1 of 3

**ENGINEERING DEVELOPMENT OF
SELECTIVE AGGLOMERATION**

FINAL REPORT

APRIL 1993

**U.S. DEPARTMENT OF ENERGY
CONTRACT NO. DE-AC22-89PC88879**

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ACKNOWLEDGMENTS

This project is funded primarily by the U.S. Department of Energy (DOE Contract No. DE-AC22-89PC88879) and cofunded by the Electric Power Research Institute (EPRI RP3118-01).

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Acronyms and Abbreviations

ABD (Availability Block Diagram)
AF (Availability Factor)
ARC (Alberta Research Council)
ARCI/CT (Acid Rain Control Initiative/Compliance Technology)

BOD (biochemical oxygen demand)
Btu (British thermal unit)

C (carbon)
cal (calorific)
calc (calculated)
CDTF (Component Development Test Facility)
COD (Chemical Oxygen Demand)
conc (concentration)
CQDC (Coal Quality Development Center)
CRSCS (coal receiving, storage, and crushing system)
CTC (Coal Technology Corporation)
cum (cumulative)

db (dry basis)
dcb (dry coal basis)
DOE (Department of Energy)

EAF (Equivalent Availability Factor)
EPRI (Electric Power Research Institute)

fcbl (feed coal basis)
FCV (flow control valve)
FLOF (Full Load Operation Factor)
flot (flotation)
flt (flotation)

ft (foot)

gal (gallon)
gpm (gallon per minute)

h (hour)
H (hydrogen)
HGI (Hardgrove Grindability Index)
HM (heavy media)
hp (horse power)
HS (high shear)
HVAC (heating ventilation air conditioning)

JWR (Jim Walter Resources, Inc.)

Acronyms and Abbreviations (Cont'd)

Kgal (1000 gallons)

Kwh (kilowatt hour)

Kva (kilovolt amp)

lb (pound)

lbm (pound mass)

m (meter)

M (mesh)

MAF (moisture and ash free)

MACRS (modified accelerated cost recovery system)

MBtu (million Btu)

MIBC (methyl iso-butyl carbinol)

min (minute)

mfc (mass flow controller)

mm (millimeter)

moist (moisture)

MTBF (mean time between failures)

MWK (M.W. Kellogg)

N (nitrogen)

NERC/GADS (North American Electric Reliability Council/Generating Availability Data System)

O (oxygen)

O&M (operations & maintenance)

OS (organic sulfur)

PETC (Pittsburgh Energy Technology Center)

PCS (process control system)

PGNAA (prompt gamma neutron activation analysis)

POC (proof-of-concept)

ppm (parts per million)

PS (pyritic sulfur)

PSD (particle size distribution)

psig (pounds per square inch gauge)

PSR (pyritic sulfur rejection)

PTI (Process Technology, Inc.)

pyr (pyrite)

rec (recovery)

rej (rejection)

ROM (run of mine)

rpm (revolutions per minute)

S (sulfur)

SDOE (sequential design of experiments)

SCFM (standard cubic feet per minute)

SCS (Southern Company Services, Inc.)

SEI (Southern Electric International, Inc.)

S/F (sink/float)

sg (specific gravity)

sol (solids)

SWPP (Stormwater Pollution Prevention Plan)

Acronyms and Abbreviations (Cont'd)

t (short ton)

TAG (EPRI Technical Assessment Guide)

t/h (tons per hour)

TOC (total organic carbon)

TS (total sulfur)

TSR (total sulfur rejection)

µm (micron)

U.S. (United States)

vac (vacuum)

val (value)

VFBD (vibrating fluidized-bed dryer)

VPI (Virginia Polytechnic Institute and State University)

wt (weight)

WWT (waste water treatment)

y (year)

EXECUTIVE SUMMARY

Scope of Work

This report presents the findings of the project entitled "Engineering Development of Selective Agglomeration." In 1989 the U.S. Department of Energy contracted with Southern Company Services, Inc. (DOE Contract No. DE-AC22-89PC88879) to develop selective agglomeration technology to a commercially acceptable level by 1993. This project is part of DOE's program to advance the state of physical coal cleaning technologies in order to accelerate the utilization of high-sulfur coals while complying with environmental regulations. Such projects assume added importance in light of the 1990 Clean Air Act Amendments. Appropriate utilization of the abundant reserves of high-sulfur coal in the United States can make a significant contribution to achieving the goal of energy independence.

This work was conducted by a multi-organization team headed by Southern Company Services, Inc. (SCS) who provided management, detailed design, and cost sharing. The Alberta Research Council (ARC) provided process development and cofunding. Southern Electric International, Inc. (SEI) was responsible for Component Development Test Facility (CDTF) and Proof-of-Concept (POC) facility construction and operations. Praxis Engineers, Inc. (Praxis) provided process engineering. The Electric Power Research Institute (EPRI) provided significant cofunding as well as expertise in coal cleaning and coal markets.

In addition to this final report, the following three companion reports have also been issued:

- A stand-alone "Executive Summary" of this final report
- A "Trace Element Removal Study" which reports on the effectiveness of the selective agglomeration technology in removing certain trace elements from coal prior to combustion that are listed as hazardous

air pollutants in the 1990 Clean Air Act Amendments, Title III

- A "Site Closecut Report" which addresses all state and federal regulations encountered in closing the selective agglomeration testing site in Wilsonville, Alabama

The selective agglomeration process, as applied to coal cleaning, is based on differences in the surface chemistry of coal and its associated impurities. Coal particles are hydrophobic (i.e., repelled from water) while the majority of its impurities are hydrophilic (i.e., stabilized in water). During selective agglomeration, a liquid (the agglomerant) that is immiscible with water is introduced into a coal-water slurry which is then mechanically agitated with sufficient intensity to disperse the agglomerant, thereby allowing it to come in contact with all particles in the slurry. This unit operation is referred to as high-shear agglomeration. The coal particles, due to their hydrophobic nature, are attracted to the agglomerant phase. The hydrophilic mineral impurities remain in the water phase. Continued high-shear agitation of the agglomerant-coated coal particles causes them to coalesce to form microagglomerates.

The size of the microagglomerates formed during this process is primarily a function of the agglomerant concentration utilized and the agglomeration residence time. Generally, as the agglomerant concentration or residence time are increased—with all other parameters remaining constant—the resulting microagglomerates increase in size.

Microagglomerates formed via high-shear agglomeration are then separated from the mineral matter-bearing aqueous phase. The manner in which this separation is achieved primarily depends on the size of the microagglomerates. If large enough, the microagglomerates can be separated by screening; otherwise, froth flotation of the microagglomerates is required to achieve the desired separation.

Following separation of the microagglomerates from the mineral matter-bearing aqueous phase, subsequent processing steps may be required to achieve a final product compatible with conventional coal handling systems. Larger agglomerates (>1 mm diameter) can be formed by adding a finely divided liquid binder, such as an asphalt/diesel blend, followed by continued low-shear mechanical agitation. These larger agglomerates are then mechanically dewatered. In lieu of low-shear agglomeration, the recovered microagglomerates can be mechanically dewatered and then pelletized using the appropriate binder to form a product of the desired size. Typically, the product from either of these final product formation processes is thermally dried to meet product moisture requirements and increase the agglomerate/pellet strength.

Project performance targets for the selective agglomeration process were to achieve 85% or greater Btu recovery at 85% or greater pyritic sulfur rejection (PSR) on a run-of-mine (ROM) coal basis, while producing a final clean-coal product with an ash content of 6% or less which is handleable by conventional coal handling systems. Engineering development of selective agglomeration included:

- Batch and continuous bench-scale process development testing
- Continuous pilot-scale (3-t/h) component development testing to evaluate the adaptation and/or modification of existing unit operations for selective agglomeration
- Continuous pilot-scale (2-t/h) POC testing to optimize the selective agglomeration process and demonstrate process reliability
- Vendor testing to evaluate pelletization and thermal drying processes as applied to selective agglomeration
- Conceptual design of a commercial-scale (200-t/h) selective agglomeration facility
- Economic analysis of the selective agglomeration process at a commercial (200-t/h) scale

The information obtained from the various modes of testing and analysis, particularly POC operations, resulted in a technical and economic design base sufficient to support construction and operation of a commercial plant.

Phase I Testing Summary

During Phase I of the project, a batch testing program was carried out by ARC and SEI. Various agglomerating liquids, ranging from heavy oils such as Maya crude to light hydrocarbons like hexane, were evaluated in this program. Initial testing indicated that diesel (considered a heavy agglomerant) and heptane (a light agglomerant) were the most appropriate high-shear agglomerants, and they were therefore investigated further in two parallel-path test programs.

The diesel process tested utilized high-shear agglomeration with diesel (approximately 1% on a dry coal basis) followed by two-stage flotation recovery of microagglomerates. The recovered microagglomerates were then either subjected to low-shear agglomeration using various blends of diesel/asphalt as a binder or mechanically dewatered and pelletized using dry-milled flour as a binder and asphalt for weatherproofing. These enlarged products were then thermally dried to produce the final product. Evaluation of these two product formation methods determined that the pelletization process provided a more handleable product at a lower cost, and it was therefore selected for further investigation.

The heptane process tested utilized heptane (approximately 20-30% on a dry coal basis) during high-shear agglomeration. Following high shear, microagglomerates were recovered by screening. These microagglomerates were then processed using low-shear agglomeration with a diesel/asphalt blend as a binder. The low-shear product was then thermally dried to recover the heptane and produce the final product.

During Phase I batch testing, Btu recoveries of approximately 90% at 75-80% PSR were achieved for all three test coals for both the diesel- and heptane-based processes on a ROM basis. Similar results were also achieved in 5-kg/h bench-continuous testing conducted during Phase I. While both the diesel- and heptane-based processes provided similar performance, the diesel/pelletization process was selected for investigation during Phase II of the project since it was ranked as a more mature process with fewer uncertainties. The diesel- and heptane-based processes, however, both possess inherent advantages and disadvantages.

A primary advantage of the diesel process is due to the nature and quantity of the agglomerant used. In the diesel or heavy-oil process, relatively small amounts (<1.0%) of diesel oil, a hydrocarbon of low volatility, are used. As such, the diesel does not have to be recovered

and remains in the final product. In contrast, the heptane or light-oil process utilizes significant concentrations (20-30%) of heptane, a highly volatile hydrocarbon, during high shear. The heptane cannot be allowed to remain in the final product and must be recovered in order to meet product safety and environmental standards. As a result, the entire process from the point of heptane introduction during high shear to heptane recovery by thermal dryers must be sealed and blanketed with an inert gas for safety reasons. This required safety precaution increases capital investment substantially.

During selective agglomeration, the high-shear unit operation is typically conducted at about 20% solids concentration. In the heptane process, no dilution of the high-shear product is required prior to screening to recover microagglomerates. While some wash water is required to assist in the screening separation, the amount of water needed is small compared to that used during the two stages of flotation to recover microagglomerates in the diesel process. For both stages of flotation in the diesel process, the feed is diluted to approximately 6% solids concentration to improve process performance.

An additional advantage of the light-oil process is that the screening operation is simpler, easier to control, and less expensive than froth flotation recovery of microagglomerates. It is also possible that with flotation the overall performance of the selective agglomeration process may be determined primarily by the microagglomerate recovery step rather than by the high-shear agglomeration unit operation.

During heptane agglomeration, final product formation is conducted via low-shear agglomeration using asphalt-based binders blended with additional heptane. While the asphalt is needed to impart strength to the final product, its presence impedes recovery of heptane to the required levels. In fact, Phase I testing indicated that when asphalt was present in the low-shear product, the quantity of heptane remaining on the final product after thermal recovery was environmentally unacceptable. In addition, the final product from the heptane process is not as strong and handleable as that from the diesel/pelletization process.

Pilot-scale (3-t/h) testing to evaluate low-shear agglomeration conditions and agglomerate dewatering methods for the diesel process was carried out during Phase I CDTF testing. It was determined during CDTF testing that a low-shear binder concentration of 13% was required to form 1-mm agglomerates with a 100-mesh topsize coal feed. Screen-bowl centrifuge dewatering of

these agglomerates achieved a product moisture content of about 14%. A similar dewatering performance was also achieved for much smaller agglomerates formed using approximately 6% binder.

A major finding of Phase I testing was that a ROM coal should be cleaned conventionally prior to agglomeration. It was found that better process performance was achieved when the project coals were cleaned prior to agglomeration than when raw coals were agglomerated directly. This conclusion is not unexpected since during conventional precleaning impurities are rejected at coarse sizes by gravity techniques with relatively high separation efficiencies. In contrast, agglomeration, a surface-based process, is carried out at fine coal sizes where separation efficiencies are lower.

Phase II Testing Summary

The primary focus of Phase II of the project centered on POC testing of the selected diesel/pelletization agglomeration process at a 2-t/h scale. In addition to POC testing, supporting bench-scale batch agglomeration tests were also completed. The following four feedstocks were tested during Phase II:

- Precleaned Upper Freeport coal from Jefferson County, Pennsylvania, obtained from P&N Coal Company and precleaned by CQ, Inc.
- Precleaned Pittsburgh No. 8 coal from Greene County, Pennsylvania, obtained from and precleaned by Cyprus Coal Company's Emerald mine preparation plant
- ROM Illinois No. 5 coal from Wabash County, Illinois, provided by AMAX Coal Company from their Wabash mine
- Reclaimed pond fines (minus 60-mesh fraction) from Jefferson County, Alabama, provided by Jim Walter Resources (JWR) from their No. 3 mine tailings pond

A photograph of the POC pilot-plant facility that was located at the SEI site in Wilsonville, Alabama, is shown in Figure 1. Figure 2 provides a simplified unit-operations block diagram of the POC facility.

During Phase II of this project, POC testing with Upper Freeport and Pittsburgh No. 8 coals was conducted using a factorial test plan design. However, determination of the effects of individual high-shear parameters on overall process performance proved to be

Figure 1. POC Facility, Wilsonville, Alabama

Figure 2. POC Facility Block Diagram

difficult since the two sequential unit operations, high shear and flotation, are coupled. Flotation dominates the process because of its sensitivity to process variables such as coal hydrophobicity and agglomerant concentration in addition to flotation equipment variables such as air flow rates and cell water levels.

POC testing showed that strongly hydrophobic coals such as Upper Freeport are sensitive to excess agglomerant addition. Diesel addition levels to high shear of 0.5% and above destabilized the air bubbles in the froth as the bubbles rose toward the surface of the flotation cell. During typical flotation froth migration, air bubbles coalesce into larger bubbles but are retained in the froth, thus assisting in drainage of process water from the flotation product.

During POC testing with Upper Freeport coal, the destabilization of air bubbles resulted in greater carryover of process water to the product and a lower froth product solids content than desired. The net effect of this phenomenon was increased product contamination by refuse particles entrained in the process water and increased difficulty in controlling coal recovery to a constant level.

In contrast, this phenomenon did not occur during testing with coals of decreasing hydrophobicity, such as Pittsburgh No. 8 and Illinois No. 5. These two coals could be agglomerated over a wide range of oil concentrations and still maintain acceptable froth properties, with corresponding improvements in flotation performance. Observations during POC operation indicate that as coal hydrophobicity decreases, the range of agglomerant concentration providing good froth

properties along with good microagglomerate recovery is as follows for each coal:

- Upper Freeport (very hydrophobic): <0.3%
- Pittsburgh No. 8 (hydrophobic): 0.125-0.5%
- Illinois No. 5 (weakly hydrophobic): 0.25-1.2%

These observations indicate that evaluation of coal-specific parameters such as hydrophobicity can help determine the range of high-shear agglomerant concentrations that will provide good froth properties during microagglomerate recovery via flotation.

Considering the high-shear and flotation unit operations as a single process, agglomerant and frother concentrations in combination control coal recovery. At any given agglomerant concentration there is an optimal frother level that will achieve the desired recovery, providing good froth structure is maintained. If the agglomerant concentration is reduced, coal recovery decreases unless frother levels are increased correspondingly. If the agglomerant concentration is increased beyond the ranges previously discussed, the presence of excess oil adversely affects the froth structure regardless of frother concentration.

Optimization of a three-component system in series (high shear followed by two stages of flotation) is difficult and requires a large number of tests to complete a factorial design. Sequential design of experiments appears to be a promising approach for optimization of these types of problems. The use of tools such as Ultramax process optimization software allows determination of the effects of several operating parameters simultaneously. This reduces the total number of experiments required to evaluate the effects of all process variables of interest over the desired ranges. For parametric testing of the Illinois No. 5 coal, EPRI made available a pre-release version of the Ultramax software program to assist in the optimization of the selective agglomeration process. This product is under co-development by EPRI and Ultramax Corporation, an industrial partner.

During Phase II POC operations, parametric testing to determine optimal operating conditions was performed for each of the three primary project coals: Upper Freeport, Pittsburgh No. 8, and Illinois No. 5. Following parametric testing of each coal, an extended demonstration run (approximately four days' duration) was conducted for each coal at selected optimal operating conditions.

Bench-scale batch agglomeration tests were completed in support of POC testing. Comparison of laboratory batch and continuous POC results for all coals tested indicates that similar performance was achieved by both modes of testing. This similarity between batch and continuous test results indicates that a well-planned series of batch agglomeration tests can provide much information to predict the performance of continuous-mode testing.

Based on the analysis of POC and batch testing results, grade/recovery curves characterizing selective agglomeration performance for each coal were developed, as shown in Figure 3. Also shown in Figure 3 are the grade/recovery curves predicted by washability analyses of POC high-shear agglomeration feed samples.

Figure 3. Phase II Testing and Washability Grade/Recovery Curves

Based on the grade/recovery curves presented in Figure 3, the ROM coal-basis PSR at 85% Btu recovery for each coal is characterized in Table 1.

Table 1
Phase II Testing and Washability
PSR at 85% Btu Recovery

Coal	Topsize	Testing PSR, %	Washability PSR, %
Upper Freeport	200 mesh	84	96
Pittsburgh No. 8	200 mesh	68	90
Illinois No. 5	100 mesh	64	80

As is evident in this table, consistently lower PSR levels were achieved in the agglomeration tests than in the centrifuge sink/float washability analyses. Agglomeration grade/recovery curves generally parallel the washability curves, but at lower PSR. At 85% Btu recovery, the differences between washability and test-result PSR are approximately 12, 22, and 16% for the Upper Freeport, Pittsburgh No. 8, and Illinois No. 5 coals, respectively.

These significant differences in PSR between agglomeration and washability results may be attributed to a number of factors. Fine coal washability results are based on gravity separations in which all particles of specific gravity greater than 1.8 report to the refuse and are therefore considered rejected. The centrifuge float/sink washability procedure utilized requires the use of a dispersant and a 20-minute residence time at a centrifuge rpm of 1500 to ensure that a good gravity separation is achieved at a 200-mesh coal topsize. In contrast, during agglomeration testing, a coal surface-based separation is made during flotation recovery of microagglomerates. Since so much care is required to guarantee a good laboratory float/sink separation at these fine coal sizes, it is not unexpected that actual batch and continuous agglomeration test results indicate poorer performance than washability predictions.

During these sink/float separations, many "locked" or unliberated particles, i.e., particles which are part carbon and part mineral matter, report to the refuse based on overall particle specific gravity. During agglomeration, however, locked particles may report to the product since any coal on the particle surface will be coated with agglomerant.

This reporting of unliberated, predominantly mineral matter-bearing particles to the product can be reduced by additional grinding to further liberate impurities from the coal. Finer grinding will, however, affect overall process costs as follows:

- Increased grinding energy costs
- Increased high-shear agglomerant requirements due to the greater coal surface area
- Increased mechanical dewatering costs to achieve the same flotation product moisture content

The use of very low (or starvation) quantities of agglomerant may help curtail the recovery of locked particles by reducing the oil available to coat the carbonaceous portions of these particles.

An additional explanation for the difference in PSR levels between the agglomeration test and washability data is the contamination of the flotation product by mineral matter carried in the water in the flotation froth. Recovery of microagglomerates by flotation results in a strong froth which can entrap ash and pyrite particles suspended in the water between the air bubbles in the froth.

One method used to prevent mineral-matter entrapment is to reduce the solids concentration in the flotation feed. A reduction in the feed solids concentration produces a corresponding reduction in the mineral-matter concentration in the flotation cells and therefore reduces the amount of impurities carried to the product with each unit of water contained in the froth. A second method practiced in other industries is to apply water to the froth as it forms on top of the flotation cells, thereby replacing the mineral matter-bearing process water in the froth with clean water.

In general, while washability data provide information about the liberation characteristics of a given coal, they cannot be used to predict actual agglomeration performance for the reasons discussed above. Prediction of continuous-mode agglomeration performance at any grind size requires, at a minimum, batch agglomeration testing which can then be extrapolated using available washability data.

Figure 4 presents a comparison of the average performance obtained during the demonstration runs for each of the three primary project coals with the grade/recovery curves presented in Figure 3.

As can be seen in Figure 4, the average POC demonstration run performance approximates the grade/recovery curve for each coal tested. Therefore, for the Upper Freeport and Pittsburgh No. 8 coals, the average demonstration run performance was used as the basis for the commercial-scale (200-t/h) plant design.

In the case of the Illinois No. 5 coal, however, POC results were not used for the commercial plant design since the POC feedstock represented the entire raw coal and the commercial design was completed for the minus 100-mesh natural fines fraction only. This natural fines agglomeration scenario was evaluated because AMAX Coal Company, the supplier of the Illinois No. 5 coal, expressed an interest in using selective agglomeration to clean the minus 100-mesh material which is currently bypassed to tailings in the design of a 1500-t/h cleaning plant now under construction. To evaluate agglomeration for this application, laboratory batch tests were

Figure 4. Characterization of Selective Agglomeration POC Performance

completed on minus 100-mesh natural fines samples of the POC feedstock. The results of these tests, along with washability data for these natural fines, are presented in Figure 5.

Figure 5. Illinois No. 5 Natural Minus 100-Mesh Fines Batch Results

These data show that batch tests using the natural minus 100-mesh fines achieved 79% PSR at 85% Btu recovery. This indicates 15% greater PSR at the same Btu recovery than POC agglomeration of the entire raw coal. This 15% increase in PSR approximates the difference in liberation indicated by washability data for the two Illinois No. 5 coal samples.

The selective agglomeration performance values used for the commercial-scale design for the three project coals are presented in Table 2.

Table 2
Selective Agglomeration Performance for Commercial-Scale Design

Coal	Topsize	Btu Rec, %	PSR, %
Upper Freeport	200 M	89.7	81.0
Pittsburgh No. 8	200 M	88.2	64.4
Illinois No. 5	100 M (natural fines)	85.0	79.0

For the fourth project feedstock tested, reclaimed pond fines from JWR No. 3 mine, both laboratory batch and POC tests achieved approximately 30% yield from the minus 60-mesh fraction while making a 15% ash product. This performance was achieved in batch and continuous testing both with and without the use of high shear prior to flotation. This indicates that for the JWR pond fines no benefit was observed from the use of agglomeration. It should be noted, however, that the benefits of agglomeration appear to be coal-dependent and for many tailings pond materials, especially those that are oxidized, agglomeration could provide superior process performance to recovery by froth flotation alone.

Pelletization of dewatered selective agglomeration products was shown to provide a "green" (wet) pellet approximately 6 mm in diameter which had sufficient strength to withstand transfer via belt conveyors. These pellets were formed using approximately 2-2.5% starch binder for strength and 0.5% asphalt binder for weatherproofing. Optimal feed moistures to achieve these results were approximately 25% and 27% for 100- and 200-mesh topsizes, respectively. Thermal drying of "green" pellets was completed in a vibrating fluidized-bed dryer at residence times of 2 to 2-1/2 minutes to achieve a moisture content of less than 4%. The resulting dried pellets are considered satisfactory for conventional coal handling facilities. Combustion testing of both mechanically dewatered agglomerates and dried pelletized products was conducted by DOE for each of the three primary project feedstocks. The results of these combustion tests were unavailable at the time of publication.

The pelletized and thermally dried product agglomerates produced from each of the three primary project test coals were characterized. The characterization data are presented in Table 3. These data show that the final thermally dried agglomerated product is low in both moisture and ash content, which indicates potential

savings during transportation of this fuel. Additional savings would be realized during power plant pulverization due to the very high grindabilities of these products. This savings would help offset the initial cost of fine grinding prior to agglomeration.

Table 3
Final Product Analysis (Dry Basis)

	Upper Freeport	Pittsburgh No. 8	Illinois No. 5
<u>Proximate Analysis</u>			
Moisture, % (as is)	0.95	1.90	2.53
Ash, %	5.56	5.28	4.72
Volatile matter, %	33.22	38.74	39.19
Fixed carbon, %	61.22	55.98	56.09
<u>Forms of Sulfur</u>			
Total sulfur, %	1.08	1.75	1.38
Pyritic sulfur, %	0.05	0.60	0.48
Sulfate sulfur, %	0.39	0.12	0.17
Inorganic sulfur, %	0.44	0.72	0.65
Organic sulfur, %	0.64	1.03	0.73
lb SO ₂ /MBtu	1.51	2.46	2.03
Btu/lb	14274	14209	13618
Hardgrove index	177	197	150

Commercial-Scale Conceptual Design

Selective agglomeration performance throughout this project has been defined by process Btu recovery and PSR calculated on a ROM basis. For POC feedstocks which were precleaned prior to agglomeration, ROM performance is based on conventional cleaning performance and subsequent agglomeration results. Thus, the conceptual design for the commercial-scale 200-t/h plant based on the Upper Freeport and Pittsburgh No. 8 coals includes conventional cleaning of the ROM coal followed by selective agglomeration of the precleaned product. This plant has been scheduled to operate 24 hours per day, 7 days per week, and 50 weeks per year. To meet this operating schedule of 8400 hours per year, excess capacity has been included in the design where necessary, along with all required spare parts.

The 200-t/h commercial agglomeration plant conceptual design for the Upper Freeport and Pittsburgh No. 8 coals is based on the average POC demonstration run performance as presented previously in Figure 4 and Table 2. A conceptual design of the same plant at a 20-t/h semiworks scale was also completed.

For the Illinois No. 5 coal, the commercial plant design is for a retrofit application to agglomerate the minus 100-mesh natural fines which are currently bypassed to plant tailings in the design of a 1500-t/h coal cleaning facility now under construction by AMAX Coal Company. In this design, the agglomerated natural fines are mechanically dewatered and blended with the conventionally cleaned coarse coal fractions. Scheduled operating hours for the design of this retrofit circuit are 5400 hours per year.

For the Illinois No. 5 coal, the commercial-scale 1500-t/h plant design for the plus 100-mesh coal is based on washability data provided by AMAX Coal Company. Commercial-scale design for the natural minus 100-mesh ROM coal fraction is based on the Phase II batch testing performance at 85% Btu recovery achieved using this natural fines fraction and presented previously in Figure 5 and Table 2.

Commercial-Scale Plant Reliability Analysis

The reliability analysis for both the 200- and 20-t/h plant designs is based on the following three criteria:

- Availability Factor (AF): The expected percentage of scheduled operating hours that the plant is in service
- Equivalent Availability Factor (EAF): The expected percentage of the design coal processing rate at which the plant will operate
- Full Load Operation Factor (FLOF): The expected percentage of scheduled operating hours at which the plant operates at the designed capacity

Table 4 presents the availability factors and expected operating rates for the conceptual designs of both the 20- and 200-t/h selective agglomeration plants.

Table 4
200-t/h Conceptual Design Availability Results

	20 t/h	200 t/h
Availability factor (AF)	95.4%	97.2%
Expected operating hours per year	8010	8164
Equivalent availability factor (EAF)	95.0%	95.2%
Expected coal processing rate (t/h)	19	190
Full load operating factor (FLOF)	93.4%	90.0%
Hours per year at design capacity	7843	7475

For reliability analyses completed on the Illinois No. 5 coal plant design, the EAF of the conventional coal cleaning portion of the plant was assumed to be 90%.

Using this assumption and reliability analysis of the natural fines agglomeration circuit, reliability results for this commercial-scale (1500-t/h feed) plant design are presented in Table 5.

Table 5
Natural Fines and Overall Plant Reliability

Equipment	EAF	Product t/h
Conventional coal cleaning only	90.0%	1083
Agglomeration process only	98.5%	69
Overall plant	90.5%	1152

Commercial-Scale Plant Economic Analysis

Economic analyses completed on the commercial-scale agglomeration plant for the Upper Freeport and Illinois No. 5 coals were based on the assumptions presented in Table 6.

Table 6
Commercial Plant (200 t/h) Economic Assumptions

Actual operating hours	7,560
Plant availability	90%
Production rate	200 t/h product
Plant life	20 years
Royalties	None
Financing	100% equity
Inflation rate	4.0%
Corporate tax rate	38.0%
Rate of return	10.0%

Based on these assumptions and the actual plant design, the annualized product costs for both coals are presented in Table 7. These costs represent total plant processing costs including capital investment, operation and maintenance costs, and Btu losses incurred during processing. However, these annualized product costs do not include the cost of the ROM coal. In this table, costs are presented for a pelletized and thermally dried product containing 4% moisture, a nonpelletized product thermally dried to 8% moisture, and a nonpelletized mechanically dewatered product containing approximately 25% moisture.

Sensitivity analyses varying the rate of return, operating hours, and plant size were performed for this 200-t/h product commercial plant design. These analyses reveal that:

Executive Summary

- An increase in the rate of return from 10% to 20% increases product costs by 20%.
- A decrease in the actual annual operating hours from 7560 to 5400 increases product costs by 16%.
- An increase in plant size from 200 t/h to 500 t/h decreases product costs by 10%.

Table 7
Commercial Agglomeration Annualized Product Cost Summary

	Upper Freeport	Pittsburgh No. 8
Pelletized Product, 4% Moisture		
\$/ton product	24.73	25.32
\$/million Btu	0.84	0.87
\$/ton SO ₂ removed	333.56	488.08
Nonpelletized Product, 8% Moisture		
\$/ton product	17.70	18.28
\$/million Btu	0.60	0.63
\$/ton SO ₂ removed	238.66	352.41
Nonpelletized Product, 25% Moisture		
\$/ton product	14.08	14.72
\$/million Btu	0.48	0.51
\$/ton SO ₂ removed	189.95	283.79

The cost of cleaning minus 100-mesh natural fines using an agglomeration circuit retrofit in a conventional cleaning plant is shown in Table 8. The natural fines circuit costs represent total processing costs including capital investment and operation and maintenance costs. Added cost for Btu losses incurred during natural fines

processing are not included since this circuit is a retrofit application to a plant which bypasses the fines to tailings. Annualized product costs for the entire plant are based on natural fines circuit product costs and an assumed value of \$2/ton product for the plant conventional cleaning circuits.

Table 8
Natural Fines Annualized Product Cost Summary

Fines Circuit	
\$/ton product	\$12.62
\$/million Btu	\$0.49
\$/ton SO ₂ removed	\$243.24
Entire Plant	
\$/ ton product	\$2.58
\$/million /Btu	\$0.10
\$/ton SO ₂ removed	\$74.68

The retrofit selective agglomeration fines recovery circuit produces approximately 378,000 tons/year of product generating additional revenue. Assuming an energy value of \$0.80 per million Btu, an after-tax profit of \$2.35 million per year is realized. This annual profit does not include credit for refuse handling cost savings such as reduced capital equipment investment, reduced operating costs, and lower ponding capital and operating costs.

Based on these revenue projections, the rate of return for the retrofit selective agglomeration fines circuit is 14.5% with a payback period of 6.5 years from startup.

INTRODUCTION

Introduction

This report presents the findings of the project entitled "Engineering Development of Selective Agglomeration." The U.S. Department of Energy contracted with Southern Company Services, Inc. in 1989 (DOE Contract No. DE-AC22-89PC88879) to develop selective agglomeration technology to a commercially acceptable level by 1993. This project is part of DOE's program to advance the development of physical coal cleaning technologies in order to accelerate the utilization of high-sulfur coals while complying with environmental regulations. Such projects assume added importance in light of the 1990 Clean Air Act Amendments. Appropriate utilization of the abundant reserves of high-sulfur coal in the United States can make a significant contribution to achieving the goal of energy independence.

Engineering development included bench-scale process development, component development, adaptation or modification of existing unit operations, proof-of-concept (POC) module design, fabrication, testing, data evaluation, and conceptual design of a commercial facility. The information obtained during POC operation resulted in a technical and economic design base sufficient to support construction and operation of a commercial plant.

Throughout this project, performance targets for the engineering development of selective agglomeration process were to achieve 85% or greater Btu recovery at 85% or greater pyritic sulfur rejection (PSR). Additional objectives included producing a final clean-coal product with an ash content of 6% or less which is suitable for conventional coal handling systems.

The selective agglomeration process, as applied to coal cleaning, is based on differences in the surface chemistry of coal and its associated impurities. Coal particles are hydrophobic (i.e., repel water) while the majority of its impurities are hydrophilic (i.e., stabilized in water). During selective agglomeration, a liquid (the agglom-

erant) that is immiscible with water is introduced into a coal-water slurry and agitated to disperse it in the slurry, thereby allowing it to come into contact with all particles in the slurry. The coal particles, due to their hydrophobic nature, are attracted to the agglomerant phase. The hydrophilic mineral impurities remain in the water phase. Continued agitation of the agglomerant-coated coal particles causes them to coalesce to form agglomerates. Once the agglomerates are formed, they are separated from the mineral matter-bearing aqueous phase by subsequent processing steps.

A major advantage of agglomeration over other coal cleaning technologies is its ability to effectively clean very finely comminuted coal. The ability to process fine particles is important since coal must be ground to fine sizes to achieve good liberation of impurities such as pyrite.

This project was conducted by a multi-organization team headed by Southern Company Services, Inc. (SCS), who provided management, detailed design, and cost sharing. The Alberta Research Council (ARC) provided process development and cofunding. Southern Electric International, Inc. (SEI) was responsible for POC facility construction and operations. Praxis Engineers, Inc. (Praxis) provided process engineering. The Electric Power Research Institute (EPRI) provided significant cofunding as well as expertise in coal cleaning and coal markets.

Objectives of the Report

This report summarizes the work completed during the project, and includes the following elements:

- Engineering analysis of process deficiencies
- Description of equipment and procedures used
- Analysis of bench-scale test results

- Analysis of POC test results
- Economic analyses
- Reliability analyses
- Conceptual design of a 20-t/h semiworks plant
- Conceptual design of a commercial-scale plant using selective agglomeration technology

An overview of the work conducted during the first phase of the project is presented, and a comprehensive evaluation of all test data and a summary of the major project findings are given.

Scope of the Report

This document constitutes the final report for this project and thus presents a summary of all work performed during the project.

In addition to this final report, the following three companion reports have also been issued:

- A stand-alone "Executive Summary" of this final report
- A "Trace Element Removal Study" which reports on the effectiveness of the selective agglomeration technology in removing certain trace elements from coal prior to combustion which are included in the Hazardous Air Pollutants list (given in the 1990 Clean Air Act Amendments, Title III)
- A "Site Closeout Report" which addresses all state and federal regulations encountered in closing the selective agglomeration test site in Wilsonville, Alabama

Report Organization

This report is organized into four main segments. The first segment (Sections 1 and 2) provides an overall introduction and presents a summary of the first phase of the project, which involved bench-scale and component development testing.

The second segment of this report (Section 3) summarizes the POC test work conducted during Phase II and provides an analysis of the POC results. Figure 1-1 shows the POC pilot plant that was located at the SEI facilities in Wilsonville, Alabama.

The third segment (Sections 4 and 5) presents the final conceptual designs for a 200-t/h commercial agglomeration plant and a 20-t/h semiworks agglomeration plant. This segment includes the techniques required to maintain 90% plant availability, plant layouts, and equipment lists.

The fourth segment of this report (Section 6) presents the projected economics for a 200-t/h commercial plant and for a cleaning plant retrofit using selective agglomeration to process the natural fines. Several sensitivity analyses are also presented.

Figure 1-1. POC Facility

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PHASE I SUMMARY

Introduction

Phase I of the project was devoted to the identification, analysis, testing, and resolution of selective agglomeration process uncertainties. Phase I consisted of seven tasks beginning with planning and identification of process uncertainties. Batch, bench-continuous, and continuous unit operations tests were performed to resolve the uncertainties and engineering deficiencies associated with the technology. Phase I concluded with the resolution of process uncertainties and selection of a commercially viable selective agglomeration process for further development in Phase II. To further illustrate the progression of Phase I, the major results of project Tasks 1-7 are summarized below:

- Task 1:** Detailed schedule and work plan for the project developed
- Task 2:** Conceptual design for a 20-t/h selective agglomeration plant prepared
- Task 3:** Areas of uncertainty identified
- Task 4:** Test plans for batch, bench-continuous, and continuous unit operation tests developed
- Task 5:** Batch and bench-continuous tests completed. Addressed process deficiencies amenable to theoretical or analytical solutions. Project coals tested and a database for their cleaning behavior developed. Optimal methods for rejecting sulfur and ash determined.
- Task 6:** Component and unit operations development tests completed
- Task 7:** Analysis of all test results was completed and previously defined process deficiencies were resolved. The final process for development in Phase II was selected.

Phase I was successful in resolving all major process uncertainties. A viable agglomeration process was developed that achieved 90% Btu recovery at approximately 80, 75, and 80% pyritic sulfur rejection (PSR) for the Upper Freeport, Pittsburgh No. 8, and Kentucky No. 9 coals respectively. Product ash contents at these performance levels were in the range 8-10% for Upper Freeport coal, 5-6% for Pittsburgh No. 8 coal, and 6-7% for Kentucky No. 9 coal. While this data indicates good process performance for all three coals, project objectives of 85% Btu recovery, 85% PSR, and 6% product ash were not met for all three coals.

The results of batch and bench-continuous tests showed that agglomeration performance was comparable for the two modes of operation. Successful operation of the 3-t/h Component Development Test Facility (CDTF) demonstrated that there were no major operating deficiencies in the process. A viable agglomeration process representing the lowest-risk, lowest-cost option meeting all major project objectives was proposed for further development in Phase II.

Test Plan

Selective agglomeration can be visualized as consisting of two distinct steps: separation of mineral matter and pyrites from coal and final product formation. The major objective of batch testing was to obtain sufficient information to better understand the process and to resolve the various uncertainties that were identified during Task 3. The Task 5 batch test program investigated the separation of coal from mineral matter via optimization of the high-shear and flotation unit operations.

A continuous bench-scale test program was also conducted as part of Task 5. The primary objectives of this continuous-mode work were to demonstrate process comparability to batch performance, determine any serious deficiencies in the continuous operation of the proc-

ess, generate information on continuous operation, and provide support to CDTF testing.

Operating conditions for product agglomerate formation and growth were investigated in Task 6. This included low-shear mixing, evaluation of binders, and tests to determine product characteristics and their environmental implications. Additional laboratory and vendor tests were performed to characterize agglomerate product handling and drying, investigate various agglomeration parameters, and determine the effects of precleaning on agglomerate quality.

The testing performed during Tasks 5 and 6 included:

- Batch agglomeration tests performed by ARC and SEI
- Bench-scale continuous testing at ARC
- Component development tests performed by SEI at the CDTF in Wilsonville, Alabama
- Vendor testing

Task 5 Batch Testing Program

The Task 5 test program began at ARC in August 1989 with initial testing to evaluate a range of agglomerants. Channel samples of three coals, Illinois No. 6, Pittsburgh No. 8, and Upper Freeport, were obtained from another DOE-sponsored coal cleaning project (DE-AC22-88PC88878). These coals were used to evaluate the agglomerants and the effects of agglomerant dosage, and to conduct preliminary tests to investigate various microagglomerate recovery methods. The remainder of the test work was conducted on the project coals, which included the three coals mentioned above. While the Upper Freeport coal was a raw coal, the other two were precleaned at commercial coal preparation plants. However, after preliminary tests, the Illinois No. 6 coal was replaced by Kentucky No. 9 coal, a raw coal, since the former contained a very low amount of pyritic sulfur.

Five agglomerating liquids were investigated initially and, following preliminary testing, diesel and heptane were selected for in-depth study during the remainder of the program. The differences in the physical and chemical properties of diesel and heptane, along with their distinct process and handling requirements, necessitated the development of separate test programs for each. The test programs for the two processes were conducted simultaneously with the objective of selecting the one

that yielded the best performance, was economically advantageous, and entailed the least environmental risk. The test program consisted of several individual test plans to evaluate the effects of various operating parameters and determine the optimum ranges of these variables. This information was used to define the parameters used in the continuous tests and also provided the basis for selection of the agglomeration process to be implemented in Phase II of the project.

The Task 6 batch program was initiated primarily to investigate the dynamics of product size enlargement and to evaluate various classes of binders. The test program focused on the evaluation of low-shear mixing conditions, agglomerate growth, agglomerate recovery methods, and various binders. The batch program was carried out using laboratory-scale mixing vessels as well as larger-scale mixers to generate samples for further work. Details of the experimental apparatus and test procedures used are provided in the Task 6 final report. Both diesel and heptane were used as agglomerating liquids.

Bench-Scale Continuous Testing Program

Bench-scale continuous testing conducted as a part of Task 5 utilized ARC's 5-kg/h bench-scale and ¼-t/h pilot-scale agglomeration units. Approximately 40 bench continuous tests (at scales of 5 and 20 kg/h) were conducted at ARC. The primary objectives of the continuous testing program were to:

- Demonstrate process comparability to batch performance
- Reveal any serious deficiencies in a continuous agglomeration operation
- Provide support to the CDTF tests

The continuous test program was developed in accordance with the overall test design philosophy of the project and is described in the Task 4 final report. In addition to generating information on continuous operation of the selective agglomeration process, these tests also provided information on the effects of several operating variables.

Continuous tests were carried out in the 5-kg/h unit with all three coals using both diesel and heptane. Three commissioning runs were also carried out using ARC's ¼-t/h pilot unit. A few low-shear tests were also

performed in ARC's 20-kg/h low-shear unit to investigate agglomerate growth characteristics.

Component Development Test Program

Component development testing was performed at the CDTF in Wilsonville, Alabama. Twenty-nine continuous component development tests were conducted by SEI. The major objectives of these CDTF test runs were to:

- Operate the 3-t/h selective agglomeration facility at steady-state conditions
- Evaluate the performance of various items of mechanical dewatering equipment
- Estimate the system response to changes in process variables
- Determine the levels of diesel and asphalt required to achieve the desired product size and moisture in the low-shear agglomeration circuit

These objectives were achieved by conducting the following test series:

- Startup and shakedown of the grinding circuit
- Startup and shakedown of the agglomeration circuit
- Evaluation of operating parameters
- Testing of mechanical product dewatering devices

The design of the 3-t/h CDTF incorporated maximum flexibility to allow various operating scenarios to be tested. The equipment procured for the CDTF was generic and based on proven, commercially available technologies so that specific pieces of equipment could be used interchangeably regardless of the final Proof-of-Concept (POC) facility design.

Vendor Test Program

Tests were conducted at the laboratories of several vendors to investigate CDTF product drying and handling characteristics. Most of the vendor testing focused on the drying of agglomerates produced using an asphalt/heptane blend during low-shear agglomeration with the primary objective of removing and recovering the heptane. Drying tests were also completed on micro-

agglomerates generated from the heptane-based process. The drying systems investigated included:

- Vibrating fluidized-bed dryer
- Rotary tray-type dryer
- Hot water treatment
- Indirect dryer for microagglomerates

Tests were also performed to investigate pelletization as a means of producing handleable agglomerates. This was explored as an alternative method of producing well-formed, strong, and handleable agglomerates since low-shear agglomeration produced high-moisture agglomerates and required high oil dosages (for the heavy-oil process). Drying tests using these pelletized products were also conducted.

Batch Test Summary

The batch test program carried out by ARC was directed at improving the separation of coal particles from mineral matter, and thus primarily focused on high-shear mixing and microagglomerate recovery. The effects of the following variables were investigated:

- Effect of feed coal precleaning
- Feed coal particle size
- Type and dosage of agglomerating liquid
- High-shear conditions
 - Solids concentration
 - Mixing intensity
 - High-shear residence time
 - Flotation residence time
- Microagglomerate recovery method
- Effect of pyrite depressants
- Effect of two-stage agglomeration

Based on continuous evaluation of these test results, optimum operating conditions were defined. While optimum conditions obtained from batch tests do not necessarily constitute optimum conditions for continuous runs, they do serve to define initial operating conditions.

The variables that exhibited a significant effect on agglomeration performance were coal particle size, agglomerant type and dosage, high-shear solids concentration, and flotation solids concentration during microagglomerate recovery. Variables found important for final product formation included low-shear binder composition and concentration, and coal particle size. Tables 2-1 and 2-2 show the various process parameters, levels tested, and selected conditions for the heavy- and light-agglomerant test programs.

The batch tests show that the best agglomeration performance in terms of Btu recovery and pyritic sulfur rejection (PSR) obtained in this project can be characterized as 90% Btu recovery and 80% pyritic sulfur rejection (PSR) for Upper Freeport and Kentucky No. 9 coals, and 90% Btu recovery and 75% PSR for Pittsburgh No. 8 coal. In general, the batch results appeared to follow the typical washability curves for these coals.

Results

The results from the batch program show that agglomeration performance, in terms of pyritic sulfur rejection and Btu recovery, is similar for the diesel and heptane

processes for all project coals. However, process conditions and the characteristics of the final product generated from the two processes are different.

Following the high-shear optimization program, tests to investigate precleaning of run-of-mine (ROM) coals prior to agglomeration were conducted. These tests were completed for the Upper Freeport and Kentucky No. 9 coals at their respective optimum high-shear conditions. The results of these tests show that integrating precleaning with agglomeration provided 5-15% greater PSR at similar Btu recoveries than direct agglomeration of raw coals. Details of these tests are presented in Section 4.3.7 of the Task 7 report.

Table 2-3 summarizes the best batch agglomeration performance achieved during Phase I. As the table shows, Btu recoveries were in the 88-96% range for all coals, and pyritic sulfur rejection varied from 68 to 74% for Pittsburgh No. 8, and from 76 to 80% for the Upper Freeport and Kentucky No. 9 coals.

These tests provided the basis for the Phase I economic analyses. Agglomeration performance at a 200-mesh coal feed topsize can be characterized as 90% Btu recovery and 80% PSR for Upper Freeport and Kentucky

Table 2-1
Phase I Testing Heavy-Agglomerant Process Optimization

<u>Variables Tested</u>	<u>Conditions Tested</u>	<u>Selected</u>
Agglomerating liquid	Diesel, Maya, M-D, FO-D*	Diesel
Agglomerant dosage	0.5, 1, 1.5, 2, 3, 6, 10%	1%
Particle size	28M, 100M, 200M	28M/100M
High-shear (HS) solids	5, 10, 20%	20%
HS residence time	0.5, 1, 2, 4 minutes	1 minute
HS mixing intensity	1600, 2000, 2400 rpm	1600/2000 rpm
Microagglomerate recovery	Screening, flotation	Flotation
Flotation solids	1.5, 3, 6%	6%
Flotation stages	1-stage, 2-stage	2-stage

*Maya: Maya crude oil, M-D: Maya-diesel mixture, FO-D: Fuel oil-diesel mixture

Table 2-2
Phase I Testing Light-Agglomerant Process Optimization

<u>Variables Tested</u>	<u>Conditions Tested</u>	<u>Selected</u>
Agglomerating liquid	Hexane and heptane	Heptane
Agglomerant dosage	10, 20, 30, 50, 100, 150%	10-30%
Particle size	28M, 100M, 200M	200M
High-shear (HS) solids	5, 10, 15, 20%	5-10%
HS residence time	0.5, 1, 2, 4 minutes	1 minute
HS mixing intensity	1600, 2000, 2400 rpm	1600/2000 rpm
Microagglomerate recovery	Screening and flotation	Screening

Table 2-3
Best Phase I Batch Test Results

Lab ID	Topsize, Mesh	Oil Type	Oil Conc., %	Clean Coal Ash, %	Total Sulfur, %	Pyritic Sulfur, %	Btu Rec. %	PSR %
<u>Pittsburgh No. 8</u>								
Pit-289	200	Heptane	20.0	5.0	3.30	0.96	91.3	73.9
Pit-317	200	Heptane	20.0	5.4	3.33	0.90	91.6	74.2
Pit-407	200	Diesel	1.0	5.1	3.37	1.03	96.0	70.3
Pit-321	200	Diesel	1.0	5.7	3.34	0.93	92.6	73.3
Pit-291	100	Diesel	1.0	5.7	3.40	1.10	92.4	69.1
Pit-404	100	Diesel	1.0	6.0	3.40	1.00	94.4	68.0
<u>Upper Freeport</u>								
Ufp-332	200	Heptane	20.0	8.2	0.92	0.37	92.9	80.4
215A	200	Heptane	20.0	9.1	1.02	0.44	94.5	76.5
226C	100	Heptane	20.0	9.1	1.03	0.45	91.5	76.7
226A	100	Heptane	20.0	10.0	1.02	0.44	92.0	76.5
Ufp-327	200	Diesel	1.0	7.8	0.94	0.39	91.5	79.7
Ufp-326	100	Diesel	1.0	7.9	0.94	0.39	91.3	79.7
Ufp-325	28	Diesel	1.0	8.3	0.98	0.43	89.2	77.9
<u>Kentucky No. 9</u>								
Ken-110	200	Heptane	20.0	6.1	2.72	1.05	90.8	76.7
217A	200	Heptane	20.0	6.9	2.80	0.98	93.0	76.3
217B	200	Heptane	20.0	6.3	2.77	0.94	91.0	78.1
Ken-109	200	Diesel	1.0	5.9	1.56	0.89	90.9	80.2
Ken-108	100	Diesel	1.0	6.1	2.59	0.92	88.5	80.0
Ken-107	28	Diesel	1.0	6.5	2.64	0.98	88.2	78.7

No. 9, and about 90% Btu recovery and 75% PSR for Pittsburgh No. 8 coal. Clean coal ash values are in the 5-6% range for Pittsburgh No. 8, 8-10% for Upper Freeport, and 6-7% for Kentucky No. 9 coal.

Table 2-4 presents Phase I batch testing agglomeration performance results and 200-mesh topsize washability data, at both 85 and 90% Btu recovery levels.

Table 2-4
Phase I 200-Mesh Testing and Washability Data

Coal	Btu Recovery, %	Testing PSR, %	Washability PSR, %
Upper Freeport	85	83	94
	90	80	92
Pittsburgh No. 8	85	78	92
	90	75	90
Kentucky No. 9	85	80	95
	90	80	93

As can be seen in Table 2-4, agglomeration testing approached but did not achieve the project objectives of

85% Btu recovery and 85% PSR. It is also evident from this table that agglomeration testing achieved consistently lower (11-15%) PSR levels than the centrifuge sink/float washability analyses. These differences in PSR between testing and washability results for all three coals are attributed to a number of factors.

Washability float/sink results represent the theoretical maximum PSR achievable by gravity separation at any given Btu recovery. In contrast, agglomeration utilizes a surface-based phenomenon to form microagglomerates which are then separated from the mineral matter-bearing process water by either flotation (surface-based) or screening (mechanical).

The fine coal washability data shown in Table 2-4 are based on gravity separations in which all particles of specific gravity greater than 1.8 report to the refuse. During these centrifuge float/sink separations, "locked" or unliberated particles, i.e., particles which are part carbon and part mineral matter, may report to the refuse due to their overall specific gravity. Since agglomeration is a

surface-based process, these "locked" particles may report to the product since any coal on the particle surface will be coated with agglomerant and will subsequently report to the product. The reporting of locked particles to the product could possibly be reduced by additional grinding to further liberate pyrite particles.

The float/sink washability procedure requires the use of a dispersant and a 20-minute residence time at a centrifuge rpm of 1500 to ensure that a good gravity separation is achieved at the fine coal sizes used (<100 mesh). Since so much care is required to guarantee a good laboratory float/sink separation at these fine coal sizes, it is not unexpected that actual agglomeration test results indicate poorer performance than washability predictions.

PSR differences between testing and washability results are due partly to contamination of the flotation product by mineral matter carried within the water in the flotation froth. Recovery of microagglomerates by flotation results in a strong froth which can entrain ash and pyrite particles suspended in the water between air bubbles in the froth. A decrease in flotation solids concentration produces a corresponding reduction in the process water mineral-matter concentration, thereby reducing the amount of impurities carried into the product for each unit of water contained in the froth. A similar effect is observed when the first-stage flotation product is repulped and a second stage of flotation separation is completed.

Washability data for the Upper Freeport and Pittsburgh No. 8 coals indicate significant increases in pyritic sulfur liberation when the coal topsize is reduced from 28 mesh to 100 or 200 mesh. However, agglomeration process performance in batch testing at the 100- and 200-mesh topsizes is only slightly better than at the 28-mesh topsize. We hypothesize that this is due primarily to the decrease in flotation separation efficiency as the particle topsize decreases. This phenomenon does not occur during washability float/sink separations.

Summary

The major conclusions that were drawn from the batch tests carried out in Task 5 are:

- Integrating precleaning and agglomeration leads to overall improvements in process performance compared to direct agglomeration of raw coals. The design of an agglomeration plant would include an

integrated process combining conventional heavy-media precleaning with selective agglomeration.

- Project batch test results at optimized high-shear conditions can be characterized as 90% Btu recovery for all coals with pyritic sulfur rejection levels of 80% for Upper Freeport and Kentucky No. 9 coals and 75% for Pittsburgh No. 8 coal. These results were obtained using an integrated precleaning and agglomeration process.
- Agglomeration performance and selectivity in high shear, as measured by Btu recovery and PSR, were similar for diesel and heptane, for all coals tested.
- The optimum feed coal particle size depends on the coal type and the process selected (i.e., diesel- or heptane-based). The optimum topsize for each coal is listed below for each process:

Coal	Diesel	Heptane
Pittsburgh No. 8	100 mesh	200 mesh
Upper Freeport	100 mesh	200 mesh
Kentucky No. 9	28 mesh	200 mesh

- Agglomerant dosage had a significant effect on Btu recovery and pyritic sulfur rejection. Increasing the diesel dosage beyond 1% reduced PSR. This PSR reduction at the higher diesel levels was due to decreased system selectivity in the presence of excess diesel, i.e., excess diesel promotes the agglomeration of mineral matter.
- Heptane dosages between 10 and 30% in the high-shear stage provided acceptable Btu recovery and PSR. Below 10% heptane, the size of the microagglomerates was too small to achieve sufficient recovery via screening. Heptane dosages of 50% and above produced no improvement in process performance and led to operating difficulties.
- The high-shear solids concentration did not have a significant effect on process performance for the diesel process. However, decreasing the solids concentration improved performance for the heptane process. The difference in the effect of the high-shear solids concentration for the two processes is due to the quantity of agglomerant used and the size of microagglomerates formed. In the heptane process, where large agglomerant dosages form sufficiently large microagglomerates for screening, inclusion of mineral matter is likely to occur. Lowering the solids concentration dilutes the mineral-matter concentration in high shear, thus effectively reducing the inclusion of

mineral matter with the microagglomerates. In the diesel process however, significantly lower quantities of agglomerant form much smaller microagglomerates into which the inclusion of mineral matter is less likely.

- High-shear conditions such as agitation intensity and residence time did not have a significant effect on product quality within the ranges tested. However, the high-shear mixing intensity had to be sufficient to provide effective agglomerant dispersion or coal losses were observed. The optimum high-shear conditions determined for each project coal are as follows:

Coal	Mixing	Res. Time
Pittsburgh No. 8	2000 rpm	1 min
Upper Freeport	2000 rpm	1 min
Kentucky No. 9	1600 rpm	1 min

- Lower flotation solids concentrations during in diesel process provided higher PSR. This phenomenon is due to the dilution of mineral matter in the process water which is carried to the product with the flotation froth. The optimum flotation percent solids based on economics was 6%, although higher PSR was obtained at lower solids concentrations.
- Two stages of flotation provided higher PSR than a single stage of flotation, at equivalent Btu recoveries. This effect is due to the dispersion of the first-stage flotation product microagglomerates and its associated mineral matter prior to recovery in the second stage of flotation.
- The flotation times required during microagglomerate recovery to provide acceptable process performance for the Upper Freeport, Pittsburgh No. 8, and Kentucky No. 9 coals were 7-8, 1, and 2 minutes respectively. The long flotation time required for the Upper Freeport coal was probably due to oxidation of the coal sample.
- Two-stage agglomeration using a grind-agglomerate-regrind process option did not improve process performance over the standard single-stage agglomeration tests performed at optimized high-shear conditions.
- For the experimental conditions tested, under-water storage of slurry for up to 4 days did not lead to any deterioration in agglomeration performance, i.e., no oxidation occurred which had a deleterious effect on process performance.

- Emulsification of the agglomerant before high shear to reduce power consumption by allowing the use of less intense high-shear mixing, resulted in poorer process performance. The effect of the subsequent lower high-shear mixing intensity was decreased Btu recovery.
- Preliminary batch low-shear testing demonstrated that a mixture of AC-5 asphalt and diesel in the ratio 1:1 provided agglomerates with a median diameter of 1 mm. This was later confirmed in testing carried out in Task 6.
- The project test results compare favorably with results from similar advanced coal cleaning projects.

Continuous Bench-Scale Results

Both the diesel- and heptane-based processes were investigated during 5-kg/h bench-scale continuous testing by ARC. Initial runs tested the Upper Freeport coal, and the information obtained from these tests was then applied to testing of the Pittsburgh No. 8 and Kentucky No. 9 coals. Most of the test conditions were selected based on optimized levels of operating variables obtained from the batch program.

Diesel Testing

The initial commissioning run performed on the 5-kg/h agglomeration unit tested the diesel-based process over the range of 1-3% diesel addition to high shear, followed by two stages of microagglomerate recovery via flotation. Conclusions drawn from this run included the following:

- Steady state was achieved in the unit, as borne out by the consistency in both feed and ash values of clean coal, settlings, and tailings from the second stage of flotation.
- Recoveries of the Upper Freeport coal were generally low, confirming batch test results that indicated possible oxidation of the coal sample, as discussed previously. However, at a 28-mesh topsize, Btu recoveries were higher than for most batch tests.
- With increased diesel and frother additions, Btu recovery increased while ash and pyrite rejections decreased.

This initial commissioning run confirmed that the unit could be used to generate bench-continuous data, and

that a run time of 45-60 minutes was sufficient to achieve steady state after a change in any input variable. With this information, 25 additional diesel-based runs were completed using all three project coals to investigate the effects of various operating variables in a continuous selective agglomerating system. Conclusions drawn from these tests are as follows:

- Diesel dosage: In general, at low levels of frother dosage, increased diesel addition led to higher Btu recoveries and lower pyritic sulfur rejections.
- Feed coal particle size: Marginal performance increases were realized when the topsize was reduced from 28 to 100 mesh for Pittsburgh No. 8 and Upper Freeport coals. At equivalent Btu recoveries, a 5% increase in PSR was observed for Kentucky No. 9 coal when the topsize was reduced from 28 to 100 mesh.
- High-shear residence time: Btu recovery for the Upper Freeport coal increased as the residence time was increased from 0.9 to 1.9 minutes and then decreased as the residence time was increased from 1.9 to about 4 minutes. PSR was comparatively insensitive to changes in high-shear residence times for low or no frother addition.
- Flotation solids concentration for microagglomerate recovery: Tests revealed improved performance and selectivity at lower solids concentrations.
- Frother dosage in the flotation step: Increasing the frother decreased process selectivity. Sensitivity to frother dosage was more pronounced for Upper Freeport coal than for the other coals tested.

Overall, the findings from the diesel-based bench-continuous runs for all three coals confirmed the findings of the batch tests. While Btu recoveries were slightly lower in the continuous process than in the batch tests, the effects of the process variables were clear and consistent for the two modes of operation, indicating that the processes are highly comparable.

Heptane Testing

Following construction of a 5-kg/h test unit to allow safe heptane-based testing, three commissioning runs were completed to demonstrate system operability. These initial tests used Upper Freeport coal at a 200-mesh topsize, heptane dosages in the 8 to 28.5% range, and microagglomerate recovery via sieve bend and vibrating

screen. No major operating problems were encountered and steady-state operation was achieved. The results of these three tests confirmed the batch test findings that low dosages of heptane allow insufficient Btu recovery.

Four additional heptane-based continuous 5-kg/h runs were completed, two using Upper Freeport coal and one each using Pittsburgh No. 8 and Kentucky No. 9 coals. The primary objectives of these tests were to investigate the effects of heptane dosage, high-shear solids concentration, and high-shear residence time on the process.

As with the batch test results, increased heptane dosages usually led to increased Btu recoveries and lower pyritic sulfur rejections. However, in the case of continuous operation, process performance was less sensitive to high-shear solids concentration. Also, Btu recoveries were lower for continuous than batch tests, especially for Pittsburgh No. 8 and Kentucky No. 9 coals. The exact reason for lower Btu recoveries from the continuous process is not known. However, evaporation of heptane during screening to recover the micro-agglomerates would decrease coal recovery. Water loading in the sieve bend did not play a significant role in determining agglomerate product quality.

Although some differences between bench-continuous and batch testing were observed for the heptane-based process, the results generally indicate that, as in the case of the diesel-based process, the batch and continuous modes of operation are comparable. Also, for both processes, the optimum levels of major variables determined during batch testing were confirmed in the bench-continuous program.

Continuous Low-Shear Testing

Tests focusing on low-shear agglomeration, specifically on agglomerate growth, were conducted in ARC's 20-kg/h continuous unit. A description of the unit is provided in the Task 5 final report. The objectives of the tests were to investigate the growth kinetics of the product agglomerates in a continuous mode and to determine binder requirements to produce an acceptable agglomerate. Comparison of continuous and batch results was also an objective of the program.

The tests were performed using Upper Freeport coal at 28- and 100-mesh topsizes and various compositions of two different binders: a mixture of Cold Lake bitumen and diesel and a mixture of AC-5 asphalt and diesel. These tests involved low-shear agglomeration both with and

without a preceding high-shear step. The test conditions are presented below:

- Low-shear mixer: 4-blade paddle impeller
- Low-shear mixer speed: 1050 rpm
- Low-shear power input: 20 watts/liter
- Solids content: 25%.

The final agglomerate size was approximately 1 mm, with no growth observed for many tests. The AC-5 asphalt/diesel blend provided larger product agglomerates than the Cold Lake bitumen/diesel blend. It was also determined that about 12% binder on a product basis was required to achieve an agglomerate size of 1 mm using 100-mesh feed coal. This observation was confirmed by the Task 6 test results.

Conclusions

The Task 5 bench-continuous test program demonstrated successful continuous operation of both the diesel- and heptane-based processes. This testing also indicated comparability between the batch and continuous modes of operation in terms of product quality and Btu recovery levels. In addition to meeting these primary objectives, the bench-continuous test program confirmed batch agglomeration process trends relating to variables such as agglomerant level, frother dosage, flotation solids concentration, and high-shear residence time. Also, the low-shear continuous tests performed to evaluate the effect of binder type and quantity on low-shear agglomerate growth confirmed trends observed during Tasks 5 and 6 batch testing.

Component Development Test Results

A total of 29 tests were conducted at the Component Development Test Facility (CDTF) in Wilsonville during Task 6. These tests comprised the following:

- Five grinding calibration runs
- Ten startup runs, eight at a 28-mesh topsize and two at a 100-mesh topsize
- Fourteen runs to investigate agglomerate product dewatering.

All 29 CDTF test runs were conducted with Upper Freeport coal. The complete data set for all CDTF operations was provided in the Task 6 final report.

Grinding Calibration

Five CDTF test runs were conducted to calibrate the ball mills. The objective of these tests was to determine the ball mill configurations and charges required to produce the desired 28- and 100-mesh topsize particle size distributions (PSD). These tests were conducted using Upper Freeport coal at a 3-t/h dry coal feed rate, primarily at a 40% solids concentration. The results of these tests indicated that a 28-mesh topsize product ($d_{50} = 126 \mu\text{m}$) was produced using the mill defined as primary at a charge loading of 5500 lb, while the mill defined as secondary yielded a 100-mesh topsize product ($d_{50} = 58 \mu\text{m}$) at a charge loading of 9750 lb.

Startup and Commissioning Tests

Following mill calibration, 10 startup test runs were carried out at the CDTF to gain operating experience, achieve continuous operation of the facility, and determine stable operating conditions for the high-shear and flotation unit operations. These runs also helped to determine initial low-shear binder requirements, low-shear residence times, and typical diesel/asphalt ratios required to generate agglomerates approximately 1 mm in size. For all 10 startup test runs, the low-shear product was dewatered using a sieve bend and vibrating screen.

Based on the initial test results, it was concluded that CDTF operating conditions to generate 1 mm topsize agglomerates were:

- For 28-mesh topsize feed
 - 1% diesel in high shear, dry feed coal basis (fcb)
 - 8-9% binder in low shear (fcb)
 - 3:2 asphalt-diesel blend
 - 40 minutes residence time in low shear
- For 100-mesh topsize feed
 - 1.5% diesel in high shear (fcb)
 - 11-12% binder in low shear (fcb)
 - 3:2 asphalt-diesel blend
 - 40 minutes residence time in low shear

It was also determined that for runs at the high binder levels, two hours were required between binder dosage step changes to ensure that the low-shear product had reached its final size.

Low-Binder-Level Agglomerate Dewatering Tests

After initial startup testing, 14 CDTF test runs were conducted to investigate the dewatering characteristics of various agglomerated products using several pieces of dewatering equipment.

Eleven of these runs investigated the dewatering of agglomerates produced with low levels of binder (0-6%) using one low-shear vessel that provided about 25 minutes of low-shear residence time. Agglomerates generated during these runs were of relatively small size, i.e., 250 and 160 μm for the 28- and 100-mesh topsize feedstocks, respectively. The conditions and dewatering equipment tested during this phase of the program were as follows (binder levels on feed coal basis):

- Plate-and-frame filter press: 28- and 100-mesh topsize with 3 and 6% low-shear binder (60:40 asphalt:diesel blend)
- Bird screen bowl centrifuge: 28- and 100-mesh topsize
 - 3 and 6% low-shear binder (60:40 blend)
 - 3% diesel in low shear
- Eimco horizontal belt vacuum filter for 28- and 100-mesh topsize
 - 3 and 6% low-shear binder (60:40 blend)
 - 3% diesel in low shear
 - No agglomerant addition in low shear
 - 100-mesh topsize with 1% asphalt in low shear

When testing the plate-and-frame filter press, a clean bank of filter press plates was filled with low-shear product and allowed to dewater until no water had been visibly removed for 30 minutes. During the Bird screen bowl centrifuge testing, the slurry feed rate and weir depth were varied while the speed of the centrifuge was kept constant at 2050 rpm. During testing of the Eimco horizontal belt vacuum filter, the slurry feed rate and belt speed were changed to provide various product retention times and filter cake thicknesses.

Dewatering tests indicated that the plate-and-frame filter presses were unsuccessful in dewatering agglomerated low-shear products, providing final product moistures in the range of 32-38%. No effect on moisture content due to coal feed size or binder content was observed for the filter presses. It is believed that the incompressible nature of the agglomerated product filter cake prevented the removal of additional surface water.

The following conclusions can be drawn from the Bird screen bowl centrifuge testing:

- A low weir setting in the solid bowl portion of the centrifuge resulted in greater fines loss to the centrate than the high weir setting.
- As the coal feed was increased, additional fines loss to the centrate was observed.
- As expected, the loss of fines to the centrate was reflected by lower product moisture contents and decreased coal recoveries.
- No significant effect of binder types on product moisture was observed.
- At a 28-mesh topsize, no effect on moisture was observed when the binder level was increased from 3 to 6%.
- At a 100-mesh topsize, increasing the binder level from 3 to 6% resulted in a 10% moisture drop.
- The lowest product moisture content achieved with the Bird screen bowl centrifuge was approximately 14% at the following conditions:
 - High weir setting
 - Greater than 96% solids recovery
 - 28-mesh coal, 3 and 6% binder
 - 100-mesh coal, 6% binder.

The following conclusions characterize the performance of the Eimco horizontal belt vacuum filter. It should be noted that virtually no solids losses were detected for this piece of dewatering equipment.

- Under all conditions tested, the filter cake thickness had no effect on product moisture, indicating that over the range of belt speeds and slurry flow rates tested, filter performance remained constant.
- Filter performance showed little sensitivity to the binder type and quantity used. Product moistures in the 22-25% range were achieved with all binders tested, as well as when no binder was used.
- The lowest product moisture of approximately 19% was achieved at the following conditions:
 - 28-mesh topsize feed
 - 3 and 6% binder addition to low shear
 - All slurry feed rates tested
 - All belt speeds tested

High-Binder-Level Agglomerate Dewatering Tests

Three CDTF runs were carried out to investigate the dewatering characteristics of large (1 mm) agglomerates produced with high levels of binder, approximately 13 and 9% for 100- and 28-mesh topsize feedstocks, respectively. These large agglomerates were dewatered using the Bird screen bowl centrifuge, the Eimco horizontal belt vacuum filter, and a static sieve bend/vibrating screen combination.

For tests carried out in the screen bowl centrifuge, the weir depth was kept constant at a high level to minimize the loss of fines to the effluent, while the feed rate was varied. The centrifuge speed was also lowered to 1750 rpm to reduce product degradation as per the manufacturer's recommendations. The horizontal belt vacuum filter was tested at both a constant belt speed (10 ft/min) and product feed rate since previous dewatering tests of low-binder-level agglomerates indicated that these variables did not affect dewatering performance within the ranges tested. For dewatering tests using the vibrating screen, the entire product stream (3-t/h plant feed) was processed across the sieve bend and vibrating screen. The following observations and conclusions were drawn from this testing:

- For all equipment tested, virtually no coal losses were observed.
- For centrifuge dewatering, moisture contents of approximately 14 and 10% were achieved for 100- and 28-mesh topsizes respectively. It was also observed that the product moisture content was independent of solids throughput.
- The horizontal belt vacuum filter achieved 13.5 and 15.8% product moisture levels for 28- and 100-mesh topsizes respectively with no product degradation.
- For sieve bend/vibrating screen dewatering, approximately 25% product moistures were achieved for both topsizes with no product degradation. Also, four hours of screen product drainage reduced the moisture by an additional 3% for both topsizes.
- Centrifuge- and vacuum filter-dewatered products had no free water present after dewatering as was confirmed by long-term drainage tests.
- The screen bowl centrifuge reduced the agglomerate product in size from a d_{50} of approximately 825 μm to 300 μm for both feed sizes.

Vendor Testing

In addition to CDTF operations and batch tests, a variety of vendor tests were conducted during Task 6. The primary objectives of the vendor tests were to:

- Evaluate various commercially available drying systems for their applicability to heptane-recovery systems.
- Investigate pelletization as a means of producing strong, handleable agglomerates.

Low-Shear Agglomerate Product Testing. Analyses were conducted for trace amounts of heptane remaining in the dried product agglomerates to determine the effectiveness of the various dryers in removing heptane and to estimate potential environmental risks due to heptane remaining in the agglomerates. The agglomerates analyzed included samples from all of the following dryer systems:

- Vibrating fluidized-bed dryer
- Rotary tray-type dryer
- Hot water treatment
- Indirect dryer for microagglomerates

Refuse samples from the heptane agglomeration process were also analyzed for their residual heptane content to estimate potential environmental risks and to determine whether a separate heptane recovery system was required for the refuse stream. In addition, samples of oven-dried agglomerates were dried at about 38°C for a period of 21 days to simulate conditions in an open stockpile. Samples of these agglomerates were taken at 0, 7, and 21 days and analyzed for their heptane content to estimate emissions to the atmosphere.

The following conclusions can be drawn from the agglomerate drying program:

- Heptane agglomerates were successfully dried using both a rotary tray dryer and a vibrating fluidized-bed dryer, with the majority of the heptane and surface water being evaporated.
- The vibrating fluidized-bed dryer resulted in less residual heptane and total volatiles content than the rotary tray dryer but at a slightly higher coal loss. Estimated capital costs were lower for the fluidized-bed dryer.

- Significant residual heptane was detected in the product agglomerates when asphalt was used as a binder. The results indicated that the amount of heptane remaining in the agglomerates could be correlated with both the level of drying and the amount of asphalt used as a binder. While the heptane content of samples containing 2 and 3% asphalt was between 2000 and 4000 ppm, microagglomerates made without asphalt contained heptane below the detection limit of 50 ppm.
- Prolonged hot water treatment of agglomerates effectively removed heptane to residual levels of approximately 900 ppm but resulted in poor green agglomerate strength. Thermal drying of these green agglomerates was required to achieve strengths comparable to those of agglomerates dried for heptane recovery.
- Steam-stripping or hot water treatment was also effective for treating process refuse, resulting in residual heptane contents of <50 ppm provided that sufficient steam stripping was applied.
- Experiments simulating heptane emissions from an open stockpile under ambient conditions indicated that residual heptane would evaporate to the atmosphere from product storage piles. Samples stored for 21 days at 38°C contained much less heptane than those stored for 0 and 7 days. This indicates that special storage requirements may be necessary for heptane agglomerates.

Pelletized Product Testing. As discussed in the Process Selection Report, pelletization of dewatered microagglomerates represents the most viable option for final product formation. This recommendation was made based on results of testing carried out at Wilsonville and vendor laboratories. Size enlargement tests using a disc pelletizer were completed at Ferro-Tech facilities in Wyandotte, Michigan. The variables tested included:

- Binder types: Wet-milled food cornstarch and dry-milled corn flour
- Binder concentrations: 0.5-5%
- Method of disc feed preparation: dry mix binder and wet mix binder
- Asphalt content for weatherproofing: 0-2%

The pelletization samples were primarily prepared in a pin mixer or Turbulator™. A paste mixer was used for several tests. It was observed that the lower the binder content in the mix, the better the mixing in the pin mixer.

Methods used to measure initial pellet strength during pellet characterization testing included compression force breakage of pellets in a narrow size distribution, 18-inch drop testing of "green" (wet) and dried pellets, and 6-foot drop testing of dried pellets. The results suggested that moisture contents of 5-6% were required to achieve maximum pellet strength, while observations made during vibrating fluidized-bed drying tests indicated that moisture contents of 2-4% were required. It was observed that increasing the starch content in the mix provided better crush strength.

Several pelletized samples were dried in the vibrating fluidized-bed dryer to determine its suitability for handling pellets and also to estimate pellet strength and the potential for dust generation. The results of these tests indicate that the heavy-oil pelletized agglomerates can be dried in a fluidized-bed dryer at higher gas velocities and temperatures than heptane agglomerates. Both types of pellets required low moisture contents to achieve maximum strength. Pelletized samples were much larger (10-12 mm) than heptane agglomerates (3-4 mm) at equivalent binder consumption.

The continuous pelletization tests successfully enlarged dewatered microagglomerates over the range of binder additions tested (1.5-4% dry-milled flour). Pellet quality in terms of size distribution and crush strength was as good as or exceeded that obtained from batch tests.

The pelletization tests proved quite successful, with pellets from the heavy oil process achieving the highest strength of all the agglomerates tested. No significant dust loss was observed during pellet drying nor was a diesel odor detected during or after the drying tests.

Related DOE Projects

In addition to the selective agglomeration process, two related advanced coal cleaning technologies are being developed under the Acid Rain Control Initiative/Compliance Technology (ARCI/CT) component of the DOE Coal Preparation Program. The ARCI/CT projects focus on developing advanced physical coal cleaning approaches to assure the continued use of coal in the traditional applications of large industrial and electric utility boilers, by addressing environmental

concerns regarding local, regional, and global impacts of coal use.¹

Advanced Froth Flotation

All three advanced coal cleaning processes utilize size reduction to fine or ultrafine particle sizes to liberate pyrite and mineral matter. The advanced froth flotation process uses froth flotation to separate the liberated pyrite and mineral matter from the remaining coal matter. Initial process development was conducted by the University of California at Berkeley, where researchers investigated the effect of surface control methods on sulfur rejection and energy recovery. Later, ICF Kaiser was awarded a contract in September 1988 to conduct engineering development of the process. The project team includes Babcock and Wilcox, Consolidation Coal Co., Eimco Corporation, and Virginia Polytechnic Institute and State University (VPI).

During process development, several advanced flotation processes were evaluated and the microbubble column technology developed by VPI was selected for POC testing. The theoretical performance of an advanced flotation circuit for Upper Freeport Coal is 81.8% pyritic sulfur rejection at 90.2% energy recovery based on the raw coal quality.² POC testing is in progress and will be completed in 1993. The POC facility is located at the Ohio Coal Development Office facility in McConnellsville, Ohio.

Advanced Cycloning

Advanced cycloning technology is an extension of conventional cyclone cleaning. The advanced technology differs from conventional technology in that alternative media such as organic liquids, aqueous solutions, and colloidal suspensions are under consideration to replace conventional dense media. Centrifugal force is also used to accelerate separation.

Previous work in advanced cycloning was conducted by Process Technology, Inc. (PTI), under DOE auspices in cooperation with EPRI. PTI's work showed that Illinois No. 6 coal with a topsize of 75 μm (200 mesh) can be cyclonically separated at 45 kg/h (100 lbm/h) using an

organic liquid, methylene chloride, to achieve 83.2% pyrite rejection at 87.7% energy recovery.³

Coal Technology Corporation (CTC) was awarded the contract to perform the engineering development of this process in September 1990. The test circuit for this process will be located at CTC's industrial site in Bristol, Virginia. Project team members include ICF Kaiser, Process Technology, Inc., and Intermagnetics General Corporation.

Conclusions

Phase I of the project was successful in meeting many major project objectives. Most of the process uncertainties identified in Tasks 2 and 3 were resolved by means of a comprehensive strategy including theoretical analysis, batch and continuous testing, and unit operations testing at Wilsonville. The remaining deficiencies, primarily related to product handling and physical characteristics such as strength, flow, degradation, and retention of agglomerant in the product, were addressed during POC operation.

Btu recoveries of 90% were achieved for all of the project coals at pyritic sulfur rejections of 80% for Upper Freeport and Kentucky No. 9 coals and 75% for Pittsburgh No. 8 coal. While these values indicate good process performance, agglomeration testing approached but did not achieve the project objectives of 85% Btu recovery and 85% PSR. This data also indicates that agglomeration testing achieved consistently lower (11-15%) PSR levels than the centrifuge sink/float washability analyses completed on the same feed coal. Batch and bench-continuous test results demonstrated that these two modes of operation are comparable.

The 3-t/h CDTF was successfully operated with no major operating deficiencies. Major unit operations, such as fine grinding in ball mills, high-shear mixing, micro-agglomerate recovery via flotation, and size enlargement by low-shear agglomeration, were successfully tested in the CDTF.

Batch test results indicated that the variables which affect agglomeration performance are the feed coal particle size, agglomerant type and dosage, high-shear solids concentration, and microagglomerate recovery method. Options that did not prove beneficial in improving agglomeration performance include the use of pyrite

¹Barnett, William P., and Thomas J. Feeley, III, "Status of Advanced Coal Cleaning as a Compliance Technology," *Mining Engineering*, October 1992, pp. 1225-1230.

²Ibid., p. 1228.

³Ibid., p. 1230.

depressants, agglomerant emulsification, and two-stage processing. These findings were confirmed in bench-continuous runs. Both diesel and heptane were used during bench-continuous testing, and provided agglomeration performance comparable to that achieved in the batch test results.

Tests conducted with diesel and heptane demonstrated similar Btu recovery and PSR levels. However, final product agglomerates formed using diesel/asphalt blends during low shear proved to be weaker and have a higher moisture content than those formed with a heptane/asphalt blend during low shear. Size enlargement of the product agglomerates via pelletization for the diesel-based process proved to be the most suitable option, providing good pellet characteristics. Starch and AC-5 asphalt were used as binders for this pelletization process.

Heptane removal tests conducted at vendors' laboratories proved that although most of the heptane was successfully removed, 0.2-0.3% heptane remained in the agglomerates. While hot-water treatment of these agglomerates resulted in residual heptane levels of about 0.1-0.2%, they proved to be weaker than thermally dried agglomerates. Tests conducted on product agglomerates indicated that residual heptane may be released to the atmosphere under ambient storage conditions. Significant amounts of heptane from an environmental standpoint were also detected in the refuse stream.

Three process alternatives were evaluated: (i) diesel high shear followed by low shear, (ii) heptane high shear followed by low shear, and (iii) diesel high shear with pelletization. In addition, two variants of the diesel and heptane processes with no size enlargement were also investigated. The processes were evaluated in terms of eight selection criteria and assigned risk factors based on maturity and feasibility. All of the processes were judged technically feasible; the heptane process was deemed a higher-risk alternative since it is not as mature as the diesel processes. Economic and reliability analyses at a 200-t/h scale were carried out to evaluate the various process options. These reliability analyses indicated that 90% availability was exceeded for all plant designs. The process combining diesel high shear with pelletization for size enlargement was recommended for POC testing in Phase II since it represented the lowest-risk option of all the processes while meeting the project objectives.

The selective agglomeration technology was evaluated with respect to criteria listed in the statement of work, and it was recommended that development of the technology proceed to the POC scale. Major accomplishments in Phase I include the development of optimized conditions for effective separation of clean coal from mineral matter and integration of precleaning with agglomeration.

3

PHASE II TESTING

Introduction

The primary objective of Phase II of the project was to generate the information required for the design of a commercial-scale selective agglomeration facility using continuous Proof-of-Concept (POC) test results. Initial work during this phase focused on the conceptual design of a commercial-scale 200-t/h selective agglomeration facility. This conceptual design was a revision of the initial 20-t/h semiworks plant design completed during Phase I and incorporated all information generated during Phase I testing and analysis.

Based on the commercial-scale conceptual design, the next project task was to complete the conceptual and detailed designs of a 2-t/h POC facility for Phase II testing, located in Wilsonville, Alabama. Upon completion of the detailed design, all necessary equipment and materials not already installed in the Phase I CDTF facility were procured and/or fabricated. The CDTF facility was then modified as required to complete POC facility construction. A POC facility test plan was developed and carried out. Analysis of POC operations and test results was carried out concurrently with execution of the test plan. The last major Phase II task was the final conceptual design of a commercial-scale selective agglomeration facility, incorporating all information generated from the Phase II work. Following the completion of POC testing, the POC module was dismantled and removed from the site. This work is described in detail in the Site Closeout Report, to be issued separately.

The following is a summary of the Phase II tasks:

- Task 8:** Conceptual design of commercial-scale (200-t/h) selective agglomeration facility
- Task 9:** POC facility design including development of design criteria, conceptual design, and detailed design

- Task 10:** Procurement and fabrication of all POC-related materials and equipment
- Task 11:** Construction, installation, and startup of the POC module
- Task 12:** POC test plan formulation
- Task 13:** Operation and testing of the POC facility
- Task 14:** Analysis of POC test results
- Task 15:** Final conceptual design of commercial-scale agglomeration facility
- Task 16:** POC module removal and site restoration

Based on Phase I testing results and analysis a diesel-based selective agglomeration process was chosen for investigation during Phase II of this project. The selected process utilizes small amounts of diesel (<1.0% on a dry coal basis) as agglomerant during high shear followed by two stages of flotation to recover the microagglomerates. The recovered microagglomerates are then mechanically dewatered, pelletized, and thermally dried to form the final product.

The other major agglomeration process investigated during Phase I of the project but not selected for continued investigation during Phase II was a heptane-based process. The heptane process tested utilized heptane (approximately 20-30% on a dry coal basis) during high-shear agglomeration. Following high shear, microagglomerates were recovered by screening. These microagglomerates were then processed through low-shear agglomeration using a diesel/asphalt blend as binder. The low-shear product was then mechanically dewatered followed by thermal drying to recover the heptane and produce the final product.

While both the diesel- and heptane-based processes provided similar performance at similar costs, the diesel

process was selected for continued investigation during Phase II of the project since it was considered to be a more mature process presenting less risk. The diesel- and heptane-based processes however, both possess inherent advantages and disadvantages.

A primary advantage of the diesel process is a result of the nature and quantity of the agglomerant used. In the diesel or heavy-oil process, relatively small amounts (<1.0%) of diesel oil, a hydrocarbon of low volatility, are used. Therefore, the diesel does not have to be recovered and remains in the final product. In contrast, the heptane or light-oil process requires significant concentrations (20-30%) of heptane, a highly volatile hydrocarbon, during high shear. As a result, the heptane cannot be allowed to remain in the final product and must be recovered. Recovery of heptane to levels that meet product safety and environmental standards is expensive. Also, with the use of heptane, the entire process from the point of heptane introduction during high shear to heptane recovery by thermal dryers must be sealed and blanketed with an inert gas for safety reasons. This required safety precaution increases capital investment substantially.

During selective agglomeration, the high-shear unit operation is typically conducted at approximately 20% solids concentration. In the heptane process, no dilution of the high-shear product is required prior to screening to recover microagglomerates. While some wash water is required to assist in the screening separation, the amount of water needed is small compared to that used during the two stages of flotation used to recover microagglomerates in the diesel process. For both stages of flotation in the diesel process, the feed is diluted to approximately 6% solids concentration.

An additional advantage of the light-oil process is that the screening operation is much simpler, easier to control, and less expensive than froth flotation recovery of microagglomerates. It is also possible that with flotation, the overall performance of the selective agglomeration process may be determined by the microagglomerate recovery step rather than by the high-shear agglomeration unit operation.

During heptane agglomeration, final product formation is conducted via low-shear agglomeration using asphalt-based binders blended with additional heptane. While the asphalt is needed to impart strength to the final product, its presence makes recovery of heptane to the required levels difficult to achieve. In fact, Phase I testing indicated that with asphalt present in the low-shear product, the quantity of heptane remaining on the final

product after thermal recovery was environmentally unacceptable. In addition, the final product from the heptane process is not as strong and handleable as that from the diesel/pelletization process.

Test Coals

During Phase I testing, it was found that selective agglomeration of the Kentucky No. 9 coal did little to enhance pyritic sulfur rejection (PSR). This finding was supported by analysis of fine coal centrifuge washability data for this coal. Based on this information, the Kentucky coal was tentatively replaced by an Upper Freeport coal middlings fraction. During Phase II testing, AMAX Coal Company expressed interest in the selective agglomeration technology as a retrofit application for a preparation plant being constructed to process Illinois No. 5 coal. Since transfer of the selective agglomeration technology to a commercial coal company was of interest to the project, Illinois No. 5 coal was considered for use during POC testing. Following washability analyses and laboratory testing, the Illinois No. 5 coal was chosen to replace the Upper Freeport middlings coal during Phase II testing.

The four coals selected by the project team for utilization during Phase II of this project were:

- Upper Freeport coal from Jefferson County, Pennsylvania
- Pittsburgh No. 8 coal from Greene County, Pennsylvania
- Illinois No. 5 coal from Wabash County, Illinois
- Pond fines from Jim Walter Resources (JWR) No. 3 mine in Jefferson County, Alabama

Both the Upper Freeport and Pittsburgh No. 8 coals were precleaned prior to project use, while the Illinois No. 5 coal sample represented approximately 80% of the run-of-mine (ROM) coal. The pond fines sample from JWR was reclaimed from a pond containing flotation tailings from the production of both steam and metallurgical coals.

Upper Freeport Coal

The Upper Freeport coal used during Phase II was obtained from P&N Coal Company's Sheesley Mine in Jefferson County, Pennsylvania, and conventionally

precleaned by CQ, Inc. in their pilot plant. This coal was sampled, precleaned, and delivered in two separate batches. Table 3-1 presents raw and washed coal properties for both batches of coal, along with corresponding precleaning performance values. Details of the precleaning of both batches of Upper Freeport coal are presented in Appendix A, Upper Freeport Precleaning Report.

Table 3-1
Upper Freeport Coal Precleaning Summary

Parameter	Batch 1		Batch 2	
	Feed	Product	Feed	Product
Ash, %	14.36	10.07	13.64	9.70
TS, %	3.04	1.81	2.83	1.67
PS, %	2.34	1.07	1.96	0.89
OS, %	0.69	0.73	0.84	0.75
SS, %	0.01	0.00	0.03	0.03
Btu/lb	13146	13908	13300	13984
Yield, %	93.0		92.0	
Btu recovery, %	98.4		96.7	
TS rejection, %	44.6		45.7	
PS rejection, %	57.5		58.2	
Ash rejection, %	34.8		34.6	

Pittsburgh No. 8 Coal

Initial project planning called for the use of a Pittsburgh No. 8 raw coal middlings fraction; however, investigations determined that use of the entire precleaned coal would be more appropriate (see Appendix B, Pittsburgh No. 8 Coal Washability Report).

The Pittsburgh No. 8 coal used during Phase II was obtained from Cyprus Coal Company's Emerald Mine coal preparation plant. The sample obtained for project use was the conventionally cleaned plant product. During cleaning, the ROM coal is crushed to a 2" topsize and the 2" x 28-mesh raw coal is cleaned in heavy-media cyclones. The minus 28-mesh fraction is cleaned via conventional flotation that provides 2-3 minutes of residence time. Both the coarse and fine clean coal fractions are then mechanically dewatered and combined to form the final plant product. Table 3-2 presents the raw and washed coal properties, along with associated precleaning performance values.

It should be noted that sampling at the Emerald Mine preparation plant to determine precleaning performance was carried out six days after the POC feedstock sample was set aside for the project's use. As such, the conventional cleaning feed and product (POC feed) analyses presented in Table 3-2 do not represent actual

feed and product stream analyses at the time when the POC feed sample was obtained. The precleaning performance values presented, however, are considered representative of typical conventional cleaning for this Pittsburgh No. 8 coal.

Table 3-2
Pittsburgh No. 8 Coal
Precleaning Summary

Parameter	Feed	Product
Ash, %	30.88	7.60
TS, %	2.16	2.19
PS, %	1.48	1.23
OS, %	0.67	0.95
SS, %	0.01	0.01
Btu/lb	10167	14089
Yield, %	67.3	
Btu recovery, %	93.3	
TS rejection, %	31.7	
PS rejection, %	44.0	
Ash rejection, %	83.4	

Illinois No. 5 Coal

The Illinois No. 5 coal utilized during Phase II testing was provided by AMAX Coal Company from their Wabash Mine in Wabash County, Illinois. Figure 3-1 shows the Illinois No. 5 coal handling procedure used at the mine site to generate the project sample which represents approximately 80% of their ROM coal.

Figure 3-1. Illinois No. 5 Coal Mine Handling Procedure

As shown in Figure 3-1, to generate this sample, the ROM coal was first screened at 1 inch. The plus 1-inch material was then reduced to a 2-1/2-inch topsize and screened at 1/2 inch. The minus 1-inch material from the initial screening was then combined with the minus 1/2-inch material from the second screening to produce the POC test sample. Table 3-3 presents the properties of this "raw" Illinois No. 5 sample.

Table 3-3
Illinois No. 5 Coal Properties

Ash, %	17.74
TS, %	1.97
PS, %	1.23
OS, %	0.70
SS, %	0.04
Btu/lb	11797

Preparation Plant Pond Fines

The sample of Jim Walter Resources (JWR) pond fines used in POC testing consisted of flotation tailings from the production of both steam and metallurgical coals from their No. 3 mine preparation plant. However, some coarse material (up to 8 mesh) was present in the sample.

At the request of JWR, POC selective agglomeration testing was carried out on only the minus 60-mesh fraction of this feedstock. Table 3-4 presents typical No. 3 mine pond fines characteristics as determined in the Wilsonville laboratory.

Table 3-4
JWR Pond Fines Characterization

Size Fraction	Wt. % Entire Sample	Wt. % Minus 60 Mesh	Ash, %	TS, %
Full sample	100.0	—	56.5	0.256
Plus 60 M	44.7	—	49.5	0.287
Minus 60 M	55.3	100.0	62.1	0.231
60 x100 M	24.8	44.9	60.0	0.255
100 M x O	30.5	55.1	64.2	0.222

Test Coal Characterization

Throughout POC testing, samples of the flotation feed and product streams were sent to Geochemical Testing Laboratories for analyses. The analyses performed on each of these samples included ash content, sulfur forms, and calorific value determinations. Coal constants

based on these data were used to define the ash/Btu relationship and the moisture- and ash-free (MAF) organic sulfur content of each coal. Samples of each coal were sent out in two separate batches for analysis. Constants were derived based on the first batch and updated based on the second batch. These constants were used to calculate POC feed and product pyritic sulfur contents and calorific values. Since the sulfatic sulfur content was low (generally <0.04%) for all three project feed coals tested, it was not included during pyritic sulfur content calculations. Thus, the calculated pyritic sulfur contents and rejections presented in this report actually represent the inorganic sulfur content, i.e., pyritic and sulfatic sulfur combined.

The MAF organic sulfur constants—assumed to remain constant for a given coal sample—that were used during POC testing are given in Table 3-5.

Table 3-5
Project Coal MAF Organic Sulfur Constants

Coal	MAF OS, %
Upper Freeport batch 1	0.667
Upper Freeport batch 2	0.674
Pittsburgh No. 8	0.998
Illinois No. 5	0.764

POC Test Plan

Background

While CDTF operations in Phase I focused on final product formation and dewatering, the primary goal of POC facility operations was to optimize the diesel agglomeration process at the pilot scale. This involved the evaluation of various selective agglomeration operating parameters and conditions with the intent of achieving the project goals of 85% Btu recovery and 85% PSR on a raw coal basis while producing clean-coal ash contents of 6% or less for all three project coals.

Various selective agglomeration process trends were identified during the Task 5 bench-scale test program and Task 6 CDTF operations. The objective of the POC test plan was to validate these trends and define the optimum operating conditions for a continuous, pilot-scale selective agglomeration operation. The trends investigated in the POC can be categorized as those relating to coal matter recovery and separation efficiency. The following subsections discuss the process variable trends

that were the basis for the design of the Task 13 POC test program.

