# PHASE I – TECHNICAL PROGRESS REPORT

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In-Plant Testing of High-Efficiency Hydraulic Separators

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# ABSTRACT

The mineral processing industry has commonly utilized hydraulic separators throughout history for classification and gravity concentration of various minerals. More commonly referred to as hindered-bed or fluidized-bed separators, these units make use of differential particle settling rates to segregate particles according to shape, size, and/or density. As with any equipment, there are inefficiencies associated with its operation, which prompted an industry driven research program to further evaluate two novel high-efficiency hindered bed separators. These units, which are commercially called the CrossFlow separator and HydroFloat separator, have the potential to improve performance (separation efficiency and throughput) and reduce operating costs (power consumption, water and reagent usage).

This report describes the results of Phase I activities (laboratory and pilot-scale tests) conducted with the CrossFlow and HydroFloat separators at several locations in the minerals and coal industries. Details of the testing programs (equipment setup, shakedown testing and detailed testing) associated with four coal plants and two phosphate plants are summarized in this work. In most of these applications, the high-efficiency units proved to provide a higher quality product at reduced costs when compared against the performance of conventional separators.

Based on promising results obtained from Phase I, full-scale prototypes will be purchased by several mining companies for use in Phase II of this project. Two of the prototype units, which will be constructed by Eriez Manufacturing, are expected to be installed by a major U.S. phosphate producer and a large eastern U.S. coal company. Negotiations are also underway to purchase and install additional prototype units by a mineral sands producer and a second phosphate producer. The data obtained from the full-scale evaluations will be used to further promote commercialization and industrial applications of these innovative technologies.

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# IN-PLANT TESTING OF CROSSFLOW IN THE COAL INDUSTRY

## 1.1 Introduction

### 1.1.1 General

The mineral processing industry has commonly utilized hydraulic separators throughout history for classification and gravity concentration of various minerals. More commonly referred to as hindered-bed or fluidized-bed separators, these units make use of differential particle settling rates to segregate particles according to shape, size, and/or density.

Conventional hindered-bed separators are inherently inefficient due to wide variations in the solids content and size distribution of the feed, which have an adverse effect of plant performance and operating costs. The traditional design consists of an open top vessel into which elutriation water is introduced through a series of distribution pipes evenly spaced across the base of the device. During operation, feed solids are injected into the upper section of the separator and are permitted to settle. The upward flow of elutriation water creates a fluidized bed of suspended particles within the separator that is automatically controlled through the use of a simple PID control loop. The control loop includes a pressure sensor mounted on the side of the separator to measure the relative bed pressure. To maintain a constant bed pressure, a single loop PID controller and a pneumatic pinch valve to control the underflow discharge are used.

The small interstices within the bed create high interstitial velocities that resist the penetration of the slow settling particles. As a result, small particles accumulate in the upper section of the separator and are eventually carried over the top of the device into a collection launder. Large particles, which settle at a rate faster than the upward current of rising water, eventually pass through the fluidized bed and are discharged out one or more restricted ports through the bottom of the separator.

As with any processing equipment, there are inherent inefficiencies associated with this design. The key operating variables that were identified as problematic with traditional hydraulic separators included: (i) turbulent feed distribution which can result in unwanted misplaced particles, (ii) limited throughput capacity due to the detrimental impact of feed water on separator performance, (iii) introduction of dead zones within the fluidization chamber caused by frequent blockage/plugging of the lateral pipes located in the base of the separation zone containing the elutriation water and (iv) maintenance of the blocked elutriation water pipes. To overcome these problems, an industry driven research program was initiated to develop a new family of innovative high-efficiency hydraulic separators that can be readily implemented in the commercial sector, called the CrossFlow Separator and HydroFloat Separator.

### 1.1.2 Advantages of the CrossFlow Separator

Figure 1.1 is a schematic drawing comparing a traditional hydraulic separator with the new CrossFlow separator. Existing hydraulic separators utilize a feed injection system which discharges through a downcomer approximately one-third of the way into the main separation chamber. The pipe discharge is usually equipped with a dispersion plate to laterally deflect the feed slurry, but this approach creates turbulence within the separator that is detrimental to both the quiescent flow of the unit and the overall separation process. The additional water added to the system at the injection point causes a secondary interface of fluidized solids to form within the separator. The CrossFlow separator minimizes this discontinuity by introducing the feed

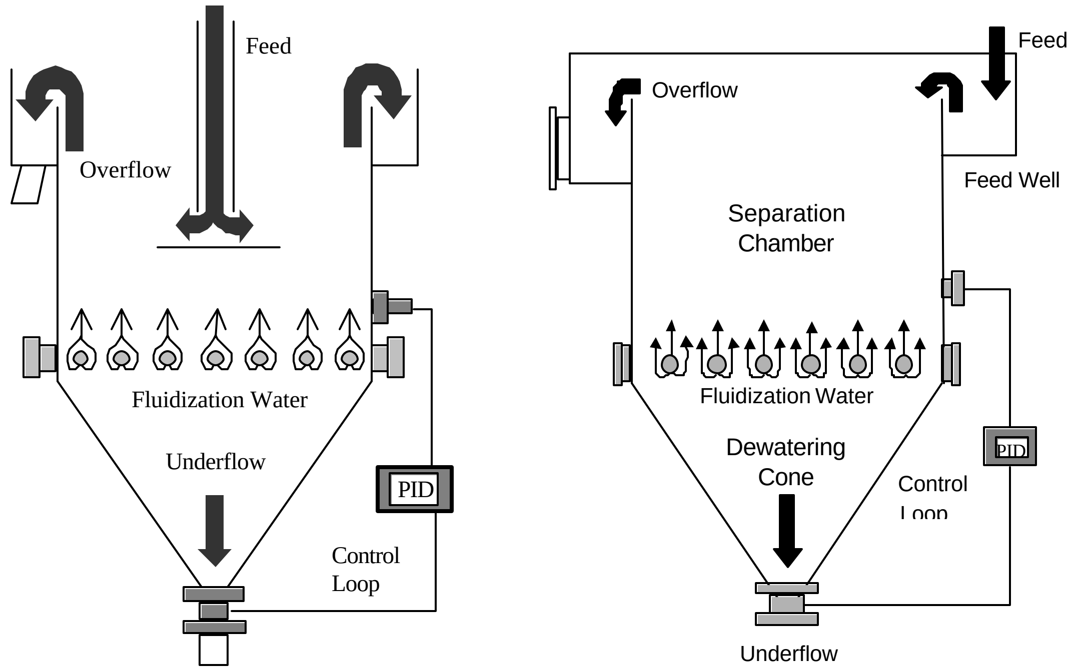



**Figure 1.1 Traditional Hydraulic Separator (Left) Versus Crossflow Separator (Right).**

stream across the top of the separator. A transition box and a baffle plate are used to reduce the feed velocity and optimize the tangential feed introduction into the top of the separator.

Another problem with the traditional design is that the water introduced with the feed solids must also report to the overflow launder. As a result, the rise velocity of the water is substantially increased at the feed injection point. The throughput capacity of existing hydraulic separators is limited by this introduction of water through the feed distribution pipe in the separation chamber and the excessive elutriation water added to the system. As previously mentioned, part of this problem was alleviated through the tangential feed distribution designed for the CrossFlow separator. A redesign of the elutriation water distribution, through use of a slotted plate at the base of the separation chamber, has minimized the amount of water used by allowing the water to better disperse through the separator. Larger diameter holes spread farther apart (6 inches versus 0.5 inches) allows for the water to be introduced into the chamber, and the baffle plate disperses the water throughout the chamber. This ultimately reduces the amount of overall elutriation water required, and increases the throughput capacity of the separator.

The improved distribution of elutriation water also minimizes dead zones within the separation chamber that were often caused by plugging of the small diameter holes in the lateral pipes at the bottom of the separation chamber. By increasing the diameter of the holes and adding the baffle plate to fully distribute the water, separation efficiency has increased due to full utilization of the separation chamber. The increase in separation efficiency and throughput capacity reduces the operating demands in terms of power, water and maintenance when reported on a per ton of concentrate basis when compared to traditional hydraulic separators.

### 1.1.3 Inefficiencies of the CrossFlow Separator

While the CrossFlow separator is a significant improvement over conventional hydraulic separators, the unit does have a few limitations. One of the significant limitations is that the unit requires a narrow particle size distribution for effective separation. Previous testing has proven that efficient concentration can only be achieved if the particles are in the size range of 200 mesh to several millimeters. The particle size ratio typically needs to be less than about 6:1 (top size to bottom size).

The other limitation of the CrossFlow separator is it requires a moderately large difference in particle densities. The separator often accumulates low density coarse particles at the top of the teeter bed, which are too light to penetrate the bed, but at the same time, too heavy to be carried by the rising water into the overflow. As a result, misplacement of low-density, coarse particles to the high-density underflow can occur. This inefficiency can be partially corrected by increasing the elutriation water, to try to carry the low density coarse particles into the overflow; however this can sometimes cause the fine, high-density particles to also report to the overflow instead of penetrating the teeter bed.

The shortcomings of the CrossFlow separator were recognized by the design team and have been overcome with the design of the HydroFloat separator, which will be discussed in detail in Chapter 2.

### 1.1.5 Project Justification

While improvements in technology have assisted the U.S. mining industry in reducing its overall energy consumption, the industry still struggles to be as efficient as possible due to the current economic climate. It is difficult for mining companies to justify huge capital investment in energy efficient technology. However, the incentive still exists to development equipment that will not only reduce costs, but improve efficiencies as well. This is due to the fact that each ton of saleable ore or coal that is recovered through an improvement in plant efficiency adds the full market price of that ton of material to the company revenue. Otherwise, the full market value is lost to waste. For a typical coal preparation plant, a one percentage point improvement in plant efficiency is roughly equivalent to a 20 percent improvement in profitability for the overall mine. As a result, the adoption of new technologies that improve efficiency is very attractive for industry representatives.

The implementation of the CrossFlow hydraulic separator will significantly reduce energy consumption and improve efficiency in the coal industry. When compared to conventional technology, the CrossFlow separator processing more material (as high as 40% solids) and operates at lower pressures (atmospheric versus 20 psig) for sizing the fine coal streams. These differences reduce the pumping requirements and minimize wear. For a typical unit, the overall savings is estimated to be 5.8 BTU per year per unit based on 3.5 million tons per year of raw coal feed to a typical preparation plant. In addition to reduction in pumping costs, the reduction in water consumption and reagent dosage associated with the higher percent solids will continue to reduce costs when compared to conventional units. Overall maintenance costs per ton of product will also be reduced.

The improved efficiency of the CrossFlow unit yields a sharper cut point, which ultimately produces additional clean coal for the same amount of raw coal processed by minimizing (i) the amount of coarse low density coal that is lost to fines and (ii) the amount of high density slimes that report to the clean coal product. As a result, coal reserves will be better utilized, productivity will be increased, and waste requirements will be reduced. These factors will allow operations to be more profitable and more competitive in domestic and international markets.

The technology is also expected to have a significant impact on the heavy mineral sands industry. The mineral sands industry currently suffers from the use of low-efficiency operations that require many stages of recleaning to achieve the required market grade. The process is considered to be very energy intensive with high operating costs. Fortunately, through the development of the CrossFlow separator, it is projected the industry can improve metallurgical efficiency tremendously during the pre-concentration step, which in turn would substantially lower the tonnage of ore that must be reprocessed in subsequent polishing stages. This would ultimately make the process more profitable by increasing performance and reducing operating costs (i.e., electrical power, diesel fuel, process water, etc.).

## 1.2 Literature Review

### 1.2.1 Hydraulic Classifiers

There are three main characteristics that distinguish a hydraulic classifier from other classifiers. First, discharge of the oversize material from the device depends upon its gravitational flow properties and not mechanical means such as a screw or rake. Coarse particles settle at a rate faster than the upward current of the elutriation water, and exit the unit through a valve or spigot at the base of the unit. The second distinctive characteristic of a hydraulic classifier is the unit is not fed under pressure; the primary source of classification is based on differential particle settling rates to segregate particles according to shape, size, and/or density. And finally, hydraulic classifiers utilize at least one, and sometimes both, of the following two mechanisms (NC State, 1992):

(i) *Hindered Settling* - An oversized particle settles against upward flowing fluid; the greater the density of the fluid, the larger the particle that will remain suspended (or teetered) in the fluid. Hindered settling is a function of particle size, density and concentration, liquid density and viscosity as well as the charge density.

(ii) *Elutriation* - An undersize particle is lifted by an upward flowing stream of water; the greater the upward velocity, the larger the particle that will be lifted.

Hydraulic classifiers are frequently used in the minerals processing industry to classify fine particles according to size. When the feed size distribution is within acceptable limits, these units can also be used for the concentration of particles based on differences in density. Over the years, various units have been developed and can be primarily categorized by the method in which the coarse material is discharged from the separation zone of the unit (Heiskanen, 1993). The two main operational categories are: (i) classifiers that operate with free and/or hindered settling that have virtually no control of the underflow (or coarse fraction) discharge and (ii) classifiers that do attempt to control the underflow discharge causing the formation of a teeter bed. Classifiers that do not attempt to control the underflow discharge can be further subdivided into mechanical and non-mechanical categories.

#### 1.2.1.1 Mechanical Hydraulic Classifiers

The *Hukki Cone Classifier* is a mechanical classifier invented by R.T. Hukki in 1967 and consists of a cylindrical tank where feed is introduced into the tank on a slowly rotating distribution disk, which causes a slight centrifugal action to it. The bottom of the tank is conical shape where water sprays are used as elutriation water. Coarse material is discharged through a pinch valve in the bottom of the cone. The key to this unit is in the conical section; where a ring of vertical, radial vanes are located to allow the pulp to rise upwards in a laminar fashion. The unit was originally designed to treat low quality sands, but is not used in practice today.

The *Sogreah Lavodune Classifier* is another mechanical classifier that consists of a cylindrical tank and a cone. Lower density counter-current classification is enhanced by laminar flow in this unit. A downcomer introduces feed material into the unit approximately one third of the distance from the top of the unit. The volume of the unit is restricted in the cone section where classification takes place in high suspension densities. The fine material rises and is discharged over the overflow lip of the unit. A plunger in the base of the unit is used to regulate the discharge rate through the bottom of the unit. As with the Hukki cone, this unit is not used in industry today.

### 1.2.1.2 Non-Mechanical Hydraulic Classifiers

Linatex classifiers have been in the industry for several years in a variety of applications. The *Linatex S Classifier* is the company's version of a non-mechanical dense flow hydraulic classifier. The pulp is fed by a downcomer into the column where it comes in contact with a deflector plate that causes the flow to turn radially outwards and upwards. The ratio of water between underflow and feed streams controls the upward current at the deflector plate and thus the cut size (Heiskanen, 1993). The unit is very inefficient for sharp separations as it inherently bypasses a large volume of material. It is best utilized for slimes removal.

The Krebs C-H Whirlsizer is another type of non-mechanical dense flow hydraulic classifier. It uses a controlled water addition to a gently swirling pulp to clean the coarse fraction from fines (Heiskanen, 1993). The upper part of the unit is cylindrical in shape, with the lower unit forming a cone as in many of the other units described thus far. The lowermost section of the cylinder contains an internal cone that forces coarse particles into the narrow gap between the wall and the cone. Elutriation water is added below this from small holes, moving the pulp in a swirling action. While no teeter bed is formed, classification takes place by means of hindered settling, allowing the coarse material to settle past the internal cone and the fines to overflow through the top of the unit. It is designed for sand classification and targets the non-spherical materials such as vermiculite, mica and kyanite (Heiskanen, 1993).

### 1.2.1.3 Fluidized Bed Hydraulic Classifiers

A simplified diagram of a fluidized bed hydraulic classifier is shown in Figure 1.2. The traditional design of a fluidized bed hydraulic classifier consists of an open top vessel into which elutriation water is introduced through a series of distribution pipes evenly spaced across the

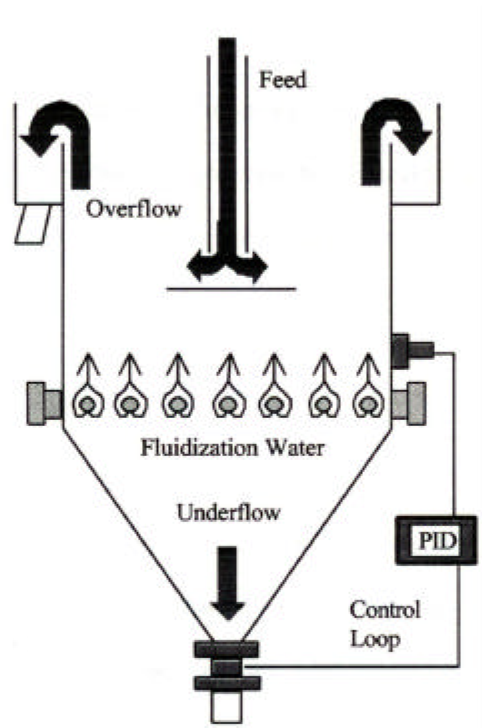



**Figure 1.2. Schematic Diagram of a Traditional Hindered Bed Separator.**

base of the device. During operation, feed solids are injected into the upper section of the separator and are permitted to settle. The elutriation fluid in a fluidized bed supports the weight of the particles within the bed by flowing between the particles. The small interstices within the bed create high interstitial liquid velocities that resist the penetration of the slow settling particles. As a result, small particles accumulate in the upper section of the separator and are eventually carried over the top of the device into a collection launder. Large particles, which settle at a rate faster than the upward current of rising water, eventually pass through the fluidized bed and are discharged out one or more restricted ports through the bottom of the separator.

One of the first hydraulic classifiers to utilize a teeter bed was the Stokes unit which was developed to sort the feed to gravity concentrators. Each teeter chamber is provided at its bottom with a supply of water under constant head which is used for maintaining a teetering condition in the solids that find their way down against the interstitial rising flow of water (Wills, 1992). Each chamber is fitted with its own pressure sensor that monitors the conditions in the chamber and automatically adjusts the discharge to maintain a balanced pressure caused by the teeter bed. A valve at the base of each compartment can be hydraulically or electrically operated to adjust the height of the teeter-bed. As the bed level increases, the pressure will also increase and the valve will open. Likewise, as the bed lowers, the pressure decreases and the valve will close. This action maintains a constant level and, therefore, constant density within the separator.

A more recent hydraulic classifier utilizing the teeter bed is the *Linatex Hydrosizer*. The Linatex Hydrosizer is a non-mechanical, hindered-settling classifier that maintains a fluidized teeter bed, but does not have the same elutriation water distribution or feed distribution as the CrossFlow separator. The pulp is fed into a central feed column where it comes in contact with a deflector plate that causes the flow to turn radially outwards and upwards. Extensive testing of a pilot-scale unit at a North Carolina phosphate plant was conducted in the early 1990's to attrition scrub and deslime flotation feed with promising results. Additional testing has been conducted at other mineral industries including mineral sands and aggregates. The Linatex Hydrosizer was marketed for sizing applications range from 28-mesh to 100-mesh, with some preliminary testing on finer material (NC State, 1992).

Phoenix Process Equipment has developed another type of fluidized bed hydraulic classifier called the Hydrosort. This separator and classifier is currently utilized in the aggregate industry, as well as some others, for separating light, harmful contaminants, such as lignite and wood, in sand washing, and for fractional sand classifications (Phoenix Process Equipment, 2003). The Hydrosort incorporates a fluidized bed created by an upward current of water flow to classify product or separate impurities in the same fashion as the Linatex Hydrosizer. A feature emphasized by Phoenix Equipment is the clog-free classifier bottom, which distributes the upward water flow equally over the separating area. Unlike in the CrossFlow where feed enters the unit tangentially, both the Phoenix Hydrosort and the Linatex Hydrosizer have a feed distribution pipe that entered the top of the unit and discharges feed into the separation chamber.

The Floatex fluidized-bed classifier (or Floatex Density Separator) is the most recent hydraulic separator designed. Like the other units, this separator utilizes a teeter bed which is formed by solids settling against an upward current of elutriation water. Coarse material settles through the teeter-bed, while finer particles report to the overflow of the unit. A differential pressure cell and discharge valve controls the bed level in the unit. This efficient unit sees very little fines bypassed to the underflow and as a result, the unit produces a relatively clean underflow. Prior to the development of the CrossFlow separator, the Floatex separator was considered to be the most advanced commercial separator for hydraulic particle classification for material whose size was between what would be considered optimal for either screens (coarse) or hydrocyclones (fine).

