DOE/PC/92550--9

EVALUATION OF HYPERBARIC FILTRATION

FOR FINE COAL DEWATERING

DOE Grant No. ~~DE-PS22-92PC92520~~

DE-FG22-92PC 92550

Ninth Quarterly Technical Progress Report
(Oct. 1, 1994 - December 31, 1994)

by

B.K. Parekh (PI)
J.W. Leonard (Co-PI)
University of Kentucky

R. Hogg (Co-PI)
The Pennsylvania State University

A. Fonseca (Co-PI)
Consol Inc.

39

DISCLAIMER

OBJECTIVES AND SCOPE OF WORK

The main objectives of the project are to investigate the fundamental aspects of particle-liquid interaction in fine coal dewatering, to conduct laboratory and pilot plant studies on the applicability of hyperbaric filter systems and to develop process conditions for dewatering of fine clean coal to less than 20 percent moisture.

The program consist of three phases, namely

Phase I - Model Development

Phase II - Laboratory Studies

Phase III - Field Testing

The Pennsylvania State University is leading efforts in Phase I, the University of Kentucky in Phase II, and Consol Inc. in Phase III of the program. All three organizations are involved in all the three phases of the program. The Pennsylvania State University is developing a theoretical model for hyperbaric filtration systems, whereas the University of Kentucky is conducting experimental studies to investigate fundamental aspects of particle-liquid interaction and application of high pressure filter in fine coal dewatering. The optimum filtration conditions identified in phase I and II will be tested in a Consol Inc. coal preparation plant using an Andritz Ruthner portable hyperbaric filtration unit.

INTRODUCTION

Most of the coal presently used by the utility industry is cleaned at preparation plants employing wet 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 inch) is easily dewatered to a 3-4 percent moisture level using conventional vibrating screens and centrifuges. However, the main problem of excess product moisture occurs in fine (minus 28 mesh) coal and refuse. Even though fines may constitute only about 20 percent of a contemporary cleaning plant feed, they account for two-thirds of the product surface moisture. This high surface moisture offsets many of the benefits of coal cleaning, and can easily undercut the ongoing programs on recovery of fine clean from refuse as well as producing an ultra-fine super clean coal fuel.

Currently, most of the coal preparation plants utilize vacuum disk or drum filter technology for dewatering of the fine coal, providing dewatered product containing about 25 percent moisture. The coal industry would prefer to have a product moisture in the range of 10 to 15 percent. Although the desired product quality can be obtained using thermal dryer, there are problems associated with this technology such as high capital costs and a source of air pollution.

In the present research project, an alternative to thermal drying, hyperbaric filtration which has shown potential in lowering moisture content in fine coal to less than 20 percent level, is being investigated in detail. The project will develop fundamental information on particle-liquid interaction during hyperbaric filtration and apply the knowledge in developing optimum conditions for the pilot plant testing of the hyperbaric filter system.

Phase 1 - Model Development

The extent of fine-coal dewatering which can be achieved by hyperbaric filtration is influenced by each of the stages in the filter cycle. The filtration rate and overall capacity are determined by the cake-formation stage which also controls cake thickness and cake structure. These two factors, in turn, play a major role in establishing the dewatering characteristics. Cake formation in constant pressure filtration has been investigated in detail and well-established models are available in the literature. For materials such as coal which generally form incompressible cakes, the models lead to the following simple expression for the filtration rate:

$$\frac{t}{V} = \frac{\mu c \alpha}{2A^2 \Delta p} + \frac{\mu \alpha_m}{A \Delta p} \tag{1}$$

where V is the volume of filtrate passing in time t through area A with applied pressure Δp. μ is the liquid viscosity, α is defined as the specific cake resistance and α_m is the filter medium resistance.

Examples of the application of Equation 1 to the filtration of fine (-100 mesh) coal from the Pittsburgh seam are given in Figures 9-1 and 9-2. Generally consistent agreement can be seen; both α and α_m increase slightly with increasing applied pressure.

The specific cake resistance α is the model parameter which accounts for the characteristics of the solid. Laboratory measurements of α can provide the basic information needed for filter design. However, general relationships between cake resistance and material characteristics such as particle shape and particle size

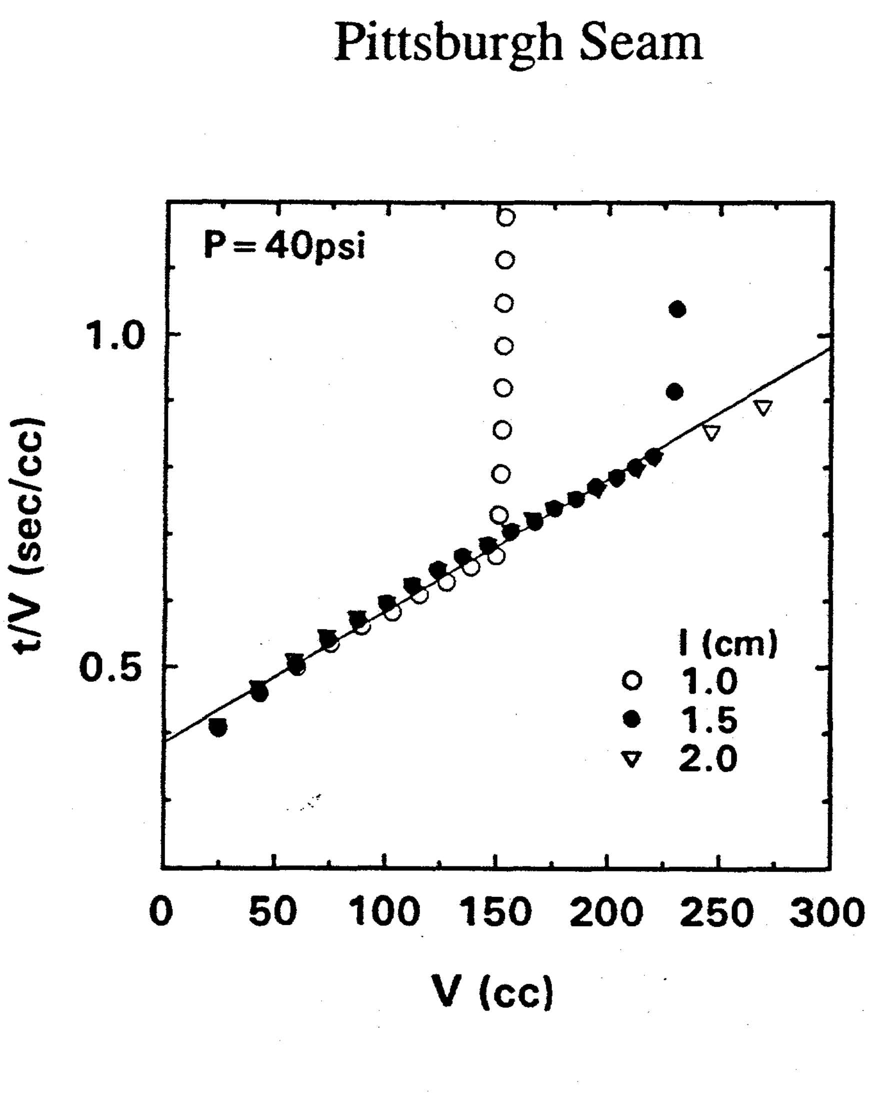

Figure 9-1. Cake formation kinetics for Pittsburgh seam coal at 40 psi.

Pittsburgh Seam

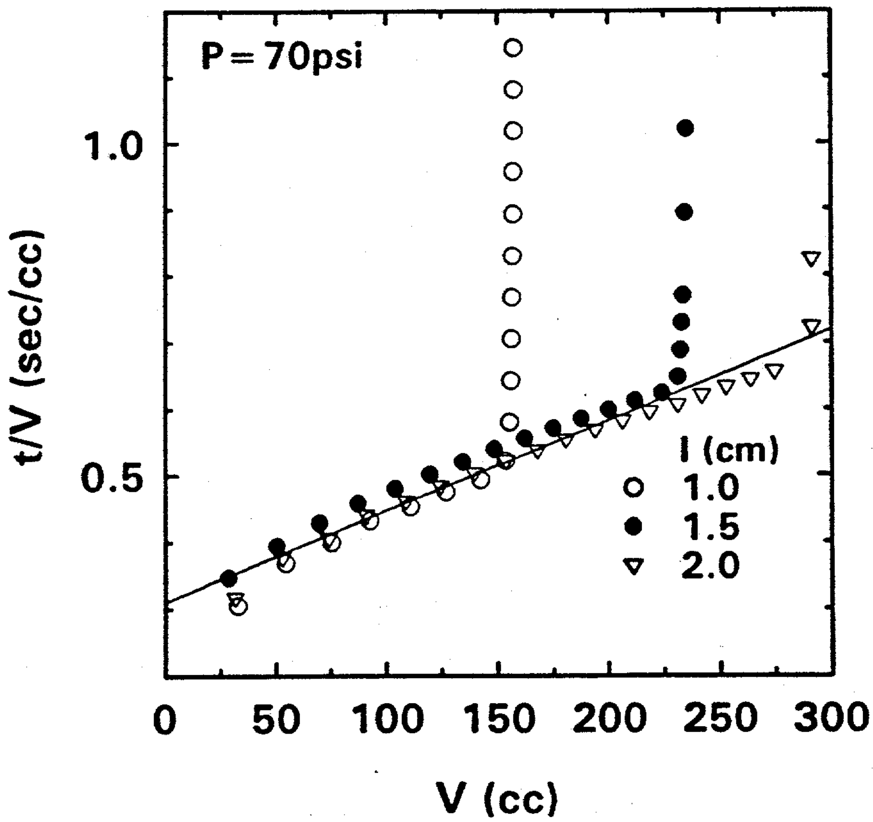

Figure 9-2. Cake formation kinetics for Pittsburgh seam coal for 70 psi.

distribution have yet to be established. Our investigations of packing behavior, discussed in previous reports, are aimed, in part, at the development of such relationships.

Application of the Carman-Kozeny expression for flow through porous media does provide some insight into the nature of the cake resistance. Thus,

$$\alpha \approx \frac{k(1-\epsilon)S_V^2}{\epsilon^3 \rho_s} \tag{2}$$

where k is a tortuosity factor (usually approximated as $k \approx 25/6$), ϵ is the cake porosity and S_V is the volume specific surface area of the particles. It should be noted, however, that Equation 2 includes some important simplifying assumptions. In particular, it is assumed that the pores in the cake can be represented using a single mean hydraulic radius, defined in the usual way by

$$r_h = 2\left(\frac{\text{Cross-sectional Area}}{\text{Wetted perimeter}}\right) \tag{3}$$

The area to perimeter ratio is equivalent to the pore volume to cross-sectional area ratio which leads to

$$r_h \approx \frac{2}{S_V}\left(\frac{\epsilon}{1-\epsilon}\right) \tag{4}$$

The use of a single effective mean radius is probably reasonable for beds of relatively uniform particles, but may be questionable for systems with the broad size distributions typically found in fine coal filtration.

An alternative approach is to obtain the effective mean pore radius by integrating the flow over the pore size distribution f(r). This leads to

$$r_{eff} \propto \left[\int_0^\infty r^2 f(r)\,dr\right]^{1/2} \tag{5}$$

The corresponding expression, based on Equation 4, is

$$r_h \propto \left[\int_0^\infty \frac{f(r)\,dr}{r}\right]^{-1} \tag{6}$$

For broad pore size distributions, these two values can differ widely. An example of the disparity between the two definitions of mean pore radius is given, for log normal pore size distributions, in Table 1.

Table 1. Ratio of effective mean pore radii for log normal pore size distributions.

