**POC-SCALE TESTING
OF AN ADVANCED FINE COAL
DEWATERING EQUIPMENT/TECHNIQUE**

Final Report

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EXECUTIVE SUMMARY

This report describes the work performed under the U.S. Department of Energy Program Contract No. DE-AC22-94PC94155, monitored by the Federal Energy Technology Center (FETC), Pittsburgh, PA. The University of Kentucky Center for Applied Energy Research led the efforts on the program with active participation by the Powell Mountain Coal Company, Andritz Ruthner Inc., WesTech Inc., Decanter Machine Inc and Outokumpu Mintec. The research program was cost-shared by all the participants.

Program Objectives and Major Tasks

Dewatering of ultra-fine (minus 150 μm) coal slurry to less than 20 percent moisture is difficult using the conventional dewatering techniques. The main objective of the project was to evaluate a novel surface modification technique, which utilizes the synergistic effect of metal ions and surfactants in combination for the dewatering of ultra-fine clean-coal slurries using various dewatering techniques on a proof-of-concept (POC) scale of 0.5 to 2 tons per hour. The addition of conventional reagents and the application of coal surface modification technique were evaluated using vacuum filtration, hyperbaric (pressure) filtration, ceramic plate filtration and screen-bowl centrifuge techniques. The laboratory and pilot-scale dewatering studies were conducted using the fine-size, clean-coal slurry produced in the column flotation circuit at the Powell Mountain Coal Company, St. Charles, VA. The pilot-scale studies were conducted at the Mayflower preparation plant in St. Charles, VA.

The program consisted of nine tasks, namely, Task 1 - Project Work Planning, Task 2 - Laboratory Testing, Task 3 - Engineering Design, Task 4 - Procurement and Fabrication, Task 5 - Installation and Shakedown, Task 6 - System Operation, Task 7 - Process Evaluation, Task 8 - Equipment Removal, and Task 9 - Reporting. The program team consisted of personnel from the

University of Kentucky Center for Applied Energy Research, Powell Mountain Coal Company, Andritz Ruthner Inc., WesTech Engineering Inc., Outokumpu Mintec, and Decanter Machine Inc. All the participating organizations contributed to the program in terms of conducting tests and providing pilot-scale equipment for testing. The cost-sharing program was for 48 months which began September 30, 1994.

Results

The Powell Mountain Coal Company (PMCC) processes high- and low-sulfur coal from five different seams. The minus 100-mesh size fine coal is cleaned using the 'Ken-Flote' column flotation technology. The ash, solid content and the average particle size (D_{50}) of the low-sulfur clean coal slurry used for the laboratory study were 7.6%, 15%, and 29.5 μm , respectively. Similarly, the ash, solid content and average particle size (D_{50}) of the high-sulfur clean-coal slurry were 8.3%, 19%, and 37.3 μm , respectively.

The dewatering studies were first conducted in the laboratory to obtain optimal operating conditions for each technique. Later on, pilot-scale dewatering studies were conducted at the Mayflower plant using the optimized operating conditions found in the laboratory studies.

The reagents utilized for the study are listed below.

Anionic surfactant - sodium 2-ethylhexyl sulfate

Cationic surfactant - cetyl pyridinium chloride

Non-ionic surfactant - octyl phenoxy polyethoxy ethanol

Cationic flocculant - Allied Colloid (Procol 371, MW 5 million)

Anionic flocculant - Allied Colloid (Procol 156, MW 12 million)

Copper chloride (CuCl_2)

Aluminum chloride (AlCl_3)

For the laboratory studies tap water, and for the pilot-scale studies the plant's recycle water was used for preparing reagent solutions.

The dewatering results were analyzed using the following criteria:

- filter cake moisture
- solids throughput
- solids capture

It was observed during the pilot-scale dewatering testing that the baseline data varied significantly. This is primarily attributed to the changing of percent solids and size distribution in the column flotation product. Because of the significant variation, it was necessary to conduct baseline testing before and after each set of tests to ensure that any improvement observed in dewatering performance was not due to change in feed conditions.

Laboratory Studies

The baseline dewatering data for the low-sulfur clean-coal slurry using the four different techniques under various operating conditions are summarized in Figure 1. In these tests, no reagents were utilized. Note, that both vacuum and pressure (hyperbaric) filter using a 3 bar pressure provided a filter cake with about 23% moisture and the screen-bowl centrifuge provided a filter cake with 23.8% moisture. The solids capture in the vacuum and pressure filters was about 99%, whereas in centrifuge the solids capture was about 65%. The solids throughput in case of pressure filter was about 400 kg/m²/hr compared to about 7 kg/m²/hr for the vacuum and about 50 kg/hr for the centrifuge. The filter cake moisture obtained using the ceramic filter was about 17.0%, which was much lower than that obtained with the pressure, vacuum or centrifuge techniques. However, the cake thickness, which relates to solids throughput was less than 2 mm; according to the equipment supplier this cake thickness is not acceptable for a commercial unit.

A thicker filter cake could only be obtained using a higher percent (>40%) solids slurry, which resulted in a high (>30%) product moisture.

Figure 1. Laboratory baseline dewatering data of the low-sulfur clean-coal slurry.

Figure 2 shows the baseline dewatering data for the high-sulfur clean-coal slurry. This figure shows that among the three dewatering techniques, the pressure filter provided the lowest filter cake moisture of 20.3% using 3 bar pressure. The solids throughput in the pressure filter

was one order of magnitude higher than the vacuum filter. The screen-bowl centrifuge provided 20.8% moisture filter cake using an excessively high retention time of 3 minutes.

Figures 3 and 4 summarize the vacuum filter dewatering data using various types and amount of reagents for the low- and high-sulfur coal slurries, respectively. Note, that for both coals the addition of reagents is beneficial in lowering the filter cake moisture as well as increasing the solids throughput. The lowest filter cake moisture of about 19.6% was obtained using 0.3 kg/t of non-ionic surfactant for the low-sulfur, and a filter cake with a moisture of 22% was obtained using 10 g/t of the anionic flocculant for the high-sulfur coal slurry. For the high-sulfur coal slurry a combination of 0.2 kg/t CuCl_2 and 0.2 kg/t of the cationic surfactant provided filter cake with 23% moisture and solids throughput of about $33 \text{ kg/m}^2/\text{hr}$.

Figures 5 and 6 shows the pressure (hyperbaric) filter dewatering data with and without the presence of reagents for the low- and high-sulfur coal slurries, respectively. Note, that the low-sulfur coal slurry dewateres to 23% moisture and the high-sulfur coal dewateres to 20% moisture without use of any reagents. Addition of reagents in most cases increased filter cake moisture and solids throughput, except for the addition of the non-ionic and anionic surfactants, which provided filter cake with moisture similar to that obtained without addition of reagent, however, at twice the solids throughput.

Figure 2. Laboratory baseline dewatering data of the high-sulfur clean coal slurry.

Figure 3. Laboratory vacuum filter dewatering data using various reagents for the low-sulfur coal slurry.

Figure 4. Laboratory vacuum filter dewatering data using various reagents for the high-sulfur coal slurry.

Figure 5. Laboratory hyperbaric filter data using various reagents for the low-sulfur coal slurry.

Figure 6. Laboratory hyperbaric filter data using various reagents for the high-sulfur coal slurry.

The centrifuge filtration tests were conducted in a batch mode using a fixed amount of coal slurry for each test. For the low-sulfur coal slurry, the addition of 0.3 kg/t of the cationic surfactant was effective in lowering the filter cake moisture from 33.8% to 29.5%. Addition of 0.2 kg/t of CuCl_2 along with 0.3 kg/t of the cationic surfactant lowered the filter cake moisture to 27%. For the high-sulfur coal slurry, the addition of 1.5 kg/t cationic surfactant was effective in lowering the filter cake moisture from 29.3% to 27.8%. Other reagents were not effective in lowering the filter cake moisture. The solids capture in all the tests was higher than 90%.

Figures 7 and 8 summarize the dewatering data using the low- and high-porosity ceramic filter. For both filters, about 20 g/t of anionic flocculant was the most effective in lowering the filter cake moisture to about 14%. The solids throughput for both ceramic filters was 80 $\text{kg/m}^2/\text{hr}$, however, high-porosity filter provided higher solids throughput and higher filter cake moisture. Even though these results appear to be encouraging, the filter manufacturer recommended against using any reagent in the ceramic filter, as it plugs up the capillaries of the filter disk.

Pilot-Scale Testing

Figures 9 and 10 shows the pilot-scale vacuum drum filter dewatering data with and without the addition of surfactants for the low- and high-sulfur clean-coal slurry, respectively. Note, that for both the coal slurries the addition of reagents lowers the filter cake moisture. For the low sulfur coal slurry the addition of anionic surfactant provides filter cake with about 17 to 18% moisture; while the addition of cationic surfactant increases solids throughput with a filter cake with about 24% moisture. For the high sulfur coal slurry the addition of anionic surfactant provides filter cake with about 23% moisture, without changing the solids throughput as obtained with the baseline data, however, in this case the addition of anionic flocculant increases solids

throughput from 70 kg/m²/hr to about 225 kg/m²/hr providing a filter cake with about 25.5% moisture.

Figure 7. Laboratory ceramic plate dewatering data using reagents for the low-porosity filters.

Figure 8. Laboratory ceramic plate dewatering data using reagents for the high-porosity filters.

Figure 9. Pilot-scale vacuum drum filter dewatering data using various reagents for the low-sulfur coal slurry.

Figure 10. Pilot-scale vacuum drum filter dewatering data using various reagents for the high-sulfur coal slurry.

Figures 11 and 12 summarize the effects of reagents on dewatering of low- and high-sulfur coal slurry, using the hyperbaric (pressure) filter, respectively. For the low-sulfur coal slurry, the addition of anionic surfactant lowers the filter cake moisture from 26.5% to 25.0%, while the addition of the nonionic surfactant increases the solids throughput from 700 kg/m²/hr to about 900 kg/m²/hr. For the high-sulfur coal slurry, both non-ionic and cationic surfactants provide a filter cake with 19% moisture. Addition of both cationic and anionic flocculants increases filter cake moisture as well as solids throughput.

Figures 13 and 14 show the pilot-scale centrifuge dewatering data for the low-sulfur and high-sulfur coal slurries, respectively. For the low-sulfur coal slurry, the addition of reagents lowers the filter cake moisture. Addition of nonionic surfactant increases solids throughput. It should be noted that using 80 kg/t of the Orimulsion (Figure 13) lowered the filter cake moisture from 36.5 to 26.5%. In this case, solids recovery from the centrifuge of 95% was obtained compared to 60% in the other tests. The ash content of clean coal was lowered from 7.0% to 4.4%, and the dewatered product even with 26.5% moisture appeared dry. For the high-sulfur coal slurry, the filter cake moisture lowers from 29.5% to 27.0% with the addition of CuCl₂ and nonionic surfactant.

In summary, the laboratory scale dewatering test results were close to the pilot-scale tests. Table I and II summarize the best results obtained using the four dewatering techniques using the high- and low-sulfur clean-coal slurries, respectively. The ceramic plate data were obtained on a laboratory leaf filter only. Note, that the pressure filter provided the lowest filter cake moisture of 19.4% using 0.5 kg/t of the cationic surfactant. For the ceramic plate filter, a high 40% solids slurry was required to obtain a filter cake with 33% moisture

Figure 11. Pilot-scale hyperbaric filter dewatering data using various reagents for the low-sulfur coal slurry.

Figure 12. Pilot-scale hyperbaric filter dewatering data using various reagents for the high-sulfur coal slurry.

Figure 13. Pilot-scale screen-bowl centrifuge dewatering data using various reagents for the low-sulfur coal slurry.

Figure 14. Pilot-scale screen-bowl centrifuge dewatering data using various reagents for the high-sulfur coal slurry.

filter cake. Table III summarizes the processing (capital and operating) cost comparison of the four dewatering techniques using the optimum test conditions listed in Table I and II. The cost was calculated for processing 25 tph of dry solids as a 15% solids slurry. Note, that vacuum filtration cost is the lowest (\$0.48/t) of all the four techniques; however, the lowest filter cake moisture of 19.4% was obtained only with the hyperbaric (pressure) technique with a cost of \$3.37/t of the clean coal product.

Table I. Selected pilot-scale dewatering test data providing optimum dewatering of high-sulfur clean-coal slurry

Technique	R.P.M.	Pressure	Reagent	Amount of Reagent	Cake Moisture	Solids Captured
Vacuum	1	--	Anionic flocculant	15 g/t	25.8%	99.9%
Pressure I	1.5	3 bar	Cationic surfactant	0.5 kg/t	19.4%	99.9%
Pressure II	1.5	3 bar	Cu ⁺² Anionic Surfactant	25 g/t 0.25 kg/t	22.0%	99.9%
Centrifuge	1000 (226 'g')	--	Orimulsion	80 kg/t	26.5%	94.1%
Ceramic plate*	1.5	--	--	--	33.0%	99.9%

*based on laboratory scale tests; requires 40% solids feed.

Table II. Selected pilot-scale dewatering test data providing optimum dewatering of low-sulfur clean-coal slurry

Technique	R.P.M.	Pressure	Reagent	Amount of Reagent	Cake Moisture	Solids Captured
Vacuum	1	--	Anionic surfactant	15 g/t	20.8%	99.9%
Pressure	1.5	3 bar	Anionic flocculated	0.005 kg/t	26.4%	99.9%
Centrifuge	1000 (226 'g')	--	Cu ⁺² Cationic surfactant	25 g/t 1.5 kg/t	29.0%	70.0%
Ceramic plate*	1.5	--	--	--	3.0%	99.9%

*laboratory scale tests data, requires 40% solids

Table III. Processing cost comparison of the various dewatering equipment.
(25 TPH of fine coal, 15% slurry)

<u>Technique</u>	<u>Cost</u> (\$/t)
Vacuum	0.49
Pressure(I)	2.99
Pressure(II)	3.58
Centrifuge	3.37
Ceramic plate	2.09

Recommendations are offered to evaluate a large scale testing of the “GranuFlow” Process using the “Orimulsion” in a commercial plant. It is recommended that research efforts should be directed at developing a domestic bitumen emulsion similar to the “Orimulsion”.