### 1.2.2 Hindered Settling

Hindered settling is an important phenomenon in all of the aforementioned hydraulic classifiers. Hindered settling considers the interaction of other particles in classification systems either on a particle-particle level or from the behavior of the particle assemblies. The interactions between two particles may be due to particles settling close to each other or to the wake effect of a larger particle on the settling of a smaller particle (Heiskanen, 1993). According to Littler (1986), the hindered settling phenomenon begins to take place at approximately 20% solids by mass. The cohesive force between two particles settling very close to one another is great enough for the particles to fall together and be treated as a single particle of greater size and lower density. A wake effect is caused when a larger particle captures a smaller particle in its wake as it is settling and as a result, the smaller particle falls at a velocity much higher than its free settling velocity. In a teeter bed, however, the high solids concentration increases the likelihood of particle collision, and these particles lose some of their settling velocity in these collisions. The fine particles, therefore, have a higher likelihood of being driven to the overflow launder by the upward current of elutriation water. And as a result, hindered settling is more efficient than free settling classification due to the decrease in fines entrained in the underflow.

An analysis of the behavior of particle assemblies can be categorized into two parts. Particle assemblies settling may occupy the whole fluid or they may be considered as clusters of particles which only fill a fractional volume of the fluid (Heiskanen, 1993). When the assemblies occupy the entire fluid they may be treated as a uniform pulp where the interactions are between the individual particles. As clusters, the particles are analyzed as large particles of reduced density and rigidity. The probability of this occurring increases with narrower particle size ranges, and is magnified in gravitational classification where high solids contents are present.

From an analysis standpoint, hydraulic classifiers are characterized by two factors: (i) the size separation and (ii) the sharpness of the separation. For theoretical analyses it is convenient to define separation size as that of particles which settle just fast enough *on the average*, to be totally collected in the underflow (Weiss, 1985). Slight variations in settling rates will occur between particles of the same size and density due to differences in shape and turbulence in the separator. The sharpness of the separation defines how the particles segregate into the product and the tails streams.

Under ideal conditions, a classifier should partition particles coarser than the cut size $d_{50}$ into the coarse stream and finer particles into the overflow (Heiskanen, 1993). The efficiency of this cut is based on the amount of misplaced particles in both streams.

### 1.2.3 Spirals

The first spirals were first utilized in the 1940's for concentrating such metals as gold, silver, tin and mineral sands by Humphreys Minerals Industries. The first use of spirals for washing coal occurred in 1947 when the Hudson Coal Company installed 48 Humphrey starts to wash anthracite fines in Eastern Pennsylvania (Denin *et al.*, 1948). Their success can be attributed to the fact that they are perceived as environmentally friendly, rugged, compact, and cost effective (Kapur *et al.*, 1998). Spirals weren't readily adapted into the industry until the 1980's when interest in recovering coal fines grew along with the introduction of fiberglass and polyurethane lined units. These units were more cost effective and efficient. Prior to that, poor performance, low capacity per unit of floor space and high capital costs kept the original cast iron or concrete units out of production. Today's spirals are able to treat material that is too fine for dense media separators but too coarse for flotation.

Some advantages of spirals include: a lightweight and simple installation process; they require no drives as they are simply pump fed, and have very low operating and maintenance costs. Their capacity and efficiency has increased over the years as twin and triple start units have been developed along with studies on the optimal number of turns to achieve the required separation.

Centrifugal forces immediately act upon the coal slurry as it is fed at between 14-15% solids into the trough of the spiral and allowed to flow downward. Lighter particles attain higher tangential velocities than the nonsuspended particles, causing them to move to the highwall of the spiral. The heavier and coarser particles will work their way towards the interior of the spiral. Any middlings present in the material will tend to report to the center of the spiral. In addition to centrifugal and gravitational forces, differential particle settling rates and interstitial trickling are all working on the particles as they work their way down the spiral trough.

Spiral performance depends largely on the characteristics of the feed coal. The most important operating parameters include feed rate, solids concentration and size and splitter positioning. The volumetric feedrate is the most important operating parameter influencing performance. As volumetric feed rate is increased, an increasing amount of entrained material will report to the outer wall and effectively reduce efficiency. Nominal dry feed rates are typically 2-4 tph per start. The feeds solids concentration has only a small impact on spiral performance compared to the other factors. Spiral can handle up to 45% solids and as little as 20% solids, but 30-35% is considered normal. Spirals have become a common method for concentration of 0.1 m to 3 mm coal, however, Leonard (1991) believes the optimum performance occurs when the top size of the feed is finer than 14 mesh (1.2 mm) and the bottom size is coarser than about 100 mesh (0.15 mm). Cutpoints generally range between 1.70 and 2.00 SG. The splitter positions and solids feed rate largely determine the SG cutpoint and the ash content of the final product.

## 1.3 Field Testing at Coal Plant A

Initial field testing of the pilot-scale CrossFlow separator was conducted at Coal Plant A. This work involved (i) equipment setup, (ii) shakedown testing, and (iii) detailed testing. The goal of this effort was to determine the anticipated product yield and grade, combustible recovery, and feed capacity of the unit in order to predict the expected performance of a full-scale unit. Approximately 3 months of effort were allocated for field-testing. Individuals from Eriez Magnetics and Virginia Tech participated in the testing at Coal Plant A with cooperation from key personnel at the processing plant.

### 1.3.1 Equipment Setup

The separator was transported from Eriez Magnetics Central Research Lab in Erie, PA to the preparation plant. With cooperation from the operators and mechanics at the plant, a 9x16 inch pilot-scale CrossFlow separator was installed at the Coal Plant A. A splitter-box, fabricated at Eriez Magnetics shop in Pennsylvania, was installed to collect the underflow of a classifying cyclone. The cyclones classify the raw feed with the overflow reporting to the froth flotation circuit and the underflow reporting to the water-only cyclones circuit. This splitter was fully adjustable and allowed for the easy regulation of feed rates. The feed sample was conveyed by gravity through a 2 inch line to the CrossFlow separator that was positioned one level below the classifying cyclone. Underflow and overflow material from the separator was discharged to sizing screens in the plant, located on a level below the unit.

Plant compressed air and 115 volt electrical power were connected to the separator for the automated control system. The separator was automatically controlled through the use of a simple PID control loop which includes a pressure sensor mounted on the side of the separator to measure the relative pressure (level), a single loop PID controller, and a pneumatic pinch valve to control the underflow discharge to maintain a constant bed pressure (level). Clarified water was connected to the separator to create the fluidized teeter bed of solids.

### 1.3.2 Shakedown Testing

After completing the installation of the test unit, preliminary shakedown testing was conducted to resolve any unexpected operational problems that could arise. These tests are normally necessary to resolve any problems that may have been overlooked in the initial engineering and to confirm that feed capabilities, pipe sizes, electrical supplies, control systems, etc., are adequate. In addition, these tests provided an opportunity to establish approximate settings for the various process variables required to provide good separation performance based on visual inspections of the product streams.

### 1.3.3 Detailed Testing

Two series of detailed test programs were conducted using the pilot-scale CrossFlow. The first series of tests were performed to investigate the effects of the key design variables on separator performance. Important test variables included: feed injection depth and distributor design. In addition to determining the optimum operating variables, the first series of test simultaneously defined the overall grade and recovery curve for the process. The subsequent round of testing was used to investigate the effects of key operating parameters. The variables examined included: (i) fluidization water rate, (ii) solids mass feed rate, (iii) volumetric slurry feed rate, and (iv) teeter bed depth. A minimum of three settings were examined for each of the listed test parameters. For each test, samples were taken from the feed, overflow, and underflow streams after conditions were stabilized. Each sample was analyzed for ash and sulfur (in many cases on a size-by-size basis).

Due to the low amount of rock present in this feed, a higher feed rate was determined to be acceptable for this application and was utilized in much of the testing. Feed rates ranged from a low of 1 tph/ft$^2$ to a high of 5 tph/ft$^2$. The feed percent solids was reasonably constant at 40%-50% throughout the test period. A significant difference in the feed for each series of testing must be noted as the average ash content for the first series was nearly 14.0% while the average ash content for the second series was only 10.5%.

### 1.3.4 Process Evaluation

To ensure the test data was reliable and self-consistent, all test data was analyzed and adjusted using mass balance software. Experimental values that were deemed by the mass balance routines to be unreliable were removed from the data set. The participating mining company used the compiled data to establish the metallurgical improvement, operating savings and economic payback that may be realized by implementing the proposed high-efficiency technologies.

The as-tested coal slurry was found to have a mean particle size of 0.631 mm during the first series of testing and 0.572 mm during the second series of testing. The solids specific gravity was measured to be 1.55 with a solids content of 50%. The feed size distribution is summarized in Table 1.1. Table 1.2 provides a summary of the operating parameters that were investigated during both rounds of testing.

**Table 1.1. Feed Size Distribution of Coal Plant A.**

| Size Class | Round 1 | | Round 2 | |
|---|---|---|---|---|
| (Mesh) | (% in Class) | (% Ash) | (% in Class) | (% Ash) |
| +10 | 2.38 | 14.02 | 1.21 | 6.67 |
| 10x14 | 8.48 | 10.94 | 6.92 | 7.09 |
| 14x28 | 38.89 | 11.77 | 34.89 | 7.27 |
| 28x60 | 26.53 | 11.45 | 29.56 | 8.60 |
| 60x100 | 11.49 | 12.99 | 10.38 | 9.69 |
| -100 | 12.23 | 28.28 | 17.04 | 21.89 |
| Overall | 100.00 | 13.83 | 100.00 | 10.39 |

**Table 1.2. Operating Parameters for On-Site Pilot-Scale Testing at Coal Plant A.**

| Test No. | Feed (gpm) | (SG) | (% Solids) | (tph/sqft) | T-Water (gpm) | Level No. | Comment |
|---|---|---|---|---|---|---|---|
| 1 | 10.0 | 1.170 | 40.9 | 1.2 | 6.0 | 64 | Round 1 |
| 2 | 10.0 | 1.170 | 40.9 | 1.2 | 6.0 | 64 | |
| 3 | 9.2 | 1.180 | 43.0 | 1.2 | 6.0 | 99 | Sensor Reset |
| 4 | 9.7 | 1.165 | 39.9 | 1.1 | 6.0 | 97 | |
| 5 | 9.8 | 1.170 | 40.9 | 1.2 | 6.0 | 95 | |
| 6 | 9.8 | 1.160 | 38.9 | 1.1 | 6.0 | 96 | |
| 7 | 12.6 | 1.165 | 39.9 | 1.5 | 6.0 | 96 | |
| 8 | 12.7 | 1.180 | 43.0 | 1.6 | 6.0 | 92 | |
| 9 | 30.0 | 1.210 | 48.9 | 4.4 | 7.0 | 94 | Round 2 |
| 10 | 35.3 | 1.200 | 47.0 | 5.0 | 7.0 | 92 | |
| 11 | 33.3 | 1.197 | 46.4 | 4.6 | 7.0 | 93 | |
| 12 | 28.6 | 1.200 | 47.0 | 4.0 | 7.0 | 94 | |
| 13 | 28.6 | 1.200 | 47.0 | 4.0 | 7.0 | 94 | |
| 14 | 20.0 | 1.185 | 44.0 | 2.6 | 7.0 | 90 | |

The results from the on-site CrossFlow separator investigation are shown graphically in Figures 1.3 through 1.5. The results are summarized as: "As-Tested" and "x 100 Mesh" with the passing 100-mesh material mathematically removed from the data. This approach is acceptable as it is expected that the clean coal product will be deslimed at approximately 0.150 mm. The material finer than 100 mesh will be upgraded by flotation at this particular plant.

As presented in Figure 1.3, this pilot-scale test work was able to define the expected grade and recovery curve for this particular coal. Specifically, the CrossFlow separator is capable of providing a clean product ranging between 6% and 8% ash at a combustible recovery of greater than 95% (when deslimed at 100 mesh). At maximum separation efficiency, the combustible recovery, for this application, approached 98%. The data presented in Figure 1.4 indicates that the sulfur content of the corresponding product will be approximately 1.75%.

Figure 1.5 is included to demonstrate the ability of the CrossFlow separator to provide high combustible recoveries even when operated at elevated throughput rates. During the second series of testing, the feed rate was increased to a very high value of 5 tph/ft$^2$. During this time, the combustible recovery remained unaffected. It must also be noted that the feed ash during this second series of testing was significantly lower than the first series of testing, resulting in product yields greater than 96%. Simply stated, there was not a significant amount of rock present in the feed stream. Regardless, the CrossFlow separator was able to produce a tailings stream with an ash content averaging 76.5% and a corresponding sulfur content averaging 12.20% for this particular feed coal.

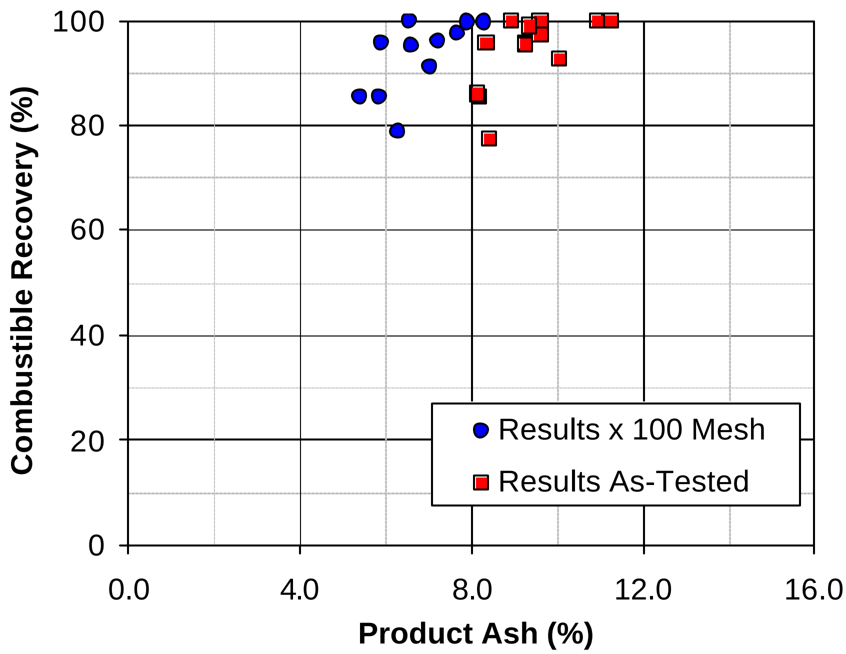



**Figure 1.3. Combustible Recovery vs. Product Ash Content.**

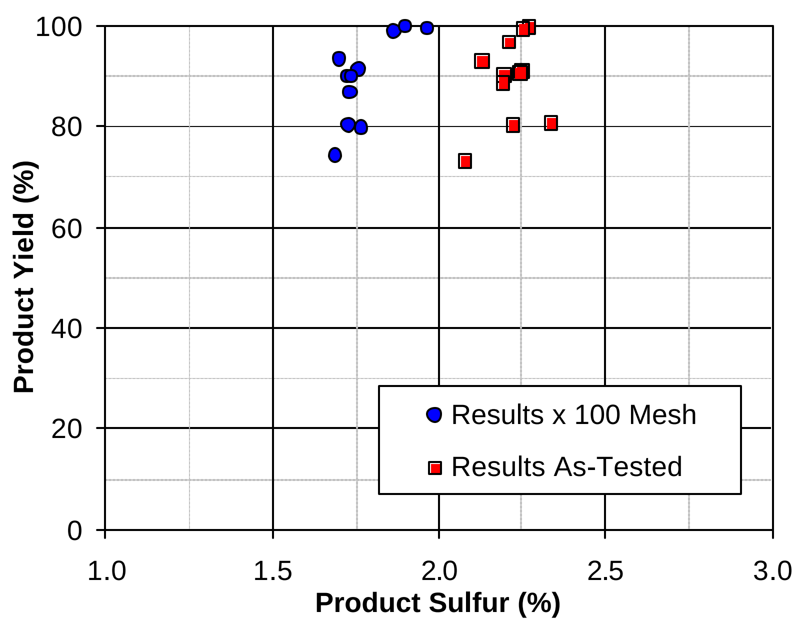



**Figure 1.4 Mass Yield vs. Product Sulfur Content.**

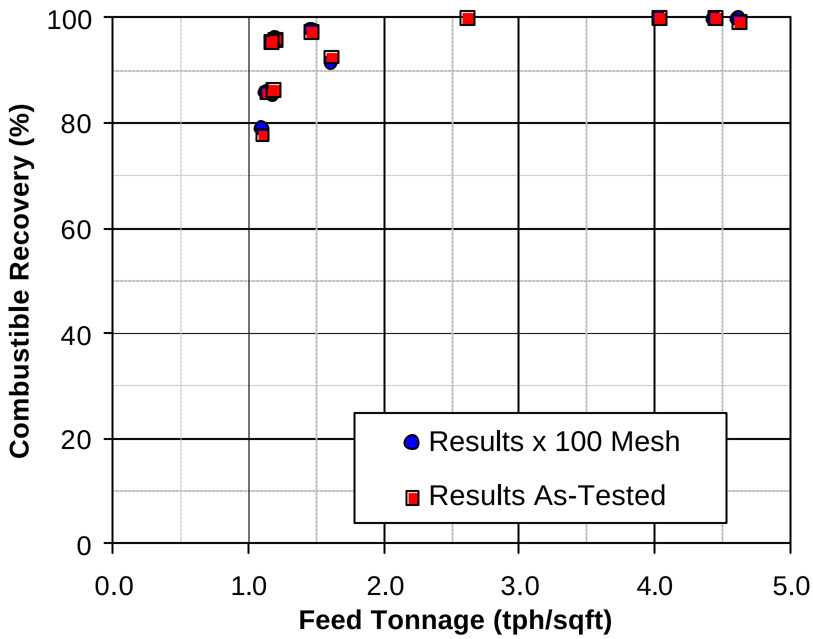



**Figure 1.5 Combustible Recovery vs. Feed Tonnage.**

The material balance presented in Figure 1.6 is included as a summary of the test work conducted at the Coal Plant A. This material balance includes all expected metallurgical results, ancillary requirements, and volumetric flows for a full-scale installation with the capacity to treat 150 tph of feed at approximately 50% solids (by weight).

For this duty, a 7x7-ft CrossFlow separator has been recommended for the operation, offering 49 $ft^2$ of cross-sectional area which results in a normalized feed rate of 3 $tph/ft^2$. The current test work has demonstrated the ability of the CrossFlow separator to handle this entire flow in a single stage circuit.

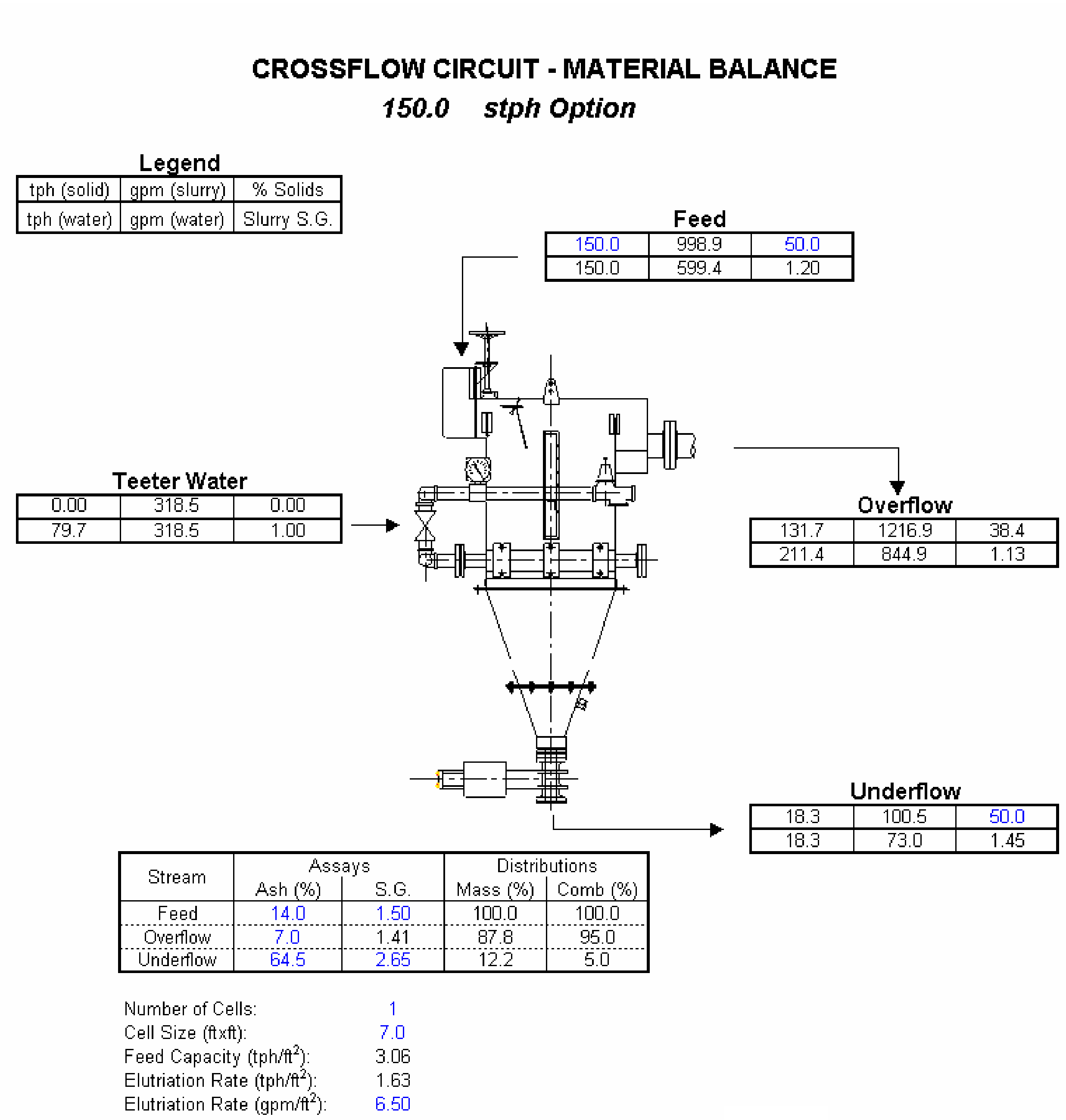



**Figure 1.6 Expected Material Balance for a Crossflow Separator Treating 150 tph.**

### 1.3.5 Sample Analysis

Detailed analysis was conducted on each of the samples collected during the testing program. The analyses were performed in accordance with ASTM procedures currently applied at Coal Plant A. Representative samples were collected around the pilot-scale unit. Slurry flow rates for the feed, underflow and overflow streams were directly measured using a stopwatch and a calibrated container. The mass and liquid flow rates were then calculated from the measured slurry flow rates and the sample assays using the two-product formula.

