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EVALUATION OF HYPERBARIC FILTRATION FOR FINE COAL DEWATERING

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Tenth Quarterly Technical Progress Report
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by

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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 model for cake formation in batch filtration, presented in the December 1994 quarterly progress report, can be applied directly to continuous filter systems. It is useful, however, to make some minor modification in order to present the relationships in terms of directly measurable quantities. The basic expression for batch filtration can be written:

$$\frac{t}{m_{wf}} = \frac{\mu_w \alpha c\, m_{wf}}{2A^2 \rho_w \Delta p} + \frac{\mu_w \alpha_m}{\rho_w A \Delta p} \tag{1}$$

where m_{wf} is the mass of liquid collected in the filtrate in time t; α and α_m are defined, in the usual way, as the specific cake resistance and filter mechanism resistance, respectively; μ_w and ρ_w are respectively the liquid viscosity and density, A is the filter area and Δp is the applied pressure. The concentration c is defined as the mass of solids in the cake per unit mass of liquid in the filtrate. For the general case where some liquid is retained in the cake and some solids pass into the filtrate, c can be related to the feed solids concentration through a simple mass balance which leads to:

$$c = \frac{(c_s - c_f)(1 - M)}{(1 - c_s - M)(1 - c_f)} \tag{2}$$

where c_s is the mass fraction of solids in the feed, c_f is the mass fraction in the filtrate and M is the moisture content (mass fraction) in the filter cake.

In the case of continuous filtration, one complete cycle (revolution) is mathematically equivalent to batch filtration for time

$$t_f = \frac{\theta_f}{2\pi N} \tag{3}$$

where θ_f is the cake formation angle (radians) and N is the rotational speed (revolutions/time). Since the solid and liquid flows occur over the entire cycle, however, the flow rates refer to the cycle time.

$$T = 1 / N \tag{4}$$

rather than to the apparent filtration time t. Thus the rate of liquid flow to the filtrate is

$$\frac{m_{wf}}{T} = m_{wf} N \tag{5}$$

The solids throughput can be expressed as

$$R_s = \frac{m_{sc}}{AT} = \frac{m_{sc} N}{A} \tag{6}$$

where R_s is the solids throughput per unit area (mass/area · time), m_{sc} is the mass of (dry) solids in the cake. From the definition of c,

$$m_{sc} = c\, m_{wf} \tag{7}$$

so that, from Equation 6,

$$\frac{m_{wf}}{A} = \frac{R_s}{cN} \tag{8}$$

Substitution from Equations 3 and 8 in Equation 1 leads to

$$\frac{\Delta p c \theta_f}{2\pi R_s} = \frac{\mu\alpha}{2\rho_w} \cdot \frac{R_s}{N} + \frac{\mu_w \alpha_m}{\rho_w} \tag{9}$$

Equation 9 provides a useful relationship between throughput R_s and the process variables Δp, c, θ_f and N.

The dewatering step can be roughly divided into two stages: displacement of bulk liquid from the (initially) saturated cake followed by further, slow removal of residual moisture during airflow through the cake. The fraction of the cycle which is taken up by the initial displacement can be estimated from the liquid flow rate at the end of the cake formation stage. Thus, the displacement time t_d can be estimated from

$$t_d = V_{wc} / Q_{wd} \tag{10}$$

where V_{wc} is the volume of liquid in the saturated cake and Q_{wd} is the flowrate.

$$V_{wc} = \varepsilon V_c \tag{11}$$

where V_c is the cake volume. A solids mass balance leads to

$$V_c \rho_s (1-\varepsilon) = c \rho_w V_{wf} \tag{12}$$

where V_{wf} is the volume of liquid passing into the filtrate during the cake formation stage. Then,

$$V_{wc} = \frac{\varepsilon\, c\, \rho_w V_{wf}}{(1-\varepsilon)\rho_s} \tag{13}$$

The filtrate volume V_{wf} can be estimated from a simplified form of Equation 1 in which the medium resistance α_m is neglected. Thus,

$$V_{wf} = \frac{m_{wf}}{\rho_w} = A\sqrt{\frac{2\Delta pt}{\mu_w \alpha c\, \rho_w}} \tag{14}$$

Assuming the liquid flowrate to be the same as at the end of the cake formation stage,

$$Q_{wd} = Q_{wf} = \frac{dV_{wf}}{dt}\bigg|_{t=t_f} \tag{15}$$

where t_f is the total cake formation time. Differentiating Equation 14,

$$Q_{wf} = A\sqrt{\frac{\Delta p}{2\mu_w \alpha c\, \rho_w t_f}} \tag{16}$$

Combining Equations 10, 13, 14 and 16.

$$t_d = \frac{2\varepsilon c \rho_w t_f}{(1-\varepsilon)\rho_s} \tag{17}$$

Expressed as a displacement angle, Equation 17 becomes

$$\theta_d = \frac{2\varepsilon c \rho_w \theta_f}{(1-\varepsilon)\rho_s} \tag{18}$$

Air consumption during the dewatering stage in the cycle is an important practical consideration. Since both cake formation and dewatering involve fluid flow through the porous cake, some correlation of throughput and air consumption is to be expected. Approximate relationships can be obtained by extension of the basic filtration model.

Expressing the permeability of the moist cake to air relative to that to water by a factor K, the air flow velocity can be written

$$u_a = K \frac{\mu_w}{\mu_a} u_w \tag{19}$$

where u_w is the velocity of water flowing through the same cake with the same applied pressure, and μ_w and μ_a are the respective viscosities. By definition,

$$u_w = \frac{Q_{wf}}{A} \tag{20}$$

so that, from Equations 3, 16 and 19

$$u_a = K \frac{\mu_w}{\mu_a} \sqrt{\frac{\pi \Delta p N}{\mu_w \alpha c \rho_w \theta_f}} \tag{21}$$

The air flowrate relative to the total filter area is given by

$$\frac{Q_a}{A} = \frac{u_a(\theta_a - \theta_d)}{2\pi} \tag{22}$$

where θ_a is the effective dewatering angle, i.e., that angle in the cycle where the cake is subjected to direct air pressure. It is convenient to express the air consumption relative to solids throughput using

$$R_a = \frac{Q_a}{R_s A} \tag{23}$$

Neglecting the medium resistance during cake formation, R_s can be obtained from Equation 9:

$$R_s = \sqrt{\frac{\Delta p c \rho_w N \theta_f}{\pi \mu_w \alpha}} \tag{24}$$

Combining Equations 21, 22, 23 and 24,

$$R_a = \frac{K \mu_w}{2c\theta_f \rho_w \mu_a}(\theta_a - \theta_d) \tag{25}$$

Finally, substitution from Equation 18 leads to

$$R_a = \frac{K \mu_w}{2\rho_w \mu_a}\left[\frac{\theta_a}{c\theta_f} - \frac{2\varepsilon \rho_w}{(1-\varepsilon)\rho_s}\right] \tag{26}$$

Equation 26 suggests a simple inverse relationship between the relative air consumption R_a and the cake formation angle θ_f. It should be noted, however, that the relative permeability K can be expected to vary with cake porosity and residual moisture content.

Equations 9 and 26 should be directly applicable to the evaluation of continuous filter performance in terms of solids throughput and air consumption relative to operating conditions and feed concentration. A preliminary evaluation of data given in the previous report (December 1994) has been conducted. As a first step in the data evaluation, an assessment of internal consistency and mass balance closure has been made.

A test of the balance of solids and liquid around the filter can be carried out by comparing the directly measured feed solids concentration with a calculated value based on throughput, cake moisture and solids loss to the filtrate. Some results of such calculations, for the Bailey Mine froth flotation product, are shown in Figure 1. The linear relationship indicates generally consistent results. However, the fact that the slope of the line is slightly greater than its expected value of one suggests some consistent bias in the measurements.

The relationship between solids throughput and measured cake thickness is also a measure of data consistency. According to Equation 6, the mass of cake per unit area should be:

$$\frac{m_{sc}}{A} = \frac{R_s}{N} \tag{27}$$

and, from simple geometry

$$m_{sc} = LA\rho_s(1-\varepsilon) \tag{28}$$

It follows that

$$L = \frac{R_s}{N\rho_s(1-\varepsilon)} \tag{29}$$

An example of results plotted according to Equation 29 is given in Figure 2. The essentially linear relationship indicates that the data are not only consistent but that the cake density is more or less constant. Based on the slope of the line and an assumed specific gravity of 1.4 for the coal, the cake porosity is about 47.4%.

An example of the correlation of data for solids throughput for a broad range of operating conditions is given in Figure 3. While there are a number of obvious outliers,

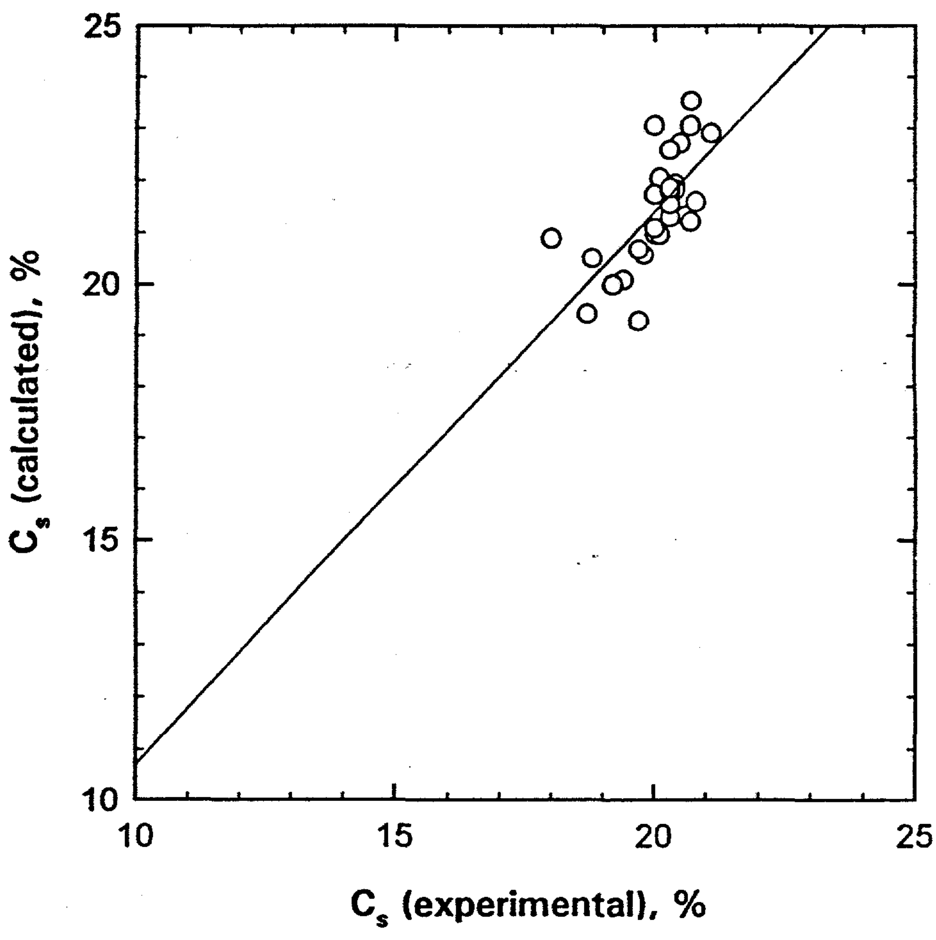

Figure 1. Comparison of measured and calculated feed solids content for Bailey Mine froth product.

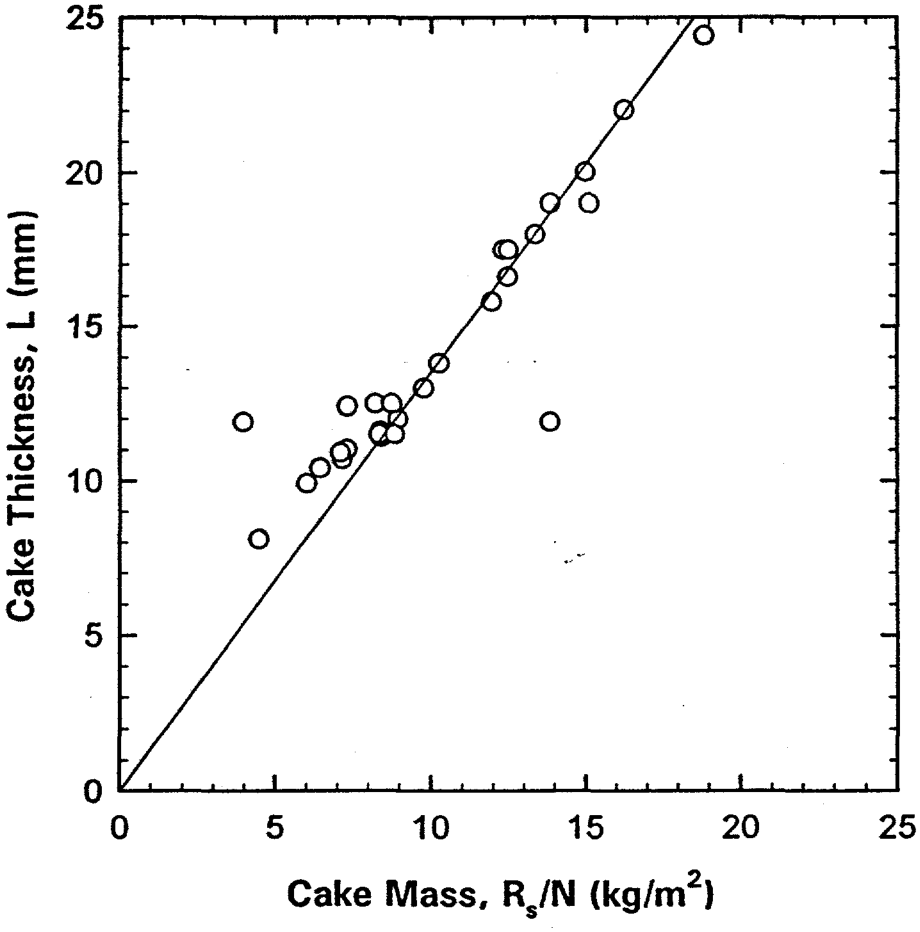

Figure 2. Relationship between measured cake thickness and solids throughput.