INTRODUCTION

For the United States of America, it is forecasted that coal will constitute a principal source of energy for the next several decades. A typical coal preparation plant produces about 20 percent of the mined coal as minus 0.5 mm (28 mesh) material. Generally, this fine fraction is discarded due to its high cost of processing. However, with the development of advanced coal cleaning technology, such as column flotation, cleaning of fine-size coal to low ash and low pyritic sulfur content is feasible at high recovery. One of the biggest hurdles in utilization of fine-coal cleaning technology by the coal industry is the economic dewatering of the fine clean-coal product. Until an economical and practical solution to dewatering of fine clean-coal is achieved, the efforts devoted to developing fine clean-coal technology will be wasted.

In the U.S.A., about 40 percent of mined coal is cleaned before it is shipped to utilities companies. Most of the coal cleaning preparation plants utilize water-based processes. Water, while being the mainstay of coal washing, is also one of the least desirable components in the final product. Coarse coal (+3/4") is easily dewatered to a 3 to 5 percent moisture level using conventional vibrating screens and centrifuges. Aqueous suspension of particles finer than 0.5 mm are the most difficult to dewater. Even though in a coal preparation plant the fines (<0.5 mm) may constitute about 20 percent of plant feed, the high surface moisture retained by the fines offsets many of the benefits of coal cleaning and can seriously undercut utilization of advanced coal cleaning technologies which can provide an ultra-clean coal product.

Even a one-percent increase in clean-coal moisture can result in significant increase in transportation costs. A power plant using 3.0 million tpy of coal from a plant located 250 miles away might spend about \$350,000 per year to transport the additional one percent moisture.⁽¹⁾

This situation clearly identifies the need for an advanced dewatering technology and such technology could result in significant cost savings for the coal-using industry.

The degree of difficulty associated with dewatering increases as the particle size decreases or the surface area of the particles to be dewatered increases. For particles larger than 0.5 mm (28 mesh) size no particular dewatering problem is encountered. These large size particles are usually dewatered using vibrating screens and centrifuge techniques. Particles finer than 0.5 mm are the most difficult to dewater. A wide variety of equipment such as solid-bowl centrifuge, screen-bowl centrifuge and vacuum disk are used for fine particle dewatering, however, moisture content of the final product is high (~30%). Moisture content of the fine coal could be lowered using thermal drying, however, this is very unattractive due to high operating costs as well as air pollution and potential fire hazards.

Dewatering of fine-coal (minus 74 μm) to obtain less than 20 percent moisture product using the conventional equipment is not feasible. Currently, most of the coal preparation plants utilize vacuum filtration for dewatering of fine clean-coal slurries, providing a dewatered product containing about 25 to 30 percent moisture. It is a common practice in coal preparation plants to utilize vacuum filtration and thermal drying to meet the final product moisture specifications.

The current project was undertaken to evaluate a combination of novel surface modification techniques developed at the University of Kentucky Center for Applied Energy Research and dewatering equipment on a pilot scale to obtain a low (less than 20 percent) moisture product. The project team was comprised of personnel from the University of Kentucky Center for Applied Energy Research, Powell Mountain Coal Company, Andritz Ruthner Inc., WesTech Engineering Inc., and Outokumpu Mintec.

OBJECTIVES AND SCOPE

The main objective of the present project was to evaluate a combination of novel surface modification techniques, which utilize the synergistic effects of metal ions combined with surfactants and various equipment for dewatering of ultra-fine clean-coal slurries on a proof-of-concept scale of 0.5 to 2 tons per hour. The conventional reagents and novel surface modification techniques developed at the UKCAER were tested using vacuum, hyperbaric, ceramic filter, and screen-bowl centrifuge techniques. Dewatering studies were conducted on the fine clean-coal slurry produced in the column flotation circuit at the Powell Mountain Coal Company's (PMCC) Mayflower Preparation Plant in St. Charles, VA. An economic comparison of the dewatering technologies was also conducted.

BACKGROUND

Although the term "dewatering" refers to removal of water, researchers have used different terms and descriptions to define water associated with particles. Tschamler and Ruiten⁽²⁾ classified five types of water associated with coal, including interior adsorption, surface adsorption, capillary, interparticle and adhesion water. The last three types of water are termed as "free" water and are potentially removable by mechanical techniques. However, the first two types of water, known as "inherent" moisture, can only be removed by energy-intensive techniques such as thermal drying.

A variety of techniques are now used in contemporary practice for dewatering coal. The type and efficiency of various dewatering equipment varies as a function of particle size. The degree of difficulty associated with mechanical dewatering increases as the surface area of the particles increases. Particles finer than 0.5 mm (28 mesh) present the greatest dewatering

difficulty. Table IV lists the theoretical amount of water present on various size coal particles. Note that 5 to 10 μm particles carry a significant amount of surface moisture.

There has been a rapid increase in research and technology involving developments to improve fine-coal dewatering. Parekh et al.⁽³⁾ have published a review on fine coal and refuse dewatering which also included new dewatering technologies being developed in dewatering of fine coal. Methods for improving filtration of fine coal can be divided into two groups, namely, equipment modification and process modifications.

Table IV. Effect of Particle Size on Amount of Surface Moisture Present on Coal Particles

Particle Size	External Surface Area	Surface Moisture* (wt. %)
2"	0.9	0.009
3/4"	2.4	0.024
28 mesh	90.2	0.90
200 mesh	601	6.18
10 microns	4511	54.7
5 microns	9023	131

*Assuming a 1 micron thick film of water

In equipment modification, Ama pressure filters have been shown to be capable of achieving a low clean-coal moisture level.⁽⁴⁾ High 'g' centrifuges capable of creating forces up to 2,000 times gravity have been reported to be successful in dewatering a froth flotation product to less than 12-percent moisture.⁽¹⁾ Coal Technology Corp. has developed a super 'g' centrifuge capable of achieving forces up to 4,000 times the force of gravity.⁽⁵⁾ Other new equipment includes The Shoe Rotary Press⁽⁶⁾, membrane pressure filter⁽⁷⁾, and electro-acoustic dewatering⁽⁸⁾ techniques.

Perhaps a more promising avenue to improve the dewatering of fine-coal lies in the category of process modifications. The dewatering of fine-coal particles has been shown to be improved through process modifications that involve viscosity reduction, flocculation and surfactant addition.

Enhanced Dewatering:

Two important aspects of the dewatering of fine-coal are the dewatering rate and the final moisture content of the product produced; the most desirable conditions would be to have a fast dewatering or filtration rate and a low product moisture. Filtrate clarity is not of primary importance in coal applications because filtrate water and any solids in it are generally recirculated within the dewatering circuit. Theoretical treatment of the dewatering process has concluded that product moisture, or residual moisture as it is sometimes termed, is reduced while the dewatering rate is increased with:

Increasing

- driving force (gravity, vacuum or pressure)
- permeability of the medium (filter cloth and filter cake)
- contact angle or hydrophobicity
- filter area, and

Decreasing

- viscosity of the filtrate
- cake thickness, and
- surface tension.

When considering these factors, the driving force, medium permeability and filter area will be determined by the dewatering device used. The other parameters are properties of the slurry to be dewatered and can be controlled by the use of chemical additives.

The interconnected voids or pores in filter cakes form irregular capillary tubes. This analogy is frequently used in soil mechanics to describe the pore structure of soils. If one considers the cake structure to simplistically be a bundle of capillary tubes, residual saturation can be related to capillary rise or the level of water within a capillary tube.⁽⁹⁾ The capillary rise formula is:

$$h = \frac{2\gamma \cos \theta}{g\rho R}$$

where h is the capillary rise, γ is the liquid/air surface tension, θ is the liquid/solid contact angle, R is the capillary radius, g is the acceleration due to gravity, vacuum or pressure and ρ is the liquid density. The primary mechanisms responsible for the improved dewatering achieved in laboratory studies with the surface modification treatment can be related to reducing the capillary rise.

Effect of Capillary Radius:

Maximizing the radii of the capillaries within the filter cake structure can be accomplished in several ways. The most obvious is to blend coarser material to effectively increase the average particle size and reduce the surface area of the solids to be dewatered. While this would decrease the filter cake moisture, the coarser material could be more effectively dewatered with equipment other than filters and the net moisture reduction would be minimal.

Flocculants can be used to increase the effective particle size by agglomerating fines which increases permeability and results in a faster filtration rate. However, during floc formation, water can be entrained within the floc structure which ultimately limits the amount of

water that can be removed and results in higher cake moisture. To reduce the floc size, surface chemical modification can be used to induce agglomeration if filtration is conducted at or near the zero point of charge (ZPC).

Effect of Contact Angle:

The contact angle is a measurement of the hydrophobicity of the solid particles to be filtered. The adsorption of surfactants onto solids can increase the contact angle and make the surface more hydrophobic.^(10,11) Increasing the contact angle can significantly lower the capillary rise as shown in the capillary rise equation. The extent to which surfactants can increase hydrophobicity is related to their structure. Dewatering efficiency of surfactants has been related to the HLB index, a calculated, dimensionless number related to the hydrophobicity of the surfactant. The HLB (Hydrophile-Lipophile Balance) System was developed by Atlas Chemical Industries in the late 1940s to choose an emulsifier for one or more ingredients. It is determined by quantification of the solubility in oil (low HLB) or water (high HLB) of the ingredients to be blended. With the assigned HLB of the ingredients and the relative amount of each, the HLB of the mixture can be calculated. Although originally conceived as a method to select appropriate emulsifiers, the HLB is frequently used to describe the hydrophobic/hydrophilic properties of surfactants. Better dewatering performance has been obtained with low HLB surfactants (HLB <14) which are more hydrophobic.

Effect of Surface Tension:

Numerous studies have shown that surfactants can indeed lower the moisture content of coal filter cake during dewatering. The general mechanism is that surfactants reduce the interfacial tension at the liquid/air interface which reduces capillary retention forces.^(12,13) Reduction of interfacial surface tension at the liquid/air interface will lower the capillary rise, hence reducing the moisture contained in the capillaries.

Literature evidence challenges the importance of surface tension in dewatering. Silverblatt and Dahlstrom⁽¹²⁾ reported that the moisture content obtained at surface tensions of 72 and 32 dynes per centimeter were essentially equal, however, moisture content was significantly decreased between 32 and 36 dynes per centimeter. The conclusion was that the improvement was due to surface reactions between the coal and the surfactants rather than a change in the liquid/air interfacial tension. Others have also shown that dewatering is more closely related to surfactant adsorption than to surface tension reduction.^(10,11,14) Dewatering aids have been developed claiming increased adsorption on solids, and measurements of the filtrate have shown very little decrease in surface tension but significant decrease in cake moisture.^(15,16,17)

Gray⁽¹⁸⁾ used flocculants, oil and surfactants and reported that each improved dewatering. The benefit of the oils may have been due to impurities of surface active components in the oils, while evidence of surfactant adsorption onto the coal was reported even though his work assumed that surfactants only lowered the liquid/air interfacial tension. Dolina and Kominski⁽¹⁹⁾ used several surfactants during vacuum filtration and found that the residual moisture content of the filter cake decreased or increased depending on which surfactant was used. This is not surprising because solution pH and electrolyte content have a significant influence on surfactant adsorption.⁽²⁰⁾ Nicol⁽¹³⁾ reported improvements in dewatering of coal using anionic surfactants

while Brooks and Bethell⁽²¹⁾ found that a cationic surfactant also improved dewatering. Keller et al.⁽¹⁰⁾ showed that surfactants added during the washing period of the filtration cycle improved dewatering by i) changing the pressure differential required for dewatering and ii) lowering the residual water content of the filter cake. The pressure differential required was correlated with a decrease in surface tension while the residual moisture content was related to surfactant adsorption at the solid/air interface. Cationic, anionic and non-ionic surfactants showed similar effectiveness at different dosages.

The use of surfactants as filtration aids shows potential benefits for lowering cake moisture, particularly in longer filtration cycles. There is contradictory evidence in the literature pertaining to the mechanism responsible for this improvement. Some evidence exists for surfactant adsorption increasing the hydrophobicity of the substrate. Additional research suggests that lowering the surface tension at the solid/liquid interface is the primary mechanism. Regardless of the mechanism responsible, removal of "free" or "surface" water is the primary objective and the addition of agglomerating as well as surface tension modifying reagents can indeed reduce the moisture content of fine coal.

Effect of Gravity, g:

In the capillary rise formula given above, increased gravitational forces minimize the capillary rise. In filtration processes, for fluid to flow through the medium (filter cake and filter medium), it is necessary that a pressure drop be applied across the medium. The driving force to achieve the pressure drop can be gravity, vacuum or pressure. Increasing the driving force will increase moisture removal from the capillary network in the filter cake. Gravity filters are not used in the mineral industry, rather a similar principle is employed by high-speed centrifuges where forces several hundred times greater than gravity are generated. Pressure, and much more

frequently, vacuum are commonly used as driving forces in the mineral industry. The basic principle remains essentially the same.

Industrial Practice:

Vacuum filtration is by far the most commonly practiced dewatering technique in the U.S. coal industry. Vacuum disc filters are generally chosen over rotary drum filters because of the higher capacity. The production rate is approximately 290 to 340 kg/m²/hr for clean coal and 90 to 150 kg/m²/hr for coal refuse.

Centrifuges are also used for dewatering minus 0.5 mm (28 mesh) coal and refuse. The centrifuges are mounted horizontally and a high speed of rotation forces a particle bed to form at the centrifuge wall so that most of the water can be decanted. For a screen-bowl centrifuge, the solids are then conveyed to a chamber where the walls are perforated and further dewatering takes place through the walls. For a solid-bowl centrifuge, the solids are advanced to the discharge point by a scroll-conveyor and additional water drains back into the decantation zone. Screen-bowl centrifuges have a much higher throughput, however solid-bowls produce a lower moisture product for capacities of 2 to 30 tons per hour.