Coal Size Effects. The desired coal grind size for the selective agglomeration process is based on the liberation properties of the feed coal. Increasing the fineness of grind leads to increased pyrite rejection and increased energy recovery provided that appropriate agglomeration procedures are used with an adequate amount of agglomerant. However, since operating costs are significantly affected by increased grinding, an attempt was made to limit the fineness of grind while meeting the project goals. Coal topsizes investigated during POC operations were 100 and 200 mesh.

Agglomerant Dosage Effects. The major process variable affecting the state of coal agglomeration within the system is the concentration of agglomerant; high-shear conditions also affect the state of agglomeration but not as significantly. The rate of agglomerate formation decreases with a reduction in agglomerant concentration. Also, if insufficient agglomeration occurs due to a low agglomerant dosage, coal matter recovery is reduced nonselectively, i.e., particles with a higher ash content are not preferentially rejected. However, if excessive amounts of oil are used, the selectivity of the coal/ash separation is adversely affected. The amount of agglomerant required for agglomerate formation is a function of the hydrophobicity of the coal and its surface area, which is a function of the fineness of the grind. The POC operations test plan was designed to allow determination of the minimum amount of agglomerant required for each project coal while meeting the project energy recovery objectives.

Effects of High-Shear Conditions. High-shear mixing affects the kinetics of agglomerate formation. Increasing the specific energy input into the mixer (by increasing the mixer impeller rpm) reduces the time required to agglomerate the coal. The Phase I laboratory test program indicated that once microagglomerate formation is achieved, additional mixing does not alter the separation performance during product recovery in the flotation cells. The batch test program also determined that a minimum residence time of one minute was required for complete agglomeration over the range of high-shear conditions tested. The POC operations test plan was designed to investigate the effect of high-shear mixing intensity on process performance while minimizing high-shear power consumption.

Frother Concentration Effects. Frothers are alcohol- or glycol-based chemicals that are used in flotation processes to stabilize air bubbles (with attached parti-

cles) in a water slurry. Use of a frother decreases the size of the air bubbles produced for a given flotation system. Project laboratory tests indicated that frothers were helpful in improving the recovery of oxidized coal (the Phase I Upper Freeport sample) since oxidation adversely affects the attachment of coal to air bubbles. However, if the test coal is very hydrophobic or excessive amounts of frother are used, an adverse effect on separation selectivity can be observed due to the attachment of unagglomerated, higher-ash-content particles to the air bubbles which then report to the agglomerate product. The POC operations test plan called for the investigation of frother levels required for each of the coals to achieve appropriate energy recovery levels.

Flotation Feed Solids Concentration and Froth Washing Effects. The recovery of microagglomerates by flotation results in a strong froth which can entrap ash particles suspended in the water between the air bubbles in the froth. A method of preventing ash entrapment which was tested during Phase I batch work at both ARC and Wilsonville is reduction of the solids concentration of the flotation feed. Such a reduction produces a corresponding reduction in the ash concentration in the flotation cells and therefore reduces the amount of ash carried into the product for each unit of water contained in the froth. Reducing the flotation solids concentration however, also decreases coal throughput while increasing process costs. Another method practiced in other industries is to apply water to the froth as it forms on top of the flotation cells thereby replacing the ash-loaded water contained between the air bubbles with clean water. Both of these methods were investigated during POC testing to determine their effect on process performance.

Summary

Initial POC testing consisted primarily of startup test runs utilizing the Upper Freeport coal. Startup testing included mechanical checks of all plant equipment, calibration of all POC facility instrumentation, and initial water balance tests. Concurrently with startup testing, the ball mills were calibrated for the Upper Freeport coal.

Following initial startup testing, the POC test plan provided for evaluation of the various selective agglomeration operating parameters discussed in the previous section for all three project coals. The same basic scenario was utilized for parametric testing of both the Upper Freeport and Pittsburgh No. 8 coals. However, for the Pittsburgh No. 8 coal, less testing was required due to the project team's increased understanding of the

selective agglomeration process and additional POC operating experience. The basic structure for POC parametric testing of the first two project coals is summarized below:

- Calibration of POC ball mills to provide both 100- and 200-mesh topsize products for subsequent agglomeration
- Initial orientation testing to determine the effect of particle topsize and to evaluate a range of agglomerant and frother concentrations
- High-shear optimization testing to investigate the effect of high-shear solids concentration and mixing intensity
- Microagglomerate recovery (flotation) optimization to characterize the effect of frother concentration, percent solids, flotation cell water level, air flow rate, and froth spraying.

During parametric testing of the third project coal (Illinois No. 5), emphasis was placed on agglomeration performance at a 100-mesh topsize. The 100-mesh grind size was of primary interest since AMAX Coal Company, the coal supplier, expressed interest in determining the applicability of the selective agglomeration process in a 1500-t/h plant currently under construction. This potential application is for treating the minus 100-mesh natural fines which the current design bypasses to the plant tailings.

For parametric testing of the Illinois No. 5 coal, EPRI made available a pre-release version of the Ultramax software program to assist in optimization of the selective agglomeration process. This product is under co-development by EPRI and Ultramax Corporation, an industrial partner. With the introduction of this optimization tool, a different approach was used for parametric testing. Initial testing generated POC performance data for a range of major test variables including diesel and frother concentrations, high-shear mixing intensity, flotation water levels, and air flow rates.

These data, along with POC data from Upper Freeport and Pittsburgh No. 8 coal testing, were then used to construct the selective agglomeration model used by Ultramax to help guide the Illinois No. 5 coal parametric testing. Primary model constraints called for maximizing PSR at a minimum 85% Btu recovery while taking relative process costs into consideration. POC testing was guided by recommendations from Ultramax rather than by a fixed matrix test design.

Prior to and during parametric testing of each coal, batch agglomeration tests were conducted in the Wilsonville laboratory. The primary objectives of these tests were to help guide POC testing and provide data to compare agglomeration performance at the two scales of testing. Batch testing on natural minus 100-mesh Illinois No. 5 coal fines was also completed to help predict continuous process performance for a natural fines feedstock.

Following the completion of parametric testing for a given coal, an extended POC test run was conducted. This demonstration run was of approximately four days' duration and was carried out using the process conditions deemed optimal from the parametric testing. A demonstration run was completed for each project coal.

The final POC operations involved the processing of pond fines generated from conventional coal cleaning. The pond fines sample was from JWR No. 3 mine tailings pond. The primary objective of this work was to determine POC performance while processing the minus 60-mesh fraction of this feedstock. Since the pond fines were relatively low in sulfur content (approximately 0.25%), performance was based on the yield of the minus 60-mesh fraction and the product ash content. Additional objectives for POC processing of this minus 60-mesh pond fines material were to:

- Determine POC performance with the application of high-shear agglomeration
- Determine POC performance without the use of high-shear agglomeration
- Generate yield vs. product cumulative percent ash performance curves for both of the above scenarios
- Produce a 15% ash product at the highest yield possible for both processes

Following required plant modifications, testing to fulfill these objectives was completed in five days of POC operations. The effect on plant performance of a number of process variables including agglomerant type, agglomerant concentration, and frother concentration was investigated during this testing.

POC Configuration

As a result of the Phase I work, a diesel-based selective agglomeration process was chosen for further study. This process involves the following scenario for cleaning ROM coals:

- ROM coal is precleaned utilizing conventional coal cleaning technologies.
- The conventionally cleaned coal is then ground to the desired topsize and subjected to high-shear agglomeration, followed by two stages of microagglomerate recovery via froth flotation.
- The recovered microagglomerates are mechanically dewatered, pelletized, and thermally dried to produce the final process product.

Based on evaluations completed early in Phase II, it was determined that conventionally precleaned coals would be obtained for POC feedstock. This precleaned coal was ground, agglomerated, and mechanically dewatered in the POC facility. The dewatered POC product was then sent to vendor laboratories for pelletization and drying evaluations.

A flowsheet for the POC facility is shown in Figure 3-2. The facility was designed to process 2-t/h of as-received feedstock for all three project coals. The following sections provide descriptions of the various POC circuits depicted in Figure 3-2.

Coal Receiving, Handling, and Crushing System

Coal was received from haulage trucks and stored on the feed coal pad. The feedstock was then recovered from storage by a front-end loader and fed to the raw coal hopper, from where it was conveyed to a 24-ton storage bin. A vibrating feeder was used to control the storage bin discharge into the coal crusher, where the topsize was reduced to 1/4 inch. The crushed coal was then conveyed to a 3-ton crushed coal bin. The coal receiving, handling and crushing system is shown in Figure 3-3.

Coal Grinding Circuit

In the POC coal grinding circuit, coal was withdrawn from the 3-ton crushed coal bin by a weigh-belt feeder that controlled the coal flow to the ball mills. Water was added to the ball mills to produce a 45% solids concentration slurry for grinding. The ball mills were calibrated for each coal such that operation of the primary mill alone produced a 100-mesh topsize product; operating both the primary and secondary mills in series produced a 200-mesh topsize product. The primary ball mill and sump are shown in Figure 3-4.

High-Shear Agglomeration Circuit

After the coal was ground to the required topsize, the ball mill product was diluted to the desired solids concentration, typically 20%, and pumped to the high-shear circuit. The high-shear circuit consisted of two high-shear agglomeration vessels. These vessels had volumes of 150 and 45 gallons which provided high-shear residence times of approximately 3 and 1 minutes respectively, at a 20% high-shear solids concentration. Figures 3-5 and 3-6 show the 150- and 45-gallon high-shear vessels, respectively.

Both of these high-shear vessels were provided with the ability to operate at various rpm rates, thus allowing determination of the high-shear conditions that require the minimum power consumption while achieving successful agglomeration. As shown in Figure 3-2, just prior to the piping split in the high-shear circuit, diesel fuel was added as the agglomerant. Following agglomeration, the high-shear product slurry was pumped to the microagglomerate recovery circuit.

Microagglomerate Recovery Circuit

The agglomerated product from the high-shear circuit was subjected to separation via flotation. The primary separation of the agglomerated coal from the mineral matter-bearing process water was made in the first stage of flotation (6 cells). The first-stage flotation product was then refloatated in a second stage of flotation (6 cells). Flotation recovery of microagglomerates was typically completed at 6% solids concentration, resulting in a flotation residence time of approximately 6 minutes per stage of flotation. Figure 3-7 depicts both the first-stage and second-stage banks of flotation cells. Four metering pumps were provided for frother addition throughout the flotation circuit. Typically, there were two frother addition points for each stage of flotation.

Product, Tailings, and Water Handling Circuits

The product handling circuit utilized for the POC facility is also shown in Figure 3-2. The second-stage flotation product was dewatered using the rotary drum vacuum filter shown in Figure 3-8. The filtrate was recycled to the clarified water tank, and the filtered product was conveyed to the product pad for loadout. Although the rotary drum vacuum filter was the primary dewatering tool used during POC testing, a high-G centrifuge was also tested.

Process tailings, consisting of flotation tailings, were collected in the tailings tank and pumped to the thickener for water clarification. The thickener is shown in Figure 3-9. Clarified water from the thickener was recycled to the head tank, and the thickened refuse was dewatered in the plate-and-frame filter presses, shown in Figure 3-10, and then transferred to the refuse pad for loadout.

Reagent Delivery Systems

The reagent delivery systems for the POC facility were designed to allow testing over a wide range of reagent levels. Reagents used in the POC facility included:

- High-shear agglomerant (diesel)

- Flotation frother (Betz M150)
- Flocculant for the water clarification circuit (Betz CM 224, anionic polymer emulsion)
- Caustic for pH control

The diesel delivery system consisted of a 7000-gallon storage tank, a centrifugal positive displacement metering pump, and associated piping. Delivery systems for both the frother and flocculant consisted of a small storage vessel, metering pumps, and associated piping. Caustic for pH control was added manually on an as-needed basis.

Figure 3-3. POC Coal Receiving Handling and Crushing System

Figure 3-4. POC Primary Ball Mill and Sump

Figure 3-6. POC 45-Gallon High-Shear Vessel

Figure 3-5. POC 150-Gallon High Shear Vessel

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Figure 3-7. POC First- and Second-Stage Flotation Cells

Figure 3-8. POC Return Conveyor System

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Figure 3-9. POC Thickener

Figure 3-10. Plate and Frame Filter Presses

POC Operations Summary

Modifications were made to the system based on recommendations from analysis of the Phase I results. These modifications were begun in November 1991 and continued through startup in March 1992. A summary of the POC operation follows. A schedule of activities for Phase II is shown in Table 3-6.

Table 3-6
POC Test Schedule

Activity (1992)	Begin	End
Modifications & startup activities	Nov. 1991	March 1992
Upper Freeport mill calibration & startup	19 March	28 April
Upper Freeport optimization testing	29 April	2 June
Upper Freeport demonstration run	3 June	8 June
Pittsburgh No. 8 mill calibration & startup	11 June	15 June
Pittsburgh No. 8 optimization testing	15 June	22 July
Pittsburgh No. 8 demonstration run	27 July	31 July
Illinois No. 5 mill calibration & startup	4 Aug.	13 Aug.
Illinois No. 5 optimization testing	14 Aug.	3 Sept.
Illinois No. 5 demonstration run	8 Sept.	12 Sept.
Pond Fines modifications & operation	14 Sept.	2 Oct.
Dismantling	5 Oct.	7 Dec. 1992

Upper Freeport Operation

A detailed log for all POC operations is provided in Appendix D, POC Operations Log. Calibration of the secondary mill was begun with Upper Freeport coal on 29 January 1992, with a target of 200-mesh topsize and a d_{50} of 25 μm , and continued through the end of February 1992. Ball charge was added incrementally as needed to approach the target grind size. Primary mill calibration began near the end of February with a target topsize of 100 mesh. A water balance was conducted on 6 February. Concurrently with mill calibration, equipment and system checks were conducted, as well as calibration of pumps, meters, and flow control valves. Orientation testing was conducted with slurry to begin shakedown of the systems, allow training of operating and laboratory personnel, and identify approximate performance. Minor modifications to system piping and instrumentation were made during this orientation period. The rotary drum vacuum filter arrived on site on 21 February and was started up on 5 March. The operability and performance of the drum filter were good throughout POC operation.

Once the mill grind size was reached and shakedown had been completed, actual test runs began. The first run was conducted on 19 March 1992. A total of 15

calibration/orientation runs were completed (Runs 1-15) with a range of diesel concentrations, various high-shear mixer configurations (150- and 45-gallon high-shear vessels), and 100- and 200-mesh coal feeds. A major accomplishment during these runs was the lining out of flotation cell operation with respect to cell levels, blower rates, and frother and solids concentrations. Sample collection and flow rate measurement methods were fine-tuned, as well as laboratory procedures. Additional modifications to piping, instrumentation, and procedures were identified and implemented.

Runs 15-19 were conducted to evaluate the optimum high-shear mixer configuration. This included residence time (150- vs. 45-gallon high-shear mixer, since the coal feed rate was kept constant at 2 t/h), solids concentration (15 and 20%), and mixer rpm. The diesel addition rate was varied somewhat during the tests, and flotation cell performance was optimized by adjustment of cell levels, blower rates, and frother addition. Several set points were obtained during each run, and the number of variables that were changed from test to test was minimized to allow reliable comparison of changes in system performance. Samples and flow rate measurements were obtained at each set point for material balances.

Optimization of flotation cell performance with Upper Freeport coal was conducted during Runs 20-29. The variables tested were flotation cell feed solids concentration, cell levels, frother addition rates, and blower air rates. In addition, various methods of froth washing via spray nozzles were tested. As in high-shear optimization testing, several set points were obtained during each run, with minimal changes in variables, to evaluate the impact of the respective changes on flotation performance. The tests were conducted with 0.5% diesel using the 45-gallon high-shear mixer. Several mixer speeds were tested during the period.

Run 30 was conducted on 1 June to demonstrate repeatability at the optimum conditions established during Runs 1-29, as determined by the results of four consecutive sets of one-hour composite samples. The run was conducted with 200-mesh slurry in the 45-gallon high-shear mixer at 20% solids and 0.5% diesel. The flotation cells were operated at 6% solids, and the frother was maintained at 0.29 lb/ton throughout the run. The 45-gallon high-shear mixer was kept at 950 rpm, and the cell levels maintained at 3 inches below the overflow. Four sets of 1-hour composite samples and flow rate measurements were collected at steady state.

The last parametric test utilizing the Upper Freeport coal, Run 31, was conducted on 2 June to simulate conventional flotation operation at lower oil levels. The run was conducted with 200-mesh slurry, 20% solids in flotation feed, and 0.1% diesel. The initial frother concentration was 0.29 lb/ton. The solids concentration to high shear was 6%. Slurry was fed to the 45-gallon high-shear mixer, where several mixer speeds were tested. In addition, the high-shear mixer was bypassed for one set point.

Upper Freeport Demonstration Run. Prior to the Upper Freeport demonstration run, it was decided to test two diesel concentrations: 0.5 and 0.125%. The majority of the demonstration run was conducted at 0.5% diesel, since more operation had been conducted under those conditions. Once sufficient demonstration run data had been generated for POC evaluation and the trace element removal study, the flotation cells were adjusted to reject more pyritic sulfur by using cells 5 and 6. The diesel concentration was then decreased from 0.5 to 0.125% to obtain two additional sets of samples at low-oil concentrations for comparison with the data generated during Run 31.

The Upper Freeport demonstration run was conducted during 3-8 June 1992 to demonstrate the operability and performance of the system, as well as to collect product for additional DOE projects. The coal feed rate was kept at 2 t/h, and the 200-mesh ball mill discharge slurry was fed to the 45-gallon high-shear mixer at 20% solids and 0.5% diesel. The mixer was operated at 950 rpm with the flotation cells at 6% solids. Frother was added at a 0.29 lb/ton rate. Steady-state samples and flow rate measurements were obtained throughout the run. The coal feed to the plant was sampled via stop-belt sampling to generate eight-hour composite samples, with a minimum of 14 increments per composite and a targeted minimum of 2 lb/increment. Ball mill discharge samples were taken every 12 hours for particle size and solids concentration measurements and control. Composite samples of the flotation cell feed and discharge streams were collected with mechanical samplers over four-hour periods, with 40 increments per sample. These samplers were located at the flotation feed, first- and second-stage flotation product, and first- and second-stage flotation tailings. Samples of the rotary drum vacuum product were obtained every four hours for moisture determinations, and the recycle water was monitored for pH. Flow rate measurements were taken every four hours at streams that did not have instrumentation, and hourly readings were taken of other plant parameters.

During the early part of the demonstration run (3 and 4 June) the primary mill pump flow began to become erratic, as if something were lodged in the suction. The suction piping was flushed with high-pressure water from time to time to help improve the flow. Also, the hammermill coal crusher continued to plug frequently with damp coal. Primary mill and hammermill operation continued to be a problem area and the primary mill pump stopped pumping early on 6 June. The unit was shut down and the pump replaced with one from the neighboring project, a coal liquefaction pilot plant. The pump motor failed upon startup, requiring location and installation of another motor that would operate with the primary mill pump variable-speed drive. The replacement motor was installed, and coal feed began after four hours of down time.

Operation continued with pluggage of the hammermill crusher occurring frequently. The vibrating feeder to the crusher was slowed down to provide a feed that could be handled by the crusher, but this feed rate was too low to keep up with the POC requirements. The plant was finally taken off coal feed on the morning of 6 June due to loss of coal inventory in the 3-ton mill feed tank. The wet coal in the 24-ton storage bin was emptied and replaced with drier coal, but the crusher continued to plug. The hammers in the crusher were worn and needed to be replaced or rebuilt with addition of welded metal on the hammers. Since no replacement hammers were in stock, the crusher discharge screen was removed so coal could be fed directly to the ball mills. This appeared to be the best method of getting back on line. An additional 500 lb of balls were added to the primary mill to ensure adequate grinding of the coal. Coal feed was once again started after about 15 hours of down time.

On 7 June, the flotation cells were adjusted to reject more pyritic sulfur. The weir bars in the overflow from cells 5 and 6 of each stage were removed, and the paddles were adjusted to recover froth from those cells. The flotation cells were lined out and two steady-state one-hour composites obtained.

The diesel concentration was decreased from 0.5 to 0.125% late on 7 June, per the demonstration run plan, to obtain two additional sets of samples at low oil concentrations for comparison with the data generated during Run 31. Two sets of steady-state samples and flow rate measurements were obtained. The unit continued to operate until the morning of 8 June, when the run was halted and the POC shut down per normal procedure. The unit operated for 96.25 hours of an available 112.95 hours for an on-stream factor of 85.2%. Approximately 100 tons of product (dry coal basis)

amounting to 765 drums was collected for testing by Combustion Engineering, and an additional 16 drums of product was collected for other tests. Upon completion of the run, all remaining Upper Freeport coal was transported to the Alabama Power Company E. C. Gaston steam plant coal pile adjacent to the facility, and the coal pad was cleaned and washed down in preparation for delivery of Pittsburgh No. 8 coal later on 8 June.

Pittsburgh No. 8 Operation

Mill calibration for Pittsburgh coal began with Runs 32 and 33 (11 and 15 June). The coal crusher hammers were rebuilt prior to startup with Pittsburgh No. 8.

Orientation and optimization with Pittsburgh No. 8 coal was conducted during Runs 34-55. A number of parameters were evaluated, as with Upper Freeport operation, including coal feed topsize (100 and 200 mesh), high-shear mixer rpm, and diesel concentration. The solids concentration to the high-shear mixer and the first-stage flotation cell feed were kept constant throughout the runs (at 20 and 6%, respectively), based on experience with Upper Freeport coal. The 45-gallon high-shear mixer was used for all runs except a portion of Run 51 and Run 55. The variables tested at the flotation cells were cell levels, frother addition rates, and blower air rates. Froth spray was not tested since no significant impact was identified during Upper Freeport operation.

The high-G centrifuge arrived from Alfa-Laval Sharples on 6 July. Run 52 was conducted on 17 July to allow shakedown of the high-G centrifuge. Slurry was fed to the centrifuge upon startup, and product samples were taken for moisture determination at 0.5, 1.0, and 1.5% diesel concentrations in the 45-gallon high-shear mixer. Moisture content for these centrifuge products were on the order of 38-40%. Additional dewatering evaluation testing was conducted during Run 54 on 21 July with 100-mesh coal. The centrifuge did not perform well during this run producing an erratic solids discharge with approximately 33-35% moisture. The rpm of the centrifuge was varied but no visible impact on product appearance was noticed. Flow rate measurements and samples were taken at the centrifuge per the manufacturer's recommendation. (An Alfa-Laval Sharples representative was on site for startup of the unit.) The plant was shut down after two sets of data were collected. The centrifuge was again tested during Run 55 on 22 July to evaluate dewatering at lower diesel rates. Once again, centrifuge cake discharge was inconsistent with moisture contents of approximately

40%. It is believed that for this Pittsburgh No. 8 coal centrifuge testing, inconsistent centrifuge feed caused the observed erratic cake discharge. The high cake moistures produced by the centrifuge were probably due to a combination of inconsistent feed and non-optimal operation of the centrifuge.

Pittsburgh No. 8 Demonstration Run. The Pittsburgh No. 8 demonstration run was conducted during 27-31 July 1992 to demonstrate the operability and performance of the system and to collect product for other DOE projects. As with the Upper Freeport demonstration run, two set points were tested during the Pittsburgh No. 8 demonstration run. The first set point consisted of 0.5% diesel and approximately 0.1 lb/ton total frother to the flotation cells. The second set point, conducted near the end of the demonstration run, used a diesel concentration of 0.125% and a total frother rate of 0.1-0.3 lb/ton.

The Pittsburgh No. 8 demonstration run began on the morning of 27 July. The run was conducted using 2 t/h of 200-mesh coal fed to the 45-gallon high-shear mixer, which was initially operated at 20% solids, 1200 rpm, and 0.5% diesel. The flotation cells were operated at 6% solids and approximately 0.1 lb/ton frother. Product dewatering was done via the rotary drum vacuum filter. Sampling and flow rate measurement procedures were the same as those used in the Upper Freeport demonstration run.

The 45-gallon high-shear mixer tripped off for about 30 minutes on 27 July. Sampling was stopped until steady state was again achieved. Operation continued until the vacuum pump for the rotary drum vacuum filter stopped working on 29 July. The shaft of the unit could not be turned, and the pump had to be removed and taken to the maintenance shop for repairs. Flotation cell product was diverted to the plate-and-frame filter presses. After a short duration, one of the filter press feed pumps stopped operating. While the reason for the failure of the pump was under investigation, the POC had to be shut down due to the inability of the single bank of presses to keep up with the full product flow. The coal feed was restarted after about four hours once the filter press feed pump had been put back in service. (Coal had entered the solenoid of the feed valve through the instrument air supply at the filters after the failure of a check valve in the instrument air supply to the core blow of the filters.) The rotary drum vacuum filter was placed back in service shortly thereafter. The vacuum pump had to be disassembled and washed with acid to remove excessive buildup of calcium from the impellers and walls of the pump housing. (Calcium had slowly built up over time

due to the hard seal water makeup.) Steady state was achieved and no further problems were encountered for the duration of the demonstration run.

The diesel concentration was lowered from 0.5 to 0.125% on 30 July, per the demonstration run plan. Correspondingly, the frother was increased from 0.1 to 0.3 lb/ton. Steady state was achieved, and five sets of samples were taken. The unit was taken off coal feed on 31 July. The coal feed was maintained for 97.16 hours of operation out of 101.45 available hours for an on-stream factor of 95.77%. Approximately 50 tons of product (dry coal basis) amounting to 500 drums was collected for testing by Combustion Engineering, and an additional 16 drums of product was collected for other tests. Upon completion of the run, all remaining Pittsburgh No. 8 coal was transported to the E. C. Gaston steam plant coal pile next to the facility, and the coal pad was cleaned and washed down in preparation for delivery of Illinois No. 5 coal.

Illinois No. 5 Operation

Mill calibration and orientation for Illinois No. 5 operation was conducted during Runs 56-61 (4-13 August). During these runs a number of parameters were tested, including coal feed topsize (100 and 200 mesh, predominantly 100 mesh), high-shear mixer rpm, and diesel addition. The solids concentration to the high-shear mixer and first-stage flotation cell feed were kept at 20 and 6%, respectively, and the 45-gallon high-shear mixer was used for all runs. The variables tested at the flotation cells were cell levels, frother addition rates, and blower air rates.

Runs 62 and 63 focused on high-shear mixer rpm testing (400 and 700 rpm) with the other conditions listed above. Runs 64-68 were all conducted with the 45-gallon high-shear mixer at 700 rpm, with changes in the diesel concentration. In all seven runs, the flotation cell levels, frother additions rates, and blower air rates were adjusted as needed to obtain the desired recovery based on operating experience.

Runs 62 and 63 (14 and 17 August, respectively) were also used to conduct additional high-G centrifuge tests. In Run 62, the performance of the centrifuge for dewatering product slurry with minimum frother concentrations and lower high-shear rpm was evaluated. The POC was operated with 100-mesh coal to the 45-gallon high-shear mixer, which was operated at 20% solids, 0.5% diesel, and an initial rpm of 700. The flotation cells were operated at 6% solids and a frother

concentration of 0.1 lb/ton. An additional 1.0% diesel was added to the second-stage flotation product tank to facilitate dewatering of the product slurry with the centrifuge. Run 63 was conducted to evaluate flotation cell and centrifuge performance with 100-mesh Illinois No. 5 coal at a 2-t/h feed rate, using the 45-gallon high-shear mixer at 400 and 700 rpm and 0.5% diesel addition. Frother concentration was 0.15 lb/ton. The solids concentration to high shear and the flotation cells were 20 and 6%, respectively. Two set points were obtained and steady-state samples and flow rate measurements were taken. The centrifuge was tested at an rpm of 4000, with no dam in the bowl (allowing the shallowest pool depth and thus the driest product). The second-stage flotation product was diverted to the centrifuge at several different flow rates. In addition, 1.5% diesel was added to the second-stage product tank to evaluate the impact of increased oil levels on product dewatering performance. During centrifuge testing with the Illinois No. 5 coal, product cake moistures in the range 36-43% moisture were observed. Attempts to achieve a uniform discharge of the centrifuge cake were not successful, probably due to unsteady-state operation of the centrifuge. The centrifuge was shipped back to the vendor on 25 August.

Runs 69-73 were used to optimize flotation cell performance. All runs were conducted at 20 and 6% solids to high shear and flotation, respectively, and with the 45-gallon high shear mixer, except a portion of Run 72 and the last part of the demonstration run, as discussed below. Diesel addition to high shear and high-shear rpm were varied per run instructions, and cell levels, frother addition, and blower air rates adjusted for desired recovery. Run 73 was conducted to repeat previous conditions that had the most promising results to help in finalizing run conditions for the Illinois No. 5 demonstration run.

Illinois No. 5 Demonstration Run. Prior to the demonstration run, the rotary drum filter vacuum pump was cleaned with acid to prevent shutdown during the run due to buildup of calcium deposits. The Illinois No. 5 demonstration run was started on 8 September with 100-mesh coal at a 2-t/h feed rate. The run was conducted with the 45-gallon high-shear mixer, which was operated at 900 rpm and 20% solids. Diesel addition to high shear was 0.8%. The flotation cells were operated at a solids concentration of 6% and 0.125 lb/ton frother. Blower rates and cell levels were adjusted to obtain the desired recovery. Sampling and flow rate measurement procedures were the same as those used in the other demonstration runs.

Sampling continued for the first six 4-hour periods until the middle of 9 September when a plant upset occurred. The frother levels in the buckets feeding the metering pumps had become low, and frother was added from a new drum. The new frother was a different type, Betz MS150 instead of M150. (An order for M150 placed with Betz had resulted in the shipment of Betz MS150 due to the sale of the mining chemicals group of Betz to Stockhausen Mining. As a result, Betz no longer carried M150, and apparently sent the item most similar to M150 in name: MS150.) The MS150 reduced froth formation dramatically, so a drum of M150 was expedited from the local Stockhausen Mining representative later that afternoon. In the meantime, there was enough M150 frother in another bucket to divide between all the frother metering pumps until the new drum arrived. Steady state was again achieved and sampling begun after about three hours.

As operation proceeded, the high-shear mixer began to leak somewhat at the seal, with no method of stopping the leak. Operation continued with the leak, as opposed to shutting down. A power outage occurred on 10 September. The unit was off-line for only a few minutes, so steady state was not jeopardized. However, the vacuum drum filter vacuum pump had locked up during the outage and would not turn, despite the acid washing prior to startup. It was acid-washed in place and put back on-line after slightly more than two hours. During the interim, product was diverted to the plate-and-frame filter presses. An additional 12 set points were obtained through the afternoon of 11 September.

The 100-mesh coal from the primary mill was diverted to the secondary mill during the afternoon of 11 September to begin processing 200-mesh coal for the second set of conditions (200 mesh and 1.2% diesel). The 45-gallon high-shear mixer rpm was increased from 900 to 1050, and the diesel addition rate was increased from 0.8 to 1.2%. Over the next 30 minutes the leakage at the 45-gallon high-shear seal increased significantly as the smaller particles began to flow through the mixer. Flow was diverted to the 150-gallon high-shear mixer at an rpm of 848. Steady state was achieved at the new set of conditions, and three-hour composite sampling periods begun. A total of five sets of samples and flow rate measurements were obtained. The unit was taken off coal feed the morning of 12 September after 96 hours of continuous coal feed. The on-stream efficiency was 100% (based on coal feed). A total of 500 drums of product was obtained for another DOE project, as well as 16 drums for project use.

Pond Fines Operation

Upon completion of Illinois No. 5 POC operation, modifications were made to the POC unit to allow the processing of pond fines from JWR. The material did not require grinding, so the ball mills and associated tanks and pumps were removed from the structure. A feed tank was then installed from which pond fines could be introduced to the system. A conveyor was leased and installed beneath the existing reclaim hopper to feed the pond fines into the feed tank, which was equipped with a mixer and pump. (The crusher and normal POC feed conveyors were not used during processing of the pond fines.) The slurry was pumped from the feed tank through a low-shear vessel (to help mix the slurry) to static sieve bends for separation into plus/minus 60 mesh. The plus 60-mesh material was discharged onto a second leased conveyor and conveyed to the product pad for disposal. Samples of the plus 60-mesh material were collected in drums for spiral testing at the JWR facility. The minus 60-mesh slurry was pumped to the 150-gallon high-shear mixer and from there to the flotation cells per normal operation.

The second-stage flotation froth product collection launder was modified to allow collection of samples and flow rate measurements from individual cells for generation of ash versus yield curves. Baffles were placed in the existing launder and discharge nozzles suitable for connecting flexible hose were installed.

A pond fines shakedown run was made on 24 September. Upon shutdown, the need for system flush water was identified, so modifications were made to the recycle water header to allow introduction of water to the low-shear vessel. Pond fines operation began on 25 September. The coal feed was approximately 2.7 t/h (dry coal basis) to the high-shear mixer. Diesel was added at approximately 0.1% and frother added as needed for recovery.

An initial run after shakedown was made on 29 September with 0.06 lb/min diesel, or approximately 0.05%, to operate in the standard flotation mode that JWR uses. Only two set points were achieved due to power outages. The first set point was made with no high shear, and the second using the 150-gallon high-shear mixer. Frother addition and blower rates were adjusted as needed for recovery based on observation of froth. Samples and flow rate measurements were taken to allow generation of performance curves.

A second run was made on 30 September to evaluate the effect of diesel on performance. Two set points were

achieved, both using the 150-gallon high-shear mixer. The first set point was conducted at 0.07% diesel and the second at 0.14% diesel. The coal feed rate was targeted at approximately 2.5-3.0 t/h. Frother addition and blower rates were adjusted as needed for recovery based on observation of froth. Flow rate measurements and samples were collected at each set point.

A third run was made on 1 October to continue evaluation of performance at different diesel concentrations. Set point 1 was conducted with 1.14% diesel to high shear, and set point 2 was conducted at 0.19% diesel. The 150-gallon high-shear mixer was used. Frother addition and blower rates were adjusted as needed for recovery based on observation of froth. Samples and flow rate measurements were taken at each set point.

The final pond fines run was conducted on 2 October to evaluate the effect of using Betz M210 type promoter with the diesel addition (90/10 diesel/M210 blend). Three set points were obtained. Set point 1 was conducted with 0.1% blend added to the sieve bend underflow tank from which the high-shear feed slurry originates, utilizing the 150-gallon high-shear mixer. Set point 2 was conducted with the same addition rate of diesel/M210, but the 150-gallon high-shear mixer was bypassed. Set point 3 was conducted with the 150-gallon high-shear mixer, but the diesel/M210 addition rate was decreased from 0.1 to 0.03%. Samples and flow rate measurements were taken at each set point.

Dismantling of the selective agglomeration POC facility began on 5 October 1992.

Operating Problems and Recommended Solutions

The bulk of the operating difficulties experienced at the selective agglomeration POC facility in Wilsonville were associated with the coal feed system, specifically due to wet coal. Coal was stored on a concrete pad and added to a 24-ton storage bin, from which it was fed to a 3-ton ball mill feed bin. Much of the feed coal was relatively small in size, with approximately 90 to 95% less than 1/4 inch in size. If the coal was too wet, it would not come out of the 24-ton storage bin despite the use of a vibrating feeder at the bottom of the bin, vibrators on the sides of the bin, and a relatively steep angle of repose in the cone bottom of the bin. Once the coal exited from the bin, it would often plug in the hammermill crusher.

One procedure which helped reduce feed problems was to keep the coal pile covered with a waterproof tarp, thereby minimizing accumulation of moisture in the coal

from rainwater, etc. Although this did not eliminate the problems of wet coal entirely, it helped somewhat. A roof over the coal storage area would be advantageous. Also, the coal was turned over with a front-end loader on dry days to allow sun and wind to help evaporate moisture. A cover was placed on the 24-ton storage bin to keep water from a full bin of coal. In addition, even when the coal was kept relatively dry, when the storage bin was filled with coal and allowed to sit for a length of time, the moisture in the coal would drain to the bottom resulting in very wet coal at startup of the next run. As a result, the amount of coal placed in the bin was kept at about 1/2 or 1/3 of its capacity to minimize the quantity of wet coal to be handled. A second vibrator was installed on the 24-ton storage bin at the start of POC operations to assist with removal of coal from the bin.

Routine inspection and proper maintenance of crusher hammers was also important in keeping coal moving through the system. As the hammers became worn, coal accumulated more readily inside the crusher and in the crusher feed chute. During POC operation, a plant air connection was installed next to the crusher for use in unplugging the hammers and chute. Plant air connections should be supplied at the crusher and other potential coal plugging locations for assistance in unplugging restrictions when they occur.

Another location where plugging occurred during POC operation was the ball-mill feed chute. Coal falling off of the weigh-belt feeder entered the top of the chute. Also, the ball mill feed water was added in the chute. If the feed water flow was interrupted or inadvertently decreased, the coal would plug in the chute at the mill inlet. The recycle water inventory and supply systems were monitored frequently during each run to ensure that loss of flow did not occur. When pluggage did occur, the coal feed was stopped immediately and the feed chute cleaned and flushed with water prior to startup. An additional thickener overflow storage tank was installed to increase the available clarified water capacity.

Frequent addition of coal to the ball-mill feed hopper was required during the course of a run, resulting in frequent starting and stopping of coal conveyors. Additional switches for the conveyors were installed to permit conveyor operation from both ends, which greatly reduced the time needed to keep adequate coal inventories. Other projects should carefully consider location of such switches to provide easier and more reliable operation, as long as safe operation is maintained.

Steady-state operation of the flotation cells was hindered during early POC runs due to erratic level control within

the cells. Turbulence and froth buildup at the level controller bubble tube (where the cell level is sensed) caused the level signal to the controller to fluctuate. After review of the problem and discussion with the flotation cell vendor, the level controller bubble tubes were relocated from the end boxes to within a representative cell. This significantly reduced the impact of turbulence and froth buildup at the bubble tube site, and therefore smoothed out the level readings and corresponding cell level control.

Grab samples and manual flow rate measurements were taken routinely throughout the unit. Several sample collection and flow rate measurement locations were such that sufficient room was not available initially to obtain the samples or measurements, requiring modifications to the piping or procedures. In future projects, consideration should be given to the procedures and space required for sampling and flow rate measurements during equipment layout and design.

During startup of the POC unit, the thickener overflow pump, which supplies recycle water to the ball mills and flotation cells, would not deliver sufficient flow to accommodate all of the water feed control valves. The POC water feed requirements were higher than those of the CDTF due to operation of the flotation cells in series and lower solids concentrations in the cells. The thickener overflow pump was replaced with one requiring less suction head, resulting in increased capacity to the recycle water head tank. The original pump required a suction head of 10 ft, while the thickener overflow tanks normally operated with a level of about 3 or 4 ft. Another pump with insufficient capacity was the plate-and-frame filter press filtrate pump. The pump used during CDTF operation was not capable of pumping recycle filtrate to the water supply system. This resulted in continued overflow of the filtrate tank to the process pad. This pump was replaced with a pump of greater capacity for POC operation. In general, pumps should be selected with sufficient capacity to accommodate startup, shutdown, and upset conditions.

Another source of pump capacity problems was the impellers. Over time, the coal slurry eroded the impeller vanes, resulting in decreased pumping capacity as the impellers became worn. The impellers were replaced when pumping capacity decreases were identified. Impellers in slurry service should be inspected regularly and replaced as needed to avoid capacity problems.

During POC operation, there were several occasions when water or coal/water slurry entered the instrument air header when air pressure was lost as a result of air

compressor shutdowns. This resulted in the loss of plant operation for several days on separate occasions until the affected instruments were cleaned and dried. Instrument air supply piping should be equipped with check valves or some other mechanism to protect instrumentation from water in the event of air compressor failures.

For POC operation, a rotary drum vacuum filter was installed downstream of the flotation cells for product dewatering. Calcium and magnesium present in the seal water to the pump deposited on the rotor vanes and rotor chamber walls during normal operation. Occasionally, the accumulation of deposits became large enough that the vacuum pump would not rotate at startup. When this occurred, acid washing of the rotor and vanes was required. Future users of this or other vacuum systems should provide for routine cleaning of the pump via acid washing, if seal water to the vacuum pump contains calcium and magnesium. If possible, soft water should be used for seal water to reduce maintenance and down time of the vacuum pump. With the exception of the calcium and magnesium buildup, the rotary drum vacuum filter performed well. The filtrate was clear and no dust was created.

During operation with Upper Freeport coal, the pH of the mill discharge dropped significantly compared to the inlet pH (4.5-6 vs. 7-8). Sodium hydroxide was added to the recycle water system to increase pH to more desirable levels (7-8). A similar drop in pH did not occur during Pittsburgh No. 8 and Illinois No. 5 operation, so the addition of sodium hydroxide was not necessary. Recycle water pH should be monitored and controlled, when necessary, to minimize corrosion of piping, vessels, etc.

Plant And Equipment Design

Prior to POC operation, the ball mill feed chute was modified with the addition of a second mill feed pipe and block valves to divert coal to the desired mill. These changes allowed switching of coal feed from one mill to the other without moving the single mill feed pipe originally installed for CDTF operation.

The flotation cells were reconfigured between CDTF and POC operation to improve product quality. The parallel flotation cells used during CDTF operation were placed in series to provide two-stage separation, with the first-stage product repulped and refloats in the second-stage cells. Also, additional cells were purchased and

added to each bank to provide increased residence time for testing of lower solids concentrations.

Sampling of the CDTF during Phase I of the project identified the need for automatic samplers on critical process streams. As a result, automatic mechanical samplers were installed on flotation feed, product, and tailings streams to provide more reliable, representative, composited samples of flotation performance. The sample volume and frequency were adjusted via an automatic controller as needed. In addition, mechanical sampling of the POC feed prior to grinding (crusher product) would have provided additional useful information. This data would help monitor the effect of grinding and high-shear agglomeration on feed coal characteristics, i.e., total sulfur and sulfur forms content.

Manual valves were used during CDTF operation for the addition of water to the ball mills, high-shear feed, and flotation cells. While this was adequate for component development, steady-state parametric studies for Phase II POC operation required more accurate, reproducible water addition rates. To provide for reliable water addition in the POC unit, automatic flow control valves and a central control panel were installed. The valves eliminated the frequent manual calibration of water flows required during CDTF operation when process conditions were changed and eliminated variation in flows due to inaccurate manual setting of valves for specific flow rates.

Accurate metering and measurement of agglomerant (diesel fuel) and other reagent flows was necessary for steady-state operation and to provide reliable comparison and analysis of various runs. The capacity of the diesel and frother pumps was designed to provide better accuracy over the range of flows desired.

During CDTF operation, the froth product from both flotation cells discharged into a microagglomerate recovery tank. The froth tended to float in the tank instead of mixing to form a consistent slurry, resulting in intermittent loss of product flow caused by pluggage of solids in the microagglomerate pump suction. In addition, the mixer installed on the microagglomerate tank was too small for adequate mixing of the tank contents, and the tank diameter-to-height ratio (2-3 to 1) was too large. For POC operation, the original common microagglomerate tank was replaced by two tanks, one for each stage of the flotation cells. The new tanks were smaller in diameter and volume, which decreased the residence time in the tank. The diameter-to-height ratio was decreased to 0.5, and mixers with increased horsepower and appropriate impeller design were

installed. These modifications eliminated the problem of froth buildup in the flotation product tanks and pluggage in the flotation pump suction piping. Slurry tanks and mixers for other DOE projects should be designed to provide sufficient mixing of the slurry to minimize plugging in pump suctions and interruption of steady-state conditions.

During the course of operation, piping changes were identified and modifications made to provide flexibility in diverting process slurry flows to refuse dewatering equipment during upsets or plant startup and shutdown. Extra tees and block valves should be provided for future use in piping layout during initial design of a facility, particularly in the refuse dewatering and tailings/refuse handling systems.

During cold weather several runs were delayed due to damage to piping in the recycle water systems. Procedures were developed to drain all piping and vessels as needed to prevent freezing, and heat tracing and insulation were added where draining of the system was not feasible. Other DOE installations should identify freeze protection requirements and design systems including the addition of heat tracing, insulation, operator procedures, etc.

The addition of on-line coal quality measurements to monitor flotation tailings stream ash content would provide helpful information. This type of data could be used for feedback control of flotation operating conditions to help maintain constant plant Btu recovery.

POC Results Summary

This section presents a summary of the Phase II test results for all four POC facility feedstocks. The results of all POC test runs are included in Appendix E, POC Results Summary.

For the Upper Freeport and Pittsburgh No. 8 coals, raw and precleaned coal characteristics and precleaning yield were used to determine precleaning performance. These precleaning performance estimates were then applied to POC selective agglomeration results to determine final POC performance on a run-of-mine (ROM) coal basis. For the Illinois No. 5 coal and JWR pond fines, POC selective agglomeration performance values are reported directly on a ROM basis since these coals were not precleaned prior to POC testing. All POC performance values reported in this section of the report are on a ROM coal basis.

Upper Freeport Coal

Upper Freeport seam coal from Jefferson County, Pennsylvania, supplied by P&N Coal Company and precleaned by CQ, Inc., was the first coal tested in the POC facility. All initial plant equipment calibration and startup testing was executed using this feedstock.

Startup and Calibration Testing. A primary objective of POC startup and calibration testing was calibration of the ball mills to produce nominal 100- and 200-mesh topsize products from the primary mill alone, and the primary and secondary mills in series. Typical 100- and 200-mesh topsize particle size distributions (PSD) from this mill calibration testing are presented in Figure 3-11. These POC feed size distributions are typical for all POC testing utilizing the Upper Freeport coal.

Figure 3-11. Upper Freeport Coal Particle Size Distributions

While initial PSD definitions called for 95% passing the desired topsize, it can be seen from Figure 3-11 that only about 80% of the material passes the 100- and 200-mesh sizes, 150 and 75 µm, respectively. The presence of this coarse fraction of coal is indicative of ball mill open-circuit grinding. Less coarse material would have been generated if a rod mill and ball mill had been used in series rather than two ball mills. The rod/ball mill configuration was not used, however, due to the impracticality of removing charge from the ball mill to switch from the 200- to 100-mesh topsize grind. Closing the mill circuit with a cyclone would have provided a narrower PSD of low solids concentration (2-4%), which would require thickening prior to high shear agglomeration. Even though these PSDs show approximately 80% rather than 95% passing the desired topsize, the same

50% passing (D_{50}) sizes of 60 and 25 µm, for 100- and 200-mesh topsizes respectively, used during Phase I testing were targeted and achieved for all project coals.

Washability Analyses. Once the POC ball mills were calibrated to produce the desired feed coal particle size distributions, primary and secondary ball mill product samples were taken for washability analyses. Washability data from these analyses are presented in Appendix C, POC Test Coal Washabilities. A plot of PSR vs. Btu recovery (precleaned coal basis) for these washability data is shown in Figure 3-12. These results indicate that about 3-4% additional PSR is achieved at Btu recoveries of 85-95% when the coal topsize is reduced from 100 to 200 mesh.

Figure 3-12. Upper Freeport Coal Washability Analyses

Orientation Testing. Concurrently with the execution of POC grinding calibration and startup tests, POC orientation runs were conducted to evaluate a range of high-shear and flotation circuit operating conditions. POC Runs 1-15 fall into this category, and their results were analyzed to develop test conditions for subsequent high-shear and flotation circuit optimization testing. These initial POC runs tested the following operating conditions:

- 100- and 200-mesh topsizes
- High-shear diesel additions from 0.5 to 2.0%
- High-shear agglomeration at 1186 rpm in the 150-gallon high-shear vessel at 15% solids
- Two-stage flotation recovery of microagglomerates at solids concentrations of 6 and 10%

- Total frother addition to flotation of 0.15- 0.34 lb/ton

The detailed test results for these initial orientation runs are presented in Appendix E, POC Results Summary. A summary plot of Btu recovery and PSR for these runs is presented on a ROM basis in Figure 3-13.

Figure 3-13. Upper Freeport Coal POC Orientation Tests

Although these initial runs were considered startup testing, the results helped define typical POC selective agglomeration performance, i.e., Btu recovery and PSR, for the Upper Freeport coal. The following trends were observed:

- POC performance at the 200-mesh topsize, as expected, was generally better than at the 100-mesh topsize.
- Higher agglomerant (diesel) and frother concentrations had minimal effects on Btu recoveries.
- Btu recoveries were generally high, indicating that the Upper Freeport coal was very hydrophobic and easily recovered and that coal recovery needed to be controlled to the levels desired.
- Increased agglomerant concentration decreased froth mobility across the surface of the flotation cells
- At a fixed agglomerant concentration, good froth mobility was most easily achieved by increasing flotation cell water levels and appeared unaffected by most other operating variables

POC performance throughout these initial tests was greatly influenced by the operation of the flotation circuit. Flotation cell water levels varied from set point to set point due to malfunctions in sensing and control of flotation cell levels. These flotation cell controller problems made it difficult to maintain consistent coal recovery during both stages of flotation. Bubble tubes used to detect flotation water levels were modified and relocated. Float cell level controller adjustments were also made to improve their control response. A satisfactory level control was ultimately achieved within a single day's operation via these modifications. However, the repeatability of flotation cell water levels between operating days remained difficult to achieve.

High-Shear Optimization Testing. Based on the results of the initial orientation testing, a nominal 200-mesh topsize was chosen for continued POC testing of Upper Freeport coal. High-shear circuit optimization testing was carried out utilizing the following ranges of plant operating conditions:

- 200-mesh topsize grind
- High-shear agglomerant (diesel) concentrations of 0.1, 0.3, and 0.5%
- 15 and 20% high-shear solids concentrations
- Agglomeration in the 150-gallon high-shear vessel at 848 and 1186 rpm
- Agglomeration in the 45-gallon high-shear vessel at 190, 500, 950, 1200, and 1800 rpm
- 6 and 8% flotation solids concentration
- Total frother concentrations of 0.08-0.16 lb/t on a dry feed coal basis

The primary objectives of high-shear optimization testing were to evaluate the effects of high-shear circuit configuration, mixer rpm, and solids concentration on POC process performance, i.e., Btu recovery and PSR. The agglomeration and flotation unit operations represent a coupled system in which any high-shear effects are determined by analysis of the flotation product to provide overall process performance. With an installed POC flotation capacity of approximately 6 minutes' residence time per stage of flotation, consistently high coal recoveries were achieved for the very hydrophobic Upper Freeport coal.

During this high-shear testing, flotation frother and solids concentrations were kept constant within a given run. Other flotation parameters, however, such as air rates and water levels in the latter cells of each stage of flotation, were adjusted based on visual observation of plant tailings to maintain similar Btu recoveries for all set points. These adjustments allowed set points with similar Btu recoveries to be compared on the basis of PSR to determine any high-shear effects.

Unfortunately, high-shear effects on process Btu recovery were undoubtedly masked by these changing flotation conditions. It also appears, based on data presented in this section of the report, that changing the high-shear conditions had no effect on process PSR. This indicates that the flotation recovery of micro-agglomerates dominated this coupled process and the data should be viewed in this light.

Table 3-7 presents the results of high-shear circuit configuration testing. These data indicate that changing the high-shear residence time from 3 minutes to 1 minute (by means of the 150- and 45-gallon high-shear vessels respectively) had no effect on PSR. These data also show that similar results were achieved at high-shear solids concentrations of 15 and 20%, indicating that an increase in solids concentration had no adverse effect on POC performance. All set points presented in Table 3-7 were completed using 0.5% diesel to high shear and 0.16 lb/t frother to the flotation circuit.

Table 3-7
Upper Freeport Coal High-Shear Residence
Time and Solids Concentration Effects

Run No.	HS Vessel	HS rpm	HS % Sol	Btu Rec, %	PSR, %
16	150	848	15	95.2	76.0
	150	848	15	95.3	75.2
	150	848	20	95.8	73.4
	150	848	20	95.7	76.4
	45	1800	20	96.2	74.8
	45	1200	20	94.6	81.0
17	150	1186	20	95.9	69.6
	45	1800	15	96.4	71.0
	45	1200	15	96.6	70.1

The 45-gallon vessel was chosen for subsequent testing because it provided a more economical high-shear unit operation and achieved results similar to the 150-gallon vessel. Similarly, 20% rather than 15% high-shear solids concentration was selected for continued testing due to the improved economics.

Testing to evaluate the effect of high-shear rpm on POC performance was also carried out, and the results are presented in Table 3-8. These runs were completed using the 45-gallon high-shear vessel at 20% solids concentration and 0.5% diesel to high shear.

Table 3-8
Upper Freeport Coal High-Shear rpm
Effects

Run No.	HS rpm	Frother, lb/t	Btu Rec, %	PSR, %
19	1200	0.16	97.8	67.1
	950		96.9	74.5
	500		95.4	68.6
22	1200	0.16	93.5	79.5
	950		93.9	79.4
	600		92.6	76.2
29	1200	0.29	93.9	88.2
	950		91.8	83.2
	500		83.7	84.2

In comparing these data within a given run rather than between runs, a consistent effect of high-shear rpm on process performance is not clearly evident. While certain set points indicate improved performance, these trends are not consistent between runs. Based on these data, it was concluded that high-shear rpm had no significant effect on process performance. As a result, 950 rpm was chosen as representative and used for subsequent demonstration run testing.

Generally, the high-shear optimization testing carried out for the Upper Freeport coal indicates that this unit operation is very robust, with process PSR being relatively insensitive to changes in major high-shear variables. As mentioned previously, however, it is possible that flotation recovery of the high-shear product dominated this system and masked any effects of high-shear variables on process PSR. The effects of high-shear changes on Btu recovery cannot be determined since, as mentioned previously, slight adjustments in flotation cell air flow rates and water levels were made to maintain similar Btu recoveries between set points. It appears that due to the hydrophobic nature of the Upper Freeport coal, combined with the relatively long (6 minutes) residence time provided per stage of flotation, high Btu recovery was achieved at all high-shear conditions tested.

Flotation Optimization Testing. Testing to optimize microagglomerate recovery via flotation investigated the effect of the following on POC process performance:

- Flotation solids concentration
- Flotation air flow rate
- Flotation cell water level
- Frother quantity
- Froth spraying
- Flotation cell baffles

Operating parameters held constant throughout the flotation optimization test work included:

- Coal topsize: 200 mesh
- High-shear agglomerant concentration: 0.5%
- High-shear solids concentration: 20%
- High-shear vessel: 45 gallons

Flotation Solids Concentration Effects. Initial flotation optimization testing investigated the effects of flotation solids concentration. Table 3-9 presents data showing the effect of solids concentration on process performance.

Table 3-9
Upper Freeport Coal POC Flotation Solids
Concentration Effects

Run No.	Flotation % Solids	Frother, lb/t	Btu Recovery, %	PSR, %
20	6	0.16	95.7	75.4
	6	0.16	95.4	72.3
	8	0.16	98.1	69.8
	8	0.16	96.5	65.4
27	6	0.29	90.6	80.4
	6	0.57	93.9	74.7
	8	0.57	95.8	74.2
	8	0.29	94.6	74.0

As can be seen from these data, the 6% flotation solids concentration provided higher PSR values than the 8% solids in Runs 20 and 27 at 0.29 lb/t frother concentration. For Run 27 at the 0.57 lb/t frother level, however, no difference in performance is observed. Improved performance at lower flotation solids concentrations is not unexpected since a more dilute flotation feed would tend to reduce entrainment of ash and pyrite in the product. Similar trends were also

observed during Phase I batch testing. Solids concentrations lower than 6% were not investigated during POC testing due to the impact of water consumption on the process economics and POC water handling capabilities. All subsequent POC testing was carried out at 6% flotation solids concentration.

Frother Concentration Effects. Tests to evaluate the effects of frother concentration on POC performance were carried out during flotation optimization testing. The results of these tests indicated that changes in the frother concentration alone had no effect on Btu recovery. This lack of effect was expected since preliminary startup testing had indicated that the Upper Freeport coal was very hydrophobic and easily recovered via froth flotation. There was also no consistent effect of frother concentration on PSR over the range tested.

The Upper Freeport froth showed poor mobility during microagglomerate recovery. This phenomenon was particularly obvious at high-shear diesel concentrations of 0.5% and higher. During typical conventional flotation, froth is of high volume with visible individual air bubbles and flows easily across the cell surface to the launder discharge. However, during flotation of Upper Freeport microagglomerates created with 0.5% or more diesel, the froth was of low volume where the individual air bubbles were not distinguishable, i.e., they failed to stay formed as the bubbles rose to the surface of the cell. As a result, this froth appeared as a matted-down mass that did not flow easily across the cell surface and had a tendency to stagnate in the back half of the cells.

A mobile froth which flows well across the surface of the flotation cell may be desirable since during movement from the back to the front of the cell time is provided for drainage of mineral matter-bearing process water. Tests using higher frother dosages were carried out to determine whether additional frother would help stabilize the air bubbles and provide a more conventional flotation froth with higher mobility. However, no such effect was observed.

Flotation Water Level and Air Flow Rate Effects. As the flotation cell water level is increased at fixed air flow rates, coal recovery increases since the froth, which has a constant thickness, is raised by the water level and is more easily discharged. When the flotation air flow rates are increased at fixed cell levels the actual thickness of the froth layer is increased, again increasing coal recovery. During typical flotation operations, these two parameters can be varied independently to achieve the desired coal recovery.

To optimize the rejection of impurities during flotation, a thick froth with low water levels is preferred. Under these conditions, a froth with good drainage properties will drain a portion its mineral matter-bearing process water prior to recovery. Generally, a lower flotation air flow rate reduces product contamination by mineral matter, compared to higher air rates, since the smaller volume of air and coal rising through the flotation cell reduces entrapment of impurities in the froth.

Froth stagnation for the Upper Freeport coal occurred primarily in the first two cells of each stage of flotation where the majority of the coal was recovered. A number of POC test runs were carried out to improve froth mobility by recovering less coal in the first two cells, thereby distributing coal recovery across all cells in each stage of flotation. It was found, however, that the two primary methods for reducing flotation recovery, i.e., lowering water levels and reducing air flow, both increased froth stagnation.

In order to assure a mobile froth for the Upper Freeport coal, water levels had to be raised to within approximately 1 inch of the product removal lip. In contrast, water levels of 3-4 inches below the product lip are desirable to allow drainage of process water from the froth. If water levels were not maintained at about 1 inch below the product lip, the froth stagnated except that removed by the froth paddles, which was then immediately replaced by froth rising from directly below. In this scenario, the froth had no opportunity to drain any of its mineral matter-bearing process water because as soon as it reached the cell surface it was removed. It appears, however, that the high flotation water levels did not provide a sufficient drainage zone since no improvement in process performance was observed over the deep stagnated froth scenario.

With float cell water levels approximately 1 inch below the product lip, coal recovery in the first two cells left little coal to float in the remaining four cells even at minimum air flow rates. This scenario was undesirable since spreading the recovery out over all six cells should provide a better separation because with less coal floating per cell, the entrapment of mineral matter is reduced. Also, better dispersion of the feed is provided as the remaining coal moves through the bank of six cells.

Froth Spraying and Float Cell Baffles. Froth spraying was incorporated into the flotation cells in an attempt to reduce the entrainment of ash and pyrite particles in the recovered froth product. In theory, the fresh froth spray water replaces the mineral matter-bearing process water

carried to the surface with the froth. This froth spraying was applied only to the first two cells of each stage of flotation since this is where most of the coal was recovered.

The initial froth spraying configuration consisted of two U-shaped 1-inch diameter pipes positioned horizontally within each other above the flotation cell water level and below the top of the froth layer. Each pipe had a series of 1/8-inch diameter holes drilled in it at regular intervals so that the spray water traveled horizontally through the froth. The open end of these U-shaped pipes faced the back of the cells with about 5 inches of clearance between the two pipes.

This froth spraying configuration, configuration 1, was tested during three POC test runs with the Upper Freeport coal as shown in Table 3-10. Froth spray water flow rates of approximately 50 and 120 lb/min/cell of water were tested at constant flotation cell water levels and air flow rates with no observed effect on process performance. In addition, the installed piping interfered with froth flow across the surface of the flotation cells and was therefore removed.

Table 3-10
Upper Freeport Coal POC Froth Spray Effects

Run No.	Frother, lb/t	Froth Spray	H ₂ O, lb/min	Btu Rec, %	PSR, %
22	0.16	1	50	93.5	79.5
23	0.16	1	50	94.7	80.8
		None	—	94.6	79.8
24	0.16	1	120	92.7	90.6
		None	—	93.7	88.0
26	0.29	2	NA	92.7	78.4
		2	NA	94.7	79.9
27	0.29	None	—	90.6	80.4

A second froth spraying configuration, configuration 2, was also tested during flotation optimization testing. This configuration consisted of a single 1-inch diameter pipe, drilled in the previous manner, and positioned within the froth layer just in front of the product removal lip. A series of spray nozzles was also mounted to spray the froth from above, assisting froth flow across the cell. POC results for this configuration are also shown in Table 3-10 and indicate no improvement in performance over tests completed without the use of froth sprays. This froth spray setup was therefore removed from the flotation cells and no further froth spray testing was conducted during POC operations.

POC testing of flotation cell baffles was carried out during Run 28 in an attempt to increase the flow of froth across the flotation cell surfaces. These baffles were installed in the first two cells of each stage of flotation. They consisted of sheet metal angled from the bottom rear edge of the cells to above the froth layer in the middle of the cells forcing the froth to rise in the front half of the cells. Although the baffles had the desired effect on froth flow, preventing froth stagnation in the back half of the flotation cells, no improvement in process performance was observed so the baffles were removed.

Additional Testing. Run 30 was executed at constant process conditions with back-to-back sampling periods representing four different set points to evaluate process performance repeatability. With the exception of the first set point, which showed a higher PSR, process performance repeatability was good, as shown in Table 3-11.

Table 3-11
Upper Freeport Coal Additional POC Test Results

Run No.	Dies, %	HS rpm	Frother, lb/t	Btu Rec, %	PSR, %
30	0.5	950	0.29	87.1	83.5
		950		89.1	78.0
		950		86.4	79.1
		950		86.1	78.5
31	0.125	0	0.12	89.0	89.4
		500		89.1	87.7
		950	0.13	87.3	90.6
		1200		83.3	90.4

The final POC Upper Freeport coal parametric test run (31) utilized a high-shear diesel concentration of 0.125%, more typical of conventional flotation than the 0.5% used for the majority of previous POC tests. During this run, all process parameters were kept constant except the high-shear rpm. Data from this run, also shown in Table 3-11, show good process performance for all set points. The data indicate that at this diesel level, increasing the high-shear rpm resulted in slight decreases in Btu recovery with corresponding increases in PSR. It is also evident that at this oil level, operations with no high shear provided similar process performance. It can also be seen from these data that the 0.125% diesel level (Run 31) provided 6-10% higher PSR than the 0.5% diesel level (Run 30) at similar Btu recoveries. This indicates that the higher diesel level resulted in lower PSR since the additional diesel, not necessary for coal recovery, increased pyrite recovery. Significant differences in froth properties also were evident during these two runs, with

the 0.125% diesel level providing a mobile froth of greater thickness typical of conventional flotation.

Parametric Testing Summary. Figure 3-14 presents all Upper Freeport coal parametric test results on a ROM coal basis. The Upper Freeport coal parametric testing was completed using two different batches of Upper Freeport coal and as such Figure 3-14 presents the data by batch and coal topsize. These data indicate that lower Btu recoveries, closer to the project goal of 85%, were achieved with the second batch of coal than with the first batch. This is probably due to better control over the flotation circuit as operating experience was gained. Generally, however, the two batches of Upper Freeport coal provided very similar process performance results.

Figure 3-14. Upper Freeport Coal POC Parametric Tests

POC performance ranging between 85 and 92% PSR at Btu recoveries greater than 85% was achieved. About half of the set points in this group of results were completed at a 0.125% high-shear diesel concentration, indicating that the lower oil level provides higher PSR, as discussed previously. The other results in this group were not reproduced and are therefore considered anomalies.

In summary, the following general observations can be made with respect to the Upper Freeport coal parametric testing:

- Being very hydrophobic, the Upper Freeport coal was easily recovered in flotation at all conditions tested.

- Any high-shear effects on process Btu recovery were masked by the consistently high Btu recoveries attained during flotation.
- Decreasing the high-shear diesel concentration from 0.5% to 0.125% increased process PSR by 6-10% at similar Btu recoveries.
- At diesel concentrations of 0.5% and higher, a stagnant flotation froth of poor mobility was observed; increasing the flotation water levels was the only method found to increase froth mobility.
- A 6% flotation solids concentration provided better PSR than an 8% solids concentration.

Demonstration Run. Based on the results of the Upper Freeport coal parametric testing, the following conditions were chosen for primary demonstration run testing:

- Nominal 200-mesh coal topsize grind
- High-shear diesel addition of 0.5%
- High-shear agglomeration using the 45-gallon vessel at 950 rpm
- High-shear solids concentration of 20%
- Microagglomerate recovery using two-stage flotation at 6% solids
- Total frother addition to the flotation circuit on a dry coal feed basis of 0.29 lb/ton

In addition, for the last two set points of the demonstration run, the high-shear diesel concentration was reduced to 0.125% with all other process conditions except frother concentration, flotation air flow rates, and cell levels remaining constant.

Detailed Upper Freeport demonstration run results are presented in Appendix E, POC Results Summary. The average Btu recovery and PSR for the 0.5% diesel demonstration run conditions were 89.7% and 78.0%, respectively. A graph of percent Btu recovery and PSR by set point is shown in Figure 3-15.

As can be seen in Figure 3-15, Btu recovery and PSR were maintained at fairly constant levels throughout the demonstration run. The slight variations in these performance indicators between set points are not unexpected since small variations in feed coal characteristics and diesel addition rates do occur. Also,

small changes in flotation air flow rates and cell water levels were made to control recovery and would result in similar variations. When the diesel concentration was lowered to 0.125%, mixed results were observed. The first set point at this lower oil level gave results similar to those with 0.5% diesel, while the last set point provided about 9% higher PSR at the same Btu recovery. Unfortunately, the demonstration run was concluded after this set point so this result could not be reproduced.

Figure 3-15. Upper Freeport Coal POC Demonstration Run Results

Figure 3-16 presents the Upper Freeport coal demonstration run results plotted with all of the 200-mesh topsize parametric test results. Also shown in Figure 3-16 is the 200-mesh washability curve for the batch 1 coal on a ROM basis. As can be seen in this figure, demonstration run results were generally not as good as those achieved during some of the parametric tests. These data indicate that optimal conditions were not chosen for the demonstration run. The average demonstration run performance values of 89.7% Btu recovery with 78.0% PSR represent good process performance. It should also be noted that the general trend (grade/recovery) curve drawn for the 200-mesh Upper Freeport coal POC test results parallels the 200-mesh washability curve.

In comparing the demonstration run results to the parametric testing data, it is apparent that some of the parametric results provided significantly higher PSR at equivalent Btu recoveries. Some of these high PSR test results correspond to testing completed at 0.125% diesel concentrations. These results were repeated only during the last set point of the demonstration run as shown in Figure 3-16. The remaining high PSR data achieved during parametric testing occurred within a 2-3 day

operating period and could not be reproduced. As such, these results are considered specific to the feed coal liberation characteristics during this particular operating period.

Figure 3-16. Upper Freeport Coal 200-Mesh POC Results

When comparing the POC testing results to the washability data, it is evident that the centrifuge sink/float washability procedure utilized achieved PSR levels approximately 15-20% higher than were achieved during the demonstration run. Differences between performances achieved during actual testing and float/sink washability analyses are not unexpected since washability results represent the theoretical maximum PSR achievable, by gravity separation, at any given Btu recovery.

Fine coal washability results, as shown in Figure 3-16, are based on gravity separations in which all particles of specific gravity greater than 1.8 report to the refuse and are therefore considered rejected. The centrifuge float/sink washability procedure utilized requires the use of a dispersant and a 20-minute residence time at a centrifuge rpm of 1500 to ensure that a good gravity separation is achieved at the fine coal sizes tested (<200 mesh). In contrast, during agglomeration testing, a coal surface-based separation is made during flotation recovery of microagglomerates. Since so much care is required to guarantee a good laboratory float/sink separation at these fine coal sizes, it is not unexpected that actual agglomeration test results indicate poorer performance than washability predictions.

During float/sink separations, many "locked" or unliberated particles, i.e., particles which are part carbon and part mineral matter, report to the refuse based on their overall particle specific gravity. During agglomeration, however, these locked particles may easily report to the product since any coal on the particle surface will be coated with agglomerant and may subsequently report to the product.

This reporting of unliberated, predominantly mineral matter-bearing particles, to the product can be reduced by additional grinding to further liberate impurities from carbon. Also, the use of very low or "starvation" quantities of agglomerant may help reduce this phenomenon by reducing the oil available to coat the carbonaceous portions of the locked particles.

An additional explanation for the difference in PSR levels between testing and washability data is the contamination of the flotation product by mineral matter carried within the water in the flotation froth. The recovery of microagglomerates by flotation results in a strong froth which can entrap ash and pyrite particles suspended in the water between the air bubbles in the froth. One method used to prevent mineral-matter entrapment is to reduce the solids concentration in the flotation feed. A reduction in feed solids concentration produces a corresponding reduction in the mineral-matter concentration in the flotation cells and therefore reduces the amount of impurities carried to the product with each unit of water contained in the froth.

Upper Freeport Coal Feed Sulfur. Inconsistencies in total sulfur measurements of the plant feed coal were observed during Upper Freeport coal POC operations. In general, a difference was observed between the sulfur content of the POC crusher product and the flotation feed sample, with the flotation feed sample having a lower total sulfur content.

POC testing from startup through the first 20 runs used batch 1 of the Upper Freeport coal which was precleaned by CQ, Inc. While typical variations in the feed coal (ball mill product) sulfur content were observed, a noticeable decrease in flotation feed total sulfur content was evident as illustrated by a 1.61% TS average, significantly lower than the 1.81% reported from CQ's head sample analysis.

Starting with POC Run 21, the second batch of precleaned Upper Freeport coal from CQ Inc. was utilized. The reported total sulfur value of this precleaned product was 1.67%. Flotation feed samples for parametric testing using this batch of coal averaged

1.42%, indicating a feed sulfur drop in the same range (0.2%) as that which occurred with the first batch of the Upper Freeport coal.

Initially these sulfur discrepancies were attributed to differences in analyses between laboratories and/or an unrepresentative head sample. During the Upper Freeport demonstration run, however, stop-belt samples indicated an average feed coal sulfur content of 1.60% while flotation feed samples averaged 1.43%.

As part of the investigation into this discrepancy, batch grinding tests were completed in the laboratory to see whether the reduction in sulfur was repeatable. Three splits of Upper Freeport demonstration run stop-belt samples were ground in the laboratory ball mill and the sulfur contents of the samples determined before and after grinding. Grinding mill product filtrates for all three tests were then sent for analysis since it was hypothesized that sulfates (oxidized from pyritic sulfur) may be leaching into the water. This test was completed three times with different pH control schemes to determine the pH effect. The results are presented in Table 3-12.

Table 3-12
Upper Freeport Coal Laboratory Grind Feed Coal Sulfur Tests

Feed % TS	Prod % TS	Diff % TS	Prod Filtrate % TS, Coal Basis	Grind pH
1.57	1.44	0.13	0.14	2.8-7.6
1.58	1.37	0.21	0.15	6.6-7.8
1.62	1.42	0.20	0.18	7.2-7.8

As can be seen in Table 3-12, the coal sulfur content dropped during grinding for all three tests. The product filtrate was analyzed for total sulfur in the form of sulfate (SO₄). Since the amounts of coal and water were known, this filtrate sulfate content was converted to percent sulfur on a coal basis. As can be seen, the sulfate present in the product filtrate accounts for the bulk of the drop in the total sulfur content during grinding. It appears that the pH during grinding has no effect on sulfur leaching into the water.

Similar analyses for total sulfur in the form of sulfate were conducted on demonstration run secondary mill product sample filtrates. A sample of POC recycle water was also analyzed for sulfate content. When the difference between the mill filtrate and recycle water analyses is converted to percent total sulfur on a dry coal basis (dcb), as was done for the batch tests, these data

indicate drops in feed coal total sulfur in the range of 0.20 to 0.31, as shown in Table 3-13.

Table 3-13
Upper Freeport Coal Demonstration Run Ball Mill Filtrate Analyses

Primary Mill Prod		Secondary Mill Prod	
SO ₄ as TS (mg/l)	Filtrate TS, % (dcb)	SO ₄ as TS (mg/l)	Filtrate TS, % (dcb)
1550	0.20	1550	0.20
1720	0.22	1830	0.23
1740	0.22	1710	0.22
2430	0.31	2440	0.31

Analysis of these data indicates that the amount of sulfur in the form of sulfate in the POC ball mill circuit accounts for, and in fact exceeds, the average difference (0.17%) in the feed coal total sulfur content between the stop-belt and the flotation feed samples.

The reported total sulfur content of the batch 2 Upper Freeport coal was 1.67% while demonstration run stop-belt and flotation feed samples had sulfur contents of 1.60 and 1.44%, respectively. It is believed that oxidation of pyrites to other forms of sulfur (primarily sulfate) occurred during coal storage. This oxidation was enhanced by the fineness of the coal. Some leaching of sulfates with coal pile runoff most likely occurred, resulting in the 1.60% stop-belt sample sulfur content. The remaining sulfates were then leached from the coal during grinding, thus accounting for the 1.44% average flotation feed sulfur content. Table 3-14, comparing the various sets of feed analyses, shows a decrease in pyritic sulfur content along with an increase in sulfate content from the precleaned product to the stop-belt samples. This shift in sulfur forms content is expected as oxidation occurs.

Table 3-14
Upper Freeport Coal Feed Sulfur Comparison

Sample	TS, %	PS, %	SS, %
Precleaned Product	1.67	0.89	0.03
Stop Belt	1.60	0.70	0.16
Flotation Feed	1.43	—	—

POC Feed Coal Adjustment. As discussed previously, POC agglomeration process performance calculations (precleaned coal basis) are based on the flotation circuit feed and product sample sulfur contents. Precleaning performance is calculated based on ROM and pre-

cleaning product sulfur contents, and precleaning yield. Total process performance, i.e., ROM Btu recovery and PSR, are then calculated by combining the individual precleaning and agglomeration performance results. Thus, any difference between the reported precleaned coal sulfur content (1.81 and 1.67% for batches 1 and 2 of the Upper Freeport coal respectively) and the average agglomeration flotation feed sulfur content (1.61 and 1.43% for batches 1 and 2 respectively) is not accounted for in the ROM performance calculation. Therefore, the calculated ROM PSR is lower than the actual ROM PSR since the sulfur drop of approximately 0.2% for each batch of Upper Freeport coal is not reflected in ROM PSR values.

Given the above, an adjustment of Upper Freeport coal demonstration run flotation feed total sulfur values is necessary to ensure that ROM PSR calculations reflect all sulfur rejection from the raw coal to the final agglomerated product. Parametric testing results presented previously, however, were not adjusted in this manner since this data set was used only for comparison purposes within itself. As such, adjustment of parametric testing results would not alter the data analysis.

The steps followed for making this data adjustment are as follows:

- Flotation feed sulfur values for each demonstration run set point are increased by a constant ratio.
- This ratio is defined as the reported precleaned coal sulfur content divided by the average demonstration run flotation feed sulfur content ($1.67/1.44 = 1.16$). This constant ratio adjustment assures that the natural feed coal sulfur content variability is not distorted.
- In a similar manner, 20% of the sulfur added to the feed coal is added to the product sulfur content. This accounts for process inefficiencies, i.e., if the sulfur had been present in the feed, only 80% of it would have been rejected and therefore 20% would still remain in the product.
- A revised process performance is then calculated using the adjusted feed and product total sulfur values.

In this manner, the adjusted average Upper Freeport demonstration run performance was determined for use in the commercial agglomeration plant design included later in this report. Table 3-15 presents the effect of this adjustment on the feed and product total sulfur contents as well as on ROM POC performance indicators.

Table 3-15
Upper Freeport Coal Demonstration Run
Feed Total Sulfur Effect

	Feed TS, %	Prod TS, %	Btu Rec, %	PSR, %
Original	1.44	1.13	89.7	78.0
Adjusted	1.67	1.18	89.7	81.0

As can be seen from this table, the net effect of this sulfur adjustment is a 3% increase in ROM PSR and no change in ROM Btu Recovery. These adjusted POC performance values are used for the commercial plant design presented in a subsequent section of this report.

Pittsburgh No. 8 Coal

The second feedstock tested in the POC facility was a Pittsburgh No. 8 coal from Greene County, Pennsylvania. This coal was obtained from Cyprus Coal Company's Emerald Mine coal preparation plant where it was conventionally precleaned prior to use by the project. Details of the conventional cleaning were presented previously in this report.

Ball Mill Calibration Testing. Initial testing with the Pittsburgh No. 8 coal involved ball mill calibration. As with the Upper Freeport coal, the ball mills were calibrated to provide nominal 100- and 200-mesh topsize products from the primary mill alone and from the primary and secondary mills in series, respectively. Since the Pittsburgh No. 8 coal (HGI = 50) was much harder than the Upper Freeport coal (HGI = 75), substantial increases in ball mill charge were required to achieve the desired coal grind sizes. Typical nominal 100- and 200-mesh topsize particle size distributions achieved in the POC for the Pittsburgh No. 8 coal are shown in Figure 3-17.

As with the Upper Freeport coal, the PSD produced for the Pittsburgh coal is approximately 80% passing the 100- and 200-mesh topsizes. The d_{50} values (50% passing size) for the Pittsburgh No. 8 coal mill product size distributions are about 62-66 and 32-35 μm for the 100- and 200-mesh topsize grinds, respectively. Additional ball charge was added to the ball mills in an attempt to reduce particle size; however, no reduction was observed. It is hypothesized that an upper limit to the ball charge available for size reduction was reached and further ball charge addition had little or no effect on resulting particle size distributions.

Figure 3-17. Pittsburgh No. 8 Coal Particle Size Distributions

During these grinding calibration tests, a discharge of spent (worn-out) grinding media was observed from both the primary and secondary ball mills. This discharge was due to substantial increases in the charge required for the Pittsburgh No. 8 coal compared to the Upper Freeport coal. When the mill charges were increased, the spent ball charge which would not discharge during Upper Freeport testing due to the small ball charges, began discharging as a result of the higher mill loadings. This coarse metal, combined with a small amount of coarse coal discharging from the primary ball mill, was captured using magnets and screens. However, some coarse material occasionally reached the mill pumps, causing pump surges and blockages.

Washability Results. As with the Upper Freeport coal, once the ball mill calibration tests were completed, samples of POC primary and secondary ball mill discharge streams were taken for washability analyses. Composite data from these washability analyses are included in Appendix C, POC Test Coal Washabilities, and summarized in a plot of PSR vs. Btu recovery (precleaned coal basis) in Figure 3-18. These results indicate that approximately 3% additional PSR is achieved at Btu recoveries of 90% when the coal topsize is reduced from 100 to 200 mesh.

Orientation and Optimization Testing. With the completion of ball mill grinding calibration tests, POC orientation and optimization tests utilizing the Pittsburgh No. 8 coal were carried out. It should be noted that

during much of the Pittsburgh No. 8 coal testing, several operational difficulties were encountered. Inconsistent primary and secondary mill pump flow rates resulted from the discharge of spent grinding media from the ball mills, as discussed above. Surges in the secondary mill pump caused fluctuations in the feed to the high-shear agglomeration circuit as well as the subsequent froth flotation circuit where microagglomerate recovery was carried out. As a result, it was difficult to maintain steady-state conditions through the flotation circuit.

Figure 3-18. Pittsburgh No. 8 Coal Washability Analyses

Inconsistent agglomerant (diesel) flow rates were recorded during initial Pittsburgh No. 8 POC test runs. These inconsistent oil flows were due to operation of the diesel pump at speeds below the ideal operating range for the pump. The diesel pump was subsequently replaced, following which more consistent oil flow rates were recorded.

Since the Pittsburgh No. 8 coal floated much more slowly than the Upper Freeport coal, any variations in the coal feed rate to the flotation circuit or the diesel flow rate had a pronounced effect on the flotation circuit tailings ash content. For example, an increase in the coal feed rate to the flotation circuit or a decrease in the oil flow rate to high shear resulted in a lower tailings ash content, i.e., an increase in coal loss to tailings and a decrease in Btu recovery.

Operating conditions tested during the Pittsburgh No. 8 coal POC parametric test runs (33-55) included:

- 100- and 200-mesh topsize particle size distributions

- 45-gallon high-shear mixer at 0, 700, and 1200 rpm
- 150-gallon high-shear mixer at 1186 rpm
- High-shear solids concentration of 20%
- High-shear diesel concentrations of 0.07-1.0%
- Solids concentrations of 6% for both stages of flotation
- Total frother concentrations of 0.04-0.29 lb/ton
- Various flotation cell water levels and air flow rates

Detailed operating conditions and performance results for all Pittsburgh No. 8 coal POC runs are presented in Appendix E, POC Results Summary.

Coal Topsize Effects. POC results comparing performance at 100- and 200-mesh topsizes with other similar operating parameters are shown in Table 3-16. Constant conditions for the set points presented below include the use of the 45-gallon high-shear mixer at 20% solids and 6% flotation solids concentration. Set points carried out at both 1200 and 700 high-shear rpm are included in this table to allow comparison of topsize data at similar Btu recoveries.

Table 3-16
Pittsburgh No. 8 Coal POC Topsizes Effects

Run	Top-size	Dies, %	HS rpm	Froth, lb/t	Btu Rec, %	PSR, %
42	200	0.5	1200	0.21	89.1	61.8
					91.5	60.3
43	100	0.5	1200	0.18	91.2	52.1
					91.5	56.5
44	100	0.5	1200	0.24	90.4	53.6
45	200	0.125	700	0.29	89.5	68.9
					89.2	64.8
44	100	0.125	1200	0.29	91.1	50.3
					91.4	53.4

As can be seen in Table 3-16, at similar Btu recoveries, consistently better (4-10%) PSR levels were achieved at the 200-mesh topsize when compared to the 100-mesh set points. While all other operating conditions were not identical between these tests and the 100-mesh tests had slightly higher Btu recoveries, the performance differences indicate coal grindsize to be the cause for this improvement. Based on these data, the nominal 200-mesh topsize grind was selected for continued testing.

High-Shear Mixing Effects. During orientation and optimization testing of the Pittsburgh No. 8 coal, various high-shear mixer rpm settings were tested, with the majority conducted in the 45-gallon high-shear vessel at 700 and 1200 rpm. Additional tests were also completed using the 150-gallon high-shear vessel at 1186 rpm and one test run with no high shear. All of the Pittsburgh coal 200-mesh topsize POC parametric test results, by high-shear conditions, are plotted in Figure 3-19.

Figure 3-19. Pittsburgh No. 8 Coal 200-Mesh High-Shear Effects

These data indicate that while all high-shear conditions tested (including the no high-shear case) achieved similar results, the 45-gallon high-shear vessel at 1200 rpm appeared to provide marginally better performance, and it was therefore chosen for subsequent demonstration testing.

Agglomerant Dosage and Flotation Cell Settings Effects.

A wide range of agglomerant (diesel) and frother concentrations were tested for Pittsburgh No. 8 coal. From this testing, it was generally found that all diesel levels resulted in similar PSR provided that Btu recovery levels were held constant. Pittsburgh coal 200-mesh topsize parametric testing results, by high-shear diesel concentration, are shown in Figure 3-20.

As can be seen in this figure, similar process performance was achieved at all diesel levels tested, i.e., all data approximate the same grade/recovery curve. However, the higher diesel levels generally achieved higher Btu recoveries, as would be expected. Control of Btu recovery to relatively constant levels was achieved

through changes in frother concentration, flotation cell water levels, and flotation air flow rates. With the interaction of these variables, it was often difficult to achieve similar coal recoveries with varying diesel levels, as is evident in Figure 3-20. The 0.5% diesel concentration was chosen for primary subsequent demonstration run conditions since it is more typical of agglomeration.

Figure 3-20. Pittsburgh No. 8 Coal 200-Mesh Topsize Oil Effects

To maintain constant Btu recovery from set point to set point, flotation tailings streams were visually monitored and POC yields controlled by changing the parameters mentioned above. Maintaining constant Btu recoveries was difficult during the Pittsburgh No. 8 coal POC operations due to high-shear feed pump surges and varying diesel additions, as discussed previously. The overall effect of these fluctuations, however, was to shift POC performance along a typical grade/recovery curve, i.e., as Btu recovery increased, PSR decreased correspondingly. Flotation cell level indices were defined for each stage of flotation and evaluated to determine the effects of water level on PSR and Btu recovery. No significant shifting off the grade/recovery curve was observed as a result of changing these variables.

Summary. POC test results for the precleaned Pittsburgh No. 8 coal indicate the following:

- A 200-mesh topsize provides better performance than a 100-mesh topsize.

- The Pittsburgh coal floats much slower than the Upper Freeport coal. As a result, Btu recovery in the flotation circuit for the Pittsburgh coal was sensitive to changes in high-shear oil concentration, frother concentration, and flotation circuit water levels, air flows, and feed rate.
- The froth flow/mobility properties of the Pittsburgh coal were much better than those of the Upper Freeport coal with no froth stagnation observed at the appropriate settings.
- Monitoring of pH during Pittsburgh coal POC operations indicated that there was little need for caustic (sodium hydroxide) addition to maintain neutral pH in the water system. In contrast, regular caustic addition was required during POC testing of the Upper Freeport coal.

Figure 3-21 presents the performance of all Pittsburgh No. 8 coal POC test runs separated by grindsize. These data cover a wide range of Btu recoveries and PSR levels. The line drawn on Figure 3-21 represents the general trend (grade/recovery) curve for the 200-mesh topsize test results.

Figure 3-21. Pittsburgh No. 8 Coal POC Parametric Test Results

Demonstration Run. Based on POC orientation and optimization testing, the operating conditions chosen for the Pittsburgh No. 8 coal demonstration run were:

- A nominal 200-mesh topsize coal feed

- High-shear diesel concentration of 0.5%, using the 45-gallon vessel at 1200 rpm for set points 1-16
- High-shear diesel concentration of 0.125% using the 45-gallon vessel at 700 rpm for set points 17-21
- Microagglomerate recovery using two-stage flotation at 6% solids
- Total frother addition to the flotation circuit on a dry coal feed basis of 0.11-0.15 lb/ton for set points 1-16 and 0.29 lb/ton for set points 17-21

Analysis of POC feed stop-belt samples taken during the demonstration run indicated that the organic sulfur content of the Pittsburgh coal was slightly higher for the demonstration run than it was during parametric testing. As such, a different MAF organic sulfur constant (1.13%) was used to calculate POC feed and product pyritic sulfur contents for the demonstration run. The use of the revised organic sulfur constant resulted in better agreement between calculated and measured POC feed and product pyritic sulfur values. Therefore, calculated PSR determinations better represent actual POC performance.

Detailed Pittsburgh No. 8 demonstration run results, utilizing the revised MAF organic sulfur constant, are presented in Appendix E, POC Results Summary. A plot of Btu recovery and PSR by set point is presented in Figure 3-22.

Figure 3-22. Pittsburgh No. 8 Coal POC Demonstration Run Results

The average Btu recovery and PSR for the first 16 set points, completed using 0.5% diesel to high shear, were 88.2% and 64.4%, respectively. Average performance

for set points 17-21, completed at the 0.125% diesel concentration, was 86.5% Btu recovery and 67.0% PSR.

The plant was operated during the demonstration run to maintain constant Btu recovery. This was achieved by making adjustments in the air flow rates for the last two cells of each stage of flotation. Slight adjustments in frother addition were also necessary to maintain constant coal recovery. Recovery was estimated by visually observing the flotation tailings stream and correlating these observations with analyses of tailings stream grab samples. This approach to process control is expected for commercial operations although a commercial plant would probably incorporate on-line coal quality measurements for feedback control.

Comparing demonstration run and parametric testing results, Figure 3-23 presents all 200-mesh topsize POC data for the Pittsburgh No. 8 coal along with the 200-mesh washability data (ROM basis). Also included in this figure is the typical 200-mesh topsize trend (grade/recovery) curve defined by the parametric testing and shown previously in Figure 3-21.

Figure 3-23. Pittsburgh No. 8 Coal 200-Mesh POC Results

As can be seen in this figure, demonstration run performance approximates optimal parametric testing results with the exception of a few points from the parametric tests considered anomalies since they could not be reproduced. In general, these results indicate that good process control was maintained near optimal conditions during the Pittsburgh No. 8 coal demonstration run.

In comparing 200-mesh topline POC testing and washability data, it is evident, as shown in Figure 3-23, that significantly greater PSR (20-25%) is achieved at similar Btu recoveries by the washability procedure than during actual POC testing. This difference in process performance is similar to, but slightly greater than, that observed for the Upper Freeport coal. As for the Upper Freeport coal the lower POC agglomeration testing PSR is attributed to a combination of several factors including:

- Washability results represent the theoretical maximum PSR achievable by gravity separation at any given Btu recovery while agglomeration performance is determined by a coal-surface based separation.
- Any "locked" or unliberated particle, i.e., a particle which is part coal and part mineral matter, may report to the refuse during washability testing based on its overall specific gravity. In contrast, during agglomeration the same "locked" particle may report to the product since any coal on its surface will be coated with agglomerant during high shear.
- Contamination of POC flotation product by mineral matter carried within the water in the flotation froth.

Illinois No. 5 Coal

The third feedstock utilized for POC testing during Phase II was an Illinois No. 5 coal from Wabash County, Illinois, supplied by AMAX Coal Company. The sample obtained for POC testing represented approximately 80% of a ROM coal. This POC feedstock sample was collected after rotary breaker treatment and screening of the coarse fraction as previously discussed. As such, this feedstock is considered a raw coal and reported POC agglomeration performance indicators are effectively equivalent to the ROM performance values reported for the two previous precleaned project coals.

Ball Mill Calibration Testing. As with the previous two coals, initial POC testing for this coal involved calibration of the POC ball mills to provide the desired nominal 100- and 200-mesh topline particle size distributions. Typical 100- and 200-mesh particle size distributions utilized during Illinois No. 5 coal POC testing are presented in Figure 3-24.

The 100-mesh grind had 82% passing the 100-mesh sieve size and the 200-mesh grind had 80% passing a 200-mesh sieve with d_{50} sizes of approximately 60 and 30 μm , respectively. These size distributions are very

similar to those produced from the previous two feedstocks.

Figure 3-24. Illinois No. 5 Coal Particle Size Distributions

Washability Analysis. Upon completion of the ball mill calibration testing, samples of the POC primary and secondary mill products, at nominal 100- and 200-mesh topline sizes respectively, were sent out for centrifuge washability analyses. The results of these washability analyses are included in Appendix C, POC Test Coal Washabilities. These data are also shown in a plot of Btu recovery vs. PSR (precleaned coal basis) in Figure 3-25.

Figure 3-25. Illinois No. 5 Coal Washability Analyses

These washability data indicate no increase in pyritic sulfur liberation when the topsize is reduced from 100 to 200 mesh in the Btu recovery range of interest.

Parametric Testing. While the POC ball mills were calibrated to produce both 100- and 200-mesh topsize products, testing for the Illinois No. 5 coal focused primarily on the 100-mesh topsize for two reasons: (i) the washability showed no improvement upon finer grinding, and (ii) AMAX Coal Company, the coal supplier, expressed interest in determining the feasibility of utilizing a selective agglomeration circuit in a 1500-t/h coal cleaning plant under construction. The potential selective agglomeration application was to clean the natural minus 100-mesh material which in the current AMAX plant design is bypassed to the refuse.

The first eight POC runs using the Illinois No. 5 coal were carried out to determine POC performance data over a range of conditions and to provide data for subsequent use by Ultramax, a process optimization software tool currently under development by EPRI and an industrial partner, Ultramax Corporation. The conditions tested included the following:

- A nominal 100-mesh topsize
- Use of the 45-gallon high-shear vessel at 400 and 700 rpm and 20 % solids concentration
- High-shear diesel concentrations of 0.25, 0.5, 1.0, and 1.5%
- Flotation frother concentrations in the range 0.05-0.42 lb/t
- First-stage flotation cell level indices of 24-89 and air flow rate indices of 45-100

The flotation cell operating parameter indices mentioned above were developed to allow Ultramax to use a single model input for each of these variables. The detailed results of these eight POC runs (57-66) are included in Appendix E. A plot showing the wide range of Btu recoveries and PSR achieved during these tests is shown in Figure 3-26.

Ultramax Model Development. Ultramax uses sequential rather than factorial (parallel) test design methods. With sequential testing, process variable settings are not restricted to a rigid test matrix. In theory, sequential design of experiments (SDOE) is more efficient than parallel testing because factorial tests are often run at combinations of extreme parameter values,

where results are known to be less than optimal. SDOE applies existing knowledge of the process to test more logical combinations of variables by constructing a statistical model of the process.

Figure 3-26. Initial 100-Mesh Illinois No. 5 Coal POC Test Results

The Ultramax optimization methodology uses a global quadratic fit based on process historical data. This quadratic fit is then refined to a local response surface model as additional data are obtained. Based on this model, local optimal conditions are then determined along with precision estimates for the data. The program also recommends test conditions that will improve the precision estimates of the model and identify possible global optimal conditions.

The Ultramax process model used results from previous testing to recommend the next appropriate test conditions. In theory, these recommended conditions always move towards more optimized conditions or propose new combinations of process variables which may also be optimal. In this manner, the selective agglomeration process for the Illinois No. 5 coal was optimized.

