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THE DEVELOPMENT OF COAL-BASED TECHNOLOGIES FOR DEPARTMENT OF DEFENSE FACILITIES

Phase III Final Report

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Work Performed Under Cooperative Agreement No. DE-FC22-92PC92162

For
U.S. Department of Energy
National Energy Technology Laboratory
P.O. Box 10940
Pittsburgh, Pennsylvania 15236

By
The Consortium for Coal-Water Slurry Fuel Technology
The Pennsylvania State University
C211 Coal Utilization Laboratory
University Park, Pennsylvania 16802

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EXECUTIVE SUMMARY

The third phase of a three-phase project investigating the development of coal-based technologies for U.S. Department of Defense (DOD) facilities was completed. The objectives of the project were to: decrease DOD's dependence on foreign oil and increase its use of coal; promote public and private sector deployment of technologies for utilizing coal-based fuels in oil-designed combustion equipment; and provide a continuing environment for research and development of coal-based fuel technologies for small-scale applications at a time when market conditions in the U.S. are not favorable for the introduction of coal-fired equipment in the commercial and industrial capacity ranges.

The Phase III activities were focused on evaluating deeply-cleaned coals as fuels for industrial boilers and investigating emissions control strategies for providing ultra-low emissions when firing coal-based fuels. This was addressed by performing coal beneficiation and preparation studies, and bench- to demonstration-scale emissions reduction studies. In addition, economic studies were conducted focused on determining cost and market penetration, selection of incentives, and regional economic impacts of coal-based technologies.

Coal Preparation/Utilization

Research conducted under Phases I and II of this project revealed a number of specific areas where continued and/or more focused effort was required in order to develop more effective, more reliable coal processing systems. The specific objectives of the Phase III research on coal preparation were centered around focused investigations into specific coal-cleaning options and their associated ancillary operations and the integration of processing/cleaning operations for overall system optimization. As in the earlier phases of the project, emphasis was on fine-coal processing for the production of a high quality, micronized coal for dry coal and coal-water mixture (MCWM) applications.

Simulations were performed for open and closed grinding circuit configurations using grinding kinetics theory to determine an optimum grinding circuit for producing MCWMs. Data obtained from investigations from the first two phases of the project, and supplemented by additional experiments carried out in Phase III, were used to perform simulations for both conventional ball milling and stirred-media milling. These simulations clearly illustrated that

single-stage grinding in a conventional ball mill does not produce size distributions with the form required for stable coal-water mixtures. A general two-mill grinding circuit simulator, previously developed at Penn State, was used to evaluate the performance of a two-mill grinding circuit under various operating conditions to obtain design criteria for a full-scale system. Data from the simulator were used to develop a conceptual integrated grinding/cleaning circuit for preparing MCWMs.

Particle size separations were carried out using a high-speed, solid-bowl centrifuge under various operating conditions to classify -100 mesh Upper Freeport seam coal. Centrifuge performance was evaluated in terms of the size selectivity curves and corresponding performance parameters: cut size, sharpness index, and apparent bypass. The results indicated that the centrifuge was able to achieve cut sizes less than 10 μm under nearly all conditions, with sharpness indices generally ranging from 0.45 to 0.60. In most cases, the apparent bypass was less than 0.3.

The solid-bowl centrifuge was also used for density separations using a magnetite-and-water dense medium under various operating conditions for treating -100 mesh coal. Centrifuge performance was evaluated in terms of the clean coal yield and ash content while, for selected tests, partition curves were generated. In nearly all cases, clean coal could be produced at an ash content of less than 7%. The clean coal yields varied over a wide range, with values over 80% possible in some cases.

Density separations were also conducted using a magnetic fluid as the dense medium. The separations were carried out using several laboratory separation cells designed to operate within a Franz electromagnet. The results demonstrated that separations could be made at different densities using both batch and continuous separators. Magnet pole design was found to be critical in achieving a constant medium density.

An integrated centrifugal/froth-flotation system was evaluated, with the aim of extending the size range of fine coal cleaning further. It was found that for a given clean coal yield, the integrated process was able to produce a clean coal with a lower ash content than either dense-medium cycloning or flotation alone.

Surface-based processes such as froth flotation are particularly attractive for deep cleaning of fine coal for MCWM applications because of their relative insensitivity to particle size. Laboratory-scale, batch testing of fine coal flotation, conducted in Phases I and II of the

project was extended to continuous, pilot-scale studies. A two-stage, rougher/cleaner circuit with a capacity of 8 gpm was used in these tests. The results indicated that performance in the continuous system roughly paralleled that in the batch tests, but that flotation rates and recoveries were consistently lower. The differences were attributed primarily to low mobility and stability of the froth. Continuous flotation tests were also conducted in a pilot-scale, fully instrumented flotation column. It was found that fine coal with an ash content of 9.4% could be cleaned to about 5% at 75 – 80% yield in a single stage operation under appropriate conditions of air flow rate, solids feed rate and concentration and froth-phase gas hold-up.

Continuous dry processing of fine coal using triboelectrostatic separation was also investigated as an extension of the batch testing carried out under Phases I and II of the project. A specially designed test unit was shown to separate fine coal selectively, but at lower efficiency than in the equivalent batch systems. Design modifications to improve separation efficiency were investigated. Preliminary evaluations of an integrated grinding/dry separation system provided encouraging results.

Emissions Reduction

The studies of emissions from coal-fired industrial boilers, that were initiated in the first two phases of the project, were extended into Phase III and involved fundamental, pilot-scale, and full-scale demonstrations and included: 1) evaluations of NO_x reduction strategies such as low-temperature selective catalytic reduction (SCR) and reburning by injecting a biomass-based product and a MCWM; 2) a pilot-scale investigation of simultaneous SO₂ removal/NO_x reduction technology injecting a biomass-based pyrolysis product into the combustion chamber; 3) a fundamental study of formation/destruction mechanisms of polycyclic aromatic hydrocarbons (PAHs) followed by demonstration-scale testing; 4) comprehensive studies of trace elements and mercury emissions investigating the effect of coal cleaning and particulate removal devices; 5) the development of a unified methodology for simultaneously measuring trace element emissions and mercury speciation; 6) an investigation of fine particulate matter removal using ceramic filters; and 7) the development of a molecular basket to capture CO₂ emissions from the flue gas of coal-fired boilers.

A low-temperature SCR study was initiated in Phase II with the objectives of identifying and/or developing a NO_x reduction catalyst that is compatible with the typical operating

conditions and the economic constraints of industrial boilers, establishing the limitations of candidate NO_x reduction catalysts so that their implementation in pilot- and demonstration-scale tests will be straightforward, and identifying maximum allowable transients that the catalyst can be exposed to before losing effectiveness. Bench-scale testing was successfully completed in Phase II and the results were scaled up for pilot-scale testing in Phase III. Pilot-scale tests demonstrated that a low-temperature SCR catalyst can achieve greater than 20% reduction at typical utility of industrial baghouse temperatures. Further work is necessary before this concept would be commercially accepted.

A reburning study was performed to understand the mechanism by which NO_x emissions are reduced when cofiring coal with MCWM. The results confirmed the potential of MCWM as a reburn fuel for significant NO_x reduction as MCWM exhibits reductions similar to that of natural gas. Reburn effectiveness depends strongly upon the reburn zone stoichiometry, which is controlled by the reburn fuel heat input and by the air staging configuration. The results indicated that the gas phase reactions did not contribute significantly to the reduction of gas phase NO_x that was produced in the initial stages of combustion but rather it was reactions with the carbon/char that resulted in the lower NO_x emissions.

A model was developed for NO reduction through homogenous gas phase reactions when a biomass-derived product, made from biomass pyrolysis products that have been reacted with air and a lime/slurry, was used as a reburn fuel. The model was validated through thermal gravimetric analysis and pilot-scale combustion testing. The study showed that the percentage of NO_x reduction using the biomass-derived product could be predicted using the gas phase reaction model. Levels of NO_x reduction were correlated with pyrolysis gas yield.

The PAH work, initiated in Phase II, was continued into Phase III to study the effect of temperature and oxygen concentration on the formation and destruction of PAHs in a laminar-flow reactor and a full-scale industrial boiler. Trends in PAH formation and destruction were observed as a function of temperature and oxygen level in the fundamental study; although, availability of oxygen is the more dominant factor in PAH formation and destruction than temperature, which was observed in both the reactor and boiler. Only naphthalene and phenanthrene were detected in the demonstration boiler flue gas whereas the laminar-flow reactor, with decreased air-to-fuel mixing, resulted in significantly higher PAH emissions.

Three mercury and trace element activities were performed, which involved evaluating ceramic filters as a control option for mercury removal, evaluating trace elements emissions from deeply-cleaned coals produced in DOE's Premium Clean Coal Program, and developing a sampling train and methodology capable of simultaneously sampling mercury species and all inorganic trace elements. As part of the ceramic filter evaluation, a long-term demonstration of the filters was performed, similar to the one conducted in Phase II of the project, to assess the technical feasibility of using ceramic filters in lieu of fabric filters.

The ceramic filter chamber was modified after the second phase of the project (first 1,000-hour ceramic filter demonstration) to enhance filter regeneration, a new set of filters was installed, and the system was operated for ≈ 930 hours. The filters were regenerable; however, the pressure drop continued to increase, which is believed to be primarily due to the plugging of the filter substrate pores following the loss of the membrane through erosion. This is a concern for commercialization in a boiler setting and needs to be addressed by the manufacturer. The ceramic member filter particulate collection efficiencies ranged from 99.80 to 99.86%, which were higher than baghouse collection efficiencies using high-performance bags. The stack particulate emissions were below DOE's LEBS target of 0.01 lb/million Btu; however, the design of the ceramic membrane filters evaluated was not able to meet DOE's HIPPS target of 0.003 lb/million Btu.

The ceramic membrane filters were more efficient in reducing trace elements in the stack gas, accounting for an additional 45 to 61% reduction (excluding mercury) over that removed by the fabric filters. The ceramic membrane filters effectively removed 79% more mercury from the gas phase than the fabric filters, even at elevated temperatures. The use of the ceramic membrane filter is a potential control option for both particulate and gas phase mercury emissions provided the high pressure drop issue can be resolved.

The effect of coal cleaning on trace elements emissions was investigated using cleaned coals from CQ, Inc. (cleaned by a combination of heavy media cyclones, water spirals, and froth flotation) and from AMAX Research & Development (cleaned either by advanced flotation or selective agglomeration) to investigate the basic assumption that there is a direct relationship between the depletion/enrichment of an element in the fuel to its depletion/enrichment in the flue gas. The data from the AMAX fuels indicate that there is not always a direct relationship between the reduction of the trace elements in a fuel and a corresponding reduction in emissions.

By contrast, the data obtained from the CQ Inc. fuels suggested that there was a direct relationship between depletion/enrichment of an element in the fuel and its concentration in the flue gas. These different results suggest that trace element emissions can not be predicted solely from their concentration in the fuel. The interaction of the inorganics within the combustion system often affects the partitioning and concentration of the trace elements in the gas and ash streams. Therefore, the effectiveness of extensive coal cleaning in the reduction of trace element emissions should be determined on an individual basis, taking into account the coal composition and system configuration.

A study was performed to develop a sample train and methodology capable of simultaneously sampling mercury species and all inorganic trace elements EPA has listed as inorganic hazardous air pollutants (IHAPs). The advantage of such a system is that it will enable IHAPs and mercury species to be sampled simultaneously, thereby reducing time and cost for stack gas sampling. The train and methodology developed, referred to as the PSU Method, consist of components of both the Ontario Hydro Mercury Speciation Method and Method 29 (total metals) sampling trains. Of particular interest was whether the differences in the PSU Method train configuration compared to the Method 29 and Ontario Hydro trains compromised the measurement of either or both of the total trace elements or mercury species. This was found not to be the case as statistical analysis of the data verified the use of the PSU Method for measuring trace elements and mercury species in combustion flue gas.

Novel CO₂ “molecular basket” adsorbents, based on polyethylenimine (PEI)-modified mesoporous molecular sieve of MCM-41 type (MCM-41-PEI), were successfully developed for CO₂ removal from combustion flue gas. Carbon dioxide was selectively separated from boiler flue gases by using the adsorbent. The adsorbent adsorbed little N₂ and O₂. The selectivity of CO₂/NO_x was 2.5 for natural gas-fired flue gas and the separation selectivity for CO₂/SO₂ and CO₂/NO_x were 10.7 and 2.86, respectively, for coal-fired flue gas. The desorption of CO₂ was complete; however, very little NO_x and SO₂ desorbed after adsorption indicating the need for pre-removal of NO_x and SO₂ from the flue gas mixture before capture of CO₂ by the PEI based “molecular basket” adsorbent.

Economic Evaluation

The objectives of this activity were to determine cost and market penetration, selection of incentives, and regional economic impacts of coal-based fuel technologies. In addition, DOD's fuel mix was determined and a national energy portfolio constructed that minimizes energy price shock effects.

A market penetration model was formulated to find the equilibrium optimal mix of boiler retrofit technology adoption (to MCWM firing) among a sample population of watertube boilers located in the Pennsylvania counties of Cambria and Indiana (two main coal-producing counties) and, after model development, for an expanded study region of the entire state. It was estimated that only 40% of Pennsylvania's $\approx 7,000$ watertube boilers can be considered for MCWM retrofitting, of which 36 would benefit from adopting the retrofit technology.

A study was performed on the selection of incentives for commercializing the MCWM technology. The study focused on the microeconomic responses generated by the firms to the incentives. The results show that in the existence of a no-risk opportunity, a tax incentive is the best alternative to offer to the industry to induce an increase in their profits.

Community sensitivity to using coal in a power facility was studied using methods which integrated economic valuation with techniques used in psychology to characterize risk perceptions to value the welfare impacts due to the presence of energy production facilities. Results from the study show that an individual's willingness to pay for risk prevention or reduction is a function of their perceptions of the health, environmental, aesthetic, and economic impacts as well as their socio-economic characteristics. Individuals appear unable to distinguish between the probability and the severity of a risk in the manner suggested by the definition of risk. Perceived environmental, health, and aesthetic impacts play a larger role in determining option prices than potential economic impacts, explaining in part why residents may oppose a facility even when it will likely bring economic benefits to an area.

A review was performed to estimate the effects on the U.S. economy and its energy sectors of conservation strategies to reduce CO₂ emissions. The analysis was undertaken with a 20-sector computable general equilibrium model by simulating various responses to command and control, carbon tax, and carbon emission permit policies. The results indicate that the characterization of energy conservation as a no-regrets strategy is too strong. In all of the simulations, energy sectors stand to lose, though, in some cases, not anywhere near as much as

would be expected. Each of the simulations of mandated conservation also lead to decline in output and employment for the U.S. economy. In contrast, some of the price-induced conservation response strategies also simulated have a neutral impact on the overall economy.

An activity to examine U.S. military fuel consumption and describe alternative technologies that could be currently or potentially applied to the military in order to comply with U.S. government directives to lower emissions, lower dependence on imported oil, and to become more energy efficient was performed. It was shown that for both power/heat generation and transport applications, there is a variety of possible alternatives that could be applied for military use. Energy use in the military could also be significantly affected by new energy management systems and implementation of modified, more efficient, conventional technology.

Portfolio theory was used to demonstrate how the energy mix consumed in the U.S. could be chosen if the goal is to reduce the risks to the domestic macroeconomy of unanticipated energy price shocks. The results indicate that the electric utility industry is operating very close to the minimum variance position with a risk aversion strategy. In contrast, overall energy consumption in the U.S. is far from an efficient mix. A shift towards coal consumption would reduce price volatility. With the inclusion of potential externality costs, the shift remains away from oil but is towards natural gas instead of coal. To achieve such shifts, policymakers could use regulation or tax incentives to industries to encourage the use of certain fuels. Of course, a minimum risk portfolio does not imply a minimum cost portfolio, and selecting a low-risk portfolio may lead to higher energy costs overall. The costs associated with an occasional energy price shock may be less than the cost associated with energy dependence or a dramatic shift towards coal, synfuels, and other alternative energy sources.

Evaluation of Deeply-Cleaned Coals as Boiler Fuels

Deeply-cleaned coals produced by Cyprus/Amax Research & Development Center were evaluated as industrial boiler fuels. The coals, produced in DOE's Premium Fuel Program using advanced column flotation and oil agglomeration cleaning techniques, were tested by Penn State to evaluate their combustion, emissions, and handling characteristics. The testing at Penn State complemented the Premium Fuel Program, which had the overall objectives of producing an alternative fuel that will emit less toxic emissions, moving advanced cleaning technologies from the laboratory to a fine coal cleaning plant, and producing highly-loaded MCWMs that are

competitive with fuel oil. The Penn State activities included modifying a MCWM circuit to handle filter cake to produce MCWMs, determining the MCWMs atomization performance, stability, and rheological characteristics, performing bench-scale evaluations of filter cake handleability, and conducting pilot-scale and demonstration-scale combustion tests to evaluate combustion performance and emissions, specifically trace elements.

The objective of the filter cake handleability study was to evaluate the flow characteristics of fine, filter-cake coal using the Jenike shear-testing procedure and to investigate the use of a pilot-scale test bin for validation of the results. The Jenike shear tester was modified to eliminate inherent weaknesses of the system to increase the sensitivity and range of results. Tests were performed using fine, filter-cake material, the corresponding parent coal, and various blends of the two. The results from this work showed that the effective yield locus, a measure of the stresses required to maintain steady flow in a consolidated powder, appears to be relatively insensitive to mixture composition and moisture content. Increasing coarseness due to the addition of the coarser, parent coal caused a slight increase in the frictional resistance to flow, while moisture addition reduced friction, probably by a lubrication effect. The effects of mixture composition, i.e., size consist, on the (static) yield locus, a measure of the stresses needed to initiate flow in a consolidated powder, revealed that blends of the filter-cake material with the parent coal tended to be more cohesive than either of the separate components. This rather surprising result is attributed to enhanced packing in the blends and indicates potential materials-handling problems if coal fines are added to the normal feed coal in an existing system. Increased moisture content increases cohesion, probably through capillary forces, but reduces internal friction, presumably by lubrication. Because of their bimodal size distributions, the coarse/fine blends generally packed to higher bulk density than the separate components. The filter-cake material was slightly compressible – bulk density increases with applied stress – while the parent coal is essentially incompressible. Flow functions for these materials indicated that while the flowabilities of the filter cake and parent coal were relatively insensitive to consolidation, those for the blends tended to decrease as consolidation stresses were increased. It is postulated that consolidation stresses applied to the blends affected the fine component disproportionately, leading to enhanced cohesion despite very little change in overall bulk density. Measurements of wall friction against a steel reference material showed very little variation with mixture composition and lead consistently to angles of wall friction of about 27°.

The filter cakes were prepared into MCWMs that exhibited satisfactory handling and short-term storage properties (the MCWMs were prepared without stabilizers or particle size manipulation) for use as industrial boiler fuels. The combustion performance of the MCWMs was good, with the exception of one of the MCWMs (i.e., prepared from the Taggart seam coal cleaned by column flotation), and combustion efficiencies for a suite of tests ranged from ≈ 97 to 98%. The filter cake MCWMs performed as well or better than MCWMs prepared in the single- or double-stage grinding circuit. The level of natural gas cofire that was required to achieve these combustion efficiencies was 23 to 30%.

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1.0 INTRODUCTION

The U.S. Department of Defense (DOD), through an Interagency Agreement with the U.S. Department of Energy (DOE), initiated a three-phase program with the Consortium for Coal-Water Slurry Fuel Technology, with the aim of decreasing DOD's reliance on imported oil by increasing its use of coal. The program was conducted as a cooperative agreement between the Consortium and DOE. The first two phases were previously reported (Miller et al., 1997; Miller et al., 2000) and the third and final phase is reported herein.

This introduction of the report discusses the overall project objectives, summarizes the results from the first two phases, and provides an outline of the third phase of the project.

1.1 DOD Project Objectives

The general objectives of the cost-shared cooperative agreement were to:

1. Establish a National Center of Excellence for Coal Utilization to decrease DOD dependence on foreign oil and increase its use of coal by pursuing a program of research and development and actively participating in technology transfer;
2. Promote both the public and private sector deployment of technologies for utilizing coal-based fuels in oil-designed combustion equipment; and
3. Provide a continuing environment for research and development of coal-based fuel technologies for small-scale applications at a time when market conditions in the U.S. were not favorable for the introduction of coal-fired equipment in the commercial and industrial capacity ranges.

To achieve the objectives of the project, a team of researchers was assembled and the members of the Consortium for the Phase III activities were Penn State (The Energy Institute, Department of Energy and Geo-Environmental Engineering, and Department of Energy, Environmental, and Mineral Economics), AMAX Research and Development Center, Corning Incorporated, and Foster Wheeler Development Corporation.

Phase I activities were focused on developing clean, coal-based combustion technologies for the utilization of both micronized coal-water mixtures (MCWMs) and dry, micronized coal (DMC) in fuel oil-designed industrial boilers. Phase II research and development activities continued to focus on industrial boiler retrofit technologies by addressing emissions control strategies for providing ultra-low emissions when firing coal-based fuels. Phase III activities

expanded upon emissions reduction strategies through the use of deeply-cleaned coals as a means of reducing air toxics. Each phase included an engineering cost analysis and technology assessment.

1.2 Summary of Phase I Activities

Phase I activities were focused on developing clean, coal-based combustion technologies for the utilization of both MCWMs and DMC in fuel oil-designed industrial boilers. These technologies were identified by the U.S. Corps of Engineers, Construction Engineering Research Laboratory, as the two top priorities from a list of six topics for initial study (Schanche 1992). The specific objective in Phase I was to deliver fully engineered retrofit options for a fuel oil-designed watertube boiler located on a DOD installation to fire either MCWM or DMC. This was achieved through fundamental, pilot-scale, and demonstration-scale activities investigating coal beneficiation and preparation, and MCWM and DMC combustion performance. In addition, detailed engineering designs and an economic analysis were conducted for a boiler located at the Naval Surface Warfare Center, near Crane, Indiana.

1.2.1 Coal Beneficiation and Preparation Studies

Samples of six coals representing a range in the relative processing complexity required to meet sulfur and ash specifications for use in fuel oil-designed boilers were procured. Each of the samples was subjected to extensive characterization by float-sink analysis to determine the required level of cleaning. Some of these coals, and much of the potential reserve base, cannot be cleaned to the <5% ash, <1% sulfur specification by conventional means. Preliminary liberation models were developed in order to establish appropriate (conventional) cleaning strategies. Fine-coal cleaning processes including fine gravity separations and surface-based processes such as froth flotation and selective agglomeration were evaluated for their applicability to those coals which require fine grinding to achieve the necessary liberation. The factors that determine the size consist required for MCWM and the grinding/classification systems needed to obtain the desired size distribution were also investigated. A dry coal cleaning system based on triboelectrostatic separation was evaluated particularly for DMC applications where dry cleaning would be preferred over wet cleaning.

1.2.2 MCWM and DMC Combustion Performance Evaluation

The combustion performance evaluation included conducting fundamental studies to determine the effect of the mineral matter on boiler tube erosion and deposition, identify the mechanism of atomizer wear (corrosion or erosion), and computationally model the burner and boiler. In addition, MCWMs and DMC produced from the candidate coals were fired in 1,000 and 15,000 lb steam/h watertube boilers. The fuels were fired in the 1,000 lb steam/h boiler in order to determine their relative performance, while one of the coals was used in two demonstrations firing DMC and MCWM in the 15,000 lb steam/h boiler (which is of similar size to the retrofit candidate at Crane). The accomplishments in this aspect of the project included: modifying and optimizing a full-scale boiler system to fire DMC and MCWM; integrating the coal storage, handling, and micronization with the burner (for DMC firing); achieving 95% combustion efficiency (DMC); determining that erosion is not significant in the convective pass; and meeting targeted NO_x emissions of <0.6 lb/MM (million) Btu.

1.2.3 Engineering Design

Two engineering studies were performed for the complete retrofit of a DOD boiler to fire either DMC or MCWM. Each design package includes a fuel preparation system (for the DMC option only; two conceptual MCWM processing circuits were prepared but not engineered), fuel delivery and handling systems, low-NO_x burner, baghouse, forced-draft fan, combustion air preheaters, induced-draft fan, ash silo, stack, and control system. The retrofit designs conform to accepted engineering practices and site requirements. The designs were based on the system in place at Penn State and the information that was learned from operating it. The design packages are intended to be used for soliciting bids from engineering/construction firms for completion of the detailed design, construction, and start-up of the candidate DOD boiler.

1.2.4 Cost/ Economic Analysis

With respect to direct capital and operating cost considerations, retrofit DMC and MCWM technology boilers are not economically viable over a broad range of parameters for the Crane Site. For boiler retrofits to become economically viable, capital cost must be reduced, transportation costs of MCWM must be reduced, higher performance must be attained, and demand for slurry fuel in any given local area must be increased. Improving these parameters

would increase the price differential of the DMC and MCWM with respect to fuel oil and natural gas.

With respect to direct capital and operating cost considerations, new DMC and MCWM technology boilers are not economically viable over a broad range of parameters for the Crane Site. There does not exist a sufficient critical mass of military boilers in one location to transform this into a viable option. For new DMC and MCWM technology boilers to become economically viable, capital costs and boiler derating must be reduced, higher performance must be attained, or fuel costs must be lowered.

Retrofit capital costs, followed by transportation costs, are the major inhibiting factors to the economic viability of MCWM technology at the Crane site. DMC delivered fuel costs are insensitive to fuel demand level. Transportation costs can be improved by developing a lower cost system of transportation dedicated solely to coal. With respect to environmental regulations, DMC and MCWM technologies are economically viable under a broad range of circumstances. Given the small size of industrial boilers, existing pollution control devices will adequately meet environmental regulations. The regional multiplier effects of DMC and MCWM technologies are not significant for cases of less than 10 and 15 Crane size-equivalent installations, respectively, in a substate area. Regional multiplier effects can be significantly enhanced by purchasing more inputs from local vendors, though the sites where this is possible are limited.

1.3 Summary of Phase II Activities

As the project progressed, Phase II was modified to incorporate results from Phase I and reflect programmatic changes by DOD. The original objectives of Phase II were to: (a) extend the Phase I boiler retrofit options by including designs to achieve further reductions in gaseous and particulate emissions, (b) prepare and characterize fuels compatible with coal precombustors, and (c) investigate precombustion as a means of using high ash, high sulfur coals. Upon investigating precombustion options for installing a system on either the demonstration boiler (15,000 lb steam/h) or research boiler (1,000 lb steam/h), it became apparent that there were limited viable options and that the complexity of the systems would likely preclude their use on small-scale, industrial boilers. A similar conclusion was presented by the U.S. Corps of Engineers regarding the use of slagging combustors in the Army (Davidson et al., 1991). Consequently, the Phase II

work was revised by eliminating the precombustion studies and focusing on emissions reduction strategies.

1.3.1 Emissions Reduction

Strategies were developed to provide for ultra-low emissions when firing coal-based fuels in industrial boilers. Emissions that were addressed were SO₂, NO_x, fine particulate matter (PM_{2.5}), and air toxics (trace elements and volatile organic compounds). To accomplish the overall objective, emissions reduction strategies were reviewed, fundamental and pilot-scale studies were conducted, and selected emissions reduction techniques were demonstrated.

It was the intent of this activity to reduce the emissions levels of SO₂, NO_x, and particulate matter from that obtained in Phase I of the project. The Phase I emissions levels were 1.2 lb SO₂/MM (million) Btu (using low-sulfur coals), 0.6 lb NO_x/MM Btu (using low-NO_x burners), and <0.1 lb particulate matter/MM Btu (Penn State's demonstration boiler's permitted level from the Department of Environmental Protection) and >99.5% collection efficiency using a fabric filter baghouse. The Phase II targets, which are those from the U.S. Department of Energy's (DOE) High Performance Power Systems program, were: SO₂ <0.1 lb/MM Btu; NO_x <0.1 lb/MM Btu; and particulate matter <0.003 lb/MM Btu. The SO₂ and NO_x targets were met; however, the particulate matter emissions were slightly greater than the target and were ≈0.005 lb/MM Btu. In addition to achieving the SO₂, NO_x, and particulate matter emissions, a detailed characterization of air toxics from coal-fired industrial boilers was conducted to assist DOE in determining the levels and composition of air toxic emissions to identify control strategies.

1.3.2 Coal Preparation/ Utilization

Research was carried out on the design of coal grinding/classification systems for the efficient production of optimally formulated MCWM, and on the development of advanced processes for beneficiating high ash/sulfur coals. Detailed investigations of the rheology and stability of bimodal (coarse/fine) slurries demonstrated that these can be used to achieve high solids loading at acceptably low viscosity provided the appropriate composition range (coarse:fine) is maintained. This constraint was especially critical as the solids loading is increased. Investigation of a two-stage grinding/classification scheme demonstrated that this

approach could be used both to permit grinding under reasonably optimum conditions and to provide considerable flexibility in product composition. General criteria for slurry formulation and for the design and operation of the grinding circuit were established. Simulations of dense-medium separations were performed to examine the effects of design and operating conditions on the separation efficiency for 28x100 mesh coal. Dense-medium separations were carried out using a solid-bowl centrifuge to process 100x150 mesh coal. Magnetic fluids were tested as an alternative medium in float-sink separations. Studies of the use of froth flotation for deep-cleaning high ash/sulfur coals revealed that single-stage separations are often inadequate. Several different grinding-flotation circuits were investigated in order to establish the role of circuit configuration in the efficiency of coal cleaning. The results indicated that the most effective circuit depends strongly on coal type. Fundamental studies of flotation kinetics were conducted in order to establish the basis for process/circuit design and control. Emphasis was on the development of procedures for evaluating the distributions of flotation rate constants for the various floatable components. These kinetics parameters represented the key to the role of coal characteristics in flotation circuit performance. Column flotation is an attractive alternative to conventional flotation for deep cleaning of coal, particularly because of the froth-washing capability. Laboratory studies were used to establish column operating conditions for the production of clean coal from different feed sources. Process simulations were used to evaluate the performance of three-stage (rougher/cleaner/scavenger) circuits for producing clean coal at <5 wt.% ash with maximum combustibles recovery. The classification and cleaning steps in slurry production must generally be carried out at relatively low solids concentration. A preliminary analysis of the use of gravity sedimentation methods for slurry density control following these steps was also performed. Studies of dry cleaning included testing of a batch triboelectrostatic separator to examine the effects of surface moisture, humidity, and particle size on the separation of Upper Freeport, Pittsburgh and Indiana seam coals. A laser-sheet generator was fabricated for use in examining particle charge after tribocharging.

1.3.3 Engineering Design and Cost; and Economic Analysis

The widespread adoption of new coal-fired boilers has economic impacts beyond initial capital expenditures and the subsequent increased demand for coal. Results indicated that modest, but far from insignificant, gains would be transmitted throughout the surrounding

regional economy via multiplier effects. In addition, despite improved combustion efficiency, which leads to a relative decrease in most pollutants, a relative increase in CO₂ emissions may ensue.

Research found the market potential for private sector retrofits of industrial boilers to MCWM to be rather modest in Pennsylvania at the energy prices in effect in the mid-1990s (note, the prospects are much better if coal refuse is utilized but much worse if the current lower price of oil is considered). The shift away from oil to coal, however, generates relatively more business in Pennsylvania, especially if it displaces foreign oil, and could translate into as many as 10,000 new jobs on a sustained basis.

Public attitudes toward new coal-fired boiler adoption were examined and it was found that such concerns were a rather low priority among a long slate of environmental issues. Finally, it was noted that the increased energy conversion efficiency and pollution reduction capability of MCWM technology relative to oil-fired units reduced all pollutants but would likely lead to a relative increase in CO₂ emissions, though of a trivial total amount.

Finally, the economic impact of a carbon tax on the Pennsylvania economy was examined and found a minimal overall economic impact, though a sizeable negative impact on the State's coal industry.

1.4 Phase III Project Outline

As the project progressed, Phase III was modified to reflect results from Phase II and programmatic changes by DOD and DOE. The original objectives of Phase III were to: (a) develop coal-based fuel/waste cofiring technologies, and (b) assist DOD in improving the combustion performance and reducing emissions from existing stoker-fired boilers. Since the initial development of the program's statements of work (Phases I through III), there has been a change in military boiler plant operating philosophy. This, coupled with recent developments in cofiring technologies and DOE coal preparation programs, necessitated the revision of the Phase III statement of work. Consequently, the Phase III work was revised by eliminating coal-based fuel/waste cofiring and stoker combustion performance analysis and evaluation, and focusing these efforts toward evaluating deeply-cleaned coals as industrial boiler fuels, and investigating fundamental, pilot-scale, and demonstration-scale emissions reduction strategies. The revised Phase III consisted of five tasks as outlined below:

Task 1. Coal Preparation/Utilization

- Subtask 1.1 Particle Size Control
- Subtask 1.2 Physical Separations
- Subtask 1.3 Surface-Based Separation Process
- Subtask 1.4 Dry Processing
- Subtask 1.5 Stabilization of Coal-Water Mixtures

Task 2. Emissions Reduction

- Subtask 2.1 SO₂ Reduction
- Subtask 2.2 NO_x Reduction
- Subtask 2.3 Study VOC and Trace Element Production, Reduction, and Capture

Task 3. Economic Evaluation

- Subtask 3.1 Cost and Market Penetration of Coal-Based Fuel Technologies
- Subtask 3.2 Selection of Incentives for Commercialization of the Coal-Using Technology
- Subtask 3.3 Community Sensitivity to Coal Fuel Usage
- Subtask 3.4 Regional Economic Impacts of New Coal Utilization Technologies
- Subtask 3.5 Economic Analysis of the Defense Department's Fuel Mix
- Subtask 3.6 Constructing a National Energy Portfolio which Minimizes Energy Price Shock Effects
- Subtask 3.7 Proposed Research on the Coal Markets and their Impact on Coal-Based Fuel Technologies
- Subtask 3.8 Integrate the Analysis

Task 4. Evaluation of Deeply-Cleaned Coals as Boiler Fuels

- Subtask 4.1 Modify MCWM Preparation Circuit
- Subtask 4.2 Fuels Characterization
- Subtask 4.3 Pilot-Scale Combustion Tests
- Subtask 4.4 Demonstration-Scale Combustion Tests

Task 5. Final Report

1.5 References

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2.0 COAL PREPARATION/UTILIZATION

Coal preparation/cleaning necessarily plays a substantial role in all aspects of coal utilization, large-scale and small-scale, including specialized applications such as coal-water mixture firing and cofiring with waste materials. Realistic applications of post-combustion treatments (e.g. flue-gas desulfurization) will continue to depend on pre-combustion processing (coal cleaning) to maintain consistent fuel quality and to minimize the load on the post-combustion system.

Research conducted under Phases I and II of this project revealed a number of specific areas where continued and/or more focused effort was required in order to develop more effective, more reliable coal processing systems. The specific objectives of the Phase III research on coal preparation were selection of specific coal-cleaning options and their associated ancillary operations and the integration of processing/cleaning operations for overall system optimization. As in the earlier phases of the project, emphasis was placed on fine-coal processing for the production of high-quality, micronized coal for dry coal and coal-water mixture applications.

2.1 Particle Size Control

2.1.1 Conventional Ball Milling

Tumbling ball mills represent an important component of fine-grinding systems used for the preparation of coal-water mixtures. The Bond Method has been used traditionally for the selection of ball mills for industrial applications. It is simple and has been quite successful for predicting required mill sizes. However, it is not suitable for optimization of mill or circuit conditions. Also, it does not predict the complete product size distribution, which is important for subsequent processes such as coal-water mixture preparation. Recently, commercial ball mill simulators have evolved based on the concepts of grinding kinetic theory using specific rates of breakage (S values) and primary breakage distributions (B values). Hence, one can examine the performance of various full-scale grinding circuits.

The method employed here follows that developed by Austin et al. (1983), which involves scale-up of the specific rates of breakage determined for the material in a laboratory test batch mill, allowing for the effects of the variables associated with milling. For ball milling, these variables typically include mill rotational speed, ball filling, powder filling, mill diameter, ball diameter, etc. Extensive work has been done to investigate the effect of these variables, and the methodology to incorporate these factors into simulations is well described by Austin et al. (1983).

The functional forms used to characterize particle breakage during grinding can be summarized as follows:

Effect of particle size (x_i)

$$S(x_i) = a \left(\frac{x_0}{x_i} \right)^a \frac{1}{1 + (x_i / \mu)^\Lambda} \quad (2.1.1)$$

where $S(x_i)$ = specific rate of breakage for particle of size x_i

a = specific rate of breakage at x_0

α = an exponential constant, characteristic of the material being ground

μ, Λ = characteristic constants for describing the decrease in the specific rates of breakage as the particles become too large to be broken effectively.

Effect of ball diameter (d) and mill diameter (D)

The parameters a and μ in Equation 2.1.1 both depend on ball size and mill diameter. Specifically:

$$a \propto 1/d \text{ and } a \propto D^{0.5}$$

$$\mu \propto d^2 \text{ and } \mu \propto D^{0.2}$$

Effect of rotational speed

$$S(x_i) \propto (\phi_c - 0.1) \frac{1}{1 + \exp[15.7(\phi_c - 0.94)]}$$

where ϕ_c = the fraction of critical speed (defined as the rotational speed at which the balls begin to centrifuge on the mill case and cease to tumble).

Effects of ball and powder loading

$$S(x_i) \propto \left[\frac{1}{1 + 6.6J^{2.3}} \right] \exp(-cU)$$

where J = the fraction of the mill volume filled by the ball bed at rest

U = the fraction of the interstices of the ball bed filled with powder (at rest)

The S values determined for the material in a laboratory-scale, batch mill can be used to simulate large-scale mills via the scale-up relations given above. On the other hand, the B values are typically independent of the breaking size and milling environment so that scale-up is not usually necessary for the B values.

For a retention grinding device such as the ball mill, the length of time that material spends in the mill controls the final degree of breakage. Moreover, some fraction of the feed stays in the mill for a short time, other fractions for a longer time, etc. Therefore, estimation of the residence time distribution (RTD) is required. It has often been shown that the experimental RTD can be matched against theoretical expressions derived for a number of fully mixed reactors in series. The one-large, two-small, fully mixed reactors in series model has been found to be appropriate for ball mills.

A grinding circuit may consist of one or more mills and classifiers. Industrial grinding circuits can be categorized into the following five types.

- a) open circuit
- b) normal closed circuit
- c) open circuit with scalped feed

- d) reversed closed circuit
- e) combined closed circuit

These circuits are illustrated in Figure 2.1.1. The combined closed circuit can be used to represent the other circuits by assigning the appropriate classification parameters for the pre-classifier and the post-classifiers.

Classification is the splitting of a stream of particles into two streams, with separation by size (and/or density) e.g., one stream with mainly large particles, the other with fine particles. The classifier performance is characterized by a set of numbers, one for each size interval, which describes how each size is split between the two product streams. The selectivity value, s_i for size i is defined as the fraction of size i in the feed which is sent to the coarse stream. A plot of the selectivity values vs. size is called a size selectivity curve. This curve generally has an S shape and can often be described by a log-logistic function. Using this functional form, the complete set of size selectivity values can be calculated using three characteristic parameters: the sharpness index, cut size (d_{50}) and apparent by-pass. Finally, the different streams in the complete circuit can be described by the mass balance equations, a mill model, and two classifier models.

Simulations were performed for the two most common circuits: open and normal closed circuit. Capacities were obtained for three different diameter (3, 4, 5 meter) mills, giving a circuit product passing through the one-point specification of 80% finer than 200 mesh (74 μm). All mills were operated with a slurry density of 70 wt% coal and at 70% of critical speed. Ball loading was set at 30% filling with makeup of 50.8 mm balls. The residence time distribution was assumed to fit the one-large/two-small fully mixed reactor model in the ratios 0.5, 0.25, 0.25, respectively. The performance of the classifier was assumed to be constant with the characteristic parameters of sharpness index of 0.7 and by-pass of 0.3. The d_{50} values were changed to give results over a range of circulating loads. The coal feed was assumed to fit the Rosin-Rammler distribution with size modulus $k_r = 9.525$ mm and $m = 1$. The breakage parameters (a , α , γ , ϕ and β) measured in a laboratory batch mill at 70% solids are shown in Table 2.1.1.

a) open

b) normal

c) open with scalped feed

d) reverse closed

e) combined closed

C: classifier

Figure 2.1.1 VARIOUS GRINDING CIRCUIT CONFIGURATIONS USED IN INDUSTRY

Table 2.1.1 Breakage Parameters for the Coals Tested at 70% Solids by Weight

Coal Seam	a	α	ϕ	γ	β
Taggart	0.85	1.37	0.52	0.44	5.0
Pittsburgh	1.90	1.33	0.45	0.58	5.0
Lower Kittanning	3.47	1.56	0.45	.36	5.0

The results of the simulations are given in Table 2.1.2. As expected, the capacity is higher for easily ground coals (high HGI). For a given coal, capacity increases with the mill diameter and is proportional to $D^{3.5}$ as shown in Figure 2.1.2. This results from the fact that for the same grinding environment, the capacity is proportional to the mill volume, which is LD^2 (for the same L/D ratio, it is D^3). Since the S values increase with the mill diameter by $D^{0.5}$ (see above), the overall effect of the mill diameter is that the capacity is proportional to $D^{3.5}$. If the circuit is closed with a classifier, the capacity increases substantially by minimizing the over-grinding of material that is already fine enough. Also, the closed circuit produces a narrower size distribution; i.e., produces less unnecessarily fine material. For a given classifier setting, increasing the feed rate coarsens the mill product, thereby causing the classifier to send more material to recycle. As a result, the circulation ratio C (ratio of the amount of recycle to that of the product) increases, and the circuit product becomes coarser. However, there is only one feed rate that can meet a particular one-point product specification. By changing the classifier setting (d_{50}), it is possible to obtain the one-point match for a range of feed rates and circulation ratios. Figure 2.1.3 shows the results. A lower value of d_{50} gives a higher circulation ratio and a steeper size distribution. Also, the production rate increases as the circulation ratio increases, as seen in Figure 2.1.4. At low circulation ratios, the product is sufficiently fine that the classifier has no effect, so the production rate is the same as that of the open circuit. On the other hand, high recycle leads

Table 2.1.2 Comparison of Circuit Capacity for Open and Closed Circuit Systems

Open Circuit									
Mill Diameter (meters)	Taggart (HGI = 47)			Pittsburgh (HGI = 56)			Lower Kittanning (HGI = 68)		
	C	TPH	d ₅₀ (μm)	C	TPH	d ₅₀ (μm)	C	TPH	d ₅₀ (μm)
3	-	8.45	-	-	14.82	-	-	22.58	-
4	-	21.13	-	-	39.37	-	-	61.66	-
5	-	44.02	-	-	81.12	-	-	131.25	-
Closed Circuit									
3	0.53	244	9.14	0.50	244	15.01	0.54	211	25.02
	1.03	137	13.13	0.97	133	21.47	1.03	133	32.6
	1.44	120	15.04	1.47	113	23.03	1.50	119	34.19
	2.10	112	17.09	1.97	104	23.17	1.98	108	33.23
	2.47	105	16.65	-	-	-	-	-	-
4	0.48	305	22.39	0.5	244	38.91	0.48	264	61.51
	0.98	141	35.06	0.99	133	59.16	1.00	141	92.41
	1.52	119	41.86	1.46	115	63.48	1.54	117	92.32
	1.91	112	45.23	2.06	106	64.38	1.74	111	91.02
	3.33	99	44.55	-	-	-	-	-	-
5	0.47	305	43.9	0.50	244	81.90	0.49	264	139.91
	0.97	139	71.31	0.95	132	118.42	0.96	141	191.20
	1.56	122	92.66	1.54	115	136.86	1.47	117	193.67
	2.04	112	97.73	2.07	105	135.48	1.97	111	197.62
	2.5	103	95.10	-	-	-	-	-	-

Figure 2.1.2 VARIATION OF BALL MILL CAPACITY WITH MILL DIAMETER

to increased overfilling, causing inefficient grinding. Thus, an optimum circulation ratio exists. Generally, the optimum circulation ratio is a function of the fineness of the final product. In this case, the optimum circulation ratio is found to be in the range of 1.5-2.0. At this optimum circulation ratio, the increase in capacity given by closing the circuit is also in the range of 1.5-2.0. The energy required for grinding to a one-point product specification is less for the closed circuit than for open-circuit grinding.

Figure 2.1.3 COMPARISON OF PRODUCT SIZE DISTRIBUTIONS RESULTING FROM OPEN AND CLOSED-CIRCUIT GRINDING

2.1.2 Stirred-Media Milling

The approach described above can also be used to predict grinding behavior in a stirred ball mill. However, the engineering and mill scale-up procedures are not as well established. It has often been suggested that the median size of the product is related to the energy input, independent of the mill size and operating conditions (Herbst and Sepulveda, 1978; Stehr and Schwedes, 1983). This suggests that the relationship of power draw to the mill dimensions and operating variables can be used to predict the product fineness. However, in recent studies (Mankosa et al., 1986, 1989; Gao and Forssberg, 1993; Zheng et al., 1995), it was reported that the energy efficiency is affected by operating conditions such as media density and size, slurry concentration, and agitation speed.

Figure 2.1.4 EFFECT OF CIRCULATION RATIO ON CAPACITY IN CLOSED-CIRCUIT GRINDING

It is clear that there are numerous apparent inconsistencies in the literature on the energy-efficiency of stirred-media milling. For the most part, these arise from interactions among many material, equipment, and operating variables. It is generally agreed, however, that there is a clear link between energy input and product fineness. Also it is known that operating conditions play an important role. Powder and media loading and the media to particle size ratio seem to be primary factors in determining energy efficiency.

