of 300 ml/minute, using the same feed input characteristics.

As can be seen from Figures 51 through 53, recovery and yield ranged from lowest values of 83.7%, 78.1% to the highest values, 85.3%, 79.7%, going from circuit AF to circuit AH, respectively. Overall circuit concentrate ash gave a similar increasing trend going from circuit AF to AH. Observing the results, it appeared that very slight differences in performance were obtained from one circuit arrangement to the other. However, with regard to product recovery, it was interesting to note that best performance was gained with circuit AH, in which the product and tailings legs of the circuit are kept more separate than in any other arrangement, with stream feedback limited to only the preceding stages within the respective process legs.

Comparing the results of circuit AH in Figure 53 with the results of the six-stage circuit AB shown in Figure 47, it can be seen that overall circuit product yield is

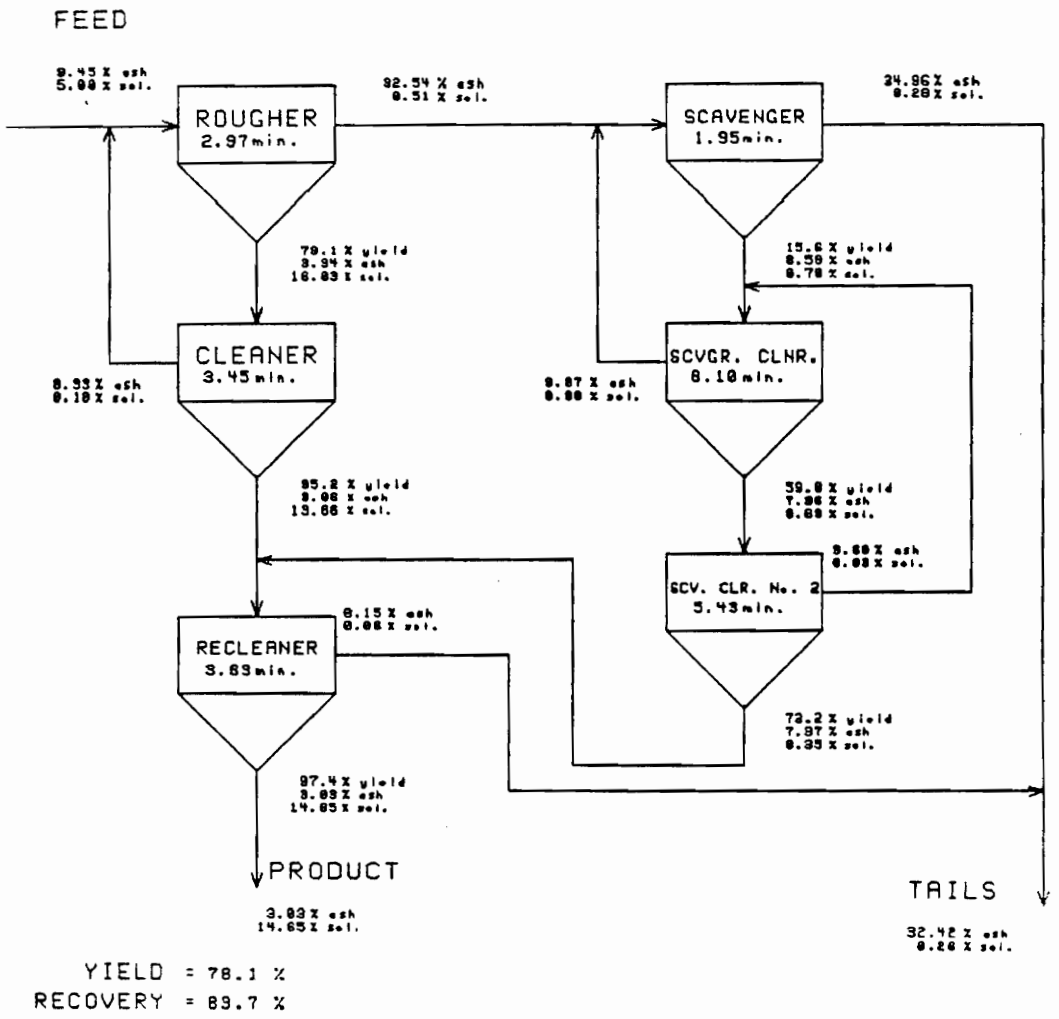


Figure 51. Simulated results for the 6-column "AF" circuit at a 300 ml/min feed flowrate.

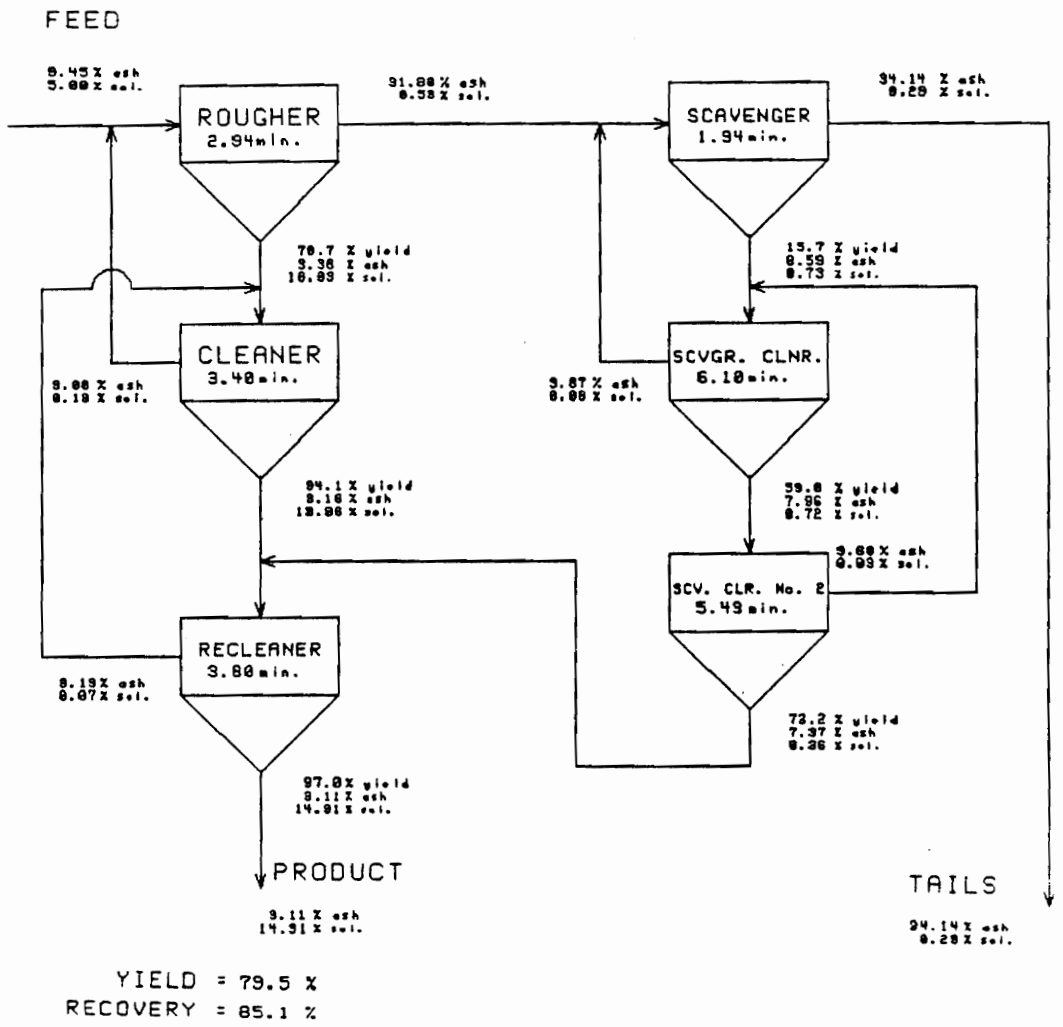


Figure 52. Simulated results for the 3-column "AG" circuit at a 300 ml/min feed flowrate.

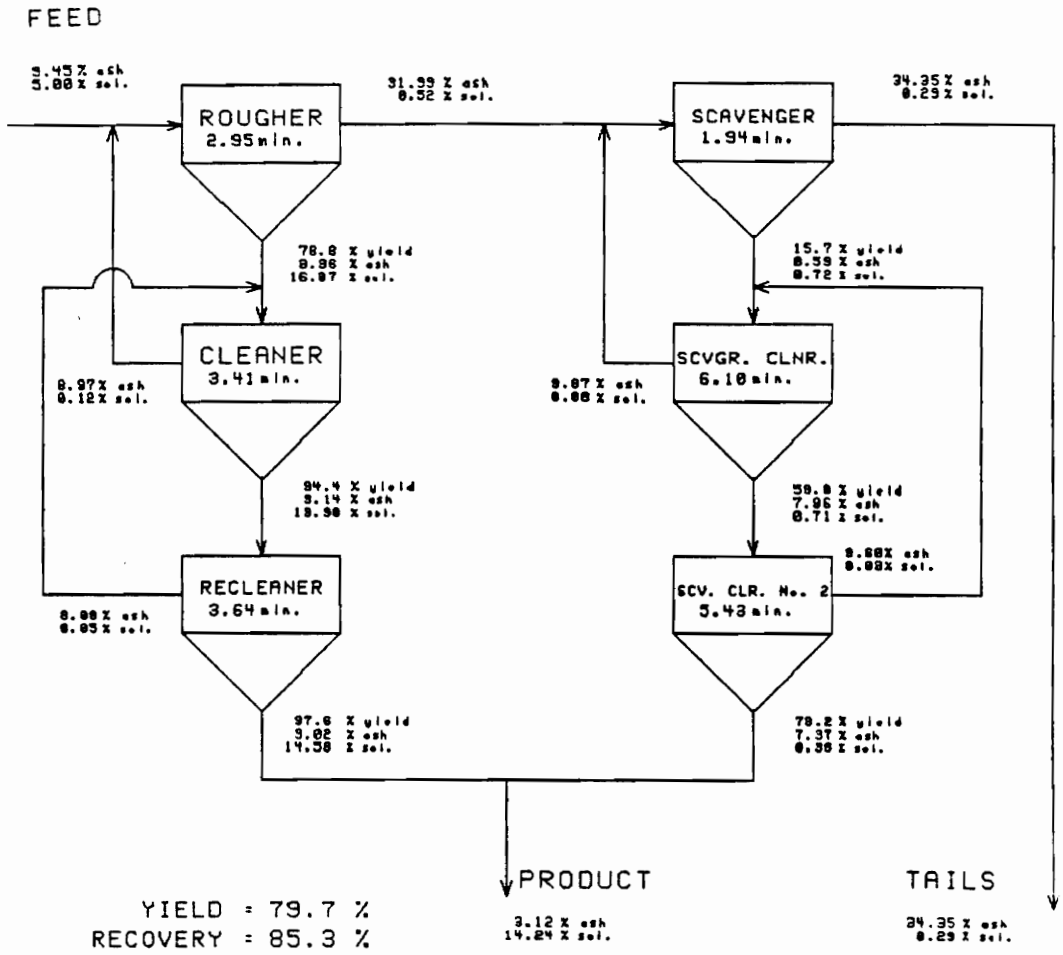


Figure 53. Simulated results for the 3-column "AG" circuit at a 300 ml/min feed flowrate.

improved but overall product quality is sacrificed. Considering overall results, not a large improvement in performance was obtained with regard to any of the criteria examined by using multiple stages or stream feedback.