Log Normal Standard Deviation, σ	Radius Ratio, r_{eff}/r_h
1	1.0
2	8.7
3	228
4	5700

The use of Equation 3 to define the mean hydraulic radius of a single pore is also subject to question. This expression is appropriate for fully turbulent conditions where shear occurs primarily at the pore walls. Under the laminar flow conditions expected in cake formation, however, shear occurs mostly in the bulk fluid and flow is concentrated in the central part of a pore with an irregular profile. It might be assumed, for example, that the effective radius for flow through a smooth-walled pore of triangular cross-section would be defined by the inscribed circle. In this case, the effective radius would be one half the mean hydraulic radius defined by Equation 3. The difference would be substantially greater for rough-walled pores between irregular particles. Further analysis of these geometric factors in flow through filter cakes is in progress. The long-term objective is to relate cake resistance directly to particle size distribution.

Pore geometry and its effects on flow through the cake also play an important role in the dewatering stage of the filter cycle. The same factors determine the rate of liquid displacement and also air consumption by flow through the cake. In addition, the capillary pressures which oppose dewatering are themselves determined by the distribution of effective pore radii. Since capillary pressures are a direct consequence of surface forces, the mean hydraulic radius should be the appropriate defining characteristic in this case. In other words, the effective pore size distribution which defines the capillary pressures may be significantly different from that which determines flow.

The limit of mechanical dewatering occurs when all of the liquid has been expelled from pores for which the applied (air) pressure is greater than the capillary pressure. In the literature, this is often taken to mean those pores for which the total pressure drop exceeds the capillary pressure. Implicitly, this assumes that all pores consist of continuous tubes extending through the entire thickness of the cake. More realistically, however, the pore structure should be represented by individual segments with a distribution of radius and length. Liquid displacement requires that the pressure drop across the segment length be greater than the capillary pressure.

If the pressure gradient across the cake is linear, the pressure drop along a pore segment of length ℓ will be

$$\Delta p = \frac{\ell \Delta p_o}{\lambda} \tag{7}$$

where Δp_o is the total pressure drop and λ is the cake thickness. The lengths of the individual pore segments are not generally known. However, it seems reasonable to assume that, on the average, the segment length will be proportional to the radius of the segment, i.e.,

$$\ell = \kappa r \tag{8}$$

where κ is a shape factor.

The limiting pore radius r_ℓ for mechanical dewatering is then given by

$$\frac{\kappa r_\ell \Delta p_o}{\lambda} = \frac{2\gamma\cos\theta}{r_\ell} \tag{9}$$

or

$$r_\ell = \sqrt{\frac{2\lambda\gamma\cos\theta}{\kappa\Delta p_o}} \tag{10}$$

where γ is the liquid surface tension and θ is the solid-liquid-air contact angle.

The limiting cake saturation S_ℓ is given by the fraction of pores whose radius is less than r_ℓ, i.e.,

$$S_\ell = F(r_\ell) \tag{11}$$

where F(r) is the pore size distribution.

Since the pore size distribution F(r) is a monotomically increasing function of the pore radius r, Equations 10 and 11 indicate that the limiting saturation should be correlated with the pressure gradient $\Delta p/\lambda$. The fractional moisture content M is often more convenient than the saturation. Since

$$S \propto \frac{M}{1-M} \tag{12}$$

the latter should also be correlated with the pressure gradient.

Some examples of the limiting moisture - pressure gradient are given in Figure 9-3 for experiments on the Pittsburgh #8, Illinois #6 and Pocahontas coals performed at the University of Kentucky and described in previous reports. Generally, there is a clear correlation, especially for the finest (Pittsburgh Seam) coal. The relatively weak correlation for the coarse (Pocohontas) coal may reflect the low overall moisture content and a masking affect due to the presence of isolated water droplets on the surfaces of coarse particles.

Phase 2 - Laboratory Studies

Laboratory dewatering studies were conducted on three different froth flotation concentrate samples obtained from Consol Inc. preparation plant processing Illinois No. 6, Pittsburgh No. 8 and Pocahontas No. 3 coal seams. In this quarterly report, the results obtained on Illinois No. 6 clean coal froth samples are summarized, in addition, data obtained on Pocahontas No. 3 coal are also listed.

Illinois No. 6 Coal:

The Illinois No. 6 froth flotation concentrate had 6.4 percent ash; and about 45 weight percent of particles were of plus 100 mesh size. For this coal a filter cake with moisture of 20 percent was easily achieved using 60 psi pressure and 2 cm cake thickness. Detailed information on hyperbaric filtration studies of the Illinois No. 6 coal samples were described in earlier quarterly reports numbered two through four. Figure 9-4 summarizes all the dewatering results obtained for the Illinois No. 6 coal slurry sample using various enhancement methods. Note, that because of large

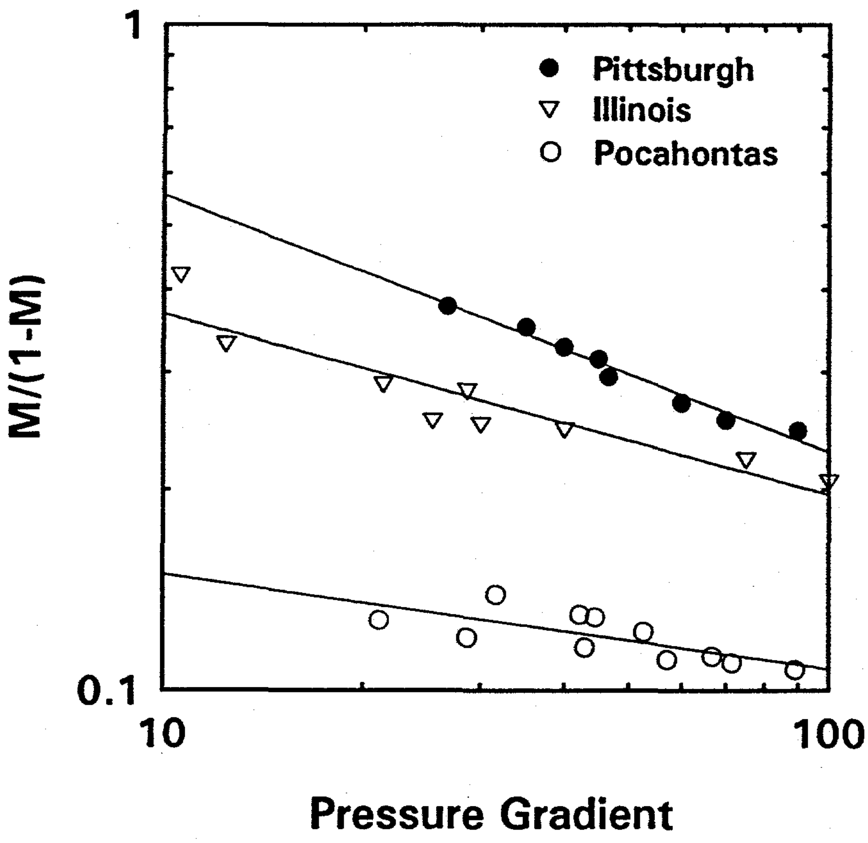

Figure 9-3. Correlation of final cake moisture content with pressure gradient.

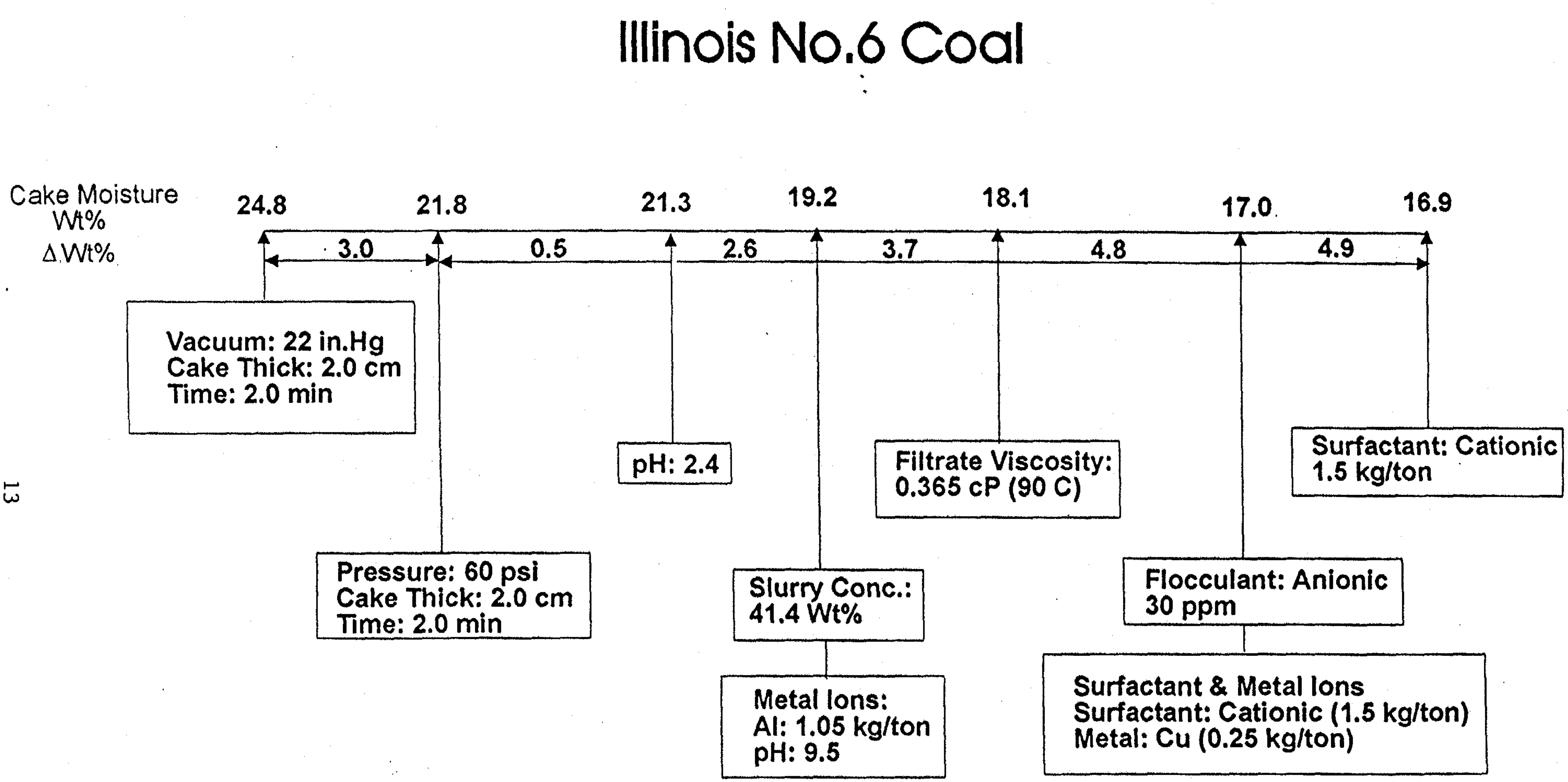

Figure 9-4. Summary of hyperbaric filter dewatering results for the Illinois No. 6 coal.

particle size hyperbaric filter provided 21.8 percent moisture filter cake which is only 3 percent additional moisture removal over that obtained using the vacuum filtration. Addition of 30 ppm of anionic flocculant or 1.5 Kg/t of the cationic surfactant provided 17 percent moisture filter cake.

Pittsburgh No. 8 Coal

Summary of the dewatering results of the Pittsburgh No. 8 sample were described in the last (eighth) quarterly report. For this coal slurry which had very fine (45% minus 500 mesh) particles, split size dewatering at 400 mesh provided 13 percent moisture filter cake compared to 25 percent moisture filter cake obtained in base line study.

