Chapter 1

In-Plant Testing of CrossFlow Separator 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
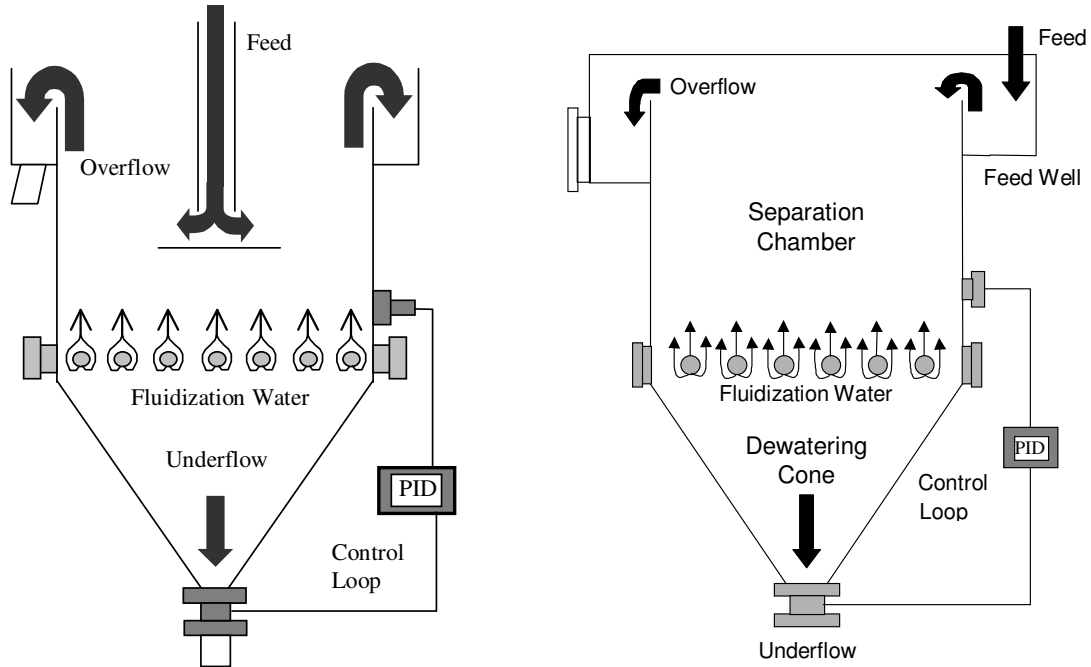
Figure 1.1 Traditional Hydraulic Separator (Left) Versus Crossflow Separator (Right).

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

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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.

1.1.4 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   <u>**Literature Review**</u>

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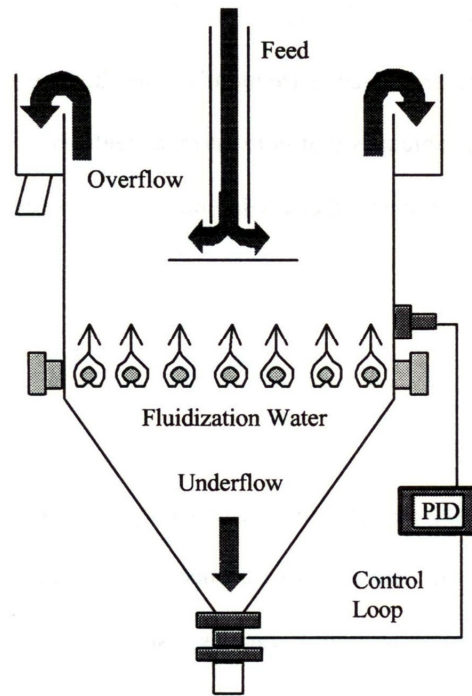
Figure 1.2. Schematic Diagram of a Traditional Hindered Bed Separator.

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				T-Water	Level No.	Comment
	(gpm)	(SG)	(% Solids)	(tph/sqft)	(gpm)		
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.
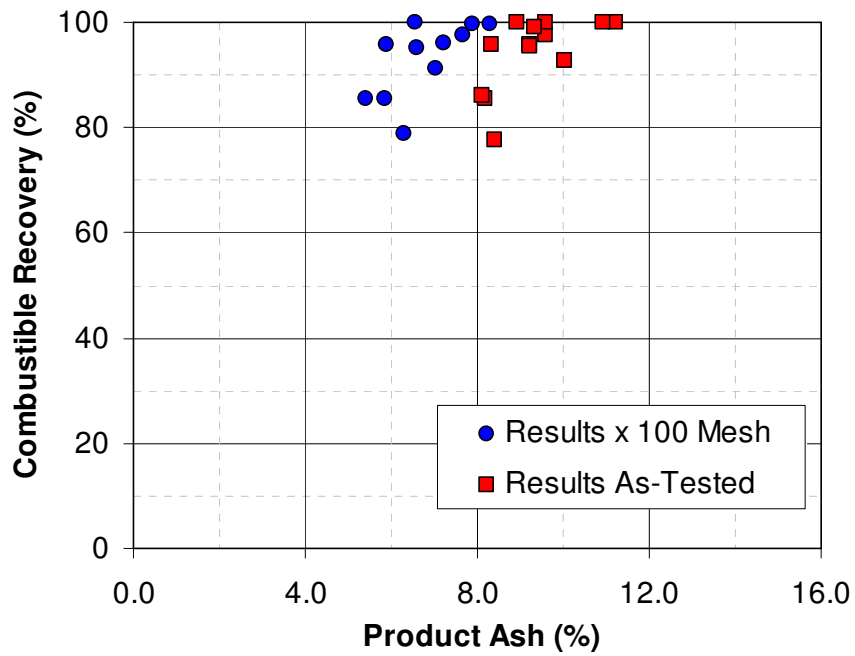
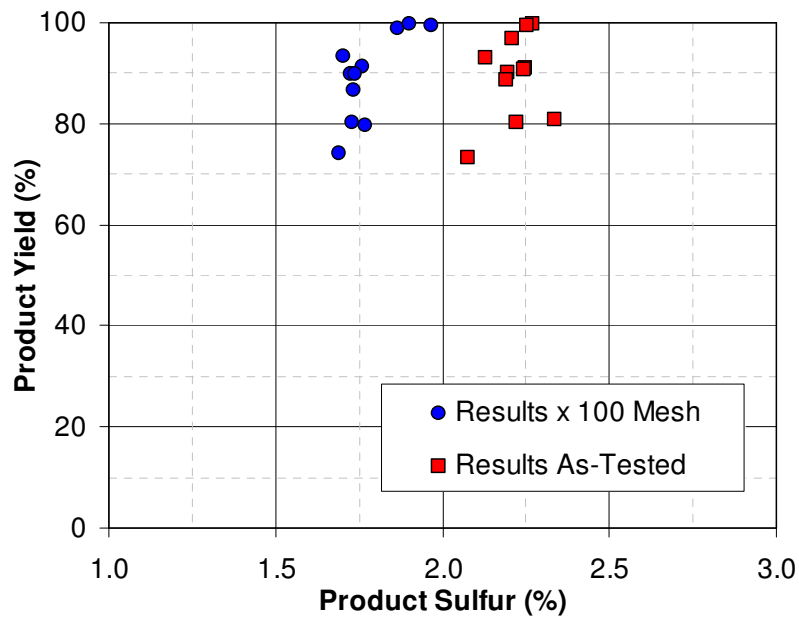**Figure 1.3. Combustible Recovery vs. Product Ash Content.**

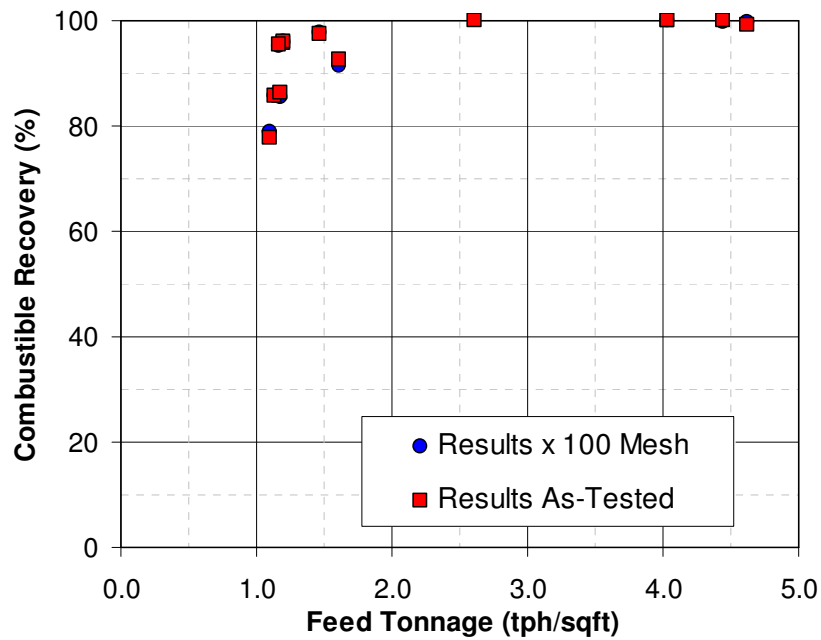**Figure 1.4 Mass Yield vs. Product Sulfur Content.**

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.

CROSSFLOW CIRCUIT - MATERIAL BALANCE
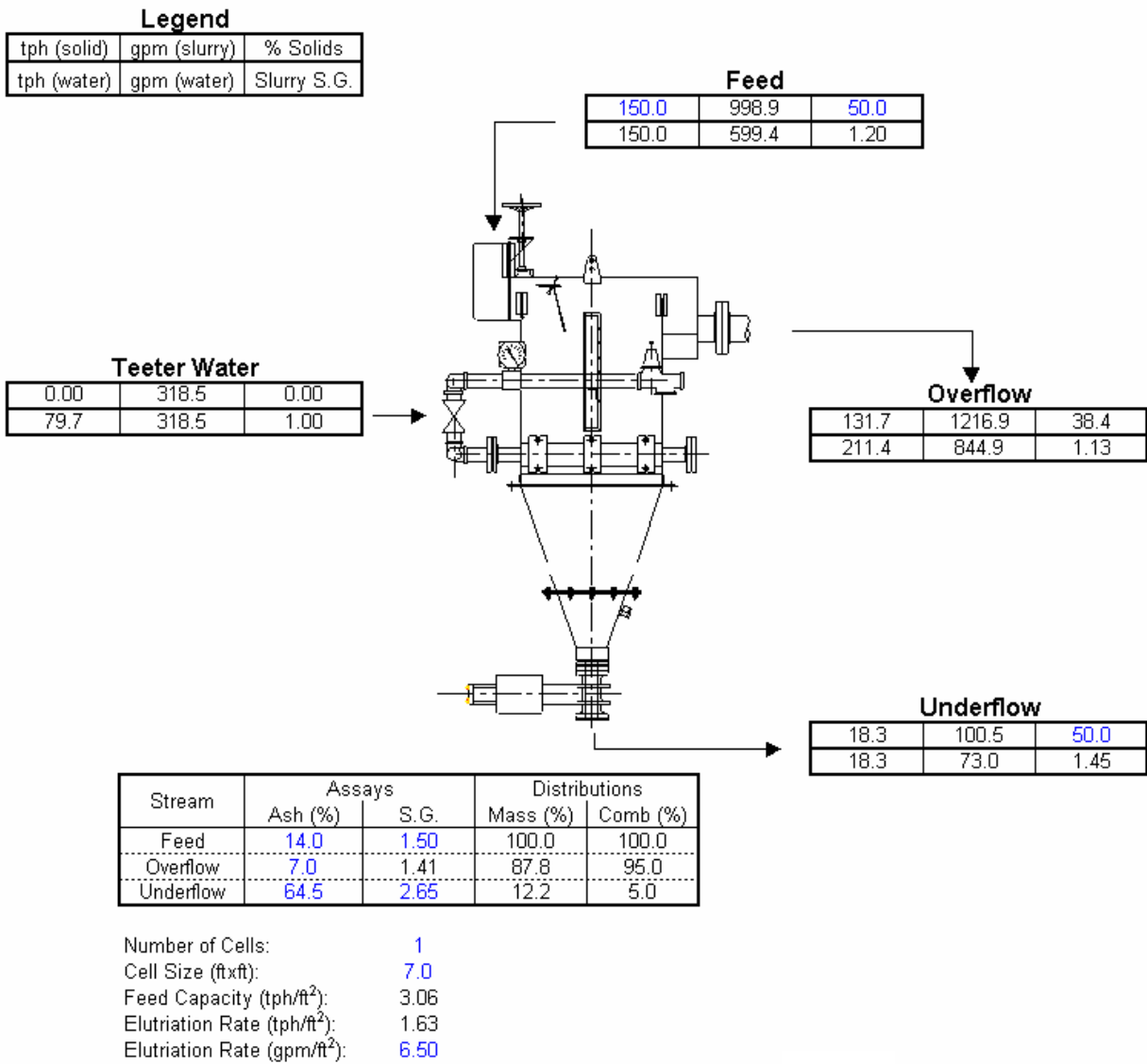## *150.0   stph Option*

Legend

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

Feed

150.0	998.9	50.0
150.0	599.4	1.20

Teeter Water

0.00	318.5	0.00
79.7	318.5	1.00

Overflow

131.7	1216.9	38.4
211.4	844.9	1.13

Underflow

18.3	100.5	50.0
18.3	73.0	1.45

Stream	Assays		Distributions	
	Ash (%)	S.G.	Mass (%)	Comb (%)
Feed	14.0	1.50	100.0	100.0
Overflow	7.0	1.41	87.8	95.0
Underflow	64.5	2.65	12.2	5.0

Number of Cells:	1
Cell Size (ftxft):	7.0
Feed Capacity (tph/ft^2):	3.06
Elutriation Rate (tph/ft^2):	1.63
Elutriation Rate (gpm/ft^2):	6.50

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   In-Plant Testing at Coal Plant B

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	Size			Weight	Ash
Description	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	Test	Feed			Level	Water
Operation	Number	% 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
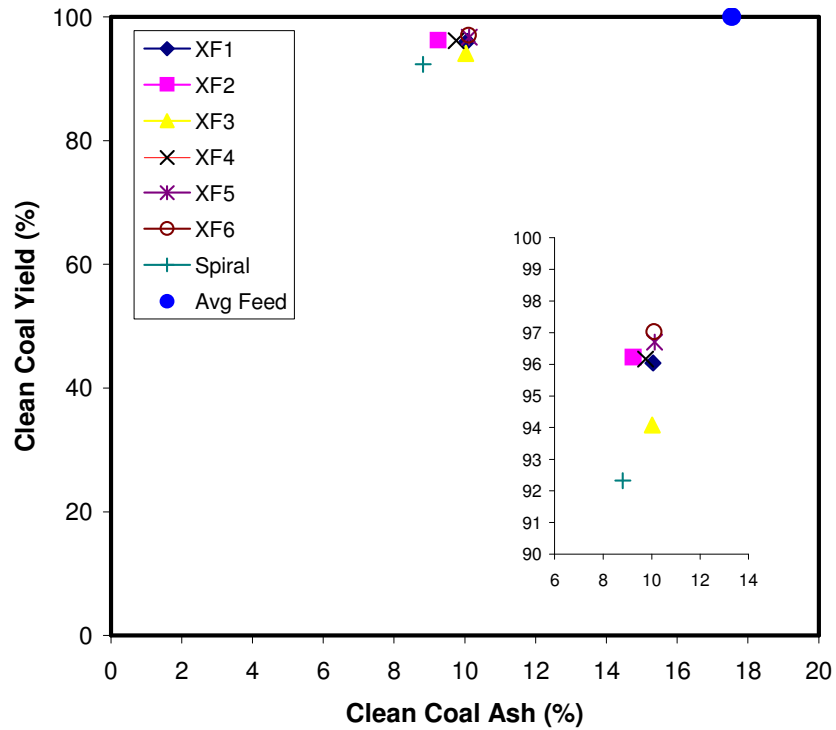
Figure 1.8 Yield vs. Clean Coal Ash for +100-Mesh Size Fraction.