5.0 DISCUSSION

5.1 Comparison of Circuit Test Results

5.1.1 Comparison of Recovery-Grade Curves

The results of the circuit tests conducted using the Pittsburgh No. 8 seam coal are summarized in Figures 54 and 55. These figures show the combustible recovery versus ash and sulfur rejections for each circuit over a wide range of feed flow rates. Although some degree of data scatter is noted, essentially the same recovery-grade relationship was obtained for each of various circuit configurations. The single-stage test results appear to be slightly inferior to those obtained with the other circuits, possibly owing to a small degree of particle entrapment (or possibly entrainment) in these tests. It is doubtful, however, that the slight improvement in performance would justify the costs associated with the multiple-cell circuit configurations.

According to the results shown in Figures 54 and 55, a combustible recovery of 90% will produce corresponding ash and sulfur rejections of approximately 73% and 58%, respectively. These rejections can be further improved to 78% and 67%, respectively, by allowing the recovery to fall to 80%. It is also interesting to note that the circuit test results fall on or just below the performance curve predicted by release analysis. The one data point which fell above the curve also had given a poor material balance, indicating that this data point was probably unreliable. The good agreement between the release analysis data and column flotation results suggests that the column was operating near its maximum expected level of performance.

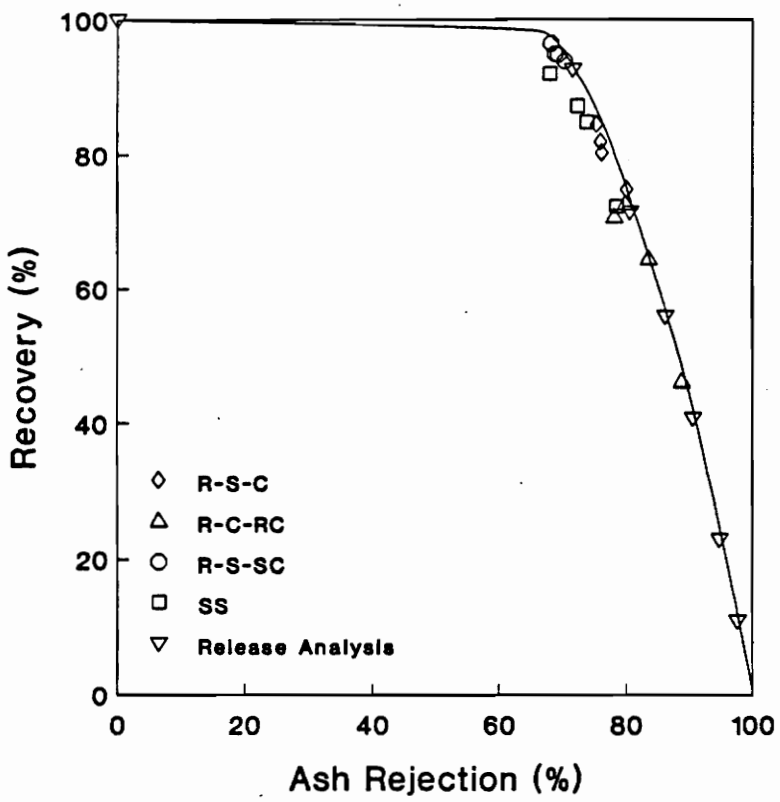


Figure 54. Effect of circuit configuration on combustible recovery and ash rejection.

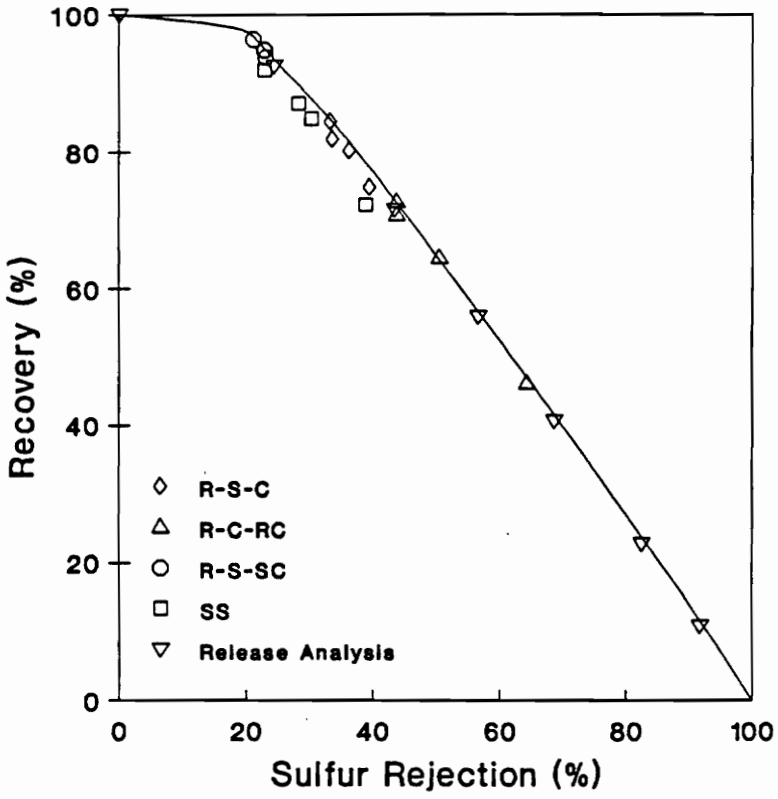


Figure 55. Effect of circuit configuration on combustible recovery and sulfur rejection.

5.1.2 Comparison of Throughput Capacity

The effect of circuit configuration on the relationship between circuit feed rate and individual column product rate is shown in Figure 56. For a given feed rate, the single-stage circuit produced the highest overall product mass rate, while the rougher-cleaner-recleaner circuit produced the lowest. With the exception of the last data point for the rougher-scavenger-cleaner circuit, all column circuits produced an increase in product capacity with increasing feed flow rate. The unit capacities (lb/hr/column) for each circuit can be ranked from highest to lowest in the following order:

$$\text{S-S} > \text{R-S-SC} > \text{R-S-C} > \text{R-C-RC}.$$

In terms of recovery and grade, the circuit configurations that produced the lowest overall recoveries produced the highest ash and sulfur rejections (as dictated by the respective recovery-grade curves).

5.2 Comparison of Actual and Simulated Results

5.2.1 Recovery and Retention Time

One of the largest deviations observed when comparing the actual column test results to simulated results, as well as to classical theories, was the trend of recovery versus retention time. First-order flotation kinetics suggests that as retention time is reduced, recovery falls off to the origin of the recovery-retention curve. However, the actual response appeared to be "skewed" or "shifted," with respect to the traditional response. Also, this effect appeared to be somewhat related to the level of solids

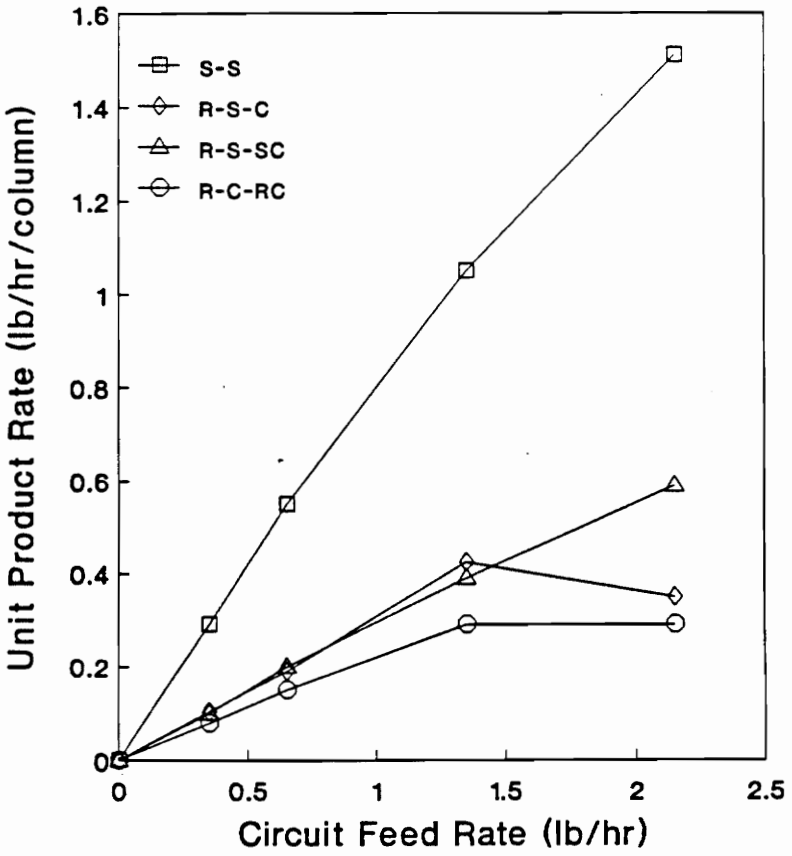


Figure 56. Effect of circuit configuration on column capacity.

concentration in the column feed slurry, i.e., the deviation increased as feed solids increased.

This style of flotation response was observed throughout the test work on into the circuit analysis testing. Also, the effect was more pronounced in the column cleaner and recleaner stages. Apparently, this effect was the result of froth overloading or inhibited flotation as mentioned by Lynch (1981). Also, it appeared that the recovery mechanism(s) present in column cells are much more sensitive to this factor than conventional flotation cells. This can be confirmed by comparing the single-stage column results of Figure 16 with the single-stage conventional flotation results of Figure 17. This is probably due to the lower surface area to volume ratio of the column as compared to the conventional cell.

As described by Lynch (1981), inhibited flotation is due to a deficiency of air bubbles or their capability to collect all recoverable or floatable material. This effect is more pronounced among the more floatable, higher grade species present within the flotation feed. Froth overloading is characterized by the inability of a developed froth phase to carry a load of recoverable components to the froth overflow lip. It is most pronounced among the coarser-sized particles that are floatable.

The carrying capacity limitation experienced with columns during this study appeared to be comparable to these conventional flotation effects, in that similar trends were noted. The presence of some form of inhibited flotation appeared to be magnified in column flotation when using these devices for successive stages of cleaning and

recleaning as shown in Figures 28 through 31 for the R-C-RC circuit test work. This effect was more apparent when greater concentrations of the more floatable, lower ash components were present in these downstream stages of cleaning. This was also reflected in the abrupt drop in final concentrate ash as the feed flow rate was increased.

Through review of the simulated results, it can be ascertained that the controlling model incorporated into the program gives a much more conventional response with respect to recovery versus retention time, with much less spread in performance for the same column or circuit throughput. Also, using the developed simulator, there seemed to be step-wise improvement with respect to yield and quality with successive stages of cleaning and recleaning as compared to the deteriorating performance during actual test work.

Frew (1981) showed that the performance of the cleaner stages deteriorated as compared to flotation response in the roughers. The more floatable, higher grade species exhibit a lower flotation rate as they were transferred from the roughing to the cleaning stages. In addition, there appeared to be a linkage of this drop in apparent flotation rate and an increase in the solids concentrations of the feeds. Frew and Trahar (1981) concluded that there was indeed inhibited flotation present in cleaner stages, and must be accounted for during the simulation of R-S-C type circuits.

Considering the already present carrying capacity problem inherent in single-stage column flotation, it becomes readily apparent that this effect is even more pronounced when progressing to successive stages of cleaning and recleaning. Also, initial

limitations of the developed model in the roughing stage would seem to lead to erroneous results during scavenger simulation even though these units could very well be operating under the mechanisms of free flotation. Thus, it becomes apparent that some means must be incorporated into the simulator to account for the presence of inhibited flotation, possibly a refined recovery model for single-stage columns and the lowering of flotation rate constants for certain floatable species relative to the stage of operation, especially for cleaners and recleaners.