Pressure filtration is widely practiced in Europe and has been attempted at several preparation plants in the U.S. Most applications are for refuse dewatering, although some clean-coal applications have been reported. Plate and frame filters are the most common and employ high pressure to force water through a filter medium while retaining the solids as a cake. While pressure filters produce very dry cakes, their application in the coal industry is limited because it is a batch process. To dewater slurry continuously requires two filters and/or a surge tank which significantly increases capital costs. Continuous pressure filtration devices have recently been developed but have not been used in the U.S. coal industry. Several European installations

report low moisture and high throughput for fine coal applications. The Andritz-Ruthner Hyperbaric Filter (HBF) is one example and consists of a high specification vacuum disc filter installed in a pressure vessel and applies dewatering pressures up to 90 psi (6.2 bar). Commercial units vary in size from 260 to 1300 ft² (24 to 120 m²) of filter area and under plant operating conditions with fine coal has produced 17.9 percent moisture versus 25.7 percent on a vacuum filter at Ruhrkohr mines in Germany.⁽²²⁾

Several other dewatering devices such as the belt filter press or the vacuum belt filter press are also used, but their primary application is for refuse dewatering. Despite relatively high capacities, high chemical consumption and high product moisture limit their use.

Dewatering of ultrafine clean-coal ($D_{50} \sim 25 \mu\text{m}$) produced in advanced flotation technology are difficult to dewater to a low (less than 20 percent) moisture due to large surface area. Table V list dewatering test data obtained on a column flotation product using various types of equipment.⁽³⁾ Note, that only pressure filters were able to provide a low moisture product.

In-Situ Dewatering/Hardening

Recently, Wen et al.⁽²³⁾ have investigated an in-situ cake hardening process for fine coal slurry. They reported that addition of 2 to 8 weight percent of asphalt emulsion lowered the moisture as well as reduced the dust formation. Wilson et al.⁽²⁴⁾ have reported success in a single stage dewatering and briquetting process in which Orimulsion is used as the dewatering and binding agent.

In summary, there is an immediate need to reduce the moisture content of fine-coal filter cakes to about or below 20 percent level. Treatment of coal slurry prior to filtration by the addition of flocculants and surfactants can provide some benefits but may not be cost effective.

The use of steam can be effective in dewatering but may not be economical. High pressure dewatering of coal may be effective in lowering filter cake moisture of the fine coal slurry, however, the capital and operating cost are high. The coal industry is seeking an economical and efficient dewatering process. Several novel surface modification approaches have been tested in the laboratory, however, systematic pilot-scale studies on the dewatering of fine coal using the reagents has not been conducted.

Table V. Dewatering Test Data on Clean Coal Slurry Using Various Types of Equipment

<u>Equipment</u>	<u>Rate</u>	<u>Moisture %</u>
Vacuum Disc Filter	4.9 lb/hr/ft ²	25
Horizontal Belt Vacuum Filter	15 lb/hr/ft ²	25
Belt Press Filter	2.8 tph/m	40
Belt Press Filter [*]	10 tph	30
Belt Press Filter [*]	20 tph	38
Plate and Frame Pressure Filter	--	22.7
Ama Filter (Continuous Pressure Filter)	21 tph	21.0

*Organic polymeric flocculant used

MATERIALS AND EQUIPMENT

The POC-Scale testing was conducted at the Mayflower Preparation Plant of the Powell Mountain Coal Company (PMCC) located in St. Charles, VA. The plant processes coals from five different seams and produces low-sulfur (compliance) and high-sulfur (noncompliance) coal. Classified minus-100-mesh (150 μm) material is cleaned using four eight-ft. (2.4 m) diameter, 22 ft. (6.7 m) high 'Ken-Flote' columns.

Coal Sample Analysis:

Ten 55-gallon drums of each of the low-sulfur (compliance) and high-sulfur (non-compliance) clean coal froth products were collected at the plant for laboratory testing. A representative sample of each coal slurry was characterized for particle size, percent solids, and ash content, etc. Figure 15 shows the particle size distribution of the low-sulfur and high-sulfur coals. The median particle size D_{50} are 29.5 μm and 37.3 μm for the low- and high-sulfur coals, respectively. Tables VI and VII list the particle size and ash distribution of the low- and high-sulfur coals, respectively. The amount of $-44 \mu\text{m}$ (-325 mesh) material in both the coal samples is in the high 50% range. The solids content and proximate analysis of the two coals are listed in Table VIII. It was observed that the solid content of the coal slurries varied during the pilot-scale testing.

Water Samples Analysis:

Samples of plant make-up water was analyzed along with the plant process water which was obtained by filtering one liter of each coal slurry. The water quality data obtained by the DC Plasma Emission Spectroscopy are summarized in Table IX. Note, that as expected the plant process water is high in Al^{+3} , Cu^{+2} , Na^{+1} , Cl^{-1} and SO_4^{-2} ions.

Figure 15. The particle size distribution of the low- and high-sulfur clean-coal slurries.

Chemicals: All the chemicals used were of technical grade. Cu^{+2} and Al^{+3} metal ions were obtained by dissolving their chloride salts in tap water for the laboratory studies and plant recycle water for the pilot-scale studies. Tables IX and X list the surfactants and flocculants used for the study.

Table VI. Particle size and ash distribution of low-sulfur clean coal froth slurry

Size (mesh)	Weight (%)	Ash (%)	Ash Distribution (%)
+100	7.77	3.00	3.06
-100+200	16.58	3.80	8.28
-200+325	16.39	4.80	10.34
-325+500	17.23	5.80	13.13
-500	42.04	11.80	65.19
Feed (calculated)	100.00	7.61	100.00
Feed (actual)		7.60	

Table VII. Particle size and ash distribution of high-sulfur clean coal froth slurry

Size (mesh)	Weight (%)	Ash (%)	Ash Distribution (%)
+100	8.96	6.30	6.65
-100+200	19.11	5.40	12.16
-200+325	16.91	6.20	12.36
-325+500	16.56	7.40	14.44
-500	38.46	12.00	54.39
Feed (calculated)	100.00	8.49	100.00
Feed (actual)		8.30	

Table VIII. Solids contents and proximate analysis of coal samples.

Coal Sample	Solids (%)	Ash (%)	Total Sulfur (%)	Volatile Matter (%)	Fixed Carbon (%)
Low-sulfur	14	7.60	0.84	31.70	59.78
High-sulfur	19	8.30	1.81	33.05	59.77

Table IX. Plant make-up and filtrate analysis

	pH	Cond (mohs)	Al ⁺³	Cu ⁺²	Fe ppm	Na ⁺¹	NH ₄ ⁺¹	K ⁺¹	Cl ⁻¹	SO ₄ ⁻²
Plant make-up water	6.98	6.04	<0.01	0.005	0.73	305	<0.01	21	23	120
Filtrate water	7.67	2.43	0.96	0.011	0.23	1261	<0.01	70	63	1035

Table X. List of surfactants investigated.

	Sodium 2-Ethylhexyl Sulfate	Octyl Phenoxy Polyethoxy Ethanol	1-Hexadecyl Pyridinium Chloride
Type	Anionic	Non-ionic	Cationic
Abbreviation	S2ES	OPPE	CPCI
Commercial Name	NAS 08	TRITON X-114	Cetyl Pyridinium Chloride
Molecular Weight	232	536	340
Manufacturer	Niacet Co.	Rohm & Hass	Sigma Chemical

Table XI. List of flocculants investigated. (Supplied by Allied Colloids).

Name	Weight Type	Molecular Density (million)
Procol 156	Anionic	12
Procol 371	Cationic	5

Experimental Apparatus:

The laboratory vacuum filtration tests were conducted with a continuous filter leaf dewatering system, as shown in Figure 16, that simulates actual plant vacuum dewatering conditions. The 6-cm diameter filter leaf was connected to a vacuum pump through a double moisture trap. A rectangular cell of 13.5 liter was used as the filtration cell to minimize vortex effects. The fine-coal slurry was continuously circulated using a peristaltic pump and a 10-liter sump. This vacuum dewatering system simulates the industrial vacuum filter system and offers

the advantage of forming filter cake without particle segregation. Unless otherwise specified, all tests were conducted using a vacuum of 63.5 cm (25 in. Hg) at pH 8.1 for the low-sulfur and pH 7.1 for the high-sulfur clean coal slurries. The slurry was conditioned with the reagent for 3 minutes before the filter leaf was submerged in the filter cell. The filter leaf was removed from the slurry after a given cake formation time (usually 40 seconds) and dried (dewatered) in air for additional 40 seconds. The filter cake was then removed from the filter leaf and the cake thickness was measured. The cake moisture was determined by drying the filter cake at about 80° C to a constant weight, using the following formula:

$$\% \text{moisture} = \frac{\text{weigh of dried filter cake}}{\text{weight of wet filter cake}} \times 100$$

Figure 16. Experimental setup for the laboratory vacuum filtration tests.

Centrifugal dewatering tests were conducted using an IEC Chemical Centrifuge (International Equipment Company) with a metal basket of 5-in. (12.5 cm) diameter, as

illustrated in Figure 17. Each test was performed using 1000 ml slurry and 30 second dewatering time which is defined as the duration of time that begins after the slurry is added to the centrifuge and ends when the centrifuge is turned off. It was observed that the centrifuge took an additional 1.5 minutes to reach a complete stop. Unless otherwise specified, all centrifugal filtration tests were conducted at 5000 rpm rotation speed.

Figure 17. Schematic illustration of the laboratory centrifuge filter.

Hyperbaric dewatering studies were conducted using the high-pressure filter system shown in Figure 18. The pressure of the filtration cell was monitored continuously. All tests were performed using 5 liter slurry and 1 minute of total dewatering time.

Figure 18. The laboratory hyperbaric filter system.

The ceramic disc filtration studies were conducted using two different porosity ceramic plates supplied by the Outokumpu Company. Figure 19 show the experimental steps utilized for the ceramic filter tests. For all the tests, the coal slurry was kept stirred during the cake formation stage to provide uniform distribution and deposition of coal particles on the filter. For the study, two different types of ceramic leaf filters were used, namely, high-porosity filter (water flow rate 3 liter/minute at 1 bar or 14.6 psig) and low-porosity filter (water flow rate 0.6 liter/minute at 1 bar or 14.6 psig).

(a)

(b)

(c)

Figure 19. (a) Ceramic plate leaf filter setup, (b) dewatering/drying of the filter cake, (c) removal of filter cake.

The cake sample from each filtration test was evaluated for moisture content. Solids capture for the centrifugal filtration was calculated by determining the weight of the solids in the filter cake and the filtrate.

Pilot-Scale Equipment:

Figure 20 shows the process flow diagram of the pilot-scale dewatering setup used at the Powell Mountain Coal Company. For all the pilot-scale testing, a slip stream of the column flotation concentrate (froth) was diverted into a 500 gallon tank. The slurry was kept stirred and fed by gravity to the pilot-scale units. Feed rate was controlled by a gate valve.

The pilot-scale vacuum drum filter utilized in this project and its operation is shown in Figure 21. The filter consisted of a 3-ft. (0.9 m) diameter by 2 ft. (0.61 m) wide rotary vacuum drum. The skid unit was supplied by the WesTech Engineering Company, Salt Lake City, Utah. The drum speed was varied from 0.2 to 1.5 rpm. Three different submerged levels (slurry level maintained in the filter tub) were evaluated e.g., full tub, -1 inch (2.5 cm) and -2 inch (5.0 cm) from the top level. The slurry feed rate was adjusted with a valve to maintain the desired submergence level.

For the high-pressure filtration testing a portable hyperbaric unit supplied by Andritz Ruthener, Inc. was utilized. The pilot unit had a disc of 4.6 ft. (1.4 m) diameter with 22 ft² (2m²) filtration area which was enclosed in a 8.2 ft. (2.5 m) diameter pressure vessel (Figure 22). The trailer-mounted unit had its own pumps and air compressor. For most of the tests, approximately 50 gallons/min. of slurry was utilized. Filter cake was discharged onto a belt conveyor every 45 seconds.

Figure 20. Schematic diagram of the flowsheet used for the pilot-scale dewatering testing.

(a)

(b)

Figure 21. (a) Pilot-scale vacuum filter. (b) Sample collection.

(a)

(b)

Figure 22. (a) Pilot-scale hyperbaric filter. (b) Filter disk inside the pressure chamber.

The pilot-scale screen-bowl centrifuge of 18 in. (0.45 m) diameter and 42 inch (1.06 m) long was supplied by the Decanter Machine Inc., Johnson City, Tennessee (Figure 23). The bowl speed was maintained constant at 1473 rpm (226 `g' force). To determine the recovery of solids in the centrifuge samples of weir overflow, centrifuge centrate and the dewatered coal were collected simultaneously for a fixed period of time.

All the dewatered samples of coal were analyzed for cake thickness and moisture content. The filtrate samples were analyzed for solids content to determine the solids capture.

(a)

(b)

Figure 23. (a)Pilot-scale screen-bowl centrifuge filter.(b) sample collection

RESULTS AND DISCUSSIONS

For the project, laboratory dewatering studies were first conducted to identify optimum operating conditions and later on the pilot-scale dewatering tests were conducted using the best operating conditions found in the laboratory studies.

Laboratory Dewatering Studies

Baseline Dewatering Tests:

The baseline dewatering tests were conducted without utilizing any reagents, while varying the operating parameters of each dewatering technique.

Vacuum Filtration:

Figure 24 shows the effect of cake formation time on solids throughput and moisture at a cake drying time of 40 seconds for the low-sulfur clean-coal slurry. Note, that both solids throughput and filter cake moisture are almost linearly dependent on the cake formation time i.e., with the increase of cake formation time from 20 to 60 seconds the solids throughput increases from 7 kg/m²/hr to 16.5 kg/m²/hr and the filter cake moisture increases from 22% to 32%. This indicates that longer cake formation time recovered more solids from the slurry, resulting in a thicker filter cake which lead to a higher cake moisture. It was observed that using a cake formation (CFT) time of 30 secs, variation in the cake drying time (CDT) had minimal effect on filter cake moisture.

Figures 25 and 26 presents solids throughput and cake moisture as a function of cake formation time and cake drying time, respectively, for the high-sulfur coal slurry. As expected, longer cake formation time gives rise to a higher solids throughput and a wetter cake (Figure 25), which was also observed in Figure 24 with the low-sulfur coal slurry. However, unlike the low-sulfur coal the filter cake moisture decreases from 26.0% to 21.8% with increasing cake drying

time from 30 to 70 secs (Figure 26). These data indicates that the high-sulfur coal dewateres to a lower moisture than the low-sulfur coal slurry. This could be due to the coarser particles

Figure 24. Effect of cake formation time on vacuum filtration of low-sulfur coal slurry.