### 1.3.6 Future Work

A final report was presented to the management of Coal Plant A. The project is currently waiting for management's approval before further testing is performed or a full scale unit is installed.

## 1.4 <u>In-Plant Testing at Coal Plant B</u>

The next set of field-tests with the pilot scale CrossFlow separator were carried out at a second coal plant (Plant B). As before, this work involved (i) equipment setup, (ii) shakedown testing, and (iii) detailed testing. In this particular case, the goal of this effort was to determine the anticipated product yield and grade, combustible recovery, and feed capacity of the unit for comparison against the existing spiral circuit. Approximately 3 months of effort were allocated for field-testing. Individuals from Eriez Magnetics and Virginia Tech participated in the testing at Coal Plant B with cooperation from key personnel at the processing plant.

### 1.4.1 Equipment Setup

The separator was transported from the Coal and Minerals Research Lab at Virginia Tech in Blacksburg, Virginia to the preparation plant. The 9x16 inch pilot-scale CrossFlow separator was installed at the Coal Plant B as shown in Figure 1.7. Feed was supplied to the CrossFlow separator through a 2 inch line connected to existing coal spiral slurry feed distributor. A slurry splitter fabricated from PVC pipe with a tee and valves was used to regulate the feed to the unit, with the remaining slurry reporting to the spiral circuit. Underflow and overflow material was discharged to sizing screens in the plant, located on a level below the unit.

Plant compressed air and 115 volt electrical power were connected to the separator for the automated control system. The separator was automatically controlled through the use of a simple PID control loop which includes a pressure sensor mounted on the side of the separator to measure the relative pressure (level), a single loop PID controller, and a pneumatic pinch valve

**Figure 1.7. The 9x16 Inch Pilot-Scale Crossflow Test Circuit at Coal Plant B.**

to control the underflow discharge to maintain a constant bed pressure (level). Clarified water was connected to the separator to create the fluidized teeter bed of solids.

### 1.4.2 Shakedown Testing

After completing the installation of the test unit, preliminary shakedown testing was conducted to resolve any unexpected operational problems that could arise. These tests are necessary to resolve any problems that may have been overlooked in the initial engineering and to confirm that feed capabilities, pipe sizes, electrical supplies, control systems, etc., are adequate.

### 1.4.3 Detailed Testing

Two series of detailed test programs were conducted using the pilot-scale test unit. The first series of tests were performed to investigate the effects of the key design variables on separator performance and to simultaneously define the overall grade and recovery curve. The subsequent series of testing was performed to investigate the effects of key operating parameters. Tests were conducted primarily as a function of teeter bed pressure and fluidization water rate. The coal/rock interface, or teeter bed, was adjusted to different levels (i.e. different bed pressure) for each steady-state test. Fluidization water was adjusted to fine tune the separation. Other variables considered were solids mass feed rate and volumetric slurry feed rate. For each test, samples were taken from the feed, overflow, and underflow streams after conditions were stabilized. The samples were analyzed for ash and sulfur (by-size).

Six test runs were completed during the on-site test work. Additionally, a set of samples was taken with regard to the existing coal spirals. The spiral samples were collected during the same time frame as tests #3, #4, and #5 of the CrossFlow separator evaluation.

### 1.4.4 Process Evaluation

To ensure the test data was reliable and self-consistent, all as-received results were analyzed and adjusted using mass balance software. Experimental values that were deemed by the mass balance routines to be unreliable were removed from the data set. The participating mining company used the compiled data to establish the metallurgical improvement, operating savings and economic payback that may be realized by implementing the proposed high-efficiency technologies.

The particles in the feed slurry were found to have a mean diameter of 0.406 mm. The solids specific gravity was measured to be 1.55. Feed percent solids ranged between 35% and 40% and the feed rate varied from 2.0-2.8 tph/ft$^2$. The feed size distribution is summarized in Table 1.3. Table 1.4 is a summary of the array of operating parameters that were investigated during testing.

**Table 1.3. Feed Size Distribution of Coal Plant B.**

| Stream Description | Size | | | Weight (%) | Ash (%) |
|---|---|---|---|---|---|
| | Passing | Retained | Mean | | |
| Plus 16 M | *** | 1.000 | 1.000 | 6.24 | 9.94 |
| 16x32 M | 1.000 | 0.500 | 0.707 | 23.87 | 11.23 |
| 32x60 M | 0.500 | 0.250 | 0.354 | 29.38 | 12.53 |
| 60x100 M | 0.250 | 0.150 | 0.194 | 22.54 | 13.75 |
| Minus 100 M | 0.150 | *** | 0.150 | 17.97 | 39.17 |
| Composite | | | 0.406 | 100.00 | 17.12 |

**Table 1.4. Operating Parameters for On-Site Pilot Scale Testing at Coal Plant B.**

| Unit Operation | Test Number | Feed | | | Level | Water |
|---|---|---|---|---|---|---|
| | | % Solids | tph | gpm | inches | gpm |
| CrossFlow | XF1 | 35.5 | 2.01 | 21.90 | 6.0 | 4.76 |
| CrossFlow | XF2 | 36.3 | 2.38 | 23.35 | 12.0 | 4.76 |
| CrossFlow | XF3 | 38.5 | 2.83 | 26.06 | 8.0 | 3.61 |
| CrossFlow | XF4 | 37.2 | 2.56 | 24.79 | 8.0 | 4.72 |
| CrossFlow | XF5 | 35.8 | 2.49 | 25.02 | 8.0 | 5.51 |
| CrossFlow | XF6 | 38.1 | 2.48 | 24.37 | 8.0 | 4.44 |
| Spiral* | 7 | 38.0 | 3.50 | 32.70 | n/a | n/a |

* Samples taken during tests 3, 4 and 5
* Multiple starts, 3 product screen feed, 1 reject screen feed

The as-received results, as analyzed and adjusted using a mass balance program, are reported in Table 1.5. The products were sized at 100 mesh so that each fraction could be evaluated separately. As expected, the minus 100 mesh product had a higher ash content than the plus 100 mesh fraction. This is expected as fine material, especially passing 150 mesh, tends to report to the separator overflow due to its relatively small mass. In essence, the teeter water overcomes the settling velocity of these particles and flushes them out of the separator. As such, the results in this report are compared on a plus 100 mesh basis. This is acceptable as the existing circuit incorporates dewatering screens for each of the product streams.

The results from the pilot-scale CrossFlow separator investigation are shown graphically in Figure 1.8 for the +100 mesh material. The results of the CrossFlow separator are comparable to the existing coal spirals. Upon close examination (Figure 1.8 inset), when compared to the coal spirals, the CrossFlow separator provides a marginally better clean coal yield at 96% vs. 92%. However, the higher product yield also generates a product with slightly higher ash content at 9.25-10.00% vs. 8.8%. Lower product ash values are possible using the CrossFlow separator and can be achieved through lower fluidization rates and/or bed pressures.

**Table 1.5. In-Plant Test Results for Coal Plant B.**

| Size Fraction | Test Number | Mass Yield (%) | Ash Assay (%) | Sulfur Assay (%) | Comb Rec (%) | Sulfur Rec (%) | Ash Rej (%) | Sulfur Rej (%) |
|---|---|---|---|---|---|---|---|---|
| Plus 100 | XF1 | 96.04 | 10.06 | 4.07 | 98.80 | 92.60 | 23.19 | 7.40 |
| Composite | XF1 | 96.54 | 15.80 | 5.16 | 98.91 | 94.54 | 14.38 | 5.46 |
| Feed | XF1 | 100.00 | 17.81 | 5.27 | 100.00 | 100.00 | 0.00 | 0.00 |
| Plus 100 | XF2 | 96.22 | 9.25 | 3.97 | 98.50 | 93.46 | 21.56 | 6.54 |
| Composite | XF2 | 96.80 | 15.00 | 4.99 | 98.67 | 95.18 | 12.59 | 4.82 |
| Feed | XF2 | 100.00 | 16.61 | 5.08 | 100.00 | 100.00 | 0.00 | 0.00 |
| Plus 100 | XF3 | 94.09 | 10.03 | 4.04 | 97.94 | 89.91 | 30.48 | 10.09 |
| Composite | XF3 | 94.87 | 15.48 | 5.16 | 98.14 | 92.40 | 19.72 | 7.60 |
| Feed | XF3 | 100.00 | 18.30 | 5.30 | 100.00 | 100.00 | 0.00 | 0.00 |
| Plus 100 | XF4 | 96.16 | 9.76 | 4.04 | 98.75 | 92.73 | 22.61 | 7.27 |
| Composite | XF4 | 96.76 | 14.75 | 5.17 | 98.90 | 94.82 | 13.98 | 5.18 |
| Feed | XF4 | 100.00 | 16.60 | 5.28 | 100.00 | 100.00 | 0.00 | 0.00 |
| Plus 100 | XF5 | 96.70 | 10.12 | 4.05 | 99.17 | 94.39 | 20.78 | 5.61 |
| Composite | XF5 | 97.21 | 16.18 | 5.33 | 99.25 | 96.07 | 12.14 | 3.93 |
| Feed | XF5 | 100.00 | 17.90 | 5.39 | 100.00 | 100.00 | 0.00 | 0.00 |
| Plus 100 | XF6 | 97.03 | 10.11 | 4.06 | 99.26 | 94.30 | 19.16 | 5.70 |
| Composite | XF6 | 97.47 | 15.19 | 5.35 | 99.34 | 96.04 | 11.79 | 3.96 |
| Feed | XF6 | 100.00 | 16.78 | 5.43 | 100.00 | 100.00 | 0.00 | 0.00 |
| Plus 100 | Spiral | 92.32 | 8.82 | 4.08 | 97.10 | 87.91 | 38.82 | 12.09 |
| Composite | Spiral | 90.56 | 13.75 | 5.00 | 96.19 | 81.84 | 33.77 | 18.16 |
| Feed | Spiral | 100.00 | 18.80 | 5.53 | 100.00 | 100.00 | 0.00 | 0.00 |

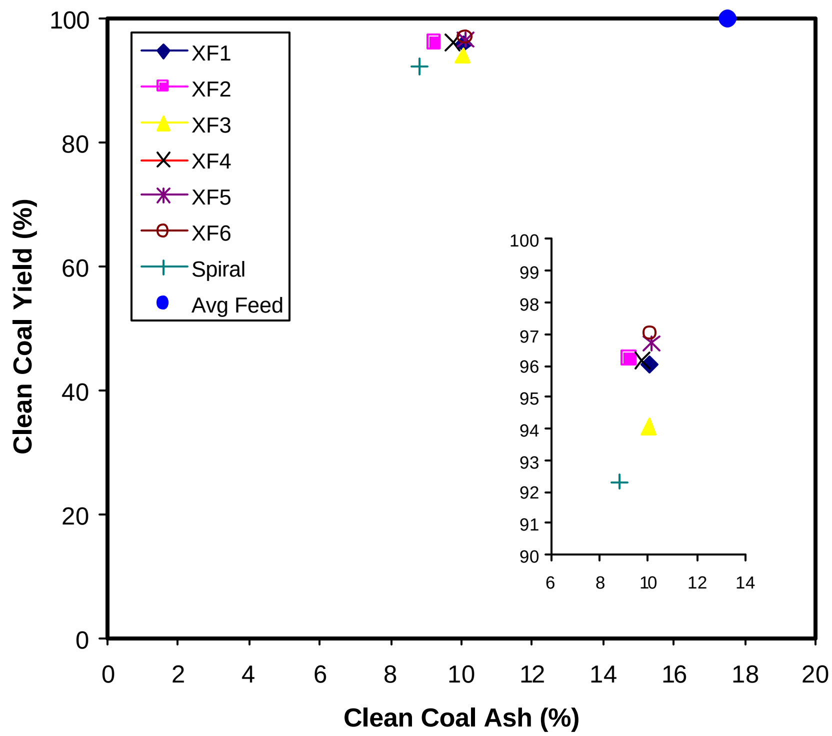



**Figure 1.8 Yield vs. Clean Coal Ash for +100-Mesh Size Fraction.**

Data in Figures 1.9 and 1.10 shows the size-by-size results from Table 1.5 graphed by size class. In these charts, the left most (i.e., lowest ash and sulfur) data points correspond to the plus 100 mesh size fraction. The data points in the middle position represents the composite (100 mesh x 0) size fraction. The right-most data points (shown at 100% yield) correspond to the feed grade.

The data demonstrate that for any given product ash content or sulfur content, the CrossFlow separator can produce a higher clean coal yield when compared to the existing coal spirals. Essentially, at 10% product ash content, the CrossFlow separator operates with a clean coal yield ranging between 96% and 97%, while the spirals produce a yield of approximately 92%. It should be noted that a 4% difference in clean coal yield for a 200 tph circuit can represent a $1,400,000 per year (i.e., 200 tph x 7000 hr/yr x $25/ton x ($Yield_{CF}$ - $Yield_S$)). A similar trend is also shown when examining the sulfur data (Figure 1.10).

Results indicate that the performance of the CrossFlow separator was equal or superior to the performance of the existing spiral circuit for this preparation plant. The material balance presented in Figure 1.11 is included as a summary of the test work conducted at the Coal Plant B. The material balance includes all expected metallurgical results, ancillary requirements, and volumetric flows for a full-scale installation capable of treating the required 200 tph flow with one 7x7-ft CrossFlow separator.

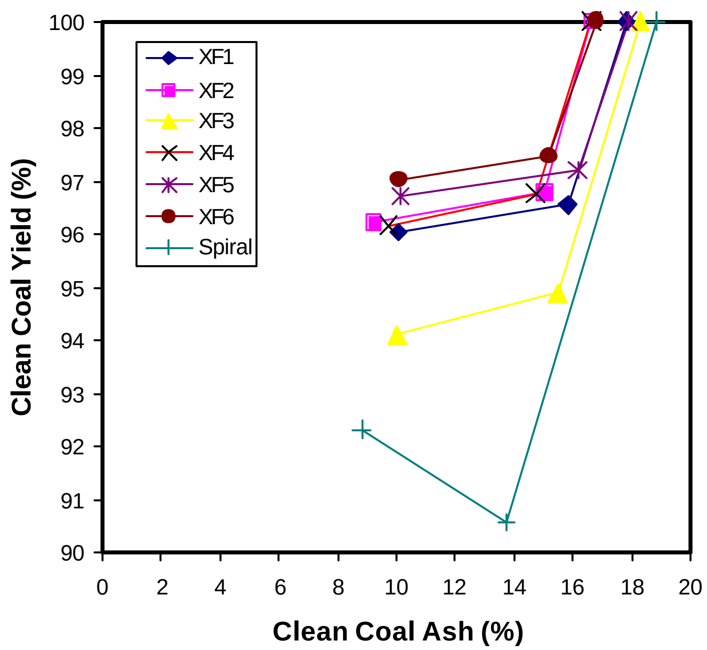



**Figure 1.9. Performance for +100 Mesh and Composite Samples.**

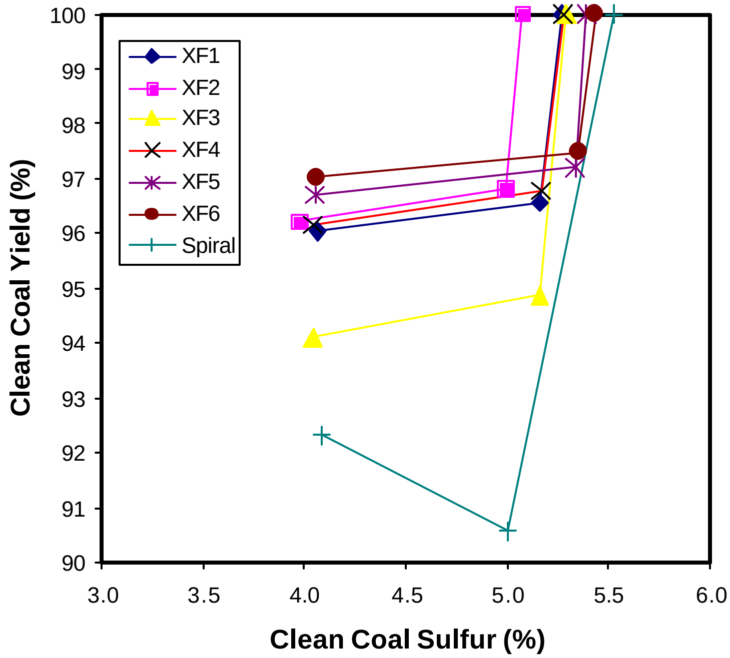



**Figure 1.10. Performance for +100 Mesh and Composite Samples.**

## CROSSFLOW CIRCUIT - MATERIAL BALANCE

### *200.0 stph Option*

**Legend**

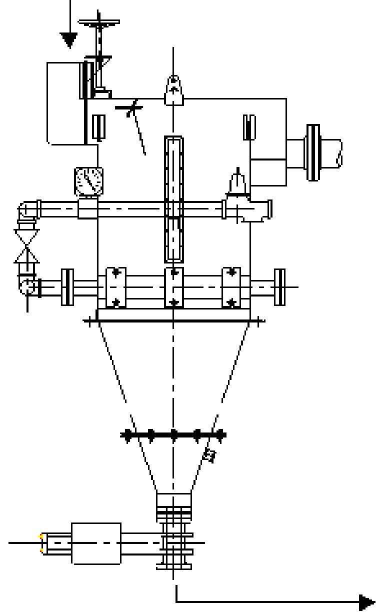
| | | |
|---|---|---|
| tph (solid) | gpm (slurry) | % Solids |
| tph (water) | gpm (water) | Slurry S.G. |

**Feed**

| | | |
|---|---|---|
| 200.0 | 1731.5 | 40.0 |
| 300.0 | 1198.7 | 1.15 |

**Teeter Water**

| | | |
|---|---|---|
| 0.00 | 232.8 | 0.00 |
| 58.2 | 232.8 | 1.00 |



**Overflow**

| | | |
|---|---|---|
| 190.4 | 1918.3 | 35.2 |
| 350.4 | 1400.1 | 1.13 |

**Underflow**

| | | |
|---|---|---|
| 9.6 | 45.9 | 55.0 |
| 7.9 | 31.4 | 1.52 |

| Stream | Assays | | Distributions | |
|---|---|---|---|---|
| | Ash (%) | S.G. | Mass (%) | Comb (%) |
| Feed | 17.5 | 1.50 | 100.0 | 100.0 |
| Overflow | 14.8 | 1.47 | 95.2 | 98.4 |
| Underflow | 72.0 | 2.65 | 4.8 | 1.6 |

Number of Cells: 1
Cell Size (ftxft): 7.0
Feed Capacity (tph/ft$^2$): 4.08
Elutriation Rate (tph/ft$^2$): 1.19
Elutriation Rate (gpm/ft$^2$): 4.75

* Ash Value of +100Mesh Clean Coal Fraction Estimated between 8% and 10%. Equivalent to Spirals.