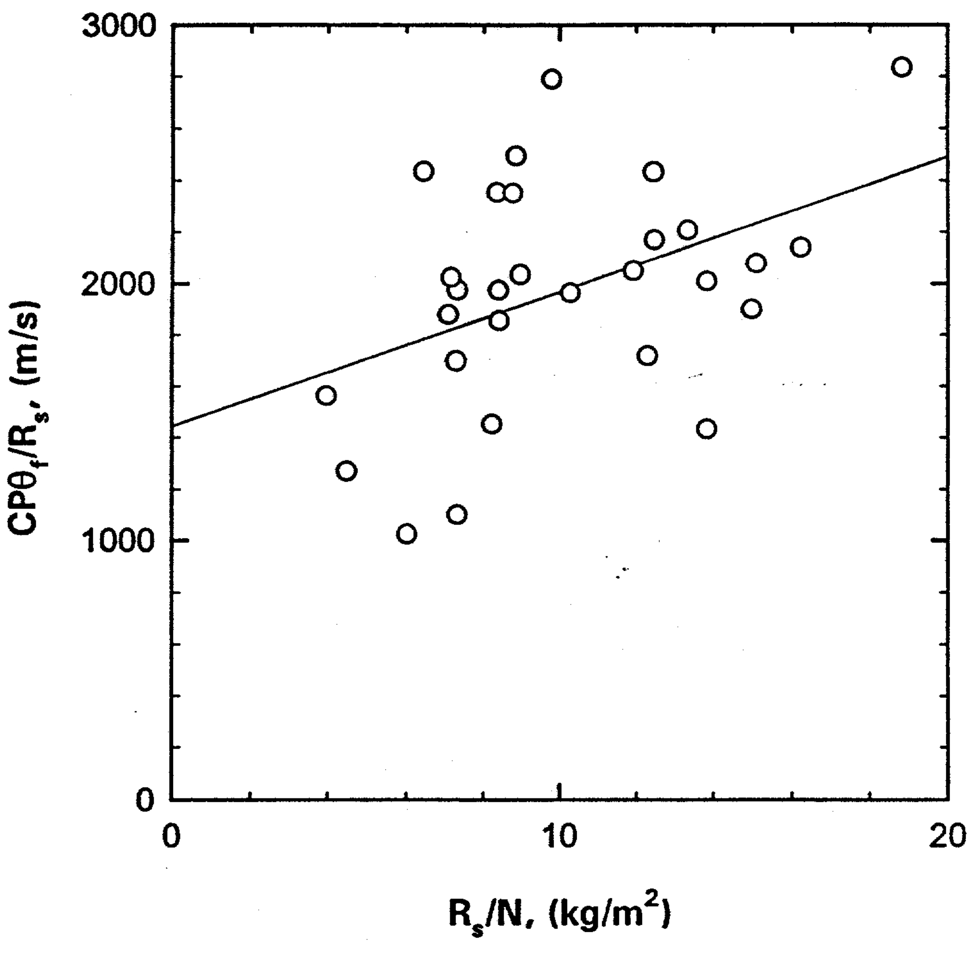

Figure 3. Data correlation for cake formation.

the results appear to be generally consistent with Equation 9. Further examination of this correlation and of specific conditions which can be associated with departures from the prediction is in progress.

A plot of relative air consumption, based on the relationship predicted by Equation 26, is given in Figure 4. While there is considerable scatter in the data, which may be due in part to variations in the permeability factor K with moisture content, the correlation with Equation 26 is reasonable. The results certainly confirm the predicted reduction in air consumption with increasing cake formation angle.

As pointed out in previous reports, the dewatering characteristics of filter cakes are determined by details of the cake structure; especially the pore size distribution. While the pore size distribution is obviously related to the particle size distribution and the overall cake density, reliable, quantitative models remain to be developed. The test results do, however, provide some insight into dewatering behavior.

Figure 5 is a plot of residual moisture content against air-flow through the cake. The relatively poor correlation indicates that passing additional air through the cake does not lead to significantly enhanced dewatering. A relationship between cake dewatering and applied pressure gradient was demonstrated both theoretically and from laboratory (batch) tests in the December 1994 report. A similar correlation for continuous filtration is given in Figure 6. In each case increasing the gradient appears to reduce the residual moisture content. The effect, however, is generally small compared to that due to differences in particle (and pore) size. The potential advantages of maintaining a high gradient by restricting cake thickness are clear. Equations 9, 26 and 29 provide a basis

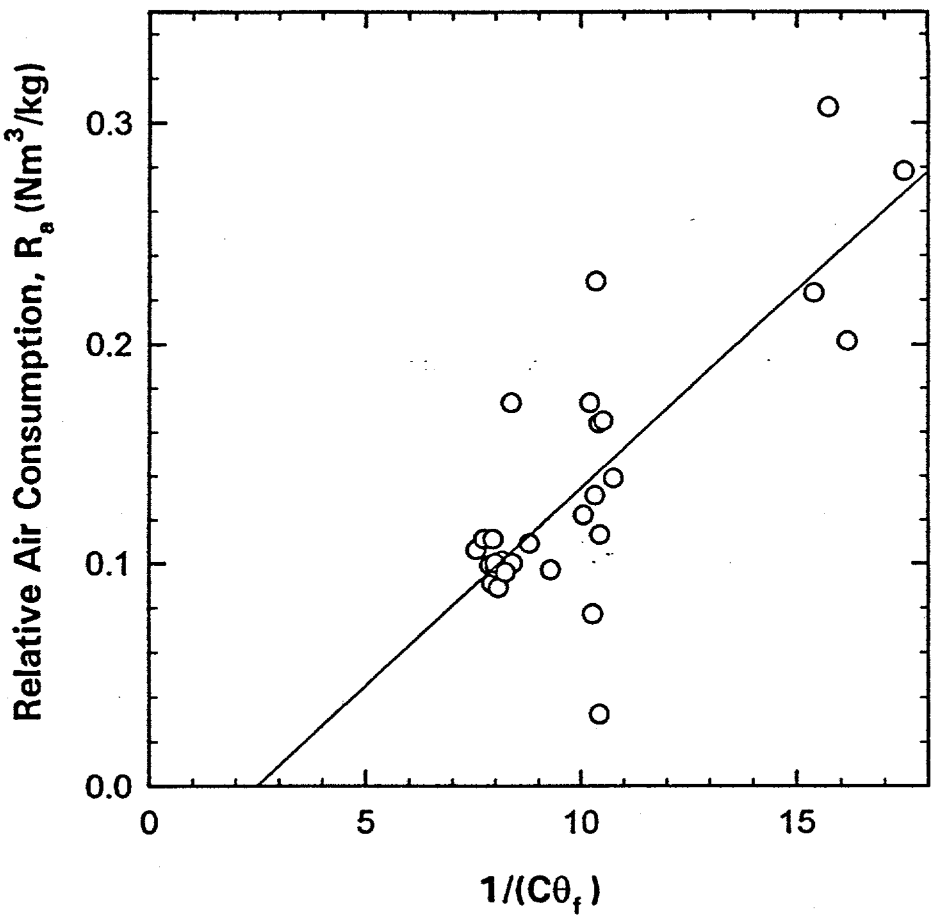

Figure 4. Data correlation for relative air consumption.

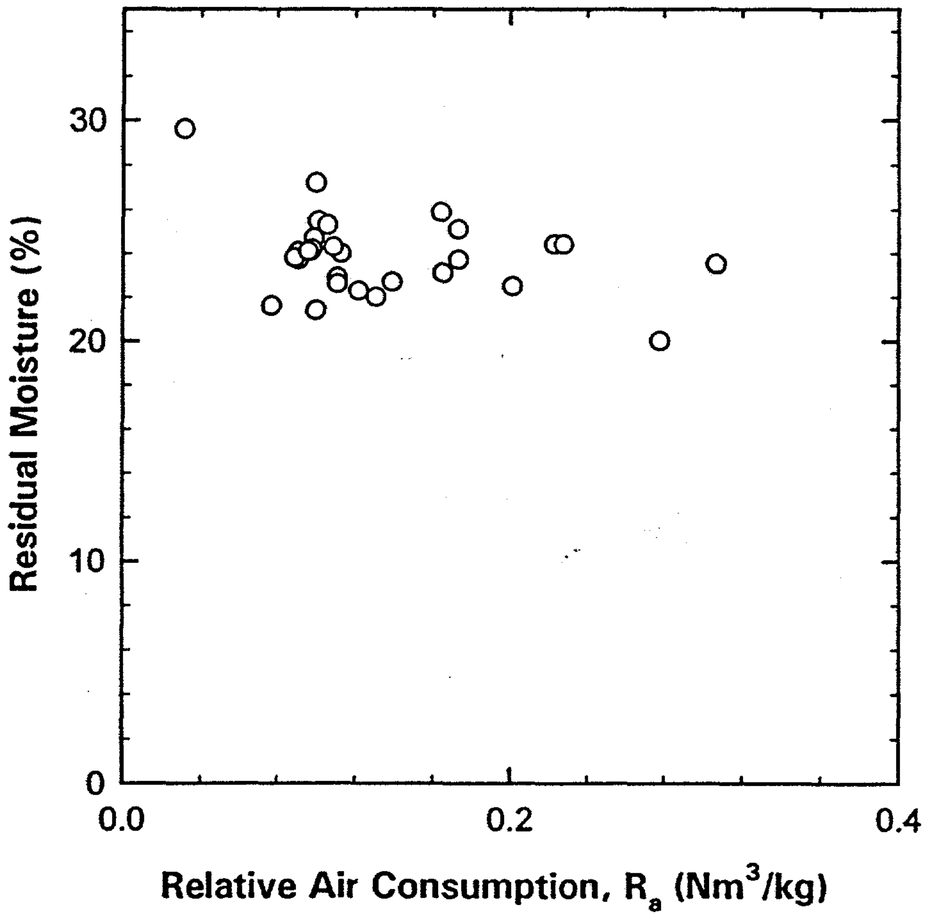

Figure 5. Relationship between air consumption and residual moisture content.

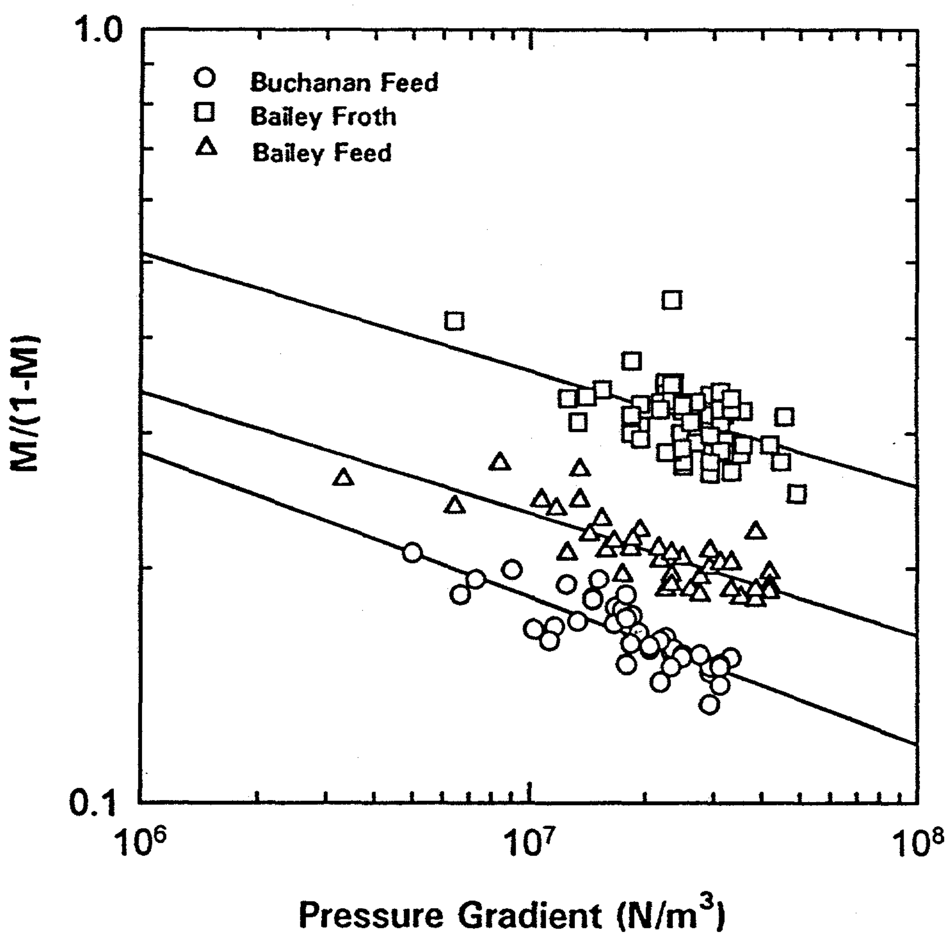

Figure 6. Correlation of residual cake moisture with pressure gradient in continuous hyperbaric filtration.

for the selection or operating conditions so as to maximize throughput while minimizing cake thickness and air consumption.

Phase II - Laboratory Studies

In this quarterly report, additional dewatering data for the Pocahontas No. 3 clean coal flotation concentrate obtained with addition of reagents are described.

Effect of Additives

Organic flocculants and surfactants are generally used as additives to enhance filtration process. These reagents enhance the filtration process through the modification of the colloidal chemistry properties of the fine coal slurry such as increasing particle sizes, reducing surface tension and surface charge.

Flocculants

Polymeric flocculants are used in many operations for increasing the particle sizes. The principal effect of flocculants is to form "bridges" from one particle to the next (Ruethrwein and Wad, 1952). This mechanism requires that the flocculant chains be adsorbed from solution onto one particle, and a physical bridge will form between the particles when another particle comes close enough for the extended flocculant chains to be adsorbed onto it. The chain length of flocculants which is directly proportional to the flocculant molecular weight, is an important factor in a flocculation process. Another important factor is the ionic characteristics of the flocculants which controls the adsorption behavior of the flocculants on particles. The addition of flocculants increases particle sizes as well as the filtrate viscosity. The increase of filtrate viscosity is detrimental to the reduction of cake moisture.

Figure 7 shows the effect of the dosage of various types of flocculants on the cake moisture of the Pocahontas No. 3 coal. For all three kinds of flocculants (anionic, cationic and non-ionic), there was an optimum dosage at which the cake moisture was the lowest. Among the three flocculants investigated, non-ionic flocculant provided the largest decrease in cake moisture. The optimum dosage of non-ionic flocculant was 80 g/t. The anionic flocculant was least effective and when the dosage was increased over 100 g/t, the cake moisture increased dramatically with increasing the dosage.