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.
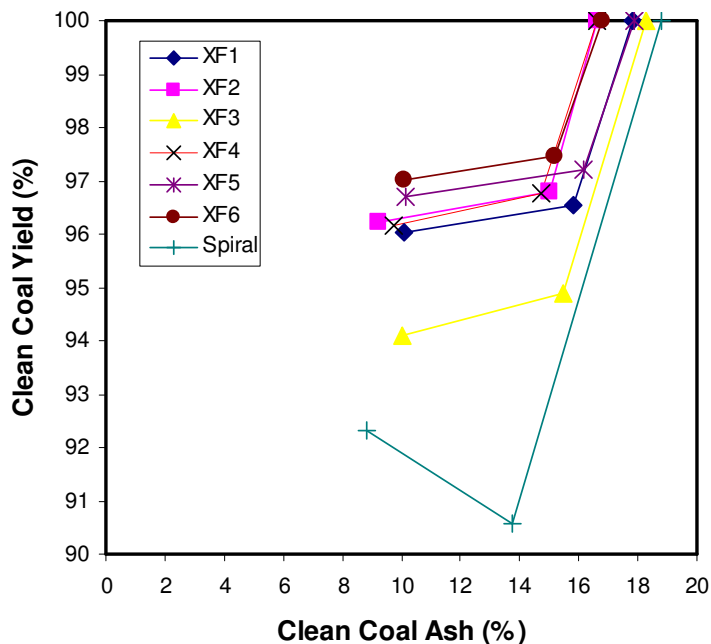
Figure 1.9. Performance for +100 Mesh and Composite Samples.
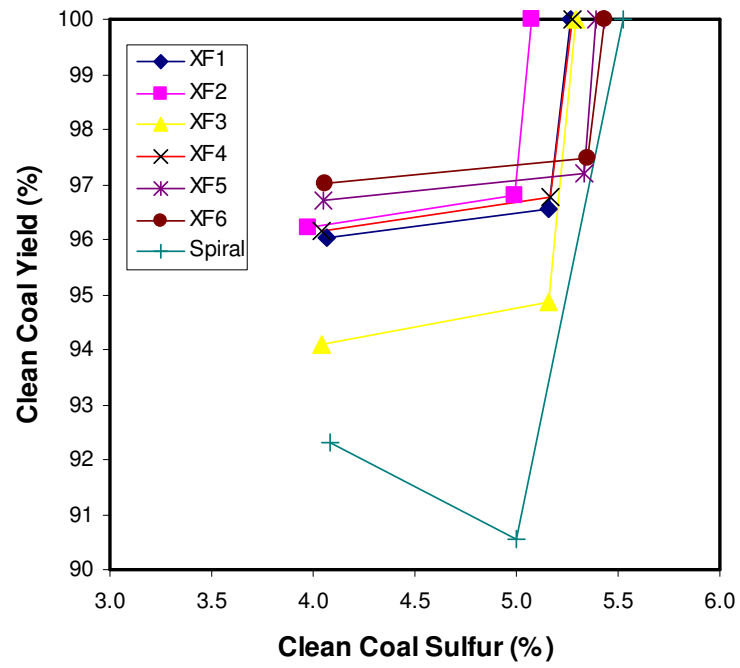
Figure 1.10. Performance for +100 Mesh and Composite Samples.

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. The12-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
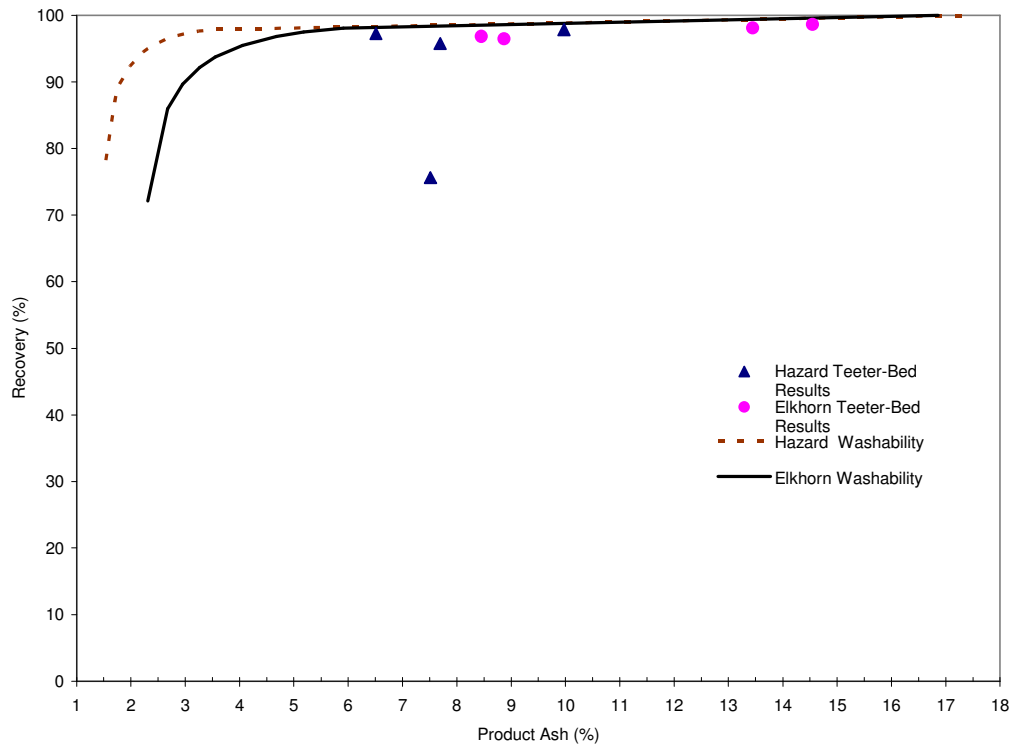
Figure 1.13. Comparison of the Teeter-Bed Separation Performances and the Washability Characteristics of the Seam A and Seam B Fine Coals at Coal Plant C.

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 | Feed | | | Level | Water |
Number	% 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
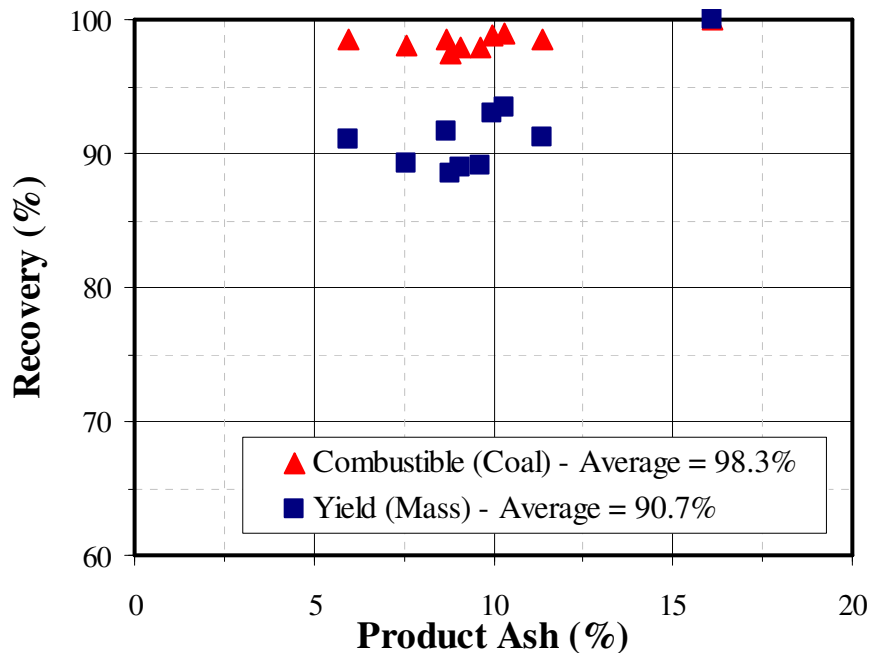
Figure 1.15. Recovery vs. Product Ash Content of 100 M Coal.
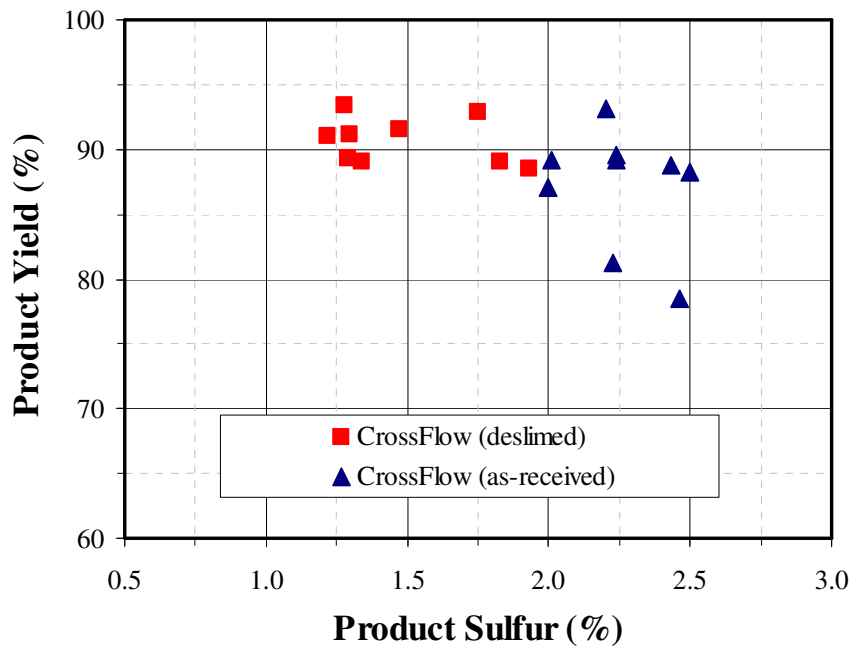
Figure 1.16. Mass Yield vs. Product Sulfur Content of 100 M Coal.

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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Chapter 2

In-Plant Testing of HydroFloat Separator in 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.
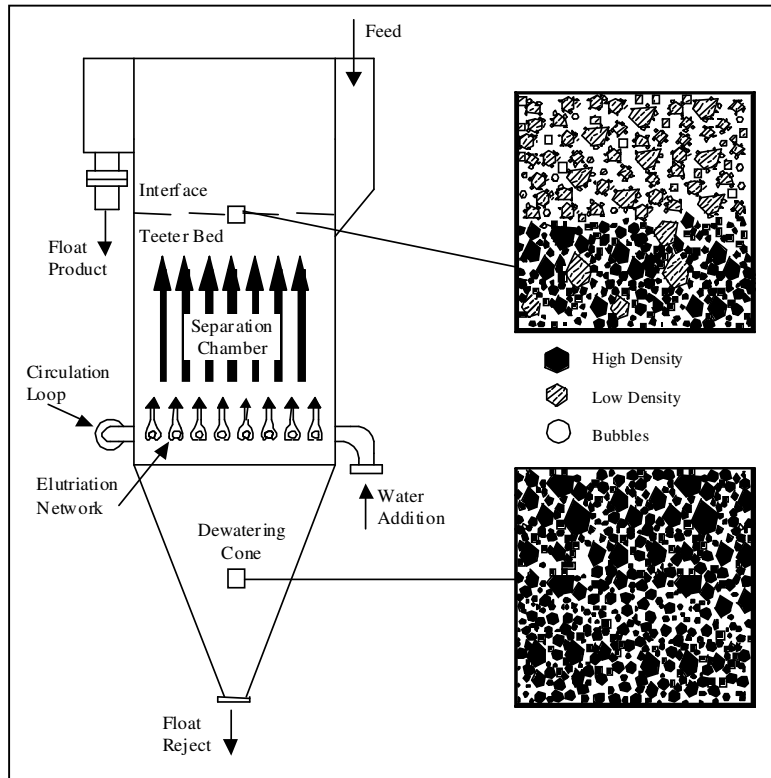
Figure 2.1. Schematic Drawing of HydroFloat Separator.

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   Literature Review

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 ($d_p < 20\mu m$) 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   In-Plant Testing at Phosphate Plant A

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.
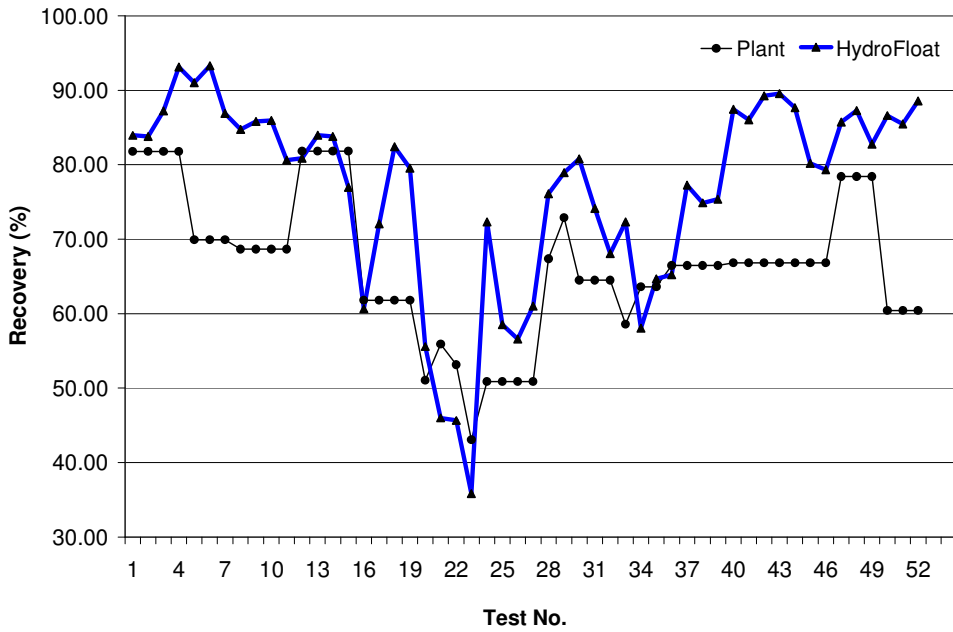
Figure 2.3 - Recovery of HydroFloat Separator versus Plant Cells.
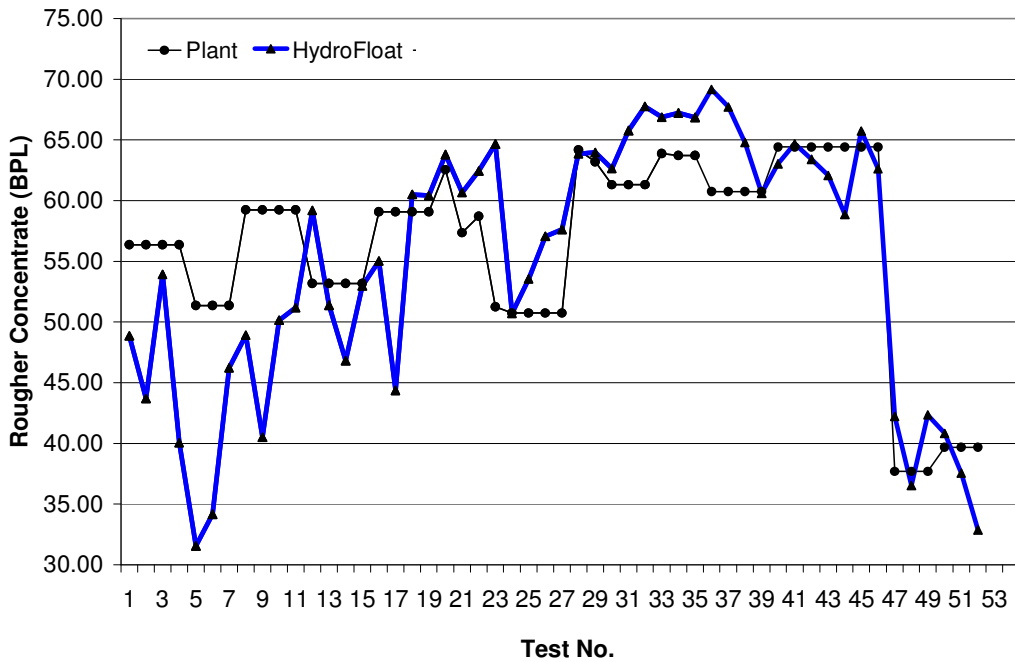
Figure 2.4. Rougher Concentrate Grade of HydroFloat Separator versus Plant Cells.