Pocahontas No. 3 Coal

Effect of filtration time and pressure:

The effects of filtration time on the moisture content of 1.4 cm and 1.9 cm thick cake for the Pocahontas No. 3 coal at various applied pressures are shown in Figures 9-5 and 9-6, respectively. As expected, the optimum filtration time and filter cake moisture increased with increasing cake thickness. Figure 9-5 shows that for 1.4 cm thick cake, the optimum filtration time was about two minutes using 80 psi (5.5 bars) pressure. Similarly, Figure 9-6 shows that for 1.9 cm thick cake, the optimum filtration time was also two minutes using 80 psi (5.5 bars).

Figure 9-7 summarizes the effect of applied pressure on the filtration time for various thickness cake. The optimum filtration time and pressure for 1.4 cm cake

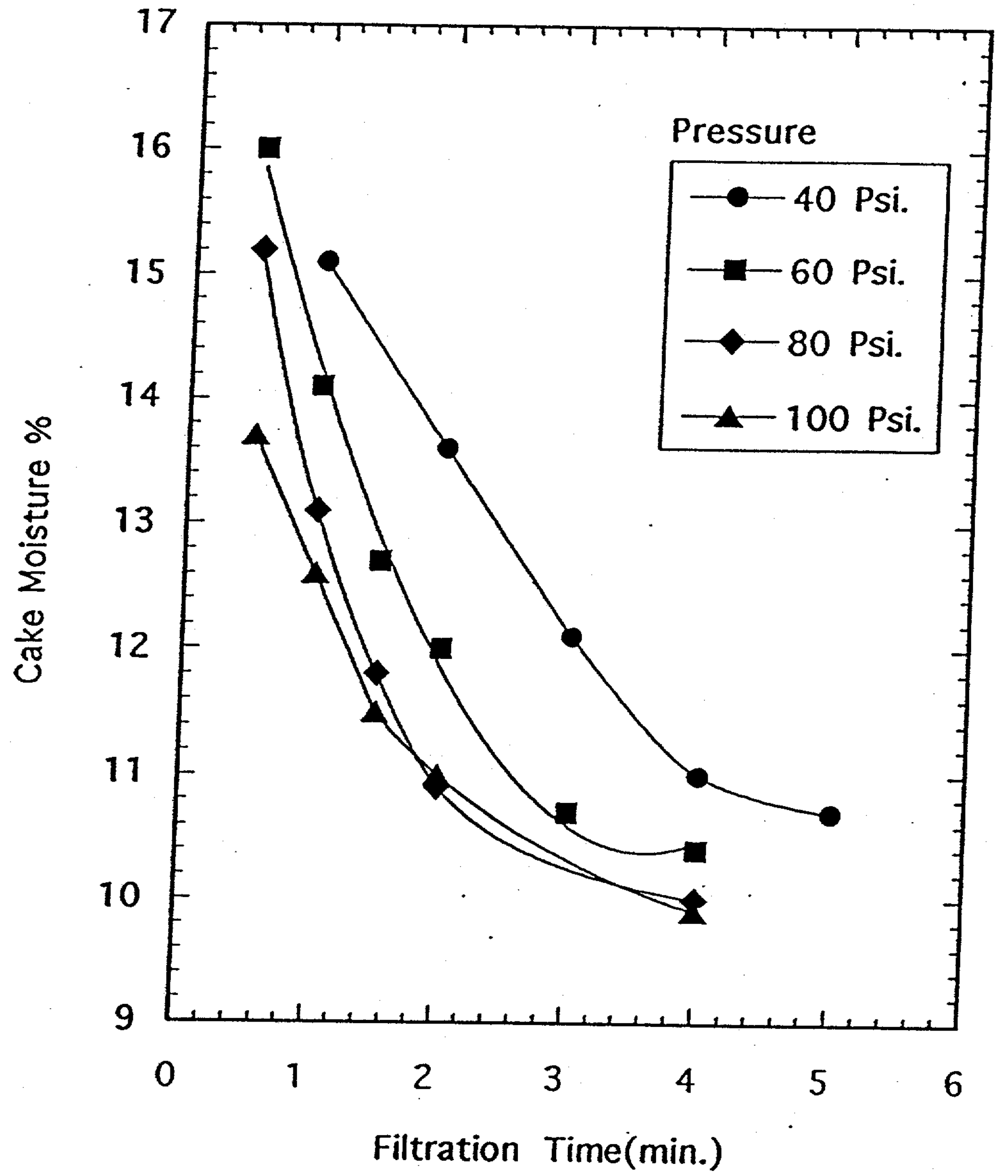

Figure 9-5. Effect of filtration time on cake moisture for the Pocahontas No. 3 coal. (cake thickness: 1.4 cm)

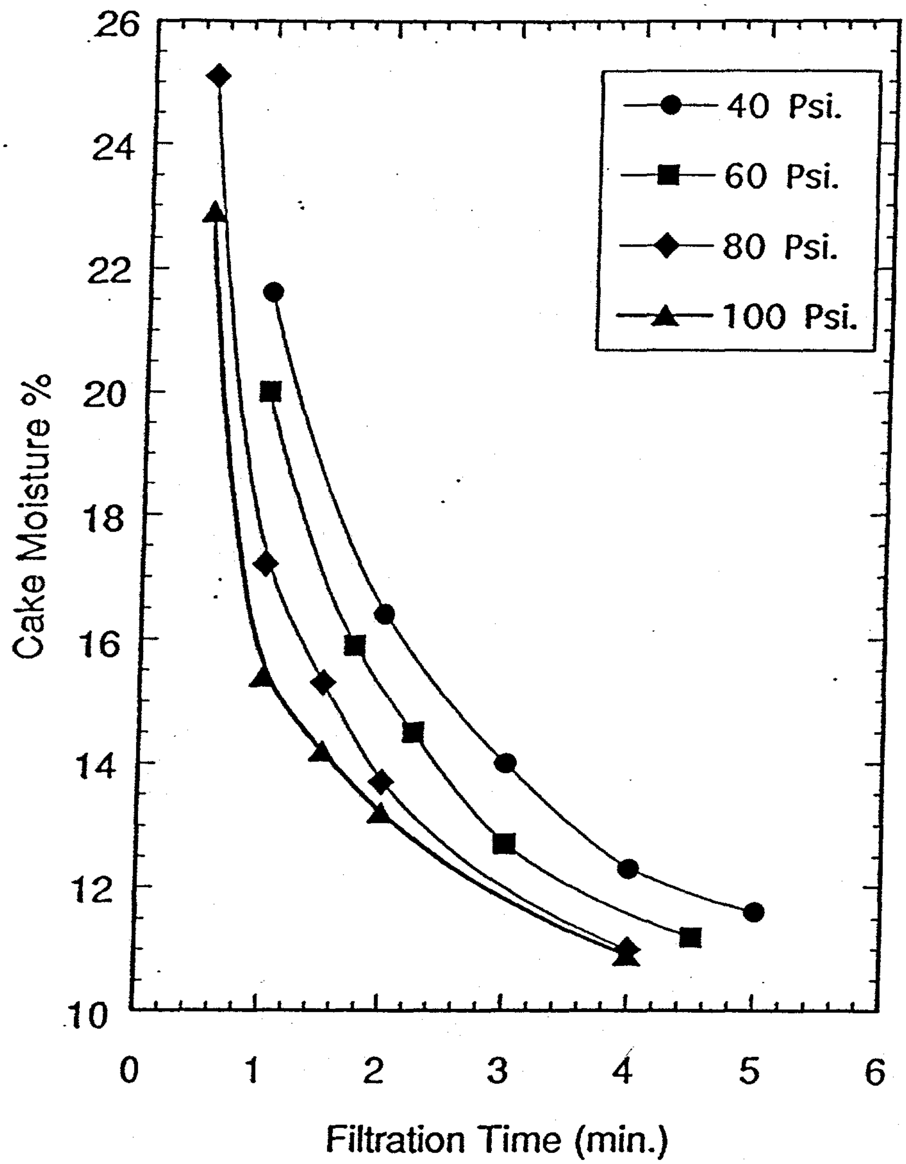

Figure 9-6. Effect of filtration time on cake moisture for the Pocahontas No. 3 Coal. (cake thickness: 1.9cm)

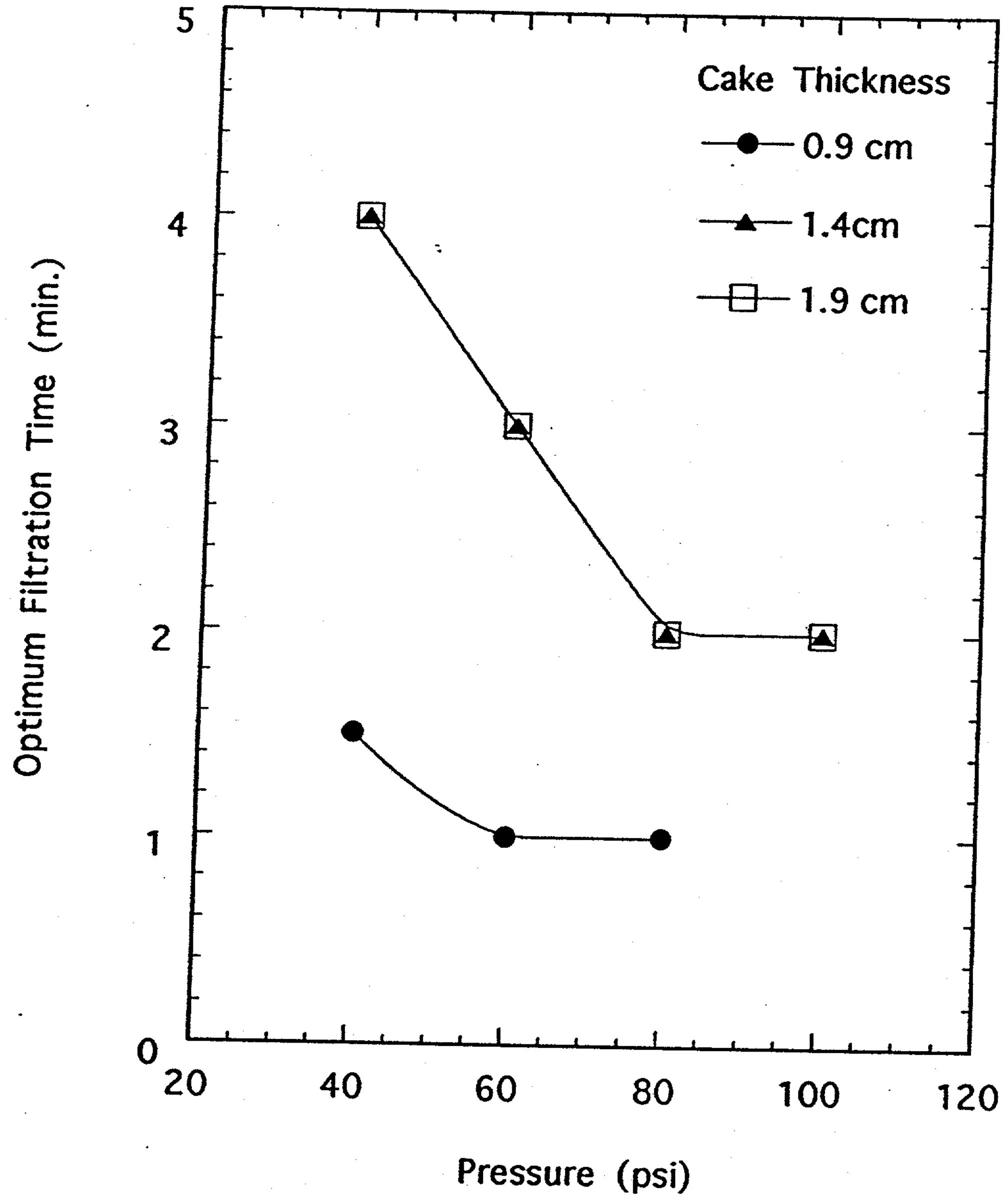

Figure 9-7. Effect of applied pressure on optimum filtration time for the Pocahontas No. 3 coal.

thickness were two minutes and 80 psi, which were the same as that for the 1.9 cm cake thickness. However, the moisture of 1.9 cm thick cake was 14 percent and that of 1.4 cm cake thickness was 11 percent. The subsequent dewatering studies on the Pocahontas No. 3 coal slurry were conducted using 1.4 cm cake thickness, 80 psi (5.5 bars) pressure and two minutes filtration time.