For comparison purpose, the effect of flocculants on cake moisture for the Pittsburgh No. 8 coal is shown in Figure 8. Generally, the effect of flocculants was more noticeable for the Pittsburgh No. 8 coal than for the Pocahontas No. 3 coal. The addition of non-ionic flocculant reduced the cake moisture from 24 to 17.5 percent at an optimum dosage of 60 g/t.

The relationship between the solution viscosity and flocculant concentration is shown in Figure 9. It can be seen that the solution viscosity increased with flocculant concentration. Thus, in the filtration process, high flocculant dosage lead to the high viscosity of filtrate, which will result in the retention of high moisture on the coal surface. It can also be observed that at the same flocculant concentration, the viscosity of anionic-flocculant solution was higher than those of non-ionic and cationic flocculant solutions. Also, the increase of anionic flocculant concentration caused the solution viscosity to increase very sharply. Therefore, the effect of anionic flocculant on cake moisture was more sensitive to the flocculant dosage in a filtration process.

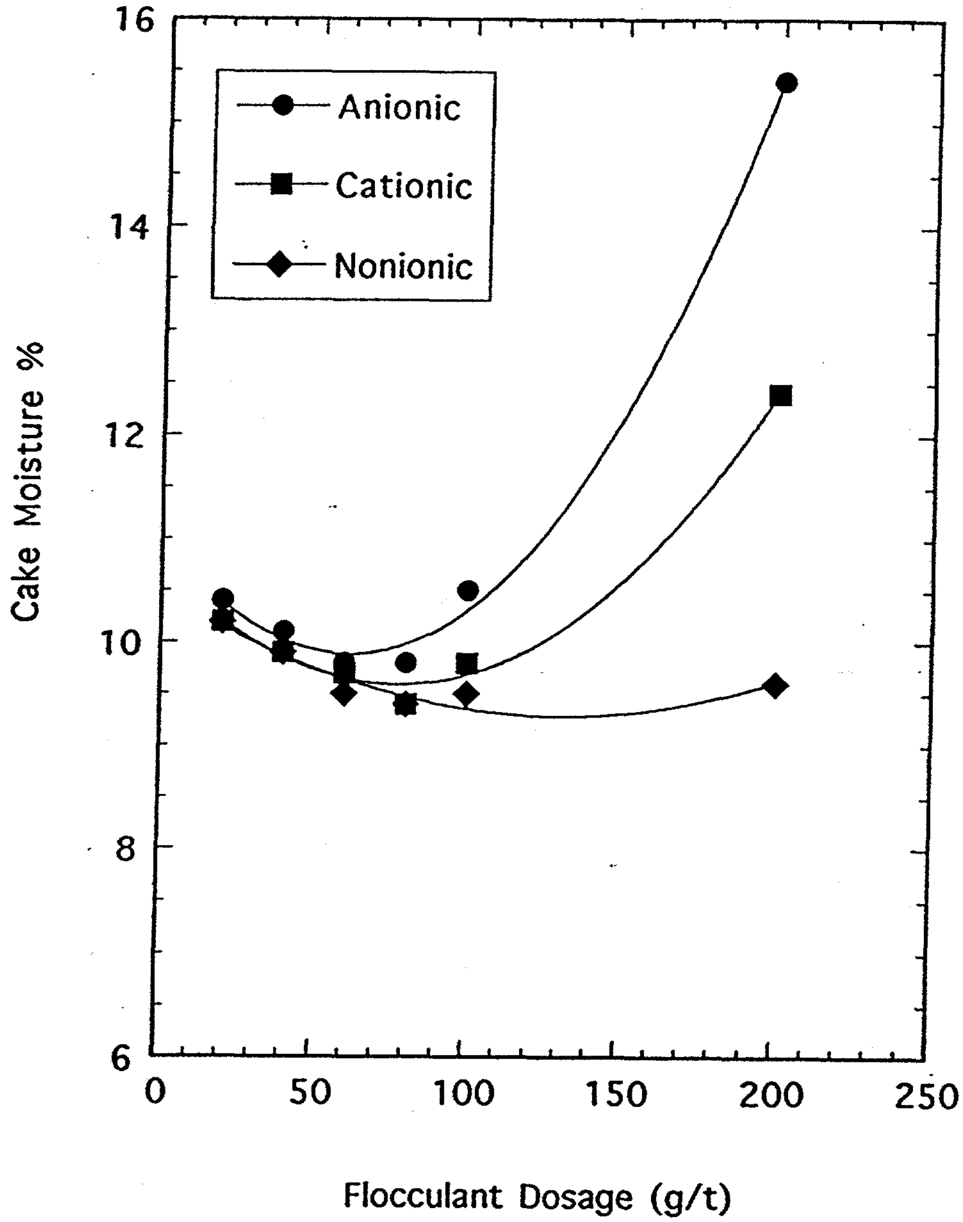

Figure 7. Effect of Flocculant Dosage on Cake Moisture for the Pocahontas No. 3 Coal. (flocculation time: 3 minutes; flocculation mixing rate: 350 rpm; flocculant MW: 4-6 million)

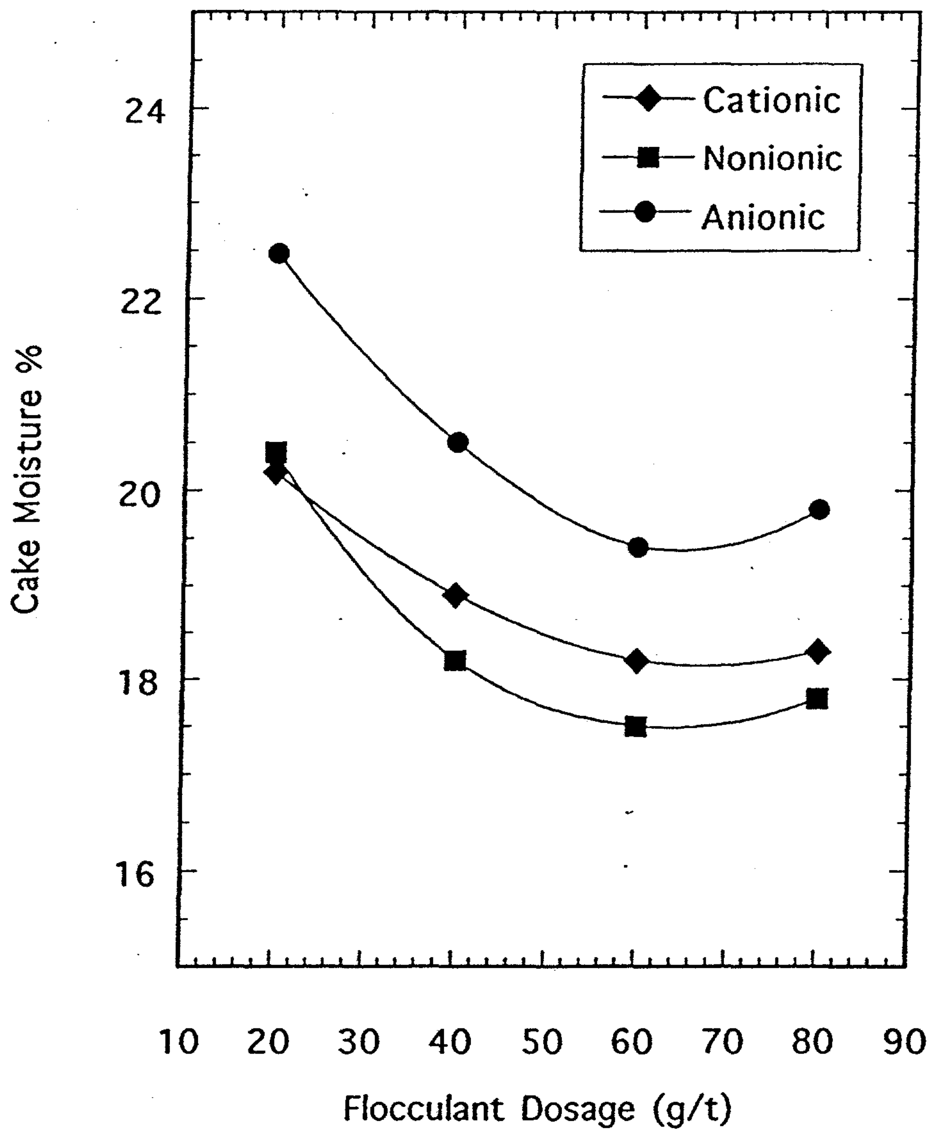

Figure 8. Effect of Flocculant Dosage on Cake Moisture for the Pittsburgh No. 8 Coal. (flocculation time: 3 minutes; flocculation mixing rate: 350 rpm; flocculant MW: 4-6 million)

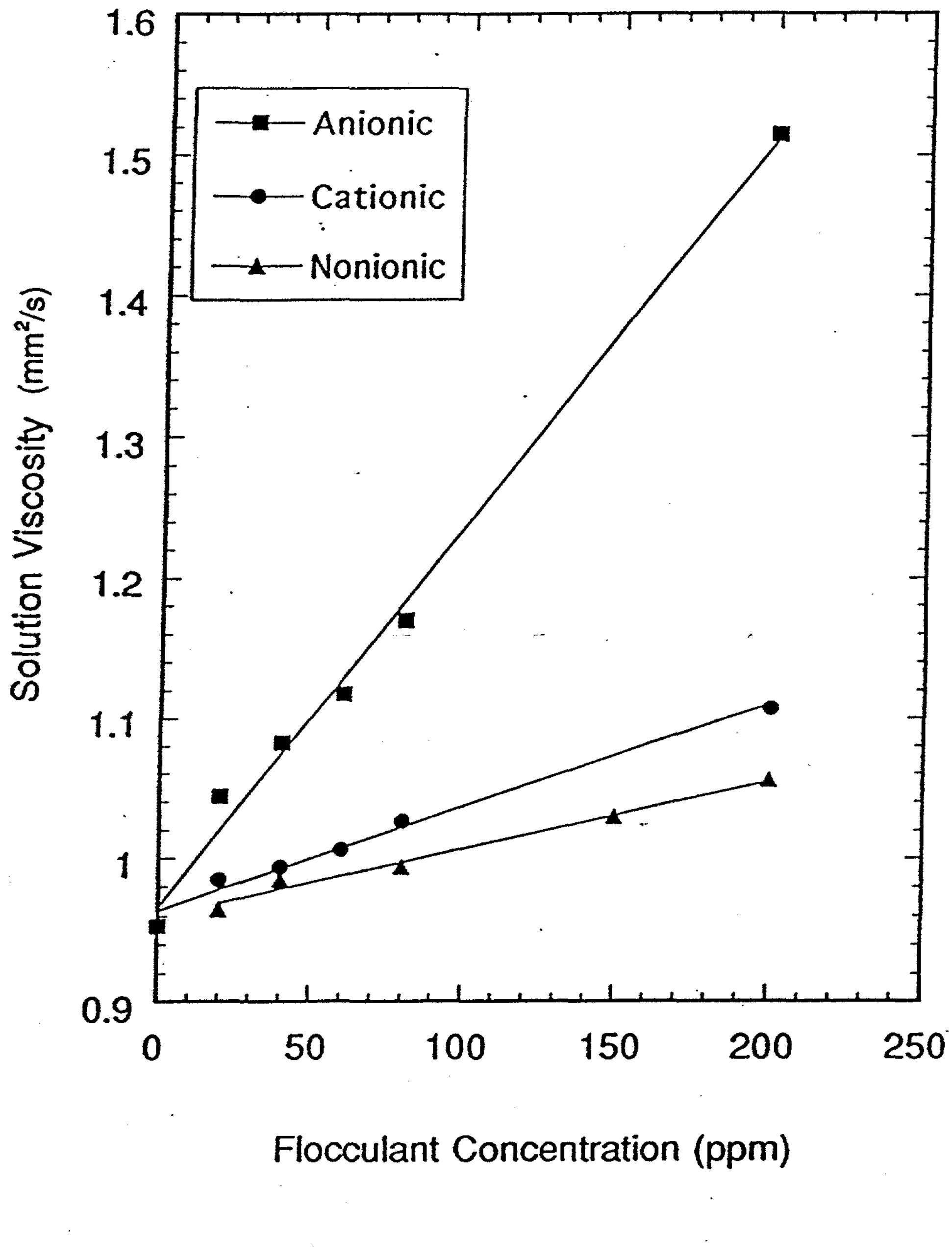

Figure 9. Relationship Between Solution Viscosity and Flocculant Concentration. (temperature: 24 ^{0}C; flocculant MW: 4-6 million)

The effect of the addition of non-ionic flocculants of different molecular weights on cake moisture is shown in Figure 10. The addition of 0.75 million molecular weight flocculant did not reduce the cake moisture because its chain length is not long enough to form suitable size flocs of the fine particles. The additions of 15 million molecular weight and 4-6 million molecular weight flocculant produced the same moisture reduction. However, the optimum dosage of the 15 million molecular weight flocculant was smaller than that of the 4-6 million molecular weight flocculant. In addition, the cake moisture was more sensitive to the dosage of 15 million molecular weight flocculant because the solution of larger flocculant has much higher viscosity than the solution of smaller flocculant. Figure 11 shows the relationships of solution viscosity and concentration of flocculants with various molecular weights. It can be seen that the viscosity of the solution of 15 million molecular flocculant is higher at the same concentration and increases more sharply as flocculant concentration increases.

In the present study, it was observed that the flocculation conditioning time and mixing rate had a negligible effect on cake moisture, as shown in Figure 12. In summary, the effects of flocculants on the filter cake moisture for the Pittsburgh No. 8 coal and for the Pocahontas No. 3 coal were different because they had different particle size distributions. The decrease of cake moisture due to the addition of the flocculants was about 7 percentage point for the Pittsburgh No.8 coal; however, it was only 1.5 percentage point for the Pocahontas No. 3 coal. The effect of flocculants on the filtration process is affected not only by flocculant properties but also by coal surface properties and especially particle size.