5.2.3 Rate Constant Selection

The methods utilized during this study for the identification of flotation feed components and the determination of their flotation rate constants appeared to be suitable and consistent with the literature given to date. Release curve construction as given by Dell (1972) can be related to the idea of perfect separation as presented by Imaizuni and Inoue (1965). Jowett and Smith (1985) and Salama (1989) furthered this analysis by the demonstrated use of M-curves for prediction of perfect separation performance.

When considering M-curves (recovery versus incremental grade), it becomes readily apparent that incremental grade axis is directly related to the dimensional grade axis of release curves. Thus, a flotation release curve should give a picture of perfect, or ultimate flotation. Furthermore, if a representation of perfect separation could be obtained, during a flotation process, the performance of individual components could be traced and identified.

The grouping of components by the relationship of similar response tendencies, or in the case of a curve by transition points, appeared reasonable, in that, similar behaving components or groups of components should give similar flotation responses with respect to the rate of flotation. This observation falls back on the work of Bushell (1962), Imaizuni and Inoue (1965), and Aplan and Rastogi (1985), where floatable fractions were identified and broken down into a fast-floating and a slow-floating component.

The deviations experienced between actual recovery-grade results (see Figure 15), and the hypothetical smoothing required to fit flotation components and components to the single-stage flotation results of coal B, does not appear to be inherent in the methodology used to determine flotation components and their respective rates of flotation. Conversely, this deviation appeared to be more the result of the inherent carrying capacity problem inherent with actual single-stage flotation. The simulator, which was used for fitting the rates, does not incorporate this effect at this time. Perhaps future release curve results, along with a modified recovery model incorporated into the developed simulator can be used for fitting more accurate rates without the need for subsequent smoothing.

5.2.4 Solids Recovery

Another area of deviation between actual and simulated results appeared to be the level of solids concentration reporting to the overflow lip of the column(s). From the

results presented in Sections 4.4 and 4.5, there appeared to be a consistently higher concentrate solids level predicted by the simulator than what was actually obtained. One reason for this deviation could be the result of the fact that in the actual tests, there was the problem of carrying capacity limitation which resulted in a lower recovery of solids as compared to what was predicted by the model. Also, as mentioned previously, the recovery model includes a simple empirical term to describe solids recovery based on the amount of counter-current wash water flow rate as described by Luttrell (1990). According to Tomlinsen and Fleming (1965) and Lynch (1981), froth phase behavior can disrupt the relationship of the first-order kinetic model substantially, especially in the presence of inhibited flotation. The simulation work by Lynch (1981) also made use of the first-order kinetic approach for the recovery of water, and assigned it a flotation rate constant.

The problem of solids recovery, along with substantial influence from the froth phase region can lead to erroneous results for column circuit simulation that could effect mass balancing results for the entire circuit. It appeared that the deviation in solids recovery for the actual tests was not the result of counter-current wash water addition level, as this was taken into account during the simulations through the use of the empirical relationship. However, several other factors could have affected froth phase behavior during the actual test work. As described by Lynch (1981), these factors could possibly include drainage, entrainment, and froth stability as related to carrying capacity.

Thus, another area for increased study in the future would be the empirical testing

of froth phase behavior in columns, the recovery of water from the pulp region through the froth region to the overflow launder, the development of suitable relationships to describe these observations, and the incorporation of these relationships into an overall, improved recovery model for column flotation.

5.2.5 Physical Circuit Operation

An additional area of consideration with respect to the deviation between actual and simulated results during this study could be in the physical operation of the circuit. As was noted previously (see Section 4.5), in order to make simulated column stage retention times match actual retention times, it was necessary to add or delete an arbitrary amount of counter-current wash water during the simulation routine. During actual circuit operation, the pressurized froth sprays had to be utilized to prevent froth build-up around the top of the columns. Not only did the sprays add extra water to feeds going to successive stages, the build-up and intermittent accumulation of froth may have affected solids transfer within the circuits even though all stages appeared to be operating at steady-state.

Secondly, considerable time had to be allowed for the feed sumps transferring material to succeeding stages to reach appropriate levels before initialization of the operation of the next column. Although steady-state operation of the circuits were assumed, the actual feed characteristics, especially solids concentration, to succeeding stages may have been different than what was actually being discharged from the

preceding column.

Third, during the course of the actual test work, it was observed that there was a continuing matter of variation in air hold-up values and frother dosages going from stage to stage within the circuit, and from feed rate to feed rate for the same column. Early on in the program, it was noticed that no applicable means were available to control air hold-up values on-line using the respective frother dosages for each column. Any deviation in air hold-up could result in the alteration of bubble size distributions, which would affect a variation in species flotation rates as pointed out by Yoon and Luttrell (1986).

Thus, it was confirmed that a more suitable means needed to be developed in order to control the parameters of an individual column to insure predictable and consistent operation. Also, it appeared early in the actual test work that overall circuit mass balance and material flow rate monitoring and control was necessary for the entire circuits to ensure more realistic results. Areas worthy of future investigation and improvement would be the monitoring of solids density at key circuit locations, automatic control of water additions, identification of various stream flow rates, and the adjustment of froth product discharge hardware to better insure uninterrupted flows.

5.3 Simulator Applicability

5.3.1 Simulator Limitations

In summary, a number of observations have been gained thus far as to the

limitations of the simulator developed in the present work. In general, the simulator appears to predict trends in flotation response reasonably well, but is incapable of accurately estimating actual recoveries or product grades. Several reasons can be given for this shortcoming. First, the greatest single limitation of the simulator appears to be in the basic recovery model, which is at the heart of the simulation routine, in its ability to predict performance in light of the substantial carrying capacity problem inherent with column flotation. Current research by Luttrell (1990) is aimed at developing a modified recovery equation for axially-dispersed plug-flow mixing that would account for froth overloading. This research is directed at improving the basic recovery model for columns through its breakdown into the fundamental effects of bubble size, and the distributions of these bubble sizes, and the relationship to flotation rates. Perhaps this approach will not only improve the basic model in the developed simulator, but also lead to a more fundamental understanding of all the mechanisms that are taking place during column flotation as described by Yoon and Luttrell (1986).

Another approach to reducing the limitation of the basic model with respect to carrying capacity could be the use of modified rate constants, much the same as Lynch (1981) described for conventional flotation in the presence of inhibited flotation. This development could prove especially useful with regard to the simulation of succeeding column stages such as cleaner and recleaner stages as was given by Frew (1981) for conventional flotation circuits.

Finally, the behavior of the froth phase and the recovery of water during column

flotation should be examined in more detail. The interaction of such factors as entrainment, drainage, solids carrying capacity, and flotation rate should be included to improve the predictive capabilities of the flotation model.

5.3.2 Optimum Circuit Configuration

The previous section outlined some of the limitations inherent in the simulator as it is developed to date. It appeared that these limitations, for the most part, are due to the inability of the basic model to take fully into account the mechanism(s) that are present during column flotation. However, as was demonstrated in Section 4.6, the simulator could prove quite useful for the analysis and optimization of complex circuit arrangements. The observations obtained during these simulations seemed to confirm past literature and studies with respect to the optimum configuration of circuits.

Imaizuni and Inoue (1965) proposed using flotation rate constants as a means of describing flotation behavior, analogous to the use of specific gravity to characterize sink-float separations. If the flotation is performed over a certain specified range of feed rates, within limits, the separation can be considered to be linear. In other words, assuming the flotation equipment is operating within an adequate capacity range, at steady-state, first-order flotation can be assumed, which is to say that the performance of individual particles are independent of the activity of other particles, or their respective amounts present in the feed. This is also the definition of a linear separation. The assumptions presented here are inherent in the first-order kinetic model for recovery

as presented by Levenspiel (1972), which was incorporated into the simulator.

Following studies by Meloy (1983), where linearity was assumed, several conclusions and observations can be noted. For example, it was observed that optimum circuit operation occurred when process legs or streams (i.e., concentrate and tailings) were not crossed or recombined. This was also confirmed in Section 4.6, where circuits AB and AH gave the best apparent results, where stream leg crossover was held to a minimum. Also, as was shown by Meloy and Williams (1989), performance of the roughing stage of a circuit has the most pronounced effect upon overall circuit performance. Again, as was shown throughout the simulations (see Figures 32 through 53), this concept seemed to hold true.

It has also been shown by some researchers that in order to achieve optimum circuit results, various process streams should be beneficiated to the same level of incremental grade as the feed stream. These observations were presented by Gottfried and Cierpisz (1977), Agar et. al. (1980), and Jowett and Smith (1985). In addition to these findings, Agar et. al. (1980) agreed with Meloy (1983) that process stream legs should not be crossed or recombined. Taggart (1945), Jowett and Sutherland (1985) and Meloy and Williams (1989) stated that best circuit performance was obtained when streams of similar grade were recombined and sent to the head of the circuit for further processing prior to discharge. This concept was confirmed when the R-S-C (Modified) circuit was simulated (see Figures 36 through 39 and Figure 50).

Finally, it was observed by Sutherland (1981) that circuit performance was

relatively insensitive to arrangement unless limiting conditions are being approached or superceeded. Sutherland (1981) also confirmed that the most important aspects of the rougher was for recovery and the cleaner was for grade performance. These observations were also confirmed with the simulator, as change in circuit performance remains relatively small, even when multiple stages of cleaning and feedback were employed. Also, the use of the rougher stage to control recovery and the cleaner stage to control final product grade seemed to be upheld through the simulations.

Even though some limitations do exist with respect to the basic recovery model, with respect to column flotation, it appeared that a suitable tool had been designed and developed in the overall simulator that can be used for the analysis of flotation circuit arrangements.

6.0 SUMMARY

An experimental test program was undertaken to investigate the effects of column circuit configuration on the flotation of finely pulverized coal. The test work involved two separate coals, both from the Pittsburgh No. 8 seam. The first coal, i.e., coal A, was utilized for establishing the flotation behavior of fine coal in both column-type and conventional flotation cells. The second coal, i.e., coal B, was used to study the performance of single-stage and multiple-stage column flotation circuits. Three different circuit arrangements utilizing three separate columns were constructed and operated in order to obtain the comparative data.

In addition to the primary test work, release analysis tests were performed on coals A and B to establish input data for a column flotation circuit simulator. It was shown that this technique may be employed for the identification of various components within a particular feed coal and the back-calculation of flotation rate constant values for each component. This information was then incorporated into a flotation circuit simulation package for both column and conventional flotation circuit arrangements. The simulator was developed to facilitate the design and evaluation of complex flotation circuits.

Several hypothetical column flotation circuits were examined with the simulator to demonstrate its potential capabilities. In almost all cases studied, the simulated results appeared to agree with recent observations relative to the analysis of flotation and other process circuit arrangements. Thus, it appeared that the simulator offered a suitable tool

for exploring the optimization of various complex circuits. However, at least two primary areas of limitation were identified with the circuit simulator as developed to date. Primarily, it was found that some form of inhibited flotation is prevalent with flotation columns as compared to what has been experienced with conventional flotation cells. In that no provision has yet been made to account for this factor in the basic recovery model, some deviations between actual and simulated values were obtained. Also, there appeared to be some discrepancy between solids recovery values as predicted by the simulator and actual values. Again, there seemed to be a need for adjustment of the basic recovery model and perhaps an accounting of performance within the froth phase region of the column flotation process.