Cake Thickness:

Cake thickness is one of the most important factors in the filtration process because it affects both the final cake moisture and solids throughput. Figure 9-8 shows the effect of cake thickness on the cake moisture for the Pocahontas No. 3 coal at different applied pressures. The relationship between cake thickness and cake moisture was almost linear when applied pressures were between 40 psi (2.8 bars) and 60 psi (4.1 bars). However, this relationship changed at 80 psi (5.5 bars) applied pressure. It can also be observed that at 80 psi (5.5 bars) applied pressure, there was little increase in cake moisture for cake thickness between 0.9 cm and 1.4 cm; the cake moisture increases considerably for cake thicker than 1.4 cm. It is indicated that the optimum cake thickness for the Pocahontas No. 3 coal should be 1.4 cm using 80 psi (5.5 bars) pressure.

The above experimental results are summarized as follows, for the Pocahontas No. 3 coal, the optimum dewatering parameters are 1.4 cm cake thickness, 80 psi (5.5 bar) pressure and two minute filtration time, to obtain 11 percent filter cake moisture.

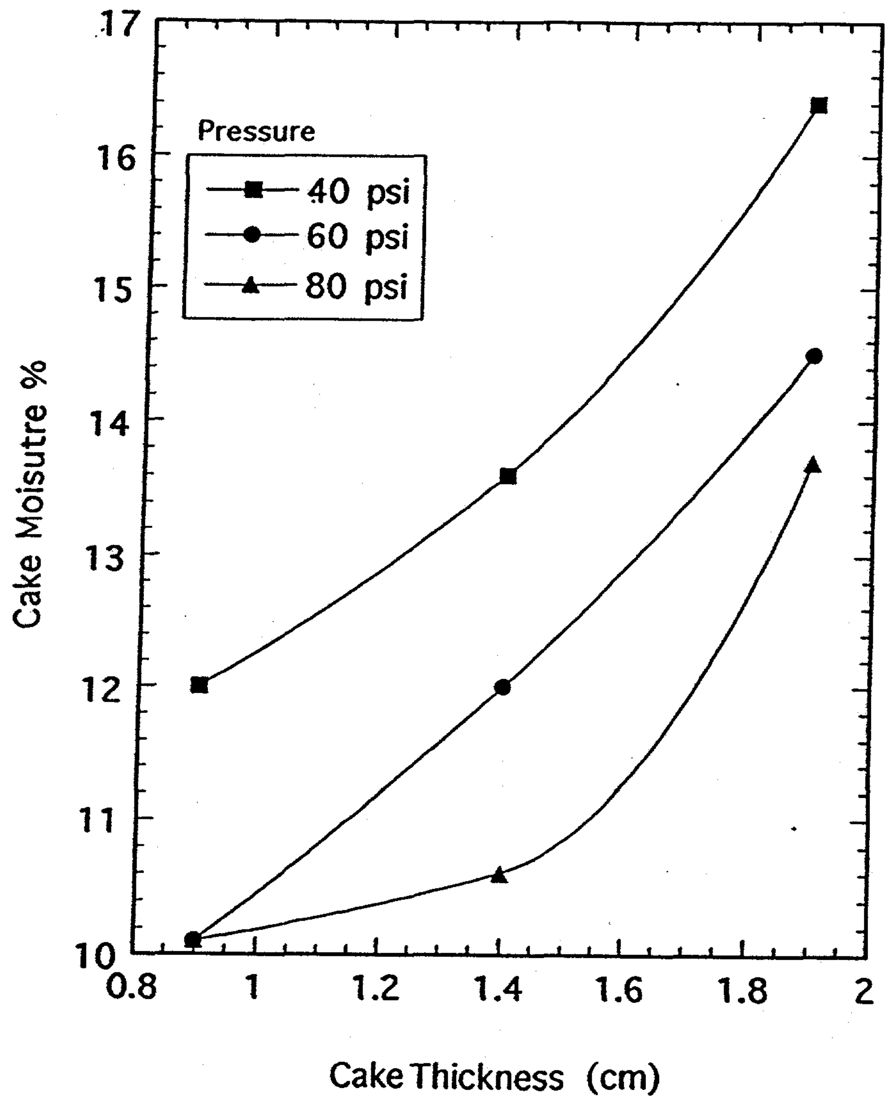

Figure 9-8. Cake moisture as a function of cake thickness under various applied pressures for the Pocahontas No. 3 coal. (filtration time: 2 minutes)

Discussion:

The above experimental results have shown that the interactions exist among cake thickness, applied pressure, and filtration time. In this study, the relationship between applied pressure gradient and cake moisture was investigated to identify the critical cake thicknesses and limits of fine coal dewatering under the set of given conditions. The applied pressure gradient is defined as:

$$\textit{Applied Pressure Gradient} = \frac{\textit{Applied Pressure}}{\textit{Cake Thickness}}$$

Figure 9-9 shows the effect of the applied pressure gradient on the cake moisture for the Pocahontas No.3 coal. It can be seen that almost all the data obtained at various cake thicknesses are on the same curve. At the same applied pressure gradient, the moisture will be the same for different cakes thickness. The lowest moisture obtained for 1.9 cm thick cake was 14 percent. In other words, to achieve a cake moisture of less than 14 percent, the cake thickness must be less than 1.9 cm. It is indicated that for a given moisture specification, there is a critical cake thickness, beyond which it will be difficult to achieve that moisture specification. The critical cake thickness for the Pocahontas No. 3 coal was 1.4 cm for moisture ranging from 10.5 to 14 percent. The applied pressure gradient is a very important parameter in a hyperbaric filtration process. The relationship between applied pressure gradient

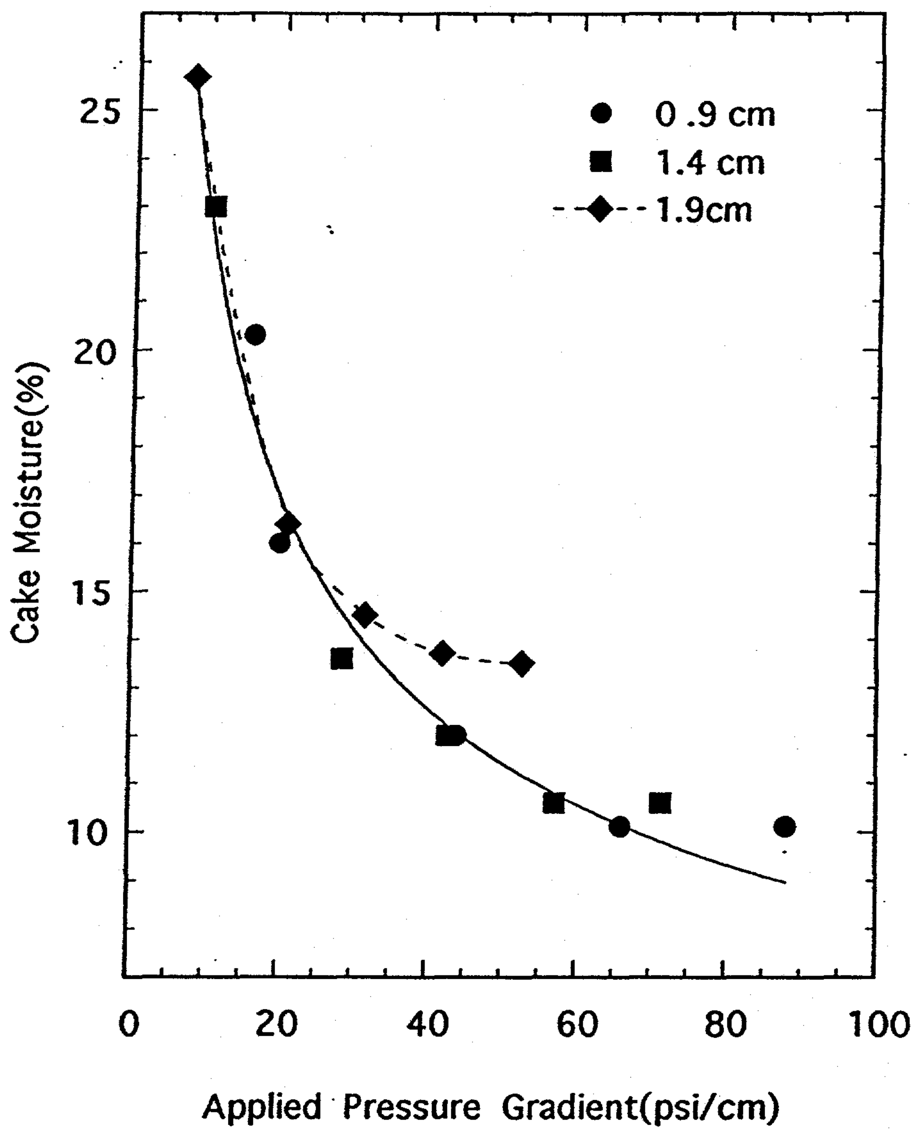

Figure 9-9. Effect of applied pressure on optimum filtration time for the Pocahontas No. 3 coal.

and cake moisture can provide information of how the applied pressure should be adjusted as the cake thickness changes. When the cake thickness changes and remains below the critical cake thickness, the applied pressure gradient should be kept constant to achieve the same cake moisture.

The equation for the curve in Figure 9-9 could be written as:

$$C_m = 61.44(\Delta P/L)^{-0.43} \quad (1)$$

where C_m is cake moisture and $(\Delta P/L)$ is applied pressure gradient. The relation between capillary number (N_{cap}) and pressure gradient ($\Delta P/L$) can be expressed as (Wakeman, 1979a):

$$N_{cap} = \{(\varepsilon^3 d^2)/[(\varepsilon-1)^2 \tau \cos\theta]\}(\Delta P/L) \quad (2)$$

For an incompressible cake, the value of $(\varepsilon^3 d^2)/[(\varepsilon-1)^2 \tau \cos\theta]$ can be regarded as a constant. Therefore, the relationship between capillary number and pressure gradient can be simplified as:

$$N_{cap} = A(\Delta P/L) \quad (3)$$

where $A = [\varepsilon^3 d^2]/[(\varepsilon-1)^2 \tau \cos\theta]$. According to Wakeman (1979b), the correlation between irreducible equilibrium saturation (S_∞) and N_{cap} is:

$$S_{\infty} = 0.155(1+0.031N_{cap}^{-0.49}) \tag{4}$$

Accordingly, the relationship between S_{∞} and pressure gradient is expressed as:

$$S_{\infty} = B+C(\Delta P/L)^{-0.49} \tag{5}$$

where B and C are the constants related to the properties of the cake and filtrate.

It can be inferred, by comparing equation 1 with equation 5, that equation 1 is a specific expression of equation 5. Therefore, the curve in Figure 9-9 can be regarded as the limitation curve of cake moisture for the Pocahontas No. 3 coal. Moisture below this curve cannot be obtained without using an enhancement method.