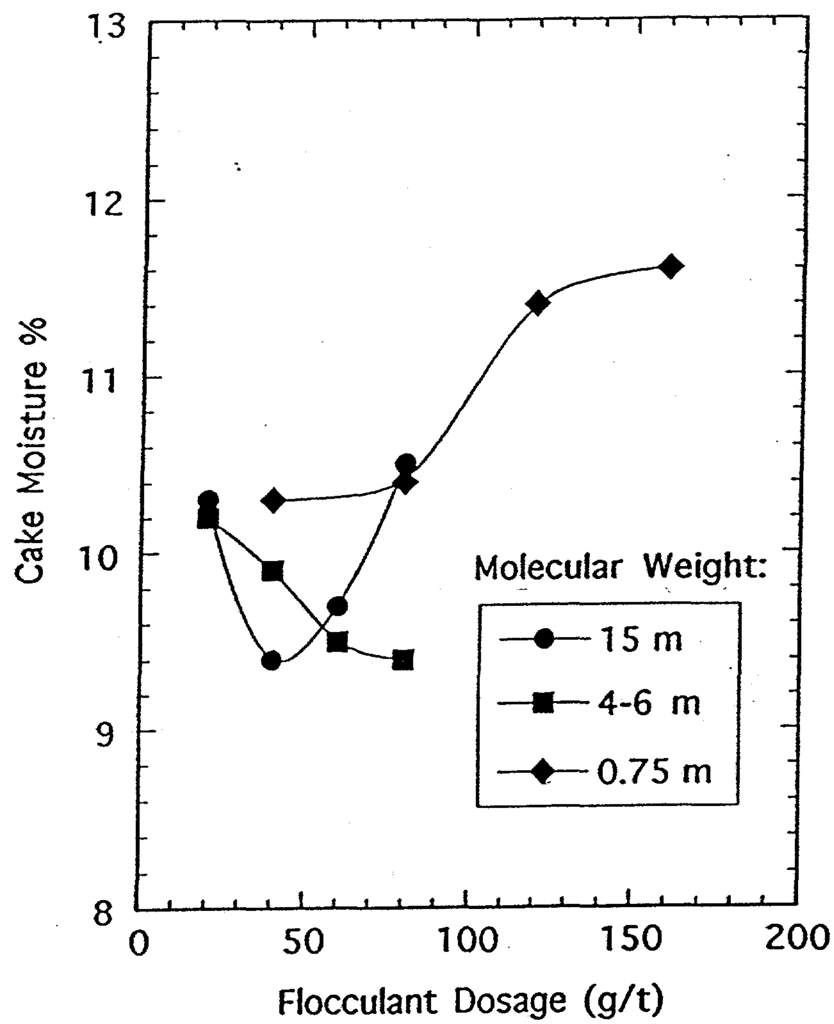

Figure 10. Effect of Flocculant Dosage on Cake Moisture for the Pocahontas No. 3 Coal. (flocculation time: 3 minutes; flocculation mixing rate: 350 rpm; non-ionic flocculants)

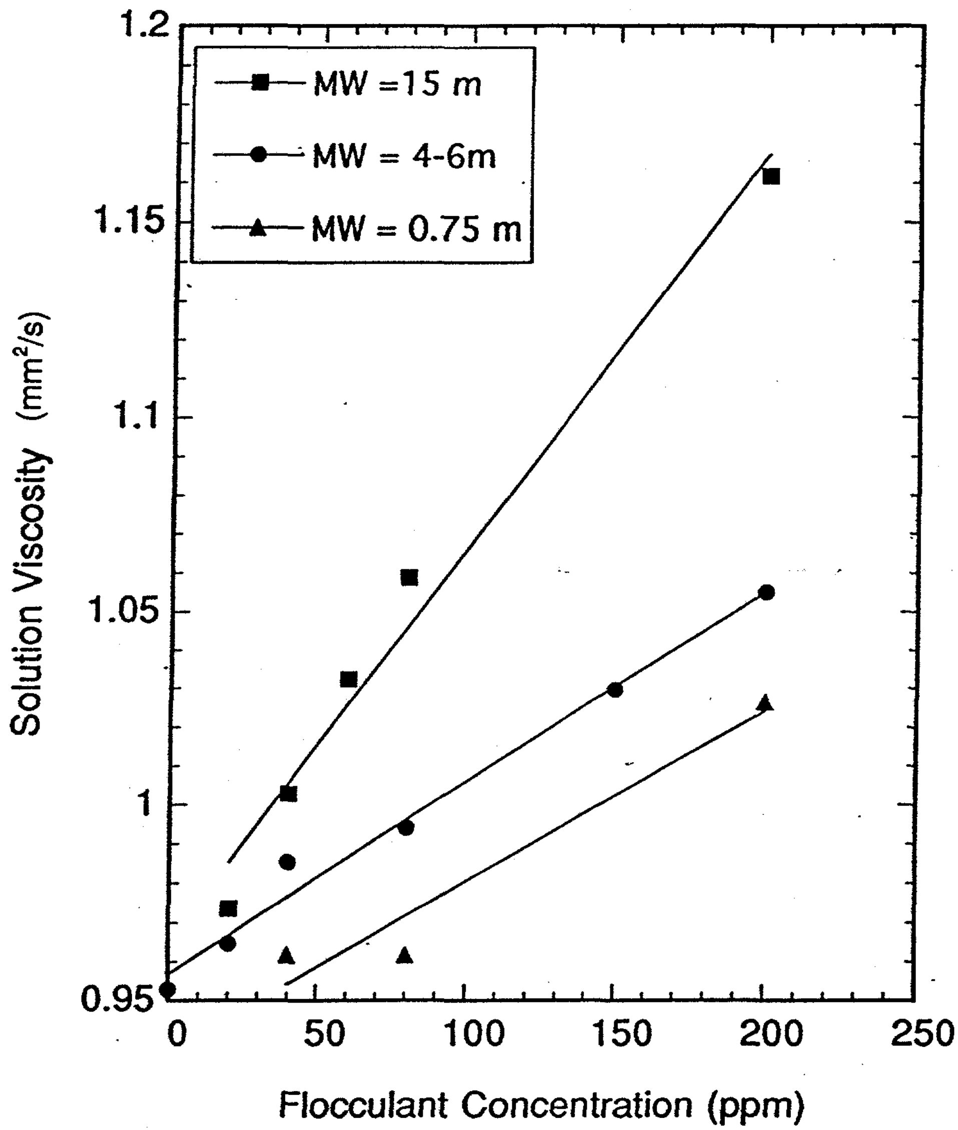

Figure 11. Relationship Between Solution Viscosity and Flocculant Concentration. (temperature: 24 ^{0}C; non-ionic flocculants)

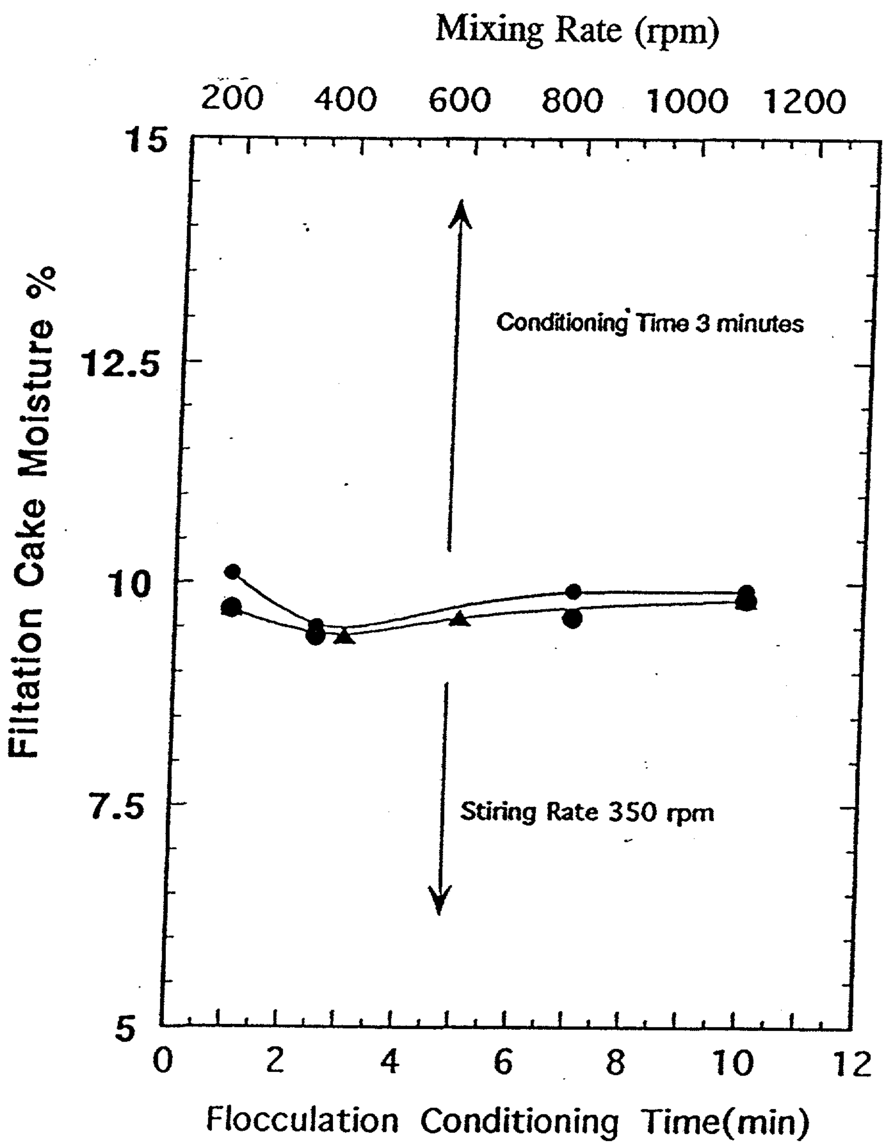

Figure 12. Effect of Flocculation Mixing Rate and Time on Cake Moisture for the Pocahontas No. 3 Coal. (non-ionic flocculant; MW=4-6 million)

Surfactants

The general mechanism of enhanced dewatering by surfactants is to reduce the liquid/air interfacial tension, leading to lower capillary retention forces and hence increasing cake drainage under a given pressure drop across the cake.

The effect of surfactant dosage on cake moisture for the Pocahontas No. 3 coal is presented in Figure 13. The addition of the anionic surfactant did not provide any improvement in cake moisture reduction. The reasons may be: (1) the addition of anionic surfactant caused much more foaming; (2) the zeta potential or surface charge on coal surface was negative at the natural pH, which prohibits the adsorption of anionic surfactant onto the coal particle surface. The lowest cake moisture was obtained by adding non-ionic surfactants. During the experiments, it was observed that non-ionic surfactant caused the minimum foaming. Non-ionic surfactant reduced the cake moisture from 11 to 9.7 percent at an optimum dosage of 800 g/t. Overall, surfactants did not profoundly decrease the cake moisture for the Pocahontas No.3 coal.

Combined Use of Flocculants and Surfactants

The effect of additives on the filtration rate for the Pocahontas No. 3 coal slurry is shown in Figure 14. Note that the addition of flocculant provided very high filtration rate due to the increase of particle size which improves the filter cake permeability. The addition of surfactant increased the filtration rate only near the end of the filtration process, which was also very small.

Flocculants and surfactants influence the different phases of a filtration process. The status of the liquid and air phase during filtration are schematically represented in

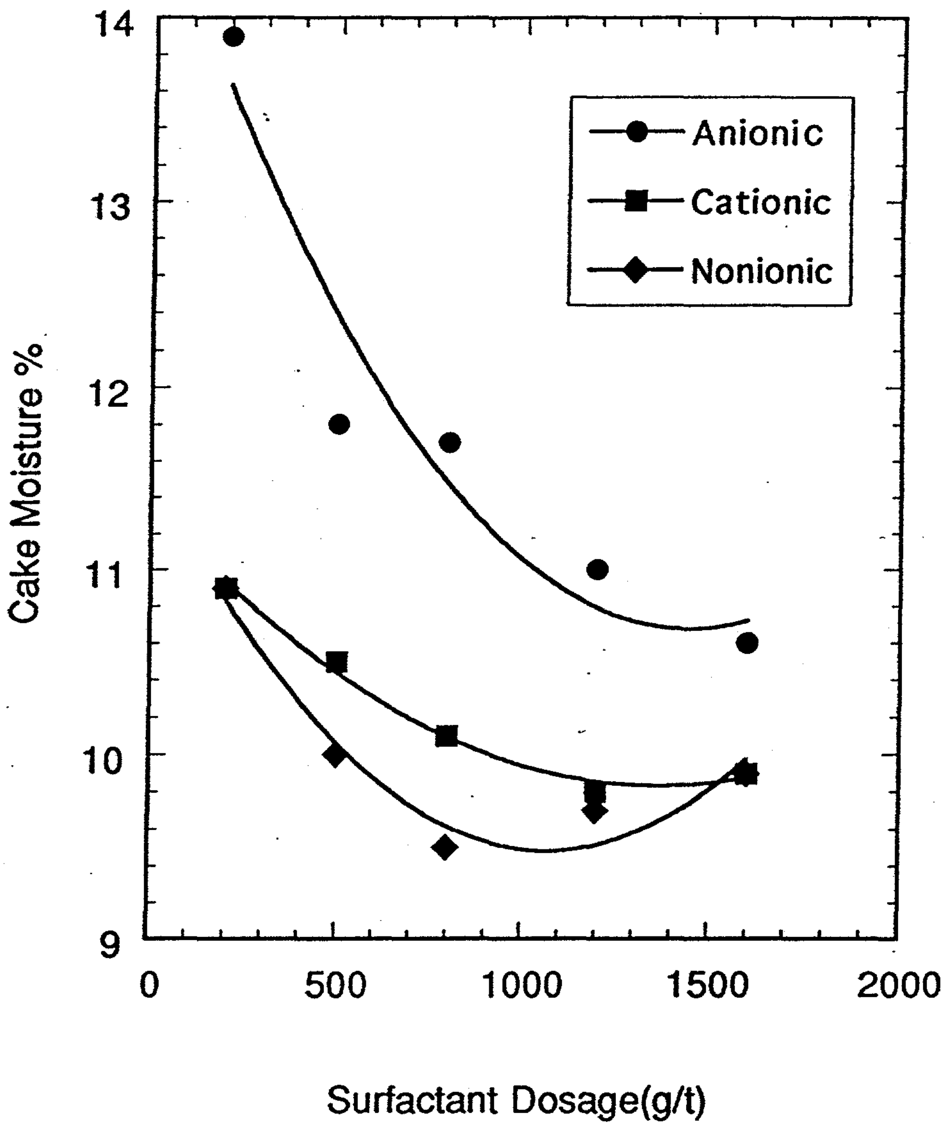

Figure 13. Effect of Surfactant Dosage on Cake Moisture for the Pocahontas No. 3 Coal. (conditioning time: 3 minutes; mixing rate: 350 rpm)