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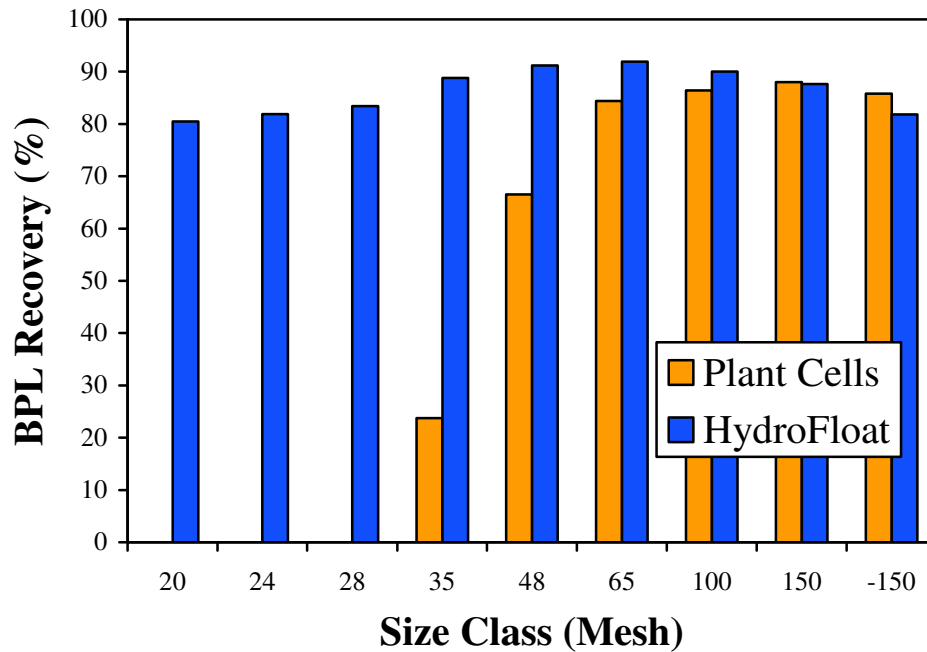
Figure 2.5. Comparison of Test Results for Fine Phosphate (Plant Circuit #2).

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.
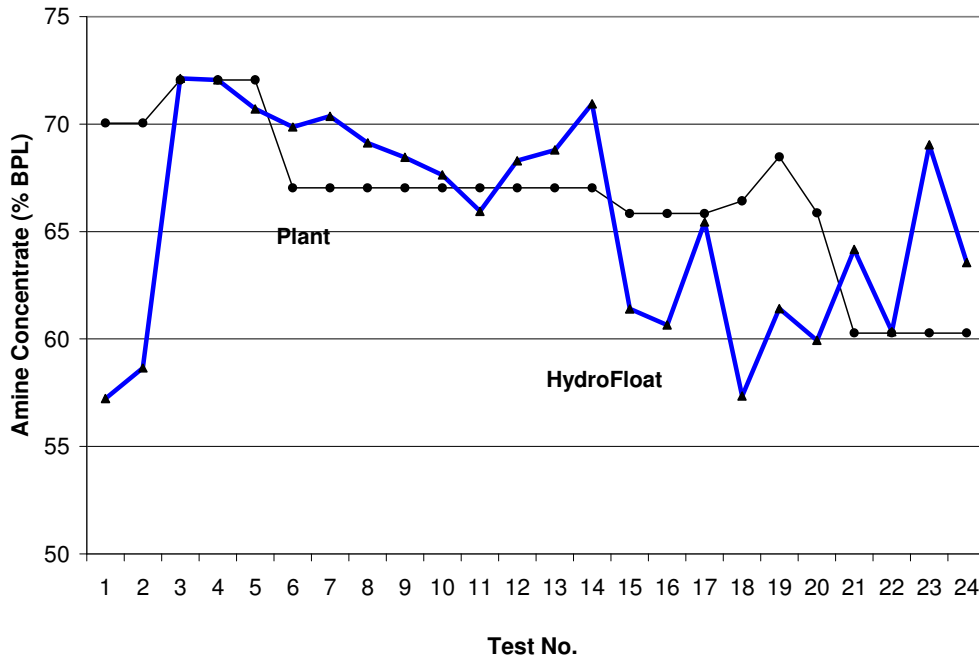
Figure 2.6. Amine Concentrate Grade Comparison of HydroFloat Separator Vs. Existing Plant Cells.
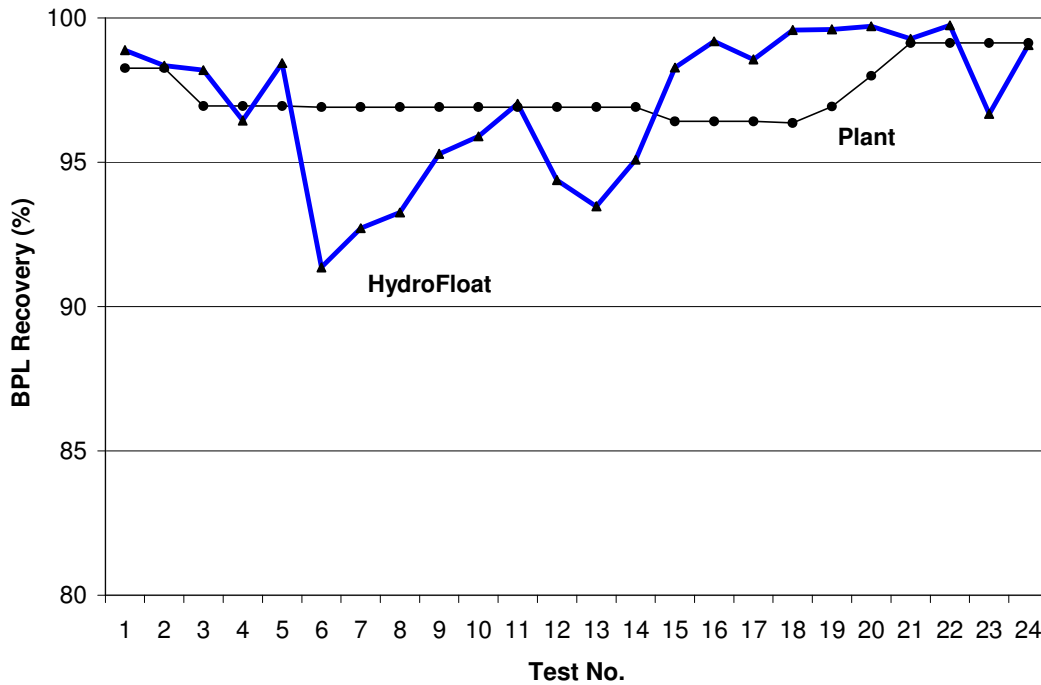
Figure 2.7. BPL Recovery Comparisons HydroFloat Separator vs. Existing Plant Cells.

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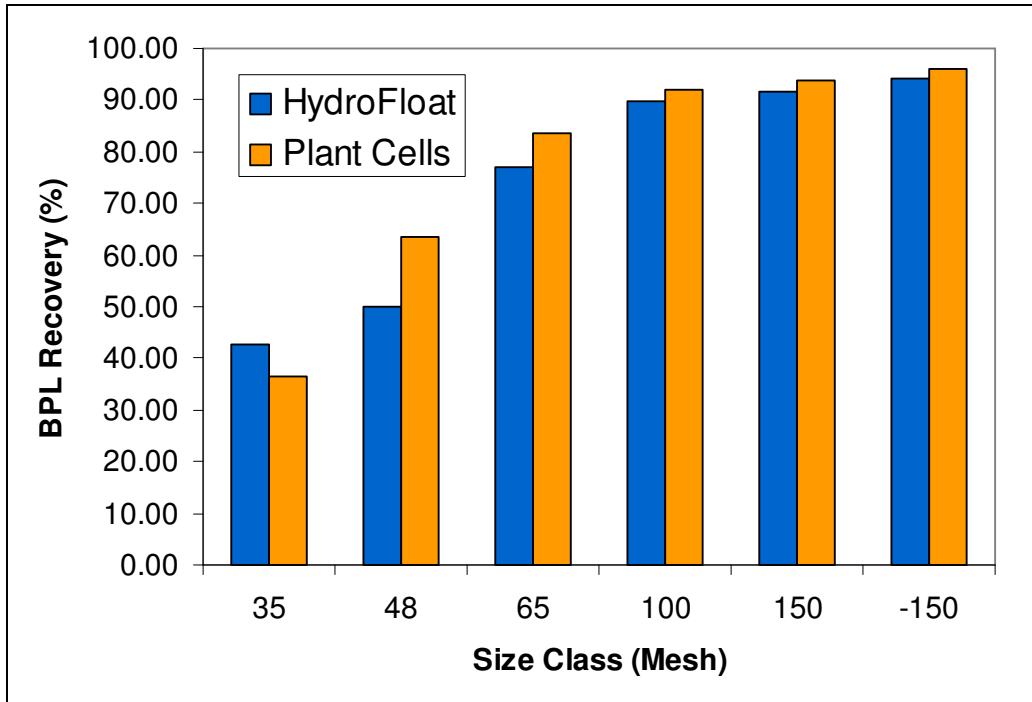
Figure 2.8. Comparison of Test Results for Amine Phosphate (Plant Circuit #2).

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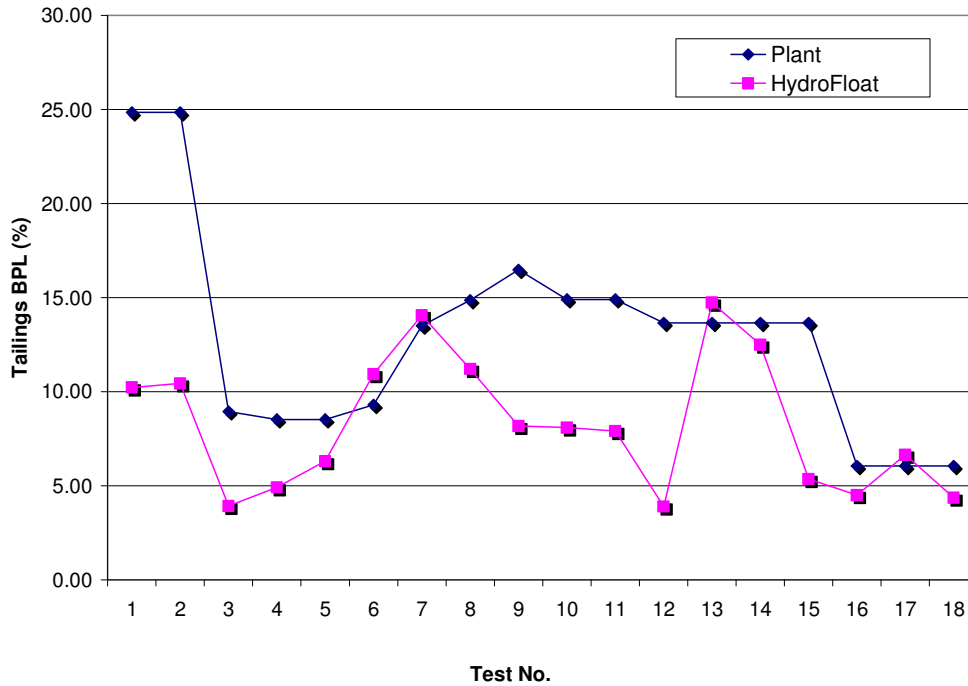
Figure 2.9. Tailings Comparison of HydroFloat Separator versus Plant Cells.

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.
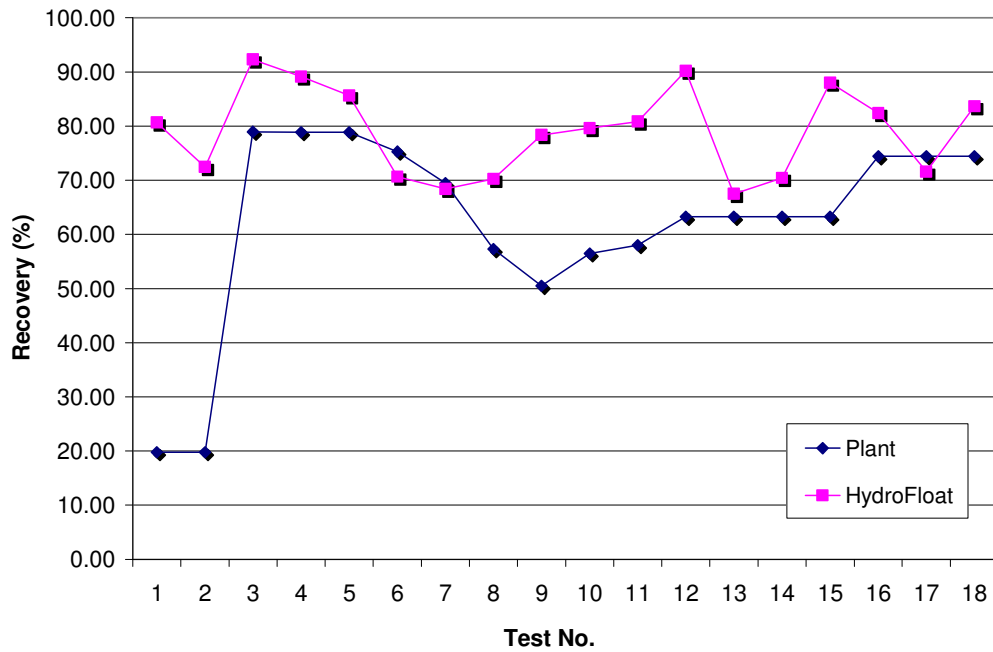
Figure 2.10. Recovery Comparison of HydroFloat Separator vs. Existing Plant Cells.
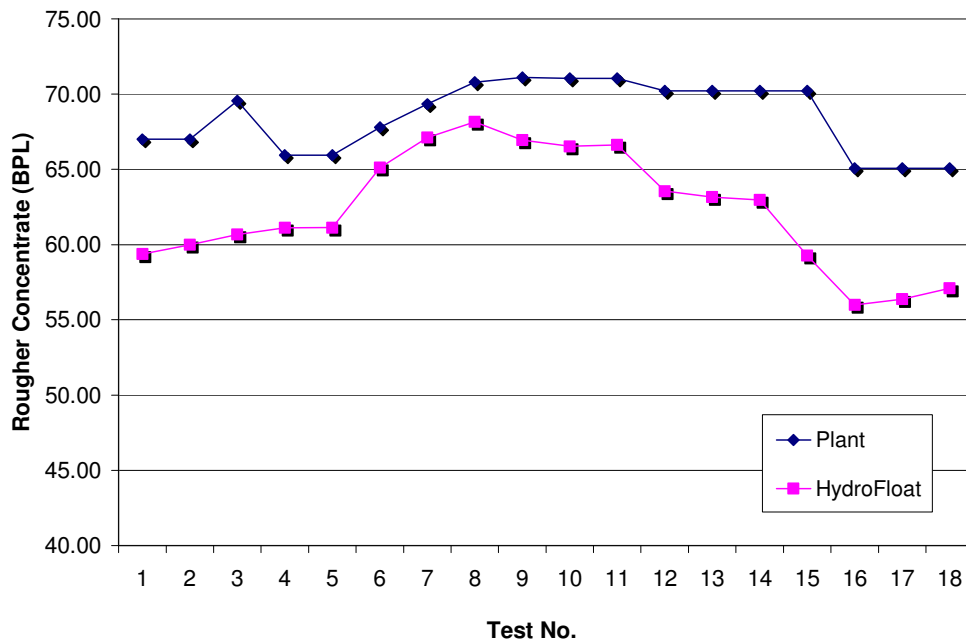
Figure 2.11. Grade Comparison of HydroFloat Separator vs. Existing Plant Cells.