Effect of pH:

pH is an important parameter in any process with respect to bulk chemistry and surface chemistry of solids suspended in an aqueous medium. In a solid/liquid/gas three phase system, pH affects the chemical species distribution in bulk, zeta potential at the interface of solid/liquid, and other surface chemical properties. The effects of pH on zeta potential and cake moisture for the Pocahontas No. 3 coal are presented in Figure 9-10. The cake moisture remained constant as the pH was changed from 3 to 9. In this pH range, even though the zeta potential of the coal changed from positive 10 mv to negative 20 mv, the effect of zeta potential was negligible in this pH range because of its low absolute value. When pH of the suspension is increased to

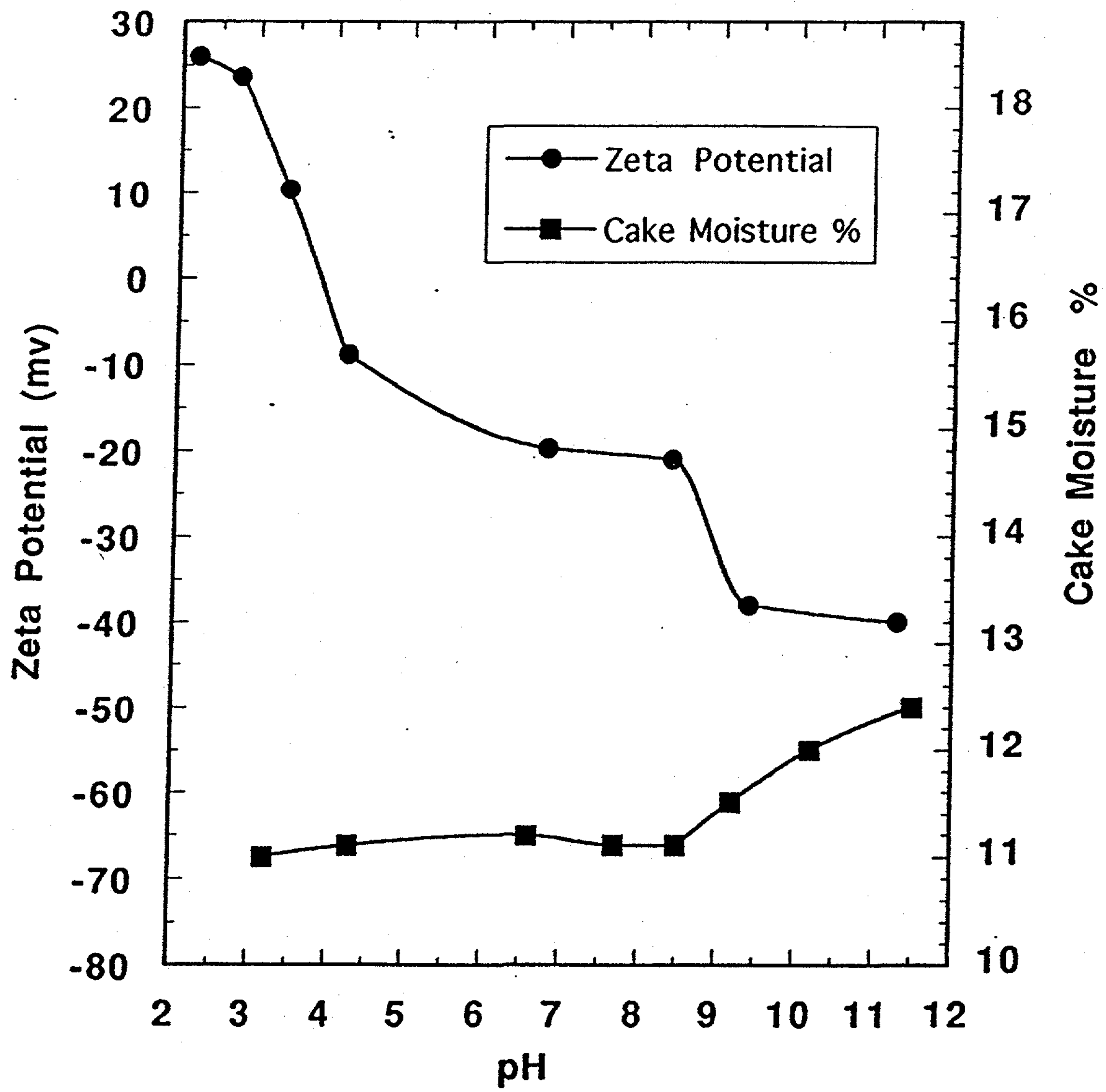

Figure 9-10. Effect of pH on zeta potential and cake moisture for the Pocahontas No. 3 coal.

9 or above, both the absolute value of the zeta potential increased and the cake moisture increased as the pH rose.

Temperature:

The relationship between cake moisture and filtration time at various temperatures for the Pocahontas No.3 coal shown in Figure 9-11 indicates that the increase of slurry temperature not only significantly reduces the cake moisture but also decreases the optimum filtration time. For the Pocahontas No.3 coal slurry, at 40 °C, the optimum filtration time is 1.5 minutes and corresponding cake moisture is about 11 percent, while at 80 °C, the optimum filtration time is one minute and corresponding cake moisture is 7.4 percent. The results show that increasing slurry temperature is very efficient in improving filtration process. Even though many researchers emphasized that the main effect of temperature on filtration process is to change filtrate viscosity, it was found in present study that the increase of temperature water from 20 ^{0}C to 80 ^{0}C lowered surface tension from 72 to 40 dynes/cm and the contact angle increased from 70^0 to 95^0, as temperature increased from 20 ^{0}C to 60 ^{0}C. Thus, it appears that higher temperature has a combined effect of lower viscosity, lower surface tension and higher hydrophobicity of coal in reducing filter cake moisture.

Particle Size:

Particle size plays a very important role in a filtration process. The dewatering data of various size fractions of the Pocahontas No. 3 coal are listed in Table 2.

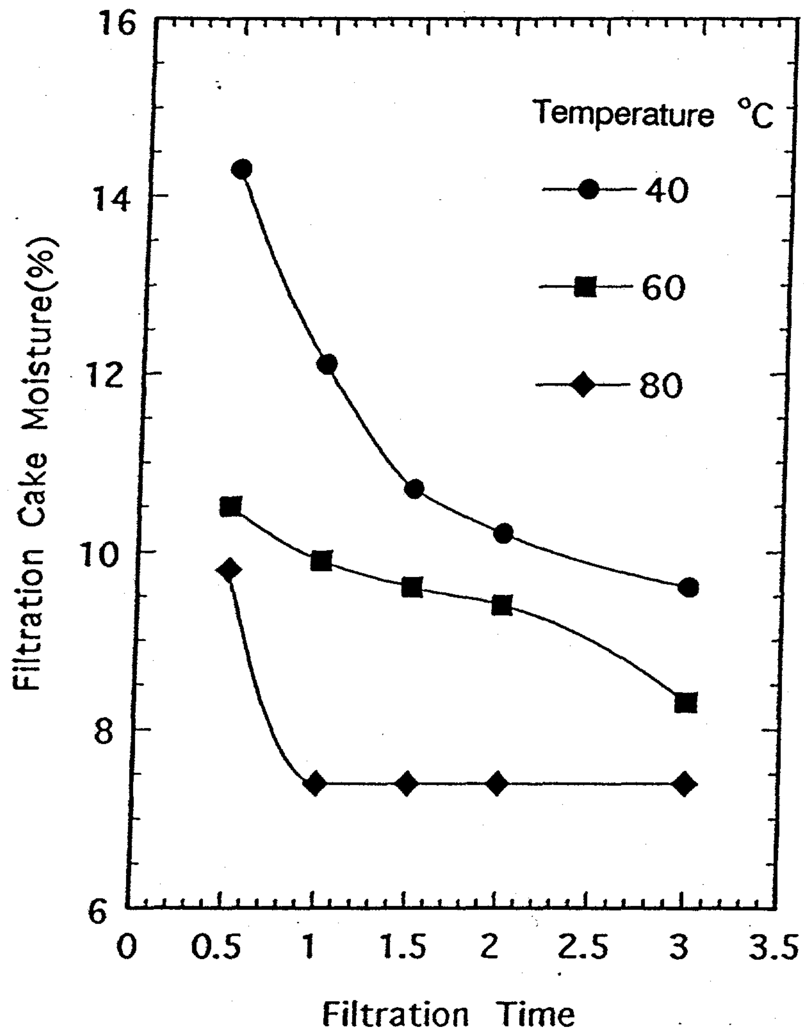

Figure 9-11. Effect of temperature on cake moisture for the Pocahontas No. 3 coal.

Table 2. Cake Moisture of Various Size Fractions of the Pocahontas No. 3 Coal Slurry Sample

Size Fraction (mesh)	Cake Moisture (%)
28x100	2.5
28x200	3.3
28x400	7.2
-100	15.42
-200	16.47
-400	23.9
Feed (as received)	11.0

As expected, the cake moisture increases as the particle size becomes smaller. It can also be noticed that plus 200 mesh size fraction has a much lower cake moisture than minus 200 mesh size fractions. The particle sizes can be artificially increased using either by flocculation or agglomeration technique.

Phase 3 - Field Testing:

Field testing with the Andritz Ruthner Inc. hyperbaric filter pilot plant unit were conducted at Consol Inc., Bailey mine located at South-west of Washington, PA in Green County. At this plant, Pittsburgh No. 8 which is a High Volatile B rank coal is being mined and cleaned. Table 3 lists the total number of pilot scale tests conducted at the Bailey plant. Note, that 38 tests were conducted on filter feed (28x0 mesh), 54 tests on froth flotation product (100x0 mesh), and 12 tests on a classified froth flotation product. The pilot scale data for all the 104 tests conducted at the Bailey

Table 3. List of the pilot scale hyperbaric filter tests conducted at the Consol Inc. Bailey Mine.

A. FILTER FEED MATERIAL[a] (38 TESTS)

1. Experimental Design (15 Tests)
2. Anionic Floc (8 Tests)
3. Cationic Surfactant (6 Tests)
4. Wood Pulp (2 Tests)

B. FROTH FLOTATION PRODUCT[b] (54 TESTS)

1. Experimental Design (15 Tests)
2. Anionic Floc (5 Tests)
3. Coagulant (6 Tests)
4. Cationic Surfactant (5 Tests)
5. Anionic Surfactant (7 Tests)
6. Anionic Surfactant and Cu Ion (6 Tests)
7. Wood Pulp (5 Tests)
8. Pressure Variation (4 Tests)
9. Other (1 Test)

C. CLASSIFIED FROTH FLOTATION PRODUCT (12 TESTS)

1. Size, Pressure, and Solids Content Variation (6 Tests)
2. Anionic Floc (2 Tests)
3. Coagulant (4 Tests)

---

[a]28x0 mesh
[b]100x0 mesh

mine are listed in Table 4. Table 5 provides a brief summary of all the data. The data shows that for the filter feed (28x0 mesh) filter cake moisture was below 20 percent, and for the forth flotation product (100x0 mesh), the filter cake moisture was above 20 percent but below 25 percent. Classified froth flotation product did not show any improvement in filter cake moisture, however, the throughput of coal in the unit was increased with little lower air requirement.

FUTURE PLANS

Additional laboratory studies on dewatering of Pittsburgh No. 8 coal are being conducted using 'oriemulsion' and specially synthesized flocculant. The pilot scale data obtained at Bailey and Buchanan mines are being evaluated in detail. A technical presentation of the pilot data will be made at Coal Prep 95 meeting in May 95.

