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EVALUATION OF HYPERBARIC FILTRATION FOR FINE COAL DEWATERING

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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 I - Model Development:

The cake formation step in continuous hyperbaric filtration can be described using the expression (Quarterly Report, April 1995):

$$\frac{\Delta p c \theta_f}{2\pi R_s} = \frac{\mu_w \alpha}{2\rho_w} \frac{R_s}{N} + \frac{\mu_w \alpha_m}{\rho_w} \quad (1)$$

in which:

Δp	=	applied pressure (Pascals)
c	=	feed solids concentration (-)
θ_f	=	cake formation angle (radians)
R_s	=	solids throughput ($\text{kg}/\text{m}^2\text{s}$)
μ_w	=	liquid viscosity (Pascal-seconds)
α	=	specific cake resistance (m/kg)
ρ_w	=	liquid density (kg/m^3)
N	=	rotational speed (sec^{-1})
α_m	=	filter medium resistance (m^{-1})

Analysis of pilot-scale performance data for the Bailey Mine tests indicates reasonable agreement with the model using a constant value of $6.8 \times 10^{10} \text{ m}^{-1}$ for the medium resistance α_m and a specific cake resistance α which varies with feed size consist and applied pressure. Curve-fitting estimates of the specific cake resistance are summarized in Table 1; their variation with pressure is shown in Figure 11-1. A direct comparison of the calculated and experimental values for solids throughput is shown in Figure 11-2.

The comparisons given in Figure 11-2 indicate that cake formation in the pilot-scale hyperbaric filter is in general accordance with the model. The scatter in the results reflects a combination of experimental error and non-idealities such as non-uniform cake structure. It should be noted that the results shown in the figure include only those tests

Table 1. Estimated specific cake resistance for hyperbaric filtration of Pittsburgh seam coal.

$$\text{Medium resistance } \alpha_m = 6.8 \times 10^{10} \text{ m}^{-1}$$

Pressure (bar)	Specific Cake Resistance α (m/kg)	
	-28 mesh	-100 mesh
2	8.4×10^8	2.1×10^{10}
3.5	2.2×10^{10}	2.6×10^{10}
5	4.0×10^{10}	3.1×10^{10}

conducted in the absence of chemical reagents such as surfactants or flocculants.

Preliminary evaluation of tests in which reagents were used indicates significant departures from the base-line results. These differences provide a measure of the effects of reagent addition on cake structure. A more detailed analysis of these effects is in progress.

It is instructive to evaluate the variations in specific cake resistance in terms of the well-known Carman-Kozeny model for fluid flow through porous media. According to this model, the specific cake resistance can be expressed as

$$\alpha = \frac{k(1-\epsilon)}{\epsilon^3 \rho_s d_p^2} \quad (2)$$

where k is a factor which includes the tortuosity of the channels (pores) in the cake, ϵ is the cake porosity, ρ_s is the solid density and d_p is the (specific surface) mean particle diameter.

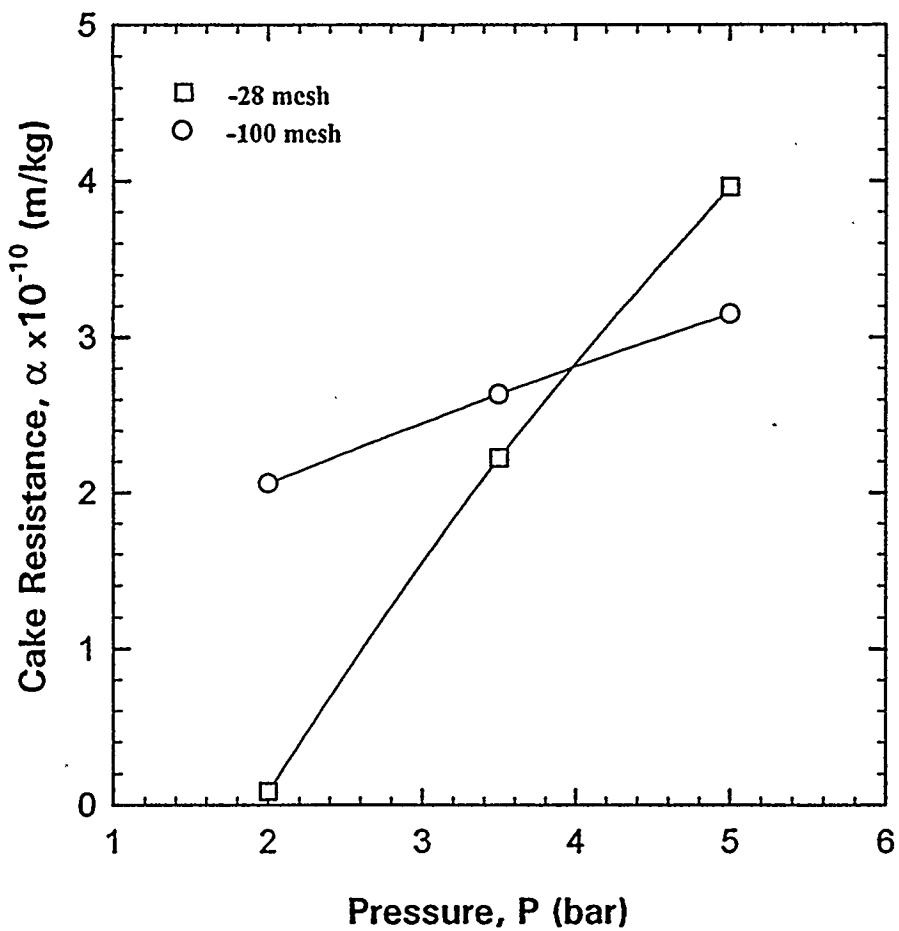


Figure 11-1. Effect of applied pressure on specific cake resistance in pilot-scale hyperbaric filtration of Pittsburgh seam coal.