Several important areas for future research have been identified from the finding of the present work. These include the development of improved mathematical relationships for predicting the carrying capacity of the froth phase, the investigation of specific areas such as regrinding, the processing of minerals other than coal, and the development of configurations combining conventional and column flotation cells within the same circuit.

7.0 CONCLUSIONS

Several general observations can be made in light of the test data collected in the present work. The most significant of these include:

1. Column flotation offers an attractive means for the beneficiation of extremely fine sized coal (-200 or -325 mesh). This capability is likely to become more important as market requirements dictate deeper levels of coal cleaning.
2. Experimental results obtained from circuit testing demonstrate that differences in the performance of individual column circuits are due to shifts along a single recovery-grade curve. This finding can be attributed to the ability of column flotation cells to eliminate the entrainment problem in a single stage of processing. Therefore, there appears to be little advantage in using multi-stage column circuits for the deep cleaning of Pittsburgh No. 8 seam coal.
3. One of the major shortcomings of column flotation is that the mass rate of froth product is limited by the froth carrying capacity. This problem becomes more significant with increases in the solids content of highly floatable, high grade particles within a particular feed.
4. The rate-based column circuit simulator constructed in the present work offers an attractive method for rapidly evaluating several alternative circuit arrangements during the design or process optimization phases. However, the predicted results are often better than those obtained from the actual test runs. This problem stems from the fact that the simulator is rate-based and, hence, cannot predict

limitations associated with the froth carrying capacity.

5. The technique of conventional flotation release analysis can be utilized for the identification of coal flotation feed components and the determination of their respective grades and rates of flotation when used in conjunction with existing single-stage flotation data. The technique provides a less time consuming alternative than microscopic analysis to obtain characteristic feed data for subsequent computer simulations.

8.0 RECOMMENDATIONS FOR FUTURE WORK

From these various conclusions and observations, some of the major areas for future consideration are as follows:

1. The basic axially-dispersed plug-flow recovery model as given by Levenspiel (1972) needs to be modified or supplemented by appropriate means to account for effects of inhibited flotation and solids recovery. Perhaps future research of the column froth phase region and elementary investigation into the relationship of column recovery with respect to bubble size distribution will reveal the fundamental relationships and mechanisms necessary for this modification.
2. The implementation of a more precise and accurate system for monitoring and control of the existing three-stage column circuit needs to be undertaken. This will affect easier physical operation and more precise actual data with which to compare to a modified form of the circuit simulator.
3. The simulator should be utilized for comparison with conventional flotation circuit results obtained from actual plants. In this manner, a check can be made to see if the principles of circuit arrangement and optimization still hold true with previous theory.
4. Other areas for future investigation should examine minerals beneficiation processes other than coal. Also, some means should be incorporated into the simulator for taking into account regrinding and combination circuits that utilize both column and conventional flotation cells.

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Appendix A:

Experimental Test Results

TABLE A1: Operating Conditions, Microbubble Column Coal A, 60 mesh x 0 Particle Size

Feed, ml/min	DF1012, kg/ton	T, min	E, %
500.0	1.20	1.90	22.96
1000.0	1.16	1.20	21.40
1500.0	1.16	0.75	25.60
2250.0	1.24	0.61	26.27

* T - Retention Time as calculated from recovery equation.

* E - Fractional Air Hold-up as calculated from transducer readings.

TABLE A2: Microbubble Flotation Results, Coal A, 60 mesh x 0 Particle Size

Size Class	T, min.	Yield, %	Ash, %	Sulfur, %
+ 100 mesh	0.61	9.95	6.63	-
+ 100 mesh	0.75	53.13	6.20	2.36
+ 100 mesh	1.20	99.99	8.71	2.26
+ 100 mesh	1.90	99.99	7.41	2.08
100 x 200	0.61	17.53	5.36	2.66
100 x 200	.75	48.55	5.82	2.69
100 x 200	1.2	98.81	7.80	2.71
100 x 200	1.9	99.99	6.69	2.28
200 x 270	0.61	16.45	4.63	2.52
200 x 270	0.75	52.54	5.37	2.78
200 x 270	1.20	97.50	7.42	3.12
200 x 270	1.90	99.99	6.89	2.56

* T - Retention Time as calculated from recovery formula.

* Yield values are calculated as percentage of respective size class.

* Sulfur measured as total sulfur.

TABLE A2 (Continued): Microbubble Flotation Results, Coal A, 60 mesh x 0 Particle Size

Size Class	T, min.	Yield, %	Ash, %	Sulfur, %
270 x 400	0.61	18.56	4.14	2.56
270 x 400	.75	58.57	5.65	2.95
270 x 400	1.2	95.84	7.42	3.37
270 x 400	1.9	99.83	6.83	2.58
- 400 mesh	0.61	20.65	4.58	2.31
- 400 mesh	0.75	51.14	7.56	2.50
- 400 mesh	1.20	83.13	9.27	2.84
- 400 mesh	1.90	93.08	10.19	2.85

* T - Retention Time as calculated from recovery formula.

* Yield values are calculated as percentage of respective size class.

* Sulfur measured as total sulfur.

TABLE A3: Operating Conditions, Microbubble Column Coal A, 200 mesh x 0 Particle Size 5.0 % Solids

Feed, ml/min	DF1012, kg/ton	T, min	E, %
150.0	1.48	3.86	18.87
200.0	1.39	3.55	18.33
300.0	1.37	3.17	17.94
400.0	1.41	2.93	17.55
600.0	1.41	2.58	19.85

* T - Retention Time as calculated from recovery equation.

* E - Fractional Air Hold-up as calculated from transducer readings.

* Counter Current Wash Water - 610 ml/min.
  Air Flowrate - 1300 ml/min.
  Froth Height - 12 in.

TABLE A4: Microbubble Flotation Results, Coal A, 200 mesh x 0 Particle Size, 5.0 % Feed Solids

T, min	Yield, %	Ash, %	Sulfur, %
2.58	70.4	5.38	2.37
2.93	78.0	4.66	2.23
3.17	82.3	4.78	2.33
3.55	84.9	4.99	2.34
3.86	87.2	5.38	2.37

* Percentage Ash and Sulfur reflect cumulative froth product results.

* Sulfur measured as total sulfur.

TABLE A5: Operating Conditions, Microbubble Column Coal A, 200 mesh x 0 Particle Size 2.0 % Solids

Feed, ml/min	DF1012, kg/ton	T, min	E, %
200.0	1.67	3.25	17.98
400.0	1.54	2.58	(18.00)
600.0	1.51	2.11	(18.00)

* T - Retention Time as calculated from recovery equation.

* E - Fractional Air Hold-up as calculated from transducer readings.

* Counter Current Wash Water - 610 ml/min.
Air Flowrate - 1500 ml/min.
Froth Height - 11.5 in.
(-) - Assumed Values.

TABLE A6: Microbubble Flotation Results, Coal A, 200 mesh x 0 Particle Size, 2.0 % Feed Solids

T, min	Yield, %	Ash, %	Sulfur, %
2.11	68.4	4.27	2.30
2.58	75.4	4.61	2.33
3.25	82.8	5.18	2.35

* Percentage Ash and Sulfur reflect cumulative froth product results.

* Sulfur measured as total sulfur.

TABLE A7: Operating Conditions, Microbubble Column Coal B, 325 mesh x 0 Particle Size 2.0 % Solids

Feed, ml/min	DF1012, kg/ton	T, min	E, %
100.0	1.19	5.31	15.18
200.0	0.12	3.78	13.43
400.0	0.50	2.89	14.05
600.0	0.61	2.41	13.55

* T - Retention Time as calculated from recovery equation.

* E - Fractional Air Hold-up as calculated from transducer readings.

* Counter Current Wash Water - 400 ml/min.
  Air Flowrate - 1500 ml/min.
  Froth Height - 15 in.

TABLE A8: Operating Conditions, Microbubble Column Coal B, 325 mesh x 0 Particle Size 5.0 % Solids

Feed, ml/min	DF1012, kg/ton	T, min	E, %
100.0	0.47	4.87	15.93
200.0	0.37	4.33	15.93
300.0	0.39	4.22	13.41
400.0	0.45	3.75	14.30

* T - Retention Time as calculated from recovery equation.

* E - Fractional Air Hold-up as calculated from transducer readings.

* Counter Current Wash Water - 400 ml/min.
  Air Flowrate - 1500 ml/min.
  Froth Height - 15 in.
  Collector - 0.45 kg/ton kerosene.

TABLE A9: Operating Conditions, Microbubble Column Coal B, 325 mesh x 0 Particle Size 10.0 % Solids

Feed, ml/min	DF1012, kg/ton	T, min	E, %
100.0	0.29	4.90	17.58
200.0	0.32	4.46	13.30
300.0	0.40	3.99	12.14
400.0	0.43	3.18	9.45

* T - Retention Time as calculated from recovery equation.

* E - Fractional Air Hold-up as calculated from transducer readings.

* Counter Current Wash Water - 400 ml/min.
  Air Flowrate - 1500 ml/min.
  Froth Height - 15 in.

TABLE A10: Operating Conditions, Semi-Continuous Conventional Cell, Coal A, 60 mesh x 0 Particle Size

Feed, ml/min	DF1012, kg/ton	T, min
500.0	1.57	11.16
1500.0	1.59	2.44
2000.0	1.55	1.60

* T - Retention Time as calculated from recovery equation.

Impeller Speed - 1200 rpm
Feed Solids - 5.0 %
Collection Paddle Speed - 60 rpm

TABLE A11: Semi-Continuous Conventional Cell Flotation Results

T, min	Yield, %	Ash, %	Sulfur, %
1.60	85.3	9.43	3.25
2.44	89.6	9.88	3.11
11.16	96.8	9.05	-

* T - Retention Time as calculated from recovery equation.

Impeller Speed	- 1200 rpm
Feed Solids	- 5.0 %
Collection Paddle Speed	- 60 rpm

* Sulfur measured as total sulfur.

TABLE A12: Operating Conditions R-S-C Circuit Coal B, 325 mesh x 0 Particle Size

Stage	DF1012, kg/ton	E, %	T, min
Rougher	1.30	17.06	5.49
Scavenger	0.33	10.20	2.97
Cleaner	0.10	14.51	4.66

* Circuit feedrate - 50 ml/min.

* E - Fractional air hold-up.

* T - Retention time as calculated from recovery equation.

	All Stages
* Counter-current Wash water	- 400 ml/min.
Air flowrate	- 1500 ml/min.
Froth height	- 15 in.

TABLE A13: Operating Conditions R-S-C Circuit Coal B, 325 mesh x 0 Particle Size

Stage	DF1012, kg/ton	E, %	T, min
Rougher	0.82	16.19	4.52
Scavenger	0.20	9.84	2.88
Cleaner	0.29	14.04	4.43

* Circuit feedrate - 100 ml/min.

* E - Fractional air hold-up.

* T - Retention time as calculated from recovery equation.

	All Stages
* Air flowrate	- 1500 ml/min.
Counter-current Wash water	- 400 ml/min.
Froth height	- 15 in.

TABLE A14: Operating Conditions R-S-C Circuit Coal B, 325 mesh x 0 Particle Size

Stage	DF1012, kg/ton	E, %	T, min
Rougher	1.06	13.75	4.34
Scavenger	0.10	10.80	2.76
Cleaner	0.12	11.28	4.03

* Circuit feedrate - 200 ml/min.