Figure 25. Effect of cake formation time on vacuum filtration of high-sulfur coal slurry

Figure 26. Effect of cake drying time on vacuum filtration of high-sulfur coal slurry.

($D_{50} = 37.3 \mu\text{m}$) present in the high-sulfur coal slurry compared to finer particles ($D_{50} = 29.5 \mu\text{m}$) in the low-sulfur coal slurry.

Centrifugal Filtration

Figure 27 shows changes in the cake moisture and solids recovery as a function of centrifuge rotation speed for the high-sulfur coal slurry. Note, as the rotation speed increased from 2500 to 5800 rpm, cake moisture reduced from 43 to 23% and solids capture increased from 78 to 97%. The higher solids recovery achieved at higher rotation speed was possibly due to the increased difference in centrifugal forces (F) exerted on large and small particles ($F = a m$, where a is acceleration and m is the mass of particles) that rendered large particles to settle much faster than small ones, creating a compacted bed of large particles that prevented small particles from passing through the filter cake.

Figure 27. Effect of centrifuge speed on dewatering of high-sulfur coal slurry.

Figure 28 shows the cake moisture as a function of filtration time for the high-sulfur coal slurry at a constant rotation speed of 5000 rpm. As seen from the figure, a significant reduction in cake moisture (from 40% to 27%) is observed when the filtration time increases from 10 to 30 seconds. There is a moderate improvement in cake moisture reduction above 30 secs filtration time.

Similar centrifugal dewatering behavior was observed with the low-sulfur clean-coal slurry. Based on these baseline tests, the centrifuge rotation speed and filtration time were fixed at 5000 rpm and 30 seconds, respectively, in all centrifugal dewatering tests conducted using various reagents.

It should be pointed out that the rotation speed of 5000 rpm and the dewatering time of 30 seconds used in the centrifuge do not apply to a commercial centrifuge. Most of the time the centrifuge speed is about 1500 rpm and the dewatering time is about 10 seconds.

Figure 28. Effect of centrifugation time on dewatering of high-sulfur clean-coal slurry at 5000 rpm.

Hyperbaric Filtration:

In the hyperbaric filtration, the effect of two variables, i.e., cake formation angle (CFA) and pressure, was investigated on filter cake moisture and solids throughput. The CFA is the angle of the filter disk rotation measured from the horizontal position where the filter elements enter the slurry, through which cake formation occurs. Thus, the CFA is the angle of filter rotation through the slurry and is analogous to the more common term 'cake formation time.'

Figure 29 shows the effect of pressure variations on dewatering of low-sulfur clean-coal slurry using an 85° CFA. Note, that the filter cake moisture decreases from 26.0% to 23.2%; solids throughput increases from 344 to 457 kg/m²/hr and air consumption increases from 468 to 670 Nm³/t, as the pressure increase from 1 to 3 bar. At a high 5 bar pressure filter cake moisture increases to 28.3% and air consumption decreases to 300 Nm³/t.

Figure 29. Effect of varying pressure on dewatering of low-sulfur coal slurry.

Figure 30 shows the effect of varying pressure on the dewatering of high-sulfur clean-coal slurry. The dewatering behavior of this slurry is similar to the low-sulfur clean-coal slurry. Increasing pressure from 1 to 3 bar decreases filter cake moisture from 25.0 to 20.3%, whereas solids throughput increases from 276 to 417 kg/m²/hr and air consumption increases from 551 to 644 Nm³/t. A higher 5 bar pressure increases the filter cake moisture. The increase of moisture observed at higher pressure for both the coal slurries could be because high pressure closed all the capillaries in the filter cake, trapping the moisture by reducing cake permeability. This premise is supported by a lower air consumption at high pressure which indicates a compact cake which does not allow air to pass through.

Figure 31 shows the effect of CFA on cake moisture, solids throughput and air consumption for the high-sulfur coal slurry using 3 bar pressure. It shows that a CFA of 125° or higher is desirable as it provides higher solids throughput and lower air consumption. Based on

the Andritz Ruthner personnel experience, the tests using various reagents were conducted using 165° CFA and 3 bar pressure.

Figure 30. Effect of varying pressure on dewatering of high-sulfur coal slurry.

Figure 31. Effect of cake formation angle on dewatering of high-sulfur coal slurry.

Ceramic Plate Filtration:

The ceramic plate filtration tests were conducted only with the low-sulfur clean-coal slurry. Figure 32 shows the effect of cake formation time on filter cake moisture and solids throughput using a 30 sec. cake drying time (CDT) using the high-porosity ceramic filter. It shows that cake moisture increases from 24% to 28% and solids throughput increases from 48 to 120 kg/m²/hr (cake thickness < 2mm) as the cake formation time increases from 15 to 60 secs. Figure 33 shows the effect of cake drying time on filter cake moisture using the low- and high-porosity ceramic filters at a 30 sec. CDT. As expected, the filter cake moisture decreases from 23.6% to 18.5% for the high-porosity and 19.8% to 17.2% for the low-porosity filter, as the cake drying time increases from 15 to 60 secs. The solids throughput for the high- and low-porosity filters were 106 and 53 kg/m²/hr, respectively. The relatively higher loading of solids obtained with the high-porosity filter results in a high moisture filter cake. The thickness of the filter cake obtained with both the filters was less than 2 mm thick. The Outokumpu Mintec U.S.A. Inc. manufacturer of the ceramic filters indicated that ceramic plate filter requires a minimum cake thickness of 2 mm, as the scraping devices are fixed on the filter's frame at a distance of 1.0 mm from the plate surface. In order to make capacity calculation, 1.0 mm must be subtracted from the laboratory filter cake thickness. Based on the manufacturer's recommendation, the slurry was thickened to about 40% solids by settling, and the dewatering data obtained using the manufacturer's suggested baseline test conditions are listed in Table XII.

Figure 32. Effect of cake formation time on dewatering of low-sulfur coal slurry using the high porosity ceramic filter.

Figure 33. Effect of cake drying time on dewatering of low-sulfur coal slurry using the ceramic filter.

Table XII. Ceramic plate filtration data of thickened low-sulfur clean coal slurry. (40% solids, low-porosity filter).

<u>Test #</u>	<u>CFT</u> (sec)	<u>CDT</u> (sec)	<u>Cake Thickness</u> (mm)	<u>Cake Moisture</u> (Wt. %)	<u>Solids Throughput</u> (kg/m ² /hr)
1	10	15.5	3.2	35.2	79.4
2	13	21.0	3.5	33.7	79.0
3	21	34.0	4.1	34.1	108.7
4	4	19.0	1.8	21.3	39.9
5	5.5	25.0	2.0	22.3	26.9
6	9.0	41.0	2.6	24.2	38.6

Note, that the thickest cake (Test Nos. 1, 2 and 3) provided a filter cake with more than 33.0% moisture, these data were utilized for a scale-up of the equipment to process 25 tph of clean coal.

Effects of Reagents

Vacuum Filtration:

Low-Sulfur Coal:

Figure 34 shows the effects of flocculant dosage on dewatering of the low-sulfur coal slurry. As the cationic flocculant dosage increases from 0 to 25 g/t, solids throughput increases from 15 to 85 kg/m²/hr while cake moisture initially decreases from 26% to 23% then increases to 27%. With the anionic flocculant the solids throughput increases from 12 to 78 kg/m²/hr with increasing the dosage from 0 to 10 g/t and then leveling off. The cake moisture, on the other

Figure 34. Effect of flocculant dosage on vacuum dewatering of low-sulfur coal slurry at 30 sec CFT and 30 sec CDT.

hand, is nearly independent of the anionic flocculant dosage. As shown earlier in Figure 24, that cake moisture increases from 23 to 32% with increasing solids throughput from 7 to 16.5 kg/m²/hr in the absence of flocculant. However, it appears that the increase in solids throughput obtained with the flocculant was much higher and the increase in moisture was lower in the presence of flocculant than in its absence. In other words, a lower moisture filter cake can be obtained at a given cake thickness by the use of a flocculant. This is primarily due to formation of flocs that are considerably larger than individual coal particles.

Figure 35 shows the effect of cationic, anionic, and non-ionic surfactants on vacuum dewatering of the low-sulfur coal slurry. With the cationic surfactant, solids throughput increases from 12 to 33 kg/m²/hr with increasing the dosage of cationic surfactant from 0 to 2 kg/t. However, the lowest cake moisture (23.2%) is achieved using 1.2 kg/t cationic surfactant. Use of the anionic surfactant increases solids throughput from 12 to 15 kg/m²/hr at a dosage of 1.5 kg/t, however, cake moisture does not show any significant changes with an increase in the surfactant dosage. For the non-ionic surfactant as the dosage increases from 0 to 0.9 kg/t, solids throughput decreases slightly from 11.8 to 9.0 kg/m²/hr and cake moisture first decreases from 24 to 19.5% then increases to 21%. Further increases in the surfactant dosage to 1.5 kg/t increased solids throughput to 11.5 kg/m²/hr and cake moisture to 23%.

Figure 36 shows the effect of addition of di-valent metal ion Cu⁺² at pH 6.5 and tri-valent metal ion Al⁺³ at pH 9 on dewatering of the low-sulfur coal slurry. pH's 6.5 and 9 are the iso-electric-point (IEP) of coal in the presence of Cu⁺² and Al⁺³ ions, respectively. It has been reported that at the IEP's the metal ions are most effective in neutralizing the surface charge on coal particles and hence, enhance dewatering^{25,26}. As shown in the figure, the addition of Cu⁺²

Figure 35. Effect of surfactant dosage on vacuum filter dewatering of low-sulfur coal slurry(30 sec CFT: 30 sec CDT).

Figure 36. Effect of metal ion dosage on vacuum filter dewatering of low-sulfur coal slurry. (30 sec CFT; 30 sec CDT)

shows no significant change of cake moisture; but it considerably increases solids throughput, presumably due to particle aggregation at the IEP. Addition of Al^{+3} ions increases solids throughput and the cake moisture shows a slight decrease, then increases to 27% at a AlCl_3 dosage of 4.5 kg/t. The increase in cake moisture at higher dosages may be associated with the increased cake thickness.

Figure 37 depicts synergistic effects of adding a combination of a fixed amount of surfactant and varying amount of Cu^{+2} ions for the low-sulfur coal. Overall, the effects on cake moisture are minimal while slight increase in solids throughput are experienced. In the presence of 1.2 kg/t cationic surfactant the solids throughput is independent of the dosage of Cu^{+2} ions but cake moisture increases with increasing the dosage. This indicates that there were adverse synergetic effects between the cationic surfactant and Cu^{+2} ions with respect to dewatering of this specific coal sample, possibly due to competitive adsorption of Cu^{+2} ions on coal surface with cationic surfactant ions, both of which are cationic, resulting in decreased surface hydrophobicity. Figure 37 also shows that solids throughput increases from 20.0 to 25.5 $\text{kg/m}^2/\text{hr}$ with increasing CuCl_2 dosage to 1.3 kg/t in the presence of 1.5 kg/t anionic surfactant while cake moisture essentially remains unchanged. Addition of CuCl_2 ions in the presence of non-ionic surfactant consistently increases solids throughput from 19 to 24.5 $\text{kg/m}^2/\text{hr}$ as the dosage of CuCl_2 increases to 1.3 kg/t. The cake moisture shows only marginal changes with increasing CuCl_2 dosages.

Figure 38 shows synergistic effects of adding surfactants at a fixed dosage and AlCl_3 (Al^{3+} ions) at varying dosages for the low-sulfur coal slurry. In general, the addition of Al^{3+} ions in the presence of surfactant has minimal effect on both cake moisture and throughput. However, addition of AlCl_3 with surfactant increases solids throughput from 18 to 21 $\text{kg/m}^2/\text{hr}$

with 1.5 kg/t anionic and from 20.0 to 21.6 kg/m²/hr in the presence of 0.9 kg/t non-ionic surfactant.

Figure 37. Synergistic effect of surfactant and varying Cu²⁺ dosage on vacuum filter dewatering of low-sulfur coal slurry(30 sec CFT;30sec CDT).

Figure 38. Synergistic effect of surfactant and varying Al^{3+} dosage on vacuum filter dewatering of low-sulfur coal slurry (30 sec CFT; 30 sec CDT).

To better understand effects of various reagents on coal dewatering behavior, a cake parameter, referred to as cake porosity, was determined for each test described above. The cake porosity, ε , was defined as:

$$\varepsilon = 1 - \frac{4 \times \text{weight of wet cake} \times (1 - \text{moisture})}{\pi \rho_m D^2 \times \text{cake thickness}}$$

where

ρ_m : density of coal;

D: diameter of filter leaf, cm;

Weight of wet cake, g;

Cake thickness, cm.

Figure 39 shows cake porosity as a function of cake thickness for the low-sulfur coal slurry. It appears that cake porosity is independent of cake thickness and more importantly, addition of various reagents did not affect the cake porosity. Although cake porosity was relatively constant in all these tests, cake moisture differed considerably, as shown in Figure 40. This suggests that cake porosity is not important in determining cake moisture for this specific coal. Relatively low cake moisture, e.g., less than 23%, can be achieved by use of the flocculant or surfactants. Flocculants appear to be the most effective reagent in enhancing coal dewatering since they generated thicker and comparatively drier cake. The results also showed that the addition of surfactant or metal ion alone has adverse effects on cake moisture but produced a thicker cake.

Figure 40. Cake moisture as a function of cake porosity for the low-sulfur coal slurry.

High-Sulfur Coal:

Figure 41 shows the effect of cationic and anionic flocculant dosage on vacuum dewatering of the high-sulfur clean-coal slurry. As the dosage of the cationic flocculant increases, cake moisture increases from 23% to 27%. The solids throughput increases significantly from 12 to 33 kg/m²/hr as the dosage increases from 0 to 25 g/t. The cake moisture at the maximum cake thickness (25 mm) at 25 g/t cationic flocculant was approximately the same as a 3 mm thick filter cake in the absence of flocculant. This clearly demonstrates that use of the cationic flocculant significantly improves the fine coal dewatering. Increasing the dosage of the anionic flocculant to 15 g/t increased cake thickness from 4 to 22 mm. However, dosages higher than 15 g/t had no effect on cake thickness. The filter cake moisture shows a maximum increase of about 2% i.e., from 26.5% to 28.5%, while the cake thickness increases several times. Apparently, the anionic flocculant also enhances fine coal dewatering.