Table 4. Andritz hyperbaric filter pilot plant data obtained at Bailey mine

Data HBF-Trailer

Projectnr.: 88-24141/0
Code: Hyperbaric filtration
customer: Consol USA
Bailey Mine

Material: Coal concentrate
filter fabric GKD Talarmesa 185/28F 07-1020-4086-USA Tatko
Filter area 2.0 m²

H.Riemer AAG / John Hughes Arus
Gary Meenan/Jim Devore
Randy Kosky
Anthony Fonseca

Test-nr.	Date DD/MM/YY	Test-time hh:mm	Type conc.	Vessel pres. bar	Pres CFA bar	Pres-CDA bar	Disc-speed rpm	feed-flow m³/h	Ash Lab %TS	FS Lab %TS	Part. > 600µ %TS	Part. < 63µ %TS	feed-temp. °C	Air Compr. Nm³/h	Air filtrate Nm³/h	filtrate flow m³/h	Filtrat %	air cor-factor Nm3/h	act. filtr.air Nm3/h	Polym. flow l/h	Polym. Type	Polym conc. %	Polym. dose g/t	Discharg time sec.	Nr. of disch. cycle	Cake weight kg	Throughp. dry kg/m²h	hroughp dry lb/ft²h	Moi-sture %	Air-consumpt. Nm³/t	Air-consump cfm/ton	cake thickness mm
Filter Feed (28x0 Mesh)																																
CFA = 85 deg																																
##1	12/09/1994	14:26	28/0	5.0	0.0	0.0	2.01	10.5		38.7			28	213	355	6.90	0.6	130	225	0	none	0.0	0	59	1	76	1936	397	16.5	58	31	21.0
##2	13/09/1994	09:48	28/0	5.0	0.0	0.0	1.93	5.2		36.6			28	227	366	3.60	0.8	130	236	0	none	0.0	0	60	1	41	1033	212	16.0	114	61	11.5
##3	13/09/1994	10:04	28/0	5.0	0.0	0.0	0.46	3.0		34.1			28	152	269	1.87	0.5	130	139	0	none	0.0	0	60	1	21.2	539	111	15.8	129	69	19.0
##4	13/09/1994	17:42	28/0	5.0	0.0	0.0	1.95	5.8		36.9			29	366	447	4.07	0.3	130	317	0	none	0.0	0	60	1	47	1197	245	15.7	132	71	12.0
##5	13/09/1994	17:59	28/0	5.0	0.0	0.0	0.47	3.4		36.7			28	236	290	2.17	0.1	130	160	0	none	0.0	0	59	1	19.2	490	101	16.3	163	87	18.0
##6	14/09/1994	13:09	28/0	3.5	0.0	0.0	1.29	5.4		36.7			30	280	339	3.36	0.2	55	284	0	none	0.0	0	58	1	42.8	1112	228	17.0	128	68	16.8
##7	14/09/1994	14:30	28/0	2.0	0.0	0.0	2.03	5.9		26.2			30	225	304	3.57	0.2	0	304	0	none	0.0	0	55	1	38.5	1030	211	18.1	148	79	12.9
##8	14/09/1994	16:15	28/0	3.5	0.0	0.0	1.29	7.0		36.3			30	306	377	4.60	0.1	55	322	0	none	0.0	0	58	1	50.8	1306	268	17.4	123	66	21.9
##9	15/09/1994	08:46	28/0	2.0	0.0	0.0	0.48	5.4		22.6			26	220	338	3.20	0.1	0	338	0	none	0.0	0	56	1	33.2	863	177	19.4	196	104	34.0
##10	16/09/1994	08:45	28/0	3.5	0.0	0.0	1.28	5.9		38.5			24	235	345	3.86	0.3	55	290	0	none	0.0	0	58	1	44.5	1132	232	18.3	128	68	19.4
##11	16/09/1994	10:20	28/0	0.7	0.0	0.0	0.48	3.9		25.9			25	99	145	1.82	0.1	0	145	0	none	0.0	0	55	1	22.6	585	120	20.7	124	66	20.4
Froth Feed (100x0 Mesh)																																
CFA = 85 deg																																
##12	16/09/1994	13:10	100/0	5.0	0.0	0.0	1.28	5.1		19.7			30	488	555	4.44	0.2	130	425	0	none	0.0	0	60	1	24.1	564	116	22.5	377	201	12.4
##13	16/09/1994	15:20	100/0	3.5	0.0	0.0	1.93	4.0		18.8			30	457	538	3.40	0.3	55	483	0	none	0.0	0	57	1	18.4	462	95	20.0	522	278	11.9
##14	19/09/1994	09:05	100/0	3.5	0.0	0.0	0.48	3.4		20.1			26	436	513	2.61	0.3	55	458	0	none	0.0	0	58	1	16.8	399	82	23.5	575	307	11.9
##15	19/09/1994	10:30	100/0	2.0	0.0	0.0	1.26	4.1		20.4			26	264	382	3.01	0.4	0	382	0	none	0.0	0	56	1	18.8	457	94	24.4	418	223	9.9
Froth Feed																																
CFA = 125 deg																																
##16	20/09/1994	08:30	100/0	5.0	0.0	0.0	1.98	7.2		20.4			27	398	479	5.64	0.3	120	359	0	none	0.0	0	60	1	37.4	852	175	24.0	211	113	10.7
##17	20/09/1994	09:55	100/0	4.6	0.0	0.0	0.50	3.6		20			28	492	565	3.20	0.2	85	480	0	none	0.0	0	59	1	17.9	404	83	25.9	594	317	18.0
##18	20/09/1994	11:55	100/0	3.5	0.0	0.0	1.29	4.5		20.5			28	379	453	3.12	0.3	25	428	0	none	0.0	0	57	1	21	500	103	24.4	427	228	10.4
##19	20/09/1994	13:15	100/0	2.0	0.0	0.0	2.02	5.0		20.3			28	248	335	3.40	0.3	0	335	0	none	0.0	0	55	1	22.5	545	112	25.9	307	164	8.1
##20	20/09/1994	14:30	100/0	3.5	0.0	0.0	1.26	5.1		21.1			28	233	314	3.96	0.3	25	289	0	none	0.0	0	57	1	25.9	635	130	22.3	228	122	11.6
##21	20/09/1994	16:40	100/0	2.0	0.0	0.0	0.48	2.9		20.7			28	182	223	2.04	0.2	0	223	0	none	0.0	0	55	1	13.8	344	71	23.7	323	173	15.8
##22	21/09/1994	09:05	100/0	3.5	0.0	0.0	1.25	5.3		20			28	248	355	4.22	0.2	25	330	0	none	0.0	0	58	1	26.3	632	129	22.7	261	139	11.4
##23	21/09/1994	11:35	100/0	0.7	0.0	0.0	0.50	2.5		20			28	58	27	1.30	0.3	0	27	0	none	0.0	0	56	1	9.66	220	45	29.6	61	32	11.0
Froth Feed																																
CFA = 125 deg																																
##24	21/09/1994	12:51	100/0	1.5	0.0	0.0	1.24	4.5		18.9			28	128	240	3.63	0.3	0	240	0	none	0.0	0	55	1	25.1	613	126	25.0	196	104	11.5
##25	21/09/1994	13:10	100/0	2.7	0.0	0.0	1.24	4.4		20.3			28	215	335	3.61	0.2	0	335	0	none	0.0	0	56	1	22	544	111	23.1	308	165	11.0
##26	21/09/1994	13:25	100/0	4.2	0.0	0.0	1.24	5.1		20.6			28	265	365	4.26	0.1	60	305	0	none	0.0	0	58	1	25.7	622	128	22.0	245	131	11.5
##27	21/09/1994	13:40	100/0	5.8	0.0	0.0	1.24	6.1		20.7			28	335	377	5.03	0.1	166	211	0	none	0.0	0	59	1	30.5	728	149	21.6	145	77	13.0
Filter Feed																																

Table 4 (continued)

Data HBF-Trailer

Projectnr.: 88-24141/0
Code: Hyperbaric filtration
customer: Consol USA
Bailey Mine

Material: Coal concentrate
filter fabric: GKD Telarease 165/28F 07-1020-k086-USA Tetko
Filter area 2.0 m²

H.Riemer AAG / John Hughes Arus
Gary Meenan/Jim Devore
Randy Kosky
Anthony Fonseca

Test-nr.	Date	Test-time	Type conc.	Vessel pres.	Pres CFA	Pres-CDA	Disc-speed	feed-flow	Ash Lab	FS Lab	Part. >500µ	Part. <63µ	feed-temp.	Air Compr.	Air filtrate	filtrate flow	Filtrat	air cor-factor	act. filtr.air	Polym. flow	Polym. Type	Polym conc.	Polym. dose	Discharg time	Nr. of disch.	Cake weight	Throughp. dry	hroughp dry	Moi-sture	Air-consumpt.	Air-onsump	cake thickness
	DD/MM/YY	hh:mm		bar	bar	bar	rpm	m³/h	%TS	%TS	%TS	%TS	°C	Nm³/h	Nm³/h	m³/h	%	Nm3/h	Nm3/h	l/h		%	g/t	sec.	cycle	kg	kg/m²h	lb/ft²h	%	Nm³/t	cfm/ton	mm
Filter Feed																																
CFA = 125 deg																																
##28	21/09/1994	14:25	28/0	3.5	0.0	0.0	1.94	4.6		38.8			29	260	342	3.24	0.3	25	317	0	none	0.0	0	57	1	33.3	873	179	16.7	182	97	12.0
##29	21/09/1994	17:00	28/0	3.5	0.0	0.0	0.43	3.3		36.2			30	204	223	2.06	0.1	25	198	0	none	0.0	0	57	1	23.7	617	126	17.3	161	86	28.8
##30	22/09/1994	09:10	28/0	5.0	0.0	0.0	1.23	5.8		34.8			28	278	378	4.16	0.2	120	258	0	none	0.0	0	59	1	44.1	1118	229	17.1	115	62	18.9
##31	22/09/1994	11:50	28/0	2.0	0.0	0.0	1.24	4.9		26.8			28	123	204	3.11	0.3	0	204	0	none	0.0	0	56	1	32	832	170	19.3	123	65	15.8
Froth Feed																																
CFA = 165 deg																																
##32	23/09/1994	08:31	100/0	2.0	0.0	0.0	1.23	5.3		19.8			28	110	231	4.32	0.2	0	231	0	none	0.0	0	56	1	25.4	608	125	25.5	190	101	12.5
##33	23/09/1994	09:50	100/0	3.5	0.0	0.0	1.95	7.0		20			28	180	283	5.77	0.1	0	283	0	none	0.0	0	58	1	35.2	832	171	24.1	170	91	10.9
##34	23/09/1994	11:00	100/0	3.5	0.0	0.0	0.47	4.6		19.4			28	285	345	3.89	0.1	0	345	0	none	0.0	0	57	1	22.5	531	109	25.1	325	173	24.4
##35	23/09/1994	12:15	100/0	5.0	0.0	0.0	1.25	7.5		20.3			28	271	318	6.27	0.1	0	318	0	none	0.0	0	49	1	33.1	935	192	23.7	170	91	16.6
Filter Feed																																
CFA = 165 deg																																
##36	21/09/1994	17:42	28/0	5.0	0.0	0.0	1.82	5.9		37.8			28	207	272	4.09	0.2	0	272	0	none	0.0	0	42	1	36.8	1310	269	17.4	104	55	16.5
##37	21/09/1994	17:49	28/0	5.0	0.0	0.0	1.24	5.4		36.6			28	220	251	3.75	0.1	0	251	0	none	0.0	0	41	1	32.5	1166	239	17.5	108	57	23.0
Filter Feed																																
CFA = 55 deg																																
##38	26/09/1994	10:30	28/0	2.0	0.0	0.0	1.25	3.7		29			28	244	418	2.26	0.3	10	408	0	none	0.0	0	47	1	18.3	580	119	17.5	351	187	12.0
##39	26/09/1994	12:30	28/0	3.5	0.0	0.0	2.01	3.9		31.8			28	441	539	2.67	0.3	120	419	0	none	0.0	0	48	1	21.9	691	142	15.5	304	162	10.9
##40	26/09/1994	13:45	28/0	3.5	0.0	0.0	0.51	2.9		29.9			28	303	390	1.90	0.1	120	270	0	none	0.0	0	48	1	16.8	526	108	16.4	257	137	19.9
##41	26/09/1994	15:00	28/0	5.0	0.0	0.0	1.30	3.8		36			28	404	482	2.60	0.3	160	322	0	none	0.0	0	49	1	24.3	753	154	15.4	214	114	12.9
##42	26/09/1994	16:10	28/0	5.0	0.0	0.0	1.22	3.9		35.9			28	247	310	2.54	0.3	160	150	0	none	0.0	0	49	1	24.7	732	150	18.8	102	55	12.5
Test with polymer Nalco 8856 0.1%																																
CFA = 55 deg																																
##43	26/09/1994	16:50	28/0	5.0	0.0	0.0	1.27	9.3		33.3			28	439	521	6.32	0.0	160	361	225	none	0.1	59	41	1	54.2	1897	389	19.7	95	51	28.0
##44	26/09/1994	17:05	28/0	3.3	0.0	0.0	1.33	8.8		32.8			28	496	621	5.99	0.1	110	511	218	none	0.1	61	39	1	48.5	1797	368	19.7	142	76	26.0
##45	26/09/1994	17:20	28/0	2.0	0.0	0.0	1.23	7.6		32.6			28	472	545	4.89	0.1	10	535	189	none	0.1	73	37	1	34.2	1306	268	21.5	205	109	24.0
Filter Feed																																
CFA = 85																																
Polymer Nalco 9810 0.1%																																
##46	27/09/1994	14:26	28/0	3.5	0.0	0.0	1.24	6.8		35.1			28	222	354	4.49	0.2	50	304	32	none	0.1	12	49	1	45	1372	281	17.8	111	59	21.0
Filter Feed																																
CFA = 85 Polymer Nalco 9810 0.1%																																