**Figure 1.11. Material Balance for Two 7x7 ft CrossFlow Separators.**

### 1.4.5 Sample Analysis

Detailed analysis was conducted on each of the samples collected during the testing program. The analyses were performed in accordance with ASTM procedures at Tra-Det Inc. laboratory in Tridelphia, West Virginia. Representative samples were collected around the pilot-scale unit. Slurry flow rates for the feed, underflow and overflow streams were directly measured using a stopwatch and a calibrated container. The mass and liquid flow rates were then calculated from the measured slurry flow rates and the sample assays using the two-product formula.

### 1.4.6 Future Work

While the results look promising, the management will not be installing a unit at this particular plant. However, these data were used by the company to justify moving ahead with the installation of this technology at another plant owned by the company.

## 1.5 In-Plant Testing of Coal Plant C

Additional field testing of the CrossFlow separator was performed for Coal Plant C. This work involved equipment setup, shakedown and detailed testing. The goal of this particular effort was to determine the anticipated product yield and grade, combustible recovery, and feed capacity of the unit. In this case, the CrossFlow separator was to be evaluated as a potential replacement for an existing single-stage spiral circuit. Approximately 3 months of effort were allocated for field-testing at this site. Individuals from Virginia Tech and University of Kentucky participated in the testing at Coal Plant C with cooperation from key personnel at the processing plant.

### 1.5.1 Equipment Setup

The CrossFlow separator was transported from the University of Kentucky in Lexington, Kentucky, to the preparation plant. With cooperation from the operators and mechanics at the plant, the 12-inch diameter pilot-scale CrossFlow separator was installed at the Coal Plant C (see Figure 1.12). Feed was supplied to the CrossFlow separator through a 2-inch line by connecting to an existing coal slurry spiral feed distributor. A slurry splitter fabricated from PVC pipe with a tee and valves was used to regulate the feed to the unit, with the remaining slurry reporting to the spiral circuit. Underflow and overflow material was discharged to the spiral underflow launders.

As with the other test sites, plant compressed air and 115 volt electrical power were connected to the separator for the automated control system. The separator was automatically controlled through the use of a simple PID control loop which includes a pressure sensor mounted on the side of the separator to measure the relative pressure (level), a single loop PID controller, and a pneumatic pinch valve to control the underflow discharge to maintain a constant bed pressure (level). Clarified water was connected to the separator to create the fluidized teeter bed of solids.

**Figure 1.12. The 12-inch Diameter Pilot-Scale CrossFlow Separator Test Circuit at Plant C.**

### 1.5.2 Shakedown Testing

After installation was complete, preliminary shakedown testing of the unit was conducted to resolve any unexpected operational problems that could arise. These tests are designed to resolve any problems that may have been overlooked in the initial engineering and to confirm that feed capabilities, pipe sizes, electrical supplies, control systems, etc., are adequate.

### 1.5.3 Detailed Testing

Two series of detailed test programs were conducted using the pilot-scale test unit. The first series of tests were performed to investigate the effects of the key design variables on separator performance and to simultaneously define the overall grade and recovery curve. The subsequent series of testing was used to investigate the effects of key operating parameters. Tests were conducted primarily as a function of teeter bed pressure and fluidization water rate. The coal/rock interface, or teeter bed, was adjusted to different levels (i.e. different bed pressure) for each steady-state test. Other variables that were considered were solids mass feed rate and volumetric slurry feed rate. For each test, samples were taken from the feed, overflow, and underflow streams after conditions were stabilized. Each sample was sized and analyzed for ash and sulfur contents.

Nine test runs were completed during the on-site test work conducted at Coal Plant C. Table 1.6 is a summary of the operating parameters that were investigated during testing. The set point transition between tests #4 and #5 is due to recalibration of the control system. The difference in the set point when treating the Seam A and Seam B is due to the particle size distribution difference and the desire to maintain a constant bed height. Additionally, samples were collected from the process streams of the existing coal spirals when treating the Seam A and Seam B fine coal.

**Table 1.6. Operating Parameters for On-Site Pilot Scale Testing at Coal Plant C.**

| Test | Seam | Set Point | Solid Density (%) | Feed | | | | |
|---|---|---|---|---|---|---|---|---|
| | | | | Pulp Density ($gm/cm^3$) | Percent Solids (%) | Rate | | |
| | | | | | | gpm | tph | $tph/ft^2$ |
| 1 | B | 46 | 1.6 | 1.088 | 21.57 | 12 | 0.7 | 0.4 |
| 2 | A | 46 | 1.6 | 1.13 | 30.68 | 14 | 1.21 | 0.69 |
| 3 | A | 46 | 1.6 | 1.09 | 22.02 | 11.65 | 0.7 | 0.4 |
| 4 | A | 45 | 1.6 | 1.09 | 22.02 | 12.32 | 0.74 | 0.42 |
| 5 | B | 78 | 1.6 | 1.1 | 24.24 | 9.83 | 0.66 | 0.37 |
| 6 | B | 79 | 1.6 | 1.13 | 30.68 | 9.49 | 0.82 | 0.47 |
| 7 | A | 87 | 1.6 | 1.125 | 29.63 | 19.47 | 1.62 | 0.92 |
| 8 | A | 88 | 1.6 | 1.1 | 24.24 | 16.22 | 1.08 | 0.61 |
| 9 | B | 80 | 1.6 | 1.13 | 30.68 | 10.5 | 0.91 | 0.52 |

### 1.5.4 Process Evaluation

To ensure the test data was reliable and self-consistent, all as-received results were analyzed and adjusted using mass balance software. Experimental values that were deemed by the mass balance routines to be unreliable were removed from the data set. The participating mining company used the compiled data to establish the metallurgical improvement, operating savings and economic payback that may be realized by implementing the proposed high-efficiency technologies.

The Coal Plant C treats coal from both the coal seams separately. As such, the teeter-bed unit was evaluated for the cleaning potential of the nominal 16 x 100 mesh fractions of both coals. Feed percent solids ranged between 22% and 30% during the test program, with variations in the mass feed rate to the unit varying from 0.37-0.92 $tph/ft^2$. Samples of the feed to the teeter-bed unit were taken and subjected to washability and particle size analysis. The washability data indicates that both coals can be classified as 'easy-to-clean' based on their relatively low contents of middling material, their cumulative float ash contents of less than 5%, and combustible recovery greater than 95%. The difference in the two coals is that the Seam B coal produces a one percentage point lower float ash content.

The particle size distribution of Seam B feed coal was significantly finer than the Seam A coal as shown in Table 1.7. The minus 100 mesh fraction was removed from the particle size analysis since the concentration on cleaning potential was isolated on the plus 100 mesh material. Both coals only had 1% to 2% by weight of plus 16 mesh material in the feed. However, the Seam B material had nearly 12 percentage points less of the coarsest plus 28 mesh size fraction. This finding explained the need to operate at this particular site at lower bed pressure settings in order to maintain the same fluidized particle bed height. The distributions of the ash-bearing material in both coals are nearly equivalent.

**Table 1.7. Feed Size Distribution for Coal Plant C.**

| Particle Size (Mesh) | Seam B | | Seam A | |
|---|---|---|---|---|
| | Weight (%) | Ash (%) | Weight (%) | Ash (%) |
| +28 | 16.98 | 15.63 | 29.20 | 17.28 |
| 28 x 48 | 35.98 | 18.38 | 31.56 | 19.30 |
| 48 x 100 | 47.04 | 19.51 | 39.24 | 19.20 |
| Total | 100.00 | 18.44 | 100.00 | 18.67 |

The teeter-bed unit achieved excellent separation performances for both feed coals as shown in Table 1.8 and Figure 1.13. For the Seam B coal, the ash content was reduced from 17.57% to a value as low as 6.51% while recovering 97% of the combustible material. Similar performances were achieved on the Seam A coal with product ash values as low as 7.51%. The performances from eight of the nine tests were very close to ideal as indicated by the comparison with the washability data in Figure 1.13. The teeter-bed performances compare favorably with those achieved by the existing spiral circuit shown in Table 1.9.

The size-by-size performance of the test unit is shown in Tables 1.10 and 1.11 for the Seam B and Seam A coals, respectively. These results indicate that the teeter-bed unit performed exceptionally well on the plus 28 mesh and the 28 x 48 mesh particle size fractions. For example, a 2.87% product ash was achieved from the plus 28 mesh Seam B coal, while the tailings ash content was maintained at a relatively high 72.26%. However, the separation density appears to shift upward significantly with a decrease in particle size as evident by the higher product ash contents in the 48 x 100 mesh particle size fractions of both coals.

**Table 1.8. Teeter-Bed Separation Performances at Coal Plant C.**

| Test Number | Ash (%) | | | Yield (%) | Recovery (%) |
|---|---|---|---|---|---|
| | Feed (%) | Product (%) | Tailing (%) | | |
| 1 | 19.96 | 9.97 | 86.67 | 86.97 | 97.83 |
| 2 | 20.54 | 14.55 | 86.88 | 91.71 | 98.63 |
| 3 | 18.99 | 8.45 | 82.06 | 85.68 | 96.83 |
| 4 | 24.05 | 10.01 | 76.51 | 78.89 | 93.47 |
| 5 | 17.57 | 6.51 | 84.08 | 85.74 | 97.25 |
| 6 | 17.57 | 7.69 | 86.43 | 87.45 | 97.93 |
| 7 | 21.44 | 13.45 | 86.43 | 89.06 | 98.11 |
| 8 | 21.21 | 8.86 | 83.25 | 83.40 | 96.47 |
| 9 | 23.43 | 7.51 | 50.09 | 62.61 | 75.63 |

**Table 1.9. Separation Performances Achieved by the Existing Spiral Circuit at Coal Plant C.**

| Test Number | Seam | Ash (%) | | | Yield (%) | Recovery (%) |
|---|---|---|---|---|---|---|
| | | Feed | Product | Tailing | | |
| 1 | SEAM B | 17.57 | 5.22 | 85.51 | 84.61 | 97.29 |
| 2 | SEAM A | 21.44 | 7.49 | 85.07 | 82.03 | 96.58 |

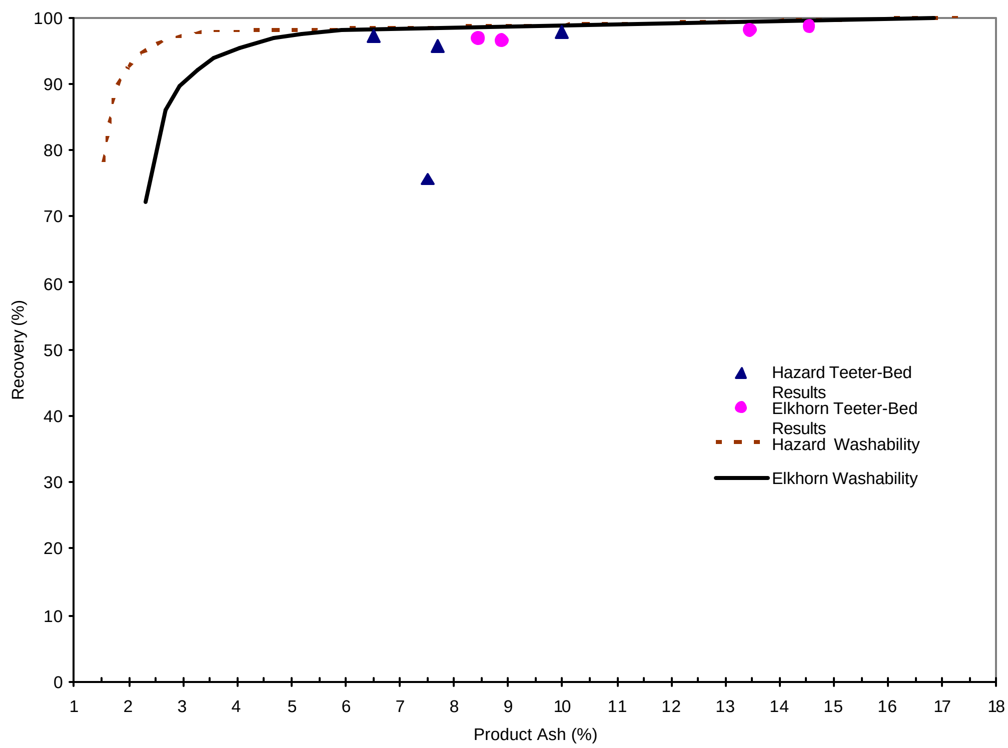



**Figure 1.13. Comparison of the Teeter-Bed Separation Performances and the Washability Characteristics of the Seam A and Seam B Fine Coals at Coal Plant C.**

**Table 1.10. Particle Size-By-Size Separation Performance Achieved from the Treatment of the Seam B Fine Coal.**

| Particle Size (Mesh) | Feed | | Product | | Tailings | | Yield (%) | Recovery (%) |
|---|---|---|---|---|---|---|---|---|
| | Weight (%) | Ash (%) | Weight (%) | Ash (%) | Weight (%) | Ash (%) | | |
| + 28 | 16.98 | 15.63 | 17.32 | 2.87 | 22.74 | 72.26 | 81.61 | 93.95 |
| 28 x 48 | 35.98 | 18.38 | 44.87 | 4.53 | 49.12 | 85.86 | 82.97 | 97.05 |
| 48 x 100 | 47.04 | 19.51 | 37.81 | 11.70 | 28.14 | 89.96 | 82.97 | 98.76 |
| Total | 100.00 | 18.44 | 100.00 | 6.95 | 100.00 | 83.92 | 85.07 | 97.06 |

**Table 1.11. Particle Size-By-Size Separation Performance Achieved from the Treatment of the Seam A Fine Coal.**

| Particle Size (Mesh) | Feed | | Product | | Tailings | | Yield (%) | Recovery (%) |
|---|---|---|---|---|---|---|---|---|
| | Weight (%) | Ash (%) | Weight (%) | Ash (%) | Weight (%) | Ash (%) | | |
| + 28 | 29.20 | 17.28 | 22.38 | 5.01 | 44.45 | 74.48 | 82.34 | 94.55 |
| 28 x 48 | 31.56 | 19.30 | 34.84 | 7.66 | 38.69 | 84.48 | 84.85 | 97.09 |
| 48 x 100 | 39.24 | 19.20 | 42.78 | 12.03 | 16.86 | 87.81 | 90.54 | 98.57 |
| Total | 100.00 | 18.67 | 100.00 | 8.94 | 100.00 | 80.60 | 86.42 | 96.76 |

### 1.5.5 Sample Analysis

Detailed analysis was conducted on each of the samples collected during the testing program. The analyses were performed in accordance with ASTM procedures at the University of Kentucky. Representative samples were collected around the pilot-scale unit. Slurry flow rates for the feed, underflow and overflow streams were directly measured using a stopwatch and a calibrated container. The mass and liquid flow rates were then calculated from the measured slurry flow rates and the sample assays using the two-product formula.

### 1.5.6 Future Work

Because of the promising results obtained from this study, a more detailed test program will be conducted at the Coal Plant C. The goal of this additional work will be (i) to obtain data needed to identify the optimum separation performances for the test unit and (ii) to compare the optimum performance data with similar results obtained from the existing spiral circuit. This work is currently scheduled to be completed sometime during the fall of 2004.

## 1.6 In-Plant Testing at Coal Plant D

The next coal plant involved in the field-testing of the pilot-scale CrossFlow separator was Coal Plant D. As with the other test sites, this work involved (i) equipment setup, (ii) shakedown testing, and (iii) detailed testing. The goal of this effort was to determine the anticipated product yield and grade, combustible recovery, and feed capacity of the test unit in order to predict the expected performance of a full-scale unit. In this particular case, the testing was performed to determine whether the installation of one or more full-scale units could be justified at a new green-field plant in Kentucky. Approximately 3 months of effort were allocated for field-testing. Individuals from Eriez Magnetics participated in the testing at Coal Plant D with cooperation from key personnel at the preparation plant.

### 1.6.1 Equipment Setup

The CrossFlow separator was transported from Eriez Magnetics Central Research Lab in Erie, Pennsylvania to the preparation plant. The 9x16 inch pilot-scale CrossFlow separator was installed at the Coal Plant D (as shown in Figure 1.14), with the cooperation from the operators and mechanics at the plant. Feed was supplied to the CrossFlow separator through a 2 inch line connected to the existing coal spiral slurry feed distributor. A slurry splitter fabricated from PVC pipe with a tee and valves was used to regulate the feed to the unit, with the remaining slurry reporting to the spiral circuit. Underflow and overflow material was discharged to sizing screens in the plant, located on a level below the unit.

Plant compressed air and 115 volt electrical power were connected to the separator for the automated control system. The separator was automatically controlled through the use of a simple PID control loop which includes a pressure sensor mounted on the side of the separator to measure the relative pressure (level), a single loop PID controller, and a pneumatic pinch valve to control the underflow discharge to maintain a constant bed pressure (level). Clarified water was connected to the separator to create the fluidized teeter bed of solids.

**Figure 1.14. The 9x16-inch Pilot-Scale CrossFlow Test Circuit at Plant D.**

### 1.6.2 Shakedown Testing

Preliminary shakedown testing was conducted after completing the installation of the test unit to resolve any unexpected operational problems that could arise. These tests are conducted to resolve any problems that may have been overlooked in the initial engineering and to confirm that feed capabilities, pipe sizes, electrical supplies, control systems, etc., are adequate.

### 1.6.3 Detailed Testing

Two series of detailed test programs were conducted using the pilot-scale test unit. The first series of tests were performed to investigate the effects of the key design variables on separator performance and to simultaneously define the overall grade and recovery curve. The subsequent series of testing was performed to investigate the effects of key operating parameters. Tests were conducted primarily as a function of teeter bed pressure and fluidization water rate. The coal/rock interface, or teeter bed, was adjusted to different levels (i.e. different bed pressure) for each steady-state test. Fluidization water was adjusted to fine tune the separation. Other variables considered were solids mass feed rate and volumetric slurry feed rate. For each test, samples were taken from the feed, overflow, and underflow streams after conditions were stabilized. The samples were analyzed for ash and sulfur contents on a size-by-size basis.

### 1.6.4 Process Evaluation

To ensure the test data was reliable and self-consistent, all test data was analyzed and adjusted using mass balance software. Experimental values that were deemed by the mass balance routines to be unreliable were removed from the data set. The participating mining company used the compiled data to establish the metallurgical improvement, operating savings and economic payback that may be realized by implementing the proposed high-efficiency technologies.

Nine test runs were completed during the on-site test work. The parameters of these tests are summarized in Table 1.12. The results from the on-site CrossFlow separator investigation are shown graphically in Figures 1.15 and 1.16. The results are summarized as: "As-Tested" and "x 100 Mesh" with the passing 100 mesh material mathematically removed from the data. This approach is acceptable as it is expected that the clean coal product will be deslimed at approximately 0.150 mm and the fine material upgraded by flotation.