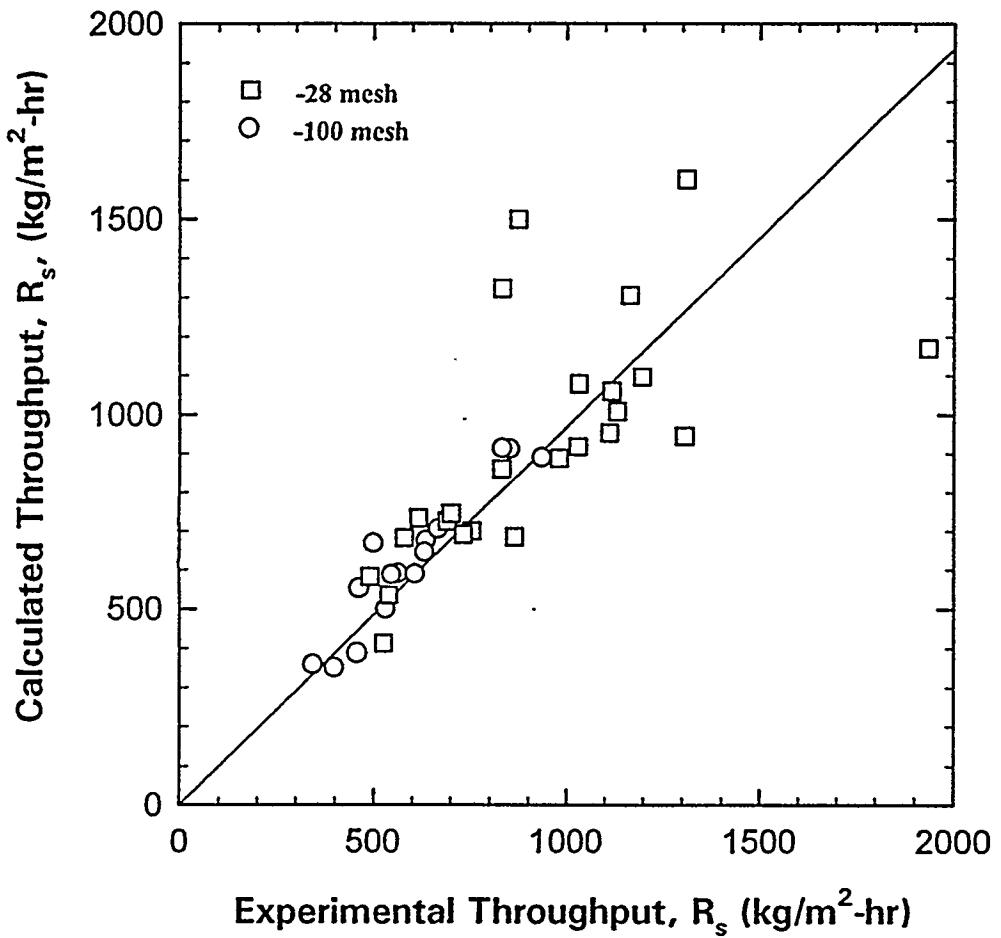


Figure 11-2. Comparison of calculated and observed solids throughput for hyperbaric filtration of Pittsburgh seam coal. Calculations based on Equation 1 with cake resistance values given in Table 1.

Referring to the results given in Table 1 and Figure 11-1, it is clear that the effects of particle size are qualitatively as expected but that the difference is somewhat less than would be predicted by Equation 2. The effects of pressure on cake resistance are usually attributed to cake compression which reduces porosity and, from Equation 2, increases α . Variations in cake porosity should be reflected in changes in cake (bulk) density ρ_b since

$$\rho_b = \rho_s(1 - \varepsilon) \quad (3)$$

It was shown in our last report (April 1995) that the bulk density of the cake can be estimated from

$$\rho_b = R_s / LN \quad (4)$$

where L is the measured cake thickness. The effects of pressure on the estimated bulk density are shown in Figures 11-3 and 11-4. In both cases there is considerable scatter in the data due, no doubt, to uncertainties in the measurements of cake thickness. While there may be some indication of cake densification with pressure, the effect would not be sufficient to account for the variations in specific cake resistance.

A more reasonable explanation for the apparent variations in cake resistance is that the cake structure is non-uniform. Flow of liquid during cake formation provides a driving force for the elutriation of fines through the relatively open structure formed from the coarser particles. Such material would accumulate at the inner surface of the cake, forming what amounts to a cake-within-a-cake. Since the inner "cake" would build-up in tandem with the total cake, the existence of such structures would not affect the applicability of the general model. The process could still be described using an overall,

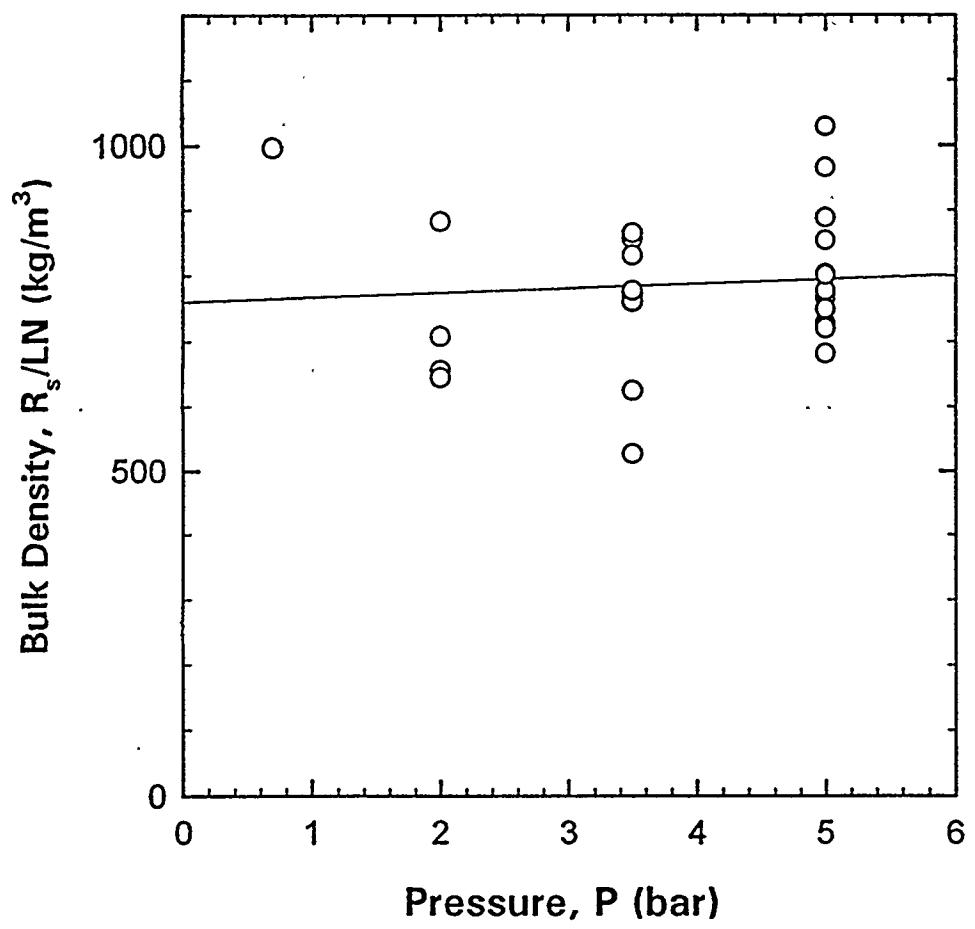


Figure 11-3. Variation in estimated cake bulk density for -28 mesh Pittsburgh seam coal.

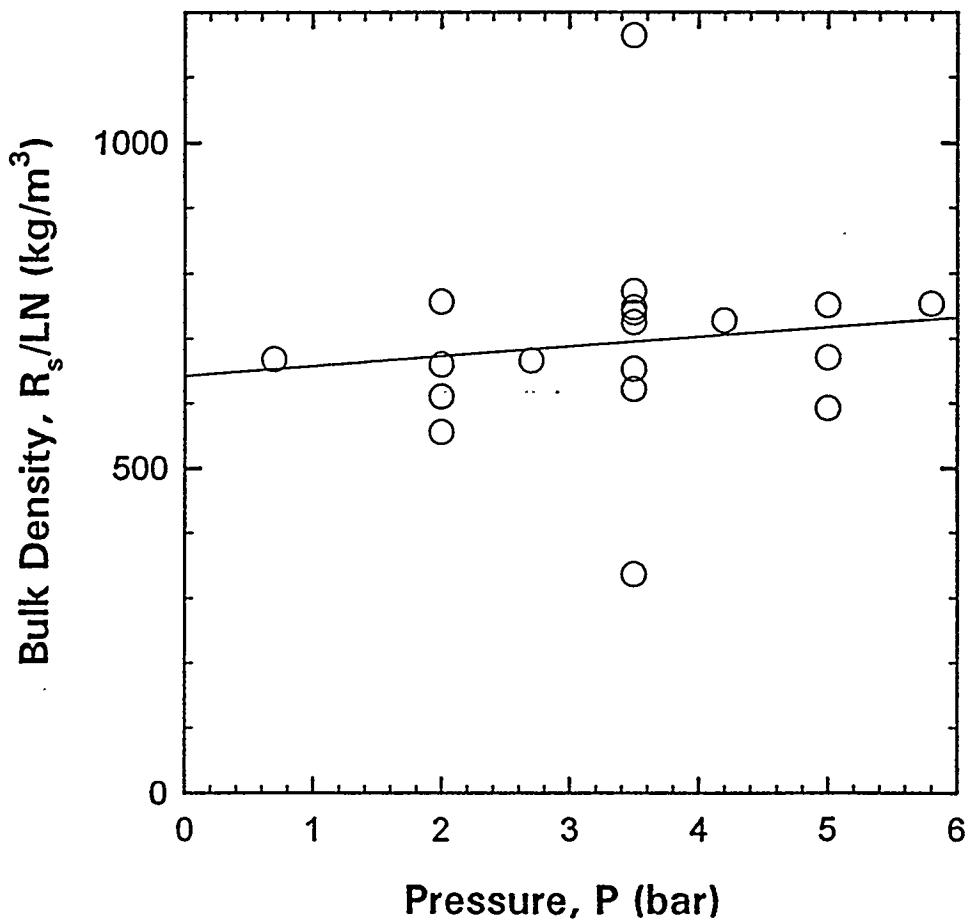


Figure 11-4. Variation in estimated cake bulk density for -100 mesh Pittsburgh seam coal.

effective cake resistance. However, the inner cake could be compressible, leading to an increased resistance with pressure but with little or no effect on the overall cake porosity.