* E - Fractional air hold-up.

* T - Retention time as calculated from recovery equation.

	All Stages
* Air flowrate	- 1500 ml/min.
Counter-current wash water	- 400 ml/min.
Froth height	- 15 in.

TABLE A15: Operating Conditions R-S-C Circuit Coal B, 325 mesh x 0 Particle Size

Stage	DF1012, kg/ton	E, %	T, min
Rougher	0.70	11.34	3.95
Scavenger	0.29	13.30	2.58
Cleaner	0.12	8.14	3.80

* Circuit feedrate - 300 ml/min.

* E - Fractional air hold-up.

* T - Retention time as calculated from recovery equation.

	All Stages
* Air flowrate	- 1500 ml/min.
Counter-current wash water	- 400 ml/min.
Froth height	- 15 in.

TABLE A16: Operating Conditions R-S-C (Modified) Circuit, Coal B, 325 mesh x 0 Particle Size

Stage	DF1012, kg/ton	E, %	T, min
Rougher	1.88	17.49	5.63
Scavenger	3.28	13.74	3.07
Cleaner	4.10	10.87	6.88

* Circuit feedrate - 50 ml/min.

* E - Fractional air hold-up.

* T - Retention time as calculated from recovery equation.

	All Stages
* Air flowrate	- 1500 ml/min.
Counter-current wash water	- 400 ml/min.
Froth height	- 15 in.

TABLE A17: Operating Conditions R-S-C (Modified) Circuit, Coal B, 325 mesh x 0 Particle Size

Stage	DF1012, kg/ton	E, %	T, min
Rougher	1.38	19.18	4.26
Scavenger	0.24	12.65	2.73
Cleaner	3.16	14.01	4.74

* Circuit feedrate - 100 ml/min.

* E - Fractional air hold-up.

* T - Retention time as calculated from recovery equation.

	All Stages
* Air flowrate	- 1500 ml/min.
Counter-current wash water	- 400 ml/min.
Froth height	- 15 in.

TABLE A18: Operating Conditions R-S-C (Modified) Circuit, Coal B, 325 mesh x 0 Particle Size

Stage	DF1012, kg/ton	E, %	T, min
Rougher	1.10	17.96	4.06
Scavenger	0.49	9.71	2.81
Cleaner	0.48	16.21	5.32

* Circuit feedrate - 200 ml/min.

* E - Fractional air hold-up.

* T - Retention time as calculated from recovery equation.

	All Stages
* Air flowrate	- 1500 ml/min.
Counter-current wash water	- 400 ml/min.
Froth height	- 15 in.

TABLE A19: Operating Conditions R-S-C (Modified) Circuit, Coal B, 325 mesh x 0 Particle Size

Stage	DF1012, kg/ton	E, %	T, min
Rougher	1.05	21.39	3.64
Scavenger	0.89	13.02	2.69
Cleaner	0.47	9.63	5.23

* Circuit feedrate - 300 ml/min.

* E - Fractional air hold-up.

* T - Retention time as calculated from recovery equation.

	All Stages
* Air flowrate	- 1500 ml/min.
Counter-current wash water	- 400 ml/min.
Froth height	- 15 in.

TABLE A20: Operating Conditions R-C-RC Circuit Coal B, 325 mesh x 0 Particle Size Particle Size

Stage	DF1012, kg/ton	E, %	T, min
Rougher	1.32	17.43	5.47
Scavenger	0.37	15.17	4.57
Recleaner	0.64	10.60	4.99

* Circuit feedrate - 50 ml/min.

* E - Fractional air hold-up.

* T - Retention time as calculated from recovery equation.

	All Stages
* Air flowrate	- 1500 ml/min.
Counter-current wash water	- 400 ml/min.
Froth height	- 15 in.

TABLE A21: Operating Conditions R-C-RC Circuit Coal B, 325 mesh x 0 Particle Size Particle Size

Stage	DF1012, kg/ton	E, %	T, min
Rougher	1.12	14.76	4.59
Scavenger	0.91	13.30	5.02
Recleaner	0.25	10.43	4.94

* Circuit feedrate - 100 ml/min.

* E - Fractional air hold-up.

* T - Retention time as calculated from recovery equation.

	All Stages
* Air flowrate	- 1500 ml/min.
Counter-current wash water	- 400 ml/min.
Froth height	- 15 in.

TABLE A22: Operating Conditions R-C-RC Circuit Coal B, 325 mesh x 0 Particle Size Particle Size

Stage	DF1012, kg/ton	E, %	T, min
Rougher	1.03	13.73	4.34
Scavenger	0.21	10.08	4.34
Recleaner	0.04	11.06	4.68

* Circuit feedrate - 200 ml/min.

* E - Fractional air hold-up.

* T - Retention time as calculated from recovery equation.

	All Stages
* Air flowrate	- 1500 ml/min.
Counter-current wash water	- 400 ml/min.
Froth height	- 15 in.

TABLE A23: Operating Conditions R-C-RC Circuit Coal B, 325 mesh x 0 Particle Size Particle Size

Stage	DF1012, kg/ton	E, %	T, min
Rougher	1.03	12.23	4.07
Scavenger	0.40	9.08	3.97
Recleaner	0.34	10.07	4.73

* Circuit feedrate - 300 ml/min.

* E - Fractional air hold-up.

* T - Retention time as calculated from recovery equation.

	All Stages
* Air flowrate	- 1500 ml/min.
Counter-current wash water	- 400 ml/min.
Froth height	- 15 in.

TABLE A24: Operating Conditions for Simulation of other Circuits, Coal B, 325 mesh x 0 Particle Size, 5.0 % Feed Solids

Stage	DF1012, kg/ton	E, %	T, min
Rougher	1.32	17.43	5.47
Scavenger	0.37	15.17	4.57
Recleaner	0.64	10.60	4.99

* Circuit feedrate - 300 ml/min.

* E - Fractional air hold-up.

* CCW - Counter-current wash water.
  Air flowrate - 1500 ml/min.(all stages).
  Froth height - 15 in. (all stages).

Appendix B:

Simulator Source Code Listing

```
'FLOTATION SIMULATOR
'
'Varibles: CNAME$ -component name
'          COL$   -stage name
'          ASH    -component ash
'          K      -component flotation rate
'          PWT    -component percent wt.
'          DIA    -column diameter, cm.
'          LGTH   -column length, cm.
'          QW     -counter current wash, ml/min.
'          QG     -gas flowrate, ml/min.
'          E      -fractional gas holdup
'          PASH   -product ash
'          TASH   -tailings ash
'          CDEST  -concentrate destination
'          TDEST  -tailings destination
'          TMASS  -total mass rate
'          MASS   -component mass rate
'          PSOL   -stream percent solids
'          TSOL   -total combined percent solids
'
CLS
DIM CNAME$(10),COL$(10)
DIM ASH$(5),K$(5),PWT$(10,3,5)
DIM K(5),ASH(5),PWT(10,3,5),PER(10)
DIM DIA$(10),LGTH$(10),QW$(10),QG$(10),E$(10),FH$(10)
DIM DIA(10),LGTH(10),QW(10),QG(10),E(10),FH(10),L(10)
DIM PASH(10,3),TASH(10)
DIM CDEST$(10),TDEST$(10)
DIM CDEST(10),TDEST(10)
DIM MASS(10,3,5),TMASS(10,3),CMASS(3)
DIM RMASS(10),SMASS(10,3,5),QSEC(10),RSEC(10)
DIM R(10),MREC(10),CREC(10),FLOW(10)
DIM PSOL(10,3),TSOL(3),QSOL(10),RSOL(10)
'
'Initial Data Input/Page Editor Set-up
'
READ FR$,SDEN$,COMP$,NCOL$
COMP=VAL(COMP$)
NCOL=VAL(NCOL$)
FOR J=1 TO COMP
   READ CNAME$(J),PWT$(1,1,J),K$(J),ASH$(J)
NEXT J
FOR N=1 TO NCOL
   READ
COL$(N),DIA$(N),LGTH$(N),QW$(N),QG$(N),E$(N),PE$(N),FH$(N),CDEST$(N),TDE
ST$(N)
NEXT N
FR$=FR$+SPACE$(10-LEN(FR$))
SDEN$=SDEN$+SPACE$(10-LEN(SDEN$))
COMP$=COMP$+SPACE$(10-LEN(COMP$))
NCOL$=NCOL$+SPACE$(10-LEN(NCOL$))
5  COLOR 7,1
   CLS
   LOCATE 10,35:PRINT "COLUMN  FLOTATION "
   LOCATE 12,35:PRINT "CIRCUIT SIMULATOR "
   R%=12:CMIN%=52:CMAX%=53:FLD$=X$:GOSUB 50000:X$=FLD$:ON NXT
   GOTO 5,5,5,5,10,5,10
```

```
10 COLOR 7,1
   CLS
   LOCATE 6,25:PRINT "Circuit Feed Rate  ";:COLOR 1,7:PRINT FR$
   COLOR 7,1
   LOCATE 8,25:PRINT "     Feed %Solids  ";:COLOR 1,7:PRINT SDEN$
   COLOR 7,1
   LOCATE 10,25:PRINT "   No. Components  ";:COLOR 1,7:PRINT COMP$
   COLOR 7,1
   LOCATE 12,25:PRINT "      No. Columns  ";:COLOR 1,7:PRINT NCOL$
20 R%=6:CMIN%=44:CMAX%=48:FLD$=FR$:GOSUB    50000:FR$=FLD$ :ON NXT GOTO
25,20,25,20,40,5,10
25 R%=8:CMIN%=44:CMAX%=48:FLD$=SDEN$:GOSUB 50000:SDEN$=FLD$: ON NXT GOTO
30,20,30,20,40,5,10
30 R%=10:CMIN%=44:CMAX%=48:FLD$=COMP$:GOSUB 50000:COMP$=FLD$ :ON NXT GOTO
35,25,35,25,40,5,10
35 R%=12:CMIN%=44:CMAX%=48:FLD$=NCOL$:GOSUB 50000:NCOL$=FLD$ :ON NXT GOTO
35,30,35,30,40,5,10
40 COLOR 7,1
COMP=VAL(COMP$)
NCOL=VAL(NCOL$)
FOR J=1 TO COMP
   CNAME$(J)=CNAME$(J)+SPACE$(10-LEN(CNAME$(J)))
   PWT$(1,1,J)=PWT$(1,1,J)+SPACE$(10-LEN(PWT$(1,1,J)))
   K$(J)=K$(J)+SPACE$(10-LEN(K$(J)))
   ASH$(J)=ASH$(J)+SPACE$(10-LEN(ASH$(J)))
NEXT J
FOR N=1 TO NCOL
   COL$(N)=COL$(N)+SPACE$(5-LEN(COL$(N)))
   DIA$(N)=DIA$(N)+SPACE$(5-LEN(DIA$(N)))
   LGTH$(N)=LGTH$(N)+SPACE$(5-LEN(LGTH$(N)))
   QW$(N)=QW$(N)+SPACE$(5-LEN(QW$(N)))
   QG$(N)=QG$(N)+SPACE$(5-LEN(QG$(N)))
   E$(N)=E$(N)+SPACE$(7-LEN(E$(N)))
   PE$(N)=PE$(N)+SPACE$(5-LEN(PE$(N)))
   FH$(N)=FH$(N)+SPACE$(5-LEN(FH$(N)))
   CDEST$(N)=CDEST$(N)+SPACE$(5-LEN(CDEST$(N)))
   TDEST$(N)=TDEST$(N)+SPACE$(5-LEN(TDEST$(N)))
NEXT N
CLS
   LOCATE 8,1:PRINT "Component          Pwt.,%              K,min(-1)
      %Ash  "
   COLOR 1,7
   FOR J=1 TO COMP
      LOCATE J+9,1:PRINT CNAME$(J)
      LOCATE J+9,20:PRINT PWT$(1,1,J)
      LOCATE J+9,40:PRINT K$(J)
      LOCATE J+9,60:PRINT ASH$(J)
   NEXT J
      42 POSIT=0
       45    R%=10+POSIT:CMIN%=1:CMAX%=10:FLD$=CNAME$(POSIT+1):GOSUB
50000:CNAME$(POSIT+1)=FLD$ :ON NXT GOTO 70,72,57,107,110,10,10
       57   R%=10+POSIT:CMIN%=20:CMAX%=30:FLD$=PWT$(1,1,POSIT+1):GOSUB
50000:PWT$(1,1,POSIT+1)=FLD$ :ON NXT GOTO 85,87,65,45,110,10,10
       65     R%=10+POSIT:CMIN%=40:CMAX%=50:FLD$=K$(POSIT+1):GOSUB
50000:K$(POSIT+1)=FLD$: ON NXT GOTO 105,107,66,57,110,10,10
       66   R%=10+POSIT:CMIN%=60:CMAX%=70:FLD$=ASH$(POSIT+1):GOSUB
50000:ASH$(POSIT+1)=FLD$: ON NXT GOTO 108,109,70,65,110,10,10
       70 IF POSIT<COMP-1 THEN POSIT=POSIT+1
```

```
        GOTO 45
     72 IF POSIT>0 THEN POSIT=POSIT-1
        GOTO 45
     85 IF POSIT<COMP-1 THEN POSIT=POSIT+1
        GOTO 57
     87 IF POSIT>0 THEN POSIT=POSIT-1
        GOTO 57      105 IF POSIT<COMP-1 THEN POSIT=POSIT+1
        GOTO 65
    107 IF POSIT>0 THEN POSIT=POSIT-1
        GOTO 65
    108 IF POSIT<COMP-1 THEN POSIT=POSIT+1
        GOTO 66
    109 IF POSIT>0 THEN POSIT=POSIT-1
        GOTO 66
110 COLOR 7,1
    CLS
   LOCATE 5,1:PRINT "Stage  Diameter  Length  C.C.Wash  Gas Flow  Air
Fraction  Pectlet No. Froth Ht."
   COLOR 1,7
   FOR N=1 TO NCOL
      LOCATE N+6,1:PRINT COL$(N)
      LOCATE N+6,8:PRINT DIA$(N)
      LOCATE N+6,18:PRINT LGTH$(N)
      LOCATE N+6,26:PRINT QW$(N)
      LOCATE N+6,36:PRINT QG$(N)
      LOCATE N+6,46:PRINT E$(N)
      LOCATE N+6,60:PRINT PE$(N)
      LOCATE N+6,72:PRINT FH$(N)
   NEXT N
     113 POSIT=0
      115    R%=7+POSIT:CMIN%=1:CMAX%=6:FLD$=COL$(POSIT+1):GOSUB
50000:COL$(POSIT+1)=FLD$ :ON NXT GOTO 145,147,120,172,180,40,10
      120     R%=7+POSIT:CMIN%=8:CMAX%=13:FLD$=DIA$(POSIT+1):GOSUB
50000:DIA$(POSIT+1)=FLD$: ON NXT GOTO 150,152,125,115,180,40,10
      125   R%=7+POSIT:CMIN%=18:CMAX%=23:FLD$=LGTH$(POSIT+1):GOSUB
50000:LGTH$(POSIT+1)=FLD$: ON NXT GOTO 155,157,130,120,180,40,10
      130     R%=7+POSIT:CMIN%=26:CMAX%=31:FLD$=QW$(POSIT+1):GOSUB
50000:QW$(POSIT+1)=FLD$ :ON NXT GOTO 160,162,135,125,180,40,10
      135     R%=7+POSIT:CMIN%=36:CMAX%=41:FLD$=QG$(POSIT+1):GOSUB
50000:QG$(POSIT+1)=FLD$ :ON NXT GOTO 165,167,140,130,180,40,10
      140    R%=7+POSIT:CMIN%=46:CMAX%=51:FLD$=E$(POSIT+1):GOSUB
50000:E$(POSIT+1)=FLD$: ON NXT GOTO 170,172,142,135,180,40,10
      142     R%=7+POSIT:CMIN%=60:CMAX%=65:FLD$=PE$(POSIT+1):GOSUB
50000:PE$(POSIT+1)=FLD$: ON NXT GOTO 175,177,144,140,180,40,10
      144     R%=7+POSIT:CMIN%=72:CMAX%=77:FLD$=FH$(POSIT+1):GOSUB
50000:FH$(POSIT+1)=FLD$: ON NXT GOTO 178,179,145,142,180,40,10
     145 IF POSIT<NCOL-1 THEN POSIT=POSIT+1
        GOTO 115
     147 IF POSIT>0 THEN POSIT=POSIT-1
        GOTO 115
     150 IF POSIT<NCOL-1 THEN POSIT=POSIT+1
        GOTO 120
     152 IF POSIT>0 THEN POSIT=POSIT-1
        GOTO 120
     155 IF POSIT<NCOL-1 THEN POSIT=POSIT+1
        GOTO 125
     157 IF POSIT>0 THEN POSIT=POSIT-1
        GOTO 125
```

```
     160 IF POSIT<NCOL-1 THEN POSIT=POSIT+1
         GOTO 130
     162 IF POSIT>0 THEN POSIT=POSIT-1
         GOTO 130
     165 IF POSIT<NCOL-1 THEN POSIT=POSIT+1
         GOTO 135
     167 IF POSIT>0 THEN POSIT=POSIT-1
         GOTO 135
     170 IF POSIT<NCOL-1 THEN POSIT=POSIT+1
         GOTO 140
     172 IF POSIT>0 THEN POSIT=POSIT-1
         GOTO 140
     175 IF POSIT<NCOL-1 THEN POSIT=POSIT+1
         GOTO 142
     177 IF POSIT>0 THEN POSIT=POSIT-1
         GOTO 142
     178 IF POSIT<NCOL-1 THEN POSIT=POSIT+1
         GOTO 144
     179 IF POSIT>0 THEN POSIT=POSIT-1
         GOTO 144
180 COLOR 7,1
    CLS
    LOCATE 5,1:PRINT "Stage        Conc. Destination   Tails Destination
  "
    COLOR 1,7
    FOR N=1 TO NCOL
      LOCATE N+7,1:PRINT  COL$(N)
      LOCATE N+7,20:PRINT CDEST$(N)
      LOCATE N+7,40:PRINT TDEST$(N)
      LOCATE N+7,60:PRINT CSOL$(N)
    NEXT N
    COLOR 7,1
    LOCATE 19,5:PRINT "NOTE:"
    LOCATE 19,15:PRINT "Indicate destination by unit no."
    LOCATE 20,15:PRINT "Indicate exit streams by (0)"
    LOCATE 21,15:PRINT "Feedback streams by -(unit no.)"
    COLOR 1,7
    183 POSIT=0
      185 R%=8+POSIT:CMIN%=1:CMAX%=8:FLD$=NCOL$(POSIT+1):GOSUB
50000:NCOL$(POSIT+1)=FLD$ :ON NXT GOTO 200,202,190,212,250,110,10
      190   R%=8+POSIT:CMIN%=20:CMAX%=21:FLD$=CDEST$(POSIT+1):GOSUB
50000:CDEST$(POSIT+1)=FLD$ :ON NXT GOTO 205,207,195,185,250,110,10
      195   R%=8+POSIT:CMIN%=40:CMAX%=41:FLD$=TDEST$(POSIT+1):GOSUB
50000:TDEST$(POSIT+1)=FLD$: ON NXT GOTO 210,212,200,190,250,110,10
      200 IF POSIT<NCOL-1 THEN POSIT=POSIT+1
          GOTO 185
      202 IF POSIT>0 THEN POSIT=POSIT-1
          GOTO 185
      205 IF POSIT<NCOL THEN POSIT=POSIT+1
          GOTO 190
      207 IF POSIT>0 THEN POSIT=POSIT-1
          GOTO 190
      210 IF POSIT<NCOL-1 THEN POSIT=POSIT+1
          GOTO 195
      212 IF POSIT>0 THEN POSIT=POSIT-1
          GOTO 195
250 GOTO 275
'
```

```
'Convert Strings to numeric values
'
275 COMP=VAL(COMP$)
    FR=VAL(FR$)
    SDEN=VAL(SDEN$)/100
    NCOL=VAL(NCOL$)
    FOR J=1 TO COMP
       K(J)=VAL(K$(J))
       ASH(J)=VAL(ASH$(J))
       PWT(1,1,J)=VAL(PWT$(1,1,J))
    NEXT J
    FOR N=1 TO NCOL
       DIA(N)=VAL(DIA$(N))
       LGTH(N)=VAL(LGTH$(N))
       QW(N)=VAL(QW$(N))
       QG(N)=VAL(QG$(N))
       E(N)=VAL(E$(N))
       PE(N)=VAL(PE$(N))
       FH(N)=VAL(FH$(N))
       CDEST(N)=VAL(CDEST$(N))
       TDEST(N)=VAL(TDEST$(N))
    NEXT N
'
'Normalize counting varibles to 0
'
    CLS
    ICOUNT=0
    TCOUNT=0
    QCOUNT=0
    CMASS(1)=FR
    TASH(1)=0
'
'Set circuit feed characteristics '
'
    FOR J=1 TO COMP
       FLOW(J)=FR*(PWT(1,1,J)/100)
       PER(J)=PWT(1,1,J)
       TASH(1)=(FLOW(J)*ASH(J))+TASH(1)
    NEXT J
    TASH(1)=TASH(1)/FR
    TSOL(1)=SDEN
    FOR N=1 TO NCOL
       RMASS(N)=0
       QSOL(N)=0
       FOR J=1 TO COMP
          SPWT(N,1,J)=0
          SMASS(N,1,J)=0
       NEXT J
    NEXT N
    COLOR 7,1
    CLS
     LOCATE    10,10:PRINT    ".................SIMULATION    IN
PROGRESS................."
'
' Begin circuit loop