POC results from Upper Freeport and Pittsburgh No. 8 coals were provided to the Ultramax process engineer as historical data for the POC selective agglomeration process. The data for the other project coals, along with data from the first eight POC Illinois No. 5 coal test runs, were used as the basis for development of the Ultramax selective agglomeration process model.

The initial Ultramax model was set up to maximize PSR at the 85% Btu recovery level. Input variables for this model included:

- Diesel concentration
- High-shear rpm
- Frother concentration
- First stage flotation air flow rates
- First stage flotation cell water levels

Based on the initial data for the Illinois coal POC testing, minimum flotation air and level settings were established and these variables were removed from the model.

An economic index was also developed for incorporation into the Ultramax model. This cost index was defined to provide a relative agglomeration process cost for various POC operating conditions. This index was based on the following assumptions:

- Coal value = \$0.80/million Btu
- Diesel cost = \$0.65/gallon
- Frother cost = \$0.55/lb
- High-shear cost = 0.15 hp/gal @ 700 rpm
- Product values based on \$250 and \$500 per ton of SO₂ removed

Using these assumed values as a basis, the cost index was defined as being equal to the sulfur rejection credit minus the process cost minus the Btu losses. It should be noted that this cost index did not account for any possible value added due to high product quality, i.e., low percent sulfur content, and only gave credit in proportion to the sulfur reduction.

Ultramax Testing. Following completion of the eight initial Illinois No. 5 coal POC test runs and development of the Ultramax model, seven more POC test runs were completed using this coal. Operating parameters were changed simultaneously at the recommendation of the Ultramax process engineer during this test series. The range of all parameters tested is as follows:

- Coal topsize: 100 and 200 mesh

- High-shear mixing intensity: 45-gallon vessel (0-1200 rpm), 150-gallon vessel (1186 rpm)
- High-shear solids concentration: 20%
- Diesel concentration: 0.15 to 1.0%
- Various flotation cell water levels and air flow rates

Minor adjustments to frother, cell level, and air flow rates were made during POC operations based on observed coal recovery, in an attempt to keep Btu recovery in the range of 85-90%. As with the previous two coals, the plant was allowed to stabilize before sampling began.

Detailed POC conditions and results for these seven additional test runs are included in Appendix E. Figure 3-27 shows POC results for all tests completed with the Illinois No. 5 coal. As can be seen in this figure, a wide range of Btu recoveries and PSR was achieved. Performance differences between the 100- and 200-mesh topsize test runs are also evident, with the 200-mesh feed coal providing marginally better results.

Figure 3-27. Illinois No. 5 Coal POC Parametric Test Results

Despite the apparent improvement when using a 200-mesh feed size rather than a 100-mesh size, the Ultramax model utilized only the 100-mesh data. The 200-mesh data were excluded because of the potential commercial application of agglomeration for natural minus 100-mesh fines, as discussed previously.

Based on data from all of the Illinois No. 5 coal POC testing, the Ultramax model recommended an optimized set of test conditions maximizing PSR at 85% Btu recovery. Table 3-17 presents these recommended POC 100-mesh topsi size operating conditions.

Table 3-17
Optimized POC Conditions
Recommended by Ultramax

Parameter	PSR Optimization
HS rpm	895
Diesel, %	0.79
Frother, lb/t	0.11
Product ash, %	4.26
Product TS, %	1.44
Btu recovery, %	85.0
PSR, %	62.6

These operating conditions, recommended by Ultramax to maximize PSR, were chosen for demonstration testing.

Demonstration Run. The POC demonstration run for the Illinois No. 5 coal consisted of 23 set points. Plant operating conditions recommended by the Ultramax model optimization were tested during the demonstration run and are listed below:

- 100-mesh topsi size grind
- 0.8% high-shear diesel concentration
- 0.118 to 0.144 lb/ton frother concentration
- Use of the 45-gallon high-shear vessel at 900 rpm
- 20% high-shear solids concentration
- 6% flotation solids concentration in both stages of flotation
- Constant flotation cell water levels and variable air flow rates

In addition to these operating conditions specified by Ultramax, a second set of operating conditions was tested during the demonstration run. These additional conditions used a 200-mesh topsi size grind, 1.2% high-shear diesel concentration, 0.091 lb/ton frother, and the 150-gallon high-shear mixer at 848 rpm.

Detailed test conditions and results for the Illinois No. 5 demonstration run are presented in Appendix E. A plot of Btu recovery and PSR by set point is shown in Figure 3-28.

Figure 3-28. Illinois No. 5 Coal Demonstration Run Results

The average Btu recovery and PSR achieved using the 100-mesh topsi size and 0.8% diesel operating conditions were 84.5% and 63.5%, respectively. For the 200-mesh topsi size and 1.2% diesel set points, average performance was 89.7% Btu recovery at 65.7% PSR. As can be seen in Figure 3-28, Btu recovery was difficult to hold constant. This is believed to be due to variations in the feed coal characteristics since the Illinois No. 5 coal was not precleaned prior to POC testing.

Figure 3-29 presents all of the 100-mesh topsi size POC Illinois coal results along with the associated general trend (grade/recovery) curve and the 100-mesh topsi size washability data.

As can be seen in this figure, the demonstration run results appear to be the optimal POC conditions tested for the Illinois No. 5 coal and, on average, validated the Ultramax model prediction for PSR at 85% Btu recovery (63.5% vs. 62.4%). As with the two previous test coals, Upper Freeport and Pittsburgh No. 8, washability data for the Illinois No. 5 coal parallels POC performance but at significantly higher PSR, approximately 15%. As for the two previous coals, the reasons for this difference in performance are the same and have been previously discussed.

Figure 3-29. Illinois No. 5 Coal 100-Mesh POC Results

These demonstration run results are not used for the commercial plant design since the POC feedstock was not a natural fines coal fraction. The commercial design for the Illinois No. 5 coal is based on batch testing results using an actual natural fines feed. The relationship between the POC results presented here and the batch test work results are discussed in a subsequent subsection of this report, "Batch Test Results Summary."

Centrifuge Testing. Testing of a Sharples solid bowl high-G centrifuge was performed during POC Runs 62, 63, and 64 to evaluate product dewatering performance. All of the centrifuge testing was performed using a nominal 100-mesh tosize PSD. Attempts were made to establish the relationship between diesel addition, solids recovery, and product moisture. The centrifuge was operated at maximum rpm and minimum pool depth to facilitate product dewatering. Initial centrifuge testing utilized typical oil addition rates for selective agglomeration (0.5%) without success. Due to the poor dewatering performance of the centrifuge at these oil levels, additional oil was added to the centrifuge feed tank with the intention of improving dewatering performance. No such improvement was observed, as can be seen in the centrifuge testing results presented in Table 3-18.

During Runs 62 and 63, the oil addition was split between the high-shear mixer and the centrifuge feed tank. The high-shear mixer oil addition rate was 0.5% diesel, with the remaining oil added to the centrifuge feed tank. During Run 64, all of the oil was added to the high-shear mixer.

Table 3-18
High-G Centrifuge Test Results

POC Run	% Diesel in Cent. Feed	Cake Moist, %	Centrifuge Yield %
62	1.6	38.4	99.8
		39.7	99.7
		38.7	99.7
63	2.2	42.6	99.6
		39.9	99.7
	2.9	43.2	99.8
		38.8	99.9
64	1.0	37.4	99.8
		37.1	99.7
		37.1	99.8
	1.5	38.1	98.7
		37.6	99.7
		35.7	99.6

The centrifuge operation was inconsistent with sporadic discharge of large cake lumps every 15 to 45 seconds. Attempts to achieve a uniform cake discharge by regulation of the inner scroll and the discharge chamber bladder, a device designed to prevent solids build-up, were not successful. It is believed that the unsteady-state operations had an adverse effect on performance.

In general, the results of the centrifuge testing were unacceptable due to the high product cake moistures (35-44%). Attempts by both a Sharples representative and SEI personnel were ineffective in achieving steady-state operations and reducing the centrifuge product moisture content. During normal POC operating conditions a rotary drum vacuum filter was used for POC product dewatering. Product moistures obtained using the drum filter were typically 25-30% for 100-mesh tosize feeds at typical agglomeration oil additions of 0.5%. High-G centrifuge testing was discontinued based on the results presented above.

Jim Walter Resources Pond Fines

The final feedstock utilized in POC testing was reclaimed fines from the JWR No. 3 mine tailings pond. The characterization data for this feedstock were presented at the beginning of this section, under the subheading "Test Coals."

Process Description. Due to the nature of the material and the test objectives, a number of POC facility modifications were required. Plant configuration changes resulted in the process flowsheet presented as a simplified POC block diagram in Figure 3-30.

Figure 3-30. POC Configuration for Pond Fines Tests

Since the pond fines did not require grinding prior to processing, the ball mills were removed from the POC structure. The pond fines were reclaimed by front-end loader and slurried in a 2000-gallon tank. Slurry was pumped from this tank to a 1200-gallon vessel equipped with a mixer in which the overflow was separated at approximately 60 mesh via static sieve bends. The plus 60-mesh material was dewatered on a vibrating screen and conveyed to the product pad. The minus 60-mesh sieve bend underflow was pumped through the 150-gallon high-shear vessel and to the flotation cells as during POC operations with the previous feedstocks. If high shear was used the agglomerant was added just prior to high shear, and if high shear was not used reagent additions were made at the sieve bend underflow tank. The second-stage flotation product launders were modified to allow sampling and flow rate determination of individual cell products so that ash/yield curves could be developed for each set point tested.

Pond Fines Testing Results. Using this POC configuration, nine set point tests were completed for this feedstock. A major objective of the tests was to determine the yield that could be obtained at a maximum product ash level of 15%. During POC pond fines testing, difficulties were encountered in maintaining steady-state conditions through the high-shear and flotation unit operations. The primary reasons for these fluctuations in plant conditions include variations in the rate of pond fines reclamation to the POC and variations in feedstock characteristics, particularly the percentage of minus 60-mesh material. As a result, the solids feed rate and solids concentration of the high-shear/flotation circuit were inconsistent. In addition to these feed rate variations, the solids concentration of the first- and

second-stage flotation feeds varied between 7 and 18% and 5 and 10%, respectively.

The POC operating conditions and results of the pond fines testing are presented in Figure 3-31 and Table 3-19. In Table 3-19, the oil type "9D:1M210" represents a 90% diesel:10% M210 (flotation collector) blend, with all oil and frother rates provided on a dry coal basis. The cumulative yield and product percent ash values in Table 3-19 represent plant performance on a flotation feed (minus 60-mesh fraction) basis. The increasing yields and their corresponding product ash contents indicate plant performance at multiple second-stage flotation residence times for each set point. The first yield and ash value for a given set point considers only the first cell product while the second incorporates both the first and second cell products, and so forth.

Figure 3-31 presents POC pond fines testing results from Table 3-19 in a plot of cumulative percent yield vs. percent ash on a flotation circuit feed basis. The large swings in the flotation circuit solids feed rate during this testing and the related changing dry coal basis oil levels resulted in operation of the flotation circuit at significantly different recovery levels from set point to set point. As a result, a wide range of yields and product ash levels was obtained for the various set points. Inconsistent operating conditions such as these makes direct comparison of individual set points difficult, and the data should be viewed in this light.

Figure 3-31. POC Pond Fines Test Results

Despite these operating problems and their effects on the data, good performance was achieved during this testing. The best results, achieved using approximately 0.1% of the diesel/M210 blend both with and without high shear, indicate that a 15% ash product can be recovered at

Table 3-19
POC Pond Fines Testing Operating Conditions And Results

Run	Set	Flot Feed	High	Oil	Target	Actual	Actual	Flot 1	Flot 2	Flot 1	Flot 2	Cum	Cum
	Point	Solids,	Shear	Type	Oil,	Oil,	Frother,	Level	Level	Air	Air	Yield,	Ash,
		lb/min			% dcb	% dcb	lb/t	Index	Index	Index	Index	%	%
P1	1	123	No	Diesel	0.1	0.05	0.067	140	40	85	55	13.6	12.8
												23.2	12.9
												27.5	13.7
												30.2	14.5
	2	110	Yes	Diesel	0.1	0.06	0.085	140	45	97	97	4.5	9.5
												9.6	9.6
												13.6	10.0
												17.4	10.7
P2	1	151	Yes	Diesel	0.1	0.00	0.000	151	45	88	60	4.3	8.9
												8.4	9.1
												10.1	9.5
												11.8	10.0
	2	199	Yes	Diesel	0.25	0.14	0.007	146	45	88	76	15.9	10.7
												20.5	10.9
												21.6	11.1
												22.5	11.3
P3	1	70	Yes	Diesel	0.5	1.14	0.099	149	58	93	76	22.6	17.7
												37.4	19.3
												45.3	21.0
												45.8	21.2
	2	83	Yes	Diesel	0.1	0.19	0.092	149	58	93	57	18.1	18.3
												29.9	19.6
												39.4	22.9
												40.9	24.0
P4	1	125	Yes	9D:1M210	0.1	0.10	0.074	160	65	62	38	20.5	14.4
												32.1	14.8
												35.4	15.3
												37.2	15.8
	2	112	No	9D:1M210	0.1	0.10	0.082	160	65	62	38	20.6	13.9
												31.5	14.2
												34.8	14.8
												36.3	15.4
	3	173	Yes	9D:1M210	0.05	0.03	0.053	160	65	62	38	5.4	11.6
												11.8	11.7
												16.7	12.4
												19.8	13.0

yields of 30% or higher. The data also show that the effect of high shear for processing of these pond fines is minimal for the conditions tested. It should also be noted that for this feedstock relatively large increases in yield (15%-20%) resulted in only small increases in product ash contents (2%-5%).

Determination of the optimal operating conditions for pond fines processing would require the achievement of steady-state flotation circuit feed throughout all testing. Steady-state test results for various operating conditions

could then be directly compared, allowing confident determination of their effects on performance. Parameter effects of interest would include oil type and concentration, frother concentration, the use of high-shear agglomeration, and flotation residence time.

A number of batch agglomeration tests were completed using the minus 60-mesh fraction of the POC pond fines feedstock. These batch tests covered the following ranges of operating conditions:

- Diesel concentration: 0.00-1.04%
- With and without high shear
- Frother concentration: 0.0-0.2 lb/t

Figure 3-32 shows these batch test results plotted with the POC results. Again, it should be noted that test conditions varied greatly but generally good agreement is observed between the batch and POC results. While this figure does not make comparisons between individual test conditions, it can be seen that the batch test results provide an excellent indication of performance from continuous pilot-scale testing of pond fines recovery.

Figure 3-32. Pond Fines Batch and POC Test Results

POC Testing Summary

Figure 3-33 presents grade/recovery curves indicating typical POC PSR on a ROM basis achieved over a range of Btu recoveries for the three project coals. Also shown in this figure are the average demonstration run performances for the three coals.

These data show that the average demonstration run performance for the Upper Freeport coal lies very close to its grade/recovery curve, indicating that near-optimal operating conditions were chosen for the Upper Freeport coal demonstration. Since there is much scatter in the data, the true optimal conditions cannot be defined with certainty. It is believed, however, based on the limited data, that lower high-shear diesel concentrations (approximately 0.125% rather than 0.5%) may have provided better performance.

Similarly, average demonstration run results for both the Pittsburgh No. 8 and Illinois No. 5 coals lie very close to

their respective grade/recovery curves. This indicates that near-optimal operating conditions were chosen for both the Pittsburgh and Illinois coal demonstration runs.

Figure 3-33. POC Operations Grade/Recovery Curves

Table 3-20 presents average POC demonstration run feed and product characteristics as determined by the Wilsonville laboratory for all three coals. In this table, all values are measured except for pyritic sulfur which is calculated from total sulfur, ash, and the appropriate moisture- and ash-free (MAF) organic sulfur content. All values in this table are on a dry basis except moisture.

Table 3-20
POC Demonstration Run Feed and Product Characteristics

Coal	POC Stream	Moist, %	Ash, %	TS, %	Calc PS, %	Btu/lb
UF	Feed	5.8	10.0	1.67	1.06	13877
	Product	33.4	6.1	1.18	0.55	14557
P8	Feed	5.2	7.3	2.12	1.07	13844
	Product	32.2	5.1	1.81	0.74	14224
I5	Feed	14.5	18.0	1.96	1.33	11716
	Product	28.3	4.5	1.40	0.67	13861

It should be noted that the calculated pyritic sulfur values presented in Table 3-20 represent the inorganic (pyritic and sulfatic) sulfur content of feed and product streams for each coal. As discussed previously in this report, any sulfatic sulfur present in these coals is included in the calculated pyritic sulfur value since sulfatic sulfur

contents were generally very low, <0.04%, for all three POC feedstocks. Upper Freeport coal values shown in Table 3-20 are adjusted, as discussed previously, to account for the oxidation and leaching of pyritic sulfur.

Based on the values shown above and the ROM coal characteristics, Table 3-21 presents selective agglomeration performance indicators on a POC feed basis, with ROM coal basis performance included where applicable. Once again it should be noted that the calculated PSR presented in this table represents inorganic (pyritic and sulfatic) sulfur rejection.

Table 3-21
POC Demonstration Run Performance

Coal	Perf Basis	Yield, %	Ash Rej, %	TS Rej, %	Calc PSR, %	Btu Rec, %
UF	POC feed	88.4	46.4	37.5	54.5	92.8
	ROM coal	81.4	64.9	66.1	81.0	89.7
P8	POC feed	92.0	35.7	21.2	36.4	94.6
	ROM coal	62.0	89.4	46.2	64.4	88.2
15	POC feed	71.4	82.0	48.5	63.5	84.5
	ROM coal	—	—	—	—	—

Batch Test Results Summary

While batch-mode testing was the primary method of testing in Phase I of the project, it was mainly used in support of POC testing during Phase II. Phase II batch testing was performed at SEI's Wilsonville laboratory.

Batch testing in Phase II was used to characterize the new project coals. Although the Pittsburgh No. 8 and Upper Freeport seam coals were tested in Phase I, the source and quality of both coals was different in Phase II. This necessitated characterization of the coals for agglomeration testing and determination of the optimum high-shear conditions for these coals.

The primary objectives of the Phase II batch tests were to:

- Conduct range-finding and confirmatory tests in support of POC runs
- Confirm and/or determine optimum Phase I high-shear conditions using the new project coals
- Provide results for comparison with POC test results under the same conditions

Batch testing was carried out on all three project coals to establish a frame of reference to compare the POC test results and to identify a correspondence between the results of batch and POC testing. In addition to the three project coals, Upper Freeport, Pittsburgh No. 8, and Illinois No. 5, batch tests were also conducted on a few other coals as part of exploratory testing.

The batch test results have been summarized in this section and compared with the POC tests to show the comparability between the two modes of testing. Detailed Phase II batch test data are contained in Appendix F, Batch Test Results Summary.

Upper Freeport Coal Testing

Several test series were carried out using Upper Freeport coal to meet the objectives described above. The initial conditions used for these tests were based on the optimal agglomeration conditions determined from the Phase I batch tests. Conditions were modified as necessary to provide optimum performance for the Phase II coal.

The test condition ranges employed for the Upper Freeport batch tests were as follows:

- 28-, 100-, and 200-mesh particle sizes
- Diesel dosages in the range of 0.3-1.0% for high-shear agglomeration
- High-shear solids concentrations of 15 and 20% were used for all tests. These levels were determined to be the optimal levels from the Phase I batch tests.
- Two stages of flotation were used for micro-agglomerate recovery. Incremental flotation products were collected for the second stage for several tests to determine flotation kinetics.
- Methyl iso-butyl carbinol (MIBC), Betz M150, and a Naico reagent were used as frothers in the flotation step. Of these, Betz M150 provided the best results and was used for most of the batch tests.

Most of the batch tests with Upper Freeport coal were carried out with samples obtained from POC runs. While some of these samples were ball mill products from the POC, the remainder were POC high-shear products. For most tests with POC high-shear products, only microagglomerate recovery using two stages of flotation was carried out and incremental timed flotation

products were collected. For tests utilizing POC ball mill products, high-shear agglomeration with 0.5% diesel at 2000 rpm was carried out in the batch mode, followed by flotation.

Figure 3-34 presents the results of the Upper Freeport coal batch and POC testing at a 200-mesh topsize. The best results of the Phase I batch tests are also plotted on the same figure for comparison, although the Phase I feed coal quality was different from that in Phase II. While operating conditions for the Phase II batch tests and POC runs cover wide ranges, Figure 3-34 provides an overall performance comparison for the two testing modes for the Upper Freeport coal at 200 mesh.

Figure 3-34. Upper Freeport Coal 200-Mesh Batch and POC Results Comparison

Comparison of results from Phase II batch tests with those from POC runs shows that the two modes of testing provide similar results for Upper Freeport coal. The curve in Figure 3-34 represents the typical 200-mesh grade/recovery POC performance curve for the Upper Freeport coal, presented earlier. The POC runs demonstrated slightly higher Btu recovery levels. In addition, the Phase II results compare quite favorably with the Phase I results.

In general, the agglomeration performance of the Upper Freeport coal in both POC and batch testing modes can be characterized as about 84% PSR at 85% Btu recovery, and about 82% PSR at 90% Btu recovery. These results indicate that batch test results are good indicators of POC performance under the conditions tested for Upper Freeport coal.

Pittsburgh No. 8 Coal Testing

Several batch tests were carried out with Pittsburgh No. 8 coal with the following objectives:

- Characterize agglomeration performance
- Establish a basis for comparison with POC results
- Conduct range-finding and confirmatory tests

Batch tests with the Pittsburgh No. 8 coal involved samples generated in the Wilsonville laboratories as well as samples obtained from the POC. The latter included samples taken from the secondary ball mill product stream and from the high-shear product stream. Test conditions for the batch tests included the following:

- 100-, 200-, and 400-mesh samples. While most of the tests were carried out with 100- and 200-mesh samples, a few POC ball mill product samples at 200 mesh were reground in the laboratory to 400 mesh.
- 0.5% diesel dosage in the batch high-shear agglomeration step. For tests run with POC high-shear product samples, no additional high shear or diesel addition was carried out in the batch testing.
- High-shear solids concentrations of 15 and 20%
- Two stages of flotation for microagglomerate recovery. For some tests, incremental timed products were collected during the second flotation stage.
- Betz M150 was added to each stage of flotation cells in the range of 0.08-0.1 lb/t for most tests.

Figure 3-35 summarizes the Phase II batch and POC test results for tests carried out with 200-mesh topsize coal. The best results from the Phase I batch testing are also included in this figure for comparison.

Analysis of the batch and POC results demonstrates that the two sets of results are quite comparable at 200 mesh. The curve in Figure 3-35 represents the typical 200-mesh POC performance grade/recovery curve presented previously for the Pittsburgh No. 8 coal. Although a few of the batch tests did provide higher PSR at equivalent Btu recovery levels, the results of the two modes are, in general, very comparable.

Comparison of the Phase II batch and POC test results with the "best" Phase I batch results shows that the latter provided slightly better agglomeration performance.

While the exact cause for this is uncertain, it may be partly attributable to the different properties and liberation characteristics of the coals used in the two phases. The washability results of the precleaned Pittsburgh No. 8 coal for the 200-mesh ball mill grind sample show 72% PSR at 1.8 specific gravity versus 77.3% for the Phase I coal, at the same gravity level. While this is not an exact comparison, differences in liberation may be a contributing factor to the slightly lower PSR values obtained in Phase II testing.

Figure 3-35. Pittsburgh No. 8 Coal 200-Mesh Batch and POC Results Comparison

Illinois No. 5 Testing

In Phase II, Illinois No. 5 coal replaced Kentucky No. 9 as the third project coal. Unlike the Upper Freeport and Pittsburgh No. 8 coals, the Illinois coal used was a raw coal. As discussed at the beginning of this section, under the subheading "Test Coals," the Illinois No. 5 coal was procured from AMAX Coal Company's mine in Wabash County, Illinois. The design of a 1500-t/h cleaning plant currently under construction calls for bypassing the minus 100-mesh natural fines to the plant tailings. One of the objectives of testing the Illinois No. 5 coal was to determine the effectiveness of agglomeration for cleaning this natural fines fraction.

The coal supplied to the project represented 80% of the ROM coal and was not subjected to any precleaning. POC and batch tests were carried out on this coal to determine the cleaning performance that can be achieved by agglomeration. Batch testing on this coal

provided agglomeration performance characterization data and confirmation of agglomeration conditions for POC testing.

The conditions for batch tests carried out on this Illinois No. 5 POC feedstock coal were as follows:

- 100- and 200-mesh topsize samples, though most tests were carried out on 100-mesh samples
- Diesel dosages of 0.2, 0.5, and 0.8% in high shear
- High-shear agglomeration in the batch mode at 2000 rpm and 15% solids concentration
- Two stages of flotation at 6% solids with Betz M150 added as a frother
- Incremental timed flotation products were collected during the second flotation stage.

The results of the Illinois No. 5 batch tests using POC primary ball mill product (100 mesh) as feed, have been compared with POC parametric and demonstration run 100-mesh topsize results in Figure 3-36. As this figure shows, the performance obtained with batch and POC tests are very comparable. The solid curve in this figure represents centrifuge washability liberation on a 100-mesh sample prepared in the POC ball mill. The close agreement between the batch and POC results are in line with that observed for the other two project coals. Agglomeration performance for the Illinois No. 5 100-mesh topsize coal is characterized as 65-68% PSR at 85% Btu recovery, while liberation data obtained from washability analyses indicates about 80% PSR at 85% Btu recovery. As discussed previously in this report, the differences in performance between washability and agglomeration test data are attributed to several factors including:

- Washability results represent the theoretical maximum PSR achievable by gravity separation at any given Btu recovery while agglomeration performance is determined by a coal surface-based separation.
- Any "locked" or unliberated particle, i.e., a particle which is part coal and part mineral matter, may report to the refuse during washability testing based on its overall specific gravity. In contrast, during agglomeration the same "locked" particle may report to the product since any coal on its surface will be coated with agglomerant during high shear.

- Contamination of POC flotation product by mineral matter carried within the water in the flotation froth.

In addition to these standard batch tests utilizing POC feedstock, three batch tests were also carried out using a minus 100-mesh natural fines sample that was generated in a commercial laboratory from the POC feedstock head sample. The average ash and total sulfur contents of this minus 100-mesh natural fines sample were approximately 33 and 1.72%, respectively.

Figure 3-36 also shows batch test results using this natural minus 100-mesh fines fraction. No POC runs were carried out with this coal. The dotted curve in the figure depicts the washability performance of the natural fines fraction. Batch test performance for these fines can be characterized as approximately 80% PSR at 85% Btu recovery, with centrifuge washability liberation indicating about 89% PSR at 85% Btu recovery. In light of the close correspondence between the batch and POC results for all three coals, POC performance with the natural fines was expected to be quite close to that obtained with the batch tests. Therefore, the batch test results obtained using this natural fines coal fraction were used for the commercial plant design, discussed in Section 4.

Figure 3-36. Illinois No. 5 Coal 100-Mesh Batch and POC Results Comparison

In addition to the tests discussed above, several other batch tests were carried out using samples of Illinois No. 5 coal from POC primary and secondary ball mill products at various topsizes. For some of these tests, the ball mill products were further ground in the laboratory to determine whether additional grinding

provides better agglomeration performance. In general, tests conducted with 100-, 200-, and 400-mesh samples showed improved performance with decreasing size, as expected.

Testing of Precleaned Illinois No. 5 Coal. In addition to batch testing with POC and natural minus 100-mesh fines samples, some tests were carried out on precleaned samples of Illinois No. 5 coal to examine the relative improvement in agglomeration performance achieved through precleaning.

It was demonstrated in Phase I that agglomeration of a precleaned coal provides better cleaning performance on a ROM basis than that provided by agglomeration of a raw coal. In order to confirm this phenomenon for the Illinois No. 5 coal, a few batch tests were conducted on simulated precleaned samples.

The batch test samples were generated from a 2000-lb 1/2-inch topsize head sample. The raw coal was initially crushed to a 1-mm topsize and the simulated precleaned samples were generated in a commercial laboratory by float/sink methods. The precleaning sample was obtained by combining weighted fractions of the following float/sink samples:

- Float 1.65 from the 1/2-inch x 1/4-inch fraction
- Float 1.65 from the 1/4-inch x 28-mesh fraction
- Float 1.80 from the 28-mesh x 100-mesh fraction
- Raw minus 100-mesh fraction

The above samples were combined to provide a simulated precleaned product from a commercial cleaning circuit. The precleaning Btu recovery was calculated to be 98.5% and the PSR was 41.5%. The batch agglomeration test conditions on these precleaned samples are as follows:

- 100- and 200-mesh sample topsizes
- Diesel dosages of 0.5 and 0.8% in high shear
- High-shear agglomeration at 2000 rpm and 15% solids concentration
- Two stages of flotation for microagglomerate recovery with Betz M150 added as a frother

Figure 3-37 presents the results of these conventionally cleaned feed coal agglomeration tests along with raw

coal tests at the same feed size consists for comparison. In general, about 3-5% greater rejection of pyritic sulfur is obtained with the precleaned samples on a ROM basis compared to the raw coals, which confirms the Phase I findings.

Figure 3-37. Illinois No. 5 Coal Raw and Precleaned Batch Results

Other Testing

Several other batch tests were carried out in Phase II with various objectives, mostly as range-finding tests for the POC or to characterize the agglomeration performance of coal samples other than the project coals. For example, raw and middlings samples of a Pittsburgh No. 8 coal from a Pennsylvania mine were tested in the batch mode. In addition, batch tests were performed using a pond fines sample from JWR and a cyclone overflow sample from Pennsylvania Electric Company's Homer City cleaning plant.

Most of these tests demonstrated high product yield and ash rejection while achieving moderate total sulfur rejection. PSR could not be calculated as pyritic sulfur measurements were not carried out in many cases.

Pelletization and Drying Results Summary

Test work and economic analysis in Phase I indicated that pelletization of microagglomerates is the most cost-effective process for formation of a product to be utilized in existing power plant coal handling facilities. The

objectives of the pelletization test work completed during Phase II were to:

- Demonstrate the viability of pelletization on a continuous scale
- Quantify pelletization feed moisture requirements
- Measure "green" (wet) and dry pellet strengths
- Verify Phase I pelletization results
- Derive the required engineering data for conceptual design of a commercial-scale pelletization process

Summary of Phase I Pelletization Work

Test work was conducted in Phase I to evaluate several cornstarch binders, various binder concentrations, use of asphalt as a weatherproofing agent, and both dry and wet methods of mixing these binders into the coal solids. The findings of this work were:

- A dried cornmeal flour, pulverized to minus 200 mesh, was a more cost-effective binder than more refined cornstarches.
- The optimum feed coal moisture for pelletization was approximately 22-25% for a 100-mesh topsize coal grind.
- A batch pin mixer (3.5 lb capacity) was tested using a 10-second residence time. Using this mixer, the binder was uniformly dispersed into the coal, but micropellet formation was evident when using high binder concentrations (3-5%). The continuous pelletization tests were performed using a ribbon blender with good results.
- A 3-ft diameter disc pelletizer was capable of forming 6-10 mm diameter pellets at a capacity of 400 lb/h. These pellets were spherical in shape with a narrow size distribution, which is indicative of a good pelletization process.
- "Green" or wet pellet strength, as measured by the number of 18-inch drops before breakage, was good for binder concentrations above 1%. All pellets tested with binder concentrations above 1% survived 5 drops.

- Dry pellet strength, as measured by the force required to crack the pellet (crush strength) under a constant rate of strain, was 6-8 lb for 6-mm diameter pellets at a moisture content of 2-4%. Reduced pellet drying or increased pellet moisture resulted in a reduction in dry pellet crush strength. It should be noted that "green" pellet crush strength was less than 1 lb.

Phase II Pelletization Test Work

In Phase II, the pelletization test work was focused on optimizing the process developed in Phase I. The basic pelletization steps selected for the Phase II work were:

- Proportioning the mechanically dewatered microagglomerates with dry cornmeal flour
- Blending the ingredients to disperse the binder
- Pelletization of the prepared solids using a drum or disc pelletizer while adding water as required to achieve acceptable pelletization kinetics.

Vacuum filter samples collected from each POC demonstration run were sent to MMC Mars Minerals for pelletization tests. The size consist of the Upper Freeport and Pittsburgh No. 8 coal samples was a nominal minus 200 mesh ($d_{50} = 30 \mu\text{m}$), while the Illinois No. 5 coal product had a size consist of minus 100 mesh ($d_{50} = 60 \mu\text{m}$). A pregelatinized dried cornmeal flour ground to 70% passing 200 mesh was used as a binder at a concentration of 2.5% on a dry coal basis. The Pittsburgh No. 8 coal was also tested at a binder concentration of 2% for comparison with the Phase I test results.

The binder was dispersed into the coal filter cake using a continuous, 12-inch diameter x 54-inch long pin mixer. The pin mixer product was then fed to a continuous 2-ft diameter deep-dish pelletizer and water was added as required to achieve good pelletization results. The deep-dish pelletizer differs from a conventional disc pelletizer in the length of the rotating dish, and therefore in the bed depth of the tumbling agglomerates, and also has a higher capacity than a conventional disc pelletizer with the same diameter. Its dish length is approximately twice the depth of a disc pelletizer, and it therefore provides a residence time approximately 1.2 times higher than equipment with the same rated capacity.

The deep-dish pelletizer was selected for the Phase II work in order to assess the effects of increased residence time and greater compaction forces resulting

from the deeper bed depth on final pellet size and strength. The advantages of producing larger pellets include increased pellet crush strength, since strength is a function of pellet size, and the generation of larger pellet fragments when breakage does occur. Anticipated limitations to deep-dish pelletizing include possible difficulties in forming large uniform pellets, increased drying residence time requirements, and increased pellet breakage when dropped from large heights, such as 6 ft.

Phase II Pelletization Test Results

The pin mixer had a very high capacity for processing the three test coals. Initially, an automatic 4-inch diameter screw feeder was used to deliver the coal to the pin mixer but this was replaced by manual feeding due to its limited feed capacity. The wet filter cake had a tendency to compact inside the screw trough, restricting the flow rate. Even with manual check feeding of the pin mixer at a capacity of 550 lb/h on a wet basis, the maximum pin mixer feed rate was not achieved as evidenced by the mixer discharge. The manufacturer stated that if the inlet port dimensions were increased, a capacity of up to 4000 lb/h was possible. The product from the pin mixer was homogeneous in appearance, with signs of coal aggregation without micropellet formation.

The deep-dish pelletizer was able to successfully produce large pellets, up to 1 inch in diameter, for all three test coals from the pin mixer product. These large pellets also had a strong green pellet strength and could survive a 6-ft drop. When operated at a drum slope of 33° from the horizontal, a pellet size of 8-15 mm was produced. The optimum pelletization moisture was approximately 27% for the two 200-mesh topsize samples (Upper Freeport and Pittsburgh No. 8 coals) and 25% for the 100-mesh topsize sample (Illinois No. 5 coal). The Illinois No. 5 feed was particularly sensitive to water addition, with excess water addition in the range of 2% resulting in a change from layering agglomeration where the fines attach to the growing agglomerates to adhesion of groups of small agglomerates, often termed "raspberry" formation. "Raspberry" agglomeration is undesirable since these agglomerates have lower strength. The sensitivity to water addition observed with the Illinois No. 5 coal is consistent with general pelletization observations which indicate that coarser feedstocks are more difficult to pelletize, requiring a narrower range of pelletization feed moisture than finer size distributions of the same material. The capacity of the 2-ft diameter deep-dish pelletizer was approximately 600 lb/h of green pellets.

Green and dry pellet strengths were tested by measuring the number of drops required to crack an 8-mm diameter pellet. A total of 20 pellets was used to compute the average number of drops to failure. Crush strength and drop-shatter tests were also used to characterize dry pellet strength. The drop-shatter test measures the change in size consist after dropping a 5-lb mass of pellets onto a metal plate from a height of 6 feet.

The results of the crush strength and drop tests are presented in Table 3-22. The green pellets had no measurable crush strength but were able to withstand between 13 and 19 drops before cracking. Drying the pellets to a moisture content of less than 4% increased their strength significantly, with crush strengths ranging from 13 to 31 lb. The oven-dried pellets had slightly lower strength than those dried in the vibrating fluidized-bed dryer (VFBD). Both of these sets of dried pellets had excellent drop strengths and could not be broken in 50 drops.

Initial drop-shatter tests conducted during the drying tests indicated that approximately 1/3 of the pellets broke when dropped from a 6-ft height. After one week of storage in sealed moisture-proof containers, however, less than 1% of the pellets sustained any breakage from a 6-ft drop. Comparison of these two sets of observations indicates that the pellets gained strength after curing for one week. Details of pellet strength over time were not quantified but could be important in determining binder requirements, especially if the pellets are stockpiled prior to reclamation by a coal handling system.

Green Pellet Drying Tests

Drying tests completed during Phase I were successful in drying disc-pelletized green pellets to less than 4% moisture with minimum degradation and good thermal efficiency. The primary objective of the Phase II drying tests was to evaluate the effects of increased pellet size on VFBD performance. The objectives of individual tests were to:

- Determine loading and operating conditions that optimize dryer efficiency
- Estimate pellet degradation during drying
- Measure residence time requirements to achieve a low-moisture product

Summary of Phase I Dryer Tests

In Phase I testing using a VFBD, drying of 6-10 mm pellets was evaluated over a range of binder concentrations and drying conditions. The conclusions of these dryer tests were as follows:

- A mean residence time of 2-3 minutes was required to achieve final pellet moisture contents of less than 4%
- Dust loss was in the range of 1.3-1.8%
- Optimal test bed coal loading was 10 lb/ft² (wet basis)

Phase II Dryer Tests

Drying tests were conducted using a 1-ft² batch VFBD with a bed loading of 10 lb/ft², wet basis. Air flow rates ranged from 200 to 300 SCFM and air flow temperatures ranged from 350 to 500°F. A summary of the final tests for each pellet sample tested is given in Table 3-23. During all tests it was necessary to raise the bed temperature above 325°F to obtain a moisture content below 4%. It was noted that pellet moisture continued to drop after switching from pellet heating to cooling. A 2-minute cooling operation resulted in a decrease in pellet moisture from 6% to 3%. This cooling operation is necessary to reduce the pellet temperature for final storage.

Table 3-22
Pelletized Agglomerate Strength Measurements

Sample	Binder, %	Green Pellets		Oven-Dried Pellets		VFBD Dryer Pellets	
		Crush, lb	Drop	Crush, lb	Drop	Crush, lb	Drop
Upper Freeport	2.5	<1	16.5	21	50+	24	50+
Illinois No. 5	2.5	<1	15.2	13	50+	19	50+
Pittsburgh No. 8	2.5	<1	19.8	25	50+	31	50+
Pittsburgh No. 8	2.5	<1	13.4	22	50+	26	50+

Note: Dried pellet moisture range, 1-2%

Table 3-23
Pelletized Agglomerate Drying Conditions (VFBD)

Sample	Air Flow SCFM/ft ²	Drying Temp	Initial Moisture	Residence Time, min ¹	Dust Generation
Upper Freeport	300	400°F	24.0%	3.0	1.8%
Illinois No. 5	200	500°F	29.7%	4.0	1.7%
Pittsburgh No. 8	300	400°F	24.9%	3.5	0.4%
2.5% binder					
Pittsburgh No. 8	200	500°F	24.6%	3.5	<0.4%
2% binder					

Note 1: Dryer residence time required to reduce pellet moisture below 6%. A further 2 minute residence time in the cooler will reduce final moisture below 4% via evaporative cooling.

The results indicate that larger pellets require a longer drying time (3.5–4 minutes total) than was the case with Phase I tests in order to achieve a moisture content of less than 4%. This is probably due to the larger pellet diameter but could also be related to higher, prolonged compaction forces achieved in the deep-dish pelletizer. These additional compaction forces would result in a denser pellet with greater strength, as discussed in the previous subsection, "Pelletization Tests," but may also inhibit the diffusion of water from the pellet interior. It was observed that partially dried pellets, when broken, had a distinct outer ring which was dry and a damp inner core. Degradation was minimal during these drying tests, with less than 2% dust generation for all tests.

As discussed in the previous subsection, the dried agglomerates were not quantitatively tested until they were sent back to the pelletizer. However, the pellets were checked for strength during the drying tests by dropping from a 3-ft height while still warm. On average, cracks appeared in the agglomerates after 1 to 2 drops, with 1/3 of the pellets breaking on the first drop. Strength measurements made after one week indicated improved pellet strength.

Final Product Characterization

In order to determine final agglomeration/pelletization product characteristics, a sample of pelletized product for each coal was sent out for proximate and ultimate analysis, sulfur forms and calorific value determination, and various other analyses. These analytical results are presented in Table 3-24.

Conclusions

Testing during the second phase of this project focused on POC operations to maximize PSR at about 85% Btu

recovery and produce a final agglomerated product with an ash content of 6% or less. An additional objective of this testing was to demonstrate continuous long-term selective agglomeration operation at the pilot scale (2 t/h). Laboratory batch tests were also conducted in support of POC testing. Pelletization and drying tests were conducted by equipment vendors using POC demonstration run product samples.

POC Operations

The POC test plan was designed to optimize the high-shear and microagglomerate recovery (flotation) unit operations separately. This proved to be difficult, however, since POC operating philosophy targeted similar Btu recoveries for all set points. This mode of operation allowed comparison of PSR at similar Btu recoveries for various operating conditions. Btu recoveries achieved during POC demonstration runs, completed at optimized operating conditions, ranged from approximately 76 to 93%. Based on demonstration run and parametric testing results, grade/recovery curves characterizing pilot-scale selective agglomeration performance were developed and are shown in Figure 3-38. Also presented in Figure 3-38 are the average demonstration run performances achieved for each of these three project coals.

Based on these grade/recovery curves, the PSR attained at 85% Btu recovery for the three primary POC feedstocks is presented in Table 3-25. Also included in this table are the average demonstration run results for each coal. As can be seen in Figure 3-38 and Table 3-25, average POC demonstration run performance approximates the grade/recovery curve for each coal tested. As such, for the Upper Freeport and Pittsburgh No. 8 coals, the average demonstration run performance was used to complete the commercial-scale (200-t/h) plant design, discussed in Section 4.

Table 3-24
Final Product Analysis (Dry Basis)

	Upper Freeport	Pittsburgh No. 8	Illinois No. 5
<u>Proximate Analysis</u>			
Moisture, % (as is)	0.95	1.90	2.53
Ash, %	5.56	5.28	4.72
Volatile matter, %	33.22	38.74	39.19
Fixed carbon, %	61.22	55.98	56.09
<u>Ultimate Analysis</u>			
Hydrogen, %	4.84	5.07	4.94
Carbon, %	80.95	80.09	77.25
Nitrogen, %	1.41	1.56	1.82
Sulfur, %	1.08	1.75	1.38
Oxygen, %	6.16	6.25	9.89
Ash, %	5.56	5.28	4.72
<u>Forms of Sulfur</u>			
Total sulfur, %	1.08	1.75	1.38
Pyritic sulfur, %	0.05	0.60	0.48
Sulfate sulfur, %	0.39	0.12	0.17
Inorganic sulfur, %	0.44	0.72	0.65
Organic sulfur, %	0.64	1.03	0.73
Lb SO ₂ /MBtu	1.51	2.46	2.03
Calorific value, Btu/lb	14274	14209	13618
Free swelling index	8.5	8.0	2.0
Hardgrove grindability index	177	197	150
<u>Elements in Ash, %</u>			
Silicon dioxide (SiO ₂)	45.72	42.33	44.99
Aluminum oxide (Al ₂ O ₃)	24.97	24.83	21.43
Iron oxide (Fe ₂ O ₃)	19.99	23.37	23.55
Titanium oxide (TiO ₂)	1.40	1.31	1.86
Phosphorus pentoxide (P ₂ O ₅)	0.48	0.34	0.65
Calcium oxide (CaO)	1.90	2.79	1.94
Magnesium oxide (MgO)	0.74	0.71	0.76
Sodium oxide (Na ₂ O)	1.37	0.66	1.40
Potassium oxide (K ₂ O)	2.36	1.44	2.41
Sulfur trioxide (SO ₃)	0.98	2.06	1.00
Manganese dioxide (MnO ₂)	0.04	0.04	0.04
<u>Ash Fusion Temperatures (°F)</u>			
Initial deformation	2170	2150	2090
Softening	2440	2450	2340
Hemispherical	2490	2460	2390
Fluid	2600	2560	2490

Figure 3-38. POC Pilot-Scale Selective Agglomeration Performance

Table 3-25
POC PSR at 85% Btu Recovery and Average
Demonstration Run Performance

Coal	Topsizes	PSR @ 85% Btu Rec, %	Avg Demo Run Btu Rec, %	Avg Demo Run PSR, %
UF	200 M	84	89.7	81.0
P8	200 M	69	88.2	64.4
I5	100 M	64	84.5	63.5

In the case of the Illinois No. 5 coal, however, average demonstration run results were not used for the commercial-scale design since the Illinois coal POC feedstock was a raw coal and the commercial design is for the minus 100-mesh natural fines fraction of this coal. Instead, batch test results achieved using a sample of the natural minus 100-mesh coal fraction were used for this design.

Relatively consistent Btu recoveries were achieved during high-shear optimization by making small adjustments in flotation parameter settings. Using this method of testing, any high-shear effects on coal yield were dominated by the flotation recovery of the microagglomerates. The hydrophobicity of the coals tested and the agglomerant levels used enhanced the flotability of the POC feedstocks, as is evident by the ease with which these coals were recovered over the range of conditions tested. Generally, it appears that the installed flotation capacity was sufficient to recover all coals to the desired levels, regardless of the high-shear conditions utilized.

Since Btu recoveries were generally held constant during much of the POC testing, process performance comparisons between different high-shear operating conditions were based on PSR determinations. For virtually all of the POC testing completed, changing the high-shear conditions had no significant impact on process PSR. High-shear parameters investigated included agglomerant concentration, solids concentration, residence time, and mixing intensity. High-shear solids concentration and residence time data clearly indicate that these parameters do not affect process PSR and, therefore, the more economical conditions are considered superior.

While the results of tests to investigate the effect of agglomerant concentration indicate that it has minimal impact on process performance, most of these data were obtained at diesel concentrations of 0.5% or higher. Though there are few results based on low diesel concentrations for the Upper Freeport coal, there is some indication that a 0.125% diesel level may provide higher PSR at equivalent Btu recoveries than the 0.5% diesel concentration. For the Pittsburgh No. 8 coal, a substantial amount of data for low diesel concentrations are available which indicate that there is no difference in process performance over the range of diesel levels tested. POC performance for Illinois No. 5 coal as a function of diesel concentration appears to be constant; however, the Ultramax optimization model recommended a 0.8% diesel concentration over lower oil levels.

The effect of high-shear mixing intensity (rpm) on process PSR is minimal for all coals over the ranges tested. In fact, results obtained with and without the use of high shear provided similar results for both the Upper Freeport and Pittsburgh No. 8 coals at 0.125% diesel concentration. Tests completed without the use of high shear for the Illinois No. 5 coal resulted in poorer process performance than similar tests utilizing high shear.

POC testing of reclaimed pond fines indicates that under the conditions tested 30% yield from the minus 60-mesh fraction was achieved at a product ash content of 15%. This level of performance was achieved both with and without the use of high shear. Similar performance was achieved during laboratory batch testing.

Batch Testing

Phase II laboratory batch-mode testing was conducted primarily in support of POC testing. The batch test results also provided data for comparing batch bench-scale and continuous POC-scale test results. In general, similar performance was achieved for both of these

modes of testing for all four POC feedstocks. This similarity between batch and continuous test results indicates that a well-planned series of batch agglomeration tests will provide much information to predict the performance of continuous-mode testing.

Batch testing results using a natural minus 100-mesh fraction of the Illinois No. 5 coal were of particular interest to the project since these data were used for the commercial-scale design, discussed in Section 4. The results of these tests indicate that approximately 79% PSR is achieved at 85% Btu recovery. This performance is a significant improvement over the 64% PSR achieved at 85% Btu recovery using the total raw coal tested during POC operations. This improvement in process performance for the natural fines fraction is due to the liberation characteristics of the coal as reflected in the washability data previously presented in this section under the subheading, "Batch Tests Results Summary."

Pelletization and Drying

Continuous pelletization tests using POC demonstration run product as feedstocks were carried out in a deep-dish pelletizer utilizing 2.5% commmeal flour as a binder. These test results indicate that given good binder dispersion and the appropriate feed moisture content (approximately 25 and 27% for 100- and 200-mesh coal topsizes, respectively), pellets of 8-15 mm diameter can be formed. The pelletization feed moisture content was found to be of primary importance, with excess moisture resulting in large, weak agglomerates, and insufficient moisture reducing the rate of pellet formation. "Green" (wet) pellet strength was sufficient to withstand transfer via belt conveyors.

Thermal drying of "green" pellets was completed in a batch vibrating fluidized-bed dryer at bed temperatures in the range 350-500°F and air flow rates of 200-300 SCFM. Three to four minutes of dryer residence time reduced pellet moistures below 6%, and two minutes of additional evaporative cooling resulted in a final moisture

content of about 4%. Dried pellet strength was such that >99% of the pellets withstood more than 50 drops from a 6-ft height. Pellet crush strengths ranged from 19 to 31 lb. From a drying perspective, the optimal pellet size is approximately 6 mm. At this size, only two minutes of dryer retention time is required; however, these smaller pellets have a lower crush strength.

Pellets formed from 200-mesh grind feedstock were generally stronger than those made with the coarser 100-mesh feed. The pelletization results also indicate that curing of pellets (long-term storage over several weeks) imparts additional strength although this phenomenon was not quantified.

Summary

The pertinent data generated during Phase II testing that are utilized in subsequent sections of this report are the agglomeration performance indicators used for the commercial-scale (200-t/h) plant design. For the Upper Freeport and Pittsburgh No. 8 coals, these process performance values represent average POC performance achieved during the respective demonstration run for each coal. For the Illinois No. 5 coal, the commercial-scale plant design is based on natural minus 100-mesh fines batch agglomeration performance at 85% Btu recovery. Table 3-26 presents the agglomeration performance indicators for the three project coals that are used in the subsequent commercial-scale plant designs.

Table 3-26
Phase II Testing Agglomeration Performance for
Commercial-Scale Design

Coal	Topsize	ROM Btu	ROM
		Rec, %	PSR, %
Upper Freeport	200 M	89.7	81.0
Pittsburgh No. 8	200 M	88.2	64.4
Illinois No. 5	100 M (natural fines)	85.0	79.0

FINAL CONCEPTUAL DESIGN OF COMMERCIAL AGGLOMERATION PLANT

Design Criteria

This section outlines the design criteria used for the final conceptual design of the commercial agglomeration plant. The conceptual design includes two flowsheet options: agglomeration and fines agglomeration. Aspects of the conceptual design including feedstock, general design basis, process considerations, product specifications, and design considerations are discussed.

Feedstock

The agglomeration flowsheet is designed to utilize either Upper Freeport or Pittsburgh No. 8 coal as the raw coal feed. The plant will operate as a mine-mouth facility. Raw coal will be transported to the plant via coal haulage trucks and stored in a sufficient quantity to assure a continuous coal supply to the plant.

The fines agglomeration flowsheet is designed to utilize an Illinois No. 5 coal feedstock. Coal will be transported to a conventional cleaning facility where the agglomeration circuit will process the minus 100-mesh size fraction of the coal feed. In the case of this flowsheet, the economics are developed for the advanced circuit only. Table 4-1 presents the raw coal properties that were used as the plant design basis for both flowsheets.

Table 4-1
Raw Coal Properties

	Upper Freeport	Pittsburgh No. 8	Illinois No. 5
Moisture, %	5.19	4.84	9.66
Ash, % db	13.64	30.88	15.50
Total sulfur, % db	2.83	2.16	1.93
Pyritic sulfur, % db	1.96	1.48	1.21
Organic sulfur, % db	0.84	0.67	0.72
Sulfatic sulfur, % db	0.03	0.01	-
Calorific value, Btu/lb	13,300	10,167	12,430

The properties presented in Table 4-1 for the Illinois No. 5 coal are not derived from POC feed analysis but represent composited washability data for this coal. The washability values were selected since they best represent the fines fraction of the coal.

Overall Design

The design of the agglomeration flowsheet incorporates conventional cleaning for economical removal of sulfur and ash from the ROM coal. The precleaned product is then processed by means of selective agglomeration. Precleaning of the coal reduces agglomeration throughput and improves overall process performance while decreasing process capital and operating costs. Table 4-2 summarizes the general design basis and major unit operations used for this flowsheet.

Table 4-2
Agglomeration Flowsheet Design Basis

Capacity	200 t/h product
Coal supply	ROM coal
Site	Greenfield
Actual operating hrs	7560 hours/year
Availability	90% minimum
Unit Operations	
Crushing	To ¼" topsize
Precleaning	96% Btu recovery minimum +100 mesh at 1.5-1.9 SG minus 100 mesh to agglomeration
Grinding	Size reduction to 95% minus 200 mesh
Agglomeration	Diesel (0.5%)
Dewatering	Vacuum filtration to 25% moisture max.
Size enlargement	Pelletization: starch and asphalt
Drying	Vibrating fluidized-bed to 4% moisture

The fines agglomeration flowsheet utilizes conventional coal cleaning equipment to reduce ash and sulfur in the plus 100-mesh coal fraction. The natural fines fraction (minus 100 mesh) is cleaned using selective agglomeration.

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eration. The cleaned fractions are then recombined to produce a clean-coal product. No size reduction of the feed to the agglomeration circuit is necessary, thereby reducing capital and operating costs. Table 4-3 shows the general design basis used for the natural fines circuit. Flowsheet elements are discussed in detail later in this section.

Table 4-3
Natural Fines Flowsheet Design Basis

Plant capacity	1500 t/h raw coal
Agg. circuit capacity	114 t/h raw coal fines (circuit feed)
Location	Within existing coal cleaning plant
Coal supply	Minus 100-mesh natural fines from conventional plant
Site	Retrofit
Actual operating hrs	5400 hours/year
Availability	90% minimum
Conventional Cleaning	
6" x 1/2" fraction	Heavy-media vessel at 1.7 SG
1/2" x 28 mesh fraction	Heavy-media cyclones at 1.7 SG
28 x 100 mesh fraction	Spiral concentrators
-100 mesh fraction	Direct to agglomeration circuit
Retrofit Unit Operations	
Crushing	None
Grinding	None
Agglomeration	0.8% diesel
Dewatering	Vacuum filtration to 25% moisture
Size enlargement	None
Drying	None

Process Conditions

Each project coal requires unique process conditions for optimal coal recovery and sulfur and ash rejection. Table 4-4 lists the optimal coal-specific agglomeration conditions for the selective agglomeration process. For Illinois No. 5 coal, it is assumed that the optimized variables determined during POC testing with a raw coal can be used for design of the natural fines circuit.

Plant Performance

The overall plant performance for the agglomeration flowsheet was estimated based on the performance of the precleaning and agglomeration circuits. Precleaning performance is based on the actual performance observed during test coal preparation. Agglomeration performance is based on POC demonstration run results using Upper Freeport and Pittsburgh No. 8 coals.

The natural fines agglomeration circuit performance was based on batch test results, derived correlations, and POC results. Pyritic sulfur rejection was estimated using correlations developed during POC testing.

Table 4-4
Agglomeration Conditions

	Upper Freeport	Pittsburgh No. 8	Illinois No. 5
High-Shear Conditions			
Feed size	95% minus 200 mesh	95% minus 200 mesh	Nat. minus 100 mesh
HS residence time, min	1	1	1
HS solids conc., %	20	20	20
Agglomeration Conditions			
Agglomerate recovery method	2-stage flotation	2-stage flotation	2-stage flotation
Agglomerant	Diesel	Diesel	Diesel
Agglomerant conc., %	0.5	0.5	0.8
Flotation Conditions - Stage 1			
Solids conc., %	6	6	6
Residence time, min	4	6	6
Frother dosage, lb/t	0.29	0.29	0.13
Flotation Conditions - Stage 2			
Solids conc., %	6	6	6
Residence time, min	4	6	6
Frother dosage, lb/t	0.29	0.29	0.13
Dewatering			
Method	Vac. filter	Vac. filter	Vac. filter
Drying	Yes	Yes	No
Pelletization	Yes	Yes	No

Precleaning, agglomeration, and overall performance values on a ROM basis are provided in Table 4-5 for each of the project coals. The agglomeration performance values for the Illinois No. 5 coal represent the natural fines agglomeration circuit only; overall plant performance for this case is discussed in Section 6, Economics of Commercial Agglomeration Plant.

Product Specifications

The commercial agglomeration plant will produce a dried pelletized product at least 6 mm in diameter containing 2% starch and 0.5% asphalt. The pelletized product will be easily handleable by conventional utility coal handling systems and suitable for utility combustion.

Table 4-5
Plant Performance

	Upper Freeport	Pittsburgh No. 8	Illinois No. 5
Precleaning Performance			
Yield, %	92.0	67.3	-
Btu recovery, %	96.7	93.3	-
Total sulfur rej., %	45.7	31.7	-
Pyritic sulfur rej., %	58.2	44.0	-
Ash rejection, %	34.6	83.4	-
Agglomeration Performance			
Yield, %	88.4	92.0	61.0
Btu recovery, %	92.8	94.6	85.0
Total sulfur rej., %	37.4	21.2	59.0
Pyritic sulfur rej., %	54.4	36.4	79.0
Ash rejection, %	46.4	35.7	88.6
Overall Performance, ROM basis			
Yield, %	81.3	61.9	-
Btu recovery, %	89.7	88.2	-
Total sulfur rej., %	66.0	46.2	-
Pyritic sulfur rej., %	81.0	64.4	-
Ash rejection, %	64.9	89.3	-

The product from the natural fines facility will consist of the conventionally cleaned coarse coal combined with the agglomerated fines fractions. Table 4-6 presents the final product characteristics expected for the three project coals. These analyses are based on the process performance assumptions discussed previously.

Table 4-6
Product Analysis

	Upper Freeport	Pittsburgh No. 8	Illinois No. 5
Moisture, %	4.0	4.0	13.5
Ash, % db	5.9	5.3	7.2
Total sulfur, % db	1.18	1.88	1.51
Pyritic sulfur, % db	0.46	0.85	0.80
Calorific value, Btu/lb	14,674	14,487	13,438
lb SO ₂ /MBtu	1.61	2.60	2.25

Design Considerations

The commercial plant shall comply with all applicable health, safety, and environmental regulations. Several regulatory permits, approvals, and reviews will be obtained from appropriate governmental agencies. Environmental requirements are discussed later.

The commercial agglomeration plant will be operated 24-hour/day, 7 days/week, for 50 weeks/year. Raw coal handling operations will be designed to accommodate one 8-hour shift per day. The natural fines circuit design will be based on typical conventional cleaning plant operating practice of 5400 hours/year since it is a retrofit application.

The commercial plant shall be a fully instrumented facility with automated startup, operation, and shutdown from a central control room. All critical process streams will be monitored and controlled to maintain optimum process conditions. Additional details on instrumentation and control are provided later in this section.

As presented above in the design basis, the plant will maintain at least 90% availability. To satisfy this criterion, equipment spares are provided as needed. Details on plant availability are provided later in this section.

Conceptual Flowsheet

This portion of the report discusses the agglomeration and fines agglomeration flowsheets. Each process flowsheet is described in detail by unit operation. Material balances and assumptions are also discussed.

Agglomeration Process Description

The flowsheet shown in Figure 4-1 depicts the entire agglomeration process including the coal precleaning circuits. Material balances are provided in Tables 4-7 and 4-8 for the Upper Freeport and Pittsburgh No. 8 feedstocks, respectively.

The first step following precleaning is grinding of the 1/4" x 0 coal to a nominal minus 200-mesh size in three stages. The ground coal slurry is agglomerated in a high-shear vessel, and the microagglomerates are recovered by flotation, dewatered, and pelletized. The pellets are then dried and stored in silos. The refuse is thickened in a water clarifier, filtered, and sent to a landfill. The clarified water is treated as necessary and recycled back to the process. This process is described in further detail below.

Coal Receiving, Storage, and Crushing. The 2" x 0 size plant feed coal is received via truck dump grates in underground feed hoppers. Vibrating feeders regulate the transfer of coal onto the feed coal belt conveyor which transports it to a stacking tube that provides live storage for 15,000 tons of coal. The stacking tube

Figure 4-1. Commercial Agglomeration Flowsheet

Table 4-7
Upper Freeport Coal Material Balance

Stream	No.	Coal t/h	Weight Percent			Water gpm	Diesel gpm	Asphalt gpm	Starch gpm	Slurry gpm	% Sol	% H ₂ O	Cal. Val. Btu/lb
			Ash	PS	TS								
Coal to precleaning	1	246	13.6	1.96	2.83	51				789	95	5	13,300
Water to precleaning	2					447				447			
Precleaning refuse	3	20	59.0	14.27	16.17	26				85	75	25	5,434
-100M coal to rod mill	4	16	15.3	3.40	4.25	192				240	25	75	12,798
Clean coal to rod mill	5	210	9.3	0.70	1.47	280				911	75	25	14,074
Water to rod mill	6					433				433			
Slurry from grinding	7	226	9.7	0.89	1.67	905				1,584	50	50	13,984
Water to high shear	8					2,711				2,711			
Diesel to high shear	9						5			5			
High shear slurry	10	226	9.7	0.89	1.67	3,617	5			4,300	20	80	13,984
Water to flotation 1	11					10,562				10,562			
Slurry to flotation 1	12	226	9.7	0.89	1.67	14,178	5			14,862	6	94	13,984
Tails from flotation 1	13	18	42.4	4.28	5.56	10,856				10,911	1	99	7,991
Froth from flotation 1	14	208	6.8	0.59	1.33	3,322	5			3,950	20	80	14,514
Water to flotation 2	15					6,776				6,776			
Slurry to flotation 2	16	208	6.8	0.59	1.33	13,025	5			13,653	6	94	14,514
Tails from flotation 2	17	8	29.8	3.87	5.05	9,830				9,854	0	100	10,484
Froth from flotation 2	18	200	5.9	0.46	1.18	3,195	5			3,800	20	80	14,674
Vacuum filter filtrate	19					2,927				2,927			
Vacuum filter cake	20	200	5.9	0.46	1.18	268	5			873	75	25	14,674
Binder to solids mixer	21							4	12	16			
Pellets to dryer	22	200	5.9	0.46	1.18	268	5	4	12	889	75	25	14,674
Dried product to storage	23	200	5.9	0.46	1.18	34	5	4	12	655	96	4	14,674
Moisture to flue	24					234				234			
Tails to clarifier	25	26	38.6	4.16	5.40	20,956				21,035	0	100	8,741
Clarifier overflow	26					20,641				20,641			
Clarifier underflow	27	26	38.6	4.16	5.40	316				395	25	75	8,741
Agglomeration refuse	28	26	38.6	4.16	5.40	45				124	70	30	8,741
Belt press effluent	29					271				271			
Refuse to landfill	30	46	47.3	8.48	10.01	71				209	72	28	7,326
Water to treatment	31					413				413			
Untreated process water	32					20,228				20,228			
Water to head tank	33					20,641				20,641			
Water from head tank	34					20,929				20,929			
Make-up water	35					288				288			

Table 4-8
Pittsburgh No. 8 Coal Material Balance

Stream	No.	Coal t/h	Weight Percent			Water gpm	Diesel gpm	Asphalt gpm	Starch gpm	Slurry gpm	% Sol	% H ₂ O	Cal. Val. Btu/lb
			Ash	PS	TS								
Coal to precleaning	1	323	30.9	1.48	2.16	63				1,032	95	5	10,167
Water to precleaning	2					568				568			
Precleaning refuse	3	106	78.8	1.99	2.10	141				458	75	25	2,095
-100M coal to rod mill	4	19	36.5	1.91	2.42	225				281	25	75	9,798
Clean coal to rod mill	5	199	4.9	1.17	2.17	265				861	75	25	14,494
Water to rod mill	6					380				380			
Slurry from grinding	7	217	7.6	1.23	2.19	870				1,522	50	50	14,089
Water to high shear	8					2,605				2,605			
Diesel to high shear	9						4			4			
High shear slurry	10	217	7.6	1.23	2.19	3,475	4			4,131	20	80	14,089
Water to flotation 1	11					10,147				10,147			
Slurry to flotation 1	12	217	7.6	1.23	2.19	13,622	4			14,279	6	94	14,089
Tails from flotation 1	13	13	35.4	5.78	5.94	10,358				10,397	1	99	9,236
Froth from flotation 1	14	204	5.8	0.94	1.95	3,264	4			3,882	20	80	14,402
Water to flotation 2	15					6,606				6,606			
Slurry to flotation 2	16	204	5.8	0.94	1.95	12,798	4			13,415	6	94	14,402
Tails from flotation 2	17	4	29.5	5.00	5.21	9,602				9,615	0	100	10,417
Froth from flotation 2	18	200	5.3	0.85	1.88	3,196	4			3,800	20	80	14,487
Vacuum filter filtrate	19					2,928				2,928			
Vacuum filter cake	20	200	5.3	0.85	1.88	268	4			872	75	25	14,487
Binder to solids mixer	21							4	12	16			
Pellets to dryer	22	200	5.3	0.85	1.88	268	4	4	12	888	75	25	14,487
Dried product to storage	23	200	5.3	0.85	1.88	34	4	4	12	655	96	4	14,487
Moisture to flue	24					234				234			
Tails to clarifier	25	17	34.0	5.59	5.76	20,140				20,192	0	100	9,527
Clarifier overflow	26					19,930				19,930			
Clarifier underflow	27	17	34.0	5.59	5.76	209				262	25	75	9,527
Agglomeration refuse	28	17	34.0	5.59	5.76	30				82	70	30	9,527
Belt press effluent	29					179				179			
Refuse to landfill	30	123	72.4	2.50	2.62	171				540	74	26	3,148
Water to treatment	31					399				399			
Untreated process water	32					19,532				19,532			
Water to head tank	33					19,930				19,930			
Water from head tank	34					20,307				20,307			
Make-up water	35					376				376			

discharge is controlled by four variable-speed vibrating feeders that maintain the proper feed rate to the conveyor belt feeding the coal crusher. This belt is equipped with a magnet for tramp iron removal. The crusher reduces the 2" x 0 raw coal feed to a 1/4" topsize product. The crushed coal is then received by a surge hopper providing one-half hour of surge capacity.

Coal Precleaning. The precleaning circuit for the commercial agglomeration plant is presented in Figure 4-2. The precleaning system will generate the following streams:

- 1/4" x 100-mesh clean coal (25% moisture)
- 100-mesh x 0 raw coal (25% solids)
- 1/4" x 100-mesh refuse (25% moisture)

Crushed coal is withdrawn from storage and fed to the precleaning circuit 24 hours/day at a rate of 246 t/h for Upper Freeport coal and 323 t/h for Pittsburgh No. 8 coal. The feed coal is slurried in a wetting box and screened at 28 mesh. The plus 28-mesh fraction reports to the heavy-media cyclone feed sump and the minus 28-

mesh fraction reports to the classifying cyclone feed sump, from which it is pumped to a classifying cyclone and separated at 100 mesh. The raw coal is thus split into three size fractions: plus 28 mesh, 28 x 100 mesh, and 100 mesh x 0.

The plus 28-mesh fraction is pumped to the heavy-media cyclone. The clean coal (overflow) from the heavy-media cyclone flows to drain-and-rinse screens for magnetite removal and dewatering. The screen oversize (plus 28-mesh clean coal) then reports to the clean coal belt for transport to the coal-grinding circuit.

The refuse (underflow) from the heavy-media cyclone reports to separate drain-and-rinse screens and then to the refuse belt for disposal. The effluent from the drain section of both the refuse and clean coal drain-and-rinse screens is recycled to the heavy-media feed sump, while the effluent from the rinse section of both the refuse and clean coal drain-and-rinse screens is recycled to the heavy-media sump via the magnetite separator.

The classifying cyclone underflow (28 x 100 mesh) is separated by gravity flow in a spiral classifier. The clean coal fraction is dewatered on 100-mesh screens. The

Figure 4-2. Precleaning Circuit

oversize (28 x 100-mesh clean coal) reports to the clean coal conveyor and is transported to the coal-grinding circuit along with the plus 28-mesh clean coal.

The refuse fraction from the spiral classifier is also dewatered on screens. The screen oversize (28 x 100-mesh refuse) goes to the refuse belt for disposal. The effluents from both the clean coal and refuse screens are recycled to the classifying cyclone feed sump. The spiral middlings and water streams are also recycled to the classifying cyclone sump.

The cyclone overflow (minus 100 mesh) is sent to a static thickener. The underflow from the thickener (~25% solids) is pumped to the agglomeration circuit rod mill feed. The static thickener overflow is recycled to the raw coal deslime screen.

Coal Grinding. The 1/4" x 100-mesh precleaning circuit product is fed to the grinding circuit at a nominal rate of 217 t/h for the Upper Freeport coal and 210 t/h for the Pittsburgh No. 8 coal. This stream is temporarily stored in a rod mill feed hopper and then delivered to the rod mill. The rod mill hopper is designed to have about one-half hour of storage capacity for the grinding system. A weigh-belt feeder reclaims the coal from the hopper to the rod mill. Precleaning static thickener underflow (minus 100-mesh coal) and water are added to the rod mill, diluting the input streams to about 50% solids. The rod mill grinds the 1/4" x 0 coal to a nominal 28-mesh topsize.

The rod mill product discharges to the ball mill sump from which it is pumped to the ball mill. The ball mill operates at the same solids concentration as the rod mill and is designed to grind the 28-mesh topsize coal to a nominal 100-mesh topsize ($d_{50} = 60$ microns). An additional ball mill operated in series further grinds the coal to a nominal 200-mesh topsize ($d_{50} = 25$ microns). The ball mill product discharges to a sump from which it is pumped to a slurry holding vessel. The slurry holding vessel is designed to hold a 10-minute supply of ground coal slurry at 50% solids and is equipped with an agitator to assure that the slurry is homogenous.

Agglomeration. Coal slurry from the holding vessel flows to a high-shear sump. Precleaning circuit static thickener underflow and water are added to the sump to dilute the slurry to 20% solids. The diluted slurry is then pumped from the high-shear sump to six high-shear vessels via a 6-way slurry distributor. The percent solids and flow rate of the high-shear feed are monitored, and these data are used to meter 0.5% diesel (on a dry coal feed basis) to the high-shear vessels. The high-shear

vessels are designed to provide one minute of retention time. During this period, phase inversion occurs and microagglomerates are formed, with liberated ash and pyrites reporting to the aqueous phase.

The high-shear vessels use a top discharge from which the agglomerated slurry flows to two banks of five flotation cells each, where it is diluted to 6% solids. Each bank of flotation cells is fed by three high-shear vessels, resulting in two parallel streams. During flotation, microagglomerates are recovered from the mineral matter-bearing aqueous phase. The retention time is coal-dependent, and was provided in Table 4-4. Tailings from flotation are sent to the plant tailings sump.

The flotation product from each of the two flotation banks, assumed to have a solids concentration of approximately 20%, flows by gravity to the flotation feed sumps where it is diluted to 6% solids. The slurry from these sumps is pumped to two banks of four flotation cells where flotation is repeated at a residence time similar to that of the first stage of flotation. The tailings are sent to the plant tailings sump. The flotation product is once again assumed to be at 20% solids.

Agglomerate Dewatering. The product from the second stage of flotation flows by gravity to conditioning tanks. These conditioning tanks are designed to provide one minute of storage and include mixers to disperse the froth. The discharge from the two conditioning tanks is pumped to two rotary vacuum disc filters. The filters are designed to yield a product with a moisture content of approximately 25%. The filtrate from the vacuum filters is recycled to the flotation feed sump.

Size Enlargement, Drying, and Product Loadout. The filter cake from each of the two vacuum filters is split into two streams, resulting in four parallel streams that flow by gravity to solids mixers. An asphalt emulsion is added to the solids mixers by means of a metering pump at a rate providing 0.5% asphalt on a solids basis. Starch (2.0% on a dry solids basis) is also fed to the mixers via a feed hopper and weigh-belt feeder. The purpose of the solids mixers is to disperse the binders (asphalt and starch) within the cake. The four solids mixers discharge onto pelletizers that are sized to produce pellets 6-8 mm in size. The pelletizers discharge onto belt conveyors which transport the pellets to the dryer feed hoppers that provide one hour of storage capacity for the "green" pellets. Each of the four hoppers feeds a vibratory fluidized-bed dryer. The dryers are designed to reduce the moisture content of the pellets to 4% in order to increase their strength. The pellet retention time in the dryer is estimated to be 2-3 minutes. The dried pellets

are then discharged onto cooling conveyors. The final clean-coal product is stored in silos which provide a 3-day storage capacity. Product loadout will be by truck.

Water Clarification and Treatment. The plant tailings streams from the agglomeration circuit are pumped from the tailings sump to the water clarifier. Flocculants and lime (for pH control) are used to facilitate water clarification. The thickened refuse stream, at approximately 25% solids, is pumped to a belt filter press. The refuse belt-press filtrate is recycled back to the water clarifier. Clarified water is circulated between the clarifier overflow tank and the head tank, with process water being removed from the head tank as required. A portion of this clarified water stream (approximately 2%) is treated to keep contamination levels within acceptable limits.

Refuse Dewatering, Handling, and Disposal. The clarifier underflow (agglomeration plant refuse) is pumped to the belt press via the thickener discharge sump. The belt press is designed to process approximately 25 t/h of solids and will dewater the refuse stream to a moisture content of 25-30%. This fine refuse is mixed with the coarse refuse from the precleaning facility and transported to a landfill site by truck.

Natural Fines Agglomeration Process Description

Figure 4-3 presents the flowsheet for the conventional plant incorporating a natural fines agglomeration circuit. The material balance is provided in Table 4-9. In this process flowsheet, the natural fines fraction of the coal (minus 100 mesh) is cleaned. Initially, 1500 t/h of raw coal is received at a conventional cleaning plant and separated into size fractions. Water is added to the natural fines fraction during processing. The natural fines slurry is then fed to the fines agglomeration circuit at a rate of 114 t/h.

Water added to the fines fraction during conventional processing is removed in a static thickener. The underflow from the static thickener flows into the high-shear sump at 20% solids. The slurry is agglomerated in the high-shear vessels and the microagglomerates are recovered by flotation, dewatered, and recombined with the conventionally cleaned coal. The solid refuse is thickened in a water clarifier, filtered, and sent to a landfill. The clarified water is treated as necessary and recycled back to the agglomeration circuit.

Conventional coal cleaning details are not addressed in the discussion of the flowsheet because the fines agglomeration circuit can theoretically be incorporated

into a variety of conventional cleaning flowsheets. A hypothetical layout for a conventional cleaning plant is included in Section 6, Economics of Commercial Agglomeration Plant, to illustrate the economics of incorporating a fines agglomeration circuit.

Process details for the agglomeration, agglomerate dewatering, and waste treatment unit operations are similar to those mentioned above, under the subheading "Agglomeration Process Description."

Thickening. Since the agglomeration circuit feed is a cyclone overflow stream of low solids concentration, a thickening step is necessary to increase the percent solids to 20% prior to high shear. Alternatively, high shear could be performed at a lower solids concentration, trading the cost of thickening for additional high-shear capacity. The thickening step was chosen because preliminary cost data indicated that increasing the high-shear capacity would be the more expensive alternative. Also, since high-shear agglomeration at the low solids concentration was not examined during POC testing, performance at these conditions is not known.

Agglomeration. This unit operation is similar to that in the agglomeration flowsheet. The differences between the two operations include the following:

- Natural minus 100-mesh fines rather than ground minus 200-mesh coal is fed to the high-shear vessels.
- Four vessels are used for high-shear mixing.
- Flotation cell capacity was reduced to accommodate lower flow rates. Four flotation cells are used for the first and second banks of flotation cells in each parallel-flotation circuit.
- Coal-dependent agglomeration conditions are changed to reflect the Illinois No. 5 feed.

Details regarding the agglomeration conditions for the Illinois No. 5 coal were presented in Table 4-4.

Other Operations. The following unit operations are similar to their counterparts in the agglomeration process description: agglomerate dewatering, water handling and treatment, and refuse dewatering, handling, and disposal. Unit operations such as coal receiving, storage, crushing, precleaning, grinding, size enlargement, drying, and product loadout are not incorporated into the fines circuit flowsheet.

Figure 4-3. Natural Fines Agglomeration Flowsheet

Table 4-9
Illinois No. 5 Coal Material Balance

Stream	No.	Coal t/h	Weight Percent			Water gpm	Diesel gpm	Asphalt gpm	Starch gpm	Slurry gpm	% Sol	% H ₂ O	Cal. Val. Btu/lb
			Ash	PS	TS								
Coal to conventional plant	1	1500	15.5	1.21	1.93	580				5,080	91	9	12,430
Added water	2					9,054				9,054			
Conventional circuit refuse	3	183	58.4	3.83	4.72	313				862	70	30	7,579
Fines to thickener	4	114	34.2	1.11	1.67	8,664				9,006	5	95	9,192
Coarse coal to product	5	1203	7.2	0.82	1.53	656				4,266	88	12	13,474
Thickener overflow	6					6,844				6,844			
Fines to high shear	7	114	34.2	1.11	1.67	1,820				2,162	20	80	9,192
Deleted	8												
Diesel to high shear	9						4			4			
High-shear slurry	10	114	34.2	1.11	1.67	1,820	4			2,166	20	80	9,192
Water to flotation 1	11					5,320				5,320			
Slurry to flotation 1	12	114	34.2	1.11	1.67	7,140	4			7,486	6	94	9,192
Tails from flotation 1	13	41	77.5	2.25	2.53	6,462				6,584	2	98	3,548
Froth from flotation 1	14	73	10.0	0.47	1.19	679	4			902	30	70	12,349
Water to flotation 2	15					3,392				3,392			
Slurry to flotation 2	16	73	10.0	0.47	1.19	4,578	4			4,801	6	94	12,349
Tails from flotation 2	17	4	80.3	2.24	2.71	3,932				3,943	0	100	3,394
Froth from flotation 2	18	70	6.4	0.38	1.11	645	4			858	30	70	12,809
Vacuum filter filtrate	19					507				507			
Vacuum filter cake	20	70	6.4	0.38	1.11	139	4			351	67	33	12,809
Deleted	21												
Deleted	22												
Product to storage	23	1273	7.2	0.80	1.51	795	4			4,617	86	14	13,438
Deleted	24												
Tails to clarifier	25	44	77.7	2.25	2.54	10,851				10,985	2	98	3,535
Clarifier overflow	26					10,318				10,318			
Clarifier underflow	27	44	77.7	2.25	2.54	534				667	25	75	3,535
Agglomeration refuse	28	44	77.7	2.25	2.54	76				210	70	30	3,535
Belt press effluent	29					457				457			
Refuse to landfill	30	227	62.2	3.52	4.30	390				1,072	70	30	6,788
Water to treatment	31					206				206			
Untreated process water	32					10,111				10,111			
Water to head tank	33					10,318				10,318			
Water from head tank	34					8,712				8,712			
Excess water	35					1,605				1,605			

Material Balances

The material balances developed for each test coal and presented previously, utilized estimates of process performance, coal quality, and precleaning performance (where applicable) provided above in the design criteria. Previous experience with conventional coal plants and selective agglomeration unit operations was used to estimate the performance within each major unit operation to complete the material balances. Several assumptions were made during the conceptual design regarding feed coal quality, plant performance, and plant operations. These assumptions are discussed below.

Material Balance Assumptions. The following assumptions were used to derive a material balance for the agglomeration flowsheet using the Upper Freeport and Pittsburgh No. 8 coals:

- Feed coal quality was based on project feed coal qualities as measured during Phase II testing.
- Precleaning performance for the commercial design will be similar to that observed during test coal preparation.
- Agglomeration performance for the commercial design is based on demonstration run performance for both coals.
- The distribution of ash and sulfur between flotation stages corresponds to POC observations.

Along the same lines, certain assumptions were made during development of the material balance for the fines agglomeration flowsheet. These assumptions are:

- Feed coal quality is based on proprietary washability analysis for the minus 100-mesh fraction of the sample tested.
- The distribution of ash and sulfur between flotation stages corresponds to POC observations.

Agglomeration performance was based on batch test results using Illinois No. 5 natural fines as the feed. The POC parametric testing and demonstration runs were performed using the bulk of the entire coal ground to fine sizes and these results are not necessarily indicative of natural fines agglomeration performance. Batch tests were plotted, and performance at 85.0% Btu recovery was interpolated.

Unit Operations and Uncertainties

Each unit operation utilized in the commercial agglomeration plant and the natural fines agglomeration circuit was discussed briefly in the process descriptions presented earlier. In this section, the design criteria and uncertainties, where applicable, are detailed for each unit operation. The design of each operation is based on test work conducted throughout the project and current engineering norms.

Each section of the commercial agglomeration plant is designed to process either Upper Freeport or Pittsburgh No. 8 coal. Use of this design basis generates one flowsheet and equipment list to handle the required processing rates for either feed coal.

It may appear that a design of this type overestimates plant equipment sizing for the feed coal requiring less processing capacity. However, the majority of the equipment costs are for circuits where sizing requirements are similar for both coals. Therefore, the final capital costs for either coal closely approximate each other using this approach.

This design approach does not apply to the natural fines circuit. This circuit is designed to process Illinois No. 5 raw coal fines and adaptability to other coals is not considered.

Coal Receiving, Storage, and Crushing (Agglomeration Flowsheet)

A number of criteria were taken into consideration during the design of the raw coal receiving, crushing, and storage circuit. These criteria include feed rate, availability requirements, and economics. In addition, a product size requirement of 1/4" x 0, with no more than 30% minus 28 mesh, was imposed on the circuit.

The design of this circuit is based on the Pittsburgh No. 8 raw coal feed rate of 323 t/h, which is more conservative than the Upper Freeport raw coal feed rate of 246 t/h. The raw coal handling system will be operated for 8 hours per day while the crushing system will be operated for 24 hours per day. A 24-hour raw coal handling system operation is not desirable due to high labor costs. Since the raw coal handling operation is scheduled for 8 hours of operation per day, the handling system is designed to process coal at the rate of 970 t/h. This design allows the circuit to meet the daily plant coal requirements of 7760 tons.

The uncertainty associated with the coal handling, storage, and crushing circuit is the ability of the hammer mill to produce a 1/4"-topsize product with no more than 30% passing minus 28 mesh.

Coal Precleaning (Agglomeration Flowsheet)

The coal precleaning circuit has been designed to meet the following objectives:

- Ability to process either Upper Freeport or Pittsburgh No. 8 coals
- Maximum raw coal feed rate of 350 t/h on dry basis
- Maximum feed-coal ash of 40%, which is equivalent to approximately 15% mine dilution
- Capability of processing 1/4"-topsize coal containing no more than 30% minus 28-mesh fines
- 90% plant availability for 24 hours/day, 7 days/week, 50 weeks/year operation
- Flexible specific gravity of separation in the range of 1.5 to 1.9

The conceptual design of the precleaning circuit utilizes heavy-media cyclones and spiral concentrators. Heavy-media cyclones were chosen to clean the 1/4" x 28-mesh coal fraction because of their ability to process a 1/4"-topsize feed, variable separation gravity, and excellent separation efficiency.

The 28 x 100-mesh coal is cleaned by means of spiral concentrators. Utilizing spiral concentrators has many benefits including:

- Good performance at high gravity separations (+1.70)
- Low water requirements
- Low operating costs
- Easily adjustable separation gravity
- No need for reagents that could subsequently affect agglomeration performance

No precleaning of the minus 100-mesh raw coal fraction is specified. This fraction is sent to the rod mills in the grinding circuit. Bypassing the precleaning circuit elimin-

ates fine coal Btu losses and reduces precleaning costs pertaining to thickening and dewatering.

The use of water-only cyclones to clean the plus 28-mesh coal was rejected because of the difficulty in changing the separation gravity and reduced efficiency at a separation gravity below 1.6. Froth flotation of the minus 28-mesh fraction was rejected due to uncertainties regarding the effects of flotation reagents on selective agglomeration performance. Also, varying the flotation froth solids concentration could make the circuit feed and solids concentration difficult to control.

The equipment design for this circuit was based on the Pittsburgh No. 8 raw coal feed rate. A higher feed rate is required since the Pittsburgh No. 8 coal contains significantly higher ash than the Upper Freeport coal.

The hardness of the Pittsburgh No. 8 coal was the controlling factor for the crusher design. However, the crusher will not significantly affect the size distribution of the softer Upper Freeport coal since material is removed from the crusher as the desired size (1/4") is reached. The net result is that with softer coals there is a greater crushing capacity at a constant power input, or less power draw at a constant capacity.

The uncertainties associated with the precleaning circuit include its ability to meet the project objectives. Though the precleaning circuit is designed to meet these objectives, no validation tests have been conducted.

Coal Grinding (Agglomeration Flowsheet)

The objectives of the coal-grinding circuit are to produce a nominal 200-mesh topsize product ($d_{50} = 25$ microns) and provide a homogenous feed to the agglomeration circuit.

Open-circuit wet grinding has been selected as the grinding operation to be used in the plant. The circuit contains three mills in series: a rod mill and two ball mills. The rod mill is used for size reduction to a nominal 28-mesh topsize. The first ball mill carries the reduction to a nominal 100-mesh topsize. The second ball mill further grinds the coal to a nominal 200-mesh topsize. Grinding is carried out in all three mills at about 45-50% solids.

Thickener underflow containing 100-mesh x 0 raw coal is used to pulp the rod mill grinding to 50% solids. This slurry concentration is then carried through the remaining two ball mills. Pulping the ball mills with this stream reduces process water consumption.

Consideration was also given to the difference in grindability between the Upper Freeport and Pittsburgh No. 8 coals. As the grindability of the coal increases, a finer product size distribution will result given the same power input and coal feed rate. To compensate for changes in grindability, the ball charge in the mill will be adjusted. Although this results in slightly less efficient grinding, the desired product size distribution is maintained.

Agglomeration (Both Flowsheets)

The coal agglomeration circuit is designed to agglomerate the coal feed and separate the micro-agglomerates from the aqueous phase to produce a clean-coal product. Slurry from the grinding circuit is fed to the high-shear mixers at 20% solids concentration. Agglomeration takes place in a high-shear mixing step. During high-shear agglomeration, a small amount of diesel is added as an agglomerant. Diesel requirements vary from 0.5%, on a dry coal weight basis, for Upper Freeport and Pittsburgh No. 8 coals to 0.8% for the Illinois No. 5 coal used in the natural fines circuit. Upon mixing, an inversion takes place in which the dispersed agglomerant coats the coal surfaces and allows the formation of microagglomerates.

The high-shear product is then fed to a bank of flotation cells to separate the microagglomerates from the mineral matter-bearing water. After the first stage of flotation, the recovered microagglomerates are floated again in a second flotation step. Frother is added in both flotation cells to enhance recovery. The recovered micro-agglomerates are then fed to the agglomerate dewatering circuit.

Agglomeration performance was assumed to be the average performance achieved during the demonstration run at the conditions considered optimal. Solids concentration and residence time in each flotation step were determined experimentally during POC parametric testing. High-shear residence times were also determined in this manner. As stated earlier, the optimal agglomeration conditions for the natural-fines circuit were assumed to be those determined during testing of the ROM coal sample. The performance for the Illinois No. 5 natural-fines circuit was determined using laboratory batch tests. The performance values and coal-dependent agglomeration conditions were discussed previously in this section.

Thickening (Natural Fines Circuit)

A static thickener was added to the natural fines circuit to increase the solids concentration of the cyclone overflow stream prior to high shear. The cyclone overflow stream is approximately 2-4% solids in typical conventional cleaning plants. Alternatives to the thickening step include lowering the solids concentration during the high-shear mixing step. This alternative would eliminate the cost of the static thickener and associated equipment and increase the capital and operating costs associated with the high-shear step. The thickening step was chosen because preliminary cost data indicated that increasing the high-shear capacity was the more expensive alternative. Also, since high-shear agglomeration at a low solids concentration was not examined during POC testing, performance at these conditions is not known.

Agglomerate Dewatering (Both Flowsheets)

The objectives of the agglomerate dewatering circuit are to mechanically dewater the microagglomerates to a level suitable for pelletization and achieve a solids recovery of more than 99%.

As discussed in the agglomeration process description, the recovered microagglomerates are sent to a holding tank for froth dispersion and then mechanically dewatered. Vacuum filtration is used to reduce the moisture to approximately 25% total moisture for the Upper Freeport and Pittsburgh No. 8 coals. A total moisture reduction to 33% was assumed in the design for Illinois No. 5 coal since the coal has a substantially higher inherent moisture.

Size Enlargement, Drying, and Product Loadout (Agglomeration Flowsheet)

The objective of size enlargement is to produce a product that is compatible with typical utility coal handling and storage systems. The following criteria have been established for the final clean-coal product:

- Larger than 4 mm in size
- Handleable for storage and loadout operations and meeting end-user requirements with respect to bin flowability and dustiness
- Moisture content acceptable to utilities

Experience with Phase II pelletization testing shows that the handleability and weatherability of the clean coal pellets are excellent. The pellets have high strength and minimal odor. It is anticipated that binder addition will be adjusted to change pellet strength as required.

Performance criteria have also been established for product storage and loadout facilities. Storage facilities will provide safe and odor-free storage of pellets. Product loadout facilities will be capable of loading conventional haulage trucks.

Water Clarification and Treatment (Both Flowsheets)

The purpose of this unit operation is to produce clarified water for recycle to the processing plant and to recover the refuse in a thickened form. In addition, a portion of the recycle water is treated to keep contamination levels within acceptable limits.

The recycle water is treated by means of an activated-carbon bed. Two beds are provided, allowing one bed to be taken out of service for maintenance or regeneration.

Refuse Dewatering, Handling, and Disposal (Both Flowsheets)

The objective of the refuse disposal circuit is to dispose of high-ash coal solids generated by the cleaning facility. Disposal of all refuse must comply with applicable environmental regulations.

Underflow from the thickener is dewatered using a belt press. The dewatered refuse is then combined with similar refuse from the precleaning circuit and transported to a landfill site by truck.

Agglomerant and Reagent Storage and Delivery (Both Flowsheets)

The objectives of this unit operation are to provide sufficient storage and delivery of all necessary oils and reagents. The reagents considered in the design include diesel, asphalt emulsion, starch, flocculant, lime, and frother. Oil and reagents will be added at various points throughout the facility including the high-shear vessel, solids mixer, pelletizer, flotation cells, and water clarifier. Asphalt emulsion and starch are not required in the natural fines circuit.

Diesel. Approximately 4.5 gpm of diesel is required for the agglomeration flowsheet based on a diesel consump-

tion of 0.5% of the dry-coal-feed rate. Less diesel is required for the natural fines circuit. An outside storage tank containing a 10-day supply of diesel will have an 80,000-gal capacity. The diesel feed tank (4000 gal) provides a 12-hour supply. The diesel feed tank is located inside the structure and is filled from outside storage as needed using a pump controlled by level indicators on the diesel feed tank. Diesel is delivered to the high-shear vessels through a metering pump.

Asphalt. Approximately 4 gpm of asphalt is required based on an asphalt consumption of 0.5% of the pelletizer dry-coal feed rate. An outside storage tank containing a 10-day supply of asphalt will have an 80,000-gal capacity. The asphalt inside storage tank (4000 gal) provides a 16-hour supply. This inside tank will be filled from the outside storage as needed using a pump controlled by level indicators on the inside asphalt storage tank. Asphalt is delivered to the solids mixer through a colloid mill and a metering pump. The colloid mill will disperse the asphalt in water, preparing an emulsion that is stored in the colloid tank. The colloid tank provides approximately one hour of storage capacity. From the colloid tank, the emulsion is fed to the solid mixers. The asphalt delivery system will be heated to about 70°C to reduce its viscosity, thus improving its flow properties.

Starch. Starch storage requirements are based on a utilization rate of 2% of the pelletizer feed coal on a dry basis, which is equivalent to 4 t/h. The dry pulverized starch (80% passing 200 mesh) is delivered in bulk by trucks and pneumatically transported to a silo. The silo has a one-week storage capacity, equivalent to about 700 tons. A receiving bin is located at the solids mixer and uses a weigh-belt feeder to proportion the starch with the filter cake.

Frother. Frother is added to the flotation cells to enhance recovery. The use of approximately 0.29 lb of frother per ton of solids has been assumed during each stage of flotation in the agglomeration plant. The natural fines agglomeration circuit will use approximately 0.13 lb per ton of solids.

Frother will be delivered to the plant site in concentrated form. The delivery tanker truck will pump to an 1800-gal storage tank holding approximately 10 days' inventory. The frother is then pumped to a 1000-gal tank and diluted to a 10% solution. This storage tank contains level indicators to measure water and concentrated reagent addition to the tank. The tank also has a mixer to ensure a uniform reagent concentration. A metering

pump controls the final addition of solution to the flotation cells. The maximum output of the frother metering pump will be 2.0 gal/min.

Lime. Approximately 0.5 lb of lime per ton of solids in the thickener feed is used for pH control in the water clarifier. An average solids flow rate of 25 t/h to the thickener is expected. Lime is received dry, in bags or in bulk, and stored inside the building. The lime is then slaked in a tank and added to the thickener feed as needed. The lime solution is added by means of a variable-speed pump controlled by pH and solids flow data.

Flocculant. Flocculant is stored and diluted in a manner similar to the frother. Delivery of flocculant is similar to that of the lime solution, but is controlled by the settling rate of the flocculated solids and the thickener solids feed rate.