Data HBF-Trailer

ANDRITZ

Projectnr.: 88-24141/0
Code: Hyperbaric filtration
customer: Consol USA
Bailey Mine

Material: Coal concentrate
filter fabric GKD Telarosa 165/ZBF 07-1020-4086-USA Tetko
Filter area 2.0 m²

H.Riemer AAG / John Hughes Arus
Gary Meenan/Jim Devore
Randy Kosky
Anthony Fonseca

Test-nr.	Date DD/MM/YY	Test-time hh:mm	Type conc.	Vessel pres. bar	Pres CFA bar	Pres-CDA bar	Disc-speed rpm	feed-flow m³/h	Ash Lab %TS	FS Lab %TS	Part. >500µ %TS	Part. <83µ %TS	feed-temp. °C	Air Compr. Nm³/h	Air filtrate Nm³/h	filtrate flow m³/h	Filtrat %	air cor-factor Nm3/h	act. filtr.air Nm3/h	Polym. flow l/h	Polym. Type	Polym conc. %	Polym. dose g/t	Discharg time sec.	Nr. of disch. cycle	Cake weight kg	Throughp. dry kg/m²h	hroughp dry lb/ft²h	Moi-sture %	Air-consumpt. Nm³/t	Air-onsump cfm/ton	cake thickness mm
Filter Feed																																
##47	29/09/1994	10:35	28/0	3.5	0.0	0.0	1.22	8.9		29.2			26	487	596	6.46	0.0	50	646	110	none	0.1	36	40	1	43.5	1543	316	21.2	177	94	26.0
##48	29/09/1994	11:11	28/0	5.0	0.0	0.0	1.22	8.7		30			27	496	562	6.42	0.1	100	462	45	none	0.1	15	42	1	44	1559	320	17.9	148	79	27.0
##49	29/09/1994	11:30	28/0	5.0	0.0	0.0	1.18	7.0		34.5			26	312	411	4.86	0.1	100	311	23	none	0.1	8	42	1	40	1443	296	15.8	108	58	22.0
Filter Feed CFA =85																																
correlation to ##47-##49																																
##50	29/09/1994	12:07	28/0	5.0	0.0	0.0	1.25	4.2		29.6			26	431	504	3.35	0.1	100	404	0	none	0.0	0	42	1	20	701	144	18.2	288	154	13.0
Filter Feed CFA= 85																																
correlation to ##51-##55																																
##56	29/09/1994	13:45	28/0	3.5	0.0	0.0	1.19	4.4		34.3			26	241	361	3.00	0.2	50	311	0	none	0.0	0	59	1	32.9	831	170	17.3	187	100	15.0
Filter Feed CFA= 85																																
Test with cationic surfactant																																
##61	29/09/1994	14:05	28/0	3.5	0.0	0.0	1.26	4.3		34.4			26	194	316	3.00	0.2	50	266	108	none	0.1	67	59	1	31.4	804	165	16.4	166	88	14.5
##52	29/09/1994	14:25	28/0	3.5	0.0	0.0	1.32	4.9		31.3			26	214	352	3.59	0.2	50	302	612	none	0.1	360	58	1	32.6	850	174	16.0	178	95	14.5
##53	29/09/1994	14:40	28/0	5.0	0.0	0.0	1.27	5.4		33.3			26	277	386	4.00	0.2	100	286	645	none	0.1	334	60	1	38.2	966	198	15.8	148	79	14.8
##54	29/09/1994	14:55	28/0	5.0	0.0	0.0	1.23	4.3		36.8			26	205	332	2.99	0.2	100	232	101	none	0.1	61	60	1	32.8	823	169	15.8	141	75	12.5
##55	29/09/1994	15:09	28/0	5.0	0.0	0.0	1.26	5.8		28.3			26	303	407	4.44	0.2	100	307	1095	none	0.1	546	60	1	39.6	1003	206	15.6	153	82	18.0
Filter Feed CFA=85																																
correlation to ##51-##55																																
##57	29/09/1994	15:30	28/0	5.0	0.0	0.0	1.15	4.7		37.3			26	302	409	3.35	0.2	100	309	0	none	0.0	0	60	1	39.3	980	201	16.9	158	84	16.0
Filter Feed CFA= 85																																
test with pulp 0.5%																																
##58	29/09/1994	16:20	28/0	5.0	0.0	0.0	1.28	4.8		34.7			26	371	469	3.53	0.2	100	369	334	none	0.5	1020	60	1	32.8	818	168	16.9	225	120	15.0
Froth Feed CFA= 165																																
correlation to ##60-##63																																
##59	03/10/1994	09:45	100/0	3.5	0.0	0.0	1.24	5.6		18			24	101	243	4.65	0.1	0	243	0	none	0.0	0	59	1	28.6	667	137	24.1	182	97	12.0
Froth Feed CFA= 165																																
Polymer Nalco 9810 0.1%																																
##60	03/10/1994	10:25	100/0	3.5	0.0	0.0	1.26	9.4		19.2			26	261	393	7.64	0.1	0	393	54	none	0.1	26	49	1	39	1047	215	27.2	188	100	19.0
##61	03/10/1994	11:15	100/0	3.5	0.0	0.0	1.26	7.7		20.3			24	216	347	6.30	0.1	0	347	34	none	0.1	18	48	1	33.2	931	191	24.7	186	99	17.5
##62	03/10/1994	11:30	100/0	5.0	0.0	0.0	1.27	8.4		20.1			26	260	373	7.06	0.1	0	373	34	none	0.1	17	50	1	37.3	1018	209	24.2	183	98	18.0
##63	03/10/1994	12:35	100/0	5.0	0.0	0.0	1.14	9.2		20.8			26	347	440	7.32	0.0	0	440	73	none	0.1	33	50	1	41.3	1111	228	25.3	198	106	22.0
Froth Feed CFA= 165																																

ANDRITZ

Data HBF-Trailer

Projectnr.: 88-24141/0
Code: Hyperbaric filtration
customer: Consol USA
Bailey Mine

Material: Coal concentrate
filter fabric GKD Telaresse 165/ZBF 07-1020-K086-USA Tecko
Filter area 2.0 m²

H.Riemer AAG / John Hughes Arus
Gary Meenan/Jim Devore
Randy Kosky
Anthony Fonseca

ANDRITZ

Test-nr.	Date DD/MM/YY	Test-time hh:mm	Type conc.	Vessel pres. bar	Pres CFA bar	Pres-CDA bar	Disc-speed rpm	feed-flow m³/h	Ash Lab %TS	FS Lab %TS	Part. >500µ %TS	Part. <63µ %TS	feed-temp. °C	Air Compr. Nm³/h	Air filtrate Nm³/h	filtrate flow m³/h	Filtrat %	air cor-factor Nm3/h	act. filtr.air Nm3/h	Polym. flow l/h	Polym. Type	Polym conc. %	Polym. dose g/t	Discharg time sec.	Nr. of disch. cycle	Cake weight kg	Throughp. dry kg/m²h	hroughp dry lb/ft²h	Moi-sture %	Air-consumpt. Nm³/t	Air-onsump cfm/ton	cake thickness mm
Filter Feed																																
correlation to ##60-##63																																
##64	03/10/1994	13:25	100/0	3.5	0.0	0.0	1.23	4.8		20.7			26	184	271	3.99	0.1	0	271	0	none	0.0	0	48	1	22.6	653	134	22.9	207	111	11.5
Froth Feed CFA= 165																																
Polymer Nalco 8856 0.1 %																																
##65	03/10/1994	14:05	100/0	3.5	0.0	0.0	1.28	5.5		20.3			26	190	279	4.53	0.1	0	279	68	none	0.1	50	48	1	23.2	673	138	22.6	207	111	12.5
##66	03/10/1994	14:20	100/0	3.5	0.0	0.0	1.30	6.5		20.3			26	203	302	5.32	0.1	0	302	154	none	0.1	96	48	1	27.2	802	164	21.4	188	100	13.8
##67	03/10/1994	14:35	100/0	5.0	0.0	0.0	1.37	8.5		20			26	272	343	7.08	0.1	0	343	200	none	0.1	97	49	1	36.7	1027	211	23.8	167	89	17.5
##68	03/10/1994	14:55	100/0	5.0	0.0	0.0	1.16	8.9		19.7			26	301	380	7.44	0.1	0	380	423	none	0.1	201	49	1	38	1052	216	24.1	181	96	19.0
##69	03/10/1994	15:12	100/0	5.0	0.0	0.0	1.20	9.6		18.7			26	367	442	8.27	0.0	0	442	941	none	0.1	435	49	1	39	1080	221	24.3	205	109	20.0
Froth Feed CFA =165																																
Surfactant Sigma 1-Hexadecylpyridinium Chloride																																
##70	04/10/1994	09:00	100/0	3.5	0.0	0.0	1.27	5.4		19.5			24	94	258	4.49	0.1	0	258	134	none	0.1	114	61	1	25.3	586	120	22.1	220	118	11.0
##71	04/10/1994	09:20	100/0	3.5	0.0	0.0	1.29	6.2		17.5			24	168	326	5.42	0.1	0	326	770	none	0.1	625	58	1	25.3	616	126	21.0	265	141	11.6
##72	04/10/1994	09:35	100/0	5.0	0.0	0.0	1.26	8.6		16			24	284	409	7.65	0.1	0	409	1068	none	0.1	660	60	1	34.2	809	166	21.1	253	135	14.6
##73	04/10/1994	09:45	100/0	5.0	0.0	0.0	1.25	8.3		17.9			26	243	365	7.20	0.1	0	365	207	none	0.1	120	60	1	36.9	860	176	22.2	212	113	16.0
Froth Feed CFA= 165																																
Correlation to ##70-##73																																
##74	04/10/1994	11:45	100/0	5.0	0.0	0.0	1.26	7.3		19.2			26	236	352	6.14	0.1	0	352	0	none	0.0	0	50	1	29.8	817	168	23.8	216	115	16.5
Froth Feed CFA= 165																																
Cu+0.05% + Surfactant anionic 0.1%																																
##75	04/10/1994	12:05	100/0	5.0	0.0	0.0	1.30	7.6		18.3			26	243	355	6.60	0.1	0	355	182	none	0.1	107	50	1	31	847	174	24.1	210	112	17.0
##76	04/10/1994	12:30	100/0	3.5	0.0	0.0	1.25	6.2		19.3			26	176	303	5.32	0.1	0	303	154	none	0.1	116	59	1	28.8	665	136	24.8	228	122	15.5
Froth Feed CFA=165																																
Cu+0.05% + Surfactant anionic 0.5%																																
##77	04/10/1994	16:00	100/0	3.5	0.0	0.0	1.30	5.6		20			26	136	273	4.92	0.1	0	273	142	none	0.5	552	59	1	28.2	642	132	25.4	213	113	14.5
##78	04/10/1994	16:15	100/0	5.0	0.0	0.0	1.30	6.7		20.2			26	182	298	5.83	0.1	0	298	166	none	0.5	525	60	1	35.2	790	162	25.2	189	101	17.0
##79	04/10/1994	16:30	100/0	5.0	0.0	0.0	1.23	5.8		20			26	184	276	5.15	0.1	0	276	152	none	0.5	488	60	1	34	778	160	23.7	177	95	15.5
##80	04/10/1994	16:45	100/0	5.0	0.0	0.0	1.34	6.6		20.1			26	136	249	6.03	0.1	0	249	324	none	0.5	1093	60	1	32.7	742	152	24.4	168	89	15.5
Froth Feed CFA= 165																																
Correlation to ##75-##80																																
##81	04/10/1994	17:05	100/0	5.0	0.0	0.0	1.37	6.8		20.8			26	217	329	5.69	0.1	0	329	0	none	0.0	0	60	1	36	806	165	25.4	204	109	16.0
Froth Feed CFA =165																																
Correlation to ##83-##86																																
##82	05/10/1994	08:55	100/0	3.5	0.0	0.0	1.28	6.6		19.7			26	136	305	5.41	0.1	0	305	0	none	0.0	0	58	1	33.3	710	146	30.9	215	115	14.8