Review of the literature indicates that the energy input alone can not correctly predict mill performance, i.e., size reduction. Nonetheless, the power requirement for stirred ball milling is important and has been the subject of many investigations, especially aimed at

scale-up. The standard treatment for the power, P , necessary to drive a stirrer in a Newtonian liquid operating in the turbulent regime starts with (Rushton et al., 1950):

$$P = N_p N^3 D^5 \rho \quad (2.1.2)$$

where N is the rotational speed, D is the diameter of the stirrer, and ρ is the density of the liquid. N_p is the power number, which depends on flow conditions and system geometry. In stirred ball milling, the mill contains the solids being ground, the liquid (in wet grinding) and, more importantly, the grinding media. It is extremely unlikely, therefore, that the behavior would be Newtonian. Furthermore, the density in Equation 2.1.2 is no longer clearly defined. Since there is a considerable difference in the densities of the liquid and the media, there is a tendency for segregation of the media in the mill. At a low rotational speed, the media tend to settle, leading to an inhomogeneous distribution of the media. On the other hand, at higher rotational speeds, a vortex is formed and the media are concentrated at the wall due to centrifugal forces. Therefore, the dependency of the power draw on the rotational speed can vary significantly. Also, the movement of the media is highly dependent upon the media loading. It might be expected that the movement of the media would be more rapid at high loading, since a higher portion of media could be in direct contact with stirrers and among themselves. Therefore, it is very difficult to analyze the mill power theoretically, and it is often necessary to resort to empirical equations based on experiment (Herbst and Sepulveda, 1978; Gao et al., 1996, Zheng et al., 1995). Unfortunately, such equations may be valid only for the particular design of the system. The values of the parameters, for the range of rotational speeds of interest, may depend on media loading, size, and density and slurry concentration, density, and viscosity.

Breakage of particles occurs as a result of applied stress, which originates from energy input to the grinding device. However, as shown above, the effectiveness of the breakage action does not depend on the level of energy input alone. The nature of the energy input also plays an important role. Kwade et al. (1996) have introduced the concept of energy intensity as a critical factor in determining whether or not stress application actually leads to breakage. At low stress intensities, several stressing events may be required to cause

fracture. As the stress intensity increases, a critical value is reached above which further increase does not lead to additional breakage. In any grinding process, particles are exposed to a distribution of stress intensities. The effectiveness of the process depends on the fraction of stressing events for which the stress intensity exceeds this critical level. On the other hand, excessive stress intensities, well above the critical value, lead to reduced grinding efficiency, i.e. increased specific energy.

Breakage rates are determined by the stress frequency, modified by the stress intensity, i.e. by the frequency of *adequate* stress applications. Relationships between the stress intensity and frequency and the operating conditions in stirred-media milling have been proposed (Blecher and Schwedes, 1996; Becker et al., 1997). The stress frequency f_s clearly depends on mill geometry and agitator speed and a general relationship can be expected with the form:

$$f_s \propto N \left(\frac{d_d}{d_b} \right)^2 \quad (2.1.3)$$

where N is the rotational speed and d_d and d_b are the respective diameters of the grinding chamber and the grinding beads. The stress intensity I_s can be expected to depend on agitation speed, grinding media size and density, and on the elasticity of the media and the material being ground. The latter determines the effectiveness of energy transfer between media and particles. The following specific relationship has been proposed (Kwade 1998; Kwade et al., 1996):

$$I_s = d_b^3 \rho_b v_t^2 \left(1 + \frac{Y_s}{Y_b} \right)^{-1} \quad (2.1.4)$$

where ρ_b is the grinding bead density, v_t is the agitator tip speed, and Y_s and Y_b are the respective elastic moduli for the solid and the grinding media.

The specific grinding energy depends on the product of the stress frequency and the stress intensity, both of which vary with grinding conditions. Becker et al. (1997) showed that, for a fixed energy input, there is an optimum stress intensity that provides the most size reduction. By varying the energy input using different combinations of speed, media size,

and media density, they demonstrated that the optimum stress intensity decreased with increased energy input. The optimum intensity presumably reflects the critical intensity for fracture discussed above. Equation 2.1.4 indicates how it varies with media size. The dependence on the size of the particles being broken is not included in Equation 2.1.4. However, it has been demonstrated that breakage rates in stirred media mills follow very similar patterns to those observed in conventional ball mills (Miller et al., 2000; Cho and Hogg, 1995; Mankosa et al., 1986; Stehr et al., 1987) and relationships of the general form of Equation 2.1.1 seem to be applicable.

Based on such data as are available and on indirect evidence from the work of Kwade et al., (1996, 1998) and others (Herbst and Sepulveda, 1978; Gao and Forssberg, 1993, 1995; Mankosa et al., 1986, 1989; and Rajamani and Bourgeois, 1990), it appears that the trade-off between stress intensity and stress frequency becomes especially critical at ultrafine sizes. Changing media size can lead to enhanced breakage in one range of sizes but simultaneously to reduced rates for slightly coarser material. Staged stirred-media milling using progressively finer media may be a useful option for improving efficiency in ultrafine grinding. The use of a distribution of media sizes, as is the common practice in tumbling ball mills, should also be appropriate.

It was noted above that breakage distributions in conventional ball milling are often relatively independent of the size being broken and the milling environment. Similar behavior has been observed in stirred-media milling for relatively coarse grinding. However, it has been found that the distributions tend to become progressively narrower as the size being broken is reduced into the micron range (Strazisar and Runovc, 1996; Cho and Hogg, 1995). This effect has been attributed to the approach to a “grind limit” (Cho et al., 1996).

2.1.3 Attrition Milling

Attrition milling operates on the same basic principle as stirred-media milling, but the size reduction is accomplished by using the coal particles themselves as the media. However, there is a major difference between the stirred-media mills and attrition mills. First, the power input is substantially less for attrition milling, since the coal is lighter than

the media and hence, less energy is required to agitate the slurry. More importantly, the breakage process in attrition milling is totally different. Large coal particles act as grinding media for smaller particles, giving breakage mechanisms similar to those obtained by normal grinding media. But they also break themselves by particle-particle contact. The breakage in any grinding mill is generally achieved by three different mechanisms - fracture, chipping and abrasion. Fracture refers to complete disintegration of a particle by massive impact, chipping involves cutting off corners, while abrasion results from the wearing of surfaces by a rubbing action. The fragments produced by fracture are distributed over a broad range of sizes. On the other hand, chipping and abrasion lead to production of fine material and a residual "core" close to the original particle size. Therefore, the shape of the product size distribution is greatly influenced by the type of the breakage mechanism. In attrition milling, chipping and abrasion will play bigger roles and the resulting products are likely to be composed of two main populations - coarse (remaining cores) and fines (chipped or abraded material). This is quite appropriate for a coal-water mixture feedstock. Therefore, attrition milling potentially offers a great advantage for coal-water mixture preparation over other grinding procedures from both economic and engineering viewpoints.

Tests were conducted to explore such a potential advantage using a vertical-type, stirred-media mill. The mill was constructed with a stainless steel cylinder 3 inches wide and 4 inches high. The stirrer shaft was fitted with 3-perforated disks and was rotated by means of a 2-Hp, variable speed motor. A torque sensor, mounted on the drive shaft between the motor and the mill, was used to provide an indication of the power input to the mill. Tests were conducted to determine the breakage characteristics of large particles alone, and of small particles of various sizes when ground together with large ("media") particles. Since the rheological properties are likely to play an important role in determining the breakage mechanisms, the effect of overall solids content was examined. Some results are shown in Figures 2.1.5 and 2.1.6.

As noted above, the coarse particles serve as grinding media for the finer material but also produce fine particles by self-breakage. Consequently, in a continuous operation, the amount of "media" in the mill decreases with time and must be replenished on a regular

Figure 2.1.5 DISAPPEARANCE PLOTS FOR COARSE (16x20 US mesh) COAL "MEDIA" IN ATTRITION GRINDING

basis. Feed disappearance plots for 16x20 US mesh fractions from three different coals are given in Figure 2.1.5. It can be seen that disappearance rates are higher for softer coals and higher solids concentrations. The self-breakage process is clearly non-linear and is rapid initially but then slows down substantially. Around 80% of the original feed remained in its original size fraction after one hour of grinding.

The size distributions of the "media" particles after grinding for various times are shown in Figure 2.1.6. It can be seen that the curves are almost flat in the fine size ranges, and are translated upward by an increase in the amount of -400 mesh material. The initially

Figure 2.1.6 SIZE DISTRIBUTION OF THE PRODUCTS OF “MEDIA” ATTRITION

high disappearance rates shown in Figure 2.1.5 together with the product size distributions (Figure 2.1.6) are consistent with a chipping/abrasion model for the process. Specifically, it appears that depletion of media-size material occurs through two mechanisms:

- Large-scale chipping of a fraction f of the feed material
- Slow abrasion of the remaining portion.

The material subject to large-scale chipping or fracture presumably includes highly irregular and/or friable particles. All particles in the mill are expected to undergo abrasion.

A model for the combined chipping/abrasion process can be developed by assuming that the chipping process is approximately first-order, leading to an exponential decay in the mass of these “breakable” particles, while the abrasion process involves a slow reduction in the mass of a fixed *number* of “non-breakable” particles. Studies of the abrasion of individual rock particles in conventional autogenous grinding (Tangsrirongkul 1993) have indicated that a first-order abrasion law, referred to as the Davis Wear Law (Davis 1919) may be appropriate for this process. The Davis Law can be expressed as

$$\frac{dx}{dt} = -kx \quad (2.1.5)$$

where x is the equivalent-sphere particle diameter and k is a first-order rate constant.

Formulation of the combined chipping/abrasion model leads to the following expression for the mass $W(t)$ of material remaining in the coarse “media” size after time t :

$$W(t) = \frac{[1 - f(1 - e^{-St})][1 - R(1 + kt)]}{(1 + kt)^3(1 - R)} \quad (2.1.6)$$

where S is the specific rate of breakage by chipping and R is the ratio of the upper and lower boundaries of the original size interval. In this formulation, it is assumed that the initial distribution of sizes within the original interval was uniform (rectangular).

The application of this model to media-size disappearance in the attrition of 16×20 US mesh Taggart seam coal is illustrated in Figure 2.1.7. It is clear that the model can describe such data well. In this case, it appears that increasing the solids loading in the mill increases the fraction susceptible to large-scale chipping, possibly by eliminating liquid films between particles which could provide a lubricating (or damping) action. The abrasion rate seems to be relatively unaffected by solids loading.

Some limited experimentation on the breakage of smaller particles by the coarse “media” has also been carried out, as part of a related but separately funded project. Some typical results, in this case for quartz particles, are shown in Figure 2.1.8. The finest feed (37-52 μm) seems to exhibit “normal” behavior consistent with massive fracture – feed

Model Parameter values used are:

Concentration	f	$S(\text{min}^{-1})$	$k(\text{min}^{-1})$	W_o
45	0.04	0.1	1.1×10^{-4}	0.95
62	0.07	0.3	2.3×10^{-4}	0.95

Figure 2.1.7 APPLICATION OF CHIPPING/ABRASION MODEL TO ATTRITION OF NOMINAL 16x20 MESH TAGGART SEAM COAL AT DIFFERENT SOLIDS CONCENTRATIONS

Figure 2.1.8 PRODUCT SIZE DISTRIBUTIONS IN THE ATTRITION MILLING OF QUARTZ USING QUARTZ MEDIA (6x8 mesh).
 (a) 270x400 mesh feed, (b) 20x30 mesh feed, (c) 6x8 mesh media

particles are broken down progressively to finer and finer sizes (Figure 2.1.8 a). Breakdown of the coarser feeds especially the media particles themselves (Figure 2.1.8 c) shows quite different behavior. The bimodal character, indicated by the broad, flat "shelf" seen over the intermediate size ranges in the cumulative distributions, follows the same pattern as that seen for coal (Figure 2.1.6). For intermediate sizes (Figure 2.1.8 b), there appears to be a larger contribution from massive fracture as indicated by the progressive broadening of the coarse mode of the distribution.

2.1.4 Grinding Circuit Simulation

It is clear from the investigations carried out in Phases I and II of this project that single-stage grinding in a conventional ball mill does not produce size distributions with the form required for stable coal-water mixtures. Stirred ball milling was also found to be unable to produce the appropriate size distribution, although the product was much finer than that produced in the ball mill. It was concluded, therefore, that a circuit arrangement involving two-stage grinding was required to produce slurry with the proper size consist. One such circuit arrangement is shown in Figure 2.1.9.

In the first stage, crushed coal (e.g., -12 mesh), is introduced into a ball mill, along with water and dispersant where it is ground in closed circuit. The product from the first-stage grinding is then classified. The fine stream is further ground in a closed stirred ball mill circuit to produce an ultrafine fraction. This product is then combined with the coarser size fraction, producing a bimodal size distribution.

The ideal mixing ratio of coarse and fine particles for maximum packing is 60:40. Therefore, the production rates of the coarse and fine size fractions must be controlled accordingly. At the same time, the maximum size of the coarse fraction must be controlled for proper combustion, while the size ratio of the coarse and the fine fractions are kept sufficiently large (at least ten) for maximum solid loading and low viscosity.

It can be expected that the mean residence time, τ , for each mill determines the characteristics of the final circuit product. The mill residence time (and hence the extent of size reduction) is controlled by material flow rates and mill operating conditions. However,

S_1 = Post-classifier for the first stage
 S_2 = Pre-classifier for the second stage
 S_3 = Post-classifier for the second stage

Figure 2.1.9 TWO-STAGE GRINDING CIRCUIT FOR PRODUCING COAL-WATER MIXTURES WITH A BIMODAL SIZE DISTRIBUTION

the operating variables are not completely independent of the design variables. Let Ω be the ratio of the hold-up of the first stage (W_1) to the hold-up of the second stage (W_2), that is, $\Omega = W_1/W_2$. For a simple two-stage closed circuit at steady state, the mass flow rate into the first stage (G_1) equals the mass flow rate into the second stage (G_2). Then, since the actual flow rate to the first mill is $F_1 = G_1(1+C_1)$, that to the second mill is $F_2 = G_2(1+C_2)$, $\tau_1 = W_1/F_1$ and $\tau_2 = W_2/F_2$:

$$\Omega = \frac{\tau_1(1+C_1)G_1}{\tau_2(1+C_2)G_2} \quad (2.1.7)$$

where C_1 and C_2 are the circulation ratios of the first and the second stages, respectively.

For a fixed set of breakage parameters, mill residence time distributions, (RTD), and classifier settings, a particular value of τ_1 fixes the entire system for a given value of Ω . Therefore, the design criteria involving two mills in series are the mean residence time for the first mill, the classifier settings and the relative sizes of the two mills. Designing such a grinding circuit requires a simulator to evaluate the performance of a two mill grinding circuit under various operating conditions.

For this purpose, a general two-mill grinding circuit simulator has been developed at Penn State. The complete circuit scheme is shown in Figure 2.1.10. In its original form, the circuit consists of two ball mills in normal closed circuit, with a circuit pre-classifier and a pre-classifier before the second mill. Additionally, the circuit has three dividers so that the streams of the first and second mills can be appropriately connected. By assigning the appropriate values to the classifiers and dividers, the general two-stage circuit can be reduced to any desired circuit varying from a simple single-stage open circuit to a very complicated two-stage circuit. Evaluating the performance of a grinding circuit is achieved by combining the mill model and the classifier model via the appropriate mass balance equations. A detailed description of the computational procedure can be found elsewhere (Yildirim 1987).

Figure 2.1.10 GENERAL TWO-MILL GRINDING CIRCUIT

Before the simulator can be used to evaluate the circuit shown in Figure 2.1.9, several modifications are needed. The scale-up procedure involving the mill sizes and operating conditions for ball milling is well established as described in the previous report. However, the design procedure for stirred ball milling has not been completely established. It is often assumed that the specific energy is constant between the test and large mill for the same reduction ratios. The power draw of a stirred ball mill is a function of mill diameter and operating conditions. If operating conditions are selected for large mills to consume the same specific energy as the test mill, the simulation can be performed using the same breakage parameters as those determined by the laboratory milling tests.

Classification is done extensively throughout the coal industry. In most cases, separations are made at sizes coarser than about 40 μm using hydrocyclones. Such operations are common and existing data are available to model the size separations. However, classification of particles down to 10 μm and perhaps finer will be required for coal-water mixture production. Although separations in this size range are made in various mineral applications, they are not generally done in the coal industry. Therefore, selected tests were conducted using hydrocyclones to develop the appropriate parameters for the classification devices used in the circuit simulator.

The make-up feed size distribution (i.e., crushed product) was chosen as 80% passing 700 μm with a Gaudin-Schuhmann slope of 0.9. The parameters for the breakage rates and the breakage distribution were obtained through laboratory testing. The complete list of these parameters is given in Table 2.1.1. The ball mill was assumed to have a residence time distribution equivalent to one-large/two-small fully mixed tanks-in-series with relative hold-up ratios of 0.5, 0.25, 0.25, whereas the stirred ball mill was assumed to be fully mixed. The simulations were conducted for a 1 meter diameter by 1.5 meter long wet-overflow ball mill operating at 70 wt. % solids, a fractional ball load of 0.35, rotational speed 70% of critical, with a make-up ball size of 31.75 mm (1.25 inches). The relative size of the stirred ball mill to the ball mill was 1:10, operating at 40 wt. % solids, 0.6 fractional ball loading and a total charge volume (balls and slurry) of 0.9.

The classifiers were assumed to be hydrocyclones. The characteristic parameters for the selectivity values at various cut sizes were determined by laboratory testing and are shown in Table 2.1.3. It was assumed that an appropriate operation could be obtained to give suitable underflow and overflow slurry densities and flow rates for the desired cut sizes.

Table 2.1.3 Characteristic Parameters for the Cyclone Size Selectivity at Different Cut Sizes

Cyclone Diameter, inches	Overflow Diameter, inches	Underflow Diameter, inches	Cut Size, μm	Sharpness Index	Apparent By-pass	Flow Rate, liter/min.
1	0.28	0.06	25.2	0.56	0.04	16.2
1	0.28	0.06	17.3	0.44	0.06	16.2
1	0.28	0.13	14.2	0.43	0.06	16.5
2	0.38	0.13	39.8	0.51	0.08	37.7
2	0.38	0.19	26.7	0.36	0.12	35.6
2	0.38	0.38	14.8	0.33	0.50	35.9
10	0.38	3.00	148.0	0.53	0.20	1128.0

The predicted size distributions of the various streams for the grinding of the Taggart seam coal are shown in Figure 2.1.11, assuming the following conditions:

- residence time for the first mill $\tau_1 = 4$ minutes
- cut sizes for the first, second and third cyclones at 150 μm , 40 μm , and 17 μm , respectively.

The required mean residence time for the second mill was determined to be 3.8 minutes.

It can be seen that the final product size distribution, resulting from combining the first and second stage products, has a bimodal shape. The bimodal shape, though, is less well defined than in the batch grinding system due to the imperfect separation by the cyclones and the back-mixing occurring in the continuous mill. The median sizes for each stage product are 52 μm and 3.3 μm , respectively, and the size ratio of the coarse and finer sizes is 16. The proportion of the coarser and the finer size fraction is about 60:40, which is an ideal composition. The size ratio could be increased by reducing the cut size for the second-stage post-classifier. This would produce a more distinct bimodal shape for the final product size

Figure 2.1.11 PRODUCT SIZE DISTRIBUTIONS IN A TWO-STAGE GRINDING CIRCUIT

distribution. However, this would also result in a decrease in the production rate for the second stage grinding, and hence, require a larger stirred ball mill.

2.1.5 Integrated Grinding/Cleaning Circuit

An important potential advantage of coal-water slurry technology over conventional coal firing lies in the opportunity for advanced coal cleaning, which is offered by the associated fine grinding but is not offset by the need for expensive product dewatering.

Integration of a multi-stage grinding operation with a physical cleaning process such as froth flotation offers several opportunities for cost-effective deep cleaning of coal. With many coals, for example, it is possible to separate a clean fraction at a relatively coarse size; grinding of the remainder of the coal can then be used to provide for sulfur and mineral matter liberation and to produce the finer fractions needed for slurry preparation. The use of an integrated grinding/cleaning process offers considerable potential for cost saving by exploiting the liberation of mineral matter obtained by fine grinding to permit extensive deep-cleaning of the coal. By integrating the grinding and cleaning steps, it should be possible to grind the raw (or pre-cleaned) feed coal only to the extent needed to achieve liberation of the sulfur and ash-forming minerals. Further, multi-stage grinding of the coal would then be carried out so as to produce the optimum size distribution for slurry formulation.

An example of an integrated grinding/cleaning process is illustrated in Figure 2.1.12.

The system consists of two basic elements:

- A liberation/cleaning step in which grinding is performed only to the extent needed for liberation of the mineral matter.
- A secondary (2-stage) grinding/classification step in which additional size reduction is carried out as needed to produce the desired final product size distribution.

The first stage consists of a grinding circuit, which is closed by means of a separator (e.g., froth flotation device) rather than by the size classifier employed in conventional grinding circuits. The flotation system could be a bank of conventional cells set up to produce a low-ash, clean-coal product, a high-ash middlings product, and a refuse stream. The middlings product is recycled to the grinding to obtain further liberation. Alternatively, the same objective could be achieved using a two-stage column-flotation set-up. By maintaining a high circulation ratio, it should be possible to obtain a high degree of liberation while minimizing overgrinding of either the combustible or refuse materials.

The second, finish-grinding step consists of a closed-circuit grinding system set to maintain the desired top-size in the product slurry. A portion of the output from this circuit is further ground to form the fine component of the final, bimodal slurry.

Figure 2.1.12 EXAMPLE OF AN INTEGRATED CLEANING/SIZE-REDUCTION PROCESS FOR MICRONIZED COAL-WATER MIXTURE (MCWM) PRODUCTION

Additional, intermediate stages of classification can be included in order to produce a true bimodal distribution. Alternative circuit configurations such as that shown in Figure 2.1.11 could also be used for the finish-grinding stage. A further advantage of the two-stage, finish-grinding step is that relatively simple dewatering procedures can be applied to the coarse product for control of the final slurry density.

2.2 Physical Separations

Separations of fine coal based on differences in particle size and density were performed. The size separations were carried out using a high-speed, solid-bowl centrifuge under various operating conditions. Centrifuge performance was evaluated in terms of the size selectivity curves and corresponding performance parameters -- cut size, sharpness index, and apparent bypass.

Density separations were carried out using both dense medium and magnetic fluid systems. The dense-medium separations were also made with the solid-bowl centrifuge using magnetite and water as the dense medium. The magnetic fluid separations were carried out in both batch and continuous laboratory separators using water-based magnetic fluids. These systems offer the advantage of changing the relative density of the fluid (medium) simply by varying the intensity of the magnetic field.

In order to extend the size range of fine coal cleaning further, testing was carried out using an integrated centrifugal/flotation system. In this case, fine coal was cleaned with a dense-medium cyclone followed by froth flotation. The addition of the flotation step not only increased the recovery and quality of the finer coal, but it also improved the quality of the recovered dense medium.

2.2.1 Fine Coal Classification

One of the key elements in fine coal cleaning and slurry preparation circuits is size classification, which involves the preparation of coal into the appropriate size ranges. Industrially, hydrocyclones are commonly used to classify coal down to approximately 75 μm (200 mesh). Although the size range can be extended down to about 10 μm , this requires the use of small diameter (and low capacity) cyclones. An alternative would be to use a continuous centrifuge. For this study, a continuous, solid-bowl centrifuge (Sharples Model P660) was used to classify -150 μm (-100 mesh) Upper Freeport seam coal. A schematic of the unit is given in Figure 2.2.1. The test variables included bowl and scroll speeds, and pond depth (weir setting). For all tests, the unit was operated at a constant volumetric slurry feed rate of 11.4 lpm (3 gpm) and a solids concentration of 10% coal (by weight).

For each test, simultaneous samples of the weir overflow (fine product) and scroll discharge (coarse product) were taken. Each sample was analyzed using a Microtrac X-100 particle size analyzer. From the size distribution data, the circulation ratio, which is defined as the solids mass flow rate of the coarse stream, T , divided by the solids mass flow rate of the fine stream, Q , was estimated by (Klimpel 1980):

Figure 2.2.1 **SCHEMATIC OF A SOLID-BOWL CENTRIFUGE**
(courtesy of Alfa Centrifuges)

$$C = \frac{\sum_{i=1}^{n-1} [p_i - q_i]}{\sum_{i=1}^{n-1} [t_i - p_i]} \quad (2.2.1)$$

where p_i , t_i , and q_i are the fractions of material in size interval i for the feed, coarse, and fine streams, respectively, and n is the number of size intervals. The circulation ratio is related to the yield to the fine stream, Y_Q , by

$$Y_Q = \frac{1}{1 + C} \quad (2.2.2)$$

The yield to the coarse stream, Y_T , is then simply $Y_T = 1 - Y_Q$.

Knowing the product size distributions and the circulation ratio, the size selectivity values were calculated by

$$s_i = \frac{t_i C}{p'_i (1 + C)} \quad (2.2.3)$$

where p'_i is the reconstituted feed for size interval i given by

$$p'_i = \frac{q_i + t_i C}{1 + C} \quad (2.2.4)$$

The size selectivity values were corrected for apparent bypass, a , by

$$c_i = \frac{s_i - a}{1 - a} \quad (2.2.5)$$

where c_i is the corrected selectivity (classification) value for size interval i . In order to determine the characteristic performance parameters -- the cut size, d_{50} , the sharpness index, κ , and the apparent bypass, a -- the classification values were fitted to the log-logistic function given by

$$c_i = \frac{1}{1 + \left(\frac{x_i}{d_{50}}\right)^{-\lambda}} \quad (2.2.6)$$

where $\lambda = \exp[-2.1972/\kappa]$. The d_{50} value is the size that corresponds to $c_i=0.5$, and $\kappa = d_{25}/d_{75}$, where d_{25} and d_{75} are the sizes that correspond to c_i values of 0.25 and 0.75, respectively. The sharpness index varies from zero for no separation to one for a perfect separation. The parameters d_{50} , κ , and a were determined using a nonlinear optimization routine contained in a spreadsheet program.

The first series of tests was conducted using the minimum weir setting, which gave the shallowest pond depth (i.e., setting 1 = shallowest pool, setting 4 = deepest pool). Table 2.2.1 summarizes the operating conditions and results for this series of tests (series a). Figure 2.2.2 shows the reconstituted feed and product size distribution curves for centrifuge tests 1a and 2a. As can be seen, the product size distribution of the fine stream for test 2a was finer than that for test 1a. For example, 85.8% of the material in the fine stream was less than 11 μm for test 2a compared to 53.2% for test 1a. Figure 2.2.3 shows the size selectivity values and fitted curves obtained for tests 1a-4a, while the parameters are given in Table 2.2.1. The cut size was the coarsest at the highest scroll speed, while the cut sizes at the other conditions were comparable. At the highest scroll speed, the speed differential between the bowl and the scroll was the smallest, resulting in a slower discharge of solids.

Table 2.2.1 Summary of the Operating Conditions and Test Results for the Solid-Bowl Centrifuge (Feed Rate = 11.4 L/min, Solids Concentration = 10% by weight)

Test	Bowl Speed, rpm	Scroll Speed, rpm	Weir Setting	Yield, %		Characteristic Parameters		
				Coarse Product	Fine Product	d ₅₀ , μm	κ	a
1a	4600	4125	1	77.4	22.6	18.9	0.36	0.23
2a	4600	3650	1	91.0	9.0	8.1	0.47	0.28
3a	4600	2700	1	92.1	7.9	4.4	0.35	0.21
4a	4600	800	1	88.9	11.1	6.4	0.44	0.00
5a	3750	1850	1	88.9	11.1	7.7	0.56	0.14
1b	4600	4125	2	86.6	13.4	9.9	0.47	0.39
2b	4600	3650	2	91.3	8.7	7.4	0.63	0.33
3b	4600	2700	2	83.3	16.7	7.9	0.52	0.19
4b	4600	800	2	88.5	11.5	6.7	0.44	0.26
5b	3750	1850	2	88.7	11.3	7.2	0.52	0.14
1c	4600	4125	3	88.8	11.2	7.4	0.52	0.41
2c	4600	3650	3	85.4	14.6	7.0	0.59	0.22
3c	4600	2700	3	80.4	19.6	10.5	0.54	0.29
4c	4600	800	3	93.5	6.5	7.2	0.50	0.65
5c	3750	1850	3	90.0	10.0	8.6	0.53	0.48

Figure 2.2.2 FEED AND PRODUCT SIZE DISTRIBUTIONS FOR CENTRIFUGE TESTS 1a AND 2a WHEN SEPARATING -100 MESH UPPER FREEPORT SEAM COAL
 (test 1a: filled symbols; test 2a: open symbols)

Figure 2.2.3 EFFECT OF SCROLL SPEED ON THE SIZE SELECTIVITY CURVES IN THE SOLID-BOWL CENTRIFUGE FOR WEIR SETTING 1

Hence, coarser solids reported to the weir (fines) discharge, leading to a coarser separation. This is also reflected in the lower coarse product yield for test 1a.

The next series of classification tests was carried out with a new weir setting. This resulted in the formation of a slightly deeper pool than used in the previous tests. The test variables were the same as in the previous series. Table 2.2.1 summarizes the operating conditions for these (series b) tests.

Figure 2.2.4 shows the size selectivity curves for tests 1b-4b. The corresponding parameters are given in Table 2.2.1. Overall, the cut size of the centrifuge ranged from about 7 to 10 μm , while the sharpness index ranged from 0.44 to 0.63. The primary impact of the bowl and scroll speeds at this weir setting was on the apparent bypass values, which ranged from 0.14 to 0.39 (Table 2.2.1).

The last series of tests (series c) was conducted using weir setting 3, which gave the second deepest pool setting. The operating conditions and results are also summarized Table 2.2.1. Figure 2.2.5 shows the reconstituted feed size distribution and product size distribution curves for centrifuge tests 1c and 2c. As can be seen, the product size distribution of the weir discharge stream for test 2c was slightly finer than that for test 1c, while the scroll discharge was coarser.

Figure 2.2.6 shows the corresponding size selectivity curves for tests 1c-4c. Overall, the cut sizes of the centrifuge again ranged from about 7 to 10 μm . The sharpness indexes were very similar and ranged from 0.50 to 0.59. As was seen at the other weir settings, the bowl and scroll speeds impacted the apparent bypass values, which ranged from 0.22 to 0.65 (Table 2.2.1). High bypass values indicate that a substantial fraction of the feed coal is reporting to the coarse stream, resulting in a finer size distribution of the coarse material. On the other hand, the bypass does not affect the size distribution of the fine stream.

Because the sharpness indexes were similar for these tests, it was possible to reduce the size selectivity curves to a single reduced classification curve. The reduced classification curve was generated by plotting the classification values versus the reduced size (defined as the size divided by the cut size). The reduced curve for centrifuge tests 1c-4c is shown in Figure 2.2.7.

**Figure 2.2.4 EFFECT OF SCROLL SPEED ON THE SIZE SELECTIVITY CURVES
IN THE SOLID-BOWL CENTRIFUGE FOR WEIR SETTING 2**

Figure 2.2.5 FEED AND PRODUCT SIZE DISTRIBUTIONS FOR CENTRIFUGE TESTS 1c AND 2c WHEN SEPARATING -100 MESH UPPER FREEPORT SEAM COAL
(test 1c: filled symbols; test 2c: open symbols)

Figure 2.2.6 EFFECT OF SCROLL SPEED ON THE SIZE SELECTIVITY CURVES IN THE SOLID-BOWL CENTRIFUGE FOR WEIR SETTING 3

Figure 2.2.7 REDUCED CLASSIFICATION CURVE FOR CENTRIFUGE TESTS 1c-4c

Except for test 1c, the values fall on a single curve. In such cases, this curve can be used to predict separator performance. For example, by selecting an appropriate cut size and corresponding apparent bypass, it is possible to generate the size selectivity values, which in turn can be used to calculate the product size distribution for a given feed size distribution.

2.2.2 Dense-Medium Separation

Dense-medium separations were also carried out using the Sharples P660 solid-bowl centrifuge. Testing was conducted to evaluate the effect of operating conditions such as bowl and scroll speeds, weir height, relative density of the medium, medium-to-coal ratio, and feed rate on separation efficiency. All tests were run using -100 mesh Upper Freeport seam coal.

For each test, samples of the clean coal (weir overflow) and refuse (scroll discharge) streams were taken. Wet screening at 25 μm (500 mesh) was used to remove the -500 mesh material. The yield of 100x500 mesh clean coal was calculated by ash balance as

$$Y_c = \frac{A_P - A_F}{A_F - A_R} \quad (2.2.7)$$

where A_F , A_P , and A_R are the ash contents of the feed, clean coal, and refuse streams, respectively.

For selected samples, a float-sink analysis was done at several relative densities on both centrifuge products. Using these data, along with the clean coal yield, the partition values for each relative density interval were calculated by

$$K_j = \frac{p_j Y_c}{p_j Y_c + r_j (1 - Y_c)} \quad (2.2.8)$$

where p_j and r_j are the fractions of clean coal and refuse in relative density interval j , respectively. These values give the fraction of feed in a given relative density interval that reports to the clean coal. In order to determine the characteristic performance parameters -- relative density of separation, ρ_{50} , and probable error, E_p -- the partition values were fitted to the logistic function given by

$$K_j = \frac{1}{1 + \exp[(1.0986/E_p)(\rho_j - \rho_{50})]} \quad (2.2.9)$$

where ρ_j is the mean relative density of the interval. The parameters were determined using a nonlinear optimization routine.

The operating conditions and results for the first set of tests are summarized in Table 2.2.2. All tests were carried out at a relative density of the medium of 1.3. Figure 2.2.8 shows the centrifuge results for various bowl speeds at a constant scroll speed (tests 1-3), along with the washability curve for the 100x500 mesh Upper Freeport seam coal. The results approached the washability curve for the lower two bowl speeds but deviated at the higher speed. No separation was obtained at a bowl speed of 2,400 rpm (Table 1.2.2). Tests 2, 4 and 5 show the effect of changing the weir height and in turn, pond depth for constant bowl and scroll speeds. In both cases, the results indicate that the centrifuge achieved very good separations at relative densities above about 1.5, even though the relative density of the medium was 1.3. This was not unexpected because the offset (i.e., the difference between ρ_{50} and the relative density of the medium) is typically greater than 0.05 relative density units for dense-medium cyclone separations.

Table 2.2.2 Summary of the Operating Conditions and Test Results for the Solid-Bowl Centrifuge (Medium Relative Density = 1.3, Feed Rate = 11.3 L/min (3 gpm), Medium-to-Coal Ratio = 10:1, Scroll Speed = 300 rpm)

Test	Bowl Speed, rpm	Weir Setting	Overflow Ash, %	Underflow Ash, %	Overflow Yield, %
1	800	2	6.9	63.0	91.7
2	1200	2	5.4	56.1	87.9
3	1500	2	4.9	27.5	70.7
4	2400	2	-	11.5	0.0
5	800	1	5.6	46.0	85.3
6	800	3	5.8	37.2	81.7

Figure 2.2.8 EFFECT OF BOWL SPEED ON THE GRADE-YIELD VALUES FOR THE SOLID-BOWL CENTRIFUGE AT A CONSTANT SCROLL SPEED OF 300 RPM AND A MEDIUM RELATIVE DENSITY OF 1.30 FOR THE 100x500 MESH UPPER FREEPORT SEAM COAL

Another set of tests was carried out with the centrifuge to evaluate the effects of bowl and scroll speeds, weir height, relative density of the medium, medium-to-coal ratio, and feed rate. Minus 100 mesh Upper Freeport seam coal was again used for all tests. Table 2.2.3 summarizes the operating conditions and results for these tests.

Figure 2.2.9 shows the partition curves obtained for several bowl speeds, with all other conditions constant. As seen, the best separation, based on the lowest probable error, was obtained for the middle bowl speed of 800 rpm (test 16). The relative density of

Table 2.2.3 Summary of the Operating Conditions and Test Results for the Solid-Bowl Centrifuge.

Test	Feed Rate,	Relative Density	Medium-to-Coal	Weir Height*	Bowl Speed,	Scroll Speed, rpm	Yield, %	Clean Coal Ash,	Refuse Ash,
1	11.4	1.3	20:1	4	500	300	56.4	8.4	15.5
2	11.4	1.3	20:1	4	700	300	69.4	6.1	23.7
3	11.4	1.3	20:1	4	800	300	66.2	6.0	22.2
4	11.4	1.3	20:1	4	900	300	62.9	6.8	19.5
5	11.4	1.3	20:1	4	1000	300	43.1	6.1	15.6
6	11.4	1.3	20:1	4	800	500	64.8	8.2	17.5
7	11.4	1.3	20:1	4	800	700	55.8	8.9	14.8
8	9.5	1.3	20:1	4	700	300	24.7	6.4	13.2
9	13.2	1.3	20:1	4	700	300	37.6	6.6	14.5
10	15.1	1.3	20:1	4	700	300	53.0	7.2	16.4
11	11.4	1.3	15:1	4	800	300	38.8	6.5	14.7
12	11.4	1.3	10:1	4	800	300	39.9	6.3	14.3
13	11.4	1.5	20:1	4	800	300	70.7	6.7	23.0
14	11.4	1.6	20:1	4	1200	300	45.9	5.8	16.4
15	11.4	1.3	20:1	1	600	300	92.3	7.7	57.7
16	11.4	1.3	20:1	1	800	300	89.4	6.0	58.1
17	11.4	1.3	20:1	1	3200	300	90.2	6.4	58.7
18	11.4	1.2	20:1	1	800	300	80.3	5.3	37.3
19	11.4	1.25	20:1	1	800	300	82.0	6.0	37.0
20	11.4	1.15	20:1	1	800	300	0.0	-	-
21	11.4	1.3	10:1	1	800	300	85.3	5.6	46.0
22	11.4	1.25	10:1	1	800	300	85.0	5.8	44.0

1 = shallowest pond depth; 4 = deepest pond depth

Figure 2.2.9 EFFECT OF BOWL SPEED ON THE PARTITION CURVES WHEN SEPARATING 100x150 MESH UPPER FREEPORT SEAM COAL IN THE SOLID-BOWL CENTRIFUGE
 (600 rpm (test 15): $E_r=0.13$, $r_{50}=2.04$; 800 rpm (test 16): $E_r=0.05$, $r_{50}=1.80$; 3200 rpm (test 17): $E_r=0.12$, $r_{50}=1.94$)

separation was also the lowest at this condition. At the highest speed (3,200 rpm), both the E_p and ρ_{50} values were higher (test 17). Because there was a lack of coal in this relative density range, the yield and corresponding ash values were similar. However, for the bowl speed of 600 rpm (test 15), the higher relative density of separation produced a slightly higher yield (92% versus 90%) but with a much higher clean coal ash content (7.7% versus 6.4%).

Overall, the ρ_{50} values were much higher than the relative density of the medium (1.3). As noted previously, a positive offset is typical for dense-medium separations. For these tests, the higher offset can be attributed, in part, to the use of the minimum weir setting, which gave the shallowest pond depth. Hence the probability that a larger amount of material would exit to the overflow stream was higher, which was reflected in the higher ρ_{50} values.

2.2.3 Magnetic Fluid Separation

Density separations were also carried out using magnetic fluids as the dense medium. The separations were carried out using one of several laboratory separation cells constructed out of Plexiglas to fit within a Franz magnetic separator. New pole pieces were fabricated for the Franz unit to accommodate the separators, while providing the necessary magnetic field gradient (Fofana 1997).

Initially, coal separations were made using a batch unit under static conditions to calibrate the device (Figure 2.2.10). Narrow size (i.e., 6x8 mesh) and narrow relative density (e.g., 1.28x1.30) fractions of bituminous coal were prepared by heavy liquid fractionation. One of the coal fractions was placed in the magnetic fluid of a known concentration. The magnetic field intensity was increased until the material just began to float. This was repeated for ten relative density fractions to establish a calibration curve (Figure 2.2.11). Subsequent tests, which were carried out using purchased plastic tracer particles of a fixed density, confirmed the calibration curve.

After calibration, a series of separations was made using a sample of the nominal -1/4" Upper Freeport seam coal, which was screened to produce 12x14 mesh and 28x100 mesh size fractions. Separations were made at 1.3, 1.4, 1.5, 1.6, and 2.0 relative densities using the magnetic fluid separator under static conditions. A parallel approach

Figure 2.2.10 SCHEMATIC OF THE BATCH MAGNETIC FLUID SEPARATOR AND MAGNET POLE PIECES

was used whereby a separate sample was used for each test, thereby providing cumulative float and sink values at each relative density.

Prior to immersion in the magnetic fluid, the coal was wetted using a small amount of 0.1% Aerosol OT solution followed by rinsing with the magnetic fluid. After setting the relative density, the coal was added to the separator and allowed to stand for one hour. At that time, the products were separated by first collecting the float material followed by removal of the sink material. The samples were washed with water to remove any remaining liquid, dried, and weighed to determine the yield at each relative density. An ash analysis was done on each fractionated sample.

The resulting values, along with those obtained using heavy-liquid fractionation, are plotted in Figures 2.2.12 and 2.2.13 for the 12x14 mesh and 28x100 mesh size fractions, respectively. Ideally, both liquids should produce the theoretically best

Figure 2.2.11 VARIATION OF THE RELATIVE DENSITY OF THE FLUID AS A FUNCTION OF THE APPLIED CURRENT FOR THE MAGNETIC-FLUID SEPARATOR

separation possible at a given relative density, i.e., the washability curve for the given size fraction. Therefore, the differences in the results can be attributed, in part, to the differences in the properties of the two liquids. That is, the magnetic fluid was water based, while the heavy liquids were organic. However, another explanation could be related to the slight variation in the magnetic field at different levels between the magnet poles. Because the resulting fluid density depends on the magnetic field gradient, any variation in the gradient would produce a non-uniform fluid density, leading to the differences in the float-sink results (Fofana 1997).

Figure 2.2.12 WASHABILITY CURVES FOR THE 12x14 MESH UPPER FREEPORT SEAM COAL
(open symbols: organic liquids; closed symbols: magnetic fluid)

Figure 2.2.13 WASHABILITY CURVES FOR THE 28x100 MESH UPPER FREEPORT SEAM COAL
(open symbols: organic liquids; closed symbols: magnetic fluid)

Separations were also carried out using a continuous flow-through separator. The separator and test circuit are shown in Figure 2.2.14. The coal was added to tank 1 containing the magnetic fluid and stirred for several minutes to ensure adequate mixing. The solids concentration was fixed at 2% solids by weight. The magnet was turned on and set to the appropriate setting to produce the desired relative density. The peristaltic pump was started and set to the desired speed. The flow rate and hence, the retention time in the device, were set by adjusting the pump speed. The valve was set to tank 1 so that only the magnetic fluid was pumped through the separator. This was done to fill the unit prior to feeding the coal slurry. Care was taken to dispel any air bubbles from the separator. Once filled, the valve was set to pump the coal slurry from tank 2. The float

Figure 2.2.14 CONTINUOUS MAGNETIC FLUID SEPARATOR TEST CIRCUIT

and sink products were split at the discharge end and collected continuously over the test period.

Upon completion of the test, the valve was set to pump the recirculated fluid from tank 1 to flush out the remaining solids in the system. After all of the float material was removed, the pump and the magnet were shut off. Any solids remaining on the bottom of the separator were combined with the sink material. The test procedure was repeated using a new pump speed or new magnet (i.e., relative density) setting.

The products were subjected to float-sink separation at several relative densities using organic liquids. Using these data, along with the clean coal yield, the partition values were calculated for each of the test conditions. These values were then fitted to the logistic function to determine the relative density of separation and the probable error (Table 2.2.4).

Table 2.2.4 Summary of Operating Conditions and Results when Separating 28x32 mesh Upper Freeport Seam Coal in the Magnetic Fluid Separator

Flow Rate, L/min	Medium Relative Density	Mean Retention Time, s	Relative Density of Separation	Probable Error
3.3	1.3	2.7	1.35	0.30
2.4	1.3	3.6	1.37	0.22
1.7	1.3	5.2	1.39	0.11
1.7	1.4	5.2	1.47	0.12
1.7	1.6	5.2	1.78	0.17

Tests were run with 28x32 mesh Upper Freeport seam coal at a relative density of 1.3 for flow rates of 1.7, 2.4, and 3.3 Lpm. This gave retention times in the device of 5.2, 3.6, and 2.7 seconds, respectively. Figure 2.2.15 shows the partition curves obtained as a function of flow rate. As expected, the curves steepened with an increase in retention time, indicating a sharper separation. A horizontal line that corresponds to the volumetric

Figure 2.2.15 PARTITION CURVES FOR THE MAGNETIC FLUID SEPARATOR AS A FUNCTION OF FLOW RATE AT A MEDIUM RELATIVE DENSITY OF 1.3 FOR THE 28x32 MESH UPPER FREEPORT SEAM COAL

split of slurry to the overflow stream would be indicative of splitting rather than separation. For these tests, the volumetric split was approximately 0.5 for all flow rates.

Additional tests were run at relative densities of 1.4 and 1.6 at a flow rate of 1.7 Lpm. Figure 2.2.16 shows the variation of the partition curves as a function of the relative density of the medium. As expected, the curves shifted to the right with an increase in the medium density. For densities of 1.3 and 1.4, the probable errors were similar (Table 2.2.4) but increased at 1.6. Figure 2.2.17 compares the grade-yield curves

Figure 2.2.16

PARTITION CURVES FOR THE MAGNETIC FLUID SEPARATOR AS A FUNCTION OF MEDIUM RELATIVE DENSITY AT A FLOW RATE OF 1.7 L/min FOR THE 28x32 MESH UPPER FREEPORT SEAM COAL

Figure 2.2.17 EFFECT OF FLOW RATE AND MEDIUM DENSITY ON THE GRADE-YIELD VALUES FOR THE 28x32 MESH UPPER FREEPORT SEAM COAL

for each flow rate and medium density to the washability curve for the 28x32 mesh coal. As can be seen, magnetic fluid separation approached the theoretical curve at the lower flow rates (longer retention times) and for the higher densities of separation.

Baseline testing of a centrifugal magnetic fluid separator was also performed. In this case, a Dynawhirpool-type separator, which was designed to fit between the pole pieces of the Frantz electromagnet, was constructed out of Plexiglas. A schematic of the flow circuit is shown in Figure 2.2.18. This device was selected over a hydrocyclone for several reasons. Because a Dynawhirpool consists only of a cylindrical portion, the unit could fit between the existing magnet poles. Also, the design lends itself to staging whereby the clean coal leaving the first device could be injected directly into a second separator. The second unit would be operated at a lower density by using a weaker magnetic field. Thus a two-stage separation, producing a high quality clean coal, a lower quality middling, and a refuse fraction, could be obtained using the same fluid. This approach is similar to the Tri-Flo separator, which uses a magnetite or ferrosilicon-based system as opposed to a magnetic fluid.