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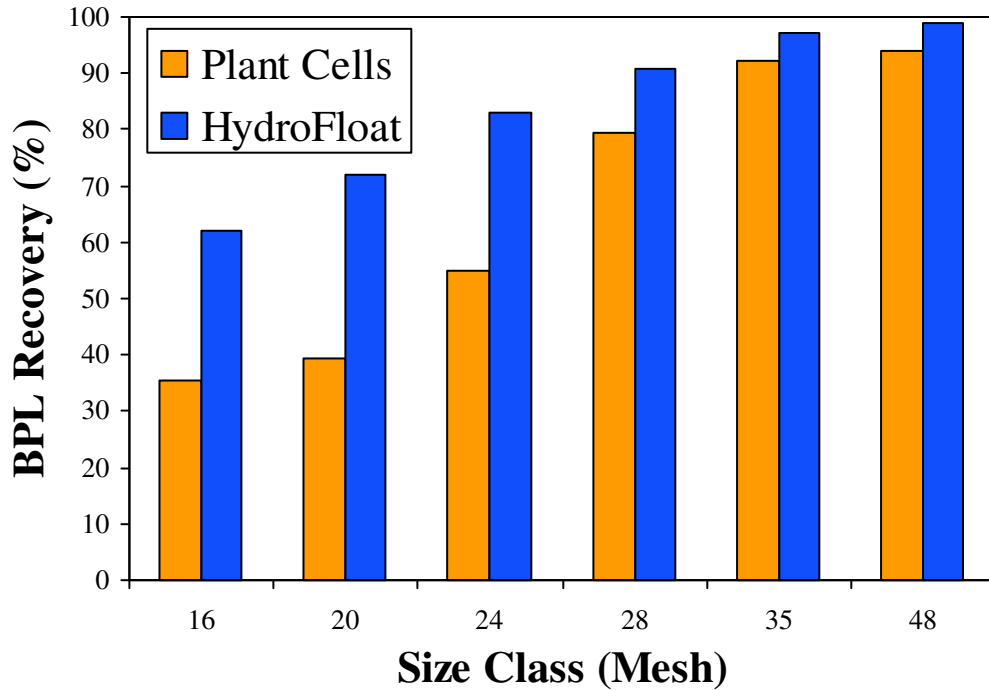
Figure 2.12. Comparison of Test Results for Coarse Phosphate (Plant Circuit #1).

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   In-Plant Testing at Phosphate Plant B

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.

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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.
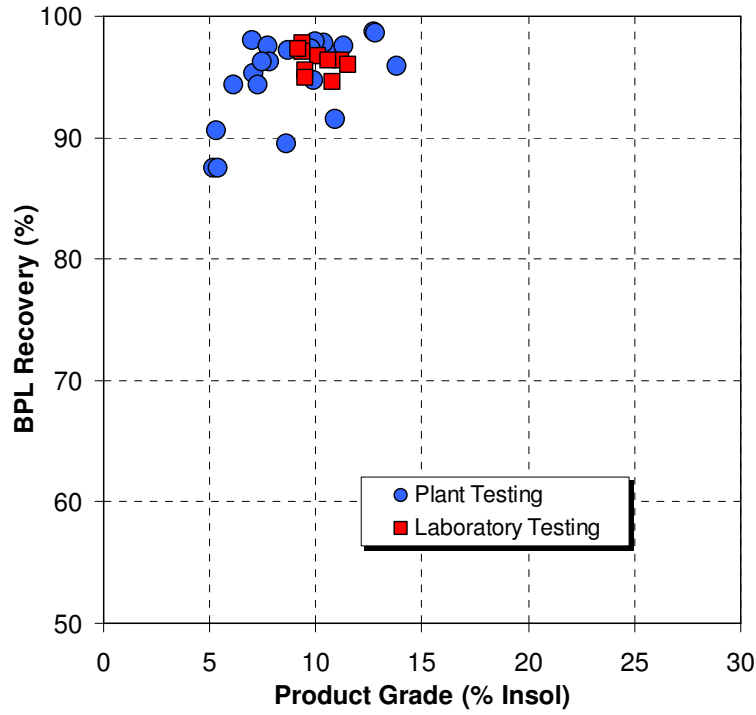
Figure 2.14. BPL Recovery vs. Product Insol Grade for Ultra-Coarse Rock Feed at Phosphate Plant B.
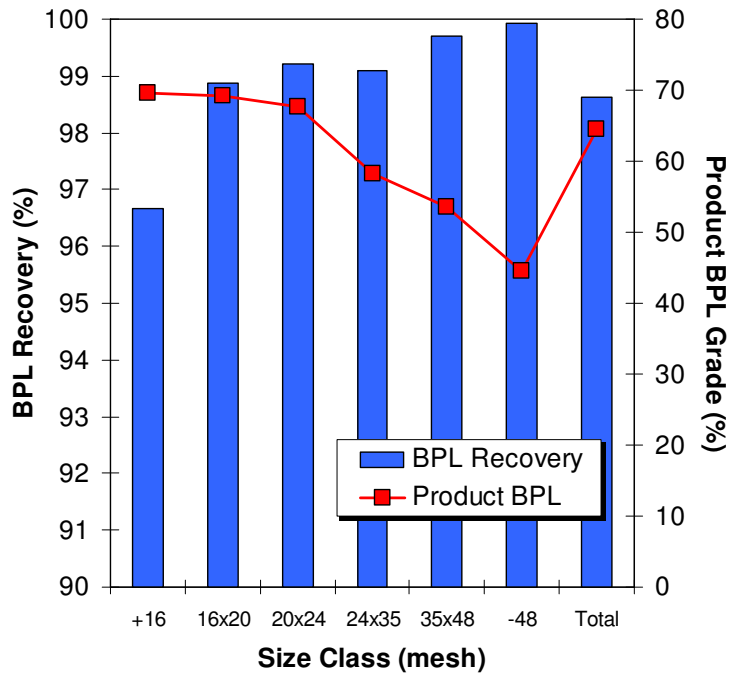
Figure 2.15. Size-by-Size Recovery and Grade for Ultra-Coarse Rock Feed.

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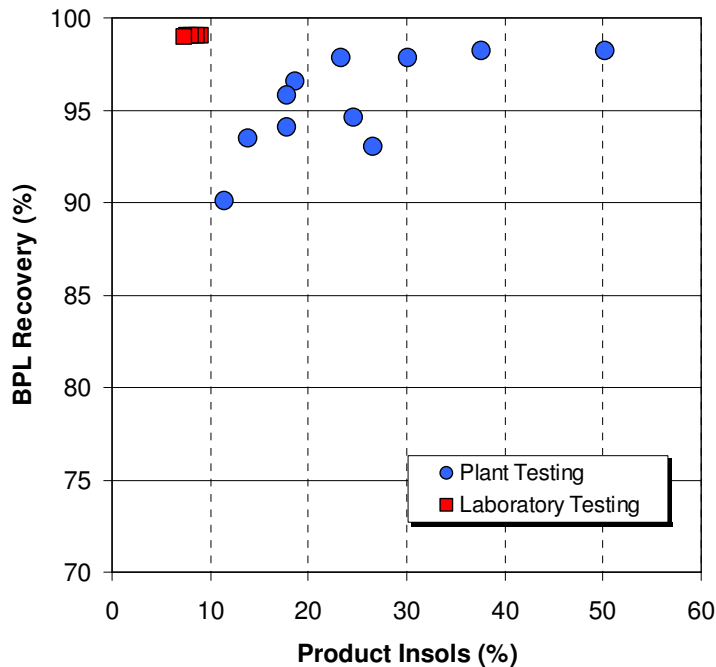
Figure 2.16. BPL Recovery vs. Product Insol Grade for Coarse Matrix.

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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.
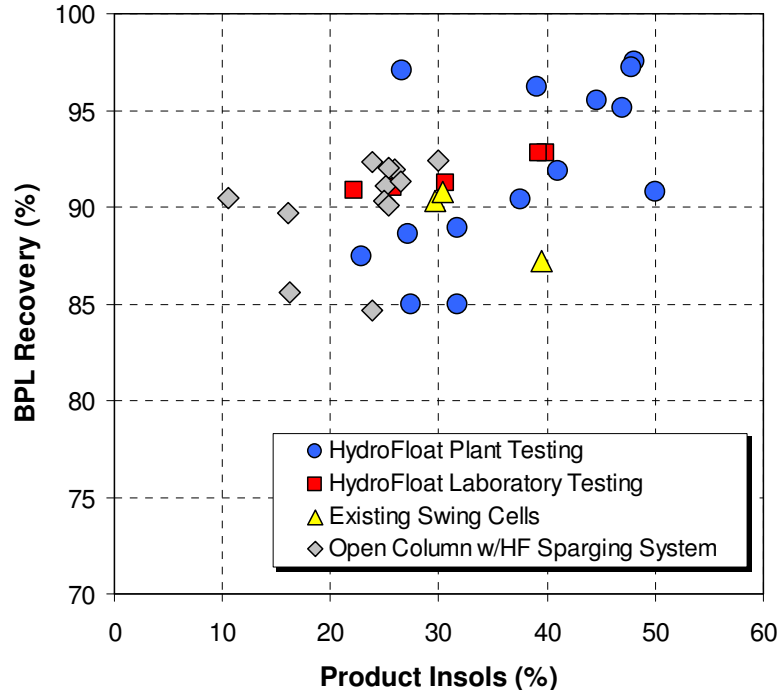
Figure 2.17. BPL Recovery vs. Product Insol Grade for Fine Matrix.
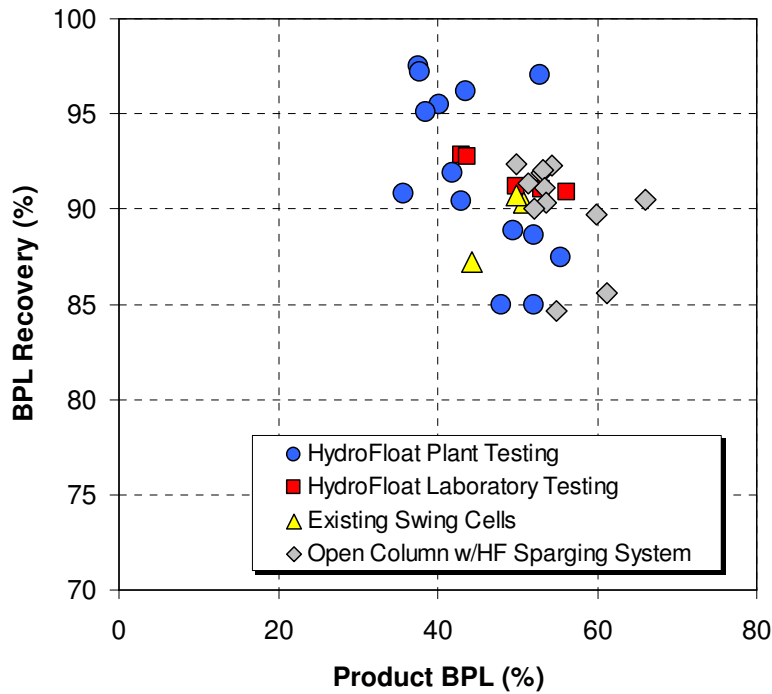
Figure 2.18. BPL Recovery vs. Product BPL Grade for Fine Matrix.

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 against existing

flotation cells in the coarse recovery circuit. Installation of the prototype is expected during September 2004.

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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Chapter 3

Beneficiation of Ultra-Fine Phosphate Streams

3.1   <u>Introduction</u>

Phosphate beneficiation plants are engineered to recover phosphate occurring in the 0.1 mm size fraction. In many cases, phosphate ore coarser than 1mm (i.e., ultra-coarse) is stockpiled and either ground and routed through the beneficiation plant or blended directly with the final product. Unfortunately, the phosphate that occurs in the finer fractions (i.e., finer than 0.1 mm) is generally rejected as waste in the clay slimes. Approximately 100,000 tons of these phosphatic clay slimes in dilute (3-5%) slurry are pumped to tailings impoundments in Florida each day. In addition to the expense of disposing of the waste clay, mining companies are throwing away millions of dollars in phosphate to the refuse stream. It is estimated that 27% of the current annual production of phosphate is lost to this refuse stream. Economic recovery of phosphate from the phosphatic clays could extend Florida's phosphate resource life by decades.

Slimes are a natural component of phosphate ore. The slimes (phosphatic clays) consist of both phosphate and clay particles. According to Zhang (2001), there are three major characteristics of refuse slimes cause extreme difficulty in recovering the phosphate economically: the ultra-fine particle size (35-50% below 1 micron), the even distribution of phosphate particles among the various size fractions and the high clay content (30-50%). As particle size is reduced, two dominating characteristics begin to emerge: the specific surface becomes large and the mass of the particle becomes very small (Abdel-Khalek, *et al.*, 1990). The ultra-fine particles in a slimes stream increase the chance of the fine particles getting carried into the froth as they are either entrained in the liquid or mechanically entrapped with the

particles being floated. Material from a slimes stream will have excess amounts of gangue minerals, which will ultimately reduce the grade of the concentrate if they are carried into the froth. Excessive slimes can significantly reduce flotation performance and simultaneously increase reagent consumption. Because of their high dolomite content, acidulation of such concentrates consumes excessive sulfuric acid and causes problems during phosphoric acid manufacture such as increased acid viscosity, precipitation of insoluble Mg-phosphates and difficulties in filtration and clarification of the final product.