Table 4 (continued)

Data HBF-Trailer

Projectnr.: 88-24141/0
Code: Hyperbaric filtration
customer: Consol USA
Bailey Mine

Material: Coal concentrate
filter fabric: GKD Tetrasse 165/ZBF 07-1020-4086-USA Tetko
Filter area: 2.0 m²

H.Riemer AAG / John Hughes Arus
Gary Meenan/Jim Devore
Randy Kosky
Anthony Fonseca

Test-nr.	Date DD/MM/YY	Test-time hh:mm	Type conc.	Vessel pres. bar	Pres CFA bar	Pres-CDA bar	Disc-speed rpm	feed-flow m³/h	Ash Lab %TS	FS Lab %TS	Part. >500µ %TS	Part. <63µ %TS	feed-temp. °C	Air Compr. Nm³/h	Air filtrate Nm³/h	filtrate flow m³/h	Filtrat %	air cor-factor Nm3/h	act. filtr.air Nm3/h	Polym. flow l/h	Polym. Type	Polym conc. %	Polym. dose g/t	Discharg time sec.	Nr. of disch. cycle	Cake weight kg	Throughp. dry kg/m²h	hroughp dry lb/ft²h	Moisture %	Air-consumpt. Nm³/t	Air-consump cfm/ton	cake thickness mm
Filter Feed																																
Froth Feed CFA = 165																																
Pulp 0.5%																																
##83	05/10/1994	10:55	100/0	3.5	0.0	0.0	1.30	5.7		19.8			26	168	317	4.79	0.1	0	317	341	none	0.5	1315	59	1	28.3	648	133	24.9	244	130	14.5
##84	05/10/1994	11:15	100/0	3.5	0.0	0.0	1.29	5.5		18.3			24	193	313	4.61	0.1	0	313	549	none	0.5	2353	59	1	25.4	583	119	24.8	268	143	13.0
##85	05/10/1994	11:35	100/0	5.0	0.0	0.0	1.31	6.3		16.5			24	221	341	5.46	0.1	0	341	627	none	0.5	2421	60	1	28.5	647	133	24.3	263	140	13.5
##86	05/10/1994	12:06	100/0	5.0	0.0	0.0	1.30	6.4		19.3			26	242	336	5.31	0.1	0	336	158	none	0.5	538	60	1	32.4	735	151	24.6	228	122	14.5
Froth Feed CFA = 165																																
0.05 % CuCO3, 0.1% Niaproof Surfactant 08 Anionic																																
##87	04/10/1994	14:45	100/0	5.0	0.0	0.0	1.29	6.2		19.6			26	260	350	5.31	0.1	0	350	157	none	0.1	110	62	1	32.6	715	146	24.5	245	131	15.0
##88	04/10/1994	15:45	100/0	3.5	0.0	0.0	1.31	6.4		19.1			26	220	335	5.35	0.1	0	335	162	none	0.1	118	58	1	30	688	141	25.8	243	130	15.0
0.05 % CuCO3, 0.5% Niaproof Surfactant 08 Anionic																																
##89	04/10/1994	16:15	100/0	3.5	0.0	0.0	1.32	5.6		20			26	157	278	4.88	0.1	0	278	425	none	0.5	1647	58	1	27.8	646	132	24.8	215	116	14.0
##90	04/10/1994	16:30	100/0	5.0	0.0	0.0	1.32	6.7		20			25	194	307	5.90	0.1	0	307	167	none	0.5	550	60	1	33.4	758	155	24.4	203	108	15.0
##91	04/10/1994	16:39	100/0	5.0	0.0	0.0	1.32	7.9		16.8			24	159	247	7.72	0.1	0	247	1381	none	0.5	4387	59	1	33.4	787	161	22.5	157	84	14.0
Froth Feed CFA = 165																																
Correlation to ##75-##80																																
##92	04/10/1994	17:00	100/0	5.0	0.0	0.0	1.37	7.2		20.4			26	221	347	5.99	0.1	0	347	0	none	0.0	0	59	1	36.7	837	172	25.0	207	111	15.0
Deslimed Froth Feed CFA = 165																																
3 cyclones open 10 PSI																																
##93	06/10/1994	09:20	100/0	3.5	0.0	0.0	1.32	7.6		14.9			20	169	308	6.75	0.1	0	308	0	none	0.0	0	59	1	29.5	712	146	21.6	217	116	14.0
##94	06/10/1994	10:50	100/0	5.0	0.0	0.0	1.28	8.1		16.4			20	233	332	7.19	0.0	0	332	0	none	0.0	0	60	1	32.5	767	157	21.6	217	116	16.5
Deslimed Froth Feed CFA = 165																																
3 cyclones open 15 PSI																																
##95	06/10/1994	11:45	100/0	3.5	0.0	0.0	1.31	8.0		13.8			20	245	357	7.10	0.1	0	357	0	none	0.0	0	58	1	30.4	735	151	22.1	243	130	15.6
##96	06/10/1994	14:00	100/0	4.9	0.0	0.0	1.24	8.0		16.2			20	211	349	7.01	0.0	0	349	0	none	0.0	0	59	1	35.1	825	169	23.0	211	113	17.0
Deslimed Froth Feed CFA = 165																																
3 cyclones open 20 PSI																																
##97	06/10/1994	14:45	100/0	3.5	0.0	0.0	1.20	5.6		24.3			24	200	249	4.04	0.1	0	249	0	none	0.0	0	58	1	35.1	827	170	24.4	150	80	17.9
##98	06/10/1994	16:40	100/0	5.0	0.0	0.0	1.18	5.6		22.2			26	242	272	4.06	0.1	0	272	0	none	0.0	0	60	1	34	769	158	24.6	177	94	19.5
Deslimed Froth Feed CFA = 165																																
Nalco 9810 0.08%																																
##99	07/10/1994	11:10	100/0	3.5	0.0	0.0	1.26	10.6		16.1			22	299	387	9.26	0.0	0	387	38	none	0.1	15	40	1	29.4	1021	209	23.1	190	101	17.9
##100	07/10/1994	12:55	100/0	5.0	0.0	0.0	1.23	12.4		15.5			22	343	388	10.69	0.0	0	388	48	none	0.1	17	42	1	33.9	1136	233	22.1	171	91	21.5

Table 4 (continued)

Data HBF-Trailer

Projectnr.: 88-24141/0
Code: Hyperbaric filtration
customer: Consol USA
Bailey Mine

Material: Coal concentrate
filter fabric GKD Telatrasse 165/ZBF 07-1020-k086-USA Tetko
Filter area 2.0 m²

H.Riemer AAG / John Hughes Arus
Gary Meenan/Jim Devore
Randy Kosky
Anthony Fonseca

Test-nr.	Date	Test-time	Type conc.	Vessel pres.	Pres CFA	Pres-CDA	Disc-speed	feed-flow	Ash Lab	FS Lab	Part. >500µ	Part. <63µ	feed-temp.	Air Compr.	Air filtrate	filtrate flow	Filtrat	air cor-factor	act. filtr.air	Polym. flow	Polym. Type	Polym conc.	Polym. dose	Discharg time	Nr. of disch.	Cake weight	Throughp. dry	hroughp dry	Moi-sture	Air-consumpt.	Air-onsump	cake thickness
	DD/MM/YY	hh:mm		bar	bar	bar	rpm	m³/h	%TS	%TS	%TS	%TS	°C	Nm³/h	Nm³/h	m³/h	%	Nm3/h	Nm3/h	l/h		%	g/t	sec.	cycle	kg	kg/m²h	lb/ft²h	%	Nm³/t	cfm/ton	mm
Filter Feed																																
Deslimed Froth Feed CFA=165																																
Nalco 8856 0.1%																																
##101	07/10/1994	14:20	100/0	3.5	0.0	0.0	1.23	10.3		14.7			22	248	297	9.17	0.0	0	297	248	none	0.1	128	40	1	27.9	964	198	22.8	154	82	16.9
##102	07/10/1994	15:40	100/0	5.0	0.0	0.0	1.16	11.7		14.7			24	390	449	10.17	0.0	0	448	275	none	0.1	136	42	1	30	1010	207	22.3	222	118	20.0
Deslimed Froth Feed CFA=165																																
Nalco 8856 0.1%																																
##103	07/10/1994	16:20	100/0	3.5	0.0	0.0	1.16	9.6		16			24	284	335	8.39	0.0	0	335	115	none	0.1	64	40	1	26.8	905	186	24.1	185	99	18.0
##104	07/10/1994	17:35	100/0	5.0	0.0	0.0	1.25	10.4		15.1			24	361	417	9.17	0.0	0	417	121	none	0.1	62	41	1	29.2	971	199	23.7	215	115	19.0

Table 5. Summary of pilot scale hyperbaric filter tests conducted at the Bailey Mine.

Material	% Moisture in Filter Cake	Throughput Dry lb/hr-ft^2	Air Consumption Scfm/t
Filter Feed (28x0 Mesh)	16-19	100-300	60-200
Froth Flotation Product (100x0 Mesh)	21-25	100-150	100-250
Classified Froth Flotation Product	21-25	130-160	100-150