Unlike a hydrocyclone, two feed ports are used in a Dynawhirpool separator. The feed coal enters at the top of the device, along with a portion of the medium (Figure 2.2.18). The remaining medium enters the separator tangentially, near the bottom of the device. This imparts the desired flow pattern in the separator to produce the necessary centrifugal force for separation. The refuse is driven outward and is carried up the separator wall where it exits tangentially near the top of the device. The clean coal is driven inward and exits at the bottom of the device.

For the test circuit, the coal was fed to the unit from a vibrating feeder. The magnetic fluid was pumped from a sump and was split prior to entering the separator. Approximately 15% of the fluid was combined with the feed coal, with the remaining fluid entering the medium inlet. The concentration of the feed coal was around 2% solids by weight. The product streams were directed into filter bags where the coal was collected. The filtrate was recovered and recirculated through the circuit, providing a continuous operation.

Initially, a 1.30 relative density zinc bromide solution was used as the separating medium to validate the operation of the device. Upper Freeport seam coal (28x32 mesh)

Figure 2.2.18 DYNAWHIRLPOOL MAGNETIC FLUID SEPARATOR TEST CIRCUIT

was passed through the device and the products were collected. Approximately 60% of the coal reported to the overflow (clean coal) stream. In comparison, no coal reported to the overflow stream when only water was used as the separating medium, as was expected.

Tests were then conducted using the magnetic fluid as the separating medium, along with the 28x32 mesh Upper Freeport seam coal. In all cases, no separation

occurred and all the coal reported to the refuse stream as when only water was used. The same results were obtained even when the concentrated magnetic fluid was used at maximum field strength. The addition of the centrifugal acceleration negated the buoyancy effect created by the magnetic fluid. To counteract this effect, a stronger magnetic field would be needed, which could not be obtained with the existing Frantz electromagnet (Fofana 1997).

2.2.4 Centrifugal/Flotation Separations

Dense-medium processes are capable of separating coal over a wide range of relative densities. Because the separations can be controlled with great precision (e.g., within ± 0.005 relative density units), dense-medium separation is the choice for processing difficult-to-clean coals such as those containing large amounts of near-density material. Moreover, it is the only technique capable of making separations at low relative densities (e.g., < 1.4). Such separations are often required to produce low-ash, low-sulfur (e.g., compliance) coal.

Dense-medium separation is used extensively for cleaning coarse coal (e.g., $+9.5$ mm) and intermediate coal down to ~ 0.6 mm (28 mesh). In some cases, the bottom size has been extended to about 100 mesh (Kempnich et al., 1993). The processing of finer size fractions of coal has not always been successful because of poor cyclone performance and inefficient dense-medium recovery circuits. This has generally led to higher than expected processing costs.

Testing of a micronized-magnetite cycloning process to treat sizes as fine as ~ 0.038 mm (400 mesh) was carried out by the U.S. Department of Energy (Klima et al., 1990) and Custom Coals International (Kindig 1992). In these cases, the separations were made using micronized (e.g., $-10 \mu\text{m}$) magnetite as opposed to the more typical coarser grades of magnetite (e.g., $-50 \mu\text{m}$). The fineness of the magnetite allows higher operating pressures and in turn greater separating forces (i.e., increased g's) to be employed, which are required to process ultrafine coal particles. However, the use of micronized magnetite adds additional problems in regards to dense-medium recovery, including the need to remove the non-magnetic material from the medium prior to reuse in the system.

Currently, froth flotation is the only technique that has found widespread commercial use for treating coal finer than about 100 mesh. This process is very effective in rejecting ash-forming mineral matter, but it often lacks the selectivity in rejecting pyritic sulfur, an important consideration when attempting to produce low-sulfur coal.

One method of overcoming the limitations of dense-medium cycloning and froth flotation is through the use of an integrated circuit. Miller (1989) proposed a combined dense-medium/flotation process for beneficiating coal while increasing magnetite recovery. His process takes advantage of the natural hydrophobicity of coal compared to the hydrophilic magnetite. Because flotation is generally more effective than dense-medium processes for cleaning fine coal size fractions, it can be used to recover fine coal that is misplaced to the medium, reducing the load on the dense-medium recovery circuit. In one study using several grades of commercial magnetite, it was found that increased coal cleaning occurred in addition to magnetite recovery (Polat et al., 1993).

Subsequent work was carried out using this integrated approach to fine coal cleaning but using ultrafine magnetite as part of the dense medium (Polat et al., 1995). Samples were taken from the overflow and underflow streams of a 25 mm diameter cyclone processing -100 mesh Upper Freeport seam coal. The samples were analyzed for ash and total sulfur content. Flotation separations were then carried out on a separate sample of each cyclone product. For comparison, additional tests were carried out using froth flotation to clean the -100 mesh Upper Freeport seam coal directly, i.e., no dense-medium separation.

Figure 2.2.19 shows the yield and ash contents when separating the -100 mesh Upper Freeport seam coal using cycloning only, flotation only, and the combined cycloning-flotation process. The washability curve, which gives the theoretical best separation, is also included. It can be seen that dense-medium cycloning is more effective than froth flotation alone for producing a specified ash content at a given clean coal yield. However, the cyclone results were still lower than the washability results obtained by float-sink testing. The results from the integrated process were much better than cycloning alone, with ash contents similar to those obtained for the float-sink tests. The improved performance of the integrated circuit can be attributed to the rejection by

flotation of the very fine mineral matter particles, which invariably report to the clean coal stream of the cyclone.

Figure 2.2.20 shows the yield and corresponding sulfur contents for the same separations. This figure demonstrates the sulfur reduction capability of the cyclone compared to froth flotation. In fact, the cyclone results are comparable to the washability curve. Integration does not provide any additional benefit in terms of sulfur rejection.

Figure 2.2.19 CLEAN COAL YIELD - ASH CURVES WHEN SEPARATING -100 MESH UPPER FREEPORT SEAM COAL USING DENSE-MEDIUM CYCLONING (DMC), FROTH FLOTATION (FF), AND THE INTEGRATED PROCESS (INT)
The washability curve is also included (Wash).

Figure 2.2.20 CLEAN COAL YIELD - SULFUR CURVES WHEN SEPARATING -100 MESH UPPER FREEPORT SEAM COAL USING DENSE-MEDIUM CYCLONING (DMC), FROTH FLOTATION (FF), AND THE INTEGRATED PROCESS (INT) (The washability curve is also included (Wash))

2.3 Surface Based Separation Processes

2.3.1 Continuous Froth Flotation

The continuous flotation tests were conducted on a pilot-plant scale froth flotation circuit. The continuous circuit has a capacity of about 8 gallons/min. The flotation circuit was designed for cleaning coal in two stages. The coal to be fed to the flotation circuit was ground in a stirred ball mill. The ball mill had a total capacity of 30 gallons and a working capacity of about 20 gallons. The operating conditions were selected so as to produce a size distribution of about 100% passing 28 mesh as shown in Figure 2.3.1.

Figure 2.3.1 SIZE DISTRIBUTION OF THE FEED

During grinding, the slurry contained about 20-35% solids and the product was transferred to the 500 gallon capacity head tank as shown in Figure 2.3.2. The slurry from the head tank was pumped by a Moyno pump at a flow rate that was set depending on the circuit requirements and the percent solids in the feed slurry. A flow meter was connected after the pump to monitor the slurry flow rate from the head tank.

The rougher circuit consisted of:

1. 20 gallon feed tank with mixer
2. Four, WEMCO cells of 1.4 cubic feet capacity each
3. 5 gallon frother tank

4. 1 gallon collector tank with mixer
5. Quinn froth pump, with a flow meter
6. Sampling system

The coal-water slurry at 20-35% solids from the head tank was fed into the feed tank where it was diluted with water to make slurry containing about 5% solids. The amount of water added to the feed tank depended on the solids content of the feed slurry from the head tank. A rotameter was used in series with the water line to measure the flow rate. The water flow rate was adjusted so as to obtain the desired total flow rate.

The MIBC frother, and the dodecane collector were each mixed with water in a ratio of 1 to 400, in the frother and collector tanks, respectively. The frother and the collector were pumped into the feed tank where the feed was diluted as required. During some tests the collector was fed to cells 3 and 4 only, to avoid overloading the froth in cells 1 and 2, and to increase recovery in cells 3 and 4.

Figure 2.3.2 ROUGHER-CLEANER FLOTATION CIRCUIT USED IN CONTINUOUS FLOTATION TESTS

The coal-water slurry containing 5% solids, the frother and collector were fed to the series of four cells. The impellers in the cells were rotated at a fixed speed of about 1,000 rpm, and the aeration was set to the maximum value. The froth containing clean coal flowed above the weir of the cells and was removed into the launders on each side of the cells by rotating paddles. The wash water washed down the froth in the launders. The wash water flow rate was about 1 gpm, and was monitored by a rotameter.

The froth product from the circuit was collected in plastic storage drums, which was subsequently taken for filtration. If an additional cleaner circuit was used, the froth was allowed to flow down into a froth sump, from where it was pumped to the cleaner circuit. The tailings from the last cell in the rougher circuit were sent to a sump.

The cleaner circuit consisted of:

1. 20 gallon feed tank with mixer
2. Two flotation cells of 1.4 cubic feet capacity each
3. 5 gallon frother tank
4. 1 gallon collector tank with mixer

The froth product from the rougher circuit was pumped to the cleaner circuit feed tank where it was diluted to about 5% solids with water. The water flow rate into the feed tank was adjusted to get the desired total flow rate, and was monitored using a rotameter. The frother and collector could be fed into the feed tank as desired. The froth product from the cleaner cells was collected in drums, and the tailings from the last cell flowed down into a sump. If required, the cleaner tailings could be recycled to the rougher cells.

The product from the cleaner circuit was filtered on a vacuum filter. The tailings in the trench sump, now containing very low percent solids, were pumped using the trench pump to an overhead tank, where it was mixed with a solution of dissolved calcium chloride, and the solids were allowed to flocculate and settle overnight. The clear water on the top was decanted, and the slurry with settled solids was pumped to the vacuum filtration system.

For the Lower Kittanning seam coal tested in this part of the investigation, the ash analysis of samples taken from the rougher and cleaner circuits during preliminary tests, showed that the froth product from the rougher circuit satisfied the ash content

requirement for subsequent combustion tests. Therefore most of the tests with this coal sample were performed using only the rougher circuit consisting of the bank of four cells.

To evaluate the performance of the circuit, slurry samples were taken from the froth product from all four cells, the feed and the tailings streams. To maximize the probability of representative sampling, whole streams were diverted into sampling containers and samples were taken simultaneously from every stream. The samples collected were subsequently filtered, dried, weighed and analyzed for ash content. Two sets of samples were collected for each test to determine whether the circuit was operating at steady state, as assumed. The samples collected from one test were sieved using 28, 48, 100, 200, and 400 mesh sieves to obtain the size distribution of each stream. From the weight and ash analysis of various samples, the flotation response of the different size fractions could be determined.

2.3.1.1 Results and Discussion

Lag Time

During laboratory tests, it is often observed that the froth flows over the weir of the cell only after a finite time has elapsed after the air is turned on. This time lapse is called 'lag time'. The time when the froth starts flowing over the weir is often set as flotation time 'zero'. The mean residence time of the slurry in a cell or a bank of cells is calculated by the relation

$$\tau = \frac{V}{Q} \quad (2.3.1)$$

where V is the effective volume and Q is the flowrate of the slurry. Here, it is assumed that the slurry acts as a liquid, and the liquid mean residence time would represent that of the whole slurry. Assuming that about 15% of the cell volume is occupied by air, the mean residence time for a 1.4 ft³ cell at a flow rate of 6 gpm, is calculated to be 1.485 minutes.

The cumulative weight percent of material remaining in the each cell is plotted in Figure 2.3.3. It is assumed that the mean residence time is the same for each cell, and the total residence time of the material in any cell is the cell number multiplied by the mean residence time for one cell. The curves joining the recovery points at different conditions

do not merge to 100 percent at time 0 or cell 0. The lag time is different, ranging from 0.2 cell to 0.5 cell, under different operating conditions. Agar et al. (1980) proposed a flotation rate model that takes the induction time into account. To calculate the lag time a modified form of the classical first order was used.

$$R = R_{\infty}(1 - e^{-k(t + t_L)}) \quad (2.3.2)$$

Figure 2.3.3 ESTIMATION OF INDUCTION TIME

where t_L is the lag time. The induction time for the rougher circuit was estimated to be 0.4 times the mean residence time of a cell, which is 0.594 minutes.

Effect of Frother Concentration

Figures 2.3.4 and 2.3.5 show the effect of frother concentration on the recovery from each cell in the bank of flotation cells. Figure 2.3.4 shows the cumulative recovery at each cell. As expected, the cumulative recovery curves shifted to higher values as the cell number increased. In each curve the cumulative recovery attains a maximum in the frother concentration range of 0.4 kg/T to 0.45 kg/T. Figure 2.3.5 shows the fractional recovery from each cell as a function of frother concentration. In the first two cells the recovery increased with increase in frother concentration in the range 0.2 kg/T to 0.4 kg/T. The recovery remained fairly constant in the range 0.4 to 0.45 kg/T, and then decreased as the frother concentration was increased further. In cell # 3, the recovery increased when the frother concentration was increased from 0.2 kg/T to 0.3 kg/T, and then remained relatively constant. In cell # 4, the fractional recovery decreased slightly when the frother concentration was increased at low concentrations. The recovery remained relatively constant at higher frother concentrations.

The flotation recovery from each cell is the maximum at a frother concentration in the range 0.4 kg/T to 0.45 kg/T. It is concluded that, under the used operating parameters this frother concentration range is optimum for the flotation of Lower Kittanning seam coal in the pilot plant flotation circuit.

Kinetics

The flotation results are plotted in Figure 2.3.6 after incorporating the lag time. Since the data fit the first order model only in the initial stages, the data up to cell 3 were used to estimate the rate constants from Figure 2.3.7. The flotation rate increased with frother concentration in the range 0.2 kg/T to 0.4 kg/T. The rate remained constant in range 0.4 kg/T to 0.45 kg/T. The flotation rate decreased with further increase in frother concentration. The flotation rate constants are given in Table 2.3.1.

Figure 2.3.4 CUMULATIVE RECOVERY FROM EACH CELL AT DIFFERENT FROTHER CONCENTRATIONS

Figure 2.3.5 FRACTIONAL RECOVERY FROM EACH CELL AT DIFFERENT FROTHER CONCENTRATIONS

Figure 2.3.6 KINETICS PLOTS AT DIFFERENT FROTHER CONCENTRATIONS

Figure 2.3.7 ESTIMATION OF FLOTATION RATE CONSTANTS

Table 2.3.1 Flotation Rate Constants at Different Frother Concentrations

Frother concentration (kg/T of coal)	Flotation rate constant (per min)
0.20	0.1148
0.30	0.1355
0.40	0.2503
0.45	0.2599
0.50	0.1910

Effect of Collector Concentration

The kinetics plots for tests performed with a fixed frother concentration of 0.3 kg/T and varying the collector concentration from 0 to 0.1 kg/T are shown in Figure 2.3.8. Most of the coal that floated was recovered in the first two cells. Though the flotation rate increased with the addition of collector, the froth in cells 3 and 4 became very stable resulting in a stagnant froth. As a result, the particles in the froth fell back into the cell, and the recovery decreased.

Independent control of the pulp level in each cell and a mechanism to selectively add frother and collector to cells 3 and 4 could improve recovery. The flotation rate constants for different collector concentrations are shown in Table 2.3.2.

Flotation Rates of Various Size Fractions

To determine the behavior of coal in various size fractions, a series of tests was conducted. The samples obtained from the test conducted with a frother concentration of 0.5 kg/T were sieved to obtain the size distribution in each stream. Each size fraction was treated independently and compared with respect to flotation response. The kinetics plots for the different size fractions are shown in Figure 2.3.9. The flotation rate and final recovery increased as the particle size decreased from 600 μm to 37 μm .

The flotation rate and recovery were lower for the -400 mesh size fraction while the 200 x 400 mesh fraction had the highest flotation rate and final recovery.

Figure 2.3.8 KINETICS PLOTS AT DIFFERENT COLLECTOR CONCENTRATIONS (frother conc.=0.3 kg/T)

Table 2.3.2. Flotation Rate Constants at Different Collector Concentrations

Collector concentration (kg/T) (frother : 0.3 kg/T)	Flotation rate constant (min ⁻¹)
0	0.1152
0.1	0.2520

The flotation rate regression lines are plotted in Figure 2.3.10. The flotation rate was calculated from the slope for each size fraction. The flotation rate constants for different size fractions are given in Table 2.3.3.

Figure 2.3.9 KINETICS PLOTS FOR DIFFERENT SIZE FRACTIONS FOR TESTS CONDUCTED AT A FROTHER CONCENTRATION OF 0.3 kg/T

The flotation rate constants for different size fractions, plotted in Figure 2.3.11, show that the flotation rate of coal particles increases as the size decreases to about 400 mesh, and decreases with further decrease in size.

The cumulative recovery of each size fraction from the four cells is shown in Figure 2.3.12. Maximum recovery was obtained for the 200 x 400 mesh fraction. The expected increase in the coarser material in cells 3 and 4 was not observed, probably because the level in these cells could not be controlled independent of cells 1 and 2. The froth mobility in cells 1 and 2 was very different from that in cells 3 and 4. It might be possible to increase recovery in cells 3 and 4 by using a frother which is more effective for coarse particles.

Figure 2.3.10 ESTIMATION OF KINETICS RATE CONSTANTS FOR EACH SIZE FRACTION

Table 2.3.3 Flotation Rate Constants for Different Size Fractions

Size fraction (mesh)	Flotation rate constant (min^{-1})
28 x 48	0.0557
48 x 100	0.1829
100 x 200	0.4079
200 x 400	0.5801
-400	0.2595

Figure 2.3.11 VARIATION OF FLOTATION RATE CONSTANT WITH PARTICLE SIZE

Evaluation of Flotation System Performance

The rate constants for the batch, batch with no paddling (batch*), and pilot-plant tests are given in Table 2.3.4 for different frother concentrations. In these tests the batch data refer to a typical kinetics test in which froth is removed continuously by manual paddling. A comparison of the rate constants for the three cases is given in Figure 2.3.13, in which the data are plotted as a function of frother concentration. The plant data refer to the flotation kinetics tests obtained in the pilot plant circuit consisting of four cells. The data marked as batch* refer to the tests in which no paddling was employed and the froth was allowed to flow on its own accord. The results confirm that froth mobility plays a very significant role in determining the overall kinetics in a continuous operation.

Figure 2.3.12 CUMULATIVE RECOVERY OF EACH SIZE FRACTION FROM EACH CELL

Table 2.3.4 Flotation Rate Constants and Froth Factors for the Batch, Batch with no Paddling (Batch*), and Pilot Plant Tests

Frother conc. (kg/T)	Rate constants (min^{-1})			Froth Factor ($k_{\text{lab}}/k_{\text{plant}}$)	
	Batch	Batch*	Plant	Batch*	Plant
0.2	1.7517	0.5995	0.1149	2.92	15.25
0.3	1.9205	0.7198	0.1356	2.67	14.61
0.4	2.1737	0.9704	0.2505	2.24	8.68
0.45	2.0228	0.9068	0.2601	2.23	7.78
0.5	2.0528	0.8636	0.1912	2.38	10.74

Figure 2.3.13 COMPARISON OF FLOTATION RATE CONSTANTS FOR THE BATCH, BATCH WITH NO PADDLING (BATCH*), AND PILOT PLANT TESTS AT DIFFERENT FROTHER CONCENTRATIONS

This effect is commonly ignored in laboratory testing. At the beginning of this project, motorized paddles were placed on top of the laboratory cell but were found to be unsatisfactory. A large number of particles was sticking to the paddles. The froth factor, defined as the ratio of the rate constant for the laboratory test to the pilot plant test or the batch flotation rate without paddling, is a function of the frother concentration. In the presence of an oily collector, such as dodecane, the froth factor was smaller as can be seen from the results in Table 2.3.5 and Figure 2.3.14. These results are not surprising since the collector is likely to increase froth stability.

Table 2.3.5 Change in Flotation Rate Constants and Froth Factors Due to the Addition of Collector for the Batch, Batch with no Paddling (Batch*), and Pilot-Plant Tests

Collector conc. (kg/T)	Rate constants (min^{-1})			Froth Factor ($k_{\text{lab}}/k_{\text{plant}}$)	
	Batch	Batch*	Plant	Batch*	Plant
0	1.9205	0.7198	0.1356	2.67	14.16
0.1	2.5493	0.9068	0.2964	2.81	8.60

Figure 2.3.14 EFFECT OF COLLECTOR ADDITION ON FLOTATION RATE CONSTANT FOR THE BATCH, BATCH WITH NO PADDLING (BATCH*), AND PILOT PLANT TESTS

From these studies, it can be concluded that the lower than expected recovery in the pilot plant flotation circuit was due to low mobility and less stability of the froth. It should be possible to improve the performance with the help of a frother, which is likely to give a more mobile froth. The use of an appropriate frother will also improve the recovery of coarse particles in the pilot plant or plant operation.

2.3.1.2 Conclusions

- The flotation rate and recovery increased as the frother concentration was increased, and reached a maximum at a concentration of 0.4 kg/T to 0.45 kg/T, and then decreased.
- The froth was stagnant in cells 3 and 4 when collector was added. Most of the material that floated was recovered in the first two cells.
- The flotation rate and recovery increased as the particle size decreased. More than 95% of the 200 x 400 mesh size fraction was recovered, while about 75% of the 28 x 48 mesh fraction was lost on the tails. This shows that MIBC is not especially suited for coarse particle flotation. The recovery of this fraction could have been greater if an alternate frother such as Dowfroth was used.

2.3.2 Column Flotation

Continuous flotation experiments were carried out in a 0.076 m x 3.55 m pilot flotation column. The flotation column was fitted with six high-accuracy pressure sensors (Cole Palmer, H-68971 series) at varying axial locations. The outputs from these sensors were fed to a 386 SX Everex computer at a frequency of 0.2 Hz. The desired hydrostatic head in the column (the liquid level) was set using the pressure sensor located at the bottom of the column. By regulating the tailings pump using an icon-based control program (COLUMNCT.VI), the hydrostatic head was maintained at the set value. The column program was developed using Labview Student Edition, provided by National Instruments. A detailed description of the column with other ancillary instrumentation has been provided in a previous report (Miller et al., 1993). Figure 2.3.15 shows a schematic diagram of the experimental set-up.

Figure 2.3.15 A SCHEMATIC REPRESENTATION OF THE EXPERIMENTAL SET-UP

The flotation experiments were carried out using - 100 mesh Lower Kittanning seam coal with feed ash and sulfur contents of 9.4% and 0.78%, respectively. For each experiment, approximately 100 liters of a 5 wt. % coal slurry was prepared. The required amount of frother was added directly to the slurry in the reservoir. Except where indicated, the frother used was methyl-iso-butyl-carbinol, MIBC.

The feed was introduced into the column at the desired flow rate with the air turned off. When the desired hydrostatic head in the column was attained, aeration was turned on. Collection of timed froth and tailings samples was initiated the moment the froth overflowed the cell lip. For the froth samples, collection was over 30 sec and 1 minute intervals for the first minute and remainder of the experimental run, respectively, while for the tailings stream, sampling was over 10 s, 15 s, 20 s, 30 s or 40 s intervals. In preliminary experiments, it was observed that froth removal would present a significant bottleneck in the accurate determination of the concentrate flow rate. This problem was overcome with the installation of a water ring spray on the outer periphery of the cell lip. For all the experiments reported here, the wash-water addition rate was fixed at 1.2 liters per minute. Figure 2.3.16 shows the variation of concentrate flow rate, percent solids in the tailings stream, and concentrate and tailings grades with flotation time. The results show that steady state conditions were achieved in about 2 to 4 minutes. Subsequently, this mode of column operation was adopted for all of the experimental runs.

The initial studies were carried out with the column equipped with a Mott porous gas sparger having an average pore size of 2 μm . In these experiments, the slurry feed was introduced into the column at varying inlet positions, which ranged from 0.43 m to 2.37 m from the gas distributor. With these experiments, the effect of recovery zone height on yield (and concentrate production rate) was determined. Additional experiments were also carried out at varying slurry concentrations of 2.5% to 10% (by weight), air flow rates of 1.5 to 6 L/min, and hydrostatic pressure settings of 1.5 m to 2.0 m.

In order to evaluate the effectiveness of a novel bubble generator that is under development at Penn State, flotation experiments were also carried out on the column equipped with the Vortactor turbulent contactor. A schematic diagram of the

Figure 2.3.16 EXPERIMENTAL VERIFICATION OF THE ATTAINMENT OF STEADY STATE: (a) CONCENTRATE; (b) TAILINGS MATERIAL (Filled symbols represent the concentrate and tailings grade)

Vortactor turbulent contactor is shown in Figure 2.3.17. In these experiments, the slurry feed and air were brought into intimate contact in the vortactor chamber where the pressure was set at 138 kPa. A detailed description of the vortactor turbulent contactor and its operation can be found elsewhere (Ityokumbul et al., 1996). The slurry-air mixture was also introduced into the column at two axial locations to provide a preliminary evaluation of the effect of recovery zone height.

The froth and tailings samples were filtered, dried and weighed. For the tailings samples, the amount of water was also measured in order to determine the tailings flow rate at any given time interval, and the per cent solids in the tailings stream. All of these values were used to determine the attainment of steady state in the experiments (see Figure 2.3.16).

The particle size distributions of the feed and tailings samples were determined. Approximately 50 g of the coal was placed in a beaker containing tap water. A small amount of Calgon and a few drops of Coal Master dispersant were added. The beaker was placed in a sonicating bath for 2 to 3 minutes to break the agglomerates. The sonicated coal slurry was then wet sieved through a 400-mesh screen, and the two fractions obtained were dried and weighed. The +400 mesh sample was dry screened. Approximately 1 g of the -400 mesh sample was dispersed in 100 ml of distilled water. After the addition of 0.1% Calgon solution and Coal Master, the particle size distribution was determined using a Leeds and Northrop SPA Microtrac. This procedure was repeated for the feed and tailings samples. The ash contents of the different fractions were determined in a Thermogravimetric Determinator (LECO TGA 501). From the weights of recovered coal and the ash contents, the yield, grade and ash rejections were calculated.

The static pressure measurements along the column height were used to compute the average gas hold-up in the recovery zone. Using the drift flux approach, the average bubble size in the recovery zone was estimated (Ityokumbul et al. 1995). From the particle size distribution analysis and the average bubble size, the ratio of projected particle surface area per unit bubble surface area was calculated for the different experimental runs.

Figure 2.3.17

**DETAILED DRAWING OF THE VORTACTOR
TURBULENT CONTACT CHAMBER**

2.3.2.1 Effect of Hydrostatic Pressure Setting

The variations of concentrate flow rate, grade, and froth phase gas hold-up with hydrostatic pressure setting are shown in Table 2.3.6. In these tests, the air flow rate was fixed at 6 L/min. The froth phase gas hold-up was estimated from the pressure sensor located 0.46 m from the cell lip. At hydrostatic pressure settings of 1.75 m and 2.0 m, the concentrate flow rate and product grade remained constant. By contrast, lowering the hydrostatic pressure setting to 1.5 m resulted in a sharp decline in the concentrate flow rate and ash content. The results obtained clearly show the importance of maintaining tight control of column operational parameters. For example, an increase of the froth phase gas hold-up from 59% to 69% reduced the concentrate flow rate by almost 50%. The reductions in the product ash content and concentrate flow rate are attributed to better drainage of the froth. In the case of the former, better drainage will mean recovery of less hydrophilic ash particles, while in the latter, better drainage will reduce the froth mobilities.

Table 2.3.6 Effect of Hydrostatic Pressure Setting on Froth Phase Gas Hold-up, Concentrate Flow Rate and Grade

Hydrostatic pressure setting, m	Average froth-phase gas hold-up, %	Concentrate flow rate, g/min	Product Ash (wt. %)
1.50	69	123	3.71
1.75	59	225	5.06
2.0	54	222	5.18

2.3.2.2 Effect of Recovery Zone Height

Figure 2.3.18 shows the effect of recovery zone height on yield and product ash content. These tests were carried out at a superficial gas velocity of 0.022 m/s (air flow rate of 6 liters per minute). The results show that the concentrate flow rate was largely independent of recovery zone height (236 ± 14 g/min). Several authors (e.g. see Bensley et al. (1985); Ounpuu and Tremblay, (1991); Ityokumbul, (1993)) have reported that the recovery zone height had negligible effect on the product yield. It is apparent from the foregoing that tall columns may not be required for the flotation of the Lower Kittanning

Figure 2.3.18 VARIATION OF CLEAN COAL YIELD AND GRADE WITH RECOVERY ZONE HEIGHT (Air Flow Rate=6 LPM)

seam coal used in the present study. This observation (i.e. negligible effect of recovery zone height on concentrate flow rate) is noteworthy as it illustrates the limitation of using statistical design in flotation column studies. For example, if the recovery zone height is one of the variables in a factorial design of column flotation experiments, the results show that no significant effect should be expected for recovery zone heights above 0.43 m.

The results show that with the exception of the lowest feed location, the product ash was fairly constant and was in the range expected for the observed yields. By contrast, Bensley et al. (1985) reported that the product ash increased as the feed location was raised. While the reasons for this are not entirely clear, it is noted that wash water was only employed in this study. With wash water addition, entrainment of fine refuse is suppressed.

2.3.2.3 Effect of Feed Solids Concentration

The effect of solids concentration on the yield, concentrate flowrate, and product ash was determined at an air flow rate of 3 L/min. In these experiments, the feed solid flow rate varied from 139 g/min (2.5% slurry) to 585 g/min (10% slurry) and the results are shown in Figure 2.3.19. The results show that the yield and product ash content decreased with increasing feed solid concentration, with the effect being more pronounced at solids concentration above 2.5%. Concurrently, a stable froth layer was observed in the collected tailing samples. Thus, the reduction in the concentrate flow rate is attributed, in part, to the loss of overloaded bubbles that lack the buoyancy to reach the froth phase (King et al., 1974; Szatkowski and Freyberger, 1984).

The results obtained at the high solids feed rates simulate column applications in cleaning duty. Typically, the separation in cleaner operations involves material that is hydrophobic. Because the transfer of particles to the bubble surface will not be rate limiting, the bubble loading process will be relatively fast. Thus, the use of tall columns will result in highly overloaded bubbles, which may lack the buoyancy to reach the froth zone. Thus, for cleaner operations, the use of short columns is strongly recommended. This may be the reason that Mines Gaspé currently operates a 3.66-m tall column in their copper cleaning circuit (Comeau 1995).

Figure 2.3.19 EFFECT OF FEED SOLIDS CONCENTRATION ON FLOTATION RESPONSE (a) CLEAN COAL YIELD AND GRADE (b) CONCENTRATE FLOW RATE (Air Flow Rate=3 LPM)

The results of the study show that the clean coal ash content decreased with increasing feed solids concentration. At feed solids concentration greater than 5%, the bubble surface availability appears to be rate limiting. Under bubble surface limiting conditions, the competition for the available surface clearly favors the more hydrophobic material (cleanest product). By contrast, at a feed solids concentration of 2.5%, the availability of bubble surface was not rate limiting. Under these conditions (free flotation), less hydrophobic material is not rejected from the froth zone. Furthermore, analysis of the froth did not show any significant differences in the product fineness. Thus, the drop in product ash content from 5.76 % to less than 3.2% as the feed solids concentration was increased from 2.5% can only be explained on the basis of limiting bubble surface area availability. Since the project objective called for a 5% ash product, operation of the column under surface limiting conditions is not recommended.

2.3.2.4 Effect of Air Flow Rate

The variation of clean coal yield and product grade with air flow rate is shown in Figure 2.3.20. These tests were carried out using a 5 wt. % feed slurry. The results show that the yield, concentrate flow rate and product ash content increased with air flow rate. This trend is expected since the available bubble surface area increases with air flow rate. Bensley et al. (1985) have reported similar results. For a solids feed rate in the range 240-320 g/min, it appears that conditions of free flotation are encountered at air flow rates above 3 L/min. This is consistent with earlier observations on the presence of air bubbles in the tailing samples collected at low air flowrates. Since the project objective called for the production of clean coal with an ash content of 5%, a single column flotation stage is sufficient with a yield of 66-80%.

2.3.2.5 Evaluation of Vortactor Turbulent Contactor

The Mott porous sparger used in the studies reported above was replaced with the Vortactor. Table 2.3.7 shows the preliminary results obtained with the Vortactor bubble generator. The results show that flotation performance (yield, product ash and concentrate flow rate) was independent of feed location. This observation is consistent

Figure 2.3.20 EFFECT OF AIR FLOW RATE ON (a) CLEAN COAL YIELD AND GRADE (b) CONCENTRATE FLOW RATE (Feed Solids Concentration = 5%)

with the results obtained with the Mott sparger. However, since there is no bulk transport of feed material above the inlet point, the results suggest that particle-bubble attachment is rather fast and takes place in the contact chamber and/or feed line.

Table 2.3.7 Preliminary Evaluation of Vortactor Bubble Generation in Fine Coal Cleaning

Air flowrate (liters/min)	Feed inlet¹ (m)	Yield (%)	Product Ash (wt. %)	Concentrate flowrate (g/min)
2	0.43	46.6	3.77	110
2	1.70	42.1	3.62	110
6 ²	1.70	70.1	5.40	181

¹ Distance from the bottom of the column

² The pulverizer used for coal preparation broke down and we had to change our coal preparation procedure.

Increasing the air flow rate from 2 L/min to 6 L/min increased the yield and concentrate flowrate to the levels obtained with the Mott porous sparger; however, the product ash was considerably higher. The reasons for the higher ash content observed with the Vortactor bubble generator at an air flow rate of 6 L/min are not entirely clear. However, the gas hold-up data suggest that the column operating conditions did not favor adequate drainage of the froth (Leffler 1997). This observation is partially supported by the recovery of coarse size fractions (i.e., +140 mesh and -140+200 mesh fractions) with unusually high ash content. The results obtained with an air flow rate of 2 L/min clearly show that increasing the recovery zone height (i.e., lowering the feed inlet point) did not have any effect on the yield and concentrate flow rate. As indicated with the Mott porous

sparger, this observation would suggest that tall columns are not required for fine coal flotation.

2.3.2.6 Correlation for Column Carrying Capacity and Yield

In flotation column separation, it is important to predict the yield and carrying capacity. These quantities are dependent on the column operational parameters. In the present case, the drift flux approach was used to estimate the average bubble size. From the measured particle size distribution, air flow rate and average bubble size, the projected particle to bubble surface area was determined. The clean coal yield was similarly determined from the measured rates of tailing and concentrate flows. From dimensional analysis, it can be shown that the column carrying capacity is related to the column operational parameters by the expression (Ityokumbul and Trubelja, 1998):

$$C_a = K \left(\frac{360 U_g}{d_b S_p} \right) \quad (2.3.3)$$

where:

K is a constant (projected particle surface area per unit bubble surface).

U_g is the superficial gas velocity (cm/s),

d_b is the bubble diameter (cm), and

S_p is the projected particle surface area (cm²/g).

The experimentally determined carrying capacities were plotted according to Equation 2.3.3 (see Figure 2.3.21). It can be seen that the data are well described by the relationship:

$$C_a = 0.5 \left(\frac{360 U_g}{d_b S_p} \right) \quad (2.3.4)$$

A comparison of Equations 2.3.3 and 2.3.4 reveal that K is equal to 0.5. This value is significant – it suggests that, in column flotation, the maximum carrying capacity will be obtained when 50% of the bubble surface in the recovery zone is covered by particles. While higher bubble loading is possible, the coalescence phenomena that occur in the froth zone would result in some of the collected particles being returned to the recovery zone. A comparison of the results of the present study with correlations currently

Figure 2.3.21 VARIATION OF CARRYING CAPACITY WITH COLUMN OPERATIONAL PARAMETERS

available in the literature gave good agreement only under free flotation conditions (Ityokumbul and Trubleja, 1998).

It is generally accepted that flotation is an interfacial phenomenon. However, availability of bubble surface is hardly considered in flotation analysis. For an interfacial phenomenon, the yield should decrease as the ratio of projected particle area to bubble surface increases. This hypothesis was tested by plotting clean coal yield as a function of

particle projected area per unit bubble surface (see Figure 2.3.22). Regression analysis of the data gave the relationship:

$$Yield = 0.445 \left(\frac{A_p}{A_b} \right)^{-1.05} \quad (2.3.5)$$

where

A_p is the total projected particle surface area flux in the feed, and

A_b is the bubble surface area flux.

The regression coefficient for Equation 1.3.5 was 0.83. As expected, the yield decreases with increasing ratio of projected particle to bubble surface area. A similar trend was reported by Bradshaw and O'Connor (1996) with pyrite flotation, even though their tests were conducted in a semi-batch manner. This observation clearly supports the well-established practice of feed desliming prior to flotation. It is noted that there is an almost linear relationship between the yield and the ratio of the surface areas, as predicted by theory.

2.3.3 Overall Evaluation

In order to determine the optimum conditions for the processing of the Lower Kittanning seam coal, the yield index is plotted as a function of the ash rejection for all the tests carried out in the present study (see Figure 2.3.23). The results show that the optimum conditions will give a yield index of about 45-48% at an ash rejection of 60%. These optimum conditions will give a clean coal yield of 75 - 80% with an ash content in the range 4.8 - 5.1 %.

The operational parameter that affects the product grade is the gas hold-up in the froth phase. For effective removal of fine refuse the use of wash water has proven successful. However, these results also show that a well-drained froth having gas hold-up of 55-70% was necessary for achieving target grades (see Figure 2.3.24). Since the objective of the current study is to produce a clean coal product with an ash content of less than 5%, the results show that this can be done in a single stage of column flotation. The processing conditions necessary for achieving these target objectives are:

Figure 2.3.22 VARIATION OF FRACTIONAL YIELD WITH PARTICLE PROJECTED SURFACE AREA PER UNIT BUBBLE SURFACE AREA

- air flow rate of 6 liters per minute
- solid feed rate 240-280 g/min, and
- froth phase gas hold-up of at least 50%.

2.3.4 Conclusion

The results suggest that flotation columns may not be effective for cleaning duties (where bubble surface area is limiting), except where a high value product is to be

produced. Since the project objective called for a 5% ash product, operation of the column under bubble surface limiting conditions is not recommended. For optimum results in fine coal cleaning, the feed solid concentration should not exceed 5 wt. %. This observation is consistent with the results of Bahr et al. (1987).

Figure 2.3.23 COMPARISON OF COLUMN FLOTATION AND WASHABILITY RESULTS

Figure 2.3.24 VARIATION OF CLEAN COAL YIELD AND GRADE WITH FROTH PHASE GAS HOLD-UP

2.4 Dry Processing

Dry separations of fine coal were carried out using triboelectrostatic separation. As part of this study, a laboratory electrostatic separator was constructed and tested to demonstrate the ability to clean fine dry coal on a continuous basis. The separations were evaluated in terms of clean coal yield and ash content. One of the potential problems of dry, fine coal separations is the tendency for particles to agglomerate. As such,

deagglomeration testing was also conducted. Finally, tests were conducted to verify the application of an integrated grinding/cleaning circuit.

2.4.1 Continuous TriboElectrostatic Separation

A continuous, laboratory triboelectrostatic separator was fabricated for use in this study. The test unit consisted of two, 600 mm diameter, rotating copper plates, which were spaced 100 mm apart (Figure 2.4.1). The plates were enclosed within a 810 mm high by 725 mm deep by 115 mm wide box constructed of 6.35 mm thick Plexiglas. The entire unit was mounted on a table. The diameter of the plates was based on the results from the batch testing, which showed that the bulk of the particles separated within a distance of 600 mm from the feed end (Weiland 1996).

The voltage was applied to each plate by attaching the positive or negative electrode from each power supply to a metal washer that "rides" on a rotating copper shaft, which was attached directly to the plate. The drive pulleys were insulated from the copper shafts. The plates were belt driven using a variable-speed motor.

The coal was fed with a vibrating feeder into a venturi feeder where it was entrained in a nitrogen stream. The solids were passed through an in-line copper static mixer to charge the particles. The solids were then discharged downward into the separator where they were collected electrostatically on the rotating plates. The separated material was scraped off at the bottom of the plates and allowed to fall into collection ports. A rotational speed of approximately 5 rpm was sufficient to facilitate product removal. After removal, the products were withdrawn under vacuum into separate Plexiglas cyclones where the solids were separated from the gas stream. The vacuum was provided by separate sweepers. The sweepers were also used to draw sweep air through the flow straighteners at the top of the separator.

Initial tests were carried out using -100 mesh Upper Freeport seam coal to establish an operating procedure and to determine limiting conditions for the separator. It was found that during operation of the separator, arcing occurred between the drive pulleys and the rotating plates at above approximately ± 15 kV, thus limiting the maximum operating voltage.

Figure 2.4.1 SCHEMATIC OF THE CONTINUOUS TRIBOELECTROSTATIC SEPARATOR

Ideally, after tribocharging, the coal and refuse would be collected onto the rotating plates and then removed as the material passed by the scraper. This material would fall into the collection trough where it would be drawn through the cyclones and then collected. However, several problems occurred during operation of the separator.

As the material was scraped from the plates, cross contamination occurred between the clean coal and refuse as the products fell toward their respective collecting troughs. The center splitter was extended to minimize this effect. Also, part of the recovered material was not carried into the cyclone collectors but remained in the collecting troughs. Although higher sweep air velocities would alleviate this problem, they could not be obtained with the circuit as tested.

Tests were carried out using Indiana, Pittsburgh, and Upper Freeport seam coals. The feed size for all coals was a nominal -100 mesh, which was obtained by stage crushing in disk and Holmes pulverizers. The Holmes pulverizer was the same as that used to integrate the grinding and batch electrostatic separation process, as discussed in a later section. However, for these tests, the -100 mesh coal was prepared separately and stored under argon in aluminum laminated bags until it was needed.

The coal was dried at approximately 100°C for two hours prior to separation. This material was then fed by a vibrating feeder into a venturi feeder and then passed through the in-line copper mixer (tribocharger) prior to electrostatic separation. Compressed nitrogen was used as the transport medium. Several hundred grams of coal were passed through the separator for a given test. Upon completion of the test, the coal and refuse were recovered from the plates, collecting troughs, and cyclones. The samples were weighed and analyzed for ash content.

Table 2.4.1 gives the weight distribution and ash values for the material collected from various parts of the separator and collection system. As can be seen, the results indicate that the device was able to separate the coal selectively. However, the overall separation was not as efficient as the batch separation because of the limitations discussed previously. Ideally, the bulk of the solids would be carried into the cyclone for collection in the cyclone underflow stream. As noted previously, some material will remain on the plates. For this unit, it was not possible or necessary to scrape all the material from the rotating plates after each revolution, and generally a thin layer of solids

remained on each plate. This buildup had a minimal effect on the collecting ability of the plates. Furthermore, the material remaining on the plates could benefit the process by reducing the wear on the plates, because the scraper would not contact the plates directly.

Table 2.4.1 Continuous Triboelectrostatic Separator Results when Processing
-100 Mesh Indiana Seam coal (feed ash = 9.1%), -100 Mesh Pittsburgh
Seam Coal (feed ash = 6.8%), and -100 Mesh Upper Freeport Seam Coal
(feed ash = 11.2%)

Product	Clean Coal		Refuse	
	Weight, %	Ash, %	Weight, %	Ash, %
<u>Indiana Seam</u>				
Electrode (Plate)	20.4	3.1	11.2	22.4
Collecting Trough	12.3	5.4	6.7	16.9
Cyclone	<u>26.3</u>	<u>6.2</u>	<u>23.2</u>	<u>10.8</u>
Total	59.0	5.0	41.0	15.0
<u>Pittsburgh Seam</u>				
Electrode (Plate)	35.2	2.0	18.2	13.9
Collecting Trough	12.1	5.3	12.9	12.9
Cyclone	<u>21.2</u>	<u>5.6</u>	<u>0.4</u>	<u>9.1</u>
Total	68.5	3.7	31.5	13.4
<u>Upper Freeport Seam</u>				
Electrode (Plate)	37.7	3.5	19.8	21.1
Collecting Trough	11.0	9.4	14.4	22.3
Cyclone	<u>15.6</u>	<u>8.0</u>	<u>1.5</u>	<u>17.7</u>
Total	64.3	5.6	35.7	21.4

In addition to the solids on the plates, some material settled in the airways leading to the collection cyclones. However, because any material in the airways was in the proper product stream, it could be collected and included with the appropriate product. For a long time run in a continuous operation, it is likely that a steady state would be

reached such that a constant amount of material would build up on the plates and throughout the collection system, allowing a steady amount of product to be collected.

As seen in Table 2.4.1 for the Indiana seam coal, the ash content was reduced from 9.1% in the feed to 5.0% in the clean coal at a yield of 59%. The corresponding ash content of the refuse was 15.0%. In the case of the Pittsburgh seam coal, clean coal with an ash content of 3.7% was obtained at a yield of 68.5%, with a refuse ash content of 13.4%. For the Upper Freeport seam coal, the ash content was reduced from 11.2% to 5.6% at a yield of 64.3%. The corresponding ash content of the refuse material was 21.4%. In all cases, middling streams could be obtained from the system. These streams would be candidates for regrinding followed by reprocessing to recover additional coal.

In all cases, the lowest ash material was found on the negative (clean coal) collecting plate and represented over 50% of the total clean coal in some cases. Similarly, the highest ash content material was typically found on the positive (refuse) collecting plate and also represented a significant portion of the respective product. Even though the splits among the plate, collecting trough, and cyclone would change depending on the total quantity of coal being fed to the unit, the overall yield and product quality should remain constant.

Some of the material for each product passed through the cyclone overflow and into the sweeper and could not be recovered. This represented between 10 and 20% of the total amount of coal that was fed to the separator. However, based on the comparison of the calculated and measured ash contents of the feed coals, it was likely that the losses were distributed in the same proportions as the product yields. Hence, the overall results would not be affected.