As a result, the phosphatic clays are continuously rejected from the beneficiation plant using both vibrating screens and hydrocyclones. The slimes, including the fine phosphate, are ultimately placed in large tailings ponds, which act as large holding cells that are used for both water clarification and refuse storage. Misplaced coarse phosphate values are also present in the refuse stream. The coarse values are predominately a result of inefficient sizing or desliming. For instance, if an excess amount of material is pumped to a hydrocyclone, the cyclone will "rope" causing vast misplacement of coarse material. "Roping" is a common occurrence in the phosphate industry due to the natural settling characteristics of the coarse ore.

A typical Florida phosphate plant discards 100,000 gpm to the tailings ponds, of which a significant portion of is fine phosphate (passing 0.1 mm). Combined with the misplaced coarse particles, the discarded phosphate represents a significant reduction in overall plant efficiency and a resultant increase in the required volume of the refuse tailings ponds. Recovering this material would increase overall plant production, while ultimately reducing operating costs. While previous research has been conducted in this area, there is still not an economical recovery system and as a result, this phosphate is currently considered unrecoverable by industry.

An urgent need exists to develop a low-cost recovery circuit to recover a significant amount of the phosphate values that are currently being thrown away. One circuit configuration that may meet this requirement is shown in Figure 3.1. The circuit, which incorporates simple low-cost unit operations such as hydraulic classifiers, hydrocyclones, and flotation cells, can be tailored to exploit particle characteristics particular to the size range in question.

The proposed circuit uses a large free-settling tank to collect the entire flow of phosphatic clays that are currently rejected from the beneficiation plant. Tanks similar to these are already extensively used by industry. The tanks allow the coarse fraction (plus 400 mesh) of the phosphatic clays to settle out in a free-settling regime. The settling tank is equipped with an overflow launder to collect the finest fraction, which is directed to plant tailings. The plus 400 mesh fraction is pumped to a bank of hydrocyclones. The hydrocyclones reject as overflow the remaining slimes that inherently remain with the coarse fraction.

At this point in the circuit, the cyclone underflow will be significantly deslimed and will have relatively high percent solids. However, as flotation feed, the cyclone underflow will still contain slimes (due to bypass) that will prove detrimental to flotation performance and reagent consumption. To overcome this problem, the cyclone underflow is further treated using a CrossFlow hydraulic separator. The teeter bed in the CrossFlow separator removes the majority of slimes via countercurrent washing. The underflow from the CrossFlow, which is optimal material for flotation feed, will be discharged through the bottom of the unit at a high percent solids (i.e., up to 75%).

The final step of the process involves column flotation to recovery the phosphate ore in the presence of the clay slimes. Unlike coarse phosphate ore, it is expected that the fine nature of
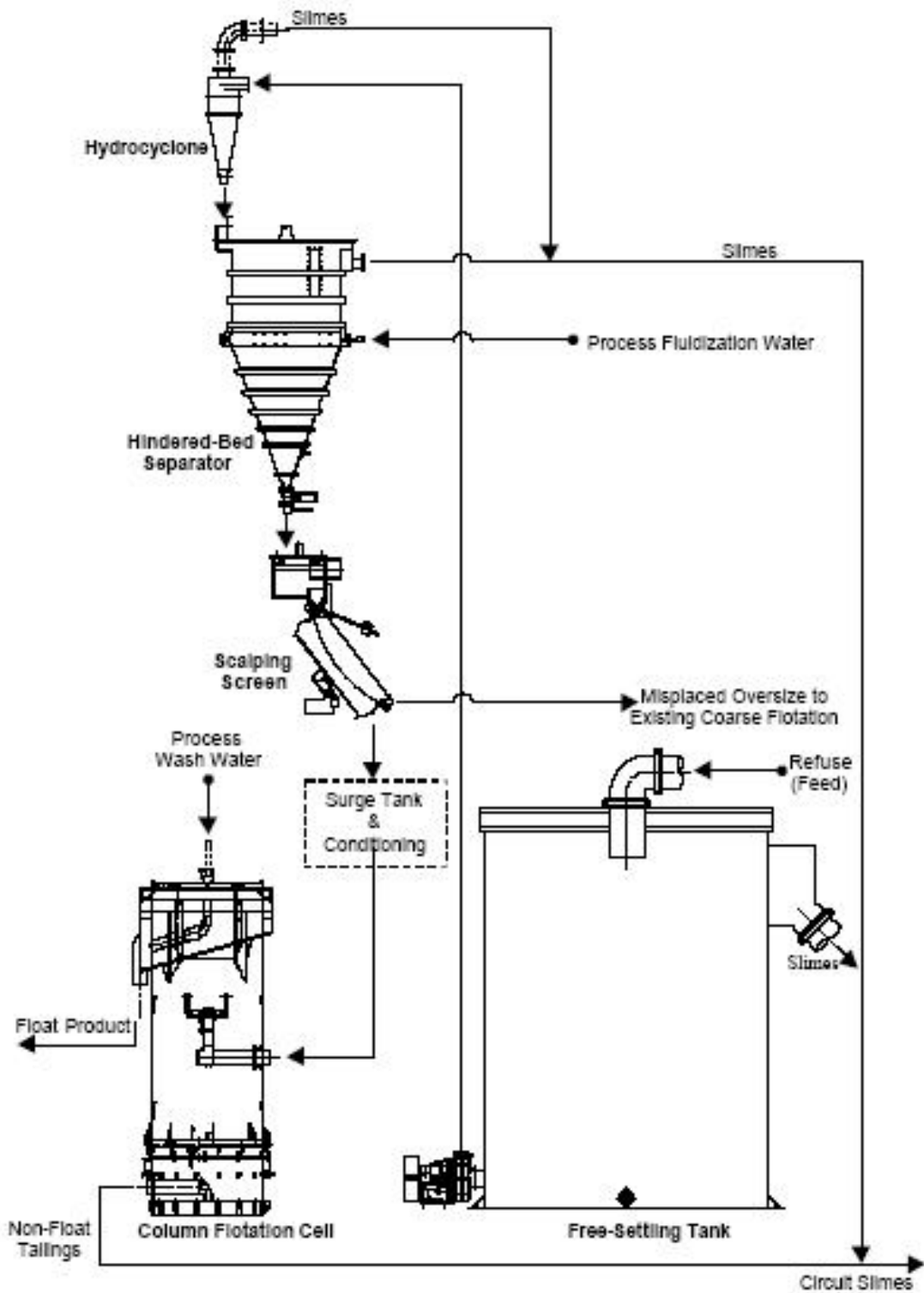
Figure 3.1. Proposed Recovery Circuit for Ultra-Fine Phosphate from Clay Tailings.

the phosphate clays will provide stable flotation froth. Most importantly, the stable froth will allow for the use of wash water. Column wash water is then used as a means to rinse any non-floatable species (i.e., clays) from the float product prior to collection. In effect, water is evenly distributed and showered over the flotation froth to force clays and slimes to the column underflow. This step ensures a high-grade phosphate product. In fact, it is anticipated that the wash water will reject the primary contaminants associated with clays (MgO and Al_2O_3).

3.2   Literature Review

3.2.1   General

Florida has enormous reserves and will continue to produce a third of the world's phosphate supply simply by making changes in the present system of mining, beneficiation, waste disposal and land reclamation (Lawver *et al*, 1984). Beneficiation plants continue to optimize separation processes to improve recoveries and minimize gangue minerals in the final product. From improved separation technologies to better flotation reagents, all aspects of the beneficiation process are being analyzed to determine where improvements can be made. As waste disposal becomes a more and more urgent issue, new techniques are being proposed to increase recovery of P_2O_5 in the clay slimes stream. Current beneficiation plants dispose of one-third of the current annual phosphate production through the clay slimes stream.

The source of the problematic recovery of phosphate from the slimes stream is the existence of almost 50% clay minerals in the stream. These slimes are removed, not only because they are refractory toward presently known upgrading processes, but they also interfere with the operation of the flotation on the remainder of the material (Hazen, 1969). As a result, researchers have been trying for decades to develop efficient and economical processes to recover the phosphate from the clay slimes stream.

Beneficiation has become increasingly more difficult as reserves continue to have more dolomite content, lower ore grades and increased clay content. As production continues into the Hawthorn Formation in Florida, the silicate contamination is compounded with additional magnesium contamination. Beneficiation plants can no longer blend lower MgO ore with higher dolomitic ores (containing the MgO) to alleviate this problem. Both calcite and dolomite consume sulfuric acid and can cause excessive foaming during rock processing. Dolomite is

particularly detrimental to the process because magnesium passes into and degrades both liquid and solid fertilizer products (McCullough *et al*., 1977).

Historically, the separation of dolomite and calcite from phosphate has primarily been conducted through various flotation processes. Three reagents are historically used for rougher flotation of phosphate from most of the siliceous gangue: a fatty acid type collector, an oil-type extender, and a pH regulator (Lawver *et al.,* 1984). Fatty acid collectors consist of crude tall oil, reconstituted tall oil, blends of tall oil with vegetable fatty acids, or tall oil soap skimmings. Oil-type extenders are commonly fuel oils, reclaimed motor oils, or mixtures of these two. The use of fuel oil extenders with fatty acids as an auxiliary collector in flotation is well known and practiced in the industry. Fuel oil does not adsorb onto phosphate and quartz in the absence of fatty acids, and as a result, flotation cannot be achieved with fuel oil alone. Fuel oil will adsorb synergistically with fatty acids on apatite. Ammonia and soda ash are the two most common pH modifiers used in the industry. Sodium silicate is one of the most widely used depressants for floating non-sulphide mineral. Sodium silicate is primarily a strong and selective depressant (Shin and Choi, 1985). The use of selective flocculation has also been studied on the phosphatic clays. While studies have been conducted with all of these reagents, today there is still no industry-wide standard of separating phosphate from dolomite, calcite and other clay constituents.

3.2.2   Advances in Flotation Reagents

3.2.2.1 Use of Sodium Silica

Recovery of phosphate from fine clay streams has historically been very problematic due to the high concentration of gangue minerals. Sun and Smit (1963) were able to recover fine

phosphate from Florida washer slime material using various fatty acid collectors and tall oil. Various conditions were tested including effects of pH, fuel oil, pulp density and addition of sodium metasilicate. Flotation results show that phosphate flotation decreases in the order of linolenic, linoleic, oleic and stearic acid, which have 3, 2, 1, and 0 double bonds respectively (Sun and Smit, 1963). Sun and Smit (1963) concluded that unsaturated acids are preferred for most flotation purposes. Duplicate test runs achieved recoveries of 29.8 and 31.0% P_2O_5 with 13.0 and 12.7% insolubles, respectively with test conditions including: 8.5 pH, 2.24 lb/ton fatty acid, 6.72 lb/ton fuel oil, 0.10 lb/ton pine oil, and 12 minutes of conditioning at 25% solids (Sun and Smit, 1963). The fatty acid used had between 1.3 and 1.5% rosin acids, with the highest overall fatty acid content of the materials tested (96.5-96.9%) to achieve the best recoveries.

Mishra (1982) was one of the first who studied the electrokinetic properties of sodium metasilicate, to be used as the modifying agent for the separation of apatite from calcite by depressing calcite when using sodium oleate as the collector. Mishra determined that the degree of polymerization of sodium silicate is influence by its SiO_2 to Na_2O ratio, where polymerization increases with the increase in SiO_2 to Na_2O ratio. The optimum rate of polymerization was found to be at pH 8.6 (Mishra, 1982). While sodium metasilicate and sodium oleate additions were kept constant at 5×10^{-3} M and 5×10^{-4} M, respectively, flotation test work was carried out at various pH. The addition of sodium metasilicate produced a depression zone between pH 6.0 and 9.0, but flotation recovery increased to 100% between pH 9.0 and 10.0. Recoveries decreased to zero by pH 11.7. Mishra related the differences in recovery to changes in zeta potential, and predicted two causes for its occurrence: (1) the adsorption of cationic species at the apatite surface or (2) the adsorption of colloidal molecular species of sodium silicate. Flotation results suggested that the selective action of sodium metasilicate on calcite, with sodium oleate

as a collector, apatite and calcite could be separated in the alkaline pH environment at about pH 10.0, depressing calcite with sodium metasilicate (Mishra, 1982).

Anazia and Hanna (1987) also found that the addition of sodium silicate improved the P_2O_5 recovery in the phosphate concentrate and increased the rejection of siliceous gangue in the tailings in Florida phosphate operations. Their research concentrated first on the removal of dolomite from the feed, followed by silica removal. Testing procedures included 250-g batches of 48 x 400 mesh flotation feed in a D-1 Denver laboratory flotation machine with the impeller speed set at 100 rpm (Anazia and Hanna, 1987). Oleic acid was used as the collector and pine oil as the frother.

Dolomite flotation was conducted by adjusting the pH to between 4.5 and 5.0 with various dilute acids. The fatty acid collector (at 3.3 lb/ton collector addition) and frother were injected into the pulp and air was immediately introduced into the pulp to float the carbonate gangue minerals. After separation of dolomite, as froth, the cell product was conditioned for three minutes with 1.1 lb/ton sodium silicate at a pH of 6.0 to 7.0. This was followed by another three minute conditioning with 1.1 lb/ton fatty acid collector and flotation of the phosphate minerals. The silica was depressed by the sodium silicate and the phosphate was floated to give concentrates of 29% P_2O_5 and 0.8% MgO, with a recovery of 76% (Anazia and Hanna, 1987). This process is also referred to as the Mineral Resource Institute (MRI) process. Since there was no phosphate depressant used, the pH had to be strictly controlled, which resulted in large acid consumption. As a result the project was considered unsuccessful.

Anazia and Hanna concluded that carbonate minerals (dolomite and calcite) readily react with inorganic acids, resulting in the preferential dissolution or removal of mixed surface contaminants on the carbonate particles and exposure of fresh clean surface sites suitable for

fatty acid adsorption. Carbon dioxide microbubbles are then generated by the carbonate minerals, allowing for enhanced oleic acid adsorption at the solid-liquid-gas interface, and 'instant' flotation of the mineral (Anazia and Hanna, 1987). Oleic acid will still have an affinity for the apatite particles, but it will not be as great as for that of the microbubble-encapsulated dolomite particles.

During the same time period, the Bureau of Mines conducted flotation tests on fine phosphate recovered from Florida operation slimes utilizing sodium silicate. The clay was first sized using hydrocyclones and hydroseparators, recovering up to 96% of the plus 400 mesh material, which was then used as a basis for the flotation test work (Zhang, 2001). The study used a fatty acid flotation at pH 9.0 with sodium silicate to depress the quartz in the rougher and cleaner flotation stages. At pH 9.0 the phosphate recovery was above 80% and the rougher concentrate grade was above 26% P_2O_5 (Jordan et al, 1982). The rougher concentrate grade fell quickly as the pH rose above 9.0. At a 3 lb/ton fatty acid and 2 lb/ton sodium silicate addition rates, a concentrate grade of 30% P_2O_5 and 90% recovery were obtained. Half the sodium silicate was added before the fatty acid collector; the rest was added prior to the cleaner flotation stage (Jordan, et al, 1982). While P_2O_5 concentrates were high, the overall P_2O_5 recovery from the total clay stream was low.