**Table 1.12. Operating Parameters for On-Site Pilot-Scale Testing at Coal Plant D.**

| Test Number | Feed | | | Level | Water |
|---|---|---|---|---|---|
| | % Solids | tph | gpm | inches | gpm |
| 1 | 32.55 | 1.83 | 20 | 14.5 | 8 |
| 2 | 34.44 | 1.95 | 20 | 20.0 | 8 |
| 3 | 35.24 | 2.00 | 20 | 10.0 | 8 |
| 4 | 35.71 | 1.73 | 17 | 10.0 | 9.5 |
| 5 | 32.90 | 2.22 | 24 | 14.5 | 9.5 |
| 6 | 32.71 | 1.84 | 20 | 20.0 | 9.5 |
| 7 | 35.21 | 2.00 | 20 | 20.0 | 6.5 |
| 8 | 34.10 | 1.93 | 20 | 14.5 | 6.5 |
| 9 | 33.55 | 1.89 | 20 | 10.0 | 6.5 |

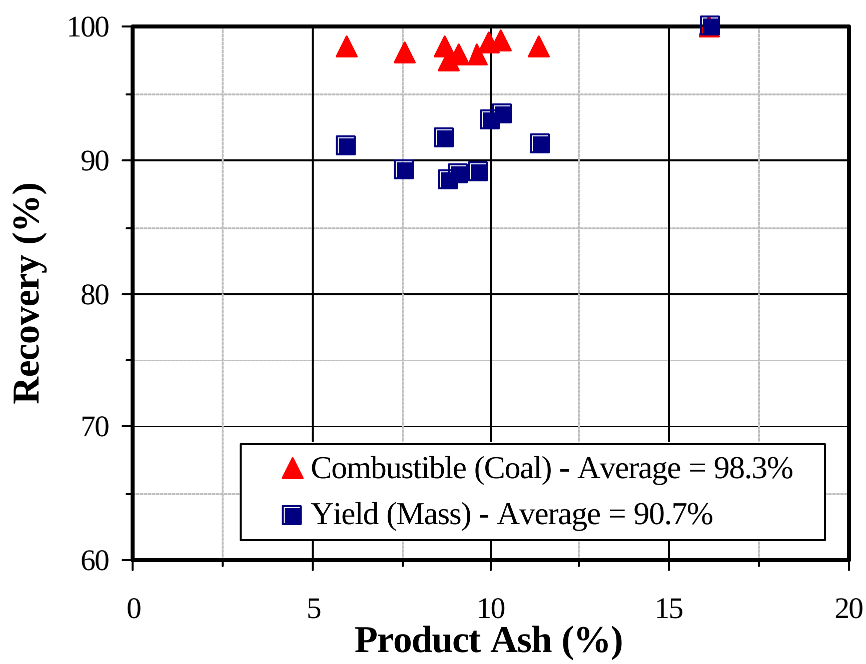



**Figure 1.15. Recovery vs. Product Ash Content of 100 M Coal.**

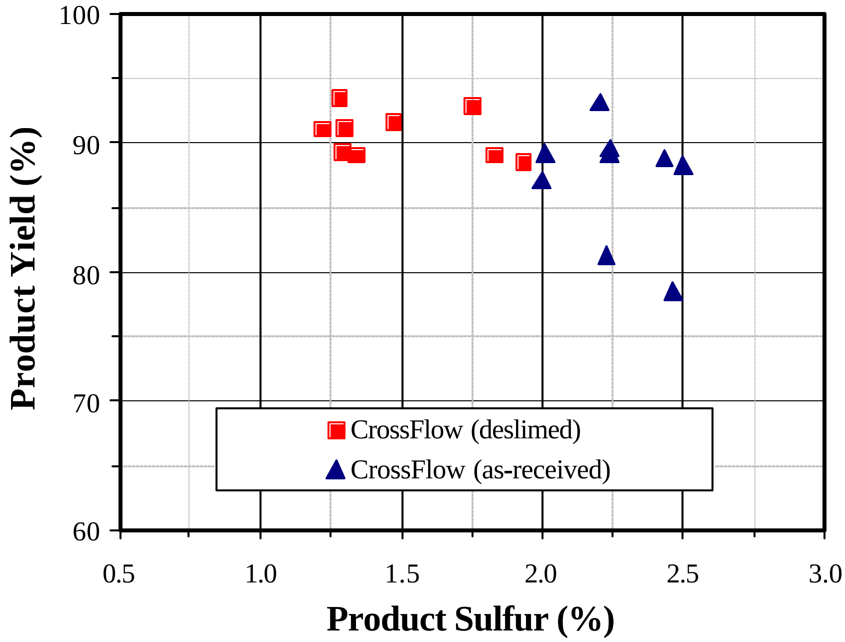



**Figure 1.16. Mass Yield vs. Product Sulfur Content of 100 M Coal.**

As shown in Figure 1.15, this pilot-scale test work was able to define the expected grade and recovery curve. Specifically, the CrossFlow separator is capable of producing a product ranging between 6% and 11% ash at a combustible recovery of greater than 97% (when deslimed at 100 mesh). At maximum separation efficiency, the combustible recovery, for this application, approached 98%. The data presented in Figure 1.16 indicates that the sulfur content of the corresponding product will be approximately 1.50%. Table 1.13 is a summary of test results of the "x 100 Mesh" material for all nine tests conducted during this series.

**Table 1.13. Test Results for x100 Mesh Coal at Coal Plant D.**

| Test No. | Feed Rate (tph) | Ash (%) | Yield (%) | Comb. Recovery (%) | Ash Rejection (%) |
|---|---|---|---|---|---|
| 1 | 1.83 | 9.08 | 88.99 | 97.92 | 53.48 |
| 2 | 1.95 | 11.35 | 91.20 | 98.49 | 42.20 |
| 3 | 2.00 | 9.96 | 92.92 | 98.79 | 39.55 |
| 4 | 1.73 | 8.67 | 91.60 | 98.52 | 47.37 |
| 5 | 2.22 | 5.95 | 91.05 | 98.52 | 58.58 |
| 6 | 1.84 | 7.59 | 89.26 | 98.13 | 57.50 |
| 7 | 2.00 | 10.32 | 93.37 | 98.97 | 37.39 |
| 8 | 1.93 | 9.66 | 89.10 | 97.87 | 51.51 |
| 9 | 1.89 | 8.82 | 88.50 | 97.47 | 54.64 |

The material balance outlined in Figure 1.17 is included as a summary of the test work conducted at the Coal Plant D. This material balance includes all expected metallurgical results, ancillary requirements, and volumetric flows for a full-scale installation with the capacity to treat 175 tph of feed at approximately 50% solids, by weight. A 9x9-ft CrossFlow separator has been recommended for the circuit, offering 81 $ft^2$ of cross-sectional area which results in a normalized feed rate of 2.1 tph/$ft^2$. The current test work has demonstrated the ability of the CrossFlow separator to handle this entire flow in a single-stage circuit.

## CROSSFLOW CIRCUIT - MATERIAL BALANCE

***175.0 stph Option***

**Legend**

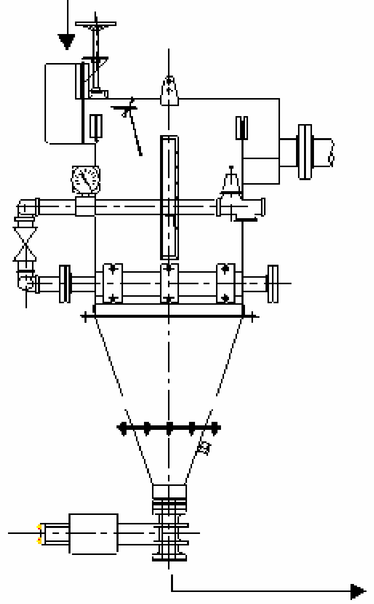
| | | |
|---|---|---|
| tph (solid) | gpm (slurry) | % Solids |
| tph (water) | gpm (water) | Slurry S.G. |

**Feed**

| | | |
|---|---|---|
| 175.0 | 958.5 | 56.1 |
| 136.9 | 547.2 | 1.30 |

**Teeter Water**

| | | |
|---|---|---|
| 0.00 | 526.5 | 0.00 |
| 131.8 | 526.5 | 1.00 |



**Overflow**

| | | |
|---|---|---|
| 140.0 | 1339.0 | 36.3 |
| 245.4 | 980.5 | 1.15 |

**Underflow**

| | | |
|---|---|---|
| 35.0 | 146.0 | 60.0 |
| 23.3 | 93.2 | 1.60 |

| Stream | Assays | | Distributions | |
|---|---|---|---|---|
| | Ash (%) | S.G. | Mass (%) | Comb (%) |
| Feed | | 1.70 | 100.0 | |
| Overflow | | 1.56 | 80.0 | |
| Underflow | | 2.65 | 20.0 | |

| | |
|---|---|
| Number of Cells: | 1 |
| Cell Size (ftxft): | 9.0 |
| Feed Capacity (tph/ft$^2$): | 2.16 |
| Elutriation Rate (tph/ft$^2$): | 1.63 |
| Elutriation Rate (gpm/ft$^2$): | 6.50 |

**Figure 1.17. Expected Material Balance for a CrossFlow Separator Treating 175 tph.**

### 1.6.5 Sample Analysis

Detailed analysis was conducted on each of the samples collected during the testing program. The analyses were performed in accordance with ASTM procedures required by the personnel at Coal Plant D. Representative samples were collected around the pilot-scale unit. Slurry flow rates for the feed, underflow and overflow streams were directly measured using a stopwatch and a calibrated container. The mass and liquid flow rates were then calculated from the measured slurry flow rates and the sample assays using the two-product formula.

### 1.6.6 Future Work

After successful completion of testing at Coal Plant D, the company has agreed to install a prototype of the CrossFlow technology at one of their processing facilities. This commercial installation is expected to be underway by the start of fall 2004.

## 1.7 In-Plant Testing at Coal Plant E

The last set of field tests with the CrossFlow unit were conducted at Coal Plant E. This effort involved equipment setup, shakedown and detailed testing. The goal of this effort was to determine the anticipated product yield and grade, combustible recovery, and feed capacity of the unit for comparison against the existing clean coal effluent cyclones at the plant. The plant personnel desired to classify minus 28 mesh clean coal slurry into plus 100 mesh and minus 100 mesh fractions. Individuals from Virginia Tech participated in the testing at Coal Plant E with cooperation from key personnel at the preparation plant.

### 1.7.1 Equipment Setup

The 9x16 inch CrossFlow separator was transported from the Coal and Minerals Research Lab at Virginia Tech in Blacksburg, Virgina to the preparation plant. With cooperation from the operators and mechanics at the plant, the separator was installed at the plant (see Figure 1.18). Feed was supplied to the separator through a 2 inch line by connecting to a sampling port located on the feed manifold for the existing clean coal effluent cyclones. Underflow and overflow material was discharged to sizing screens in the plant, located on a level below the unit.

Plant compressed air and 115 volt electrical power were connected to the separator for the automated control system. The separator was automatically controlled through the use of a simple PID control loop which includes a pressure sensor mounted on the side of the separator to measure the relative pressure (level), a single loop PID controller, and a pneumatic pinch valve to control the underflow discharge to maintain a constant bed pressure (level). Clarified water was connected to the separator to create the fluidized teeter bed of solids.

**Figure 1.18. The 9x16-inch Pilot-Scale CrossFlow Separator Test Circuit at Plant E.**

### 1.7.2 Shakedown Testing

After completing the installation of the test unit, preliminary shakedown testing was conducted to resolve any unexpected operational problems that could arise. These tests are normally necessary to resolve any problems that may have been overlooked in the initial engineering and to confirm that feed capabilities, pipe sizes, electrical supplies, control systems, etc., are adequate. In addition, the shakedown tests provided an opportunity to roughly determine the ranges of operating conditions that would be most appropriate for this particular application.

### 1.7.3 Detailed Testing

Two series of detailed test programs were conducted using the pilot-scale test unit. The first series of tests were performed to investigate the effects of the key design variables on separator performance and to simultaneously define the overall grade and recovery curve. The subsequent series of testing was used to investigate the effects of key operating parameters. Tests were conducted primarily as a function of teeter bed pressure and fluidization water rate. The coal/rock interface, or teeter bed, was adjusted to different levels (i.e. different bed pressure) for each steady-state test. Fluidization water was adjusted to fine tune the separation. For each test, samples were taken from the feed, overflow, and underflow streams after conditions were stabilized. Five test runs were completed during the on-site test work.

### 1.7.4 Process Evaluation

Due to the low percent solids, the fine size distribution, and the low specific gravity of the material, bed development in the CrossFlow separator was very difficult for this particular application. Initial plans to feed the unit at 1 tph/ft$^2$ could not be obtained due to the turbulence occurring in the bed formation area. Feed rates were slowly reduced over time until a 0.10 tph/ft$^2$ feed rate with a water addition rate of 1.5 gpm produced a stable bed in the unit. Even at this low feed rate, an appreciable amount of plus 100 mesh material was still reporting to the overflow. Five sets of samples were collected of the feed, overflow, and underflow streams. However, laboratory analyses were not conducted on these samples because visual observations of the product streams indicated poor performance at attempting to classify the feed stream.

The coal slurry evaluated in this series of experiments possessed a mean particle size of 0.075 mm. Table 1.14 is a summary of the array of operating parameters that were investigated during testing. The feed slurry specific gravity was measured to be 1.05 with an average of 12% solids. The feed rate was varied from 0.02-0.09 tph/ft$^2$. Due to the poor levels of separation performance, the classification of very fine clean coal slurry using the CrossFlow separator is not recommended at this time.

**Table 1.14 Operating Parameters for On-Site Pilot Scale Testing at Coal Plant E.**

| Unit Operation | Test Number | Feed | | | Level | Water Gpm |
|---|---|---|---|---|---|---|
| | | % Solids | tph | gpm | | |
| CrossFlow | F1 | 12 | 0.09 | 2.8 | 80 | 4.0 |
| CrossFlow | F2 | 12 | 0.09 | 2.8 | 80 | 4.0 |
| CrossFlow | F3 | 12 | 0.05 | 1.6 | 80 | 3.0 |
| CrossFlow | F4 | 12 | 0.04 | 1.2 | 80 | 1.5 |
| CrossFlow | F5 | 12 | 0.02 | 0.5 | 80 | 1.5 |

### 1.7.5 Sample Analysis

Detailed analysis of these samples was not conducted due to the poor results obtained at the plant (as indicated by the visual observations of the solids in the product streams).

### 1.7.6 Future Work

There is currently no future work planned at this location due to the poor preliminary results.

## 1.8 Conclusions

1. A comprehensive study of the CrossFlow separator was conducted at four coal preparation plants on the east coast. In-plant testing of a 9 x 16 inch unit resulted in separation efficiencies at or above existing classification equipment in the size class of 0.2 to 1.0 mm.

2. The data demonstrate that for any given product ash content or sulfur content, the CrossFlow separator can produce a higher clean coal yield and higher combustible recoveries at higher feed rates when compared to the existing coal spirals. The CrossFlow also demonstrated its ability to handle the entire flow of multiple spirals in a single-stage circuit.

3. In the instance where the ultimate goal was to compare results against the existing clean coal effluent cyclones (28 mesh by zero material at 100 mesh), it was determined that the material was too fine to develop the necessary teeter-bed, and the project was therefore abandoned.

4. The test work conducted in this series of tests supports the replacement of spirals with the CrossFlow technology for several applications. As a result, several full scale installations of the unit are being planned in the near future. Based on the successful installation of these full scale units, further implementation of additional units can be utilized in a broad spectrum of companies and industries.

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# IN-PLANT TESTING OF HYDROFLOAT IN THE PHOSPHATE INDUSTRY

## 2.1 Introduction

### 2.1.1 General

Teeter bed technologies can only be applied for gravity concentration when the particles in the feed stream have a relatively narrowly size distribution and moderately large difference in component densities. These units inherently accumulate low density coarse particles at the top of the teeter bed which are too light to penetrate the bed, but at the same time, too heavy to be carried by the rising water into the overflow. As a result, misplacement of low-density, coarse particles to the high-density underflow can occur. This inefficiency can be partially corrected by increasing the elutriation water, to try to carry the low density coarse particles into the overflow. However, this action often causes the fine, high-density particles to also report to the overflow, thereby impacting the quality of the products. As a result, the widespread application of traditional hydraulic separators is greatly limited by these physical constraints.

The limitations of traditional hydraulic separators were recently recognized and overcome through the design of the HydroFloat separator. This technology effectively combines the flexibility of a flotation process with the high capacity of a density separator to overcome barriers that commonly limit conventional teeter bed separators. The HydroFloat can theoretically be applied to any mineral classification system where differences in apparent density can be created by the selective attachment of air bubbles. Figure 2.1 provides a schematic drawing of the HydroFloat separator.

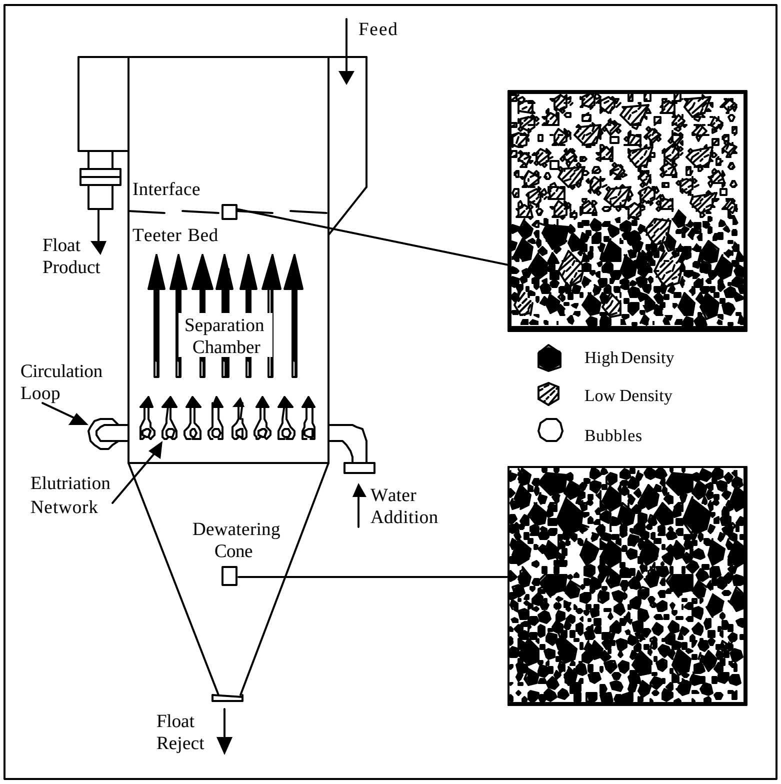


**Figure 2.1. Schematic Drawing of HydroFloat Separator.**

The HydroFloat operates similar to a traditional teeter bed separator with the feed settling against an upward current of fluidization water. However, unlike other hydraulic separators, the HydroFloat utilizes compressed air and a small amount of frothing agent in the fluidization water to produce fine air bubbles. Reagentized feed is introduced into the top of the separator where the feed particles are allowed to settle based on their size and/or density. Previously treated with a collector to make one or more of the minerals hydrophobic, the particles within the separation chamber attach to the small bubbles, reducing their effective density. These lighter bubble-particle aggregates rise through the separation chamber, through the teeter bed and overflow the top of the HydroFloat separator into the product launder. The hydrophilic particles move down through the teeter bed and are eventually discharged through the control valve at the bottom of the separator.

### 2.1.2 Advantages of a Hydraulic Separator

Compared to traditional froth flotation, the use of a fluidized bed within the HydroFloat significantly improves the recovery of particles by (i) reducing turbulence, (ii) enhancing buoyancy, (iii) increasing particle retention time, and (iv) improving bubble-particle contact. The presence of the high-solids teeter bed reduces the turbulence commonly associated in traditional flotation units and therefore enhances the buoyancy of the particles. The teetering effect of the hindered-bed relinquishes the need for bubble-particle aggregates to have sufficient buoyancy to rise to the top of the cell. The low density agglomerates can easily overflow into the product launder, where as the hydrophilic particles move through the teeter bed and eventually discharge through the control valve at the bottom of the separator.

Other benefits of the HydroFloat separator versus traditional froth flotation cells include increases in particle retention time by producing a counter-current flow of particles settling in a hindered state against an upward rising current of water, and the increased probability of bubble-particle contacting in the teeter-bed due to the high-solids content. A higher production rate is possible with the HydroFloat separator than in traditional froth flotation cells due to the high percent solids in the compact teeter bed.

The HydroFloat separator is ideally suited to recover coarse particles that traditional froth flotation cells cannot efficiently recover for several reasons. One reason for the improved recovery of coarse particles is the upward flow of elutriation water in the HydroFloat separator helps lift the larger particles into the product launder. The teeter bed also produces ideal conditions for bubble-particle interactions by maintaining high solids content and quiescent flow conditions. In addition, the high solids content within the teeter-bed separator makes it possible to treat large tonnages in a very compact volume as compared to conventional flotation separations which are conducted at very low solids contents using large volume cells.

### 2.1.3 Project Justification

One of the driving forces behind the HydroFloat separator is the phosphate industry's need to recover coarse particle phosphate (28 x 35 M size fraction) from the feed matrix. It is estimated that 10% of feed material to a Florida phosphate plant is in the plus 35 mesh fraction, which is virtually impossible to recover with present classification equipment. An improvement in coarse particle recovery with the HydroFloat alone corresponds to an additional $7.5-15 million of revenues.

As in the coal industry, the energy benefits of the HydroFloat over conventional equipment are related to the reduction in pumping requirements and water usage which is a direct result of the higher feed ton rate. The lower operating and maintenance cost per ton of product is significantly reduced with the HydroFloat versus conventional equipment. Overall, the implementation of the HydroFloat separator will allow operations to become more profitable and more competitive by utilizing reserves more effectively, reducing waste and increasing productivity.

## 2.2 <u>Literature Review</u>

### 2.2.1 General

The recovery of minerals by flotation is one of the most versatile mineral-processing techniques used in industry today. Flotation methods are utilized throughout the mining industry to treat sulfide ores such as copper, lead and zinc, oxide ores such as hematite and cassiterite and non-metallic ores such as phosphate and coal (Wills, 1992). Since its inception in the early 1900's, improvements in the flotation process have long been a goal within the industry and numerous studies have been financed to overcome the inefficiencies inherent in the process. Industry and government sponsored research programs have focused on all areas of the flotation process to improve recoveries including advancements in chemical reagents, adaptations to existing equipment and introduction of novel equipment.