2.4.2 Charging/Deagglomeration Testing

Charging/deagglomeration testing was carried out using the batch triboelectrostatic separator. A fluidized-bed unit was used for deagglomerating and charging the coal simultaneously. The fluidized-bed unit was constructed out of a 1-5/8 inch diameter steel pipe, which was 8 inches long. The unit was sealed at both ends using threaded steel caps. The fluidized bed consisted of a layer (~3/4 inch thick) of either steel or copper beads, which were supported on a screen located in the middle of

the tube. A second screen was located one inch above the lower screen. The nitrogen-entrained coal particles were injected into the fluidized bed from below through a pipe in the side of the steel tube. Contact with the fluidized bed was used to promote deagglomeration and particle charging simultaneously. After passing through the fluidized bed, the coal particles were carried out through a tube, which extended from above fluidized bed down through the bottom cap. The particles discharged into either the copper in-line mixer or a copper transfer pipe prior to injection into the separator. For this testing, -6 mm Upper Freeport seam coal was pulverized to nominal -100 mesh using disk and Holmes pulverizers. The size reduction was done immediately before the testing. After size reduction, the coal was dried for two hours at 100°C. The dried coal was then fed into a venturi feeder, which was connected to the fluidized bed unit. Nitrogen was used as the transport gas. In the first test, the coal was passed through the fluidized bed (containing the copper beads) and then through the in-line mixer before being injected into the batch electrostatic separator. The batch separator consisted of two 48 inch long by 12 inch wide parallel copper plates, which were spaced 4 inches apart (Figure 2.4.2). This unit was the same as that used in previous testing (Weiland, 1996). A voltage of approximately ± 20 kV was used for the current tests. Approximately 100 g of coal were used for each test.

Upon completion of the test, the material was recovered from sections spaced along the length of the separator plates. The samples were weighed and the clean coal and refuse yields were calculated. Ash analyses were then performed on all samples. Additional tests were conducted using the following conditions: fluidized bed with steel beads and in-line mixer; fluidized bed with copper beads and transfer pipe; and, in-line mixer only.

Figure 2.4.3 shows the variation of clean coal and refuse yields as a function of distance along the plate for all conditions. As seen, the bulk of the clean coal and refuse material was collected along the first 40 cm of the separator plates. Up until this distance, the highest clean coal yield was obtained using only the in-line mixer, while the lowest yield was for the copper beads/transfer pipe combination. However, over the entire length of the separator, the yields were comparable.

Figure 2.4.2 SCHEMATIC OF THE BATCH TRIBOELECTROSTATIC SEPARATOR

Figure 2.4.3 CLEAN COAL AND REFUSE YIELDS AS A FUNCTION OF DISTANCE ALONG THE BATCH SEPARATOR FOR DIFFERENT CHARGING / DEAGGLOMERATION TECHNIQUES WHEN PROCESSING -100 MESH UPPER FREEPORT SEAM COAL

Figure 2.4.4 shows the variation in the cumulative ash as a function of distance for the same tests. The ash content of the clean coal was very similar in all cases. However, the ash contents of the refuse streams were different, especially for the copper beads/transfer pipe combination. The grade-yield curves for these tests are shown in Figure 2.4.5. As can be seen, for a given yield, the lowest ash content was obtained for the copper beads/in-line mixer combination, followed by the steel beads/in-line mixer combination. This indicates that the fluidized-bed unit may be helping to deagglomerate the particles prior to charging. Furthermore, it also demonstrates that contact with the beads without the in-line mixer may not be sufficient to charge the particles adequately.

Figure 2.4.4 CUMULATIVE ASH CONTENT FOR THE CLEAN COAL AND REFUSE STREAMS AS A FUNCTION OF DISTANCE ALONG THE BATCH SEPARATOR FOR DIFFERENT CHARGING / DEAGGLOMERATION TECHNIQUES WHEN PROCESSING -100 MESH UPPER FREEPORT SEAM COAL

2.4.3 Integrated Grinding/Separation

Integrating a grinding device prior to triboelectrostatic separation was also investigated. In such cases, the goal would be to achieve sufficient liberation of the coal from the mineral matter with minimal grinding, while adequately charging the particles prior to separation. A Holmes high-speed pulverizer was used for size reduction in combination with the batch triboelectrostatic separator. As noted previously, this pulverizer was also used to prepare the -100 mesh coal that was used in the previous triboelectrostatic separation tests. However, in those cases, the pulverized coal was

Figure 2.4.5 CLEAN COAL ASH-YIELD CURVES FOR DIFFERENT CHARGING / DEAGGLOMERATION TECHNIQUES WHEN PROCESSING -100 MESH UPPER FREEPORT SEAM COAL IN THE BATCH SEPARATOR

stored under argon for up to several weeks prior to tribocharging and separation. In this case, the coal was used directly after size reduction.

Nominal -28 mesh Upper Freeport seam coal was fed to the pulverizer. As the coal passed through the device, it was pulverized by rotating steel hammers. A stainless steel screen, having nominal 0.2 mm openings, was used to produce a product of approximately -100 mesh. Contact with the hammers and screen provided the opportunity for tribocharging the coal.

In order to provide a direct feed to the separator, the collection bin was removed and a funnel was attached to the pulverizer discharge. The funnel discharge was placed over the venturi feeder of the batch electrostatic separator, which had the in-line tribocharger removed. Nitrogen was used as the transport medium. As in the previous tests, the solids were separated and collected along the copper plates. Upon completion of the test, the material was removed in increments along each plate. Each sample was weighed to determine the incremental yield. Ash and total sulfur analyses were performed on each sample. These results are plotted in Figure 2.4.6.

As can be seen, the overall clean coal yield was about 55% with an ash content of about 4%. The clean coal sulfur content was about 1%. As was found in the previous tests, the lowest and highest ash fractions were obtained near the feed end of the separator. Based on these results, separation immediately after grinding should be possible.

2.5 References

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3.0 EMISSIONS REDUCTION

The objective of this task was to extend the studies of emissions from coal-fired industrial boilers that were conducted in the first two phases of the project. The activities that were performed involved fundamental studies, pilot-scale investigations, and full-scale demonstrations and included:

- Evaluations of NO_x reduction strategies such as low-temperature selective catalytic reduction and reburning by injecting a biomass-based product and a micronized coal-water mixture;
- A pilot-scale investigation of simultaneous SO₂ removal/NO_x reduction technology injecting a biomass-based product of pyrolysis in the combustion chamber;
- A fundamental study of formation/destruction mechanisms of polycyclic aromatic hydrocarbons followed by demonstration-scale testing;
- Comprehensive studies of trace elements and mercury emissions investigating the effect of coal cleaning, particulate removal devices, and operating conditions on trace element emissions;
- The development of a Unified Methodology for simultaneously measuring trace element emissions and mercury speciation;
- Continuing the investigation of fine particulate matter removal using ceramic filters; and
- The development of a novel molecular basket to capture CO₂ emissions from the flue gas of coal-fired boilers.

Each of these activities is discussed in detail in the following sections. Some of the activities expanded upon work that began in Phase II, which has been reported previously (Miller et al., 2000).

3.1 NO_x Reduction/ SO₂ Removal

In this activity, NO_x reduction and SO₂ removal strategies were investigated. NO_x reduction was investigated using low-temperature, selective catalytic reduction (SCR), BioLimeTM injection was studied for simultaneous NO_x/SO₂ control, and reburning tests for NO_x control were conducted. Sorbent injection tests were originally planned for SO₂/SO₃ control (as part of the low-temperature SCR work); however, the sorbent injection results from Phase II were so successful, with up to 98% SO₂ capture observed, that DOE and Penn State agreed not to pursue this work any further (Miller et al., 2000).

3.1.1 Low-Temperature Selective Catalytic Reduction

A low-temperature SCR activity was initiated in Phase II and extended into the Phase III of the project. The overall objectives for the low-temperature SCR activity were to:

- 1) To identify and/or develop a NO_x reduction catalyst that is compatible with the typical operating conditions and the economic constraints of industrial boilers, specifically:
 - Flue gas temperatures of 550°F (288°C);
 - O₂ concentrations of 3-5 vol.%;
 - H₂O concentration of 10-20 vol.%;
 - SO₂ concentrations of 500-1,000 ppm;
 - NO concentrations of 100-500 ppm;
 - No regeneration of sorbent/catalyst required; and
 - Low maintenance and operating costs.
- 2) To establish the limitations of the candidate NO_x reduction catalyst so that its implementation in pilot and demonstration-scale tests will be straightforward, for example, determining the relationship between space velocity and NO_x conversion efficiency for scale-up purposes.
- 3) To identify maximum allowable transients that the catalyst can be exposed to before losing effectiveness, such as swings in flue gas temperature and sulfur and unburned hydrocarbon concentrations.

Bench-scale and preliminary pilot-scale work was performed in Phase II to address these objectives and the results are discussed in detail in the Phase II final report (Miller, et al., 2000). In the previous testing, pilot-scale tests examined the operation of Engelhard Corporation's NOxCatTM920LT catalyst while firing natural gas. These tests confirmed the observations of the bench-scale tests with simplified (i.e., SO₂ free) flue gas. The objective of the next set of tests, those reported in this section, was to examine the behavior of the catalyst during simultaneous deSO_x-deNO_x operation, wherein a dry sorbent and ammonia were injected into the flue gas.

The overall concept for the low-temperature SCR technology was to reduce NO_x emissions using SCR technology either in conjunction with particulate removal (i.e., simultaneous NO_x control and fly ash removal) or separately from the particulate removal. This would involve either coating the clean side of a ceramic filter with a catalyst or coating ceramic monoliths located on the clean side of a particulate control device (e.g., plenum of a baghouse).

The bench-scale testing was to precede a full-scale demonstration using Penn State's demonstration boiler system with the ceramic filter chamber (described in Section 3.2); however, only pilot-scale testing was performed because the level of industrial interest in this work waned. Engelhard Corporation informed Penn State that a corporate decision had been made not to proceed with catalyzed filter activities. Likewise, Babcock and Wilcox, who had initially expressed interest in this activity because it was complementary to their SO_x - NO_x -Rox-BoxTM (SNRB) concept (B&W, 1999), later informed Penn State that the SNRB concept was not a primary business concern. Corning, Inc. was willing to participate in these experiments but with limited involvement so further work with the catalyst at the pilot-scale was conducted. The demonstration-scale tests were not performed however, because of the high cost to coat a large number of ceramic filters or monoliths, while industrial interest for this concept declined.

The pilot-scale testing was performed using Penn State's down-fired combustor (DFC) system. Activities included modifying the DFC system to accommodate ceramic monoliths, performing shakedown testing, and conducting the test program, which are discussed in the following sections.

3.1.1.1 System Modifications and Shakedown

A schematic diagram of the DFC showing the baghouse is given in Figure 3.1.1. The DFC is designed to evaluate the combustion performance and pollutant emissions from various fuels (natural gas, coal, coal-water slurry fuels). The combustor has a 20 inches internal diameter, and is 20 feet high. Each module consists of five 21 inch tall and 20 inch diameter circular refractory sections. A divergent refractory cone, commonly called a quarl, is positioned on top of the four circular refractory sections. The divergent cone top has a half-angle of approximately ten degrees and is fitted with a multi-fuel burner. The combustor is lined with a 3 inch thick refractory material to withstand temperatures up to 1,650 °C. An 8-inch-thick, light weight, high alumina insulation layer was used to minimize heat loss and produce a flame temperature of between 1,450 and 1,500°C. The flue gases pass through a modular heat exchanger that cools them to about 100°C before they exit. Several 4-inch sampling ports are located along the combustor. Sample ports are numbered 1 through 10 starting at the top. Gas samples for compositional analysis were extracted at the exit of the heat exchanger. Wall temperatures are monitored with type-S thermocouples at six locations along the combustor. The

Figure 3.1.1 SCHEMATIC DIAGRAM OF THE 500,000 Btu/h DOWN-FIRED COMBUSTOR AND AUXILIARY COMPONENTS

temperature readings T1 through T6 in the main combustor section are wall temperatures as read by type-S thermocouples. The temperatures of the flue gas entering and leaving the heat exchanger are monitored with type-K thermocouples (T11 and T12). Coal is transported using 10% of the total combustion air through the central pipe. The air that is sent with the coal stream is called primary air. Secondary air is introduced into the annular secondary pipe, which usually constituted 30% of the total air. The balance of the combustion air is introduced as tertiary air. All air streams are introduced at the same height.

In its original configuration, the baghouse contains a 3 by 3 array of cloth bag filters suspended from bag cups on the bottom of a steel tube sheet, with each bag supported internally by a wire cage. Venturis are mounted on the bottom side of the tube sheet, concentric with the bag cups. Blowtubes mounted in the plenum box above the venturis provide a back pulse of compressed air through the filters, thus removing particulate matter from the exterior of the bags.

The modifications to the DFC system included removing the bags and cages from the baghouse, increasing the height of the plenum, relocating the blow pipes, fabricating retaining rings to hold the ceramic monoliths in place, installing sorbent and ammonia injections systems, and installing additional continuous emissions monitors.

Prototype I

Figure 3.1.2 shows the configuration of the baghouse for the first series of tests. In the first round of modifications, the existing tube sheet along with the venturis, bag cups, cages, and bag filters were removed. A new tube sheet (identical to the original) was fabricated and a new set of cages, cups, and filter bags was purchased and installed. The ceramic monoliths were mounted within the bag filters for the first few tests. A platinum catalyzed monolith occupied the central position in the filter matrix while uncatalyzed monoliths were used at the other eight positions. Three-inch diameter Lo-Flo Pitot traverse stations were mounted on top of the tube sheet, above the catalyzed monolith and above one of the uncatalyzed monoliths. Details of the sensors and the traverse stations are provided in the following section (Prototype II), which shows the final configuration of the baghouse modifications. Six inch long flanged extension tubes were mounted above the remaining seven monoliths. The Lo-Flo stations were modified by the addition of 3/8 inch OD ports to permit sampling of flue gas immediately downstream of the monoliths. Venturis were not used in this configuration.

Figure 3.1.3 shows a cross-section of a monolith and its retaining tube. The uncatalyzed monoliths were turned in a lathe in order to reduce their outer diameter to approximately 2 9/16 inches. The platinum-catalyzed monolith was used without modification. The retaining tubes were fabricated from 3 5/16" id, 16-gauge, stainless steel tubing. A flange was welded to the upper end of the retaining tube (the bolt pattern in this flange matched the patterns in the tube sheet and bag cups) and a ring was welded to the bottom of the tube to support the monoliths. The retaining tubes were split in half lengthwise to permit insertion of the monoliths. Two sections of monolith with a total length of 19.2 inches were used in each tube. Pins (2 per monolith) fabricated from 1/16 inch OD welding rod were used to align the hole patterns of the two monolith sections. The monoliths were wrapped in a blanket of Saffil mat insulation and then placed between the two halves of the retaining tube. Stainless steel hose clamps were used to compress the matting, firmly holding the monoliths in place. The retaining tubes and

PROTOTYPE I

- NO SPACES BETWEEN MONOLITHS
- BAGS OVER RETAINER TUBES
- SMALL SPACER

Figure 3.1.2 CROSS SECTION OF BAGHOUSE CONTAINING MONOLITHS IN FILTER BAGS

PROTOTYPE I 2 PIECES WITH NO SEPARATION BETWEEN MONOLITHS

Figure 3.1.3 EXPLODED VIEW OF MONOLITH CHAMBER WITH RETAINING RING

monoliths were then mounted on the bottom side of the tube sheet, concentric with the bag filter mounting cups.

Ports were added to the 8 inch duct between the outlet of the heat exchanger and the inlet of the bag house. The first, located approximately 10 feet upstream of the baghouse inlet was for the injection of sodium bicarbonate into the flue gas stream. The second port, located 5 feet down stream of the sodium bicarbonate injection point, was for the injection of ammonia.

A dry material feeder (Schenck Accurate 102M) was used to inject NaHCO_3 into a transport air stream. In order to improve dispersion of the sodium bicarbonate in the flue gas stream, it was injected into the duct against the flow direction. Feed rates could be measured by using a load cell to follow the decreasing mass of the feeder (plus sodium bicarbonate) as a function of time. Initial calibrations of the dry material feeder produced a reliable mapping of feed rate versus control dial setting. However, these tests indicated that the feed rates required for the NO_x reduction study were all compressed in the bottom 10% of the motor speed control potentiometer. The manufacturer of the dry material feeder provided a new feed tube and auger, better suited to the low feed rates required in this study. However, attempts to calibrate the feeder in this configuration proved to be unreliable (repeated measurements of what was ostensibly the same feed rate varied greatly). Thus, all sodium bicarbonate feed rates were measured using the load cell.

Figure 3.1.4 shows the ammonia injection system installed in the 8 inch duct. The flow rate of the ammonia mixture (15% NH_3 , balance nitrogen) was measured with a mass flow meter. A nozzle with two outlets was used to inject the ammonia into the flue gas flow. Like the NaHCO_3 injection system, the ammonia injector introduced its flow against the flow direction in the duct to improve mixing.

Backpulsing of the baghouse causes a significant increase in the draft pressure in the DFC followed by several seconds of pressure oscillations as the induced draft fan controller recovers from the upset. In order to minimize the impact of baghouse pulsing on the analytical measurements, the portion of the timing circuitry which controls the time between pulses of compressed air was modified. The modified circuit allowed off-times of up to 10 minutes, although typical off-times of approximately 2 minutes (with a pulse duration of 0.5 seconds) were used in this study.

Maintaining the temperature of the monoliths near the values specified in the NO_x reduction test matrix proved to be difficult. Initial tests showed significant heat loss between the inlet and outlet of the baghouse. This problem was alleviated via improvements to the insulation on the exterior of the unit. Regulation of the temperature of the monoliths presented a greater problem. The only temperature control in the baghouse was by varying the flow of water through the heat exchanger. Due to the significant thermal inertia of the system, varying the flow of water usually led to undershooting the desired temperature upon increasing the flow and

overshooting (with the accompanying potential to damage the bag filters) upon decreasing the flow.

AMMONIA INJECTION SYSTEM

Figure 3.1.4 AMMONIA INJECTION SYSTEM INSTALLED IN 8-INCH INLET DUCT

Initial testing while firing natural gas revealed a large pressure drop (up to 8 inches w.c.) across the baghouse (typically pressure drops of 1 to 2 inches w.c. had been observed in previous studies). Due to this large pressure drop, the induced draft fan was unable to maintain a negative draft pressure even when operating at maximum revolutions per minute. Switching from firing natural gas to firing coal allowed us to decrease the flow of combustion air, thus decreasing the load on the fan. However, this proved to be insufficient and the combustor was shut down due to positive draft pressures and possible leakage of combustion products into the high bay. It was unclear at that time whether this pressure drop was due to clogging of the monoliths or clogging of the bag filters. At the time of these first tests, the only pressure measurement taps were located in the plenum box and in the main body of the baghouse. Thus, only the total pressure drop across the filter bags and monoliths could be measured at this time. A later modification added a pressure measuring tap which permitted the measurement of the pressure drop across the bag filters.

Following the first few days of testing, the unit was disassembled and inspected. The monoliths were found to be relatively clean and it was assumed that the observed pressure drop was due to ineffective cleaning of the bag filters by the compressed air pulse. Venturis were attached to the end of each of the monolith retaining tubes with silicone adhesive. Caked-on ash was removed from the exterior of the bag filters and the system was reassembled. A test firing of the system on natural gas revealed similar results. The pressure drop across the system increased with time presumably due to inefficient removal of ash from the exterior of the bag filters.

Modifications were made in an attempt to improve the effectiveness of the baghouse pulsing system (see Figure 3.1.5). The original tube sheet, bag cups, and venturis, along with the bag filters used in the series of tests described above, were installed in the baghouse. A 30-inch high galvanized steel spacer (see Figure 3.1.6) with blowdown tubes was fabricated and installed on top of this tube sheet. Then the tube sheet used for the previous series of tests (along with the monoliths and retaining tubes, low-flow stations, and extension tubes) was installed on top of the spacer. Finally, the 6-inch spacer from the previous tests and the plenum box were installed on top of this assembly.

PROTOTYPE II

Figure 3.1.5 CROSS SECTION OF BAGHOUSE WITH MONOLITHS ABOVE FILTER BAGS

Figure 3.1.6 30-INCH SPACER USED IN HOLDING MONOLITHS

An additional step was taken to mitigate the pressure drop problem by increasing the capacity of the induced draft fan. A larger motor (15 HP versus 10 HP) and a new inverter sized for the 15 HP motor were installed. In addition, the rotational speed of the fan was increased by changing the pulleys on the shaft of the motor and the shaft of the fan.

Following subsequent testing, ash build up was found in the plenum box region. The system was torn down and holes were found in the bag filters. Apparently these filters were damaged while inserting the tube sheet/bag filter assembly into the baghouse. The filters were replaced with a used set of bags, which were removed prior to the NO_x reduction testing.

Throughout most of these early tests, reliable readings were not obtainable from the low-flow Pitot stations. The readings were frequently zero or fluctuated wildly (between reasonable values, based upon the experimental conditions, and zero). Apparently, the flows were near the minimum flow threshold required to produce a stable reading on these flow stations. The tubing and fittings between the flow stations and the transmitter (including all connections within the plenum box) were checked for leaks and none were found.

Prototype II

The modifications to the baghouse for the second prototype centered around the flow stations. Two airflow stations were installed to measure the flow and space velocities through two of the nine ceramic monoliths, as shown in Figure 3.1.7. The low-flow stations, fabricated of type 316 stainless steel pipe and a Veltron II transmitter, were purchased from Air Monitor Corporation. The low-flow stations provide highly accurate measurement of low air volumes for monitoring, controlling, and indicating applications when connected to the Veltron II transmitter.

One of the flow stations was mounted on the pipe containing the catalyzed monolith while the other one was mounted on one of the eight remaining uncatalyzed monoliths. Three stainless steel tubes were connected to each flow station. One tube was used to withdraw a flue gas sample. The other two tubes were used to read static and total pressures. Each of the two tubes were plumbed into an inline three-way valve and then into the Veltron transmitter where the flow is read in ACFM. The 3/8-inch three-way valves were used to switch between the catalyzed and uncatalyzed monoliths, as the transmitter was equipped to read only one flow station at a time.

**PROTOTYPE II
SPACE BETWEEN MONOLITHS**

Figure 3.1.7 SIDE AND TOP VIEW WITH LO-FLO STATIONS AND TUBING

Several problems were encountered when testing resumed. The transmitter did not record a reading for the catalyzed monolith and varied between zero and 14 actual cubic feet per minute (ACFM). After several tests, it was determined that the detection limit for the transmitter was 14 ACFM and would not yield any value below this limit. It was also determined that the

catalyzed monolith presented a higher resistance to flow than the uncatalyzed monolith. The solution was to cap four of the eight uncatalyzed monoliths in order to increase the flow through the remaining monoliths. Four brass pipe fittings were secured in place using Dow Corning RTV sealant 736, a sealant resistant to heat up to 500°F. The monoliths on each corner were capped to maintain a symmetric flow regime through the remaining monoliths.

The next test yielded better results in that the flow station had sufficient flow and attainable transmitter readings. As the monolith temperature steadily rose from 100°F to approximately 375°F, the flow through each of the two monoliths increased steadily from 31 to 38 ACFM for the uncatalyzed monoliths and from 15 to 24 ACFM for the catalyzed monoliths. This test confirmed that the lower detection limit for the transmitter was approximately 14 ACFM.

The parameters chosen for the testing included space velocities on the order of 12,000 hr⁻¹. At a space velocity of 12,000 hr⁻¹, the flow through the five monoliths that were left uncapped was too high. Appropriate flow through each monolith should be approximately 16 ACFM. It was decided to uncap two of the four monoliths before the next set of tests. To maintain a symmetric flow regime, two monoliths on opposite corners were uncapped while the remaining two were left capped. After several more tests, there was inconsistency in maintaining the desired flow through the seven open monoliths. Problems with flow rates were also encountered when sodium bicarbonate was fed via an airswept system into the 8-inch duct before entering the baghouse (see Figure 3.1.4). Slide gates were then added to two of the monoliths thus allowing manual control of the flow. Because of the tight arrangement, only two of the monoliths were fitted with slide gates. The slide gates were installed on monoliths located at opposite corners and could be manually manipulated by handles located outside the plenum box. The final arrangement contained seven uncapped monoliths, two of which could be left open or closed.

Several other modifications were also made at this time. Three-inch collars made from aluminum flashing were attached to the top of the low-flow stations. This extended the length of the flow stations and helped to alleviate the flow from mixing back down into the top of the stations, which could result in a false reading or noise in the transmitter reading. The insulation on the two spacers and plenum box was greatly improved at this time.

Several more tests were conducted with these modifications. More problems existed with the flow through the catalyzed monolith. The transmitter indicated a value of zero for most of

the testing and indicated an average of 28 ACFM for the uncatalyzed monoliths. The pressure drop across the bags gradually increased with the addition of each new test. The pressure drop increased from approximately 1 to 3 inches w.c. to 8 to 9 inches w.c. This led to a visual inspection of the filter bags, which had been damaged.

The filter bags were replaced with new P84 felt filter bags and were properly seeded with crushed limestone per the manufacturer's instructions. After installation, the average pressure drop was 2.4 inches w.c., which was expected for typical operation.

The monolith configuration was changed per a conversation with a representative from Engelhard, who coated the monoliths with the catalyst. The testing to this point consisted of two monolith sections stacked atop one another to accommodate a length of 19.3 inches and were supported with a pipe and retaining ring as shown in Figure 3.1.3. It was recommended by Engelhard that there be at least a one-inch space between each section of monolith if they are to be stacked. To allow for this spacing, the retaining ring was removed from the bottom of the pipe. Two monolith sections, with a one-inch space, were held in place with the pressure exerted from the retaining pipe and Saffil insulation. A small piece of rigid wire was used for added support at the bottom of the pipe. The wire was molded around the bottom of the pipe and held in place with a hose clamp as can be seen in Figure 3.1.8. Lastly, to assure the monoliths were not contaminated, an eight-inch section was returned to Engelhard Corporation for evaluation and determined to be worthy for use.

3.1.1.2 Preliminary Testing and Results

Preliminary low-temperature SCR testing was performed and included experiments to establish the range of test parameters such as flue gas temperature, space velocity, and to shake down the NaHCO_3 and NH_3 injection systems. A test matrix was developed based on the preliminary testing where the flue gas flow rate through the monoliths (i.e., space velocity) and temperature limitations of the fabric bags were determined. The test matrix is given Table 3.1.1

The commercially available platinum-based NOxCat LT 920 catalyst used in this study was supplied by the Engelhard Corporation. It contained 200 cells per square inch (cpsi). Testing performed in an earlier evaluation of the SCR process at higher temperatures utilized the same catalyzed monoliths. To eliminate concern over its continued reactivity, a sample of the catalyzed monolith was returned to Engelhard Corporation for evaluation. This evaluation was

PROTOTYPE II 2 PIECES OF MONOLITH WITH A 1-INCH GAP BETWEEN PIECES

Figure 3.1.8 EXPLODED VIEW OF MONOLITH CHAMBER WITH RETAINING WIRE

performed at 392 and 482°F and space velocities of 20,000 and 60,000 hr^{-1} using a synthetic flue gas containing 10% oxygen, 10% water, and 500 ppm NO and NH_3 . The evaluation conducted at 392°F reported 17.7 and 7.8% conversions of NO at space velocities of 20,000 and 60,000 hr^{-1} , respectively. These results indicate 5% lower conversions than reported in the Phase II final report for a catalyst containing only 100 cpsi tested under similar conditions (Miller et al., 2000).

Table 3.1.1 Low-Temperature SCR Test Matrix

Test Number	Space Velocity (hr ⁻¹)	Flue Gas Temperature (°F)	NH ₃ /NO Molar Ratio
1 - 4	12,000	325	0.6, 0.8, 1.0, 1.2
5 - 8	12,000	375	0.6, 0.8, 1.0, 1.2
9 - 12	12,000	425	0.6, 0.8, 1.0, 1.2
13 - 16	24,000	325	0.6, 0.8, 1.0, 1.2
17 - 20	24,000	375	0.6, 0.8, 1.0, 1.2
21 - 24	24,000	425	0.6, 0.8, 1.0, 1.2
25 - 28	48,000	325	0.6, 0.8, 1.0, 1.2
29 - 32	48,000	375	0.6, 0.8, 1.0, 1.2
33 - 36	48,000	425	0.6, 0.8, 1.0, 1.2

Possible poisoning or fouling of some catalytic sites may have occurred in the earlier testing of this catalyst. At a higher temperature of 482°F, 53 and 37% conversions were measured for 20,000 and 60,000 hr⁻¹, respectively. The evaluation also indicated that greater than 50% of the NO was converted to N₂O, yielding a low selectivity.

The testing in this subtask was to examine the simultaneous deSO_x-deNO_x operation by injecting ammonia and dry sodium bicarbonate into the duct upstream of the DFC baghouse. Initial tests to examine the use of dry NaHCO₃ injection for the removal of SO₂ were performed while maintaining an average baghouse inlet temperature of 500°F. The goal was to demonstrate 90% SO₂ reduction at the baghouse outlet. Reducing the SO₂/SO₃ concentrations to very low levels minimizes the formation of ammonium bisulfate (ABS) during SCR operation. The deposition of ABS results in fouling of the catalyst surface. Earlier testing injecting sodium bicarbonate in the demonstration boiler system had shown greater than 90% reduction in SO₂ emissions at a Na₂/S molar ratio of 1.65. A time interval of 1.5 minutes between backpulses for cleaning the bags was used for this testing. A greater time interval was not possible with the backpulse controls. The timer for backpulsing of the DFC baghouse has been modified to allow up to 10 minutes between backpulses. This should allow greater control in optimizing the feed rate of NaHCO₃ and its retention on the filter bags. Initial testing of this system indicated that greater than 90% reduction of the SO₂ was achievable at lower Na₂/S molar ratios in the DFC system.

The preliminary testing with NH₃ injection provided results that compared well with those provided by Engelhard Corporation. Consequently, the test program listed in Table 3.1.1 was performed.

3.1.1.3 Low-Temperature SCR Testing and Results

Pilot-scale testing was performed to evaluate the use of the low temperature selective SCR catalyst with NH₃ injection for the control of NO_x emissions from a coal-fired industrial boiler. As previously mentioned, this testing utilized an approach similar to the SO_x-NO_x-Rox-Box™ technology. The objective of this testing was to evaluate the modified SNRB approach for controlling NO_x emissions at conventional baghouse temperatures, typical of those observed in the flue gas from an industrial boiler. The goal of this phase of testing was to evaluate this technology for simultaneous control of SO₂, NO_x and particulate emissions while burning coal in the DFC to determine if bench-scale results could be matched, in which significant conversion of NO was obtained at a catalyst temperature as low as 480°F (Miller et al., 2000).

SO₂ Emissions Control

Control of the SO₂/SO₃ emissions was accomplished through the injection of a dry sorbent, sodium bicarbonate (NaHCO₃), upstream of the fabric filter. Church and Dwight supplied the sorbent grade—extra fine sodium bicarbonate used in this study. The bicarbonate had a particle size distribution of greater than 95% passing 200 mesh and a mass median diameter of 35 µm. Sodium bicarbonate was chosen as the sorbent since it endothermically decomposes at relatively low temperatures, 250°F and above, forming a surface of sodium carbonate to react with acidic gases such as SO₂. This decomposition proceeds by the following reaction:

Several factors including flue gas temperature, humidity, and concentration of CO₂ affect the rate of decomposition. Although the flue gas velocity and the location of the injection site provided little residence time before the entrance to the fabric filter, evaluations of dry sorbent injection have shown high percentages of decomposition with less than 1 second at temperatures as low as

275°F (EPRI, 1990). After decomposition, the sodium carbonate reacts with SO₂ by the following reactions:

and

The purpose of sodium bicarbonate injection was not to perform a detailed investigation of the sorbent's performance in capturing sulfur. This examination was performed in Phase II of the project (Miller et al., 2000). Instead, the goal of the dry sorbent injection was to reduce the concentrations of SO₂ and SO₃ in the flue gas, thus preventing the possible formation of ammonium bisulfate, (NH₄)₂SO₂ or NH₄HSO₄. When sulfur is present in the flue gas, a small percentage of the SO₂ formed undergoes further oxidation to SO₃. The SO₃ can react with H₂O to form sulfuric acid, which can then combine with NH₃, resulting in the formation of ammonium bisulfate (ABS). The ABS, at temperatures typical of those inside a fabric filter (300 – 450°F), can deposit on and foul downstream equipment including the SCR catalyst. Therefore, it is important to remove as much of the SO₂ and SO₃ as possible from the gas stream. The goal for SO₂ removal was a minimum of 90%.

NO_x Emissions Control

SCR is a widely accepted technology for the control of NO_x emissions from fossil fuel combustion. The most commonly applied type of catalyst is a monolith coated with vanadia-titania. This type of catalyst displays a high reactivity at temperatures between 500-800°F. However, at the lower temperatures (338-482°F), typical of the flue gas leaving an industrial boiler, the reactivity drops off considerably. As a result, platinum, which has a higher reactivity at lower temperatures, becomes the active catalytic component of choice.

For the testing, a Middle Kittanning seam coal was burned and ammonia was injected for the NO_x control. Upon contacting the catalyst, the NO reacts with the ammonia by the following desired reactions:

and

Particulate Emissions Control

A commercially-available baghouse was used for removing the particulate matter from the flue gas. Nine teflon coated fiberglass felted bags removed the fly ash and dry sorbent prior to the flue gas contacting the SCR catalyst. This type of fabric has a maximum temperature rating of 500°F, typical for most boiler applications. The test conditions were chosen to eliminate the need for high-temperature filter bags, which are significantly more expensive than conventional filter bags. Also, consideration was not given to the use of flue gas reheating since this option would certainly decrease the overall system efficiency.

Backpulsing of the baghouse determines the exposure time of the sodium bicarbonate to the flue gas while retained on the filter bags. In order to increase the exposure time and minimize the impact of baghouse pulsing on the analytical instrumentation, modifications were made to the timing circuit board. These modifications allowed up to 10 minutes in off-times, although typical off-times of 1.75 minutes were used in this study. No attempt was made to optimize these operating conditions for greater sorbent utilization.

The particulate dust loading present in the baghouse outlet was not measured during this evaluation. However, it is assumed that a typical collection efficiency for this type of fabric, approximately 99.2%, was observed. Occasional inspections of the baghouse plenum and the ceramic monoliths indicated only small amounts of ash penetration. Therefore, blinding of the active catalytic sites with fly ash was not expected.

Experimental

The DFC was preheated on natural gas until the wall temperature inside the divergent cone on top of the combustor reached approximately 1,800°F. The coal was then progressively introduced and the natural gas decreased until the combustor was firing only coal. An average firing rate of 350,000 Btu/h was used for most of the testing. The firing rate, combined with separate control of the inlet cooling water to the heat exchanger, provided the desired flue gas temperatures entering the baghouse. A cross section of the baghouse showing the location of the monoliths is given in Figure 3.1.9. Care was exercised not to exceed the maximum temperature rating of the filter bags.

Figure 3.1.9 CROSS SECTION OF BAGHOUSE WITH MONOLITHS ABOVE FILTER BAGS

A Middle Kittanning seam coal was burned for all tests performed in this evaluation. A typical composition of the coal is given in Table 3.1.2.

Table 3.1.2 Typical Composition of the Middle Kittanning Seam Coal

Proximate Analysis, (wt.%, dry)	
Volatile Matter	31.0
Ash	4.5
Fixed Carbon	64.5
Ultimate Analysis, (wt.%, dry)	
Carbon	81.9
Hydrogen	5.1
Nitrogen	1.6
Sulfur	0.7
Oxygen	6.3
Ash	4.5
Higher Heating Value, Btu/lb	14,050

The following flue gas composition was produced at the combustor exit (dry basis) when the Middle Kittanning seam coal was burned in the DFC at an excess air level of approximately 20%:

Oxygen	3.5 – 4.5 %
Carbon Monoxide	75 – 150 ppm
Carbon Dioxide	14.5 – 15.5 %
Sulfur Dioxide	400 – 475 ppm
Nitrogen Oxides	650 – 750 ppm

After achieving stable operation on coal, dry sorbent injection was started and the rate adjusted to obtain 90% SO₂ reduction.

Prior to injecting ammonia, the flow rate through the catalyzed monolith was measured using a lo-flo traverse station mounted above the monolith retainer tube. The flow rate was then adjusted to achieve the desired space velocity by opening or closing the slide gate valves over the uncatalyzed monoliths. Side and top views of the ceramic monoliths are shown in Figure 3.1.10.

The ammonia was supplied from compressed gas cylinders containing 15 vol% NH₃ and a balance of nitrogen. A manual valve was used to adjust the flow rate, and it was measured by a series 8000 thermal mass flow meter manufactured by Eldridge Products, Inc. The required

injection rate of ammonia was determined from the desired NH_3/NO molar ratio, the volumetric flue gas flow rate, and the NO_x concentration in the flue gas entering the baghouse. This system, once adjusted, provided a stable flow rate of injected ammonia into the system.

Figure 3.1.10 SIDE AND TOP VIEW WITH LO-FLO STATIONS AND TUBING

Data collection was initiated once the desired test conditions were reached. Two emissions monitoring systems (CEMS) measured O_2 , CO , CO_2 , SO_2 , and NO_x concentrations

prior to the dry sorbent injection location and at the baghouse outlet. The chemiluminescent NO_x analyzers have the ability to measure the NO and NO_2 separately. This allows for the determination of the fate of NO during the injection of the sodium bicarbonate or ammonia. N_2O and NH_3 analyzers were added to the baghouse outlet to measure the catalyst selectivity and NH_3 slip.

Results

Sulfur Capture

Initial shakedown tests injecting sodium bicarbonate at varying rates upstream of the fabric filter were performed to determine the minimum injection rate required to yield 90% reduction in the SO_2 concentration. The results of these tests are plotted in Figure 3.1.11. An average normalized stoichiometric ratio (NSR) of 2.5 was required to achieve an SO_2 reduction of greater than 90% at the conditions tested. The average percent SO_2 reduction at a given injection rate was lower than that achieved by other researchers and during earlier tests evaluating this process on the demonstration boiler facility (Miller et al., 2000), probably because the system was not optimized.

Sodium Utilization

Sodium utilization calculated for these tests ranged between 25-47%. This indicates a very low level of effectiveness when compared to typical sodium utilizations of 70%. Previous evaluations have shown a direct correlation between increasing the flow rate or air-to-cloth ratio (A/C) and increasing sodium utilization (EPRI, 1990).

An examination of the fabric filter specifications indicated that it was designed for an air-to-cloth ratio of 4.2:1 at a flue gas volume of 275 ACFM. An average flue gas volumetric flow rate of 200 ACFM and an A/C ratio of 3.0:1 were used in this study to maintain an acceptable catalyst space velocity. This lowering of the A/C ratio may account for approximately a 5 – 10% decrease in the sodium utilization. However, sorbent particle fallout in the duct and fabric filter hopper are more likely to be the cause of the lower level of SO_2 removal and poor sodium utilization.

Typical duct velocities for an industrial coal-fired system are between 30 – 60 feet/second. Due to the large diameter ducts used on the down-fired combustor, flue gas

velocities of less than 10 feet/second are common with a flue gas volumetric flow rate of 200 ACFM. This low duct velocity results in sorbent fallout in the duct and in the hopper, where lower temperatures inhibit decomposition and where contact with, and capture of SO_2 are severely diminished.

I.D. Code	MMD, μm	Symbol
BA7	9	▲
BA6	9	○
BA5	27	■
BA2	32	◇
BA1	46	●
BA3	75	△
BA4	89	□
PSU-Demonstration Boiler	35	●
PSU-DFC	35	★

3.1.11 COMPARISON OF SO_2 REMOVAL AS A FUNCTION OF NSR AND PARTICLE SIZE PSU Flue gas temperature: (Demonstration Boiler ~380°F; DFC ~425°F) - All others tested at ~300°F

An additional reason that lower sodium utilizations were experienced is a shorter residence time for sorbent retention in the form of filter cake on the bags. During testing, an interval of 105 seconds between pulses was used in cleaning the fabric filter. No attempt was made to optimize the utilization by lengthening the delay between pulses, thus extending the sorbent's exposure to the flue gas. Previous testing indicated that a decrease in the NSR from 1.82 to 1.55 was achievable, while still maintaining greater than 90% SO₂ removal, by increasing the sorbent's retention time within the fabric filter (Miller et al., 2000). No significant impact on the filter performance was noted in either test series in which sodium bicarbonate was injected

NO_x Reduction by Sodium Bicarbonate Injection

Some level of NO_x reduction was observed with sodium bicarbonate injection, although the mechanism by which this occurs is not well understood. One possible reaction mechanism is that the combustion products, SO₂ and NO, are catalytically oxidized on the surface to SO₃ and NO₂, respectively. The SO₃ then reacts with available sodium carbonate forming sodium sulfate while the NO₂ reacts to form sodium nitrate. A likely reaction for the capture of NO₂ is the following:

Unlike the SO₃, the NO₂ does not have the ability to diffuse into the particle's interior. Therefore, while the production of SO₃ and NO₂ continues on the surface, the NO₂ is not completely captured, leaving some residual level in the gas stream.

Typical percentages of NO_x removal ranged from 5 – 10% at a NSR of 2.75. These removals, measured with a flue gas temperature of 325°F, indicate some level of NO oxidation and subsequent capture of the NO₂ on the sodium carbonate surface. Additional screening tests performed with a flue gas temperature of 425°F indicated between 10 – 15% of the NO had been oxidized to NO₂, however, no net decrease in NO_x emissions were observed. No explanation is offered for these results since additional studies report that NO_x removals do not decrease until significantly higher temperatures are reached (i.e., 800°F) (Babcock & Wilcox Co., 1999)).

SCR NO_x Conversion

Proposed initial testing included evaluating the low temperature SCR catalyst for NO conversion at several different flue gas temperatures, NH₃/NO molar ratios, and space velocities ranging between 12,000 and 48,000 h⁻¹. However, when a test series performed at a space velocity of 21,000 h⁻¹ yielded very low conversions of NO, it was decided that higher space velocity or a shorter residence time would not yield improved levels of NO conversion. Although greater conversions would be expected at lower space velocities, the geometry of the modified baghouse and limited available catalyzed monolith prevented these tests from being conducted.

The inlet and outlet baghouse emissions were recorded during testing and the percent conversion of NO was determined by the following equation:

$$\text{Conversion}(\%) = \frac{([NO]_{\text{initial}} - [NO]_{\text{final}})}{[NO]_{\text{initial}}} \cdot 100$$

The results from the initial series of tests evaluating the NO conversion as a function of catalyst temperature at a space velocity of approximately 21,000 h⁻¹ and a NH₃/NO molar ratio of 1.0 are given in Figure 3.1.12.

Figure 3.1.12 PERCENT NO_x CONVERSION AS A FUNCTION OF CATALYST TEMPERATURE (space velocity ~21,000 h⁻¹, NH₃/NO molar ratio ~ 1.0)

A maximum NO conversion of 24.3% was achieved at a catalyst temperature of 374°F. Additional test results were measured at temperatures as high as 410°F. However, difficulties in maintaining a steady NO concentration in the flue gas exiting the combustor made these results unreliable. Significant heat loss from the baghouse, combined with the temperature limits of the filter bag fabric, prevented temperatures above 410°F from being achieved. A list of the general operating conditions are provided in Table 3.1.3.

Table 3.1.3 Operating Parameters for SCR Catalyst/Dry Sorbent Injection

Sodium Bicarbonate Injection	
Normalized Stoichiometric Ratio (NSR)	2.75
Air-to-Cloth Ratio	3.0
Time Between Backpulses (sec)	105
Backpulse Pressure (psi)	80
Pressure Drop (inch w.c.)	1.5-2.2
Injection Velocity (ft/sec)	70
Injection Temperature (°F)	300-500
SCR Catalyst	
Cells Per Square Inch (CPSI)	200
Catalyst Volume (ft ³)	0.0478
Catalyst Temperature (°F)	290-410
Pressure Drop (inches water column)	1.2-2.0
NH ₃ /NO _x Molar Ratio	0.6-1.0

The initial objective of this work was to demonstrate this technology at the industrial scale on Penn State's 15 MM Btu/h watertube boiler. Given that the average temperature inside the baghouse of this facility is 365 °F, while burning the Middle Kittanning seam coal, these results accurately represent the percent NO_x conversion that could be achieved with this technology.

Figure 3.1.13 shows the percent conversion of NO_x versus catalyst temperature measured in the DFC and during the earlier bench scale evaluations. This figure shows that when the space velocities are considered, good agreement exists between the bench scale and the DFC results.

Figure 3.1.13 NO_x CONVERSION VERSUS MONOLITH TEMPERATURE FOR 21,000 H⁻¹ (■) IN THE DFC AND 12,000 h⁻¹ (◆) AND 24,000 h⁻¹ (Δ) IN THE BENCH-SCALE TESTING

Selectivity

The potential exists for the injected ammonia and the nitric oxide (NO) to undergo an additional reaction as follows:

This secondary reaction is not desired, since the product, N₂O, is a greenhouse pollutant. The chemiluminescent NO_x analyzer that was used has the ability to analyze for either NO or NO₂, a further oxidized state of NO. To determine the percent of NO converted by the above reaction, an analyzer was added to measure the N₂O at the baghouse outlet. The selectivity to N₂ was then determined by the following equation:

$$\text{Selectivity}(\%) = \frac{([\text{NO}]_{\text{initial}} - [\text{NO}]_{\text{final}} - [\text{N}_2\text{O}])}{[\text{NO}]_{\text{initial}} - [\text{NO}]_{\text{final}}} \cdot 100$$

Mechanical failures of the analyzer prevented N₂O measurements for several of the tests conducted. However, the selectivity determined for the higher temperatures reported in Figure 1.1.12 was approximately 80%. Earlier work conducted on the bench scale suggests that using lower space velocities would yield higher selectivities (Anand et al., 1997).