Figure 42 depicts the effect of surfactants addition on vacuum dewatering of the high-sulfur coal. Use of cationic surfactant at a dosage of 1.5 kg/t increases solids throughput from 22 to 33 kg/m²/hr and reduces cake moisture from 26 to 24%. Similarly, the use of 1.5 kg/t of non-ionic surfactant reduces cake moisture from 26 to 24%. Use of anionic surfactant marginally increases solids throughput and the filter cake moisture.

Figure 43 presents effect of addition of di-valent Cu⁺² metal ions at pH 6.5 and tri-valent Al⁺³ metal ions at pH 9 on vacuum dewatering of the high-sulfur coal slurry. The addition of 0.2 kg/t of CuCl₂ ions decreases moisture from 25.3 to 24%, however, further increasing CuCl₂ ions dosage increases filter cake moisture to 26.2%. The solids throughput increases from 21 to 24.5 kg/m²/hr as the CuCl₂ dosage increases from 0.2 to 1.5 kg/t. The addition of 3.5 kg/t of AlCl₃ increases solids throughput from 19 to 24.5 kg/m²/hr and cake moisture from 25.5% to 31.0%.

Figure 41. Effect of flocculant dosage on vacuum filter dewatering of high-sulfur coal slurry (30 sec CFT; 30 sec CDT).

Figure 42. Effect of Surfactant dosage on vacuum filter dewatering of high-sulfur coal slurry. (30 sec CFT; 30 sec CDT)

Figure 43. Effect of metal ions dosage on vacuum filter dewatering of high-sulfur coal slurry coal slurry (30 sec CFT; 30 sec CDT).

Figure 44 shows the synergistic effect of the addition of cationic, anionic and non-ionic surfactant at fixed dosages of 0.3 kg/t, 0.9 kg/t and 0.6 kg/t, respectively, while varying CuCl_2 dosages for the high-sulfur clean-coal slurry. Although the anionic and non-ionic surfactants had little effect, the cationic surfactant along with about 0.2 kg/t of CuCl_2 is effective in lowering the filter cake moisture from 24.5 to 23%, and increasing solids throughput from 23 $\text{kg/m}^2/\text{hr}$ to 34 $\text{kg/m}^2/\text{hr}$.

The addition of surfactants with AlCl_3 showed no effect on moisture reduction.

Figure 45 shows the cake moisture versus cake porosity for the high-sulfur clean coal slurry. The wide scattered data points indicate there is no correlation between cake moisture and cake porosity. In general, the cationic flocculant data shown with ($\diamond$), indicates it to be the most effective in lowering the filter cake moisture for the high-sulfur coal slurry.

Figure 46 presents the relation between cake thickness and cake moisture for the high-sulfur clean coal-slurry. It shows that flocculants as expected provide the highest cake thickness with a somewhat increase in moisture; however, as explained before, a thicker cake obtained without addition of a flocculant has much higher moisture than the one obtained with the flocculant, or in other words, flocculant in general provides a lower moisture filter cake.

Centrifugal Filtration

Low-Sulfur Coal:

The effects of cationic and anionic flocculants addition on cake moisture and solids capture in the centrifugal dewatering of the low-sulfur clean-coal slurry is shown in Figure 47. Addition of 15 g/t of the cationic flocculant decreases cake moisture from 31% to 27%; however,

Figure 44. Synergistic effect of surfactant and varying Cu²⁺ dosage on vacuum filter dewatering of high-sulfur coal slurry (30 sec CFT; 30 sec CDT).

Figure 45. Cake moisture as a function of cake porosity for the high-sulfur coal slurry

Figure 46. Cake moisture as a function of cake thickness for the high-sulfur coal slurry.

Figure 47. Effect of flocculant dosage on screen-bowl centrifuge dewatering of low-sulfur coal slurry(speed 5000 rpm; filtration time 30 sec).

further increase in the cationic flocculant dosage increases cake moisture. Solids capture, on the other hand, increased from 88% to 92% as the dosage of the cationic flocculant increased. Similarly, addition of 30 g/t anionic flocculant lowers cake moisture from 32% to 29.0% but also reduced solids recovery slightly. High solids recovery in the laboratory centrifuge was due to the smaller opening (~ 100 mesh) screen used for the study.

Figure 48 presents the effect of cationic, anionic, and non-ionic surfactants on the centrifuge dewatering of the low-sulfur coal slurry. Of the three surfactants, only the cationic surfactant (0.3 kg/t) was effective in lowering the filter cake moisture from 33.8% to 29.5%, a 4.3% absolute moisture reduction. The higher dosage of the cationic surfactant was detrimental to moisture reduction. Solids capture for all the surfactant remains more or less constant.

Figure 49 shows the effect of adding either di-valent metal ions Cu^{+2} at pH 6.5 or tri-valent metal ions Al^{+3} at pH 9 on dewatering of the low-sulfur coal slurry. Addition of 0.1 kg/t of CuCl_2 reduces cake moisture from 29.0% to 26.3% and increases solids capture from 85.2% to 86.9%. Further increase in CuCl_2 dosage increases filter cake moisture. Addition of 2.8 kg/t of AlCl_3 marginally lowers cake moisture from 29.0% to 28.0%. It appears from the data that Cu^{+2} ions are more effective than Al^{+3} ions in lowering the filter cake moisture.

The synergistic effect of adding a combination of surfactants at a fixed dosage of 0.3 kg/t of cationic, 0.9 kg/t of anionic and 0.6 kg/t of non-ionic and varying CuCl_2 dosage is shown in Figure 50. A combination of 0.5 kg/t of CuCl_2 and the cationic surfactant lowers the filter cake moisture from 31% to 27%; 0.5 kg/t of CuCl_2 and the anionic surfactant combination lowers the cake moisture from 31% to 28.2%. The non-ionic surfactant and 0.8 kg/t of CuCl_2 combination reduces cake moisture from 31.0% to 28.2%. Increasing Cu^{+2} ions concentration was

Figure 48. Effect of surfactant dosage on screen-bowl centrifuge dewatering of low-sulfur coal slurry (speed 5000rpm; filtration time 30 sec).

Figure 49. Effect of metal ion dosage on screen-bowl centrifuge dewatering of low-sulfur coal slurry (speed 5000 rpm; filtration time 30 sec).

Figure 50. Synergistic effect of surfactant and varying Cu^{2+} ion dosage on screen-bowl centrifuge dewatering of low-sulfur coal slurry (speed 5000 rpm; filtration time 30 sec).

detrimental for all the three surfactants. The solids recovery for all the three surfactants remains high and constant for all the dosages of Cu^{2+} .

Synergistic effect of surfactants with varying Al^{3+} ions concentration for the low-sulfur clean-coal slurry is shown in Figure 51. A combination of 0.9 kg/t AlCl_3 and 0.3 kg/t of the cationic surfactant lowers cake moisture from 31.0 to 26.8%. Addition of 1.8 kg/t of AlCl_3 and 0.3 kg/t of the anionic surfactant reduces cake moisture from 31.0% to 28.6% and 0.6 kg/t of the non-ionic surfactant and 1.8 kg/t of AlCl_3 lowers the cake moisture from 31.0% to 28.2%. It appears that in this case all the three surfactants, along with Al^{+3} ions, were effective in lowering the moisture while maintaining high solids recovery.

High-Sulfur Coal:

Addition of either cationic or anionic flocculant to the high-sulfur coal slurry had a minimal effect in filter cake moisture reduction and solids recovery.

Addition of cationic, anionic and non-ionic surfactants on cake moisture and solids recovery is shown in Figure 52. Addition of 1.5 kg/t of the cationic surfactant lowers cake moisture from 29.3% to 27.8%. Both anionic and non-ionic surfactants are marginally effective in lowering the filter cake moisture at 0.3 kg/t dosage. Solids recovery shows marginal improvement when non-ionic surfactant is used.

Figure 53 depicts the effect of addition of di-valent Cu^{+2} ions at pH 6.5 and tri-valent Al^{+3} ions at pH 9 on cake moisture and solids capture for the high-sulfur clean-coal slurry. At a dosage of 0.2 kg/t of CuCl_2 , the filter cake moisture decreases from 30.5% to 27.6%. The addition of 4.4 kg/t of AlCl_3 ions lowers filter cake moisture from 30.5% to 28.5%. Solids capture improves from 86-90% as the dosage of AlCl_3 increases.

Figure 51. Synergistic effect of surfactant and varying Al^{3+} ion dosage on screen-bowl centrifuge dewatering of low-sulfur coal slurry (speed 5000 rpm; filtration time 30 sec).

Figure 52. Effect of surfactant dosage on centrifugal dewatering of high-sulfur coal slurry. (speed 5000 rpm; filtration time 30 sec).

Figure 53. Effect of metal ion dosage on centrifugal dewatering of high-sulfur coal slurry. (speed 5000 rpm; filtration time 30 sec).

The synergistic effect of the concentration of metal ions and surfactants on dewatering of the high-sulfur coal slurry is shown in Figure 54. The addition of 1.5 kg/t cationic and 0.2 kg/t of CuCl_2 lowers filter cake moisture from 29.5% to 26.8% and 0.6 kg/t of anionic and 0.55 kg/t of CuCl_2 ions lowers filter cake moisture from 29.7% to 27.2%. Similarly, use of 0.9 kg/t of non-ionic and 0.8 kg/t of CuCl_2 lowers the filter cake moisture from 29.8% to 27.8%.

Figure 55 shows the synergetic effect of surfactant and Al^{+3} ions addition on centrifuge dewatering of the high-sulfur coal slurry. It is interesting to note that in this case, with 1.8 kg/t AlCl_3 and 1.5 kg/t cationic, the filter cake moisture reduces from 29.7% to 27.6%; with 0.6 kg/t anionic the moisture reduces from 29.7% to 28.0%; and with 0.9 kg/t non-ionic surfactant, the filter cake moisture reduces from 29.7% to 27.2%. Solids recovery for all the test conditions remains nearly unchanged.

Hyperbaric Filtration:

Low-Sulfur Clean Coal:

Figure 56 shows the effect of addition of surfactants on hyperbaric (pressure) dewatering of the low-sulfur clean coal slurry. For all of the tests the pressure was kept constant at 3 bar (43.5 psi) and cake formation angle at 165° . For the anionic and non-ionic surfactants filter cake moisture reduces from 24.8% to 23.0% at a high dosage of 1500 g/t of the surfactants; changes in the solids throughput and air consumption are insignificant. For the cationic surfactant, a 500 g/t dosage lowers the filter cake moisture from 25.0% to 23.5%. Solids throughput increase from $955 \text{ kg/m}^2/\text{hr}$ to $1113 \text{ kg/m}^2/\text{hr}$ as the surfactant dosage increases from 0 g/t to 1500 g/t; however, the air consumption for all the tests remains nearly constant.

Figure 54. Synergistic effect of surfactant and varying Cu^{2+} ion dosage on screen-bowl centrifuge dewatering of high-sulfur coal slurry (speed 5000 rpm; filtration time 30 sec)

Figure 55. Synergistic effect of surfactant and varying Al^{3+} ion dosage on screen-bowl centrifuge dewatering of high-sulfur coal slurry (speed 5000 rpm; filtration time 30 sec).

Figure 56. Effect of surfactant dosages on hyperbaric filter dewatering of low-sulfur coal slurry (pressure 3 bar; speed 1.5 rpm).

Effect of addition of anionic and cationic flocculants on the hyperbaric dewatering of the low-sulfur coal slurry is shown in Figure 57. As expected, for the addition of the anionic flocculant, the filter cake moisture increases from 24.8% to 29.1% and for the cationic flocculant moisture increases from 24.8% to 28.0% as the flocculant dosage increases from 0 to 33 g/t. Similarly, solids throughput increases from 842 kg/m²/hr to 1521 kg/m²/hr for the anionic and from 842 kg/m² to 1262 kg/m²/hr for the cationic. Using the anionic flocculant, air consumption increases from 170 m³/t to 488 m³/t; however, for the cationic flocculant, air consumption increases from 170 m³/t to 321 m³/t.

The effect of addition of metal ions (Cu⁺², Al⁺³) alone and with cationic surfactant (CPCL) are listed in Table XIII. Note, that the addition of metal ions alone was not effective in lowering the filter cake moisture. However, the synergistic effect of combining 230 g/t of CuCl₂ and the cationic surfactant CPCL (1500 g/t) lowers the filter cake moisture from 24.8% to 23.3% and combining 60 g/t of AlCl₃ and CPCL (1500 g/t) lowers the filter cake moisture from 24.8% to 22.9%. The solids throughput for both cases increases marginally, however, air consumption reduces from 170 m³/t to 118 m³/t for the Cu⁺²-CPCL and from 170 m³/t to 100 m³/t for the Al⁺³-CPCL system. Lowering of filter cake moisture in these tests appears to be due to the CPCL addition, as shown earlier, metal ions alone were not effective in lowering the cake moisture.

High-Sulfur Coal Slurry:

Figure 58 shows the effect of addition of surfactant on the hyperbaric dewatering of the high-sulfur coal slurry. Compared to the low-sulfur clean coal slurry, the high-sulfur coal slurry shows moisture reduction from 25.1% to 22.3% for the non-ionic and from 25.1% to 21.7%

TABLE XIII. Synergistic effect of addition of metal ions and surfactants on dewatering of low-sulfur clean coal slurry. (Pressure = 3 bar; filter speed = 1.5 rpm).

<u>Reagent</u>	<u>Amount</u> (g/t)	<u>Filter Cake</u> <u>Moisture</u> (%)	<u>Solids</u> <u>Throughput</u> (kg/m ² /hr)	<u>Air</u> <u>Consumption</u> (m ³ /t)
---	---	24.8	842	170
CuCl ₂	600	25.1	894	158
CuCl ₂ + CPCI	600 + 1500	23.3	899	118
AlCl ₃	500	24.9	889	145
AlCl ₃ + CPCI*	500 + 1500	22.9	855	100

*Cetyl pyridinium chloride

Figure 57. Effect of flocculant dosages on hyperbaric filter dewatering of low-sulfur coal slurry (pressure 3 bar ; speed 1.5rpm).