Process Instrumentation and Control

A process control system (PCS) is recommended for the agglomeration plant to ensure smooth, continuous plant operation at near-optimal conditions. The system also ensures that all equipment is operating within specified conditions.

For the purposes of this design, a preliminary level of process control has been assumed. The commercial agglomeration plant PCS is described in detail below, including discussion of hardware, process parameters, and selection of instrumentation.

The overall process instrumentation and control for the natural fines circuit is not discussed here. The overall process control philosophy and extent of process instrumentation and control required for the circuit will be dictated by the existing process control system at the retrofit site. However, control of unit operations such as agglomeration and agglomerate dewatering for the fines circuit is discussed below.

Description

The PCS is designed to maintain a record of the state of the plant by means of instrumentation, measurements, and data acquisition. All process parameters relevant to process control and event-logging are included in the PCS.

All relevant process parameters are continuously tracked and monitored to ensure uninterrupted operation of all units. The PCS provides automated startup and shutdown control. In case of disturbances, the PCS alarms the operator. Manual action and/or computer control strategies are then implemented to bring the plant back to steady-state operation. In case of equipment failure, the plant automatically shuts down for safety purposes.

These requirements are achieved using a combination of distributed and centralized controls. Centralized control systems are designed to supervise all of the distributed control systems, implement the overall control strategy, and automate startup and shutdown procedures. Figure 4-4 illustrates the general layout of the PCS.

Figure 4-4. Process Control System Layout

The control strategy is aimed at ensuring that the quality of the product and refuse streams is within specifications. On-line measurements will be periodically compared with laboratory analyses to assure proper analyzer calibration.

The PCS is configured such that all changes in the control strategy or configuration of the unit operations can be implemented by revising the computer software. The design provides enough spares and redundancies to safeguard against failures caused by the PCS or by external factors such as power or computer failures.

Data Acquisition and Control Hardware

The data acquisition and control hardware for the PCS is capable of performing the functions shown in Figure 4-5. In addition, the PCS has other capabilities that are discussed below.

A system with physically distributed or separate hardware elements handling data acquisition and control functions for small groups of control loops is preferred for overall system reliability. The system also has a data bus which makes all data available for plant control, as well as for supervisory, data-logging, and analysis functions.

Figure 4-5. Data Acquisition and Control Architecture

A central operator's station provides facilities for process monitoring and control, programming of loop or local controllers, data trend display, data archiving, alarm log printing, and report printing.

This central control management station, or a separate host computer linked to it, performs supervisory control computations for real-time process optimization. The system also performs real-time analysis of process performance and material balance and allows data entry and reduction for integration of off-line data such as laboratory results.

Individual controllers perform data acquisition and control functions.

Process Parameters

Process parameters relevant to process control and event-logging are discussed in this section. These parameters have been selected on the basis of the conceptual design flowsheet for the commercial agglomeration plant. The discussion follows the main unit operations as indicated in the flowsheet, and is divided into seven sections based on the circuit descriptions presented earlier.

Coal Receiving, Storage, and Crushing. The objectives of the PCS in the coal receiving, storage, and crushing system (CRSCS) are to ensure an adequate supply of raw coal for continuous 24-hour operation of the agglomeration plant. The preliminary design of the CRSCS is shown in Figure 4-6. The coal receiving system operates for 8 hours per day but provides coal for 24-hour operation.

Figure 4-6. Instrumentation and Control for Receiving, Storage, and Crushing

The crushing operation is a 24-hour operation. The four vibrating feeders at the outlet of the two underground hoppers under the stacking tube regulate the coal supply to a belt conveyor that transfers the coal to a coal crusher. In the event that there is a deviation in flow rates or a change of set point, the amplitude of these feeders is adjusted to yield the desired flow rates. The coal from the crusher is transferred by a belt conveyor to a surge hopper. This hopper is equipped with a level indicator that warns of imminent interruption of the process (when empty) or overfilling of the hoppers (when full).

The function of the control system is to ensure the orderly shutdown of the CRSCS, whether routinely or in emergency situations such as overfilling of the crushed coal hopper. A mechanical sampling system is installed under the crusher to obtain coal samples periodically. The sample sulfur, ash, moisture, and Btu contents are analyzed in the laboratory to provide a history of the feed coal quality before it enters the beneficiation process.

In summary, 17 analog signals and 25 digital signals are required to monitor and control the CRSCS. The process parameters and number of signals required are summarized in Table 4-10. To determine the total num-

ber of signals required, variables that are controlled in addition to being measured are counted twice, once for measurement and once for control.

Table 4-10
Control Variable Summary
Coal Receiving, Storage, and Crushing

Variable	Signal	No.
<i>Receiving</i>		
Vibrating feeder flow rate ^c	Analog (4)	8
Vibrating feeder on/off status ^c	Digital (4)	8
Conveyor on/off status ^c	Digital	2
<i>Crushing</i>		
Vibrating feeder flow rate ^c	Analog (4)	8
Vibrating feeder on/off status ^c	Digital (4)	8
Conveyor on/off status ^c	Digital	2
Crusher power consumption	Analog	1
Crusher on/off status ^c	Digital	2
Conveyor on/off status ^c	Digital	2
Surge hopper level measurement	Digital	1
Total number of analog signals		17
Total number of digital signals		25

^cVariable is measured and controlled.

Coal Precleaning. Since the control architecture for the precleaning circuits is well-defined it is not outlined in this report; existing plant control systems will be implemented for this circuit.

Coal Grinding. The objective of the PCS in the coal-grinding circuit is to ensure the production of minus 200-mesh coal from a 1/4" x 0 feed. Since the mass flow rates of the coal and water feeding the mills affect the particle size distribution of the circuit discharge, these rates are measured and controlled. The preliminary design of the grinding operation, as shown in Figure 4-7, is an open circuit consisting of a rod mill, two ball mills, and a holding vessel.

The rod mill and ball mill product size distributions are controlled by varying the water addition rate to the rod mill and the ball charge in the ball mills. At a constant coal feed rate, as the water addition rate is increased the coal residence time decreases, resulting in coarser grind sizes. Similarly, as the water flow rate is decreased the coal residence time increases, producing finer product size distributions.

The mass flow rate to the rod mill is measured by a belt weigh feeder as this flow rate determines the rate of water addition to the rod mill to produce a discharge of a specified particle size. An automatic sampling system is

installed to obtain periodic coal samples in order to check the particle size distribution in the laboratory and take appropriate control actions. The holding vessel requires a level indicator to alarm overflow and empty tank conditions.

Figure 4-7. Instrumentation and Control for Coal Grinding

In summary, 8 analog signals and 22 digital signals are required for control of the coal-grinding circuit based on the preliminary design. These signals are summarized in Table 4-11.

Table 4-11
Control Variable Summary
Coal Grinding System

Variable	Signal	No.
<i>Rod Mill</i>		
Precleaned coal feed rate ^c	Analog	2
Belt feeder on/off status ^c	Digital	2
Rate of water addition ^c	Analog	2
Rod mill on/off status ^c	Digital	2
Rod mill sump level indicators	Digital	1
Rod mill pump on/off status ^c	Digital	2
<i>Ball Mill</i>		
Ball mill mass flow rate ^c	Analog(2)	4
Ball mill on/off status ^c	Digital(2)	4
Ball mill sump level indicators	Digital(2)	2
Ball mill pump on/off status ^c	Digital(2)	4
<i>Holding Vessel</i>		
Holding vessel level measurement	Digital	1
Holding vessel stirrer on/off status ^c	Digital	2
Holding vessel pump on/off status ^c	Digital	2
Total number of analog signals		8
Total number of digital signals		22

^cVariable is measured and controlled.

Coal Agglomeration. The objectives of the PCS in the coal agglomeration circuit are to control the addition of diesel into the process at specified amounts and control the production of recoverable microagglomerates to meet specifications. The preliminary design of the coal agglomeration circuit, shown in Figure 4-8, begins with a high-shear vessel followed by two stages of flotation.

Figure 4-8. Instrumentation and Control for Coal Agglomeration

Agitation intensity, residence time, diesel dosage, and pulp density are important parameters that affect high-shear performance. Slurry flowmeters and nuclear density gauges are required to monitor the mass flow rates and percent solids of the slurry feeding each high-shear vessel. The diesel dosage is a function of the coal feed rate, and water addition is a function of the pulp density. Control valves are used to regulate diesel and slurry distribution to the high-shear vessels. The high-shear agitation intensity is also controlled. The high-shear vessels are equipped with level indicators to alarm empty conditions.

Aeration, pulp level, and froth level are important parameters that affect flotation performance. The yield from the flotation cells is controlled by controlling the aeration rate and froth level. The flow rate to the flotation stages is measured to determine the amount of water required to maintain the desired solids concentration in the flotation banks. The flotation solids concentration is adjusted to control the flotation residence time in order to obtain the desired froth yield that meets specifications.

Automatic sampling systems are installed at various points in the agglomeration circuit to obtain periodic slurry and tailings samples for laboratory analysis to evaluate the performance of the PCS.

In summary, the preliminary design of the coal agglomeration circuit requires a total of 68 analog signals and 39 digital signals. These signals are summarized in Table 4-12.

Table 4-12
Control Variable Summary
Coal Agglomeration System

Variable	Signal	No.
High-Shear Vessels		
Slurry flow rate to sump ^c	Analog	2
Rate of water addition ^c	Analog	2
High-shear sump level indicator	Digital	1
High-shear pump on/off status ^c	Digital	2
Slurry flow rates	Analog (6)	6
Density measurement	Analog (6)	6
Rate of oil addition ^c	Analog (6)	12
Stirrer speed ^c	Analog (6)	12
High-shear vessel stirrer on/off status ^c	Digital (6)	12
High-shear vessel level measurement	Digital (6)	6
Flotation Stages		
Rate of water addition ^c	Analog (4)	8
Flotation cell impeller on/off status ^c	Digital (4)	8
Flotation cell level measurement	Digital (4)	4
Flotation cell level control ^c	Analog (4)	8
Rate of aeration ^c	Analog (4)	8
Flotation sump level indicator	Digital (2)	2
Flotation pump on/off status ^c	Digital (2)	4
Second-stage flow rate ^c	Analog (2)	4
Total number of analog signals		68
Total number of digital signals		39

^cVariable is measured and controlled.

Agglomerate Dewatering. The objectives of the PCS in the agglomerate dewatering circuit are to ensure continuous operation of the vacuum filters and to produce filter cake at a specified moisture. The preliminary design of the agglomerate dewatering system, as shown in Figure 4-9, consists of two conditioning tanks in parallel followed by two vacuum filters also in parallel.

The froth flow to the conditioning tanks is controlled by a flow control valve. The tank is equipped with a level indicator to alarm overflow and empty tank conditions. The flow to the vacuum filter is controlled by a flow meter. The moisture content in the filter cake is controlled by varying the vacuum filter disc speed.

In summary, the preliminary design of the agglomeration dewatering circuit requires a total of 14 analog signals

and 20 digital signals. These signals are summarized in Table 4-13.

Figure 4-9. Instrumentation and Control for Agglomerate Dewatering

Table 4-13
Control Variable Summary
Agglomerate Dewatering System

Variable	Signal	No.
<i>Conditioning Tank</i>		
Slurry flow rates	Analog (2)	2
Conditioning tank stirrer on/off status ^c	Digital (2)	4
Conditioning tank level indicator	Digital (2)	2
<i>Vacuum Filter</i>		
Slurry feed pump on/off status ^c	Digital (2)	4
Slurry flow control ^c	Analog (2)	4
Vacuum filter on/off status ^c	Digital (2)	4
Vacuum filter rpm control ^c	Analog (2)	4
Vacuum filter level indicator	Digital (2)	2
Vacuum pump on/off status ^c	Digital (2)	4
Filtrate flow rate ^c	Analog (2)	4
Total number of analog signals		14
Total number of digital signals		20

^cVariable is measured and controlled.

Size Enlargement, Drying, and Product Loadout. The purpose of the PCS system in the size enlargement, drying, and product loadout circuit is to regulate binder addition to the solids mixers and control the size and quality of the clean coal pellets. The size enlargement, drying, and product loadout control system is shown in Figure 4-10.

Since the vacuum filters discharge directly into the solids mixers, it is difficult to monitor the solids mixer coal feed rate which is required to monitor the flows of both asphalt and starch to the solids mixers. Instead, the pelletizer

coal output (which is equal to the solid mixers coal feed rate) is monitored to meter starch and asphalt requirements.

Figure 4-10. Instrumentation and Control for Size Enlargement, Drying, and Product Loadout

A complete dryer package with instrumentation and process control equipment is anticipated for pellet drying. The drying system will have the following features:

- Automatic controls that are self-compensating for variations in the feed rate, feed moisture, and residual product moisture
- Measurement and control of oxygen content in the drying chamber
- Self-contained programmable controller that can communicate with the control computer management station
- Local free-standing control panel and central control room, information display, and set point adjustment
- Production and regulation of flue gas from natural gas burners

The feed rate to and from the dryer is monitored. Samples are taken from the product belt for elemental analysis. The silos are equipped with level indicators to alarm overflow and empty conditions.

In summary, the preliminary design of the size enlargement, drying, and product loadout circuit requires a total of 37 analog signals and 75 digital signals. These signals are summarized in Table 4-14.

Table 4-14
Control Variable Summary
Size Enlargement, Drying and Product Loadout

Variable	Signal	No.
Solids Mixer		
Asphalt flow rates ^c	Analog (4)	8
Starch flow rates ^c	Analog (4)	8
Solid mixer stirrer on/off status ^c	Digital (4)	8
Solid mixer rpm control ^c	Digital (4)	8
Pelletizer		
Pelletizer on/off status ^c	Digital (4)	8
Pelletizer rpm control ^c	Digital (4)	8
Dryer		
Dryer feed hopper level indicator	Digital (4)	4
Dryer feeder flow rate ^c	Analog (4)	8
Dryer feeder on/off status ^c	Digital (4)	8
Dryer on/off status ^c	Digital (4)	8
Dryer temperature	Analog (4)	4
Dryer oxygen level	Analog (4)	4
Dryer flue gas flow rate ^c	Digital (4)	8
Dryer rpm control ^c	Digital (4)	8
Product Loadout		
Product belt conveyor on/off status ^c	Digital	2
Sulfur, ash, Btu, and moisture analyzers	Digital (4)	4
Sulfur, ash, Btu, and moisture analyzers	Analog (4)	4
Mass flow rate	Analog	1
Silo level indicator	Digital	1
Total number of analog signals		37
Total number of digital signals		75

^cVariable is measured and controlled.

Water Clarification and Refuse Handling. The PCS in the water clarification and refuse handling system regulates the processing of tailings rejected by the agglomeration process so as to form a handleable refuse for disposal and recover and recycle the process water. The system also controls the treatment of recycled water to maintain it within permissible levels of contamination and pH. The preliminary design of the refuse handling control system is shown in Figure 4-11.

The thickener is provided with torque instrumentation and a lifting device on the rake drive to protect the mechanism from damage. The settling rate of the feed slurry is determined by a settling rate meter and flocculant addition is adjusted accordingly. The overflow pH is monitored and controlled to aid in flocculation. The measurements that are used in automating the thickener are the feed pulp mass flow, solids settling time, overflow turbidity, and solids level. Some of the water recovered from the thickener overflow tank is then processed in the water treatment unit.

Figure 4-11. Instrumentation and Control for Water Clarification and Refuse Handling

The water treatment unit requires eight analog signals to control the flow of input water and four digital signals to switch activated-carbon beds when the quality of the water is not within specifications. The water quality is determined by taking samples from the output stream of the unit and measuring the total dissolved solids in the laboratory. The output stream of treated water is then pumped back to the head tank for recycling.

The preliminary design of the water clarification and refuse handling system requires 26 analog signals and 14 digital signals. These signals are listed in Table 4-15.

Selection of Instrumentation

Based on the analysis of the process parameters presented earlier, a wide array of instrument types using various measurement techniques was investigated for each parameter to be measured and/or controlled. Selection of instruments for the detailed design of the process control system will require further study. However, in this section, a preliminary selection of instrumentation has been made for each process parameter. Areas of uncertainty involved in this selection are also discussed.

General Items. A partial list of instrumentation selected for some of the general parameters is given in Table 4-16. An uncertainty in terms of instrument selection affects the choice of an ultrasonic level gauge for all level measurements. Pressure-sensitive load cell sensors may be adequate for level measurement in the crushed coal bin and a differential-pressure meter may be adequate for level measurements in all vessels holding slurry.

Table 4-15
Control Variable Summary
Water Clarification and Refuse Handling

Variable	Signal	No.
<i>Water Clarifier</i>		
Tails flow rate	Analog	1
Flow rates of reagents and flocculants ^c	Analog (3)	6
Belt press effluent flow rate	Analog	1
Rake height	Analog	1
Rake torque	Analog	1
Rake on/off status ^c	Digital	2
Clarifier solids level	Analog	1
Settling rate	Analog	1
Underflow pump on/off status ^c	Digital	2
Underflow flow rate ^c	Analog	2
Underflow solids concentration	Analog	1
Overflow pump on/off status ^c	Digital	2
Overflow flow rate	Analog	1
Overflow turbidity	Analog	1
Overflow pH	Analog	1
<i>Water Treatment</i>		
Adsorption bed on/off status ^c	Digital (2)	4
Flow control of input water ^c	Analog (4)	8
<i>Belt Press</i>		
Belt press on/off status ^c	Digital	2
Refuse belt conveyor on/off status ^c	Digital	2
Total number of analog signals		26
Total number of digital signals		14

^cVariable is measured and controlled.

Table 4-16
List of Instruments

Parameter	Recommended Instrument
On/off status	Two-position controller
Power consumption	Wattmeter (ammeter)
Level measurement	Ultrasonic gauge
Mixer speed	Tachometer
Rate of oil/reagent addition	Positive displacement metering pump
Rate of water addition	Flow control valve and flowmeters
pH	pH electrode
Temperature	Thermocouple
Pressure	Pressure gauge (Bourdon, bellow)
Slurry density	Nuclear density gauge
Flow control of steam	Control valve
Liquid flow rate	Ultrasonic flowmeter
Solids flow rate	Belt conveyor scale

Mass Flow Rate. It is recommended that two solids flow meters be used for the grinding circuit. This type of meter applies the principle of force measurement without using an impact plate by allowing the material to slide down onto the curved measuring chute; thus, the effects of free-fall heights, impact shock, and friction forces are eliminated. The reported accuracy is $\pm 2\%$.

Vibratory Feeder Flow Rates. Vibratory feeders are typically driven using a fixed amplitude and variable frequency or a fixed frequency and variable amplitude. Fixed-amplitude feeders are generally mechanically driven systems with a variable-speed motor, while fixed-frequency feeders use an alternating magnetic field to generate the vibration. Fixed-frequency feeders have the advantage of not using any mechanical linkages to convey motion and can be tuned to operate near their harmonic frequencies, thus increasing their efficiencies. Therefore, a fixed-frequency feeder is recommended in the coal handling system.

On-Line Elemental Analyzer. This application requires detailed analysis of the coal, including its sulfur content and heating value. An elemental analyzer, based on prompt gamma neutron activation analysis (PGNAA), is the only coal quality instrument currently available that has the capability of performing the required analyses on-line. At least two elemental analyzers currently on the market, the MDH Elan and the Gamma-Metrics 1812C, can sample the entire 200-t/h product stream without requiring a sampling system.

One of the main areas of uncertainty in the use of these analyzers is whether they will be capable of handling the agglomerated coal without plugging. Both analyzers are constructed around a vertical chute filled with coal to be analyzed. If the agglomerates being analyzed are sticky, they may plug the chute. In addition, a unique calibration curve may be required for the analyzer to give accurate readings in the presence of the oils used for agglomeration. Different calibration curves may also be required for each coal.

A microwave moisture meter will be used for on-line moisture measurement. The effects of oil in the sample path are uncertain. The effects of moisture trapped within the agglomerates are also uncertain. It is expected that the microwave moisture meter will require calibration for each of the three coals due to their differing inherent moisture content.

Settling Rate Meter. An on-line settleometer would be very useful for minimizing the use of chemical flocculants.

This device would measure the effect of adding flocculants to the slurry. One instrument for this application is currently on the market; however, it has high capital and operating costs. This settling rate meter consists of a 6-ft-tall settling cylinder with automatic photo cells tracking the liquid-solid interface on a discrete batch basis.

Turbidity Detector. An on-line overflow turbidity monitor can be used to determine the turbidity of the clarified water. This monitor uses visible and near-infrared light to determine turbidity levels. The output from the turbidity gauge is connected to an audible alarm to signal potential problems.

Ancillary Requirements

The ancillary requirements for the commercial agglomeration plant are identified below. These include a definition of the battery limits and identification of utility requirements, traffic needs, emissions, and refuse disposal needs. The ancillary requirements for the existing conventional coal cleaning plant with retrofit agglomeration circuit are not addressed in this discussion.

Battery Limits

The commercial agglomeration plant is assumed to be located such that all utility requirements are available at the plant boundary. This assumption has a significant impact on the capital cost estimate. The plant is assumed to be fairly close to sources of power, water, and gas that are supplied by local utility companies at the required voltage, flow rate, and pressure, respectively.

Power

Power is drawn from an existing facility to a substation housed inside the complex from where it is distributed to various areas of the plant at a suitable voltage. Before delivery to the plant, a voltage stepdown from the main supply is implemented by means of a transformer. The kVA rating is calculated from the installed horsepower of the equipment, the lighting load, and other miscellaneous requirements. Power requirements for specific equipment items are detailed in Appendix G.

Water

Water requirements can be classified as:

- Process water

- Potable water
- Firefighting requirements

It is assumed that water is available from a nearby location. Pipelines may have to be installed to draw the water from the source; however, the cost of developing the water resources is not included in the cost estimate. It is estimated that two pumps are used to draw process water requirements. A complete, standalone fire protection piping and pump system is also assumed.

Process Water. Process water is stored in a storage pond with a 675,000-gallon capacity, providing sufficient process make-up water for one day, which amounts to about 550,000 gallons, including losses. The remaining water is stored in the pond for emergency and firefighting requirements. It is estimated that three to four hours of pumping may be required per day to fill the storage pond. Two pumps are allocated to draw water from the outside water supply.

Potable Water. The daily requirement for potable water is estimated to be 2000 gallons per day, assuming 30 gal/day of consumption per person. A storage tank with a 5000-gallon capacity is included to supply this water. Potable water disposal is connected to sewage lines.

Firefighting Requirements. Emergency firefighting water requirements of 125,000 gallons have been allocated. This water is drawn from the process water storage pond and is equivalent to two hours of demand at a pumping rate of 1500 gal/min. Two or three pumps are allocated for firefighting requirements.

Natural Gas

Natural gas requirements for the dryer burners in the pelletization circuit are estimated to be 211,000 ft³/h (heating value of 1000 Btu/ft³). Natural gas will be supplied by a local utility. Provisions will also be made to store a one-day supply of gas. The cost of the burners is included as part of the drying system.

Compressed Air

Compressed air is available in the plant for use in various pneumatic control valves and other pneumatically operated equipment and parts. An air compressor and receiver is installed to generate the compressed air. The compressed air delivery pressure is assumed to be 80-90 psig. The distribution system is designed to handle 20 ft³/min of compressed air.

Traffic Needs

In addressing traffic needs for the commercial agglomeration plant, the major considerations are the coal supply, product shipping, coal handling equipment, plant maintenance equipment, incoming deliveries, transportation of plant operating personnel, and parking.

It is assumed that conventional haulage is used for coal supply and product shipping and that gravel roads are sufficient to support heavy vehicles and handling equipment. The plant layout will dictate overall traffic patterns; however, heavy equipment movement for coal supply and handling will be routed on a separate road from incoming deliveries so as not to impede the flow of traffic. The needs of the plant operating personnel are also addressed in terms of providing for adequate parking lots. Safety and other instructional signs are displayed on all roads to alert plant personnel, subcontract staff, suppliers, and visitors to traffic regulations.

Water Quality

Plant water management is important for many reasons. Environmental damage may result from poor water management practices. Examples of possible environmental damage include modification of local drainage patterns, silting of nearby lakes and streams, contamination of surface and ground water, and disruption of aquatic habitats.

A number of liquid effluents may be generated during plant operation. These include:

- Plant site sewage
- Boiler plant blowdown
- Spent solution from ion exchange or carbon bed regeneration
- Filter backwash from drinking water treatment
- Oily waste water from maintenance shops
- Fuel oil storage area drainage
- Laboratory waste water

Process water treatment facilities will treat some of the waste water, e.g., boiler plant blowdown and regeneration liquid from spent beds. In addition, treatment facilities to handle surface runoff wastes are included.

Surface Runoff

Surface runoff and underground plant drainage are two major drainage sources anticipated at the plant site. These streams are treated to meet effluent quality standards or for possible reuse in plant operations.

Surface runoff is the residual portion of rainfall not absorbed in the ground and filtered down to the water table. It may be categorized as the overflow after all losses have been satisfied. These losses are:

- Interception storage, i.e., interception by vegetation to be returned via evaporation
- Depression storage or ponding
- Infiltration or movement of water through soil zones
- Evaporation from the surface

Surface runoff is a function of the intensity, duration, and distribution of precipitation, the permeability of the soil, surface temperature, and the slope of the land surface. It is especially important to control surface runoff in the presence of oily water on the ground or on open stockpiles and noticeable gradations of the ground. The extent of surface runoff may only be estimated at this time since the plant site is unknown.

The EPA has recently issued a rule concerning surface runoff that is expected to affect permitting requirements for the agglomeration plant. The new permitting mechanism for industrial stormwater regulation is known as a General Permit. Under the plan, an initial permit application is filed and a Stormwater Pollution Prevention Plan (SWPPP) is prepared. It is anticipated that stormwater monitoring will include flow rate measurement and analysis for pH, total dissolved solids, iron, magnesium, oil and grease, chemical oxygen demand, and biochemical oxygen demand.

As a conservative estimate, an average annual rainfall of 40 inches or approximately 1000 mm is assumed. This would lead to a surface runoff of approximately 82,000 m³ per year (assuming 10% runoff and a plant area of about 820,000 m²). More important, however, the runoff control scheme should be based on peak rainfall times. A storm with 100 mm/h rainfall would generate about 8,200 m³ of surface runoff per hour (assuming 10% runoff). It is necessary to control such large amounts of water, especially during severe storms and periods of heavy rainfall. Surface runoff can be effectively managed using a network of dikes, channels, and ponds.

Sedimentation ponds need to be constructed to take care of surface runoff water. Settling ponds used for process waste water treatment can also be utilized for this purpose; however, they will require excess capacity to handle peak loads of surface runoff volume.

The land at the plant site should be graded such that surface runoff water can be easily directed and conveyed to sedimentation ponds through dikes and channels.

Hydrocarbon and Oil Runoff

Although the likelihood of oil and hydrocarbon runoff at the plant site is minimal, the possibility must be taken into consideration. Hydrocarbon runoff may be generated from leakage and spillage from machinery, and leakage from vehicles at the plant site. No runoff is expected from hydrocarbon storage tanks since the tanks will be surrounded by concrete dikes for spill containment.

There is also a possibility of leakage from warehouse storage of reagents and chemicals. Such leakage, however, is not anticipated to be major and should not be a problem of daily occurrence. In the event of a major spill or leak, remedial action will be taken to clean up the soil and/or water. It is difficult to quantify the amount of oil and hydrocarbon runoff that could occur at the plant at this stage.

Air Quality

The commercial agglomeration plant is not anticipated to experience any serious air pollution problems in the absence of coal combustion. The major sources of air pollution are identified as:

- Dust emissions from the raw coal live storage stockpile
- Hydrocarbon vapors from product silos
- Flue gas from dryers
- Minor hydrocarbon pollution from fuel oil storage tanks and internal combustion engines
- Dust emissions from fine coal storage silos
- Dust emissions from crusher operation
- Fugitive dust from drying operations
- Dust emissions from coal haulage roads

Precise quantification of possible air pollution from the various sources is difficult at this time. Detailed emissions and weather data, etc., are needed to quantify emissions. An emissions inventory can be taken by enumerating emission factors for the different sources. An emission factor is a statistical average of the rate of contaminant release from a particular operation. This inventory is an estimate of the amounts of contaminants released from all sources in a given time period and enables plant design personnel to assess the degree of overall control necessary at the site to meet federal and state air quality standards.

Dust emissions resulting from open stockpiling of raw coal, crusher operation, and fine coal storage silos constitute the major emissions sources of dust in the plant. Unconfined emissions of dust may occur when the open stockpiles are exposed to air and subjected to strong winds. The dust produced from such stockpiles is a result of various activities such as:

- Loading of material on the pile
- Removal of material from the pile
- Wind erosion
- Movement of vehicles in the storage area

The amount of dust emissions from open stockpiles depends on the volume of material being handled and its moisture content, the fines preparation method, wind speed and direction, degree of compaction of the pile, and age of the material in the stockpile.

Dust emissions from fine-coal storage silos may also be significant, and are affected by such factors as the proportion of fines, their moisture content, and wind speed and direction. Solid waste disposal and topsoil management also generate a great deal of dust. Other sources of dust emission include the transfer points of conveyors. Wind erosion of unclaimed land is a major source of dust emissions.

Evaporation of hydrocarbons from the product during storage is a potential source of air pollution. The degree of emissions is unknown at this time, but is estimated to be negligible due to the low consumption of diesel in the process.

Flue gas from dryers also contributes to atmospheric pollution. Emissions from gas-fired burners are minimal and require no downstream installations like scrubbers.

Emissions Control

Particulate matter and dust (fine particles) are the major targets for emissions control at the plant site. Dust emissions from open stockpiles and unreclaimed ground can be mitigated or controlled by one or more of the following steps:

- Spraying of water
- Use of wetting agents
- Chemical stabilization
- Physical stabilization
- Vegetative stabilization
- Wind speed reduction

The most common and least expensive of these methods is wet suppression with water. The drawback of this method is that its effect is temporary and it is required at least two or three times a day. Chemicals may be used as wetting agents or to form a protective crust to shield surface particles from the wind but may be expensive. Use of chemicals carries the potential hazard of contaminating the coal or the surface or ground water through seepage. Planting of vegetative covers can be effective for dust control on barren land. Physical stabilization or compaction may be used for open coal stockpiles. Windbreaks may be impractical in the plant due to cost considerations and doubts about their effectiveness.

The final choice of the emissions control mechanism depends on the location of the plant and the prevailing weather conditions. It can be assumed that wet suppression of dust with water along with some wetting agents will be used for controlling dust emissions from open stockpiles.

Refuse Disposal

The requirements and methods of disposing of process-generated refuse were discussed previously in this section, under the subheading "Unit Operations." Dewatered refuse from the plant will be trucked to an appropriate landfill site. The disposal requirements for refuse generated by non-process activities need to be identified. The majority of this refuse is generated by regeneration of carbon beds, sludge from potable water filtration, surface runoff sludge, and oily waste water from analytical laboratories.

The composition and rate of generation of these wastes varies. The major components of these wastes are a combination of organic and inorganic compounds. The varying nature of the waste materials may warrant separate treatment for groups of compounds. Settling ponds may be used to separate sludge from waste water. Ponds, in addition to those used for process waste water, may have to be constructed. Organic sludges, if high in volatiles, may be suitable for microbial treatment to convert contaminants into CO_2 and H_2O . Other sludges may either be stabilized at the bottom of the pond or recovered, dewatered, and sold to outside contractors, depending on their composition and value.

Techniques to Maintain 90% Plant Availability (Commercial Plant)

The goal for the plant is to process at least 90% of the design process rate which is 200 t/h of product on a dry basis. The purpose of the reliability analysis is to model the plant design using component failure rate and repair time data to estimate the achievable coal processing rate. The modeling tool used is an EPRI computer program, UNIRAM.

Plant Description and Assumptions

The plant is scheduled to operate 24 hours per day, 7 days per week and 50 weeks per year. It is assumed that adequate spare parts will be kept on hand. Maintenance personnel will be available to perform corrective maintenance promptly during scheduled service hours. During the two annual one-week overhauls, major maintenance including preventive maintenance will be performed. Unforeseen major catastrophes have not been considered in the analysis. However, equipment failures such as shaft breaks and motor burnouts are included in the failure and repair data.

In the plant, the coal is crushed and precleaned using conventional coal cleaning methods. Water is added to the precleaned coal and it is ground to a fine size. Oil and reagents are added to the fine coal slurry in the agglomeration process, which utilizes conventional flotation for coal recovery. Starch and asphalt are then added during the pelletization process which is followed by thermal drying.

The conventional precleaning circuits will operate as an integral part of the selective agglomeration plant. Therefore, these circuits will not be bypassed. Problems that render the conventional precleaning circuits unavail-

able will cause the plant to be unavailable. The general assumptions regarding spare parts and operation and maintenance schedules that apply to the agglomeration advanced cleaning circuit also apply to the conventional precleaning circuits.

The concentrating spirals are sorting chutes with no moving parts. If a concentrating spiral component were to fail (usually due to worn liners), its corresponding classifying cyclone would be shut off by means of a valve. Therefore, each set of three parallel spirals (which correspond to one classifying cyclone) was treated as one component with triple the failure rate of one spiral. In actual practice, spiral liner wear rates are monitored and liners are refurbished during scheduled plant overhauls when wear rates indicate maintenance is needed.

No quality restrictions have been placed on the agglomerate product. The results reflect the estimated amount of coal that can be processed. The percentage of the coal that becomes an acceptable product from a quality standpoint is beyond the scope of this analysis.

The availability block diagram (ABD) model, Figure 4-12, shows the general layout of the selective agglomeration process. The ABD does not necessarily represent the physical and functional relationships of the components in the plant; however, it does represent the relationship of the components with respect to plant availability. The capacities shown above each block on the ABD indicate the maximum plant capacity (in percent) that can be supported by each component. The circles with an "X" indicate an origination point outside the boundaries of the model. Triangles indicate transfer points between process trains.

Data Sources and Assumptions

The sources for much of the data used in this analysis are published Electric Power Research Institute (EPRI) reports containing failure rates and repair times for components similar to those used in the plant. Some data were supplied by Praxis Engineers, Inc. and Southern Company Services, Inc.

Some components are expected to have very high reliability and can be bypassed or run manually in emergencies. These components (listed below) were either not modeled, bypassed, or assigned "perfect" reliability. These components will have an insignificant impact on plant availability.

Coal unloading feed hoppers
Reagent system
Tramp iron magnet
Feed coal stacking tube
Refuse storage area
Slurry splitter
Truck scale and weigh station
Truck dump grates
Piping
Product silo
Distributors
Tanks
Colloid mill
Starch storage and addition system
Activated carbon bed
Wet box

The following parallel-configured components listed in Table 4-17 have a capacity in excess of the nominal capacity shown on the availability block diagram. If one or more of these parallel components fail, the parallel train can operate at the capacities listed below.

Table 4-17
Excess Capacity

	1 fail	2 fail	3 fail
First-stage flotation cells (10% nominal)	95%	90%	85%
Second-stage flotation cells (12.5% nominal)	90%	80%	70%
Pelletizer and dryer train (25% nominal)	85%	60%	30%

Note: The pelletizer train can run in the excess capacity state for only an 8-hour period per failure. After eight hours, product quality would degrade. Only eight hours credit was given for this condition.

Surge Capacity. Some components in the plant are storage points that provide hours of surge time or buffer time during which certain components upstream (or downstream) of the storage point can fail and be repaired without shutting the plant down. The method used to give credit for this storage is a function of the storage time and mean down time of the upstream components. The storage times are listed in Table 4-18. The mean time between failure (MTBF) is adjusted in the model to reflect these storage times.

Table 4-19 is a list of plant components and their associated MTBFs, failure rates, mean down times, and reference sources. The mean down times represent not only the amount of time it takes to repair the equipment but also the logistics time involved in shutting down and restarting the plant or equipment train.

Figure 4-12. Availability Block Diagram for Commercial Agglomeration Plant

Figure 4-12. Availability Block Diagram for Commercial Agglomeration Plant (cont.)

Figure 4-12. Availability Block Diagram for Commercial Agglomeration Plant (cont.)

Table 4-18
Storage Times

Refuse thickener	4 hours
Colloid tank	1 hour
Diesel storage tank	10 days
Diesel feed tank	12 hours
Asphalt storage tank	10 days
Asphalt feed tank	16 hours
Magnetite thickener	3 hours
Static thickener	3 hours

- Availability Factor (AF)
- Equivalent Availability Factor (EAF)
- Full Load Operation Factor (FLOF)

For purposes of this evaluation, the terms above apply only to scheduled operating times. Therefore, they should not be compared to standard North American Electric Reliability Council/Generating Availability Data System factors for other plants.

Results

Three factors can be used to express the reliability of the selective agglomeration plant.

The AF is the expected percent of scheduled service hours that the plant will actually be in service. The plant is scheduled to operate 8424 hours per year. The AF is

Table 4-19
Component Failure Rate and Repair Time Data: Selective Agglomeration Plant

Component	MTBF (Service Hours)	Failure Rate (Failures per yr)**	Mean Down Time (Hours)	Source*
Conveyor	17,520	0.5	18.0	5276
Rod mill feed hopper	91,104	0.1	11.0	5276
Ball mill	10,000	0.8	40.0	SCS
Slurry sump	47,304	0.2	7.5	5276
Slurry pump	8,760	1.0	40.0	5276/SCS
Flotation cell	8,424	1.0	8.0	SCS
Conditional mix tank	17,520	0.5	4.0	SCS
Pumps (transfer, feed, metering, or recycle)	33,376	0.3	8.0	2205/SCS
High-shear mixer	8,000	1.1	15.3	2205/SCS
Screen	17,520	0.5	18.1	2205
Tramp iron magnet	10,000	0.8	4.0	SCS
Coal crusher	8,424	1.0	4.0	SCS
Weigh-belt feeder	17,520	0.5	7.0	5276
Rod mill	7,000	1.2	40.0	SCS
Vibrating feeder	17,520	0.5	7.0	5276
Refuse thickener	35,040	0.2	83.0	PRAX
Water pump	87,600	0.1	40.0	PRAX/SCS
Storage hopper	7,513	1.1	4.4	MWK
Rotary disc vacuum filter	5,896	1.4	24.0	PRAX
Solids mixer (Turbulator™)	17,520	0.5	15.3	2205
Disc pelletizer	10,000	0.8	24.0	PRAX
Vibrating fluidized-bed dryer	5,000	1.7	24.0	PRAX
Belt filter	17,520	0.5	6.0	PRAX
Static thickener	17,520	0.5	24.0	PRAX
Magnetite thickener	17,520	0.5	24.0	PRAX
Magnetic separator	10,000	0.8	4.0	SCS
Classifying cyclones	13,140	0.7	4.0	PRAX
Concentrating spirals	17,520	0.5	8.0	PRAX
Heavy-media cyclones	8,760	1.0	4.0	PRAX

* 5276 Availability Analysis of an IGCC, EPRI AP-5276, June 1987

PRAX Praxis Engineers, Inc.

MWK M. W. Kellogg (1990 IGCC Studies)

2205 Component Failure and Repair Data: GCC Units, EPRI AP-2205, February 1982

SCS Southern Company Services, Inc.

** Based on 8424 service hours per year

97.19%. Therefore, the expected service hours per year are (8424 hours) (0.9719) = 8187 service hours per year.

The EAF is the average expected coal processing rate. The plant is designed to process 200 t/h of product. The EAF is 95.16%. Therefore, the average expected coal processing rate is (200 t/h product) (0.9516) = 190 t/h.

The FLOF is the percentage of scheduled service hours during which the plant is expected to be running at full load (200 t/h). The FLOF is 88.99%. With 8424 scheduled service hours the expected number of service hours at full load is (8424 hours) (0.8899) = 7497 hours per year.

Critical Components. The critical components are those components that cause the most lost production (equivalent coal processing time). The two ball mills (in series) account for 19% of lost production (9.5% each). The single most critical component in the plant is the rod mill which accounts for about 14% of lost production. The other components which cause at least 10% of lost production are the four fluidized-bed dryers (11%) and the two rotary disk vacuum filters (10%).

Table 4-20 shows the critical component ranking list. The ranking factor is interpreted to mean that if that component could be made perfectly reliable, the plant would increase its equivalent availability factor approximately by the number of percentage points equal to the ranking factor.

Table 4-20
Critical Component Ranking List

Component	Ranking Factor
Rod mill	0.55
Fluidized-bed dryers	0.44
Ball mills	0.38 (each)
Rotary disk vacuum filters	0.38

Summary

Given the plant configuration, component failure and repair data and assumptions used in this analysis, the commercial-scale selective agglomeration plant will meet or exceed its performance goal of achieving 90% of its design process rate (200 t/h product).

The EAF for the plant design is 95.16%. With plant EAF well above the design goal, no further improvements are necessary to the design to increase EAF further. This reliability model and the UNIRAM computer program may

be used to study the impact of trading off parallel-configured components with spares and maintenance philosophy.

Techniques to Maintain 90% Plant Availability (Natural Fines)

The purpose of the reliability analysis is to model the design of the natural fines circuit using component failure rate and repair time data to estimate the achievable coal processing rate. As in the previous case, the modeling tool used is an EPRI computer program, UNIRAM.

Plant Description and Assumptions

The conventional coal cleaning system will have 1500 t/h of coal delivered to it for processing. It was assumed that the conventional coal cleaning system has a mechanical operating equivalent availability of 90%. The conventional coal cleaning system can produce 1203 t/h product. The 90% operating availability assumption includes the effects of coal receiving, storage, and crushing but does not include the effects of water treatment. In addition to product, 114 t/h of fines are assumed to be fed to the agglomeration circuit for further processing. Without any reliability problems, the agglomeration circuit would produce 70 t/h of product.

It is assumed that adequate spare parts will be kept on hand. Maintenance personnel will be available to perform corrective maintenance promptly during scheduled service hours. Unforeseen major catastrophes have not been considered in the analysis. However, equipment failures such as shaft breaks and motor burnouts are included in the failure and repair data.

The natural fines circuit design is based on a typical conventional coal cleaning plant operating practice of 5400 operating hours per year.

No quality restrictions have been placed on the agglomerate product. The results reflect the estimated amount of coal that can be processed. The percentage of the processed coal that becomes an acceptable product from a quality standpoint is beyond the scope of this analysis.

The ABD model, Figure 4-13, shows the general layout of the natural fines agglomeration plant. The ABD does not necessarily represent the physical and functional relationships of the components in the plant; however, it does represent the relationship of the components with respect to plant availability. The capacities shown above

Figure 4-13. Availability Block Diagram for Natural Fines Circuit

each block on the ABDs indicate the maximum plant capacity (in percent) that can be supported by each component. The circles marked with an "X" indicate an origination point outside the boundaries of the model. Triangles indicate transfer points between process trains.

Data Sources and Assumptions

Table 4-21 is a list of plant components and their associated MTBFs, failure rates, mean down times, and reference sources used in the availability analysis for the conventional cleaning plant with the retrofit agglomeration circuit. The following components were either not modeled, bypassed, or assigned "perfect" reliability and will have an insignificant impact on plant availability.

Reagent system
Refuse storage area
Piping
Product silo
Distributor
Tanks
Activated-carbon bed

The flotation cells have a capacity in excess of the nominal capacity shown on the availability block diagram. If one or more of these parallel components fail, the

parallel train can operate at 90, 80, and 70% capacity for 1, 2, and 3 failures, respectively.

Surge Capacity. Storage times used to adjust the MTBFs shown in Table 4-21 are listed in Table 4-22.

Table 4-22
Storage Times

Refuse thickener	4 hours
Diesel storage tank	10 days
Diesel feed tank	12 hours

Results

The EAF is used to express the coal processing rates of the natural fines agglomeration plant system. As stated in the previous analysis, the EAF applies only to scheduled operating times. Therefore, it should not be compared to standard North American Electric Reliability Council/Generating Availability Data System factors for other plants.

The EAF for conventional coal cleaning was assumed to be 90%. Based on this assumption and the results of the agglomeration model, the total plant results are presented in Table 4-23.

Table 4-21
Component Failure Rate and Repair Time Data: Natural Fines Circuit

Component	MTBF (Service Hours)	Failure Rate (Failures per yr)**	Mean Down Time (Hours)	Source*
Conveyor	17,520	0.3	18.0	5276
Slurry sump	47,304	0.1	7.5	5276
Slurry pump	8,760	0.6	40.0	5276/SCS
Flotation cell	8,424	0.6	8.0	SCS
Conditioning mix tank	17,520	0.3	4.0	SCS
Transfer, feed, metering, or recycle pumps	33,376	0.2	8.0	2205/SCS
High-shear mixer	8,000	0.7	15.3	2205/SCS
Refuse thickener	35,040	0.2	83.0	PRAX
Water pump	87,600	0.1	40.0	PRAX/SCS
Rotary disc vacuum filter	5,896	0.9	24.0	PRAX
Belt filter	17,520	0.3	6.0	PRAX

* 5276: Availability Analysis of an IGCC, EPRI AP-5276, June 1987

PRAX: Praxis Engineers, Inc.

2205: Component Failure and Repair Data: GCC Units, EPRI AP-2205, February 1982

SCS: Southern Company Services, Inc.

** Based on 5400 service hours per year

Table 4-23
Plant Performance Reliability Results

Equipment	EAF	Product t/h
Conventional coal cleaning	90.00%	1083
Agglomeration process only	98.52%	69
Combination (1)	90.51%	1152

1. Total effect of conventionally cleaned and agglomerated products:
(1083 tons @ 90%) + (69 tons @ 98.52%)
1152 tons

Critical Components. The critical components are those components which cause the most lost production (equivalent coal processing time). The critical components are expressed here relative to the agglomeration system only. If the conventional coal cleaning facility were included, it would dominate the critical component list (89.7% of total lost processing time).

Table 4-24 shows the critical component ranking list. The ranking factor is interpreted to mean that if that component could be made perfectly reliable, the plant would increase its equivalent availability factor approximately by the number of percentage points equal to the ranking factor. The percentage of plant processing time lost is also shown.

Table 4-24
Critical Component Ranking List for
Agglomeration System Only

Component	Ranking Factor	Percent of Total Losses
Rotary vacuum filter	0.40	35
Refuse thickener	0.22	20
High-shear mixers	0.19	16
Flotation cells	0.15	13
Static thickener	0.12	10

Summary

Given the plant configuration, component failure and repair data, and assumptions used in this analysis, the processing rate of the total plant is 90.51% or 1152 t/h. This includes both the agglomerated and conventionally cleaned coal products. The agglomeration circuit itself has an EAF of 98.52% or 69 t/h from a possible product output of 70 t/h.

Plant Layout

The conceptual layout of the commercial plant is shown in Figure 4-14. The site plan has been developed based on space requirements for the process areas, material handling equipment, water supply pond, and traffic flow patterns including raw coal truck deliveries and product shipments. No site plan was developed for the natural fines circuit since it is assumed that the circuit will be incorporated into a conventional coal cleaning plant as a retrofit.

The 500,000-gal process water pond will require an area of 8000 ft². The pond will supply make-up water to the facility and serve as a source of fire protection water.

The traffic pattern for the plant will allow for 40-ton trucks (legal gross weight) to deliver coal to the raw-coal receiving station and proceed to the product silos to pick up the clean-coal product.

Plant layout details will depend on the site topography, available surface area, and the underlying geological structure.

Process Structure Layout

The process structure arrangement is shown in Figure 4-15. The layout is based on the following criteria:

- Equipment type
- Equipment operation and maintenance access requirements
- Relationship between individual unit operations

The system design is based on a two-story cleaning plant with strong consideration given to minimizing the number of pumps. The ground coal slurry from the holding vessel is pumped to high-shear vessels on the second floor. From these vessels, the slurry flows by gravity to the first flotation stage. The froth from the first flotation bank flows by gravity to a sump on the first floor and is pumped to the second flotation bank on the second floor. The froth from the second flotation bank also flows by gravity to the conditioning tanks on the first floor. The tails from the flotation cells flow by gravity to the thickener. The conditioning tank output is pumped to the vacuum filters on the second floor. Filter cake flows by gravity to the solids mixers and pelletizers located on the first floor. The pellets are transferred to the dryers by conveyors.

The conceptual design arrangement for the natural fines circuit is similar to the layout in the commercial agglomeration plant. Since this circuit is a retrofit to an existing conventional cleaning plant, the actual detailed layout would be affected by tie-ins to existing equipment and utilities and will therefore not be discussed here. In general, process equipment such as the high-shear vessels, flotation cells, vacuum filters, and control room facilities would be located on an upper level of the

structure, and flotation sumps, pumps, and tanks would be located on the ground level of the existing plant.

Equipment List

Major plant equipment and specifications are listed by flowsheet and process area in Appendix G.

Figure 4-14. Commercial Plant Layout

Figure 4-15. Process Layout: Agglomeration Structure

FINAL CONCEPTUAL DESIGN OF SEMIWORKS AGGLOMERATION PLANT

Design Criteria

This section outlines the design criteria used for the final conceptual design of the semiworks agglomeration plant. The semiworks plant flowsheet is similar in many respects to the agglomeration flowsheet presented in Section 4. The topics discussed below include feedstock, overall design basis, process conditions, plant performance, product specifications, and design considerations.

Feedstock

The semiworks flowsheet design utilizes either Upper Freeport or Pittsburgh No. 8 as the raw coal feed. The plant will operate as a mine-mouth facility. Mined raw coal will be transported to the plant via coal haulage trucks and stored in a sufficient quantity to assure a continuous coal supply to the plant.

The properties of the raw coal feed assumed in the semiworks design are the same as those used in the commercial agglomeration plant design in the previous section. These properties for the two raw coals are given in Table 5-1.

Table 5-1
Raw Coal Properties

	Upper Freeport	Pittsburgh No. 8
Moisture, %	5.19	4.84
Ash, %db	13.64	30.88
Total sulfur, %db	2.83	2.16
Pyritic sulfur, %db	1.96	1.48
Organic sulfur, %db	0.84	0.67
Sulfatic sulfur, %db	0.03	0.01
Calorific value, Btu/lb	13,300	10,167

Overall Design

The design of the semiworks agglomeration flowsheet incorporates a precleaning circuit to economically remove sulfur and ash from the ROM coal. Coal precleaning is accomplished by means of conventional cleaning techniques, and the precleaned product is then processed using selective agglomeration. Precleaning the coal reduces agglomeration throughput and improves overall process performance while decreasing process capital and operating costs. Table 5-2 summarizes the general design basis and major unit operations used for this flowsheet.

Table 5-2
Semiworks Flowsheet Design Basis

Capacity	20 t/h product
Coal supply	ROM coal
Site	Greenfield
Actual operating hrs	7560 hours/year
Availability	90% minimum
Unit Operations	
Crushing	To 1/4" topsize
Precleaning	96% Btu recovery minimum +100 mesh at 1.5-1.9 SG minus 100 mesh to agglomeration
Grinding	Size reduction to 95% minus 200 mesh
Agglomeration	Diesel (0.5%)
Dewatering	Vacuum filtration to 25% moisture max.
Size enlargement	Pelletization: starch and asphalt
Drying	Vibrating fluidized-bed to 4% moisture

Process Conditions

As stated earlier, each project coal requires unique process conditions for optimal coal recovery and sulfur and ash rejection. The process conditions selected for the semiworks plant are the same as those used in the design of the commercial plant. These conditions are repeated in Table 5-3.

Plant Performance

The plant performance for the semiworks agglomeration plant was estimated based on POC observations and demonstration run results. Similar results were used for the commercial plant design.

Table 5-3
Agglomeration Conditions

	Upper Freeport	Pittsburgh No. 8
High-Shear Conditions		
Feed size	95% minus 200-mesh	95% minus 200-mesh
HS residence time, min	1	1
HS solids conc., %	20	20
Agglomeration Conditions		
Agglomerate recovery method	2-stage flotation	2-stage flotation
Agglomerant	Diesel	Diesel
Agglomerant conc., %	0.5	0.5
Flotation Conditions - Stage 1		
Solids conc., %	6	6
Residence time, min	4	6
Frother dosage, lb/t	0.29	0.29
Flotation Conditions - Stage 2		
Solids conc., %	6	6
Residence time, min	4	6
Frother dosage, lb/t	0.29	0.29
Dewatering		
Method	Vac. filter	Vac. filter
Drying	Yes	Yes
Pelletization	Yes	Yes

Precleaning, agglomeration, and overall performance values on a ROM basis are provided in Table 5-4 for each of the project coals.

Product Specifications

The semiworks agglomeration plant will produce a dried pelletized product identical to that produced by the commercial agglomeration plant. The pellets are at least 6 mm in diameter and contain 2% starch and 0.5% asphalt. The pelletized product is easily handled by conventional coal handling systems and is suitable for utility combustion. Table 5-5 presents the final product characteristics expected for Upper Freeport and Pittsburgh No. 8 coals. These analyses are based on the process performance assumptions discussed previously.

Table 5-4
Plant Performance

	Upper Freeport	Pittsburgh No. 8
Precleaning Performance		
Yield, %	92.0	67.3
Btu recovery, %	96.7	93.3
Total sulfur rejection, %	45.7	31.7
Pyritic sulfur rejection, %	58.2	44.0
Ash rejection, %	34.6	83.4
Agglomeration Performance		
Yield, %	88.4	92.0
Btu recovery, %	92.8	94.6
Total sulfur rejection, %	37.4	21.2
Pyritic sulfur rejection, %	54.4	36.4
Ash rejection, %	46.4	35.7
Overall Performance, ROM basis		
Yield, %	81.3	61.9
Btu recovery, %	89.7	88.2
Total sulfur rejection, %	66.0	46.2
Pyritic sulfur rejection, %	81.0	64.4
Ash rejection, %	64.9	89.3

Table 5-5
Product Analysis

	Upper Freeport	Pittsburgh No. 8
Moisture, %	4.0	4.0
Ash, %db	5.9	5.3
Total sulfur, %db	1.18	1.88
Pyritic sulfur, %db	0.46	0.85
Calorific value, Btu/lb	14,674	14,487
lb SO ₂ /MBtu	1.61	2.60

Design Considerations

The semiworks plant shall comply with all applicable health, safety, and environmental regulations. Several regulatory permits, approvals, and reviews will be obtained from appropriate governmental agencies. Environmental requirements are discussed later.

The semiworks agglomeration plant is designed to operate 7,560 hours per year. This assumes 24-hour/day operation, 7 days/week, for 50 weeks/year. It is anticipated that the plant life of the semiworks facility will be 3 years maximum. The economics of the agglomeration process at the semiworks scale dictates

that the facility be operated for research purposes, and it would have limited usefulness for commercial production.

The semiworks plant shall be a fully instrumented facility with automated operation from a central control room. Startup and shutdown operations will be automatically controlled. All critical process streams will be monitored and controlled to maintain optimum process conditions. Additional details on instrumentation and control are provided later in this section.

As presented above in the design basis, the plant will maintain at least 90% availability. To satisfy this criterion, equipment spares are provided as needed. Details on plant availability are provided in a subsequent section.

Conceptual Flowsheet

This portion of the report discusses the semiworks agglomeration flowsheet. This flowsheet is similar to the commercial agglomeration flowsheet, with the significant differences being the throughput and amount and size of equipment. The flowsheet is described in detail by unit operation. Material balances and assumptions are also discussed.

Agglomeration Process Description

The semiworks flowsheet shown in Figure 5-1 depicts the entire agglomeration process and coal precleaning circuits. Material balances are provided in Table 5-6 and Table 5-7 for the Upper Freeport and Pittsburgh No. 8 feedstocks, respectively.

The first step following precleaning is grinding of the 1/4" x 0 coal to a nominal minus 200-mesh size in three stages. The ground coal slurry is agglomerated in a high-shear vessel, and the microagglomerates are recovered by flotation, dewatered, and pelletized. The pellets are then dried and stored on the product pad. The refuse is thickened in a water clarifier, filtered, and sent to a landfill. The clarified water is treated as necessary and recycled back to the process. This process is described in further detail below.

Coal Receiving, Storage, and Crushing. The 2" x 0 size plant feed coal is received via truck dump grates onto a coal storage pad. Coal from the pad is loaded into a hopper using a front-end loader. Vibrating feeders under the hopper regulate the transfer of coal onto the crusher feed coal belt conveyor. This belt is equipped with a magnet for tramp iron removal. The crusher

reduces the 2" x 0 raw coal feed to a 1/4" topsize product. The crushed coal is then received by a surge hopper providing one-half hour of surge capacity.

Coal Precleaning. The precleaning circuit for the commercial agglomeration plant is presented in Figure 5-2. The precleaning system will generate the following streams:

- 1/4" x 100-mesh clean coal (25% moisture)
- 100-mesh x 0 raw coal (25% solids)
- 1/4" x 100-mesh refuse (25% moisture)

Crushed coal is withdrawn from storage and fed to the precleaning circuit 24 hours/day at a rate of 25 t/h for Upper Freeport coal and 33 t/h for Pittsburgh No. 8 coal. The feed coal is slurried in a wetting box and screened at 28 mesh. The plus 28-mesh fraction reports to the heavy-media cyclone feed sump and the minus 28-mesh fraction reports to the classifying cyclone feed sump, from which it is pumped to a classifying cyclone and separated at 100 mesh. The raw coal is thus split into three size fractions: plus 28 mesh, 28 x 100 mesh, and 100 mesh x 0.

The plus 28-mesh fraction is pumped to the heavy-media cyclone. The clean coal (overflow) from the heavy-media cyclone flows to drain-and-rinse screens for magnetite removal and dewatering. The screen oversize (plus 28-mesh clean coal) then reports to the clean coal belt for transport to the coal grinding circuit.

The refuse (underflow) from the heavy-media cyclone reports to separate drain-and-rinse screens and then to the refuse belt for disposal. The effluent from the drain section of both the refuse and clean coal drain-and-rinse screens is recycled to the heavy-media feed sump, while the effluent from the rinse section of both of these screens is recycled to the heavy-media sump via the magnetite separator.

The classifying cyclone underflow (28 x 100-mesh) is separated by gravity flow in spiral classifiers. The clean-coal fraction is dewatered on 100-mesh screens. The oversize (28 x 100-mesh clean coal) reports to the clean-coal conveyor and is transported to the coal grinding circuit along with the plus 28-mesh clean coal.

The refuse fraction from the spiral classifiers is also dewatered on screens. The screen oversize (28 x 100-mesh refuse) goes to the refuse belt for disposal. The

Figure 5-1. Semiworks Agglomeration Flowsheet

Figure 5-2. Semiworks Precleaning Circuit

effluents from both the clean coal and refuse screens are recycled to the classifying cyclone feed sump. The spiral middlings and water streams are also recycled to the classifying cyclone sump.

The cyclone overflow (minus 100 mesh) is sent to a static thickener. The underflow from the thickener (~25% solids) is pumped to the agglomeration circuit rod mill feed. The static thickener overflow is recycled to the raw coal deslime screen.

Coal Grinding. The 1/4" x 100-mesh precleaning circuit product is fed to the grinding circuit at a nominal rate of 23 t/h for the Upper Freeport coal and 22 t/h for the Pittsburgh No. 8 coal. This stream is temporarily stored in a rod mill feed hopper and then delivered to the rod mill. The rod mill hopper is designed to have about one-half hour of storage capacity for the grinding system. A weigh-belt feeder reclaims the coal from the hopper to the rod mill. Precleaning static thickener underflow (minus-100-mesh coal) and water are added to the rod mill, diluting the input streams to about 50% solids. The rod mill grinds the 1/4" x 0 coal to a nominal 28-mesh topsize.

The rod mill product discharges to the ball mill sump from which it is pumped to the ball mill. The ball mill operates

at the same solids concentration as the rod mill and is designed to grind the 28-mesh topsize coal to a nominal 100-mesh topsize ($d_{50} = 60$ microns). An additional ball mill operated in series further grinds the coal to a nominal 200-mesh topsize ($d_{50} = 25$ microns). The ball mill product discharges to a sump from which it is pumped to a slurry holding vessel. The slurry holding vessel is designed to hold a 10-minute supply of ground coal slurry at 50% solids and is equipped with an agitator to assure that the slurry is homogenous.

Agglomeration. Coal slurry from the holding vessel flows to a high-shear sump. Precleaning circuit static thickener underflow and water are added to the sump to dilute the slurry to 20% solids. The diluted slurry is then pumped from the high-shear sump to the high-shear vessel. The percent solids and flow rate of the high-shear feed are monitored, and these data are used to meter 0.5% diesel (on a dry coal feed basis) to the high-shear vessel. The high-shear vessel is designed to provide one minute of retention time. During this period, phase inversion occurs and microagglomerates are formed, with liberated ash and pyrites reporting to the aqueous phase.

The high-shear vessel uses a top discharge from which the agglomerated slurry flows to a bank of five flotation

cells where it is diluted to 6% solids. During flotation, microagglomerates are recovered from the mineral matter-bearing aqueous phase. The retention time is coal-dependent, and was provided in Table 5-3. Tailings from flotation are sent to the plant tailings sump.

The first-stage flotation product (solids concentration of approximately 20%) flows by gravity to the flotation feed sump where it is diluted to 6% solids. The slurry from the sump is pumped to the second-stage flotation cells where flotation is repeated at a residence time similar to that of the first stage of flotation. The tailings are sent to the plant tailings sump. The flotation product is once again assumed to be at 20% solids.

Agglomerate Dewatering. The product from the second stage of flotation flows by gravity to the conditioning tank. The conditioning tank is designed to provide one minute of storage and includes a mixer to disperse the froth. The discharge from the conditioning tank is pumped to a rotary vacuum disc filter. The filter is designed to yield a product with a moisture content of approximately 25%. The filtrate from the vacuum filter is recycled to the flotation-feed sump.

Size Enlargement, Drying, and Product Loadout. The filter cake from the vacuum filter flows by gravity to a solids mixer. An asphalt emulsion is added to the solids mixer by means of a metering pump at a rate providing 0.5% asphalt on a solids basis. Starch (2% on a dry solids basis) is also fed to the mixers via a feed hopper and weigh-belt feeder. The purpose of the solids mixer is to disperse the binders (asphalt and starch) within the cake. The solids mixer discharges onto a pelletizer that produces 6-8 mm pellets. The pelletizer discharges onto a belt conveyor that transports the pellets to the dryer feed hopper. The hopper provides one hour of storage capacity for the "green" pellets. The hopper feeds a vibratory fluidized-bed dryer designed to reduce the moisture content of the pellets to 4% in order to increase their strength. The pellet retention time in the dryer is estimated to be 2-3 minutes. The dried pellets are then discharged onto a cooling conveyor. The final clean-coal product is stored on a storage pad that provides a 5-day storage capacity. Product loadout will be by loader and truck.

Water Clarification and Treatment. The plant tailings streams from the agglomeration circuit are pumped from the tailings sump to the water clarifier (thickener). Flocculants and lime (for pH control) are used to facilitate water clarification. The thickened refuse stream, at approximately 25% solids, is pumped to a belt filter

press. The refuse belt press filtrate is recycled back to the water clarifier. Clarified water is circulated between the clarifier overflow tank and the head tank, with process water being removed from the head tank as required. A portion of this clarified-water stream (approximately 2%) is treated to keep contamination levels within acceptable limits.

Refuse Dewatering, Handling, and Disposal. The clarifier underflow (agglomeration plant refuse) is pumped to the belt press via the thickener discharge sump. The belt press is designed to process approximately 3 t/h of solids and will dewater the refuse stream to a moisture content of 25-30%. This fine refuse is mixed with the coarse refuse from the precleaning facility and transported to a landfill site by truck.

Material Balances

Material balances have been developed and presented in Tables 5-6 and 5-7 for the Upper Freeport and Pittsburgh No. 8 feed coals, respectively. Estimates of process performance, coal quality, and precleaning performance (where applicable) were provided above under the subheading, "Design Criteria." Previous experience with conventional coal plants and selective agglomeration unit operations was used to estimate the performance within each major unit operation to complete the material balance. Several assumptions were made during the conceptual design regarding feed coal quality, plant performance, and plant operations. These assumptions are discussed below.

Material Balance Assumptions. The following assumptions were used to derive a material balance for the agglomeration flowsheet using the Upper Freeport and Pittsburgh No. 8 coals:

- Feed coal quality was based on project feed coal qualities as measured during Phase II testing.
- Precleaning performance for the semiworks design will be similar to that observed during test coal preparation
- Agglomeration performance for the semiworks design is based on demonstration-run performance for both coals.
- The distribution of ash and sulfur between flotation stages corresponds to POC observations.

Table 5-6
Upper Freeport Coal Material Balance

Stream	No.	Coal t/h	Weight Percent			Water gpm	Diesel gpm	Asphalt gpm	Starch gpm	Slurry gpm	% Sol	% H ₂ O	Cal. Val. Btu/lb
			Ash	PS	TS								
Coal to precleaning	1	24.6	13.6	1.96	2.83	5				79	95	5	13,300
Water to precleaning	2					45				45			
Precleaning refuse	3	2.0	59.0	14.27	16.17	3				9	75	25	5,434
-100M coal to rod mill	4	1.6	15.3	3.40	4.25	19				24	25	75	12,798
Clean coal to rod mill	5	21.0	9.3	0.70	1.47	28				91	75	25	14,074
Water to rod mill	6					43				43			
Slurry from grinding	7	22.6	9.7	0.89	1.67	91				158	50	50	13,984
Water to high shear	8					271				271			
Diesel to high shear	9						0.5			0			
High-shear slurry	10	22.6	9.7	0.89	1.67	362	0.5			430	20	80	13,984
Water to flotation 1	11					1,056				1,056			
Slurry to flotation 1	12	22.6	9.7	0.89	1.67	1,418	0.5			1,486	6	94	13,984
Tails from flotation 1	13	1.8	42.4	4.28	5.56	1,086				1,091	1	99	7,991
Froth from flotation 1	14	20.8	6.8	0.59	1.33	332	0.5			395	20	80	14,514
Water to flotation 2	15					678				678			
Slurry to flotation 2	16	20.8	6.8	0.59	1.33	1,303	0.5			1,365	6	94	14,514
Tails from flotation 2	17	0.8	29.8	3.87	5.05	983				985	0	100	10,484
Froth from flotation 2	18	20.0	5.9	0.46	1.18	320	0.5			380	20	80	14,674
Vacuum filter filtrate	19					293				293			
Vacuum filter cake	20	20.0	5.9	0.46	1.18	27	0.5			87	75	25	14,674
Binder to solids mixer	21							0.4	1.2	2			
Pellets to dryer	22	20.0	5.9	0.46	1.18	27	0.5	0.4	1.2	89	75	25	14,674
Dried product to storage	23	20.0	5.9	0.46	1.18	3	0.5	0.4	1.2	65	96	4	14,674
Moisture to flue	24					23				23			
Tails to clarifier	25	2.6	38.6	4.16	5.40	2,096				2,104	0	100	8,741
Clarifier overflow	26					2,064				2,064			
Clarifier underflow	27	2.6	38.6	4.16	5.40	32				39	25	75	8,741
Agglomeration refuse	28	2.6	38.6	4.16	5.40	5				12	70	30	8,741
Belt press effluent	29					27				27			
Refuse to landfill	30	4.6	47.3	8.48	10.01	7				21	72	28	7,326
Water to treatment	31					41				41			
Untreated process water	32					2,023				2,023			
Water to head tank	33					2,064				2,064			
Water from head tank	34					2,093				2,093			
Make-up water	35					29				29			

Table 5-7
Pittsburgh No. 8 Coal Material Balance

Stream	No.	Coal t/h	Weight Percent			Water gpm	Diesel gpm	Asphalt gpm	Starch gpm	Slurry gpm	% Sol	% H ₂ O	Cal. Val. Btu/lb
			Ash	PS	TS								
Coal to precleaning	1	32.3	30.9	1.48	2.16	6				103	95	5	10,167
Water to precleaning	2					57				57			
Precleaning refuse	3	10.6	78.8	1.99	2.10	14				46	75	25	2,095
-100M coal to rod mill	4	1.9	36.5	1.91	2.42	22				28	25	75	9,798
Clean coal to rod mill	5	19.9	4.9	1.17	2.17	26				86	75	25	14,494
Water to rod mill	6					38				38			
Slurry from grinding	7	21.7	7.6	1.23	2.19	87				152	50	50	14,089
Water to high shear	8					261				261			
Diesel to high shear	9						0.4			0			
High-shear slurry	10	21.7	7.6	1.23	2.19	347	0.4			413	20	80	14,089
Water to flotation 1	11					1,015				1,015			
Slurry to flotation 1	12	21.7	7.6	1.23	2.19	1,362	0.4			1,428	6	94	14,089
Tails from flotation 1	13	1.3	35.4	5.78	5.94	1,036				1,040	1	99	9,236
Froth from flotation 1	14	20.4	5.8	0.94	1.95	326	0.4			388	20	80	14,402
Water to flotation 2	15					661				661			
Slurry to flotation 2	16	20.4	5.8	0.94	1.95	1,280	0.4			1,342	6	94	14,402
Tails from flotation 2	17	0.4	29.5	5.00	5.21	960				962	0	100	10,417
Froth from flotation 2	18	20.0	5.3	0.85	1.88	320	0.4			380	20	80	14,487
Vacuum filter filtrate	19					293				293			
Vacuum filter cake	20	20.0	5.3	0.85	1.88	27	0.4			87	75	25	14,487
Binder to solids mixer	21							0.4	1.2	2			
Pellets to dryer	22	20.0	5.3	0.85	1.88	27	0.4	0.4	1.2	89	75	25	14,487
Dried product to storage	23	20.0	5.3	0.85	1.88	3	0.4	0.4	1.2	65	96	4	14,487
Moisture to flue	24					23				23			
Tails to clarifier	25	1.7	34.0	5.59	5.76	2,014				2,019	0	100	9,527
Clarifier overflow	26					1,993				1,993			
Clarifier underflow	27	1.7	34.0	5.59	5.76	21				26	25	75	9,527
Agglomeration refuse	28	1.7	34.0	5.59	5.76	3				8	70	30	9,527
Belt press effluent	29					18				18			
Refuse to landfill	30	12.3	72.4	2.50	2.62	17				54	74	26	3,148
Water to treatment	31					40				40			
Untreated process water	32					1,953				1,953			
Water to head tank	33					1,993				1,993			
Water from head tank	34					2,031				2,031			
Make-up water	35					38				38			

Unit Operations

The unit operations utilized in the semiworks agglomeration plant are similar to their counterparts in the commercial agglomeration plant. Each unit operation was described in detail in Section 4, under the subheading "Unit Operations." Design aspects relevant only to the semiworks plant are presented below.

Equipment sizing was selected to reflect commercial-scale operations. Parallel processing of streams was eliminated to allow larger equipment sizes which will facilitate scale-up to commercial operations.

Coal Receiving, Storage, and Crushing

A storage pad was selected for coal storage for the semiworks plant because it is anticipated that the operating costs associated with loader operations would be lower than the capital cost of a storage silo or stacking operation given the short duration of the semiworks operation.

Coal Precleaning

A split-screen unit was chosen to dewater both the heavy-media cyclone overflow and underflow streams to reduce structural and piping costs. The split-screen unit makes efficient use of plant space and is less expensive than two separate screens.

Coal Grinding

Consideration was given to replacing the two ball mills with one larger ball mill. Two ball mills were used in both the POC facility and the conceptual design for the commercial plant. However, the two-ball-mill design was retained for better control of coal particle size distribution since it allows the use of two different size distributions of grinding media that improves grinding energy efficiency. The final grind size can be better controlled because the grinding media in the second ball mill are not subjected to the wear caused by the 28-mesh to 100-mesh size reduction step.

Agglomeration and Agglomerate Dewatering

The parallel circuits containing the high-shear mixers, flotation cells, and dewatering equipment were eliminated and only one circuit is used. This allows better scale-up from semiworks operation to commercial-scale operation.

Size Enlargement, Drying, and Product Loadout

Only one circuit containing a solids mixer, pelletizer, and dryer are used in the semiworks plant design. Improved scale-up and lower capital expenditures are two reasons for the change.

Agglomerant and Reagent Storage and Delivery

The equipment list and cost estimate assume small tanks used for reagent storage. Delivery and storage of reagents in 50-gallon drums is another possible alternative. Metering pumps would then be portable and would be attached to reagent drums as needed. An actual semiworks plant may consider either option for reagent handling depending on reagent availability and the cost benefits associated with bulk storage.

Process Instrumentation and Control

A process control system is needed for smooth, continuous plant operations at near-optimal conditions. The system is also included to study instrumentation and control of the agglomeration process for scale-up purposes. All instrumentation provided reflects the instrumentation for the commercial-scale process with a few exceptions. Therefore, the discussion in Section 4 provides an adequate description of the process control system for the semiworks plant. The exceptions are detailed below.

The on-line analyzers included in the commercial-scale conceptual design are not expected to be included within the semiworks instrumentation. On-line analysis is not expected to be economic given the throughput of the plant and its limited operating life. The analyzers measuring product properties will be replaced with laboratory analysis.

Process optimization computation capabilities will not be required for the semiworks process control system. Process optimization is not expected to be beneficial from an economic standpoint given the expected operating life of the plant. Investigation of process optimization for research purposes in the semiworks plant would be of limited value given the current knowledge base of process behavior and process control.

Finally, the number of signals requiring processing will be reduced since the semiworks plant utilizes less equipment than the commercial-scale plant. The number

of control variables for the semiworks plant is listed by circuit in Table 5-8.

Table 5-8
Control Signal Summary

Circuit	Analog Signals	Digital Signals
Receiving, storage, and crushing	5	13
Grinding	8	22
Agglomeration	16	12
Agglomerate dewatering	7	10
Size enlargement, drying, and loadout	7	18
Water clarification/Refuse handling	26	14
<i>Total</i>	69	89

Ancillary Requirements

The ancillary requirements for the semiworks agglomeration plant are similar to those for the commercial-scale plant. Items such as definition of the battery limits, identification of utility requirements, traffic needs, emissions, and refuse disposal needs are addressed below.

Battery Limits

The semiworks agglomeration plant is assumed to be located such that all utility requirements are available at the plant boundary. This assumption has a significant impact on the capital cost estimate. The plant is assumed to be fairly close to sources of power, water, and gas, which are supplied by a local utility company at the required voltage, flow rate, and pressure, respectively.

This assumption is consistent with the premise that the semiworks plant will be used primarily for research purposes.

Utilities

Electrical. Power is supplied to the semiworks plant using the same methods as were employed in the commercial design. However, power requirements will be lower given the reduced throughput and subsequent power draw from the process equipment. A portion of the power requirements will not change significantly. This portion includes lighting and power to supporting facilities.

Water. Distribution system requirements for the potable, sanitary, and firefighting water systems for the semiworks plant are expected to be similar to those of the commercial plant since the water load is based on operating and support personnel. Process water requirements will be reduced given the reduction in process throughput. Process water storage will have a 185,000-gallon capacity, of which 125,000 gallons will be required for firefighting and process water make-up for one day.

Natural Gas. Natural gas requirements for the dryer burners in the pelletization circuit are estimated to be 21,100 ft³/h (heating value of 1,000 Btu/ft³). Natural gas will be supplied by a local utility. No provisions will be made to store natural gas at the site.

Compressed Air. The compressed air delivery pressure is assumed to be 80-90 psig. The distribution system is designed to handle 20 ft³/min of compressed air.

Traffic Needs

The traffic needs of the semiworks plant will be similar to those of the commercial plant. Two differences regarding traffic anticipated for the semiworks plant include the loading and shipment of raw materials and product and traffic related to plant personnel.

Raw coal and product storage will be on storage pads located close to the facility. The area surrounding the pad will be sized for unloading and loading operations using a front-end loader. This area will be restricted to other traffic for safety and operational purposes.

The semiworks facility will employ approximately one-half the number of employees required by a commercial-scale operation. This assumes that some administrative and engineering duties will occur off-site. Given this reduction in personnel and the expected plant life of three years, construction of temporary parking lots using compacted baserock or other suitable material will satisfy parking requirements.

Environmental

Environmental requirements including water quality, air quality, surface runoff, hydrocarbon runoff, emissions control, and refuse disposal will be the same as those discussed earlier for the commercial agglomeration plant.

Techniques to Maintain 90% Plant Availability

A reliability analysis was performed on the 20-t/h semiworks selective agglomeration plant. Since there is a close similarity between the configuration of this plant and that of the 200-t/h commercial-scale plant, only the major differences and assumptions are noted below.

Major Differences and Assumptions

The size difference of 20-t/h product versus 200-t/h product had no direct impact on reliability because essentially the same types of components and operating philosophy were assumed. The configuration of the components was almost identical except for the following:

- One set of vibrating coal feeders instead of two
- The vibrating feeder is 1 x 100% vs. 4 x 33.3%
- The high-shear vessel is 1 x 100% vs. 6 x 16.7%
- The first flotation cells are 5 x 20% vs. 10 x 10%
- The second flotation cells are 4 x 20% vs. 8 x 12.5%
- The conditioning tank and vacuum filter are 1 x 100% vs. 2 x 50%
- The pelletization train is 1 x 100% vs. 4 x 25%
- There is no coal stacking tube but a stockpile of feed coal is assumed to be available.
- The coal precleaning circuit has only minor design differences.

For purposes of this study, the reliability was assumed to be the same as in the 200-t/h case.

Two banks of flotation cells have a runout capacity in excess of the nominal capacity of each cell, as shown in Table 5-9.

Availability Block Diagram. Because the 20-t/h semiworks plant design is very similar to that of the 200-t/h commercial plant, an ABD was not prepared. Please refer to the ABD for the 200-t/h plant, noting the differences described in the preceding section. However, a complete computer model was created using these differences, and the results are given below.

Table 5-9
Excess Capacity

	1 Fail	2 Fail	3 Fail
First flotation cells (20% nominal)	90%	75%	60%
Second flotation cells (25% nominal)	80%	60%	35%

Results

The results presented in Table 5-10 are shown in direct comparison to the 200-t/h commercial-scale plant. Please refer to the results section for the larger plant for any interpretation.

Table 5-10
Availability Results

	20 tph	200 tph
Availability factor (AF)	95.36%	97.19%
Equivalent availability factor (EAF)	94.95%	95.16%
Full load operating factor (FLOF)	93.37%	88.99%
Service hours per year	8033	8187
Coal processing rate, t/h	19	190
Service hours at full load per year	7865	7497

The EAF (or coal processing rate) of the 20-t/h plant is very close to the 200-t/h plant performance. Both plants easily meet the goal of 90% availability.

The main differences lie in the AF and FLOF. Because the 20-t/h plant is composed of 1 x 100% trains compared to 2 x 50% or 4 x 25% trains as in the 200-t/h case, there are more times when the 20-t/h plant is completely inoperable. Therefore, the AF is lower and the expected service hours per year are lower.

Conversely, the 200-t/h plant has more components (in parallel) which can fail. The 200-t/h plant will run more service hours but those extra service hours are usually at a reduced capacity. Therefore, the 20-t/h plant has a higher FLOF and more service hours per year at full load.

Availability Block Diagram. The critical components for the 20-t/h plant are essentially the same as those for the 200-t/h plant, with the differences as noted previously.

Summary

Given the plant configuration, component failure and repair data, and assumptions used in this analysis, the 20-t/h semiworks selective agglomeration plant will meet or exceed its performance goal of 90% (18 t/h) of its design process rate (20 t/h product). The EAF is expected to be 94.95% or 19 t/h.

This analysis used most of the data and assumptions presented in Section 4 for the 200-t/h commercial-scale plant. Please refer to that section for details. This reliability model and EPRI's UNIRAM computer program may be used to study the impact of trading off parallel-configured components with spares and maintenance philosophy.

Plant Layout

The conceptual layout of the semiworks agglomeration plant is not expected to differ substantially from the layout presented for the commercial plant. Size requirements for the precleaning and agglomeration structures would be reduced to reflect the removal of parallel circuits of equipment provided in the commercial plant and the downsizing of equipment due to reduced throughput. Plot size requirements will be approximately equal to the commercial plant requirements since space requirements for roads and spaces between structures remain constant and the additional space required for loader operations will offset structure downsizing.

Equipment List

Major plant equipment and specifications are listed for the semiworks plant by process area in Appendix G, Tables G-13 through G-19.

ECONOMICS OF COMMERCIAL AGGLOMERATION PLANT

The assumptions, cost estimation, cost analysis, and economic analysis for the commercial agglomeration plant and natural fines circuit are detailed below. Sensitivity analyses analyzing the economics without size enlargement and the rate of return, operating hours, and plant size are presented as well.

Assumptions

Several assumptions regarding the design basis for the commercial agglomeration plant were presented in Section 4. These assumptions are summarized in Table 6-1 along with the general economic assumptions used for this economic analysis.

Table 6-1
Economic Assumptions

Actual operating hours	7,560
Plant availability	90%
Production rate	200 t/h product
Plant life	20 years
Royalties	None
Financing	100% equity
Inflation rate	4.0%
Corporate tax rate	38.0%
Rate of Return	10.0%

Land requirements for the commercial agglomeration plant are based on mass balance calculations for the volume of refuse going to landfill. Sufficient land has been allocated to the landfill to hold the refuse for the 20-year life of the plant. The land cost is assumed to be \$5,000/acre.

It is also assumed that plant construction will take place during a two-year period. It is estimated that one-half of the construction expenditures will be incurred in the first year and the remaining half in the second year.

Plant depreciation is based on the Modified Accelerated Cost Recovery System (MACRS). This system allows assets to be depreciated over a seven-year period utilizing the double declining balance method and switching to a straight line method in the taxable year that maximizes the depreciation allowance. This approach is typical for a stand-alone coal preparation facility.

Cost Estimation

The costs for major equipment items and their power requirements are based on vendor quotations, prices of similar equipment, and estimating guides. The costs are reported in mid-1991 dollars. A detailed listing of the equipment in the commercial agglomeration plant and the natural fines agglomeration circuit is included in Appendix H, Economic Details. The listing includes estimated costs, spares, and estimated power consumption. Spare equipment requirements were determined during the availability analysis which is presented in Section 4.

Direct costs such as structures, piping, foundations, HVAC, electrical, instruments, and buildings have been estimated from equipment costs using factors based on previous estimating experience. A cost for replacement parts was assumed to be 10% of the total major plant equipment cost.

Indirect costs such as temporary construction, design engineering, construction supervision, and project contingency are then estimated from the total direct fixed costs. The factors used to estimate the direct and indirect costs are presented in Appendix H.

A process contingency factor of 10% has been applied to the purchase price of each item of equipment in an attempt to quantify the uncertainty in technical performance and equipment cost. A 20% project contingency factor was applied to the total process capital cost to cover the cost of additional equipment or additional costs resulting from a more detailed site-specific design. This

factor is based on the level of design and cost estimating performed. These factors are in accordance with EPRI Technical Assessment Guide (TAG) recommendations.

The preproduction costs used in calculating the total capital requirement are shown in Appendix H. Preproduction costs include one month each of variable operating costs (excluding fuel) and fixed operating costs (labor and maintenance material), 25% of the fuel cost for one month, and 2% of the total plant investment to cover startup modifications.

Inventory capital is estimated using the sum of raw material costs for seven days of plant operation at full capacity. Spare parts are also included in the inventory capital estimate and are assumed to be 2% of the total plant cost.

Annual operating and maintenance costs were estimated assuming the unit costs for raw materials, labor rates, and utilities as shown in Appendix H. These costs include raw materials, utilities, operating manpower, maintenance labor, maintenance material, and administrative and support labor.

Operating costs are based on mass balances, vendor quotations, literature references, and SCS in-house data. The electrical power requirement is based on the installed horsepower of the operating electrical equipment (excluding the horsepower of spare equipment). The maintenance material cost is estimated to be 6% of the process capital cost as recommended by the EPRI TAG. Administrative and support labor is assumed to cost 30% of operating and maintenance labor costs.

Commercial Agglomeration Economics

The overall plant investment for the commercial agglomeration plant is approximately \$55,600,000. The total capital requirement excluding working capital is approximately \$59,500,000. The total capital requirement includes the overall plant investment, preproduction costs, and land purchase costs. Each of these costs is shown by component in Appendix H. The breakdown of the purchased equipment cost by circuit is shown in Figure 6-1.

Annual operating and maintenance costs are presented in Table 6-2. Of the total O&M cost of each plant, approximately 45% is allocated to materials, 38% to utilities, and 10% to labor.

Figure 6-1. Equipment Cost Breakdown

Table 6-2
Annual Operating and Maintenance Costs

	Upper Freeport	Pittsburgh No. 8
Raw Material		
Diesel	\$1,385,748	\$1,518,426
Asphalt	\$907,200	\$907,200
Starch	\$7,257,600	\$7,257,600
Flocculant	\$79,532	\$143,338
Frother	\$272,890	\$298,337
Lime	\$68,433	\$74,814
Magnetite	\$286,176	\$311,215
Preclean flocculant	\$48,384	\$56,548
Total	\$10,305,963	\$10,567,478
Utilities		
Electricity	\$4,609,937	\$4,609,937
Natural gas	\$5,556,600	\$5,556,600
Process water	\$16,353	\$21,342
Total	\$10,182,890	\$10,187,879
Labor		
Operating	\$1,215,000	\$1,215,000
Maintenance	\$607,500	\$607,500
Administrative/support	\$546,750	\$546,750
Total	\$2,369,250	\$2,369,250
Maintenance Material	\$2,182,132	\$2,182,132
Insurance/Taxes	\$1,668,339	\$1,668,339
Total O&M Cost	\$25,040,235	\$25,306,739

The economics for the Upper Freeport and Pittsburgh No. 8 operating cases were analyzed using methods described in the Coal Preparation Plant Evaluation Model User Guide and Technical Model prepared by Eos Technologies, Inc. under DOE Contract No. DE-AC22-87PC79338, Subcontract No. PTS-88-8244.

For modeling purposes, a value of \$0.80 per million Btu was assumed for the feed coal. This value is used to determine the after-tax cost of Btu loss during process-

ing. Btu loss is one component of the incremental product costs calculated in the economic analysis.

The incremental product cost is defined as the total plant processing cost including capital investment and operation and maintenance costs. The incremental cleaning costs exclude feedstock coal costs, with the exception of the cost of Btu lost during processing. The value of the feed coal can be added to the incremental product cost to determine a total product cost.

The payback period for the agglomeration plant for either coal is approximately 7.9 years. The payback period assumes undiscounted cash flows.

The annualized product costs in \$/ton of product, \$/million Btu, and \$/ton of SO₂ removed are presented in Table 6-3. Desulfurization costs are presented on a before-tax basis.

Table 6-3
Commercial Agglomeration Economic Summary

	Upper Freeport	Pittsburgh No. 8
<i>Annualized Product Costs</i>		
Incremental, \$/ton	\$24.73	\$25.32
Incremental, \$/million Btu	\$0.84	\$0.87
Desulfurization, \$/ton SO ₂	\$333.56	\$488.08

The annualized product cost for both cases is influenced primarily by variable costs (diesel, etc.). The operating and maintenance costs are approximately 70% of the final product cost.

There is little difference between the Upper Freeport and Pittsburgh No. 8 product costs; the slight difference in cost is due to differences in agglomeration yield. The Pittsburgh No. 8 coal yields less product per ton of agglomeration feed, and thus more feed is processed and costs are slightly higher.

The desulfurization costs are greater for the Pittsburgh No. 8 coal due to the fact that the agglomeration process rejects less total sulfur when processing Pittsburgh No. 8 coal than is the case with Upper Freeport coal.

Economic Analysis Without Size Enlargement

Costs have also been calculated for two process options which do not include size enlargement. The first option, shown in Table 6-4, produces a product thermally dried

to 8% moisture. Such a product could be stored prior to combustion at a power plant using silos designed for fine coal storage that incorporate an inerting system.

Table 6-4
8% Moisture Economic Summary

	Upper Freeport	Pittsburgh No. 8
<i>Annualized Product Costs</i>		
Incremental, \$/ton	\$17.70	\$18.28
Incremental, \$/million Btu	\$0.60	\$0.63
Desulfurization, \$/ton SO ₂	\$238.66	\$352.41

This option results in an incremental cost savings of \$7.03 and \$7.04 per ton of product for Upper Freeport and Pittsburgh No. 8 coals, respectively. The savings is due primarily to the elimination of capital equipment associated with the pelletization process and with the operating cost savings associated with the elimination of asphalt and starch. There is also a savings in utility costs because less natural gas is required in the drying process.

Costs have also been calculated for a product with no size enlargement and no thermal drying. This yields a product with approximately 25% moisture after filtration. The economic estimates resulting from this option are shown in Table 6-5. Further cost reductions are due primarily to the lack of drying costs.

Table 6-5
No Size Enlargement or Thermal Drying

	Upper Freeport	Pittsburgh No. 8
<i>Annualized Product Costs</i>		
Incremental, \$/ton	\$14.08	\$14.72
Incremental, \$/million Btu	\$0.48	\$0.51
Desulfurization, \$/ton SO ₂	\$189.95	\$283.79

Sensitivity Analysis

Sensitivity analyses were performed for the agglomeration plant economics to determine the effects of changing rate of return, operating hours and plant size.

Rate of Return

The base case analysis presented in Table 6-3 is based on 10% rate of return. A sensitivity analysis was performed on the commercial agglomeration plant for

15% and 20% rates of return. The annualized product costs are presented in Tables 6-6 and 6-7.

Table 6-6
15% Rate of Return Economic Summary

	Upper Freeport	Pittsburgh No. 8
<i>Annualized Product Costs</i>		
Incremental, \$/ton	\$27.10	\$27.68
Incremental, \$/million Btu	\$0.92	\$0.96
Desulfurization, \$/ton SO ₂	\$365.45	\$533.69

Table 6-7
20% Rate of Return Economic Summary

	Upper Freeport	Pittsburgh No. 8
<i>Annualized Product Costs</i>		
Incremental, \$/ton	\$29.94	\$30.53
Incremental, \$/million Btu	\$1.02	\$1.05
Desulfurization, \$/ton SO ₂	\$403.82	\$588.56

Operating Hours

Figure 6-2 shows the effect of operating hours on costs. From the illustration, it is evident that costs increase as operating hours decrease. Reducing the number of operating hours from 7,560 hours/year to 5,400 hours/year results in cost increases of about \$5.41 per ton of product. The increase in operating hours also increased desulfurization costs by 23%. This is expected since capital costs are a constant and at lower operating hours the burden of the capital charges on the total product cost increases.

The product costs are only moderately sensitive to changing operating hours. This is expected since the majority of the costs are in operating costs. Additional sensitivity would be observed if a larger portion of the final product cost was represented by the capital investment.

Plant Size

Equipment costs were scaled up using an exponential method according to the formula below.

$$\text{Cost}_{(500\text{-t/h})} = \text{Cost}_{(200\text{-t/h})} \times (500/200)^{0.70}$$

An exponential factor of 0.7 was chosen for scale-up since it is consistent with published references for the type of plant equipment selected. Electrical power

requirements were scaled up using a value of 0.9 for the exponent. Other utilities and costs were scaled up in direct proportion to plant throughput. Product and desulfurization costs are presented for the 500-t/h agglomeration plant in Table 6-8.

Figure 6-2. Operating Hours Sensitivity Analysis

Table 6-8
500-t/h Case

	Upper Freeport	Pittsburgh No. 8
<i>Annualized Product Costs</i>		
Incremental, \$/ton	\$22.03	\$22.62
Incremental, \$/million Btu	\$0.75	\$0.78
Desulfurization, \$/ton SO ₂	\$297.11	\$435.78

The product costs can be decreased by \$2.70 per ton of product produced if the plant is scaled-up to 500 t/h. There is also an 11% reduction in desulfurization costs for both coals. Plant scale sensitivity is dependent on the percentage of annualized capital investment in relation to the incremental product cost. The larger the portion that the capital investment represents, the greater the savings associated with larger scale operations.

Economics of Natural Fines Circuit

The economic evaluation for processing the natural fines circuit processing 114 t/h of feed is presented below. It is assumed that the conventional plant processes 1386 tons of the 1500 tons of the hourly raw coal feed while the minus 100-mesh fraction (114 tph) is processed using agglomeration.

The design basis for the natural fines circuit is discussed in Section 4. Assumptions regarding plant performance

and operating conditions are also detailed in Section 4. The natural fines circuit is designed to process 114 tons of minus 100-mesh fines per hour. The circuit is expected to operate at 85.0% Btu recovery and 79.0% pyritic sulfur rejection. The circuit yield is 61.0%. All circuit performance values are on a dry basis.

Capital and operating costs were estimated for the retrofit installation of the circuit into a conventional coal cleaning plant. No land costs are included for the circuit. It is expected that the retrofit circuit will be located at an existing plant on land owned by the host site.

The estimated plant equipment costs for the circuit are \$4,353,600. Approximately 46% of the total represents the cost of equipment in the agglomeration circuit, 25% represents water clarification and refuse handling equipment, 22% represents the coal thickening equipment required for dewatering prior to agglomeration, and the remaining 7% represents product loadout and reagent storage/delivery equipment.

The estimated plant capital investment required to incorporate a natural fines agglomeration circuit into a conventional coal cleaning plant is \$14,645,510. The operating and maintenance costs associated with the operation is presented in Table 6-9.

Table 6-9
Annual Operating and Maintenance
Costs for Natural Fines Circuit

Consumables	
Diesel	\$673,920
Flocculant, clarification	\$56,808
Frother	\$82,952
Lime	\$20,801
Flocculant, for feed	\$246,240
Subtotal	\$1,080,721
Utilities	
Electricity	\$714,096
Process water	\$11,680
Subtotal	\$725,776
Labor	
Operating	\$121,500
Maintenance	\$81,000
Administrative/support	\$60,750
Subtotal	\$263,250
Maintenance Material	\$574,675
Insurance/Taxes	\$439,365
Total O&M Cost	\$3,083,787

Product costs for the fines circuit and the entire plant are summarized in Table 6-10. The incremental product cost for the fines circuit has been calculated assuming a rate of return equal to 10%. This cost should be compared to the processing costs of other coal cleaning alternatives. The coal cleaning costs are offset by the additional revenue generated from the circuit; however, this offset is not included in the product costs presented in the table.