Ammonia Slip

The ammonia slip measured at the baghouse outlet was below the detection limits of the ammonia analyzer for all tests performed injecting at a NH_3/NO_x molar ratio of less than or equal to 1.0

Conclusions

Typical combustion control techniques for limiting NO_x emissions from coal-fired utility boilers, such as low NO_x burners, have demonstrated 15-50% reduction in NO_x emissions. However, the application of these techniques in coal-fired industrial boilers is limited due to the small combustion chamber. The short flue gas residence time makes completing combustion when applying staged burners or reburn fuels very difficult. The technology of utilizing a low-temperature SCR catalyst in combination with ammonia injection could provide comparable or additional reductions of NO_x emissions when applied post combustion. Pilot-scale tests have demonstrated that a low temperature SCR catalyst can achieve better than 20% reduction at typical temperatures (365°F) observed inside an industrial or utility-scale baghouse. Although a selectivity of approximately 80% was determined for some of the higher temperatures, earlier work suggests that decreasing the space velocity increases the selectivity. These measurements were made with little or no ammonia slip at space velocities of approximately $21,000 \text{ h}^{-1}$. The use of dry sorbent injection, such as sodium bicarbonate, can provide not only significant reductions in SO_2 emissions, but provides some added reduction in the NO_x emissions. Greater than 90% reduction in the SO_2 emissions and 5-10% reduction in NO_x emissions was demonstrated at the pilot-scale using a NSR of 2.75. Optimizing the operation of the baghouse backpulsing could lead to further decreases in the NSR. The use of dry sorbent injection further eliminated the possible formation of ammonium bisulfate, a potential source of fouling for the SCR catalyst and down stream equipment.

3.1.2 BioLime™ Injection

The Clean Air Act Amendments (CAAA) of 1990 established the Acid Rain Program under Title IV to achieve significant environmental and public health benefits through reductions in emissions of SO_2 and NO_x from power plants. NO_x emission limits are expected to be

decreased to 0.15 lb/MMBtu from coal-fired units by year 2007 under the Ozone Transport Regulation (OTR). In order to achieve very low emissions, a combination of two or more techniques may have to be used. Therefore, there is a need for an inexpensive and simple method to control NO_x . Reburning is an in-situ combustion modification in which NO_x produced in the main combustion zone is reduced by using a reburning fuel as a reducing agent. The reburning configuration divides the combustion zone into three subzones:

- Main combustion zone, where the main fuel is fired in oxidizing or slightly reducing conditions with approximately 80-85% of the heat released;
- Reburn zone, where a reburn fuel (natural gas, oil, pulverized coal, or biomass) is injected to create a fuel-rich, oxygen deficient zone for NO reduction, and the reburn fuel rate is normally 15-20% of the total heat input; and
- Burnout zone, where additional air is introduced further downstream to oxidize any remaining fuel fragments and complete the combustion.

NO formed in the main combustion zone is partially destroyed by the action of hydrocarbon radicals to form molecular nitrogen and nitrogenous intermediate species such as HCN and NH_3 .

Various studies (Chen and Ma, 1996; Spliethoff et al., 1996; Liu et al., 1997; Burch et al., 1991) have documented the role of both homogenous and heterogeneous reactions for NO_x reduction when firing coal. However, Wendt (1995) and Mereb and Wendt (1994) have discounted heterogeneous reactions showing that they have minor contribution towards reburning. The reduction potential of a fuel depends on its ability to produce CH_i radicals to react with NO. Assuming that only homogeneous gas phase reactions participate in the reduction, a higher volatile matter content of the fuel will enhance the NO_x reduction.

The reburning process is reasonably well understood when using coal and natural gas as the reburn fuels (Spliethoff et al., 1996; Liu et al., 1997; Bilbao et al., 1994, 1997; Ashworth et al., 1996; Smart and Morgan, 1994; Kercherer et al., 1994; Adams and Harding, 1998; Burch et al., 1991). However, pyrolysis products of biomass and other biomass-based materials have also been shown to cause NO_x reduction by mechanisms similar to reburning (Rüdiger et al., 1997; Pisupati et al., 1997; Brouwer, 1995). Biomass and biomass-based products offer the advantage of not only reducing SO_2 and NO_x emissions because of low nitrogen and sulfur contents, but also reducing greenhouse gas emissions (N_2O and CO_2).

The composition of the biomass or biomass-based product (lignin, cellulose and hemicellulose content) influences the distribution of pyrolysis products (gas, liquid, char) and hence the reduction of NO_x (Samolada et al., 1990; Zanzi et al., 1996). Hence, a deeper understanding of the composition will help in selection of feedstock to increase the yield of pyrolysis gases that are responsible for reducing NO_x emissions.

From a critical analysis of the literature, the concept emerges that the components of biomass (lignin, cellulose, and hemicellulose) and bio-oils themselves play significant roles in determining the pyrolysis behavior of the feed material (Raveendran et al., 1999; Bhalla and Pisupati, 1999). Also, the basic structure is less significant than the composition. In other words, the way in which these components are bound is not as important as the actual amounts of individual components present in the biomass or bio-oils. Thermo-analytical techniques such as TGA, provide information on chemical composition in a simple and rapid manner. ¹³C-NMR is also a technique that can be used to study the structure of biomass and biomass based materials. Through a combination of techniques like cross-polarization (CP) and magic angle spinning (MAS), it is possible to obtain high-resolution ¹³C-NMR spectra of solid samples without the necessity of modifying or degrading the sample.

In this phase of the program, the gas composition from flash pyrolysis of BioLime™, which was produced in an earlier study (Bhalla and Pisupati, 1999) was used to model NO reduction through homogeneous gas phase reactions when BioLime™ was used as a reburn fuel. The numerical predictions were then compared to NO_x emission levels from Penn State's down-fired combustor (DFC) testing to validate the model. A difference in NO_x reduction was observed by using the two different BioLime™ products as reburn fuels under similar operating conditions. This is believed to be due to differences in yields of flash pyrolysis products from the different BioLime™ feedstocks. With this in mind, the model was further used to study the relative contribution of each of the pyrolysis gas species in NO reduction through homogeneous reactions. The predictions were then verified by experimental results from the flow reactor. The composition of two biomass-based materials was studied using TGA and CP/MAS ¹³C-NMR techniques. Samples were then flash pyrolyzed in a flow reactor to relate the yield of pyrolysis gases and char with the composition.

3.1.2.1 Experimental

Raw Materials

Two biomass-based materials (BioLime™ I and III) were obtained from DynaMotive Technologies Corporation, Vancouver, Canada. BioOil produced from pyrolysis of biomass was simultaneously reacted with air, and lime/water slurry in a stirred reactor to prepare BioLime™. BioLime™ I has approximately 7 wt. % calcium, whereas BioLime™ III has a higher calcium content of approximately 14 wt %. Both samples, as obtained, were highly viscous liquids requiring progressive cavity pumps to induce flow, and had a decomposition temperature of about 60°C (160°F), at which congealing began to occur. The samples were first dried in an oven at 44°C (111°F) overnight. The dried samples were ground to pass through a 60 mesh (US standard) screen and stored in sample bottles. Compositional analysis of the samples is shown in Table 3.1.4.

Table 3.1.4 Compositional Analysis of the Two BioLime™ Samples (wt.%)

	BioLime™ I	BioLime™ III
Carbon (daf ^a)s	48.5	38.7
Hydrogen (daf)	7.7	8.4
Nitrogen (daf)	0.23	0.21
Sulfur (daf)	0.01	0.02
Oxygen (daf by diff)	43.5	52.6
Ash (db ^b)	20.9	48.5
Moisture (wt. %)	7.97	4.11

^a dry ash-free basis

^b dry basis

TGA Study

A Perkin-Elmer 7 series Thermal Analysis System was used for TGA studies on the two BioLime™ samples. Approximately 10 mg of BioLime™ sample was placed in a platinum crucible of the TGA furnace. The quantity of sample was chosen such that the volatile species were removed efficiently from the sample and thereby vapor-phase mass transfer did not control the BioLime™ pyrolysis. The furnace was heated from 30 to 1,000°C at a rate of 50°C min⁻¹ and then held isothermally at 1,000°C for 10 minutes. The methodology used was similar to that

reported by Ghetti et al., (1996), Vitolo and Ghetti (1994), and Raveendran et al., (1996). All of the pyrolysis studies were carried out in an inert atmosphere of flowing nitrogen (150 sccm)

Each test yielded a plot of weight loss versus time and temperature. The ash-free weight loss curve for each sample was calculated by subtracting the weight of ash (obtained from the proximate analysis) from the sample weight at each temperature. It was assumed that the weight of ash remains constant during pyrolysis. The percentage weight loss was calculated from the ash-free weight as a function of temperature.

CP/MAS ^{13}C -NMR Study

The solid state ^{13}C NMR spectra, employing the techniques of cross polarization/magic angle spinning (CP/MAS) were collected for BioLime™ samples on a Chemagnetics M100 NMR spectrometer operating at a field strength of 2.3 T. Approximately 250 mg of samples was loaded into a 9 mL ceramic rotor for all the experiments. The rotor was then placed in a spectrometer and accelerated to a spinning rate of 3.5 kHz. All spectra were collected in 0.6 K and zero-filled to 4K over a sweep width of 10 kHz. A contact time of 2 ms and a recycle delay of 1 s were applied with 230,000 transients collected. All shifts were measured relative to tetramethylsilane (TMS). The peaks were then quantified using GRAMS/32 software involving a mixture of Gauss and Lorz integration methods

Flow Reactor Set-up

The experimental setup used for flash pyrolysis is shown in Figure 3.1.14. Flash pyrolysis was performed in a flow reactor of length 137 cm (54 inches) and 3.8 cm (1.5 inches) ID Schedule 40 stainless steel 316 pipe. The reactor was heated using a Lindberg® model 55342 single-zone furnace. The furnace was interfaced to a Lindberg® controller capable of achieving and maintaining any temperature between 200 and 1,200°C. Samples were pyrolyzed at peak temperatures of 1,100 and 1,200°C. The peak temperatures were selected to be consistent with the temperatures found in the reburn zone of the pulverized coal-fired combustor. In each run, approximately 2 grams of sample were pyrolyzed using helium as carrier gas at a flow rate of 0.707 m³(std.)/h (25 scfh) to achieve a heating rate of 600°C/second when passed through the hot zone of the reactor. The pyrolysis products leaving the reactor were passed through a 2µm filter paper to collect the char. The char was collected from the filter paper and weighed. The gases

were collected in a 10-liter TEDLAR® bag and were later analyzed using an Arnel 0158 Series S GC. The peaks were integrated and quantified (based on 1 ml sample injected) using Chrom Perfect, version 3.5 software. The gas composition was then calculated as milligrams/gram of BioLime™ sample using gas densities at 25°C.

Figure 3.1.14 SCHEMATIC DIAGRAM OF THE FLOW REACTOR USED FOR FLASH PYROLYSIS

3.1.2.2 Down-Fired Combustor (DFC)

The DFC, shown in Figure 3.1.1 and discussed in Section 3.1.1.1, was used to evaluate SO₂ and NO_x emissions when using BioLime™ in a pilot-scale combustor. BioLime™ was introduced through port 5 into the reburn zone. Wall temperatures were monitored with type-S thermocouples at six locations along the combustor. Table 3.1.5 provides the ultimate analysis of the Middle Kittanning seam coal used in the study. The reburn zone parameters are shown in

Table 3.1.6. The BioLime™ was prepared, stored in a day tank, and pumped to the burner by a Moyno™ progressive cavity pump. The flue gas composition (O₂, CO₂, CO, SO₂, and NO_x) was recorded using a Texas Micro Systems™ Data Acquisition System with ViewDac™ software. The data collected during steady-state operation was then averaged.

Table 3.1.5 Analysis of the Middle Kittanning Seam Coal used in the Testing

Constituent	Weight % (Dry basis)
Carbon	77.32
Hydrogen	4.89
Nitrogen	1.22
Sulfur	2.13
Oxygen	1.53

Table 3.1.6 DFC Reburn Zone Parameters

DFC Reburn zone parameters	
Diameter	50.8 cm
Total length	96.52 cm
Volume	195,600 cm ³
Pressure	1 atm
Residence time	1.127 seconds
Inlet temperature at the center	1415 K
Inlet temperature at the wall	1426 K

3.1.2.3 Results and Discussion

TGA Results

Proximate analysis showed that both BioLime™ samples have different ash contents and hence, the unsubtracted weight loss curves obtained directly from the TGA are not directly comparable for determining volatiles yield. The weight loss curves on an ash-free basis shown in Figures 3.1.15 and 3.1.16 for BioLime™ samples I and III, respectively. The curves can be divided into four zones (Ghetti et al., 1996; Raveendran et al., 1996). These four zones can thus be related to the decomposition of individual components of BioLime™ as shown in Table 3.1.7. Various zones arise from the fact that lignin, having an aromatic structure, is expected to undergo pyrolysis at a higher temperature than that for cellulose and hemicellulose, which are saturated compounds.

Figure 3.1.15 ASH-FREE WEIGHT AS A FUNCTION OF TEMPERATURE FOR BioLime™ I

Figure 3.1.16 ASH-FREE WEIGHT AS A FUNCTION OF TEMPERATURE FOR BioLime™ III

Table 3.1.7 Various Zones of Weight Loss during the Pyrolysis of BioLime™ Samples

Zones	Corresponding Component
Zone I: ~100°C	Mainly moisture evaporation
Zone II: 166-400°C	Predominately hemicellulose and cellulose decomposition
Zone III: 400-600°C	Mainly lignin decomposition
Zone IV: 600-900 C	Thermal cracking of char

The distribution of moisture and volatile matter as a percent of total ash-free weight during pyrolysis is shown in Table 3.1.8. Table 3.1.9 shows the weight loss for both samples in Zones II and III as a fraction of initial weight.

Table 3.1.8 Cumulative Volatile Matter Yield (wt.%) during Pyrolysis of BioLime™ I and III

Zones	BioLime™ I	BioLime™ III
Zone I (~100°C)	8.36	5.77
Moisture		
Zone II (166-400°C)	35.95	19.09
Hemicellulose and Cellulose		
Zone III (400 –600°C)	24.26	34.30
Lignin		
Zone IV (600-900°C)	31.40	40.83
Secondary thermal decomposition of char		
Total volatiles released in (Zone II + Zone III)	60.21	53.40
Total volatiles released (Zone II +Zone III + Zone IV)	91.61	94.23

Evaluation of the weight loss of the samples between 166 and 900°C indicated that approximately 83% of the volatile matter, as determined from the proximate analysis, was released in the interval. Knowing that the individual components (hemicellulose, cellulose, lignin) play a more significant role in determining the pyrolysis characteristics of BioLime™ than the basic structure or degree of polymerization, the zonal distribution of volatiles released can be used to compare the amounts of individual components present in the BioLime™. Data

Table 3.1.9 Percent Volatile Matter Released in Zones II and III for the BioLime™ Samples

	BioLime™ I (weight loss/Initial Weight x 100)	BioLime™ III (weight loss/Initial Weight x 100)
Zone II	22.46	8.35
Zone III	15.15	15.0
Total	37.61	23.35

presented in Table 3.1.8 indicate that BioLime™ I has higher cellulose and hemicellulose contents, whereas BioLime™ III has higher lignin content. A higher amount of weight loss in Zone IV for BioLime™ III indicates that the higher lignin content forms more char during pyrolysis. More volatiles are released by primary pyrolysis for BioLime™ with a lower lignin content. A higher sum of percent distribution for zones II and III for BioLime™ I corroborates this point.

CP/MAS ¹³C-NMR Results

The NMR spectra for BioLime™ I and III are shown in Figures 3.1.17 and 3.1.18, respectively. The signal assignments for the chemical shifts in the spectra are summarized in Table 3.1.10, based upon values reported by others (Pfeffer and Gerasimonwicz, 1989; Macial et al., 1984; Haw et al., 1984; Sterk, 1987; Cyr et al., 1988; Almendros et al., 1992; Landuci et al., 1992; Love et al., 1992; Pan and Lachenal, 1995; Haw and Schultz, 1985). The relative intensities of the peaks as obtained from GRAMS/32 software are also shown in Table 3.1.10.

The following observations can be made from the data in Table 3.1.10:

- The peak at 148-149 ppm in the spectra is due to C₃ and C₅ of syringyl units, but only those with a free phenolic group at C₄. The peak at 153 ppm, which is also due to C₃ and C₅ of syringyl units but with ether linkages at C₄ (principally C_β-O-C₄ bond), is completely absent from the spectra. This indicates that aryl-alkyl ether linkages of the lignin are almost completely cleaved during the pyrolysis;
- The peaks at 181 (ester and carboxylic acids) and 148-149 (phenolic carbon) ppm show that both the BioLime™ samples have significant concentrations of oxygenated compounds;
- No cleavage of lignin methoxy groups and major reorganization of the aromatic rings takes place during the pyrolysis of biomass to form BioLime™; and

- The signals due to the aromatic carbon of lignin are well separated from the signals due to carbohydrates (although minor overlap with the signal due to C-1 of cellulose is present). A relative measure of lignin aromatic carbon content can be obtained by integrating from approximately 160-ppm to the low shielding side of the C-1 peak (approximately 109-ppm) (Haw et al., 1984).

The intensity of the region from 160-109 ppm can be calculated (Haw et al., 1984) for both BioLime™ I and III and is referred to as I_{low} . A formula of $C_9H_{8.35}O_{2.83}(OCH_3)_{1.43}$ with a formula weight of 206 g/mol is assumed to be representative of the lignin (Rydholm, 1965). Carbohydrates repeat units have an average empirical formula of $C_6H_{10}O_5$ (formula weight 162 g/mol). This assumption is completely valid for cellulose and approximately valid for hemicellulose. With this assumption, total integrated ^{13}C intensities due to all lignin and all carbohydrate carbons, I_{lig} and I_{carb} were calculated as:

Figure 3.1.17 ^{13}C -NMR SPECTRA FOR BioLime™ I

Figure 3.1.18 ^{13}C -NMR SPECTRA FOR BioLime™ III

Table 3.1.10 Signal Assignment and Percent Intensity of Chemical Shifts in ^{13}C -CP/MAS NMR Spectra of BioLime™ Samples

Peak Shift (ppm)	Assignment	BioLime™ I (Intensity %)	BioLime™ III (Intensity %)
181	Esters & carboxylic acids	17.0	8.55
168-169	carbonyl atoms of hemicellulose	2.3	5.64
148-149	Aromatic lignin C bonded to O	5.6	10.6
120-109	Aromatic lignin bonded to C or H	11.7	18.1
69-103	Cellulose carbon	42.4	38.1
54-58	OCH ₃ group of lignin	1.8	2.6
30-40	alkyl C of lignin not attached to O	10.1	18.9
22-23	CH ₃ of acetyl group of hemicellulose	9.3	3.0

$$I_{\text{lig}} = (10.43/6) \times I_{\text{low}} \quad 3.1.1$$

Equation 3.1.1 follows directly from the assumption that the average lignin repeat units has 10.43 carbons, 6 of which account for the total intensity measured by I_{low} .

$$I_{\text{carb}} = (100 - I_{\text{lig}}) \quad 3.1.2$$

The wt.% of lignin can then be calculated as follows:

$$\text{Wt \% lignin} = 100 \times \left[\frac{(206/10.43) \times I_{\text{lig}}}{(206/10.43) \times I_{\text{lig}} + (162/6) \times I_{\text{carb}}} \right] \quad 3.1.3$$

The I_{lig} and weight % of lignin calculated from equation 3.1.3 is shown in Table 3.1.11.

Table 3.1.11 I_{lig} and Weight Percent of Lignin Calculated for BioLime™ I and III

	I_{lig}	Weight % Lignin
BioLime™ I	30.07	23.9
BioLime™ III	50.1	42.0

Both ^{13}C -NMR and TGA analyses show that BioLime™ I has a lower lignin content (higher cellulose and hemicellulose content) than BioLime™ III. As seen from the TGA analyses, lower lignin content releases more volatiles by primary pyrolysis. Since BioLime™ is flash pyrolysed upon injection into the combustor, primary pyrolysis is thus predominately responsible for release of volatiles, which then react with NO reducing it to N_2 .

Flow Reactor Results

The composition of pyrolysis products from the flow reactor, calculated in milligrams/gram of BioLime™ is shown in Table 3.1.12. The total yield as well as the yield of the individual gases is higher for BioLime™ I. BioLime™ III is observed to produce more char. This pyrolysis behavior of both BioLime™ I and III is consistent with the lignin, cellulose, and hemicellulose contents calculated from the TGA and ^{13}C -NMR analyses. In other words, BioLime™ I, which has higher cellulose and hemicellulose content, produced more volatiles by primary pyrolysis. A higher lignin content in BioLime™ III forms more char during pyrolysis. Also, the weight percent of the total gas yield from flash pyrolysis closely matches the weight loss of BioLime™ in Zone II and Zone III in Table 3.1.9. This proves that primary pyrolysis is mainly responsible for the volatiles evolved during the flash pyrolysis. The weight percent of total gas yield for BioLime™ I is 1.7 times that for BioLime™ III, which also closely matches the lower lignin content (1.75 times) calculated from ^{13}C -NMR (Table 3.1.11)

Table 3.1.12 Composition of Pyrolysis Products at 1,100°C

Product	BioLime™ I (mg/g of BioLime™)	BioLime™ III (mg/g of BioLime™)
Hydrogen (H ₂)	3.37	2.95
Carbon dioxide (CO ₂)	278.21	158.77
Carbon monoxide (CO)	41.22	25.29
Methane (CH ₄)	8.79	6.34
Ethane (C ₂ H ₆)	1.46	0.86
Ethylene (C ₂ H ₄)	8.19	5.53
Acetylene (C ₂ H ₂)	0.77	0.51
Total	342.0	200.3
Total wt. % gas yield	34.20	20.02
Char	409	572
Total wt. % char yield	40.9	57.2

The composition of gases and char in milligrams/gram of BioLime™ III at 1,100 and 1,200°C is shown in Table 3.1.13. The decrease in char yield with increasing temperature is due to secondary decomposition of the char residue. The gaseous product yield increased with peak pyrolysis temperature. The increase in gaseous products can be attributed predominately to secondary decomposition of the char at higher temperatures. However, secondary decomposition of the char at higher temperatures may also give non-condensable gaseous products. The yields of CO, CO₂, and H₂ increased in general with increasing temperature. However, the increase in yield of CO₂ is more pronounced than that of CO or H₂. A slight decrease in yield of hydrocarbon gases is observed with increased temperature. The results are consistent with work reported in the literature by Rüdiger et al., (1997) Horne and Williams (1996).

Modeling NO_x Reduction

All numerical calculations were performed using a PSR computer code (Glarborg et al., 1989), which runs in conjunction with the Chemkin library (Kee et al., 1989). The reverse rate constants were obtained from the forward rate constants and the thermodynamic data was mainly taken from the Sandia Thermodynamic Database (1987). The code computes species concentrations from the balance between the net rate of production of each species by chemical reaction and the difference between the input and output flow rates of species. The model used to predict NO emissions is taken from the studies of Kilpinen et al., (1992) and Glarborg et al.,

Table 3.1.13 Composition of Pyrolysis Products for BioLime™ III at 1,100 and 1,200°C

Product	1,100°C (mg/g of BioLime™)	1,200°C (mg/g of BioLime™)
Hydrogen (H ₂)	2.95	3.76
Carbon dioxide (CO ₂)	158.77	191.27
Carbon monoxide (CO)	25.29	44.61
Methane (CH ₄)	6.34	6.08
Ethane (C ₂ H ₆)	0.86	0.86
Ethylene (C ₂ H ₄)	5.53	5.39
Acetylene (C ₂ H ₂)	0.51	0.799
Total	200.2	252.7
Total wt. % gas yield	20.02	25.27
Char	57.2	51.7
Total wt. % char yield	57.2	51.7

(1989). In general, the mechanism includes generation of hydrocarbon radicals from C₁ and C₂ parent hydrocarbons, oxidation mechanisms for HCN, and NH₃, together with a subset of the interactions between the hydrocarbon radicals and the nitrogenous species. The nitrogen-hydrocarbon chemistry is essentially an extension of the Miller and Bowman mechanism (1989).

The procedure and assumptions in the BioLime™ study are:

- Only homogeneous gas phase reactions are responsible for NO reduction. The effect of heterogeneous reactions is neglected. Gases from the pyrolysis of BioLime™ react with NO_x in the reburn zone to form N₂ and nitrogenous intermediate species such as HCN and NH₃;
- Gas concentrations are calculated from the yields reported in an earlier study (Bhalla and Pisupati, 1999) and are shown in Table 3.1.12 from flash pyrolysis for BioLime™ I and III;
- Reburn fuel is mixed rapidly and perfectly with the products of combustion from the primary zone. This is achieved by use of a high-pressure injection nozzle for feeding the reburn fuel;
- Reburn fuel is pyrolyzed instantly on entering the combustor and therefore pyrolysis kinetics do not control the NO_x reduction kinetics;
- The kinetics of reburning are controlled by the rate constants for the hydrocarbon-NO reaction system. The effect of mass transfer is neglected. This follows from the previous two assumptions;
- An average of the wall and bulk temperatures at the point of reburn fuel injection is used as the inlet temperature for the calculations;
- Gas densities are calculated at the reburn zone inlet temperatures;
- The flue gas in the reburn zone consists of 14% CO₂, 2.0% O₂, and 497 ppm of CO in nitrogen. These concentrations of CO₂, O₂, and CO are chosen to be consistent with those of a coal primary flame operated at a stoichiometric ratio of 1.0-1.1; and

- A residence time of 1.127 seconds is used for the calculations, which is consistent with the residence time in the reburn zone of the DFC.

Discussion of Results

Predictions of NO_x emissions from the model along with the DFC data are shown in Table 3.1.14. The reburn zone stoichiometry is based on the total amount of fuel and oxidizer that enters the DFC in the first two zones of the combustor divided by the stoichiometry requirement for the primary and reburn fuel. The stoichiometry requirements for coal and BioLime™ were calculated by subtracting the amount of air equivalent to oxygen in the fuel from the amount of air required for complete oxidation of carbon and hydrogen (determined by the ultimate analysis) content of the fuel. The predictions from the model match closely with the DFC data. The results show a higher NO_x reduction for BioLime™ I than for BioLime™ III. However, the stoichiometry was different for different runs when BioLime™ I and III were used as reburn fuels. Therefore, to further compare the performance of BioLime™ I and III, NO_x reduction was estimated for both BioLimes™ for each run. The results are shown in Figure 3.1.19. It can be seen that BioLime™ I, shows a higher NO_x reduction than BioLime™ III, which is attributed to the higher yield of total pyrolysis gases for BioLime™ I, which is in agreement with the studies of Kicherer et al., (1994). Kircherer et al. showed that the main reduction effect is due to the volatiles of the reburning fuel. Therefore, a higher yield of pyrolysis gases will result in higher NO_x reduction.

Table 3.1.14 Model Predictions for Percentage NO_x Reduction with NO_x Reduction Observed in the DFC

Run (BioLime™)	Stoich- iometry	% of Hydro- carbon added	% CO ₂ added	% CO added	% H ₂ added	% NO _x Reduction (DFC)	% NO _x Reduction (Model)
Run 1(I)	0.501	0.13	14.88	0.25	0.23	17.0	16.0
Run 2(I)	0.517	0.14	14.92	0.26	0.24	16.2	16.8
Run 3(III)	0.549	0.09	14.50	0.18	0.20	13.6	11.2
Run 4(III)	0.652	0.04	14.20	0.10	0.08	11.6	4.4
Run 5(III)	0.656	0.05	14.27	0.12	0.11	12.5	7.2
Run 6(III)	0.717	0.05	14.27	0.12	0.11	12.6	6.3

Figure 3.1.19 INFLUENCE OF BioLime™ USED ON NO_x REDUCTION. THE PERCENT NO_x REDUCTION IS PREDICTED FROM THE MODEL FOR BOTH BioLime™ FOR EACH RUN CONDITION. (Series 1 – BioLime™ I, Series 2 - BioLime™ III)

Different pyrolysis gases have different NO_x reduction potential. However, as seen from Table 3.1.14, the relative percentage of pyrolysis gases added is different for different runs. Knowledge of the relative contribution of pyrolysis gases in NO_x reduction will help in the selection of a biomass feedstock to increase the yield of desired species. With this in mind, a parametric analysis was performed using the model to study the effect of varying concentration of hydrocarbons, CO₂, CO, and H₂. The results were then verified using the flow reactor. The setup of the flow reactor is discussed in an earlier study (Bhalla and Pisupati, 1999). The results are shown in Figures 3.1.20 to 3.1.23. The results show that CO₂, CO, and H₂ have very little effect on NO_x reduction. Hydrocarbons are seen to be mainly responsible for causing reductions in emissions of NO_x. Kilpinen et al., (1992), Chen et al., (1986), and Mereb and Wendt (1990) have also shown that for natural gas reburning CO, and H₂ are useable as reburn fuels, even though the rates of reaction for NO reduction by CO and H₂ are significantly slower compared to NO-hydrocarbon reactions. Two main stages are considered to be mainly responsible for nitric oxide reduction. First is the destruction of NO to HCN, through the reaction with CH_i and HCCO.

Figure 3.1.20 EFFECT OF CO₂ CONCENTRATION ON NO_x REDUCTION (■ Experimental, ♦ Model A)

Figure 3.1.21 EFFECT OF H₂ CONCENTRATION ON NO_x REDUCTION (using Model A)

Figure 3.1.22 EFFECT OF CO CONCENTRATION ON NO_x REDUCTION (using Model A)

Figure 3.1.23 INFLUENCE OF HYDROCARBON CONCENTRATION ON PERCENT NO_x REDUCTION (■ Experimental, ♦ Model A)

The second stage is the reaction of HCN with the oxidizing species (O, OH) to form NH₃, N₂, and NO. HCN is first converted to isocyanic acid, HNCO primarily by

and to a lesser extent also through the direct reaction

The HNCO reacts further to NH_2 . This occurs primarily by the reaction with a H radical

The NH_2 radical, depending on the conditions, can be converted to N_2 , NH_3 , or NO .

Burch et al., (1991) from their study of different fuels (methane, hexane, benzene, and coal) have also shown that NO_x reduction efficiency can be related to C/H ratio for the fuel. Carbon rich fuels produce more CH_i fragments, leading to a lower NO_x concentration. BioLime™ I has a C/H ratio of 6.3 as compared to BioLime™ III which has a C/H ratio of 4.6. Therefore, a BioLime™ with a higher C/H ratio will help to increase the NO_x reduction.

3.1.2.4 Conclusions

The following conclusions were made from the BioLime™ study:

- ^{13}C -NMR and TGA analytical techniques can be effectively used to estimate various components (lignin, cellulose, and hemicellulose content) of biomass-based materials;
- The NMR spectra of the oils indicated that aryl-alkyl ether linkages of the lignin structure are extensively cleaved and carbonyl functional groups were dominating in the BioLime™;
- The TGA weight loss curves of BioLime™ samples in an inert atmosphere showed two distinct peaks between 160-400°C and 400-600°C, corresponding to the decomposition of cellulose, hemicellulose and lignin structures, respectively. The third distinct peak between 600-900°C corresponds to secondary thermal decomposition of the char;
- The total gas yield from flash pyrolysis for both BioLime™ I and III closely matched with the primary pyrolysis weight loss in TGA showing that primary pyrolysis reactions were mainly responsible for the volatiles being evolved in flash pyrolysis;
- Lower lignin and higher cellulose and hemicellulose contents are desirable in choosing a feedstock for NO reduction. Lower amounts of lignin evolved more volatiles and produced less char during flash pyrolysis of bio-oils in the combustor. Higher yields of pyrolysis gases increased the NO reduction potential of the BioLime™ through homogeneous gas phase reactions;

- An increase in pyrolysis temperature increased the gas yield and decreased the yield of char. This is mostly attributed to the increased primary decomposition of the char residue;
- Percentage of NO_x reduction using BioLime™ as a reburn fuel can be accurately predicted using the homogeneous gas phase reaction model. For most of the conditions, the percentage NO_x reduction predicted from the model is well within the allowable variation from the experimental results from the DFC and flow reactor;
- The higher NO_x reduction potential of BioLime™ I over BioLime™ III can be attributed to the higher yield of total pyrolysis gases; and
- The parametric study from the flow reactor shows that the CO, CO₂, and H₂ have very little effect on NO_x reduction and hydrocarbon reactions are mainly responsible for NO_x reduction. Hence a fuel with higher C/H ratio will produce more CH_i fragments, and result in a better NO_x reduction potential.

3.1.3 Reburning

3.1.3.1 Objectives

The main objective of this study was to understand the mechanism by which NO_x emissions are lowered when cofiring coal and coal-water slurry fuel (CWSF) as compared to firing coal only. This was accomplished by obtaining quantitative data to:

- Examine the extent of reburning in the gas phase, and
- Understand the heterogeneous char NO contribution to the reduction of NO.

3.1.3.2 Background

NO_x Fundamentals

Nitrogen oxides emissions from coal combustion can occur from three sources – thermal NO_x, fuel NO_x, and prompt NO_x. Thermal NO_x primarily forms from the reaction of nitrogen and oxygen in the combustion air. Fuel NO_x is a component that forms mainly from the conversion of nitrogen in the fuel to nitrogen oxides. Prompt NO_x is formed when hydrocarbon radical fragments in the flame zone react with nitrogen to form nitrogen atoms, which then form NO.

No definite rules exist to determine which nitrogen oxide formation mechanism dominates for a given stationary combustor configuration because of the complex interactions between burner aerodynamics, and both fuel oxidation and nitrogen species chemistry. But in general, fuel nitrogen has been shown to dominate pulverized coal fired boilers, although thermal

NO is also important in the post-flame regions where over-fire air is used. Thermal NO contributions only become significant at temperatures above 2,500°F in coal flames. Prompt NO formation is not typically an important mechanism during coal combustion. In the absence of fuel nitrogen, fuel NO is not a problem for natural gas flames. Prompt NO, however, is important in the vicinity of the inlet burners where reacting fuel fragments mix with the oxidizing air. Thus, in natural gas burners, both prompt and thermal NO contribute to the formation of nitrogen oxides.

Thermal NO

The principal reactions governing the formation of thermal NO are:

These two reactions are usually referred to as the thermal-NO formation mechanism or the Zeldovich mechanism. In fuel-rich environments, it has been suggested that at least one additional step should be included in this mechanism:

The reactions (3.1.4), (3.1.5) and (3.1.6) are usually referred to as the extended Zeldovich mechanism. Experiments have shown that as the complexity of the reactors and fuels increase, it is difficult to evaluate the rate forms (Bonvini et al., 1991). The general conclusion is that although the non-equilibrium effects are important in describing the initial rate of thermal-NO formation, the accelerated rates are still sufficiently low that very little thermal NO is formed in the combustion zone and that the majority is formed in the post flame region where the residence time is longer.

Prompt NO_x

Prompt NO_x occurs by the fast reaction of hydrocarbons with molecular nitrogen in fuel-rich flames. This mechanism accounts for rates of NO formation in the primary zone of the reactor, which are much greater than the expected rates of formation predicted by the thermal NO mechanism alone. NO measurements in both hydrocarbon and non-hydrocarbon flames have been interpreted to show that prompt NO is formed only when hydrocarbon radicals are available. The amount of NO formed in the fuel-rich systems is proportional to the concentration of N₂ and also to the number of carbons present in the gas phase. Hence, this mechanism is much more significant in fuel-rich hydrocarbon flames or in the reburning flames. Two reactions are believed to be the most significant to this mechanism (Miller and Bowman, 1989):

Reaction (3.1.7) was originally suggested and reaction (3.1.8) is considered to be only a minor, but non-negligible, contributor to prompt NO with its importance increasing with increasing temperature (Miller and Bowman, 1989).

Fuel NO

Fuel NO is by far the most significant source of nitric oxide formed during the combustion of nitrogen-containing fossil fuels. Fuel NO accounts for 75 to 95% of the total NO_x accumulation in coal flames and greater than 50% in fuel oil combustors. The reason for fuel NO dominance in coal systems is because of the moderate temperatures (1,500-2,000 K) and the locally fuel-rich nature of most coal flames. Fuel NO is formed more readily than thermal NO because the N-H and N-C bonds common in fuel-bound nitrogen are weaker than the triple bond in molecular nitrogen which must be dissociated to produce thermal NO.

The main step, at typical combustion temperatures, consists of conversion of fuel nitrogen into HCN, a step which according to some investigators (Fenimore, 1976; Rees et al.,

1981) is independent of the chemical nature of the initial fuel nitrogen. Once the fuel nitrogen has been converted to HCN, it rapidly decays to NH_i ($i = 1, 2, 3$) which reacts to form NO and N_2 .

3.1.3.3 NO_x Control Technologies

In order to comply with the regulations for nitrogen oxides emissions, various abatement strategies have been developed. These strategies can be divided into the following categories: 1) modification of the combustion configuration, 2) injection of reduction agents into the flue gases, and 3) treatment of the flue gas by post-combustion de-nitrification processes. A determination of the most effective and least expensive abatement technique depends on specific boiler firing conditions and the emission standards. A combination of techniques may be necessary to achieve certain mandated limits. Hence, the best NO_x control strategy for a certain unit is highly site dependent. Also, the selection of a pollution control system must also include the impact on boiler efficiency since many of the possible changes affect the fuel oxidation process. The large variety of combustion configurations and different fuel types requires that the abatement strategy be applied on a case-by-case basis. Due to complex interactions among fluid dynamics, chemistry, and energy transport processes, there is no guarantee that a particular alternative will achieve the desired control level. For this reason, a combination of theoretical, practical, and computational techniques is highly desirable in the development of a NO_x abatement strategy.

Modifications recognized as being technically viable and effective in reducing nitrogen oxide emissions from 50-80 % include: low excess air, flue gas recirculation, staged air combustion and staged fuel combustion. Also, advanced techniques like AGR-advanced gas reburning (Folsom et al., 1994), advanced low- NO_x burners (De Jong, 1993; Penterson and Ake, 1998) and COMBINOX (Gallo, 1994) have been developed. They usually involve a combination of techniques to reduce NO_x emissions.

Reburning Concept Development and Evolution

Reburning is a combustion modification technology, which removes NO_x from combustion products by using reburn fuel as the reducing agent. This technology is alternately referred to as “in-furnace NO_x reduction” or “staged fuel injection”.

The concept was originally developed by the John Zinc Company, and Wendt et al. (1973) and was based on the principle of Myerson et al. (1957) that CH fragments can react with NO. More recently, detailed studies have been conducted on the reburning process, and its potential for large-scale applications has been demonstrated (Chen et al., 1986; Lanier et al., 1986; Mereb and Wendt, 1990; Folsom et al., 1996; Maly et al., 1997; Falcone Miller, et al., 1996; Farzan et al., 1997). Despite these successful demonstrations, implementation of the technology has not been extensive on full scale in power plants, particularly in coal-fired units due to inherent difficulties in optimization of the reburning process. Concerted efforts to develop reburning for application to large boilers started in Japan during the late 1970's and early 1980's. An extensive development effort concerned the Mitsubishi Advanced Combustion Technology, which was applied to full scale boilers and resulted in 50% NO reduction, independent of fuel type and initial NO level. A review of the most representative reburning studies is presented in Table 3.1.15.

Table 3.1.15 Parameters and Characteristics of Reported Reburning Experiments

Recent Citation	Reburn fuels^a	Combustor Type / No.of exp.	NO meas. and reduction	Numerical modeling - NO prediction	RZ^b residence time (sec)	RZ reburn temp. (°F)
Chen et al. (1986)	NG	3 MW / 6	yes, 70 %	none	> 0.4	2500
Lanier et al. (1986)	NG	0.88 MW / 5	yes, 65 %	none	0.075	2100
Kokkinos (1992)	NG, coal, CWSF	1.8 MW / 3	yes, 40-75 %	none	1.5	-
Hura and Breen (1993)	NG	18 MW / 5	yes, 68 %	yes, 68 %	0.5	2200-2800
Li et al. (1993)	NG, coal	158 MW / 4	yes, 75 %	yes	-	-
May et al. (1994)	NG	33-172 MW / 8	yes, 60-70 %	none	-	-
Smart and Morgan (1994)	NG, coal, fuel oil	2.5 MW / 10	yes, 85 %	none	0.1	2300
Kicherer et al. (1994)	NG, coal, fuel oil	0.5 MW / 8	yes, 75 %	none	1.1	-
Mereb and Wendt (1994)	coal	17 kW / 14	yes, 50 %	none	0.18	2200
Payne and Moyeda(1994)	NG	33-158 MW / 4	yes, 50-68 %	yes	0.25-0.5	> 2500
Moyeda et al. (1995)	NG, coal	0.3 MW / 10	yes, 50 %	yes	0.2-0.6	2600

Spliethoff et al. (1996)	methane, syngas	30 kW / 4	yes, 70 %	none	1.2	2400
Liu et al. (1997)	coal	drop-tube	yes, 55 %	none	0.36	2100
Farzan et al. (1997)	NG	60 MW / 6	yes, 71 %	yes	-	-
Rudiger et al. (1995)	biomass, pyro-gas	30 kW / 5	yes, 50 %	none	1 - 2	2200
Ashworth et al. (1997)	NG, coal, CWSF	3 MW / 5	yes, 60-70 %	none	0.42-0.53	2700
Maly et al. (1997)	NG, PC, cws, BM, Orim	0.6 MW / 5	yes, 75 %	none	0.4-0.7	2600
Falcone Miller et al., (1996)	CWSF	30MW / 6	Yes,	None	N/A	N/A
Rostorfer et al. (1998)	Orim	74 MW / 3	yes, 67 %	none	-	-
Kluger et al. (1998)	coal	30 kW / 21	yes, 70 %	none	3	2400
Ashworth et al. (1998)	CWSF	3 MW / 4	yes, 50-77 %	none	0.53	2700

^a Reburn fuel: NG = natural gas; CWSF = coal-water slurry fuel; PC = pulverized coal; BM = biomass; Orim = orimulsion

^b RZ – recirculation zone

The primary advantages of reburning over other available NO_x control technologies are that it: 1) provides high levels of NO_x control; 2) can be implemented without significant impact on boiler performance; 3) produces no measurable by-product emissions; and 4) can be applied to all types of boilers, including cyclone-fired boilers, a design that is difficult to adapt to low-NO_x firing configurations. Other notable advantages, other than NO_x control include: SO₂ reduction (Chagger et al., 1997) in proportion to the gas fired (typically 15-20%), reduced ash disposal, also in proportion to the gas fired, reduced maintenance, and improved availability (Folsom, 1994).

The most current reburning techniques rely on using the entire furnace, although the burner fuel-staging concept has been successfully demonstrated (Smart and Morgan, 1994). However, in the current furnace and burner designs using reburning, the required primary, reburn and burnout zones are created by physically separating the three zones within the boiler. This

requires additional expensive operations and it is partly for this reason that the application of reburning on the full scale is currently not more widespread.

Reburning Mechanisms

Mechanisms governing the inter-conversion and destruction of nitrogen species in the fuel-rich reburning zone are the driving force behind reburning technology and have long been the subject of research by several workers (Wendt et al., 1973; Wendt, 1995). Comprehensive parametric studies have been conducted (Chen et al., 1986), delineating the factors that affect reburning effectiveness. Subsequent work by Lanier et al. (1986) identified critical mechanisms controlling the destruction of NO during reburning in a package boiler simulator. The destruction of N₂ by hydrocarbon reactions to form HCN was proposed as a mechanism that would limit the effectiveness of reburning, especially at lower primary NO levels. Using a model that did not account for hydrocarbon reactions, Knill and Morgan (1989) were able to predict the destruction of NO and HCN for selected experiments in an isothermal plug flow reactor. This mechanism was able to predict profiles of all nitrogenous species (NO, HCN and NH₃) in the reburning zone for all tests. Mechanisms governing the destruction of NO in the far fuel rich post flame of a pulverized coal staged combustion system were documented by Bose and Wendt (1988), Mereb and Wendt (1990) and Burch et al. (1994). Their work expands the reburning configuration analysis to predict both HCN and NH₃ concentrations, using measured values of NH₃ in their mechanism.

Alternative Fuels and Coal-Water Slurry Potential as a Reburning Fuel

In the past several years, various testing programs have shown that a wide range of fuels can effectively be used for reburning (Jones, 1997). Coal water slurry (McHale, 1985; Falcone Miller et al., 1996; Huettenhein and Chari, 1998), biomass (Kartak et al., 1996; Juchelkova, 1998) and Orimulsion (Schimmoller, 1998; Rostorfer, et al., 1998) are the ones receiving the highest interest due to their potential for replacing natural gas and coal. Basic reburning performance for coal fuels and biomass were found to be similar or slightly better than that of natural gas, with over 70 % NO_x reduction available at reburn heat inputs above 20 % (Maly et al., 1997). While reburning with alternative fuels has the potential to have boiler impacts, such as increased slagging with biomass, these impacts are generally minimized because the reburn

fuel comprises only a small fraction of the total boiler heat input. Reburning techniques with these fuels are flexible since they can be readily applied to a wide variety of waste products and off-specification fuels having low cost. Therefore, alternative fuel reburning can provide great economic benefits while offering the potential for high levels of NO_x reduction for a wide range of new and retrofit power plant applications.

Coal-water slurry fuel (CWSF) has been regarded for many years as an advantageous alternative for coal burning (Battista and Zawadski, 1995; Battista and Ashworth, 1998; Allen and Beal, 1998). It is estimated that there are over two billion tons of coal fines currently stored in active and inactive ponds in the United States (Falcone Miller, et al, 1995). Furthermore, some 50 million additional tons of fines are added to these ponds each year. These stored coal fines are a byproduct stream from coal cleaning plants and represent a significant fuel loss for coal companies and also an additional operating cost and environmental problem as ponds are filled and new pond permits become necessary. However, if the coal pond fines can be economically recovered and used, the material is no longer a liability but represents a significant stockpile of fuel (Falcone Miller, et al, 1995). These coal fines represent a huge energy resource, enough fuel to generate 3,600 MW of electricity every hour for the next 100 years (Ashworth et al., 1997). Hence, recovery of the fines for use as a low solids coal water slurry is a very attractive option from both an economical and an environmental standpoint.

For reburning application, CWSF needs to be of the same quality as that for the wall-fired and tangentially-fired units. With CWSF reburning, NO_x reductions of up to 75% could be obtained on cyclone units (Ashworth et al., 1998). This is important for example to cyclone-fired boiler operators because NO_x emissions from cyclone units range from 1.0 to 2.0 lb/million Btu and there are not many viable options available to reduce emissions to the U.S. EPA regulated limit of 0.86 lb NO_x /million Btu for units greater than 155 MWe. CWSF reburning is applicable to all types of coal-fired boilers and offers electric utilities a potential low cost option to achieve significant NO_x emissions reduction.