The relatively large value for the medium resistance α_m is also somewhat surprising. The relative values of α and α_m imply that typically, about 25% of the pressure drop is across the filter medium with a considerably larger contribution for the coarser feed at low pressure. Partial blinding of the medium by penetration of fines during continuous operation may be responsible for this observation.

A simplified extension of the model to describe air consumption during the dewatering stage was outlined in our April 1995 report. The principal simplification involved neglect of the medium resistance which, on the basis of the results described above, is questionable. A more rigorous analysis of the process has now been carried out.

The mass flow rate of air through unit area of the cake can be expressed as:

$$\frac{G}{A} = \frac{k_a \bar{\rho}_a \Delta P}{\mu_a L} \quad (5)$$

where k_a is the permeability to air, μ_a is the viscosity of air and $\bar{\rho}_a$ is the average density of air in the cake. Expressed as a standard volumetric flow rate, Equation 5 becomes

$$\frac{Q_{st}}{A} = k_a \frac{\bar{P}}{P_{st}} \frac{\Delta P}{\mu_a L} \quad (6)$$

with

$$\frac{\bar{P}}{P_{st}} = \frac{\Delta P + 1}{2} \quad (7)$$

for hyperbaric filtration against an ambient pressure of one atmosphere.

At the end of the cake formation stage, the liquid flow can be similarly expressed as

$$\frac{Q_w}{A} = \frac{k_w \Delta P}{\mu_w L} \quad (8)$$

so that, from Equations 6, 7 and 8

$$\frac{Q_{st}}{A} = K \frac{\mu_w}{\mu_a} \left(\frac{\Delta P + 1}{2} \right) \frac{Q_w}{A} \quad (9)$$

where K is a relative permeability factor whose value should be unity for an idealized dry cake and less than unity for cakes with residual moisture (which causes partial blockage of pores).

The flow velocity Q_w/A can be directly related to the rate of solids accumulation R_s through

$$\frac{Q_w}{A} = \frac{1}{\rho_w c N} \frac{dR_s}{dt} \Big|_{t=t_f} \quad (10)$$

or, in terms of angles

$$\frac{Q_w}{A} = \frac{2\pi}{\rho_w c} \frac{dR_s}{d\theta} \Big|_{\theta=\theta_f} \quad (11)$$

Differentiation of Equation 1 leads to

$$\frac{dR_s}{d\theta} = \frac{\rho_w \Delta pcN}{2\pi \mu_w (\alpha R_s + N\alpha_m)} \quad (12)$$

The pressure term in Equation 12 can be eliminated using Equation 1. The result is

$$\frac{dR_s}{d\theta} \Big|_{\theta=\theta_f} = \frac{R_s}{2\theta_f} \left(\frac{\alpha R_s + 2N\alpha_m}{\alpha R_s + N\alpha_m} \right) \quad (13)$$

Substitution into Equations 9 and 11 and using the relative air consumption R_a defined, as before, by

$$R_a = \frac{Q_{st}(\theta_a - \theta_d)}{2\pi A R_s} \quad (14)$$

where θ_a is the dewatering angle and θ_d is that part of the dewatering angle taken up by the expulsion of bulk liquid, gives

$$R_a = \frac{K\mu_w(\theta_a - \theta_d)}{2\rho_w \mu_a c \theta_f} \left(\frac{\alpha R_s + 2N\alpha_m}{\alpha R_s + N\alpha_m} \right) \left(\frac{\Delta P + 1}{2} \right) \quad (15)$$

Using the previously-obtained approximation:

$$\theta_d \approx \frac{2\epsilon c p_w \theta_f}{(1-\epsilon) \rho_s} \quad (16)$$

the final expression for relative air consumption is:

$$R_a = \frac{K \mu_w}{2 \rho_w \mu_a} \left(\frac{\Delta P + 1}{2} \right) \left(\frac{\theta_a}{c \theta_f} - \frac{2\epsilon p_w}{(1-\epsilon) \rho_s} \left(\frac{\alpha R_s + 2N\alpha_m}{\alpha R_s + N\alpha_m} \right) \right) \quad (17)$$

Equation 17 is equivalent to the expression obtained by the simplified treatment but modified by the pressure correction and the term on the right hand-side. By inspection, it is clear that this term is relatively invariant, ranging in value from about 1.1 to 1.3 for the results being analyzed here.

Experimental data on relative air consumption are compared with the predictions of Equation 17 in Figures 11-5 and 11-6. It can be seen that, typically, the air consumption is close to the predicted value. However, the results show considerable scatter, especially for the finer coal. In some cases, the air consumption is as much as three times larger than predicted while for other conditions the observed value is substantially lower than the prediction.

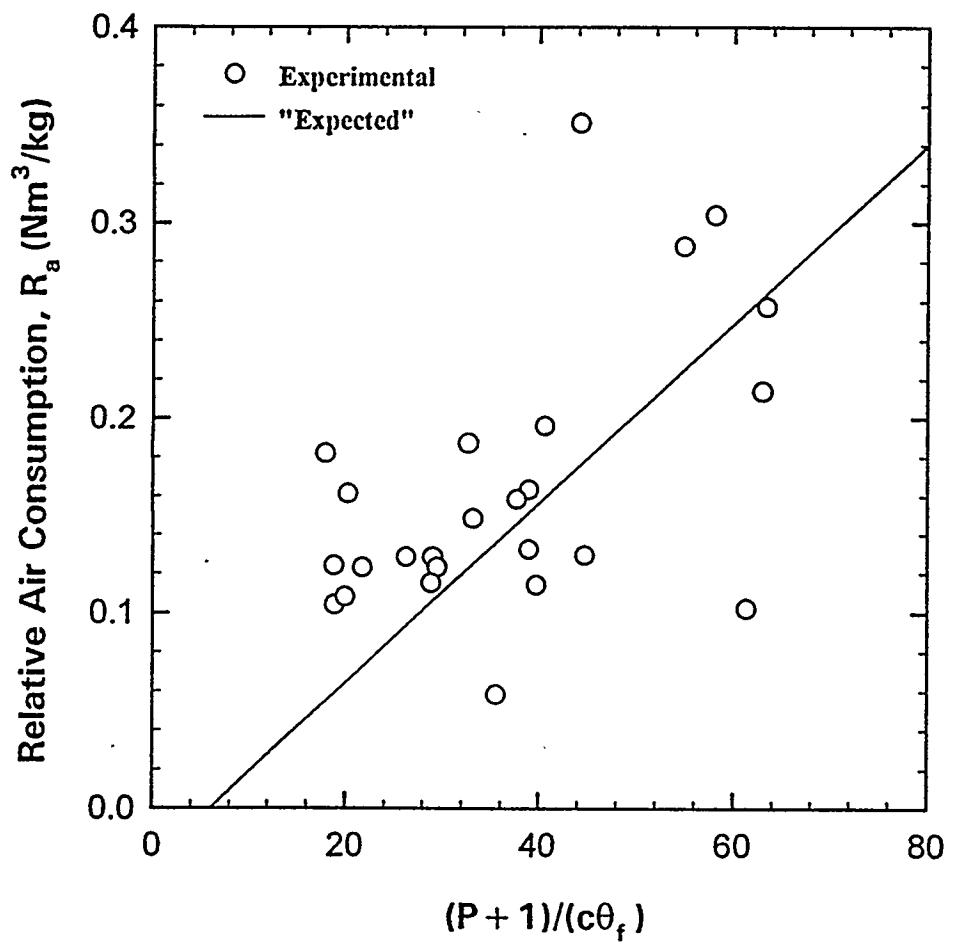


Figure 11-5. Relative air consumption in hyperbaric filtration of -28 mesh Pittsburgh seam coal.