### 2.2.2 History of Flotation

Although a subject of considerable debate, flotation was believed to be first utilized in the mining industry by T.J. Goover, who in 1909 patented (British Patent No. 27-02-1909) the first multi-cell impeller-type apparatus for froth flotation (Rubinstein, 1995). However, research into the relationship between particle size and floatability did not begin until 1931, when Gaudin, *et al.* (1931) showed that coarse and extremely fine particles are more difficult to recover as compared to intermediate size particles. Twenty years after this original work, Morris (1952) arrived at the same conclusion, that particle size is one of the most important factors in the recovery of ores by flotation. Intermediate size particles will achieve the highest recovery, where as very fine particles ($d_p$<20µm) will have the lowest recovery. In addition, as the particle diameter begins to increase, the recovery will start to decline. This reduction in recovery on the fine and coarse size fractions is indicative of a reduction in the flotation rate of the particles (Jameson, 1977). It can be seen that the efficiency of the froth flotation process deteriorates rapidly when operating in the extremely fine or coarse particle size ranges, which is considered between 10 µm and 200 µm. This is evidence that conventional flotation practices are optimal for the recovery of particles between 65 to 100 mesh.

According to Soto and Barbery (1991), conventional flotation cells operate with two contradictory goals. First, a conventional cell has to provide enough agitation to maintain particles in suspension, shear and disperse air bubbles, and promote bubble-particle collision. However, for optimal recovery, a quiescent system is required to reduce detachment and minimize entrainment. As a result, coarse particle flotation is more difficult since increased agitation is required to maintain particles in suspension. Furthermore, coarse particles are more likely to detach under turbulent conditions. To compensate for the lack of recovery, some installations are using relatively small flotation devices operated at low feed rates (Lawver, 1984).

As particle size is reduced, two dominating characteristics begin to emerge, i.e., the specific surface becomes large and the mass of the particle becomes very small (Abdel-Khalek, *et al*., 1990). These are the dominating factors affecting fine particle recovery in flotation systems. Virtually all ores are associated with a clay mineral, which is ultimately transferred to the preparation plant with the mineral of interest. The clay minerals associated with the fine fractions will reduce mineral recovery by inhibiting bubble-particle attachment, and consuming flotation reagents.

The variety of flotation machines available on the market today can be classified into two distinct groups: pneumatic and mechanical machines (Wills, 1992). Pneumatic machines commonly utilize air that is blown in or induced, where it must be dissipated through a series of baffles or some form of permeable base within the cell. Since air is used not only to produce the froth and create aeration but also to maintain the suspension and to circulate it, an excessive amount is usually introduced (Wills, 1992). Complications directly related to the excessive amount of air limited the use of pneumatic machines until the development of the flotation column.

Mechanical flotation machines are the most common and widely used flotation machine on the market today. The units are characterized by a mechanically driven impeller which agitates the slurry and disperses the incoming air into small bubbles (Wills, 1992). Air addition into the cell can either be forced through an external blower, or self-aerating. Typically most mechanical flotation cells are set up in a series of "banks", where several cells will allow free flow from one cell to the next down the bank.

Performance is generally based on three factors including: (i) metallurgical performance, i.e., product recovery and grade, (ii) capacity, and (iii) operating and maintenance costs (Wills, 1992). An analysis of the effectiveness of the various types of flotation machines was made by Young (1982), who discusses performance with regard to the basic objectives of flotation, which are the recovery of the hydrophobic species into the froth product, while still achieving a high selectivity by retaining as much as the hydrophilic species as possible in the slurry. Recovery is directly related to particle-bubble attachment and requires quiescent conditions, which is not found in conventional mechanical flotation devices. The mechanical impellers found in typical flotation cells are not ideal for particle-bubble contact, which has led the industry to utilize column cells for a variety of mineral applications that, up until the past decade or two, was unheard of.

Column cells are considered to be ideal displacement machines, where as mechanical cells are ideal mixers (Wills, 1992). A column cells improves flotation performance by minimizing turbulence within the cell and reducing entrainment using froth washing. In 1914, G.M. Callow patented the first apparatus with air sparging through a porous false bottom, (Rubinstein, 1995), which would become the basis for future column cell designs. By 1919, M. Town and S. Flynn had developed the first design involving a countercurrent of slurry and air within a column. While pneumatic Callow apparatuses were very popular in the early 1920's and 30's, the lack of technological progress in the area of reliable pneumatic air spargers and lack of process control systems forced the introduction of impeller-type apparatuses. It wasn't until the mid 1960's that column cells began to be intensively developed and extensively introduced into the industry, when practically all the work on updating other types of flotation cells ceased (Rubinstein, 1995).

The advantages of column cell technology over conventional mechanical cells are directly related to the direction of flow of the slurry and air. The counter-current regime provides for more ideal bubble-particle attachment and enhanced aggregate stability. The likelihood of bubble-particle detachment is minimized due to low turbulence of slurry flows within the column. These benefits have prompted the phosphate industry to implement column flotation cells into the industry for fine and coarse particle flotation.

### 2.2.3 Phosphate Flotation

Phosphate beneficiation plants are designed to process run-of-mine ore, typically called the ore matrix, into a sellable product for use in either the fertilizer market or as an integral part or the production of phosphoric acid. The ore matrix is upgraded by separating the phosphate grains from other impurities such as clay and silica. Beneficiation plants in the southeastern United States (Florida and North Carolina) generally use sizing and classification processes to concentrate the phosphate rock and separate it from impurities.

Florida beneficiation plants typically wash and deslime the ore matrix at 150 mesh. The material finer than 150 mesh is considered tailings and is pumped to settling ponds. Approximately 30% of the phosphate contained in the original ore matrix is lost to the tailings ponds. The remaining rock is separated into three size classes, a pebble size fraction, coarse and fine size fractions. The pebble is a high phosphate content rock (-3 ¼ x 14 mesh) that requires no further processing. The coarse size fraction (14 x 35 mesh) and fine size fraction (35 x 150 mesh) are treated separately in different flotation circuits.

Historically, fine phosphate flotation is an efficient process with recoveries from conventional froth flotation in excess of 90% for most ores. Recoveries will vary depending on the ore type, with recoveries dropping slightly for some high manganese or dolomitic ores. In contrast, froth flotation recoveries for coarse phosphate are generally much lower than those of fine phosphate ores. Typical recoveries for coarse flotation are less than 50%. Historically, hammer mills were used for size reduction, but due to high maintenance costs and loss of fines, this practice has been discontinued (Soto, 1992).

The industry, however, has taken other approaches to circumvent the problem of low floatability of coarse particles. For instance, such approaches are exemplified by the use of gravitational devices such as spirals, tables, launders, sluices and belt conveyors modified to perform a "skin flotation" of the reagentized pulp. Although a variable degree of success is obtained with these methods, they have to be normally supplemented by scavenger flotation. In addition, some of them require excessive maintenance, have low capacity, or involve high operating costs. Their performance is less than satisfactory and in certain cases their use has been discontinued.

Previous laboratory and pilot-scale testing of the HydroFloat separator has proven its capabilities as an effective flotation device for recovering fine and coarse phosphate. The unit has especially proven successful in the 35 mesh fraction of the phosphate ore matrix in Florida. This size fraction was previously discarded to the tailings when detachment and buoyancy limitations in traditional flotation methods failed to recover the material.

## 2.3 <u>In-Plant Testing at Phosphate Plant A</u>

The Phase I field-testing of the HydroFloat separator involved equipment setup, shakedown and detailed testing at the Phosphate Plant A. The goal of this effort was to compare the unit to existing conventional cells in several different areas of the plant by analyzing the anticipated product grade and recovery, insol content, reagent consumption, and feed capacity at, and above, design feed rates of the unit. The three areas of the plant where the HydroFloat separator was tested included the fine feed, amine flotation and coarse feed circuits.

The main objective of the fine and coarse phosphate testing was to demonstrate the potential of the unit as a candidate for the process equipment in a proposed plant design with both fine and coarse circuits. The main objective of the amine flotation testing was to demonstrate the feasibility of using the unit for silica flotation and to develop data to determine its potential application for use in the amine flotation circuit at Phosphate Plant A. Approximately 6 months was allocated to this task.

Individuals from Eriez Magnetics and Virginia Tech participated in the testing at Phosphate Plant A with cooperation from key personnel at the processing plant. Additional tests were conducted by Phosphate Plant A representatives to expand the data base for evaluating the potential of incorporating the HydroFloat separator into proposed circuit upgrades.

### 2.3.1 Equipment Setup

#### 2.3.1.1 Fine Circuit

The installation of the pilot-scale unit in the fine feed circuit at Phosphate Plant A was the main objective of this task. The separator was transported from the Eriez Magnetics Central Research Lab in Erie, Pennsylvania to the processing plant. With cooperation from the operators and mechanics at the plant, the 18-inch diameter pilot-scale HydroFloat separator was installed at the fine circuit at Phosphate Plant A as shown in Figure 2.2. Reagentized feed was supplied to the HydroFloat separator through a 2 inch line connected to the existing plant conditioning tanks. Concentrate and tailings streams were discharged into floor sumps.

The unit was operated as a column flotation cell, utilizing the HydroFloat separator air sparging system. The test unit included 3 compartments that allowed more water and air to be added (up to 60 gpm water and 10 cfm air). There was no teeter-bed required in this system. Plant compressed air and 115 volt electrical power were connected to the separator for the automated control system. The separator was automatically controlled through the use of a simple PID control loop which includes a pressure sensor mounted on the side of the separator to measure the relative pressure (level), a single loop PID controller, and a pneumatic pinch valve to control the underflow discharge to maintain a constant bed pressure (level).

**Figure 2.2. The 18-inch Diameter Pilot-Scale HydroFloat Separator Test Circuit at Plant A.**

### 2.3.1.2 Amine Circuit

The same separator used in the fine circuit was also used in the amine flotation circuit. With cooperation from the operators and mechanics at the plant, the 18-inch diameter pilot-scale HydroFloat separator was installed in the amine circuit at Phosphate Plant A. Reagentized feed was supplied to the HydroFloat separator through a two-inch line connected to the existing plant conditioning tanks. Concentrate and tailings streams were discharged into floor sumps.

The unit was operated as a column flotation cell, utilizing the HydroFloat separator air sparging system. The test unit included 3 compartments that allowed more water and air to be added (up to 60 gpm water and 10 cfm air). There was no teeter-bed required in this system. Plant compressed air and 115 volt electrical power were connected to the separator for the automated control system. The separator was automatically controlled through the use of a simple PID control loop which includes a pressure sensor mounted on the side of the separator to measure the relative pressure (level), a single loop PID controller, and a pneumatic pinch valve to control the underflow discharge to maintain a constant pressure (level).

#### 2.3.1.3 Coarse Circuit

The same separator used in the fine and amine flotation circuits was also used in the coarse circuit, with one modification. The center compartment was removed from the unit, so as to allow the unit to operate with a typical teeter-bed (a total of 2 compartments). With cooperation from the operators and mechanics at the plant, the 18-inch diameter pilot-scale HydroFloat separator was installed in the coarse circuit at Phosphate Plant A. Reagentized feed was supplied to the HydroFloat through a 2-inch line connected to existing plant conditioning tanks. Concentrate and tailings streams were discharged into floor sumps.

Electrical power at 115 volt and plant compressed air were connected to the separator for the automated control system. The separator was automatically controlled through the use of a simple PID control loop which includes a single loop PID controller, a pressure sensor mounted on the side of the separator to measure the relative pressure, and a pneumatic pinch valve to control the underflow discharge to maintain a constant bed pressure.

### 2.3.2 Shakedown Testing

Preliminary shakedown testing was conducted after completing the installation of the test HydroFloat unit to resolve any unexpected operational problems that could arise. These tests are normally necessary to resolve any problems that may have been overlooked in the initial engineering and to confirm that feed capabilities, pipe sizes, electrical supplies, control systems, etc., are adequate. An average of six shakedown tests per circuit was conducted with the unit.

### 2.3.3 Detailed Testing

Two series of detailed test programs were conducted using the pilot-scale test unit. The first series of test were performed to investigate the effects of the key design variables on separator performance and to simultaneously define the overall grade and recovery curve.

The HydroFloat separator is designed for feed rates of 2 tph/ft$^2$ and 1 tph/ft$^2$ rougher concentrate, which allows the test unit to operate at 4 tph feed and 2 tph concentrate, respectively. The initial testing in the fine and coarse circuit evaluated the unit at loading rates much higher than design to establish the *recovery fall-off*. The design rates for the amine flotation circuit were not precisely known going into the testing, but were thought to be similar to those for rougher flotation. Part of the amine testing program was devoted to determining the design rates and evaluating the HydroFloat separator performance across the board, both at the design rate and above.

With the *recovery fall-off* determined for each circuit and unit configuration, the subsequent series of testing was used to investigate the effects of key operating parameters. Tests were conducted to establish reagent consumption (fatty acid, surfactant, amine and diesel oil), to investigate the bed levels and sparger water required for the best unit operation and to investigate the variability associated with the overall system. For each test, samples were taken from the feed, concentrate and tailings streams after conditions were stabilized. The samples were analyzed for BPL, MgO and insol contents.

### 2.3.4 Process Evaluation

All as-received results were analyzed and adjusted using mass balance software to ensure the test data was reliable and self-consistent. Any experimental values that were deemed by the mass balance routines to be unreliable were removed from the data set. The participating mining company used the compiled data to establish the metallurgical improvement, operating savings and economic payback that may be realized by implementing the proposed high-efficiency technologies.

The process evaluation has been divided into three sections including (i) fine feed circuit, (ii) amine flotation circuit, and (iii) the coarse feed circuit.

#### 2.3.4.1 Fine Feed Circuit

Fifty-three tests were conducted during the fine circuit testing at Phosphate Plant A. Testing in the fine circuit produced an average of 10% higher BPL recoveries with a 0.8% lower BPL rougher tail in the HydroFloat separator than in the plant Wemco cells. Figure 2.3 displays the HydroFloat separator and plant tails percent BPL for each test. The plant Wemco cells averaged only about 0.7% BPL higher-grade rougher concentrates than the HydroFloat as shown in Figure 2.4. An average HydroFloat separator rougher concentrate grade of 54.9% BPL is satisfactory considering the test feed grade only average 8% BPL through most of the testing.

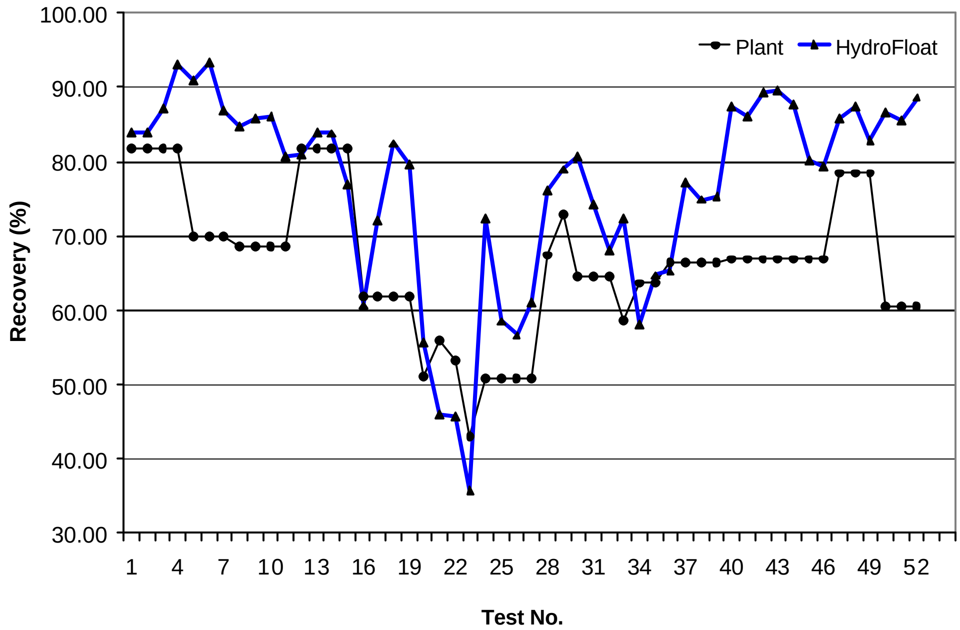


**Figure 2.3 - Recovery of HydroFloat Separator versus Plant Cells.**

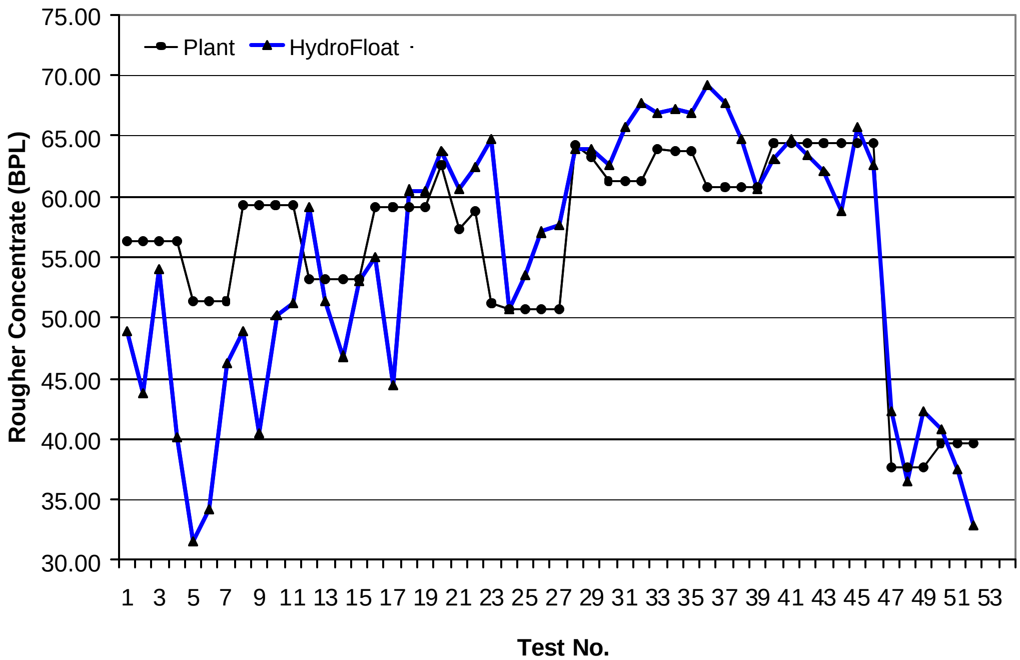


**Figure 2.4. Rougher Concentrate Grade of HydroFloat Separator versus Plant Cells.**

During testing, several attempts were made to obtain final grade concentrates (7% insol) with one stage of flotation. The results show that insol concentrates between 9-10% produced only 74-76% recoveries, and dropping the insol to 7-8% reduced the recoveries to 70% or less. Further testing in this area needs to be conducted utilizing more selective reagents or higher feed grades to achieve the desired 7% insol concentrates in a one step flotation process with the HydroFloat separator.

One of the most important operating parameter to consider for fine flotation is the ability of the process equipment to recover coarser material into an acceptable concentrate: i.e., recover coarse phosphate without recovering fine silica. Comparison testing of the HydroFloat separator with the Wemco Cell produced promising results. As shown in Figure 2.5, the HydroFloat separator recovered 80%, 83%, and 88% of the plus 20 mesh, 20 x 28 mesh, and 28 x 35 mesh phosphate, respectively. The performance values were well above those established for the plant; the plant recovered only 24% of the plus 35 mesh and 67% of the plus 48 mesh phosphate.

Percent solids in the tailings averaged between 20-30% at optimum testing conditions. During less than optimum conditions, the solids were as high as 53%. Optimum conditions occurred at 70-75 bed levels, with between 50-60 gpm of sparger water, and 4 tph feed. While higher bed levels and less sparger water could produce a slightly higher percent solids in the tailings, this adversely affected the recovery and concentrate grades. Using the unit with 3 compartments and with bed levels of 70-75, the optimum froth depths were 15-20 inches.