3.1.3.4 Test Program

The study was conducted using the DFC described in Section 3.1.1.1 and shown in Figure 3.1.1. Figure 3.1.24 shows the schematic of the Dalavan nozzle that was used to atomize the CWSF. Pulverized coal (PC) was fired through the central pipe of the burner, and CWSF

was injected in Port 2 to simulate the reburn conditions. CWSF was cofired with coal at 0, 10, 20, 30% of the total thermal input. Fifteen tests were performed in the DFC. The baseline flue gas concentration and temperature profile along the combustor were generated for coal firing. A continuous emission monitoring system (CEM) was used to measure O_2 , NO_x , and CO at various port levels. Axial oxygen concentration measurements were also obtained.

Experimental Procedure

A nominal firing rate of ≈ 0.5 million (MM) Btu/h was used in the study. Ports were installed as per EPA methods for stack gas sampling for the sulfur oxides evaluation. The CWSF atomizing gun was placed at port 2 for all tests. Combustion air for CWSF was introduced in port 3 during all reburn tests. An atomizing air pressure of 100 psig was used for the CWSF.

Figure 3.1.24 SCHEMATIC DIAGRAM OF THE DELAVAN NOZZLE

Table 3.1.16 provides a summary of combustion test conditions and emissions data. Data on the gas temperatures using a LAND suction pyrometer, and oxygen, CO, and NO_x concentration profiles using a CEM were also obtained during the tests. Char samples were collected from various ports in the DFC using an isokinetic sampling probe. The char samples were collected in wet condition by washing the probe after each test and were analyzed for carbon, hydrogen, nitrogen and carbon burnout.

Table 3.1.16 Summary of the Test Matrix

Test No.	Thermal Input Slurry (MMBtu/h)	Thermal Input-coal (MMBtu/h)	lb NO _x / MM Btu	NO _x (corrected, ppm)	Pct. O ₂ in the Flue Gas	Comments
1	0.154	0.353	0.71	519	5.39	All combustion air through burner
2	0.151	0.356	0.84	618	4.79	Use 100% PC, 5% O ₂ air in burner Port 3, make up air
3	NA ^a	0.357	0.89	654	9.89	100% PC, 0.5 MM Btu air
4	NA	0.360	0.64	469	5.03	5% O ₂ , 100% PC
5	0.165	0.364	0.83	613	5.26	Use 100% PC, 5% O ₂ air in burner Port 3, make up air
6	NA	0.407	0.76	560	5.48	5% O ₂ , 100% PC
7	NA	0.416	0.94	692	7.91	100% PC, 0.5 MM Btu air
8	0.101	0.410	0.80	587	5.39	All combustion air through burner
9	0.106	0.413	0.93	684	5.49	Use 100% PC, 5% O ₂ air in burner Port 3, make up air
10	NA	0.452	0.96	709	6.61	100% PC, 0.5 MM Btu air
11	NA	0.459	0.74	527	5.35	5% O ₂ , 100% PC
12	0.063	0.457	0.74	541	5.00	All combustion air through burner
13	0.065	0.458	0.87	638	5.38	Use 100% PC, 5% O ₂ air in burner Port 3, make up air
14	11.1 lb/h	0.508	0.77	564	5.64	All combustion air through burner
15	NA	0.515	0.89	656	5.00	Baseline test 5% O ₂ stack 100% PC

^a Not Applicable

Analysis of Results

The gas composition (oxygen, CO and NO_x) at various port levels was measured at the center of the combustor, midway between the center and the wall (called "Mid") and at the wall. These profiles are shown in Figure 3.1.25 when firing 100% PC with all the air through the burner. This plot shows that the NO_x concentration in the gas phase was maximum near the burner and as the flue gases passed down the combustor, the NO_x levels decreased. This suggests that the rate of destruction exceeded the rate of production of NO_x. An increase in the destruction is possibly accomplished by the reburn mechanism. Reburning can take place either by homogeneous gas phase reactions or by NO_x - char reactions. Upon heating a coal particle, nitrogen is also distributed between the volatile and char phases. NH₃ and HCN species in the volatiles, depending on the temperature, concentration and mixing, can form NO_x or N₂. Simultaneously the char nitrogen will be released to form either NO_x or finally to N₂ as shown in Figure 3.1.26. A balance between the two processes results in a net increase or decrease in the NO_x emissions. Figures 3.1.27-3.1.29 provide NO_x profiles as a function of distance from the

burner when all the air is admitted through the burner, although the amount of fuel is decreased in the case cofiring. This was done in order to keep the amount of air and thus the aerodynamics similar to that of the other tests (especially in the near-burner zone). Figures 3.1.30-3.1.32 show NO_x profiles when cofiring 10, 20, and 30% CWSF but with 90, 80, and 70% of the air respectively, introduced through the burner and the balance introduced through Port 3. It can be seen from Figures 3.1.27-3.1.29 that the initial NO_x levels when cofiring CWSF are lower than the values for 100% PC. This suggests that the formation of NO_x in the initial phases is lower. Subsequently, there is a minor decrease in the NO_x concentration for the tests with CWSF cofiring. These results indicate that the NO_x production and destruction rates, after the primary zone, are almost equal leading to no net production of NO_x .

Figure 3.1.25 NO_x PROFILE WHEN FIRING 100% PC WITH ALL OF THE AIR INTRODUCED THROUGH THE BURNER

Conceptually, the reburning process can be divided into three zones: a) the Primary Combustion Zone: This is the main heat release zone which accounts for approximately 80 % of the total heat input to the system and is operated under fuel-lean conditions. The level of NO_x exiting this zone is defined as the input to the reburning process; b) the Reburning Zone: The reburning fuel (10-30 % of the total fuel input) is injected downstream of the primary zone to create a fuel-rich, NO_x -reduction zone. Reactive nitrogen enters this zone from two sources: the primary NO_x input and the fuel nitrogen in the reburning fuel. These reactive nitrogen species

Figure 3.1.26 SCHEMATIC DIAGRAM OF NITROGEN TRANSFORMATIONS DURING PULVERIZED COAL COMBUSTION

Figure 3.1.27 NO_x PROFILE IN THE COMBUSTOR WHEN FIRING 90% PC AND 10% CWSF WITH ALL OF THE AIR INTRODUCED THROUGH THE BURNER

Figure 3.1.28 NO_x CONCENTRATION PROFILE WHEN FIRING 80% PC AND 20% CWSF WITH ALL OF THE AIR INTRODUCED THROUGH THE BURNER

Figure 3.1.29 NO_x PROFILE WHEN FIRING 70% PC AND 30% CWSF WITH ALL OF THE AIR INTRODUCED THROUGH THE BURNER

Figure 3.1.30 NO_x EMISSION PROFILE WHEN FIRING 90% PC AND 10% CWSF WITH MAKE-UP AIR INTRODUCED THROUGH PORT 3

Figure 3.1.31 NO_x EMISSION PROFILE WHEN FIRING 80% PC AND 20% CWSF WITH MAKE-UP AIR INTRODUCED THROUGH PORT 3

Figure 3.1.32 NO_x EMISSION PROFILE WHEN FIRING 70% PC AND 30% CWSF WITH MAKE-UP AIR INTRODUCED THROUGH PORT 3

react with hydrocarbon radicals, primarily CH from the reburning fuel, to produce intermediate species like NH₃ and HCN (Lanier et al., 1986). The HCN then decays through several reaction intermediates, ultimately producing N₂ via the reverse Zeldovich reaction; and c) the Burnout Zone: In this final zone, air is added to produce overall lean conditions and oxidize all remaining fuel fragments. The total fixed nitrogen species (TFN = NH₃ + HCN + NO + Char N) will either be oxidized to NO_x or reduced to molecular nitrogen.

The amount of nitrogen from the fuel contained in the volatiles and char will change as a function of temperature, nitrogen content and size of the coal particle. Figure 3.1.33 shows the wall temperature profiles for various cofiring configurations. A test was performed with 100% PC but water, equivalent to a 30% CWSF test, was injected in port 3 to separate the effect of NO_x changes due to a temperature decrease and any reburning reactions. Since all the fuel and combustion air was admitted through the burner, the resulting temperature, in the case of 100% PC, was higher than for the rest of the tests. This contributes to higher thermal and fuel NO_x formation in the primary combustion zone. Partitioning of nitrogen between the volatiles and char phase is a function of temperature, heating rate, coal type and equipment used. It has been

shown in the literature that total volatile and nitrogen yields are approximately proportional at comparatively low pyrolysis temperatures (1,800°F) and short heating times, but pyrolysis at temperatures greater than 2,500°F results in the evolution of a significant amount of additional nitrogen (Gibbons et al., 1995). Since the temperature for 100% PC in the Port 1 region is high (>2500 °F), it is likely that more nitrogen is released into the gas phase which, with more air, tends to form higher amounts of NO_x . However, there is a reduction in NO_x concentration between Ports 1 and 3. This is primarily due to gas phase reaction involving hydrocarbon species from the volatiles.

Figure 3.1.33 TEMPERATURE PROFILE FOR VARIOUS COFIRING CONFIGURATIONS

Figure 3.1.34 shows the carbon burnout as a function of distance from the burner. These data were obtained from the isokinetic char samples that were collected during the tests. It is clear that the carbon burnout during cofiring tests was lower than that for the 100% PC test. This implies that a lesser amount of carbon was available in the form of HC radicals during the initial stages of cofiring than for the 100% PC run. Figure 3.1.35 shows the percentage of

Figure 3.1.34 CARBON BURNOUT AS A FUNCTION OF DISTANCE FROM THE BURNER

Figure 3.1.35 PERCENT NITROGEN REMAINING IN THE SOLID SAMPLE AS A FUNCTION OF DISTANCE FROM THE BURNER

original nitrogen that remained in the solid phase (char). The chars collected show higher amounts of carbon, and up to 18% of total nitrogen remaining in the char phase at Port 3. Wendt (1995) shows that under rich conditions, HCN plays a critical role in driving the nitrogen cycle to form N_2 and that one source of HCN formation is the destruction of NO by hydrocarbon radicals, which is the main reburning reaction destroying NO. In addition, Lanier et al. (1986) and Mereb and Wendt (1990) have shown that the N_2 fixation reaction also produces HCN, especially under natural gas reburning conditions. Mereb and Wendt (1994) confirmed the fact that reburning with pulverized coal is also effective but less than with gas because natural gas produces more hydrocarbon species in the NO_x -reduction zone and reactions between NO and hydrocarbons are important. Their data showed that with coal reburning, HCN was destroyed more rapidly than it was being formed and its values decayed to levels below 20 ppm within 0.6 seconds. This result supported their hypothesis that it is important to maintain HCN formation rates during long time scales in order to keep N_2 formation mechanisms going, that this formation is mostly due to hydrocarbon reactions with NO or N_2 and that any slow release of nitrogen from the coal residue is a minor contributor to the process. The results in the present study also confirm these observations since the residence time between Ports 1-4 is approximately 0.6-0.7 seconds. The gas phase reburning after this port appears to be insignificant.

NO is expected to be the primary product of char nitrogen combustion (Thomas, 1997). However, the NO-carbon reaction can reduce the NO to N_2 . NO reduction is enhanced by the presence of CO (Thomas, 1997). From the CO measurements in the combustor in this study, no significant difference in the CO in the gas phase emissions was observed. In most of the tests the CO concentration was around 100-120 ppm. Therefore, it is likely that the NO produced from oxidation of char nitrogen reacts with the carbon in the pore structure. Char produced from the CWSF is texturally different from char produced from PC (Pisupati et al., 1996). Therefore, local NO-char reactions are responsible for the small reduction in NO above and beyond the reduction obtained due to the temperature reduction caused by water addition.

The reburning zone stoichiometry is controlled by the reburn fuel heat input and different air staging configurations. Figure 3.1.36 shows the effect of reburn fuel fraction on NO_x reduction for CWSF and other fuels, combined with two different air staging configurations: all combustion air at the burner and make-up air addition downstream at Port 3, respectively. Each

Figure 3.1.36 EFFECT OF REBURN FUEL HEAT INPUT ON NO_x REDUCTION FOR DIFFERENT FUELS AND AIR-STAGING CONFIGURATIONS

separate case proved effective in NO_x reduction, with reduction levels between 20 and 40 %.

Propane and natural gas are the most effective reburn fuels at the same residence times because they need no time for devolatilization. The CWSF and pulverized coal are also effective in reducing NO_x from the primary zone; however they are less effective than the gaseous fuels, most likely because they add additional reactive fuel nitrogen. The maximum NO_x reduction for CWSF was achieved at 30% reburn fuel input. Reburning combined with burner air staging gave greater NO_x reduction than combined with Port 3 air staging. This may be regarded as a result of changing stoichiometry in the reburning zone due to different air staging configurations. Also, in both CWSF cases, there is a minimum in the NO_x reduction at 20% reburn fuel input. The explanation for this is due to the different stages of the reburning mechanism, as explained by Smart and Morgan (1994). For all the Port 3 air staging tests, the primary fuel was pulverized coal. It can be seen that under the studied test conditions, the effect of reburn fuel type on the NO_x reduction efficiency is important, differences of up to 25% being noticed. Also, it was observed that a greater reduction in NO_x emissions occurred when reburning CWSF fuel than

when injecting the same quantity of water at the same firing rate. This indicates that the reduction in NO_x emissions is not due solely to the lower temperature caused by the water present in the CWSF. The CWSF nitrogen content, volatile matter, and ability to form CH fragments appear to be parameters of most importance.

From Figures 3.1.26-3.1.32, it can be seen that there is a significant difference between the NO_x concentrations measured at the center and the wall. This indicates the lack of mixing for about 7-8 feet in the combustor. If the NO_x molecules do not mix with hydrocarbon radicals, the reduction in NO_x will be reduced. It has been shown (Mereb and Wendt, 1990) that the reburning mechanisms occur in two regimes: one in which fast reactions between NO and hydrocarbons are usually limited by mixing; the other in which reactions have slowed and in which known gas phase chemistry controls. A separate effort was launched to model the fluid dynamics in the combustor and its role in the reduction of NO_x in addition to the homogeneous and heterogeneous phase chemical reactions.

3.1.3.5 Conclusions

The results of this study confirm the potential of CWSF as a reburn fuel for significant NO_x reduction. Basic reburning performance of CWSF was found to be close to that of natural gas, with 39% NO_x reduction achievable at a CWSF reburn heat input of 30%. Reburning effectiveness depends strongly upon the reburn zone stoichiometry, which is controlled by the reburn fuel heat input and by the air staging configuration.

The results indicate that NO_x is produced during initial phases in pulverized coal flames. The NO_x concentration profile indicates that the concentration decreases rapidly by reburning in the gas phase by the radicals in the volatile phase. Excess air levels (air staging) and mixing appeared to be the most important parameters in reducing the NO_x in the primary zone. When CWSF was cofired with coal, the temperature in the upper part of the combustor was lower because of water addition. The NO_x concentration near the Port 1 region of the combustor was also lower. This reduction was attributed to the lower temperature and lower fuel nitrogen split into the gas phase. Reduction of NO_x due to reburning in the primary zone was not observed for the CWSF tests. The carbon burnout was lower for the CWSF cofiring tests. Therefore, lower amounts of carbon were available in the volatile phase to reduce NO_x initially. It is believed that

gas phase reactions did not play a significant role in the reduction of gas phase NO_x that was produced initially.

3.2 Polycyclic Aromatic Hydrocarbons

3.2.1 Introduction

The Clean Air Act Amendments (CAAA) of 1990 contain provisions which will set standards for the allowable emissions of 188 analytes designated as hazardous air pollutants (HAPs). This list of HAPs was used to establish an initial list of source categories for which EPA would be required to establish technology-based emission standards, which would result in regulated sources sharply reducing routine emissions of toxic air pollutants. Polycyclic organic matter (POM) has also been referred to as polynuclear or polycyclic aromatic compounds (PACs). Nine major categories of POM have been defined by the EPA (Brooks, 1989). The study of organic compounds from coal combustion is complex and the results obtained so far are inconclusive with respect to emission factors (Curtis et al., 1993). The most common organic compounds in the flue gas of coal-fired power plants are polycyclic aromatic hydrocarbons (PAHs). Furthermore, EPA has specified 16 PAH compounds as priority pollutants. These are naphthalene, acenaphthylene, acenaphthene, fluorene, phenanthrene, anthracene, fluoranthene, pyrene, chrysene, benz[a]anthracene, benzo[b]fluoranthene, benzo[k]fluoranthene, benzo[a]pyrene, indeno[1,2,3-cd]pyrene, benzo[ghi]perylene, and dibenz[ah]anthracene.

3.2.2 Objectives

The objectives of this portion of the study were to measure the emissions of PAHs from a laminar flow, bench-scale drop-tube reactor and from Penn State's demonstration-scale facility. The effects of temperature and oxygen concentrations on the formation and destruction of PAHs were studied. Monitoring was limited to the 16 EPA-specified priority hazardous pollutants. Results from similar studies undertaken in Penn State's pilot-scale research boiler, down-fired combustor, and bench-scale reactor were reported in the Phase II final report (Miller et al., 2000) and elsewhere (Pisupati et al., 2000).

3.2.3 Description of Equipment and Experimental Matrix

3.2.3.1 Drop-Tube Reactor

The electrically-heated drop-tube reactor (DTR) can simulate the heating rate, temperature profile and particle residence time of a utility boiler. The DTR has an alumina muffle tube, which is 6.35 cm in internal diameter and 95.25 cm long, with a maximum temperature ($1,550^{\circ}\text{C}$) zone of 50.8 cm. Coal is delivered vertically into the reactor. Selected operating temperatures for this study were between 800°C and $1,400^{\circ}\text{C}$. A coal feed rate of 0.33 ± 0.01 g/min and a primary air flow rate (to entrain the coal into the reactor) of 1.0 ± 0.1 l/min were used. An additional 3.0 ± 0.2 l/min of secondary air preheated to 250°C was introduced making the total combustion air flow 4.0 ± 0.2 l/min, which corresponds to an air to fuel ratio of approximately 1.25/1 for the selected coal feed rate. A second series of tests was conducted at the same coal feed rates and temperatures with the combustion air reduced to produce an air to fuel ratio of approximately 1/1. Residence time for individual particles in the DTR was approximately one second. The DTR is shown schematically in Figure 3.2.1.

3.2.3.2 Demonstration Boiler

Penn State operates a 15,000 lb steam/h boiler system in which demonstrations of coal-based fuels fired in industrial boilers are conducted, hardware development is performed, and emissions characterization and their control are tested. The boiler is a D-type design watertube boiler manufactured by Tampella Power Corporation, rated for 15,000 lb/h saturated steam (@300 psig), and integrated into the University's steam distribution system. The boiler was designed to fire fuel oil but the overall system has been modified to fire coal-based fuels (i.e., coal-water slurry fuel (CWSF) and dry, micronized coal). Modifications were made to the original system to provide combustion air and CWSF preheating, fly ash removal from the flue gas using either a conventional fabric filter baghouse or a ceramic filter system, coal unloading, storage, and crushing, and either CWSF preparation or dry coal micronization. Figure 3.2.2 shows the micronized coal-fired boiler system with both the pulse-jet baghouse and ceramic membrane filter chamber. The on-line coal handling and processing equipment is shown in Figure 3.2.3 and described in detail in the Phase II final report (Miller, et al., 2000)

The system is located in two buildings at University Park's East Campus Steam Plant except for the baghouse and ceramic filter chamber, which are located outdoors. The boiler and

Figure 3.2.1 SCHEMATIC DIAGRAM OF THE DROP-TUBE REACTOR

Figure 3.2.2 GENERALIZED DIAGRAM OF PENN STATE'S MICRONIZED COAL-FIRED DEMONSTRATION BOILER SYSTEM WITH A CERAMIC MEMBRANE FILTER CHAMBER AND PULSE-JET BAGHOUSE

Figure 3.2.3 GENERALIZED DIAGRAM OF PENN STATE'S MICRONIZED COAL-FIRED DEMONSTRATION BOILER SYSTEM WITH A PULSE-JET BAGHOUSE

heat pipe heat exchanger are located in the boilerhouse, which is adjacent to Penn State's East Campus Steam Plant. The coal processing and handling equipment is located in the Fuel Preparation Facility and includes a 25-ton coal hopper, a cage mill, several magnets, a Redler conveyor, a 3-ton surge bin, a weigh-belt feeder, a MicroEnergy Systems, Inc. (TCS) industrial mill, and a booster fan to pneumatically transport the coal approximately 150 feet from the booster fan to the burner front in a four inch schedule 40 pipe contained in an underground trench.

The boiler's combustion chamber and convective pass have not been modified specifically for coal firing except that several ports were installed for flame observation and sampling. The furnace is approximately 8 ft long by 6 ft wide, with a volume of 370ft³. The bulk gas residence time in the furnace is less than one second. The boiler contains 1,730 ft² of effective heating surface.

Four sets of operating conditions involving changes in excess oxygen and fuel firing rate were studied. Tests performed at 5.2% excess oxygen were termed 'High-Air' (HA). Tests performed at 3.7% excess oxygen were termed 'Low-Air' (LA). Tests performed at 14.5 million (MM) Btu/h feed rate were termed 'High-Fire' (HF). Tests performed at 11.0 MM Btu/h feed rate were termed 'Low-Fire' (LF). The four sets of tests were conducted by varying these parameters. For each test, flue gas samples were simultaneously collected from sampling ports located both upstream and downstream of the cloth filter baghouse. All other parameters involving routine operation of the demonstration boiler were held constant and standard boiler operating procedures were followed.

3.2.3.3 Fuel Analysis

The fuel used for these studies was a Middle Kittanning seam coal micronized to $\approx 80\%$ passing US Standard 325 mesh (commonly referred to as a micronized coal, or "MC grind"). Tables 3.2.1 and 3.2.2 show the analysis of the Middle Kittanning feedstock coal.

3.2.4 Analytical Methodology

A majority of the analytes listed in the CAAA, including PAHs, can be collected by what is commonly referred to as a Modified Method 5 (MM5) sampling train. The train is used to

Table 3.2.1 Proximate Analysis and Higher Heating Value

Sample	% Moisture (Air Dried)	% Ash (Dry)	% V.M. (Dry)	% Fixed C (Dry)	HHV(Btu/lb) (Dry)
Middle Kittanning	0.80	5.24	30.72	64.04	14,350

Table 3.2.2 Ultimate Analysis

Sample	% Carbon (Dry)	% Hydrogen (Dry)	% Nitrogen (Dry)	% Sulfur (Dry)	% Oxygen (Dry)
Middle Kittanning	80.41	4.84	1.41	0.72	7.51

isokinetically collect samples at desired locations in flue gas streams. The collected sample can be separated into solid, condensed liquid and gaseous phases. The PAHs of interest are extracted from the collected sample, concentrated, then separated and quantified by gas chromatography/mass spectrometry (GC/MS).

A modular sampling train meeting all of the requirements of EPA Methods 5 and 0010 was utilized for sampling. EPA Methods 1 through 4 were utilized to determine various parameters needed for isokinetic sampling of the DTR. EPA Method 8270-C was selected for determining the concentrations of PAHs in the collected samples. Extraction of PAHs from the collected particulate matter and the XAD-2 polymeric resin fractions was done by EPA Method 3540-C, Soxhlet Extraction. Extraction of PAHs from the collected liquid fraction was done by EPA Method 3510-C, Separatory Funnel Liquid-Liquid Extraction. Sample cleanup was accomplished by EPA Method 3630-C, Silica Gel Cleanup.

The EPA Method 5 sampling train is designed to isokinetically sample a flue gas stream. A portion of the gas stream was withdrawn from the bottom of the DTR through a heated probe where the particulate matter was filtered out of the stream. The remaining portion of the gas stream was passed through a condenser, a module containing the polymeric resin, and a series of glass impingers immersed in ice water where the remaining condensable analytes were collected.

Specifically, the sampling train contained the following components. They are listed from the most upstream component to the most downstream component.

A straight stainless steel nozzle is located at the end of a sampling probe. The nozzle opening is placed into the flue gas stream facing upstream with pitch and yaw angles of 0°. The inside diameter of the nozzle is precisely known, as its size affects the sampling rate.

Connected to the nozzle by a Teflon® ferrule is a probe of borosilicate glass surrounded by a stainless steel jacket. The tip of the nozzle was introduced 5 cm into the bottom of the DTR. The stainless steel jacket contains heating elements. While sampling, the probe was maintained at 250°F by a temperature controller. The seal between the DTR/probe interface was made airtight from the surroundings using insulation.

The probe was connected to a borosilicate glass filter assembly located within a heated oven. The filter assembly was also maintained at 250°F. A filtering media was selected to retain particles larger than 0.2 µm.

Gases exiting the filter assembly were passed through a water-cooled glass condenser to condense PAHs that may be in the vapor phase. This condensate and remaining gas phase components were passed over the water-cooled porous polymeric resin contained in a glass module.

Gases and condensed liquids were passed through the glass resin module into a chilled condensate knockout trap. This trap was similar in design to an empty impinger. Many analytes not retained on the resin were collected in this knockout impinger.

The next three downstream chilled impingers were used to collect additional water condensate. The final impinger contained indicating-type silica gel to protect the pump from unwanted moisture. The temperature of the gas stream exiting this final impinger was maintained at or below 68°F. All glass ball-and-socket connections were clamped and made airtight with Teflon® O-rings.

The gases were passed through an air tight pump, dry gas meter, across a manometer, (gas samples may be collected at this point if desired) and vented. These devices, along with thermocouple readouts, temperature controllers, manometers, valves, timer and other equipment were contained in a metering console.

3.2.5 Operation of the MM5 Sampling Train

Prior to sampling, data sheets were prepared and leak checks were performed on the sampling system. Proper temperatures and flow rates were maintained. Leak checks were also

performed at the conclusion of the sampling runs. All other appropriate EPA sampling procedures were adhered to, and all samples were collected within the acceptable sampling range of 90 to 110% of isokineticity.

3.2.6 Sample Recovery and Preparation for Analysis

The collected sample can be separated into six (or more) fractions. The subsamples contained materials collected from these particular sections of the sampling train. Typically, the first subsample contained recovered materials from the nozzle to the filter assembly. The second subsample contained the filtering media, filtered solids, and other solids removed from the filter assembly. The third subsample contained recovered materials from the back half of the filter paper assembly and the condenser. The fourth subsample contained recovered materials from the resin module. The fifth subsample contained recovered materials from the condensate knockout trap. The sixth and final subsample contained recovered materials from the silica gel-filled impinger. These six subsamples can be processed and analyzed separately or combined in any manner and analyzed. Five of the six subsamples were combined into two samples in the following manner.

The first of the two samples was generated using the following procedure. The recovered particulate matter from the front half rinse and the recovered particulate matter from the filter and the filter itself were spiked with a known amount of a surrogate spiking solution and transferred into a glass soxhlet extraction thimble. The surrogate compounds are compounds chemically similar to the analytes of interest, but not expected to be contained in the extract. The surrogate compounds were utilized to monitor unusual matrix effects or sample processing errors during the extraction and recovery process. The glass thimble was placed into a soxhlet extractor and connected to a round bottom flask containing approximately 10 clean Teflon® boiling chips and enough methylene chloride to maintain extraction cycling. A bulb type condenser was connected atop the extractor. A heating mantle was placed under the flask and adjusted to cycle the extractor about once every thirty minutes for approximately 18 hours.

The remaining front half rinse was quantitatively transferred to a separatory funnel, and serially extracted three times with methylene chloride. The pH of the rinse was then adjusted to at least 11 with sodium hydroxide and serially extracted three more times. The pH was then adjusted to below 2 using sulfuric acid and serially extracted three additional times. This extract

was combined with the soxhlet extracted material. A Kuderna-Danish (K-D) concentrator was assembled by attaching a 10 ml concentrator tube with Teflon ® boiling chips to a 500 ml evaporatory flask. The extracts were passed through a sodium sulfate filter to remove any residual water and transferred into the concentrator. A prewet three-ball macro Snyder column was placed atop the concentrator. The concentrator assembly was placed in a hot water bath and concentrated to 6-8 ml within 30 minutes. The apparatus was allowed to cool. The three-ball macro column and evaporatory flask were removed and replaced with a two-ball micro Snyder column and the extraction was further concentrated to approximately 4 ml. The extract was then quantitatively transferred to a vial, diluted to a final volume of 5 ml, and stored at or below 4°C until analyzed.

The second sample was generated using the following procedure. The condensate and condensate rinse samples were combined and transferred to a separatory funnel, spiked, and serially extracted with methylene chloride exactly as was done with the front half rinse.

The porous polymeric resin was transferred to a glass extraction thimble, spiked and processed identical to the particulate matter sample. The condensate and resin extracts were combined, residual water removed, and then concentrated as described above.

Blank samples were extracted concurrently with the collected samples. All samples were spiked with internal standards just prior to analysis.

3.2.7 GC Analysis

Prior to GC/MS analysis, a 1 µl aliquot of the samples was analyzed on a Perkin Elmer 8500 GC system. The system was equipped with a J&W Scientific DB-5 capillary column. This screening procedure checked the concentration of the analytes and determined if dilution or further concentration of the samples was necessary. The GC screening showed that the samples needed to be concentrated to a final volume of 1 ml. Final concentration of the samples from 5 ml to 1 ml was performed as described earlier.

3.2.8 GC/MS System, Calibration and Operation

A GC/MS system was used for extract analysis. The system was temperature programmable and had a splitless injection option. A fused-silica capillary column capable of PAH separation was used. Other GC/MS equipment met specifications, including GC/MS

interface, data acquisition system and other ancillary equipment as described in EPA Method 8270.

The GC/MS system was properly hardware-tuned. Background subtraction techniques designed to eliminate column bleed or instrument background ions were available but not utilized as they were not necessary. GC column performance was evaluated and injection port inertness was confirmed. Calibration standards were analyzed and their response factors were calculated and evaluated. A system performance check using approved System Performance Check Compounds was performed and the results evaluated. The percent relative standard deviation of the compounds was calculated as was the linearity of the calibration curve. When the system tuning, the calibration check, the performance check, and the internal standard response data met minimum method requirements, analysis of the samples began. An external calibration curve was generated as a check against the internal calibration.

A Hewlett Packard (HP) Model 5890 Series II Gas Chromatograph was used for analyte separation. The system was equipped with a 30 m x 0.25 mm ID, 1 μ m film thickness silicone-coated, fused-silica capillary column; a J&W Scientific DB-5 capillary column. The GC temperature program held the initial temperature at 40° C for 5 minutes to remove the methylene chloride solvent. The temperature was ramped to 280° C at a rate of 6°C per minute and held there for 18 minutes. The Helium carrier gas flow rate was 30 ml/min. This flow rate is lower than that recommended by Method 8270C, causing the analytes of interest to elute at a later time than published in the Method, but this did not present a problem for either identification or quantification. After the system parameters were set, a spiked 1 μ l aliquot of concentrated extract was injected (splitless) into the GC/MS system. A HP 5971A Mass Selective Detector was interfaced with the GC scanning up to 500 amu per second at 70 volts. The qualitative identifications of compounds determined by this method are based on retention time in the GC and on comparison of the sample's mass spectrum with characteristic ions in a reference mass spectrum. The reference library used was the NBS49K library. Compound identification was not hampered by component coelution; that is, all analytes of interest were sufficiently separated chromatographically. When a compound had been identified, the determination of analyte concentration was based on the integrated abundance from the Extracted Ion Current Profile (EICP) of the primary characteristic ion.

3.2.9 Results

3.2.9.1 Drop-Tube Reactor Results

PAHs in quantifiable amounts were found in all samples collected. Additionally, PAHs were also found in a sample generated by extracting 1g of Middle Kittanning feedstock coal. Table 3.2.3 shows the concentrations in micrograms per megajoule ($\mu\text{g}/\text{MJ}$) of PAHs in the extracts from the recovered solid subsamples at an air to fuel ratio of 1.25/1, and from the feedstock coal. Note that multiplying $\mu\text{g}/\text{MJ}$ by 2.33 converts the results to pounds per trillion Btu, another common basis the results are reported in. Table 3.2.4 shows the concentrations in $\mu\text{g}/\text{MJ}$ of the PAHs from the recovered liquid and condensed gaseous phase subsamples at an air to fuel ratio of 1.25/1. Table 3.2.5 shows the concentrations in $\mu\text{g}/\text{MJ}$ of PAHs in the extracts from the recovered solid subsamples at an air to fuel ratio of 1/1. Table 3.2.6 shows the concentrations in $\mu\text{g}/\text{MJ}$ of the PAHs from the recovered liquid and condensed gaseous phase subsamples at an air to fuel ratio of 1/1.

Table 3.2.3 PAH Concentrations ($\mu\text{g}/\text{MJ}$) from Solid Phase Extracts at Air-to-Fuel Ratio of 1.25/1

Analyte	800°C	900°C	1,000°C	1,100°C	1,200°C	1,300°C	1,400°C	Coal
Naphthalene	49	44	50		17		17	39
Acenaphthylene			39					
Phenanthrene	21	23	27	11	18	9	24	17
Fluoranthene			17	5	12			
Pyrene			17	15	23			

Table 3.2.4 PAH Concentrations ($\mu\text{g}/\text{MJ}$) from Condensed Vapor Phase Extracts at Air-to-Fuel Ratio of 1.25/1

Analyte	800°C	900°C	1,000°C	1,100°C	1,200°C	1,300°C	1,400°C
Naphthalene	420	4,871	4,674	7,027	11,371	11,736	4,518
Acenaphthylene			232	737	954	491	73
Fluorene					240	209	92
Phenanthrene				39	264	475	
Anthracene						288	
Fluoranthene				27	35	523	30
Pyrene				26	30	376	
Chrysene						253	103

Table 3.2.5 PAH Concentrations ($\mu\text{g}/\text{MJ}$) from Solid Phase Extracts at Air-to-Fuel Ratio of 1/1

Analyte	800°C	900°C	1000°C	1100°C	1200°C	1300°C	1400°C
Naphthalene	5	69	5	9	12	3	3
Acenaphthylene					1		
Flourene	1	1		1	1		
Phenanthrene	33	13	42	62	51	2	13
Anthracene	1	1		3	3	1	1
Fluoranthene	19	3	4	19	14	2	26
Pyrene	18	3	12	18	15	1	6
Benzo[a]anthracene	16		15	21	13	1	7
Chrysene	39	3	46	76	47	5	16
Benzo[b]fluoranthene	14		17	31	10	1	15
Benzo[k]fluoranthene	14		9	14			9
Benzo[a]pyrene	18		9	46	6		
Indeno[1,2,3-cd]pyrene	19			5			
Dibenzo[a,h]anthracene	1			18	1		
Benzo[g,h,i]perylene	33			39	1		

Table 3.2.6 PAH Concentrations ($\mu\text{g}/\text{MJ}$) from Condensed Vapor Phase Extracts at Air-to-Fuel Ratio of 1/1

Analyte	800°C	900°C	1000°C	1100°C	1200°C	1300°C	1400°C
Naphthalene	1185	875	1090	964	934	652	1023
Acenaphthylene	12			6	12	9	28
Acenaphthene				3	3		1
Flourene	12	3	15	11	19	9	13
Phenanthrene	45	5	64	54	113	59	91
Anthracene	5	1		5	15	6	11
Fluoranthene	61	8	12	6	25	18	56
Pyrene	58	7	16	6	14	17	30
Benzo[a]anthracene	23	2	15	9	30	11	6
Chrysene	32	5	33	22	60	24	10
Benzo[b]fluoranthene	22			6	21	13	
Benzo[k]fluoranthene	23			3	16		
Benzo[a]pyrene		24			12		
Indeno[1,2,3-cd]pyrene	32						
Benzo[g,h,i]perylene	63				15		

Combustion efficiencies were also calculated from analysis of the feedstock coal samples and the collected char samples. Ash content was determined for all samples in a LECO MAC 400 Thermogravimetric Analyzer. Combustion efficiencies were calculated using ash as the tracer. The equation used is:

$$\Delta W = \{1 - [A_o(100 - A_c)] / [A_c(100 - A_o)]\} * 100$$

Where ΔW = Per cent weight loss calculated on a dry ash free basis (% Burnout)

A_o = ASTM ash of the dry feedstock

A_c = ASTM ash of the dry char

The assumption involved in using this equation is that the mineral matter in the coal does not undergo transformations which would change the quantity of ash produced upon ashing the chars. Table 3.2.7 shows the combustion efficiencies for the two air to fuel ratios investigated in the DTR.

Table 3.2.7 DTR Combustion Efficiencies for Varying Air to Fuel Ratios

Temperature	800°C	900°C	1000°C	1100°C	1200°C	1300°C	1400°C
% Burnout, 1.25/1	48.5	65.6	71.4	82.4	77.2	82.1	79.7
% Burnout, 1/1	39.9	55.2	64.3	69.1	70.2	68.4	70.1

3.2.9.2 Demonstration Boiler Results

PAHs in quantifiable amounts were found in all samples collected. Table 3.2.8 shows the concentrations in $\mu\text{g}/\text{MJ}$ of PAHs in the extracts from the samples collected upstream of the baghouse. Table 3.2.9 shows the concentrations in $\mu\text{g}/\text{MJ}$ of the PAHs from the samples collected downstream of the baghouse.

Table 3.2.8 PAH Concentrations ($\mu\text{g}/\text{MJ}$) from Condensed Vapor Phase Extracts Collected Upstream of the Baghouse

Analyte	HF-HA	HF-LA	LF-HA	LF-LA
Naphthalene	74	31	58	100

Table 3.2.9 PAH Concentrations ($\mu\text{g}/\text{MJ}$) from Condensed Vapor Phase Extracts Collected Downstream of the Baghouse

Analyte	HF-HA	HF-LA	LF-HA	LF-LA
Naphthalene	92	72	76	43
Phenanthrene	2	2	2	1

3.2.10 Discussion

3.2.10.1 Discussion of Drop-Tube Reactor Results

Of the 16 PAH species listed by the EPA as HAPs, only naphthalene and phenanthrene were found to be extractable from the feedstock coal. Since most PAH species are not present as such in the feed, it is not possible to conduct a material balance. It was necessary to investigate the effects of temperature and oxygen concentration on the formation and subsequent destruction of the PAHs. Two series of experiments were conducted across a temperature range of 800°C-1400°C at two different air to fuel ratios. The PAH species were measured from samples in both the solid and condensed vapor phases as described above. PAHs were found in quantifiable amounts in all samples collected.

For the experiments conducted at an air to fuel ratio of 1.25/1, seven of the sixteen HAP PAHs were detected (Miller et al., 2000; Pisupati et al., 2000). It was observed that the vapor phase PAH emission concentrations increased with temperature to about 1,300°C and then dropped dramatically. The emission values reported for the vapor phase PAHs, and in particular those for naphthalene, are extremely high. At 1,300°C, the naphthalene concentration of 11,736µg/MJ represents just under 0.04% (by weight) of the thermal input of the coal. Since the carbon burnout values for these samples are much lower than found in utility boilers, it is not surprising to find the bench scale PAH emissions much higher than those reported from utility boilers (Szpunar, 1992) and other earlier studies including pilot scale combustors (Davies et al., 1992; Pisupati, et al., 1998; Miller et al., 2000).

In contrast to the vapor phase PAHs concentrations, solid phase char sample extracts contained small amounts of PAHs throughout the temperature range. This suggests that when PAHs either form in the solid phase or are liberated from the solid phase, they easily lose their affinity for the solid phase and rapidly enter the vapor phase. It has been reported (Miller et al., 2000; Pisupati, et al., 2000) that up to approximately 1,300°C, the rate of PAH formation and release from the solid phase for this coal exceeds the rate of PAH oxidation in the vapor phase.

There were many monomethyl-, dimethyl- and trimethylnaphthalenes and other similar PAH derivatives identified in the vapor phase of these samples. Their concentrations were not quantified. As the temperature was increased from 800 to 1,200°C, the methyl- aliphatic side chains on the PAHs derivatives were the predominant functional group. At 1,300°C these side chains were converted predominantly to carboxylic acid functional groups, signaling the first

stages of PAH oxidation. Above 13,00°C, the rate of PAH oxidation had exceeded the rate of PAH formation, and the PAH concentrations decreased rapidly. Different phenomena were observed at an air to fuel ratio of 1/1.

For the experiments conducted at an air to fuel ratio of 1/1, thirteen of the sixteen HAP PAHs were detected. With the exception of the naphthalene, there are not major differences found in PAH concentrations between the solid phase and vapor phase samples at particular temperatures. There also does not seem to be a trend of individual PAH concentrations varying with temperature. The total amount of quantified PAHs at all investigated temperatures (with the exception of 800°C) is lower in the 1/1 air to fuel ratio sample suite than in the 1.25/1 air to fuel ratio sample suite in both concentration and total mass.

Along with the PAH concentrations, concentrations of the derivative PAH compounds are important to consider. In the suite of experiments where the air to fuel ratio was 1.25/1, the most predominant side chain functional group at lower temperatures was the methyl group. As the temperature increased, the methyl group was oxidized to a carboxylic acid group. In the suite of experiments where the air to fuel ratio was 1/1, the predominant non-oxygen containing side chain functional group at the lower temperatures was 2-propylene. As the temperature increased, the predominant non-oxygen containing side chain functional group became ethylene. These types of side chains can help create larger PAH molecules from smaller ones, and can do so more readily in low oxygen environments. The laminar flow profile of the combustor and the 1/1 air to fuel ratio should create localized areas in the combustor with low concentrations of available free oxygen. Derivative PAH compounds can then more easily become larger PAHs.

In addition to the 16 EPA HAP PAHs, other large PAHs and their derivatives were detected. Anthanthrene, benzo(a)chrysene, and dibenzo(ah)pyrene and their derivatives were detected in appreciable, but unquantified amounts. These large PAHs are hazardous but are not in the EPA HAPs list, probably because they have not been detected in utility boiler flue gas streams.

Oxygen concentration is an important variable in PAH formation. The data suggests that lower oxygen concentrations during combustion leads to higher molecular weight PAH formation along with their precursors and derivatives. The data also suggests that lower oxygen concentrations also retard the release of PAHs and their derivatives from the solid phase to the vapor phase.

Quantified PAH emissions reported from a pilot scale DFC showed only fluoranthene and acenaphthene in detectable limits for the same coal (Miller et al., 2000; Pisupati et al., 2000). These two sets of data show that, at least above 1300°C, the vapor-phase PAH concentrations continue to decrease rapidly with increased particle residence time and increased mixing of air and fuel.

3.2.10.2 Discussion of Demonstration Boiler Results

None of the 16 analytes of interest were found in quantifiable amounts in the solid phase extracts from the samples collected either upstream or downstream of the baghouse. Previously reported data from pilot-scale combustors show slight increases in PAH concentrations with increasing excess-air to fuel ratios, and slight decreases in PAH concentrations with increased firing rate (Miller et al., 2000; Pisupati et al., 2000). Data from the demonstration boiler show slight decreases in PAH concentrations with increasing excess-air to fuel ratios, and slight increases in PAH concentrations with increased firing rate. It is important to note that two of the four data set pairs from the demonstration boiler (i.e., Low Air and High Fuel) are within EPA sampling reproducibility limits. The other two pairs are just above EPA sampling reproducibility limits. The numerical differences from the pairs of demonstration boiler data may be statistically insignificant, and if so the reported trends may also be insignificant. The important differences are found in the types of PAH derivatives present in the demonstration boiler samples and total demonstration boiler PAH concentrations.

As discussed above, lower excess oxygen concentration in the DTR combustor leads to predominantly longer side chains on the PAHs. The only PAH side chains identified for the High Air samples were the methyl and carboxylic acid functional groups. These unquantified PAH derivatives were detected at relatively low concentrations. The predominant PAH side chains identified for the Low Air samples were the methyl, ethyl, carboxylic acid and acetic acid functional groups. These unquantified PAH derivatives were detected at significantly higher concentrations than their High Air counterparts. Therefore similar trends were noted in the types of PAH derivatives between the low (1/1) air to fuel ratio data from the DTR and the Low Air demonstration boiler data, as well as between the high (1.25/1) air to fuel ratio data from the DTR and the High Air demonstration boiler data.

The total amount of PAHs reported from the demonstration boiler is well within the range reported for the down-fired combustor and only slightly higher than the research boiler (Miller et al., 2000; Pisupati et al., 2000). It may well be prudent not to try to over-interpret the demonstration boiler PAH data. There are only minor, possibly statistically insignificant differences between the PAH concentrations reported in many of the samples collected from the demonstration boiler. Other factors that affect the interaction with oxygen to the fuel (particle residence time, gas and temperature profiles), important to PAH formation and release to the vapor phase were not investigated.

3.2.11 Conclusions

A Middle Kittanning seam coal was fired in a bench-scale, laminar flow, drop-tube furnace at a variety of temperatures. Solid, liquid and condensed vapor-phase samples were withdrawn isokinetically from the furnace and analyzed for PAH concentrations using EPA methods. For an air to fuel ratio of 1.25/1, results showed near constant PAH concentrations in the solid-phase samples, while the PAH concentrations in the vapor-phase samples increased as a function of temperature. At a temperature around 1,300°C, it appears that the rate of PAH formation was exceeded by the rate of PAH oxidation, and PAH concentrations in the vapor-phase began to decrease (Miller et al., 2000; Pisupati et al., 2000). For an air to fuel ratio of 1/1, results showed little difference in PAH concentration between the solid phase and vapor phase samples, with the exception of naphthalene. A higher number of PAHs were detected in the lower air to fuel ratio. Increasing temperatures between 800°C and 1400°C did not dramatically affect either PAH formation or destruction. At lowered oxygen concentrations, higher molecular weight PAHs and many more types of PAH derivatives were formed.

A Middle Kittanning seam coal was fired in a demonstration-scale boiler at various air and fuel rates. Solid, liquid and condensed-vapor phase samples were withdrawn isokinetically from the boiler and analyzed for PAH concentrations using EPA methods. Only naphthalene and phenanthrene were detected. The amount of PAHs detected in the demonstration boiler are similar those found in other pilot scale units.

Availability of oxygen is a more dominant factor in PAH formation and destruction than temperature for all units tested. The laminar flow design of the DTR decreases air to fuel mixing as compared to the other combustors investigated. This effectively lowers the amount of

available oxygen to the fuel in the DTR, resulting in significantly higher PAH to fuel ratios than in pilot or demonstration scale combustors.