The plant product costs are calculated assuming an incremental product cost of \$2.00 per ton for the conventionally processed coal. In practice, conventional cleaning costs vary significantly between operating facilities. Costs are presented to illustrate the increase in plant processing costs when the retrofit circuit is included. The cost of conventional coal cleaning is assumed to exclude profit. Profits and rate of return from the circuit are discussed later in this section.

Table 6-10
Natural Fines Case Economic Summary

<i>Annualized Product Costs (Fines Circuit)</i>	
Incremental, \$/ton	\$12.62
Incremental, \$/million Btu	\$0.49
Desulfurization, \$/ton SO ₂	\$243.24
<i>Annualized Product Costs (Entire Plant)</i>	
Incremental, \$/ton	\$2.58
Incremental, \$/million Btu	\$0.10
Desulfurization, \$/ton SO ₂	\$74.68

Table 6-10 shows that the incremental product cost increases when the natural fines processing is incorporated. This is expected since a fraction of the coal is being processed using a method more costly than the conventional methods employed. Other fines processing methods such as flotation similarly increase product costs. As mentioned earlier, the increase in product cost brings about an increased utilization of a larger fraction of the coal and a corresponding increase in overall yield and plant output.

Additional revenue is generated through recovering natural coal fines discarded as plant refuse. The circuit is expected to generate an additional 378,000 tons of cleaned coal annually with an energy content of 12,809 Btu/lb on a dry oil-free basis. Assuming a price of \$0.80 per million Btu, this will generate an additional \$2.35 million of after-tax profits per year.

This additional profit does not account for cost savings derived from a decrease in refuse handling costs due to reduced refuse generation. Costs associated with refuse

handling include but are not limited to operating costs, capital equipment costs of refuse handling equipment, and ponding capital and operating costs.

Two criteria were examined to evaluate the economic feasibility of a natural fines retrofit circuit, rate of return and payback period. Based on revenue projections, the actual rate of return associated with the addition of the retrofit circuit is 14.5%. The payback period for the project will be 6.5 years after startup. Both criteria indicate that a retrofit project would provide a reasonable return on investment.

Operations at Increased Recovery

As stated in Section 4, the fines circuit is assumed to operate at 85% Btu recovery which corresponds to a circuit yield of 61%. Circuit operations at 90% Btu recovery are possible and were examined to ascertain the effects on overall plant economics.

Circuit operations at the higher recovery level increases profitability while decreasing product quality. Conventional plants may consider operations at higher recovery levels which would give product quality compatible to existing clean coal.

Increasing Btu recovery to 90% will increase circuit yield to 65% while circuit ash and sulfur rejection will fall to 87.7% and 46.0%, respectively. Table 6-11 illustrates the changes in the circuit and plant product. All values are given on a dry-basis. The final circuit product is still of higher quality than the conventionally cleaned product and does not degrade the final plant product. Operation above 90% was not examined since data at this level of operation are limited.

Table 6-11
Energy Recovery Effects on Product Quality

	85% Energy Recovery	90% Energy Recovery
<i>Plant Product</i>		
Ash, %	7.2	7.2
Total sulfur, %	1.51	1.52
Energy content, Btu/lb	13,438	13,431
<i>Circuit Product</i>		
Ash, %	6.4	6.5
Total sulfur, %	1.12	1.39
Energy content, Btu/lb	12,809	12,727

From an economic standpoint, operations at 90% Btu recovery rather than 85% has the net effect of reducing

the annualized product cost for the circuit by 5.0%. The product basis cost is reduced because of improved yield. Costs on a feed basis would essentially remain constant. Table 6-12 illustrates the differences between the circuit recoveries.

Table 6-12
Energy Recovery Effects on Plant Economics

	85% Energy Recovery	90% Energy Recovery
Output, million t/y	6.87	6.90
Additional after-tax profits, million \$/year	2.35	2.58
Payback period, years	6.5	5.9
Rate of return, %	14.5	16.2

Summary

The incremental product cost for the pelletized coal product was determined to be \$24.73/ton and \$25.37/ton for the Upper Freeport and Pittsburgh No. 8 coals, respectively. Additional savings of about 30% result from no size enlargement and thermal drying to 8% moisture. The incremental product cost can be reduced by 45% if no thermal drying is used in the flowsheet.

Sensitivity analyses for the agglomeration plant show that an increase in rate of return from 10% to 20% increases incremental product costs 21%. Two operating hour cases were examined for the agglomeration plant, 7560 and 5400 hours per year. The latter case increased product costs about 17%. It was also shown that reductions of 11% in product cost can be achieved by increasing the plant size to 500 t/h.

The economics of retrofitting a conventional cleaning plant with an agglomeration circuit to clean the natural fines portion of the coal was also presented. The cost of processing the natural fines product was \$12.62 per ton of product from the natural fines circuit or \$243.24 per ton of SO₂ removed. These costs are substantially less than the commercial agglomeration plant processing costs.

The cost of a conventional coal cleaning process incorporating recovery of natural fines using agglomeration was determined to be \$2.58 per ton of product or a \$0.58 per ton increase in overall cost. While the addition of an agglomeration circuit represented an increase in overall processing costs, a 14.5% rate of return is expected from the additional revenues generated making the project a

reasonable investment. Operations at higher recovery rates improve the economics slightly.

Of the applications of agglomeration technology discussed, the best economic results are obtained when it is used to process the natural fines fraction of coal in a conventional cleaning plant. Processing in this manner

yields a reasonable economic rate of return. It also allows utilities to recover coal matter and reject ash and sulfur from the fines stream at higher separation efficiencies than other available methods. Recovery of the fine coal matter also reduces costs associated with disposal of the fines as refuse.

CONCLUSIONS AND RECOMMENDATIONS

Phase I Testing

During the Phase I batch test program, various agglomerating liquids, ranging from heavy oils such as Maya crude to light hydrocarbons like hexane, were evaluated. Initial testing indicated that diesel and heptane were the most appropriate agglomerants for further investigation.

The diesel process tested involved high-shear agglomeration with diesel (approximately 1% on a dry coal basis) followed by two-stage flotation recovery of microagglomerates. The recovered microagglomerates were then either subjected to low-shear agglomeration using diesel/asphalt blends as a binder or mechanically dewatered and pelletized using dry-milled flour as a binder and asphalt for weatherproofing. These enlarged products were then thermally dried to produce the final product. Evaluation of these two product formation methods determined that the pelletization process provided a more handleable product at a lower cost, and it was therefore investigated further.

The heptane process tested utilized heptane (approximately 20-30% on a dry coal basis) during high-shear agglomeration. Following high shear, microagglomerates were recovered by screening. These microagglomerates were then processed through low-shear agglomeration using a diesel/asphalt blend as a binder. Finally, the low-shear product was thermally dried to recover the heptane and produce the final product.

During Phase I batch testing, Btu recoveries of approximately 90% at 75-80% PSR were achieved for all three Phase I test coals for both the diesel- and heptane-based processes on a run-of-mine (ROM) basis. Similar results were also achieved in bench-continuous testing. While both the diesel- and heptane-based processes provided similar performance, the diesel/pelletization process was selected for investigation during Phase II of the project since it was ranked as a more mature process with fewer uncertainties. However, the diesel- and heptane-based

processes both possess inherent advantages and disadvantages.

A primary advantage of the diesel process is a result of the nature and quantity of the agglomerant used. In the diesel or heavy-oil process, relatively small amounts (<1.0%) of diesel oil, a hydrocarbon of low volatility, are used. The diesel does not have to be recovered and remains in the final product. In contrast, the heptane or light-oil process requires significant concentrations (20-30%) of heptane, a highly volatile hydrocarbon, during high shear. As a result, the heptane cannot remain in the final product and must be recovered to meet product safety and environmental standards. Also, when using heptane, the entire process from the point where heptane is introduced during high shear until it is recovered by thermal dryers must be sealed and blanketed with an inert gas as a safety precaution. This required safety measure increases capital investment substantially.

During selective agglomeration, the high-shear unit operation is typically conducted at about 20% solids concentration. In the heptane process, no dilution of the high-shear product is required prior to screening to recover microagglomerates. While some wash water is required to assist in the screening separation, the amount of water needed is small compared to that used during the two stages of flotation to recover microagglomerates in the diesel process. For both stages of flotation in the diesel process, the feed is diluted to approximately 6% solids concentration.

An additional advantage of the light-oil process is that the screening operation is simpler, easier to control, and less expensive than froth flotation recovery of microagglomerates. It is also possible that with flotation the overall performance of the selective agglomeration process may be determined by the microagglomerate recovery step rather than by the high-shear agglomeration unit operation.

During heptane agglomeration, final product formation is achieved via low-shear agglomeration using asphalt-based binders blended with additional heptane. While the asphalt is needed to impart strength to the final product, its presence makes recovery of heptane to the required levels difficult to achieve. In fact, Phase I testing indicated that when asphalt is present in the low-shear product, the quantity of heptane remaining in the final product after thermal recovery is environmentally unacceptable. In addition, the final product from the heptane process is not as strong and handleable as that from the diesel/pelletization process.

Pilot-scale (3-t/h) testing to evaluate the diesel process low-shear agglomeration conditions and agglomerate dewatering methods was carried out during Phase I Component Development Test Facility (CDTF) testing. CDTF testing found that 13% low-shear binder concentration was required to form 1-mm agglomerates with a 100-mesh topsize coal feed. Screen bowl centrifuge dewatering of these agglomerates achieved a product moisture content of about 14%. Similar dewatering performance was also achieved for much smaller agglomerates formed using approximately 6% binder.

A major finding of Phase I testing was that a ROM coal should be cleaned conventionally prior to agglomeration. It was found that process performance was better when the project coals were cleaned prior to agglomeration than when raw coals were agglomerated directly. This conclusion is not unexpected since during conventional precleaning impurities are rejected at coarse sizes by gravity techniques with relatively high separation efficiencies. In contrast, agglomeration—a surface-based process—is carried out at fine coal sizes where separation efficiencies are lower. Processing costs are also lower when precleaned coals are used.

Phase II Testing

The Phase II work primarily involved Proof-of-Concept (POC) testing of the selected diesel agglomeration process at a 2-t/h scale. Batch agglomeration, continuous pelletization, and batch drying tests were also carried out during Phase II. Based on Phase II test results, a commercial-scale (200-t/h) selective agglomeration plant was designed, and the economics of this plant was determined.

POC testing of Upper Freeport and Pittsburgh No. 8 coals was conducted using factorial test plan designs. As such, determination of the effects of individual high-

shear parameters on overall process performance proved to be difficult since the two sequential unit operations, high shear and flotation, are coupled. Flotation appeared to dominate the process because of its sensitivity to process variables such as coal hydrophobicity, agglomerant concentration, and flotation equipment variables such as air flow rates and cell water levels. As a result, the effects of high-shear parameters on process performance were difficult to determine.

The high-shear mixers tested during POC operations were designed to produce intensive microshearing forces on the process slurry. During Phase I CDTF operations, this intense mixing was found to be effective in minimizing the amount of viscous binder required during low-shear agglomeration. However, the POC tests using diesel suggest that less intensive mixing could be utilized with similar results.

POC testing showed that strongly hydrophobic coals, such as Upper Freeport, are sensitive to excess agglomerant addition. Diesel addition levels to high shear of 0.5% and greater destabilized the air bubbles in the froth as the bubbles rose towards the surface of the flotation cell. During typical flotation froth migration, air bubbles coalesce into larger bubbles but are retained in the froth, thus assisting in drainage of process water from the flotation product.

During POC testing with Upper Freeport coal, the destabilization of air bubbles resulted in greater carryover of process water to the product and a lower froth product solids content than desired. The net effect of this phenomenon is increased product contamination by refuse entrained in the process water and increased difficulty in controlling coal recovery to a constant level.

In contrast, testing with coals of decreasing hydrophobicity such as Pittsburgh No. 8, and even more so with Illinois No. 5, did not exhibit this phenomenon. These two coals could be agglomerated over a wide range of oil concentrations and still maintain acceptable froth properties with corresponding improvements in flotation performance. Observations during POC operation indicate that as coal hydrophobicity decreases, the range of agglomerant concentrations providing good froth properties along with good microagglomerate recovery is as follows for each coal:

- Upper Freeport (very hydrophobic) – <0.3%
- Pittsburgh No. 8 (hydrophobic) – 0.125-0.5%
- Illinois No. 5 (weakly hydrophobic) – 0.25-1.2%

These observations indicate that evaluation of coal-specific parameters such as hydrophobicity can help determine the range of high-shear agglomerant concentrations that will provide good froth properties during microagglomerate recovery via flotation.

POC data indicate that lower flotation solids concentrations yield higher pyritic sulfur rejection (PSR). Froth spraying to reduce mineral matter contamination in the flotation product was found to be ineffective for the Upper Freeport coal when tested at a 0.5% high-shear agglomerant level. However, with the proper froth structure, i.e., lower agglomerant levels, froth spraying might prove effective. This would allow a reduction in flotation cell capacity by using higher flotation solids concentrations or allow elimination of the second stage of flotation while achieving similar performance.

Considering the high-shear and flotation unit operations as a single process, agglomerant and frother concentrations in combination control coal recovery. For any given agglomerant concentration there is an optimal frother level that will achieve the desired recovery, providing good froth structure is maintained. If the agglomerant concentration is reduced, coal recovery decreases unless frother levels are increased correspondingly. If the agglomerant concentration is increased beyond the ranges previously discussed, the presence of excess oil adversely affects the froth structure regardless of the frother concentration.

Optimization of a three-component system in series (high shear followed by two stages of flotation) is difficult and requires a large number of tests to complete a factorial design. Sequential design of experiments appears to be a promising approach for optimization of these types of problems. The use of tools such as Ultramax process optimization software allows determination of the effects of several operating parameters simultaneously. This reduces the total number of experiments required to evaluate the effects of all process variables of interest over the desired ranges. For parametric testing of the Illinois No. 5 coal, EPRI made available a pre-release version of the Ultramax software program to assist in the optimization of the selective agglomeration process. This product is under development by EPRI and Ultramax Corporation, an industrial partner.

Selective agglomeration performance achieved during Phase II POC testing for the three primary project coals is characterized by the grade/recovery curves shown in Figure 7-1. Also presented in this figure are the average demonstration run performances achieved for each of these feedstocks.

Figure 7-1. Characterization of Selective Agglomeration POC Performance

As can be seen in Figure 7-1, the average POC demonstration run performance approximates the grade/recovery curve for each coal tested. As such, for the Upper Freeport and Pittsburgh No. 8 coals, the average demonstration run performance was used to complete the commercial-scale (200-t/h) plant design.

For the Illinois No. 5 coal, however, POC results were not used for the commercial plant design since the Illinois coal tested in the POC represented the entire raw coal and the commercial design was completed for the minus 100-mesh natural fines fraction. The results of batch tests using the natural minus 100-mesh coal fraction were used for this design. These batch tests indicated selective agglomeration performance of 79% PSR at 85% Btu recovery. This 15% increase in PSR for batch agglomeration of the minus 100-mesh fines over that achieved during POC testing of the whole raw coal approximates the difference in liberation indicated by washability data for these two Illinois No. 5 coal samples.

The selective agglomeration performance values used for the commercial-scale design are shown in Table 7-1.

**Table 7-1
Selective Agglomeration Performance for
Commercial-Scale Design**

Coal	Topsize	Btu Rec, %	PSR, %
Upper Freeport	200 M	89.7	81.0
Pittsburgh No. 8	200 M	88.2	64.4
Illinois No. 5	100 M (natural fines)	85.0	79.0

In addition to the three primary project coals, laboratory batch and POC tests were conducted using reclaimed pond fines from Jim Walter Resources (JWR) No. 3 mine. In general, good agreement was observed between the batch and POC results. For both the laboratory-batch and POC-continuous testing modes, approximately 30% yield of the minus 60-mesh fraction was achieved while making a 15% ash product. During the POC pond fines testing, these results were achieved using approximately 0.1% of a blend of 90% diesel/10% Betz M210 (flotation collector), both with and without high shear. These data, along with batch testing results, indicate that for this particular pond fines sample under the conditions tested, the effect of high shear on performance was minimal. It should be noted, however, that additional agglomeration testing, particularly with oxidized pond fines samples, may well indicate that selective agglomeration provides better performance than more conventional coal cleaning processes.

Laboratory batch-scale testing was conducted in support of continuous POC tests. Similar performance was achieved with both of these modes of testing for all coals tested during Phase II, provided that comparisons are made at the same feed coal size consist. This similarity between batch and continuous test results indicates that a well-planned series of batch agglomeration tests will provide much information predicting the performance of continuous-scale testing.

During laboratory batch testing it was often difficult to compare results between individual tests because it was hard to terminate each test at the same energy recovery value, i.e., 85% Btu recovery. This problem is overcome by collection of second-stage flotation froth products in a series of timed intervals for each test. The results of various tests can then be interpolated to the same energy recovery values.

The reproducibility of batch tests proved difficult because of test sensitivity to small variations in flotation procedures from test to test, indicative of tests performed by different technicians. The batch flotation procedure is very subjective and dependent on operator observations of the flotation froth. To ensure more consistent batch procedures, particularly between operators, an automated flotation cell could be utilized. This type of cell would include continuous flotation water level control and motorized paddles for froth removal. Use of a cell of this type would ensure more reproducible batch testing results, thereby minimizing the number of duplicate tests required.

Coal topsizes of 100 mesh ($d_{50} = 60$ microns) and 200 mesh ($d_{50} = 30$ microns) were investigated during both POC and batch testing. The results of both modes of testing indicate that size reduction from 100 to 200 mesh provided marginally better PSR at equivalent Btu recoveries. For the Upper Freeport and Pittsburgh No. 8 coals, a 200-mesh topsize was chosen to maximize PSR. For the Illinois No. 5 coal, a 100-mesh size was tested since the primary objective for testing of this coal was to generate data for the agglomeration of the minus 100-mesh natural fines.

In general, agglomeration results do not reach the level of separation efficiency demonstrated by centrifuge sink/float analyses. For both POC and batch testing, agglomeration grade/recovery curves parallel washability curves, but at lower PSR. This indicates that while washability data provide information about the liberation characteristics of a given coal, they cannot be used to predict agglomeration performance. Prediction of continuous-mode agglomeration performance at any grind size requires, at a minimum, that batch agglomeration tests be conducted, the results of which can then be extrapolated using available washability data.

During Phase II testing, it proved difficult to achieve improvements in selective agglomeration performance such that test results more closely approximated washability data. Continued investigation of agglomeration, particularly at lower agglomerant concentrations (about 0.1%), may provide better PSR at equivalent Btu recoveries. In particular, testing to optimize the flotation recovery of microagglomerates at these low agglomerant levels is needed. Agglomeration testing at finer coal grind sizes, where greater liberation is achieved, will also provide improved results but at increased process costs.

All continuous coal grinding during this project was carried out using an open circuit. This type of grinding produces broad size distributions containing significant amounts of both fine and coarse material. In contrast, closed-circuit grinding to produce the same product d_{50} , requires less grinding energy while producing a narrower size distribution with less fine and coarse material. The effect of closed- vs. open-circuit grinding on overall mineral-matter liberation is unknown for the three project coals tested. However, it is possible that closed-circuit grinding may increase mineral-matter liberation such that agglomeration performance improves at the same overall process cost.

Pelletization and drying tests indicate that the following conditions are capable of producing a satisfactory product for conventional coal handling systems:

- Use of a disc pelletizer for size enlargement of the dewatered product to approximately 6 mm
- Optimal moisture for pelletization is approximately 25% and 27% for 100- and 200-mesh topsizes, respectively
- Binder requirements consist of a starch binder at 2-2.5% for pellet strength and an asphalt binder at 0.5% for weatherproofing (required asphalt concentration not optimized)
- Thermal drying in a vibrating fluidized-bed dryer
- Dryer residence time of 2 to 2-1/2 minutes to achieve a moisture content of less than 4%

The pelletization feed moisture content was found to be of primary importance, with excess moisture resulting in large, weak agglomerates, and insufficient moisture reducing the rate of pellet formation. The strength of properly formed "green" (wet) pellets was found to be sufficient to withstand transfer via belt conveyors.

Pellets formed from 200-mesh grind feedstock were generally stronger than those made with the coarser 100-mesh feed at the same binder concentration. The pelletization results also indicate that curing of pellets (long-term storage of several weeks) imparts additional strength although this phenomenon was not quantified.

The characteristics of pelletized and thermally dried POC demonstration run products are presented in Table 7-2.

It can be seen from these data that the final thermally dried agglomerated product is low in both moisture and ash content, indicating potential savings during transportation of this fuel. Additional savings would be realized during power plant pulverization due to the very high grindabilities of these products. This savings would help offset the initial cost of fine grinding prior to agglomeration.

Commercial Design and Economics

The conceptual design completed for the 200-t/h commercial agglomeration plant was based on Phase II test results achieved with the Upper Freeport and

Pittsburgh No. 8 coals. The commercial design was based on average POC demonstration run results of 89.7% Btu recovery and 81.0% PSR for the Upper Freeport coal and 88.2% Btu recovery and 64.4% PSR for the Pittsburgh No. 8 coal. The commercial selective agglomeration plant design includes conventional cleaning of the ROM coal, followed by agglomeration, pelletization, and thermal drying to produce the final product. The plant has been scheduled to operate 24 hours per day, 7 days per week, and 50 weeks per year. To meet this operating schedule of 8400 hours per year, excess capacity has been included in the design where necessary, along with all required spare parts. A conceptual design of the same plant at a 20-t/h semi-works scale was also completed.

Table 7-2
Final Product Analysis (Dry Basis)

	Upper Freeport	Pittsburgh No. 8	Illinois No. 5
<u>Proximate Analysis</u>			
Moisture, % (as is)	0.95	1.90	2.53
Ash, %	5.56	5.28	4.72
Volatile matter, %	33.22	38.74	39.19
Fixed carbon, %	61.22	55.98	56.09
<u>Forms of Sulfur</u>			
Total sulfur, %	1.08	1.75	1.38
Pyritic sulfur, %	0.05	0.60	0.48
Sulfate sulfur, %	0.39	0.12	0.17
Inorganic sulfur, %	0.44	0.72	0.65
Organic sulfur, %	0.64	1.03	0.73
 lb SO ₂ /MBtu	 1.51	 2.46	 2.03
Btu/lb	14274	14209	13618
Hardgrove index	177	197	150

The reliability analysis for both the 200- and 20-t/h plant designs is based on the following criteria:

- Availability Factor (AF): The expected percentage of scheduled operating hours that the plant is in service
- Equivalent Availability Factor (EAF): The expected percentage of the design coal processing rate at which the plant will operate
- Full Load Operation Factor (FLOF): The expected percentage of scheduled operating hours at which the plant operates at the designed capacity

Table 7-3 presents the availability factors and expected operating rates for the conceptual designs of both the 20- and 200-t/h selective agglomeration plants.

Table 7-3
200-t/h Conceptual Design Availability Results

	20 t/h	200 t/h
Availability factor (AF)	95.4%	97.2%
Expected operating hours per year	8010	8164
Equivalent availability factor (EAF)	95.0%	95.2%
Expected coal processing rate, t/h	19	190
Full load operating factor (FLOF)	93.4%	90.0%
Hours per year at design capacity	7843	7475

A separate commercial plant design was executed for the retrofit of a selective agglomeration circuit to process the minus 100-mesh natural fines coal fraction in a 1500-t/h feed plant under construction by AMAX Coal Company. Process performance for this design is based on batch agglomeration results achieved using the minus 100-mesh natural fines fraction of the Illinois No. 5 coal. In this design, the agglomerated natural fines are mechanically dewatered and blended in with the conventionally cleaned coarse coal fractions. The scheduled operating hours for the design of this retrofit circuit is 5400 hours per year.

For the purposes of reliability analyses on this plant, the EAF of the conventional coal cleaning portion of this plant was assumed to be 90%. Using this assumption and reliability analysis of the natural fines agglomeration circuit, the reliability results for this commercial-scale (1500-t/h feed) plant design are presented in Table 7-4.

Table 7-4
Natural Fines Commercial Plant Reliability

Equipment	EAF	Product t/h
Conventional coal cleaning only	90.0%	1083
Agglomeration process only	98.5%	69
Overall plant	90.5%	1152

Economic analyses of the commercial-scale agglomeration plants were based on the assumptions presented in Table 7-5. Using these assumptions and the actual plant design, the annualized product costs for the Upper Freeport and Pittsburgh No. 8 coals are presented in Table 7-6. These cost represent total plant processing costs including capital investment, operation and maintenance costs, and Btu losses incurred during processing. However, these annualized product costs do

not include the cost of the ROM coal. As such, these costs represent the total cost incurred over direct sale of the ROM coal. In this table, costs are presented for a pelletized and thermally dried product of 4% moisture, a nonpelletized product thermally dried to 8% moisture, and a mechanically dewatered product with no additional processing of about 25% moisture.

Table 7-5
Commercial Plant (200 t/h) Economic Assumptions

Actual operating hours	7,560
Plant availability	90%
Production rate	200 t/h product
Plant life	20 years
Royalties	None
Financing	100% equity
Inflation rate	4.0%
Corporate tax rate	38.0%
Rate of return	10.0%

Table 7-6
Commercial Agglomeration Annualized Product Cost Summary

	Upper Freeport	Pittsburgh No. 8
Pelletized Product, 4% Moisture		
\$/ton product	24.73	25.32
\$/million Btu	0.84	0.87
\$/ton SO ₂ removed	333.56	488.08
Nonpelletized Product, 8% Moisture		
\$/ton product	17.70	18.28
\$/million Btu	0.60	0.63
\$/ton SO ₂ removed	238.66	352.41
Nonpelletized Product, 25% Moisture		
\$/ton product	14.08	14.72
\$/million Btu	0.48	0.51
\$/ton SO ₂ removed	189.95	283.79

Sensitivity analyses varying the rate of return, operating hours, and plant size were performed for the commercial plant. These analyses reveal that:

- An increase in the rate of return from 10% to 20% increases product costs by 20%.
- A decrease in the actual annual operating hours from 7560 to 5400 increases product costs by 16%.
- An increase in plant size from 200 t/h to 500 t/h decreases product costs by 10%.

The cost of cleaning minus 100-mesh natural fines using an agglomeration circuit retrofit in a conventional cleaning plant is shown in Table 7-7. The natural fines circuit costs represent total processing costs including capital investment and operation and maintenance costs. Added costs for Btu losses incurred during natural fines processing are not included since this circuit is a retrofit application to a plant which bypasses the fines to tailings. Annualized product costs for the entire plant are based on natural fines circuit product costs and an assumed value of \$2/ton product for the plant conventional cleaning circuits.

The retrofit selective agglomeration fines recovery circuit produces approximately 378,000 tons/year of product generating additional revenue. Assuming an energy value of \$0.80 per million Btu, an after-tax profit of \$2.35 million per year is realized. This annual profit does not include credit for refuse handling cost savings such as

reduced capital equipment investment, reduced operating costs, and lower ponding capital and operating costs.

Table 7-7
Natural Fines Annualized Product Cost
Summary

Fines Circuit	
\$/ton product	\$12.62
\$/million Btu	\$0.49
\$/ton SO ₂ removed	\$243.24
Entire Plant	
\$/ton product	\$2.58
\$/million /Btu	\$0.10
\$/ton SO ₂ removed	\$74.68

Based on these revenue projections, the rate of return for the retrofit selective agglomeration fines circuit is 14.5% with a payback period of 6.5 years from startup.

Appendix A

UPPER FREEPORT PRECLEANING REPORT

REPORT MEMORANDUM

To Southern Clean Fuels
Southern Company Services

From Doug McCollough

Date May 1, 1992

Subject Precleaning of the Upper Freeport Seam Coal
92F0166

OVERVIEW

The goal of this project was to produce a precleaned coal for Southern Company Services to be used in its test work on selective oil agglomeration. This document addresses the discrepancies between the proposed flowsheet and the actually run.

The precleaned coal produced for this project was to have a high heating value-- approximately 95 percent energy recovery. The raw coal for this project was Upper Freeport Seam coal, locally know as the Sheesley "E" Seam coal. This coal from P&N Mining Company's Sheesley Mine near Punxsutawney, Pennsylvania, was purchased by Southern Clean Fuels through IUE Coal Sales Company. A heavy-media cyclone and high recovery water-only cyclone/spiral concentrator circuit was used to preclean the coal. The heavy-media cyclone was operated at a 1.70 circulating gravity. CQ Inc. shipped 377 tons of precleaned coal from 485 tons of raw coal received. In addition there were approximately one ton of coal samples and five tons of precleaned coal left that could not be loaded onto the last truck.

PROCEDURES AND RESULTS

Over a six-day period beginning on January 17, 1992, 485 tons of raw coal from P&N Mining Company's Sheesley Mine were received at CQ Inc.'s Coal Quality Development Center (CQDC) near Homer City, Pennsylvania. A sample of the first truckload of coal received was taken to assure Southern Company Services that the coal being shipped was of the quality ordered. The quality of the coal received was satisfactory to Southern Company Services. Also a 468-pound representative sample of the complete shipment was taken before crushing to obtain a whole coal analysis of the coal received. The as-received raw coal analysis is shown in Table 1.

Table 1. As-received Raw Coal Analysis. (Dry Basis, Except Moisture)

Quality (Dry Basis)	First Truck Load	Complete Shipment
Ash (Wt %)	13.93	13.67
Moisture (as-received)	5.61	7.16
Total Sulfur (Wt %)	2.93	2.95

Flowsheet Setup

On January 21, 1992, CQ Inc. technicians began to process the coal. On the first day of processing a flowsheet setup test was performed to get a feel of the flowsheet operation, flag any potential problems, and determine the expected theoretical yield. At the request of Southern Company Services no flocculants were used in the thickener to aid in solids settling. It was thought that flocculants might hinder the oil agglomeration process to be performed on this coal. As a precaution one slight change in the proposed flowsheet was made in case flocculants had to be used to settle solids in the thickener to recover fine coal. The rapped sieve underflow that was to be sent to the thickener was thickened instead in four-in. thickening cyclones with the underflow; the rapped sieve overflow reported to a screen-bowl centrifuge for dewatering. The thickening cyclone overflow reported to the thickener. If flocculants had to be used this small size fraction of material (minus 325 mesh) would not report as product but would be rejected with the refuse. Pump gland water, deslime sprays, and heavy-media rinse sprays contained solids but these did not cause any operational problems. In the end, no flocculants were required and the flowsheet remained as changed from the setup test. During mass production, solids in the recirculating process water would have to be addressed. Continued operation with solids in the process water would be detrimental to the maintenance and life of equipment.

Only one minor problem occurred during the initial setup run: a residual freeze conditioning chemical in the plant water caused foaming in the plant. To resolve this problem all the plant water was drained to ponds, the plant was filled with fresh water, and the plant makeup water was valved to the fresh water well.

Two truck loads of precleaned coal from the setup test were shipped to Southern Company Services for use in a grinding circuit setup.

Precleaning Production

CQ Inc. technicians proceeded to clean the Upper Freeport Seam coal using the following flowsheet. To prepare precleaned coal, the Upper Freeport Seam coal was crushed to 3/4-in. in a two-stage four-roll crusher. It was then conveyed to 100-ton bins for storage prior to cleaning. From there the coal was fed to the processing plant at a rate of 17-20 tph and screened on the deslime screen at 28 mesh. The plus 28 mesh coal was fed to a 14-in. heavy-media cyclone. A specific gravity of 1.70 was maintained in the circulating media of the heavy-media cyclone circuit. The precleaned coal and the refuse extracted by the heavy-media cyclone were drained and rinsed of magnetite. The precleaned coal was dewatered further in a CMI basket centrifuge. The refuse extracted by the heavy-media cyclone was discarded directly to refuse. The heavy-media cyclone geometry is shown in Table 2.

The setup test demonstrated that the precleaning could proceed.

Table 2. Heavy-Media Cyclone Geometry

Body Diameter (in.)	14
Orifice Diameters:	
Feed (in.)	4 x 2-3/4
Overflow (in.)	6
Apex (in.)	5
Cone Angle (°)	20
Mounting Angle (°)	10

Note: The cyclone was made by Roberts & Schaefer Company, Inc., under an original Dutch State Mines license.

The minus 28 mesh material removed from the plant feed by a deslime screen was processed in a two-stage, water-only cyclone/spiral concentrator recycle circuit. This circuit consisted of a 10-in. water-only cyclone and a five-turn, single start spiral concentrator. The cyclone and concentrator were used in a two-stage configuration: the underflow from the water-only cyclone was recleaned in the spiral concentrator and product from the spiral concentrator, containing misplaced coal from the primary cyclone, was recycled to the primary cyclone for recleaning. The primary cyclone overflow is the product from this circuit. The water-only cyclone geometry and spiral concentrator configuration are shown in Table 3.

Table 3. Water-only Cyclone Geometry and Spiral Concentrator Configuration

Cyclone Geometry

Body Diameter (in.)	10
Orifice Diameters:	
Feed (in.)	4
Overflow (in.)	4
Apex (in.)	1-1/4
Vortex Finder (in.)	11-1/2
Cone Angle (°)	75

Spiral Configuration

Number of Turns	5
Number of Starts	1

Note: The primary cyclone was made by Krebs Engineers. The spiral concentrator was made by Mineral Deposits Ltd.

To remove the water from the water-only cyclone/spiral concentrator product, a rapped sieve, thickening cyclones, and static thickener were employed. The precleaned coal from this circuit was first sized on a rapped sieve to remove the plus 100 mesh material. Secondly, the underflow from the rapped sieve was thickened in four-in. cyclones to remove the plus 325 mesh material. Thirdly, the minus 325 mesh material and the remaining water reported to the static thickener. The rapped sieve overflow (plus 100 mesh), thickening cyclone underflow (100 mesh x 325 mesh), and the thickener underflow (minus 325 mesh) all reported to a Bird Machine Company screen-bowl centrifuge for final dewatering. Refuse produced from the water-only cyclone/spiral concentrator was collected in a storage tank and dewatered in a batch operation after the coal was processed.

Sampling

Samples were collected to determine overall flowsheet yield and energy recovery by Southern Company Services representatives. During setup approximately 50 sample increments were collected at five-minute intervals at each of nine sample points. During the precleaning operation approximately 50 sample increments were collected at 30-minute intervals at each sample point. The weight of each sample is included in the laboratory analysis data included in the attachment to this report. Sampling points during the setup test and precleaning production are as indicated in the precleaning flowsheet.

Sampling Points

Originally there were eight sampling points proposed. CQ Inc. engineers added one additional sampling point, the screen-bowl centrifuge effluent, during the final dewatering. This sample was taken to determine the quality of the very fine material that more than likely was not recovered in the process. A flowrate of this effluent is not a normal measurement recorded during plant operation and therefore was not taken during this operation. If this flowrate had been available it would have been helpful in determining the amount of material not recoverable.

During the sampling, the as-received sample and the plant clean coal sample collection containers were located outdoors. Special care was taken to prevent moisture from entering these samples: a plastic sheet was draped around the sample discharge point and out over the sample collection drum. If collection occurred over several days the drums were tightly covered overnight. Some moisture may nevertheless have entered these samples. The composite as-received sample indicated a slightly higher moisture content (7.16 percent) than the composite plant feed sample (5.27 percent). The plant clean coal sample moisture content was also slightly higher (9.20 percent versus 7.90 percent) than the pile sample taken at completion of the work.

Sample Analysis

Each sample was analyzed for moisture or percent solids, ash, sulfur forms, and calorific value (Btu/lb). The gross weight of each sample was also reported. In addition, the plant feed sample and deslime screen overflow and underflow samples were sized at 16 mesh, 28 mesh, and 100 mesh to determine the quantity of coal processed in the heavy-media cyclone and the water-only cyclone/spiral concentrator circuits. The laboratory analyses for the setup test and the production run are included in the supporting data.

Coal Storage

Two truck loads (45 tons) of precleaned coal from the setup test were shipped to Southern Company Services. This precleaned coal was for setup of the grinding circuit at Southern Company Services. The precleaned coal from the actual production run (331 tons) was stored for Southern Company Services on paved ground area and covered with tarp and plastic to reduce the rate of weathering. Precleaned coal was loaded on trucks arranged by Southern Clean Fuels periodically over two months time with the last truck being loaded on April 9, 1992. Approximately five tons remained at CQDC because it could not be loaded on the last truck without overloading the vehicle. This remaining five tons of precleaned coal was discarded because Southern Company Services felt that it might be different enough in quality from the next scheduled production of precleaned coal to cause different results in the selective oil agglomeration process.

OVERALL PRECLEANING RESULTS

The results of the overall precleaning operation are shown in Table 4, which includes the laboratory analyses of the raw coal and clean coal produced. The two refuse streams from the process were not combined and therefore there is no composite analysis of the refuse.

The performance reported in Table 4 was determined using the following approach. First, from the screen analysis of the plant feed sample and estimations of the plus one-in. oversize, the percentage of the plant feed to the heavy-media circuit (83 percent), water-only/spiral concentrator cyclone circuit (15 percent), and plus one-in. (two percent) oversize were determined. The two percent plus one-inch material was reject from the cleaning circuit to prevent plant pipes from plugging. Next the yields of the heavy-media cyclone cleaning circuit and the water-only cyclone/spiral concentrator circuit were determined using ash balance equations. The heavy-media cyclone circuit produced a yield of 95 percent and the water-only cyclone/spiral concentrator circuit produced a 97 percent yield.

The total plant yield (93 percent) was determined by applying each circuit's yield to the percentage of feed to each circuit and then adding them together.

Table 4. Overall Precleaning Data. (Dry Basis, Except Moisture)

Conditions			
Average Feed Rate (wet t/h)		17	
Coal Size		3/4-in x 0	
Performance			
Yield (Wt %)		93	
Energy Recovery (%)		98	
Ash Reduction (Wt %)		30	
Total Sulfur Reduction (Wt %)		40	
Pyritic Sulfur Reduction (Wt %)		54	
		Clean	
Quality (Dry Basis)	Feed	Coal	Refuse *
Ash (Wt %)	14.36	10.07	62.0
Moisture (as-received)	5.27	9.20	N/D
Total Sulfur (Wt %)	3.04	1.81	16.72
Pyritic Sulfur (Wt %)	2.34	1.07	16.46
Heating Value (Btu/lb)	13,146	13,908	4,672
SO ₂ Emission Potential (lb/MBtu)	4.62	2.60	9.34

* Calculated values

During the precleaning operation CQ Inc. plant operators monitored the plant operating variables such as plant feed rate, heavy-media circulating gravity, process pump flow rates and pressures, and thickener pH. Plant operating data in the attached supporting data are shown in two forms: one is a computer printout, the other is a hand written form filled out by CQ plant operators.

A total of 377 tons of precleaned coal was shipped to Southern Clean Fuels. Along with the estimated five tons discarded at the CQDC, a total of 382 tons of precleaned coal was produced from the 485 tons of raw coal shipped to CQ Inc.

REPORT MEMORANDUM

To	Southern Clean Fuels Southern Company Services
From	Doug McCollough
Date	June 22, 1992
Subject	Precleaning Upper Freeport Seam Coal Second Batch (92F0166-02)

OVERVIEW

The goal of this project was to produce a second batch of precleaned coal for Southern Company Services to be used in its test work on selective oil agglomeration. The first batch (377 tons) was delivered between January 21, 1992, and April 9, 1992; the second batch (346 tons) was delivered between April 27, 1992, and June 5, 1992. This document addresses the operational differences between the first and second batches. Also included in an attachment are the laboratory and operational data for this work.

Both precleaned coals produced for this project were to have a high heating value--approximately 95 percent energy recovery. The raw coal for both batches was the Upper Freeport Seam coal, locally known as the Sheesley "E" Seam coal. This coal from P&N Mining Company's Sheesley Mine near Punxsutawney, Pennsylvania, was purchased by Southern Clean Fuels through IUE Coal Sales Company. Both batches of precleaned coal were produced utilizing heavy-media cyclone and high recovery water-only cyclone/spiral concentrator circuits. The heavy-media cyclone was operated at a 1.70 circulating gravity in both precleaning batches. For the second batch, 376 tons were produced from 452 tons of raw coal received. From the 376 tons of precleaned coal produced, 346 tons were shipped to Southern Company Services for test work on selective agglomeration.

PROCEDURES AND RESULTS

Over a three-day period beginning on April 21, 1992, CQ Inc. received 452 tons of raw coal from P&N Mining Company's Sheesley Mine at CQ Inc.'s Coal Quality Development Center (CQDC) near Homer City, Pennsylvania. A two-ton representative sample of the complete shipment was taken before crushing to obtain a whole coal analysis of the coal received and to retain material for possible laboratory middlings washability testing at a later time. The as-received raw coal analysis is shown in Table 1.

Table 1. As-received Raw Coal Analysis. (Dry Basis, Except Moisture)

Quality (Dry Basis)	Complete Shipment
Ash (Wt %)	13.44
Moisture (as-received)	4.07
Total Sulfur (Wt %)	2.66

Precleaning Production

On April 21, 1992, CQ Inc. technicians began to process the Upper Freeport Seam coal using the following flowsheet. Because both the first and second batches were produced from the same coal source, there were no setup tests performed for the second batch. There was only one flowsheet change between batches: for the second batch, heavy-media refuse dewatering in a CMI basket centrifuge was added. This centrifuge was not available for use when the first batch was produced.

To prepare the coal for precleaning, CQ Inc. technicians crushed the Upper Freeport Seam coal to 3/4-in. in a two-stage four-roll crusher. The crushed coal was then conveyed to 100-ton bins for storage prior to cleaning. From there the coal was fed to the processing plant at a rate of 15 tph and sized on the deslime screen at 28 mesh. The plus 28 mesh coal was fed to a 14-in. heavy-media cyclone. A specific gravity of 1.70 was maintained in the circulating media of the heavy-media cyclone circuit. The precleaned coal and the refuse extracted by the heavy-media cyclone were drained and rinsed of magnetite. The precleaned coal and refuse were dewatered further in CMI basket centrifuges. The heavy-media cyclone geometry is shown in Table 2.

Table 2. Heavy-Media Cyclone Geometry

Body Diameter (in.)	14
Orifice Diameters:	
Feed (in.)	4 x 2-3/4
Overflow (in.)	6
Apex (in.)	5
Cone Angle (°)	20
Mounting Angle (°)	10

Note: The cyclone was made by Roberts & Schaefer Company, Inc., under an original Dutch State Mines license.

The minus 28 mesh material removed from the plant feed by the deslime screen was processed in a two-stage, water-only cyclone/spiral concentrator recycle circuit, shown in Table 3. The water-only cyclone geometry and spiral concentrator configuration remained as it was in the first batch of precleaning. This circuit consisted of a 10-in. water-only cyclone and a five-turn, single-start spiral concentrator. The water-only cyclone and spiral concentrator were used in a two-stage configuration: the underflow from the water-only cyclone was recleaned in the spiral concentrator and product from the spiral concentrator, containing misplaced coal from the water-only cyclone, was recycled to the water-only cyclone for recleaning. The water-only cyclone overflow is the product from this circuit. As with the first batch, operators removed the water from the water-only cyclone/spiral concentrator product using a rapped sieve, thickening cyclones, and static thickener.

The precleaned coal from this circuit was sized on a rapped sieve to remove the plus 100 mesh material. Secondly, the underflow from the rapped sieve was thickened in four-in. cyclones to remove the plus 325 mesh material. Thirdly, the minus 325 mesh material and the remaining water reported to the static thickener. The rapped sieve overflow (plus 100 mesh), thickening cyclone underflow (100 mesh x 325 mesh), and the thickener underflow (minus 325 mesh) all reported to a Bird Machine Company screen-bowl centrifuge for final dewatering.

Table 3. Water-only Cyclone Geometry and Spiral Concentrator Configuration

Cyclone Geometry	
Body Diameter (in.)	10
Orifice Diameters:	
Feed (in.)	4
Overflow (in.)	4
Apex (in.)	1-1/4
Vortex Finder (in.)	11-1/2
Cone Angle (°)	75
Spiral Configuration	
Number of Turns	5
Number of Starts	1

Note: The primary cyclone was made by Krebs Engineers. The spiral concentrator was made by Mineral Deposits Ltd.

Refuse produced from the water-only cyclone/spiral concentrator was collected in a storage tank and dewatered in a batch operation after the coal was processed.

Sampling

Samples were collected by automatic samplers and CQ Inc. operators. These samples were then analyzed by a commercial laboratory so that Southern Company Services could determine overall clean coal yield and energy recovery. During the precleaning operation approximately 50 sample increments were collected at 30-minute intervals at each of nine sampling points. Sampling points and types of samples taken during the precleaning production are as indicated in the precleaning flowsheet.

Sampling Points

The location of one sampling point was changed from the first batch to obtain a sample that was more representative of the minus 28 mesh portion of the clean coal. A sample of clean coal from the water-only cyclone/spiral concentrator circuit was taken after the coal was dewatered in a screen-bowl centrifuge. As in the first batch of precleaning, the screen-bowl centrifuge effluent was sampled during the final dewatering. This sample was taken to determine the quality of the very fine material (minus 100 mesh) that more than likely was not recovered in the process. A flowrate of this effluent is not normally measured and recorded during plant operation and therefore was not taken during this operation. If this flowrate had been available it would have been helpful in determining the amount of material not recoverable.

Sample Analysis

Each sample was analyzed for moisture or percent solids, ash, sulfur forms, and calorific value (Btu/lb). The gross weight of each sample was also reported. In addition, the plant feed sample and deslime screen overflow and underflow samples were sized to determine the quantity of coal processed in the heavy-media cyclone and the water-only cyclone/spiral concentrator circuits. In addition, the clean coal product was sized to help determine the yield produced in each circuit.

Coal Storage

As was the first batch, the precleaned coal from the second batch was stored as the first batch for Southern Company Services on paved ground and covered with tarp and plastic to reduce the rate of weathering. Precleaned coal was periodically loaded into trucks arranged by Southern Clean Fuels over a six-week period, with the last truck being loaded on June 5, 1992. The dates and quantities of coal shipped are shown in Table 4.

Table 4. Coal Shipments. (As-received Basis)

Date Shipped	Number of Trucks	Total Tons
April 27, 1992	1	22.13
April 29, 1992	1	21.385
April 30, 1992	1	21.45
May 20, 1992	2	48.68
May 22, 1992	1	25.47
June 1, 1992	2	45.58
June 2, 1992	2	44.34
June 3, 1992	1	24.70
June 4, 1992	3	67.23
June 5, 1992	<u>1</u>	<u>24.98</u>
TOTAL	15	345.945

Approximately 30 tons of precleaned coal remain at CQ Inc. as this coal was not needed for Southern Company Services's test program. This remaining coal will be discarded at Southern Company Services's direction.

OVERALL PRECLEANING RESULTS

The results of the overall precleaning operation are shown in Table 5, which includes the laboratory analyses of both the raw coal and clean coal. The two refuse streams from the two cleaning circuits and the plus one-in. deslime oversize were not combined and therefore there is no composite analysis of the refuse.

The performance reported in Table 5 was determined using the following approach. First, from the screen analysis of the plant feed sample and percent weight of the plus one-in. oversize, the percentages of the plant feed to the heavy-media circuit (82 percent), water-only/spiral concentrator cyclone circuit (18 percent), and plus one-in. (0.1 percent) oversize were determined. The 0.1 percent plus one-in. material was reject from the cleaning circuit to prevent plant pipes from plugging. Next the yields of the heavy-media cyclone cleaning circuit and the water-only cyclone/spiral concentrator circuit were determined using ash balance equations. The heavy-media cyclone circuit produced a 92 percent yield and the water-only cyclone/spiral concentrator circuit produced a 99 percent yield. The total plant yield (92 percent) was determined by applying each circuit's yield to the percentage of feed to each circuit and then adding them together.

Table 5. Overall Precleaning Data. (Dry Basis, Except Moisture)

Conditions			
Average Feed Rate (wet t/h)		15	
Coal Size		3/4-in x 0	
Performance			
Yield (Wt %)		92	
Energy Recovery (%)		97	
Ash Reduction (Wt %)		29	
Total Sulfur Reduction (Wt %)		41	
Pyritic Sulfur Reduction (Wt %)		55	
Quality (Dry Basis)		Clean	Refuse *
Ash (Wt %)	Feed	Coal	
	13.64	9.70	61
Moisture (as-received)	5.19	6.00	N/D
Total Sulfur (Wt %)	2.83	1.67	16.8
Pyritic Sulfur (Wt %)	1.96	0.89	14.8
Heating Value (Btu/lb)	13,300	13,984	5,078
SO ₂ Emission Potential (lb/MBtu)	4.26	2.39	66.1

* Calculated values

During the precleaning operation CQ Inc. plant operators monitored the plant operating variables such as plant feedrate, heavy-media circulating gravity, process pump flow rates and pressures, and thickener pH.

A total of 346 tons of precleaned coal was shipped to Southern Clean Fuels. Along with the estimated 30 tons stored at the CQ Inc., a total of 376 tons of precleaned coal was produced from the 452 tons of raw coal shipped to CQ Inc.

Appendix B

PITTSBURGH NO. 8 MIDDLINGS WASHABILITY

**ENGINEERING DEVELOPMENT OF
SELECTIVE AGGLOMERATION**

**REPORT ON WASHABILITY OF
PITTSBURGH No. 8 COAL
(GREENE COUNTY, PA)**

**CONTRACT No.
DE-AC22-89PC88879**

JUNE 1992

PREPARED FOR:

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**Engineering Development of Selective Agglomeration
DE-AC22-89PC88879**

PITTSBURGH No. 8 WASHABILITY ANALYSIS

1.0 INTRODUCTION

The POC test plan submitted to DOE calls for the use of three coals during POC operations, one of which is to be a middlings coal. In view of the economic attractiveness of middlings agglomeration and the potential of many Appalachian coals to be used in such applications, it was decided to test a middlings coal sample from the Pittsburgh No. 8 or Upper Freeport seams in the POC.

Several issues influence the selection of a coal to be used to generate a middlings sample for agglomeration in this project, including its amenability to the process scheme, its economic attractiveness, and the cost to the project of preparing a middlings sample. Previous project reports prepared on this subject indicated that Upper Freeport was a good candidate for middlings agglomeration due to its high liberation, the considerable amount of middlings that can be generated, and overall enhanced performance. However, a comparable estimate was not available for Pittsburgh No. 8 coal—especially that obtained from Greene County, Pennsylvania, which has a low organic sulfur content—since projections of middlings agglomeration were made based on existing data in the absence of middlings washability data.

It was therefore decided to conduct middlings washability testing of Pittsburgh No. 8 coal to obtain a better estimate of the overall suitability of this fraction for agglomeration. A 1-ton raw coal sample of Pittsburgh No. 8 coal was procured from Emerald mines, and middlings washability testing was carried out at Geochemical Testing laboratories in Somerset, Pennsylvania, in accordance with a test plan which is discussed below.

2.0 MIDDLINGS WASHABILITY TEST OBJECTIVES

The objectives of the middlings washability test plan were to :

- Evaluate the potential for middlings agglomeration of Pittsburgh No. 8 coal
- Determine the liberation characteristics of the middlings fraction
- Determine the amount of middlings that can be generated from this coal
- Use the washability data to select the coal types and quantities for middlings agglomeration runs in the POC

Two coals are to be used for this testing and evaluation: Pittsburgh No. 8 coal from Emerald mines in Greene County, Pennsylvania, and Upper Freeport coal from Sheesly mine in Jefferson County, Pennsylvania, with initial tests focusing on Pittsburgh No. 8 coal.

3.0 MIDDLEINGS WASHABILITY TEST PLAN

3.1 Test Plan Philosophy

The middlings washability test plan has been developed such that data from the washability tests can be used to calculate the amount of middlings that can be generated for each coal and estimate the middlings liberation characteristics as well as predict middlings agglomeration performance. This necessitates a fairly comprehensive set of tests. In view of time and cost constraints, however, a phased testing approach has been adopted to minimize the number of tests required to obtain the necessary information.

This work will start with Pittsburgh No. 8 coal. If it is determined upon analysis of the resulting data that the Pittsburgh coal is not amenable to middlings agglomeration, then Upper Freeport coal will not be tested and will be used to generate middlings samples for the POC work.

Since the testing must be carried out interactively, the test plan has been developed accordingly. Data from the washability tests must be obtained and analyzed in order to allow further testing to proceed in the right direction. As discussed in the next section, the test procedure used for the 28 x 100-mesh size fraction will depend both on the amount of material as a proportion of the whole and on the relative concentration of pyrite in that fraction. Therefore, test data need to be provided to the project team as testing progresses so that decisions can be made about the eventual processing of this fraction.

Most of the testing will be carried out at 1-1/2" topsize. This will help estimate the liberation of coal at the size at which the Emerald plant cleans their coal. However, it is also realized that middlings yield and liberation may be different when the coal is crushed to 3/4", as will be done when the large middlings sample for POC operation will be generated at CQ, Inc. Therefore, a summary washability test at 3/4" topsize is also included.

The testing will be carried out in two phases. The first phase will produce the required washability data on the size fractions and gravities detailed in this test plan. The second phase will involve sample generation using gravities determined from analysis of results from the first series of tests. The second phase of testing will be shorter than the first series of tests.

3.2 Data Requirements

All laboratory data and analyses must be reported to the project team to conduct analyses and make recommendations both on the future course of testing and on selection of the coal for agglomeration testing. Also, samples from sink-float testing must be saved by the laboratory performing the work since they will be used for batch agglomeration testing at Wilsonville and Alberta Research Council (ARC).

3.3 Test Procedure

Figures 1 (a) and (b) present block diagrams showing the washability test procedure for the two coals. A 2-ton sample of raw coal will be procured from the mines for each coal. Of this, a 1000-lb subsample will be extracted for test purposes.

The following steps will be performed to carry out the proposed work in the first phase:

1. The entire coal sample will be crushed to 1-1/2" nominal topline using a jaw crusher. A 200-lb subsample will be split from this sample of 1000 lb and crushed to 3/4" nominal topline. A sample will be taken from this fraction (3/4" x 0) and analyzed for moisture, ash, total sulfur, forms of sulfur, and calorific value. This will constitute the head sample for the analysis.
2. The sample crushed to 3/4" topline will then be screened at 28 mesh. The +28-mesh sample will be subjected to sink-float testing according to ASTM D4371 at the following specific gravities: 1.3, 1.35, 1.4, 1.45, 1.6, 1.7, and 1.8. Each gravity fraction will be dried, weighed, and analyzed for ash, total sulfur, pyritic sulfur, and calorific value.

The minus 28-mesh sample from the 3/4" topline material will be wet- screened at 100 mesh. The 28 x 100 mesh and 100 mesh x 0 fractions will then be weighed and analyzed for ash, total sulfur, pyritic sulfur, and calorific value. These data must be reported by the laboratory to the project team as soon as they become available. All samples from the screened fractions and sink-float gravity fractions will be saved by the laboratory for possible future testing.

Figure 1 (b). Middlings Washability Scheme for Pittsburgh and Upper Freeport Coals

3. The sample crushed to 1-1/2" topsize in Step 1 will be screened at 28 mesh. The +28 mesh fraction will be split to produce 4 subsamples, of which one subsample will be stored in the laboratory as a save sample. This sample needs to be stored carefully in the laboratory since it will be used to generate middlings samples later in the program.

The processing of the minus 28-mesh sample is described in Step 6.

4. The other three subsamples of +28 mesh topsize from Step 3 will be homogenized. Washability testing will be carried out on this sample according to ASTM D4371 at the following gravities: 1.3, 1.35, 1.4, 1.45, 1.6, 1.7, and 1.8. Each individual gravity fraction will be dried to remove the organic liquids and weighed. The sink 1.8 fraction will be analyzed for ash, total sulfur, pyritic sulfur, and calorific value. These data must be reported to the project team as soon as possible.

Based on the amount of material recovered in each gravity fraction, splits of the remaining gravity fractions (other than the sink 1.80) will be ground to achieve 90% passing 200 mesh. The gravity fraction that constitutes the greatest proportion of the total coal, which is expected to be one of float 1.3, 1.3-1.35 or 1.35-1.4 for the Pittsburgh coal, will be ground initially to achieve the desired size. Samples will be taken during the grinding step and screened to determine the percent passing 200 mesh. Once the desired size is obtained, the time required for grinding will be noted for this fraction. Other gravity fractions will be ground for the same length of time, since this is representative of a real-life situation where the entire middlings fraction will be ground for a specified length of time.

Samples ground to 200 mesh will then be sampled for analysis of ash, total sulfur, pyritic sulfur, and calorific value according to ASTM procedures.

5. The 200-mesh topsize material from all gravity fractions (except for the sink 1.80 fraction) will be subjected to centrifuge sink-float testing at 1.6 and 1.8 specific gravities. The centrifuge sink-float tests will be carried out soon after the grinding is performed. A copy of the suggested centrifuge sink-float procedure is attached in Appendix A. The three gravity fractions arising from each sink-float test, i.e. float 1.6, 1.6-1.8, and sink 1.8, will be dried and weighed. Each fraction will be sampled for ash, sulfur, pyritic sulfur, and calorific value analysis. The remaining sample for each gravity fraction will be saved at the laboratory since these will be shipped to Wilsonville and ARC for batch agglomeration tests after analysis of the test data.
6. The minus 28-mesh fraction from 1-1/2" topsize (Step 3) will be sampled for screening at 100 mesh. The remaining material will be saved at the laboratory. Each size fraction from the screening, i.e., 28 x 100 mesh and 100 mesh x 0, will be analyzed for ash, total sulfur, pyritic sulfur, and calorific value. **This analysis, as well as the relative weights of the size fractions, will be reported by the laboratory to the project team as soon as possible since the test procedure will differ based on this analysis, as outlined below. Thus, the following work will not be carried out until the project team has analyzed the data and recommended a course of action.**

If the weight and/or pyritic sulfur content of the 28 x 100-mesh fraction relative to those of the entire coal are significant, i.e., above 10%, washability testing will be performed on this fraction. If the amount of 28 x 100-mesh material and/or the

relative pyritic sulfur content is not significant, the minus 28-mesh save sample will be split into two subsamples. One subsample will be saved, while the other will be split again into two subsamples. Of this, one subsample will be subjected to sink-float testing at 1.3, 1.4, 1.6, and 1.8 gravities, after being reduced to 200-mesh topsize. The other subsample will be subjected to sink-float testing at 1.3, 1.4, 1.6, and 1.8 specific gravities. Each fraction will be analyzed for ash, total sulfur, pyritic sulfur, and calorific value.

In the event that the relative pyritic sulfur content is greater than 10% in the 28 x 100-mesh fraction, a washability test will be carried out to estimate the liberation of this fraction. The 28 x 100-mesh fraction will initially be subjected to sink-float testing at 1.6 and 1.8 specific gravities. Based on the results of this analysis, i.e., the pyrite content of the float 1.6 or 1.8 fraction, an alternative course of action will be taken. If either of the float fractions meets the project's standards for cleanability, no further work will be performed. However, if the float fractions contain a significant amount of pyrite, the float 1.6 or 1.8 (depending on the amount of material in the respective fractions) will be ground to 200-mesh topsize. The minus 200-mesh material will then be subjected to centrifuge sink-float testing at the following gravities: 1.3, 1.4, 1.6, and 1.8. All samples will be saved at the laboratory.

In addition, if the relative pyrite content in the 28 x 100-mesh fraction is significant, the 100 mesh x 0 fraction from the screening step in Step 5 will be subjected to sink-float testing at the same gravities, i.e., 1.3, 1.4, 1.6, and 1.8. The gravity fractions will be weighed and analyzed.

Based on analysis of the data from these tests, a gravity cutpoint will be recommended to generate middlings samples for agglomeration. Selection of the gravity cutpoint will take into consideration the capability to produce a significant fraction of middlings as well as amenability to optimization of the overall process performance. The 1-1/2" x 28-mesh and 28-mesh x 0 save samples described in Steps 4 and 6 will be used in this phase of testing. Each of these samples will be subjected to sink-float testing at two gravities to produce the middlings fraction. The middlings thus produced will be analyzed and then shipped to Wilsonville and ARC for batch agglomeration tests.

4.0 ANALYSIS OF PITTSBURGH NO. 8 MIDDLEINGS

Upon completion of the middlings washability test work on Pittsburgh No. 8 coal, an analysis was carried out to evaluate the potential for middlings agglomeration of Pittsburgh No. 8 and Upper Freeport coals, taking into account the potential performance of the two coals as well as costs. The washability work was carried out only on the Pittsburgh No. 8 coal; the performance of the Upper Freeport coal was projected based on existing washability data—which do not include middlings washability—and is similar to that reported in middlings washability reports in November 1991.

Data from the washability tests are attached in Appendix B. No specific middlings washability tests were conducted on Upper Freeport coal; its performance was projected from the available washability data from Phase I. However, to compare both coals on the same basis, the process scheme was maintained the same. Data for comparison of the costs to generate samples of each coal were obtained from Mr. W. T. Henshaw's letters dated March 18 and 24, 1992.

Evaluation of the washability data reveals that the raw Pittsburgh No. 8 coal sampled at Emerald mines contains a low amount of middlings—i.e., coal in the 1.35-1.80 specific gravity range—as a fraction of the total coal. This is partly due to the high level of ash in the coal (32%). In addition, a significant fraction of the coal floats at 1.35 specific gravity. The middlings fraction of the coal (1.35-1.80 specific gravity fraction) amounts to about 13% at 1½" x 28M and about 17% at ¾" x 28M. The small amount of middlings in the coal may not make it particularly attractive for a middlings agglomeration process scheme.

The agglomeration feed to the POC will consist of the minus 28-mesh fraction as well as the coarse middlings fraction. It is assumed in this case that the 28 x 100-mesh fraction of the Pittsburgh No. 8 raw coal will not be precleaned since it represents a relatively small proportion of the total ROM coal and contains only 6% of the total pyrite. For the Pittsburgh No. 8 coal, the total feed to the agglomeration circuit is estimated to be about 25-30% of the total coal. In contrast, the amount of middlings that could be generated from the Upper Freeport coal is estimated to be about 30-35% of the total coal at ¾" topsize, with the total feed to the POC (including the minus 28-mesh fraction) amounting to about 40%.

The liberation potential of the Pittsburgh No. 8 coal is significant, as indicated by the middlings washability data. When ground to 200-mesh topsize, the middlings fraction achieves about 85-88% PSR at 1.8 specific gravity. However, due partly to the inefficiency of agglomeration in separating minus 200-mesh particles, an overall process PSR of about 64% and Btu recovery of 94% are achieved, assuming that the precleaning and deep cleaning are carried out at 1½" topsize as is likely in a commercial situation. This estimate is based on the assumption that 80% of the liberated pyrite particles at 200 mesh will be rejected by agglomeration. This estimate of agglomeration separation efficiency is based on project agglomeration data and has been assumed to be the same for both coals. While agglomeration tests have not been carried out on the middlings samples, it is felt that the efficiency of separation via agglomeration may be lower for the Pittsburgh No. 8 coal as it usually contains a significant amount of fine pyrite.

The estimated performance of the two coals for the overall process involving middlings agglomeration as well as for the individual processes is presented in Table 1. Btu recovery and PSR values presented for

each individual process are based on that particular process; for example, PSR for agglomeration is the percentage of pyritic sulfur rejected based on the agglomeration feed rather than on the total coal. As seen from this table, the estimated performance for the two coals is close, with Upper Freeport providing a slightly higher PSR. However, in the absence of detailed middlings washability data, projections of the Upper Freeport middlings performance cannot be made as confidently as those of the Pittsburgh No. 8 coal. The coal quality at each stage of the process is presented in Table 2. As seen from this table, the final product quality is better for the Upper Freeport coal and its SO₂ emissions output is 1.75 lb SO₂/MMBtu compared to 2.5 lb SO₂/MMBtu for the Pittsburgh No. 8 coal.

Table 1
MIDDLEINGS PROCESS PERFORMANCE
FOR PITTSBURGH No. 8 AND UPPER FREEPORT COALS

Process Performance:	Pittsburgh No. 8	U. Freeport
Precleaning:		
- Btu Recovery, %	96.0	98.0
- PSR, %	39.9	25.4
Deep Cleaning:		
- Btu Recovery, %	77.6	56.0
- PSR, %	55.0	71.2
Agglomeration:		
- Btu Recovery, %	92.3	94.0
- PSR, %	67.0	73.0
Overall Performance:		
- Btu Recovery, %	94.0	96.0
- PSR, %	64.0	69.0

Table 2
PRODUCT QUALITY

Category	Pittsburgh No. 8	U. Freeport
Raw Coal Quality:		
- Ash	32.2	16.2
- Total Sulfur	2.12	2.18
- Pyritic Sulfur	1.46	1.56
- Btu/lb	9950	12714
- lb SO ₂ /MMBtu	4.26	3.43
Deep Cleaned Coal Quality:		
- Ash	5.3	6.7
- Total Sulfur	1.7	1.32
- Pyritic Sulfur	0.74	0.64
- Btu/lb	14385	15230
Agglomeration Feed Quality (incl. -28M)		
- Ash	21.3	19.7
- Total Sulfur	3.1	2.36
- Pyritic Sulfur	2.4	1.76
- Btu/lb	11290	11400
Agglomeration Product Quality:		
- Ash	13.6	12.5
- Total Sulfur	1.90	1.22
- Pyritic Sulfur	1.00	0.58
- Btu/lb	13500	13450
Final Combined Product Quality:		
- Ash	7.9	9.5
- Total Sulfur	1.76	1.26
- Pyritic Sulfur	0.82	0.60
- Btu/lb	14112	14370
- lb SO ₂ /MMBtu	2.49	1.75

4.1 Economic Estimates

For purposes of comparison, a factored cost estimate has been developed for the middlings agglomeration scenario for the two coals. This estimate is based on the following unit cost figures (levelized capital and operating costs):

- Precleaning and deep cleaning: \$5.50/ton raw coal processed
- Agglomeration without size enlargement: \$11.00/ton coal processed

These costs are approximate, and are the same as those used in previous middlings washability reports. Table 3 presents the cost estimates for the entire process (i.e., starting from raw coal to final product) in terms of \$/ton of raw coal processed, \$/ton of clean coal processed, \$/MMBtu clean coal, and \$/ton SO₂ removed. As is expected, the processing cost of Pittsburgh No. 8 coal is lower than that of Upper Freeport coal since a much higher fraction of the total coal is processed in the agglomeration step for the Upper

Freeport coal. As the agglomeration step costs significantly more than conventional cleaning, the costs of processing a larger amount of coal make the Upper Freeport coal processing costs higher. However, while the difference in processing costs is large when compared on the basis of \$/ton raw coal processed, it narrows considerably when compared on the bases of \$/ton clean coal and \$/MMBtu clean coal, respectively.

Table 3
COST ESTIMATES FOR MIDDLEINGS AGGLOMERATION

Cost Basis	Pittsburgh No. 8	Upper Freeport
\$/ton raw coal processed	\$7.81	\$10.91
\$/ton clean coal processed	\$11.79	\$12.79
\$/MMBtu clean coal	\$0.42	\$0.45
\$/ton SO ₂ removed	\$475	\$536

4.2 Costs of POC Sample Generation

Due to project budget constraints, the costs of POC middlings sample generation for the two coals were compared. In the case of the Upper Freeport coal, only raw coal can be procured from the mine and both precleaning and deep cleaning must be carried out at CQ, Inc. However, in the case of the Pittsburgh No. 8 coal, precleaned coal can be purchased from Emerald mines. As expected, procurement of precleaned coal provides a significantly cheaper option. The relative costs are compared in Table 4.

Table 4
SAMPLE GENERATION COST COMPARISON
MIDDLEINGS SAMPLE OF 500 TONS

Cost Category	Pittsburgh No. 8		Upper Freeport
	Raw Coal Purchase	Clean Coal Purchase	Raw Coal Purchase
Amt. of purchased coal	1,800 t	1,400 t	1,250 t
Cost of coal	\$45,000 (@ \$25/t)	\$39,200 (@ \$28/t)	\$38,125 (@ \$30.5/t)
Transportation cost: mine to CQ	\$18,000 (@\$10/t)	\$14,000 (@\$10/t)	Included in coal cost
Cleaning cost	\$104,400 (@ \$58/t)	\$50,400 (@ \$36/t)	\$72,500 (@ \$58/t)
Transportation cost: CQ to SCS	\$30,000 (@ \$60/t)	\$30,000 (@ \$60/t)	\$30,000 (@ \$60/t)
Total Cost	\$197,400	\$133,600	\$140,625

From the above discussion, it is apparent that while Upper Freeport coal provides better performance in terms of Btu recovery and PSR, it entails a higher cost for middlings agglomeration. In addition, the costs of generating a middlings sample are slightly higher for the Upper Freeport coal. The estimated costs of middlings agglomeration are higher for Upper Freeport coal primarily because it produces a significant amount of middlings, while the Pittsburgh No. 8 coal has a small middlings fraction. The smaller middlings fraction in the Pittsburgh No. 8 coal may render it less attractive for a middlings agglomeration process scheme. In this context, other issues such as utility interest in the two coals must also be factored into the evaluation. If utilities are indeed more interested in Pittsburgh No. 8 coal, then, in the absence of a strong

technical and economic incentive to use Upper Freeport coal, the Pittsburgh No. 8 coal appears to be more suitable in spite of its low middlings generation potential and inferior performance. On the other hand, if the project objective is to demonstrate the concept of middlings agglomeration, it would be more appropriate to select the Upper Freeport coal in view of its projected superior performance and the fact that it contains a significant amount of middlings.

APPENDIX A
DOE/PETC CENTRIFUGAL FLOAT-SINK EXPERIMENTAL PROCEDURE (12/89)

APPENDIX A

DOE/PETC Centrifugal Float-Sink Experimental Procedure (12/89)

1. Dry the appropriate number of 40-gram samples in a convection oven, which is set at 100° C, for 2 hours to achieve a moisture content of essentially 0%. Then place them in a desiccator.
2. Place the required number of 40-gram samples of the dried coal in enough empty 500 ml centrifuge cups to make up the proper total starting sample weight.
3. Make a solution of 10 g of Aerosol OT-100 per 500 ml of the proper organic heavy liquid. Add 20 ml of this solution to the coal in the cup (0.4 g of actual Aerosol OT-100) to obtain a dosage of 20 lb Aerosol OT-100 per ton of coal. Make sure the coal is wetted.
4. Add organic heavy liquid (i.e., the proper amount of Certigrav) at the proper specific gravity to reach about the 350-ml level in the cup to obtain about 5 to 10 percent of solids.
5. Mix this sample with a mechanical stirrer for the proper conditioning time of about 30 seconds.
6. Wash the sample down into the cup with the heavy organic liquid set at the separating specific gravity to reach a level of 400 ml in the cup.
7. Ultrasonically condition the sample for the desired amount of time of 2 minutes, at a power setting of 100 amperes. The temperature of the sample should be monitored since it increases during conditioning. Mix the sample for 1 minute using a magnetic stirrer and cool in an ice bath to ensure the sample is uniformly at room temperature prior to centrifugation.
8. Centrifuge the sample for 20 minutes at 2850 rpm at room temperature to minimize specific gravity fluctuation.
9. Allow the centrifuge to come to a stop on its own momentum because braking disturbs the separation.
10. The float products are removed, filtered, dried, weighed, and chemically analyzed for ash, pyritic and total sulfur, and Btu/lb.
11. Filter the sink material, and then using media of the next higher gravity, repeat the procedure above from step 4 through step 10 until the desired specific gravity separations are completed.
12. The results are tabulated and a washability report is generated showing direct and cumulative data by specific gravity increment.

APPENDIX B
PITTSBURGH No. 8 MIDDINGS WASHABILITY DATA

APPENDIX B

PITTSBURGH No. 8 COAL MIDDLINGS WASHABILITY

Table B-1
1-½" x 0 Analysis

	Wt, %	Ash, %	Tot Sulfur, %	Pyr Sulfur, %	Btu/lb
1-½" x 28M	89.70	32.71	2.07	1.39	9883
28M x 100M	5.56	21.30	2.53	2.17	11643
100M x 0	4.74	35.58	2.48	1.85	9239
Composite	100.0	32.21	2.12	1.46	9950

Table B-2
¾" x 0 Analysis

	Wt, %	Ash, %	Tot Sulfur, %	Pyr Sulfur, %	Btu/lb
¾" x 28M	87.80	33.39	2.13	1.49	9723
28M x 100M	6.40	22.48	2.38	1.77	11439
100M x 0	5.80	36.50	2.42	1.91	9117
Composite	100.0	32.87	2.16	1.53	9798

WASHABILITY: RAW COAL
SEAM: PITTSBURGH No. 8
GREENE COUNTY, PA

Table B-3

1-1/2" X 28M COMPOSITE

SPECIFIC GRAVITY		(%)	(%)	(%)	(%)		(%)	# SO2/	%PYR SUL
SINK	FLOAT	WT%	ASH	TOT SUL	PYR SUL	Btu	Btu REC	M Btu	REJECTED
	1.30	34.3	4.14	1.37	0.44	14567	50.55	1.88	89.16
1.30	1.35	53.8	5.30	1.70	0.74	14384	78.36	2.37	71.33
1.35	1.40	60.1	6.00	1.86	0.89	14270	86.73	2.60	61.44
1.40	1.45	62.3	6.40	1.93	0.96	14204	89.55	2.71	56.82
1.45	1.60	66.0	7.41	2.05	1.10	14035	93.65	2.92	48.09
1.60	1.70	67.4	7.99	2.12	1.17	13935	95.03	3.04	43.37
1.70	1.80	68.5	8.55	2.17	1.22	13837	95.83	3.13	39.89
1.80		100.0	32.71	2.07	1.39	9883	100.00	4.18	0.00

Table B-4

SAMPLE CRUSHED TO 3/4" TOPSIZE
3/4" X 28M COMPOSITE

SPECIFIC GRAVITY		(%)	(%)	(%)	(%)		(%)	# SO2/	%PYR SUL
SINK	FLOAT	WT%	ASH	TOT SUL	PYR SUL	Btu	Btu REC	M Btu	REJECTED
	1.30	29.7	3.82	1.20	0.33	14653	44.73	1.64	93.42
1.30	1.35	48.0	4.87	1.50	0.60	14475	71.49	2.07	80.72
1.35	1.40	57.0	5.72	1.71	0.80	14327	83.99	2.39	69.43
1.40	1.45	59.9	6.19	1.81	0.90	14250	87.83	2.53	63.82
1.45	1.60	64.6	7.21	1.97	1.08	14081	93.56	2.80	52.99
1.60	1.70	66.1	7.82	2.03	1.16	13979	95.07	2.90	48.67
1.70	1.80	67.2	8.33	2.08	1.21	13891	96.01	2.99	45.20
1.80		100.0	33.39	2.13	1.49	9723	100.00	4.38	0.00

Table B-5

MIDDLEINGS LIBERATION DATA
1-1/2" X 28M GRAVITY FRACTIONS GROUND TO 200 M

Specific Gravity:	Wt %	Ash %	TS %	PS %	Btu/lb	Btu Rec %	PSR %
Gravity Fraction: 1.30 Float Wt. %=34.3							
-1.60	95.75	2.95	1.04	0.09	14807	96.83	78.77
1.60 x 1.80	99	3.25	1.07	0.13	14754	99.76	67.88
+1.80	100	3.81	1.35	0.41	14641	100.00	0.00
Gravity Fraction: 1.30 x 1.35 Wt. %=19.54							
-1.60	92.85	4.73	1.47	0.42	14371	95.08	62.04
1.60 x 1.80	97.1	5.15	1.50	0.45	14327	99.13	57.90
+1.80	100	6.83	2.10	1.03	14034	100.00	0.00
Gravity Fraction: 1.35 x 1.40 Wt. %=6.23							
-1.60	90.46	6.77	1.48	0.38	14010	96.28	83.07
1.60 x 1.80	92.71	7.34	1.52	0.43	13915	98.00	80.48
+1.80	100	11.48	3.14	2.03	13163	100.00	0.00
Gravity Fraction: 1.40 x 1.45 Wt. %=2.24							
-1.60	83.64	8.49	1.54	.044	13884	92.74	87.22
1.60 x 1.80	88.33	9.81	1.61	0.52	13659	96.35	83.96
+1.80	100	16.06	3.95	2.88	12522	100.00	0.00
Gravity Fraction: 1.45 x 1.60 Wt. %= 3.64							
-1.60	72.64	10.79	1.47	0.50	13502	88.15	88.90
1.60 x 1.80	80.57	13.38	1.54	0.61	13068	94.63	85.02
+1.80	100	23.60	4.18	3.27	11127	100.00	-0.00
Gravity Fraction: 1.60 x 1.70 Wt. %= 1.45							
-1.60	53.16	12.29	1.68	0.64	13354	75.48	92.47
1.60 x 1.80	65.5	16.89	1.82	0.85	12572	87.56	87.72
+1.80	100	33.81	5.44	4.52	9405	100.00	0.00
Gravity Fraction: 1.70 x 1.80 Wt. %= 1.05							
-1.60	33.72	12.61	1.37	0.55	13244	59.92	96.31
1.60 x 1.80	47.71	18.20	1.49	0.74	12287	78.66	92.99
+1.80	100	42.57	5.49	5.02	7453	100.00	0.00

Appendix C

POC TEST COAL WASHABILITIES

3 of 3

Table C-1
Precleaned Upper Freeport Coal Washability Data
100 mesh x 0

Specific Gravity		Elementary Data					Computed Data: Cumulative Recovery						
Sink	Float	Wt %	Ash %	TS %	PS %	Btu/lb	Wt %	Ash %	TS %	PS %	Btu/lb	Btu Rec %	PSR %
	1.30	44.71	2.03	0.68	0.05	15131	44.7	2.03	0.68	0.05	15131	49.02	97.73
1.30	1.40	28.09	5.78	0.85	0.22	14403	72.8	3.48	0.75	0.12	14850	78.34	91.45
1.40	1.50	14.77	11.59	1.06	0.48	13434	87.6	4.85	0.80	0.18	14611	92.72	84.24
1.50	1.60	2.87	21.45	1.48	0.94	11686	90.4	5.37	0.82	0.20	14518	95.15	81.50
1.60	1.70	3.86	26.78	1.74	1.19	10718	94.3	6.25	0.86	0.24	14363	98.15	76.83
1.70	1.80	0.99	40.36	2.33	1.99	8476	95.3	6.60	0.87	0.26	14302	98.75	74.83
1.80		4.71	65.1	16.38	15.63	3650	100.0	9.36	1.60	0.98	13800	100.00	0.00

Table C-2
Precleaned Upper Freeport Coal Washability Data
200 mesh x 0

Specific Gravity		Elementary Data					Computed Data: Cumulative Recovery						
Sink	Float	Wt %	Ash %	TS %	PS %	Btu/lb	Wt %	Ash %	TS %	PS %	Btu/lb	Btu Rec %	PSR %
	1.30	35.84	1.69	0.66	0.05	15227	35.8	1.69	0.66	0.05	15227	39.55	98.16
1.30	1.40	27.87	4.87	0.74	0.14	14651	63.7	3.08	0.69	0.09	14975	69.15	94.15
1.40	1.50	20.30	7.87	0.86	0.23	14260	84.0	4.24	0.73	0.12	14802	90.13	89.36
1.50	1.60	7.15	18.06	1.25	0.64	12287	91.2	5.32	0.78	0.16	14605	96.50	84.66
1.60	1.70	1.94	32.23	1.71	1.24	9889	93.1	5.88	0.79	0.19	14507	97.89	82.19
1.70	1.80	1.03	41.64	2.07	1.67	8329	94.1	6.27	0.81	0.20	14439	98.51	80.42
1.80		5.87	66.62	14.04	13.34	3504	100.0	9.82	1.59	0.97	13797	100.00	0.00

Table C-3
Precleaned Pittsburgh No. 8 Coal Washability Data
 100 mesh x 0

Specific Gravity		Elementary Data					Computed Data: Cumulative Recovery						
Sink	Float	Wt %	Ash %	TS %	PS %	Btu/lb	Wt %	Ash %	TS %	PS %	Btu	Btu Rec %	PSR %
	1.30	42.14	2.25	1.10	0.09	14848	42.1	2.25	1.10	0.09	14848	45.10	96.71
1.30	1.40	38.00	5.56	1.42	0.36	14187	80.1	3.82	1.25	0.22	14535	83.95	84.85
1.40	1.50	11.15	10.29	1.76	0.75	13212	91.3	4.61	1.31	0.28	14373	94.57	77.60
1.50	1.60	3.92	17.94	2.47	1.47	12052	95.2	5.16	1.36	0.33	14277	97.97	72.61
1.60	1.70	1.05	29.91	4.28	3.40	10100	96.3	5.43	1.39	0.37	14232	98.74	69.51
1.70	1.80	0.43	35.28	6.24	5.18	9022	96.7	5.56	1.41	0.39	14209	99.02	67.58
1.80		3.31	60.49	24.40	23.55	4119	100.0	7.38	2.18	1.15	13875	100.00	0.00

Table C-4
Precleaned Pittsburgh No. 8 Coal Washability Data
 200 mesh x 0

Specific Gravity		Elementary Data					Computed Data: Cumulative Recovery						
Sink	Float	Wt %	Ash %	TS %	PS %	Btu/lb	Wt %	Ash %	TS %	PS %	Btu/lb	Btu Rec %	PSR %
	1.30	29.85	1.81	1.01	0.06	14771	29.9	1.81	1.01	0.06	14771	31.93	98.23
1.30	1.40	39.95	4.50	1.22	0.21	14241	69.8	3.35	1.13	0.15	14468	73.14	89.96
1.40	1.50	19.69	8.07	1.40	0.40	13754	89.5	4.39	1.19	0.20	14311	92.75	82.19
1.50	1.60	5.46	14.62	1.97	0.91	12599	95.0	4.98	1.23	0.24	14212	97.73	77.29
1.60	1.70	1.40	27.34	3.25	2.22	10355	96.4	5.30	1.26	0.27	14156	98.78	74.23
1.70	1.80	0.65	36.18	4.78	3.61	8927	97.0	5.51	1.29	0.29	14121	99.20	71.52
1.80		3.00	63.60	25.16	24.31	3668	100.0	7.25	2.00	1.01	13808	100.00	0.00

Table C-5
Illinois No. 5 Coal Washability Data
100 mesh x 0

Specific Gravity		Elementary Data					Computed Data: Cumulative Recovery						
Sink	Float	Wt %	Ash %	TS %	PS %	Btu/lb	Wt %	Ash %	TS %	PS %	Btu/lb	Btu Rec %	PSR %
	1.30	6.73	1.48	0.91	0.13	14105	6.7	1.48	0.91	0.13	14105	7.93	99.23
1.30	1.40	29.29	3.77	1.00	0.25	13940	36.0	3.34	0.98	0.23	13971	42.06	92.82
1.40	1.50	30.24	6.95	1.02	0.35	13556	66.3	4.99	1.00	0.28	13782	76.32	83.55
1.50	1.60	13.18	11.48	1.18	0.54	12850	79.4	6.07	1.03	0.33	13627	90.47	77.31
1.60	1.70	4.53	22.19	1.62	1.11	10946	84.0	6.94	1.06	0.37	13482	94.62	72.91
1.70	1.80	2.58	32.23	1.84	1.29	9279	86.6	7.69	1.08	0.40	13357	96.62	69.99
1.80		13.45	72.41	5.94	5.94	3007	100.0	16.39	1.74	1.14	11965	100.00	0.00

Table C-6
Illinois No. 5 Coal Washability Data
200 mesh x 0

Specific Gravity		Elementary Data					Computed Data: Cumulative Recovery						
Sink	Float	Wt %	Ash %	TS %	PS %	Btu/lb	Wt %	Ash %	TS %	PS %	Btu/lb	Btu Rec %	PSR %
	1.30	1.05	1.58	0.99	0.11	14289	1.1	1.58	0.99	0.11	14289	1.30	99.90
1.30	1.40	16.99	3.32	0.98	0.19	13753	18.0	3.22	0.98	0.19	13784	21.50	97.21
1.40	1.50	20.03	8.32	1.11	0.35	13002	38.1	5.90	1.05	0.27	13373	44.02	91.35
1.50	1.60	41.17	10.79	1.18	0.38	12593	79.2	8.44	1.12	0.33	12968	88.85	78.27
1.60	1.70	5.48	16.98	1.67	0.90	11721	84.7	8.99	1.15	0.37	12887	94.40	74.15
1.70	1.80	0.91	27.33	2.17	1.50	9997	85.6	9.19	1.16	0.38	12856	95.19	73.01
1.80		14.37	66.70	6.17	6.08	3872	100.0	17.45	1.88	1.20	11565	100.00	0.00

Table C-7
Illinois No. 5 Coal Washability Data
1" x 0 Composite

Specific Gravity		Elementary Data					Computed Data: Cumulative Recovery						
Sink	Float	Wt %	Ash %	TS %	PS %	Btu/lb	Wt %	Ash %	TS %	PS %	Btu/lb	Btu Rec %	PSR %
	1.30	57.66	4.56	1.23	0.51	13944	57.7	4.56	1.23	0.51	13944	68.19	76.58
1.30	1.40	14.86	7.64	1.74	0.94	13286	72.5	5.19	1.34	0.60	13809	84.93	65.52
1.40	1.50	6.53	10.05	1.84	1.10	13004	79.1	5.59	1.38	0.64	13743	92.14	59.86
1.50	1.60	3.61	15.18	1.96	1.28	12172	82.7	6.01	1.40	0.67	13674	95.87	56.19
1.60	1.70	1.22	25.38	3.15	2.50	10571	83.9	6.29	1.43	0.70	13629	96.96	53.79
1.70	1.80	1.21	47.17	2.87	2.28	7058	85.1	6.88	1.45	0.72	13535	97.69	51.60
1.80		14.89	80.95	4.62	4.38	1832	100.0	17.91	1.92	1.26	11792	100.00	0.00

Table C-8
Illinois No. 5 Coal Washability Data
1" x 1/4" Size Fraction: 25.56% of Sample

Specific Gravity		Elementary Data					Computed Data: Cumulative Recovery						
Sink	Float	Wt %	Ash %	TS %	PS %	Btu/lb	Wt %	Ash %	TS %	PS %	Btu/lb	Btu Rec %	PSR %
	1.30	61	4.59	1.22	0.46	13920	61.0	4.59	1.22	0.46	13920	72.57	78.37
1.30	1.40	18.8	10.2	2.19	1.31	13001	79.8	5.92	1.45	0.66	13704	93.40	59.44
1.40	1.50	3.39	20.2	4.08	3.15	11386	83.2	6.50	1.56	0.76	13609	96.70	51.21
1.50	1.60	1.03	29.3	4.73	3.95	9953	84.2	6.78	1.59	0.80	13565	97.57	48.07
1.60	1.70	0.49	36.4	6.37	5.49	8727	84.7	6.95	1.62	0.83	13537	97.94	46.00
1.70	1.80	0.58	37.6	6.11	5.49	8606	85.3	7.16	1.65	0.86	13503	98.37	43.54
1.80		14.8	85.7	3.83	3.83	1297	100.0	18.74	1.97	1.30	11703	100.00	0.00

Table C-9
Illinois No. 5 Coal Washability Data
1/4" x 28M Size Fraction: 51.45% of Sample

Specific Gravity		Elementary Data					Computed Data: Cumulative Recovery						
Sink	Float	Wt %	Ash %	TS %	PS %	Btu/lb	Wt %	Ash %	TS %	PS %	Btu/lb	Btu Rec %	PSR %
	1.30	77.5	4.72	1.25	0.55	13946	77.5	4.72	1.25	0.55	13946	88.00	65.72
1.30	1.40	6.69	12.4	2.43	1.63	12687	84.2	5.33	1.34	0.64	13846	94.91	56.96
1.40	1.50	2.06	20.4	3.44	2.58	11385	86.3	5.69	1.39	0.68	13787	96.82	52.68
1.50	1.60	1.07	27.8	4.56	3.77	10019	87.3	5.96	1.43	0.72	13741	97.69	49.44
1.60	1.70	0.71	36.2	5.38	4.46	8777	88.0	6.20	1.46	0.75	13701	98.20	46.89
1.70	1.80	0.47	40.9	5.87	5.03	7975	88.5	6.39	1.49	0.77	13671	98.50	44.99
1.80		11.5	82.5	4.92	4.87	1602	100.0	15.13	1.88	1.24	12284	100.00	0.00

Table C-10
Illinois No. 5 Coal Washability Data
28M x 100M Size Fraction: 10.47% of Sample

Specific Gravity		Elementary Data					Computed Data: Cumulative Recovery						
Sink	Float	Wt %	Ash %	TS %	PS %	Btu/lb	Wt %	Ash %	TS %	PS %	Btu/lb	Btu Rec %	PSR %
	1.30	20.29	1.41	0.99	0.24	14074	20.3	1.41	0.99	0.24	14074	23.64	96.61
1.30	1.40	43.29	3.30	1.14	0.40	13785	63.6	2.70	1.09	0.35	13877	73.05	84.53
1.40	1.50	16.69	7.07	1.43	0.74	13340	80.3	3.61	1.16	0.43	13766	91.49	75.92
1.50	1.60	4.42	17.35	2.51	1.82	11604	84.7	4.32	1.23	0.50	13653	95.73	70.32
1.60	1.70	1.63	27.61	3.39	2.77	9894	86.3	4.76	1.27	0.55	13582	97.07	67.17
1.70	1.80	0.82	35.33	3.79	3.17	8776	87.1	5.05	1.30	0.57	13537	97.67	65.36
1.80		12.86	74.50	8.15	7.29	2192	100.0	13.98	2.18	1.43	12078	100.00	0.00

Table C-11
Illinois No. 5 Coal Washability Data
100M x 0 Size Fraction: 12.54% of Sample

Specific Gravity		Elementary Data					Computed Data: Cumulative Recovery						
Sink	Float	Wt %	Ash %	TS %	PS %	Btu/lb	Wt %	Ash %	TS %	PS %	Btu/lb	Btu Rec %	PSR %
	1.30	0.60	1.97	0.89	0.06	14349	0.6	1.97	0.89	0.06	14349	0.89	99.97
1.30	1.40	16.75	3.39	0.90	0.14	13844	17.4	3.34	0.90	0.14	13861	24.74	97.92
1.40	1.50	22.82	4.97	0.82	0.14	13888	40.2	4.27	0.85	0.14	13877	57.35	95.12
1.50	1.60	18.64	10.18	0.93	0.29	13042	58.8	6.14	0.88	0.19	13612	82.37	90.39
1.60	1.70	4.45	15.14	0.89	0.46	12367	63.3	6.77	0.88	0.21	13524	88.03	88.60
1.70	1.80	5.88	52.54	1.12	0.63	6246	69.1	10.67	0.90	0.24	12905	91.81	85.35
1.80		30.86	76.25	3.72	3.16	2580	100.0	30.91	1.77	1.14	9719	100.00	0.00

Appendix D

POC OPERATING LOG

Upper Freeport Operating Log

December 21 - January 3

The flotation tanks arrived and were installed. Piping from these tanks to the pump suctions were completed. The flotation cell additions arrived on site, and preparation of surfaces for installing the additions began. Electrical work for the flotation cells and other equipment continued. Control valves, orifices, transmitters and associated piping modifications were completed for water addition to the ball mills and No. 1 Flotation Cells. Work continued on the systems feeding water to the secondary mill tank and No. 2 Flotation Cells. Fabrication of filter press chutes was completed.

January 6

New flotation cell additions and end boxes were installed. An orifice was installed in the secondary mill tank water inlet piping. Pump suctions were cleaned, and the plate and frame filter presses cycled to check operation.

January 7

The secondary ball mill was flushed with water in preparation for removing the mill head and a portion of the balls. Dirt was moved from the coal storage pad to the product pad area in construction of ramps at that pad. The thickener and secondary ball mill fittings were greased. Work continued with installation of a cover at the 24 ton storage bin. A mount for a new vibrator for that tank was fabricated. Electrical work continued. Instrumentation technicians worked on the filter presses and the primary mill pump variable speed drive. Clyde worked on the wiring for the Reliance variable speed controller for the small high shear mixer. Wiring for the new Flotation Tank Mixers continued. Water piping to the No. 2 Flotation Cells and No. 1 Flotation Tank continued.

January 8

The head was removed from the secondary ball mill so the operators could remove 3500 lbs of the ball charge in preparation for running Upper Freeport coal at 2 tph. About half of the required volume of balls were removed. Flow control valves were installed in the ball mill and secondary mill tank water feed piping. The cover for the 24 ton storage tank was completed. The filter press optical portion of one of the light curtains was sent to the manufacturer for repair.

January 9

Instrumentation technicians began connecting transmitters and running tubing to the flow control orifices. Ball charge removal from the secondary mill was completed.

January 10

Piping from the No. 1 Flotation Pump to the No. 2 Flotation Cells was completed. The secondary ball mill head was replaced. Instrumentation work continued on the FCV's, tubing, etc.

January 13

Electrical work continued. A bracket was made for support of the FCV box at the flotation cell deck. Instrumentation work continued.

January 14

A training session was held for operators. The process was described and the POC test plan reviewed.

January 15

Instrumentation work continued for FCV's and panel. Flotation Tank mixers with shafts and impellers were installed. Installation of new filtrate pump and piping continued. Electrical work continued.