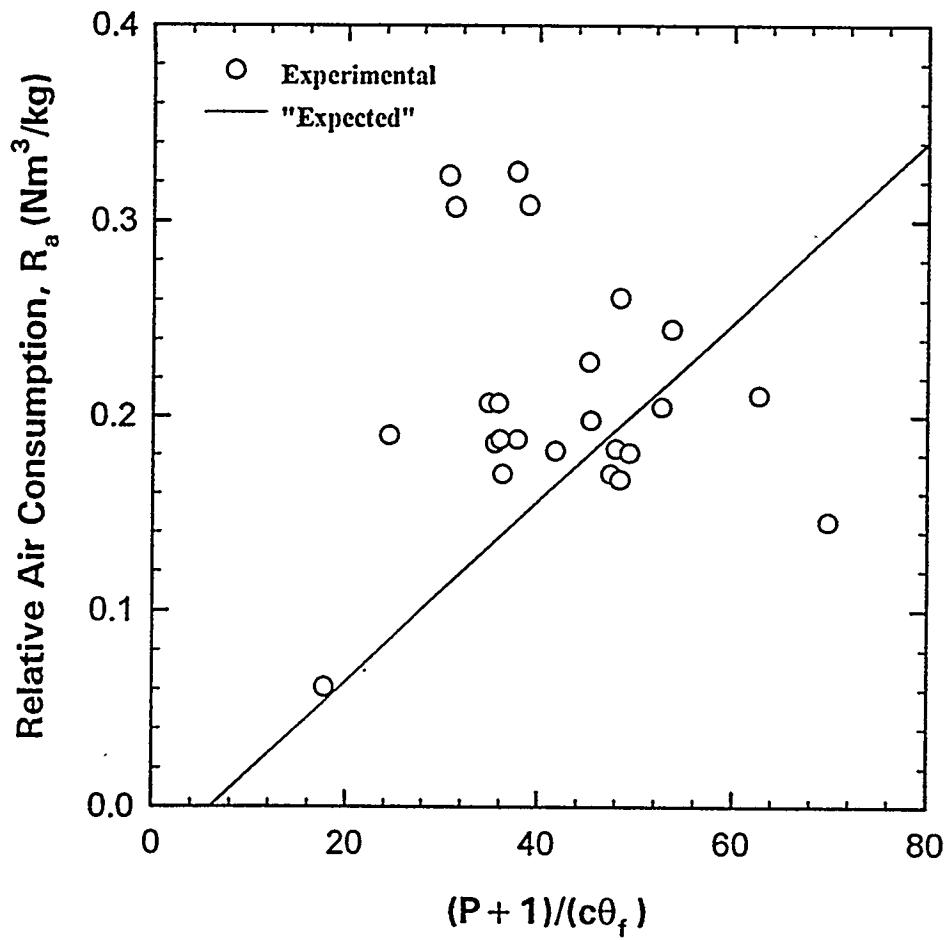


Figure 11-6. Relative air consumption in hyperbaric filtration of -100 mesh Pittsburgh seam coal.

Since the "expected" line shown in Figures 11-5 and 11-6 is based on the ideal case where the cake is equally permeable to air or water, lower air consumption could easily result from reduced permeability due to residual moisture. Partial collapse of the pore structure on dewatering could also be a contributing factor. The most likely explanation for the "high" values probably lies in non-uniformity of the cake and particularly in the development of microcracks in the dewatering stage.

In hyperbaric filtration, the cake is subjected to substantial stresses especially during the initial phase of the dewatering stage as bulk liquid is being expelled from the pores. Any slight flexing of the cake can lead to the development of shear stresses or even tensile stresses which could exceed the cohesive strength of the cake. While such stresses also exist during cake formation, cracks formed during this stage would tend to be self-healing, thereby not resulting in increased solids throughput. It is well known that macroscopic cracking is common in the dewatering of fine solids. However, it is not necessary for cracks to be large to impact on air consumption. A simple calculation shows that a single crack 5 μm wide across unit area of a uniform cake composed of 1 μm particles would contribute as much to air flow as the entire bed. A crack 8 μm wide would triple the air flow but still be essentially invisible. Further evaluation of the data to ascertain the specific conditions associated with abnormal air consumption is in progress.

Phase II - Laboratory Studies:

During this quarter, studies were conducted in developing modified techniques to improve dewatering of fine coal.

Effect of Medium Support:

Perforated plates are conventionally used as filter medium support. The open area of the perforated plate is much smaller than that of the filter medium. The effective filtration area is also much smaller than the medium open area when a perforated plate is employed. In this study, a modification was made by inserting a piece of 30 mesh opening screen between the perforated plate and filtration medium, as shown in Figure 11-7. Springs were placed between the sieving screen and the perforated plate to separate them in order to make the effective filtration area as large as possible. The filtration results obtained using the conventional medium support and the modified support are compared in Table 2. The data shows that the modified medium support produced a moisture reduction ranging from 2 to 5.2 percentage point, depending on the particle size. It can be seen that for the Pocahontas No. 3 coal, the cake moisture of the plus 400 mesh size fraction is reduced by 5.6 percentage points (from 7.2 to 1.6 percent), while the cake moisture of the minus 400 mesh size fraction is reduced by 3 percentage points (from 23.9 to 20.9 percent). The effect of medium support on the cake moisture for coarser size fraction is more significant than for the finer size fraction.

The enhancement obtained by using the modified support could be explained in terms of the capillary model. According to the capillary filtration model, a filter cake is considered as a bundle of tortuous capillaries with various diameters and the outlets of