3.3 Trace Elements

Title III of the Clean Air Act Amendments of 1990 designated 189 hazardous air pollutants (HAPs) (later revised to 188 HAPs. Inorganic HAPs, herein referred to as IHAPs, include the following elements: arsenic (As), beryllium (Be), cadmium (Cd), cobalt (Co), manganese (Mn), nickel (Ni), lead (Pb), antimony (Sb), selenium (Se), and total mercury (Hg). Mercury has been of primary concern given its significant negative health effects on humans and wildlife. The United States Environmental Protection Agency (EPA) initiated an information collection request (ICR) on January 1, 1999 for coal-fired electric utility steam generating units (> 25 MW) to document mercury levels in their fuels and emissions. The issuance of an ICR generally precedes the implementation of regulatory action by the EPA.

In December 2000, the U.S. EPA decided that regulation of mercury from coal-fired power plants was appropriate and necessary under Section 112 of the Clean Air Act due to the significant health hazards it presented to the public. EPA is scheduled to propose regulation of mercury emissions by December 2003 and promulgate them by December 2004. In order to come to this determination, the accurate measurements of mercury and its speciation are necessary.

Penn State conducted three activities in Phase III addressing issues EPA and DOE identified as requiring further study. These activities included evaluating ceramic filters for mercury removal as a control option, evaluating trace elements emissions from deeply-cleaned coals produced through DOE's Premium Clean Coal Program, and developing a sampling train and methodology capable of simultaneously sampling mercury species and all inorganic trace elements. As part of the ceramic filter evaluation, a long-term demonstration of the filters was performed, similar to the one in Phase II of the project, to assess the technical feasibility of using ceramic filters in lieu of fabric filters. The results of these activities are discussed in the following sections.

3.3.1 Fine Particulate Matter/Mercury

Additional efforts in Phase III, focusing on the control of fine particulate emissions when firing coal in an industrial boiler, were added to complement the trace element testing. In Phase II, approximately 900 hours of operation were obtained using ceramic membrane filters on the demonstration boiler (Miller et al., 1999). Although their performance was evaluated in Phase II, additional work was added (in conjunction with DOE) because it was determined near the end of the Phase II demonstration that there were design problems in the system, which resulted in the inability to regenerate the filters. If ceramic filters are to be considered as a control option for fine particulate matter (and mercury capture on the fine particles), it was decided to conduct a long-term demonstration of the filters along with mercury testing. Additional ceramic filter operation was conducted to assess Corning Incorporated worked closely with Penn State to redesign the venturis and chamber. Following modifications of the chamber, Corning Incorporated donated 80 new ceramic filters for a second demonstration testing. The results from the operability assessment and effect of particle control device on trace elements emissions are presented in this section.

3.3.1.1 Ceramic Filter Testing

A near-commercial ceramic filtering system was identified in Phase II of this project, in conjunction with DOE, for installation on the demonstration boiler. The demonstration boiler system is shown in Figures 3.2.2 and 3.2.2 and described in Section 3.2.3.2. The ceramic filtering system consisted of the ceramic membrane filters, chamber to house the filters, components to regenerate the filters, structural supports, walkways, steps and ladders, ducting valves, induced draft fan, and associated controls. The ceramic filtering system was described in detail in the Phase II final report (Miller et al., 2000). Brief descriptions of the major components, including the ceramic filters and filter chamber, as they relate to their continued evaluation in Phase III, are provided in this section.

Description of Ceramic Filter System

Ceramic Filters

The membrane filters, manufactured by Corning Incorporated (Corning), were formed by coating the walls of a cellular ceramic substrate with a patented inorganic membrane. The

membrane has a nominal pore size of $0.5\ \mu\text{m}$, which provides a barrier to particle penetration into the substrate while allowing for easy cleaning during the backpulse cycle.

The substrate is extruded and baked to form a cordierite composition. It has a multitude of parallel square passageways extending from one end face to the other. Ceramic plugs are applied to alternate ends of the substrate channel passageways, preventing the straight-through passage of combustion products. During operation, particle laden flue gas flows into the open passageways at the inlet end face. Particles are collected on the membrane surface and the filtered gas passes through the wall and exits the filter via the downstream end face open passageways. Figure 3.3.1 illustrates the details of the ceramic membrane gas filter.

The ceramic membrane filters are wrapped in a ceramic fiber mat to prevent by-pass of dirty gas and mounted in metal housings to support the filters within the filter chamber. The filters are 7 inches in diameter and 15 inches in length with a filter area of approximately 2,000 square feet.

Figure 3.3.1 DETAILS OF CORNING INCORPORATED'S CERAMIC MEMBRANE GAS FILTER

Ceramic Filter Chamber

The original filter chamber was designed by Penn State with guidance from Corning, manufacturer of the ceramic filters. Installation of the main components of the system was conducted from mid-July through December 1996, with final installation, for such items as insulating the ceramic filter chamber, occurring in January and February 1997.

The chamber houses 80 filters, arranged in two groups of 40, with each group consisting of a common compressed air header for back pulsing. Figure 3.3.2 is a schematic diagram of the interior of the ceramic filter chamber showing the ceramic membrane filters with respect to the flue gas inlet and outlet. Backpulse hardware, similar to that used in pulse jet baghouses, is used to clean the filters. Each row of four filters is serviced by a single blowtube for backpulsing. The system is configured to backpulse two rows (one on each side of the chamber) of filters at a time.

Figure 3.3.2 INTERIOR OF THE CERAMIC FILTER CHAMBER SHOWING BAFFLES AND CERAMIC MEMBRANE FILTERS

Modifications to the Ceramic Filter Chamber

Following the Phase II demonstration it was determined that design problems existed in the original system, which prevented regeneration of the filters. Therefore, modifications were

made to the chamber to correct these problems. The modifications included the following activities:

- Raised the roof to accommodate the new venturis, which are approximately 8 inches taller;
- Added a new roof that is rain and snow proof;
- Cleaned the tube sheet by sand blasting and then added a water-proof epoxy to the metal surfaces on the clean side of the tubesheet;
- Cleaned the compressed air headers and valve bodies;
- Replaced the venturis with new ones manufactured from 304 stainless steel;
- Replaced the blowtubes with ones constructed from 304 stainless steel;
- Replaced the filters and cans (cans are made from 304 stainless steel);
- Installed baffles in the bottom of the chamber;
- Determined, and subsequently used, an optimum regeneration protocol with the new design;
- Used an RTV seal on the roof doors, which will be used each time the chamber is closed;
- Added the capability to check face pressures on regeneration for two filters;
- Developed, and subsequently used, an optimum startup and shutdown protocol; and
- Installed sensors to measure actual flow rate into the chamber.

Ceramic Filter Operation – Second Demonstration

Following modification of the filtering system, preliminary testing of the ceramic filters was resumed in October 1999. Similar to Phase II, a number of operational parameters were monitored/ measured during the continued evaluation, including:

- Pressure drop history of the ceramic filter chamber. This included monitoring both the pressure drop across the chamber as a function of hours of firing coal as well as the pressure drop recovery (i.e., return to a baseline value). The pressure drop, when firing natural gas at 12 million (MM) Btu/h at 4% O₂ and 400 acfm of transfer air, was measured prior to firing coal at the beginning of a day of operation (normal operation consists of either one or two 8-hour shifts of operation) and after coal feed had been terminated and purged from the coal feed lines;
- Periodic checks on the filter face pressure and flow;
- Ceramic filter chamber fine particulate (i.e., PM₁₀ and PM_{2.5}) and condensable particulate matter (CPM) emissions measured by EPA Methods 201A (PM₁₀) modified to include PM_{2.5}, and 202 (CPM);
- Ceramic filter chamber fine particulate collection efficiency; and
- Comparison of particulate emissions and collection efficiency of the ceramic filter chamber and the fabric filter baghouse.

A summary of the burner settings and daily averages of the operating conditions, combustion performance, and boiler efficiencies is contained in Appendix 3A for all Phase III

demonstration boiler operation. This includes days when the fabric filter baghouse was used as well as the ceramic filter chamber.

As part of the second demonstration using the filters, Corning requested that filters be removed periodically (i.e., after 10, 70, 170 hours operation, etc.) and replaced with new ones. Corning was to perform a series of analyses on the filters, some of which were destructive. Likewise, Penn State periodically removed filters and analyzed them by nondestructive X-ray computerized tomography (CT) to: characterize the filters before and after use; evaluate the ability to map out filled channels in the filters; investigate the ability to detect erosion in the filters; and assess the topology of the end plugs.

The chamber was operated in October and November 1999 and approximately 80 hours of operation were obtained. During this time, two filters were removed for Corning at 10 and 80 hours of operation. When the chamber was opened to remove the filters, it was noted that the epoxy on the clean side of the tubesheet was discolored and a liquid residue was dripping onto the venturis and blowtubes. Samples of the material were collected and analyzed. It was found to be rich in zinc and chlorine. Further investigation found that the epoxy had a zinc base. Apparently the chlorine in the coal was reacting with the zinc-based epoxy. After discussions with Corning and the epoxy manufacturer, it was decided to remove the venturis and the filters, send the filters back to Corning for pressure drop determinations (to ensure that material was not trapped in the clean side of the filters which would negatively affect pressure drop), sandblast the upper plenum chamber to remove the remaining epoxy, add a different epoxy, and install a steam coil to keep the temperature of the upper plenum chamber above the dew point of the gases to ensure that moisture will not condense when the unit is shutdown at night. The boiler facility was being operated on a two-shift per day basis and the chamber cooled down at night.

During this investigation, another item was discovered that needed to be replaced. The boiler system has been engineered so that the entire flue gas stream leaving the boiler can be diverted to either the baghouse or ceramic filter. Isolation dampers are located at the inlet and outlet of each particulate removal device. When one particulate removal device is used, the dampers to the other device are closed. This is to ensure that all flue gas flow is being diverted to the device being tested. However, it was found upon close examination (by operating personnel crawling into the 24 inch diameter ducting), that the dampers were not zero-flow

dampers as requested during the design and construction phase of the filter chamber. Consequently, it was decided to replace the dampers with zero-flow valves.

Ceramic Filter Operability

Approximately 930 hours of coal-fired operation were obtained using the ceramic filter chamber during Phase III. An important filtering system operating requirement is the ability to regenerate, i.e., clean the filters so that the pressure drop across them returns to the baseline value. During operation, the particulate matter accumulates on the membrane surface and the pressure drop increases. The ceramic membrane filter is then cleaned by backpulsing with compressed air in a manner similar to cleaning fabric filters in a pulse jet baghouse.

The pressure drop of the ceramic filter chamber was monitored as part of the system evaluation. It was recorded by both the taking of manual readings and using a computerized data acquisition system (DAS).

The typical operating procedure of the boiler begins by backpulsing each filter five times without the forced draft and induced draft fans operating. The boiler system is then preheated firing natural gas at a rate of 12 MM Btu/h until the ceramic filter chamber outlet temperature is ~300°F. The boiler is then transitioned from natural gas to firing micronized coal by reducing the quantity of natural gas in increments while simultaneously increasing the micronized coal feed. With the natural gas valve closed, the coal feed is adjusted to achieve a typical firing rate of 15 MM Btu/h. The system is allowed to stabilize then the DAS begins recording the filter pressure drop (data points are recorded every 30 seconds), while manual readings are taken every half hour. The boiler is typically operated 8 to 12 hours per day. Upon conclusion of testing, the coal feed is turned off, the coal line is purged of coal, and the natural gas rate increased until the firing rate is returned to 12 MM Btu/h. After the coal line has been purged, the natural gas is shut off and the ceramic filter chamber allowed to cool with the forced draft and induced fans running. When the chamber outlet temperature has decreased to 300°F, the fans are shut down and the backpulsing system allowed to cycle until each filter has been backpulsed at least five times.

Figure 3.3.3 is a plot of the ceramic filters' pressure drop as a function of time firing coal from October 25, 1999 through June 17, 2003. The data points are the manual readings taken

- A** Second period of modifications to chamber performed
- B** Stopped testing to re-evaluate increasing ΔP
- C** Slipstream chamber operated for 100 hours
- D** Period of 4 weeks of baghouse operation

Figure 3.3.3 CERAMIC FILTER PRESSURE DROP AS A FUNCTION OF TIME FIRING COAL (Second demonstration)

every half hour. Similar to the set of filters tested in Phase II, an initial increase in the pressure drop from ~ 3.0 to between 4.0 to 4.5 inches W.C., during the first 100 hours of operation, represents the conditioning of the filters (i.e., the approach to steady state, baseline fly ash loading on the membrane). Not all the fly ash is removed from the filters during backpulsing.

Following a second period of modifications to the ceramic filter chamber (Point A), the pressure drop continued to increase, approaching 7.5 to 8.0 inches W.C. This additional increase may be attributed to further conditioning of the filters. However, it is still below the design pressure drop of 8 to 10 inches W.C. predicted for a flow of 8,100 acfm @ 400°F. The system was designed based on a natural gas firing rate of ~ 19 MM Btu/h while the actual flue gas flow firing dry micronized coal (DMC) is approximately 5,600 acfm at 365°F. The DMC firing rate is maintained at approximately 15 MM Btu/h to reduce ash deposition in the convective pass. This lower flue gas flow rate would yield a lower pressure drop across the filters.

It should be noted that pressure drops that are significantly higher than the trend are due to operating the ceramic filters backpulse system. The backpulse system, used to remove ash from the filters' inlet surfaces, is operated at a selected frequency during boiler operation. The

frequency was varied from 15 to 45 seconds between backpulses during evaluation of this set of filters. The frequency and the duration, or length of backpulse, were occasionally varied in an attempt to improve cleaning efficiency and to lower the filters' pressure drop. When the backpulse system is operated, the additional air that is introduced into the filter chamber causes the measured pressure drop to increase several inches. Depending on when the pressure drop reading is recorded, higher than usual values can be obtained as the pressure drop is returning to its normal operating value. Greater detail of the pressure drop as a function of time firing coal is illustrated in Figures 3B-1 through 3B-19. These figures, contained in Appendix 3B, are plots of the pressure drop as a function of time firing coal for over 50 hours of operation. The use of the boiler's convective pass steam sootblower is also indicated in these figures.

A greater amount of scatter was observed in Figure 3.3.3 beginning at the 165 run hour (Point B). The primary difference between this demonstration and that performed in phase II, was that a slipstream system was installed on the demonstration boiler to evaluate ceramic filters for an industrial sponsor concurrent with the large ceramic filter chamber. Approximately 10% of the flue gas flow was removed from the boiler system at the boiler outlet, prior to entering the ceramic filter chamber, and routed to the slipstream system. At this time, 100 hours of shakedown operation were performed with 90% of the flue gas directed through the baghouse. A second 100-hour shakedown period was then conducted, Period C, but at this time the ceramic filter chamber was used to filter 90% of the flue gas. To ensure that the filter area (i.e., air-to-cloth ratio) of the ceramic filter chamber was consistent with the first demonstration and the first 165 hours of operation of the second demonstration, 10% of the ceramic filters (i.e., 8 filters) were removed. Testing that was conducted from Run Hour 165 to 770 was performed in this manner. The scatter in the data appear to be strongly influenced by the simultaneous operation of the slipstream system.

Another variable encountered during the second demonstration was varying coal ash content. Typically, the coal ash is maintained at 4.5 to 6 wt.% since this is a natural gas/ fuel oil-designed boiler retrofitted to fire coal-based fuels. However, the coal received during this testing contained ash contents ranging from approximately 3.5 to 10 wt.%. The coal company had laid out the coal in their coal yard to dry it (upon receiving it from the mine) and inadvertently picked up dirt prior to stockpiling the coal for Penn State. Consequently, the ash content fluctuated greatly from day-to-day during the latter part of this demonstration. The significant increase in

the pressure drop observed following testing of the slipstream system is believed to be a result of the higher coal ash contents. Changes in the backpulse frequency and duration were not able to maintain the pressure drop at the levels observed prior to the increase in the coal ash percent.

The loss of the filter's ceramic membrane coating on the inlet side, discussed in the following section, may also be a contributor to the filter's increased pressure drop. The loss of the membrane coating would allow the ash particles to penetrate the substrate pores, possibly preventing passage of the flue gas, thus increasing the overall pressure drop of the filters.

The pressure drops observed with the second set of filters from Corning were similar to those observed in the Phase II of the program (i.e., first demonstration). In Phase II, the pressure drop varied from 6-9 inches W.C., with much of the data between 7-8 inches W.C. for the first 700 hours of operation (excluding the filter conditioning during the first ≈ 150 hours). The pressure drops then increased in two increments, 9-10 inches W.C. and 13-15 inches W.C., when problems arose with moisture infiltration. The pressure drops during the first 800 hours of the Phase III operation (Figure 3.3.3) varied between 6-9 inches W.C. and then increased to 10-12 inches W.C., which appears to be due to the loss of the membranes and capture of fine particles in the pore structure. This is a ceramic membrane design/manufacturing issue.

Filter Evaluation by Nondestructive X-Ray Computerized Tomography

Each new filter, following its manufacturing, was evaluated by Corning, Inc. using several measurement techniques. These measurements included the determination of pressure drop at relevant face velocities and testing for cracks using powdered graphite and alumina. A select sampling of the filters was removed during the first 770 hours of testing at time intervals chosen by Corning and returned to them for further examination. This examination included the physical cross-sectioning of filters, measurement of the depth of ash retained within the filter cells, and the examination of the cell walls by scanning electron microscope.

To support these diagnostic measurements and to gain a better understanding of factors affecting the filter's performance, an additional twelve filters were purchased from Corning. Five of these filters were scanned by a medical X-ray scanner in the Center for Quantitative Imaging at Penn State's Energy Institute. This nondestructive technique permits an internal examination of not only a filter's construction, but also how exposure to the flue gas affects a

filter's performance (i.e., map out the degree of channel filling and detect any erosion within the filter).

A Universal System's model HD250 computerized tomography scanner (CT scanner) was used to examine the filters. This is a fourth generation medical system that takes a single slice in one rotation. The filters were supported in a sample chuck mounted on the patient table for scanning. The CT scanner, with a filter centered within the scanning aperture, is shown in Figure 3.3.4.

Figure 3.3.4 HD250 CT SCANNER WITH A FILTER SUPPORTED WITHIN THE SCANNING APERTURE

CT scanning passes X-rays through the filter. The X-rays are absorbed or attenuated (weakened) at differing levels by the filter creating a matrix or profile of X-ray beams of different strength. A set of detectors, mounted on the opposite side of the aperture from the X-ray tube or source, measures this profile. A computer then reconstructs the "raw CT data" into an image.

Each filter manufactured by Corning is canned in a shell made of 316 stainless steel. There was initial concern that this shell would interfere with the X-ray beam. However, the metal resulted in a sharpening of the beam producing a relatively clear image.

The HD250 CT scanner has a maximum energy level of 130kV at 125 mA. An energy level of 130 kV and 65 mA was used to scan the ceramic filters.

A baseline scanning was performed on each of the five new filters. This included a coarse scanning sequence performed over the filter's length using a 2 mm scan thickness and a 20 mm spacing between scans. This was followed with a more detailed scanning of the filter's inlet and outlet using a 2 mm scan thickness and spacing.

A digital radiograph was produced for each filter as part of the scanning sequences. The radiograph for filter #8122 is shown in Figure 3.3.5. The radiographs produced from the nondestructive technique of CT scanning provide details of the filters internal structure. The extruded cell walls, inlet and outlet plugs, filter matting, and exterior metal shell are visible in the radiograph.

Figure 3.3.5 A DIGITAL RADIOGRAPH OF FILTER #8122

A series of CT slices (CT range: -1000 to 2000) produced for both the inlet and outlet of filter #8122 are shown in Figures 3.3.6 and 3.3.7, respectively. An examination of these series shows that the inlet and outlet plugs are approximately 12-14 mm in length. Cells that were found to leak during quality control testing were sealed on the inlet and outlet ends. These patches are visible as breaks in the alternating open and closed pattern within the cells.

Figure 3.3.6 CT SCANS (2mm scan thickness and spacing) FOR THE INLET OF FILTER #8122 – CT Range: -1000 to 2000 (Scans progress from left to right starting at top left corner)

Figure 3.3.7 CT SCANS (2mm scan thickness and spacing) FOR THE OUTLET OF FILTER #8122 – CT Range: -1000 to 2000 (Scans progress from left to right starting at top left corner)

Following the baseline scanning, the five filters were installed on March 10, 2000 in the ceramic filter chamber and exposed to the ash laden flue gas produced from coal-fired operation of the boiler. The existing filters within the chamber had been exposed to approximately 80 hours of coal-fired operation prior to the installation of the five “CT” filters. The five filters were then removed following selected time intervals of operation. Corning had also chosen time intervals during which additional filters would be exchanged and those removed sent back to Corning for follow up diagnostic testing. To minimize the number of times the ceramic filter chamber would be opened, time intervals were chosen to correspond approximately to time intervals selected by Corning for the removal of filters. A summary of the filter exchange schedule including those exchanged or removed for Corning is given in Table 3.3.1.

Table 3.3.1 Filter Exchange Schedule

Date	Total Hours of Operation	Filters Exchanged	
		For Corning	For PSU (Filter Number)
9/28/99	0	Initial set of 80 filters installed	
10/26/99	10	2	0
11/04/99	70	2	0
3/10/00	80	0	5 PSU filters installed
3/16/00	92	0	1 – (#8125)
4/20/00	165	8 filters blanked off for slip-stream testing	
6/13/00	165	5	1 – (#8122)
7/28/00	270	5	1 – (#8112)
9/11/00	510	0	1 – (#8114)
9/25/00	705	2	0
10/30/00	770	3	1 – (#8115)

The filters used in the chamber and evaluated by the CT scanner consist of a cordierite porous substrate with two membrane coatings on the inlet (dirty side) and one membrane coating on the outlet cell (clean side). Filter #8112 was sectioned, plugs were made from filter pieces from outlet end, middle, and inlet end of the filter, scanning electron microscope (SEM) photomicrographs were taken to compare to the CT scanner results. The one membrane coating on the clean side and the two membrane coatings on the dirty side, along with an ash layer on the membrane coatings in the inlet cell, can be seen in Figure 3.3.8.

(a)

(b)

Figure 3.3.8 SEM PHOTOGRAPH OF THE OUTLET END OF FILTER #8112 (a-Inlet Cell; b-Outlet Cell)

On April 20, 2000 eight filters were removed from the ceramic filter chamber and their openings blanked off. This was done in preparation for the beginning of testing using a slipstream system installed on the demonstration boiler system, which is part of a separately funded project and not reported herein. During operation of the slipstream system, approximately 10% of the total flue gas flow was diverted from the large ceramic filter chamber to a slipstream chamber that houses ceramic filters. To maintain an air/cloth ratio that was consistent with the previous testing on the large chamber, 10% of the filters (i.e., eight filters) were removed from the large chamber. This was done in order to accumulate relevant data on the large chamber while operating the slipstream system.

Following exposure to the flue gas, each removed filter was scanned in the same manner and orientation in the scanner as the baseline condition (i.e., virgin filter prior to installation in the chamber). All five of the Penn State filters were removed and rescanned. The comparison results are discussed here

A detailed comparison of the filter inlet CT scans before and after use indicated a loss of material within the inlet cells. This effect can be seen in the Figure 3.3.9 where slightly larger, darker areas represent larger openings within the inlet cells. Comparison of the filter outlet CT scans, however, showed an increase in the material detected within the inlet cells. This increase can be attributed to the retention of ash within the cells at the filter's outlet.

Figure 3.3.9 COMPARISON INLET CT SCANS OF FILTER #8122 BEFORE (Left Scan) AND AFTER USE (Right scan) (Note that scans were conducted at same location along filter length)

The loss of material at the filter's inlet is due to a loss of the membrane coatings on the dirty side of the inlet cells, which appears to be caused by erosion from the ash particles during backpulsing. The above observations are also evident in Figure 3.3.10 where an average CT number is plotted as a function of location in filter #8122 for the coarse CT scans. A greater CT number indicates an increase in filter material, while a drop in CT number indicates material loss. The ceramic plugs closing off inlet or outlet cells cause the large increases at either end of the plot.

Figure 3.3.10 AVERAGE CT NUMBER AS A FUNCTION OF LOCATION IN THE FILTER (Zero represents filter inlet, scan spacing is 2 cm)

The above observations are also evident in Figures 3.3.11 and 3.3.12, which are SEM photomicrographs of the middle and inlet end of filter #8112, respectively. Figure 3.3.11 shows that the membrane coating in a dirty side (inlet) cell is still retained in the corners but has been eroded from the wall. Figure 3.3.12, on the other hand, shows that all of the membrane coating in the dirty side (inlet) cells has been removed. The sequence of photomicrographs shows that the membrane coating in the dirty side cells is progressively removed as one transits from the outlet end of the filter to the inlet end. As Figure 3.2.8 shows, there are no signs of erosion on the outlet end of the filter. One possible cause for the erosion is the high backpulse pressure utilized. This was done to maximize backpulse effectiveness.

Figure 3.3.11 SEM PHOTOMICROGRAPH OF A DIRTY CELL (Inlet Cell) IN THE MIDDLE OF FILTER #8112

Figure 3.3.12 SEM PHOTOMICROGRAPH OF A DIRTY CELL (Inlet Cell) IN THE INLET END OF FILTER #8112

The loss of the filter's membrane coatings on the inlet cells allows the ash particles to penetrate into the substrate pores, resulting in an increased pressure drop across the filter. The average inlet CT number, as shown in Table 3.3.2, decreased dramatically for filter #8125 and then slowly for filters run for longer durations suggesting that the erosion occurred early on. It is not possible to quantify the amount of membrane lost throughout the filter's length because some ash is retained, especially at the outlet end. An increasing CT number with greater hours of operation suggests that additional ash is being retained at the outlet end of the filters. The phenomenon of ash filling the pores is verified in Figure 3.3.12 and also in Table 3.3.3.

Table 3.3.2 Change in the Average Inlet/Outlet CT Number as a Function of Hours of Operation

Filter Number	Hours of Operation	<u>Change in Average CT Number</u>	
		Inlet	Outlet
8125	12	-30	+19
8122	85	-36	+38
8112	190	-36	+51
8114	430	NM	NM
8115	690	NM	NM

Table 3.3.3 Bench-Top Filter Pressure Testing

Filter Number	Hours of Operation	<u>Initial ΔP (in. W.C.)</u>			<u>Final ΔP (in. W.C.)</u>		
		3 ft/min	4 ft/min	5 ft/min	3 ft/min	4 ft/min	5 ft/min
8125	12	2.1	2.9	3.6	5.1	7.0	8.9
8122	85	2.0	2.8	3.5	9.0	12.8	15.7
8112	190	2.1	3.0	3.8	7.1	9.9	12.4
8114	430	2.2	3.3	4.3	11.9	NA ^a	NA
8115	690	2.2	2.9	3.6	9.4	13.5	NA

^aNA-Not able to achieve set point for that flow.

In summary, through repeated scans, the CT scanning technique for examining filters has proven to be consistent and reproducible. It provides a nondestructive method to examine a filter's structure with a relatively high degree of accuracy. This technique also provides added information on factors affecting a ceramic filter's performance.

Stack Particulate Emissions and Ceramic Membrane Filter Collection Efficiency

Similar to the measurements performed in Phase II on the initial set of ceramic membrane filters, EPA Modified Method 201A (modified to include PM_{2.5}) particulate emissions tests were conducted to determine the dust loadings and collection efficiency for the second set of filters when firing micronized coal. The combined results for both sets of filters are shown in Table 3.3.4. The Modified Method 201A collection efficiency measured for the second set of filters ranged from 99.80 to 99.85%, indicating their performance in removing particulate matter from the flue gas was approximately that of the initial set of filters.

Also evident in Table 3.3.4 is that the particulate emissions from the ceramic membrane filters have met the Pennsylvania Department of Environmental Protection's permitted level of 0.1 lb/MM Btu, the New Source Performance Standards (NSPS) of 0.03 lb/MM Btu, and the LEBS target of 0.01 lb/MM Btu. The emissions, however, exceed the HIPPS target of 0.003 lb/MM Btu. An objective of the project was to meet the HIPPS target, however, the current design of the ceramic membrane filter evaluated has not been unable to accomplish this during the final phase of testing

Comparison of Particulate Emissions between the Ceramic Filter Chamber and Baghouse. The stack particulate emissions from the fabric filter baghouse were not measured in Phase III, however, a comparison with those measured on January 20, 1998 (Phase II) indicate a significantly lower collection efficiency was observed for the fabric filter baghouse while firing the Middle Kittanning Seam DMC. The combustion efficiency for the four tests Method 201A tests performed on the ceramic filter chamber (0502/2001, 06/12/2001, 05/15/2003, and 06/17/2003) ranged from 96.5 to 97.0%, with an average of 96.7%. Although the combustion efficiency for the baghouse test was slightly lower, 96.4%, the inlet particulate emissions were not higher than those measured for the ceramic filter chamber. Thus, an increased level of particulate within the inlet did not cause the lower collection efficiency for the fabric filter. The baghouse collection efficiency of 99.17% is the lowest efficiency obtained in all of the micronized coal testing.

Fine Particulate and Condensable Particulate Matter Emissions. The Modified EPA Method 201A provides information on the nature of the fine particulate emissions from the

Table 3.3.4. Particulate Emissions as Measured by EPA Methods 5 and Modified 201A (fly ash) when Firing Middle Kittanning Seam DMC. The Ceramic Filter Chamber was the Particulate Collection Device used Except where Noted

DATE	Inlet			Outlet			Collection Eff. (%)	Combustion Eff. (%)
	gr/dscf	lb/MM Btu	g/Nm ³	gr/dscf	lb/MM Btu	g/Nm ³		
Method 5 Sampling								
07/09/97	2.2791	4.0845	5.2148					94.8
07/21/97				0.0056	0.0101	0.0128	99.77 ^a	92.4
07/24/97	2.4609	4.2990	5.6308					93.2
09/03/97	1.6707	3.3096	3.8227					97.6
09/08/97	1.6654	3.3343	3.8106	0.0028	0.0060	0.0064	99.83	97.7
09/12/97	1.2621	2.7982	2.8878	0.0025	0.0053	0.0057	99.80	98.0
09/22/97				0.0024	0.0053	0.0055	99.86 ^b	96.0
09/23/97	1.7068	3.3650	3.9053					97.6
11/05/97	2.2223	4.3580	5.0848	0.0045	0.0088	0.0103	99.80	Not Measured
11/18/97	1.9750	3.9139	4.5190	0.0050	0.0105	0.0114	99.75	96.1
Method 201A Sampling								
11/25/97				0.0037	0.0071	0.0085		96.4
12/04/97				0.0037	0.0072	0.0085		96.5
12/09/97	1.6364	3.3442	3.7442	0.0038	0.0073	0.0087	99.77	97.1
12/11/97	1.2436	2.4624	2.8455					97.4
12/22/97	1.7584	3.4878	4.0234					97.0
01/15/98	3.8487	7.4084	8.8062	0.0048	0.0095	0.0110	99.88	92.2
02/17/98	1.6194	3.1838	3.7053	0.0053	0.0105	0.0121	99.67	97.0
02/19/98	2.3802	4.7504	5.4461	0.0044	0.0089	0.0101	99.82	97.1
05/02/01	2.3777	5.2713	5.4404	0.0042	0.0089	0.0096	99.80	97.0
06/12/01	2.2141	4.3277	5.0661	0.0045	0.0096	0.0103	99.80	96.6
05/15/03	2.8531	5.8943	6.5282	0.0041	0.0108	0.0096	99.86	96.6
06/17/03	2.0665	4.4428	4.7281	0.0039	0.0090	0.0096	99.81	96.5
01/20/98 (BH) ^c	1.5420	3.3491	3.5283	0.0126	0.0263	0.0288	99.17	96.4

↑

Phase II

↓

Phase III

^a Collection efficiency was determined using inlet results from 07/09/97.

^b Collection efficiency was determined using inlet results from 09/23/97.

^c This test was performed on the fabric filter baghouse when firing Brookville seam MCWM to obtain an operating permit from the Pennsylvania Department of Environmental Protection. The emissions cannot exceed 0.1 lb/MM Btu

↑
Phase II

Phase III
↓

ceramic membrane filters. The particulate is separated into three size fractions during sampling. These fractions are reported in Table 3.3.5 as $> 10\ \mu\text{m}$ (initial cyclone), 2.5 to $10\ \mu\text{m}$ (second cyclone), $<2.5\ \mu\text{m}$ (filter paper). An EPA Method 202 (condensable particulate matter (CPM)) stack test was performed with each Modified Method 201A and the results are also given in Table 3.3.5. Similar levels of particulate emissions were measured for the ceramic membrane filter inlet. Similar to the Phase II testing, the largest portion of the fine particulate emissions was found in the $>10\ \mu\text{m}$ and 2.5 to $10\ \mu\text{m}$ fractions with the $>10\ \mu\text{m}$ fraction being slightly greater than the 2.5 to $10\ \mu\text{m}$ fraction. The smallest fraction was contained in the CPM with the $<2.5\ \mu\text{m}$ fraction being slightly larger.

The outlet ash concentrations from the ceramic filter chamber measured prior to 2003 exhibited similar trends in that the $>10\ \mu\text{m}$ and 2.5 to $10\ \mu\text{m}$ fractions were nearly identical and were approximately twice that of the $<2.5\ \mu\text{m}$ fraction. However, a significant decrease in the 2.5 to $10\ \mu\text{m}$ fraction was measured during tests performed in 2003. The reason for this change in the particle size distribution at the outlet of the ceramic filter chamber is not known. However, the overall collection efficiency remained unchanged and was similar to that measured for the initial set of ceramic membrane filters evaluated in Phase II. There is continued variability within the CPM concentrations with this fraction being the largest measured at the outlet of the ceramic filter chamber. The Modified Method 201A and 202 results are provided in Figures 3.3.13 and 3.3.14 in increasing order of combustion efficiency. No trends in the results as a function of combustion efficiency were observed.

Fine Particulate Penetration of the Particulate Devices. The percent of each fraction that penetrated the ceramic membrane filters when burning DMC is shown in Figure 3.3.15. Similar to the results observed for the first demonstration, the CPM fraction had the greatest penetration. Also, the 2.5 to $10\ \mu\text{m}$ fraction is the second largest fraction that is emitted from the collection device.

Concluding Remarks

In Phase III modifications were performed to the ceramic membrane filter chamber, a new set of ceramic membrane filters was installed, and their performance evaluated and

Table 3.3.5 Fine Particulate Emissions for the Ceramic Membrane Filter Chamber, as Determined by EPA Methods 201A (modified) and 202, when Firing Middle Kittanning Seam DMC

Date:	11/25/97	12/4/97	12/9/97	12/11/97	12/22/97	1/15/98	2/17/98	2/19/98	01/20/98 (BH) ^a	05/01/01	06/12/01	5/15/03	6/17/03
Inlet Part.													
Concen.													
(gr/dscf)													
> 10 µm			0.6784	0.6888	0.8863	2.3592	0.7132	0.9979	0.6282	1.1463	0.9445	1.7160	1.2005
2.5-10 µm			0.6001	0.4459	0.7058	0.6668	0.7430	0.8676	0.4907	0.7930	0.9322	0.9629	0.7395
< 2.5 µm			0.3579	0.1089	0.1663	0.8227	0.1632	0.5148	0.4051	0.4384	0.3374	0.1742	0.1265
total ash			1.6364	1.2436	1.7584	3.8487	1.6194	2.3803	1.5240	2.3777	2.2141	2.8531	2.0665
CPM ^b			0.0418	0.1635	0.0019	0.0600	0.0092	0.0050	0.0382	0.0348	0.0074 ^d	0.0201	0.1396
total fine part.			1.6782	1.4071	1.7603	3.9087	1.6286	2.3853	1.5622	2.4125	2.2215	2.8732	2.2061
Outlet Part.													
Concen.													
(gr/dscf)													
> 10 µm	0.0017	0.0016	0.0015			0.0019	0.0021	0.0018	0.0054	0.0015	0.0019	0.0020	0.0018
2.5-10 µm	0.0014	0.0014	0.0016			0.0017	0.0024	0.0018	0.0043	0.0016	0.0018	0.0009	0.0003
< 2.5 µm	0.0006	0.0007	0.0007			0.0012	0.0008	0.0008	0.0029	0.0011	0.0008	0.0012	0.0018
total ash	0.0037	0.0037	0.0038			0.0048	0.0053	0.0044	0.0126	0.0042	0.0045	0.0041	0.0039
CPM	0.0009	0.0019	0.0051			0.0044	0.0022	0.0011	0.0066	0.0076	0.0070	0.0022	0.0184
total fine part.	0.0046	0.0056	0.0089			0.0092	0.0075	0.0055	0.0192	0.0118	0.0115	0.0063	0.0223
Collection													
Eff.^c (%)			99.77			99.88	99.67	99.82	99.17	99.80	99.80	99.86	99.81
Combustion													
Eff. (%)	96.4	96.5	97.1	97.4	97.0	92.2	97.0	97.1	96.4	97.0	96.6	96.6	96.5
SO₂ (ppm)													
502 451 517 508 494 359 479 483 497										417 442 465 476			
NO_x (ppm)													
373 318 308 425 425 315 452 435 418										611 623 581 527			
O₂ (%)													
3.2 3.2 3.2 3.0 3.0 3.5 3.5 3.5 3.6										3.8 3.4 3.3 3.5			

^a Baghouse

^b Condensable particulate matter

^c Collection efficiencies do not include CPM

^d This value is suspect

← Phase II Phase III →

Figure 3.3.13 INLET FINE PARTICULATE EMISSIONS FROM THE CERAMIC MEMBRANE FILTER CHAMBER WHEN FIRING MIDDLE KITTANNING SEAM DMC COAL

Figure 3.3.14 OUTLET FINE PARTICULATE EMISSIONS FROM THE CERAMIC MEMBRANE FILTER CHAMBER WHEN FIRING MIDDLE KITTANNING SEAM DMC COAL

Figure 3.3.15 PERCENT OF EACH FINE PARTICULATE FRACTION THAT PENETRATES COLLECTION DEVICE

compared with DOE emissions targets and the performance of a fabric filter baghouse outfitted with high-performance bags. Specifically:

- The ceramic membrane filter system was operated for ~930 hours;
- The filters were regenerable through backpulsing with compressed air; however, the pressure drop continued to increase. The increase is believed to be due to higher ash percents in the feed coal and plugging of the substrate pores following loss of the membrane;
- The pressure drop data for Phase III were similar to those observed during Phase II, both of which were nearly 1,000 hours in length. The elevated pressure drop data for both demonstrations indicate that this is a ceramic membrane design/manufacturing issue and not operator induced. This is a concern for commercialization in a boiler setting and needs to be addressed by the manufacturer;

- The ceramic membrane filter particulate collection efficiencies ranged from 99.80 to 99.86% when firing Middle Kittanning seam DMC, which were higher than that observed when using the pulse-jet baghouse; and
- The stack particulate emissions are below Pennsylvania DEP's permit conditions (0.1 lb/MM Btu), the NSPS (0.03 lb/MM Btu), and DOE's LEBS target (0.01 lb/MM Btu), and approach DOE's HIPPS target (0.003 lb/MM Btu).

3.3.1.2 Effect of Particle Control Device on Trace Elements Emissions

The objective of this activity was to observe the effect of particulate control device (PCD), specifically baghouse and ceramic membrane filter, on trace element emissions. Trace element emission testing was performed in conjunction with DMC combustion tests conducted at the demonstration boiler, which are described in Section 5.4.1. These tests were performed as part of the study of the particulate capture characteristics and capacity of the ceramic membrane filters and fabric filters during combustion of a micronized Middle Kittanning seam coal.

As reported in the previous section (3.3.1.1), the fine particulate concentration to the baghouse was similar to that entering the ceramic filter chamber in previous tests. The emissions from the baghouse, however, were much greater than from the ceramic membrane filter. The collection efficiency of the baghouse was 99.2% compared to 99.67 to 99.88% for the ceramic filter chamber. The particulate emissions from the baghouse test were triple those from the ceramic filter chamber for similar combustion efficiencies. The combustion efficiency during the baghouse test was 96.4%, which is identical to the ceramic filter chamber tests.

The outlet ash concentrations from the ceramic filter chamber exhibited similar trends in that the >10 and 2.5 to 10 μm fractions were nearly identical and were approximately twice that of the <2.5 μm fraction. The condensable particulate material (CPM) concentrations were more variable and, in some cases, were the largest fractions. At the outlet of the baghouse the CPM fraction is by far the largest fraction that is emitted. With one exception, the 2.5 to 10 μm fraction is the next largest fraction that is emitted from the collection devices. Each size fraction emitted from the baghouse was greater than that emitted from the ceramic filter, with the exception of the CPM. In general, the ceramic membrane filter had greater capture efficiency and removed a greater percentage of fines than the fabric filter. The capture surface area of the two PCDs was approximately equal (2,000 sq. ft). A detailed discussion of the particulate

emissions and performance of the two PCDs is given in Section 3.3.1.1 and by Miller et al. (1998, 1999).

Duplicate inlet and outlet tests were performed on each particulate collection device, i.e., ceramic membrane filter chamber and baghouse. EPA Method 29 (Code of Federal Regulations 1996) was used to determine trace element concentrations. Method 29 was used since the evaluation of the unified sampling train/methodology (PSU Method), which is described in Section 3.3.3 was not yet completed. Extensive testing has shown that EPA Method 29 may not speciate mercury correctly; however, it may still be used to measure total mercury (EPRI 1996, 1997). Mercury is completely volatilized during coal combustion and forms elemental mercury (Hg^0) and may form mercury chloride (HgCl_2), mercuric oxide (HgO) and other mercury species as the flue gas cools. Due to their high vapor pressure, these species will remain in the gas phase unless they interact with a solid that traps the mercury through sorption. Currently numerous carbon- and noncarbon-based solid sorbents are being investigated as to their ability to react with gas phase mercury species. The potential of carbon-based sorbents for mercury capture is based on early observations that mercury species were removed from the gas phase via sorption onto the carbonaceous portion of char present in the gas stream and PCD. The effectiveness of different chars to remove mercury from the gas stream is not consistent even under similar combustion conditions and gas compositions. In general, the physical characteristics (e.g., surface area, porosity and permeability, particle size) of the char (carbonaceous and inorganic component) as well as the amount of residual carbon determine its effectiveness. Also important is the contact time between the sorbent, as a suspended particle in the gas stream or as a captured particle in the PCD.

The Energy and Environmental Research Center at The University of North Dakota investigated sorbent injection for mercury control in coal-fired boilers (Dunham et al. 1998). They investigated the effect of PCDs, i.e., pulse-jet baghouse, ESP and a compact hybrid particulate collector (COHPAC), on mercury emissions as determined by Method 29. They observed that for two subbituminous coals there was some inherent mercury capture even by fly ash having an LOI less than 1%. Some of this capture is due to the cooling of the stack gas to allow for some condensation of certain trace elements. They also note that total mercury and removal across a baghouse decreases as baghouse temperature increases. The percent removal of total mercury ranged from 81% at 250°F to approximately 10% at 400°F for an Absaloka coal

Interestingly a Blacksville coal showed only an 18% to 10% removal of mercury in the same temperature range. Dunham et al. (1998) showed that a lignite-based activated carbon removed over 80% of the total mercury at temperatures at or below 250°F. At 300°F, vapor-phase removal of mercury was reduced to 70, 50, and 0% for the Comanche, Absaloka, and Blacksville coals, respectively. Their results show that there is considerable variation in the effectiveness of a lignite-based activated carbon due to temperature and coal type. In addition, mercury speciation is affected by temperature.

It should be stated that temperature effects were not an issue in this study as the temperature in the ceramic and bag filters was approximately 400°F. Based on the work by Dunham et al. (1998), a PCD temperature of 400°F would represent a worst-case scenario with regards to the percent mercury removal across the PCDs. In addition, the capture surface areas of the two PCDs were approximately equal at 2,000 ft². Given that the particle size distribution of the captured fly ash and efficiency of removal of the two PCDs differ suggests the possibility that inherent capture of mercury and other trace elements might also differ.

Test Results

Trace Elements Excluding Mercury

The percent penetration for the more volatile elements through the two PCDs during the duplicate test is shown in Figure 3.3.16. The concentration of trace elements in the coal and emissions measured are given in Table 3.3.6. In general, the baghouse removed approximately 94.7 to 99.1% of all the trace elements excluding mercury. The ceramic membrane filter removed 97.3 to 99.7% of the trace elements excluding mercury. The ceramic filter accounted for an additional 45 to 61% reduction in emissions of the trace elements. This corresponds to removal levels observed by Dunham et al. (1998), in which greater efficiency of removal leads to a corresponding greater removal of trace elements associated with particulate capture. Variation in the percent penetration of selenium measured during the duplicate tests through the ceramic filter was greater than that measured through the baghouse but is not considered significant. There was good agreement between the duplicate tests based on the demonstrated reproducibility of Method 29.

Figure 3.3.16 PERCENT FILTER PENETRATION OF SELECTED TRACE ELEMENTS FOR THE MIDDLE KITTANNING COAL (FABRIC FILTER VS. CERAMIC FILTER)

Table 3.3.6 Middle Kittanning Coal Analysis and Trace Element Emissions as a Function of PCD

Element	Coal mg/g	Outlet Emissions (10 ¹² Btu/lb)			
		Fabric Filter 1	Fabric Filter 2	Ceramic Filter 1	Ceramic Filter 2
As	2.43	1.51	2.12	0.94	1.08
Be	4.02	1.36	1.52	0.54	0.45
Cu	15.08	5.65	6.51	3.72	2.98
Hg	0.033	0.56	0.56	0.07	0.17
Pb	8.03	2.34	2.94	1.52	1.68
Se	1.33	0.70	0.99	0.74	0.82
Sr	27.16	10.83	9.89	3.63	4.09
Sb	4.59	1.39	1.91	0.64	0.82

Mercury Emissions

Duplicate inlet and outlet tests were performed on each particulate collection device, i.e., ceramic membrane filter chamber and baghouse, and the results for total mercury are shown in Figure 3.3.17. The inlet concentration of mercury, for the duplicate baghouse tests, was 1.90 and 2.06 lb/10¹² (trillion) Btu while the inlet concentration of mercury, for the duplicate ceramic filter tests, was 1.54 and 2.14 lb/10¹² Btu. The concentration of mercury in the outlet of the filters was 0.56 and 0.56 lb/10¹² Btu for the baghouse and 0.07 and 0.16 lb/10¹² Btu for the ceramic filter. A total of 2.38 lbs/10¹² Btu of mercury was present in the coal. The fabric filters retained 70.5 to 72.8% of the mercury that entered the baghouse. The ceramic membrane filters retained 92.3 to 95.3% of the mercury that entered the PCD. Overall, mercury emissions from the ceramic filter were 79% lower than those measured at the baghouse outlet.