January 16

Installation of the filtrate pump and piping continued. Modifications were made to the water feed piping downstream of all FCV's to allow calibration of the valves to drums. Work continued on fabrication of ball mill feed piping with block valves. Instrumentation and electrical work continued.

January 20

Most roads were closed due to snow and ice, so very little work was done. Some instrumentation work was done in the shop, and freeze damage repair was begun. Installation of the second vibrator for the 24 ton storage bin was completed. Piping for calibration of flow control valves was completed.

January 21

The Thickener was drained to allow repair of High Pressure Water Pump and Thickener Bottoms Pump suction valves, which had been damaged by freezing. The Thickener pipe to the Filter Press Feed Tank was repaired where it had split.

January 22

Instrumentation work continued on the FCV control panel to get the system operational. Work continued on freeze damage repair. One truckload of precleaned Upper Freeport coal from CQ, Inc. arrived at about 21:00, and was placed on the pad. The larger capacity filtrate pump was set in place, and electrical hooked up. Instrument air piping was installed to the flotation cell level controllers.

January 23

The Flotation Cell launders arrived and were installed. The recirculation piping for the Thickener Bottoms and High Pressure Water pumps was completed. The valves were replaced in the water supply system. Bearings were replaced in three flotation cell motor blowers on the south side, the first two cells. The blowers were checked for proper rotation. Instrumentation work at the FCV control panel continued. The vendor made a service call to the plant. The I to P's were replaced in the panel, and replacement controllers were ordered for the two flotation cell water feed valves. The Flotation Cell level control valves were connected.

January 24

Installation of the mechanical samplers began.

January 28

Instrument technicians worked with the weigh belt feeder and got it working in the automatic mode of operation. Both coal conveyors were realigned.

January 29

Conducted secondary mill calibration from 07:30 to 11:00. Stopped calibration due to failure of the west filter press plate shifter (a broken trailing wheel had gotten off track), and resulting overflow of the filter press feed tank. Two sets of mill discharge composite samples were collected and taken to the lab. The weigh belt feeder

worked well in automatic throughout the run. The No. 2 flotation pump discharge piping was installed to the feed valve to the vacuum drum filter system. The old microagglomerate flowmeter was relocated to this piping.

January 30

Put the west filter press trailing wheel back on track and temporarily operational. Replacement wheels were ordered, delivery of January 31. The secondary mill calibration was continued, with operation from 09:30 to 13:30. Composite samples were taken every 45 minutes beginning at 10:15. The weigh belt feeder operated well in the automatic mode, controlling the coal feed rate at 4000 lbs/hr for the entire run. The primary ball mill was flushed to clean the ball charge of rust, etc. Installation of the mechanical samplers was completed.

January 31

An additional 1000 pounds of balls were added to the secondary mill, bringing the charge to about 6800 lbs, in preparation for the next calibration run. Trailing wheels were received and the broken wheel replaced on the west filter bank. Automatic operation of the weigh belt feeder was checked by feeding feeder coal discharge into drums. A total coal quantity of 1515.6 lbs of coal was metered into the drums, with the feeder indicating 1517 lbs metered. The total calibration time was 22 minutes, for a feed rate of 4133 lbs/hr at a set point of 4000 lbs/hr. A sufficient quantity of feeder discharge coal was fed through the system before calibration began to ensure steady state data was obtained. The calibration check indicated that automatic operation is good. All four water feed flow control valves were calibrated. Electrical and instrument air was hooked up to the mechanical samplers.

February 3

Calibration of the secondary ball mill at a 6800 lb charge was conducted from 07:20 to 10:30, when the run was stopped due to coal slurry in the instrument air header. A check valve in the instrument air core blow to the west bank of the filter press had failed, and the filter press feed slurry was pumped into the instrument air system. The check valve was replaced, and cleanout of the instrument air header begun. The head was removed from the primary mill, 1815 lbs balls removed, and the head replaced. The total primary mill charge, including estimated "heel", is about 3700 lbs. Tailings piping was installed from both flotation cells to the tailings tank.

February 4

Cleaning of the instrument air header and instruments continued.

February 5

Calibrated secondary mill from 07:40 to 12:00, when the coal feed chute plugged. Samples from the run were taken to the laboratory for percent solids and particle size analysis. The weigh belt feeder failed to operate in automatic; maintenance could correct the problem. An electrical hazard was discovered beneath the ball mills where an open-ended cable was energized and went to ground. The problem was corrected. The 150-gallon high shear mixer was opened and inspected. The impellers and vessel internals were in good shape.

February 6

A water balance was begun through the flotation cells, tailings tank, and the No. 2 flotation product tank. Water supply was lost due to failure of the thickener overflow pump, which was pulled and repaired. The impeller nut had been eaten away by low pH water during CDTF operation, and the impeller was not able to deliver full flow. In addition, water flow control valves were not operating properly, and the instruments were cleaned more thoroughly of coal slurry. The calibration of the diesel metering pump to high shear was checked, and found to be different from that seen during CDTF operation. Data was collected to allow flow rate determination for the range of flows to be used during POC operation. Repair of the dump truck bed began to make it leakproof for disposal of products to the Gaston coal pile.

February 7

Calibration of the secondary mill was conducted again to achieve steady state operation for a long enough period of operation to verify the particle size distribution of the mill product. The secondary mill pump stopped pumping several times, finally maintenance was called to repair an electrical problem at the pump motor. Continued mill calibration by letting secondary mill tank overflow to mill area sump, then pumping sump to filters via tailings tank. Samples of mill discharge were taken every hour, and delivered to the lab for % solids and PSD. A calibration of the diesel metering pump to high shear was conducted.

February 10

Added 500 lbs balls to secondary mill. Total in secondary mill now at about 7300 lbs. A calibration run was conducted at the secondary mill, with flow through high shear and the flotation cells. Lost water to the ball mill feed several times due to the low flow rate of water to the head tank. The pump was pulled and checked by maintenance, but no reason could be found for the pump not to deliver the expected 250 gpm. The pump curve was checked, and it was found that the NPSH requirement of the pump was listed at 10 ft. head, and that the available suction head is only about 3 - 4 ft. head. The pump delivers only 105 gpm with a low suction head. Since the flow rate required by the POC is higher than that of the CDTF, water to the mill feed piping was intermittent as the head tank level fluctuated.

February 11

The flow rate of the thickener overflow pump was checked after it was reinstalled by maintenance. The flow was measured at about 110 gpm. A replacement pump was ordered with sufficient flow rate and discharge pressure and a suction head requirement of only 3 ft. The filter cloths on the east bank of filter plates was replaced with the cloths with the rubber core.

February 12

Calibration of the secondary mill was done with a ball charge of about 7300 lbs. Mill discharge was sent directly to the filter press feed tank. The unit was operated from 07:30 to 11:30. Samples were taken to the lab for PSD and percent solids. The pH of the process water was measured at 7.6. The flow through the mechanical samplers at each tailings stream was checked with no sampler discharge piping to check for backing up of flow in the sampler, and resulting sample contamination and overflow of the samplers to the pad.

The replacement pump for the thickener overflow system arrived at the site. Maintenance began removing the original pump and preparing the base of the new pump.

February 13

Approximately 800 lbs balls were added to the secondary mill, bringing the total to about 8100 lbs. A replacement thickener overflow pump was installed. The pump was started, and delivered 730 and 725 lb/min (87.5 and 86.9 gpm) flow, respectively, to No. 1 and No. 2 flotation cells.

February 14

Operated secondary mill with discharge through the large high shear mixer to the flotation cells, which were operated in series. Diesel to high shear was 1.0 lb/min (1 wt %). No frother was added to the flotation cells. The flotation cells were initially operated at a solids concentration of 6%, but was changed to 9% solids due to the flow restrictions in the tailings mechanical samplers. Purged excess water to WWT. The excess water came from a washdown header installed beneath the ball mills. This water, designed to keep solids flushed under the mills, added an excessive amount of fresh make-up process water to the POC unit, resulting in high inventories throughout the plant. The thickener would not operate properly, and the filter press feed and tailings tanks overflowed much of the day.

February 17

Too much water in system, but could not purge to WWT due to high rainfall amounts over the weekend. The froth paddles on the new flotation cells were adjusted for better froth recovery. Frother metering pumps were set up at the cells.

February 18

Did not run due to high water inventory. Excess water was purged to WWT as that unit allowed. Changed a trailing wheel on the east bank of filter plates.

February 19

Work continued on CDTF revisions. Began secondary mill calibration at 8100 lb ball charge at 07:20. The hammermill crusher plugged, and the run was stopped at 08:15 due to loss of coal. The 1/4" screen was removed from the hammermill crusher discharge.

February 20

Started up secondary mill for calibration run at 8100 lbs at 07:15. The mill discharge was sent through the 150 gallon high shear mixer to the flotation cells, which were operated in series. This allowed continued shakedown of the system and checks of the mechanical samplers. Diesel agglomerant addition to high shear was set at 1% throughout the run. The flotation cells were operated at 9% solids, and two set points were reached, one with no frother to the flotation cells, and one with the minimum possible with the "old" metering pumps at a 50/50

mixture of frother and water. The mechanical samplers were operated for one hour at each of the two set points after steady-state was obtained. For the set point with no frother addition, the samplers were set at one cycle every two minutes, resulting in 60 passes of the hose over the sample opening (one cycle is "over and back"). The set point with frother addition used a sampling frequency of one cycle every one minute.

The mill discharge was routed through the small high shear mixer from 07:30 to 08:00 while maintenance repaired leaking packing on the 150 gallon high shear mixer. The first set point samples were obtained, then the ball mill feed chute plugged at 11:00. The plug was removed and coal feed resumed at 11:15. Steady-state was reobtained before the second set point samples were obtained. The sampling was begun at 11:40. The mill plugged again at 12:30, due to trouble with the water FCV to the mill. Coal feed was reobtained at 12:40, and the unit shut down at 12:40. Samples were taken at all five of the mechanical samplers at each of the two steady-state conditions. These were sent to the laboratory for % solids, ash, and PS. The relative performance of the flotation cells with and without frother at 1 wt % diesel to high shear could be compared. However, it must be noted that the hammermill crusher discharge screen had been removed prior to the run.

February 21

The vacuum drum filter unit arrived at the site, set in place and electrical and piping connections begun. A filter cloth and manuals for the unit arrived in the mail.

Started up secondary mill for calibration run at 8100 lbs at 07:15. The mill discharge was sent through the 150 gallon high shear mixer to the flotation cells, which were operated in series. This allowed continued shakedown of the system and checks of the mechanical samplers. Diesel agglomerant addition to high shear was set at 1.5% throughout the run, to allow comparison of performance with the 1% diesel run on February 20. The flotation cells were operated at 9% solids, and two set points were reached, one with no frother to the flotation cells, and one with the minimum possible with the "old" metering pumps at a 50/50 mixture of frother and water. The mechanical samplers were operated for one hour at each of the two set points after steady-state was obtained. For the set point with no frother addition, the samplers were set at one cycle every two minutes, resulting in 60 passes of the hose over the sample opening (one cycle is "over and back"). The set point with frother addition used a sampling frequency of one

cycle every one minute. There were no system upsets during the run. Samples were taken at all five of the mechanical samplers at each of the two steady-state conditions. These were sent to the laboratory for % solids, ash, and PS.

February 24

Calibration of the primary ball mill at a charge of 3700 lbs balls was conducted. The calibration run started at 07:30, with mill discharge sent directly to the filter presses. The run continued until 12:30, with samples obtained every hour. Installation of the vacuum drum filter continued.

February 25

Added 2300 lbs balls to the primary mill in preparation for operating the mills in series for at 200 mesh grind target. Installation of the vacuum drum filter piping continued. Piping to operate mills in series was installed.

February 26

The 1/4" screen was replaced in the hammermill discharge. Data from runs on February 20 and 21 indicated that the size consist of the mill feed and product had changed significantly. It was decided to put the screen back in and work harder on keeping the coal pile dry.

February 27

Maintenance removed every other finger in the back third of the hammermill discharge screen. The manufacturer suggested this as the first measure to keep screen discharge unplugged. By removing every other finger, it may be possible to stay unplugged and still obtain a size consist necessary for mill feed (1/4").

February 28

A mill calibration was conducted at the secondary mill with the partial hammermill screen. The purpose of the run was to evaluate the hammermill and ball mill discharge PSD to determine if the hammermill could be operated with reduced number of discharge fingers, without producing excessive fluctuations in the PSD's. In addition, the ball mill discharge sampling method was tested by obtaining the same "composite" as previous runs, as well as a "grab" five-gallon bucket as was done during CDTF operation.

The secondary mill was started at 08:15 and the coal in the three ton surge bin was purged. The bin was emptied and new crushed coal began feeding the mill at 09:45. The run continued for over five hours, to allow collection of samples every hour by both sample collection methods. The run was shut down at 14:55 and the unit cleaned. The ball mill feed chute plugged at 14:10, and was unplugged and operational at 14:20. Data from the run indicated that the hammermill discharge size consist varied from about 800 to 1400 microns, too much variability for feeding the mills. The hammermill screen discharge fingers will all have to be reinstalled.

March 2

Maintenance replaced all the fingers in the hammermill discharge screen. Data from calibrations on February 28 indicated too much variability in the crusher discharge size. The high pressure pump head was replaced. The thickener underflow pump would not operate well. The coal was covered with the new tarp.

March 4

Tried to filter the water from the thickener, but the thickener underflow pump would not pump. Water was found in the instruments. Liquefaction personnel had blocked off the instrument air header to the site, and water had backed up into the purges on some of the instruments. Calibration of new frother metering pumps began.

March 5

Added 1000 lbs of balls to the primary mill, then operated the mills in series. This allowed calibration of the primary mill as a 100 mesh mill, and the mills in series as 200 mesh operation. The run was started at 08:30, with the secondary mill discharge pumped to the filter presses. Collection of mill discharge samples began at 11:30, and continued to 14:30, when the mills were shut down. The samples were taken to the laboratory for PSD and percent solids. At 10:15 a portion of the mill discharge was fed to the rotary drum vacuum filter, which was operated until completion of the run. The samples of the mill discharge were collected with by a new method - a five-gallon bucket was used to collect three increments of mill discharge over a three minute period.

March 6

A calibration of the primary mill was conducted, with the mill discharge sent through the 150-gallon high shear

vessel to the flotation cells and rotary drum vacuum filter. The run was started at 07:05, but was stopped at 10:00 due to the inability to get the thickener underflow pump operating. Work began draining the thickener to allow replacement of the thickener underflow pump with the pump which was used as the thickener overflow pump during CDTF operation. This pump has a larger suction size, which should help prevent some of the solids pluggage problems. Also, the thickener rake was overloaded with solids, and kept tripping off.

March 7

Water was pumped from the thickener in preparation for maintenance to replace the underflow pump on March 9. Excess water was purged to WWT.

March 9

The thickener underflow pump was replaced.

March 10

The thickener was filled with water. Added 700 pounds of balls to primary mill.

March 11

The mills were operated in series to calibrate for 100 and 200 mesh. Coal feed was started at 09:15. The new frother metering pumps were used. The mills were shut down at 13:30 due to solids in the recycle water caused by holes in the head cloth of the large bank of the plate and frame filter presses. Replacement cloths were ordered. The mill discharge was sent to the 150-gallon high shear mixer and flotation cells, so product could be dewatered via the vacuum drum filter, to relieve the solids loading on the one filter press with all good cloths.

March 12

Water was circulated to remove solids. The clarified water tanks were flushed, and floating material on the thickener cleaned.

March 13

The filter cloth replacement for the head plate of the large bank of plates had not arrived, so cleanup of water continued using the small bank.

March 16

The ball charge to the primary mill was increased by 800 lbs, then the mills were operated in series to calibrate for 100 and 200 mesh. The run was started at 08:00, but was interrupted for 30 minutes at 08:45 due to pluggage of the mill feed chute caused by low water levels in the thickener. The chute plugged again at 10:20, and the unit was shut down. The hole in the head cloth of the large bank of filter plates continued to contaminate the recycle water with solids.

March 17

Calibration of the new frother metering pumps was begun. The filter cloths for the head end on the large bank of filter plates arrived and were installed.

March 19

Run 1 was conducted with the mills started in series at 08:30 to continue calibration at 100 and 200 mesh. A 200 mesh grind was sent to high shear. The 150-gallon high shear mixer and flotation cells were used, and the diesel addition rate to high shear checked with a calibration pot at the diesel pump suction. Flotation cell solids were 10%. The pump would deliver only about 1.2 lb/min at maximum speed, or about 1% diesel. The run was stopped at 11:30.

March 20

The high shear diesel pump was inspected, but no problem could be found. The discharge tubing from the calibration pot was replaced with 1" pipe to reduce suction friction.

March 23

Run 2A was started at 07:30 to calibrate the ball mills and diesel addition to high shear. Ball mill slurry (200 mesh) was sent to the 150-gallon high shear and flotation. The flotation cells were operated at 10% solids. The diesel pump would not provide more than about one lb/min, so the pump was replaced with another one. Flow still would not increase, indicating pluggage in the diesel piping going into the bottom of the 150-gallon high shear mixer. The run was stopped at 10:45 due to the inability to obtain an adequate range of diesel addition and pluggage of the ball mill inlet chute. The diesel piping to high shear was unplugged.

March 24

Run 2B (the run planned for March 23) continued, with the mills operated in series for a 200 mesh grind coal, and diesel calibration to the 150-gallon high shear mixer. The run was started at 08:00, with an initial diesel addition of 1%. The flotation cells were operated at 10% solids. The flotation cells were lined out and samples taken of product and tailings. The diesel addition was increased to 2% at 10:00, and steady state again achieved in the flotation cells. Samples were obtained, and the diesel decreased to 0.5% at 11:30. The cells would not line out at 0.5%, so the diesel was increased to 0.7% at 12:35. The cells were lined out and samples taken.

March 25

The ball charge to the primary mill was increased by 500 lbs. The diesel pump to high shear was replaced with a smaller capacity pump to increase the range of the pump used for metering.

March 26

Run 3 was conducted with the mills started up at 07:30 in series for 200 mesh operation, with flow through high shear and flotation. Flotation cell solids were 6%. (No flow restriction problems occurred with the increase in volume throughput when the solids were decreased from 10 to 6%) The diesel pump would not pump, so maintenance reinstalled the larger, original pump removed on March 25. Ball mill discharge slurry was diverted to the filter presses while the diesel pump was being replaced to continue purging of the mills to achieve the new mill steady-state after the mill addition. The run continued at 10:00 at 1% diesel to high shear. Samples were taken, then the diesel changed to 0.7%. Flow rates and samples were taken at 0.7% diesel. The run was stopped at 14:30.

March 27

A new 4" line was installed from the tailings tank to the filter press feed tank, and replaced the diesel pump to high shear with a new, smaller pump. The high pressure water pump was repacked. The flotation cell level controller bubbler tubes were reworked to provide a continuous air purge. Frother pump calibrations continued. An attempt was made to pump water only through the 45-gallon high shear vessel to evaluate possible secondary mill pump capacity restrictions through the vessel and in-

line mixers, but the mixer would not operate. Maintenance began investigating the problem.

March 30

Run 4 was conducted with the mills operated in series to produce a 200 mesh slurry, and the 150-gallon high shear was used. The purpose of the run was to conduct a water level/froth height test to evaluate impact on flotation performance. The run was started at 07:30 with 0.5% diesel to high shear, and 6% solids in flotation. Three set points were obtained, and samples at each set point were obtained of flotation streams. Manual flow rate measurements were taken at the third set point of flotation products and tailings, as well as vacuum drum filter product. The run was stopped at 13:00.

March 31

A calibration of the levels in the cells at various level controller settings was conducted.

April 1

Run 5 was conducted with 100 mesh (primary mill only) to obtain baseline data for flotation cell operation at that grind. The diesel addition was 0.5%, and the flotation cell solids was 6%. The frother rates were adjusted to obtain three set points, under Steve's direction. Samples and flow rates were taken at each set point. The run started at 07:15, and continued until about 12:00. The run was stopped at 13:00 due to continued loss of water. Flocculant was added the thickener via a new flocculant addition system. The small bank of the filter presses began malfunctioning in the plate shifter assembly, and maintenance began investigation.

April 2

Added 500 lb balls to the primary mill. Installed sight glass on the 3000 gal water storage tank.

April 3

Run 6 was conducted with 100 mesh coal, 150-gallon high shear mixer, 6% solids in flotation, and 0.5% diesel addition to high shear. The purpose of the run was to obtain baseline data for 100 mesh coal and evaluate the effect of level changes in the flotation cells. The unit was started at 09:00. Steady state was obtained and samples pulled at 10:10. Flow rates were also taken. A second set point was obtained and sample/flow rates

taken at 12:45. The unit was shut down at 13:30. Manual flow rates compared well with calculated flow rates from material balance. The pH was measured at 7.1, and some caustic added.

April 6

Run 7 was conducted with 100 mesh coal, 150-gallon high shear mixer, and 6% solids in high shear. The purpose of the run was to evaluate the effect of varying the diesel concentration to high shear. Three set points were obtained - 1.0, 0.75, and 0.5% diesel. The run was started at 07:30 with 1.0% diesel. Steady state was obtained and samples taken with the mechanical samplers from 08:50 to 09:20. Flow rates were taken at the end of the sampling period. The diesel concentration was changed to 0.7% at 09:35, and steady-state obtained. Samples were taken from 10:15 to 10:45, then flow rate measurements were obtained. The diesel was changed to 0.5% at 11:00. Samples were obtained from 11:30 to 12:00, then flow rate measurements taken. The unit was shut down at 12:15.

April 7

Water was circulated through the system to clean solids. All remaining drums of project coals were disposed of at Gaston (with exception of 6 drums each of UF, Pitt #8, and V.I. Ky # 9). The baffles in the flotation cells were lowered to separate the cells. Caustic was added to the system to increase pH.

April 8

Run 8 was conducted with 100 mesh coal, 150-gallon high shear mixer, 0.5% diesel, and 6% solids to flotation. The purpose of the run was to evaluate the effect of varying frother addition rates, as well as levels. The run was started at 07:20. Nick Moro and Harold varied the frother and levels to obtain steady-state, and the first set of samples were collected from 11:05 to 11:35. Flow rate measurements were taken. The gear for the paddle on the first stage of flotation began to slip and stop. The run was stopped at 12:20, and unit shutdown commenced. The reduction gear had stripped some teeth. A new gear box was ordered, with delivery expected on April 9.

April 9

Product was transported to Gaston, and the product pad was cleaned. Circulation of water and purging to WWT

was done. The material balance data for runs 7 and 8 indicated that the percent solids in the high shear feed was not right. Therefore, the water control valves to the ball mills and the secondary mill sump were checked for proper calibration. Both checked out good. The data also indicated the weigh belt feeder was beginning to act up. The reduction gear for the paddles on the first two cells of the first stage flotation was replaced. The teeth on a portion of the brass gear had become worn. The packing and lantern rings were replaced on the 150-gallon high shear mixer. A shaft sleeve for the high pressure water pump was fabricated and the packing replaced on the pump.

April 10

The weigh belt feeder was repaired. Coal was cleaned from beneath the belt.

Run 9 was started at 09:40 with 100 mesh coal, 150-gallon high shear, 0.5% diesel, and 6% solids in flotation. The purpose of the run was to evaluate the effect of varying the frother addition rates and levels, similar to Run 8 on April 8. The first set of steady-state samples were obtained from 12:00 to 12:30. Flow rate measurements were also taken.

April 13

Maintenance pulled the flocculant mixer to replace bearings. Replacement bearings were ordered, due in a day or two. They also installed the piping required to run the mills in series. A check of the flotation cell level controllers was conducted.

Run 10 was conducted with 200 mesh coal, the 150-gallon high shear mixer, 0.5% diesel, and 6% solids in flotation. The purpose of the run was to evaluate frother additions. The run was started at 08:50. Steady-state was difficult to obtain due to ornerly level control valves. The level would be steady, then would move up or down for no reason. A set of samples and flow rate measurements were taken at 12:00. Another set was taken at 13:30, but the sampling period was cut short due to an upset caused by one of the level controllers.

April 14

Representatives from Wemco visited the site and made suggestions pertaining to flotation cell operation, including level control operation. They suggested that the bubbler tubes for the No. 1 and 2 cells of both stages

of flotation be moved from the tailings discharge box to the No. 2 cell area. The operators conducted a test of the throughput of the secondary mill pump through the 45-gallon high shear mixer with and without the in-line mixers in series. The flow through the 45-gallon high shear mixer was 432 lb/min or more, at an inlet pressure of 9 psig and an outlet pressure of 3 psig. The flow through the 45-gallon mixer and in-line mixer was 245 lb/min, with an inlet pressure of 13 psig and an outlet pressure of 11 psig. Water was circulated and the pH was adjusted by the addition of caustic.

April 15

Run 11 was conducted with 200 mesh coal using the 150-gallon high shear mixer. Diesel addition to high shear was 0.5%, and the flotation cell solids concentration was 6%. The purpose of the run was to evaluate flotation level and air flow rates on performance. The run was started at 07:15, and continued until 11:00. The level controllers were not working properly. A set of samples and flow rates was taken at 10:15.

April 16

Work began relocating the bubbler tubes to the No. 2 cell positions, instead of the tailings box locations. The weir bars were pulled from the No. 6 cell tailing boxes to clean the slots and bars for reinstallation at various levels for optimization. Piping modifications were made to decrease the pressure drop from the secondary mill pump to the flotation cells to enable the full throughput to the 45-gallon high shear mixer with in-line mixers.

April 20

The remainder of the weir bars were pulled from the No. 2 cell tailings boxes and all weir bars taken to the shop for brushing to remove rust, etc. The level controllers were checked with water and operated correctly. A test was conducted of the throughput from the secondary mill pump to the flotation cells via the 45-gallon high shear mixer with the in-line mixers in series. The maximum flow possible was only 250 lb/min. Excess water was purged to VWT.

April 21

Run 12 was conducted at 200 mesh, with 0.5% diesel in high shear and 6% solids in flotation. The purpose of the run was to evaluate flotation level and air flow rates on performance. Also, the high shear configuration (150-

vs. 45-gallon high shear mixer) was tested. The run was started at 07:25 with the 150-gallon high shear mixer. The hammermill crusher plugged at 08:00 and had to be unplugged. The unit was started back up at 09:45. Steady-state was obtained and samples pulled at 11:20. Flow was then switched to the 45-gallon high shear, and steady-state obtained. Samples and flow rates were taken.

April 22

Run 13 was conducted to evaluate the impact of varying the diesel concentration to high shear with 200 mesh coal. The 150-gallon high shear mixer was used, with 6% solids in flotation. Although diesel concentration had been tested previously, the new location of the level controller bubbler tubes in the flotation cells warranted another look. After startup, the primary mill discharge tank level continued to increase, indicating that the primary mill discharge pump was not keeping up with flow requirements. The test was changed from a 200 to a 100 mesh run. Three set points were obtained - 0.5, 1.0, and 0.75% diesel to high shear. Steady-state samples and flow rates were taken at each set point. During the continuation of Run 13, the piping feeding the secondary mill was pulled apart, and removed some pluggage. No other restrictions were found.

April 23

The plant was started up at 200 mesh coal to conduct Run 14, testing the diesel concentration to high shear. After startup, the level in the primary mill tank continued to increase, indicating that the primary mill pump was not keeping up with the flow requirements due to the pressure drop in the discharge piping. About 4 or 5 gallons of slurry were overflowing to the ball mill area sump. The run was stopped, and maintenance installed a new section of piping directly from the primary mill pump discharge into the secondary mill, replacing the previous pipe which had split off of the secondary mill bypass line. This revision lowered the actual height of the piping, and used fewer elbows, resulting in a lowering of the pump discharge head. In addition, the pump impeller clearance was decreased.

April 24

Run 14 was conducted after the piping changes were made to test the impact of varying the diesel concentration to high shear at 200 mesh coal. The 150-gallon high shear mixer was used, with 6% solids in flotation.

Steady-state was obtained at the first set point - 0.75% diesel to high shear, and samples and flow rates taken. The diesel was then changed to 0.5%, and steady-state samples and flow rates obtained. An attempt was made to force more coal into the tailings at a 0.5% diesel concentration, but while this was occurring, the ball mill feed chute plugged, and the run was stopped.

April 27

The unit was started up to conduct Run 15, a confirmation run for 200 mesh coal operation with the 150-gallon high shear mixer, 15% solids and 0.5% diesel in high shear, and 6% solids in flotation. The weigh belt feeder was not working properly, the belt would operate at too fast a speed. Maintenance was called and work conducted on the feeder. The unit was started after 1.5 hours of down time, but the feeder still would not work correctly. In addition, the water flow control valve to the secondary mill discharge tank did not work properly. Maintenance corrected the valve problem, and got the weigh feeder working. A calibration of the weigh belt feeder was conducted by sending coal off the belt to drums for a set period of time. The coal was then weighed, and found to agree with the weight indicated on the feeder display.

April 28

Run 15 was started again at 07:45 at a 200 mesh grind size to confirm the results of the 200 mesh operation during Run 14. The 150-gallon high shear mixer was used, with diesel addition at 0.5% and solids to high shear at 15%. The flotation solids concentration was 6%. The unit was started up and allowed to achieve steady-state. Three sets of samples were collected for 30 minutes each at the same conditions to check for reproducibility and steady-state. Flow rates were collected only after the third set of samples so that steady-state was not affected. Upon completion of the flow rate measurements at 15% solids, the solids concentration to high shear was increased to 20%. The unit was operated for another 1.25 hours to obtain steady-state, but this was not achieved due to an erratic water flow control valve to the secondary mill tank. In addition, the coal plugged in the hammermill crusher (due to moisture content - the coal used is from the last on the pad, and slightly higher in moisture than that used in the last few runs) and the crusher could not be cleared in time to keep up with unit feed coal demands. The unit was shut down at 12:00. After the unit was shut down, the high shear mixer was flushed well. A test was then conducted with water only to the high shear mixer at a 0.5% diesel concentration to allow collection of oil/water

samples from the high shear discharge. Two 16 oz samples were collected of the stream at each of two flow rates, the same water flow rate fed to the mixer during operation with 15 and 20% solids to high shear.

The sheaves on the 150-gallon high shear mixer were changed to decrease the rpm from 1186 to 848 in preparation for a high shear optimization test on April 9. In addition, an air line was routed to the hammermill crusher deck to hasten the process of clearing the crusher upon pluggage. Work began on the installation of an air-operated vibrator on the side of the ball mill feed chute, which could be operated to help eliminate pluggage of the mill inlet. The coal was spread out as much as possible to allow it to dry. The wet coal in the 24 ton storage bin was removed so it could be replaced with dryer coal for Run 16. The first truckload of the second batch of Upper Freeport precleaned coal was delivered to the site. The coal was placed on the coal pad separate from the original coal.

April 29

Run 16 was started at 08:00, at the conditions shown below, but the froth would not discharge from the froth launders. The run was stopped, and the discharge nozzles from the launders replaced with larger nozzles.

April 30

Run 16 was conducted primarily to test two solids concentrations to high shear - 15 and 20%, and collect data operating the 150-gallon high shear mixer at a lower rpm. The rpm of the mixer had been decreased from 1186 to 848 to conduct high shear optimization tests. In addition, the 45-gallon high shear mixer was operated at two rpm, 1800 and 1200, at 20% solids. The mills were operated in series to produce a 200 mesh feed to high shear. The solids concentration of the flotation cells was 6% throughout the run, and the diesel addition to high shear was 0.5%.

The run was started at 08:00 with 200 mesh slurry to the 150-gallon high shear mixer at 15% solids and 0.5% diesel. Steady-state was obtained and two sets of composite samples taken. Routine flow rate measurements were taken. The second set point was started at 09:45, when the solids concentration to high shear was changed from 15 to 20%. Steady-state was obtained and another two sets of composite samples taken, as well as flow rate measurements. Ball mill slurry was diverted from the 150-gallon high shear mixer to the 45-gallon mixer at 11:45, at 20% solids. The rpm was set at 1800. Steady-

state was obtained and one set of samples and flow rates taken. The rpm of the 45-gallon high shear mixer was decreased from 1800 to 1200 at 13:10, and steady-state samples and flow rates taken.

May 1

Run 17 was conducted to test impact on unit performance of using the 150-gallon high shear mixer at 20% solids and 1186 rpm, as well as to evaluate the 45-gallon high shear mixer at 15% solids at both 1800 and 1200 rpm. Flotation cell solids concentration was set at 6% throughout the run, and diesel addition to high shear was set at 0.5%. The run was started at 07:15, and steady-state obtained at the first set point - with the 150-gallon high shear mixer at 20% solids. Two sets of samples were collected, as well as routine flow rate measurements. At 09:45, the high shear feed was diverted from the 150-gallon high shear mixer to the 45-gallon mixer, and the solids concentration changed to 15%. Steady-state was obtained at an rpm of 1800, and samples and flow rates taken. The 45-gallon mixer rpm was changed to 1200 at 11:10. An instrument air compressor failure shut all water flow control valves, resulting in pluggage of the ball mill feed chute at 11:45. The run was started back at 11:55, but another air failure occurred at 12:20. The unit was started again at 12:25, and steady-state samples and flow rate measurements taken at 13:25.

May 4

Run 18 was conducted to simulate conventional flotation circuit operation for comparison to POC operation. The run was conducted at 200 mesh, with a 15% solids slurry fed to the 45-gallon high shear mixer, which was used for minimal mixing of the diesel with the slurry. The mixer was initially operated at 190 rpm, the lowest possible with the variable speed drive. Diesel concentration to the mixer was initially set at 0.1%, and increased to 0.25% later in the run. Flotation cell solids was maintained at 8% throughout the run. Four set points were obtained. Set points 1 and 2 were obtained at a diesel addition of 0.1%, with rpm settings of 190 and 1800, respectively. Set points 3 and 4 were obtained at a 0.25% diesel addition, at 1800 and 190 rpm, respectively. Steady-state was obtained at each set point, and samples and flow rate measurements taken.

May 5

Work was conducted on the secondary ball mill, which needed to be aligned on the discharge end. In addition, installation of a header for froth spray was begun.

May 6

Run 19 was conducted to test several rpm's with the 45-gallon high shear mixer at 20% solids. The diesel addition to high shear was 0.5%, and the flotation cells were operated at 6% solids. Steady-state set points were obtained at 1200, 950, and 500 rpm, and samples and flow rate measurements taken.

May 7

Run 20 was conducted to evaluate the impact on flotation performance of varying air rates and flotation solids concentrations at a 0.16 lb/ton frother rate. The 45-gallon high shear mixer was used at 20% solids and an rpm of 950. Diesel addition to high shear was 0.5%, and the ball mills were operated in series to obtain a 200 mesh grind. A total of four set points were obtained. The first two set points were obtained at a 6% flotation cell solids concentration. Set point 1 was conducted at a blower air rate normally used prior to the run. Set point 2 was conducted with an air rate lower than that previously used. Steady-state was obtained at both conditions, and samples and flow rate measurements taken. Set points 3 and 4 were conducted at a flotation solids concentration of 8%, at normal and lower blower rates, respectively. Samples and flow rates were obtained.

May 8

Run 21 was conducted to evaluate the impact on flotation performance of varying air rates and flotation solids, similar to Run 20, but at a 0.32 lb/ton frother rate. The 45-gallon high shear mixer was used at 20% solids and an rpm of 950. Diesel addition to high shear was 0.5%, and the ball mills were operated in series to obtain a 200 mesh grind. A total of four set points were obtained. The first two set points were obtained at a 6% flotation cell solids concentration. Set point 1 was conducted at the normal blower air rate indicated in Run 20. Set point 2 was conducted with an air rate somewhat lower. Steady-state was obtained at both conditions, and samples and flow rate measurements taken. Set points 3 and 4 were conducted at a flotation solids concentration of 8%, at normal and lower blower rates, respectively. Samples and flow rates were obtained.

May 11

Installation of froth spray headers began per the design of Steve Hadley of Praxis.

May 12

Installation of froth spray headers was completed.

May 13

Run 22 was scheduled. Prior to startup calibrations were conducted to check flow to the froth sprays. The water flow control valves to both stages of flotation were manually blocked, and the flow of water to the froth spray headers (cells 1 and 2 of both stages of flotation) adjusted with manual valves until the respective orifices indicated a flow of just over 100 lb/min total to each stage. The feed to the first stage was set at about 120 lb/min, split as equally as possible to cells 1 and 2, and the second stage set at about 110 lb/min. Flow was then opened to the flow control valves, which were set at normal flow rates. This allowed control of the total flow of water to each stage as normally done, except a portion of the feed water was directed to the froth spray headers. The flow passing through each control valve was then manually measured to check the accuracy of the orifice measurement to the headers. Feed to the first stage checked well, with about 95 lb/min indicated to the headers. This flow was expected to drop slightly from the initial 120 lb/min due to differences in head. Flow to the second stage could not be verified, due to incorrect operation of the valve to that stage. The valve was pulled by maintenance personnel, reworked, and reinstalled. The valve stem had been worked loose, and the butterfly would not seat properly when closed. In addition, the butterfly was loose during normal operation, and fluttered back and forth within the valve opening. Run 22 could not be completed, so it was delayed to the next day.

May 14

Run 22 was conducted at a 200 mesh coal grind, with 20% solids and 0.5% diesel to the 45-gallon high shear mixer. The purpose of the run was to evaluate froth spray header performance at about 100 lb/min flow, and evaluate three rpm's of the mixer - 1200, 950, and 600. The flotation cells were operated at 6% solids. Prior to the run the water flow control valve to the second stage of flotation was calibrated at about 110 lb/min water flow to the froth spray header in cells 1 and 2. Steady-state was obtained using froth washing at each rpm, and

normal samples and flow rates collected. One observation during the run was that the froth did not flow well out of the cells, apparently due to the orientation of the spray headers within the froth.

May 15

Run 23 was conducted at conditions similar to Run 22, except the 45-gallon high shear mixer rpm was maintained at 1200 throughout the run. The purpose of the run was to obtain steady-state with and without froth washing at the optimum blower rates and levels for evaluation of the froth washing effect on system performance. Set points 1 and 2 were obtained without and with froth washing, respectively. A third set point was obtained at even lower cell levels and increased air rates to improve performance. This third set point used froth washing. Steady-state flow rates and samples were taken at each set of conditions.

May 18

Run 24 was planned, but no run was conducted. Upon arrival at the site, it was found that the firewater sprinkler system for the No. 1 coal conveyor had tripped. This had resulted in the coal in the 24 ton storage tank becoming too wet to feed to the crusher. Upon investigation the only source of the trip that could be found was a blocked instrument air supply valve to the trip system and a suspected slow leak of air from that trip system due to rubbing of conveyor No. 1 against the piping. The block valve was opened, the conveyor realigned, and wet coal removed from the 24-ton storage tank. The reason the block valve was closed could not be determined.

May 19

Run 24 was conducted at 200 mesh, 0.5% diesel and 20% solids to the 45-gallon high shear mixer, and 6% solids in the flotation cells. The purpose of the run was to continue evaluation of froth washing and frother addition rates. The first set point was obtained with no froth spray, the second with froth spray, and the third with froth spray and a higher frother rate. The cell levels and air rates were kept constant throughout the run. Steady-state samples and flow rate measurements were taken at each point for run evaluation.

May 20

The froth spray headers within the cells were removed due to poor flow of the froth. The headers will be

replaced with a different design that provides washing while allowing adequate froth flow.

Run 25 was conducted with 200 mesh, 0.5% diesel and 20% solids to the 45-gallon high shear mixer, and 6% solids to flotation. No froth washing was used. The run was conducted to evaluate the impact of varying cell levels and resulting froth on system performance. Air rates were kept constant and low throughout the run, at the first mark on the air intake piping. The rpm of the high shear mixer was kept at 1200. Two set points were obtained, at higher and lower cell levels. Samples and flow rate measurement were obtained.

A representative of Stockhausen visited the site to check the performance of the CM-224 flocculant used in the thickener feed stream. The installation and operation of the flocculant addition system looked good, and the system appears to be producing water of sufficient quality. (Performance of the thickener has improved over the last few weeks as the operators gained more experience with the system, and the flocculant concentration was kept at the recommended 0.5 or so volume percent in the flocculant mixing and storage tank.

May 21

Operations added 300 lbs of 1" diameter balls to each of the ball mills. Continued wearing of the ball charge, as well as the change to the second batch of precleaned Upper Freeport coal, had resulted in a slight change in the particle size distribution.

May 22

Run 26 was conducted to evaluate a new design of froth spray, which provide improved flow of the froth within the cells. The run was made with 200 mesh coal slurry to the 45-gallon high shear mixer, which was operated at 1200 rpm. Diesel and solids concentration to high shear were 0.5 and 20%, respectively. Flotation cell solids were kept at 6% throughout the run. Two set points were obtained, at two air rates to the first two cells of each stage. (The second set point had a higher air rate to cells 1 and 2 than the first, to obtain a better recovery.) Steady-state samples and flow rate measurements were taken.

May 26

Run 27 was conducted to evaluate the impact of percent solids in flotation (6 and 8%) at two frother rates at each solids concentration. The 200 mesh slurry was sent to

the 45-gallon high shear mixer at 20% solids concentration and 0.5% diesel. The mixer rpm was set at 1200 throughout the run, and there was no froth spray. The first two set points used 6% solids, with frother rates of 0.29 and 0.57 lb/ton, respectively. The third and fourth set points were with 8% solids, at 0.57 and 0.29 lb/ton, respectively. Steady-state samples and flow rate measurements were taken at each set point. Blower air rates at cells 1 and 2 of both stages was kept at the first mark throughout the run. Cell levels were kept constant, also.

May 27

Run 28 was conducted to evaluate performance at three different levels in the first two cells of each stage of flotation. The blower air rates were changed at each level to optimize froth movement. In addition, baffles had been installed in cells 1 and 2 of both stages to divert the froth from the back of the cells toward the front. Three set points were obtained, with medium (water level about 3" below overflow), high (2"), and low (4") levels. The conditions of the run were 200 mesh slurry at 20% solids and 0.5% diesel to the 45-gallon high shear mixer, 6% solids in the mixer, and the mixer rpm maintained at 1200. The air rates for the three set points were set at mark 1, 1/2, and 1, respectively. Frother was maintained at 0.29 lb/ton throughout the run. Steady-state samples and flow rate measurements were obtained at each set point.

May 28

Run 29 was conducted to test the impact of high shear mixer rpm on flotation optimization. Four rpm's were tested: 500, 950, 190, and 1200. The 200 mesh slurry was sent to the 45-gallon high shear mixer at 0.5% diesel and 20% solids. The flotation cells were operated at 6% solids and a frother rate of 0.29 lb/ton. The cell baffles used in Run 28 did not appear to improve froth flow or cell performance, so they were removed prior the Run 29. No froth washing was used. The cell levels of cells 1 and 2 of both stages of flotation were maintained at the normal 3" below the overflow, with constant blower rates throughout the run.

May 29

No run was conducted. The operators completed a number of tasks needed for preparation of the demonstration run. Orders were placed for delivery of 8 truckloads of Upper Freeport precleaned coal, with 2 trucks each on June 1, 2, 4, and 5 for the demonstration run.

June 1

Run 30 was conducted to demonstrate repeatability of the unit for four consecutive sets of one-hour composite samples. The run was conducted with 200 mesh slurry to the 45-gallon high shear mixer at 20% solids and 0.5% diesel addition. The flotation cells were operated at 6% solids, and the frother was maintained at 0.29 lb/ton throughout the run. The 45-gallon high shear mixer was kept at 950 rpm, and the cell levels at 3" below the overflow. Four sets of one-hour composite samples were collected at steady-state, as well as flow rate measurements.

June 2

Run 31 was conducted to simulate conventional flotation operation at lower oil levels for comparison with POC operation at higher oil levels. The run was conducted with 200 mesh slurry, 20% solids in flotation feed, and 0.1% diesel addition. Initial frother concentration was 0.29 lb/ton. The run was started up bypassing both high shear mixers with the flotation cells at 8% solids. The cells would not line out, so feed was diverted to the 45-gallon high shear mixer, frother reduced to 0.1 lb/ton, and the solids concentration reduced to 6%. Four set points were obtained. The first set point was lined out with the 45-gallon high shear mixer at 900 rpm. The second set point was made by bypassing both high shear mixers. Ball mill slurry was diverted back to the 45-gallon mixer for the third and fourth set points, which were operated at 500 and 1200 rpm, respectively. The three ton coal surge bin was emptied at the end of the run, and then refilled using stop belt sampling in preparation of the demonstration run beginning on Wednesday, June 3.

June 3

A conference call was held with SCS (Tom Johnson), Praxis (Steve Hadley), SEI (Darrell Moore), and DOE (Tom Feeley) to discuss and finalize conditions for the Upper Freeport demonstration run. The discussions centered around the diesel concentration which should be used during the demonstration run, and whether or not the flotation cells should be operated utilizing froth recovery in cells five and six of both stages, which normally do not recovery much froth. It was finally decided to operate with 0.5% diesel and Btu recoveries similar to that proven in Run 30, since more operation has been conducted under those conditions. Once sufficient demonstration run data has been generated for POC evaluation and air toxics, the flotation cells would be adjusted to reject more pyritic sulfur by using cells five

and six, and two sets of steady-state composite samples obtained. The diesel concentration would then be decreased from 0.5 to 0.125% to obtain two additional sets of samples at the low oil concentrations for comparison to the data generated during Run 31.

Upper Freeport coal feed was started at 11:20 at 2 tph with 200 mesh slurry to the 45-gallon high shear mixer at 20% solids and 0.5% diesel addition. The mixer was operated at 950 rpm, with the flotation cells at 6% solids. Frother was added at a 0.29 lb/ton rate.

Steady-state samples and flow rate measurements were obtained throughout the run. Coal feed to the plant was sampled via stop-belt sampling to generate eight-hour composite samples, with a minimum increments per composite of 14, and a targeted minimum of 2 lb/increment. Ball mill discharge samples were taken every 12 hours for particle size and solids concentration measurements and control. Composite samples of flotation cell feed and discharge streams were collected with mechanical samplers over four-hour periods, with 40 increments per sample. These samplers were located at the flotation feed, first and second stage flotation product, and first and second stage flotation tailings. Samples of rotary drum vacuum product were obtained every four hours for moisture determinations, and the recycle water was monitored for pH. Flow rate measurements were taken every four hours at streams which did not have instrumentation, and hourly readings taken of other plant parameters.

The unit was lined out and the first set of four-hour composite samples collected from 14:00 to 18:00. Samples and flow rates were collected every four hours. Drumming of the rotary drum vacuum filter product began to collect samples for Combustion Engineering for another DOE project.

June 4

The demonstration run continued, with samples and flow rates collected every four hours. The primary mill pump flow began to be erratic, as if something was lodging in the suction. The suction piping was flushed with high pressure water from time to time to help in lining out this flow. Also, the hammermill coal crusher continued to plug frequently with the damp coal.

Drumming of the rotary drum vacuum filter product continued to collect the required samples for Combustion Engineering for another DOE project. That project initially required 20 tons, but notification was received

that 100 tons were required. Additional drums and pallets were placed on rush order, and arrangements made for additional manpower.

June 5

The demonstration run continued. Primary mill and hammermill operation continued to be a problem area.

June 6

The unit operated until 00:43 on June 6, when the primary mill pump stopped pumping. The unit was shut down and the pump replaced with one from the Liquefaction unit. The motor then went out, and another motor found in Liquefaction that would operate with the primary mill pump variable speed drive. This motor was installed, and coal feed started up again at 04:43. The unit was off stream for four hours.

Operation continued, with pluggage of the hammermill crusher occurring frequently. The vibrating feeder to the crusher was slowed down to provide a feed that could be handled by the crusher, but this feed rate was too low to keep up with the POC requirements. The plant was taken off coal feed at 08:30 due to loss of coal inventory in the three ton mill feed tank. The wet coal in the 24-ton storage bin was emptied and replaced with drier coal, but the crusher continued to plug. The hammers were worn and needed to be replaced or built up. Since no replacement hammers were in stock, the crusher hammers and discharge screen were removed so coal could be fed directly to the ball mills. This appeared to be the best method of getting back on line, due to the fine size of the crusher feed coal (about 95% -1/4"). An additional 500 lbs of balls were added to the primary mill to ensure adequate grinding of the coal for feed to high shear at 200 mesh. Coal feed was started back at 21:12, and the unit allowed to achieve steady-state.

June 7

Steady-state was achieved and sampling and flow rate measurement began again at 00:01. Four more sample periods were obtained (through 16:00), allowing the stop-belt sampling to achieve the desired 10 eight-hour composite samples for the air toxics material balance.

At 16:10 adjustment of the flotation cells began to reject more pyritic sulfur. The weir bars in the overflow from cells five and six of each stage were removed, and the paddles adjusted to recover froth from those cells. The

flotation cells were lined out and steady-state sampling began at 20:00. Two one-hour composites were obtained.

The diesel concentration was decreased from 0.5 to 0.125% at 22:25 to obtain two additional sets of samples at the low oil concentrations for comparison to the data generated during Run 31. Steady-state was obtained and samples and flow rate measurements began at 01:00. Two sets of samples were obtained. The unit continued to operate until 04:17, resulting in 96.25 hours of operation between 11:20 on June 3 and 04:17 on June 8 (total available hours of 112.95), an on-stream factor of 85.2%.

June 8

The Upper Freeport demonstration run was completed at 04:17. The unit was shut down per normal procedures. Unused Upper Freeport coal was transported to Gaston to make room for the first two truckloads of Pittsburgh No. 8, which were delivered. The pad was washed down prior to dumping the Pittsburgh No. 8. Flushing of the mills began, and cleanup of the site continued. The crusher was disassembled and repair of the crusher hammers begun. The hammers were worn down significantly, building up the hammers and coating with a hardcoat material. New replacement hammers were ordered.

Pittsburgh No. 8 Operating Log

June 9

Approximately 1500 lbs "small" (3/4 - 1" diameter) balls were added to the secondary mill. The mills were flushed, and water purged to WWT. Site cleanup continued. Repair of the crusher hammers continued.

June 10

No run was conducted. Operations added 2000 lbs balls, approximately a 50/50 split of medium (1 - 1.5" diameter) and small (3/4 - 1" diameter) balls, to the primary mill. The sumps and trenches were cleaned, and purging of water to WWT continued. Site cleanup continued. Repair of the crusher hammers continued. In addition, the ball mill feed chute and inlet piping to the primary mill were repaired where erosion had worn holes. The primary mill pump clearance was adjusted to make it tighter.

June 11

The crusher was installed, with rebuilt hammers.

Run 32 was conducted to calibrate the ball mills for Pittsburgh No. 8 coal, and conduct shakedown of the plant with that coal. This was the first run with Pittsburgh No. 8 coal. The run was started at 10:50. The primary mill pump would not keep up with flow requirements, resulting in shutdown of the run after about two hours. Two sets of ball mill discharge samples were taken for percent solids and particle size distribution analyses. The primary mill pump head was pulled and the impeller inspected. The impeller had become worn and eroded away, resulting in insufficient capacity.

June 12

A new impeller was ordered for the primary mill pump. The second stage flotation product pump and the tailings pump were inspected. A replacement impeller was ordered for the tailings pump.

June 15

Run 33 was started conducted with the ball mills in series to check the calibration of the ball mills and conduct a shakedown of the entire plant with Pittsburgh No. 8 coal. The target grind was 200 mesh, with a 20% slurry fed to the 45-gallon high shear mixer. Diesel addition to the mixer was 0.5%, and the rpm was 700. The flotation cells were operated at 6% solids, with a frother concentration of 0.18 lb/ton. Two steady-state set points were obtained, and samples and flow rate measurements taken.

June 16

Run 34 was conducted with 200 mesh slurry to continue calibration of the ball mills and evaluate performance of the unit with low oil concentrations with Pittsburgh No. 8 coal. The 45-gallon high shear mixer was operated at 20% solids, 0.125% diesel, and 700 rpm. The flotation cells were operated at 6% solids and 0.18 lb/ton frother. The blower rates and cell levels were varied between set points for evaluation on performance. The second set point had high air rates to cells one and two of each stage, compared to set point one, and the third set point had higher air rates to cells three through six, compared to set point two, as well as higher cell levels. Steady-state was obtained at each of the three set points, and samples and flow rate measurements taken.

June 17

Run 35 was conducted with 200 mesh coal to evaluate performance with varying air rates and cell levels, achieving several recovery points with 0.5% diesel to high shear. The 45-gallon mixer was used, at 20% solids and 1200 rpm. The flotation cells were operated at 6% solids and 0.18 lb/ton frother. Three set points were achieved and steady-state samples and flow rate measurements obtained. The first set point was lined out at the cell levels and air rates that appeared to provide the best performance based on observation. The second set point was operated at lower levels and blower rates in cells three through six of both stages. The third set point was obtained at higher cell levels and lower blower rates in cells one and two of both stages, compared to set points one and two.

June 18

An additional 600 lbs of medium ball were added to the primary mill, and 400 lbs small and medium balls added to the secondary mill.

Run 36 was conducted to continue ball mill calibrations and obtain several recovery points with the 45-gallon mixer at 700 rpm, 0.5% diesel, and 20% solids. The flotation cells were operated at 6% solids and 0.18 lb/ton frother. Two set points were obtained, with constant levels in all cells throughout the run, but lower air rates to cells three through six of both stages. Steady-state samples and flow rate measurements were taken.

June 19

Run 37 was started at 0.125% diesel and 700 rpm to the 45-gallon high shear mixer, but the plant was shut down due to the inability of the primary mill pump and thickener overflow pump to keep up. The flow supplied by the primary mill pump was erratic due to the discharge of small pieces of metal from the mills into the pump suction. The water feed requirements to the second stage flotation cells is higher with Pittsburgh No. 8 than with Upper Freeport coal, due to the higher solids concentration of the first stage product. The thickener overflow pump, which supplies water to the head tank feeding the POC, is not of sufficient capacity to meet this requirement.

Both high shear mixers were opened to allow inspection of the impellers. The 150-gallon mixer impellers were in good shape. The bottom 45-gallon mixer impeller had some wear, from the metal exiting from the mills (a good handful of small metal pieces was drained from the 45-

gallon vessel). The thickener overflow pump impeller was inspected, and found to be in good shape, however, the impeller was not the maximum possible, so a larger impeller was ordered (13" vs. 9.5"). The primary mill pump suction was inspected for pluggage, but none was found. Obstructions which may have been causing pump flow to be erratic had been flushed out upon shutdown.

June 22

Operations added 400 lbs 50/50 small/medium balls to the primary mill and 300 lbs small balls to the secondary mill. Maintenance completed reinstallation of the thickener overflow pump which had been inspected on June 19, tightening the impeller clearance as much as possible. The No. 1 flotation pump impeller was replaced. The motor on the rotary drum vacuum filter blower was reinstalled, after being rebuilt due to water damage inside the motor. The 45-gallon high shear mixer was put back together, to allow operation until a replacement impeller arrives for the bottom impeller. The tailings pump impeller, which was very worn, was replaced. The secondary mill pump impeller was inspected, and found to be worn somewhat, so a new impeller will be ordered.

June 23

Run 38 was conducted to continue calibration of the ball mills and orientation testing with Pittsburgh No. 8 coal. The run was made using mesh coal to the 45-gallon high shear mixer, at 20% solids, 1200 rpm, and 0.125% diesel. The flotation cells were operated at 6% solids and 0.18 lb/ton frother. All cell levels were kept constant, with the variable between set points the air rates. The air was increased from set point one to set point two in cells one through four of both stages, then again for the third set point. All blowers were at maximum air rate for the third set point. Steady-state samples and flow rate measurements were taken at each set point.

June 24

Run 39 was started up with 200 mesh coal to continue orientation testing for Pittsburgh No. 8, as a repeat of the third set point of Run 38. The 45-gallon high shear mixer was used, at 20% solids, 0.125% diesel, and 1200 rpm. The flotation cells were operated at 6% solids and an initial frother concentration of 0.18 lb/ton. About 45 minutes after startup the rotary drum vacuum filter blower motor stopped working. Product was diverted to the plate and frame filter press, and maintenance pulled the head off the blower to inspect impeller. The impeller was

too close to the back head of the casing, causing it to hit on the head, resulting in the blower tripping out. Some shims were placed behind the impeller to keep the impeller clear. The run continued, but a hole in the rotary drum filter cloth resulted in solids in the filtrate. Operation continued by allowing the filtrate to return to the process sump. A steady-state set of samples and flow rate measurements was collected.

The frother was increased for the second set point, but before steady state was achieved the ball mill feed chute plugged. The vibrator on the side of the ball mill feed chute had been turned down too low. The chute was cleared, and the vibrator turned up. Shortly afterwards a trailing wheel on one of the banks of the plate and frame filter presses broke, making it difficult to keep up with the filter feed material, which had more solids than normal due to the hole in the rotary drum filter cloth. Finally, the run was stopped after only about 3.5 hours of operation when one of the secondary mill motor belts broke from rubbing on the belt guard. The belts were replaced with a new set, but the unit was not started back up because it was so late in the day. During the run, the discharge of metal from the mills continued to be a problem. Screens were placed at the mill discharge to catch these - 1/4" pieces until magnets which had been ordered arrived.

June 25

Run 40 was conducted with 200 mesh coal to continue evaluation of Pittsburgh No. 8 coal with changes in flotation cell conditions. The 45-gallon high shear mixer was operated at 1200 rpm, 20% solids, and two diesel concentrations - 0.125 and 0.25%. The flotation cells were operated at 6% solids and varying frother rates (0.25 and 0.18 lb/ton). Set point one was conducted at 0.125% diesel to high shear and 0.25 lb/ton frother in the flotation cells. Set point two was conducted at 0.25% diesel to high shear and 0.18 lb/ton frother. Finally, set point three was made at 0.25% diesel and 0.25 lb/ton frother, but reduced blower rates in all cells. Steady-state samples and flow rate measurement were collected. The filter cloth on the rotary drum vacuum filter was replaced, which had continued to have solids in the filtrate despite repair.

June 26

Run 41 was conducted to continue flotation optimization by varying the frother and air rates at 0.25% diesel to high shear. The 45-gallon high shear mixer was used, at 20% solids and an initial rpm of 1200. The diesel rate was maintained at 0.25% throughout the run. The

flotation cells were operated at 6% solids, and an initial frother rate of 0.18 lb/ton. Samples and flow rates were taken for set point one, but the mill pumps continued to have fluctuations due to metal discharged from the mills in the impellers. The air rates were increased slightly from set point one to set point two. The rpm was decreased from 1200 to 700 for set points three and four, and no frother was added in cells three of either stage during set point three, for a total frother of 0.07 lb/ton. Frother was returned to a 0.18 lb/ton rate for set point four, which was conducted at 700 rpm in the 45-gallon high shear mixer, and air rates reduced due to the restored frother rates.

June 29

Run 42 was conducted to continue evaluation of flotation performance. The ball mills were operated in series to produce a 200 mesh slurry, which was sent to the 45-gallon high shear mixer at 20% solids. The mixer was operated at 1200 rpm and varying diesel rates. The flotation cell solids concentration was 6% throughout the run, and the frother rate was 0.21 lb/ton. Three set points were obtained, with 0.5% diesel to the high shear mixer in set points one and three, and 0.35% diesel in set point two. Flotation air rates and cell levels were optimized at each set point based on observation. Steady-state samples and flow rate measurements were taken for each set point. During the run, consistent diesel rate could not be achieved at a particular flow rate. The lower oil concentrations required are in the lower range of the pump capacity.

June 30

Run 43 was conducted with the primary mill only, for a 100 mesh grind slurry, to evaluate performance of Pittsburgh No. 8 coal at the coarser grind. A 20% solids slurry was fed to the 45-gallon high shear mixer, operated at 1200 rpm and varying diesel addition rates. The flotation cells were operated at 6% solids and varying frother rates. The first set point was conducted with 0.5% diesel and a total frother concentration of 0.18 lb/ton. The second set point was obtained at the same diesel addition rate and frother concentration, but at lower air rates in the second stage and lower levels in cells three through six of the second stage. The third set point was at a target of 0.125% diesel to high shear and 0.25 lb/ton frother. The second stage levels were decreased slightly from set point two, and the diesel rate was lower than targeted. Set point four was conducted at 0.07% diesel (target of 0.125%), at 0.28 lb/ton frother and cell levels and air rates similar to set point three.

Samples and flow rate measurement were taken at each set point.

July 1

The flocculant addition pump was relocated for use in adding diesel to the high shear mixers, due to the lower capacity rating of that pump. The caustic addition pump was installed for use with the flocculant, resulting in caustic addition to the process sump.

Run 44 was conducted to evaluate operation with the relocated diesel pump, and continue evaluation of Pittsburgh No. 8 coal at 100 mesh. The 45-gallon high shear mixer was used, with 20% solids and varying diesel addition rates. The flotation cells were operated at 6% solids for the entire run. Three set points were obtained. The first two set points were identical to evaluate reproducibility of the data as well as to check the diesel pump rate. The frother rate for set points one and two was 0.29 lb/ton. Set point three was conducted at 0.5% diesel and slightly lower frother rates (0.24 lb/ton). Steady-state samples and flow rate measurements were taken at each set point.

The asphalt tank was emptied to a vendor-supplied truck. Several inches of asphalt remain in the bottom of the tank below the take-off point.

July 2

Impellers on the secondary mill pump and the thickener overflow pump were replaced (feeding the water head tank). Two trucks were loaded with drums of product from the Upper Freeport demonstration run for shipment to Combustion Engineering.

July 6

Two trucks were loaded with drums of product from the Upper Freeport demonstration run for shipment to Combustion Engineering. The high-G centrifuge arrived from ALFA-LAVAL Sharples, and was set on the process pad.

July 7

Run 45 was conducted at 200 mesh, using the 45-gallon high shear mixer at 20% solids, 700 rpm, and 0.125% diesel. The flotation cells were operated at 6% solids and 0.29 lb/ton frother. Set point one was operated at high air rates to most cells of both stages. The cell levels

were reduced slightly for set points two and three in cells three through six of both stages. The air rates were changed slightly for these two set points to obtain different recovery points. Steady-state samples and flow rate measurements were taken at each set point.

July 8

Run 46 was conducted at 200 mesh, with the 45-gallon high shear mixer at 700 rpm, 20% solids, and 0.5% diesel. The flotation cells were operated at 6% solids. Four set points were obtained at varying frother rates and cell conditions. Set points one, two, and three were at 0.07 lb/ton frother. The cell levels remained constant for these three set points. Set points one and two were identical, while set point three was made with slightly higher air rates in cells three and four of both stages. Set point four was conducted at the same air rates as set point three, but at slightly higher levels in cells one and two of the second stage. Steady-state samples and flow rate measurements were taken at each set point.

July 9

Run 47 was conducted with no high shear mixer to evaluate performance of Pittsburgh No. 8 coal with that configuration. The ball mills were operated in series to provide a 200 mesh slurry, which was sent to the flotation cells at 20% solids. The diesel concentration was maintained at 0.125% for the entire run, and the flotation cells were operated at 6% solids and 0.27 lb/ton frother. Flotation cell levels were constant throughout the run, while the air rates were varied to obtain several recovery points. Steady-state samples and flow rate measurements were obtained at each set point.

July 10

Run 48 was conducted to evaluate Pittsburgh No. 8 performance with 200 mesh at low frother rates. The 45-gallon high shear mixer was used, with 700 rpm, 0.5% diesel and 20% solids. The flotation cells were operated at 6% solids. The frother rate was 0.07 lb/ton for set point one, and 0.14 lb/ton for set point two. Cell levels and air rates were kept constant throughout the run. Steady-state samples and flow rate measurements were taken at each set point.

July 13

The small pieces of metal exiting the mills had resulted in the removal of metal over time, and the mill discharge

particle size distribution had slowly increased over time. To keep the grind at the desired 100 and 200 mesh, additional balls were added to the mill prior to Run 49. Approximately 400 lbs small/medium balls were added to the primary mill, and 450 lbs small/medium balls to the secondary mill.

Run 49 was conducted to continue evaluation of Pittsburgh No. 8 at low frother concentrations. Run conditions were the same as Run 48, with mesh slurry fed to the 45-gallon high shear mixer, which was operated at 700 rpm, 20% solids, and 0.5% diesel. The flotation cells were operated at 6% solids, with a frother addition rate of 0.1 lb/ton for set point one, and 0.16 lb/ton for set points two and three. The frother addition point was changed for Run 49. Instead of the normal addition at cells one and three of each stage, the frother was added at cells three and five. Initial air rates were the same as set point two of Run 48. Cell levels remained relatively constant between all three set points, as did the air rates. Steady-state samples and flow rate measurements were taken at each set point. Installation of the high-G centrifuge began.

July 14

Run 50 was started to allow comparison of the 45-gallon and 150-gallon high shear mixers at low diesel concentrations to high shear (0.125%). The 45-gallon mixer was used for the first set point, at 20% solids and 1200 rpm. The flotation cells were operated at 6% solids, with a 0.27 lb/ton frother addition rate. After 1.5 hours of operation, the inlet piping to the 45-gallon high shear mixer plugged with metal. The mixer was shut down and the metal flushed, then the unit was started back up. A set of samples were taken, but the recovery was very low. The cell levels and air rates were increased to improve recovery, but it still appeared to be low. Shortly thereafter the primary mill pump stopped pumping due to metal in the suction piping. The unit was shut down and flushed out. Only one set of samples was taken.

July 15

A 3/16" mesh screen was installed at the mill discharge baskets, over the existing 1/4" screen. Water was routed to a spray nozzle inside each basket to keep fine coal from plugging the screen. In addition, fine mesh baskets were installed at the discharge hose from each mill into the respective discharge tanks. Installation of the high-G centrifuge continued.

July 16

Run 51 was conducted with 200 mesh coal to allow comparison of the two high shear mixers, as planned in Run 50. The first mixer tested was the 45-gallon, with 20% solids, 0.5% diesel, and 1200 rpm. The flotation cells were operated at 6% solids, at various frother addition rates. Frother was added to cells three and five of each stage. The first set point was obtained at 1200 rpm and 0.11 lb/ton frother. The second set point was obtained at 1200 rpm and 0.12 lb/ton frother. High shear feed was switched to the 150-gallon high shear mixer for set point three, with an rpm of 1186 and a frother rate of 0.11 lb/ton. Steady-state samples and flow rate measurements were taken at each set point. Between set points one and two the 45-gallon mixer rpm was changed from 1200 to 1800, then 700, then back to 1200 to allow collection of high shear product samples. After set point two, the 45-gallon mixer rpm was changed from 1200 to 300, then 150, and high shear product samples taken. Mill discharge pump performance was much better during the run due to the installation of the finer screens and baskets at the mill discharges. The finer mill discharge basket at the primary mill collected about 7-8 lbs metal balls, while the basket entering the mill discharge tank collected about 3 lbs metal. About 3 lbs metal was collected discharging from the secondary mill. Installation of the high-G centrifuge continued.

July 17

Installation of the high-G centrifuge continued. The product conveyor was modified to allow placement under the centrifuge, and a centrifuge discharge chute installed.

Run 52 was conducted to allow shakedown of the high-G centrifuge. The unit was operated at 200 mesh, with the 45-gallon high shear mixer at 700 rpm, 20% solids, and various diesel addition rates. The flotation cells were operated at 6% solids and 0.13 lb/ton frother. Frother was added to cells three and five of both stages. Slurry was fed to the centrifuge upon startup, and product samples taken for moisture determination at 0.5, 1.0, and 1.5% diesel to the 45-gallon high shear mixer.

July 20

Run 53 was conducted to test low diesel concentrations in the 45-gallon high shear mixer. Coal was fed to the mixer at 20% solids, with 200 mesh grind. The diesel addition to high shear was 0.125%, with an initial rpm of 1200. The flotation cells were operated at 6% solids, and the frother was added at a 0.32 lb/ton rate. A mass flow

rate test was conducted at cells 1 and 2 of the first stage. One set of samples and flow rate measurements were taken. The rpm was then changed to 700, and a second set of mass flow rate measurements taken. The primary mill pump began operating erratically shortly thereafter, and the run was stopped. No cause could be found for the erratic operation of the pump.

July 21

Although no problems could be identified with the primary mill pump, it was replaced prior to Run 54. **Run 54** was conducted at 100 mesh grind to evaluate dewatering of the high-G centrifuge. The 45-gallon high shear mixer was used at 20% solids, 1.0% diesel, and 700 rpm. The flotation cells were operated at 6% solids and 0.27 lb/ton frother. Steady-state was achieved and routine samples and flow rate measurements taken of flotation feed and discharge streams. Product flow was sent to the centrifuge, which was operated for the first time. The unit did not perform well, with solids discharge erratic and wet. The rpm of the centrifuge was varied, but no visible impact was noticed on product appearance. Flow rate measurements and samples were taken at the centrifuge per the manufacturer's recommendation. (An Alfa-Laval Sharples representative was on site for startup of the unit.) The plant was shut down after two sets of data were collected.

July 22

The ball charge to each mill was increased; with an additional 400 and 600 lbs small/medium balls added to the primary and secondary mill, respectively. A calibration of the weigh feeder was conducted, which indicated accurate performance of the unit.

Run 55 was conducted at 200 mesh with the 150-gallon high shear mixer, which was operated at 1186 rpm and 20% solids, at a diesel addition rate of 0.5%. The flotation cells were operated at 6% solids, with a frother concentration of 0.11 lb/ton. Three recovery points were obtained by varying the blower rates. Steady-state samples and flow rate measurements were obtained at each point. In addition, the centrifuge was again tested to evaluate dewatering at lower diesel rates. Once again, the unit did not perform well, with a product cake that was high in moisture. The cake discharge rate from the centrifuge was inconsistent. During the run, the 150-gallon high shear mixer packing smoked badly.

July 23

No run was conducted. The 45-gallon high shear mixer packing had begun to leak during Run 55, and a replacement seal kit was ordered. The unit was prepared for the Pittsburgh No. 8 demonstration run. A method of recycling balls discharged from the mills was added to the secondary mill feed chute. A leaking seal was repaired on the secondary mill pump. Solids which had accumulated in the sumps and trenches from centrifuge concentrate solids during Run 55 were removed. Water was circulated to remove solids.

July 24

The 45-gallon high shear mixer packing was replaced. Also, work began replacing the pillow block bearing and packing of the 150-gallon high shear mixer.

July 27 - 28

Coal feed for the Pittsburgh No. 8 demonstration run began at 09:23. The run was conducted with 200 mesh coal to the 45-gallon high shear mixer, which was initially operated at 20% solids, 1200 rpm, and 0.5% diesel. The flotation cells were operated at 6% solids and approximately 0.1 lb/ton frother. Product dewatering was done via the rotary drum vacuum filter. Stop belt samples were taken of coal from the crusher discharge, with eight-hour composites. Composite samples were taken of flotation cell inlet and outlet streams over four-hour periods via mechanical samplers. Rotary drum vacuum filter product cake was sampled via grab for moisture determination. Recycle water samples were collected and system pH measurements taken throughout the run. Flow rates were recorded from instrumentation and obtained via manual rate measurements. Steady-state was achieved and the first set of samples collected from 14:00 - 18:00. The mechanical sampler cutter opening was changed to better control sample volume, and the sampling of the second set point started at 18:30. At approximately 19:00, it was noticed that the high shear mixer had tripped off. Flotation cell operation indicated that it had been off only about 30 minutes. Steady-state was reestablished, and three set points obtained. Water supply in the clarified water surge tank was lost for about 10 minutes from 08:05 to 08:15. Flow was again established, and steady state sampling began at 09:00. Steady-state sampling continued into July 29.

July 29

At approximately 10:20, the vacuum pump at the rotary drum vacuum filter skid stopped working. The shaft of the unit could not be turned, and the pump had to be removed and taken to the maintenance shop for repairs. Flotation cell product was diverted to the plate and frame filter presses. One of the filter press feed pumps stopped operating. During the time required to locate the reason for the failure of the pump, the POC had to be shut down at 12:15 due to the inability of the single bank of presses to keep up with the full product flow. The unit was placed back on coal feed at 16:22, once the filter press feed pump had been put back in service. Coal had entered the solenoid of the feed valve through the instrument air supply at the filters, resulting from coal entering the supply from the core blow of the filters. The rotary drum vacuum filter was placed back in service at 19:15. The vacuum pump was taken apart and an excessive buildup of calcium removed from the impellers and walls of the pump housing. The calcium had slowly built up over time due to the hard seal water makeup. Steady state was achieved and no further problems were encountered for the duration of the demonstration run. A total of 16 composite samples were collected at 0.5% diesel.

July 30

At 15:30 on July 30 the diesel concentration to high shear was lowered from 0.5 to 0.125%, per the demonstration run plan. Correspondingly, the frother was increased from 0.1 to 0.3 lb/ton. Steady-state was achieved, and five sets of samples taken. The unit was taken off coal feed at 14:50 on July 31. Coal feed was maintained for 97.16 hours of operation out of 101.45 available hours, for an on-stream factor of 95.77%. Approximately 50 dry tons of product was collected for testing by Combustion Engineering (500 drums), and an additional 16 drums of product was collected for other tests. Upon completion of the run, all remaining Pittsburgh No. 8 coal was transported to the E. C. Gaston coal pile, and the coal pad was cleaned and washed down in preparation for delivery of Illinois No. 5 coal on August 3.

Illinois No. 5 Operating Log

August 3

The first two truckloads of Illinois No. 5 coal arrived. Approximately 600 gals diesel oil no. 2 was added to the diesel tank. A stand was fabricated at the end of each

ball mill for a five-gallon bucket to catch balls exiting from the ends of the mills. Instrument air to the filter press inlet three-way valves was rerouted from a location in the instrument air header upstream of the check valves feeding air to the filter press system, reducing the potential for coal fines to enter the solenoid valves on the valves via the core blow.

The interior of both mills was inspected, and the walls and lifters found to be in good condition. The depth of ball charge was measured at approximately 15.5 " (ID of the mills about 56 - 57") in both mills. The hammermill crusher fingers were inspected and found to be in good condition. The mill was lubricated. The system water was circulated and cleaned of solids, all pads and trenches were cleaned. General unit housekeeping was done. The rotary drum filter blower air intake filter was cleaned.

August 4

The first Illinois No. 5 run, **Run 56**, was conducted with the mills operated in series to check calibration of the mills and begin orientation testing of the system with that coal. The 45-gallon high shear mixer was used at 700 rpm, 0.5% diesel, and 20% solids. The flotation cell solids concentration was 6%, and 0.2 lb/ton frother was added (0.1 lb/ton to each stage). Ball mill samples were taken for particle size analysis, as well as grab samples of individual flotation cells and first and second stage tailings.

August 5

No run was conducted. All insulation at the site except the asphalt tank insulation was stripped and disposed of in the Liquefaction dismantling insulation dumpsters. Several ball valves in the ball mill area were replaced. Site cleanup was done.

August 6

Run 57 was conducted with 100 mesh coal using the 45-gallon high shear mixer, which was operated at 700 rpm, 0.5% diesel, and 20% solids. The flotation cell solids concentration was 6%, and 0.2 lb/ton frother was added (0.1 lb/ton to each stage). Two steady-state periods were obtained, at different frother and blower air rates, and samples and flow rate measurements taken. In addition, mass flow rate measurements of cells 1 and 2 of the first stage were taken.

August 7

Run 58 was conducted to evaluate two different high recovery points with 100 mesh Illinois No. 5 coal. The 45-gallon high shear mixer was used, at 20% solids, 0.5% diesel, and 700 rpm. The flotation cells were operated at 6% solids, with a frother concentration of 0.24 lb/ton for the first recovery point, and 0.2 lb/ton for the second, lower recovery point. Two sets of composite samples were collected at each recovery point, and flow rate measurements taken.

August 10

Run 59 was conducted to evaluate the effects of high shear rpm on coal recovery by measuring the froth mass flow rate of the first two cells of the first stage of flotation. The run was made with 100 mesh coal, using the 45-gallon high shear mixer at 20% solids, 0.5% diesel, and varying rpm's. The first stage of flotation was operated at 6% solids, with no frother. At 1200, 700, and 400 rpm two blower rates at each of two cell levels were obtained, for a total of 12 set points. The flow rate of froth from cells 1 and 2 of the first stage was measured. Samples were collected of the two cells, total flow from both cells, and the feed to the first stage. During the run cells 4 - 6 of the first stage were operated at very low levels to prevent interference with the froth flow measurement of cells 1 and 2. As a result, solids loading in the tailings was much higher than normal operation. Both filter presses were needed to keep up with the solids loading.

August 11

No run was conducted because only one filter press feed pump could be operated. The second pump would not start, due to a trip signal from within the filter press control panel. A froth mass flow test similar to Run 59 was planned (but with frother addition), but without both filter presses in operation, solids removal could not be obtained. Maintenance worked on the filter control panel all day, but the specific cause of the problem could not be identified.

August 12

A loose solenoid contact was found in the filter press feed valve, which had not allowed the control panel to sense when the valve was in the correct position for feed to enter the band of filter elements. This was corrected, and the condition satisfied for the panel to advance to the next step in the sequence of filter operation.

Run 60 was conducted with 100 mesh coal to evaluate use of high frother rates. The 45-gallon high shear mixer was used, at 20% solids, 0.5% diesel, and 700 rpm. The flotation cells were operated at 6% solids, with frother at 0.45 lb/ton. Three set points were obtained, at different recovery points with higher frother rates. Steady-state samples and flow rate measurements were taken at each set point.

August 13

Run 61 was conducted to continue measurements of flotation rate within cells 1 and 2 of the first stage. The POC was operated with 100 mesh coal to the 45-gallon high shear mixer. The mixer was operated at 20% solids, 0.5% diesel, and a range of rpm's (1200, 700, 400, and 0). The flotation cells were operated at 6% solids, with a frother rate to cell 1 of 0.07 lb/ton. Cell levels and air rates were varied throughout the run. Samples of cells 1 and 2 were taken at each set point, as well as a composite of the total froth from the two cells. Froth flow and height were measured for each point.

August 14

Run 62 was conducted to evaluate minimum frother concentration and lower high shear rpm, as well as performance of the high-G centrifuge for dewatering product slurry. The POC was operated with 100 mesh coal to the 45-gallon high shear mixer, which was operated at 20% solids, 0.5% diesel, and an initial rpm of 700. The flotation cells were operated at 6% solids, and a frother concentration of 0.1 lb/ton. An additional 1.0% diesel was added to the second stage flotation product tank to facilitate dewatering of the product slurry with the centrifuge. Two set points were obtained, at 700 and 400 rpm in the 45-gallon high shear mixer. Steady-state samples and flow rate measurements were taken.

August 17

Run 63 was conducted with 100 mesh Illinois No. 5 coal at a 2 tph feed rate, using the 45-gallon high shear mixer at 400 and 700 rpm and 0.5% diesel addition. A major objective of the run was to evaluate dewatering of the product with the high-G centrifuge. Frother concentration was 0.15 lb/ton. The solids concentration to high shear and the flotation cells was 20 and 6%, respectively. Two set points were obtained, and steady-state samples and flow rate measurements taken. The centrifuge was tested at an rpm of 4000, with no dam in the bowl (allowing the shallowest pool depth, and therefore driest

product). Second stage flotation product was diverted to the centrifuge at several different flow rates. In addition, 1.5% diesel was added to the second stage product tank to evaluate the impact of increased oil levels on product dewatering performance.

August 18

Run 64 was started with 100 mesh Illinois No. 5 coal using the 45-gallon high shear mixer, which was operated at 700 rpm and 0.25% diesel. The flotation cells were operated at 6% solids, with frother addition varied as needed to achieve recovery. The run was stopped after about 2 hours due to pluggage of coal in the bottom of the 3-ton surge bin. The coal in the 24-ton storage tank had gotten too wet. The coal was emptied from the storage bin and the coal turned over as much as possible to improve drying.

August 19

Run 64 was continued with 2 tph 100 mesh Illinois No. 5 coal using the 45-gallon high shear mixer, which was operated at 20% solids, 700 rpm and an initial diesel addition rate of 1.0 %. The flotation cells were operated at 6% solids and the frother rates changed as needed for recovery (frother was 0.152, 0.107, and 0.054 lb/ton for the three set point, respectively). Steady-state was achieved for the first two set points at different recovery points, and samples and flow rate measurements taken. In addition, the dewatering performance of the high-G centrifuge was tested, with a pool depth as low as possible (achieved by removing the dam) and an rpm of 4048. Three different feed rates to the centrifuge were tested at 1% diesel. Diesel addition to high shear was then increased to 1.5% to facilitate dewatering of the product with the centrifuge, and steady-state samples and flow rate measurements taken for the third set point. The centrifuge was again tested at 4048 rpm and a low pool depth, with three different feed rates.

August 20

Run 65 was conducted with 100 mesh Illinois No. 5 coal (2 tph), using the 45-gallon high shear mixer. The mixer was operated at 700 rpm, 20% solids, and 0.25% diesel. The flotation cells were operated at 6% solids and 0.25 lb/ton frother. Three recovery points were obtained by adjusting the blower rates on the cells. Steady-state samples and flow rate measurements were obtained at each set point.

August 21

Run 66 was conducted with 100 mesh Illinois No. 5 coal to obtain several recovery points at 1% diesel. Coal was fed to the 45-gallon high shear mixer at the normal 2 tph rate. The high shear was operated at 700 rpm, 1% diesel, and 20% solids. The flotation cells were operated at 6% solids and an initial frother addition rate of 0.15 lb/ton. Two recovery points were obtained at the initial settings, with collection of routine steady-state samples and flow rate measurements. The recovery was then decreased somewhat by reducing the frother to 0.12 lb/ton, and samples and flow rate measurements collected for a third set point.

August 24

Run 67 was conducted with 200 mesh Illinois No. 5 coal to evaluate the unit performance at various diesel addition rates. Coal was fed to the 45-gallon high shear mixer at a 2 tph rate. The mixer was operated at 700 rpm and 20% solids, and diesel rates of 0.5, 0.75, and 1.0%. The flotation cells were operated at 6% solids, and varying frother addition rates as needed for recovery. Frother addition for the three set points was 0.21, 0.10, and 0.09 lb/ton, respectively. Steady-state was achieved at each set of conditions and samples and flow rate measurements taken.

August 25

Run 68 was conducted with 100 mesh Illinois No. 5 coal to evaluate three diesel addition rates. Coal was fed to the 45-gallon high shear mixer at a 2 tph rate. The mixer was operated at 700 rpm, 20% solids, and diesel addition rates of 0.7, 0.9, and 0.8%, respectively. The flotation cells were operated at 6% solids, and frother and blower rates as needed for recovery. Frother addition for the three set points was 0.14, 0.12, and 0.11 lb/ton, respectively. Steady-state samples and flow rate measurements were taken at each of the three set points. The high-G centrifuge was shipped back to the vendor.

August 26

Run 69 was conducted with 2 tph 100 mesh Illinois No. 5 coal. Initially, the run was scheduled to test performance of the 45-gallon high shear mixer at 600 rpm and low diesel addition rates. However, the 45-gallon high shear mixer could not be started. While the maintenance personnel repaired the mixer, the unit was started up bypassing high shear, and two set points obtained.

Solids concentration at the diesel addition point was maintained at the normal 20% level, and diesel added at 0.2 and 0.15% for set points one and two, respectively. The solids concentration to the flotation cells was kept at 6%, and frother addition, blower rates, and cell levels adjusted as needed to obtain required recovery. Frother for set points one and two was 0.36 and 0.43 lb/ton. During set point two, the 45-gallon high shear mixer became operational after replacing a board in the controller. A third set point was obtained using the 45-gallon mixer at an rpm of 600, and a diesel addition of 0.3%. Frother addition was 0.24 lb/ton. Steady-state samples and flow rate measurements were taken at each set point.

August 27

The filter press feed pumps were repacked.

August 28

Run 70 was conducted with 200 mesh Illinois No. 5 coal to evaluate various rpm and diesel rates. Coal was fed to the 45-gallon high shear mixer at 2 tph. The mixer was operated at the normal 20% solids concentration, and the flotation cells were operated at 6% solids. Three recovery points were obtained. Set point one was made with a mixer rpm of 500, diesel addition of 0.15%, and a frother rate of 0.41 lb/ton. Set point two was conducted at 700 rpm, 0.3% diesel, and 0.25 lb/ton frother. Set point three was obtained at 600 rpm, 0.6% diesel, and 0.20 lb/ton frother. Steady-state samples and flow rate measurements were taken at each of the set points.

August 31

Run 71 was started with 100 mesh Illinois No. 5 coal, but had to be stopped due to an instrument air compressor failure. The compressor had to be serviced, and unit orifice taps blown out.

September 1

Run 71 was conducted with 100 mesh Illinois No. 5 coal to continue collection of data for recovery at a wide range of conditions for Ultramax. The 45-gallon high shear mixer was operated at 20% solids and various rpm and diesel addition rates. The flotation cells were operated at 6% solids. Set point one was made with a mixer rpm of 600, diesel addition of 0.8%, and a frother rate of 0.07 lb/ton. Set point two was conducted at 700 rpm, 0.5% diesel, and 0.16 lb/ton frother. Set point three was obtained at 800 rpm, 0.25% diesel, and 0.21 lb/ton frother.

Steady-state samples and flow rate measurements were taken at each of the set points.

September 2

Run 72 was conducted with 100 mesh Illinois No. 5 coal at various diesel addition rates and rpm to high shear. In addition, both the 150-gallon and 45-gallon high shear mixers were operated, both with 20% solids. The flotation cells were operated at 6% solids. A total of five set points were obtained, and steady-state samples and flow rate measurement taken at each set point. The first four set points, with the 45-gallon high shear mixer, were obtained with rpm's of 400, 600, 500, and 1200, respectively. The diesel addition was 0.15, 0.25, 0.3, and 0.3%, and frother rates 0.35, 0.27, 0.22, and 0.27 lb/ton. The fifth set point, using the 150-gallon mixer, was obtained at an rpm of 848, diesel of 0.3%, and a frother addition of 0.26 lb/ton.

September 3

Run 73 was conducted with 100 mesh Illinois No. 5 coal to repeat previous conditions that had the most promising results, to help in finalizing run conditions for the Illinois No. 5 demonstration run. The 45-gallon mixer was used throughout the run, with 20% solids to high shear, and 6% solids to flotation. Four set points were obtained, and steady-state samples and flow rate measurements taken. Set point one was obtained at an rpm of 700, diesel addition to high shear of 0.7%, and a total frother rate of 0.14 lb/ton. Set point two was made at 600 rpm, 0.6% diesel, and 0.18 lb/ton frother. Set point three had an rpm of 700, diesel addition of 0.9%, and a frother rate of 0.13 lb/ton. Set point four was obtained at 500 rpm, 0.35% diesel, and 0.21 lb/ton frother.

September 4

No run was conducted. Preparations were made for the demonstration run. The vacuum pump was cleaned with acid to prevent shutdown during the run due to buildup of calcium deposits, the tailings pump suction was cleared of trash, and the recycle water was cleaned.

September 8 - 10

The Illinois No. 5 demonstration run was started at 0755 with 100 mesh coal at a 2 tph feed rate. The run was conducted with the 45-gallon high shear mixer, which was operated at 900 rpm and 20% solids. Diesel addition to high shear was 0.8%. The flotation cells were

operated at a solids concentration of 6%, with 0.125 lb/ton frother. Blower rates and cell levels were adjusted to obtain desired recovery. The coal feed to the plant was sampled via stop belt, with eight-hour composites. Composite samples of flotation feed, products, and tailings were collected for each four-hour steady-state period using mechanical samplers. A grab sample of vacuum drum filter product was collected for product moisture, and the pH of the recycle water was monitored. Flow rate measurements were taken via instrumentation or manual rate checks for all streams during each four-hour sampling period. In addition, vacuum drum filter product was drummed for Combustion Engineering.

Steady-state was achieved and the first four-hour sampling period obtained from 1115 to 1515. Sampling continued for the first six four-hour periods until 1215 on September 9, when a plant upset occurred. The frother levels in the buckets feeding the metering pumps had gotten low, and frother was added from a new drum. This drum was Betz MS150, not M150. (An order for M150 placed with Betz had resulted in the shipment of Betz MS150, due to the sale of the mining chemicals group of Betz to Stockhausen Mining. As a result, Betz no longer carried M150, and apparently sent the item most similar to M150 in name: MS150.) The MS150 did not work at all, so a drum of M150 was expedited from the local Stockhausen Mining representative, and arrived about 1700. In the meantime, there was enough M150 frother in another bucket to divide up to all frother metering pumps until the new drum arrived. Steady-state was again achieved, and sampling begun at 1515. The high shear mixer began to leak somewhat at the seal, with no method of stopping it. Operation continued with the leak, as opposed to shutting down. A power outage occurred at about 1900 on September 10. The unit was off line for only a few minutes, so steady-state was not jeopardized. However, the vacuum drum filter vacuum pump had locked up during the outage, and would not turn. It was acid washed in place, and put back on line at 2115. During the interim, product was diverted to the plate and frame filter presses. An additional 12 set points were obtained through 1515 on September 11, for a total of 18 set points at the initial conditions with 0.8% diesel. Drumming of vacuum drum filter product for Combustion Engineering was completed at 0530 on September 10. A total of 500 drums was obtained. In addition, 16 drums of product was obtained for project use.

September 11 - 12

The coal grind was diverted to the secondary mill at 1420 on September 11 to begin processing 200 mesh coal for the second set of conditions. The 45-gallon high shear mixer rpm was increased from 900 to 1050, and the diesel addition rate increased from 0.8 to 1.2%. Over the next 30 minutes the leakage at the 45-gallon high shear seal increased significantly as the smaller particles began to flow through the mixer. Flow was diverted to the 150-gallon high shear mixer at 1510, at an rpm of 848. Steady-state was achieved at the new set of conditions, and three-hour composite sampling periods begun at 1630 on September 11. A total of five sets of samples and flow rate measurements were obtained. The unit was taken off coal feed at 0800 on September 12, after 96 hours of continuous coal feed.

Pond Fines Operating Log**September 14 - 23**

Upon completion of Illinois No. 5 POC operation, modifications were made to the unit to allow the processing of pond fines. The material did not require grinding, so the ball mills and associated tanks and pumps were removed from the structure. A feed tank was then installed into which pond fines could be introduced to the system. A conveyor was leased and installed beneath the existing reclaim hopper to feed the pond fines into the feed tank, which was equipped with a mixer and pump. The slurry was pumped from the feed tank through a low shear vessel (to help mix the slurry) to static sieve bends for separation into +/- 60 mesh. The + 60 mesh was discharged onto a second leased conveyor and conveyed to the product pad for disposal. The - 60 mesh slurry was pumped to the 150-gallon high shear mixer, and from there to the flotation cells per normal operation.

September 24

A pond fines shakedown run was made to test the equipment and piping. Upon shutdown, the need for system flush water was identified, so modifications were made to the recycle water header to allow introduction of water to the low shear vessel.

September 25

A pond fines run was made to determine performance of the system. Coal feed was approximately 2.7 tph (dcb)

to the high shear mixer. Diesel was added at approximately 0.1%, and frother added as needed for recovery.

September 28

No run was made.

September 29

A run was made with 0.06 lb/min diesel, or approximately 0.05%, to operate in standard flotation mode that Jim Walter Resources uses. Only two set points were achieved due to power outages. The first set point was made with no high shear, the second with the 150-gallon high shear mixer. Frother addition and blower rates were adjusted as needed for recovery, based on observation of froth. Samples and flow rate measurements were taken to allow generation of performance curves.

September 30

A run was made to evaluate the effect of diesel on performance. Two set points were achieved, both using the 150-gallon high shear mixer. The first set point was conducted at 0.1% diesel, and the second at 0.25% diesel. Coal feed rate was targeted at approximately 2.5-3 tph. Frother addition and blower rates were adjusted as needed for recovery, based on observation of froth. Flow rate measurements and samples were collected at each set point.

October 1

A run was made to continue evaluation of performance at different diesel concentrations. Set point one was conducted with 0.5% diesel to high shear, and set point two was conducted at 0.1% diesel. The 150-gallon high shear mixer was used. Frother addition and blower rates were adjusted as needed for recovery, based on observation of froth. Samples and flow rate measurements were taken at each set point.

October 2

The final pond fines run was conducted, to evaluate the effect of using Betz M210 type promoter with the diesel addition (90/10 diesel/M210 blend). Three set points were obtained. Set point one was conducted with 0.12 lb/min blend, added to the sieve bend underflow tank from which high shear feed slurry originates, and the 150-gallon high shear mixer. Set point two was

conducted with the same addition rate of diesel/M210, but the 150-gallon high shear mixer was bypassed. Set point three was conducted with the 150-gallon high shear mixer, but the diesel/M210 addition was decreased from 0.12 to 0.06 lb/min. Samples and flow rate measurements were taken at each set point.

October 5

Dismantling of the Selective Agglomeration unit began.