'
300  FOR L=1 TO 3
        FOR N=1 TO NCOL
```

```
            TMASS(N,L)=0
            PSOL(N,L)=0
            QSEC(N)=0
            RSEC(N)=0
            FOR J=1 TO COMP
               PWT(N,L,J)=0
               MASS(N,L,J)=0
            NEXT J
         NEXT N
      NEXT L
      FOR L=2 TO 3
         CMASS(L)=0
         TASH(L)=0
         TSOL(N)=0
      NEXT L
      TMASS(1,1)=CMASS(1)
      PSOL(1,1)=TSOL(1)
      SCOUNT=0
      FOR J=1 TO COMP
         PWT(1,1,J)= PER(J)
         MASS(1,1,J)=(PWT(1,1,J)/100)*TMASS(N,1)
      NEXT J
      IF QCOUNT>0 THEN
         FOR N=1 TO NCOL
            IF QMASS(N)=0 AND TMASS(N,1)=0 THEN
               GOTO 350
            END IF
PSOL(N,1)=(QMASS(N)*QSOL(N)+TMASS(N,1)*PSOL(N,1))/(QMASS(N)+TMASS(N,1))
            TMASS(N,1)=QMASS(N)+TMASS(N,1)
            FOR J=1 TO COMP
               MASS(N,1,J)=PMASS(N,1,J)+MASS(N,1,J)
            NEXT J
350      NEXT N
      END IF
      IF TCOUNT>0 THEN
         FOR N=1 TO NCOL
            IF RMASS(N)=0 AND TMASS(N,1)=0 THEN
               GOTO 375
            END IF
PSOL(N,1)=(RMASS(N)*RSOL(N)+TMASS(N,1)*PSOL(N,1))/(RMASS(N)+TMASS(N,1))
            TMASS(N,1)=RMASS(N)+TMASS(N,1)
            FOR J=1 TO COMP
               MASS(N,1,J)=SMASS(N,1,J)+MASS(N,1,J)
            NEXT J
375      NEXT N
       END IF
'
'Begin column loop
'
      FOR N=1 TO NCOL
'
'Calculate column recovery
'
      GOSUB 700
'
'Route column output and combine streams
'
      IF CDEST(N)<0 THEN
```

```
        IF ABS(CDEST(N))>1 THEN
          IF QSEC(ABS(CDEST(N)))>0 THEN
QSOL(ABS(CDEST(N)))=(TMASS(N,2)*PSOL(N,2)+QMASS(ABS(CDEST(N)))*QSOL(ABS(
CDEST(N))))/(TMASS(N,2)+QMASS(ABS(CDEST(N))))
            QMASS(ABS(CDEST(N)))=TMASS(N,2)+QMASS(ABS(CDEST(N)))
            FOR J=1 TO COMP
            PMASS(ABS(CDEST(N)),1,J)=MASS(N,2,J)+PMASS(ABS(CDEST(N)),1,J)
            NEXT J
          ELSE
            QMASS(ABS(CDEST(N)))=TMASS(N,2)
            QSOL(ABS(CDEST(N)))=PSOL(N,2)
            FOR J=1 TO COMP
            PMASS(ABS(CDEST(N)),1,J)=MASS(N,2,J)
            NEXT J
           END IF
           QCOUNT=QCOUNT+1
           QSEC(ABS(CDEST(N)))=QSEC(ABS(CDEST(N)))+1
        END IF
        TMASS(ABS(CDEST(N)),1)=TMASS(N,2)+FR
PSOL(ABS(CDEST(N)),1)=(TMASS(N,2)*PSOL(N,2)+FR*SDEN)/TMASS(ABS(CDEST(N))
,1)
        IF SCOUNT>0 THEN
          GOTO 400
        END IF
        FOR J=1 TO COMP
        MASS(ABS(CDEST(N)),1,J)=MASS(N,2,J)+FLOW(J)
PWT(ABS(CDEST(N)),1,J)=(MASS(ABS(CDEST(N)),1,J)/TMASS(ABS(CDEST(N)),1))*
100
        NEXT J
400    SCOUNT=SCOUNT+1
            I F     S C O U N T > 1     T H E N
TMASS(ABS(CDEST(N)),1)=TMASS(N,2)+CMASS(1)
PSOL(ABS(CDEST(N)),1)=(TMASS(N,2)*PSOL(N,2)+CMASS(1)*TSOL(1))/TMASS(ABS(
CDEST(N)),1)
          FOR J=1 TO COMP
             MASS(ABS(CDEST(N)),1,J)=MASS(N,2,J)+MASS(ABS(CDEST(N)),1,J)
PWT(ABS(CDEST(N)),1,J)=(MASS(ABS(CDEST(N)),1,J)/TMASS(ABS(CDEST(N)),1))*
100
         NEXT J
         END IF
         CMASS(1)=TMASS(ABS(CDEST(N)),1)
         TSOL(1)=PSOL(ABS(CDEST(N)),1)
         FOR J=1 TO COMP
            MASS(ABS(CDEST(N)),1,J)=(PWT(ABS(CDEST(N)),1,J)/100)*CMASS(1)
         NEXT J
      END IF
         IF CDEST(N)=0 THEN
TASH(2)=(TMASS(N,2)*PASH(N,2)+CMASS(2)*TASH(2))/(TMASS(N,2)+CMASS(2))
TSOL(2)=(TMASS(N,2)*PSOL(N,2)+CMASS(2)*TSOL(2))/(TMASS(N,2)+CMASS(2))
            CMASS(2)=TMASS(N,2)+CMASS(2)
         END IF
         IF CDEST(N)>0 THEN
PSOL(CDEST(N),1)=(TMASS(N,2)*PSOL(N,2)+TMASS(CDEST(N),1)*PSOL(CDEST(N),1
))/(TMASS(N,2)+TMASS(CDEST(N),1))
            TMASS(CDEST(N),1)=TMASS(N,2)+TMASS(CDEST(N),1)
            FOR J=1 TO COMP
PWT(CDEST(N),1,J)=((MASS(N,2,J)+MASS(CDEST(N),1,J))/TMASS(CDEST(N),1))*100
              MASS(CDEST(N),1,J)=(PWT(CDEST(N),1,J)/100)*TMASS(CDEST(N),1)
```

```
          NEXT J
        END IF
        IF TDEST(N)<0 THEN
          IF ABS(TDEST(N))>1 THEN
            IF RSEC(ABS(TDEST(N)))>0 THEN
RSOL(ABS(TDEST(N)))=(TMASS(N,3)*PSOL(N,3)+RMASS(ABS(TDEST(N)))*RSOL(ABS(
TDEST(N))))/(TMASS(N,3)+RMASS(ABS(TDEST(N))))
              RMASS(ABS(TDEST(N)))=TMASS(N,3)+RMASS(ABS(TDEST(N)))
              FOR J=1 TO COMP
SMASS(ABS(TDEST(N)),1,J)=MASS(N,1,J)+SMASS(ABS(TDEST(N)),1,J)
              NEXT J
            ELSE
              RMASS(ABS(TDEST(N)))=TMASS(N,3)
              RSOL(ABS(TDEST(N)))=PSOL(N,3)
              FOR J=1 TO COMP
                 SMASS(ABS(TDEST(N)),1,J)=MASS(N,3,J)
              NEXT J
            END IF
            TCOUNT=TCOUNT+1
            GOTO 475
          END IF
          TMASS(ABS(TDEST(N)),1)=TMASS(N,3)+FR
PSOL(ABS(TDEST(N)),1)=(TMASS(N,3)*PSOL(N,3)+FR*SDEN)/TMASS(ABS(TDEST(N))
,1)
          IF SCOUNT>0 THEN
             GOTO 450
          END IF
          FOR J=1 TO COMP
             MASS(ABS(TDEST(N)),1,J)=MASS(N,3,J)+FLOW(J)
PWT(ABS(TDEST(N)),1,J)=(MASS(ABS(TDEST(N)),1,J)/TMASS(ABS(TDEST(N)),1))*
100
          NEXT J
450       SCOUNT=SCOUNT+1
          IF SCOUNT>1 THEN
            TMASS(ABS(TDEST(N)),1)=TMASS(N,3)+CMASS(1)
PSOL(ABS(TDEST(N)),1)=(TMASS(N,3)*PSOL(N,3)+CMASS(1)*TSOL(1))/TMASS(ABS(
TDEST(N)),1)
            FOR J=1 TO COMP
               MASS(ABS(TDEST(N)),1,J)=MASS(N,3,J)+MASS(ABS(TDEST(N)),1,J)
PWT(ABS(TDEST(N)),1,J)=(MASS(ABS(TDEST(N)),1,J)/TMASS(ABS(TDEST(N)),1))*
100
            NEXT J
          END IF
          CMASS(1)=TMASS(ABS(TDEST(N)),1)
          TSOL(1)=PSOL(ABS(TDEST(N)),1)
          FOR J=1 TO COMP
             MASS(ABS(TDEST(N)),1,J)=(PWT(ABS(TDEST(N)),1,J)/100)*CMASS(1)
          NEXT J
        END IF
        IF TDEST(N)=0 THEN
TASH(3)=(TMASS(N,3)*PASH(N,3)+CMASS(3)*TASH(3))/(TMASS(N,3)+CMASS(3))
TSOL(3)=(TMASS(N,3)*PSOL(N,3)+CMASS(3)*TSOL(3))/(TMASS(N,3)+CMASS(3))
           CMASS(3)=TMASS(N,3)+CMASS(3)
        END IF
        IF    TDEST(N)>0    THEN
PSOL(TDEST(N),1)=(TMASS(N,3)*PSOL(N,3)+TMASS(TDEST(N),1)*PSOL(TDEST(N),1
))/(TMASS(N,3)+TMASS(TDEST(N),1))
          TMASS(TDEST(N),1)=TMASS(N,3)+TMASS(TDEST(N),1)
```

```
            FOR J=1 TO COMP
PWT(TDEST(N),1,J)=((MASS(N,3,J)+MASS(TDEST(N),1,J))/TMASS(TDEST(N),1))*100
               MASS(TDEST(N),1,J)=(PWT(TDEST(N),1,J)/100)*TMASS(TDEST(N),1)
            NEXT J
         END IF
475      NEXT N
   CMASS(1)=TMASS(1,1)
   TSOL(1)=PSOL(1,1)
   FOR J=1 TO COMP
      PER(J)=PWT(1,1,J)
   NEXT J
'
'Check for circuit convergence
'
   ICOUNT=ICOUNT+1
   IF ICOUNT>250 THEN
       PRINT "250 ITERATIONS EXCEEDED!"
      END
     ELSE
      IF ABS((FR-CMASS(2)-CMASS(3))/FR)>.001 THEN
         GOTO 300
       END IF
   END IF
 '
 'Calculate combustible recovery for each column
 '
   FOR N=1 TO NCOL
       CREC(N)=((100-PASH(N,2))/(100-PASH(N,1)))*MREC(N)
      FOR L=2 TO 3
       PSOL(N,L)=PSOL(N,L)*100
      NEXT L
    NEXT N
'
'Calculate circuit recovery
'
    TMREC=(CMASS(2)/FR)*100
    TCREC=((100-TASH(2))/(100-TASH(1)))*TMREC
    FOR L=2 TO 3
       TSOL(I)=TSOL(L)*100
    NEXT L
'
'Print results
'
500   CLS
    LOCATE 5,25:PRINT "      TOTAL CIRCUIT RESULTS"
    LOCATE 6,25:PRINT "                  Yield=":LOCATE 6,46:PRINT USING
"###.##";TMREC
    LOCATE 7,25:PRINT "               Recovery=":LOCATE 7,46:PRINT USING
"###.##";TCREC
    LOCATE 8,25:PRINT "   Concentrate %Grade=":LOCATE 8,46:PRINT USING
"###.##";TASH(2)
    LOCATE 9,25:PRINT "      Tailings %Grade=":LOCATE 9,46:PRINT USING
"###.##";TASH(3)
    LOCATE 10,25:PRINT "   Concentrate %Solid=":LOCATE 10,46:PRINT USING
"###.##";TSOL(2)
    LOCATE 11,25:PRINT "      Tailings %Solid=":LOCATE 11,46:PRINT USING
"###.##";TSOL(3)
    LOCATE 23,10,0
```

```
    PRINT "N - NEXT PAGE   P - PREVIOUS PAGE   R - RUN AGAIN   Q - QUIT"
525 A$=INKEY$: IF (A$="N") OR (A$="n") THEN 530
               IF (A$="Q") OR (A$="q") THEN 610
               IF (A$="P") OR (A$="p") THEN 500              IF (A$="R")
OR (A$="r") THEN 10
               IF A$="" THEN 525
530 FOR N=1 TO NCOL
    CLS
    LOCATE 5,40:PRINT COL$(N)" RESULTS"
    LOCATE 6,25:PRINT "                Yield=":LOCATE 6,46:PRINT USING
"###.##";MREC(N)
    LOCATE 7,25:PRINT "             Recovery=":LOCATE 7,46:PRINT USING
"###.##";CREC(N)
    LOCATE 8,25:PRINT "  Concentrate %Grade=":LOCATE 8,46:PRINT USING
"###.##";PASH(N,2)
    LOCATE 9,25:PRINT "      Tailings %Grade=":LOCATE 9,46:PRINT USING
"###.##";PASH(N,3)
    LOCATE 10,25:PRINT "  Concentrate %Solid=":LOCATE 10,46:PRINT USING
"###.##";PSOL(N,2)
    LOCATE 11,25:PRINT "     Tailings %Solid=":LOCATE 11,46:PRINT USING
"###.##";PSOL(N,3)
    LOCATE 15,25:PRINT "      Retention Time=":LOCATE 15,46:PRINT USING
"###.##";RET(N)
    LOCATE 16,25:PRINT "      Tailings Flow =":LOCATE 16,46:PRINT USING
"####.##";TFLOW(N)
LOCATE 23,10,0
    PRINT "N - NEXT PAGE   P - PREVIOUS PAGE   R - RUN AGAIN   Q - QUIT"
535 A$=INKEY$: IF (A$="N") OR (A$="n") THEN 540
               IF (A$="Q") OR (A$="q") THEN 610
               IF (A$="P") OR (A$="p") THEN 542
               IF (A$="R") OR (A$="r") THEN 10
               IF A$="" THEN 535
540 IF N<NCOL THEN
       GOTO 545
       ELSE
        GOTO 610
    END IF
542 N=N-2
    IF N<0 THEN
    GOTO 500
    END IF
545 NEXT N
610 END
'
'Subroutine for recovery calculation
'
700 IF PSOL(N,1)>SDEN THEN
    PSOL(N,1)=SDEN
    END IF
    QF=TMASS(N,1)/PSOL(N,1)
    QT=QF+QW(N)
    L(N)=LGTH(N)-FH(N)
750 T=((3.14*DIA(N)^2)*L(N)*(1-E(N)))/(4*QT)
    RET(N)=T
    FOR J=1 TO COMP
        MASS(N,1,J)=TMASS(N,1)*(PWT(N,1,J)/100)
        A=(1+(4*K(J)*T/PE(N)))^0.5
        B=A*EXP(PE(N)/2)
```