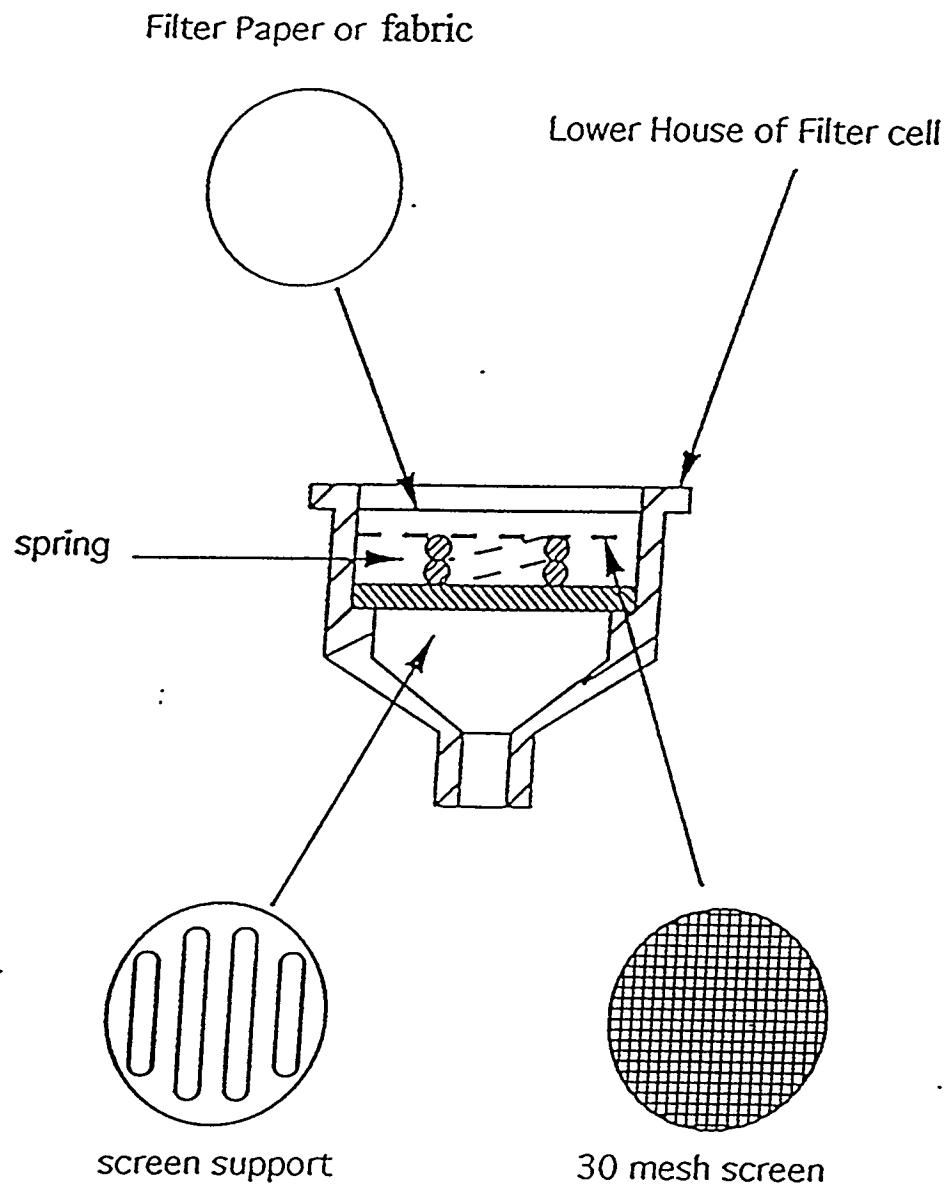


Figure 11-7. Schematic diagram of modified medium support.

Table 2. Effect of Medium support on Cake Moisture of Various Samples

Sample Name	Particle Size (mesh)	Filtration Condition	Cake Moisture	
			Conventional Support	Modified Support
Pocahontas No. 3	28x400	P=80psi, Time=2 min. Thickness=1.4 cm	7.2	1.6
	-28	P=80psi, Time=2 min. Thickness=1.4 cm	11.2	9.2
	-400	P=80psi, Time=2 min. Thickness= 1.4 cm	23.9	20.9

the capillaries are located at the bottom of the cake. If the filtration medium is placed directly on a perforated plate, some of the capillary outlets may be blocked by the imperforate area of the perforated plate support, which will obstruct the flow of water from capillaries. The water in the capillaries obstructed must travel horizontally to the nearest hole. The flow of water in horizontal direction is minimum due to very low pressure drop. When the filtration medium is put on a screen instead of a perforated plate, the capillary water is not blocked due to the larger opening area.

Effect of Filter Medium:

In this study, the effects of two kinds of filtration media, paper and fabric, have been investigated. The experimental results showed that for a given filtration condition, the fabric medium provided lower cake moisture as compared to the paper medium. During the experiments, it was observed that whenever a filter paper was used as a medium, a thin layer of ultra-fine particles formed at the bottom of the cake. This thin layer, combining with the filter paper, formed a high resistant layer to the flowing of the

filtrate and air. When the fabric was employed as filtration medium, there was no fine particle layer formed at the bottom of the cake because the ultra fines passed through the filter cloth.

Effect of The Combination of Medium and Support:

A comparison of the effect of the combination of filter medium (paper and fabric) with medium support (conventional and modified) for the Pittsburgh No. 8 coal is shown in Figure 11-8. Among the four combinations, the fabric/the modified medium support produced the lowest cake moisture and dramatically reduced the filtration time. The highest cake moisture is obtained from the filter paper with the conventional support. The average filtration rate in the first 30 seconds for the various combination of medium and medium support is shown in Table 3. Note, that compared to the conventional support, the modified support increased the filtration rate by 25 percent using the paper and 11 percent using a fabric filter medium.

Split Size Filtration:

As mentioned earlier, particle size is the most important factor in a filtration process. Unfortunately, in a preparation plant, making given particle size coarser is not possible. During this study, a new approach "split size filtration" was investigated in which the feed was separated into two different size fractions and each of the fractions was dewatered separately and final moisture was calculated by combining the moisture of both size fractions. Figures 11-9 and 11-10 show the results of split size filtration at different split particle sizes for the Pocahontas No. 3 and Pittsburgh No. 8 coals,