Shin and Choi (1985) continued the investigation of sodium silicate to better understand its selective flotation behavior among calcium minerals. Materials tested were wet-ground and sized to obtain a minus 270 mesh to plus 400 mesh fraction. The maximum amount of adsorption of sodium silicate was found to occur at pH 9.8. The amount of adsorption of sodium silicate increased with the increase in temperature of sodium silicate solutions (Shin and Choi, 1985).

Several years later, another in-depth study of the role of sodium silicate was conducted by Dho and Iwasaki (1990) on Florida phosphate ore. Their research led them to conclude sodium silicate can enhance flotation efficiency through: (1) the removal of impurity minerals and calcium-bearing precipitates from quartz surfaces by dispersion, (2) drier and more persistent froths stabilized by oily droplets containing calcium silicate precipitates, and (3) higher specific flotation rates of phosphate relative to quartz, leading to faster flotation rates and increased selectivity of separation (Dho and Iwasaki, 1990). Their research involved frothability tests and continuous and batch flotation tests to compare flotation results with and without sodium silicate.

Testing was conducted with a (1:1) fatty acid-fuel oil mixture, a 3.22 SiO_2/Na_2O sodium silicate ratio and ammonia as the pH modifier. The material was conditioned for 90 seconds at pH between 9.2 and 9.4 and diluted to 25% solids prior to flotation in a 2-liter Denver laboratory flotation cell. Several observations were noted during the flotation testing, including:

- The tests showed more stable froths in the presence of sodium silicates than in the absence of sodium silicates.

- The use of sodium silicate improved both the specific flotation rates and the coefficients of mineralization of phosphate, thereby leading to increased relative floatability of the phosphate.

- Increasing the amount of sodium silicate prevented the entrainment of quartz particles in the froth, making it more stable (Dho and Iwasaki, 1990).

Factors adversely affecting the efficiency of anionic flotation of phosphate in the absence of sodium silicate can decrease the overall concentrate grade (Dho and Iwasaki, 1990). $CaCO_3$ is a common precipitate in plant water, caused by the presence of Ca^{++} ions, which adversely affects flotation. When Ca^{++} is mixed with sodium silicate, calcium and silicate ions interact, resulting in the formation of calcium silicate precipitates. The main dispersive action of sodium silicate on quartz is produced by electrostatic repulsion due to calcium silicate precipitates formed on the surface of quartz (Dho and Iwasaki, 1990). The calcium silicate eventually detaches as the zeta potential of calcium silicate and quartz decreases, leaving the quartz surfaces virtually free of precipitates and thus is depressed (Dho and Iwasaki, 1990).

The aforementioned test work was conducted using the conventional froth flotation technique. Several researchers have successfully floated fine phosphate from clay slimes in a column flotation cell. Fine phosphate particles (<45µm) from an Egyptian phosphate mine were recovered utilizing a laboratory flotation column with a diameter of 5.04 cm by 361 cm high. Testing included the use of oleic acid as the phosphate collector and sodium silicate as the silica depressant. Test results included P_2O_5 concentrate of 25.3% with a 51.52% recovery (Abdel-Khalek *et al.*, 2000). Similar results could not be attained with conventional flotation techniques.

3.2.2.2 Dolomite Recovery

More recently, the Florida Institute of Phosphate Research (FIPR) conducted a study to optimize the Chinese dolomite collector, "PA-31" for use in the United States phosphate pebble industry. The Chinese Lianyungang Design and Research Institute (CDRI) had previously demonstrated the ability to reduce dolomite content in mainland China phosphate ores and were willing to assist the United States phosphate industry with their findings. The dolomite

collectors USPA-31 and FAS-40A, developed by two Florida local reagent producers, produced concentrates of more than 30% P_2O_5 and less than 1.0% MgO, with overall P_2O_5 recoveries averaging approximately 79% (FIPR 02-150-197, 2003).

In 2003, another FIPR project (No. 00-02-145, 2003) concluded it was possible to separate dolomite from apatite by coating it with a surfactant and immersing it in a dilute acid solution where the dolomite generates carbon dioxide gas that is trapped by the surfactant and floated to the surface (FIPR 00-02-145, 2003). The preliminary study examined the effects of over a dozen different surfactants and dosages on their amenability to dolomite flotation. El-Shall and Stana found polyvinyl alcohol (PVA) to be the most promising surfactant for coating the mixture of phosphate rock and dolomite to effectively separate the two minerals. The ore is first immersed in a 3% PVA solution and mixed well prior to being added to acidic water (3% sulfuric acid). The sink product of dolomite flotation was further upgraded by either silica or phosphate flotation, achieving concentrates near 64% BPL and 1-2% MgO. While this research looks promising, additional testing will be needed on finer material to determine its feasibility in the industry.

3.2.3   Other Recovery Mechanisms

One of the first tests conducted on phosphate slimes material that didn't incorporate solely flotation testing was in 1992 at PCS Phosphate in North Carolina (TexasGulf). The North Carolina Minerals Research Laboratory (MRL) conducted a preliminary study on the effectiveness of using hydraulic classifiers (Linatex Hydrosizer) to improve the beneficiation process of North Carolina phosphate ores. The main objective of the research was to efficiently deslime feed ore (14 mesh x 0) at a cut size of 200 mesh. Less than 2% of the P_2O_5 value was

lost to the overflow, while over 96% of the minus 200 mesh was removed at a throughput of 0.27 stph of feed per ft^2 of overflow area (Schlesinger, L.; Hutwelker, J, 1992). Prior to this study, the Linatex Hydrosizer had only been fully tested on 28 x 100 mesh material. The initial test work was deemed a success and additional testing was proposed but never carried out.

While flotation is the most typical separation process, there have been several other mechanisms proposed to facilitate the separation of phosphate and gangue minerals. One of the earliest accounts of this is a patent issued by Hazen Research in 1969 (Patent No. 3,425,799) to leach the slimes with sulfuric acid under conditions that allowed crystals of calcium sulfate to form and function as a filter aid. The leach liquor was then treated with an amine solvent and processed through a solvent extraction step using ammonia. The phosphate values were recovered as diammonium phosphate.

The Florida Institute of Phosphate Research sponsored a project that utilized an autoclave acidulation technique to recover phosphate from the clay stream (Zhang, 2001). At high temperature and pressure, P_2O_5 was quickly recovered and minimal clay residue (35-45%) remained as a by-product. The project was never brought to full scale due to the high capital investment cost and inconsistent acid availability.

3.3   Testing

3.3.1   Equipment Setup and Sample Acquisition

Sample acquisition took place during January 2004 at PCS Phosphate in White Springs, Florida. A 2-inch line from the clay launder located fifty feet above the washer floor fed a 6-inch diameter Krebs hydrocyclone and sump as shown in Figure 3.2. The hydrocyclone was initially set up with a 1.25-inch apex and a 2.5-inch vortex finder.   The cyclone/sump configuration was a semi-closed system as a majority of the cyclone overflow was circulated through the system while samples were collected.

Eight 5-gallon buckets of fine clay refuse stream were collected over the four day period. Three 55-gallon drums of cyclone underflow were collected and dewatered during the 4-day testing period for the subsequent conditioning and flotation test work. Additional samples of the cyclone underflow were collected, to be used as a composite of the larger bulk sample and for subsequent "settling" tests. These tests were used to determine the size of the required tanks for a full-scale operation. The material was screened at 400 mesh and analyzed for BPL and insol content.

After sampling was completed, all samples were shipped to Eriez Magnetics Central Research Lab (CRL) in Erie, Pennsylvania. Here, the samples were thoroughly characterized to determine the size distribution of the phosphate and gangue minerals.   The concentrated underflow material was run through a 2 x 6 inch CrossFlow separator to wash the slurry of the minus 400 mesh clay material as shown in Figure 3.3. The plus 400 mesh material obtained from the CrossFlow unit was dried, riffle-split and divided into 250 gram charges to be used as flotation feed.

Figure 3.2. Hydrocyclone Set-Up for Slimes Collection.

Figure 3.3. Laboratory CrossFlow Unit (2x6 Inch) Utilized for Sample Concentration.

3.3.2 Plant Hydrocyclone Testing

A 6-inch diameter Krebs Hydrocyclone was utilized during the on-site testing and sample acquisition for future evaluations of a full-scale unit. The hydrocyclone was placed in series with the large mixing sump, and the circuit was set up to run continuously in a closed-loop configuration. The cyclone was situated such that the overflow and underflow could return to the feed sump by gravity.

The cyclone was optimized to give the optimum size separation by varying feed pressure, vortex finder and apex parameters. The test rig was run at pressures between 20 – 28 psig with three different apexes and two different vortex finders. An analysis of these tests parameters would determine the best suited configuration to maximize the recovery of plus 400 mesh phosphate values to the underflow while simultaneously rejecting the maximum amount of clay slimes to the overflow. Five tests were conducted using the parameters as defined in Table 3.1.

Table 3.1: HydroCyclone Testing Parameters at Florida Phosphate Plant.

Test No.	Operating Pressure (psi)	Apex Size (inch)	Vortex Finder (inch)	Feed Flow (gpm)	U/F (gpm)	O/F (gpm)
1	27	1.00	2.5	190.7	16.70	174.00
2	20	1.00	2.5	na	27.30	na
3	26	0.75	2.5	na	7.70	na
4	20	0.75	2.0	160.30	10.30	150.00
5	28	0.75	2.0	250.00	10.30	239.70

Underflow and overflow samples were taken for each set of test parameters and assayed for a complete characterization and material balance at plus 150 mesh, 150x270, 270x325, 325x400 and minus 400 mesh. Table 3.2 summarizes the size distribution for each test parameter.

Table 3.2: Size Distribution for Hydrocyclone Tests 1-5.

Size					Cyclone Underflow Percent (%) Retained					
					Test Number					
Passing	Retained	Passing	Retained	Mean	1	2	3	4	5	Average
***	100	250	150	193.6	1.52	0.53	1.13	0.19	0.13	0.70
100	150	150	106	126.1	3.48	1.00	1.01	0.58	0.26	1.27
150	270	106	53	75.0	82.25	19.70	9.54	7.30	6.41	25.04
270	325	53	45	48.8	2.51	15.30	18.45	4.42	5.51	9.24
325	400	45	37	40.8	1.24	6.88	2.02	1.92	1.15	2.64
400	***	37	***	37.0	8.99	56.58	67.85	85.59	86.54	61.11

As described later, the cyclone underflow collected in this step was further processed in a hindered-bed classifier.

3.3.3   Detailed Testing

The concentrated bulk cyclone underflow material was next run through a hindered-bed classifier. The objective of this task was to remove any residual slimes that were inherently misplaced to the coarse underflow by the cyclone. Several hindered-bed evaluations were considered to optimize the condition of the underflow material, ideally producing the minimum amount of slimes. The operating parameters were bed level (pressure), fluidization rate, and feed characteristics (i.e., rate and percent solids).

The underflow from the hindered-bed classifier was then upgraded through conditioning and flotation evaluations. During conditioning evaluation, the optimum residence time, agitation intensity, and reagent dosage was established. The flotation testing includes determination of the optimum operating parameters, including aeration rate, feed rate and froth level. While ultimately the circuit will include a column cell, the initial test work was conducted with conventional flotation cells.

A column cell will offer better results for a variety of reasons. A conventional cell is normally run with an overflowing pulp level, which will increase entrainment and hydraulic carry-over of silica gangue, which must then be removed in subsequent amine flotation steps. However, in a column cell, where a froth level can be maintained due to the ultra-fine material, carry-over will be eliminated and minimize the silica entrainment. In addition, wash-water can be utilized which will rinse the product of entrained silica or remaining clays, prior to discharge.

Flotation test work was carried out during two separate testing periods. Initial flotation test work based on standard flotation techniques was not successful. Additional research was conducted on previous flotation testing of ultra-fine phosphate and adjustments to the testing program were made. All initial testing was conducted in a Denver Model D-12 laboratory batch flotation cell. Approximately 150 grams (dry basis) of the phosphate ore sample was conditioned at 50% solids by shaking it by hand in a sealed 250-mL Nalgene bottle. The pulp pH was adjusted with sodium carbonate ($NaCO_3$) and hydrochloric acid from 5.5 to 9.7 depending on the test. A total of 21 tests were conducted using various reagents and testing criteria. A summary of the test data is summarized in Appendix A.

As indicated in the literature review, a variety of techniques and reagents have been utilized during flotation testing of ultra-fine phosphate. Reagents showing the most promising results were collected and used in this testing program. A summary of these reagents is listed in Table 3.3.

Table 3.3: Reagents Used in Phosphate Flotation Test Work.

Reagent	Source	Description
Fatty acids:	Arizona Chemical	Sylva FA-1. Contains a low rosin percent
	PCS Phosphate	Standard FA/FO. Current flotation reagent
Amine:	PCS Phosphate	Current reagent
Na_2SiO_3:	PQ Industries	Formula "D" and "N". Used as a silica depressant
Alum:	Fischer Scientific	Used as a dolomite depressant.
H_2O:	Erie, PA water system	Used in all test work.

3.3.4   Process Evaluation

A representative sample of the fine clay refuse stream was screened at plus 150, 150x270, 270x325, 325x400 and minus 400 mesh. The size distribution of the feed is summarized in Table 3.4. A screen analysis of the cyclone underflow material was performed and size fractions were assayed for BPL, acid insol, Fe_2O_3, Al_2O_3, and MgO content as shown in Table 3.5. Settling tests were then conducted on this material for future scale up evaluation. The results of the settling tests are summarized in Table 3.6.

Table 3.4: Average Size Distribution of Feed Sample.

Size				Individual	Cumulative
Mesh		Microns		Mass	Mass
Passing	Retained	Passing	Retained	(%)	(%)
***	100	***	150	1.0	1.0
100	150	150	106	3.3	4.2
150	270	106	53	3.5	7.8
270	325	53	45	0.8	8.6
325	400	45	37	0.9	9.5
400	***	37	***	90.5	100.0

Table 3.5: Cyclone U/F Analysis by Size Fraction (Flotation Feed).

Size (Mesh)		Size (Microns)		BPL	Insol	Fe_2O_3	Al_2O_3	MgO
Passing	Retained	Passing	Retained		(%)	(%)	(%)	(%)
***	100	***	150	26.64	51.61	1.16	5.94	2.00
100	150	150	106	15.11	74.28	0.98	2.98	0.58
150	270	106	53	30.25	49.64	1.09	5.86	1.45
270	325	53	45	27.57	48.02	0.89	5.34	2.70
325	400	45	37	25.79	50.18	0.61	4.89	2.97
400	***	37	***	18.31	58.86	2.07	8.21	2.14

Table 3.6: Cyclone Underflow Settling Tests.