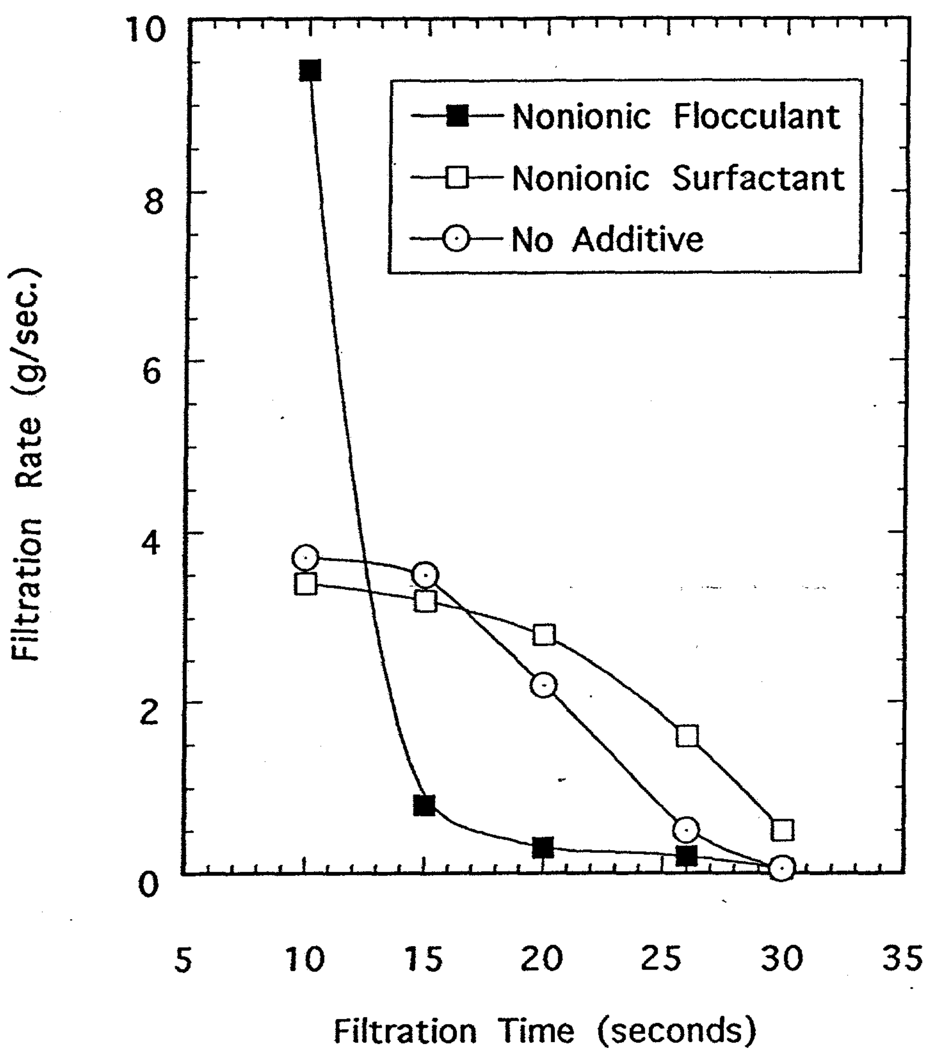

Figure 14. Effect of Additives on the Filtration Rate for the Pocahontas No. 3 Coal Slurry.

Figure 15. In phase a, b, c, only water passes through the channels in the cake. Flocculants increase the particle size and therefore increase the diameters of the channels or pores. Therefore, the addition of flocculants dramatically increases filtrate flow rate from the very beginning of the process. After most of the water has drained out of the cake, air begins to flow through the cake, as shown in phase d. The remaining water still covers the surface of the solids and isolates the solids from air at the beginning of phase d. As air continues to flow through the cake, the water covering on the surface of solids is replaced by air, which determines the residual moisture of the cake. The replacement is affected by several factors, such as diameters of the pores in the cake and the surface tension of air-water interface. Therefore, the effect of surfactant in filtration process can take place only in the phase d. The above analysis can help to understand why flocculants and surfactants affect the different phases of a filtration process.

Figure 16 shows the effect of combined use of flocculant and surfactant on the filtration process. A combination of non-ionic flocculant and non-ionic surfactant produced the best filtration result, decreasing the cake moisture to as low as 7.8 percent. When the flocculants and surfactants, having the same ionic charge, are mixed together, an electrostatic repulsion occurs between the flocculants and surfactants, which may affect the adsorption of additives on the coal surface. When the additives with different kinds of electric charges are used together, they will neutralize each other in the bulk phase. The interaction between non-ionic flocculant/surfactant is weaker than that between anionic/cationic flocculant/surfactant. This may be the reason why the combined

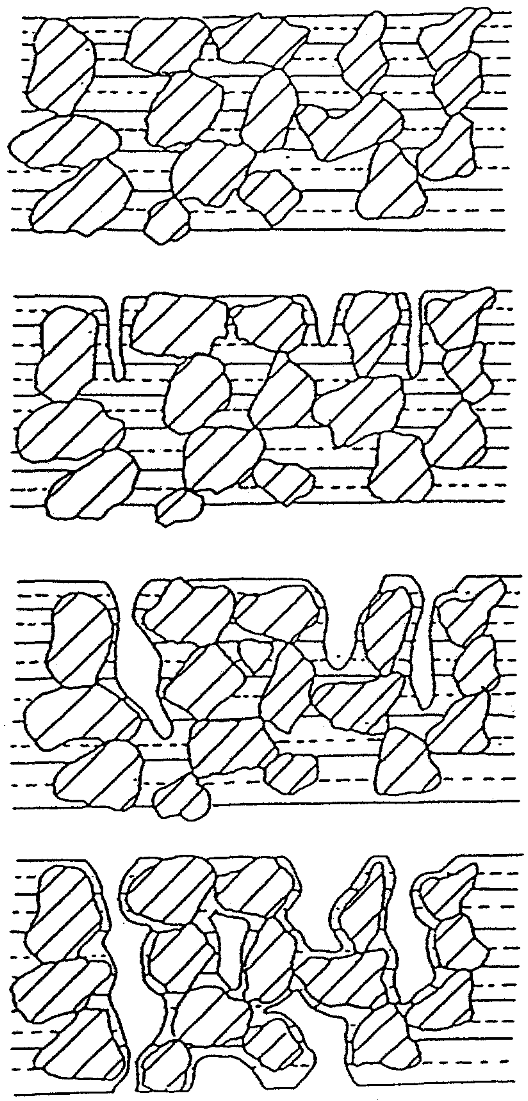

Figure 15. State of Liquid and Air During Filtration (Wakeman, 1975a).

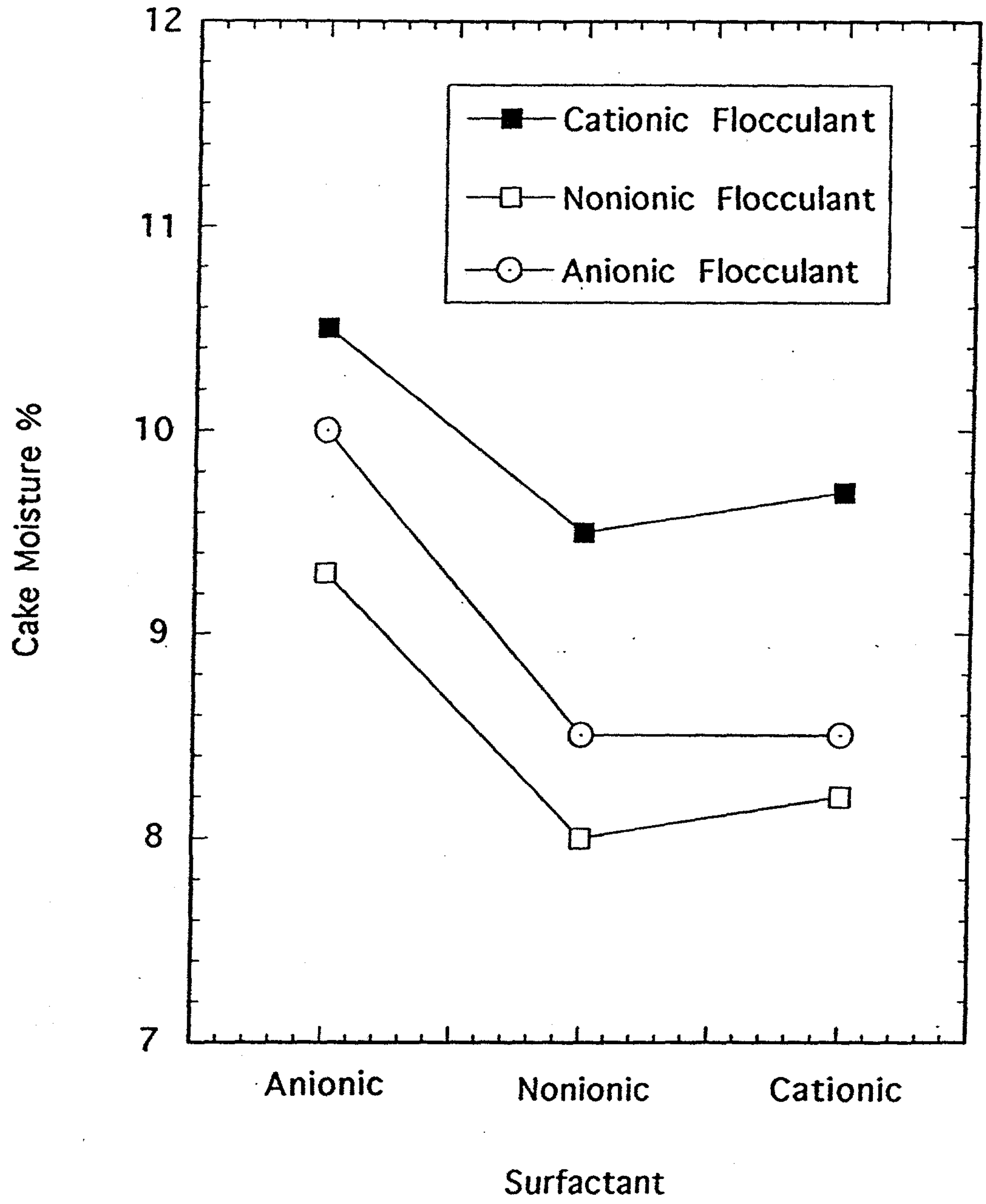

Figure 16. Effect of Combined Use of Flocculant and Surfactant for the Pocahontas No. 3 Coal. (flocculant and surfactant were added simultaneously and conditioned for 3 minutes)

use of non-ionic flocculant and non-ionic surfactant gave the largest reduction in cake moisture.

Metal Ions

Addition of metal ions, such as copper and aluminum ions, have been reported to have a beneficial effect on dewatering of fines (Parekh, 1979; Groppo, 1992). In this study, no noticeable reduction in the cake moisture due to the addition of metal ions for the Pocahontas No. 3 coal was observed, as shown in Figure 17. However, for the Pittsburgh No.8 coal, the effect of metal ion addition on cake moisture was more significant, as shown in Figure 18. Both copper and aluminum ions provided a 5 percentage point reduction in the cake moisture. The dosage of aluminum ion required was less than that of copper ion because aluminum ion is tri-valent, while copper ion is di-valent. Metal ions affect filtration processes mainly through changing electric charge at the interfaces of liquids/solids. Groppo (1992) and Parekh (1979) reported that the addition of metal ions at certain pH value coagulates fine particles. However, this is not applicable to larger size particles. Thus, in the present case, the Pittsburgh No. 8 coal having finer particle size was able to coagulate to larger size, providing lower filter cake moisture.

Phase III - Pilot Scale Testing

The pilot scale testing with the Andritz Ruthner Inc. hyperbaric filter unit were conducted at Consol Inc.'s two preparation plants, namely, Bailey mine (Pittsburgh No. 8) and Buchanan mine (Pocahontas No. 3). The Bailey mine is located in Green County near Washington, PA and the Buchanan mine is located in Buchanan County near

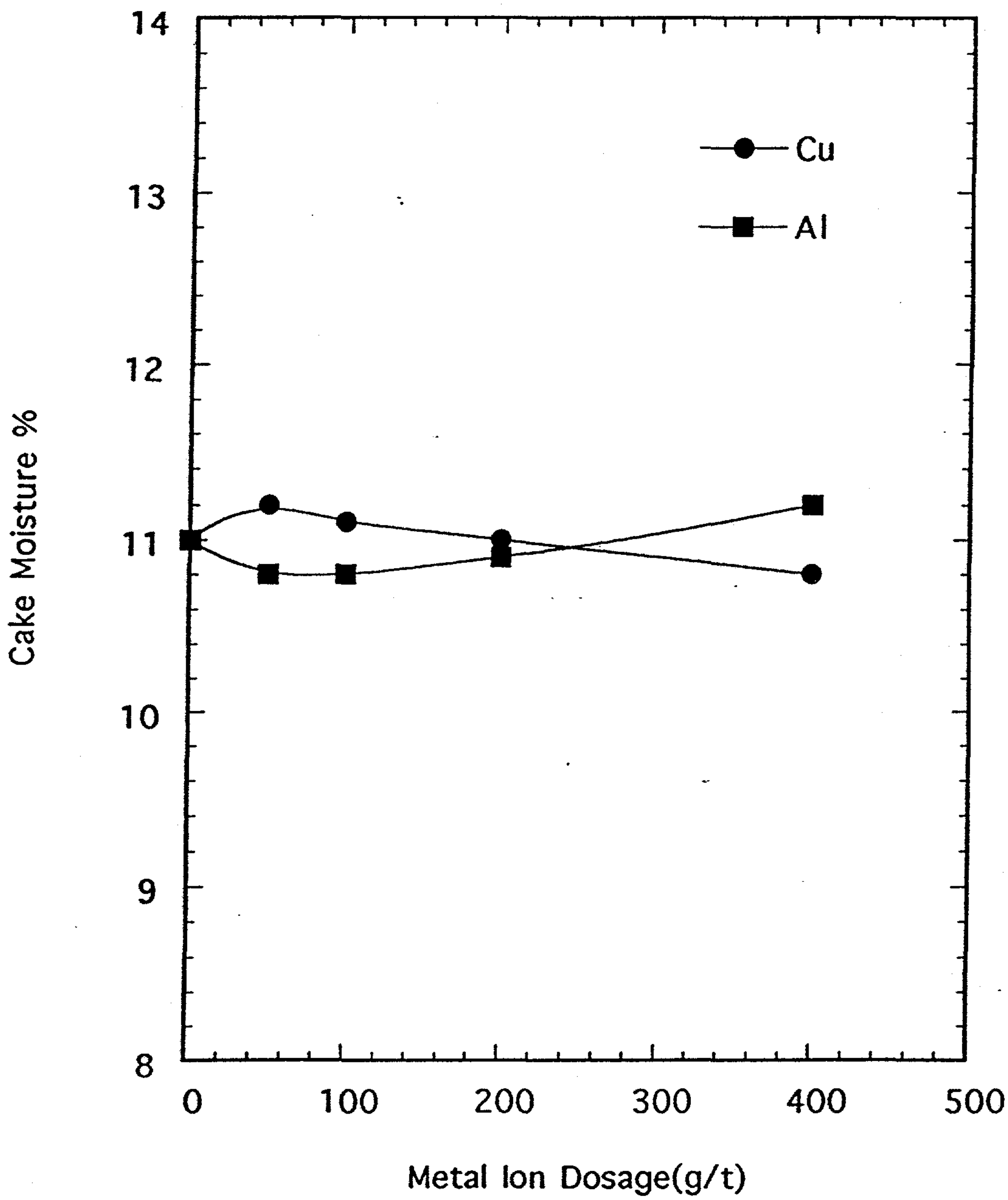

Figure 17. Effect of Metal Ion on Cake Moisture for the Pocahontas No.3 Coal. (conditioning time: 3 minutes)