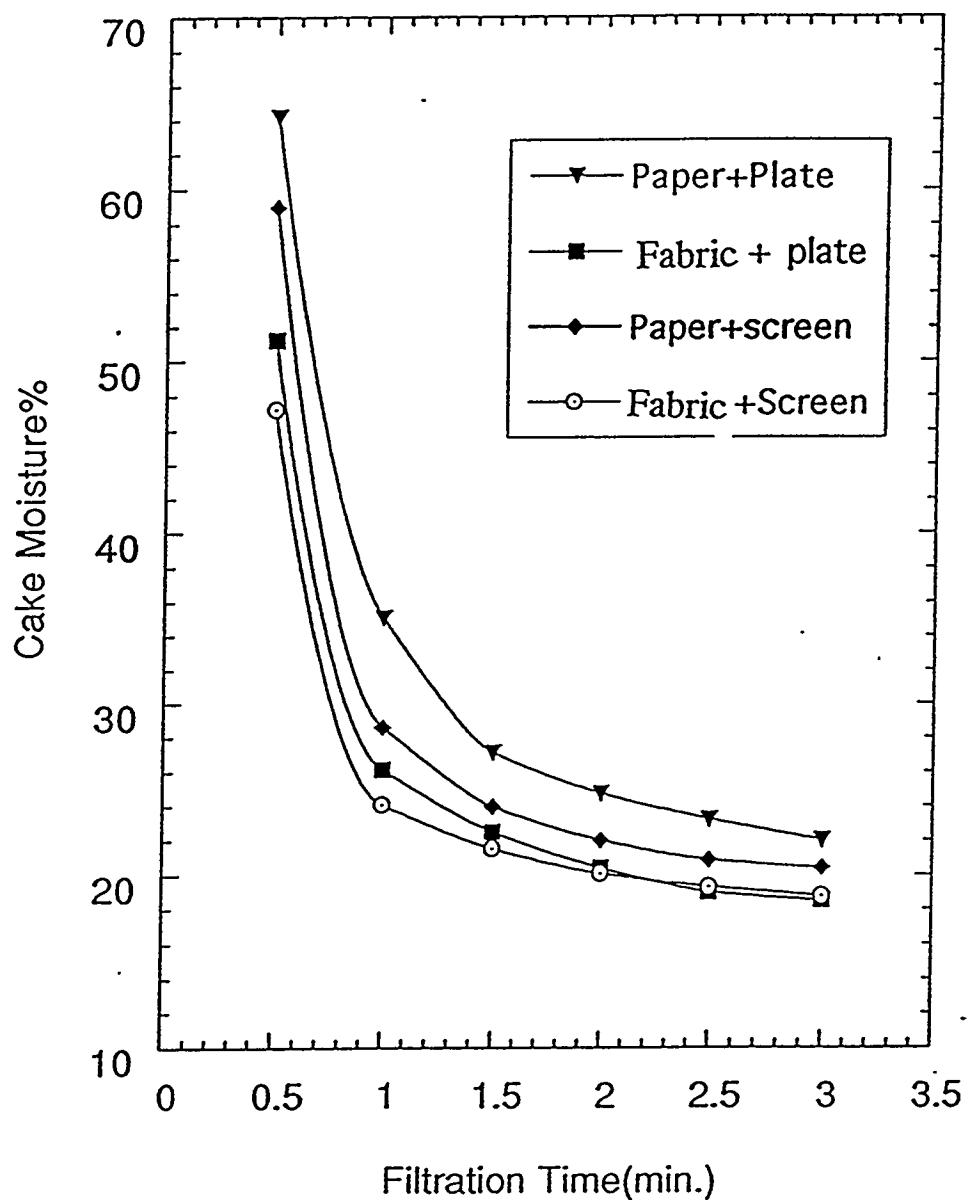


Figure 11-8. Effect of Filtration Medium and Medium Support on Filtration for the Pittsburgh No. 8 Coal.

Table 3. Filtration Rate In First 30 Seconds

Filtration Medium + Support	Filtration Rate (g/s)	Relative Filtration Rate
Paper + Conventional Support	1.67	1.00
Paper + Modified Support	2.09	1.25
Fabric + Conventional Support	2.21	1.32
Fabric + Modified Support	2.46	1.47

respectively. For the Pocahontas No. 3 coal slurry, split size filtration at the split size of 200 and 100 mesh provided a 7.5 percent moisture product, a 3.5 percentage points absolute or 31 percent relative moisture reduction over the baseline product moisture (11 percent). For the Pittsburgh No. 8 coal, split size filtration at the split size of 500 mesh yielded a 17 percent moisture product, a 7 percentage point absolute or about 33 percent relative moisture reduction. Even at 200 mesh split size dewatering provided a 20.5 percent product moisture which is about a 15 percent improvement in moisture reduction.

The principle of the moisture reduction obtained using the split size filtration could be explained in terms of the particle packing arrangement in a filter cake. The study of systematic packing of uniform spherical particles has shown that there are six different packing arrangements. Two of them (tight and loose packing) are shown in Figure 11-11. In these packing arrangement, each sphere is surrounded by a complex void space, as schematically described in Figure 11-12. It can be imagined that in a

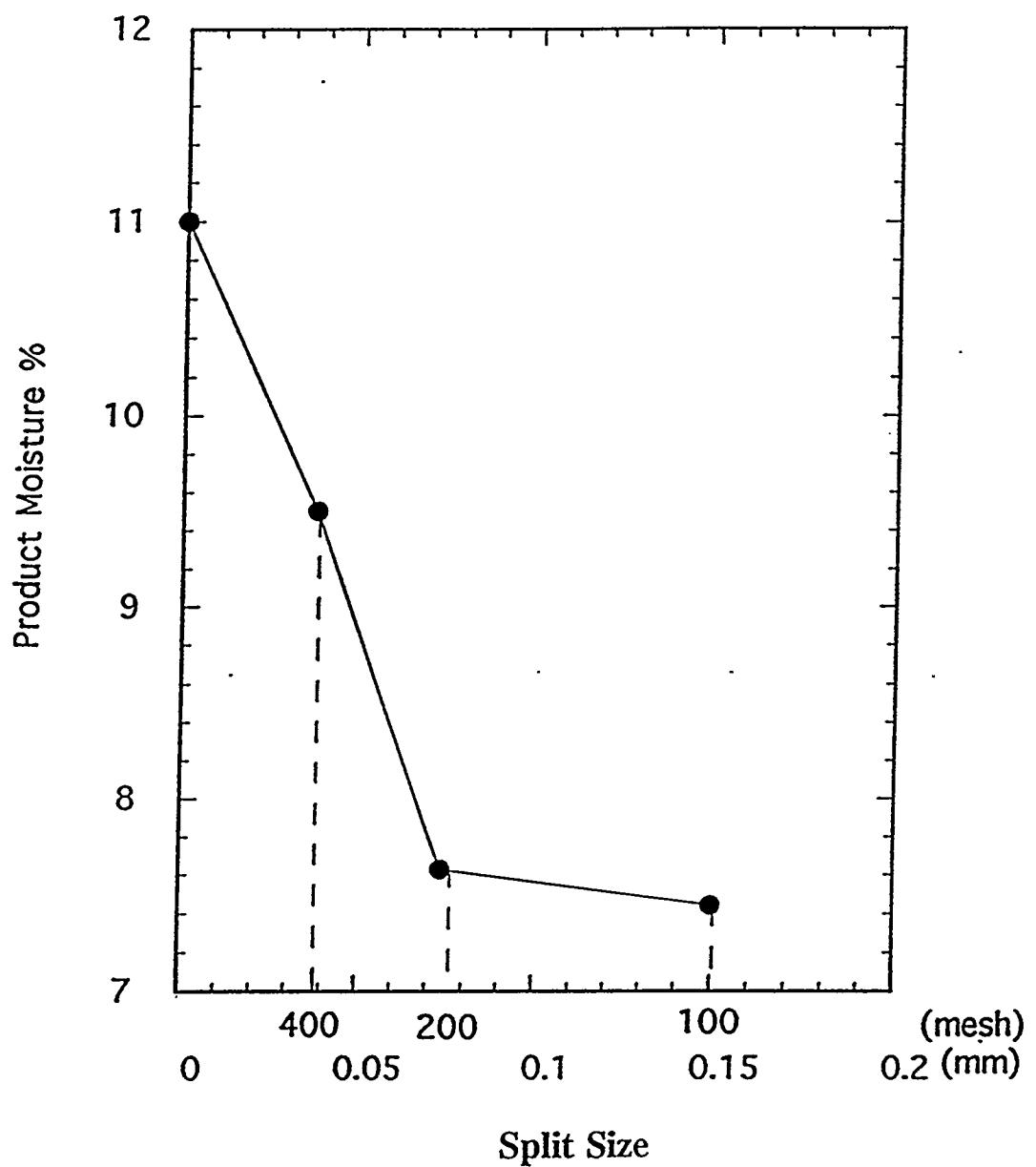


Figure 11-9. Result of Split Size Filtration for the Pocahontas No. 3 Coal.