Size					Pre Settling Average	Post Settling Average	Post Settling Normalized Average	Percent Passing
Mesh		Microns						
Passing	Retained	Passing	Retained	Mean	(%)	(%)	(%)	(%)
***	100	265	150	199	4.1	2.8	6.0	100.0
100	150	150	106	126	0.9	0.6	1.3	94.0
150	270	106	53	75	3.5	2.4	5.1	92.7
270	325	53	45	49	2.1	1.4	3.0	87.6
325	400	45	37	41	1.0	0.7	1.4	84.6
400	***	37	***	37	88.4	38.7	83.1	83.1

Of the initial five tests conducted, an unacceptable 36% BPL was achieved with very high insol and magnesium content. The second series of testing was conducted several weeks later with alternative reagents and flotation techniques that provided more promising results. A total of 21 tests were conducted, with the best test achieving a 58% BPL, 1.75% MgO and 5.5% Al_2O_3. A summary of the concentrate and tailings BPL is shown in Figure 3.4, showing none of the tests achieved the 65% BPL minimum concentrate grade. As a result, a mineralogical study was undertaken to develop a better understanding of the potential reasons for the poor results (Section 3.3.6).
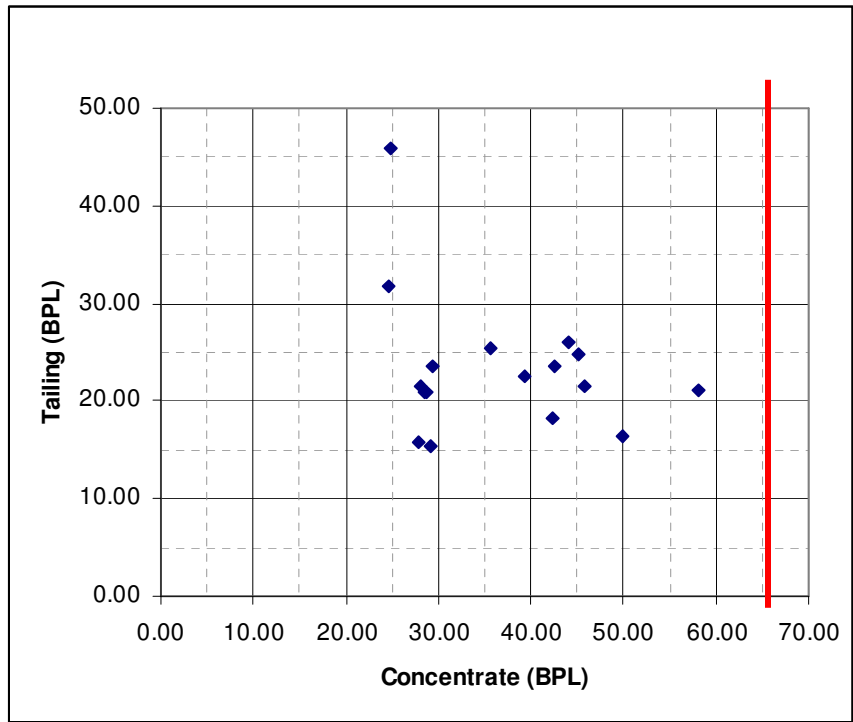
Figure 3.4: Concentrate and Tailings BPL for Second Series of Flotation Testing.

3.3.5 Sample Analysis

Detailed analyses were conducted on each sample collected during the test program. ASTM procedures and standards were employed for all determinations. Both Virginia Tech personnel and industrial participants of this project were responsible for complete assay determination and particle size analysis.

3.3.6   Mineralogical Investigation

3.3.6.1 SEM/EDX Introduction and Setup

The clay tailings sample acquired at PCS Phosphate in White Springs, Florida in January 2004 was analyzed by at the Department of Geological Sciences at Virginia Tech on the SEM/EDX. The Geological Sciences Department at Virginia Tech owns and operates a Cambridge Instruments Camscan II Scanning Electron Microscope (SEM) outfitted with an American Nuclear Systems System 4001 EDX spectroscopy system. The system produces a qualitative, not quantitative image by either back-scattered or secondary electrons. Over a dozen different particles were thoroughly analyzed during the 2 day period, of which this report describes nine of the most common. For each particle, a snapshot was taken along with the corresponding spectrum plot. A software program called Quantum Excalibur was utilized to view the spectrum. The spectrum plot measures the energy of each peak (in keV) versus the intensity of each mineral.

3.3.6.2 Results

Some general comments about what was seen on the SEM. First of all, most of the individual particles were covered with fine clay, making it difficult to get the exact components of the particle. The coating caused a significant amount of interference in the analysis. There are few samples that could be easily identified as exactly one mineral; virtually everything was a composition of Al, Si, P, and Ca, with small amounts of Mg, Cl, K, Fe and Ti.

Aluminum was associated with virtually all the particles that were analyzed, as most commonly aluminum phosphate and aluminum silicate clay. There were several calcium particles that were obviously dead marine life that were virtually all calcium. Some particles

were identified as potassium feldspar.   While these particles were not very common (approximately less than 20% of overall sample composition), there unique honeycomb shape made them easily identifiable during the analysis. The following is an explanation of the particle analyzed and identified by the SEM and Dr. Robert Tracy.

Figures 3.5 and 3.6 are *silica* particles. Silica was found to be present in both round, pebble-like material and very sharp, angular particles as can be seen by the variations in the two pictures.
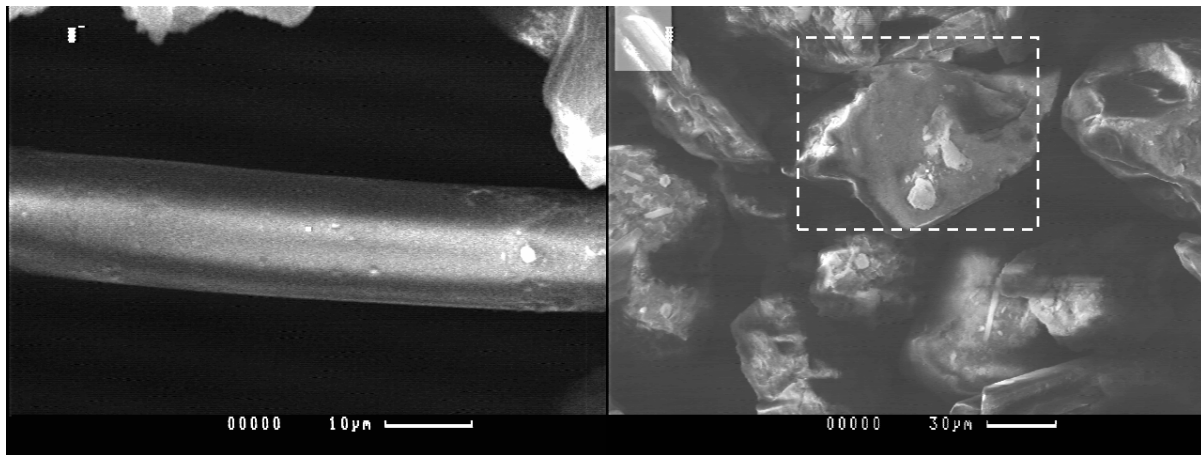
Figure 3.5: Silica Particle #1.   **Figure 3.6: Silica Particle #2.**

Figure 3.7: Spectrum Plot of Silica Particle #1.

ID(1):

Figure 3.8: Spectrum Plot of Silica Particle #2.

The honeycomb particles in Figure 3.9 are *potassium feldspar*. There is significant amount of aluminum associated with particle #2 as shown in the spectrum plot in Figure 3.10, but this was not the case with all of the potassium feldspar particles that were viewed.

Aluminum phosphate was the most prominent mineral (along with apatite) found during the analysis. Several pictures and spectrum were acquired to demonstrate the variations in which the aluminum phosphate is present in the sample. Figure 3.11 contains two particles, the top particle is apatite, and the bottom particle is aluminum phosphate. Figure 3.12 is another aluminum phosphate particle that contains more silica and calcium, which is believed to be interference from surrounding particles. A small peak of iron was also found on a similar particle of aluminum phosphate not shown here.
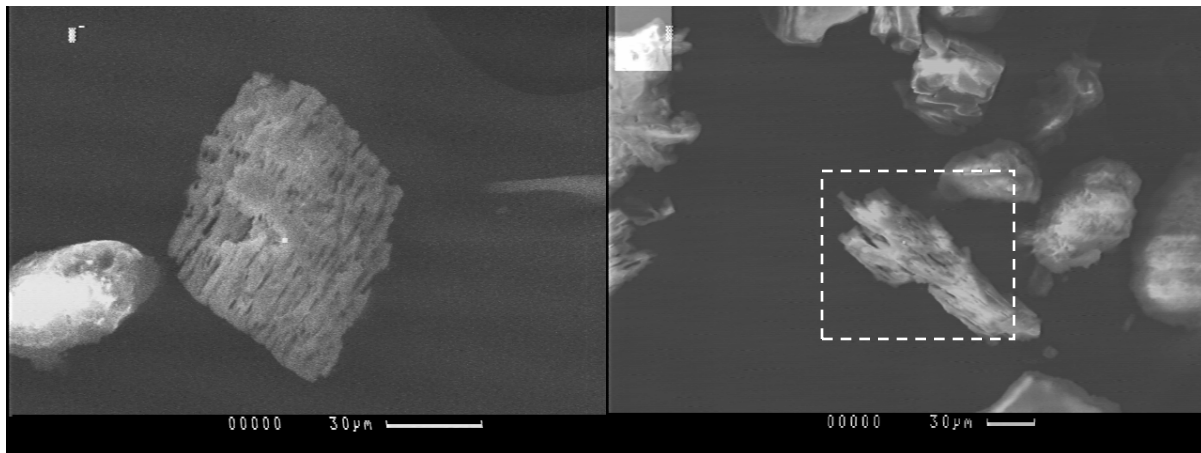
Figure 3.9: Potassium Feldspar Particles #1 and #2.
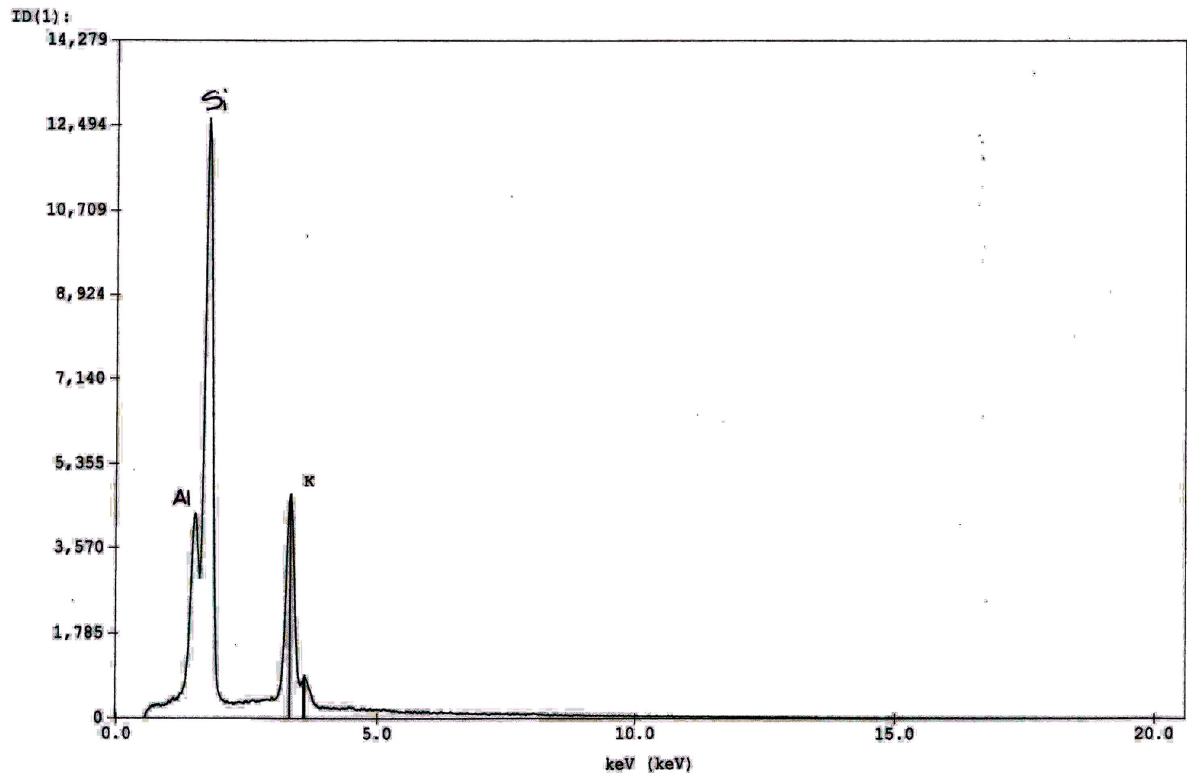
Figure 3.10: Spectrum Plot of Potassium Feldspar Particle #2.
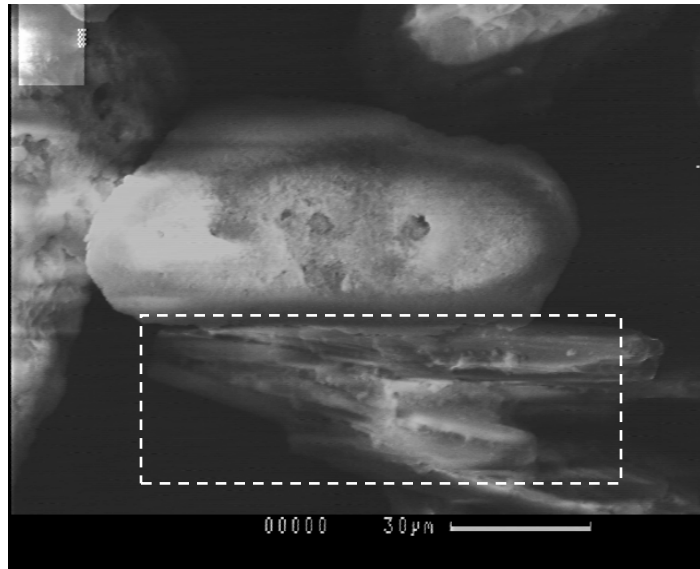
Figure 3.11: Aluminum Phosphate Particle #1 (with apatite).
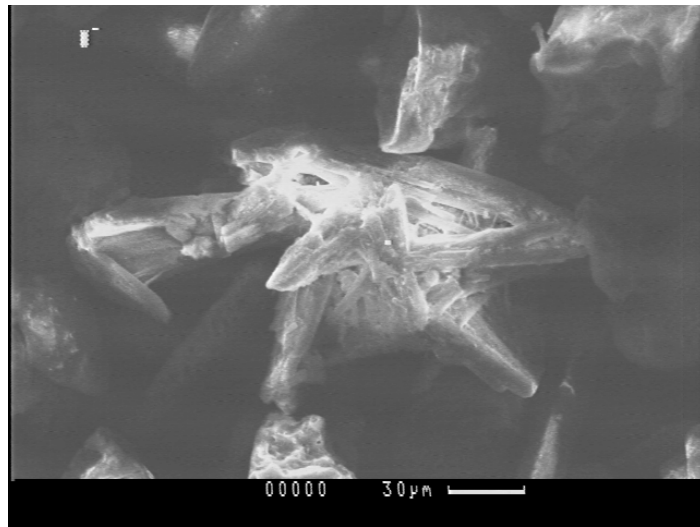
Figure 3.12: Aluminum Phosphate Particle #2.