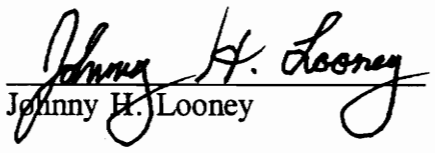
```
        C=(1+A)^2*EXP(PE(N)*A/2)
        D=(1-A)^2*EXP(-PE(N)*A/2)
        R(J)=1-(4*B/(C-D))
        MASS(N,2,J)=R(J)*MASS(N,1,J)
        MASS(N,3,J)=MASS(N,1,J)-MASS(N,2,J)
    NEXT J
    FOR J=1 TO COMP
        TMASS(N,2)=MASS(N,2,J)+TMASS(N,2)
        PWT(N,2,J)=(MASS(N,2,J)/TMASS(N,2))*100
    NEXT J
    TMASS(N,3)=TMASS(N,1)-TMASS(N,2)
    FOR J=1 TO COMP
       PWT(N,3,J)=(MASS(N,3,J)/TMASS(N,3))*100
    NEXT J
    PSOL(N,2)=PSOL(N,1)
    IF PSOL(N,2)<0.02 THEN
       PSOL(N,2)=0.02
    END IF
    QC=0.2*QW(N)
    QS=(QF+QW(N))-QC
'
'Check for flow convergence
'
  IF ABS((QS-QT)/QT)>0.0001 THEN
      QT=QS
      TMASS(N,2)=0
      GOTO 750
    END IF
    TFLOW(N)=QT
    PSOL(N,2)=TMASS(N,2)/QC
    PSOL(N,3)=TMASS(N,3)/QT
    775 FOR L=1 TO 3
          SUM(L)=0
        NEXT L
'
'  Calculate product & tailings grade
'
  FOR L=1 TO 3
      FOR J=1 TO COMP
         SUM(L)=ASH(J)*MASS(N,L,J)+SUM(L)
      NEXT J
      PASH(N,L)=SUM(L)/TMASS(N,L)
    NEXT L
    MREC(N)=(TMASS(N,2)/TMASS(N,1))*100
    RETURN
'
5 0 0 0 0   C % = C M I N % : L O C A T E   R % , C % , 1
'EDITOR---------------------------------------
50010 DEF SEG=0:POKE 1050,PEEK(1052)
50020 I$=INKEY$:IF I$="" THEN 50020
50030 IF LEN(I$)=2 THEN I$=RIGHT$(I$,1):I=ASC(I$):GOTO 50220
50040 I=ASC(I$):IF I=9 OR I=13 THEN GOSUB 50430:NXT=3:RETURN 'Tab/Return
50050 IF I<>8 THEN 50080 'Backspace
50060    IF    C%>CMIN%    THEN    B$=MID$(FLD$,C%-CMIN%+1,CMAX%-C%+1)+"
":FLD$=MID$(FLD$,1,C%-CMIN%-1)+B$:C%=C%-1:LOCATE ,C%:PRINT B$;:LOCATE ,C%
ELSE FLD$=MID$(FLD$,2,CMAX%-C%+1)+" ":PRINT FLD$;:LOCATE ,CMIN%
50070 GOTO 50020
50080 IF I=27 THEN GOSUB 50430:FLD$=SPACE$(CMAX%-CMIN%+1):LOCATE
```

```
,CMIN%:PRINT FLD$;:LOCATE ,CMIN%:C%=CMIN%'Esc 50090 IF I<32 THEN 50020
'Unprintable character
50100 IF INS THEN 50140   'Branch for insert mode
5 0 1 1 0    ' - - - - N o n    i n s e r t    m o d e    d a t a
entry-------------------------------------
50120 PRINT I$:C%=C%+1:IF C%<CMAX%+1 THEN LOCATE R%,C%:CI%=C%-1 ELSE
LOCATE R%,C%-1:CI%=C%-1:C%=C%-1
50130 MID$(FLD$,CI%+1-CMIN%,1)=I$:GOTO 50020
5 0 1 4 0     ' - - - - I n s e r t     m o d e     d a t a
entry-----------------------------------------
50150 IF ASC(RIGHT$(FLD$,1))<>32 THEN BEEP:GOTO 50020
50160 B$=MID$(FLD$,C%-CMIN%+1,CMAX%-C%)
50170 FLD$=LEFT$(FLD$,C%-CMIN%)+I$+B$
50180 PRINT I$;:C%=C%+1:LOCATE ,C%:PRINT B$;:IF C%>CMAX% THEN C%=C%-1
50190 LOCATE ,C%:GOTO 50020
5 0 2 0 0    ' - - - - P r o c e s s    s p e c i a l    f u n c t i o n
key-----------------------------------
50210 I$=RIGHT$(I$,1)
50220 IF I=77 THEN 50230 ELSE 50240          'CR
50230 IF C%<CMAX% THEN C%=C%+1:LOCATE ,C%:GOTO 50020 ELSE BEEP:GOTO 50020
50240 IF I=75 THEN 50250 ELSE 50260          'CL
50250 IF C%>CMIN% THEN C%=C%-1:LOCATE ,C%:GOTO 50020 ELSE BEEP:GOTO 50020
50260 IF I=72 THEN GOSUB 50430:NXT=2:RETURN 'CU
50270 IF I=15 THEN GOSUB 50430:NXT=4:RETURN 'Shift+Tab
50280 IF I=80 THEN GOSUB 50430:NXT=1:RETURN 'CD
50290 IF I=71 THEN GOSUB 50430:NXT=7:RETURN 'Home
50300 IF I<69 AND I>58 THEN GOSUB 50430:NXT=8:RETURN 'F-keys
50310 IF I=73 THEN GOSUB 50430:NXT=6:RETURN 'PgUp
50320 IF I=81 THEN GOSUB 50430:NXT=5:RETURN 'PgDn
50330 IF I=115 THEN C%=CMIN%:LOCATE ,C%:GOTO 50020 'Ctrl+CL
50340 IF I=116 THEN GOSUB 50430:LF=CMAX%-CMIN%+1 ELSE 50380 'Ctrl+CR
50350 FOR L=0 TO LF-1:B$=MID$(FLD$,LF-L,1):IF ASC(B$)<>32 THEN 50360 ELSE
NEXT L
50360 C%=CMAX%-L+1:IF C%>CMAX% THEN LOCATE ,CMAX% ELSE LOCATE ,C%
50370 GOTO 50020
50380 IF I=82 THEN 50390 ELSE 50410    'Insert
50390 IF INS THEN INS=0:LOCATE ,,,6,7 ELSE INS=1:LOCATE ,,,4,7
50400 GOTO 50020
50410    IF    I=83    THEN    B$=MID$(FLD$,C%-CMIN%+2,CMAX%-C%)+"
":FLD$=MID$(FLD$,1,C%-CMIN%)+B$:PRINT B$;:LOCATE ,C%:GOTO 50020 'Delete
50420 GOTO 50020                             'Undefined key
50430 IF INS THEN INS=0:LOCATE ,,,6,7
5     0     4     4     0                R     E     T     U     R     N
'-------------------------------------------------------------
'
DATA  15.00,5.0,4,7
DATA "Coal",70.00,4.50,2.50
DATA "Middling",10.00,0.75,7.50
DATA "Reject",14.50,0.07,10.00
D A T A    " N o n f l o a t " , 5 . 5 0 , 0 . 0 0 , 1 0 0 . 0 0    D A T A
"RGHR",5.08,152.4,385,1500,0.1900,4,38.1, 2, 0
DATA "CLNR",5.08,152.4,430,1500,0.1200,4,38.1, 3, 0
DATA "SCVGR",5.08,152.4,390,1500,0.1225,4,38.1, 0, 0
DATA "SCLR",5.08,152.4,278,1500,0.1275,4,38.1,0,0
DATA "SCLR2",5.08,152.4,400,1500,0.1200,4,38.1,0,0
DATA "SCLR3",5.08,152.4,400,1500,0.1200,4,38.1,0,0
DATA "RECLR",5.08,152.4,400,1500,0.1200,4,38.1,0,0
```

VITA

Johnny H. Looney was born on August 21, 1959, in Richlands, Virginia, and graduated from Grundy Senior High School in 1977. He then attended Virginia Polytechnic Institute and State University and graduated with a B.S. in Mining and Minerals Engineering in 1981. He spent eight years working for United Coal Company in Bristol, Virginia, after completing his undergraduate studies. In 1989, he returned to Virginia Polytechnic Institute and State University to work on a M.S. degree in Mining and Minerals Engineering in the speciality area of mineral and coal processing.

After completion of his graduate work in 1991, Mr. Looney accepted employment with Ceramic Technologies, Inc., in Richlands, Virginia. The following year, he accepted an offer to join Union Zinc, Inc., at their eastern Tennessee mining operations, where he is currently employed as a Senior Metallurgical Engineer.



Johnny H. Looney