Figure 58. Effect of surfactant dosages on hyperbaric filter dewatering of high-sulfur coal slurry. (pressure 3 bar; speed 1.5rpm).

using the cationic surfactant at a dosage of 500 g/t; anionic surfactant was marginally effective in lowering the filter cake moisture. The solids throughput increases from 541 kg/m²/hr to 686 kg/m²/hr for the anionic; for the non-ionic increases from 541 kg/m²/hr to 678 kg/m²/hr; and for the cationic surfactant increases from 541 kg/m²/hr to 815 kg/m²/hr as the dosage of surfactant increases to 1500 g/t. The air consumption decreases from 325 Nm³/t to 179 Nm³/t using the anionic, from 325 Nm³/t to 166 Nm³/t using the non-ionic, and 325 Nm³/t to 228 Nm³/t using the cationic surfactant as the dosage increases to 1500 g/t.

The effect of anionic and cationic flocculant dosages on the hyperbaric dewatering of high-sulfur coal slurry is shown in Figure 59. As expected, the filter cake moisture increases from 25.1% to 28.9% at a dosage of 85 g/t of anionic flocculant and at the same dosage the filter cake moisture increases from 25.1% to 27.3% for the cationic flocculant. Both the solids throughput and air consumption increases with increase in dosages of both the flocculants.

The effect of addition of Cu⁺² and Al⁺³ metal ion alone and with CPCL are summarized in Table XIII. Note, that addition of 600 g/t of CuCl₂ or 500 g/t of AlCl₃ are effective in lowering the filter cake moisture from 25.1% to 24.2%; for both metal ions, the solids throughput increases and air consumption lowers. The synergistic effect of metal ions with 1500 g/t of CPCL lowers the filter cake moisture from 25.17% to 22.4% for both the metal ions. This combination also increased solids throughput (541 kg/m²/hr to 840 kg/m²/hr) and the air consumption remained close to the previous tests where metal ions alone were used.

Figure 59. Effect of flocculant dosages on hyperbaric filter dewatering of high-sulfur coal slurry (pressure 3 bar; speed 1.5 rpm).

TABLE XIV Synergistic effect of metal ions and surfactant on hyperbaric dewatering of high-sulfur clean coal slurry. (pressure = 3 bar, filter speed = 1.5 rpm).

<u>Reagent</u>	<u>Amount</u> (g/t)	<u>Filter Cake</u> <u>Moisture</u> (%)	<u>Solids</u> <u>Throughput</u> (kg/m ² /hr)	<u>Air</u> <u>Consumption</u> (Nm ³ /t)
---	---	25.17	541	325
CuCl ₂	600	24.2	711	224
CuCl ₂ + CPCL*	600 + 1500	22.7	831	221
AlCl ₃	500	24.6	703	228
AlCl ₃ + CPCL*	500 + 1500	22.4	841	220

*Cetyl pyridinium chloride.

Ceramic Plate Filtration:

The ceramic plate filtration tests were conducted using a portable vacuum pump which provided a constant 25 inches Hg of vacuum. Even though the manufacturer recommended using a pre-thickened slurry for the tests, however, for uniform comparison with other techniques, the 'as-received' PMCC low-sulfur clean coal slurry was used in these tests.

Of the three surfactants, i.e., cationic, anionic, and non-ionic surfactants, only the non-ionic (octyl phenoxy polyethoxy ethanol) and the cationic surfactant (cetyl pyridinium chloride CPCL) showed any significant effect in lowering of the filter cake moisture. Figure 60 shows the effect of addition of the two surfactants on the filter cake moisture and solids throughput. In this case, only the low-porosity ceramic filter shows lowering of filter cake moisture from 19.5%

Figure 60. Effect of surfactant dosages on ceramic plate dewatering of low-sulfur coal slurry.

to 16.8% at a surfactant dosage of 0.4 kg/t. The solids throughput for both the filters remains unaffected with increasing surfactant dosage. The figure also shows addition of the cationic surfactant (CPCL) lowers the filter cake moisture significantly from 19% to 15.8% for the low-porosity filter at a surfactant dosage of 1 kg/t.

Figure 61 shows the effect of flocculant dosage on dewatering of the low-sulfur coal slurry using the ceramic filter. Note, that addition of 10 g/t of the anionic flocculant is effective in lowering the filter cake moisture from 22% to 14% for the high-porosity and 18% to 13.5% for the low-porosity ceramic filter while solids throughput remained constant. Increasing anionic flocculant dosage did not provide further lowering of the cake moisture. Addition of cationic flocculant shows a slight decrease in filter cake moisture from 18% to 16% at a dosage of 15 g/t using the low-porosity filter. For the high-porosity filter, the filter cake moisture remains constant, however, the solids throughput increases from 100 kg/m²/hr to 360 kg/m²/hr when flocculant dosage increases to 25 g/t.

A note of caution on the test results mentioned above is in order. The low moisture filter cake obtained in these tests was better than that obtained in the vacuum or the pressure filters. Manufacturer of the ceramic filter however, advised against using any reagent in their filter, as it hinders the capillary function.

Figure 61. Effect of flocculant dosages on ceramic plate filter dewatering of low-sulfur coal slurry.

Pilot-Scale Dewatering Studies

The pilot-scale dewatering tests were conducted using hyperbaric, vacuum, and centrifuge technologies. The pilot-scale tests were conducted by processing clean coal at a rate of 0.2 to 2 ton/hr. For sampling, after the appropriate test conditions were maintained for at least 30 minutes, the entire dewatered product was collected for one minute time period in a tared container. The dewatered product was weighed, thoroughly mixed, and one kilogram representative sample was retained in a sealed plastic bag for moisture determination. Timed samples of the feed and effluent streams were also collected in a calibrated vessel to determine the flowrates. A one liter of the representative samples of the feed and effluent was retained in a sealed plastic jars for percent solids determination and in some cases particle size distributions. Table XV lists the baseline dewatering data for the low- and high-sulfur clean coal slurries using the three techniques. In these tests no reagents were utilized and the test conditions listed are the optimum selected from a series of tests conducted using various variables for each technique. Note that for the high-sulfur coal slurry, the hyperbaric (pressure) filter provided the lowest moisture product containing 23.6% moisture. Surprisingly, the filter cake moisture for the low-sulfur coal obtained using the three techniques is similar. This may be due to the fine particle size of the slurry. The solids capture for both pressure and vacuum filter are 99%, however, for the screen-bowl centrifuge it is only 64% for the low-sulfur and 42% for the high-sulfur coal slurries. Also, the solids throughput in the pressure filter is about eight times greater than that processed in the vacuum filter, and more than two times greater than that processed in the centrifuge.

TABLE XV. BASELINE PILOT SCALE DEWATERING DATA

Filter Type		Pressure		Vacuum		Screen-bowl Centrifuge	
Filter Area (ft ²) (m ²)		22 2.044		18.8 1.747		18-in. diam.	
Baseline Operation Conditions	R.P.M	1.5		1		1000	
	Pressure or Vacuum	43.5 psi (3 bar)		15" Hg		---	
	Cake Formation Angle	165°					
	Filter Submergence			30%		---	
Dewatering Data		High Sulfur	Low Sulfur	High Sulfur	Low Sulfur	High Sulfur	Low sulfur
	Cake Moisture (%)	23.6	27.1	27.8	27.2	33	27.7
	Throughput (lb/ft ² /hr) (kg/m ² /hr)	165 805.7	140 683.6	25 122.1	18 87.9	---	
	Throughput (tons/hr)	1.8	1.8	0.23	0.23	0.75	0.75
	Air Consumption (scfm/t) (Nm ³ /t)	460 13.03	530 15.01	---		---	
	Solids Capture (%)	99	99	99	99	42	64

High-Sulfur Clean-Coal Slurry:

Effects of cationic surfactant dosage on the filter cake moisture and solids throughput using the three techniques are shown in Figures 62 and 63, respectively. Note, that only the pressure filter shows a lowering of moisture from 22.4% to 19.4% at a cationic surfactant dosage of about 0.4 kg/t. In vacuum and centrifuge dewatering, the surfactant is not effective. Solids throughput (Figure 63) for all the three technologies shows only marginal change with surfactant dosage increase.

Figures 64 and 65 shows the effect of anionic surfactant dosage on the filter cake moisture and the solids throughput, respectively. In this case, only the vacuum filtration shows lowering of filter cake moisture from 29.6% to 23.0% at a surfactant dosage of 1.5 kg/t. Solids throughput in vacuum shows an increase from 41 kg/m²/hr to 86 kg/m²/hr with a surfactant dosage of 1.5 kg/t. For the pressure and centrifuge filter, there was no change in solids throughput with the surfactant addition.

It should be noted that the base-line data points in terms of moisture and solids throughput did not remain constant and in some cases changes significantly. This was attributed to the variation in feed solid contents of the column product, which was difficult to control in a plant. Thus, for each reagent, a base-line data point was obtained and the data obtained with reagent addition are compared with that point.

Figure 62. Effect of cationic surfactant dosage on pilot-scale dewatering of high-sulfur coal slurry.

Figure 63. Effect of cationic surfactant dosage on the solids throughput in pilot-scale dewatering of high-sulfur coal slurry.

Figure 64. Effect of anionic surfactant dosage on pilot-scale dewatering of high-sulfur coal slurry.

Figure 65. Effect of anionic surfactant dosage on the solids throughput in pilot-scale dewatering of high-sulfur coal slurry.

Effects of non-ionic surfactant dosage on filter cake moisture and solids throughput are shown in Figures 66 and 67, respectively. Addition of 1.5 kg/t of the surfactant lowers the filter cake moisture from 21.1% to 17.4%, using the pressure filter and from 29.6% to 25.0% using the vacuum filter. As the surfactant dosage increases, solids throughput decreases from 600 kg/m²/hr to 375 kg/m²/hr in the case of the pressure filter and from 700 kg/hr to 600 kg/hr for the centrifuge. The solids throughput decrease for the pressure filter could be due to a low solids content in the feed which happened several times during the pilot-scale testing. In case of the vacuum filter the solids throughput increases from 40 kg/m²/hr to 50 kg/m²/hr with increase in the surfactant dosage to 1.5 kg/t.

Effect of addition of anionic and cationic flocculants on filter cake moisture is shown in Figure 68. With increase in the anionic flocculant dosage from 0 to 20 g/t the filter cake moisture increases from 22.5% to 25.2% for the pressure and from 29.5% to 31.2% for the centrifuge techniques. In case of vacuum filtration, the moisture increases from 26.0 to 27.9% then decreases to 25.4% with increase in anionic flocculant dosage to 20 g/t. For the cationic flocculant only the pressure filter shows a lowering of moisture from 23.4% to 21.1% using a dosage of 20 g/t. For the vacuum and centrifuge, the filter cake moisture increases with an increase in flocculant dosage.

Figure 66. Effect of nonionic surfactant dosage on the pilot-scale dewatering of high-sulfur coal slurry.

Figure 67. Effect of nonionic surfactant dosage on the solids throughput in pilot-scale dewatering of high-sulfur coal slurry.

Figure 68. Effect of flocculant dosage on pilot-scale dewatering of high-sulfur coal slurry.

Figure 69 shows the solids throughput for the anionic and cationic flocculant dosage for all the three dewatering techniques, respectively. For both flocculants, the solids throughput increases with increase in flocculant dosage for all the three techniques. As expected, the amount of solids throughput in pressure filter was almost 4 to 6 times more than that processed in the vacuum Filter. In all the centrifuge tests, the amount of solids captured averaged 70 percent.

Effect of addition of Cu^{+2} ions dosage on dewatering of the high-sulfur clean-coal slurry is shown in Figure 70. It shows that using pressure filter, 250 g/t of CuCl_2 lowers the filter cake moisture from 23.0% to 21.2% and, in the case of vacuum filter about 500 g/t of CuCl_2 lowers the filter cake moisture from 27.1% to 25.6%. No moisture reduction was observed in the centrifuge. This could be due to the high 'g' produced in the centrifuge that may have destroyed the weak flocs produced by the metal ions.

The solids throughput with respect to CuCl_2 dosage is shown in Figure 71. Except for the vacuum filter, the solids throughput declines with the increase in CuCl_2 dosage. For the vacuum filter, the solids throughput increases from 100 $\text{kg/m}^2/\text{hr}$ to 200 $\text{kg/m}^2/\text{hr}$ at a dosage of about 500 g/t of CuCl_2 .

A combination of 25 g/t CuCl_2 and 0.25 kg/t of the anionic surfactant using the hyperbaric filter provided a filter cake with 22.0% moisture with solids throughput of 776 $\text{kg/m}^2/\text{hr}$. Similarly, 25 g/t of CuCl_2 and 0.5 kg/t of the cationic surfactant provided a filter cake with 21.0% moisture at a solids throughput of 863 $\text{kg/m}^2/\text{hr}$.

Low-Sulfur Coal Slurry:

Figure 72 shows the filter cake moisture as a function of surfactant dosage for the low-sulfur coal slurry. The addition of about 1.5 kg/t of the anionic surfactant lowers the filter cake

Figure 69. Effect of flocculant dosage on the solids throughput in pilot-scale dewatering of high-sulfur coal slurry.

Figure 70. Effect of CuCl_2 dosage on pilot-scale dewatering of high-sulfur coal slurry.

Figure 71. Effect of CuCl_2 dosage on the solids throughput in pilot-scale dewatering of high-sulfur coal slurry.

Figure 72. Effect of surfactant dosage on pilot-scale dewatering of low-sulfur coal slurry.

moisture from 24.0% to 16.8% using the vacuum filter; while no reduction in filter cake moisture is observed using either the pressure or centrifuge filter. Using 0.5 kg/t of the cationic surfactant lowers the filter cake moisture from 24.0% to 23.0% using the vacuum filter and from 36.8% to 34.3% using the centrifuge filter. Due to unavailability of the low-sulfur clean coal slurry no tests were conducted using the hyperbaric filter. Addition of 1.5 kg/t of the non-ionic surfactant lowers the filter cake moisture from 24.0% to 19.0% using the vacuum filter, whereas with centrifuge filter the cake moisture decreases from 36.8% to 34.0% using only 1.0 kg/t of the surfactant. With the pressure filter a slight increase in filter cake moisture occurs with increasing the surfactant dosage.