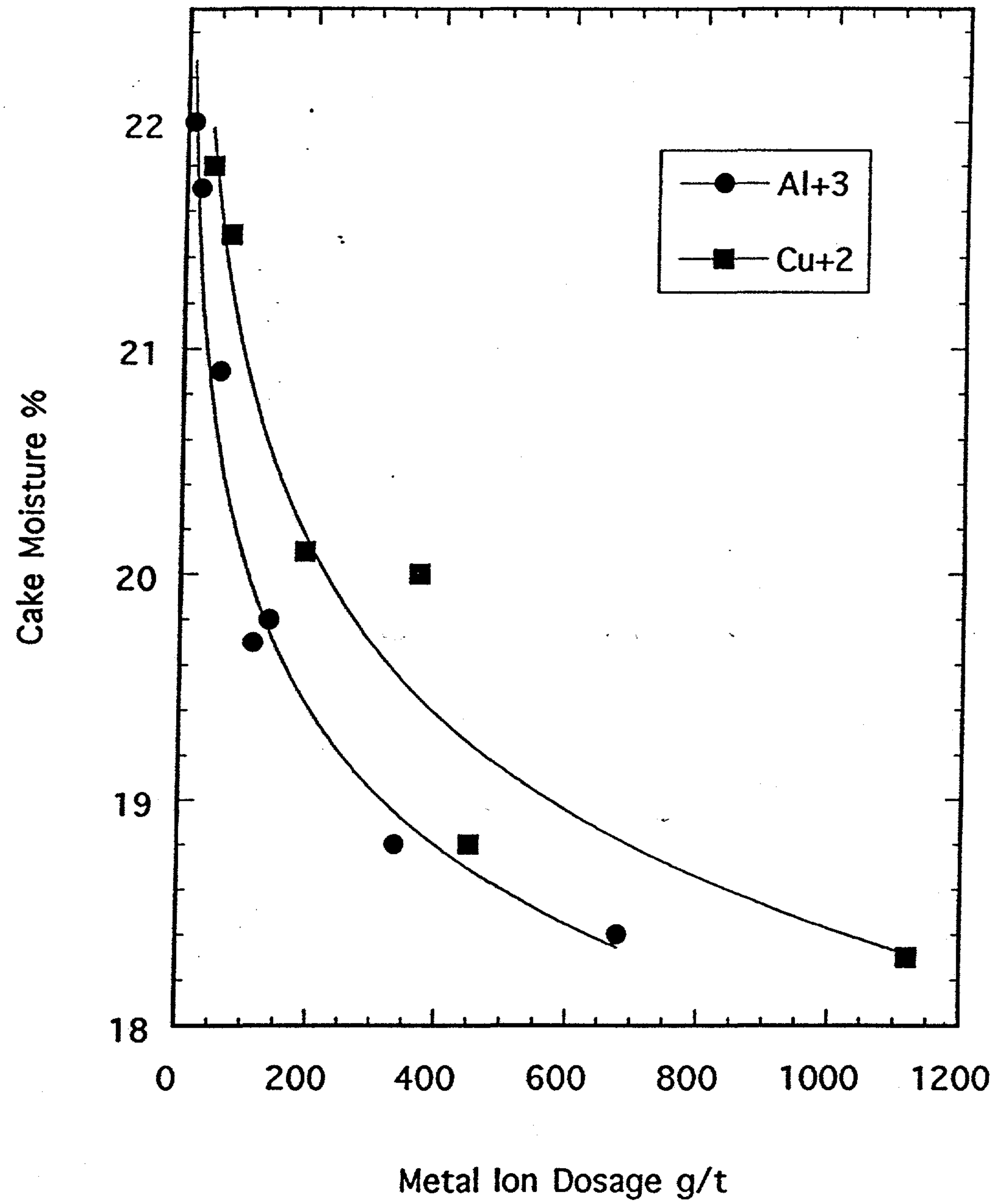

Figure 18. Effect of Metal Ion on Cake Moisture of the Pittsburgh No. 8 Coal. (conditioning time: 3 minutes)

Mavisdale, VA. Tables 1 and 2 list the pilot scale tests conducted at Bailey and Buchanan mines, respectively. As seen from the table that at Bailey 104 tests, and at Buchanan 41 tests were conducted. Pilot scale test data for the Bailey mine were given in last quarterly report and the Buchanan mine are listed in the appendix.

Figure 19 shows filter cake moisture, solids throughput, and air consumption at various pressure for the Bailey mine (Pittsburgh No. 8) clean coal concentrate. Note, that as expected filter cake moisture lowers with increasing pressure. Using 4.2 bar (62 psi) filter cake with 22 percent moisture was obtained. Increasing pressure to 5.8 bar (85 psi) did not lower the moisture significantly. At 4.2 bar (62 psi) pressure, the solids throughput and air consumption wre 128 lb/hr-ft^2 and 130 scfm/tph.

Figure 20 shows filter cake moisture, solids throughput and air consumption with respect to pressure for the Buchanan mine (Pocahontas No. 3) clean coal. It shows that filter cake moisture of 15.2 percent is achieved using 2 bar (30 psi) pressure. Increasing pressure to 5 bar (74 psi) moisture was lowered to 14.2 percent; at this pressure the solids throughput and air consumption were 340 lb/hr.ft^2 and 30 scfm/tph.

It is very surprising that with fine coal the air consumption was 130 scfm/tph compared to 30 scfm/tph for the coarse coal. The baseline data for the Buchanan mine clean coal slurry are summarized in Table 3.

FUTURE PLANS

A M.S. thesis on the dewatering topic will be finalized. One paper on the pilot scale studies will be prepared for publication. The dewatering model will be refined and finalized.

Table 1. 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

Table 2. List of Pilot Scale Tests Conducted at the Consol Inc. Buchanan Mine

A. FILTER FEED (32 Tests)

1. Experimental Design (15 tests)
2. Anionic Flocculant (5 tests)
3. Cationic Surfactant (5 tests)
4. Anionic Surfactant (5 tests)
5. Anionic Surfactant and Cu^2 Ions (5 tests)
6. Other (1 test)

B. CLASSIFIED FILTER FEED

1. Size, Pressure and Solid Content (5 tests)

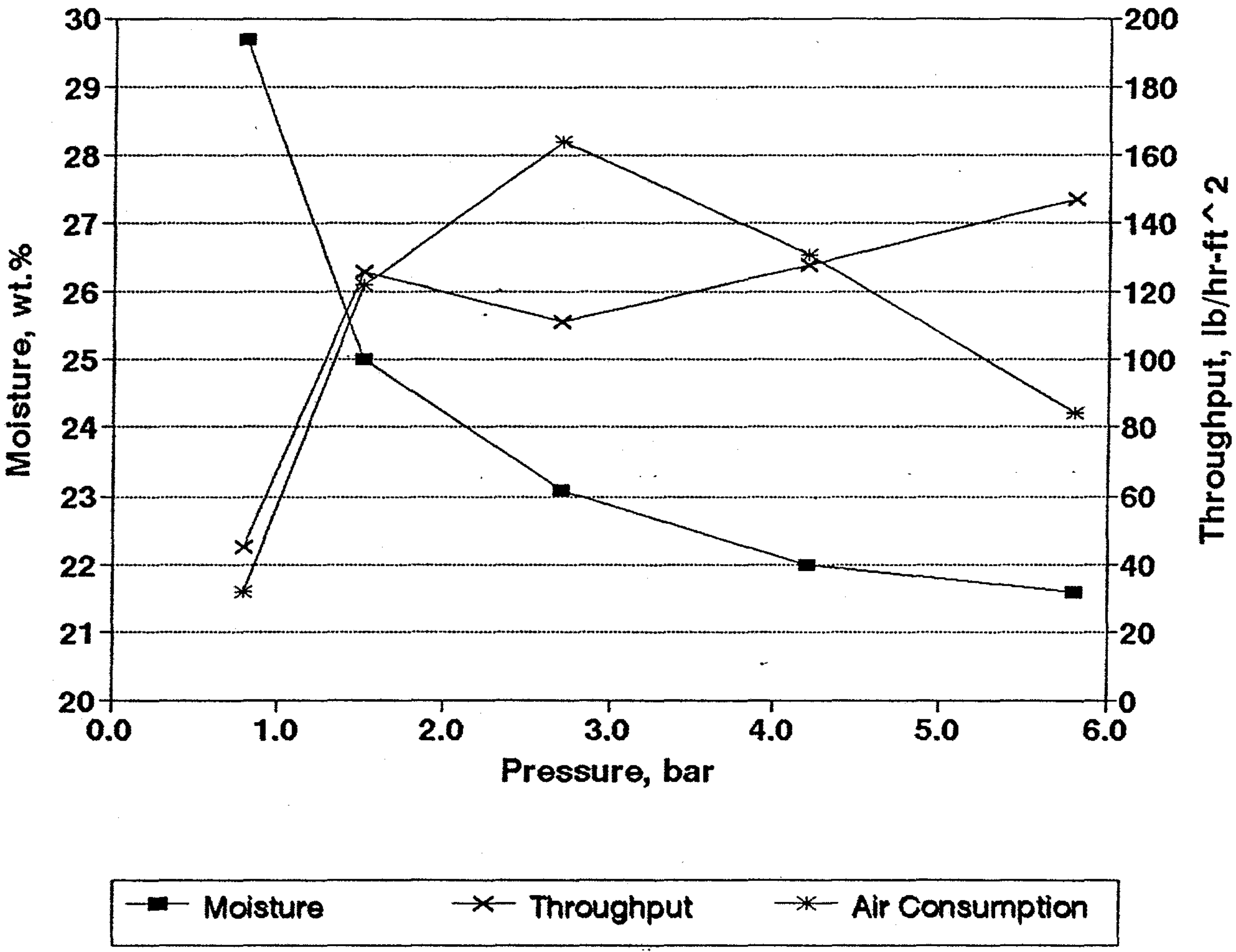

Figure 19. Effect of applied pressure on filter cake moisture, solid throughput and air consumption for the Bailey mine Pittsburgh No. 8 clean coal slurry (cake thickness ~ 15 mm).

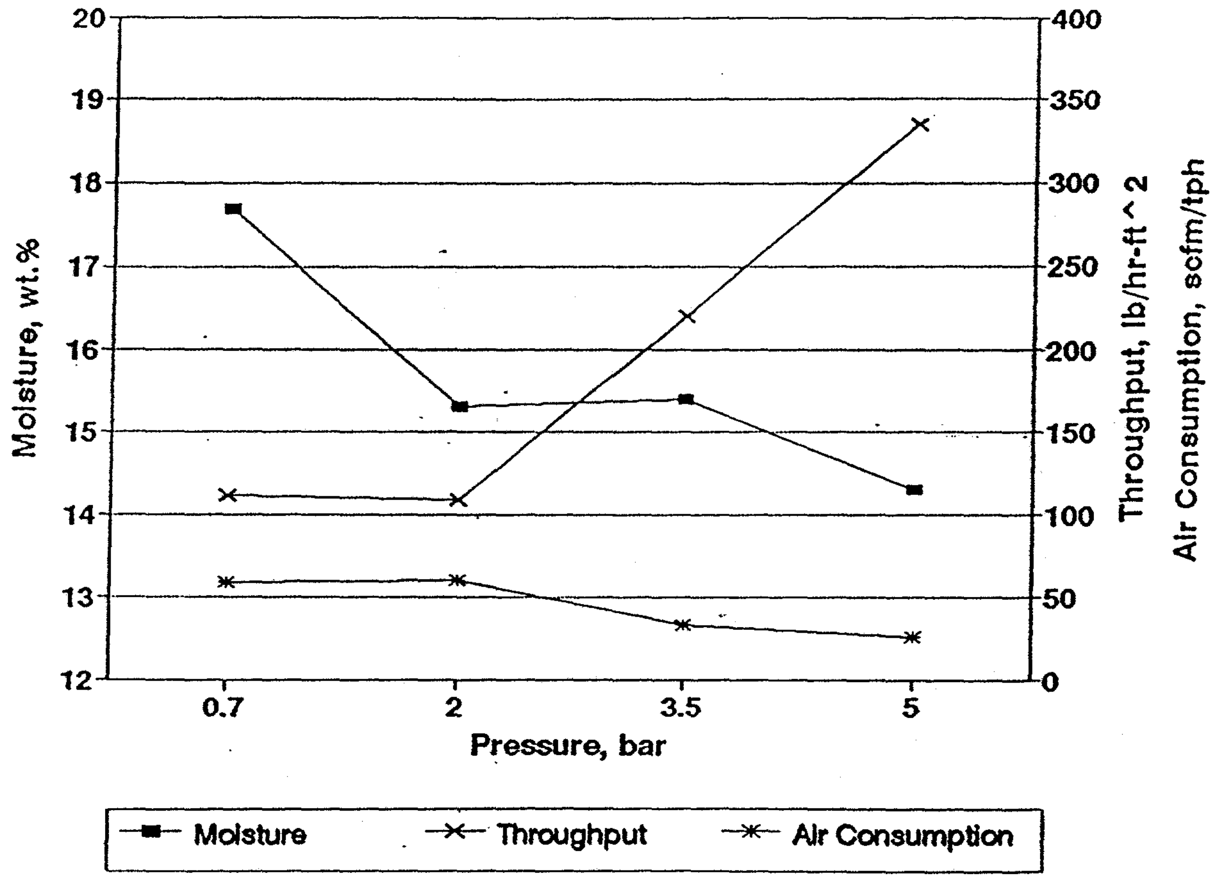

Figure 20. Effect of applied pressure on filter cake moisture, air consumption and solid throughput for the Buchanan mine Pocahontas No. 3 clean coal slurry slurry (cake thickness ~ 15 mm).