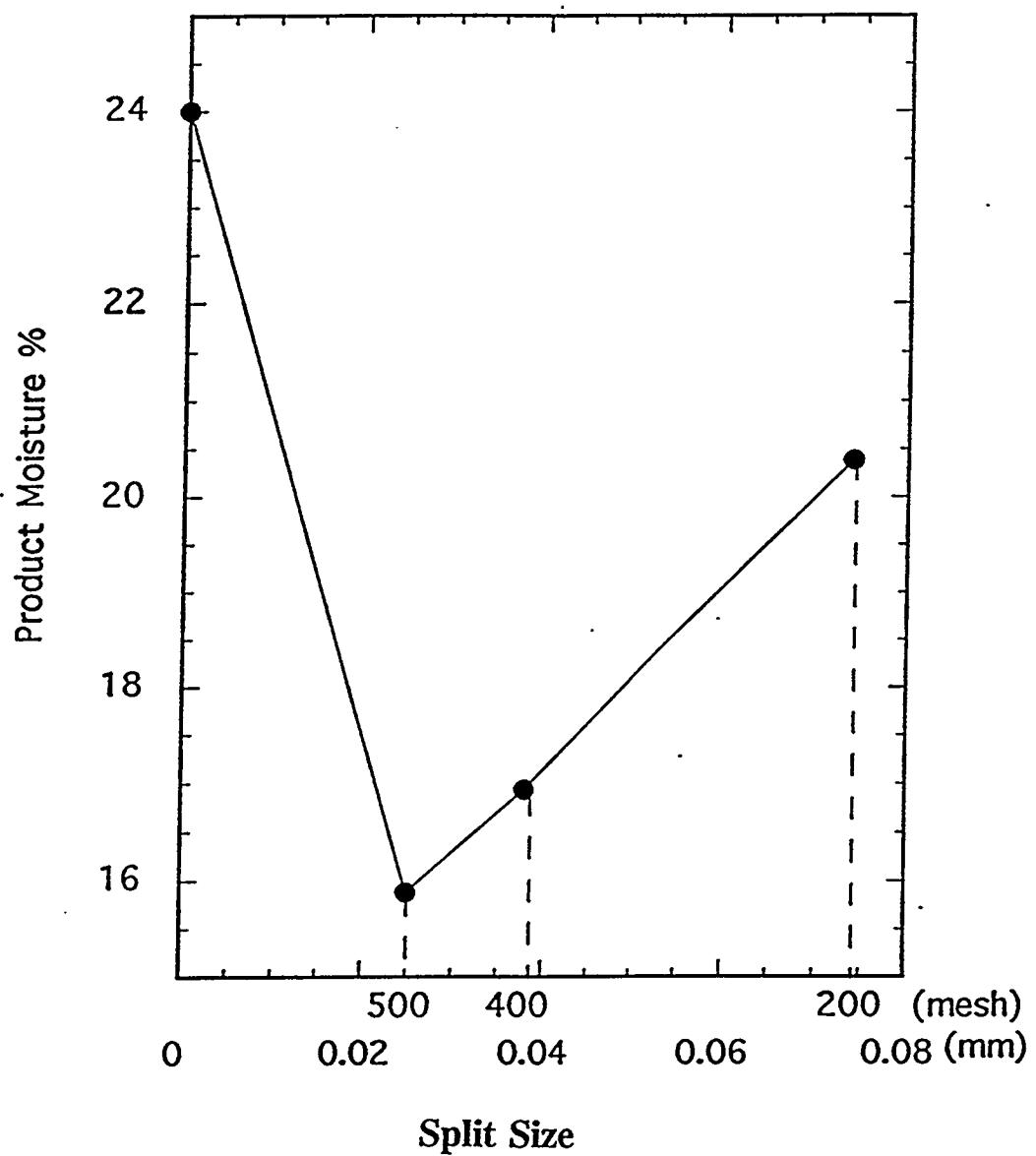
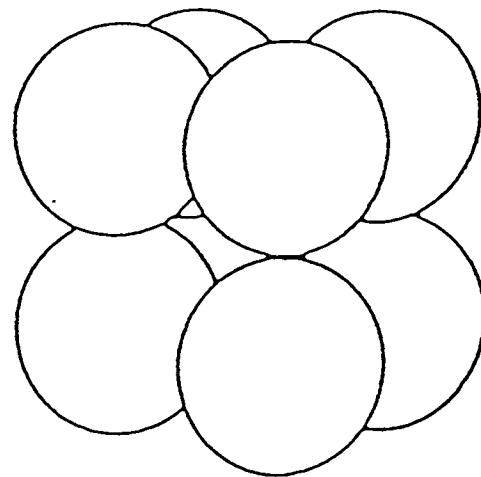
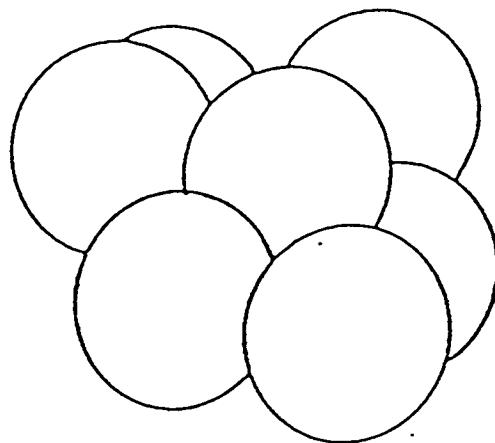


Figure 11-10. Result of Split Size Filtration for the Pittsburgh No. 8 Coal.

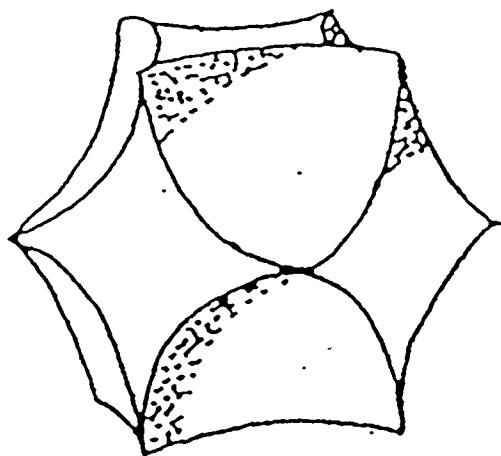


Cubic

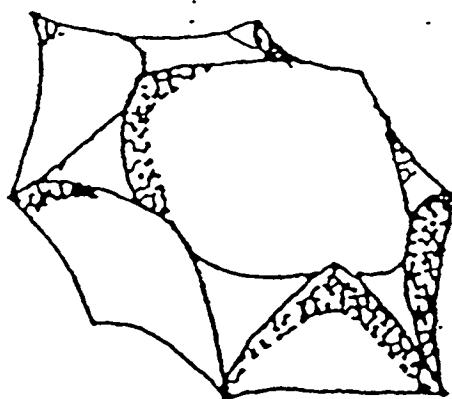


Rhombohedral

Figure 11-11. Packing Arrangement for Uniform Spheres.



Cubic



Rhombohedral

Figure 11-12. Pore Spaces in Packings of Uniform Spheres.

granular particle packing bed, each particle is also surrounded by a group of voids which may form the draining channels of water. In this way, a filter cake can be regarded as a porous matrix formed by large particles, in which the voids may be filled by finer particles, water or air. When a cake is formed only by coarse particles, the pores in the cake only are occupied by water and no pores in the cake will be blocked. For a multi-size particle cake, the finer particles and water simultaneously occupy the void spaces formed by coarser particles. The smaller particles in the void spaces will block either completely or partially the void spaces to prevent water from draining out of the cake. The number and diameters of the capillaries in a multi-size particle cake are definitely smaller than in a uniform coarser particle cake. The cake permeability mainly depends on particle packing arrangement for a coarser particle cake, while the permeability of a multi-size particle cake is mainly determined by the distribution and the amount of finer particles. The moisture of a coarser particle cake is much lower, as compared to a multi-size particle cake. However, there is not much difference in the moisture between multi-size particle cake and finer particle cake because in both cases, cake permeability is principally influenced by finer particles.