Reagent dosages were affected by the poor water quality and excessive slimes in the feed during the testing program. The fatty acid dosage in the plant ranged from 0.80 to 1.20 lb per ton of fine feed during testing, whereas the fuel oil dosage in the plant ranged from 0.35 to 0.55 lb

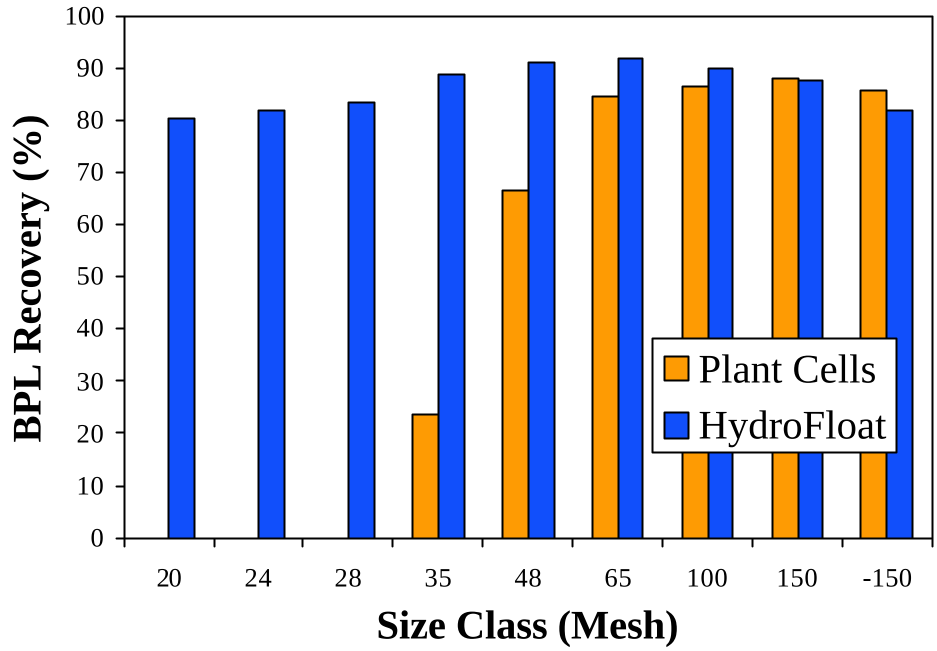


**Figure 2.5. Comparison of Test Results for Fine Phosphate (Plant Circuit #2).**

per ton fine feed. The fuel oil dosage was slightly higher than average dosage at Phosphate Plant A, which is partially to blame for the poorer than expected recoveries.

The recommended surfactant dosage was 0.13 lb per ton at design rates, with actually results being slightly higher. Dosage in the HydroFloat separator ranged from 0.20 to 0.32 lb per ton of feed (6.9 to 10.4 cc per minute). Projected surfactant dosage for the fine circuit can not be determined at this time, but it is estimated that it is just slightly higher than the recommended dosage.

While the operation of the HydroFloat separator for fine flotation was difficult to optimize due to various outside variables affecting the system, a significant number of tests were conducted at differing operating variables under varying operating conditions to achieve optimum operating conditions. The optimum conditions for the HydroFloat separator for use in fine flotation as defined by this testing program are: 3 compartment unit, with bed level between 70-75, a froth depth of 15-20 inches, sparger water between 55-60 gpm, air flow of 10 cfm, and a surfactant dosage of at least 0.2 lb per ton of feed. The measured recovery values and concentrate grade at these design rates were acceptable. Based on this data, the HydroFloat separator can successfully be implemented into the Phosphate Plant A fine flotation circuit.

#### 2.3.4.2 Amine Circuit

Twenty-four tests were conducted during the amine flotation circuit testing at Phosphate Plant A. HydroFloat separator testing in the amine flotation circuit produced an average of 1.3% higher insol concentrate and recovered about 8% less insol to the amine tailings than in the Plant Wemco Cell. Figure 2.6 displays the concentrate grade for the HydroFloat separator and the plant for each test. The plant Wemco cells averaged only about 0.5% higher BPL recovery than the HydroFloat separator as shown in Figure 2.7.

The HydroFloat separator performed virtually the same as the plant Wemco cell for amine flotation over the range 3 to 18% concentrate insol and 95 to 99% BPL concentrate recovery. The unit demonstrated it could effectively recover coarse silica. The HydroFloat separator insol recovery values were about 3% lower on average than those in the plant at above design feed rates. The differences ranged from 6% to 11% in the 35 mesh and 48 mesh fractions to 2% in the finer fractions. The HydroFloat separator insol recovery values were about 2% higher on average than those in the plant at the lower feed rates.

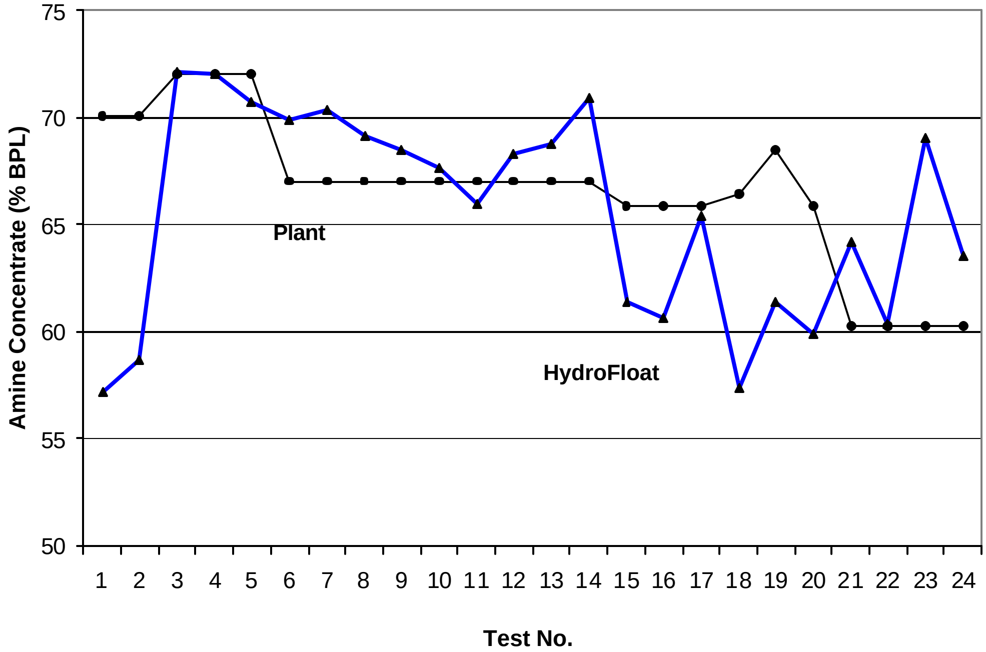


**Figure 2.6. Amine Concentrate Grade Comparison of HydroFloat Separator vs. Existing Plant Cells.**

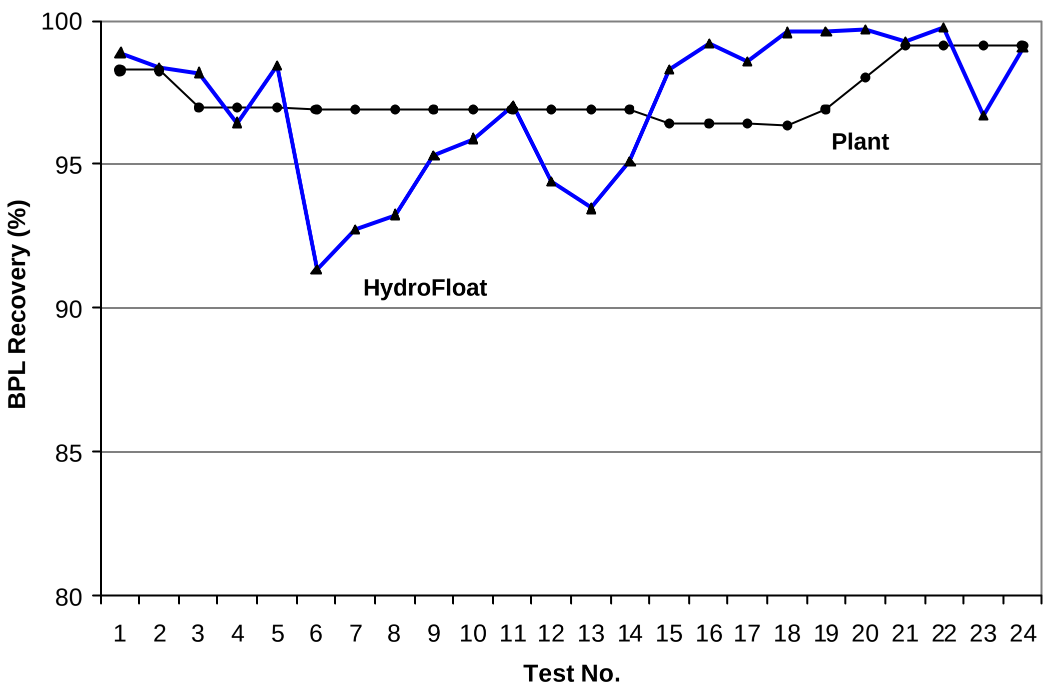


**Figure 2.7. BPL Recovery Comparisons HydroFloat Separator vs. Existing Plant Cells.**

One of the most important operating parameters to consider for amine flotation is the ability of the process equipment to recover coarse silica without recovering phosphate. Comparison testing of the HydroFloat separator with the Wemco Cell produced promising results. As shown in Figure 2.8, the HydroFloat separator had just slightly less recoveries than the plant for all of the size fractions except the 35 mesh, where it had a nearly 6% increase in BPL recovery than the plant.

Reagent dosages were affected by the poor water quality and excessive slimes in the feed during the testing program. The surfactant dosage for the HydroFloat separator ranged from 0.13 to 0.40 lb per ton of feed. The recommended dosage was 0.14 lb per ton at design rates.

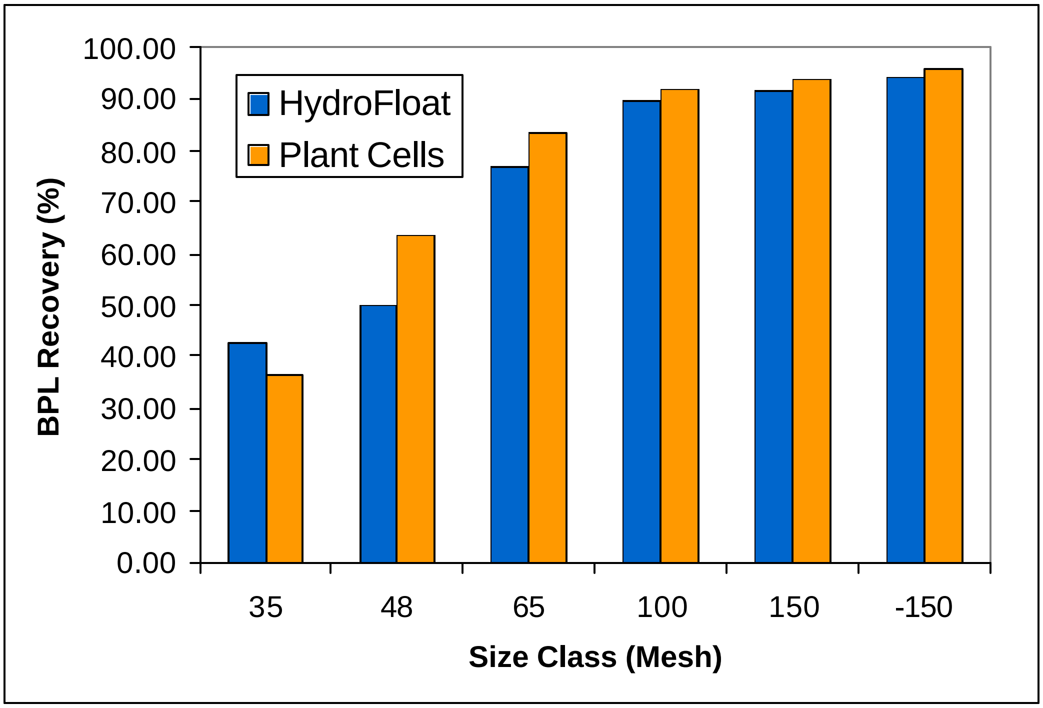


**Figure 2.8. Comparison of Test Results for Amine Phosphate (Plant Circuit #2).**

The interactions of varying diesel fuel dosage rates were studied during the amine circuit testing. Amine flotation circuits use diesel oil or polymer occasionally to modify the froth when slimy water is present. Froth stability was investigated, but was difficult to determine due to the lack of air flow measurement available at the time of testing. Exact diesel fuel dosage rates are unknown at this time.

While the operation of the HydroFloat separator for amine flotation was difficult to optimize due to various outside variables affecting the system, a significant number of tests were conducted at differing operating variables under varying operating conditions to achieve optimum operating conditions. The optimum conditions for the HydroFloat separator for use in amine flotation as defined by this testing program are: 3 compartment sections, with bed level between 70-75, a froth depth of 15-20 inches, sparger water at 25 gpm, air flow of 10 cfm, and a surfactant dosage of at least 0.2 lb per ton of feed. Additional testing will be needed in the future to validate these recommendations. The measured silica recovery values and concentrate grades at these design rates were acceptable. Based on this data, the HydroFloat separator can successfully be implemented into the Phosphate Plant A amine flotation circuit.

#### 2.3.4.3 Coarse Circuit

Twenty-four tests were conducted during the coarse circuit testing at Phosphate Plant A. Testing in the coarse circuit produced an average 12% higher BPL recovery with a 3.5% lower BPL rougher tail in the HydroFloat separator than in the Plant Wemco Cell. Figure 2.9 displays the HydroFloat separator and plant tails percent BPL for each test. Figure 2.10 displays the concentrate recovery for the HydroFloat separator and Plant Wemco Cell. The plant average about 6% BPL higher-grade rougher concentrates than the HydroFloat separator as shown in Figure 2.11. However, the average concentrate grade of 62.6% BPL was still considered satisfactory for the testing.

As with the fine and amine flotation circuits testing, poor water quality played an important role in the overall performance of the reagents during testing. Fatty acid dosage in the plant ranged from 2.04 to 3.61 lb per ton of coarse feed during testing, while fuel oil dosage ranged from 1.06 to 1.68 lb per ton of feed. Both of these values are considered high for Phosphate Plant A, and hindered recoveries as a result.

Surfactant dosage for the HydroFloat ranged from 0.23 to 0.77 lb per ton of feed, which was also considered to be a high dosage, mostly attributable to the high fatty acid-fuel oil dosage in the plant. Other contributing factors were the poor water quality and the need to set the surfactant dosage rates higher than normal in the plant to maintain an adequate froth bed depth. (This was the case in the fine and amine flotation circuits testing as well.)

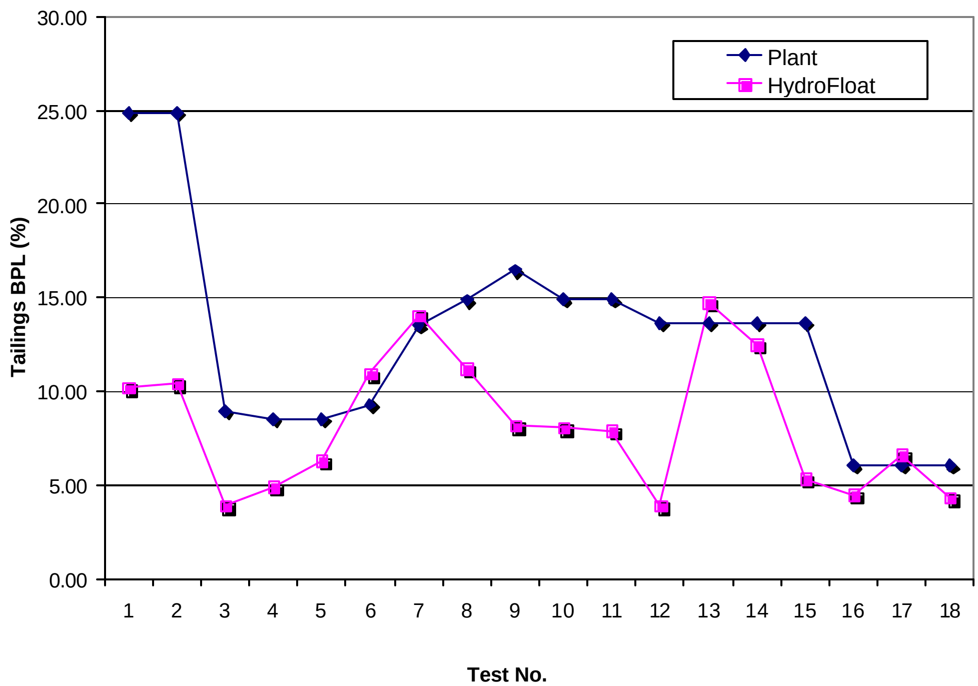


**Figure 2.9. Tailings Comparison of HydroFloat Separator versus Plant Cells.**

The ability of the unit to recover coarse material into an acceptable concentrate proved to be successful during the testing program. One test achieved an overall BPL of 92% at a feed rate of 3.92 tph (98% of design) and a concentrate overflow froth rate of 1.56 tph (78% of design). The associated concentrate grade was 61% BPL.

Screen and chemical analyses were conducted on selected tests to determine the recovery values for various mesh sizes. The HydroFloat separator recovery values are considered to be excellent as shown in Figure 2.12.

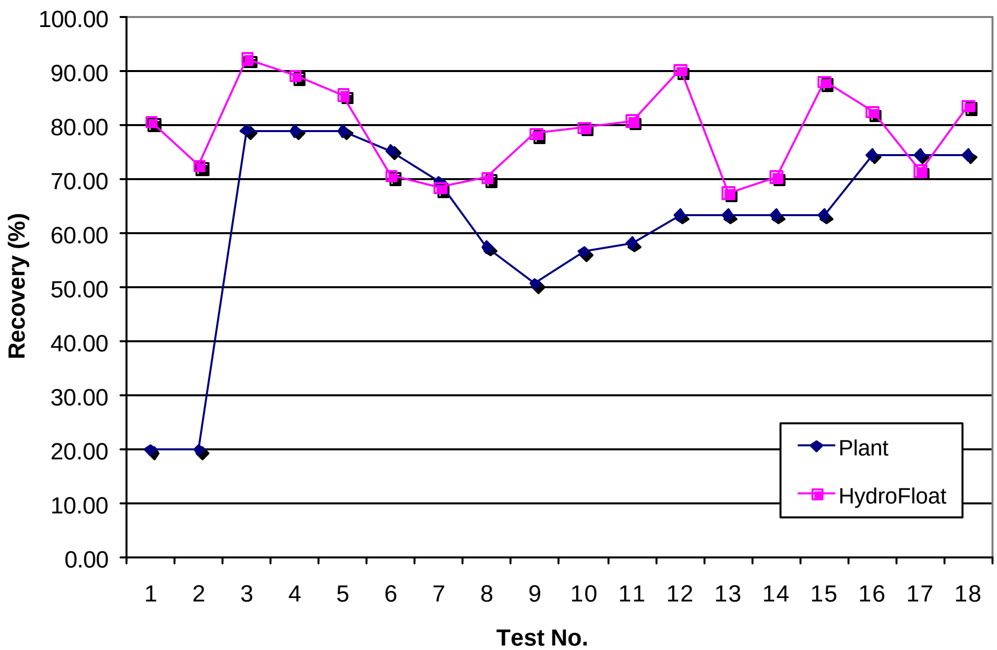


**Figure 2.10. Recovery Comparison of HydroFloat Separator vs. Existing Plant Cells.**

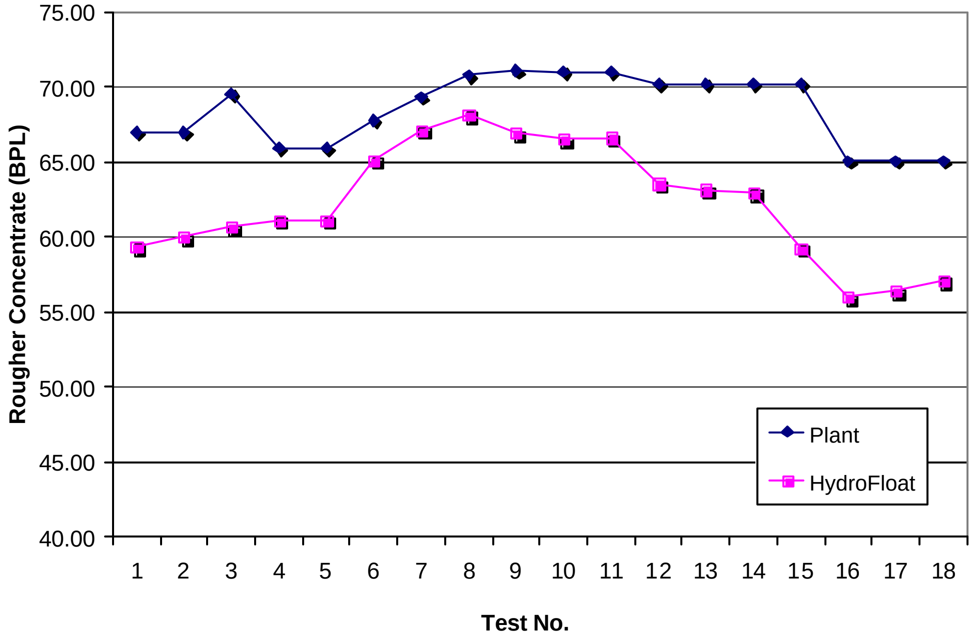


**Figure 2.11. Grade Comparison of HydroFloat Separator vs. Existing Plant Cells.**

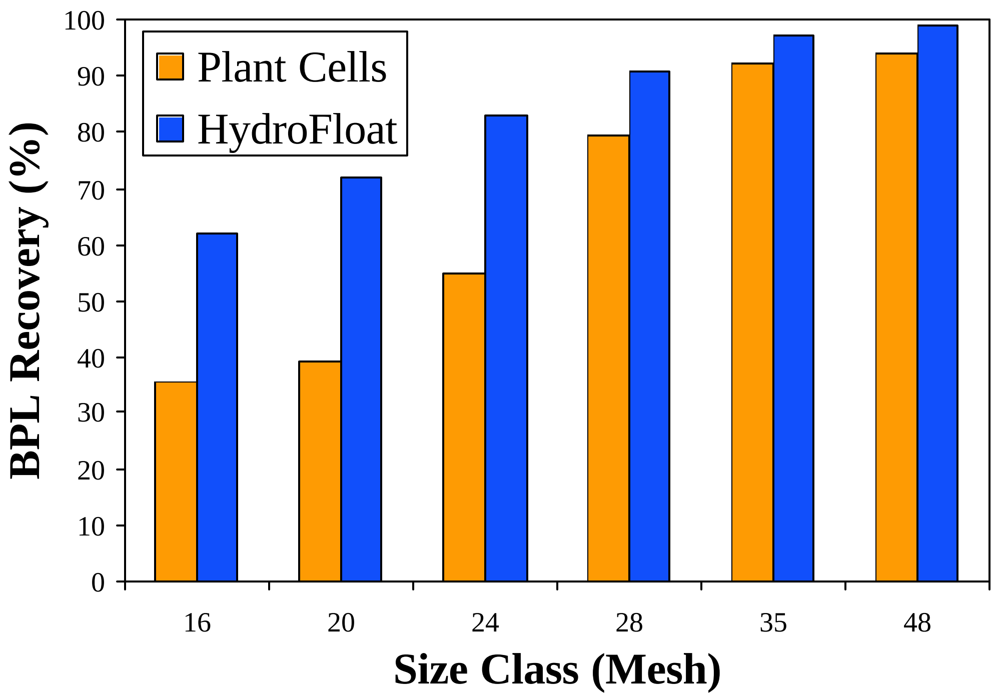


**Figure 2.12. Comparison of Test Results for Coarse Phosphate (Plant Circuit #1).**

Percent solids in the tailings averaged 75.8% for all tests. The HydroFloat separator was configured with 2 compartments, with bed levels between 82 and 87, and with a recommended level of 85. This resulted in optimum condition of: froth depths between 15 and 20 inches, sparger water near 20 gpm, and air flow at 5.0 cfm. The measured recovery values and concentrate grade at these design rates were acceptable. Based on this data, the HydroFloat can successfully be implemented into the Phosphate Plant A coarse flotation circuit.