Figure 73 shows the solids throughput for the three techniques with the surfactant dosage. With the anionic surfactant addition the solids throughput shows a modest increase with increasing surfactant concentration using the vacuum and centrifuge filters. In the case of pressure filter a decline in solids throughput is observed, which could be due to the varying solids content of the feed. With the cationic surfactant addition the solids throughput increases with increasing surfactant dosage for the vacuum and centrifuge filters, which indicates formation of flocs. In case of non-ionic surfactant only the screen-bowl centrifuge shows increase in the solids throughput from 700 kg/hr to 900 kg/hr as the surfactant dosage increase to a dosage of 1.0 kg/t. The pressure and vacuum filters show a decline in solids throughput, possibly due to variation in feed solids.

The effects of flocculants addition on the filter cake moisture and solids throughput are shown in Figures 74 and 75, respectively. Using 5 g/t of the anionic flocculant the cake moisture decreases from 28.6% to 26.6% using the pressure filter and from 25.5% to 24.2% using the

Figure 73. Effect of surfactant dosage on the solids throughput in pilot-scale dewatering of low-sulfur coal slurry.

Figure 74. Effect of flocculant dosage on pilot-scale dewatering of low-sulfur coal slurry.

vacuum filter. The centrifuge shows a decrease in moisture from 31.3% to 30.3%, a 3% relative percentage reduction, using a high dosage of 15 g/t of the anionic flocculant. The addition of 15 g/t of the cationic flocculant decreases filter cake moisture from 25.2% to 21.3% using the vacuum filter. With the centrifuge, no moisture change is noticed when the cationic flocculant is used.

Using the anionic flocculant, the solids throughput (Figure 75) for all the three dewatering techniques shows an increase with increase in flocculant dosage, for example, for the pressure filter it increases from 664 kg/m²/hr to 776 kg/m²/hr, for the vacuum filter it increases from 71.7 kg/m²/hr to 116 kg/m²/hr, and for the centrifuge it increases from 500 kg/hr to 700 kg/hr, as the flocculant dosage increases from 0 to 15 g/t. With the cationic flocculant solids throughput increases from 89 kg/m²/hr to 103 kg/m²/hr and for the centrifuge it increases from 500 kg/hr to 700 kg/hr at a flocculant dosage of 15 g/t.

Figure 75. Effect of flocculant dosage on the solids throughput in pilot-scale dewatering of low-sulfur coal slurry.

The effect of metal ion addition on the filter cake moisture using the vacuum and centrifuge techniques is shown in Figure 76. It shows that in vacuum filtration 100 g/t of AlCl_3 and 500 g/t of CuCl_2 lowers the filter cake moisture from 28.0% to 26.5% and 26.0%, respectively. In case of the screen-bowl centrifuge about 250 g/t of either CuCl_2 or AlCl_3 lowers the filter cake moisture from 36.8% to 33.2%. The solids throughput in the vacuum increases from 122 $\text{kg/m}^2/\text{hr}$ to 170 $\text{kg/m}^2/\text{hr}$ as the metal salt dosage increases to 100 g/t. The solids capture in the centrifuge improved from 40.5% to 44.0% at a metal salt dosage of 500 g/t.

Figure 76. Effect of metal salt dosage on pilot-scale dewatering of low-sulfur coal slurry.

The synergistic effect of combining 500 g/t of CuCl_2 and various amount of surfactant on dewatering of the low-sulfur coal slurry is shown in Figure 77. It shows that using the vacuum filter, addition of 0.5 kg/t of either cationic (CPCL) or anionic (S2ES) surfactant along with 500 g/t of CuCl_2 lowers the filter cake moisture, from 28.0% to 25%. Addition of 1.0 kg/t of the anionic along with 0.5 kg/t of CuCl_2 lowers the filter cake moisture to 23%. The solids throughput at this dosage increases from 122 $\text{kg/m}^2/\text{hr}$ to 156 $\text{kg/m}^2/\text{hr}$.

In the centrifuge, the combination of the metal ions and surfactant did not show any reduction in filter cake moisture.

During the course of the pilot-scale testing using the screen-bowl centrifuge, it was decided to test the USDOE developed 'GRANUFLOW' process which utilizes 'Orimulsion' for dewatering of fine coal. Orimulsion is a low cost, high Btu, bitumen-in-water emulsion produced in Venezuela. The emulsion consists of 70% bitumen, 30% water, and trace amount of surfactants. According to Bitor America Corporation, supplier of the 'Orimulsion' in the U.S.A., the cost of Orimulsion is close to the cost of coal on an equal Btu basis.

Figure 78 shows the dewatered product moisture and solids captured with respect to the Orimulsion dosage. Note, that the product moisture lowers from 35.7% to 26.5%, a 9.2 absolute percent reduction, when the Orimulsion dosage is increased to 6.4 weight percent (80 kg/t). Also, note that the solids captured in the screen-bowl centrifuge increased from 65% to 95%.

The Orimulsion treatment dramatically reduces the solids content in both the screen and main effluents, and as a result, the coal recovery as shown in Table XV, increases from 64.7% to 94.1%, a 45% increase at an Orimulsion dosage of 6.4 percent. The amount of solids reduction in the main effluent alone accounted for about a 17.5% increase in recovery of the dewatered product.

Figure 77. Synergistic effect of Cu^{2+} and surfactant dosage on the pilot-scale vacuum drum filter dewatering of low-sulfur coal slurry (CuCl_2).

Figure 78. Centrifuge product moisture as a function of Orimulsion dosage.

Table XVI. Approximate mass balance of the centrifuge products. (1000 rpm and 226 g-force)

Test No.	Orimulsion wt%	Solids Balance, wt%			
		Feed	Product	Main Effluent	Screen Effluent
MF 1-1	0	100	64.7	22.5	12.8
MF 1-2	0.7	100	73.1	18.0	8.9
MF 1-3	3.2	100	82.2	15.3	2.5
MF 1-4	4.8	100	93.5	6.1	0.4
MF 1-5	6.4	100	94.1	5.0	0.9
MF 1-6	4.8	100	86.7	12.4	0.9
MF 1-7	3.2	100	83.5	14.1	2.4

Table XVII shows the Granuflow process test data in terms of the product ash contents, and effluent ash and solids content. It is interesting to note that the average ash content of the screen-bowl dewatered product is 4.4%, which is significantly lower than the feed ash content of 6.5%. Evidently, addition of 'Orimulsion' provided some coal cleaning resulting in ash reduction. The results also indicate that the bitumen in the Orimulsion selectively agglomerated coal particles but not ash-forming particles, resulting in an increase in the effluent solids ash content and product recovery.

Table XVII. GranuFlow process test data.

Test No.	Orimulsion wt%	Product Moisture wt%	Product Ash wt%	Main Effluent Solids wt%	Main Effluent Ash wt%	Screen Effluent Solids wt%	Screen Effluent Ash wt%	Dust Index	Dust Reduction Efficiency %
MF 1-1	0	35.7	4.4	3.4	14.0	44.7	9.3	82	10
MF 1-2	0.7	35.5	4.4	3.0	16.1	33.7	9.3	56	38
MF 1-3	3.2	32.8	4.4	2.8	17.5	9.6	11.3	14	85
MF 1-4	4.8	28.3	4.3	1.1	31.8	1.5	16.9	3	97
MF 1-5	6.4	26.5	4.4	NA*	NA*	3.1	11.8	2	98
MF 1-6	4.8	31.4	4.4	2.5	16.8	3.3	13.2	7	92
MF 1-7	3.2	32.4	4.5	3.3	16.9	8.5	11.1	9	90

*No sample available.

The handleability of the centrifuge product was greatly improved with the additions of the Orimulsion. Free flowing granules, as opposed to wet lumpy material, were clearly observed at an Orimulsion dosage of 3.2 weight percent and above (Figure 79). The handleability improvement of the product was also indicated by the formation of a dry product discharged from the centrifuge. During the tests, it was observed that the Orimulsion treated dewatered coal showed a smaller angle of repose, while the untreated coal showed a larger angle of repose. The angle of repose is the angle between the horizontal and the slope of a heap of material dropped from some elevation. The smaller the angle of repose the more flowable is the material.

To evaluate the performance of the Granuflow Process for dust control, FETC adopted a simple Ro-Tap dry screening process to experimentally measure the dust index (li) of the cakes with a constant amount of stress applied. A dust reduction efficiency (E) is therefore calculated based on the following equation:

$$E = \frac{lo - li \times 100}{lo}$$

where,

E = percent efficiency of dry cake dust reduction, %.

lo = dust index of coal, cumulative weight percent of feed coal finer than 150 mesh (106 m) by wet screening.

li = dust index of cake, cumulative weight percent of dry cake finer than 150 mesh (106 m) after Ro-Tapping for 5 minutes.

The dust index of the feed coal, lo, was 91% passing 150 mesh (106 m) obtained from a wet screen analysis. The average dust indices of the Orimulsion treated dry product, li, were 82, 56, 12, 5 and 2 with Orimulsion dosages of 0, 0.7, 3.2, 4.8 and 6.4%, respectively. Dust

(a)

(b)

Figure 79. Dewatred coal product (a) before and (b) after the addition of Orimulsion.

reduction efficiency as shown in Table XVII indicated that over 85% of the dust (material finer than 150 mesh) was reduced by agglomeration using 3.2% Orimulsion. The dust reduction efficiency reached 92% and 98% at 4.8% and 6.4% Orimulsion addition, respectively.

Economic Evaluation

Upon completion of the POC-scale testing of the vacuum filter, the pressure filter and the screen-bowl centrifuge, an economic comparison of the three dewatering techniques was made. In order to make such a comparison, it was necessary to select the conditions from a single test from the results obtained with each device that represented the best dewatering performance. The criteria used for selection of the best performance was based on obtaining the lowest product moisture with the highest solids throughput. The actual data recorded from the selected tests were scaled to a feed rate of 25 tons of dry solids per hour for comparative purposes.

For the vacuum filter testing, the filter operating conditions selected were 1 rpm with 15 g/t anionic flocculant as an additive which provided a throughput of 229 kg/m².hr. The resulting cake moisture has 25.8% moisture with a solids capture of 100% as shown in Figure 80. For the screen bowl centrifuge testing, a speed of 1000 rpm was selected with 6.0 weight % (80 kg/t) Orimulsion as an additive. These conditions provided a product with 26.0% moisture and a solids capture of 94% (Figure 81).

Two hyperbaric (pressure) filter test conditions were selected. The first (Figure 82) used 1.5 rpm filter speed and 0.5 kg/t cationic surfactant (CPCL) as an additive. The filter throughput with these conditions was 612 kg/m²/hr with an air consumption of 296 Nm³/t. The product contained 19.4% moisture with 100% solids capture. The second pressure filter test selected (Figure 83) again used 1.5 rpm filter speed, but the additives were 25 g/t CuCl₂ and 0.25 kg/t anionic surfactant (S2ES). The resulting throughput was 776 kg/m²/hr with an air consumption

of 285 Nm³/t. The product obtained with these conditions contained 22.0% moisture with 100% solids capture.

Figure 80. Mass balance and the operating conditions for a vacuum filter processing 25 tph of dry solid slurry.

Figure 81. Mass balance and the operating conditions for a screen-bowl centrifuge processing 25 tph of dry solid slurry.

Figure 82. Mass balance and the operating conditions for a hyperbaric filter processing 25 tph of dry solid slurry using 0.5 kg/t of the cationic surfactant.

For the ceramic disk filter, because of the unavailability of the pilot-scale unit, the laboratory baseline data were utilized to determine the size of a commercial filter to process 25 tph of coal. The baseline conditions used were: 40% solids slurry, cake formation time of 13 secs, and cake drying time of 21 secs. These conditions provided a filter cake with 33% moisture at a solids throughput of $79 \text{ kg/m}^2/\text{hr}$ at 100% solids capture. The ceramic filter manufacturer utilized these data to estimate the size and cost of a commercial filter to process 25 tph of coal.

Copies of the flowsheets (Figures 79 to 82) were sent to the appropriate manufacturers with a request to provide and estimate the size of equipment required to dewater 25 tons/hr of feed solids under the conditions specified. Manufacturers were also requested to provide capital and operating cost estimates. The information received from the various manufacturers was compiled into a spreadsheet for the purposes of economic comparison. The resulting comparison is shown in Table XVIII.

The feed, product and filtrate rates, along with additive dosages, operating conditions and throughput were derived from the pilot-plant experimental testing. Additive costs were obtained from various suppliers. Electrical costs were assumed to be $0.04 \text{ \$/kWhr}$. To compare the total cost of each device on a per ton basis, it was necessary to distribute the equipment costs over a 10-year period with 4160 hours/year of operation. The total cost is the sum of equipment cost, electrical cost, labor and materials, maintenance and additives.

Figure 83. Mass balance and the operating conditions for a hyperbaric filter processing 25 tph of dry solid slurry using 0.25 kg/t of the cationic surfactant and 25 g/t of CuCl₂.

Table XVII. Economic and Operating Data Comparison of Various Dewatering Techniques

		Vacuum	Centrifuge	Pressure	Pressure	Ceramic
Feed	tph solids	25	25	25	25	25
	% solids	15	15	15	15	15
	gpm slurry	638	638	638	638	638
Product	tph solids	25	23.5	25	25	25
	% solids	74.6	74	80.6	78	67
	gpm slurry	106	100	98	99	100
Filtrate	tph solids	0	1.5	0	0	0
	% solids	0	1.1	0	0	0
	gpm slurry	532	538	540	539	538
Additive	anionic flocculant (g/t)	20				20
	cationic surfactant (kg/t)			0.5		
	anionic surfactant (kg/t)				0.25	
	Cu (g/t)				25	
	Orimulsion (wt/%)		6			
Conditions	rpm	1	1000	1.5	1.5	1.0
	air consumption (Nm ³ /t)			296	285	
Throughput	kg/m ² /hr	229	270**	612	776	278
hp		225	400	100	125	10
Equipment size		1064 ft ²	44"x132"	36 m ²	36 m ²	90 m ²
Equipment cost	\$	275,000	255,000	1,900,000	1,900,000	1,524,000
Electric cost	\$/hr	3.18	5.66	1.41	1.77	1.40
Labor & materials	\$/hr	0.40	0.21	0.30	0.30	8.90
Maintenance	\$/hr	1.18	4.80	1.00	1.00	8.83
Additive cost	\$/lb	0.90	0.0225	0.75	0.775	0.90
Equipment cost*	\$/ton	0.2644	0.2452	1.8269	1.8269	1.4654
Electric cost	\$/ton	0.1272	0.2264	0.0564	0.0708	0.0560
Labor & materials	\$/ton	0.0160	0.0084	0.0120	0.0120	0.3560
Maintenance	\$/ton	0.0472	0.1920	0.0400	0.0400	0.3530
Additive cost	\$/ton	<u>0.0396</u>	<u>2.7000</u>	<u>1.6520</u>	<u>1.0460</u>	<u>0.0396</u>
Total cost	\$/ton	0.4944	3.3720	3.5873	2.9957	2.0950

Assumptions

*25 tph, 4160 hr/yr, 10 yr

Electrical cost 0.04 \$/kWhr

Orimulsion cost 45 \$/ton (delivered)=0.0225 \$/#

Cationic surfactant cost 1.50 \$/#

Anionic surfactant cost 0.90 \$/#

Anionic flocculant cost 0.90 \$/#

Cu Cost 10 \$/#

**kg/hr

The lowest total operating cost of \$0.49/t was obtained with the vacuum filter. Although the vacuum filter equipment cost was not the lowest, the very low additive cost (\$0.04/ton) was the primary reason for the low operating cost. While the total cost with the vacuum filter was the lowest in this evaluation, the vacuum filter also provided the lowest throughput and a product moisture of 26.4%.