Appendix E

POC RESULTS SUMMARY

Table E-1
UPPER FREEPORT POC DEMONSTRATION RUN CONDITIONS AND RESULTS SUMMARY

RUN	TOP SIZE	SET PNT	HS VESS	HS RPM	HS		FLT %	TOTAL lb/ft	FEED			FINAL PRODUCT			COMBINED TAILINGS			AGGLOMERATION PERFORMANCE			ROM PERFORMANCE					
					%	SOL			%	ASH	TS	%	ASH	TS	%	ASH	TS	%	YIELD	PSR	%BTU REC	%	YIELD	PSR	%	%BTU REC
DEM	200M	1	45g	950	20	0.5	6	0.29	11.0	1.42	5.9	1.06	48.5	3.77	88.0	53.8	93.7	81.0	80.7	90.6	81.0	80.7	90.6			
		2	45g	950	20	0.5	6	0.29	10.9	1.44	6.5	1.09	57.9	4.33	91.5	49.3	96.5	84.1	78.8	93.3	84.1	78.8	93.3			
		3	45g	950	20	0.5	6	0.29	10.0	1.39	5.9	1.04	46.8	3.70	89.8	53.6	94.5	82.6	80.6	91.4	82.6	80.6	91.4			
		4	45g	950	20	0.5	6	0.29	10.0	1.47	6.1	1.10	51.4	4.33	91.5	50.9	95.8	84.1	79.5	92.7	84.1	79.5	92.7			
		5	45g	950	20	0.5	6	0.29	10.1	1.41	6.2	1.11	43.2	3.52	89.3	47.4	93.7	82.2	78.0	90.6	82.2	78.0	90.6			
		6	45g	950	20	0.5	6	0.29	10.1	1.50	6.0	1.15	41.2	3.84	88.4	49.1	92.9	81.3	78.7	89.8	81.3	78.7	89.8			
		7	45g	950	20	0.5	6	0.29	10.0	1.42	6.0	1.19	39.6	3.32	88.2	40.0	92.5	81.2	74.9	89.5	81.2	74.9	89.5			
		8	45g	950	20	0.5	6	0.29	10.3	1.42	6.0	1.13	32.2	2.79	83.8	49.3	88.2	77.1	78.8	85.3	77.1	78.8	85.3			
		9	45g	950	20	0.5	6	0.29	10.4	1.42	5.9	1.08	32.9	2.71	83.3	54.6	88.0	76.6	81.0	85.1	76.6	81.0	85.1			
		10	45g	950	20	0.5	6	0.29	10.1	1.48	5.9	1.12	40.6	3.57	88.1	51.4	92.6	81.0	79.7	89.6	81.0	79.7	89.6			
		11	45g	950	20	0.5	6	0.29	10.2	1.41	6.5	1.22	36.5	2.88	87.6	36.2	91.6	80.6	73.3	88.6	80.6	73.3	88.6			
		12	45g	950	20	0.5	6	0.29	9.4	1.47	6.1	1.15	39.3	3.10	90.0	45.5	93.7	82.8	77.2	90.6	82.8	77.2	90.6			
		13	45g	950	20	0.5	6	0.29	9.8	1.41	6.0	1.06	32.2	2.69	85.7	54.9	89.7	78.8	81.1	86.7	78.8	81.1	86.7			
		14	45g	950	20	0.5	6	0.29	9.8	1.41	5.9	1.17	31.6	2.87	84.8	44.0	88.9	78.0	76.6	86.0	78.0	76.6	86.0			
		16	45g	950	20	0.5	6	0.29	9.9	1.54	6.1	1.14	30.9	3.08	84.4	53.8	88.5	77.7	80.7	85.6	77.7	80.7	85.6			
		17	45g	950	20	0.5	6	0.29	9.3	1.43	5.7	1.17	38.4	3.55	89.1	42.4	93.0	82.0	75.9	90.0	82.0	75.9	90.0			
		18	45g	950	20	0.5	6	0.29	9.4	1.37	6.0	1.09	43.4	3.58	90.8	45.7	94.7	83.6	77.3	91.6	83.6	77.3	91.6			
		19	45g	950	20	0.5	6	0.29	9.6	1.47	6.1	1.16	49.2	4.13	91.7	44.0	95.7	84.4	76.6	92.6	84.4	76.6	92.6			
		20	45g	950	20	0.5	6	0.29	9.8	1.44	6.1	1.18	51.7	4.12	92.0	39.6	96.2	84.7	74.8	93.0	84.7	74.8	93.0			
		21	45g	950	20	0.5	6	0.29	9.9	1.45	6.1	1.19	43.0	3.73	89.7	40.4	93.9	82.6	75.1	90.9	82.6	75.1	90.9			
		22	45g	950	20	0.5	6	0.29	10.4	1.48	6.3	1.13	45.9	3.65	89.7	48.8	94.3	82.5	78.6	91.2	82.5	78.6	91.2			
		23	45g	950	20	0.1	6	0.08	10.3	1.43	6.2	1.10	45.7	3.80	89.7	48.9	94.2	82.5	78.6	91.1	82.5	78.6	91.1			
		24	45g	950	20	0.1	6	0.09	10.1	1.25	5.7	0.86	41.5	3.67	87.6	68.8	92.4	80.6	87.0	89.4	80.6	87.0	89.4			

E-3

Table E-2 (Continued)
UPPER FREEPORT COAL POC PARAMETRIC TESTING CONDITIONS AND RESULTS SUMMARY

RUN	SIZE	TOP SET PNT	HS	VESEL	RPM	HS		SOL	DIES	FLT	TOTAL	FEED		FINAL PRODUCT		COMBINED TAILINGS		AGGLOMERATION PERFORMANCE		ROM PERFORMANCE	
						%	%	%	%	%	%	%	%	%	%	%	%	%	%	%	%
						ASH	TS	ASH	TS	ASH	TS	ASH	TS	YIELD	PSR	REC	%BTU	YIELD	PSR	REC	%BTU
11	200M	1	150g	1186	15	0.5	6	0.18	10.1	1.66	6.6	1.32	39.5	4.9	89.5	41.1	93.5	83.2	75.0	92.0	
12	200M	1	150g	1186	15	0.5	6	0.15	10.3	1.58	6.5	1.24	63.4	6.4	93.4	41.6	97.9	86.8	75.2	96.3	
		2	45g	1800	15	0.5	6	0.15	10.3	1.65	6.9	1.30	60.5	5.0	93.6	39.2	97.8	87.1	74.1	96.2	
13	100M	1	150g	1186	15	0.5	6	0.15	10.0	1.51	6.4	1.22	42.8	3.7	90.0	41.2	94.2	83.7	75.0	92.7	
		2	150g	1186	15	1.0	6	0.15	10.3	1.57	6.7	1.30	42.8	3.2	90.1	36.8	94.2	83.8	73.1	92.7	
		3	150g	1186	15	0.8	6	0.15	9.4	1.39	7.0	1.34	48.9	3.2	94.3	13.5	97.2	87.7	63.2	95.6	
14	200M	1	150g	1186	15	0.8	6	0.15	9.5	1.59	6.4	1.30	53.3	6.6	93.4	36.5	97.1	86.9	73.0	95.5	
		2	150g	1186	15	0.5	6	0.15	10.1	1.51	6.4	1.11	54.2	6.0	92.3	50.8	96.7	85.9	79.1	95.1	
15	200M	1	150g	1186	15	0.5	6	0.15	9.4	1.55	6.3	1.22	46.3	5.2	92.3	42.2	95.9	85.8	75.4	94.4	
		2	150g	1186	15	0.5	6	0.15	10.0	1.68	6.6	1.30	45.9	4.8	91.2	42.7	95.2	84.8	75.6	93.7	
		3	150g	1186	15	0.5	6	0.15	10.6	1.53	6.9	1.23	46.1	4.3	90.6	40.9	94.9	84.3	74.9	93.4	
16	200M	1	150g	848	15	0.5	6	0.16	9.8	1.59	6.6	1.22	51.1	5.1	93.2	43.6	96.8	86.7	76.0	95.2	
		2	150g	848	15	0.5	6	0.16	9.7	1.57	6.6	1.23	52.1	4.9	93.1	41.7	96.8	86.5	75.2	95.3	
		3	150g	848	20	0.5	6	0.16	9.8	1.53	6.8	1.24	55.3	5.1	93.8	37.5	97.4	87.2	73.4	95.8	
		4	150g	848	20	0.5	6	0.16	10.0	1.61	6.8	1.22	55.2	5.3	93.5	44.6	97.3	86.9	76.4	95.7	
		5	45g	1800	20	0.5	6	0.16	9.8	1.63	6.7	1.27	58.6	6.1	94.0	40.8	97.8	87.4	74.8	96.2	
		6	45g	1200	20	0.5	6	0.16	10.1	1.73	6.8	1.17	49.3	5.8	92.2	55.3	96.1	85.8	81.0	94.6	
17	200M	1	150g	1186	20	0.5	6	0.16	10.2	1.62	7.3	1.39	55.3	4.5	94.0	28.5	97.5	87.4	69.6	95.9	
		2	45g	1800	15	0.5	6	0.16	10.0	1.58	7.1	1.32	59.6	4.2	94.4	31.8	98.0	87.8	71.0	96.4	
		3	45g	1200	15	0.5	6	0.16	10.2	1.68	7.3	1.42	61.6	5.2	94.6	29.7	98.2	87.9	70.1	96.6	

Table E-2 (Continued)
UPPER FREEPORT COAL POC PARAMETRIC TESTING CONDITIONS AND RESULTS SUMMARY

RUN	TOP SIZE	SET PNT	HS VESSEL	HS RPM	HS		HS DIES	FLT SOL	TOTAL lb/t	FEED		FINAL PRODUCT		COMBINED TAILINGS		AGGLOMERATION PERFORMANCE			ROM PERFORMANCE							
					%	SOL				%	ASH	%	TS	%	ASH	%	TS	%	YIELD	PSR	%	%	YIELD	PSR	%	%
18	200M	1	45g	190	15	0.1	8	0.08	10.3	1.57	5.3	1.09	29.3	2.8	79.1	62.7	84.2	73.5	84.1	82.8						
		2	45g	1800	15	0.1	8	0.08	9.7	1.53	4.7	0.97	25.0	2.6	75.6	72.6	80.4	70.3	88.4	79.1						
		3	45g	1800	15	0.3	8	0.11	10.3	1.53	7.1	1.42	64.6	4.4	94.4	18.9	98.3	87.8	65.5	96.7						
		4	45g	190	15	0.3	8	0.11	10.3	1.42	7.7	1.34	66.1	3.6	95.5	15.8	98.7	88.9	64.2	97.2						
19	200M	1	45g	1200	20	0.5	6	0.16	8.1	1.44	7.2	1.27	60.9	4.7	98.3	22.6	99.4	91.4	67.1	97.8						
		2	45g	950	20	0.5	6	0.16	9.8	1.63	7.0	1.27	63.9	4.4	95.0	40.0	98.5	88.4	74.5	96.9						
		3	45g	500	20	0.5	6	0.16	10.1	1.56	7.6	1.37	48.8	3.9	94.0	26.2	97.0	87.4	68.6	95.4						
20	200M	1	45g	950	20	0.5	6	0.16	10.0	1.51	7.1	1.18	53.7	5.1	93.9	42.2	97.3	87.3	75.4	95.7						
		2	45g	950	20	0.5	6	0.16	9.8	1.49	6.9	1.24	51.6	4.8	93.4	34.9	96.9	86.9	72.3	95.4						
		3	45g	950	20	0.5	8	0.16	9.6	1.44	7.5	1.23	78.7	3.9	97.1	28.9	99.7	90.3	69.8	98.1						
		4	45g	950	20	0.5	8	0.16	10.2	1.55	7.7	1.43	57.6	4.3	95.1	18.6	98.1	88.4	65.4	96.5						
21	200M	1	45g	950	20	0.5	6	0.30	10.0	1.52	6.9	1.20	76.0	5.5	95.5	40.2	99.2	88.8	75.0	95.9						
		2	45g	950	20	0.5	6	0.30	11.3	1.46	6.4	1.17	55.9	4.1	90.1	43.6	95.7	83.8	76.4	92.5						
		3	45g	950	20	0.5	8	0.30	12.1	1.60	6.6	1.12	64.8	3.8	90.5	56.0	96.9	84.1	81.6	93.7						
		4	45g	950	20	0.5	8	0.30	12.6	1.52	6.9	1.16	53.5	3.5	87.6	49.9	94.1	81.5	79.1	91.0						
22	200M	1	45g	1200	20	0.5	6	0.16	11.9	1.40	6.2	1.07	64.0	3.5	90.1	51.0	96.6	83.8	79.5	93.5						
		2	45g	950	20	0.5	6	0.16	11.9	1.43	6.3	1.09	66.0	3.5	90.6	50.7	97.0	84.3	79.4	93.9						
		3	45g	600	20	0.5	6	0.16	11.2	1.36	6.2	1.11	56.4	3.5	90.1	43.1	95.7	83.8	76.2	92.6						
23	200M	1	45g	1200	20	0.5	6	0.16	12.3	1.52	6.7	1.12	70.9	4.2	91.3	51.7	97.8	84.9	79.8	94.6						
		2	45g	1200	20	0.5	6	0.16	12.0	1.45	6.5	1.06	71.4	4.4	91.6	54.1	97.9	85.1	80.8	94.7						
		3	45g	1200	20	0.5	6	0.26	11.5	1.26	6.3	0.87	62.3	3.8	90.6	67.4	96.6	84.3	86.4	93.5						

Table E-2 (Continued)
UPPER FREEPORT COAL POC PARAMETRIC TESTING CONDITIONS AND RESULTS SUMMARY

RUN	TOP SIZE	SET PNT	HS VESSEL	HS RPM	HS		FLT SOL	TOTAL LBS	FEED		FINAL PRODUCT		COMBINED TAILINGS		AGGLOMERATION PERFORMANCE				ROM PERFORMANCE						
					%	SOL			%	DIES	%	%	%	%	%	%	%	%	%	%	%	%	%	%	%
					ASH	TS			ASH	TS	ASH	TS	ASH	TS	YIELD	PSR	REC	%BTU	YIELD	PSR	REC	%BTU	YIELD	PSR	REC
24	200M	1	45g	1200	20	0.5	6	0.16	11.0	1.29	6.1	0.85	62.4	4.6	91.3	71.3	96.9	84.9	88.0	93.7					
		2	45g	1200	20	0.5	6	0.16	10.9	1.33	5.8	0.82	57.1	4.5	90.2	77.6	95.9	83.9	90.0	92.7					
		3	45g	1200	20	0.5	6	0.29	11.1	1.35	5.9	0.89	41.0	2.7	85.2	70.6	90.8	79.2	87.7	87.8					
25	200M	1	45g	1200	20	0.5	6	0.29	10.4	1.32	6.0	1.11	50.3	4.0	90.2	40.0	95.0	83.8	74.9	91.9					
		2	45g	1200	20	0.5	6	0.29	9.2	1.43	5.5	1.06	20.7	2.4	75.8	60.8	79.2	70.5	83.6	76.6					
26	200M	1	45g	1200	20	0.5	6	0.29	9.9	1.29	6.0	1.02	51.2	4.2	91.5	48.2	95.8	85.1	78.4	92.7					
		2	45g	1200	20	0.5	6	0.25	10.3	1.32	6.5	1.03	65.2	4.6	93.5	51.8	97.9	87.0	79.9	94.7					
27	200M	1	45g	1200	20	0.5	6	0.29	10.5	1.45	6.1	1.08	45.1	3.8	88.8	53.1	93.7	82.6	80.4	90.6					
		2	45g	1200	20	0.5	6	0.57	10.6	1.32	6.6	1.10	59.8	4.4	92.4	39.5	97.1	85.9	74.7	93.9					
		3	45g	1200	20	0.5	8	0.57	10.7	1.35	7.4	1.11	75.3	4.2	95.1	38.2	99.1	88.5	74.2	95.8					
		4	45g	1200	20	0.5	8	0.29	10.2	1.34	7.2	1.1	60.1	3.7	94.3	37.9	97.8	87.7	74.0	94.6					
28	200M	1	45g	1200	20	0.5	6	0.29	9.6	1.45	6.1	1.19	40.0	4.0	89.7	40.5	93.6	83.5	75.1	90.5					
		2	45g	1200	20	0.5	6	0.29	9.5	1.43	6.2	1.28	49.0	4.9	92.2	27.2	96.0	85.7	69.6	92.9					
		3	45g	1200	20	0.5	6	0.29	9.8	1.47	6.5	1.18	37.2	3.3	89.1	43.2	92.8	92.9	76.3	89.8					
29	200M	1	45g	500	20	0.5	6	0.29	9.3	1.51	5.5	1.05	27.7	3.0	82.6	62.3	86.6	76.8	84.2	83.7					
		2	45g	950	20	0.5	6	0.29	9.6	1.62	5.9	1.08	46.1	4.8	90.8	59.8	94.9	84.4	83.2	91.8					
		3	45g	190	20	0.5	6	0.29	9.3	1.32	6.0	0.81	34.2	3.4	88.2	78.0	91.8	82.0	90.8	88.8					
		4	45g	1200	20	0.5	6	0.29	9.7	1.47	6.7	0.89	54.0	4.7	93.6	71.6	97.1	87.1	88.2	93.9					
30	200M	1	45g	950	20	0.5	6	0.29	9.9	1.42	5.7	1.01	34.7	3.9	85.6	60.6	90.0	79.6	83.5	87.1					
		2	45g	950	20	0.5	6	0.29	10.0	1.44	6.2	1.13	37.4	3.9	88.0	47.4	92.1	81.8	78.0	89.1					
		3	45g	950	20	0.5	6	0.29	10.0	1.42	6.0	1.11	32.7	3.3	85.2	50.1	89.4	79.2	79.1	86.4					
		4	45g	950	20	0.5	6	0.29	10.4	1.42	6.1	1.13	34.1	3.4	84.5	48.6	89.1	78.6	78.5	86.1					

Table E-2 (Continued)
UPPER FREEPORT COAL POC PARAMETRIC TESTING CONDITIONS AND RESULTS SUMMARY

RUN	TOP SIZE	SET PNT	HS VESSEL	HS RPM	HS		TOTAL FROTH	FEED			FINAL PRODUCT			COMBINED TAILINGS			AGGLOMERATION PERFORMANCE			ROM PERFORMANCE		
					%	SOL	%	lb/r	%	ASH	%	TS	%	ASH	%	TS	%	YIELD	%	PSR	%	%BTU REC
31	200M	1	45g	950	20	0.11	6	0.12	10.5	1.36	5.3	0.84	39.5	3.3	84.8	77.4	90.3	78.8	90.6	87.3		
		2	NA	0	20	0.11	6	0.12	10.7	1.46	5.4	0.89	43.8	3.9	86.3	74.6	92.0	80.3	89.4	89.0		
		3	45g	500	20	0.11	6	0.12	10.5	1.40	5.4	0.91	43.7	4.0	86.6	70.5	92.1	80.5	87.7	89.1		
		4	45g	1200	20	0.11	6	0.13	11.2	1.53	5.0	0.91	35.9	3.3	79.9	77.1	86.1	74.3	90.4	83.3		

Table E-3
PITTSBURGH NO. 8 POC DEMONSTRATION RUN CONDITIONS AND RESULTS SUMMARY

RUN	TOP SIZE	SET PNT	HS VESS	HS RPM	HS		FLT %	TOTAL lb/t	FEED			FINAL PRODUCT			COMBINED TAILINGS			AGGLOMERATION PERFORMANCE			ROM PERFORMANCE									
					%	SOL			%	DIESEL	%	SOL	%	ASH	%	TS	%	ASH	%	TS	%	YIELD	%	PSR	%	YIELD	%	PSR	%	REC
DEM	200M	1	45g	1200	20	0.5	6	0.14	7.16	2.08	5.48	1.83	26.49	4.34	92.0	32.4	93.9	61.9	62.2	87.6										
		2	45g	1200	20	0.5	6	0.10	7.52	2.16	5.24	1.81	26.98	4.51	89.5	40.5	92.0	60.3	66.7	85.9										
		3	45g	1200	20	0.5	6	0.10	7.17	2.03	4.84	1.77	28.42	4.21	90.1	36.4	92.7	60.7	64.4	86.5										
		4	45g	1200	20	0.5	6	0.10	7.44	2.21	4.89	1.85	29.91	4.39	89.8	39.9	92.6	60.5	66.4	86.4										
		5	45g	1200	20	0.5	6	0.10	7.60	2.28	5.14	1.92	32.86	4.95	91.1	37.7	93.9	61.4	65.2	87.6										
		6	45g	1200	20	0.5	6	0.12	7.21	2.07	5.12	1.78	32.18	4.81	92.3	36.3	94.6	62.1	64.4	88.3										
		7	45g	1200	20	0.5	6	0.12	7.35	2.04	5.26	1.71	41.90	5.76	94.3	39.0	96.7	63.5	65.9	90.2										
		8	45g	1200	20	0.5	6	0.12	7.45	2.03	5.31	1.78	38.98	5.53	93.6	32.2	96.1	63.1	62.0	89.7										
		9	45g	1200	20	0.5	6	0.12	7.25	2.09	5.19	1.78	39.43	5.68	94.0	35.5	96.4	63.3	63.9	89.9										
		10	45g	1200	20	0.5	6	0.12	7.08	2.11	5.12	1.82	38.39	5.39	94.1	33.2	96.4	63.4	62.6	89.9										
		11	45g	1200	20	0.5	6	0.12	7.07	2.09	5.18	1.84	40.87	6.08	94.7	29.8	96.9	63.8	60.7	90.4										
		12	45g	1200	20	0.5	6	0.14	7.64	2.12	5.26	1.79	33.39	5.01	91.5	39.3	94.2	61.6	66.0	87.9										
		13	45g	1200	20	0.5	6	0.12	7.24	2.14	5.07	1.77	32.64	4.85	92.1	41.3	94.6	62.0	67.1	88.3										
		14	45g	1200	20	0.5	6	0.12	7.24	2.09	4.91	1.84	31.82	4.63	91.3	32.9	94.0	61.5	62.5	87.7										
		15	45g	1200	20	0.5	6	0.12	7.33	2.12	4.78	1.81	30.00	4.61	89.9	38.5	92.7	60.5	65.6	86.5										
		16	45g	1200	20	0.5	6	0.12	7.81	2.26	5.25	1.90	38.38	5.50	92.3	36.8	95.2	62.1	64.6	88.6										
		17	45g	700	20	0.125	6	0.28	7.33	2.11	4.93	1.77	31.25	4.17	90.9	40.8	93.6	61.2	66.9	87.3										
		18	45g	700	20	0.125	6	0.27	7.71	2.12	4.50	1.75	28.03	4.24	86.4	46.5	89.8	58.1	70.1	83.8										
		19	45g	700	20	0.125	6	0.27	7.23	2.16	4.96	1.81	33.49	4.98	92.0	39.2	94.6	62.0	66.0	88.3										
		20	45g	700	20	0.125	6	0.12	7.32	2.08	4.75	1.75	27.67	4.49	88.8	41.9	91.6	59.8	67.5	85.5										
		21	45g	700	20	0.125	6	0.12	7.61	2.18	5.15	1.85	33.20	4.75	91.2	37.0	94.0	61.4	64.7	87.7										

POC Results Summary

Table E-4
PITTSBURGH NO. 8 POC PARAMETRIC TESTING CONDITIONS AND RESULTS SUMMARY

RUN	TOP SIZE	SET PNT	HS VESS	HS RPM	HS SOL	HS %	HS DIES	FLOT SOL	TOTAL lb/r	FEED				FINAL PRODUCT				COMBINED TAILINGS				AGGLOMERATION PERFORMANCE				ROM PERFORMANCE							
										%		%		%		%		%		%		%		%		%		%		%		%	
										ASH	TS	ASH	TS	ASH	TS	ASH	TS	YIELD	PSR	%	YIELD	PSR	%	YIELD	PSR	%	YIELD	PSR	%	YIELD	PSR	%	YIELD
33	200M	1	45g	700	20	0.5	0.18	6	0.18	7.8	2.13	6.6	1.61	39.8	6.1	96.4	45.8	97.8	64.9	69.7	91.3												
		2	45g	700	20	0.5	0.18	6	0.18	8.0	2.17	5.7	1.69	50.8	8.1	94.9	43.1	97.6	63.9	68.1	91.1												
34	200M	1	45g	700	20	0.125	0.18	6	0.18	8.2	2.23	4.2	1.50	13.6	3.0	57.9	75.8	60.7	39.0	86.5	56.6												
		2	45g	700	20	0.125	0.18	6	0.18	8.4	2.36	4.7	1.65	16.2	3.4	67.7	67.3	70.9	45.6	81.7	66.1												
		3	45g	700	20	0.125	0.18	6	0.18	9.1	2.49	5.2	1.70	17.3	3.6	68.1	67.6	71.4	45.8	81.9	66.6												
35	200M	1	45g	1200	20	0.5	0.18	6	0.18	8.0	2.25	5.6	1.68	30.3	5.1	90.4	50.0	93.0	60.8	72.0	86.8												
		2	45g	1200	20	0.5	0.18	6	0.18	8.3	2.39	5.4	1.82	23.3	4.4	84.1	50.5	87.1	56.6	72.3	81.2												
		3	45g	1200	20	0.5	0.18	6	0.18	8.1	2.29	5.5	1.88	22.0	4.2	84.0	42.3	86.8	56.6	67.7	81.0												
36	200M	1	45g	700	20	0.5	0.18	6	0.18	8.4	2.37	6.0	1.96	35.5	5.7	91.9	35.5	94.7	61.9	63.9	88.3												
		2	45g	700	20	0.5	0.18	6	0.18	8.2	2.29	5.9	1.94	29.3	5.3	89.9	34.5	92.6	60.6	63.3	86.4												
38	200M	1	45g	1200	20	0.125	0.18	6	0.18	8.1	2.18	4.4	1.58	18.8	3.9	74.2	63.3	77.6	50.0	79.4	72.4												
		2	45g	1200	20	0.125	0.18	6	0.18	8.1	2.12	4.6	1.60	17.4	3.5	72.0	61.0	75.2	48.5	78.2	70.2												
		3	45g	1200	20	0.125	0.18	6	0.18	8.2	2.21	5.2	1.58	19.2	3.8	79.0	61.5	81.9	53.2	78.5	76.4												
39	200M	1	45g	1200	20	0.125	0.18	6	0.18	7.9	2.09	5.0	1.61	15.4	3.3	72.0	59.3	74.6	48.5	77.2	69.6												
40	200M	1	45g	1200	20	0.125	0.25	6	0.25	7.9	2.17	5.1	1.65	28.0	4.6	87.7	50.7	90.8	59.1	72.4	84.7												
		2	45g	1200	20	0.250	0.18	6	0.18	8.2	2.21	6.2	1.82	48.5	9.2	95.3	34.6	97.6	64.1	63.4	91.1												
		3	45g	1200	20	0.250	0.25	6	0.25	7.9	2.21	6.3	1.95	65.5	11.3	97.3	23.3	99.2	65.5	57.1	92.6												
41	200M																																
		1	45g	1200	20	0.250	0.18	6	0.18	7.7	2.12	5.1	1.73	27.3	4.7	88.2	42.3	91.0	59.4	67.7	84.9												
		2	45g	1200	20	0.250	0.07	6	0.07	8.0	2.16	5.4	1.73	30.7	5.2	89.8	42.8	92.7	60.4	68.0	86.5												
		3	45g	700	20	0.250	0.07	6	0.07	7.9	2.03	5.3	1.65	23.3	4.5	85.4	45.8	88.2	57.5	69.6	82.3												
		4	45g	700	20	0.250	0.18	6	0.18	8.1	2.17	6.0	1.79	36.8	6.9	93.2	36.3	95.7	62.8	64.4	89.2												

Table E-4 (Continued)
PITTSBURGH NO. 8 POC PARAMETRIC TESTING CONDITIONS AND RESULTS SUMMARY

RUN	TOP SIZE	SET PNT	HS VESS	HS RPM	HS SOL	HS %	HS DIES	FLOT SOL	TOTAL FROTH lb/ft	FEED			FINAL PRODUCT			COMBINED TAILINGS			AGGLOMERATION PERFORMANCE				ROM PERFORMANCE			
										%	ASH	TS	%	ASH	TS	%	ASH	TS	%	YIELD	PSR	%	%	%	%	%
42	200M	1	45g	1200	20	20	0.50	6	0.21	7.8	2.19	5.8	1.87	35.5	5.7	93.2	31.8	95.5	62.7	61.8	89.1					
		2	45g	1200	20	20	0.35	6	0.21	7.7	2.12	5.6	1.81	33.6	5.8	92.5	33.4	94.9	62.3	62.7	88.6					
		3	45g	1200	20	20	0.50	6	0.21	8.3	2.33	6.7	1.98	47.3	9.2	98.2	29.1	98.0	64.7	60.3	91.5					
43	100M	1	45g	1200	20	20	0.50	6	0.18	7.7	1.98	6.5	1.88	39.3	5.5	96.3	14.4	97.7	64.8	52.1	91.2					
		2	45g	1200	20	20	0.50	6	0.18	7.9	2.03	6.1	1.83	49.3	4.5	95.9	22.3	98.0	64.6	56.5	91.5					
		3	45g	1200	20	20	0.125	6	0.25	7.9	2.10	5.6	1.77	17.8	2.8	80.7	43.2	83.1	54.4	68.2	77.5					
		4	45g	1200	20	20	0.070	6	0.28	8.0	2.26	6.4	1.79	23.2	3.8	90.9	41.9	92.6	61.2	67.5	86.4					
44	100M	1	45g	1200	20	20	0.125	6	0.29	7.8	2.08	6.5	2.01	39.6	5.2	96.1	11.2	97.6	64.7	50.3	91.1					
		2	45g	1200	20	20	0.125	6	0.29	7.7	2.08	6.1	1.94	47.0	6.8	96.1	16.7	98.0	64.7	53.4	91.4					
		3	45g	1200	20	20	0.500	6	0.24	8.0	2.07	6.5	1.94	37.2	4.8	95.2	17.0	96.9	64.1	53.6	90.4					
45	200M	1	45g	700	20	20	0.125	6	0.29	8.1	2.22	5.6	1.72	41.6	5.4	93.0	44.4	96.0	62.6	68.9	89.5					
		2	45g	700	20	20	0.125	6	0.29	8.2	2.24	5.7	1.84	40.4	6.9	92.7	37.2	95.6	62.4	64.8	89.2					
		3	45g	700	20	20	0.125	6	0.29	8.1	2.09	5.3	1.82	26.4	4.6	86.5	35.6	89.6	58.2	63.9	83.8					
46	200M	1	45g	700	20	20	0.5	6	0.07	7.7	2.04	5.5	1.82	24.4	4.3	88.1	30.8	90.6	59.3	61.3	84.5					
		2	45g	700	20	20	0.5	6	0.07	8.0	2.06	5.5	1.76	26.7	4.2	88.3	37.0	91.1	59.5	64.7	85.0					
		3	45g	700	20	20	0.5	6	0.07	7.8	2.02	5.6	1.77	24.3	4.3	88.3	33.6	90.7	59.5	62.8	84.7					
		4	45g	700	20	20	0.5	6	0.04	8.2	2.20	5.8	1.84	24.8	4.3	87.2	38.6	89.9	58.7	65.6	83.9					

Table E-4 (Continued)
PITTSBURGH NO. 8 POC PARAMETRIC TESTING CONDITIONS AND RESULTS SUMMARY

RUN	TOP SIZE	SET PNT	HS VESS	HS RPM	HS SOL	HS %	HS DIES	FLOT SOL	TOTAL FROTH lb/ft	FEED				FINAL PRODUCT				COMBINED TAILINGS				AGGLOMERATION PERFORMANCE				ROM PERFORMANCE							
										% ASH		% TS		% ASH		% TS		% ASH		% TS		% YIELD		% PSR		% %BTU REC		% YIELD		% PSR		% %BTU REC	
										%	ASH	%	TS	%	ASH	%	TS	%	ASH	%	TS	%	YIELD	%	PSR	%	%BTU	%	YIELD	%	PSR	%	%BTU
47	200M	1	45g	0	20	0.125	6	0.27	8.1	2.13	5.8	1.75	29.6	5.1	90.0	39.9	92.7	60.6	66.3	86.5													
		2	45g	0	20	0.125	6	0.27	8.2	2.21	5.5	1.76	29.3	6.7	88.4	44.3	91.5	59.5	68.8	85.3													
		3	45g	0	20	0.125	6	0.27	7.7	1.90	5.2	1.67	43.9	6.7	93.7	30.7	96.5	63.1	61.2	90.0													
48	200M	1	45g	700	20	0.5	6	0.07	7.7	2.10	5.5	1.76	30.3	4.5	90.9	37.4	93.4	61.2	65.0	87.2													
		2	45g	700	20	0.5	6	0.14	8.1	2.26	5.7	1.84	44.5	6.9	94.0	37.2	96.7	63.3	64.9	90.3													
49	200M	1	45g	700	20	0.5	6	0.10	7.8	2.14	5.7	1.79	30.7	5.2	91.3	36.5	93.8	61.5	64.4	87.5													
		2	45g	700	20	0.5	6	0.16	7.9	2.24	6.0	1.85	55.9	9.3	96.2	33.6	98.5	64.8	62.8	91.9													
		3	45g	700	20	0.5	6	0.16	8.1	2.23	6.0	1.90	42.4	7.0	94.4	30.8	96.8	63.6	61.3	90.3													
50	200M	1	45g	1200	20	0.125	6	0.27	7.9	2.15	3.9	1.47	12.1	2.9	51.6	78.6	54.2	34.7	88.0	50.5													
51	200M	1	45g	1200	20	0.5	6	0.11	8.4	2.32	5.5	1.95	34.4	4.9	90.2	35.6	93.4	60.7	64.0	87.1													
		2	45g	1200	20	0.5	6	0.12	8.2	2.47	5.7	1.78	36.3	5.1	91.8	50.4	94.6	61.8	72.2	88.3													
		3	150g	1186	20	0.5	6	0.11	8.4	2.29	5.8	1.91	37.0	5.4	91.7	35.7	94.7	61.7	64.0	88.3													
53	200M	1	45g	1200	20	0.125	6	0.32	8.4	2.39	5.3	1.87	26.7	4.8	85.5	46.3	88.9	57.6	70.0	82.9													
54	100M	1	45g	700	20	1	6	0.27	8.4	2.34	6.2	2.12	37.5	5.2	92.7	22.8	95.4	62.4	56.8	89.0													
55	200M	1	150g	1186	20	0.5	6	0.11	8.0	2.14	6.0	1.87	45.6	6.6	94.8	27.6	97.2	63.8	59.5	90.7													
		2	150g	1186	20	0.5	6	0.11	7.5	2.01	5.3	1.77	34.9	5.7	92.8	29.8	95.2	62.5	60.7	88.8													
		3	150g	1186	20	0.5	6	0.11	7.7	2.08	5.5	1.81	31.9	5.1	91.5	31.8	94.1	61.6	61.8	87.8													

Table E-5
ILLINOIS NO. 5 POC DEMONSTRATION RUN CONDITIONS AND RESULTS SUMMARY

RUN	TOP SIZE	SET PNT	HS VESS	HS RPM	HS % SOL	HS % DIESEL	FLT % SOL	TOTAL FROTH lb/t	FEED			FINAL PRODUCT			COMBINED TAILINGS			AGGLOMERATION PERFORMANCE		
									% ASH	% TS	% TS	% ASH	% TS	% TS	% ASH	% TS	% TS	% YIELD	% PSR	% REC
DEM	100M	1	45g	900	20	0.8	6	0.144	18.3	1.96	1.96	5.1	1.59	1.59	65.6	2.80	2.80	78.2	49.4	92.2
		2	45g	900	20	0.8	6	0.120	18.4	1.95	1.95	4.6	1.42	1.42	52.5	3.31	3.31	71.2	62.9	84.5
		3	45g	900	20	0.8	6	0.120	18.7	2.08	2.08	4.5	1.38	1.38	53.1	3.30	3.30	70.8	68.4	84.6
		4	45g	900	20	0.8	6	0.120	17.6	1.78	1.78	4.6	1.38	1.38	57.9	3.29	3.29	75.5	57.3	88.8
		5	45g	900	20	0.8	6	0.118	17.8	1.88	1.88	4.6	1.36	1.36	57.1	3.39	3.39	74.9	62.2	89.2
		6	45g	900	20	0.8	6	0.118	16.3	1.80	1.80	4.6	1.42	1.42	54.4	3.22	3.22	76.4	54.5	88.3
		7	45g	900	20	0.8	6	0.120	19.4	2.28	2.28	4.3	1.32	1.32	45.3	2.68	2.68	63.1	77.6	76.3
		8	45g	900	20	0.8	6	0.120	18.2	2.05	2.05	4.2	1.35	1.35	43.5	2.84	2.84	64.2	72.1	76.5
		9	45g	900	20	0.8	6	0.126	18.1	1.88	1.88	4.2	1.34	1.34	45.9	2.86	2.86	66.8	67.6	79.3
		10	45g	900	20	0.8	6	0.126	17.4	1.82	1.82	4.3	1.34	1.34	42.5	2.73	2.73	65.6	66.4	77.2
		11	45g	900	20	0.8	6	0.132	17.6	1.94	1.94	4.2	1.42	1.42	43.9	2.54	2.54	66.2	65.2	78.2
		12	45g	900	20	0.8	6	0.136	18.2	1.98	1.98	4.8	1.42	1.42	57.1	3.15	3.15	74.4	62.0	87.9
		13	45g	900	20	0.8	6	0.123	18.5	1.97	1.97	5.1	1.54	1.54	67.0	3.93	3.93	78.4	52.6	92.7
		14	45g	900	20	0.8	6	0.120	18.8	2.01	2.01	4.5	1.39	1.39	54.1	3.05	3.05	71.2	66.4	85.1
		15	45g	900	20	0.8	6	0.120	17.5	1.96	1.96	4.1	1.34	1.34	43.4	2.88	2.88	66.1	69.8	77.9
		16	45g	900	20	0.8	6	0.120	17.4	1.95	1.95	4.5	1.39	1.39	54.0	3.22	3.22	74.0	62.8	86.8
		17	45g	900	20	0.8	6	0.120	18.1	1.94	1.94	4.6	1.42	1.42	55.3	2.80	2.80	73.3	61.3	86.8
		18	45g	900	20	0.8	6	0.120	18.2	1.99	1.99	4.7	1.38	1.38	60.2	3.35	3.35	75.7	63.7	89.5
	200M	19	150g	848	20	1.2	6	0.091	18.7	1.96	1.96	4.5	1.37	1.37	65.3	3.62	3.62	76.7	63.3	91.5
		20	150g	848	20	1.2	6	0.091	18.5	1.95	1.95	4.4	1.32	1.32	63.4	4.84	4.84	76.0	66.4	90.7
		21	150g	848	20	1.2	6	0.091	18.8	2.12	2.12	4.4	1.32	1.32	62.6	4.26	4.26	75.3	70.4	90.1
		22	150g	848	20	1.2	6	0.091	16.6	1.87	1.87	4.2	1.35	1.35	57.1	4.03	4.03	76.7	61.7	89.2
		23	150g	848	20	1.2	6	0.091	18.3	1.97	1.97	4.2	1.34	1.34	55.9	3.65	3.65	72.8	66.8	86.7

Table E-6
ILLINOIS NO.5 POC PARAMETRIC TESTING OPERATIONS AND RESULTS SUMMARY

RUN	TOP SIZE	SET PNT	HS VESS	HS RPM	HS		HS SOL	HS DIESEL	FLT SOL	TOTAL lb/t	FEED		FINAL PRODUCT		COMBINED TAILINGS		AGGLOMERATION PERFORMANCE		
					%	%					%	%	%	%	%	%	%	%	%
											ASH	TS	ASH	TS	ASH	TS	YIELD	PSR	%BTU REC
											TS	TS	TS	TS	TS	TS			
57	100M	1	45g	700	20	0.5	20	0.5	6	0.203	16.9	1.78	4.6	1.42	44.7	2.24	69.3	57.8	80.7
		2	45g	700	20	0.5	20	0.5	6	0.203	19.0	1.99	3.7	1.31	31.6	2.27	45.3	81.1	54.8
58	100M	1	45g	700	20	0.5	20	0.5	6	0.238	18.2	1.85	4.9	1.47	59.0	2.41	75.3	54.1	89.0
		2	45g	700	20	0.5	20	0.5	6	0.238	19.4	2.15	5.1	1.51	64.8	2.50	76.0	60.9	91.1
		3	45g	700	20	0.5	20	0.5	6	0.202	17.1	1.71	4.7	1.45	56.1	2.74	75.8	49.5	88.4
		4	45g	700	20	0.5	20	0.5	6	0.202	16.4	1.74	4.5	1.46	55.3	2.62	76.6	49.3	88.6
60	100M	1	45g	700	20	0.5	20	0.5	6	0.416	17.4	1.76	6.7	1.57	77.8	3.09	85.0	35.9	97.2
		2	45g	700	20	0.5	20	0.5	6	0.339	19.3	2.22	6.2	1.74	75.2	3.35	81.0	48.3	95.6
		3	45g	700	20	0.5	20	0.5	6	0.339	18.3	2.06	5.9	1.66	81.6	3.39	83.6	45.5	97.7
62	100M	1	45g	700	20	0.5	20	0.5	6	0.095	18.3	1.89	3.4	1.20	26.7	1.94	36.1	86.7	43.4
		2	45g	400	20	0.5	20	0.5	6	0.095	15.7	1.84	3.8	1.31	28.7	2.19	52.2	75.1	60.3
63	100M	1	45g	400	20	0.5	20	0.5	6	0.149	18.2	1.87	3.9	1.24	25.8	2.24	34.8	85.7	41.6
		2	45g	700	20	0.5	20	0.5	6	0.149	18.2	1.76	3.7	1.20	28.0	2.08	40.4	83.5	48.3
64	100M	1	45g	700	20	1.0	20	1.0	6	0.152	18.8	1.86	4.8	1.44	58.9	3.29	74.1	57.5	88.3
		2	45g	700	20	1.0	20	1.0	6	0.107	20.6	2.07	4.3	1.36	48.0	2.94	62.7	72.9	77.0
		3	45g	700	20	1.5	20	1.5	6	0.054	16.8	1.87	4.1	1.38	37.5	2.68	61.9	67.6	72.4

Table E-6 (Continued)
ILLINOIS NO.5 POC PARAMETRIC TESTING OPERATIONS AND RESULTS SUMMARY

RUN	TOP SIZE	SET PNT	HS VESS	HS RPM	HS % SOL	HS % DIESEL	FLT % SOL	TOTAL lb/ft	FEED			FINAL PRODUCT			COMBINED TAILINGS			AGGLOMERATION PERFORMANCE		
									% ASH	% TS	% %	% ASH	% TS	% %	% ASH	% TS	% %	% YIELD	% PSR	% %BTU REC
65	100M	1	45g	700	20	0.25	6	0.256	19.4	2.01		5.5	1.58		58.3	2.34		73.8	54.9	87.9
		2	45g	700	20	0.25	6	0.256	20.0	2.25		5.2	1.50		59.4	3.23		72.7	65.5	87.7
		3	45g	700	20	0.25	6	0.256	17.1	1.74		4.4	1.35		43.7	3.05		67.7	62.3	79.2
66	100M	1	45g	700	20	1	6	0.156	16.9	1.77		4.5	1.42		56.0	3.43		76.0	53.6	88.5
		2	45g	700	20	1	6	0.156	17.6	1.71		4.6	1.38		59.4	2.34		76.3	53.7	89.7
		3	45g	700	20	1	6	0.125	16.3	1.86		4.2	1.36		38.8	3.27		64.9	66.5	75.4
67	200M	1	45g	700	20	0.5	6	0.210	19.4	2.03		4.2	1.38		56.8	3.31		71.1	67.6	86.0
		2	45g	700	20	0.75	6	0.102	16.4	1.88		4.2	1.38		50.1	2.92		69.1	64.3	82.4
		3	45g	700	20	1	6	0.091	20.4	2.11		4.4	1.44		58.8	3.05		70.7	66.8	86.5
68	100M	1	45g	700	20	0.7	6	0.136	17.4	1.83		4.5	1.33		51.4	2.69		72.5	63.7	85.1
		2	45g	700	20	0.9	6	0.120	16.7	1.84		4.2	1.41		42.3	2.55		67.2	62.2	78.4
		3	45g	700	20	0.8	6	0.112	18.8	1.86		4.1	1.34		39.6	2.19		58.4	71.6	70.2
69	100M	1	45g	0	20	0.2	6	0.360	15.5	1.67		4.3	1.33		33.5	2.03		61.5	64.0	70.5
		2	45g	0	20	0.15	6	0.430	15.2	1.76		4.9	1.51		46.7	2.78		75.4	46.7	85.5
		3	45g	600	20	0.3	6	0.239	18.0	2.02		4.2	1.40		35.1	2.17		55.3	73.3	65.7
70	200M	1	45g	500	20	0.15	6	0.407	18.8	1.81		5.0	1.40		60.9	3.08		75.3	57.1	89.5
		2	45g	700	20	0.3	6	0.253	19.0	1.86		4.6	1.38		57.6	3.06		72.8	61.6	87.2
		3	45g	600	20	0.6	6	0.200	16.4	1.83		4.4	1.37		53.1	2.89		75.2	59.4	87.3
71	100M	1	45g	600	20	0.8	6	0.070	17.8	1.75		3.9	1.31		35.5	2.26		56.2	70.9	66.7
		2	45g	700	20	0.5	6	0.157	17.3	1.72		4.7	1.45		48.4	2.56		71.1	53.1	83.2
		3	45g	800	20	0.25	6	0.212	16.2	1.70		4.8	1.42		43.4	2.44		70.4	53.7	81.0

Table E-6 (Continued)
ILLINOIS NO.5 POC PARAMETRIC TESTING OPERATIONS AND RESULTS SUMMARY

RUN	TOP SIZE	SET PNT	HS VESS	HS RPM	HS % SOL	HS % DIESEL	FLT % SOL	TOTAL lb/l	FEED		FINAL PRODUCT		COMBINED TAILINGS		AGGLOMERATION PERFORMANCE	
									ASH	TS	ASH	TS	ASH	TS	YIELD	% PSR REC
72	100M	1	45g	400	20	0.15	6	0.349	17.7	1.80	4.9	1.48	42.3	2.70	65.8	57.6 77.1
		2	45g	600	20	0.25	6	0.271	18.9	2.06	4.8	1.46	43.8	2.27	63.8	67.5 76.2
		3	45g	500	20	0.3	6	0.219	17.5	1.91	5.2	1.52	59.9	2.70	77.6	51.3 90.4
		4	45g	1200	20	0.3	6	0.270	16.9	1.76	4.9	1.45	52.6	2.42	74.9	51.5 86.9
		5	150g	848	20	0.3	6	0.259	16.0	1.74	4.6	1.47	45.9	2.37	72.4	51.5 83.3
73	100M	1	45g	700	20	0.7	6	0.137	18.2	1.90	4.7	1.44	58.5	4.17	74.9	58.2 88.6
		2	45g	600	20	0.6	6	0.178	18.4	1.86	5.3	1.52	70.0	3.23	79.8	48.8 94.0
		3	45g	700	20	0.9	6	0.133	17.8	1.84	4.9	1.49	66.1	3.51	79.0	50.2 92.7
		4	45g	500	20	0.35	6	0.211	17.3	1.88	5.3	1.53	61.0	2.90	78.4	48.9 91.1

Nomenclature:

TOP SIZE: Particle topsize (mesh)
 SET PNT: Set point
 HS VESS: High-shear vessel
 HS RPM: High-shear intensity (RPM)
 HS Sol %: High-shear percent solids
 HS % DIES: Weight percent diesel in high-shear
 FLOT SOL %: Percent solids in flotation
 ASH %: Weight percent ash
 TS %: Weight percent total sulfur
 % YIELD: Weight percent clean coal yield
 % BTU REC: Percent Btu Recovery
 % PSR: Percent pyritic sulfur rejection
 ROM: Run-of-mine

Appendix F

BATCH TEST RESULTS SUMMARY

Batch Test Results Summary

BATCH TEST CONDITIONS AND RESULTS SUMMARY

Test No	Coal	Source	Batch	Sample ID	Size mesh	Diesel %	HS RPM	HS Sol %	Frother	Flot. Sol%	Feed			Final Product			Tailings			Agglomeration Performance						ROM Performance					
											Ash %	TS %		Ash %	TS %		Ash %	TS %		Yield %	Butu Rec %	TSR %	PSR %		Butu Rec %	TSR %	PSR %		Butu Rec %	TSR %	PSR %
2	UF1	LAB	1	1-6	200	0.99	2000	15.3	NONE	6.2	8.7	1.64	7.0	1.41	66.0	9.18	97.0	99.2	16.5	25.7	97.6	53.8	68.4								
3	UF1	LAB	1	2-7	200	1.00	2000	14.9	NONE	6.0	8.7	1.79	6.2	1.48	66.8	8.83	95.8	98.9	20.7	30.6	97.3	56.1	70.5								
5	UF1	LAB	1	3-4	200	0.99	2000	15.6	MIBC	6.3	8.7	1.87	6.7	1.63	67.9	8.72	96.6	99.1	15.9	23.0	97.5	53.4	67.3								
6	UF1	LAB	1	1-4	200	1.05	2000	15.0	NONE	6.1	8.7	1.67	6.9	1.41	67.1	9.71	96.9	99.2	18.2	28.0	97.6	54.7	69.4								
7	UF1	LAB	1	4-4	100	1.02	2000	15.1	NONE	6.1	8.7	1.52	7.2	1.39	57.6	5.89	97.0	98.9	11.5	18.3	97.3	51.0	65.2								
8	UF1	LAB	1	3-10	100	1.00	2000	15.4	NALCO	6.3	8.7	1.87	6.4	1.55	58.7	8.72	95.5	98.3	20.9	30.0	96.8	56.2	70.2								
9	UF1	LAB	1	3-13	100	1.00	2000	15.5	BM150	6.3	8.7	1.91	6.6	1.66	58.6	7.64	95.9	98.5	16.3	23.1	96.9	53.7	67.3								
10	UF1	LAB	1	2-10	100	0.99	2000	15.8	NONE	6.4	8.7	1.87	6.9	1.71	60.1	6.38	96.5	98.8	11.8	16.8	97.2	51.2	64.6								
11	UF1	LAB	1	2-11	100	0.98	2000	15.9	NONE	6.5	8.7	1.81	7.2	1.64	61.8	7.63	97.2	99.1	11.9	17.4	97.5	51.2	64.9								
12	UF1	LAB	1	3-3	100	1.00	2000	15.6	MIBC	6.3	8.7	1.89	7.3	1.73	59.9	7.63	97.3	99.1	10.8	15.3	97.5	50.6	64.0								
13	UF1	LAB	1	3-1	28	1.01	2000	15.4	NONE	6.3	8.7	1.87	7.3	1.74	51.6	5.79	96.7	98.5	10.1	14.1	96.9	50.2	63.5								
14	UF1	LAB	1	3-2	28	1.00	2000	15.4	NONE	6.2	8.7	1.63	7.4	1.62	51.3	2.03	96.9	98.6	3.8	5.1	97.0	46.7	59.6								
15	UF1	LAB	1	3-5	28	0.97	2000	16.0	MIBC	6.5	8.7	1.87	6.9	1.71	50.6	5.40	95.7	98.0	12.4	17.2	96.4	51.5	64.8								
20	UF1	POC	38	38-2	100	1.01	2000	18.1	NONE	7.4	9.6	1.77	5.5	1.32	43.4	5.43	89.1	93.8	33.4	47.2	92.3	63.1	77.5								
21	UF1	POC	38	38-9	100	1.20	2000	13.6	NONE	5.5	9.6	1.75	5.9	1.36	51.7	6.23	92.0	96.3	28.5	41.2	94.7	60.4	75.0								
22	UF1	POC	38	38-8	100	1.13	2000	16.1	NONE	6.5	9.6	1.72	4.6	1.12	30.9	4.27	81.0	86.2	47.2	64.9	84.8	70.8	85.1								
23	UF1	POC	38	38-1	100	1.00	2000	18.2	NONE	7.4	9.6	1.79	3.7	0.93	16.6	2.83	54.5	58.6	71.8	86.8	57.6	84.4	94.4								
24	UF1	LAB	1	4-1	325	1.02	2000	15.0	NONE	6.1	8.7	1.55	6.1	1.25	69.0	8.37	95.7	99.0	23.1	37.1	97.4	57.4	73.2								
25	UF1	LAB	1	4-2a	325	0.00	1200		BM150	6.2	8.7	1.50	4.0	0.92	17.9	2.63	65.9	69.8	59.7	79.4	68.7	77.7	91.2								
26	UF1	LAB	1	4-2b	325	0.36	2000	14.9	NONE	6.1	8.7	1.50	6.2	1.21	76.6	9.47	96.4	99.5	22.4	37.0	97.9	57.0	73.2								
27	UF1	LAB	1	3-7a	200	0.00	0		BM150	6.5	8.7	1.80	4.0	1.11	22.8	3.87	74.9	79.4	53.8	70.5	78.1	74.4	87.4								
28	UF1	LAB	1	3-7b	200	0.17	2000	15.2	BM150	6.2	8.7	1.82	5.4	1.24	76.2	13.58	95.3	99.3	35.2	52.2	97.7	64.1	79.7								
29	UF1	LAB	1	3-6a	200	0.00	0		BM150	6.7	8.7	1.85	3.6	0.99	15.2	2.94	55.7	59.4	70.2	84.5	58.4	83.5	93.4								
30	UF1	LAB	1	3-6b	200	0.30	2000	15.2	NONE	6.2	8.7	1.82	6.8	1.59	66.4	8.65	96.7	99.1	15.8	23.0	97.5	53.4	67.3								
31	UF1	POC	38	38-10	100	1.40	2000	13.0	NONE	5.2	9.6	1.72	6.6	1.44	74.0	7.71	95.6	99.2	19.7	29.8	97.6	55.6	70.1								
32	UF1	POC	38	38-11	100	1.17	2000	15.9	BETZ	6.5	9.6	1.71	4.5	1.08	28.7	4.10	78.9	84.0	50.6	68.9	82.7	72.6	86.8								
33	UF1	LAB	1	4-9	200	1.01	2000	15.3	NALCO	6.2	8.7	1.59	6.8	1.36	66.2	8.49	96.7	99.1	17.4	27.5	97.5	54.3	69.1								
34	UF1	LAB	1	4-13	200	1.02	2000	15.1	NALCO	6.1	8.7	1.49	6.9	1.30	64.4	7.36	96.8	99.1	15.6	25.6	97.5	53.3	68.3								
35	UF1	LAB	1	5-1	200	1.01	2000	15.1	BETZ	6.1	8.7	1.57	6.6	1.34	57.5	6.74	95.8	98.4	18.2	28.5	96.8	54.7	69.6								
36	UF1	LAB	1	5-2	200	1.01	2000	15.1	MIBC	6.1	8.7	1.54	6.6	1.30	59.8	7.34	96.0	98.6	18.9	30.1	97.0	55.1	70.3								
37	UF1	LAB	1	5-3	200	1.02	2000	15.1	MIBC	6.1	8.7	1.65	6.7	1.42	56.1	7.02	95.8	98.3	17.7	26.9	96.7	54.4	68.9								
38	UF1	LAB	1	5-4	200	1.02	2000	15.0	NALCO	6.1	8.7	1.59	6.6	1.37	52.0	6.06	95.2	97.8	18.4	28.1	96.2	54.8	69.4								

Batch Test Results Summary

BATCH TEST CONDITIONS AND RESULTS SUMMARY (Continued)

Test No	Coal	Source	Batch	Sample ID	Size mesh	Diesel %	HS RPM	HS Sol%	Frother	Flot. Sol%	Feed			Final Product			Tailings			Agglomeration Performance						ROM Performance		
											Ash %	TS %	%	Ash %	TS %	%	Ash %	TS %	%	Yield %	Btu Rec %	TSR %	PSR %	%	Btu Rec %	TSR %	PSR %	%
39	UF1	LAB	1	5-5	200	1.02	2000	15.1	BETZ	6.1	8.7	1.52	7.2	1.38	62.4	6.29	97.1	99.1	11.9	19.1	97.5	51.2	65.6					
40	UF1	LAB	1	5-6	200	1.02	2000	15.0	NONE	6.1	8.7	1.54	6.7	1.32	54.5	6.53	95.8	98.2	17.8	28.1	96.7	54.5	69.4					
41	UF1	LAB	1	5-7	100	1.02	2000	15.0	NONE	6.1	8.7	1.58	6.5	1.33	34.1	4.46	92.0	94.5	22.7	33.2	93.0	57.2	71.6					
42	UF1	LAB	1	5-8	28	1.02	2000	15.0	NONE	6.1	8.7	1.69	7.4	1.51	43.8	6.21	96.2	97.9	14.1	20.7	96.3	52.4	66.3					
43	UF1	LAB	1	5-9	200	1.02	2000	15.0	BETZ	6.1	8.7	1.57	6.9	1.36	61.4	7.41	96.5	98.8	16.3	25.7	97.3	53.7	68.4					
44	UF1	LAB	1	5-10	200	1.02	2000	15.0	MIBC	6.1	8.7	1.57	7.1	1.40	65.9	7.74	97.3	99.2	13.4	21.3	97.6	52.1	66.5					
45	UF1	LAB	1	5-11	200	1.00	2000	15.2	NALCO	6.2	8.7	1.58	5.8	1.23	51.7	6.74	93.6	97.1	27.3	42.2	95.5	59.8	75.4					
46	UF1	LAB	1	5-12	200	1.01	2000	15.1	BETZ	6.1	8.7	1.54	6.6	1.30	61.4	7.57	96.1	98.7	19.1	30.4	97.1	55.2	70.4					
47	UF1	LAB	1	6-1	200	1.00	2000	15.3	MIBC	6.2	8.7	1.63	5.9	1.23	48.9	7.27	93.4	96.7	29.5	44.9	95.2	61.0	76.6					
48	UF1	LAB	1	6-2	200	1.02	2000	19.1	NONE	6.1	8.7	1.62	6.4	1.33	57.8	7.64	95.5	98.3	21.4	33.1	96.7	56.5	71.5					
49	UF1	LAB	1	6-3	100	1.00	2000	19.3	NONE	6.1	8.7	1.67	7.0	1.50	46.7	5.39	95.6	97.7	14.3	21.0	96.1	52.5	66.4					
50	UF1	POC	118	118-1	200	0.00	0		NONE	8.5	10.6	1.78	3.0	0.83	21.1	3.10	58.0	63.7	73.0	91.1	62.7	85.0	96.2					
51	UF1	POC	118	118-2	200	0.00	0		BETZ	8.3	10.6	1.71	3.2	0.85	22.7	3.12	62.0	68.0	69.2	88.7	66.9	83.0	95.2					
52	UF1	POC	118	118-3	200	1.01			NONE	8.6	10.6	1.73	5.9	1.27	51.0	5.71	89.6	95.0	34.5	49.6	93.5	63.7	78.6					
53	UF1	POC	118	118-4	200	1.02			BETZ	8.4	10.6	1.71	6.0	1.24	53.9	6.23	90.4	95.8	34.8	50.7	94.2	63.9	79.0					
54	UF1	POC	118	118-5	200	1.01	2000	20.8	NONE	8.5	10.6	1.68	5.0	1.04	38.6	4.87	83.3	89.3	48.6	69.0	87.9	71.5	86.8					
55	UF1	POC	118	118-6	200	0.20	2000	20.5	NONE	8.4	10.6	1.68	4.7	1.04	33.3	4.17	79.4	85.4	51.1	70.6	84.1	72.9	87.5					
56	UF1	POC	118	118-7	200	0.61	2000	21.2	NONE	8.7	10.6	1.68	6.3	1.33	58.0	5.61	91.7	96.8	27.5	40.6	95.2	59.9	74.7					
57	UF1	POC	118	118-8	200	0.97	2000	21.0	NONE	8.6	10.6	1.80	7.0	1.45	68.2	7.24	94.1	98.5	23.9	34.7	96.9	57.9	72.2					
58	UF1	POC	144	144-HS	200	0.00	0			5.3	10.6	1.58	7.8	1.31	56.6	6.02	94.3	97.7	21.6	33.1	96.1	56.6	71.6					
59	UF1	POC	121	121-1	200	0.19	2000	19.3	BETZ	7.9	10.6	1.56	5.8	1.03	40.6	4.83	86.2	91.5	43.0	63.9	90.1	68.4	84.7					
60	UF1	POC	121	121-2	200	0.49	2000	21.0	BETZ	8.6	10.6	1.54	6.8	1.20	44.9	4.60	89.9	94.4	30.0	45.0	92.9	61.3	76.6					
61	UF1	POC	121	121-3	200	0.52	2000	19.9	BETZ	8.1	10.6	1.55	6.1	1.14	46.0	4.77	88.8	93.9	34.6	52.1	92.4	63.8	79.6					
62	UF1	POC	121	121-4	200	0.86	2000	18.1	BETZ	7.4	10.6	1.55	6.3	1.16	60.9	6.14	92.2	97.3	31.0	48.2	95.7	61.8	78.0					
63	UF1	POC	121	121-5	200	0.17	2000	16.7	BETZ	6.8	10.6	1.61	4.3	0.87	26.9	3.54	72.1	78.0	61.1	83.6	76.7	78.5	93.0					
64	UF1	POC	121	121-6	200	0.54	2000	18.6	BETZ	7.6	10.6	1.55	6.3	1.16	61.7	6.21	92.3	97.4	31.0	48.4	95.8	61.8	78.1					
65	UF1	POC	121	121-7	200	0.51	2000	19.0	BETZ	7.8	10.6	1.67	6.1	1.26	44.4	4.68	88.1	93.3	33.3	47.7	91.8	63.1	77.8					
66	UF1	POC	121	121-8	200	0.54	2000	18.8	BETZ	7.7	10.6	1.68	6.3	1.28	49.5	5.29	90.1	95.1	31.3	45.5	93.5	62.0	76.8					
67	UF1	LAB	1	4-12	100	1.04	2000	14.9	NONE	6.0	8.7	1.64	3.3	0.85	15.6	2.65	55.9	59.7	71.1	88.9	58.8	84.0	95.3					
68	UF1	POC	118	118-9	200	0.47	2000	20.2	BM150	8.3	10.6	1.67	6.1	1.24	46.7	5.13	88.8	94.0	34.2	49.5	92.5	63.6	78.5					
69	UF1	POC	118	118-10	200	0.49			BM150	7.9	10.6	1.65	5.3	1.11	40.0	4.65	84.7	90.5	43.0	61.5	89.1	68.4	83.6					
70	UF1	POC	403	403-2A	200	0.40	2000	21.8	BM150	9.0	10.0	1.51	5.7	1.04	26.4	3.30	79.2	83.6	45.5	64.2	82.2	69.8	84.8					

BATCH TEST CONDITIONS AND RESULTS SUMMARY (Continued)

Test No	Coal	Source	Batch	Sample ID	Size mesh	Diesel %	HS RPM	HS Sol %	Frother	Flot. Sol%	Feed			Final Product			Tailings			Agglomeration Performance						ROM Performance		
											Ash %	TS %	%	Ash %	TS %	%	Ash %	TS %	%	Yield %	Blu Rec %	TSR %	PSR %	%	Blu Rec %	TSR %	PSR %	%
71	UF1	POC	403	403-2B	200	0.40	2000	21.8	BM150	9.0	10.0	1.50	5.8	1.04	41.6	5.01	88.3	93.1	38.9	59.7	91.6	66.2	82.9	91.6	66.2	82.9		
72	UF1	POC	422	422-1A	200	0.00	0		BM150	7.2	9.6	1.54	5.2	1.04	18.8	2.58	67.7	71.5	54.2	70.5	70.3	74.7	87.4	70.3	74.7	87.4		
73	UF1	POC	422	422-1B	200	0.00	0		BM150	7.3	9.6	1.54	5.7	1.10	39.9	4.92	88.5	93.0	36.7	55.5	91.5	65.0	81.1	91.5	65.0	81.1		
74	UF1	POC	422	422-1C	200	0.00	0		BM150	7.3	9.6	1.53	5.7	1.10	42.3	5.20	89.4	93.8	35.9	54.8	92.3	64.5	80.8	92.3	64.5	80.8		
75	UF1	POC	456	456-1A	200	0.00	0		BM150	10.0	10.0	1.47	5.1	0.98	18.3	2.30	62.7	66.7	58.2	75.0	65.6	76.9	89.4	65.6	76.9	89.4		
76	UF1	POC	456	456-1B	200	0.00	0		BM150	9.9	10.0	1.47	5.5	1.02	30.7	3.54	82.1	86.9	43.0	63.2	85.5	68.5	84.4	85.5	68.5	84.4		
77	UF1	POC	447	447-1A	200	0.00	0		BM150	9.6	9.8	1.44	4.9	0.92	14.8	1.98	50.6	53.7	67.8	82.8	52.9	82.2	92.7	52.9	82.2	92.7		
78	UF1	POC	447	447-1B	200	0.00	0		BM150	9.8	9.8	1.44	5.4	0.97	28.5	3.45	80.9	85.5	45.6	67.4	84.1	69.9	86.1	84.1	69.9	86.1		
79	UF1	POC	501	501A	200	0.00	0		BM150	7.0	10.2	1.55	5.3	1.04	25.8	3.19	76.3	81.1	48.8	67.2	79.7	71.7	86.1	79.7	71.7	86.1		
80	UF1	POC	501	501B	200	0.00	0		BM150	7.1	10.2	1.55	5.6	1.05	36.1	4.31	84.8	89.8	42.5	62.5	88.4	68.1	84.0	88.4	68.1	84.0		
81	UF1	POC	501	501C	200	0.00	0		BM150	7.1	10.2	1.55	5.6	1.06	39.8	4.72	86.6	91.7	40.7	60.7	90.2	67.2	83.3	90.2	67.2	83.3		
82	UF1	POC	404	404-2A	200	0.00	0		BM150	7.7	10.0	1.55	5.2	0.99	23.6	3.14	73.8	78.3	53.0	72.4	77.1	74.0	88.2	77.1	74.0	88.2		
83	UF1	POC	404	404-2B	200	0.00	0		BM150	7.7	10.0	1.56	5.4	1.01	35.3	4.56	84.6	89.6	45.1	66.5	68.2	69.6	85.7	68.2	69.6	85.7		
84	UF1	POC	404	404-2C	200	0.00	0		BM150	7.6	10.0	1.56	5.5	1.02	39.0	5.03	86.5	91.5	43.4	64.9	90.1	68.7	85.1	90.1	68.7	85.1		
85	UF1	POC	528	528A	200	0.00	0		BM150	8.4	10.3	1.48	3.7	0.81	11.7	1.62	17.1	18.6	90.6	96.8	18.3	94.8	98.6	18.3	94.8	98.6		
86	UF1	POC	528	528B	200	0.00	0		BM150	7.9	10.3	1.49	3.8	0.81	13.5	1.82	33.2	35.9	81.9	93.7	35.3	90.0	97.3	35.3	90.0	97.3		
87	UF1	POC	528	528C	200	0.00	0		BM150	7.6	10.3	1.48	4.1	0.81	19.5	2.47	59.7	64.5	67.3	88.5	63.4	81.9	95.1	63.4	81.9	95.1		
88	UF1	POC	528	528D	200	0.00	0		BM150	7.6	10.3	1.49	4.3	0.83	25.9	3.20	72.3	77.8	59.8	84.4	76.6	77.7	93.4	76.6	77.7	93.4		
89	UF1	POC	537	537A	200	0.00	0		BM150	2.4	10.3	1.02	3.8	0.79	27.4	1.62	72.6	78.6	43.6	74.3	77.4	68.8	89.1	77.4	68.8	89.1		
90	UF1	POC	537	537B	200	0.00	0		BM150	3.8	10.3	1.02	4.0	0.79	34.0	1.86	78.9	85.4	38.6	71.7	84.0	66.0	88.0	84.0	66.0	88.0		
91	UF1	POC	537	537C	200	0.00	0		BM150	5.1	10.3	1.02	4.2	0.81	47.7	2.33	86.0	92.8	31.8	65.2	91.3	62.2	85.2	91.3	62.2	85.2		
92	UF1	POC	537	537D	200	0.00	0		BM150	5.8	10.3	1.03	4.6	0.83	64.1	2.88	90.4	97.0	27.0	59.2	95.5	59.6	82.7	95.5	59.6	82.7		
93	UF1	POC	464	464A	200	0.00	0		BM150	10.2	9.8	1.49	4.4	0.94	12.6	1.79	34.8	37.2	78.1	88.2	36.6	87.9	95.0	36.6	87.9	95.0		
94	UF1	POC	464	464B	200	0.00	0		BM150	9.8	9.8	1.50	4.7	0.96	18.4	2.42	63.1	67.1	59.6	77.2	66.0	77.6	90.3	66.0	77.6	90.3		
95	UF1	POC	464	464C	200	0.00	0		BM150	9.7	9.8	1.50	5.1	0.98	29.9	3.77	81.2	86.1	47.1	68.8	84.7	70.7	86.7	84.7	70.7	86.7		
96	UF1	POC	464	464D	200	0.00	0		BM150	9.7	9.8	1.50	5.4	0.99	36.7	4.61	86.0	90.8	43.2	65.6	89.3	68.5	85.4	89.3	68.5	85.4		
97	PIT	POC	1037	1037-2A	200	0.00	0		BM150	7.0	8.2	2.21	3.4	1.41	11.1	2.71	38.2	40.5	75.6	86.8	37.8	83.4	92.6	37.8	83.4	92.6		
98	PIT	POC	1037	1037-2B	200	0.00	0		BM150	6.4	8.2	2.22	3.7	1.45	15.2	3.40	60.8	64.2	60.2	77.2	59.9	72.9	87.2	59.9	72.9	87.2		
99	PIT	POC	1037	1037-2C	200	0.00	0		BM150	6.2	8.2	2.22	4.1	1.49	21.3	4.55	76.1	80.1	48.9	68.9	74.7	65.1	82.6	74.7	65.1	82.6		
100	PIT	POC	1084	1084-1A	200	0.00	0		BM150	6.4	8.2	2.32	3.6	1.54	12.3	2.95	44.7	47.4	70.3	81.6	44.3	79.7	89.7	44.3	79.7	89.7		
101	PIT	POC	1084	1084-1B	200	0.00	0		BM150	6.1	8.4	2.32	3.8	1.56	16.7	3.69	64.3	68.0	56.8	72.6	63.5	70.5	84.7	63.5	70.5	84.7		
102	PIT	POC	1084	1084-1C	200	0.00	0		BM150	5.8	8.4	2.29	4.5	1.62	20.5	4.36	75.6	79.3	46.5	63.4	74.0	63.5	79.5	74.0	63.5	79.5		

Batch Test Results Summary

BATCH TEST CONDITIONS AND RESULTS SUMMARY (Continued)

Test No	Coal	Source	Batch	Sample ID	Size mesh	Diesel %	HS RPM	HS Sol %	Frother	Flot Sol%	Feed			Final Product			Tailings			Agglomeration Performance			ROM Performance		
											Ash %	TS %	%	Ash %	TS %	%	As _h %	TS %	%	Yield %	Bitu Rec %	TSR %	PSR %	%	Bitu Rec %
103	PIT	POC	2020	2020-1A	200	0.13	2000	20.8	BM150	8.5	8.1	2.17	4.0	1.59	13.4	2.93	57.0	59.9	58.1	71.1	55.9	71.4	83.8		
104	PIT	POC	2020	2020-1B	200	0.13	2000	20.5	BM150	8.4	8.1	2.16	3.3	1.61	20.4	3.97	76.5	80.1	43.1	59.8	74.8	61.1	77.5		
105	PIT	POC	2020	2020-1C	200	0.13	2000	20.4	BM150	8.3	8.1	2.17	4.5	1.64	27.5	5.06	84.5	88.3	36.1	53.6	82.4	56.4	74.0		
106	PIT	POC	1020	1020-A	200	0.00	0		BM150	5.2	7.8	2.17	4.4	1.66	21.3	4.19	80.0	83.4	38.6	54.6	77.8	58.1	74.6		
107	PIT	POC	1020	1020-B	200	0.00	0		BM150	5.2	7.8	2.16	4.6	1.66	31.5	5.90	88.2	91.7	32.3	49.7	85.5	53.8	71.9		
108	PIT	POC	1020	1020-C	200	0.00	0		BM150	5.2	7.8	2.16	4.7	1.67	35.2	6.54	89.8	93.3	30.7	48.1	87.1	52.7	71.0		
109	PIT	POC	1037	1037-1A	200	0.00	0		BM150	7.1	8.2	2.23	3.4	1.40	9.6	2.48	23.3	24.6	85.4	92.3	23.0	90.0	95.7		
110	PIT	POC	1037	1037-1B	200	0.00	0		BM150	6.5	8.2	2.23	3.6	1.40	11.0	2.73	37.8	40.0	76.2	87.4	37.3	83.8	92.9		
111	PIT	POC	1037	1037-1C	200	0.00	0		BM150	6.2	8.2	2.22	3.9	1.45	14.0	3.28	57.7	60.7	62.4	78.4	56.7	74.4	87.9		
112	PIT	POC	1037	1037-1D	200	0.00	0		BM150	6.1	8.2	2.23	4.1	1.48	15.5	3.54	63.8	67.1	57.6	74.6	62.6	71.1	85.8		
113	PIT	POC	1060	1060-A	200	0.00	0		BM150	5.6	8.0	2.24	3.1	1.36	10.2	2.65	32.0	33.0	80.6	90.5	31.5	86.7	94.7		
114	PIT	POC	1060	1060-B	200	0.00	0		BM150	5.7	8.0	2.24	3.3	1.41	12.6	3.06	49.8	52.7	68.6	83.2	49.2	78.6	90.6		
115	PIT	POC	1060	1060-C	200	0.00	0		BM150	5.7	8.0	2.24	3.6	1.46	16.2	3.70	65.2	68.8	57.5	75.4	64.2	71.0	86.2		
116	PIT	LAB	1	P8-2-A	400	0.47	2000	15.9	BM150	6.5	7.7	2.23	4.1	1.58	14.1	3.36	63.6	66.4	54.9	69.6	62.0	69.2	83.0		
117	PIT	LAB	1	P8-2-B	400	0.48	2000	15.6	BM150	6.3	7.7	2.23	4.3	1.59	24.5	5.35	82.9	86.4	41.0	60.0	80.6	59.7	77.6		
118	PIT	LAB	1	P8-2-C	400	0.49	2000	15.5	BM150	6.3	7.7	2.23	4.4	1.59	29.8	6.41	86.7	90.3	38.2	57.9	84.3	57.8	76.5		
119	PIT	LAB	1	P8-3-A	200	0.19	2000	10.0	BM150	4.0	7.7	2.23	2.7	1.25	8.8	2.43	16.9	18.0	90.5	96.5	16.8	93.5	98.0		
120	PIT	LAB	1	P8-3-B	200	0.17	2000	11.4	BM150	4.6	7.7	2.22	3.1	1.29	9.2	2.51	23.4	24.8	86.4	94.2	23.1	90.7	96.7		
121	PIT	LAB	1	P8-3-C	200	0.15	2000	12.7	BM150	5.1	7.7	2.23	3.1	1.30	10.0	2.68	32.9	34.8	80.8	91.6	32.5	86.9	95.3		
122	PIT	LAB	1	P8-3-D	200	0.14	2000	13.7	BM150	5.5	7.7	2.23	3.3	1.35	11.9	3.04	47.8	50.5	71.1	86.0	47.1	80.3	92.2		
123	PIT	LAB	1	P8-4-A	400	0.49	2000	15.6	BM150	6.3	7.7	2.24	4.1	1.51	14.0	3.50	63.0	65.9	57.7	73.9	61.4	71.1	85.4		
124	PIT	LAB	1	P8-4-B	400	0.50	2000	15.1	BM150	6.1	7.7	2.25	4.2	1.52	22.6	5.29	80.7	84.3	45.4	65.7	78.6	62.8	80.8		
125	PIT	LAB	1	P8-4-C	400	0.51	2000	14.9	BM150	6.0	7.7	2.24	4.4	1.56	46.6	10.15	92.1	95.9	36.0	57.8	89.5	56.3	76.4		
126	PIT	LAB	1	P8-4-D	400	0.51	2000	14.9	BM150	6.0	7.7	2.25	4.5	1.59	56.4	12.16	93.8	97.5	33.6	54.9	91.0	54.7	74.8		
127	PIT	LAB	1	P8-5-A	400	0.15	2000	12.5	BM150	5.0	7.7	2.15	2.7	1.28	9.7	2.49	28.0	29.7	83.4	93.0	27.7	88.6	96.1		
128	PIT	LAB	1	P8-5-B	400	0.14	2000	13.5	BM150	5.5	7.7	2.15	2.8	1.26	11.3	2.79	42.0	44.6	75.4	90.1	41.6	83.2	94.5		
129	PIT	LAB	1	P8-5-C	400	0.14	2000	14.1	BM150	5.7	7.7	2.15	3.0	1.27	15.4	3.56	61.7	65.3	63.5	84.8	60.9	75.1	91.5		
129	PIT	LAB	1	P8-5-D	400	0.13	2000	14.3	BM150	5.8	7.7	2.14	3.3	1.30	18.3	4.12	70.1	74.0	57.5	80.8	69.0	71.0	89.3		
130	PIT	LAB	1	P8-6-A	100	0.50	2000	15.1	BM150	6.1	7.7	2.27	5.3	1.83	30.2	6.25	90.1	92.9	27.2	40.8	86.7	50.3	66.8		
131	PIT	LAB	1	P8-6-A	100	0.51	2000	15.1	BM150	6.1	7.7	2.27	5.4	1.84	43.6	8.68	93.8	96.5	23.9	37.7	90.1	48.1	65.1		
132	PIT	LAB	1	P8-6-B	100	0.51	2000	15.1	BM150	6.1	7.7	2.27	5.4	1.84	43.6	8.68	93.8	96.5	23.9	37.7	90.1	48.1	65.1		
132	PIT	LAB	1	P8-6-B	100	0.51	2000	15.1	BM150	6.1	7.7	2.26	5.4	1.85	46.1	9.06	94.3	97.0	22.8	36.1	90.5	47.3	64.3		
133	PIT	LAB	1	P8-6-C	100	0.51	2000	15.1	BM150	6.1	7.7	2.26	5.4	1.85	46.1	9.06	94.3	97.0	22.8	36.1	90.5	47.3	64.3		
134	PIT	LAB	1	P8-7-A	100	0.13	2000	14.7	BM150	5.9	7.7	2.26	3.6	1.56	11.5	2.88	47.2	49.7	67.4	78.9	46.3	77.7	88.2		

BATCH TEST CONDITIONS AND RESULTS SUMMARY (Continued)

Test No	Coal	Source	Batch	Sample ID	Size mesh	Diesel %	HS RPM	HS Sol %	Frother	Flot. Sol%	Feed			Final Product			Tailings			Agglomeration Performance						ROM Performance		
											Ash %	TS %	%	Ash %	TS %	%	Ash %	TS %	%	Yield %	Blu Rec %	TSR %	PSR %	%	Blu Rec %	TSR %	PSR %	%
135	PIT	LAB	1	P8-7-B	100	0.13	2000	14.8	BM150	6.0	7.7	2.25	3.7	1.56	14.6	3.43	63.2	66.3	56.2	71.5	61.9	70.1	84.1					
136	PIT	LAB	1	P8-7-C	100	0.13	2000	14.8	BM150	6.0	7.7	2.25	3.8	1.57	16.7	3.81	69.6	73.0	51.5	68.1	68.1	66.9	82.1					
137	PIT	LAB	1	P8-7-D	100	0.13	2000	14.8	BM150	6.0	7.7	2.25	4.1	1.61	18.1	4.08	73.9	77.3	47.2	63.8	72.1	64.0	79.8					
138	UF1	POC	118	118-1A	200					12.9	10.6	1.78	2.4	0.76	11.0	1.83	4.7	5.2	98.0	99.6	5.2	98.9	99.8					
139	UF1	POC	118	118-1B	200					9.9	10.6	1.79	2.4	0.77	11.8	1.93	12.5	13.8	94.6	98.8	13.6	97.0	99.5					
140	UF1	POC	118	118-1C	200					9.1	10.6	1.78	2.5	0.78	13.0	2.08	23.0	25.4	89.9	97.5	25.0	94.4	98.9					
141	UF1	POC	118	118-1D	200					8.5	10.6	1.78	3.0	0.83	21.1	3.10	58.0	63.7	73.0	91.1	62.7	85.0	96.2					
142	UF1	POC	118	118-2A	200				BM150	9.8	10.6	1.75	2.5	0.79	11.2	1.82	6.8	7.5	96.9	99.2	7.4	98.3	99.7					
143	UF1	POC	118	118-2B	200				BM150	9.0	10.6	1.75	2.5	0.79	11.8	1.89	13.0	14.3	94.1	98.4	14.1	96.7	99.3					
144	UF1	POC	118	118-2C	200				BM150	8.6	10.6	1.75	2.6	0.79	13.1	2.05	23.5	26.0	89.4	97.1	25.6	94.1	98.8					
145	UF1	POC	118	118-2D	200				BM150	8.1	10.6	1.75	3.2	0.85	22.7	3.22	62.0	68.0	69.9	89.0	66.9	83.3	95.3					
146	UF1	POC	118	118-3A	200					9.0	10.6	1.73	5.4	1.20	17.1	2.40	55.7	59.4	61.4	72.1	58.5	78.6	88.1					
147	UF1	POC	118	118-3B	200					8.6	10.6	1.73	5.8	1.25	34.0	4.05	82.9	88.0	40.1	54.6	86.6	66.9	80.7					
148	UF1	POC	118	118-3C	200					8.6	10.6	1.73	5.9	1.26	48.8	5.50	88.9	94.4	35.2	50.4	92.9	64.1	78.9					
149	UF1	POC	118	118-3D	200					8.6	10.6	1.73	5.9	1.27	51.0	5.71	89.6	95.0	34.3	49.4	93.5	63.6	78.5					
150	UF1	POC	118	118-4A	200				BM150	8.5	10.6	1.72	5.8	1.20	50.8	3.88	80.8	85.8	43.5	58.7	84.4	68.7	82.5					
151	UF1	POC	118	118-4B	200				BM150	8.4	10.6	1.71	6.0	1.23	50.9	5.92	89.7	95.1	35.6	51.6	93.5	64.3	79.4					
152	UF1	POC	118	118-4C	200				BM150	8.4	10.6	1.72	6.0	1.24	53.9	6.23	90.4	95.8	34.7	50.5	94.3	63.8	79.0					
153	UF1	POC	118	118-5A	200	0.87	1200	23.8		9.8	10.6	1.68	3.7	0.91	12.4	1.88	21.0	22.8	88.6	94.8	22.5	93.7	97.8					
154	UF1	POC	118	118-5B	200	0.96	1200	21.8		8.9	10.6	1.68	4.0	0.94	15.6	2.24	43.1	46.8	75.9	88.1	56.0	86.6	94.9					
155	UF1	POC	118	118-5C	200	0.99	1200	21.0		8.6	10.6	1.68	4.6	0.99	25.4	3.36	71.1	76.6	58.0	76.7	75.4	76.7	90.1					
156	UF1	POC	118	118-5D	200	1.00	1200	20.8		8.5	10.6	1.68	5.0	1.04	38.5	4.86	83.2	89.3	48.5	68.8	87.9	71.5	86.7					
157	UF1	POC	118	118-6A	200	0.19	2000	22.5		9.3	10.6	1.68	3.8	0.92	12.7	1.91	23.1	25.1	87.4	94.1	24.7	93.0	97.5					
158	UF1	POC	118	118-6B	200	0.20	2000	21.2		8.7	10.6	1.69	4.0	0.96	16.0	2.28	44.9	48.8	74.4	86.8	48.0	85.8	94.4					
159	UF1	POC	118	118-6C	200	0.20	2000	20.6		8.5	10.6	1.68	4.4	1.00	25.0	3.26	69.8	75.4	58.5	76.7	74.2	77.0	90.1					
160	UF1	POC	118	118-6D	200	0.20	2000	20.5		8.4	10.6	1.68	4.7	1.04	33.3	4.17	79.4	85.5	51.0	70.5	84.1	72.9	87.5					
161	UF1	POC	118	118-7A	200	0.60	2000	21.5		8.8	10.6	1.68	6.1	1.29	26.9	3.10	78.3	82.8	40.0	52.2	81.5	66.8	79.7					
162	UF1	POC	118	118-7B	200	0.61	2000	21.2		8.7	10.6	1.69	6.2	1.32	45.5	4.62	88.9	93.9	30.4	43.4	92.4	61.5	75.9					
163	UF1	POC	118	118-7C	200	0.61	2000	21.2		8.7	10.6	1.68	6.3	1.33	58.0	5.61	91.8	96.8	27.5	40.4	95.3	59.9	74.7					
164	UF1	POC	118	118-8A	200	0.95	2000	21.4		8.8	10.6	1.80	6.7	1.41	24.6	3.20	78.0	82.0	39.1	49.2	80.6	66.3	78.4					
165	UF1	POC	118	118-8B	200	0.96	2000	21.1		8.7	10.6	1.81	6.7	1.43	42.2	4.89	89.1	93.6	29.5	40.5	92.1	61.0	74.7					
166	UF1	POC	118	118-8C	200	0.97	2000	21.0		8.6	10.6	1.80	7.0	1.45	67.2	7.24	94.0	98.4	24.3	35.2	96.8	58.1	72.4					

Batch Test Results Summary

BATCH TEST CONDITIONS AND RESULTS SUMMARY (Continued)

Test No	Coal	Source	Batch	Sample ID	Size mesh	Diesel %	HS RPM	HS Sol %	Frother	Flot. Sol%	Feed			Final Product			Tailings			Agglomeration Performance			ROM Performance		
											Ash %	TS %	%	Ash %	TS %	%	Ash %	TS %	%	Yield %	But Rec %	TSR %	PSR %	But Rec %	TSR %
167	UF1	POC	121	121-5A	200	0.17	2000	17.3	BM150	7.1	10.6	1.61	3.8	0.83	14.3	2.03	34.9	37.9	82.0	93.5	37.3	90.1	97.3		
168	UF1	POC	121	121-5B	200	0.17	2000	16.9	BM150	6.9	10.6	1.61	4.0	0.84	19.2	2.62	56.5	61.4	70.8	88.9	60.4	83.7	95.3		
169	UF1	POC	121	121-5C	200	0.17	2000	16.7	BM150	6.8	10.6	1.61	4.3	0.87	26.9	3.54	72.1	78.0	61.1	83.6	76.7	78.5	93.0		
170	UF1	POC	121	121-6A	200	0.54	2000	18.7	BM150	7.6	10.6	1.55	6.2	1.14	43.3	4.58	88.2	93.2	35.0	52.3	91.7	64.0	79.7		
171	UF1	POC	121	121-6B	200	0.54	2000	18.6	BM150	7.6	10.6	1.55	6.3	1.16	61.7	6.19	92.3	97.4	30.9	48.2	95.8	61.8	78.0		
172	UF1	POC	118	118-9A	200	0.46	2000	20.3	BM150	8.3	10.6	1.68	5.9	1.23	31.5	3.66	81.7	86.6	40.1	54.4	85.2	66.8	80.6		
173	UF1	POC	118	118-9B	200	0.47	2000	20.2	BM150	8.3	10.6	1.68	6.0	1.24	43.9	4.87	87.9	93.1	35.0	50.1	91.6	64.0	78.8		
174	UF1	POC	118	118-9C	200	0.47	2000	20.2	BM150	8.3	10.6	1.68	6.1	1.24	46.7	5.14	88.8	94.0	34.3	49.5	92.5	63.6	78.5		
175	UF1	POC	118	118-10A	200	0.48	2000	19.5	BM150	8.0	10.6	1.65	5.0	1.06	23.5	3.00	69.6	74.7	55.2	71.8	73.5	75.2	88.0		
176	UF1	POC	118	118-10B	200	0.49	2000	19.3	BM150	7.9	10.6	1.65	5.3	1.10	37.4	4.38	83.4	89.1	44.3	62.8	87.7	69.2	84.2		
177	UF1	POC	118	118-10C	200	0.49	2000	19.3	BM150	7.9	10.6	1.65	5.3	1.11	40.0	4.65	84.7	90.5	43.0	61.5	89.1	68.4	83.6		
178	PIT	POC	1084	1084-2A	200				BM150	6.0	8.4	2.25	4.6	1.71	18.0	3.64	71.8	75.2	45.6	59.4	70.1	62.8	77.3		
179	PIT	POC	1084	1084-2B	200				BM150	5.9	8.4	2.26	4.8	1.72	29.0	5.36	85.2	89.0	35.1	51.2	83.1	55.7	72.7		
180	PIT	POC	1084	1084-2C	200				BM150	5.9	8.4	2.25	5.0	1.74	36.1	6.40	89.0	92.8	31.4	47.5	86.6	53.1	70.6		
181	PIT	POC	2020	2020-2	200	0.49	2000	18.3	BM150	7.5	8.1	2.16	5.2	1.80	47.3	7.05	93.2	96.5	22.3	35.9	90.1	46.9	64.1		
182	PIT	POC	2163	2163-1A	200				BM150	6.0	8.1	2.13	2.8	1.36	9.4	2.32	19.5	20.8	87.6	93.7	19.4	91.5	96.5		
183	PIT	POC	2163	2163-1B	200				BM150	6.0	8.1	2.13	3.0	1.39	10.6	2.48	32.2	34.2	79.0	88.8	31.9	85.7	93.7		
184	PIT	POC	2163	2163-1C	200				BM150	5.9	8.1	2.13	3.2	1.43	12.2	2.71	45.3	48.1	69.6	82.7	44.9	79.2	90.3		
185	PIT	POC	2163	2163-1D	200				BM150	5.9	8.1	2.13	3.3	1.45	12.8	2.80	49.6	52.5	66.3	80.2	49.0	77.0	88.9		
186	PIT	POC	2190	2190-A	200				BM150	6.3	8.0	2.12	4.3	1.65	19.0	3.53	75.0	78.4	41.6	56.6	73.2	60.2	75.7		
187	PIT	POC	2190	2190-B	200				BM150	6.2	8.0	2.12	4.4	1.66	27.3	4.63	84.6	88.3	33.6	50.1	82.4	54.7	72.1		
188	PIT	POC	2190	2190-C	200				BM150	6.2	8.0	2.12	4.5	1.67	30.5	5.06	86.7	90.4	31.7	48.3	84.4	53.4	71.1		
189	PIT	POC	2163	2163-2A	200				BM150	4.7	8.1	2.17	5.9	1.81	9.0	2.30	27.6	28.4	76.9	80.6	26.5	84.2	89.2		
190	PIT	POC	2163	2163-2B	200				BM150	5.3	8.1	2.17	5.1	1.71	10.2	2.47	40.1	41.6	68.4	75.5	38.8	78.4	86.3		
191	PIT	POC	2163	2163-2C	200				BM150	5.6	8.1	2.16	4.4	1.61	13.4	2.96	59.0	61.7	56.1	69.0	57.5	70.1	82.6		
192	PIT	POC	2163	2163-2D	200				BM150	5.6	8.1	2.17	4.2	1.59	16.0	3.35	66.9	70.2	51.0	66.2	65.5	66.6	81.1		
193	PIT	POC	2163	2163-3A	200				BM150	22.9	8.1	2.16	2.5	1.29	8.3	2.18	2.7	2.9	98.4	99.3	2.7	98.9	99.6		
194	PIT	POC	2163	2163-3B	200				BM150	9.2	8.1	2.15	2.7	1.35	9.0	2.28	14.0	14.9	91.2	95.7	13.9	94.0	97.6		
195	PIT	POC	2163	2163-3C	200				BM150	7.4	8.1	2.15	2.9	1.38	10.1	2.45	27.9	29.8	82.1	90.7	27.8	87.8	94.8		
196	PIT	POC	2163	2163-3D	200				BM150	6.4	8.1	2.14	3.7	1.50	16.3	3.33	64.8	68.4	54.7	71.6	63.8	69.1	84.1		
197	PIT	POC	2209	2209-1A	200	0.51	2000	19.7	BM150	8.1	7.9	2.02	4.2	1.45	17.5	3.47	71.9	75.3	48.3	67.7	70.2	64.7	81.9		
198	PIT	POC	2209	2209-1B	200	0.51	2000	19.8	BM150	8.1	7.9	2.02	4.4	1.49	40.2	6.94	90.2	94.2	33.5	56.1	87.9	54.6	75.4		

BATCH TEST CONDITIONS AND RESULTS SUMMARY (Continued)