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Spectrum Plot Routine
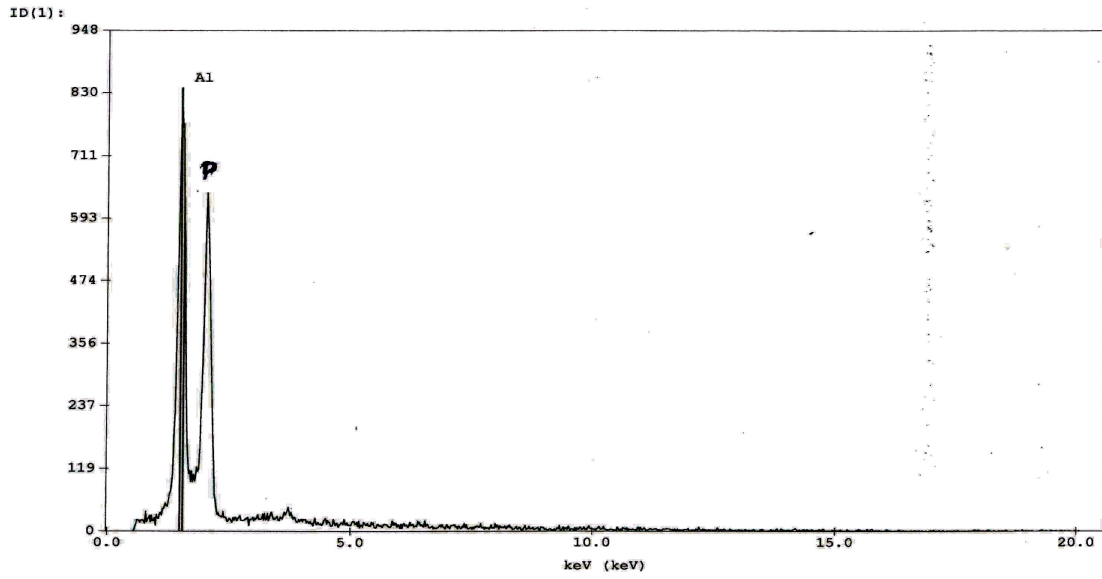
Figure 3.13: Spectrum Plot of Aluminum Phosphate Particle #1.

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ANS Quantum Software Report
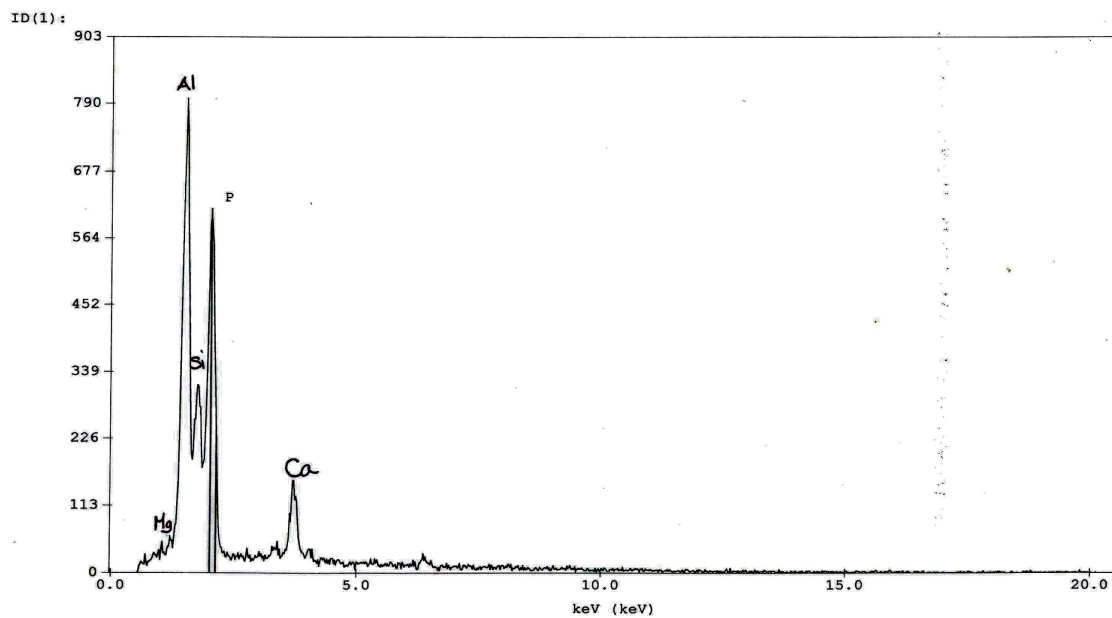Friday, July 16, 2004

Spectrum Plot Routine

Figure 3.14: Spectrum Plot of Aluminum Phosphate Particle #2.

Figure 3.15 below is another example of aluminum phosphate, with a very different structure than the previous two samples. The two samples to the right of the aluminum phosphate are apatite.
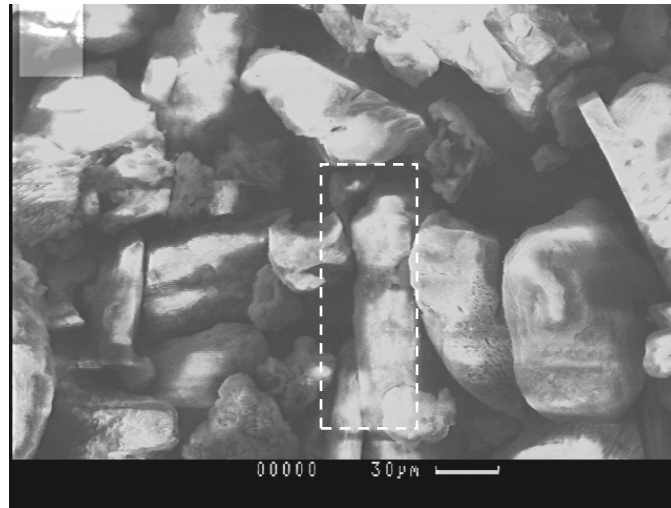
Figure 3.15: Aluminum Phosphate Particle #3.

The majority of *calcium phosphate*, or apatite, particles found were round and pebble-like in shape. Figure 3.16 is a typical example of an apatite mineral. Just to the left of the apatite particle is another example of potassium feldspar. Figure 3.11 is shown again (as Figure 3.17), this time with the apatite particle highlighted. It is believed aluminum interfered with the spectrum (Figure 3.18) on the analysis of this particle.
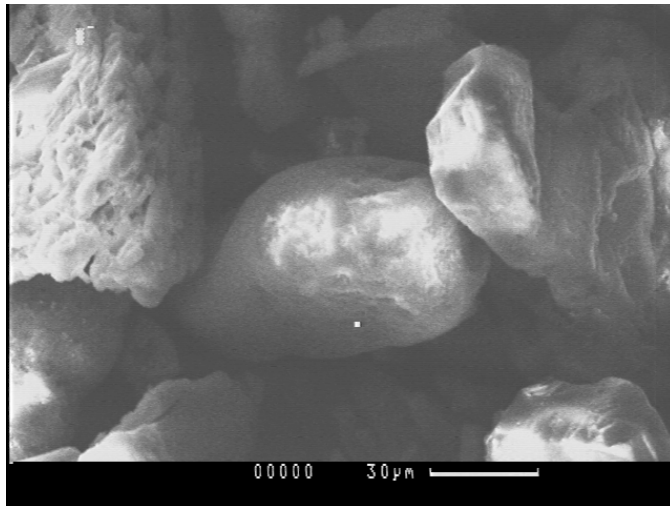
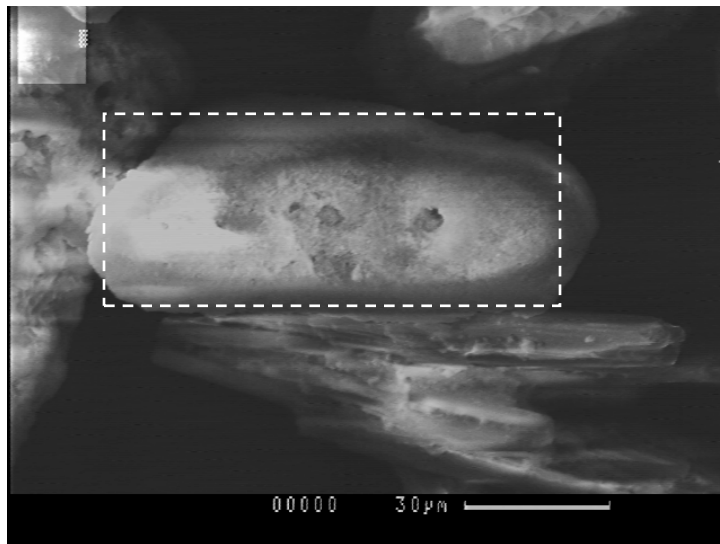**Figure 3.16: Apatite Particle #1**

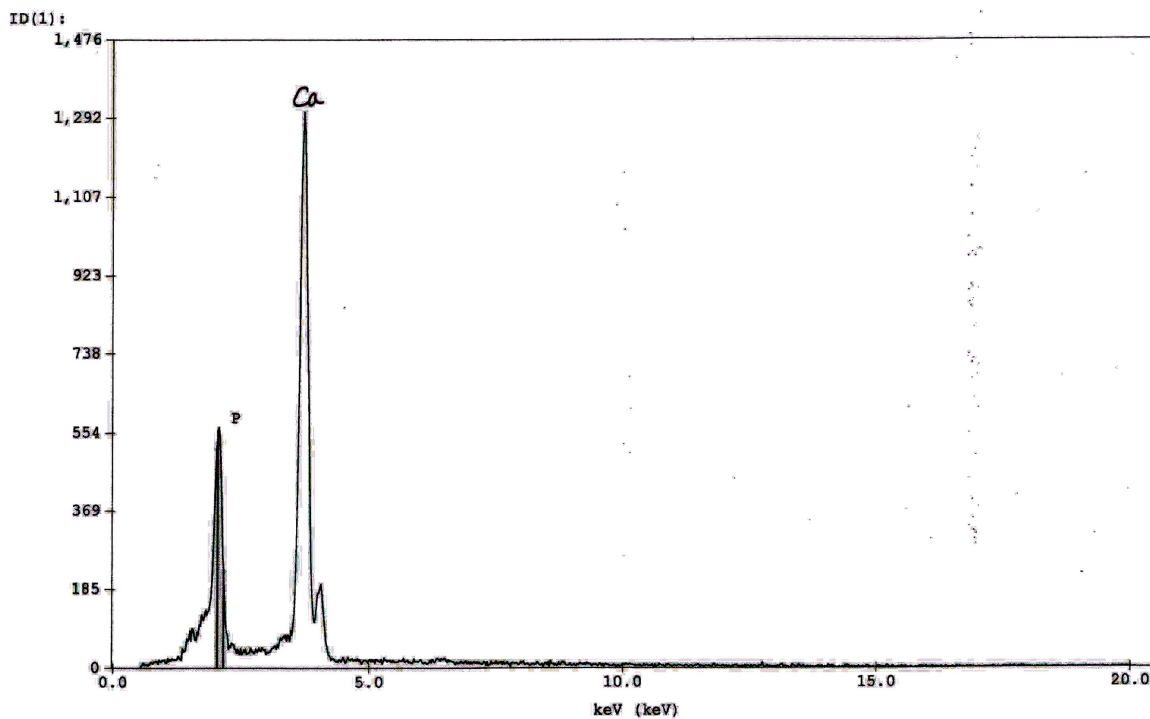**Figure 3.17: Apatite Particle #2 (with aluminum phosphate)**

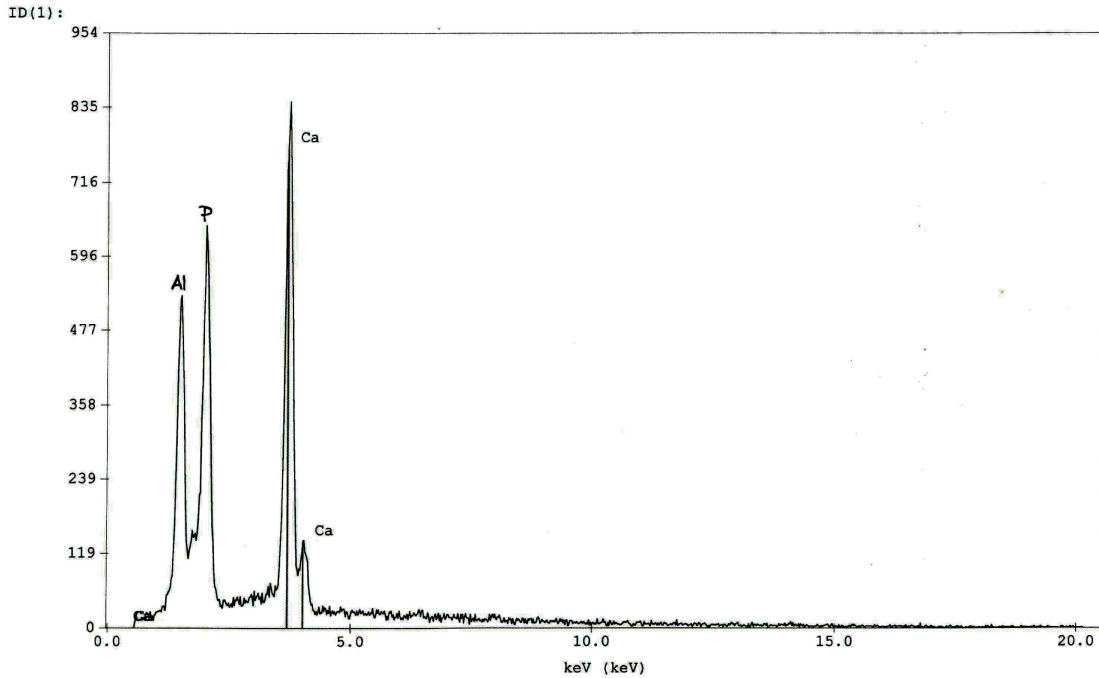**Figure 3.18: Spectrum Plot of Apatite Particle #1.**

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Figure 3.19: Spectrum Plot of Apatite Particle #2.

This mineralogical analysis supports the difficulties in the flotation testing described in Section 3.3.3. The presence of clay particles through the sample and the absence of pure apatite particles are detrimental to the successful concentration of phosphate particles by flotation methods. The sample was not as uniform as originally thought and as a result, further research is needed to better understand the interactions between the minerals.

3.4   **Future Work**

Further research should be conducted on the interactions of aluminum particles and apatite and/or clay. A better understanding of the effects of aluminum on phosphate flotation needs to be conducted to determine its exact interaction in the ore matrix and the existing reagents.

3.5   <u>Conclusions</u>

1. Based on the SEM/EDX analysis, the clay refuse stream sample appears to be composed of about equal amounts of calcium phosphate, aluminum phosphate and silica. The presence of potassium feldspar and calcium carbonate (seashell material) was visible, but not as prominent as the aforementioned. The presence of aluminum was significant in virtually all of the particles identified, where as very little magnesium was found in comparison. As expected, the presence of "pure" particles is minimal as the ultra-fine material has coated all particles of any significant size.

2. The flotation results follow what is shown in the mineralogical analysis. The absence of pure apatite particles and the presence of an assortment of clay particles inhibit the successful concentration of phosphate particles by flotation. A better understanding of the interactions between the clay and apatite particles is necessary before successful completion of this task can occur.

3. After a better understanding of the clay interactions is conceived, it is believed that an acceptable concentrate grade can be achieved with further refinement of the test criteria to minimize the clay/apatite interactions. Additional flotation test work is scheduled for the near future while further research is currently being conducted on the clay interactions.

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APPENDIX