Combined Use of Various Dewatering Enhancement Methods:

As described earlier, the addition of flocculants, the use of the modified medium support and split size filtration improved the reduction of the cake moisture for both the Pittsburgh No. 8 coal and Pocahontas No. 3 coal. Tables 4 and 5 show the data obtained utilizing the combination of nonionic flocculant (4-6 million molecular weight) addition, fabric filter, modified medium support and split size filtration for the Pittsburgh No. 8

Table 4. Result of Combined Enhancement Methods for the Pittsburgh No. 8 Coal (Filter Cloth; Modified Support; Split Size Filtration; 60 g/t Non-ionic flocculant for the Finer Size Fractions)

Split Size (mesh)	Size Fraction (mesh)	% Wt.	Cake Moisture
200	100×200	21.8	1.3
	200×0	78.2	17.9
	Composite	100	14.28
400	100×400	46.6	1.6
	400×0	53.4	18.96
	Composite	100	10.87
500	100×500	57.7	2.6
	500×0	42.3	21.2
	Composite	100	10.47

(The baseline cake moisture obtained without using any enhancement approach was 24%)

Table 5. Result of Combined Enhancement Methods for the Pocahontas No.3 Coal (Filter Cloth; Modified Support; Split Size Filtration; 80 g/t Non-ionic Flocculant for the Finer Size Fractions)

Split Size (mesh)	Size Fraction (mesh)	% Wt.	Cake Moisture
100	28×100	49.8	1.01
	100×0	50.2	13.2
	Composite	100	7.13
200	28×200	67.1	1.4
	200×0	32.9	15.59
	Composite	100	6.06
400	28×400	78.0	1.6
	200×0	22.0	16.4
	Composite	100	4.85

(The baseline cake moisture obtained without using any enhancement approach was 11%)

coal and Pocahontas No. 3 coal slurry, respectively. Note that for the Pittsburgh No. 8 coal, the combined enhancement approach produced filter cakes of 10.5 to 14.3 percent moisture, a 13.5 to 9 percentage point moisture reduction over the baseline data of 24 percent moisture. This represents a 56 to 40 percent relative moisture reduction. Similarly, for the Pocahontas No. 3 coal slurry, the combined enhancement approach provided filter cake with 4.8 to 7.1 percent moisture, a 6.15 to 3.8 percentage point of total absolute moisture reduction over the baseline of 11 percent moisture. This represents 56 to 35 percent of relative moisture reduction.

Summary of the Dewatering Enhancement Methods:

The improvements in the moisture reduction obtained for the Pocahontas No. 3 and the Pittsburgh No. 8 coals are summarized in Figures 11-7 and 11-8, respectively. Note that for the Pocahontas No. 3 coal slurry (Figure 11-7), vacuum filtration provides a 24 percent moisture product, which is close to that obtained in the preparation plants processing the Pocahontas No. 3 coal. Hyperbaric filter using 80 psi pressure reduces the moisture to 11 percent, a more than 50 percent relative moisture reduction. The use of the combined dewatering enhancement approaches reduces the moisture to as low as 4.85 percent, which is a 56 percent relative moisture reduction over the baseline moisture obtained using the hyperbaric filter and a 80 percent relative moisture reduction over the filter cake moisture obtained using the vacuum filter. For the Pittsburgh No. 8 coal slurry (Figure 11-8), the vacuum filter produces a 73 percent moisture product due to the very fine size particles, while the hyperbaric filter using 70 psi (4.8 bars) pressure reduces the moisture to 24 percent. The combined use of various types of enhancement

approaches reduces the moisture to 10.5 percent. The final moisture reduction achieved is 56 percent over the baseline moisture obtained using the hyperbaric filtration alone.

Phase III - Pilot Scale Testing:

The pilot scale testing was conducted using the Andritz Ruthner Inc. hyperbaric filter which consisted of a single disc of 1.4m diameter or 2m² (22 ft²) filtration area. the pilot scale tests were conducted at Bailey mine (Pittsburgh No. 8) and the Buchanan mine (Pocahontas No. 3). In the last quarterly report, baseline dewatering data for both the coal slurries describing moisture, consumption and solid feed rate were presented.

The effect of a flocculant and a cationic surfactant dosage on dewatering of the Pittsburgh No. 8 flotation concentrate is given in Tables 6 and 7. Note, that with increase in flocculant dosage (Table 6), both the filter cake moisture and solids throughput increases, however, the air consumption at 3.5 bar (50 psi) pressure shows no change, whereas at 5.0 bar (74 psi) pressure it increases from 87 to 115 scfm/tph. In contrast, increase in addition of a cationic surfactant (Table 7) shows that at 3.5 bar (50 psi) pressure the filter cake moisture decreases and solids throughput increases, however, air consumption does not vary significantly. Using a higher 5.0 bar (74 psi) pressure, the filter cake moisture, solids throughput, and air consumption, shows an increase with increase in surfactant dosage.

Effect of a flocculant and a cationic-anionic surfactant dosages for the Pocahontas No. 3 coal slurry are listed in Tables 8 and 9, respectively. Note, that with increasing flocculant dosage (Table 8) at 3.5 bar or 5.0 bar pressure, the filter cake moisture and solids throughput shows an increase. Addition of various dosage of cationic surfactant

Table 6. Effect of Flocculant Dosage on Filter Cake Moisture, Solids Throughput and Air Consumption (minus 100 mesh froth product).

Pressure (bar)	Flocculant Dosage (g/t)	Moisture (wt. %)	Throughput (lb/hr-ft ²)	Air Consumption (scfm/tph)
3.5	0	24.1	135	105
	13	24.6	187	102
	27	27.2	215	102
5.0	0	23.7	187	87
	13	24.2	208	98
	33	25.3	225	115

Table 7. Effect of Cationic Surfactant on Filter Cake Moisture, Throughput and Air Consumption (minus 100 mesh froth product).

Pressure (bar)	Surfactant Dosage (g/t)	Moisture (wt. %)	Throughput (lb/hr-ft ²)	Air Consumption (scfm/tph)
3.5	0	22.8	132	108
	50	22.6	139	111
	100	21.4	162	102
5.0	0	23.7	192	87
	100	23.8	212	90
	200	24.1	218	97
	450	24.3	221	112

Table 8. Effect of Flocculant Dosage on Filter Cake Moisture, Throughput and Air Consumption (minus 28 mesh).

Pressure (bar)	Flocculant Dosage (g/t)	Moisture (wt. %)	Throughput (lb/hr-ft ²)	Air Consumption (scfm/tph)
3.5	0	15.0	130	55
	9	15.4	152	62
	17	16.0	172	67
5	0	13.4	113	63
	4	14.2	161	52
	9	14.7	180	50

Table 9. Effect of Surfactant Dosage on Filter Cake Moisture, Throughput and Air Consumption (minus 28 mesh, pressure 5 bar).

Surfactant Type	Surfactant Dosage (g/t)	Moisture (wt. %)	Throughput (lb/hr-ft ²)	Air Consumption (scfm/tph)
Cationic	0	13.0	89	84
	80	12.8	88	55
	380	12.4	85	65
Anionic	0	13.4	113	63
	70	13.4	113	68
	300	13.3	99	79

(Table 9) shows a slight lowering of filter cake moisture, solids throughput and air consumption. On the other hand, addition of anionic did not show any effect on the filter cake moisture.

Conclusions

Pilot scale testing of the Andritz hyperbaric filter of the Pittsburgh No. 8 flotation product (minus 100 mesh) provided a 23 percent moisture in the filter cake using about 3 bar (44 psi) pressure. Air consumption for this condition was about 160 scfm/tph and solids throughput was about 110 lb/hr-ft². Addition of flocculant increase moisture content of filter cake, however, addition of a cationic surfactant (100 g/t) lowered filter cake moisture from 22.8 percent to 21.4 percent using 3.5 bar (51 psi) pressure.

For the Pocahontas No. 3 froth product which was 28x0 mesh size, a 15 percent moisture product could be obtained using about 2 bar (30 psi) pressure. Under these conditions, the air consumption was 50 scfm/tph and solids throughput was about 100 lb/hr-ft². Addition of flocculant did not lower the filter cake moisture, however, addition of cationic surfactant (380 g/t) lowered filter cake moisture from 13 percent to 12.4 percent.

FUTURE PLANS

Prepare the final report. Obtain additional data to help improve the Penn State dewatering model.