Table 3. Summary of baseline pilot scale hyperbaric filter tests conducted at the Buchanan mine (Pocahontas No. 3).

Product	% Moisture in Filter Cake	Throughput lb/hr.ft^2	Air Consumption scfm/tph
Filter feed (28x0 mesh)	13-17	150-400	40-80
Classified feed (28x100 mesh)	12-14	150-400	30-100

Appendix

Table A. Andritz Hyperbaric Filter Pilot Plant Data for the Buchanan Mine Clean Coal Slurry

ANDRITZ

Data HBF-Trailer

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

Material: Coal concentrate
filter fabric: 796-07-1020-k068-USA Tetko
Filter area: 2.0 m²

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

ANDRITZ

Test-nr.	Date	Test-time	Type conc.	Vessel pres.	Pres CFA	Pres-CDA	Disc-speed	feed-flow	Ash Lab	FS Lab	FS Calc.	Part. <63µ	feed-temp.	Air Compr.	Air filtrate	filtrate flow	filtrate	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	%DS	Nm3/h	Nm3/h	l/h		%	g/t	sec.	cycle	kg	kg/m²h	lb/ft²h	%	Nm³/t	cfm/ton	mm
Filter Feed																																
CFA=85 degree																																
#100	31/10/1994	14:30	28/0	3.5	0.0	0.0	1.24	8.8		26.6	32.1		25	259	286	6.00	0.1	125	161	0	none	0.0	0	41	1	48.4	1802	369	15.8	45	24	26.3
#101	31/10/1994	16:30	28/0	5.0	0.0	0.0	1.95	10.5		30.7	28.9		26	305	329	7.48	0.1	170	159	0	none	0.0	0	42	1	53.8	1995	409	14.2	40	21	20.6
#102	01/11/1994	10:05	28/0	2.0	0.0	0.0	2.01	7.3		22.6	29.7		24	78	249	5.16	0.2	25	224	0	none	0.0	0	47	1	37	1187	243	15.4	94	50	12.0
#103	01/11/1994	11:30	28/0	3.5	0.0	0.0	1.27	7.7		24.2	27.8		24	184	315	5.58	0.1	125	190	0	none	0.0	0	51	1	42.9	1267	260	15.8	75	40	21.5
#104	01/11/1994	13:00	28/0	2.0	0.0	0.0	0.53	5.6		22.2	32		23	131	241	3.77	0.0	25	216	0	none	0.0	0	48	1	27.6	880	180	15.2	123	65	31.9
#105	01/11/1994	14:50	28/0	0.7	0.0	0.0	1.25	5.4		21.9	37.8		22	28	172	3.37	0.1	0	172	0	none	0.0	0	46	1	23.9	770	158	17.7	111	59	14.9
#106	01/11/1994	16:10	28/0	5.0	0.0	0.0	0.51	7.5		23.2	25.8		23	237	316	5.57	0.0	170	146	0	none	0.0	0	46	1	37.9	1278	262	14.5	57	30	43.0
#107	01/11/1994	17:30	28/0	3.5	0.0	0.0	1.24	8.1		20.9	26.4		24	217	305	5.94	0.1	125	180	0	none	0.0	0	44	1	40.4	1405	288	14.6	64	34	23.4
CFA = 85 degree Deslimed Filter Feed																																
1 cyclone operating at 20-30 PSI																																
#108	02/11/1994	11:10	28/0	3.5	0.0	0.0	1.25	11.3		27.4	25.1		16	251	338	8.45	0.0	125	213	0	none	0.0	0	43	1	53.6	1920	394	14.0	55	30	30.8
#109	02/11/1994	12:35	28/0	5.0	0.0	0.0	1.24	10.5		26.9	23.3		17	313	359	8.02	0.0	170	189	0	none	0.0	0	41	1	47.5	1791	367	13.2	53	28	28.4
CFA = 85 degree Deslimed Filter Feed																																
1 cyclone operating at 20-30 PSI more dilute than Test Nr:109																																
#110	02/11/1994	14:05	28/0	5.0	0.0	0.0	1.26	9.9		18.6	14		18	371	409	8.51	0.0	170	239	0	none	0.0	0	41	1	28.8	1106	227	12.4	108	58	16.0
#111	02/11/1994	15:30	28/0	3.5	0.0	0.0	1.26	9.0		19.7	14.8		22	312	371	7.65	0.0	125	246	0	none	0.0	0	40	1	25.5	1006	206	12.3	122	65	15.4
CFA = 85 degree Deslimed Filter Feed																																
2 cyclones operating at 10 PSI																																
#112	02/11/1994	16:55	28/0	3.5	0.0	0.0	1.26	7.9		26.7	21.6		24	249	306	6.24	0.0	125	181	0	none	0.0	0	40	1	33.8	1299	266	13.6	70	37	20.5
Filter Feed																																
CFA=55 deg																																
#113	03/11/1994	13:45	28/0	3.5	0.0	0.0	1.97	7.4		23.5	27.1		28	289	371	5.36	0.1	100	271	0	none	0.0	0	43	1	36.5	1330	273	13.0	102	54	14.6
#114	03/11/1994	14:55	28/0	3.5	0.0	0.0	0.48	4.8		27.8	29.7		28	290	315	3.40	0.0	100	215	0	none	0.0	0	52	1	28	824	169	14.2	130	70	32.6
#115	03/11/1994	16:10	28/0	5.0	0.0	0.0	1.24	7.3		26.9	26.5		28	360	420	5.38	0.1	140	280	0	none	0.0	0	41	1	35.8	1342	275	14.2	104	56	20.6
#116	04/11/1994	08:10	28/0	2.0	0.0	0.0	1.27	5.1		20.9	29.2		28	162	316	3.58	0.1	10	306	0	none	0.0	0	39	1	18.8	747	153	14.4	205	109	13.4
CFA = 55 degree Deslimed Filter Feed																																
1cyclone operating at 30 PSI																																
#117	04/11/1994	10:10	28/0	3.5	0.0	0.0	1.25	6.3		20	16.3		22	322	397	5.31	0.1	100	297	0	none	0.0	0	41	1	20.3	781	160	12.4	190	101	15.5
CFA = 55 degree Filter Feed																																
Sigma surf. I-Hexadecylpyridinium Chloride																																
#118	08/11/1994	16:10	28/0	5.0	0.0	0.0	1.26	6.7		24	25.3		28	340	361	5.04	0.1	140	221	338	none	0.2	293	41	1	31.2	1194	245	12.0	92	49	18.6

Table A (continued)

Data HBF-Trailer

Projectno.: 88-24141/0
Code: Hyperbaric filtration
customer: Consol USA
Buchanan

Material: Coal concentrate
filter fabric: ZBF 07-1020-4068-USA Teka
Filter area: 2.0 m²

H.Riemer AAG / John Hughes Ana
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	FS Calc.	Part. <63µ	feed-temp.	Air Compr.	Air filtrate	filtrate flow	filtrate	air corr-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	%DS	Nm3/h	Nm3/h	l/h		%	g/t	sec.	cycle	kg	kg/m²h	lb/ft²h	%	Nm³/t	cfm/ton	mm
Filter Feed																																
CFA = 55 degree Deslimed Filter Feed																																
1 cyclone operating at 30 PSI																																
#119	09/11/1994	16:10	28/0	5.0	0.0	0.0	1.25	7.1		19.5	20.7		24	379	418	5.67	0.1	140	278	0	none	0.0	0	41	1	25.3	969	199	12.2	144	77	16.7
CFA = 55 degree Filter Feed																																
Sigma surf. 1-Hexadecylpyridinium Chloride Deslimed Filter Feed																																
#120	10/11/1994	08:50	28/0	5.0	0.0	0.0	1.24	7.9		19	17.8		20	337	391	6.49	0.1	140	251	396	none	0.2	390	42	1	27	1015	208	11.8	124	66	17.1
CFA = 55 degree Deslimed Filter Feed																																
1cyclone operating more concentrated																																
#121	10/11/1994	10:35	28/0	5.0	0.0	0.0	1.29	7.4		27.9	26		20	338	389	5.51	0.0	140	249	0	none	0.0	0	42	1	36.8	1366	280	13.1	91	49	19.4
CFA = 55 degree Deslimed Filter Feed																																
1cyclone operating at 30 PSI more concentrated																																
#122	14/11/1994	09:30	28/0	3.5	0.0	0.0	1.25	6.7		27	32		22	224	335	4.52	0.1	100	235	0	none	0.0	0	44	1	37	1313	269	14.0	89	48	19.1
CFA = 55 degree Deslimed Filter Feed																																
1cyclone operating at 30 PSI -decreased apex diameter to 5/8"																																
#123	14/11/1994	13:45	28/0	3.5	0.0	0.0	1.23	9.5		29.1	35.3		24	358	418	6.13	0.1	100	318	0	none	0.0	0	38	1	40	1627	334	13.3	98	52	26.9
CFA = 55 degree Deslimed Filter Feed																																
1cyclone operating at 30 PSI -decreased apex diameter to 5/8" slurry diluted																																
#124	14/11/1994	15:25	28/0	5.0	0.0	0.0	1.27	9.3		19.9	23.8		22	435	482	7.08	0.0	140	342	0	none	0.0	0	40	1	33.1	1312	269	11.9	130	70	19.5
CFA=55 degree Filter Feed																																
Surfactant Arquad T50 Activity 50 %																																
#125	15/11/1994	10:00	28/0	5.0	0.0	0.0	1.26	6.4		22.9	24.6		28	276	363	4.81	0.1	140	223	76	none	0.4	136	45	1	31.8	1110	228	12.7	100	54	15.0
#126	15/11/1994	10:15	28/0	5.0	0.0	0.0	1.26	6.4		20.9	13.8		26	249	329	5.54	0.1	140	189	326	none	0.4	609	46	1	31	1070	219	12.3	88	47	15.5
#127	15/11/1994	10:30	28/0	5.0	0.0	0.0	1.26	6.5		21	6.9		26	240	284	6.04	0.1	140	144	648	none	0.4	1235	45	1	29.8	1049	215	12.0	68	37	15.5
CFA =55 degree Filter Feed																																
Baseline to Test #125 - #127																																
#128	15/11/1994	11:10	28/0	5.0	0.0	0.0	1.23	8.1		27.3	28.2		28	376	459	5.79	0.0	140	319	0	none	0.0	0	45	1	43.3	1495	306	14.3	107	57	22.0
CFA=55 degree Filter Feed																																
Surfactant Ethoquad Cl 12 activity 70 %																																
#129	15/11/1994	13:40	28/0	5.0	0.0	0.0	1.22	6.7		23.5	24.2		28	334	406	5.05	0.0	140	266	80	none	0.2	68	45	1	33.5	1173	240	13.4	114	61	17.0
#130	15/11/1994	14:20	28/0	5.0	0.0	0.0	1.26	6.3		23	11.8		28	320	354	5.54	0.1	140	214	94	none	0.6	272	44	1	31.4	1111	228	12.9	96	51	17.0
#131	15/11/1994	14:40	28/0	5.0	0.0	0.0	1.26	6.1		23.2	4.07		28	310	323	5.83	0.1	140	183	186	none	0.6	562	45	1	30.3	1068	219	12.5	86	46	16.5

Table A (continued)

Data HBF-Trailer

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

Material: Coal concentrate
filter fabric 786-07-1020-k088-USA Teske
Filter area: 2.0 m²

H.Riemer AAG / John Hughes Anus
Gary Meenen/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	FS Calc. %TS	Part. < 63µ %TS	feed-temp. °C	Air Compr. Nm³/h	Air filtrate Nm³/h	filtrate flow m³/h	filtrate %DS	air corr-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																																
CFA = 65 degree Filter Feed																																
Baseline to Test #129 - #130																																
#132	15/11/1994	15:25	28/0	5.0	0.0	0.0	1.26	7.0		27.9	28.8		28	375	427	5.02	0.1	140	287	0	none	0.0	0	44	1	35	1226	251	13.7	117	62	18.0
CFA = 65 degree Filter Feed																																
Filter Feed diluted with water																																
#133	15/11/1994	16:05	28/0	5.0	0.0	0.0	1.25	6.6		23.4	25.4		26	405	454	4.93	0.1	140	314	663	none	0.0	0	45	1	30.6	1063	218	13.2	148	79	16.5
CFA = 125 degree Filter Feed																																
standard test																																
#134	16/11/1994	09:20	28/0	5.0	0.0	0.0	1.25	11.0		28.7	29.1		30	412	485	7.77	0.0	317	168	0	none	0.0	0	42	1	54.4	1964	403	15.6	43	23	27.4
#135	16/11/1994	10:45	28/0	3.5	0.0	0.0	1.97	12.2		29.4	30.1		30	344	438	8.54	0.1	300	138	0	none	0.0	0	40	1	59.9	2255	462	16.3	31	16	22.0
#136	16/11/1994	12:25	28/0	3.5	0.0	0.0	0.48	7.4		24.1	31.1		30	318	378	5.08	0.0	300	78	0	none	0.0	0	41	1	35.7	1325	272	16.4	30	16	46.0
#137	16/11/1994	13:35	28/0	2.0	0.0	0.0	1.24	8.4		23.4	32.1		30	258	331	5.73	0.1	268	63	0	none	0.0	0	38	1	39.8	1558	319	16.8	20	11	22.7
CFA = 125 degree Filter Feed																																
Test with Cetyl Pyridinium																																
#138	16/11/1994	15:15	28/0	5.0	0.0	0.0	1.24	11.3		27.9	24.9		28	344	401	9.50	0.0	300	101	566	none	0.2	280	40	1	53	2023	415	15.1	25	13	28.5