### 2.3.5 Sample Analysis

Detailed analysis was conducted on each of the samples collected during the testing program. The analyses were performed in accordance with ASTM procedures onsite at the Phosphate Plant A. Representative samples were collected around the pilot-scale unit. Slurry flow rates for the feed, concentrate and tailings streams were directly measured using a stopwatch and a calibrated container. The mass and liquid flow rates were then calculated from the measured slurry flow rates and the sample assays using the two-product formula.

### 2.3.6 Future Work

While the results of the testing look promising, the project has been temporarily been put on hold for administrative reasons.

## 2.4 <u>In-Plant Testing at Phosphate Plant B</u>

Equipment setup, shakedown testing, and detailed testing comprised the phase I field-testing of the HydroFloat separator at Phosphate Plant B. The goal of this effort was to compare the unit to existing hydroclassifiers and conventional cells by analyzing the anticipated product grade and recovery, insol content, reagent consumption and feed capacity at, and above, design feed rates of the unit. The main objective of testing was to determine if the HydroFloat separator could achieve higher recoveries of the ultra-coarse particles than the existing second-stage hydroclassifer at the plant. Further investigations of the coarse and fine matrices were conducted, comparing results against the existing conventional cells currently in operation at the plant. Approximately 12 months was allocated to this task. Individuals from Eriez Magnetics and Virginia Tech participated in the testing at Phosphate Plant B with cooperation from key personnel at the processing plant.

### 2.4.1 Equipment Setup

The separator was transported from the Eriez Magnetics Central Research Lab in Erie, PA to the processing plant. With cooperation from the operators and mechanics at the plant, the 1-foot diameter pilot-scale HydroFloat separator was installed at each circuit (ultra-coarse, coarse and fine) for a period of several weeks for each circuit at Phosphate Plant B as shown in Figure 2.13. Reagentized feed was supplied to the HydroFloat separator through a 2-inch line connected to the existing plant conditioning tanks. Concentrate and tailings streams were discharged into floor sumps.

Plant compressed air and 115 volt electrical power were connected to the separator for the automated control system. The separator was automatically controlled through the use of a simple PID control loop which includes a pressure sensor mounted on the side of the separator to measure the relative pressure (level), a single loop PID controller, and a pneumatic pinch valve to control the underflow discharge to maintain a constant bed pressure (level). Clarified water was connected to the separator to create the fluidized teeter bed of solids.

**Figure 2.13. Pilot-Scale HydroFloat Separator Test Circuit at Plant B.**

### 2.4.2 Shakedown Testing

After completing the installation of the test HydroFloat unit in each circuit, preliminary shakedown testing was conducted to resolve any unexpected operational problems that could arise. Shakedown test are commonly utilized to resolve any problems that may have been overlooked in the initial engineering and to confirm that feed capabilities, pipe sizes, electrical supplies, control systems, etc., are adequate.

### 2.4.3 Detailed Testing

Two series of detailed test programs were conducted for each circuit using the pilot-scale test unit. The first series of test were performed to investigate the effects of the key design variables on separator performance and to simultaneously define the overall grade and recovery curve.

The HydroFloat separator is designed for feed rates of 2 tph/sqft and 1 tph/sqft rougher concentrate, which allows the test unit to operate at 4 tph feed and 2 tph concentrate, respectively. The initial testing in the coarse circuit evaluated the unit at loading rates much higher than design, to establish the *recovery fall-off*.

With the *recovery fall-off* determined for each circuit and unit configuration, the subsequent series of testing was used to investigate the effects of key operating parameters. Tests were conducted to establish reagent consumption (fatty acid, surfactant, and diesel oil), to investigate the bed levels and sparger water required for the best HydroFloat separator operation, and to investigate the variability associated with the overall system. For each test, samples were taken from the feed, concentrate and tailings streams after conditions were stabilized. The samples were analyzed for BPL, MgO, and insol contents.

### 2.4.4 Process Evaluation

To ensure the test data was reliable and self-consistent, all as-received results were analyzed and adjusted using mass balance software. Experimental values that were deemed by the mass balance routines to be unreliable were removed from the data set. The participating mining company used the compiled data to establish the metallurgical improvement, operating savings and economic payback that may be realized by implementing the proposed high-efficiency technologies.

The process evaluation has been divided into three sections including the (i) ultra-coarse rock feed, (ii) the coarse rock feed, and (iii) the fine feed circuits.

#### 2.4.4.1 Ultra Coarse Feed

Grade versus recovery data for the in-plant evaluation of the HydroFloat had BPL recoveries of 87% to 99% with product grades ranging between 5% and 14% insols. The resulting products contained, on average, 67% BPL. Figure 2.14 is a graph of the grade versus recovery data for the in-plant testing and earlier laboratory-scale testing. Size-by-size analysis of the HydroFloat was conducted and results are presented in Figure 2.15. The HydroFloat is capable of high BPL recoveries for even the coarsest size fractions, where 96.7% of the available BPL in the +16 mesh size class was recovered.

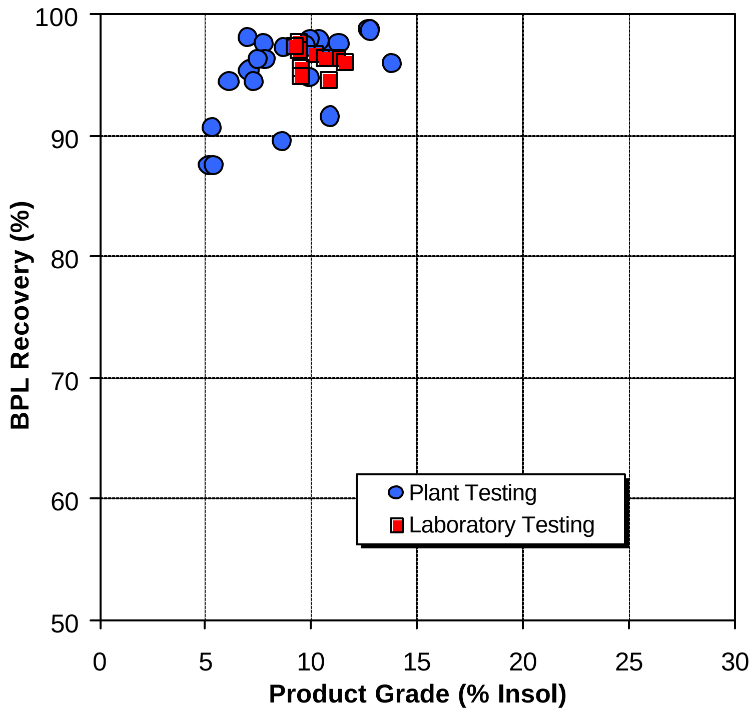


**Figure 2.14. BPL Recovery vs. Product Insol Grade for Ultra-Coarse Rock Feed at Phosphate Plant B.**

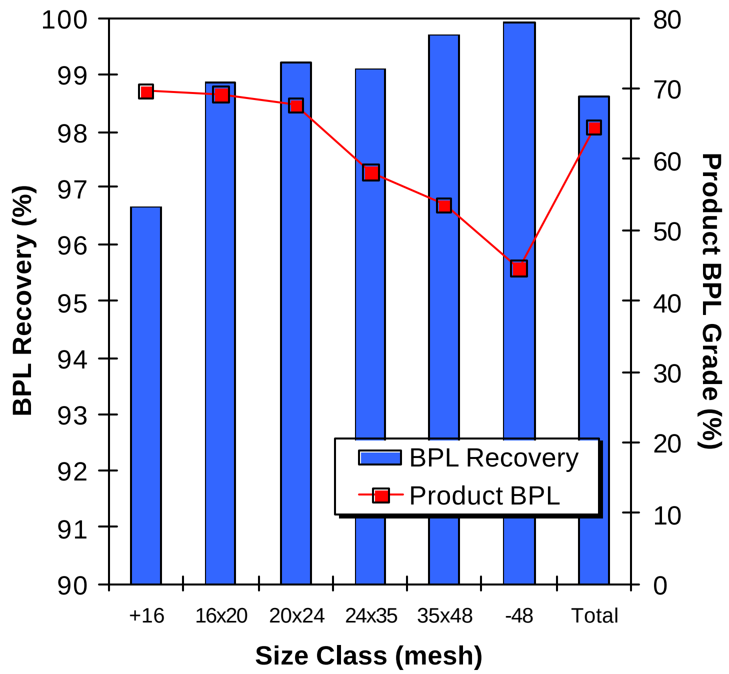


**Figure 2.15. Size-by-Size Recovery and Grade for Ultra-Coarse Rock Feed.**

### 2.4.4.2 Coarse Feed

Figure 2.16 summarizes the grade and recovery data for the coarse feed test work. BPL recoveries ranged from 90% to 98% while product grades averaged 24.7% insols. The resulting products contained, on average, 55% BPL by weight. Figure 2.16 also illustrates that the results for the laboratory evaluations were superior to those produced for the in-plant trials. This occurrence is a direct result of the mean particle size difference found between the samples used for the laboratory and in-plant testing. It was calculated that the sample used for the coarse matrix laboratory testing was as coarse (mean size: 0.706 mm) as the sample provided for the ultra-coarse testing (mean size: 0.721 mm). During the in-plant trials, it was observed that the coarse matrix was significantly finer, amplifying any occurrence of hydraulic carry-over or activation of fine floatable insols.

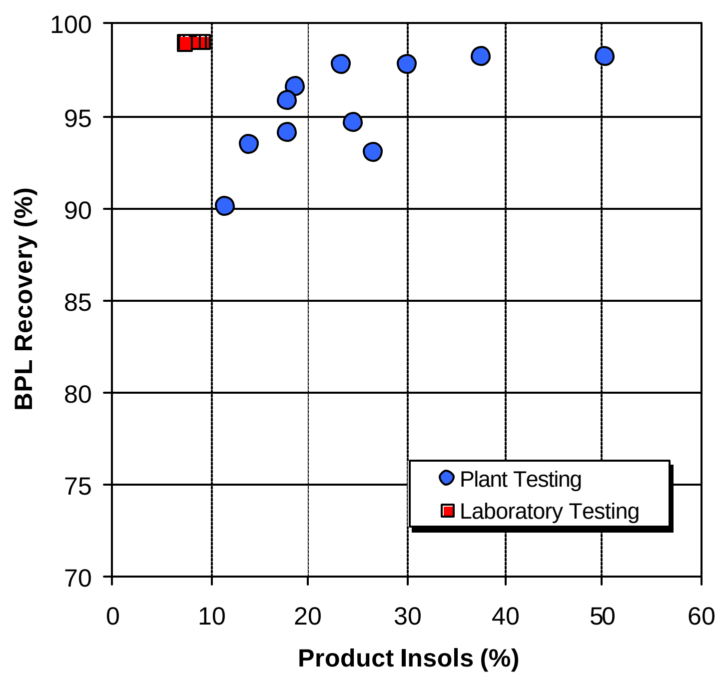


**Figure 2.16. BPL Recovery vs. Product Insol Grade for Coarse Matrix.**

### 2.4.4.3 Fine Feed

The results from the in-plant testing on the fine matrix are shown in Figure 2.17. BPL recovery ranged from 88% to 97% using the HydroFloat. When operated as an open column, BPL recoveries ranged from 85% to 92%, though at a significantly lower product insol (37% vs. 22%, respectively). Results from samples collected around the existing plant rougher-scavenger swing circuit are also presented in Figure 2.17 for comparison. The findings indicate that the open column cell (w/ HydroFloat sparging system) is able to achieve incrementally higher BPL recoveries at lower product insol grades compared to either the HydroFloat or the existing column technology. The corresponding product grade (%BPL) averaged 55% for the open column system as seen in Figure 2.18. As with the ultra-coarse and coarse circuits, the HydroFloat achieved an acceptable product grade and recovery in the fine circuit.

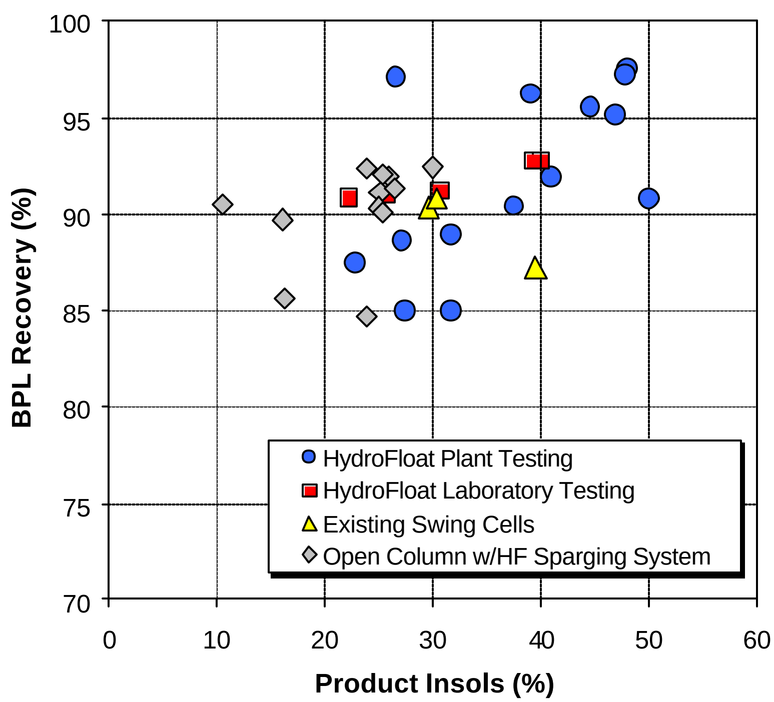


**Figure 2.17. BPL Recovery vs. Product Insol Grade for Fine Matrix.**

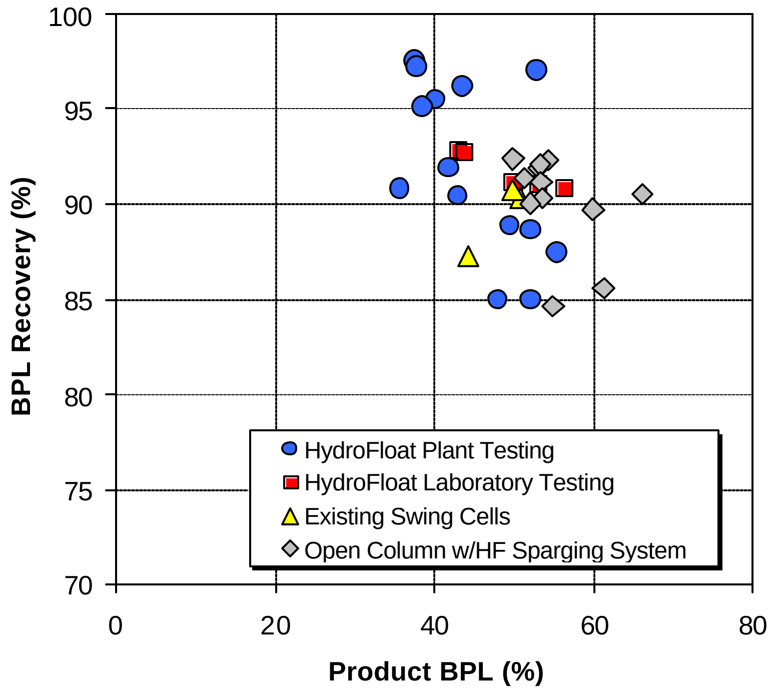


**Figure 2.18. BPL Recovery vs. Product BPL Grade for Fine Matrix.**

### 2.4.5 Sample Analysis

Detailed analyses were performed in accordance with ASTM procedures onsite at the Phosphate Plant B. Representative samples were collected around the pilot-scale unit. Slurry flow rates for the feed, underflow and overflow streams were directly measured using a stopwatch and a calibrated container. The mass and liquid flow rates were then calculated from the measured slurry flow rates and the sample assays using the two-product formula.

### 2.4.6 Future Work

Pilot-scale testing at Phosphate Plant B in Florida proved to be successful, and the company has agreed to purchase a prototype HydroFloat separator for testing and evaluation. The prototype unit will be compared to existing flotation cells in the coarse recovery circuit.

## 2.5 Conclusions

1. The in-plant evaluation of the HydroFloat separator demonstrated that this novel separation device can successfully treat the three different size fraction in a typical phosphate processing plant. For the ultra-coarse rock, the separator produced a high grade phosphate product (+66% BPL) at BPL recoveries exceeding 95%. For the coarse sized feed fraction, the separator produced a 99% BPL recovery at an 8% insol grade. Significant improvements were also achieved in the fine feed fractions where a BPL recovery greater than 90% was achieved with product insoles ranging between 22-25%.

2. Several advantages can be realized through implementation of the HydroFloat system. The system can provide a higher product mass recovery, superior metallurgical results, lower reagent costs and lower power requirements, with the greatest advantage being the higher separation efficiency. A higher product mass recovery with a better product quality is a significant achievement for this application. The HydroFloat has a substantially lower operating cost due to reduced reagent consumption and power requirements compared to conventional equipment.

3. One of the goals of this project is to successfully prove the technology in a sufficient period of time to minimize the financial risk that will be taken by industry. The previous years test work has eliminated the uncertainties associated with the HydroFloat separator by proving plant scale units do in fact work. This can be seen by the fact that industry leaders have submitted purchase requests for full scale units in their preparation plants.

Based on the successful installation of these full scale units, further implementation of additional units can be utilized in a broad spectrum of companies and industries.

4. Key design and operating variables have been established based on the performance capabilities of the HydroFloat separator. From here, proof-of-concept (POC) tests using a production-scale unit can be implemented at the various test locations where full scale prototypes are being installed. The POC-scale tests will identify critical scale-up criteria for the design of industrial applications. The POC-scale tests will also be used to define the performance capabilities of the high-efficiency processes in an industrial setting and to fully demonstrate the potential economic benefits that can be realized with the HydroFloat separator.

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