The total costs obtained using the pressure filter were \$3.00 to \$3.58/ton, with the lower cost resulting from using cationic surfactant alone and the higher cost from using anionic surfactant with the copper salt. The primary cost component of the pressure filter (61 to 68% of the total cost) is the equipment cost. Despite the relatively high equipment cost, the pressure filter undoubtedly provided the lowest product moisture (19.4%) while simultaneously providing the highest solids throughput (612 to 776 kg/m²/hr).

The total processing cost using the centrifuge was estimated to be \$3.37/ton. While the electrical and maintenance costs are slightly higher than those for vacuum and pressure filters, the primary reason for the high processing cost with the centrifuge is the additive cost (\$2.70/t) which comprised 80% of the total cost.

The processing cost of the ceramic filter was estimated to be \$2.09/ton; however, it provided a filter cake with 33% moisture, which will not be acceptable to the coal industry. In addition, use of ceramic filter requires thickening of the froth slurry from 15% to 40% solids. In order to properly estimate the processing costs, the capital cost of a 50 ft diameter thickener (\$91,000) and 150 gpm underflow pump (\$3000) were included, as well as the additional horsepower requirement to operate the thickener rakes and motor.

CONCLUSIONS

Based on the above-mentioned results and discussions, the following conclusions are made:

- The low- and high-sulfur clean-coal slurry produced at the Powell Mountain Coal Company has an average particle size (D_{50}) of 29.5 μm and 37.3 μm , respectively. The solids content of the low- and high-sulfur slurries averaged 14% and 19%, respectively.
- The baseline dewatering studies conducted with the low-sulfur clean-coal slurry showed that
 - with the vacuum filter a 30 seconds cake formation time and 30 seconds cake drying time provided a 2.0 mm thick cake (6.98 $\text{kg/m}^2/\text{hr}$ solids throughput) with 23% filter cake moisture.
 - with the pressure filter an 85° cake formation angle and 3 bar pressure provided a filter cake with 23.2% moisture and 457 $\text{kg/m}^2/\text{hr}$ solids throughput. The air consumption was about 670 Nm^3/t .
 - using the high-porosity ceramic filter, a 30 seconds cake formation (CFT) and 30 seconds cake drying time (CDT) provided a filter cake with 21.8% moisture at a solids throughput of 0.53 $\text{kg/m}^2/\text{hr}$. A low-porosity ceramic filter using the 30 seconds cake formation time and 30 seconds cake drying time provided filter cake with 18% moisture at a solids throughput of 0.45 $\text{kg/m}^2/\text{hr}$.

The solids throughput obtained with the ceramic disk filter was very low, because the filter is not suitable for a low solids content slurry. The minimum solids requirement for the filter is about 40% compared to 15% available with

the fine-coal slurry. Using a 40% solids slurry a 15 seconds CFT and 20 seconds CDT a filter cake with 33% moisture was obtained.

- The baseline dewatering studies conducted with the high-sulfur clean-coal slurry showed that
 - using the vacuum filter a 30 seconds cake formation time and 30 seconds cake drying time provided a filter cake of 5.0 mm thickness ($17.5 \text{ kg/m}^2/\text{hr}$ solids throughput) with 26% moisture.
 - using the laboratory centrifuge at 5000 rpm (2000 g) filter cake with about 26% moisture at about 94% solids recovery was obtained.
 - using 3 bar pressure and 125° cake formation angle the hyperbaric filter provided a filter cake with 23% moisture at a solids throughput of $600 \text{ kg/m}^2/\text{hr}$ and air consumption of $400 \text{ Nm}^3/\text{t}$.
- Vacuum filter dewatering of the coal slurries using the various reagents showed that
 - for the low-sulfur clean-coal slurry, the use of anionic or cationic flocculant increased filter cake moisture as well as cake thickness (solids throughput). However, the moisture content of a given thickness filter cake obtained without reagent was much higher than that obtained using the flocculant. For example, a cake of 4.5 mm thickness ($15.7 \text{ kg/m}^2/\text{hr}$ solids throughput) obtained without the flocculant addition had a moisture content of 32%, whereas the same thickness filter cake obtained with flocculants had about 24% moisture; an absolute eight percent reduction in moisture.
 - for the high-sulfur clean-coal slurry, the addition of about 5 g/t of the cationic flocculant lowered moisture content from 26% to 22%; however, addition of

anionic flocculant increased the filter cake moisture. With both cationic and anionic flocculant the cake thickness increased with increase in flocculant dosage.

- for the low-sulfur coal slurry, addition of cationic and non-ionic surfactants was effective in lowering the filter cake moisture. 1.2 kg/t of cationic surfactant lowered filter cake moisture from 24.5 to 23.2% and 0.9 kg/t of non-ionic surfactant lowered filter cake moisture from 26.5% to 21.0%. Significant increase of cake thickness from 3.5 mm to 9 mm (12 kg/m²/hr to 31.4 kg/m²/hr solids throughput) was also achieved using the cationic surfactant.
- for the high-sulfur coal slurry, addition of 1.5 kg/t of the cationic or non-ionic surfactant lowered the filter cake moisture from 28% to 24%. Again, increase in the cake thickness was significant with the increase of the cationic surfactant.
- for the low-sulfur coal slurry, addition of the Cu⁺² or Al⁺³ metal ions alone was not effective in lowering the filter cake moisture, however, the cake thickness (solids throughput) increased with increase in metal ions dosage indicating flocs formation.

Synergistic effect of adding metal ions and surfactant combination also did not show any lowering of moisture. The increase in cake thickness was marginal for all the surfactants.

- for the high-sulfur coal slurry, addition of 0.5 kg/t of CuCl₂ lowered the filter cake moisture from 26.0% to 24.8%.

- Synergistic effect of addition of metal ions alone with surfactants did not show any improvement in filter cake moisture reduction.
- Hyperbaric filtration of low-sulfur clean-coal slurry with the addition of reagents showed that
 - a high dosage (1.5 kg/t) of anionic or non-ionic surfactant reduced filter cake moisture from 24.8% to 23.0%. 0.5 kg/t of cationic surfactant was effective in lowering moisture from 25.0% to 23.5%. The solids throughput increased from 955 kg/m²/hr to 1113 kg/m²/hr as the anionic surfactant dosage increased to 1.5 kg/t; however, air consumption remained constant.
 - addition of about 30 g/t of the flocculants increased cake moisture by absolute 5% and also increased solids throughput by 80% (from 842 kg/m²/hr to 1520 kg/m²/hr).
 - addition of metal ion alone was not effective in lowering the filter cake moisture; however, a combination of metal ion and surfactant showed lowering of filter cake moisture by about 1.5% to 1.9% absolute, which could be due to the presence of the surfactants.
- Hyperbaric filtration of high-sulfur clean coal slurry with the addition of reagents showed that
 - using 0.5 kg/t of non-ionic lowered the filter cake moisture from 25.1% to 22.3% and 0.5 kg/t of cationic surfactant lowered moisture from 25.1% to 21.7%. All three surfactants showed increase in solids throughput and lowering of air consumption.

- addition of flocculants increased cake moisture by about three absolute percent at a dosage of about 80 g/t and also increase solids throughput and air consumption.
- addition of 600 g/t of CuCl_2 or 60 g/t of AlCl_3 lowered filter cake moisture from 25.1% to 24.4%. A combination of 600 g/t of CuCl_2 and 1.5 kg/t of CPCL lowered filter cake moisture to 22.7% and 500 g/t of AlCl_3 and 1.5 kg/t CPCL lowered it to 22.4%. The solids throughput increased from 541 $\text{kg/m}^2/\text{r}$ to 700 $\text{kg/m}^2/\text{hr}$ using the metal ions alone; and with metal ion-surfactant combination it increased to 835 $\text{kg/m}^2/\text{hr}$. The air consumption decreased from 325 Nm^3/t to 220 Nm^3/t .
- Ceramic plate filtration of the 'as-received' low-sulfur clean-coal slurry with the addition of reagents provided a very low moisture filter cake, however, the solids throughput (cake thickness) was much lower than that obtained with the vacuum or pressure filters. For example, 0.4 kg/t of nonionic or 1 kg/t of cationic surfactant lowered the filter cake moisture from 22.0% to about 15.8%, the solids throughput ranged from 100 $\text{kg/m}^2/\text{hr}$ to 360 $\text{kg/m}^2/\text{hr}$, providing a cake thickness of <2 mm.

The Outokumpu Mintec U.S.A. Inc., manufacturer of the ceramic filter, strongly recommended not to use any reagents with their filter as it hinders the capillaries operation.

- The baseline pilot-scale dewatering of the clean-coal slurry showed that
 - using 3 bar pressure, 165° cake formation angle and 1.5 rpm, the hyperbaric filter provided filter cakes with 23.6% moisture for the high-sulfur and 27.1%

moisture for the low-sulfur clean-coal slurries. Solids capture in the process was 99%.

- using 1 rpm and 30% submergence of the vacuum drum filter provided filter cakes with 27.8% moisture for the high-sulfur and 27.2% moisture for the low-sulfur slurries. Solids capture in vacuum filtration was 99%.
- using a screen-bowl centrifuge rotating at 1000 rpm (230 g) speed, filter cakes with 33% moisture obtained for the high-sulfur and 27.7% moisture for the low-sulfur clean coal slurries. Solids capture for the high-sulfur and low-sulfur coal slurries were 42% and 64%, respectively.
- Pilot-scale dewatering of high-sulfur clean coal slurry with the addition of reagent indicated that
 - with the hyperbaric filter, 0.5 kg/t of the cationic surfactant lowered the filter cake moisture from 22.4 to 19.4% and 1.5 kg/t of the non-ionic surfactant lowered the filter cake moisture from 22.6% to 17.4%. With the vacuum filter 1.5 kg/t of either anionic or non-ionic surfactant lowered the filter cake moisture from 29.6 to 25.0%.
 - addition of flocculant in general increased filter cake moisture, as well as solids throughput for all the three dewatering techniques. The amount of solids throughput in pressure filter was almost 4 to 6 times that processed in the vacuum. The solids capture in the centrifuge tests averaged 60 percent.
 - addition of 250 g/t of CuCl_2 lowered the filter cake moisture from 23.0% to 21.2% using the hyperbaric filter and from 27.1% to 25.6% using 500 g/t of

CuCl₂ utilizing the vacuum filter. A combination of 25 g/t of CuCl₂ and 0.25 kg/t of the cationic surfactant provided a filter cake with 22.0% moisture.

- Pilot-scale dewatering data on low-sulfur clean-coal using various reagents showed that
 - using the vacuum filter, addition of 1 kg/t of either anionic or cationic or 1.5 kg/t of non-ionic surfactant lowered the filter cake moisture from 24% to 16.8%, 24.0% and 21.1%, respectively. 1.0 kg/t of either cationic or non-ionic surfactant also lowered the filter cake moisture from 36.8% to 34.0% using the screen-bowl centrifuge. Addition of surfactant provided a modest increase in solids throughput.
 - addition of 5 g/t of the anionic flocculant decreased the filter cake moisture from 28.6% to 26.6% using the pressure filter; from 26.8% to 24.2% using the vacuum; and from 31.3% to 30.3% using the centrifuge filter.

Solids throughput for the pressure filter increased from 664 kg/m²/hr to 776 kg/m²/hr; for the vacuum it increased from 71.7 kg/m²/hr to 116 kg/m²/hr; and for the centrifuge it increased from 500 kg/hr to 700 kg/hr; with increase of flocculant dosage to 15 g/t.

- Addition of 6 weight percent (80 kg/t) of the Orimulsion lowered the filter cake moisture from 36% to 26% using the screen-bowl centrifuge. The solids capture increased from 64% to 95%; ash in the final dewatered product reduced from 6% to 4%; and the handleability and dust reduction efficiency index improved to 98%.
- The operating cost (\$/t of clean coal) of the four dewatering techniques to process 25 tph of dry coal in 15 weight percent solids slurry was as follows:

<u>Technique</u>	<u>Reagent</u>	<u>Filter Cake Moisture</u>	<u>Cost \$/t</u>
Hyperbaric	0.5 kg/t cationic surfactant	19.4	3.58
Hyperbaric	0.25 kg/t anionic surfactant 25 g/t of CuCl ₂	22.0	2.99
Vacuum	20 g/t flocculant	25.4	0.48
Centrifuge	80 kg/t Orimulsion	26.0	3.37
Ceramic filter	---	33.0	2.09*

*(40% solids slurry needed).

Hyperbaric filter provided the lowest moisture product, at a processing cost of about \$2.70 to \$3.00 per ton of clean coal.

RECOMMENDATIONS

Based on the above conclusions, the following recommendations are offered:

- To obtain a more definite technical and economical evaluation, a large-scale continuous fine coal dewatering test should be conducted using the hyperbaric filter or the screen-bowl centrifuge with the Orimulsion.
- Efforts should be directed toward understanding the basic chemistry of Orimulsion and to develop an economical substitute using domestic bitumen.

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