Test No	Coal	Source	Batch	Sample ID	Size mesh	Diesel %	HS RPM	HS Sol %	Frother	Flot. Sol%	Feed			Final Product			Tailings			Agglomeration Performance				ROM Performance			
											Ash %	TS %	%	Ash %	TS %	%	Ash %	TS %	%	Yield %	Blu Rec %	TSR %	PSR %	Blu Rec %	TSR %	PSR %	%
200	PIT	POC	2209	2209-1C	200	0.51	2000	19.8	BM150	8.1	7.9	2.02	4.6	1.51	56.7	9.46	93.6	97.5	30.0	52.6	91.0	52.2	73.4	91.0	52.2	73.4	
201	PIT	POC	2209	2209-1A	200	0.51	2000	19.5	BM150	8.0	7.9	1.95	4.0	1.36	13.9	2.84	60.4	63.4	57.8	76.3	59.1	71.2	86.8	59.1	71.2	86.8	
202	PIT	POC	2209	2209-1B	200	0.50	2000	19.7	BM150	8.0	7.9	1.94	4.1	1.38	25.4	4.54	82.1	86.0	41.7	66.1	80.3	60.2	81.0	80.3	60.2	81.0	
203	PIT	POC	2209	2209-1C	200	0.50	2000	19.7	BM150	8.1	7.9	1.94	4.4	1.44	66.2	10.37	94.3	98.4	30.1	55.3	91.9	52.3	75.0	91.9	52.3	75.0	
204	PIT	POC	2232	2232-1A	200	0.54	2000	19.3	BM150	7.9	7.8	2.06	4.5	1.59	42.2	6.93	91.1	94.9	29.8	49.3	88.6	52.1	71.6	88.6	52.1	71.6	
205	PIT	POC	2232	2232-1B	200	0.56	2000	18.4	BM150	7.5	7.8	2.07	4.6	1.61	58.5	9.18	94.0	97.8	26.7	46.0	91.2	50.0	69.8	91.2	50.0	69.8	
206	PIT	POC	2232	2232-2	200	0.51	2000	17.5	BM150	7.1	7.8	2.08	5.2	1.79	60.2	7.95	95.3	98.4	18.0	30.7	91.8	44.0	61.2	91.8	44.0	61.2	
207	PIT	POC	2270	2270-A	200				BM150	17.6	8.0	2.13	3.0	1.36	8.3	2.18	6.0	6.3	96.2	98.1	5.9	97.4	98.9	5.9	97.4	98.9	
208	PIT	POC	2270	2270-B	200				BM150	8.2	8.0	2.13	3.1	1.39	9.9	2.41	27.8	29.5	81.9	90.3	27.5	87.6	94.6	27.5	87.6	94.6	
209	PIT	POC	2270	2270-C	200				BM150	6.8	8.0	2.12	3.5	1.45	14.0	3.02	57.1	60.3	61.0	77.0	56.3	73.4	87.1	56.3	73.4	87.1	
210	PIT	POC	2270	2270-D	200				BM150	6.5	8.0	2.12	4.1	1.56	20.2	3.89	75.9	79.5	44.2	62.0	74.2	61.9	78.7	74.2	61.9	78.7	
211	PIT	POC	2208		200	0.43	2000	19.6	BM150	8.0	8.0	2.15	5.7	1.97	50.3	5.60	95.0	97.6	13.0	20.6	91.1	40.6	55.5	91.1	40.6	55.5	
212	PIT	POC	2232	2232-3A	200	0.50	2000	18.1	BM150	7.4	7.8	2.02	4.3	1.51	30.9	5.32	86.7	90.5	35.1	56.1	84.5	55.7	75.4	84.5	55.7	75.4	
213	PIT	POC	2232	2232-3B	200	0.50	2000	18.1	BM150	7.4	7.8	2.01	4.4	1.52	41.2	6.79	90.7	94.5	31.5	53.0	88.2	53.2	73.7	88.2	53.2	73.7	
214	PIT	POC	2232	2232-4A	200	0.51	800	18.1	BM150	7.4	7.8	2.07	5.1	1.77	43.3	6.00	92.9	96.0	20.5	33.4	89.6	45.7	62.8	89.6	45.7	62.8	
215	PIT	POC	2232	2232-4B	200	0.51	800	18.1	BM150	7.4	7.8	2.07	5.1	1.77	43.3	6.00	92.9	96.0	20.5	33.4	89.6	45.7	62.8	89.6	45.7	62.8	
216	PIT	POC	2232	2232-5A	200	0.51	800	16.8	BM150	6.8	7.8	2.11	4.9	1.73	36.0	5.79	90.6	93.9	25.7	40.5	87.6	49.3	66.7	87.6	49.3	66.7	
217	PIT	POC	2232	2232-5B	200	0.50	800	16.8	BM150	6.8	7.8	2.12	5.0	1.74	43.1	6.77	92.5	95.8	23.9	38.7	89.4	48.1	65.7	89.4	48.1	65.7	
218	PIT	POC	2232	2232-6A	200	0.50	800	16.6	BM150	6.8	7.8	2.06	4.6	1.68	26.9	4.29	85.6	89.0	30.1	45.2	83.0	52.3	69.3	83.0	52.3	69.3	
219	PIT	POC	2232	2232-6B	200	0.49	800	16.8	BM150	6.9	7.8	2.06	4.7	1.69	26.9	4.29	85.7	89.1	29.7	44.5	83.1	52.0	69.0	83.1	52.0	69.0	
220	PIT	POC	2232	2232-7A	200	0.50	800	17.4	BM150	7.1	7.8	2.10	4.8	1.74	32.4	5.02	89.1	92.4	26.2	40.3	86.2	49.6	66.6	86.2	49.6	66.6	
221	PIT	POC	2232	2232-7B	200	0.50	800	17.3	BM150	7.1	7.8	2.09	4.9	1.74	39.5	5.86	91.4	94.8	24.0	38.4	88.4	48.1	65.5	88.4	48.1	65.5	
222	PIT	POC	2281	2281-1A	200				BM150	6.5	8.4	2.28	4.9	1.84	21.4	3.96	79.2	82.6	36.1	48.4	77.0	56.4	71.1	77.0	56.4	71.1	
223	PIT	POC	2281	2281-1B	200				BM150	6.4	8.4	2.28	5.2	1.86	32.7	5.44	88.4	91.9	27.8	40.8	85.7	50.7	66.8	85.7	50.7	66.8	
224	PIT	POC	2281	2281-1C	200				BM150	6.4	8.4	2.28	5.3	1.88	39.3	6.34	91.0	94.5	25.0	37.8	88.1	48.8	65.2	88.1	48.8	65.2	
225	PIT	POC	2278	2278-1A	200	0.50	700	18.0	BM150	7.4	8.5	2.25	5.6	1.92	38.6	5.68	91.3	94.6	22.0	33.0	88.3	46.7	62.5	88.3	46.7	62.5	
226	PIT	POC	2278	2278-1B	200	0.50	700	18.0	BM150	7.4	8.5	2.25	5.7	1.93	45.1	6.43	92.9	96.1	20.4	31.4	89.7	45.6	61.6	89.7	45.6	61.6	
227	PIT	POC	2278	2278-2A	200	0.49	1200	17.2	BM150	7.0	8.5	2.29	5.3	1.89	30.6	5.06	87.5	90.9	27.7	39.8	84.9	50.7	66.3	84.9	50.7	66.3	
228	PIT	POC	2278	2278-2B	200	0.49	1200	17.1	BM150	7.0	8.5	2.29	5.4	1.90	42.7	6.62	91.8	95.3	23.7	36.0	88.9	47.9	64.2	88.9	47.9	64.2	
229	PIT	POC	2278	2278-2C	200	0.49	1200	17.1	BM150	7.0	8.5	2.29	5.5	1.91	45.7	7.00	92.5	96.0	22.9	35.1	89.6	47.3	63.7	89.6	47.3	63.7	
230	PIT	POC	2278	2278-3A	200	0.49	1800	16.7	BM150	6.8	8.5	2.32	5.6	1.97	44.7	6.77	92.8	96.1	21.1	32.0	89.6	46.1	61.9	89.6	46.1	61.9	
231	PIT	POC	2278	2278-3B	200	0.49	1800	16.6	BM150	6.8	8.5	2.31	5.7	1.97	60.9	8.77	95.0	98.3	19.0	30.0	91.7	44.7	60.8	91.7	44.7	60.8	

Batch Test Results Summary

BATCH TEST CONDITIONS AND RESULTS SUMMARY (Continued)

Test No	Coal	Source	Batch	Sample ID	Size mesh	Diesel %	HS RPM	HS Sol %	Frother	Flot. Sol%	Feed			Final Product			Tailings			Agglomeration Performance			ROM Performance		
											Ash %	TS %	%	Ash %	TS %	%	Ash %	TS %	%	Yield %	Blu Rec %	TSR %	PSR %	Blu Rec %	TSR %
232	PIT	POC	2280	2280-A	200		1200		BM150	7.5	8.4	2.28	3.9	1.65	11.5	2.73	41.5	43.8	70.0	79.0	40.9	79.5	88.3	62.8	67.9
233	PIT	POC	2280	2280-B	200		1200		BM150	7.1	8.4	2.28	4.1	1.68	15.9	3.35	63.9	67.3	53.0	66.3	62.8	67.9	81.1	78.2	58.9
234	PIT	POC	2280	2280-C	200		1200		BM150	7.0	8.4	2.28	4.4	1.72	24.1	4.50	79.9	83.8	39.8	55.2	78.2	58.9	74.9	82.6	55.7
235	PIT	POC	2280	2280-D	200		1200		BM150	7.0	8.4	2.28	4.6	1.75	28.9	5.21	84.6	88.6	35.1	50.6	82.6	55.7	72.4	46.1	76.5
236	PIT	POC	2281	2281-2A	200		1800		BM150	6.8	8.4	2.31	4.3	1.69	12.0	2.86	47.0	49.4	65.6	75.3	46.1	76.5	86.2	68.8	64.7
237	PIT	POC	2281	2281-2B	200		1800		BM150	6.7	8.4	2.31	4.4	1.70	17.7	3.75	70.3	73.8	48.2	62.4	68.8	64.7	78.9	51.5	56.2
238	PIT	POC	2281	2281-2C	200		1800		BM150	6.7	8.4	2.31	4.7	1.75	28.8	5.45	84.8	88.7	35.8	51.5	82.8	56.2	72.8	46.7	87.1
239	PIT	POC	2281	2281-2D	200		1800		BM150	6.7	8.4	2.31	4.9	1.78	37.8	6.85	89.5	93.4	31.0	46.7	87.1	52.9	70.2	61.6	67.7
240	PIT	POC	2279	2279-A	200		700		BM150	7.1	8.5	2.31	4.5	1.74	15.2	3.29	62.9	66.0	52.7	64.7	61.6	67.7	80.2	78.5	58.0
241	PIT	POC	2279	2279-B	200		700		BM150	7.0	8.5	2.31	4.7	1.77	23.7	4.53	80.3	84.1	38.5	53.0	78.5	58.0	73.7	84.5	93.9
242	PIT	POC	2279	2279-C	200		700		BM150	7.0	8.5	2.31	4.9	1.79	31.8	5.73	86.7	90.6	33.0	48.0	84.5	54.2	70.9	66.7	83.7
243	PIT	POC	2279	2279-D	200		700		BM150	7.0	8.5	2.31	5.0	1.80	35.6	6.31	88.6	92.5	31.0	46.0	86.3	52.9	69.8	65.1	66.7
245	ILL	LAB	1	I-5-1A		0.57	2000	13.5	BM150	5.5	18.1	1.79	5.7	0.97	23.1	2.12	28.6	33.4	84.5	93.9	33.4	84.5	93.9	66.7	83.7
246	ILL	LAB	1	I-5-1B		0.58	2000	13.1	BM150	5.3	18.1	1.77	6.1	1.05	33.5	2.68	56.0	65.1	66.7	83.7	65.1	66.7	83.7	75.6	60.8
247	ILL	LAB	1	I-5-1C		0.57	2000	13.5	BM150	5.5	18.1	1.79	6.2	1.08	40.4	3.11	65.0	75.6	60.8	79.7	75.6	60.8	79.7	81.3	56.6
248	ILL	LAB	1	I-5-1D		0.57	2000	13.5	BM150	5.5	18.1	1.79	6.3	1.11	45.8	3.39	70.1	81.3	56.6	76.3	81.3	56.6	76.3	24.6	88.1
249	ILL	LAB	2	I-5-2A		0.62	2000	12.3	BM150	5.0	18.5	1.92	3.6	1.12	22.3	2.13	20.5	24.6	88.1	94.0	24.6	88.1	94.0	37.3	81.9
250	ILL	LAB	2	I-5-2B		0.62	2000	12.2	BM150	4.9	18.5	1.92	3.5	1.12	25.2	2.28	31.0	37.3	81.9	90.9	37.3	81.9	90.9	51.2	74.4
251	ILL	LAB	2	I-5-2C		0.62	2000	12.3	BM150	4.9	18.5	1.92	3.6	1.16	29.5	2.49	42.5	51.2	74.4	86.2	51.2	74.4	86.2	62.1	68.5
252	ILL	LAB	2	I-5-2D		0.62	2000	12.3	BM150	4.9	18.5	1.92	3.6	1.17	34.3	2.73	51.6	62.1	68.5	82.8	62.1	68.5	82.8	18.1	90.7
253	ILL	POC	25221	2522-1A	100	0.51	1200	15.0	BM150	6.1	17.8	1.86	3.3	1.15	20.4	1.99	15.1	18.1	90.7	95.0	18.1	90.7	95.0	29.8	84.2
254	ILL	POC	25221	2522-1B	100	0.51	1200	15.0	BM150	6.1	17.8	1.86	3.4	1.18	22.6	2.09	24.9	29.8	84.2	91.1	29.8	84.2	91.1	45.5	75.2
255	ILL	POC	25221	2522-1C	100	0.51	1200	15.0	BM150	6.1	17.8	1.87	3.5	1.21	26.7	2.27	38.2	45.5	75.2	85.4	45.5	75.2	85.4	66.2	61.7
256	ILL	POC	25221	2522-1D	100	0.51	1200	15.0	BM150	6.1	17.8	1.86	3.9	1.28	35.4	2.59	55.7	66.2	61.7	75.4	66.2	61.7	75.4	23.2	87.0
271	ILL	POC	25222	2522-2A	100	0.50	2000	14.1	BM150	5.7	17.7	1.93	3.6	1.29	21.1	2.09	19.5	23.2	87.0	91.7	23.2	87.0	91.7	38.1	78.5
272	ILL	POC	25222	2522-2B	100	0.50	2000	14.2	BM150	5.7	17.7	1.93	3.5	1.30	24.4	2.23	32.0	38.1	78.5	86.2	38.1	78.5	86.2	56.8	66.9
273	ILL	POC	25222	2522-2C	100	0.50	2000	14.2	BM150	5.7	17.7	1.94	3.7	1.34	30.5	2.48	47.7	56.8	66.9	77.9	56.8	66.9	77.9	78.7	51.1
274	ILL	POC	25222	2522-2D	100	0.50	2000	14.1	BM150	5.7	17.7	1.94	4.2	1.42	44.5	2.97	66.6	78.7	51.1	64.9	78.7	51.1	64.9	10.1	94.4
275	ILL	POC	25231	2523-1A	200	0.49	2000	14.9	BM150	6.0	17.9	1.86	4.2	1.21	19.2	1.92	8.5	10.1	94.4	96.7	10.1	94.4	96.7	26.7	85.3
276	ILL	POC	25231	2523-1B	200	0.49	2000	14.9	BM150	6.0	17.9	1.86	4.2	1.21	21.9	2.05	22.6	26.7	85.3	91.3	26.7	85.3	91.3	60.4	66.0
277	ILL	POC	25231	2523-1C	200	0.49	2000	14.9	BM150	6.0	17.9	1.86	4.1	1.24	32.2	2.50	50.9	60.4	66.0	79.0	60.4	66.0	79.0	63.9	47.4
278	ILL	POC	25231	2523-1D	200	0.49	2000	14.9	BM150	6.0	17.9	1.86	4.4	1.34	54.4	3.26	73.0	86.3	47.4	63.9	86.3	47.4	63.9	63.9	47.4

Batch Test Results Summary

BATCH TEST CONDITIONS AND RESULTS SUMMARY (Continued)

Test No	Coal	Source	Batch	Sample ID	Size mesh	Diesel %	HS RPM	HS Sol %	Frother	Flot. Sol%	Feed			Final Product			Tailings			Agglomeration Performance			ROM Performance		
											Ash %	TS %	Ash %	TS %	Ash %	TS %	Yield %	Blu Rec %	TSR %	PSR %	Ash %	TSR %	Blu Rec %	TSR %	PSR %
279	ILL	POC	25232	2523-2A	200	0.50	2000	14.3	BM150	5.8	17.8	1.85	4.0	1.29	21.7	2.01	22.3	26.4	84.5	89.9	26.4	84.5	89.9	26.4	84.5
280	ILL	POC	25232	2523-2B	200	0.50	2000	14.3	BM150	5.8	17.8	1.85	4.1	1.33	25.8	2.15	37.1	43.9	73.3	81.8	43.9	73.3	81.8	43.9	73.3
281	ILL	POC	25232	2523-2C	200	0.50	2000	14.3	BM150	5.8	17.8	1.85	4.0	1.32	35.4	2.53	56.3	66.7	59.8	72.9	66.7	59.8	72.9	66.7	59.8
282	ILL	POC	25232	2523-2D	200	0.50	2000	14.3	BM150	5.8	17.8	1.85	4.5	1.38	62.7	3.43	77.2	91.0	42.3	58.8	91.0	42.3	58.8	91.0	42.3
283	ILL	POC	25233	2523-3A	400	0.49	2000	15.8	BM150	6.4	18.5	1.88	9.3	1.44	21.5	2.02	24.4	27.5	81.3	85.5	27.5	81.3	85.5	27.5	81.3
284	ILL	POC	25233	2523-3B	400	0.49	2000	15.8	BM150	6.4	18.5	1.88	8.8	1.40	23.8	2.14	35.1	39.7	73.9	80.4	39.7	73.9	80.4	39.7	73.9
285	ILL	POC	25233	2523-3C	400	0.49	2000	15.8	BM150	6.4	18.5	1.88	7.7	1.36	31.3	2.50	54.3	62.3	60.8	71.8	62.3	60.8	71.8	62.3	60.8
286	ILL	POC	25233	2523-3D	400	0.49	2000	15.8	BM150	6.4	18.5	1.88	7.5	1.43	84.4	4.57	85.6	98.6	34.9	50.8	98.6	34.9	50.8	98.6	34.9
287	ILL	POC	25234	2523-4A	400	0.52	2000	15.0	BM150	6.1	18.1	1.80	6.1	1.08	19.0	1.85	6.4	7.4	96.2	98.0	7.4	96.2	98.0	7.4	96.2
288	ILL	POC	25234	2523-4B	400	0.53	2000	15.0	BM150	6.1	18.1	1.80	5.8	1.09	19.5	1.88	10.2	11.9	93.8	96.8	11.9	93.8	96.8	11.9	93.8
289	ILL	POC	25234	2523-4C	400	0.53	2000	15.0	BM150	6.1	18.1	1.80	8.4	1.35	67.3	4.06	83.5	94.5	37.3	53.7	94.5	37.3	53.7	94.5	37.3
290	ILL	POC	25234	2523-4D	400	0.53	2000	15.0	BM150	5.8	18.8	1.83	6.2	1.23	23.1	2.04	25.4	29.8	82.9	89.3	29.8	82.9	89.3	29.8	82.9
291	ILL	POC	25236	2523-6A	400	0.80	2000	14.3	BM150	5.8	18.8	1.84	5.8	1.20	28.2	2.30	41.9	49.3	72.7	83.5	49.3	72.7	83.5	49.3	72.7
292	ILL	POC	25236	2523-6B	400	0.80	2000	14.3	BM150	5.8	18.8	1.84	5.6	1.20	37.6	2.74	58.6	69.2	61.7	77.0	69.2	61.7	77.0	69.2	61.7
293	ILL	POC	25236	2523-6C	400	0.80	2000	14.3	BM150	5.8	18.8	1.84	5.4	1.21	62.0	3.86	76.4	90.3	49.7	69.4	90.3	49.7	69.4	90.3	49.7
294	ILL	POC	25236	2523-6D	400	0.80	2000	14.3	BM150	5.8	18.8	1.83	5.6	1.22	68.7	4.15	79.1	93.4	47.4	67.5	93.4	47.4	67.5	93.4	47.4
295	ILL	POC	25236	2523-6E	400	0.80	2000	14.3	BM150	5.8	18.8	1.83	5.0	1.33	18.7	2.74	74.6	77.8	41.2	54.4	77.8	41.2	54.4	77.8	41.2
296	ILL	LAB	3	MIX-1A		0.52	2000	14.6	BM150	5.9	8.5	1.69	5.0	1.44	85.7	8.56	96.6	99.7	17.5	29.5	99.7	17.5	29.5	99.7	17.5
297	ILL	LAB	3	MIX-1B		0.52	2000	14.6	BM150	5.9	8.5	1.68	5.7	1.44	85.7	8.56	96.6	99.7	17.5	29.5	99.7	17.5	29.5	99.7	17.5
298	ILL	LAB	4	MIX-2A		0.53	2000	14.4	BM150	5.8	9.0	1.69	3.8	1.32	11.9	1.89	35.4	37.7	72.3	79.1	37.7	72.3	79.1	37.7	72.3
299	ILL	LAB	4	MIX-2B		0.53	2000	14.4	BM150	5.8	9.0	1.68	3.9	1.31	14.9	2.11	53.3	56.6	58.6	69.0	56.6	58.6	69.0	56.6	58.6
300	ILL	LAB	4	MIX-2C		0.53	2000	14.4	BM150	5.8	9.0	1.69	5.5	1.45	84.0	6.71	95.5	99.6	17.8	29.8	99.6	17.8	29.8	99.6	17.8
301	ILL	LAB	5	MIX-3A		0.53	2000	14.4	BM150	5.8	8.9	1.65	3.2	1.17	9.8	1.72	13.1	14.0	90.7	94.1	14.0	90.7	94.1	14.0	90.7
302	ILL	LAB	5	MIX-3B		0.53	2000	14.4	BM150	5.8	8.9	1.65	3.2	1.18	11.3	1.84	29.1	31.1	79.2	86.5	31.1	79.2	86.5	31.1	79.2
303	ILL	LAB	5	MIX-3C		0.53	2000	14.4	BM150	5.8	8.9	1.65	3.6	1.24	19.7	2.48	66.8	71.2	49.8	64.8	71.2	49.8	64.8	71.2	49.8
304	ILL	LAB	5	MIX-3D		0.53	2000	14.4	BM150	5.8	8.9	1.65	4.7	1.31	69.6	6.46	93.5	98.2	25.6	42.7	98.2	25.6	42.7	98.2	25.6
305	ILL	LAB	6	RAW15-6A	100	0.85	2000	14.4	BM150	5.8	29.5	1.62	4.5	0.95	35.6	1.78	19.6	27.5	88.5	96.0	27.5	88.5	96.0	27.5	88.5
306	ILL	LAB	6	RAW15-6B	100	0.85	2000	14.4	BM150	5.8	29.5	1.61	4.5	0.96	46.1	2.05	39.9	56.0	76.3	91.5	56.0	76.3	91.5	56.0	76.3
307	ILL	LAB	6	RAW15-6C	100	0.86	2000	14.4	BM150	5.8	29.5	1.61	6.0	1.29	86.1	2.39	70.7	97.3	43.5	62.4	97.3	43.5	62.4	97.3	43.5
308	ILL	LAB	7	RAW15-7A	100	0.54	2000	14.2	BM150	5.8	29.2	1.67	4.6	0.88	33.4	1.80	14.5	20.2	92.3	98.0	20.2	92.3	98.0	20.2	92.3
309	ILL	LAB	7	RAW15-7B	100	0.54	2000	14.2	BM150	5.8	29.2	1.66	4.7	0.88	40.1	2.01	30.7	42.8	83.7	95.8	42.8	83.7	95.8	42.8	83.7
310	ILL	LAB	7	RAW15-7C	100	0.54	2000	14.2	BM150	5.8	29.2	1.66	5.0	1.08	65.9	2.55	60.2	83.6	60.9	81.0	83.6	60.9	81.0	83.6	60.9

Batch Test Results Summary

BATCH TEST CONDITIONS AND RESULTS SUMMARY (Continued)

Test No	Coal	Source	Batch	Sample ID	Size mesh	Diesel %	HS RPM	HS Sol %	Frother	Flot Sol%	Feed			Final Product			Tailings			Agglomeration Performance						ROM Performance		
											Ash	TS	%	Ash	TS	%	Ash	TS	%	Yield	Blu Rec	TSR	PSR	Blu Rec	TSR	PSR	Blu Rec	TSR
311	ILL	LAB	7	RAWI5-7D	100	0.54	2000	14.2	BM150	5.8	29.2	1.66	5.5	1.17	79.4	2.70	67.9	93.6	52.2	72.9	93.6	52.2	72.9	93.6	52.2	72.9		
312	ILL	LAB	8	RAWI5-8A	100	0.27	2000	14.3	BM150	5.8	29.7	1.68	4.3	0.87	31.4	1.73	6.3	8.9	96.7	99.2	8.9	96.7	99.2	8.9	96.7	99.2		
313	ILL	LAB	8	RAWI5-8B	100	0.27	2000	14.2	BM150	5.8	29.7	1.67	5.3	0.90	34.3	1.82	15.9	22.1	91.5	97.5	22.1	91.5	97.5	22.1	91.5	97.5		
314	ILL	LAB	8	RAWI5-8C	100	0.27	2000	14.2	BM150	5.8	29.7	1.68	5.3	0.96	42.2	2.04	33.8	47.1	80.6	93.0	47.1	80.6	93.0	47.1	80.6	93.0		
315	ILL	LAB	8	RAWI5-8D	100	0.27	2000	14.2	BM150	5.8	29.7	1.68	5.3	1.07	57.7	2.37	53.4	74.4	65.9	83.7	74.4	65.9	83.7	74.4	65.9	83.7		
316	ILL	LAB	8	RAWI5-8E	100	0.27	2000	14.2	BM150	5.8	29.7	1.67	6.0	1.15	71.5	2.60	63.8	88.2	56.2	75.8	88.2	56.2	75.8	88.2	56.2	75.8		
317	ILL	LAB	9	RAWI5-1A	200	0.91	2000	13.5	BM150	5.4	17.1	1.78	3.6	1.05	21.4	2.01	23.9	28.3	85.9	93.5	28.3	85.9	93.5	28.3	85.9	93.5		
318	ILL	LAB	9	RAWI5-1B	200	0.91	2000	13.5	BM150	5.4	17.1	1.78	3.6	1.07	26.1	2.24	39.7	46.9	76.1	88.4	46.9	76.1	88.4	46.9	76.1	88.4		
319	ILL	LAB	9	RAWI5-1C	200	0.91	2000	13.5	BM150	5.4	17.1	1.78	3.6	1.11	38.6	2.85	61.4	72.5	61.8	80.0	72.5	61.8	80.0	72.5	61.8	80.0		
320	ILL	LAB	9	RAWI5-1D	200	0.91	2000	13.5	BM150	5.4	17.1	1.78	3.8	1.13	50.3	3.39	71.3	84.1	54.7	75.4	84.1	54.7	75.4	84.1	54.7	75.4		
321	ILL	LAB	9	RAWI5-1E	200	0.91	2000	13.5	BM150	5.4	17.1	1.78	4.0	1.17	57.3	3.65	75.4	88.6	50.5	71.3	88.6	50.5	71.3	88.6	50.5	71.3		
322	ILL	LAB	10	RAWI5-2A	100	0.57	2000	13.5	BM150	5.5	17.4	2.02	3.6	1.09	18.9	2.12	9.7	11.5	94.8	97.5	11.5	94.8	97.5	11.5	94.8	97.5		
323	ILL	LAB	10	RAWI5-2B	100	0.56	2000	13.6	BM150	5.5	17.4	2.03	3.5	1.08	20.9	2.26	19.9	23.6	89.4	95.1	23.6	89.4	95.1	23.6	89.4	95.1		
324	ILL	LAB	10	RAWI5-2C	100	0.56	2000	13.6	BM150	5.5	17.4	2.02	3.4	1.11	26.8	2.64	40.3	47.8	77.9	89.2	47.8	77.9	89.2	47.8	77.9	89.2		
325	ILL	LAB	10	RAWI5-2D	100	0.58	2000	13.6	BM150	5.5	17.4	2.03	3.5	1.14	30.2	2.84	47.9	56.9	73.0	86.1	56.9	73.0	86.1	56.9	73.0	86.1		
326	ILL	LAB	11	RAWI5-3A	200	1.34	2000	13.7	BM150	5.5	17.9	1.84	3.8	1.15	28.2	2.34	42.3	50.3	73.5	85.5	50.3	73.5	85.5	50.3	73.5	85.5		
327	ILL	LAB	11	RAWI5-3B	200	1.34	2000	13.7	BM150	5.5	17.9	1.84	3.9	1.19	44.0	3.04	65.1	77.4	57.8	75.4	77.4	57.8	75.4	77.4	57.8	75.4		
328	ILL	LAB	11	RAWI5-3C	200	1.34	2000	13.7	BM150	5.5	17.9	1.84	4.5	1.27	76.4	4.32	81.4	96.1	43.7	63.7	96.1	43.7	63.7	96.1	43.7	63.7		
329	ILL	LAB	12	RAWI5-4A	200	0.89	2000	13.8	BM150	5.6	18.1	1.66	4.1	0.85	21.3	1.84	18.5	22.1	90.5	97.9	22.1	90.5	97.9	22.1	90.5	97.9		
330	ILL	LAB	12	RAWI5-4B	200	0.89	2000	13.8	BM150	5.6	18.1	1.65	3.9	0.90	27.3	2.14	39.4	46.9	78.6	93.6	46.9	78.6	93.6	46.9	78.6	93.6		
331	ILL	LAB	12	RAWI5-4C	200	0.89	2000	13.8	BM150	5.6	18.1	1.66	4.1	0.93	48.3	3.23	68.4	81.3	61.6	86.9	81.3	61.6	86.9	81.3	61.6	86.9		
332	ILL	LAB	12	RAWI5-4D	200	0.89	2000	13.8	BM150	5.6	18.1	1.65	4.7	0.95	68.2	4.27	78.9	93.2	54.6	83.0	93.2	54.6	83.0	93.2	54.6	83.0		
333	ILL	LAB	13	RAWI5-5A	100	0.88	2000	13.9	BM150	5.6	18.3	1.87	3.9	0.97	25.2	2.30	32.4	38.7	83.2	93.9	38.7	83.2	93.9	38.7	83.2	93.9		
334	ILL	LAB	13	RAWI5-5B	100	0.88	2000	13.9	BM150	5.6	18.3	1.87	5.0	1.23	65.8	4.14	78.1	92.2	48.5	68.3	92.2	48.5	68.3	92.2	48.5	68.3		
335	ILL	LAB	13	RAWI5-5C	100	0.88	2000	13.9	BM150	5.6	18.3	1.87	5.8	1.35	79.2	4.40	82.9	97.1	40.1	58.1	97.1	40.1	58.1	97.1	40.1	58.1		
336	I5P	LAB	1	PCI5-1A	100	0.88	2000	13.9	BM150	5.6	7.8	1.45	3.0	1.09	9.1	1.54	20.9	22.1	84.2	90.1	21.8	84.2	90.1	21.8	84.2	90.1		
337	I5P	LAB	1	PCI5-1B	100	0.88	2000	13.9	BM150	5.6	7.8	1.44	2.9	1.09	10.4	1.63	34.6	36.6	73.8	83.7	36.1	73.8	83.7	36.1	73.8	83.7		
338	I5P	LAB	1	PCI5-1C	100	0.88	2000	13.9	BM150	5.6	7.8	1.44	3.1	1.13	13.3	1.81	53.9	56.9	57.8	71.6	56.1	57.8	71.6	56.1	57.8	71.6		
339	I5P	LAB	1	PCI5-1D	100	0.88	2000	13.9	BM150	5.6	7.8	1.45	3.1	1.16	20.8	2.25	73.6	77.7	41.1	58.5	76.5	41.1	58.5	76.5	41.1	58.5		
340	I5P	LAB	1	PCI5-1E	100	0.88	2000	13.9	BM150	5.6	7.8	1.45	3.4	1.19	29.9	2.75	83.5	87.8	31.4	49.2	86.5	31.4	49.2	86.5	31.4	49.2		
341	I5P	LAB	2	PCI5-2A	200	0.54	2000	14.3	BM150	5.8	10.5	1.47	4.6	1.00	11.7	1.57	17.6	18.9	88.0	93.9	18.6	88.0	93.9	18.6	88.0	93.9		
342	I5P	LAB	2	PCI5-2B	200	0.54	2000	14.3	BM150	5.8	10.5	1.47	4.6	1.00	12.8	1.66	28.6	30.7	80.5	90.1	30.3	80.5	90.1	30.3	80.5	90.1		

BATCH TEST CONDITIONS AND RESULTS SUMMARY (Continued)

Test No	Coal	Source	Batch	Sample ID	Size mesh	Diesel %	HS RPM	HS Sol %	Frather	Flot. Sol%	Feed		Final Product		Tailings		Agglomeration Performance				ROM Performance			
											Ash	TS	Ash	TS	Ash	TS	Yield	Blu Rec	TSR	PSR	Blu Rec	TSR	PSR	TSR
343	I5P	LAB	2	PCI5-2C	200	0.54	2000	14.3	BM150	5.8	10.5	1.47	4.4	1.04	16.6	1.91	50.4	54.1	64.4	80.2	53.3	64.4	88.4	
344	I5P	LAB	2	PCI5-2D	200	0.54	2000	14.3	BM150	5.8	10.5	1.47	4.4	1.07	25.2	2.43	70.8	76.1	48.3	69.3	75.0	48.3	82.0	
345	I5P	LAB	2	PCI5-2E	200	0.54	2000	14.3	BM150	5.8	10.5	1.47	4.6	1.10	38.4	3.20	82.6	88.6	38.0	60.8	87.2	38.0	77.1	
346	I5P	LAB	3	PCI5-3A	200	1.31	2000	14.1	BM150	5.7	8.3	1.53	3.6	1.14	12.3	1.85	45.7	48.3	65.9	77.7	47.5	65.9	86.9	
347	I5P	LAB	3	PCI5-3B	200	1.31	2000	14.1	BM150	5.7	8.3	1.52	3.6	1.17	17.9	2.24	67.2	71.0	48.3	64.4	69.9	48.3	79.2	
348	I5P	LAB	3	PCI5-3C	200	1.31	2000	14.1	BM150	5.7	8.3	1.52	3.9	1.23	40.8	3.69	88.0	92.7	29.0	47.0	91.3	29.0	69.0	
349	I5P	LAB	3	PCI5-3D	200	1.31	2000	14.1	BM150	5.7	8.3	1.52	4.1	1.25	57.8	4.68	92.2	96.8	24.2	41.7	95.4	24.2	65.9	
350	I5P	LAB	4	PCI5-4A	200	0.85	2000	14.3	BM150	5.8	9.3	1.66	3.6	1.03	10.4	1.77	15.5	16.6	90.3	95.3	16.3	90.3	97.2	
351	I5P	LAB	4	PCI5-4B	200	0.85	2000	14.4	BM150	5.8	9.3	1.65	3.5	1.06	11.8	1.90	29.3	31.4	81.2	90.1	30.9	81.2	94.2	
352	I5P	LAB	4	PCI5-4C	200	0.85	2000	14.4	BM150	5.8	9.3	1.66	3.6	1.14	18.7	2.50	61.9	66.2	57.4	74.1	65.2	57.4	84.9	
353	I5P	LAB	4	PCI5-4D	200	0.85	2000	14.4	BM150	5.8	9.3	1.65	3.9	1.23	39.8	4.00	84.7	90.4	36.9	56.3	89.0	36.9	74.4	
354	I5P	LAB	5	PCI5-5A	100	0.84	2000	14.5	BM150	5.9	8.7	1.52	3.8	1.22	12.4	1.76	43.6	46.2	65.1	74.4	45.5	65.1	85.0	
355	I5P	LAB	5	PCI5-5B	100	0.84	2000	14.5	BM150	5.9	8.7	1.53	4.6	1.31	55.2	4.04	92.0	96.5	21.2	35.7	95.0	21.2	62.4	
356	I5P	LAB	5	PCI5-5C	100	0.84	2000	14.5	BM150	5.9	8.7	1.53	4.8	1.34	76.3	4.82	94.6	99.0	17.1	30.2	97.5	17.1	59.2	
357	I5P	LAB	5	PCI5-5D	100	0.84	2000	14.5	BM150	5.9	8.7	1.53	4.8	1.34	76.3	4.82	94.6	99.0	17.1	30.2	97.5	17.1	59.2	
360	PEN	LAB	1	PEN3-1A	200	0.50	2000	14.6	BM150	5.9	20.4	1.08	5.6	1.13	67.4	0.93	76.0		20.6				20.6	
361	PEN	LAB	1	PEN3-1B	200	0.50	2000	14.6	BM150	5.9	20.4	1.09	5.7	1.14	83.6	0.86	81.1		15.0				15.0	
362	PEN	LAB	2	PEN3-2A	200	0.25	2000	14.0	BM150	5.7	20.5	1.07	5.3	1.09	41.4	1.04	57.9		41.0				41.0	
363	PEN	LAB	2	PEN3-2B	200	0.25	2000	14.0	BM150	5.7	20.5	1.07	5.3	1.10	63.0	0.99	73.6		24.4				24.4	
364	PEN	LAB	2	PEN3-2C	200	0.25	2000	14.0	BM150	5.7	20.5	1.07	5.5	1.11	77.8	0.92	79.2		17.9				17.9	
365	PEN	LAB	2	PEN3-2D	200	0.25	2000	14.0	BM150	5.7	20.5	1.07	5.8	1.12	86.7	0.84	81.8		14.3				14.3	
366	PEN	LAB	3	PEN3-3A	200	0.13	2000	14.8	BM150	6.0	20.6	1.10	4.9	1.08	32.6	1.12	43.4		57.5				57.5	
367	PEN	LAB	3	PEN3-3B	200	0.13	2000	14.8	BM150	6.0	20.6	1.10	5.0	1.07	51.7	1.16	66.7		35.1				35.1	
368	PEN	LAB	3	PEN3-3C	200	0.13	2000	14.8	BM150	6.0	20.6	1.10	5.2	1.08	72.1	1.16	77.1		24.2				24.2	
369	PEN	LAB	3	PEN3-3D	200	0.13	2000	14.8	BM150	6.0	20.6	1.10	5.6	1.10	82.9	1.12	80.6		19.6				19.6	
370	PEN	LAB	4	PEN3-4A	200	0.00	0	350.0	BM150	6.1	22.3	1.08	4.6	1.02	26.2	1.09	17.9		83.1				83.1	
371	PEN	LAB	4	PEN3-4B	200	0.00	0	350.0	BM150	6.1	22.3	1.08	4.5	1.03	31.9	1.10	34.9		66.6				66.6	
372	PEN	LAB	4	PEN3-4C	200	0.00	0	350.0	BM150	6.1	22.3	1.08	4.6	1.05	47.7	1.12	58.9		42.7				42.7	
373	PEN	LAB	4	PEN3-4D	200	0.00	0	350.0	BM150	6.1	22.3	1.07	5.1	1.07	65.3	1.08	71.4		28.8				28.8	
374	PEN	LAB	5	PEN3-5A	200	0.99	2000	16.3	BM150	6.6	20.6	1.07	5.7	1.11	59.6	0.96	72.2		25.0				25.0	
375	PEN	LAB	5	PEN3-5B	200	0.99	2000	16.3	BM150	6.6	20.6	1.07	5.7	1.12	82.2	0.87	80.5		15.9				15.9	
376	PEN	LAB	5	PEN3-5C	200	0.99	2000	16.3	BM150	6.6	20.6	1.07	5.8	1.12	85.8	0.84	81.4		14.6				14.6	

Batch Test Results Summary

BATCH TEST CONDITIONS AND RESULTS SUMMARY (Continued)

Test	No	Coal	Source	Batch	Sample ID	Size mesh	Diesel %	HS RPM	HS Sol %	Frother	Flot. Sol%	Feed			Final Product			Tailings			Agglomeration Performance			ROM Performance		
												Ash	TS	%	Ash	TS	%	Ash	TS	%	Btu Rec	TSR	PSR	Btu Rec	TSR	PSR
377	EMR	LAB	1	EMR-1A	400	0.52	2000	14.7	BM150	6.0	30.8	2.45	5.2	1.65	45.7	2.92	36.8	75.3	75.3	75.3	75.3	75.3	75.3	75.3	75.3	75.3
378	EMR	LAB	1	EMR-1B	400	0.52	2000	14.7	BM150	6.0	30.8	2.45	5.2	1.63	58.3	3.33	51.8	65.6	65.6	65.6	65.6	65.6	65.6	65.6	65.6	65.6
379	EMR	LAB	1	EMR-1C	400	0.52	2000	14.7	BM150	6.0	30.8	2.45	5.3	1.66	82.7	4.06	67.1	54.5	54.5	54.5	54.5	54.5	54.5	54.5	54.5	54.5
380	EMR	LAB	1	EMR-1D	400	0.52	2000	14.7	BM150	6.0	30.8	2.45	5.6	1.71	88.6	4.14	69.7	51.3	51.3	51.3	51.3	51.3	51.3	51.3	51.3	51.3
381	EMR	LAB	2	EMR-2A	28	0.52	2000	15.0	BM150	6.1	30.5	2.78	7.8	2.38	72.1	3.51	64.7	44.6	44.6	44.6	44.6	44.6	44.6	44.6	44.6	44.6
382	EMR	LAB	2	EMR-2B	28	0.52	2000	15.0	BM150	6.1	30.5	2.78	8.0	2.42	85.3	3.66	70.9	38.3	38.3	38.3	38.3	38.3	38.3	38.3	38.3	38.3
383	EMR	LAB	2	EMR-2C	28	0.52	2000	15.0	BM150	6.1	30.5	2.78	8.8	2.50	88.4	3.52	72.8	34.5	34.5	34.5	34.5	34.5	34.5	34.5	34.5	34.5
384	EMR	LAB	3	EMR-3A	200	0.52	2000	14.9	BM150	6.1	30.7	2.46	5.5	1.84	61.7	3.21	55.1	58.7	58.7	58.7	58.7	58.7	58.7	58.7	58.7	58.7
385	EMR	LAB	3	EMR-3B	200	0.52	2000	14.9	BM150	6.1	30.7	2.45	5.6	1.86	78.7	3.59	65.7	50.2	50.2	50.2	50.2	50.2	50.2	50.2	50.2	50.2
386	EMR	LAB	3	EMR-3C	200	0.52	2000	14.9	BM150	6.1	30.7	2.45	6.1	1.95	88.1	3.63	70.0	44.4	44.4	44.4	44.4	44.4	44.4	44.4	44.4	44.4
387	EMR	LAB	4	EMR-4A	28	0.26	2000	15.2	BM150	6.2	30.5	2.64	8.1	2.65	72.7	2.63	65.3	34.8	34.8	34.8	34.8	34.8	34.8	34.8	34.8	34.8
388	EMR	LAB	4	EMR-4B	28	0.26	2000	15.2	BM150	6.2	30.5	2.65	8.3	2.66	83.0	2.61	70.3	29.3	29.3	29.3	29.3	29.3	29.3	29.3	29.3	29.3
389	EMR	LAB	4	EMR-4C	28	0.26	2000	15.2	BM150	6.2	30.5	2.64	8.8	2.72	86.9	2.44	72.2	25.7	25.7	25.7	25.7	25.7	25.7	25.7	25.7	25.7
390	EMR	LAB	5	EMR-5A	28	0.10	2000	14.8	BM150	6.0	30.8	2.30	7.7	2.16	68.9	2.53	62.2	41.6	41.6	41.6	41.6	41.6	41.6	41.6	41.6	41.6
391	EMR	LAB	5	EMR-5B	28	0.10	2000	14.8	BM150	6.0	30.8	2.30	7.8	2.18	81.2	2.55	68.7	34.8	34.8	34.8	34.8	34.8	34.8	34.8	34.8	34.8
392	EMR	LAB	5	EMR-5C	28	0.10	2000	14.8	BM150	6.0	30.8	2.29	8.0	2.20	85.1	2.52	70.5	32.4	32.4	32.4	32.4	32.4	32.4	32.4	32.4	32.4
393	EMR	LAB	7	EMR-7A	200	0.52	2000	15.0	BM150	6.1	30.7	2.57	6.9	2.25	74.6	3.16	64.8	43.2	43.2	43.2	43.2	43.2	43.2	43.2	43.2	43.2
394	EMR	LAB	7	EMR-7B	200	0.52	2000	15.0	BM150	6.1	30.7	2.57	7.0	2.26	86.5	3.30	70.2	38.2	38.2	38.2	38.2	38.2	38.2	38.2	38.2	38.2
395	EMR	LAB	7	EMR-7C	200	0.52	2000	15.0	BM150	6.1	30.7	2.57	7.7	2.42	88.7	2.95	71.6	32.6	32.6	32.6	32.6	32.6	32.6	32.6	32.6	32.6
396	EMM	LAB	1	EMM-1A	200	0.50	2000	15.0	BM150	6.1	26.1	4.13	14.0	3.12	43.1	5.56	58.7	55.7	55.7	55.7	55.7	55.7	55.7	55.7	55.7	55.7
397	EMM	LAB	1	EMM-1B	200	0.50	2000	15.0	BM150	6.1	26.1	4.13	15.1	3.27	62.1	6.96	76.6	39.3	39.3	39.3	39.3	39.3	39.3	39.3	39.3	39.3
398	EMM	LAB	1	EMM-1C	200	0.50	2000	15.0	BM150	6.1	26.1	4.13	16.8	3.33	79.6	8.78	85.3	31.3	31.3	31.3	31.3	31.3	31.3	31.3	31.3	31.3
399	EMM	LAB	2	EMM-2A	28	0.51	2000	15.5	BM150	6.3	26.4	4.38	20.0	4.51	40.5	4.09	68.6	29.3	29.3	29.3	29.3	29.3	29.3	29.3	29.3	29.3
400	EMM	LAB	2	EMM-2B	28	0.51	2000	15.5	BM150	6.3	26.4	4.41	20.2	4.48	52.6	4.14	80.7	18.1	18.1	18.1	18.1	18.1	18.1	18.1	18.1	18.1
401	EMM	LAB	2	EMM-2C	28	0.51	2000	15.5	BM150	6.3	26.4	4.38	22.2	4.40	70.7	4.14	91.2	8.3	8.3	8.3	8.3	8.3	8.3	8.3	8.3	8.3
402	EMM	LAB	3	EMM-3A	100	0.51	2000	15.2	BM150	6.2	26.7	4.26	18.0	4.03	48.7	4.85	71.7	32.2	32.2	32.2	32.2	32.2	32.2	32.2	32.2	32.2
403	EMM	LAB	3	EMM-3B	100	0.51	2000	15.2	BM150	6.2	26.7	4.26	18.5	4.02	60.9	5.28	80.8	23.8	23.8	23.8	23.8	23.8	23.8	23.8	23.8	23.8
404	EMM	LAB	3	EMM-3C	100	0.51	2000	15.2	BM150	6.2	26.7	4.26	20.5	3.98	79.9	6.65	89.6	16.3	16.3	16.3	16.3	16.3	16.3	16.3	16.3	16.3
405	EMM	LAB	4	EMM-4A	28	0.25	2000	15.3	BM150	6.2	26.9	4.58	19.0	4.65	43.1	4.44	67.0	31.9	31.9	31.9	31.9	31.9	31.9	31.9	31.9	31.9
406	EMM	LAB	4	EMM-4B	28	0.25	2000	15.3	BM150	6.2	26.9	4.58	19.7	4.54	53.6	4.73	78.7	22.0	22.0	22.0	22.0	22.0	22.0	22.0	22.0	22.0
407	EMM	LAB	4	EMM-4C	28	0.25	2000	15.3	BM150	6.2	26.9	4.58	22.9	4.62	74.2	4.15	92.2	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1
408	EMM	LAB	5	EMM-5A	28	0.10	2000	15.7	BM150	6.4	26.1	4.57	18.9	4.72	40.6	4.26	66.6	31.2	31.2	31.2	31.2	31.2	31.2	31.2	31.2	31.2
409	EMM	LAB	5	EMM-5B	28	0.10	2000	15.7	BM150	6.4	26.1	4.57	19.4	4.62	50.5	4.37	78.2	20.8	20.8	20.8	20.8	20.8	20.8	20.8	20.8	20.8
410	EMM	LAB	5	EMM-5C	28	0.10	2000	15.7	BM150	6.4	26.1	4.57	21.4	4.54	64.5	4.77	89.0	11.4	11.4	11.4	11.4	11.4	11.4	11.4	11.4	11.4

Nomenclature:

HS RPM: High-shear intensity (RPM)
HS Sol %: High-shear percent solids
Flot Sol %: Percent solids in flotation
Ash %: Weight percent ash
TS %: Weight percent total sulfur
Yield %: Weight percent clean coal yield
Btu Rec %: Percent Btu Recovery
TSR %: Percent total sulfur rejection
PSR %: Percent pyritic sulfur rejection
ROM: Run-of-mine

Coal types:

EMM: Emerald Mine Middlings Coal (Pittsburgh No. 8)
EMR: Emerald Mine Raw Coal (Pittsburgh No. 8)
ILL: Illinois No. 5
ISP: Illinois No. 5 Precleaned
PEN: Penelec Sample
PIT: Pittsburgh No. 8
UF1: Upper Freeport

Appendix G

EQUIPMENT LISTS

200-t/h Commercial Agglomeration Plant

Table G-1
Coal Receiving, Storage, and Crushing Circuit Equipment List

Equipment Item	Capacity	Size	No.	Cost FOB	Power hp	Description
Truck Scale & Weigh Station	1000 t/h	20 t truck	1	\$100,000		Coal delivery receipts
Truck Dump Grate	1000 t/h	20 t truck	4	Civil		Loads the belt conveyor
Underground Feed Hopper	200 t	100 t	2	Civil		Temporary storage of coal to feed vibrating feeders
Vibrating Feeder	1000 t/h	100-250 t/h	4	\$15,000	5	Regulates coal onto the feed coal belt conveyor
Feed Coal Belt Conveyor	1000 t/h	48"W x 150'	1	\$150,000	25	Transfers coal to coal stockpile
Feed Coal Stacking Tube	350 t/h	15,000 t	1	Civil		Distributes coal to stockpiles complete with reclaim tunnel
Vibrating Feeder	350 t/h	25-100 t/h	4	\$6,000	5	Regulates coal onto the crusher feed coal belt conveyor
Crusher Feed Belt Conveyor	350 t/h	36"W x 100'	1	\$75,000	20	Transfers 2" x 0 feed coal from the stockpiles to the crusher
Tramp Iron Magnet	350 t/h	48"W belt	1	\$15,000		Protects crusher from tramp iron
Cage Packtor Coal Crusher	350 t/h	350 t/h	1	\$170,500	700	Reduces feed coal from a topsize of 2" to ¼", Steadman G-62 2-Row, 52000 lb
Hopper Belt Conveyor	350 t/h	36"W x 50'	1	\$38,000	15	Transfers ¼" x 0 coal either to grinding mill storage hopper or to the preclean deslime screen storage hopper
Surge Hopper	175 t	175 t	1	\$56,000		Provides 1/2-hour crushed coal storage

Table G-2
Coal Precleaning Circuit Equipment List

Equipment Item	Capacity	Size	No. (spare)	Cost FOB	Power hp	Description
Weigh Belt Feeder	350 t/h	350 t/h	1	\$40,000	10	Weigh and regulate raw coal
Deslime Screen	350 t/h	8' x 16'	3	\$45,000	10	Size separation at 28 mesh
Heavy Media Cyclone Feed Sump	12,000 gal	12,000 gal	1	Civil		Provides temporary storage for +28 mesh raw coal slurry
Heavy Media Cyclone Feed Pump	3,600 gpm	@40' head	1(1)	\$29,000	175	Transfers +28 mesh slurry to heavy media cyclones
Heavy Media Cyclone	250 t/h	24"φ	4	\$25,000		Specific gravity separation of raw coal to clean coal & refuse
Coal Sieve Bend & Drain & Rinse Screen	170 t/h	8' x 16'	2	\$45,000	10	Dewaters heavy media cyclones overflow stream (clean coal)
Refuse Sieve Bend & Drain & Rinse Screen	90 t/h	8' x 16'	1	\$45,000	10	Dewaters heavy media cyclones underflow stream (refuse)
Dilute Media Sump	3,000 gal	3,000 gal	1	Civil		Collects clean coal and refuse D&R screen effluents
Dilute Media Pump	1,250 gpm	@40' head	1(1)	\$13,000	40	Transfers D&R screen effluents to magnetic separator
Magnetite Separator	400 gal	30"φ x 6'L	1	\$38,000	5	Recovers magnetite from slurry
Magnetite Thickener	1,200 gal	25"φ	1	\$100,000	5	Thickens magnetite slurry
Classifying Cyclone Feed Sump	6,000 gal	6,000 gal	1	Civil		Provides temporary storage for -28 mesh raw coal slurry
Classifying Cyclone Feed Pump	3,500 gpm	@40' head	1(1)	\$19,000	150	Transfers -28 mesh coal slurry to classifying cyclones
Classifying Cyclone	100 t/h	15"φ	8	\$10,000		Splits -28 mesh slurry stream into 28 x 100 mesh underflow and -100 mesh overflow
Spiral Concentrators	70 t/h	3-turn	24	\$4,000		Specific gravity separation of raw coal to clean coal & refuse
Clean Coal Screen	40 t/h	5' x 8'	3	\$30,000	5	Dewaters spiral coal stream
Refuse Screen	30 t/h	5' x 8'	2	\$30,000	5	Dewaters spiral refuse stream
Static Thickener	3,500 gpm 26 t/h	95"φ	1	\$500,000	45	Produces clarified water from -100 mesh slurry
Thickener Overflow Pump	3,500 gpm	@40' head	1(1)	\$28,000	150	Transfers thickener overflow, clarified water to the deslime screen
Thickener Underflow Pump	225 gpm	@40' head	1(1)	\$15,000	15	Transfers thickener underflow, -100 mesh raw coal slurry to surge tank
Surge Tank	15,000 gal	15,000 gal	1	\$25,000		Storage of -100 mesh raw coal slurry
Surge Tank Pump	500 gpm	@80' head	1(1)	\$12,000	15	Transfers -100 mesh slurry to rod mill
Clean Coal Conveyor	220 t/h	30"W x 200'L	1	\$175,000	50	Transfers coal to mill hopper
Refuse Conveyor	110 t/h	30"W x 70'L	1	\$60,000	30	Transfers refuse for disposal

Table G-3
Coal Grinding Circuit Equipment List

Equipment Item	Capacity	Size	No. (spare)	Cost FOB	Power hp	Description
Rod Mill Feed Hopper	175 t	6,363 ft ³	1	Civil		Temporary storage of crushed coal for feed to rod mill feeder
Rod Mill Belt Weigh Feeder	250 t/h	100-250 t/h	1	\$40,000	10	Regulates feed into the rod mill
Rod Mill	230 t/h	12.5'φ x 18'L	1	\$800,000	1,150	Reduces 1/2" x 0 feed coal to minus 28 mesh, 50% solids coal slurry. Complete with rod charges, motor, and lube system
Rod Mill Discharge Sump	3,000 gal	3,000 gal	1	Civil		Gathers the rod mill slurry discharge, the 28M x 0, 50% solids coal slurry for feed to the ball mill
Rod Mill Discharge Pump	1,600 gpm	@ 40' head	1(1)	\$15,000	30	Transfers 28M x 0, 50% solids coal slurry to holding tank
Ball Mill	230 t/h	15'φ x 16'	2	\$1,400,000	2,300	Reduces 28M x 0, 50% coal slurry to 95% passing 200M coal slurry, complete with initial ball charge, motor and lube system
Holding Tank Feed Sump	3,000 gal	3,000 gal	1	Civil		Gathers ball mill slurry discharge, 200M x 0, 50% solids coal slurry for feed to holding tank
Holding Tank Feed Pump	1,600 gpm	@ 30' head	1(1)	\$18,000	25	Transfers 200M x 0, 50% coal slurry to slurry holding tank
Slurry Holding Vessel	16,000 gal	16,000 gal	1	\$40,000	20	Stores 200M x 0, 50% solids coal slurry providing 10-min capacity and homogenous output to high-shear vessel, includes the mixer

Table G-4
Coal Agglomeration Circuit Equipment List

Equipment Item	Capacity	Size	No. (spare)	Cost FOB	Power hp	Description
High-Shear Feed Sump	13,000 gal	13,000 gal	1	Civil		Gathers discharge from slurry holding vessel for feed to high shear
High-Shear Feed Pump	4,300 gpm	@60' head	1(1)	\$14,100	65	Transfers slurry from holding tank to high shear
High-Shear Distributor	4,300 gpm	6 way	1	Civil		Distributes slurry to the high-shear vessels
High-Shear Vessels & Mixer	4,300 gal	750 gal	6	\$100,000	125	Provides high agitation for a 1-min retention time to enable formation of the flocs/microagglomerates
Flotation Bank 1	15,000 ft ³	1,500 ft ³ /cell	10	\$70,000	100	Enables recovery of microagglomerates, approx. 6-min residence time, at 6% solids
Flotation Bank 2 Feed Sump	22,000 gal	11,000 gal	2	Civil		Gathers discharge froth from first-stage flotation for feed to second-stage flotation
Flotation Bank 2 Feed Pump	13,700 gpm	@ 30' head	2(1)	\$13,500	75	Transfers slurry from first- to second-stage flotation
Flotation Bank 2	12,000 ft ³	1,500 ft ³ /cell	8	\$70,000	100	Enables separation of microagglomerates, approx. 6-min residence time, at 6% solids
Conditioning Tank	3,800 gal	3,800 gal	2	\$25,000	30	Redisperses froth before addition to vacuum filter, about 1-min residence time
Vacuum Filter Feed Pump	3,800 gpm	@ 40' head	2(1)	\$12,000	75	Transfers dispersed froth from conditioning tank to vacuum filter
Rotary Vacuum Filter	200 t/h	12'6"φ 9 disc	2	\$400,000	450	Reduction of froth moisture

Table G-5
Size Enlargement, Drying and Product Loadout Circuit Equipment List

Equipment Item	Capacity	Size	No.	Cost FOB	Power hp	Description
Solid Mixers	200 t/h	3.5' ϕ x 12'L	4	\$115,000	100	Dispersion of binder within the cake
Pelletizer	200 t/h	20' ϕ	4	\$220,000	125	Pelletization of coal, about 6 mm to 8 mm
Dryer Belt Conveyor	200 t/h	30"W x 50'L	4	\$25,000	25	Transfers wet pellets to dryer feed storage hopper
Storage Hopper	200 t/h	50 t	4	Civil		Provides 1-hour storage of wet pellets for feed to dryer
Vibrating Bed Dryer	200 t/h	10' x 36'	4	\$500,000	120	Pellet drying to harden pellets and reduce moisture, includes heating system
Cooling Conveyor	200 t/h	48"W x 60'L	4	\$60,000	25	Cools dried coal pellets
Product Belt Conveyor	200 t/h	30"W x 100'L	1	\$175,000	50	Transfers dried coal pellets to silos
Storage Silos	15,000 t	15,000 t	1	\$1,500,000	10	Provide 3-day storage for product

Table G-6
Clarified Water and Refuse Disposal Circuit Equipment List

Equipment Item	Capacity	Size	No. (each)	Cost FOB	Power hp	Description
High Capacity Thickener	21,000 gpm	90' ϕ	1	\$550,000	25	Produces clarified water from tails
Refuse Slurry Pump	400 gpm	@ 20' head	1 (1)	\$8,500	5	Transfers clarifier underflow to belt filter
Belt Filter	25 t/h	3-meter W	2	\$270,000	25	Removal of water from refuse slurry
Refuse Belt Conveyor	25 t/h	24" x 70'	1	\$25,000	20	Transfers refuse to refuse bins (landfill)
Carbon Bed Feed Pump	420 gpm	@ 40' head	1	\$8,000	15	Transfers part of clarified water to activated carbon bed
Activated Carbon Bed	420 gpm		2	\$15,000		Water treatment to reduce dissolved solids and other impurities
Thickener Overflow Tank	60,000 gal	60,000 gal	1	\$80,000		Collect thickener overflow
Head Tank Pump	21,000 gpm	@ 70' head	1	\$43,000	350	Transfers process water from process water tank to head tank
Head Tank			1	\$3,000		Head tank to provide a constant flow of process water

Table G-7
Reagent Storage and Delivery Circuit Equipment List

Equipment Item	Capacity	Size	No. (spare)	Cost FOB	Power hp	Description
Diesel Transfer Pump	1,000 gpm	@ 30' head	1	\$2,000	20	Transfers diesel from delivery system to storage tank
Diesel Storage Tank	80,000 gal	26' x 26'H	1	\$34,000		Storage for 10-day supply of diesel
Diesel Feed Pump	500 gpm	@ 25' head	1(1)	\$3,000	10	Transfers diesel from storage tank to diesel feed tank
Diesel Feed Tank	4,000 gal	9' x 9'H	1	\$16,000		Storage for process feed diesel, 12-hour storage
Diesel Metering Pump	10 gpm	@ 60' head	1(1)	\$1,500	1	Transfers diesel from storage tank to high-shear mixtures
Asphalt Transfer Pump	1,000 gpm	@ 50' head	1	\$3,000	30	Transfers hot asphalt from delivery system to storage tank
Asphalt Storage Tank	80,000 gal	26' x 26'H	1	\$102,000	10	Storage for 10-day supply of asphalt, includes internal plate coils to heat the asphalt to 70°C
Asphalt Feed Pump	500 gpm	@ 30' head	1(1)	\$3,000	8	Transfers asphalt from storage tank to asphalt feed tank
Asphalt Feed Tank	4,000 gal	9' x 9'H	1	\$32,000	5	Storage for 16-hour supply of asphalt, includes internal plate coils to heat asphalt to 70°C
Asphalt Metering Pump	10 gpm	@ 60' head	1(1)	\$2,500	5	Transfers asphalt from asphalt feed tank to solids mixers via colloid mill
Colloid Mill	50 gpm	15" rotor	1(1)	\$65,000	75	Preparation of asphalt emulsion
Colloid Tank	500 gal	5' x 5'H	1	\$5,000	5	Heated storage of asphalt emulsion
Starch Transfer Pump	25 t/h	@ 60' head	1	\$3,000	25	Transfers starch pneumatically from delivery system to storage silo
Starch Storage Silo	500 t	500 t	1	\$80,000		Storage for 10-day supply of starch
Starch Feed Pump	25 t/h	@ 60' head	1(1)	\$3,000	25	Transfers starch pneumatically from starch storage to starch receiver
Starch Receiver	35 t	35 t	1	\$28,000		Storage of 8-hour supply of starch, will feed pneumatically to solid mixers
Starch Weighbelt Feeder	4 t/h	4 t/h	1	\$6,000	5	Pneumatically meters starch from starch receivers to solid mixers
Reagent Transfer Pump	10 gpm	@ 10' head	5	\$1,000	1	Transfers reagents (surfactant, flocculant, and pH-reagents) from delivery system to storage tanks
Reagent Storage Tank	500 gal	5' x 5'H	5	\$1,300		Storage for 1-week supply of concentrated reagents
Reagent Feed Pump	10 gpm	@ 10' head	5(1)	\$1,000	1	Transfers reagents from reagent storage tanks to reagent feed tanks
Reagent Feed Tank	50 gal	2' x 2'H	5	\$200		Storage for 1-day supply of diluted reagents
Reagent Metering Pump	1 gpm	@ 60' head	5(1)	\$1,000	1	Transfers reagents from feed tank to agglomeration circuit

Natural Fines Agglomeration Circuit

Table G-8
Coal Thickening Circuit Equipment List

Equipment Item	Capacity	Size	No. (spare)	Cost FOB	Power hp	Description
Static Thickener	9,000 gpm 114 t/h	125'φ	1	\$850,000	25	Produces clarified water from -100 mesh slurry
Thickener Overflow Pump	6,900 gpm	@40' head	1(1)	\$38,000	300	Transfers thickener overflow, clarified water to conventional plant water system
Thickener Underflow Pump	2,200 gpm	@40' head	1(1)	\$25,000	100	Transfers thickener underflow, -100 mesh raw coal slurry to surge tank

Table G-9
Coal Agglomeration Circuit Equipment List

Equipment Item	Capacity	Size	No. (spare)	Cost FOB	Power hp	Description
High-Shear Feed Sump	11,400 gal	11,400 gal	1	Civil		Gathers discharge from slurry holding vessel for feed to high shear
High-Shear Feed Pump	2,200 gpm	@60' head	1(1)	\$12,000	50	Transfers slurry from holding tank to high shear
High-Shear Distributor	2,200 gpm	4 way	1	Civil		Distributes slurry to the high-shear vessels
High-Shear Vessels & Mixer	2,200 gal	650 gal	4	\$90,000	100	Provides high agitation for a 1-min retention time to enable formation of the flocs/microagglomerates
Flotation Bank 1	8,000 ft ³	1,500 ft ³ /cell	8	\$70,000	50	Enables recovery of microagglomerates, approximately 6-min residence time, at 6% solids
Flotation Bank 2 Feed Sump	22,000 gal	11,000 gal	2	Civil		Gathers discharge froth from first-stage flotation for feed to second-stage flotation
Flotation Bank 2 Feed Pump	4,800 gpm	@ 30' head	2(1)	\$12,000	40	Transfers slurry from first- to second-stage flotation
Flotation Bank 2	8,000 ft ³	1,000 ft ³ /cell	8	\$46,500	50	Enables the separation of microagglomerates, approximately 6-min residence time, at 6% solids
Conditioning Tank	900 gal	900 gal	2	\$15,000	15	Redisperses froth before addition to vacuum filter, about 1-min residence time
Vacuum Filter Feed Pump	900 gpm	@ 40' head	2(1)	\$8,000	25	Transfers dispersed froth from conditioning tank to vacuum filter
Rotary Vacuum Filter	70 t/h		2	\$280,000	350	Reduction of froth moisture

Table G-10
Product Loadout Circuit Equipment List

Equipment Item	Capacity	Size	No.	Cost FOB	Power hp	Description
Product Belt Conveyor	180 t/h	30"W x 100'L	1	\$175,000	50	Transfers coal to silos
Storage Silos				-0-	10	Existing storage

Table G-11
Clarified Water and Refuse Disposal Circuit Equipment List

Equipment Item	Capacity	Size	No. (spare)	Cost FOB	Power hp	Description
High Capacity Thickener	11,000 gpm	65'φ	1	\$370,000	20	Produces clarified water from tails
Refuse Slurry Pump	550 gpm	@ 20' head	1(1)	\$12,000	10	Transfers clarifier underflow to belt filter
Belt Filter	45 t/h	4-meter W	2	\$270,000	25	Removal of water from refuse slurry
Refuse Belt Conveyor	45 t/h	24" x 70'	1	\$25,000	20	Transfers refuse to refuse bins (landfill)
Carbon Bed Feed Pump	225 gpm	@ 40' head	1	\$8,000	15	Transfers part of clarified water to activated carbon bed
Activated Carbon Bed	225 gpm		2	\$11,000		Water treatment to reduce dissolved solids and other impurities
Thickener Overflow Tank	60,000 gal	60,000 gal	1	\$80,000		Collect thickener overflow
Head Tank Pump	11,000 gpm	@ 70' head	1	\$35,000	200	Transfers process water from process water tank to head tank
Head Tank			1	\$3,000		Head tank to provide a constant flow of process water

Table G-12
Reagent Storage and Delivery Circuit Equipment List

Equipment Item	Capacity	Size	No. (spare)	Cost FOB	Power hp	Description
Diesel Transfer Pump	1,000 gpm	@ 30' head	1	\$2,000	20	Transfers diesel from delivery system to storage tank
Diesel Storage Tank	80,000 gal	26'φ x 26'H	1	\$34,000		Storage for 10-day supply of diesel
Diesel Feed Pump	500 gpm	@ 25' head	1(1)	\$3,000	10	Transfers diesel from storage tank to diesel feed tank
Diesel Feed Tank	4,000 gal	9'φ x 9'H	1	\$16,000		Storage for process feed diesel, 12-hour storage
Diesel Metering Pump	10 gpm	@ 60' head	1(1)	\$1,500	1	Transfers diesel from storage tank to high-shear mixtures
Reagent Transfer Pump	10 gpm	@ 10' head	5	\$1,000	1	Transfers reagents (surfactant, flocculant, and pH-reagents) from delivery system to storage tanks
Reagent Storage Tank	500 gal	5'φ x 5'H	5	\$1,300		Storage for 1-week supply of concentrated reagents
Reagent Feed Pump	10 gpm	@ 10' head	5(1)	\$1,000	1	Transfers reagents from reagent storage tanks to reagent feed tanks
Reagent Feed Tank	50 gal	2'φ x 2'H	5	\$200		Storage for 1-day supply of diluted reagents
Reagent Metering Pump	1 gpm	@ 60' head	5(1)	\$1,000	1	Transfers reagents from feed tank to agglomeration circuit

20-t/h Semiworks Agglomeration Plant

Table G-13
Coal Receiving, Storage, and Crushing Circuit Equipment List

Equipment Item	Capacity	Size	No.	Description
Truck Scale & Weigh Station	150 t/h	20 t truck	1	Coal delivery receipts
Coal Storage Pad	4000 t	4000 t	1	Temporary storage of coal to feed vibrating feeders
Reclaim Hopper	50 t/h	0-50 t/h		Regulates coal onto the feed coal belt conveyor
Vibrating Feeder	35 t/h		1	Regulates coal onto the crusher feed coal belt conveyor
Crusher Feed Belt Conveyor	35 t/h	24"W x 60'L	1	Transfers 2" x 0 feed coal from the stockpiles to the crusher
Tramp Iron Magnet	35 t/h	24" W belt	1	Protects crusher from tramp iron
Cage Packtor Coal Crusher	35 t/h	35 t/h	1	Reduces 2" topsize coal to 1/2" topsize
Hopper Belt Conveyor	35 t/h	24"W x 80'L	1	Transfers 1/4" x 0 coal either to grinding mill storage hopper or to the preclean deslime screen storage hopper
Surge Hopper	20 t	20 t	1	Provides 1/2-hour crushed coal storage

Table G-14
Coal Precleaning Circuit Equipment List

Equipment Item	Capacity	Size	No. (spare)	Description
Weigh Belt Feeder	35 t/h	35 t/h	1	Weigh and regulate raw coal
Deslime Screen	35 t/h	4' x 12'	1	Size separation at 28 mesh
Heavy Media Cyclone Feed Sump	1,200 gal	1,200 gal	1	Provides temporary storage for +28 mesh raw coal slurry
Heavy Media Cyclone Feed Pump	360 gpm	@40' head	1(1)	Transfers +28 mesh slurry to heavy media cyclones
Heavy Media Cyclone	25 t/h	14"φ	1	Specific gravity separation of raw coal to clean coal & refuse
Coal Sieve Bend & Refuse Drain & Rinse Screen	17 t/h 8 t/h	6' x 16"	1	Split screen dewater HM cyclones overflow and underflow streams
Dilute Media Sump	300 gal	300 gal	1	Collects clean coal and refuse D&R screen effluents
Dilute Media Pump	125 gpm	@40' head	1(1)	Transfers D&R screen effluents to magnetic separator
Magnetite Separator	40 gal	30"φ x 18"L	1	Recovers magnetite from slurry
Magnetite Thickener	120 gal	10'φ	1	Thickens magnetite slurry
Classifying Cyclone Feed Sump	600 gal	600 gal	1	Provides temporary storage for -28 mesh raw coal slurry
Classifying Cyclone Feed Pump	350 gpm	@40' head	1(1)	Transfers -28 mesh coal slurry to classifying cyclones
Classifying Cyclone	10 t/h	14"φ	1	Splits -28 mesh slurry stream into 28 x 100 mesh underflow and -100 mesh overflow
Spiral Concentrators	7 t/h	3-tum	3	Specific gravity separation of raw coal to clean coal & refuse
Clean Coal Screen	4 t/h	3' x 5'	1	Dewater spiral coal stream
Refuse Screen	3 t/h	3' x 5'	1	Dewater spiral refuse stream
Static Thickener	350 gpm 3 t/h	30'φ	1	Produces clarified water from -100 mesh slurry
Thickener Overflow Pump	350 gpm	@40' head	1(1)	Transfers thickener overflow, clarified water to the deslime screen
Thickener Underflow Pump	25 gpm	@40' head	1(1)	Transfers thickener underflow, -100 mesh raw coal slurry to surge tank
Surge Tank	1,500 gal	1,500 gal	1	Storage of -100 mesh raw coal slurry
Surge Tank Pump	50 gpm	@80' head	1(1)	Transfers -100 mesh slurry to rod mill
Clean Coal Conveyor	21 t/h	24"W x 100'L	1	Transfers coal to mill hopper
Refuse Conveyor	11 t/h	24"W x 50'L	1	Transfers refuse for disposal

Table G-15
Coal Grinding Circuit Equipment List

Equipment Item	Capacity	Size	No. (spare)	Description
Rod Mill Feed Hopper	25 t	900 ft ³	1	Temporary (1 hour) storage of crushed coal for feed to rod mill feeder
Rod Mill Belt Weigh Feeder	25 t/h		1	Regulates feed into the rod mill
Rod Mill	25 t/h	4'φ x 6'L	1	Reduces ¼" x 0 feed coal to minus 28 mesh, 50% solids coal slurry. Complete with rod charges, motor, and lube system
Rod Mill Discharge Sump	320 gal	320 gal	1	Gathers the rod mill slurry discharge, the 28M x 0, 50% solids coal slurry for feed to the ball mill
Rod Mill Discharge Pump	160 gpm	@ 40' head	1(1)	Transfers 28M x 0, 50% solids coal slurry to holding tank
Ball Mill	25 t/h	8.5'φ x 8'L	2	Reduces 28M x 0, 50% coal slurry to 95% passing 200M coal slurry, complete with initial ball charge, motor and lube system
Holding Tank Feed Sump	320 gal	320 gal	1	Gathers ball mill slurry discharge, 200M x 0, 50% solids coal slurry for feed to holding tank
Holding Tank Feed Pump	160 gpm	@ 30' head	1(1)	Transfers 200M x 0, 50% coal slurry to slurry holding tank
Slurry Holding Vessel	1600 gal	1600 gal	1	Stores 200M x 0, 50% solids coal slurry providing 10-min capacity and homogenous output to high- shear vessel, includes the mixer

Table G-16
Coal Agglomeration Circuit Equipment List

Equipment Item	Capacity	Size	No. (spare)	Description
High-Shear Feed Sump	2,000 gal	2,000 gal	1	Gathers discharge from holding vessel for feed to high shear
High-Shear Feed Pump	450 gpm	@60' head	1(1)	Transfers slurry from holding tank to high shear
High-Shear Vessel & Mixer	450 gal	450 gal	1	Provides high agitation for a 1-min retention time to enable formation of the flocs/microagglomerates
Flotation Bank 1	1,500 ft ³	300 ft ³ /cell	5	Enables recovery of microagglomerates, approx. 6-min residence time, at 6% solids
Flotation Bank 2 Feed Sump	2,200 gal	1,100 gal	1	Gathers discharge froth from first-stage flotation for feed to second-stage flotation
Flotation Bank 2 Feed Pump	1,370 gpm	@ 30' head	1(1)	Transfers slurry from first- to second-stage flotation
Flotation Bank 2	1,200 ft ³	300 ft ³ /cell	4	Enables separation of microagglomerates, approx. 6-min residence time, at 6% solids
Conditioning Tank	380 gal	380 gal	1	Redisperses froth before addition to vacuum filter, about 1-min residence time
Vacuum Filter Feed Pump	380 gpm	@ 40' head	1(1)	Transfers dispersed froth from conditioning tank to vacuum filter
Rotary Vacuum Filter	20 t/h	8'10"φ, 4 disc	1	Reduction of froth moisture

Table G-17
Size Enlargement, Drying and Product Loadout Circuit Equipment List

Equipment Item	Capacity	Size	No.	Description
Solid Mixer	20 t/h	33"φ x 90"L	1	Dispersion of binder within the cake
Pelletizer	20 t/h	14'φ	1	Pelletization of coal, about 6 mm to 8 mm
Dryer Belt Conveyor	20 t/h	24" x 40'L	1	Transfers wet pellets to dryer feed storage hopper
Storage Hopper	20 t	20 t	1	Provides 1-hour storage of wet pellets for feed to dryer
Vibrating Bed Dryer	20 t/h	4' x 9'	1	Pellet drying to harden pellets and reduce moisture, includes heating system
Cooling Conveyor	20 t/h	24"W x 40'	1	Cools dried coal pellets
Product Belt Conveyor	20 t/h	24"W x 80'	1	Transfers dried coal pellets to storage pad
Storage Pad	2,500 t	2,500 t	1	Provide 5-day storage for product

Table G-18
Clarified Water and Refuse Disposal Circuit Equipment List

Equipment Item	Capacity	Size	No. (spare)	Description
High Capacity Thickener	2,100 gpm	30'φ	1	Produces clarified water from tails
Refuse Slurry Pump	40 gpm	@ 20' head	1(1)	Transfers clarifier underflow to belt filter
Belt Filter	3 t/h	1m Wide	1	Removal of water from refuse slurry
Refuse Belt Conveyor	3 t/h	24"W x 60'	1	Transfers refuse to refuse bins (landfill)
Carbon Bed Feed Pump	50 gpm	@ 40' head	1	Transfers part of clarified water to activated carbon bed
Activated Carbon Bed	50 gpm		2	Water treatment to reduce dissolved solids and other impurities
Thickener Overflow Tank	6,000 gal	6,000 gal	1	Collect thickener overflow
Head Tank Pump	2,100 gpm	@ 70' head	1	Transfers process water from process water tank to head tank
Head Tank			1	Head tank to provide a constant flow of process water

Table G-19
Reagent Storage and Delivery Circuit Equipment List

Equipment Item	Capacity	Size	No. (spare)	Description
Diesel Transfer Pump	100 gpm	@ 30' head	1	Transfers diesel from delivery system to storage tank
Diesel Storage Tank	8,000 gal		1	Storage for 10-day supply of diesel
Diesel Feed Pump	50 gpm	@ 25' head	1(1)	Transfers diesel from storage tank to diesel feed tank
Diesel Feed Tank	400 gal		1	Storage for process feed diesel, 12-hour storage
Diesel Metering Pump	1 gpm	@ 60' head	1(1)	Transfers diesel from storage tank to high-shear mixtures
Asphalt Transfer Pump	100 gpm	@ 50' head	1	Transfers hot asphalt from delivery system to storage tank
Asphalt Storage Tank	8000 gal		1	Storage for 10-day supply of asphalt, includes internal plate coils to heat the asphalt to 70°C
Asphalt Feed Pump	50 gpm	@ 30' head	1(1)	Transfers asphalt from storage tank to asphalt feed tank
Asphalt Feed Tank	400 gal		1	Storage for 16-hour supply of asphalt, includes internal plate coils to heat asphalt to 70°C
Asphalt Metering Pump	10 gpm	@ 60' head	1(1)	Transfers asphalt from asphalt feed tank to solids mixers via colloid mill
Colloid Mill	5 gpm		1(1)	Preparation of asphalt emulsion
Colloid Tank	50 gal		1	Heated storage of asphalt emulsion
Starch Transfer Pump	5 t/h	@ 60' head	1	Transfers starch pneumatically from delivery system to storage silo
Starch Storage Silo	50 t	50 t	1	Storage for 10-day supply of starch
Starch Feed Pump	5 t/h	@ 60' head	1(1)	Transfers starch pneumatically from starch storage to starch receiver
Starch Receiver	4 t	4t	1	Storage of 8-hour supply of starch, will feed pneumatically to solid mixers
Starch Weighbelt Feeder	800 lb/h	800 lb/h	1	Pneumatically meters starch from starch receivers to solid mixers
Reagent Transfer Pump (delete for drum storage)	10 gpm	@ 10' head	5	Transfers reagents (surfactant, flocculant, and pH-reagents) from delivery system to storage tanks
Reagent Storage Tank (delete for drum storage)	50 gal		5	Storage for 1-week supply of concentrated reagents
Reagent Feed Pump	1 gpm	@ 10' head	5(1)	Transfers reagents from reagent storage tanks to reagent feed tanks
Reagent Feed Tank	5 gal		5	Storage for 1-day supply of diluted reagents
Reagent Metering Pump	0-1 gpm	@ 60' head	5(1)	Transfers reagents from feed tank to agglomeration circuit

Appendix H

ECONOMIC DETAILS

Table H-1
Major Plant Equipment and Energy Consumption - Agglomeration Plant

Equipment Item	Numbers Reqd	Spare	FOB Factory Unit Cost	Equipment Cost	Power (HP) Unit	Total (HP)
Coal receiving, storage and crushing circuit						
Truck scale and weigh station	1		\$100,000	\$100,000		0
Truck dump grate	4		\$10,000	\$40,000		0
Underground feed hopper	2		\$30,000	\$60,000		0
Vibrating feeder from hopper	4		\$15,000	\$60,000	5	20
Feed coal belt conveyor	1		\$150,000	\$150,000	25	25
Vibrating feeder from stacking tube	4		\$6,000	\$24,000	5	20
Crusher feed belt conveyor	1		\$75,000	\$75,000	20	20
Tramp iron magnet	1		\$15,000	\$15,000		0
Cage Packton coal crusher	1		\$170,500	\$170,500	700	700
Hopper belt conveyor	1		\$38,000	\$38,000	15	15
Surge hopper	1		\$56,000	\$56,000		0
Subtotal				\$788,500		800
Coal precleaning circuit						
Weigh belt feeder	1		\$40,000	\$40,000	10	10
Sieve bend	3		\$45,000	\$135,000	10	30
Heavy media cyclone feed sump	1		\$16,000	\$16,000		0
Heavy media cyclone feed pump	1	1	\$29,000	\$58,000	175	175
Heavy media cyclone	4		\$25,000	\$100,000		0
Sieve bend & clean coal D&R screen	2		\$45,000	\$90,000	10	20
Sieve bend & refuse D&R screen	1		\$45,000	\$45,000	10	10
Magnetite separator	1		\$38,000	\$38,000	5	5
Magnetite thickener	1		\$100,000	\$100,000	5	5
Dilute media sump	1		\$8,000	\$8,000		0
Dilute media pump	1	1	\$13,000	\$26,000	40	40
Classifying cyclone feed sump	1		\$13,500	\$13,500		0
Classifying cyclone feed pump	1	1	\$19,000	\$38,000	150	150
Classifying cyclone	8		\$10,000	\$80,000		0
Spiral concentrators	24		\$4,000	\$96,000		0
Clean coal screen	3		\$30,000	\$90,000	5	15
Refuse screen	2		\$30,000	\$60,000	5	10
Static thickener	1		\$500,000	\$500,000	45	45
Thickener overflow pump	1	1	\$28,000	\$56,000	150	150
Thickener underflow pump	1	1	\$15,000	\$30,000	15	15
Surge tank	1		\$25,000	\$25,000		0
Surge tank pump	1	1	\$12,000	\$24,000	15	15
Clean coal conveyor	1		\$175,000	\$175,000	50	50
Refuse conveyor	1		\$60,000	\$60,000	30	30
Subtotal				\$1,903,500		775

Table H-1 (cont.)
Major Plant Equipment and Energy Consumption - Agglomeration Plant

Equipment Item	Numbers Reqd	Spare	FOB Factory Unit Cost	Equipment Cost	Power (HP) Unit	Total (HP)
Coal grinding circuit						
Rod mill feed hopper	1		\$60,000	\$60,000		0
Rod mill belt weigh feeder	1		\$40,000	\$40,000	10	10
Rod mill	1		\$800,000	\$800,000	1,150	1,150
Rod mill discharge sump	1		\$8,000	\$8,000		0
Rod mill discharge pump	1	1	\$15,000	\$30,000	30	30
Ball mill	2		\$1,400,000	\$2,800,000	2,300	4,600
Holding tank feed sump	1		\$8,000	\$8,000		0
Holding tank feed pump	1	1	\$18,000	\$36,000	25	25
Slurry holding vessel	1		\$40,000	\$40,000	20	20
Subtotal				\$3,822,000		5,835
Agglomeration circuit						
High-shear feed sump	1		\$14,200	\$14,200		0
High-shear feed pump	1	1	\$14,100	\$28,200	65	65
High-shear distributor	1		\$1,500	\$1,500		0
High-shear vessels and mixer	6		\$100,000	\$600,000	125	750
Flotation bank 1	10		\$70,000	\$700,000	100	1,000
Flotation bank 2 feed sump	2		\$14,200	\$28,400		0
Flotation bank 2 feed pump	2	1	\$13,500	\$40,500	75	150
Flotation bank 2	8		\$70,000	\$560,000	100	800
Conditioning tanks	2		\$25,000	\$50,000	30	60
Vacuum filter feed pump	2	1	\$12,000	\$36,000	75	150
Rotary vacuum filter	2		\$400,000	\$800,000	450	900
Subtotal				\$2,858,800		3,875
Size enlargement, drying and product loadout circuit						
Solids mixer	4		\$115,000	\$460,000	100	400
Pelletizer	4		\$220,000	\$880,000	125	500
Dryer belt conveyor	4		\$25,000	\$100,000	25	100
Storage hopper	4		\$4,000	\$16,000		0
Vibrating bed dryer	4		\$500,000	\$2,000,000	120	480
Cooling conveyor	4		\$60,000	\$240,000	25	100
Product belt conveyor	1		\$175,000	\$175,000	50	50
Storage silos	1		\$1,500,000	\$1,500,000	10	10
Subtotal				\$5,371,000		1,640

Table H-1 (cont.)
Major Plant Equipment and Energy Consumption - Agglomeration Plant

Equipment Item	Numbers Req'd	Spare	FOB Factory Unit Cost	Equipment Cost	Power (HP) Unit	Total (HP)
Water clarification and refuse handling circuit						
High capacity thickener	1		\$550,000	\$550,000	25	25
Refuse slurry pump	1	1	\$8,500	\$17,000	5	5
Belt filter	2		\$270,000	\$540,000	25	50
Refuse belt conveyor	1		\$25,000	\$25,000	20	20
Carbon bed feed pump	1		\$8,000	\$8,000	15	15
Activated carbon bed	2		\$15,000	\$30,000		0
Thickener overflow tank	1		\$80,000	\$80,000		0
Head tank pump	1		\$43,000	\$43,000	350	350
Head tank	1		\$3,000	\$3,000		0
Subtotal				\$1,296,000		485
Reagent delivery and storage circuit						
Diesel transfer pump	1		\$2,000	\$2,000	20	20
Diesel storage tank	1		\$34,000	\$34,000		0
Diesel feed pump	1	1	\$3,000	\$6,000	10	10
Diesel feed tank	1		\$16,000	\$16,000		0
High shear diesel metering pump	1	1	\$1,500	\$3,000	1	1
Asphalt transfer pump	1		\$3,000	\$3,000	30	30
Asphalt storage tank	1		\$102,000	\$102,000	10	10
Asphalt feed pump	1	1	\$3,000	\$6,000	8	8
Asphalt feed tank	1		\$32,000	\$32,000	5	5
Asphalt metering pump	1	1	\$2,500	\$5,000	5	5
Colloid mill	1	1	\$85,000	\$130,000	75	75
Colloid tank	1		\$5,000	\$5,000	5	5
Starch transfer pump	1		\$3,000	\$3,000	25	25
Starch storage silo	1		\$80,000	\$80,000		0
Starch feed pump	1	1	\$3,000	\$6,000	25	25
Starch receiver	1		\$28,000	\$28,000		0
Starch weighbelt feeder	1		\$6,000	\$6,000	5	5
Reagent transfer pump	5		\$1,000	\$5,000	1	5
Reagent storage tank	5		\$1,300	\$6,500		0
Reagent feed pump	5	1	\$1,000	\$6,000	1	5
Reagent feed tank	5		\$200	\$1,000		0
Reagent metering pump	5	1	\$1,000	\$6,000	1	5
Subtotal				\$491,500		239
Total major plant equipment costs				\$16,531,300		13,629

Table H-2
Cost Estimation Factors

Direct Cost Factors		
Construction	40%	of plant equipment
Foundations	5%	of plant equipment
Piping	10%	of plant equipment
Structures and buildings	20%	of plant equipment
HVAC	5%	of plant equipment
Electrical	20%	of plant equipment
Instruments	10%	of plant equipment
Replacement parts	10%	of plant equipment
Indirect Cost Factors		
Temporary construction	5%	of process capital
Design engineering	20%	of plant equipment
Field supervision	5%	of process capital
Process contingency	10%	of process capital
Project contingency	20%	of direct and indirect capital

Table H-3
Preproduction Costs - Commercial Plant

	Upper Freeport	Pittsburgh No. 8
Variable costs		
Raw materials	\$858,830	\$880,623
Electricity	\$384,161	\$384,161
Process Water	\$1,363	\$1,779
Total	\$1,244,354	\$1,266,563
Fixed costs		
Labor	\$197,438	\$197,438
Maintenance material	\$181,844	\$181,844
Total	\$379,282	\$379,282
Fuel costs	\$115,763	\$115,763
Startup modifications	\$1,112,226	\$1,112,226
Total	\$2,851,625	\$2,873,834

Table H-4
Inventory Capital (7 day) - Commercial Agglomeration Plant

	Unit	Upper Freeport		Pittsburgh No. 8	
	Cost	Consumption	Cost	Consumption	Cost
Diesel	\$0.65 /gal	47,376	\$30,794	51,912	\$33,743
Asphalt	\$120 /t	168	\$20,160	168	\$20,160
Starch	\$240 /t	672	\$161,280	672	\$161,280
Flocculant	\$2.00 /lb	884	\$1,768	1,593	\$3,186
Frother	\$0.55 /lb	11,026	\$6,064	12,054	\$6,630
Lime	\$0.04 /lb	38,018	\$1,521	41,563	\$1,663
Magnetite	\$0.06 /lb	105,991	\$6,359	115,265	\$6,916
Preclean Flocculant	\$2.00 /lb	538	\$1,076	628	\$1,256
Total plant cost	2%	\$12,299,287	\$245,986	\$12,299,287	\$245,986
Total inventory costs			\$475,008		\$480,820

Table H-5
Unit Costs

Raw Material	
Diesel	\$0.65 /gal
Asphalt	\$120 /t
Starch	\$240 /t
Flocculant	\$2 /lb
Frother	\$0.55 /lb
Lime	\$0.04 /lb
Magnetite	\$0.06 /lb
Preclean Flocculant	\$2 /lb
Utilities	
Electricity	\$0.06 /kWh
Natural gas	\$3.5 /MBtu
Process water	\$0.5 /kgal
Labor	
Operating/Supervisory	\$22.5 /h
Maintenance	\$22.5 /h
Administrative/Support	30% of O&M labor cost
Other	
Maintenance Material	6% of process capital
Insurance/Taxes	3% of direct fixed capital

Table H-6
Major Plant Equipment and Energy Consumption - Natural Fines Circuit

Equipment Item	Numbers Req'd	Spare	FOB Factory Unit Cost	Equipment Cost	Power (hp) Unit	Total
Coal Thickening circuit						
Static thickener	1		\$850,000	\$850,000	25	25
Thickener overflow pump	1	1	\$38,000	\$76,000	300	300
Thickener underflow pump	1	1	\$25,000	\$50,000	100	100
Subtotal				\$976,000		425
Agglomeration circuit						
High-shear feed sump	1		\$14,200	\$14,200		0
High-shear feed pump	1	1	\$12,000	\$24,000	50	50
High-shear distributor	1		\$1,500	\$1,500		0
High-shear vessels and mixer	4		\$90,000	\$360,000	100	400
Flotation bank 1	8		\$70,000	\$560,000	50	400
Flotation bank 2 feed sump	2		\$14,200	\$28,400		0
Flotation bank 2 feed pump	2	1	\$12,000	\$36,000	40	80
Flotation bank 2	8		\$46,500	\$372,000	50	400
Conditioning tanks	2		\$15,000	\$30,000	15	30
Vacuum filter feed pump	2	1	\$8,000	\$24,000	25	50
Rotary vacuum filter	2		\$280,000	\$560,000	350	700
Subtotal				\$2,010,100		2,110
Product loadout circuit						
Product belt conveyor	1		\$175,000	\$175,000	50	50
Storage silos	1		\$0	\$0	10	10
Subtotal				\$175,000		60
Water clarification and refuse handling circuit						
Hi capacity thickener	1		\$370,000	\$370,000	20	20
Refuse slurry pump	1	1	\$12,000	\$24,000	10	10
Belt filter	2		\$270,000	\$540,000	25	50
Refuse belt conveyor	1		\$25,000	\$25,000	20	20
Carbon bed feed pump	1		\$8,000	\$8,000	15	15
Activated carbon bed	2		\$11,000	\$22,000		0
Thickener overflow tank	1		\$80,000	\$80,000		0
Head tank pump	1		\$35,000	\$35,000	200	200
Head tank	1		\$3,000	\$3,000		0
Subtotal				\$1,107,000		315
Reagent delivery and storage circuit						
Diesel transfer pump	1		\$2,000	\$2,000	20	20
Diesel storage tank	1		\$34,000	\$34,000		0
Diesel feed pump	1	1	\$3,000	\$6,000	10	10
Diesel feed tank	1		\$16,000	\$16,000		0
HS diesel metering pump	1	1	\$1,500	\$3,000	1	1
Reagent transfer pump	5		\$1,000	\$5,000	1	5
Reagent storage tank	5		\$1,300	\$6,500		0
Reagent feed pump	5	1	\$1,000	\$6,000	1	5
Reagent feed tank	5		\$200	\$1,000		0
Reagent metering pump	5	1	\$1,000	\$6,000	1	5
Subtotal				\$85,500		46
Total major plant equipment costs				\$4,353,600		2,956

Table H-7
Preproduction Costs - Fines Circuit

	Illinois No. 5
Variable costs	
Raw materials	\$90,060
Electricity	\$59,508
Process Water	\$973
Total	\$150,541
Fixed costs	
Labor	\$21,938
Maintenance material	\$47,890
Total	\$69,828
Fuel costs	\$0
Startup modifications	\$292,910
Total	\$513,279

Table H-8
Inventory Capital (7 day) - Fines Circuit

	Unit Cost	Illinois No. 5 Consumption	Cost
Diesel	\$0.65 /gal	32,256	\$20,966
Flocculant	\$2.00 /lb	884	\$1,768
Frother	\$0.55 /lb	4,692	\$2,581
Lime	\$0.04 /lb	16,178	\$647
Thickener Flocculant	\$2.00 /lb	3,830	\$7,660
Total plant cost	2%	\$3,239,078	\$64,782
Total inventory costs			\$98,404

**DATE
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10/27/93

END