Data HBF-Trailer

ANDRITZ

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

Material: Coal concentrate
filter fabric: 28F-07-1020-4066-USA Tetko
Filter area 2.0 m²

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

ANDRITZ

Test-nr.	Date	Test-time	Type conc.	Vessel pres.	Pres CFA	Pres-CDA	Disc-speed	Feed-flow	Ash Lab	FS Lab	FS Calc.	Part. <63µ	Feed-temp.	Air Compr.	Air filtrate	filtrate flow	filtrate	air corr-factor	act. filtr.air	Polym. flow	Polym. Type	Polym conc.	Polym. dose	Discharg time	Nr. of disch.	Cake weight	Throughp. dry	hroughp dry	Moisture	Air-consumpt.	Air-consump	cake thickness
	DD/MM/YY	hh:mm		Bar	Bar	Bar	rpm	m³/h	%TS	%TS	%TS	%TS	°C	Nm³/h	Nm³/h	m³/h	%DS	Nm3/h	Nm3/h	l/h		%	g/t	sec.	cycle	kg	kg/m²h	lb/ft²h	%	Nm³/t	cfm/ton	mm
Filter Feed																																
CFA=65 deg																																
##200	31/10/1994	14:30	28/0	3.5	0.0	0.0	1.24	8.8		26.8	32.1		25	259	286	6.00	0.1	125	161	0	none	0.0	0	41	1	48.4	1808	371	15.5	44	24	26.3
##201	31/10/1994	16:30	28/0	5.0	0.0	0.0	1.95	10.5		31.3	28.9		28	305	329	7.48	0.1	170	159	0	none	0.0	0	42	1	53.8	2000	410	14.0	40	21	20.6
##202	01/11/1994	10:05	28/0	2.0	0.0	0.0	2.01	7.3		21.4	29.7		24	78	249	5.16	0.1	25	224	0	none	0.0	0	47	1	37	1191	244	15.1	94	50	12.0
##203	01/11/1994	11:30	28/0	3.5	0.0	0.0	1.27	7.7		21.1	27.8		24	184	315	5.58	0.1	125	190	0	none	0.0	0	51	1	42.9	1279	262	15.0	74	40	21.5
##204	01/11/1994	13:00	28/0	2.0	0.0	0.0	0.53	5.6		21.7	32		23	131	241	3.77	0.0	25	216	0	none	0.0	0	48	1	27.6	876	180	15.6	123	66	31.9
##205	01/11/1994	14:50	28/0	0.7	0.0	0.0	1.25	5.4		20.7	37.8		22	28	172	3.37	0.2	0	172	0	none	0.0	0	46	1	23.9	773	159	17.3	111	59	14.9
##206	01/11/1994	16:10	28/0	5.0	0.0	0.0	0.51	7.5		22.9	25.8		23	237	316	5.57	0.0	170	146	0	none	0.0	0	46	1	37.9	1279	262	14.4	57	30	43.0
##207	01/11/1994	17:30	28/0	3.5	0.0	0.0	1.24	8.1		21.8	26.4		24	217	305	5.94	0.1	125	180	0	none	0.0	0	44	1	40.4	1407	288	14.5	64	34	23.4
CFA = 85 degree Deslimed Filter Feed																																
1 cyclone operating at 20-30 PSI																																
##208	02/11/1994	11:10	28/0	3.5	0.0	0.0	1.25	11.3		28.7	25.1		16	251	338	8.45	0.0	125	213	0	none	0.0	0	43	1	53.6	1922	394	13.9	55	30	30.8
##209	02/11/1994	12:35	28/0	5.0	0.0	0.0	1.24	10.5		26.9	23.3		17	313	359	8.02	0.0	170	189	0	none	0.0	0	41	1	47.5	1793	368	13.1	53	28	28.4
CFA = 85 degree Deslimed Filter Feed																																
1 cyclone operating at 20-30 PSI more dilute than Test Nr:109																																
##210	02/11/1994	14:05	28/0	5.0	0.0	0.0	1.26	9.9		18	14		18	371	409	8.51	0.0	170	239	0	none	0.0	0	41	1	28.8	1114	228	11.8	107	57	16.0
##211	02/11/1994	15:30	28/0	3.5	0.0	0.0	1.26	9.0		18.2	14.8		22	312	371	7.65	0.0	125	246	0	none	0.0	0	40	1	25.5	1003	206	12.5	122	65	15.4
CFA = 85 degree Deslimed Filter Feed																																
2 cyclone operating at 10 PSI																																
##212	02/11/1994	16:55	28/0	3.5	0.0	0.0	1.26	7.9		0	21.5		24	249	306	6.24	0.0	125	181	0	none	0.0	0	40	1	33.8	1299	266	13.6	70	37	20.5
Filter Feed																																
CFA=55 deg																																
##213	03/11/1994	13:45	28/0	3.5	0.0	0.0	1.97	7.4		24.4	27.1		28	289	371	5.36	0.1	100	271	0	none	0.0	0	43	1	36.5	1323	271	13.5	103	55	14.6
##214	03/11/1994	14:55	28/0	3.5	0.0	0.0	0.48	4.8		28.1	29.7		28	290	315	3.40	0.0	100	215	0	none	0.0	0	52	1	28	823	169	14.3	130	70	32.5
##215	03/11/1994	16:10	28/0	5.0	0.0	0.0	1.24	7.3		27.6	26.5		28	360	420	5.38	0.1	140	280	0	none	0.0	0	41	1	35.8	1352	277	13.6	104	55	20.6
##216	04/11/1994	08:10	28/0	2.0	0.0	0.0	1.27	5.1		20.6	29.2		28	162	316	3.58	0.1	10	306	0	none	0.0	0	39	1	18.8	745	153	14.6	205	109	13.4
CFA = 55 degree Deslimed Filter Feed																																
1cyclone operating at 30 PSI																																
##217	04/11/1994	10:10	28/0	3.5	0.0	0.0	1.25	6.3		21.9	16.3		22	322	397	5.31	0.1	100	297	0	none	0.0	0	41	1	20.3	776	159	13.0	192	102	15.5
CFA=55 degree Filter Feed																																
1-Hexadecylpyridinium Chloride																																
##218	07/11/1994	10:55	28/0	3.5	0.0	0.0	1.24	5.4		13.4	29		22	163	289	3.83	0.1	100	189	65	none	0.2	75	58	1	32.4	864	177	13.4	109	58	14.0
##219	07/11/1994	13:44	28/0	5.0	0.0	0.0	1.22	6.2		24.1	28.4		22	224	338	4.44	0.1	140	198	74	none	0.2	79	60	1	36.4	952	195	12.8	104	55	17.0

Table A (continued)

Data HBF-Trailer

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

Material: Coal concentrate
filter fabric: 2BF-07-1020-4066-USA Tedo
Filter area: 2.0 m²

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

Test-nr.	Date	Test-time	Type conc.	Vessel pres.	Pres CFA	Pres-CDA	Disc-speed	Feed-flow	Ash Lab	FS Lab	FS Calc.	Part. < 63µ	Feed-temp.	Air Compr.	Air filtrate	filtrate flow	filtrate	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 thicknes
	DD/MM/YY	hh:mm		Bar	Bar	Bar	rpm	m³/h	%TS	%TS	%TS	%TS	°C	Nm³/h	Nm³/h	m³/h	%DS	Nm3/h	Nm3/h	l/h		%	g/t	sec.	cycle	kg	kg/m²h	lb/ft²h	%	Nm³/t	cfm/ton	mm
Filter Feed																																
##220	07/11/1994	14:00	28/0	5.0	0.0	0.0	1.27	6.4		22.2	24.6		22	241	360	4.79	0.1	140	220	315	none	0.2	347	60	1	34.3	906	186	12.4	121	65	15.0
##221	07/11/1994	14:20	28/0	3.5	0.0	0.0	1.26	5.8		21.8	24.8		22	223	310	4.36	0.1	100	210	291	none	0.2	348	58	1	30.8	834	171	13.0	126	67	15.0
CFA=55 degree Filter Feed																																
comp. to ##218-##221																																
##222	07/11/1994	14:45	28/0	5.0	0.0	0.0	1.26	6.3		21.6	27.3		22	319	400	4.60	0.1	140	260	0	none	0.0	0	59	1	36.4	966	198	13.0	135	72	17.0
CFA=55 degree Filter Feed																																
Polymer Nalco anionic 9810																																
##223	08/11/1994	10:15	28/0	3.5	0.0	0.0	1.23	10.2		29	28.7		26	453	549	7.30	0.0	100	449	64	none	0.1	17	40	1	48.7	1841	377	16.0	122	65	28.0
##224	08/11/1994	10:35	28/0	3.5	0.0	0.0	1.22	8.8		26.3	27.3		27	381	472	6.43	0.0	100	372	27	none	0.1	8	40	1	42.8	1629	334	15.4	114	61	23.0
##225	08/11/1994	11:00	28/0	5.0	0.0	0.0	1.24	10.1		28.3	26		28	438	513	7.57	0.0	140	373	31	none	0.1	8	42	1	51.7	1904	390	14.8	98	52	26.5
##226	08/11/1994	11:25	28/0	5.0	0.0	0.0	1.24	9.4		28.8	24.8		28	403	477	7.04	0.0	140	337	14	none	0.1	4	41	1	45.9	1729	354	14.2	97	52	26.0
CFA=55 degree Filter Feed																																
comparison to ##223- ##226 and ##228- ##231																																
##227	08/11/1994	11:50	28/0	5.0	0.0	0.0	1.24	6.7		27	26.1		28	378	429	4.99	0.1	140	289	0	none	0.0	0	42	1	33	1231	252	13.6	117	63	18.0
CFA=55 degree Filter Feed																																
Anionic Surfactant																																
##228	08/11/1994	14:00	28/0	5.0	0.0	0.0	1.22	6.7		28.5	25.1		28	399	448	5.00	0.1	140	308	80	none	0.2	66	41	1	32	1208	248	13.4	128	68	20.0
##229	08/11/1994	14:15	28/0	5.0	0.0	0.0	1.22	6.3		26.8	17.8		28	415	455	5.17	0.1	140	315	302	none	0.2	277	42	1	29.2	1091	224	13.3	144	77	15.0
##230	08/11/1994	14:35	28/0	3.5	0.0	0.0	1.22	5.7		24.4	20.3		28	386	424	4.53	0.1	100	324	272	none	0.2	278	40	1	25.2	980	201	13.6	165	88	17.0
##231	08/11/1994	14:55	28/0	3.5	0.0	0.0	1.22	6.0		22.8	30.5		28	351	399	4.16	0.1	100	299	67	none	0.2	63	40	1	27.6	1071	219	13.8	140	74	18.0
CFA=55 degree Deslimed Filter Feed																																
Anionic Surfactant 0.2%+CuCl2*2 H²O 0.1%																																
##232	09/11/1994	13:44	28/0	5.0	0.0	0.0	1.26	6.6		25.4	28.4		26	378	429	4.71	0.1	140	289	78	none	0.2	68	41	1	30.5	1161	238	13.3	125	66	20.0
##233	09/11/1994	14:06	28/0	3.5	0.0	0.0	1.26	5.9		22.9	30.1		26	326	388	4.09	0.1	100	288	70	none	0.2	67	40	1	26.9	1032	212	13.9	140	74	16.0
Anionic Surfactant 0.6%+CuCl2*2 H²O 0.1%																																
##234	09/11/1994	14:30	28/0	3.5	0.0	0.0	1.25	5.6		22.8	28.2		26	334	406	4.05	0.1	100	306	67	none	0.6	207	40	1	25.2	972	199	13.7	157	84	16.5
##235	09/11/1994	14:30	28/0	5.0	0.0	0.0	1.24	5.9		26.8	26		26	404	460	4.39	0.1	140	320	71	none	0.6	199	41	1	28	1068	219	13.1	150	80	16.0
CFA=55 degree Deslimed Filter Feed																																
comp. to ##232- ##235																																
##236	09/11/1994	15:15	28/0	5.0	0.0	0.0	1.24	5.9		24.3	31.9		26	370	425	4.05	0.1	140	285	0	none	0.0	0	41	1	26.3	997	204	13.0	143	76	16.0
CFA=125 degree Filter Feed																																

Table A (continued)

Data HBF-Trailer

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

Material: Coal concentrate
filter fabric 28F-07-1020-k066-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	FS Calc.	Part. < 63µ	Feed-temp.	Air Compr.	Air filtrate	filtrate flow	filtrate	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	%DS	Nm3/h	Nm3/h	l/h		%	g/t	sec.	cycle	kg	kg/m²h	lb/ft²h	%	Nm³/t	cfm/ton	mm
Filter Feed																																
Standard Test																																
##237	16/11/1994	09:20	28/0	5.0	0.0	0.0	1.25	11.0		29	29.1		30	412	485	7.77	0.0	317	169	0	none	0.0	0	42	1	54.4	1971	404	15.3	42	23	27.4
##238	16/11/1994	10:45	28/0	3.5	0.0	0.0	1.97	12.2		29.4	30.1		30	344	438	8.54	0.1	300	138	0	none	0.0	0	40	1	59.9	2258	463	16.2	31	16	22.0
##239	16/11/1994	12:25	28/0	3.5	0.0	0.0	0.48	7.4		23.9	31.1		30	318	378	5.08	0.0	300	78	0	none	0.0	0	41	1	35.7	1328	272	16.2	29	16	46.0
##240	16/11/1994	13:35	28/0	2.0	0.0	0.0	1.24	8.4		24	32.1		30	258	331	5.73	0.1	268	63	0	none	0.0	0	38	1	39.8	1562	320	16.6	20	11	22.7
CFA = 125 degree Filter Feed																																
Test with Cetyl Pyridinium																																
##241	16/11/1994	15:15	28/0	5.0	0.0	0.0	1.24	11.3		27.4	24.9		28	344	401	8.50	0.0	300	101	566	none	0.2	279	40	1	53	2032	417	14.7	25	13	28.5