Figure 3.3.17 MERCURY EMISSIONS FROM THE DEMONSTRATION BOILER WHEN FIRING MIDDLE KITTANNING SEAM MICRONIZED COAL

Checks were done on the reproducibility of the sampling and analyses. In addition, closure on the material balances across the filters and system was calculated. Assuming that the inlet mercury concentrations are identical for the four tests, one standard deviation of the mean

was determined and is shown in Figure 3.3.17 along with the mean inlet mercury concentration ($1.91 \text{ lb}/10^{12} \text{ Btu}$). This was done as a check on reproducibility and shows that it was good as three of the four inlets are within one standard deviation while one of the ceramic filter inlets is only slightly lower than one standard deviation. The material balances on mercury through the system and across the filters were also very good. Ash samples were collected every half hour from the ceramic filter chamber or baghouse during the testing. A composite was made for each test and analyzed for mercury (as well as the other inorganic trace elements). Filter balances, defined as

$$(\text{Hg}_i - \text{Hg}_o) / \text{Hg}_a$$

where Hg_i = flue gas inlet mercury

Hg_o = flue gas outlet mercury

Hg_a = mercury in the ash collected by the filtering devices

had closures of 1.05 and 1.10 for the baghouse and 0.95 and 0.98 for the ceramic filter. A value of 1.00 would be perfect closure. Similarly, the system balance, defined as

$$((\text{Hg}_{\text{fuel}} - \text{Hg}_i) + \text{Hg}_a + \text{Hg}_o) / \text{Hg}_{\text{fuel}}$$

where Hg_{fuel} = mercury in the fuel

had closures of 0.92 and 0.87 for the baghouse and 1.08 and 1.03 for the ceramic filter. This is quite good for total mercury using Method 29.

The percent of the mercury that penetrated the particulate collection devices is shown in Figure 3.3.18. Approximately 30% of the mercury that entered the baghouse was emitted to the atmosphere, whereas less than 10% of the mercury entering the ceramic filter was emitted. Since the ceramic filter captures more of the fine particulate, it is likely that the mercury is condensed on the fine particles.

Figure 3.3.18 PERCENT MERCURY PENETRATION THROUGH THE FABRIC AND CERAMIC FILTERS WHEN FIRING MIDDLE KITTANNING SEAM MICRONIZED COAL

Conclusions

The ceramic membrane filters were not only more efficient in removing particulate matter but also in reducing trace elements in the stack gas. Total mercury emissions were reduced by 95%, as compared to 77.8% through the system via the ceramic filter and fabric filter, respectively. Refractory or less volatile trace elements are generally associated with particulate matter; therefore, their concentration would decrease in the stack gas as particulate removal increases. However, more volatile elements such as mercury and arsenic and selenium, to a lesser degree, are generally present in the gas phase. Their removal relies on the interaction of the gas phase species with other phases in the char. Given that measurements were made at the PCD inlet and outlet, the data represent reductions that occurred as a result of interactions between the gas and solid phases in the PCDs.

Both devices operated at approximately 400°F and had the approximately the same capture surface area (2,000 ft²). The average burnout for the duplicate tests for the fabric filter

and ceramic filter were essentially the same at 94.6%. This translates to the char having an ash content of $\approx 40\%$ or a carbon content of $\approx 60\%$ (which is a significant amount of carbon). Combustion efficiency is determined using the ash tracer technique discussed in Section 3.2.9.1. The only variable that differed between the tests that has been shown to impact the interaction of unburned carbon, acting as an activated carbon, with gas phase mercury is the reactive surface area of the char. The extent of sorption is a function of the amount of reactive surface area available. The fact that the ceramic membrane filter captures a greater percentage of fine particulate, as well as the fine particles being imbedded in the filters' pores, could result in a greater amount of reactive surface area than in the fabric filter. In addition, there has been speculation that gas phase species may react with the inorganic portion of the char (Dunham et al. 1998). Speciation measurements were not conducted at this time (due to lack of consensus of methodology in the literature at the time of testing).

The data are compelling and the duplicate runs show good agreement. The ceramic membrane filter effectively removed 79% more mercury from the gas phase than the fabric filter even at elevated temperature. The performance of the ceramic membrane filter is a promising environmental technology for both particulate and gas phase mercury emission control provided the high pressure drop issue can be resolved.

3.3.2 Effect of Coal Cleaning on Trace Element Emissions

The objective of this activity was to investigate the effect of coal cleaning on trace elements emissions using several coals provided by Cyprus-AMAX Research & Development Center and CQ Inc. The cleaned coals were produced via projects funded under the U.S. Department of Energy's Premium Fuel Program. For this work, filter cakes produced by AMAX from cleaned coal from the Taggart seam (Virginia) and Hiawatha seam (Utah) and Pittsburgh coal cleaned by CQ Inc. were used. The AMAX coals were cleaned using advanced column flotation and oil agglomeration to produce a filter cake. Elemental analysis of the parent coal and filter cake and the corresponding stack gas and fly ash was used to determine the percent penetration of various trace elements through the system. Of interest is whether extensive cleaning of the coals results in a benefit in reducing trace element emissions. The AMAX fuels were fired as MCWMs.

3.3.2.1 Background

AMAX Fuels

The AMAX cleaned fuels were produced as part of a Department of Energy funded study titled “Engineering Development of Advanced Physical Fine Coal Cleaning for Premium Fuel Application” (DOE Contract No. DE-AC22-92PC92208) conducted by AMAX Research and Development Center from October 1992 to September 1997 (Jha et al. 1997). The objective of the project was the engineering development of two advanced physical fine coal cleaning processes, column flotation and selective agglomeration, for premium fuel applications. The project included laboratory research and bench-scale testing on six coals and the design, construction and operation of a 2 t/h process development unit (PDU). The secondary objective of the project was to determine the removal of toxic trace elements during fine coal cleaning. Depending upon the coal, AMAX reported that advanced coal cleaning reduced the concentrations of toxic trace elements such as mercury and could be a part of a strategy to control hazardous air pollutants. A detail discussion of the cleaning technologies, flotation and agglomeration studies, and the physical characteristics of the filter cake produced are presented in the project final report by Jha et al. (1997).

According to Jha et al. (1997), the residual amounts of trace elements in the clean coals after cleaning were dependent upon the source coal. Generally, when prepared from the same coal, the advanced flotation and selective agglomeration clean coal contained about the same amounts of trace elements. The residual mercury and selenium analyses were interesting. Mercury was not detected (<0.01 ppm) in some clean coal samples and ranged up to 0.03 ppm in other samples.

There were substantial reductions (25-76%) in the concentration of ash, arsenic and manganese (lb/MM Btu) from the amount in the as-received test coals. There was little or no reduction ($<25\%$ or negative) in the amounts of antimony, beryllium, cobalt, nickel, and selenium in the as-received coal on the same basis. Little or no reduction in the concentration of an impurity means that the impurity is associated with the carbonaceous components of the coal. A substantial reduction signifies an association with the mineral matter in the coal. The reduction of other impurities, such as total sulfur, pyritic sulfur, cadmium, chromium, mercury, lead and chlorine varied from coal to coal.

According to Jha et al. (1997), the fine coal cleaning was effective for reducing the lb/MM Btu concentration of mercury in the Taggart coal by 39%. Since the Hiawatha coal contained less mercury to begin with, coal cleaning had less impact upon the final concentration of the mercury in the cleaned product.

The authors concluded that the two cleaning processes appeared to be equally effective for reducing the concentrations of impurities in the coal and were equally effective for cleaning coal to premium fuel specifications. However, for certain coals, cleaning resulted in substantially reducing the concentrations of some of the hazardous air pollutant trace elements, especially arsenic, chromium, cobalt, manganese, mercury, selenium and chlorine. Analysis of the filter cake at Penn State, reported on an lb/MM Btu basis, shows that the Taggart Oil Agglomeration fuel had approximately four times as much mercury as the Taggart Column Flotation fuel.

The AMAX project did not have a combustion test component to verify if the cleaned coals would produce reduced levels of trace element emissions as a function of the cleaning process. Significant quantities of cleaned coal were produced during the project, which could be used for pilot- or demonstration-scale testing. Penn State acquired several of the filter cakes produced during the AMAX project for this purpose. The filter cakes were used to formulate MCWMs, which were tested in Penn State's research and demonstration boilers.

CQ Inc. Fuels

CQ Inc. provided Penn State with coal from the Pittsburgh, Middle Kittanning, and Freeport (h_vA bituminous coal) seams (Pennsylvania), which had been cleaned using a combination of heavy media cyclones, water spirals, and froth flotation. The Freeport and Kittanning coals were studied and reported on in the Phase II Final Report (Miller et al., 2000).

The Freeport seam parent and cleaned coals were fired as pulverized coal in the 500,000 Btu/h down-fired combustor described in Section 3.1.1. In general, coal cleaning was seen to have a beneficial effect on reducing the concentration of trace element emissions during combustion of the Freeport seam.

The Middle Kittanning seam coal was used to study the effect of fuel form on trace element emissions. Approximately 250 gallons of highly loaded MCWM were prepared from the Middle Kittanning seam parent coal. The Middle Kittanning DMC and MCWM were fired

in Penn State's research boiler. The boiler is a 1,000 lb saturated (@200 psig) steam/h watertube package boiler of A-frame construction, which was designed and built by Cleaver Brooks.

Coal cleaning was found to have a beneficial effect on reducing the concentration of trace element emissions during combustion of the Freeport coal. All of the trace elements in the flue gas from the clean coal were present at levels between 38 to 81% less than in the raw coal flue gas. The level of depletion of As, B, Cr, Hg, Mn, Pb, V, and Zn concentrations in the flue gas from the clean coal versus the raw coal was similar to that seen between the original fuels. The mercury concentration in the flue gas from the clean coal was 38% less than in that from the raw coal. This corresponds to the 32% depletion of mercury in the clean coal due to the cleaning process. Elements that were enriched in the clean coal, i.e., Ba, Be, Cd, and Se, were present in lower concentrations in the flue gas from the clean coal than in the flue gas from the raw coal.

Fuel form, i.e. DMC versus MCWM, was found to have little effect on emission levels during combustion of the Middle Kittanning seam coal. In general, the partitioning behavior of the elements during combustion of the Middle Kittanning seam coal was consistent with earlier tests conducted at Penn State and reported in the literature. All of the trace elements were depleted in the stack gas by 71.4 to 99.9% during DMC and MCWM combustion. There was no significant difference in the percent of penetration by most of the trace elements during combustion of the DMC and MCWM, however, the percent penetration by mercury and boron did vary between the two fuels. The partitioning of elements for both fuels was more of a function of their occurrence, i.e., mineral versus organic, in the fuel than the form in which they were fired. The formation of large agglomerates may have resulted in enhanced volatility of certain elements contributing to increased concentrations in submicron ash particles or the gas phase. Further details can be found in Miller et al. (1998, 2000).

In Phase III, the Pittsburgh seam coal was studied to determine the effects of coal cleaning on emissions using the PSU Method sample train during several replicate tests. The results are discussed in Section 3.3.3.

3.3.2.2 Analysis of AMAX Fuels

Four fuels were tested in this work. The Taggart seam (Virginia) and Hiawatha seam (Utah) filter cakes were produced using AMAX's PDU that was developed as part of the original DOE funded project. The PDU was an integrated 2 ton process that contained advanced flotation

and selective agglomeration modules for cleaning the test coals. Trace element analyses of the feed coal and cleaned products for the Taggart and Hiawatha seam coals are given in Tables 3.3.7 and 3.3.8. Unfortunately the feed coal used to produce the cleaned coals was unavailable to Penn State for complete elemental analysis or testing. AMAX Research and Development Center provided the feed coal trace element analysis.

Table 3.3.7 Fuel Analysis (ppm)

Element	Taggart			Hiawatha		
	^a Feed Coal	Oil Agglomeration	Column Flotation	^a Feed Coal	Oil Agglomeration	Column Flotation
As	3.69	1.64	1.54	0.71	0.76	1.00
Ba	na	16.61	20.37	na	10.62	18.77
Be	2.0	0.86	1.31	0.3	< 0.07	1.25
Cd	< 0.1	< 0.06	0.04	< 0.1	< 0.07	< 0.03
Co	8.7	12.3	5.26	0.8	5.31	6.47
Cr	6	8.24	8.54	4.8	7.04	4.69
Cu	na	6.89	8.67	na	3.19	8.34
Hg	0.02	0.027	< 0.01	0.01	0.011	0.014
Mn	7	5.17	7.23	8	3.85	6.05
Mo	na	5.78	1.97	na	4.25	3.02
Ni	11	9.23	8.02	1	4.38	8.34
Pb	4	1.32	2.35	< 2	1.14	2.15
Sb	0.8	0.68	0.60	0.09	0.50	0.67
Se	1.52	0.81	1.12	1.12	1.00	0.54
Sr	na	18.45	23.00	na	33.86	22.42
V	na	14.76	7.23	na	9.30	9.91
Zn	na	4.31	4.20	na	2.66	2.92

^a Data provided by AMAX (Jha et al. 1997)
na: data not available

The MCWM formulation is provided in detail in Section 5.2.3. The proximate and ultimate analyses and physical properties of the filter cakes and MCWMs are given in Table 3.3.9.

Figures 3.3.19 and 3.3.20 present the data in Tables 3.3.7 and 3.3.8 graphically. The Taggart Oil Agglomeration fuel had at least a four times greater level of mercury than measured in the Column Flotation fuel (3.03 versus <0.69 lbs/10¹²Btu). The Column Flotation fuel was lower in Co, Hg, Mo, Ni and V. The Oil Agglomeration had lower levels of Ba, Pb, Be and Sr. Both fuels had similar levels of As, Sb and Zn.

The Hiawatha fuels had similar levels of Hg, Zn, V and Sb. In general, the remaining elements were slightly elevated in the Column Flotation fuel compared to the Oil Agglomeration

Table 3.3.8 Fuel Analysis for Cleaned Coals (lb/MM Btu)

Element	Taggart		Hiawatha	
	Oil Agglomeration	Column Flotation	Oil Agglomeration	Column Flotation
As	184.15	176.00	97.93	129.14
Ba	1865.07	2328.00	1368.46	2424.01
Be	96.57	149.71	< 9.02	161.43
Cd	< 6.74	4.57	< 9.02	< 3.87
Co	1381.11	601.14	684.23	835.55
Cr	925.23	976.00	907.15	605.68
Cu	773.65	990.86	411.05	1077.05
Hg	3.03	< 0.69	1.42	1.81
Mn	580.52	826.29	496.10	781.31
Mo	649.01	225.14	547.64	390.01
Ni	1036.40	916.57	564.39	1077.05
Pb	148.22	268.57	146.90	277.66
Sb	76.35	68.57	64.43	86.53
Se	90.95	128.00	128.86	69.74
Sr	2071.67	2628.57	4363.10	2895.38
V	1657.34	826.29	1198.37	1279.81
Zn	483.95	480.00	342.76	377.10

fuel. The only exception was that the Oil Agglomeration fuel had a significantly higher concentration of Sr.

3.3.2.3 AMAX Combustion Tests and Results

Each of the cleaned coals from Cyprus-AMAX was tested in Penn State's demonstration boiler (15,000 lb saturated (@300 psig) steam/h watertube package boiler), which has been described elsewhere (Section 5.4 for MCWM firing). Each filter cake was formulated into a MCWM. The formulation procedure is described in detail in Section 5.2.3. General physical characteristics of the MCWMs are given in Table 3.3.9. All MCWMs were cofired with 30% (thermal input) natural gas support (Table 3.3.10). Pressures indicated under test variable refer to atomizing air pressure. Stack gas sampling was conducted according to Method 29 (described in detail in Section 3.3.3).

Two sequential tests were run during a test period in which Method 29 trains were run simultaneously at the particle control device inlet and outlet. The sampling protocol for EPA Method 29 was followed as outlined in the Code of Federal Register (1996). The emission levels of trace elements for the Taggart and Hiawatha seam fuels are shown in Figures 3.3.21 and 3.3.22.

Table 3.3.9 Filter Cake And MCWM Analysis

	Hiawatha Selective Agglomeration	Hiawatha Column Flotation	Taggart Selective Agglomeration	Taggart Column Flotation
Filter cake Properties				
Proximate Analysis (wt. %, dry basis)				
Volatile Matter	43.8	41.7	33.3	33.3
Ash	2.2	2.5	1.5	1.7
Fixed Carbon	54.0	55.8	65.2	65.0
Ultimate Analysis (wt. %, dry basis)				
Carbon	80.0	83.2	87.3	88.2
Hydrogen	5.8	6.1	6.4	5.7
Nitrogen	1.5	1.5	1.4	1.5
Sulfur	0.5	0.7	0.8	0.7
Oxygen	10.0	6.0	2.6	2.2
Higher Heating Value (Btu/lb, dry)	14,130	14,255	15,012	14,975
HGI	44 ^a	44 ^a	49 ^a	49 ^a
MCWM Properties				
Dispersant Conc.(wt.%)	0.35	0.35	0.30	0.30
Stabilizer Conc. (wt.%)	None	None	None	None
pH	7.2	7.1	5.7	6.2
Solids Loading (wt.%)	54.3	55.2	59.5	57.7
Apparent Viscosity (cp, @ 100 sec ⁻¹)	372	384	1,088	291
Particle Size Distribution (μm)				
Top Size	207	189	119	250
D(v,0.9)	70.7	73.2	57.7	101.0
D(v,0.5)	21.5	28.4	15.0	34.0
D(v,0.1)	5.5	6.1	4.6	6.1

^a From Jha *et al.*, 1997

Previous tests conducted in Phase II showed reductions in emissions similar to the reduction of trace elements seen in coal as a result of coal cleaning (Falcone Miller et al., 1998a; Miller et al., 2000). In Figures 3.3.23 and 3.3.24, the percent depletion/enrichment of various trace elements in the column flotation MCWM as compared to the oil agglomeration MCWM is plotted against their relative depletion/enrichment in the stack gas (e.g., $100 \times [As_{\text{Oil Agglomeration}} - As_{\text{Column Flotation}}] / [As_{\text{Oil Agglomeration}}]$). The percent difference for each element in the fuel and

Figure 3.3.19 FUEL ANALYSIS OF CLEANED TAGGART FUELS

Figure 3.3.20 FUEL ANALYSIS OF CLEANED HIAWATHA FUELS

Table 3.3.10 Combustion Tests Completed on the Demonstration Boiler

Fuel	Cleaning Technique	Particulate Control Device	Test Variable
Taggart MCWM	Agglomeration	Baghouse	135 psig
			80 psig
Taggart MCWM	Flotation	Baghouse	135 psig
Hiawatha MCWM	Agglomeration	Baghouse	135 psig
Hiawatha MCWM	Agglomeration	Ceramic	
Hiawatha MCWM	Flotation	Baghouse	30 % Gas Support
Hiawatha MCWM	Flotation	Baghouse	26 % Gas Support

Figure 3.3.21 EMISSIONS MEASURED BY METHOD 29 AT PCD OUTLET DURING COMBUSTION OF TAGGART CLEAN COALS

emissions was calculated in this manner for each fuel. In theory, if there is a direct relationship between the depletion/enrichment of an element in a fuel and its depletion/enrichment in the flue gas then one would expect all of the elements to plot in the I and III quadrants. Ideally, the points should lie upon the solid line showing a 1:1 relationship between the extent of depletion/enrichment of an element in a fuel to its depletion/enrichment in the flue gas.

Those elements in Quadrant I are enhanced in the Oil Agglomeration Fuel and corresponding emissions. Those elements in Quadrants III are enhanced in the Column Flotation

Figure 3.3.22 EMISSIONS MEASURED BY METHOD 29 AT PCD OUTLET DURING COMBUSTION OF HIAWATHA CLEAN COALS.

Figure 3.3.23 RELATIONSHIP OF ELEMENT CONCENTRATION IN FUEL AS A FUNCTION OF CLEANING AND CORRESPONDING ELEMENT EMISSIONS FOR THE TAGGART OIL AGGLOMERATION AND COLUMN FLOTATION FUELS

$$\% \text{ Difference} = (\text{Element}_{\text{Taggart Oil Agglomeration}} - \text{Element}_{\text{Taggart Column Flotation}}) / \text{Element}_{\text{Taggart Oil Agglomeration}}$$

Figure 3.3.24 RELATIONSHIP OF ELEMENT CONCENTRATION IN FUEL AS A FUNCTION OF CLEANING AND CORRESPONDING ELEMENT EMISSIONS FOR THE HIAWATHA OIL AGGLOMERATION AND COLUMN FLOTATION FUELS

$$(\% \text{ Difference} = (\text{Element}_{\text{Hiawatha Oil Agglomeration}} - \text{Element}_{\text{Hiawatha Column Flotation}}) / \text{Element}_{\text{Hiawatha Oil Agglomeration}})$$

fuel and corresponding emissions. Elements that lie in these quadrants suggest a direct relationship with enhancement and reduction in the fuel to emissions. Elements in Quadrants II and IV do not demonstrate a direct relationship between enhancement or reduction of an element in the fuel to corresponding emissions.

In the case of the Taggart Fuels (Figure 3.3.23), Co, V, and Mn lie near the solid line in Quadrant I and Mn in Quadrant III suggesting that there is a direct relationship with enhancement and reduction in the fuel to emissions. However, many of the elements show inconsistent relationships. For example, Pb is significantly enhanced in the Column Flotation Fuel but present at roughly the same concentration in the emissions measured for the two fuels. Mercury is significantly higher in the Oil Agglomeration Fuel (77%) but the concentration of mercury measured in the Oil Agglomeration Fuel was less than in the Column Flotation Fuel (-33%).

In the case of the Hiawatha fuels (Figure 3.3.24), only Mo, Cr and Ba showed a proportional relationship with enhancement and reduction in the fuel to emissions. Most of the

elements are located in Quadrant IV. Elements are enhanced in the Column Flotation Fuel but are depleted in the corresponding emissions (or conversely enhanced in the Oil Agglomeration emissions. The variable or inconsistent nature of the data suggests that the amount of an element in a fuel is not the only variable that determines its concentration in the flue gas.

3.3.2.4 Combustion Tests and Results of the CQ, Inc. Coals

Pittsburgh seam raw and cleaned coals were fired in a 2 million Btu/hr A-frame watertube boiler. A schematic diagram of the 1,000 lb/h steam boiler is shown in Figure 3.3.25. The boiler is a 200 psig working pressure, A-Frame water-tube boiler, designed and built by Cleaver Brooks. The combustion chamber is a 3 ft x 3 ft x 7 ft chamber with a maximum heat release rate of 42,000 Btu/ft³-h. It contains 288 ft² of heating surface and the maximum fuel firing rate is two million Btu/h. It has been modified to fire any combination of natural gas, liquid fuels, or solid fuels.

Figure 3.3.25 SCHEMATIC DIAGRAM OF THE RESEARCH BOILER

The boiler is equipped with eighteen side ports for gaseous and particulate sampling. Fourteen of the ports have diameters of 3 inches and four have diameters of 4 inches. The combustion gases split into two convective passes, one on each side of the radiant combustion chamber. There are access doors into each of the convective sections. There are also two ash hoppers under each convective section and there is a doorway into the radiant combustion chamber.

The steam pressure inside the boiler is maintained constant during tests by a back pressure regulator. The steam flow rate is measured by a steam flow meter before passing through a condenser. Condensed water flows through a deaerator before returning to the boiler.

To promote and enhance combustion, a ceramic burner throat extends the combustion chamber by two feet. This ceramic section, termed a quarl, is preheated by a natural gas flame prior to pulverized coal combustion. This section aids in the support of hard to ignite fuels by storing some of the radiant energy released by the flame. When firing DMC, it is screw fed from an 80 lbs capacity hopper to an eductor and transported by compressed air into the burner. When firing MCWM, it is gravity fed from a day tank into a progressive cavity pump.

Duplicate simultaneous sampling of the flue gas was conducted at the inlet and outlet of the baghouse using the PSU Method Train to sample combustion flue gas mercury species and inorganic trace elements. A description of the PSU Method can be found in Section 3.3.3. Analysis of the parent and cleaned coal is given in Table 3.3.11.

The process of cleaning the coal resulted in a reduction of all the trace elements on a lbs/10¹² Btu basis except for Sr which did not change appreciably. The percent reduction ranged from 15 (for Sb) to 82% for As. Mercury decreased by $\approx 37\%$.

The emission of trace elements measured at the baghouse outlet is given in Figure 3.3.26. The cleaned coal showed a reduction in all of the elements measured except for mercury. Interestingly mercury emission was $\approx 72\%$ higher in the cleaned coal stack gas.

A plot of the depletion/enrichment of an element in the fuel to its depletion/enrichment in the flue gas is given in Figure 3.3.27. In theory, all of the data points should fall in Quadrants I or III as close to the solid line (indicating a 1:1 relationship between reduction in the fuel and emissions). This assumes that there is a direct relationship between the depletion/enrichment of an element in the fuel to its depletion/enrichment in the flue gas. Most of the elements plot in Quadrant I. Therefore most of the elements are present in the cleaned fuel at lower levels than in

Table 3.3.11 Trace Element Analysis Of Parent And Cleaned Pittsburgh Coals

Element	Pittsburgh Parent Coal		Pittsburgh Cleaned Coal	
	ppm	Lbs/10 ¹² Btu	ppm	Lbs/10 ¹² Btu
As	9.88	813.0	2.09	145.7
Ba	59.70	4912.5	42.29	2948.2
Be	1.49	122.6	1	69.7
Cd	0.10	8.2	0.07	4.9
Co	15.17	1248.3	11.69	815.0
Cr	26.12	2149.3	14.68	1023.4
Cu	12.94	1064.8	6.47	451.1
Hg	0.120	9.9	0.089	6.2
Mn	16.17	1330.6	10.95	763.4
Mo	4.98	409.8	4.23	294.9
Ni	16.67	1371.7	8.96	624.6
Pb	7.39	608.1	1.62	112.9
Sb	0.75	61.7	0.75	52.3
Se	2.44	200.8	1.64	114.3
Sr	48.51	3991.7	58.46	4075.5
V	32.34	2661.1	19.9	1387.3
Zn	28.61	2354.2	15.92	1109.9

the parent raw coal and show a corresponding reduction in the flue gas. Several elements lie along or within $\pm 20\%$ of the solid line within Quadrant I suggesting that they follow approximately a direct relationship between reduction in the fuel and emissions, after accounting for sampling and analytical errors.

Elements, which significantly deviate from this relationship, are antimony and mercury. Mercury is singular in its behavior in that it is the only element that showed an increase in emission in the flue gas of the cleaned coal. One would expect to see a reduction in emissions if there was a corresponding reduction of trace elements in the fuel due to cleaning. In the case of mercury, there was a 38% reduction of mercury in the cleaned fuel yet there was 72%, i.e., 0.4 versus 0.69 lbs/10¹² Btu, more mercury emitted at the baghouse outlet. The material balance for mercury across the baghouse was 1.02 and 0.93 for the Pittsburgh raw and cleaned coal, respectively. This suggests that the measured levels of mercury in the flue gas are reliable.

One possible explanation may lie in the difference in the amount and/or nature of carbon present in the baghouse ash. Significant amounts of carbon in ash has resulted in increased retention of mercury in the particulate collected at the baghouse inlet during tests performed on the research and demonstration boilers at Penn State. The percent carbon measured in baghouse ash plotted against the percent of mercury measured in the particulate for a series of tests

Figure 3.3.26 TRACE ELEMENTS EMISSIONS OF PITTSBURGH RAW AND CLEANED COAL

conducted using one fuel in the research and demonstration boilers is shown in Figure 3.3.28. In general, an increase in percent carbon results in an increase in the percent mercury in the particulate phase. The correlation (r) between percent carbon and percent particulate mercury for the research boiler tests was 0.628 and 0.996 for the demonstration boiler tests.

The percent carbon in the Pittsburgh raw coal and cleaned coal baghouse ash was quite similar, i.e., 13.9 and 16.4%, respectively. The percent burnout of the raw and cleaned Pittsburgh coal was 94.5 and 98.6%, respectively. The percent of particulate mercury measured

Figure 3.3.27 THE RELATIONSHIP BETWEEN REDUCTION IN TRACE ELEMENTS DUE TO COAL CLEANING AND THEIR REDUCTION IN STACK EMISSIONS FOR THE PITTSBURGH COAL
(Solid line represents theoretical 1:1, Dash lines represent $\pm 20\%$ of theoretical).

at the baghouse inlet was also similar for the raw and cleaned coal, i.e., 98.4 and 90%, respectively. However, the percent of particulate mercury measured at the baghouse outlet was quite different for the raw and cleaned coal, i.e., 85.6 and 36.8%, respectively. The data suggest that there was a significantly greater percentage of gas phase mercury exiting the baghouse during combustion of the cleaned coal than during combustion of the raw coal. During combustion of the raw coal 93.4% of the mercury that entered the baghouse was retained and presumably associated with ash and char particles. During combustion of the cleaned coal only 74.6% of the mercury that entered the baghouse was retained. The data support the work of

Figure 3.3.28 THE RELATIONSHIP BETWEEN CARBON IN ASH AND THE RETENTION OF MERCURY ON PARTICULATES

others showing the importance of the interaction of mercury in the gas phase with ash and char in particle control devices (Livengood et al., 1994; Senior et al., 1997). The percent carbon contained in the baghouse ash was essentially the same for the two fuels. It is possible that a difference in the inorganic chemistry of the ashes contributed to the difference in the retention of mercury in the baghouse.

3.3.2.5 Effect of Atomization Quality on Trace Element Emissions

The Taggart seam MCWM (oil agglomeration) was fired at two different atomizing pressures, i.e., 135 and 80 psig to determine what effect this had on trace element emissions. Numerous studies have suggested that the presence of carbon in char acts as a medium that absorbs mercury on active sites (Senior et al., 1997; Senior et al., 1998). Reducing the atomizing air pressure resulted in a decrease in the carbon burnout of the fuel from 97.8 to 94.8 %. In theory, the char character of the two fuels should be similar and only the amount of carbon present should differ by virtue of the reduced carbon burnout. This could enhance the retention of mercury in the baghouse.

The percent penetration of selected elements for the two tests is given in Figure 3.3.29. In general, there was a greater penetration of the highly volatile elements, i.e., Hg, in the test

Figure 3.3.29 PERCENT SYSTEM PENETRATION OF SELECTED IHAPS FOR THE TAGGERT MCWM (Oil Agglomeration) AS A FUNCTION OF ATOMIZING AIR PRESSURE

conducted at 135 psig having the greater burnout. There was a 93% reduction in mercury penetration through the system at 80 psig and 94.8 % burnout as compared to the baseline test at 135 psig. The data suggest that the extent of burnout has less effect on the retention of elements that are less volatile than mercury.

3.3.2.6 Conclusions

The objective of this activity was to investigate the effect of coal cleaning on trace elements emissions. A basic assumption was made that there was a direct relationship between the depletion/enrichment of an element in the fuel to its depletion/enrichment in the flue gas. Earlier results reported in the Phase II final report using the Freeport seam coal cleaned by CQ Inc. suggested that there was a relationship between the extents to which trace elements were

removed during cleaning and their reduction in stack gas (Miller et al., 2000). Further studies were conducted to validate this assumption.

In the case of the AMAX fuels there was significant scatter in the data. In some cases, a percent reduction of elements due to cleaning made up a greater percentage of the trace elements analyzed for in the stack gas. The data indicate that there is not always a direct relationship between the reductions in trace elements in a fuel and a corresponding reduction in emissions. The atomization tests with the Taggart Oil Agglomeration Fuel were conducted to see if a decrease in burnout due to poorer atomization quality would affect trace element emissions. In general, there was a greater penetration of the highly volatile elements, i.e., Hg, in the test conducted at 135 psig having the greater burnout. There was a 93% reduction in mercury penetration through the system at 80 psig and 94.8 % burnout as compared to the baseline test at 135 psig. The data suggest that the extent of burnout had less effect on the retention of elements that are less volatile than mercury.

By contrast, the data obtained during combustion of the Pittsburgh seam coal and cleaned coal (provided by CQ Inc.) suggested that there was a direct relationship between the depletion/enrichment of an element in the fuel to its depletion/enrichment in the flue gas. Most of the elements plotted in Quadrant I, therefore, most of the elements were present in the cleaned fuel at lower levels than in the parent raw coal and showed a corresponding reduction in the flue gas. One exception was mercury. Mercury concentration decreased by $\approx 37\%$ in the cleaned coal, however, mercury concentration was $\approx 72\%$ higher in the cleaned coal stack gas. Comparing the inlet and outlet data showed that there was a significantly greater percentage of gas phase mercury exiting the baghouse during combustion of the cleaned coal than during combustion of the raw coal. During combustion of the raw coal 93.4% of the mercury that entered the baghouse was retained and presumably associated with ash and char particles. During combustion of the cleaned coal only 74.6% of the mercury that entered the baghouse was retained. The data support the work of others showing the importance of the interaction of mercury in the gas phase with ash and char in particle control devices (Livengood et al., 1994; Senior et al., 1997). No specifics identifying the mechanism that caused this difference are given at this time.

The variation in the results obtained suggest that trace element emissions can not be predicted solely based on their concentration in the fuel. The interaction of the inorganics within

the system during combustion often affects the partitioning and concentration of trace elements in the gas and ash stream. Therefore, the effectiveness of extensive coal cleaning in the reduction of trace element emissions is dependent on coal composition and the system configuration.

3.3.3 Methodology Development

3.3.3.1 Introduction

In January 1999, the U.S. Environmental Protection Agency (EPA) initiated an Information Collection Request (ICR) for mercury emissions from electric utility steam generating units, which are 25 MW_e or greater in capacity (EPA ICR No. 1858.01). The ICR was conducted to assist the EPA in determining whether it is appropriate and necessary to regulate mercury emissions from electric utility steam generating units under Title III, Section 112 (n)(1)(A) of the Clean Air Act Amendments of 1990. The issuance of an ICR generally precedes the implementation of regulatory action by the EPA.

In December of 2000, the U.S. EPA decided that regulation of mercury from coal-fired power plants was appropriate and necessary under Section 112 of the Clean Air Act due to the significant health hazards it presented to the public health. The EPA is scheduled to propose regulation of mercury emissions by December 2003 and promulgate them by December 2004. In order to come to this determination, the accurate measurement of mercury and its speciation was necessary. These data could also be used to confirm measurements by online analyzers being developed for use in monitoring emissions and determining the effectiveness of new control technologies. It was stated by State and Local Stakeholders involved in the working group for the utility MACT formed under the Clean Air Act Advisory Committee that if multiple metals were to be regulated by the EPA, Method 29 should be the preferred method for measuring mercury until CEMs are commercially available.

Extensive testing has shown that EPA Method 29 may not speciate mercury correctly; however, it may still be used to measure total mercury (Electric Power Research Institute 1996, 1997). Early work by The University of North Dakota Energy and Environmental Center involved a self-validation test (EPA Method 301) to evaluate the precision and bias of Method 29 for the determination of mercury (Dunham et al. 1998; Heidt and Laudal, 1995). The overall pooled standard deviation for the inlet, outlet, and baghouse ash taken from nine pairs of

simultaneous tests (three tests on three coals) were 0.85, 0.97 and 0.60 mg/m³, respectively. The self-validation test included twelve pairs of simultaneous Method 29 samples, and resulted in a standard deviation of 0.58 mg/m³, which compares well with the 0.85 and 0.97 mg/m³. They concluded that about 60% of the variability is due to the sampling/analysis and 40% of the variability is due to the process.

Studies have shown that the speciation of mercury is extremely important in its environmental impact. Oxidized mercury (Hg⁺²) is highly reactive with surface water and is methylated and is incorporated into the food chain via ingestion by fish. The amount of oxidized mercury emitted from a particular power plant for a given coal can be variable.

The EPA had designated the Ontario Hydro Mercury Speciation Method for stack gas sampling to measure total, elemental and oxidized mercury. This recommendation is based on extensive work by the University of North Dakota Energy and Environmental Research Center (UNDEERC) in evaluating a series of methodologies for determining mercury speciation (Laudal et al. 1997). EPA had also designated the Method 29 sampling train for the sampling and measurement of IHAPs.

The Method 29 procedure is outlined in the Code of Federal Regulations, Protection of the Environment, 40, Part 60 (Code of Federal Regulations 1996). The Ontario Hydro Mercury Speciation Method used in this study is based on the draft procedure titled "Standard Test Method for Elemental, Oxidized, Particulate, and Total Mercury in Coal-Fired Stationary Sources (Ontario Hydro Mercury Speciation Method)" recommended by UNDEERC to the American Society for Testing and Materials (ASTM) and to EPA (U.S. EPA 1999). This is a version of the Ontario Hydro Method originally developed by Keith Curtis and others at Ontario Hydro Technologies.

Not all IHAPS elements are of equal environmental concern given their varied behavior in combustion systems. Elements that tend to volatilize during combustion, e.g., Hg, Se, As, and remain totally or partially in the gas phase and/or are readily re-emitted from the solid phase are of particular environmental importance. Currently there are no emission restrictions on these elements from power plants, however studies are being conducted as to their health risk and environmental impact as small particulate in the gas effluent and fly ash deposited in ash pits.

The objective of this study was to develop a sample train and methodology capable of simultaneously sampling mercury species and all inorganic trace elements listed as IHAPs. The

train and methodology, referred to as the PSU Method, consists of components of both the Ontario Hydro Mercury Speciation Method and Method 29 sampling trains. Of interest is whether the difference in the PSU Method train configuration compared to the Method 29 and Ontario Hydro trains compromises the measurement of either or both of the total trace elements or mercury species. Also at issue is whether the sample preparation and analytical methods outlined in Method 29 and the Ontario Hydro Mercury Speciation Method are appropriate for a multielemental train. The advantage of the PSU Method is that it will enable IHAPS and mercury species to be sampled simultaneously. This reduces the time required to change out sampling trains and obtain and recover samples, thereby allowing for a complete suite of data to be collected during one continuous run. The following discussion is taken from several conference papers by Falcone Miller et al. (1999a, 1999b, 1999c).

The trace element testing was performed on three test units at Penn State's Energy Institute – the 400,000 Btu/h DFC, the 1,000 lb saturated (@200 psig) steam/h research boiler, and the 15,000 lb saturated (@300 psig) steam/h demonstration boiler). These units have been described earlier in Section 3.0.

3.3.3.2 Early Development of the PSU Method

It was determined that the PSU Method would adapt features of the Method 29 (used for multielements) and the Ontario Hydro Mercury Speciation Method into a unified method and address any analytical limitations. A schematic of the Method 29, Ontario Hydro and PSU Method Trains is given in Figure 3.3.30. The sampling and recovery procedure used for the PSU Method is a combination of those used in the Method 29 and Ontario Hydro Mercury Speciation Methods (U.S. EPA 1999; Code of Federal Regulations 1996). Modification of the Ontario Hydro train included omission of one KCl impinger and one $\text{H}_2\text{SO}_4/\text{KMnO}_4$ impinger and the addition of a $\text{HNO}_3/\text{H}_2\text{O}_2$ impinger (Figure 3.3.30). The inclusion of an extra impinger ($\text{HNO}_3/\text{H}_2\text{O}_2$) in the PSU Method train should have no effect on the amount of mercury trapped compared to the Ontario Hydro train. The configuration was based on discussions with UNDEERC and earlier testing conducted at PSU. Tests conducted at PSU using one and two $\text{HNO}_3/\text{H}_2\text{O}_2$ impingers showed the presence of IHAPs in the second impinger. Therefore it was determined that two $\text{HNO}_3/\text{H}_2\text{O}_2$ impingers should be used to ensure the recovery of multielements. The recovery protocol for the filter, filter rinse, $\text{HNO}_3/\text{H}_2\text{O}_2$ and $\text{H}_2\text{SO}_4/\text{KMnO}_4$

Figure 3.3.30 SAMPLE TRAIN CONFIGURATIONS FOR METHOD 29, ONTARIO HYDRO MERCURY SPECIATION METHOD AND PSU METHOD

samples are the same for Method 29 and the Ontario Hydro Mercury Speciation Method. The recovery and sample preservation of the KCl solution followed the Ontario Hydro Mercury Speciation Method. The sample preparation, *i.e.*, digestion techniques, and analytical techniques for the multielements and mercury for the current PSU Method are shown in Figure 3.3.31.

Early tests showed poor agreement between the data generated for samples following the digestion procedures in Method 29 and the Modified Ontario Hydro Method. Therefore, additional digestion techniques were evaluated including EPA Methods 7471, 3050 and 3051 (Mercury in Solid or Semisolid Waste – Manual Cold Vapor Technique; U.S. EPA 1988). Splits were made of each sample so that different sample preparations and analyses could be performed as appropriate. All of the analytical techniques used meet instrument detection levels outlined in

Figure 3.3.31 SAMPLING PREPARATION AND ANALYTICAL PROTOCOL FOR THE PSU METHOD (Falcone Miller, et al., 1999a)

both Method 29 and the Modified Ontario Hydro Method.

Solid samples consisted of the filter and rinse samples combined. One sample split was digested according to Method 7471 and the other by Method 29 Par Bomb digestion. The Method 7471 digested sample was analyzed for total mercury by CVAAS. GFAAS and DCP were used to analyze the sample digested by Method 29 for multielements.

The KCl solution (Impingers #1 and 2) was analyzed for multielements and mercury. Given that the KCl containing impinger precedes the H₂O₂ and the KMnO₄ solutions, it was a concern that multielements could be captured in the KCl solution. Therefore, analysis of only the H₂O₂ solution for multielements could result in under reporting the amount of gas phase multielements. There is no provision for analyzing multielements in the Modified Ontario Hydro Method in any of the impinger reagents. Therefore, the KCl solutions in Impingers #1 and 2 were combined and then split into two samples. One split was recovered according to the Modified Ontario Hydro Method using KMnO₄ as a stabilizer. The sample was then digested by Method 7471 and successfully analyzed for mercury (oxidized) by CVAAS. The second split was divided into two samples to be analyzed for multielements. One sample was digested by a modified Method 29 procedure using nitric acid and analyzed by DCP and GFAAS. Most of the multielements were present in concentrations below instrument detection limits. This was due to the fact that the high concentration of chlorine (contributed by the KCl reagent) in the samples

required excessive dilution to avoid interference effects in the DCP and GFAAS. The second sample was digested by Method 3050 and analyzed directly by an ICP-MS (quadrupole), DCP and GFAAS. DCP and GFAAS could not be used to analyze the second sample due to chlorine interference and dilution problems. However, analysis for multielements by ICP-MS was successful due to the increased resolution of the instrument and its ability to factor out interference due to chlorine present in the sample. One limitation of the quadrupole ICP-MS used is the high detection limit of mercury. The levels of mercury in the samples were at or near the detection limit of the ICP-MS (quadrupole) used.

The hydrogen peroxide ($\text{HNO}_3/\text{H}_2\text{O}_2$) solutions (Impingers #3 and 4) were combined and then split. One split was further divided into three samples: no digestion, digestion by the Modified Ontario Hydro Method, and digestion by Method 7470. Each sample was subsequently analyzed for elemental mercury by CVAAS. The samples digested by Method 7471 (indicated on Figure 3.3.31) gave the most reproducible results. The second split was digested according to Method 3050 and successfully analyzed for multielements by GFAAS and DCP. The potassium permanganate ($\text{H}_2\text{SO}_4/\text{KMnO}_4$) solutions (Impingers # 5 and 6) were combined and digested as outlined by Method 7470. The solution was then analyzed for mercury (elemental) by CVAAS.

3.3.3.3 Preliminary Test Results -- Multielements

A series of tests was conducted in the 20 million Btu/h demonstration boiler and the 2 million Btu/h research boiler using a deeply-cleaned Taggart seam hvA bituminous coal from Virginia. The coal was cleaned by AMAX using an advanced oil agglomeration process and then processed into a filter cake (Jha et al. 1997). A MCWM was prepared from the filter cake at Penn State. Details as to the preparation process, atomization character and combustion performance of this fuel are provided in Section 5.0. Samples were collected simultaneously at the baghouse inlet and outlet during each test.

Method 29 versus PSU Method

Emissions measured by Method 29 and the PSU Method at the inlet and outlet of the baghouse on the demonstration boiler and at the inlet of the baghouse on the research boiler are given in Figures 3.3.32 through 3.3.34. Emissions are reported in $\text{lb}/10^{12}$ Btu for each of the

Figure 3.3.32 EMISSIONS MEASURED FOR THE TAGGART FUEL AT THE BAGHOUSE INLET OF THE DEMONSTRATION FACILITY

Figure 3.3.33 EMISSIONS MEASURED FOR THE TAGGART FUEL AT THE BAGHOUSE OUTLET OF THE DEMONSTRATION FACILITY

Figure 3.3.34 EMISSIONS MEASURED FOR THE TAGGART FUEL AT THE BAGHOUSE INLET OF THE RESEARCH BOILER

inorganic HAPs previously identified. In general there is good agreement between the Method 29 and PSU Method data for each test based on the reproducibility of Method 29.

Prior experience with Method 29 has shown that there is variability in the data due to sampling, recovery and sample preparation error (Dunham et al., 1998; Heidt and Laudal, 1995). In order to address this issue, a test series was conducted on the Middle Kittanning seam coal (hV A bituminous) from Jefferson County, Pennsylvania to determine the reproducibility of the data. Emission data were obtained for four identical runs conducted in the demonstration boiler. Samples at the baghouse inlet were analyzed and compared. A mean, standard deviation, margin of error and percent margin of error (based on a 95% confidence interval) for each element were calculated (Table 3.3.12).

The percent margin of error is taken to represent the range in values possible during a replicate run attributed to sampling, recovery, and instrument errors combined. The percent margin of error was then applied to the mean values obtained for each element measured by the Method 29 and PSU Method sampling trains. The mean emission of each element measured by the two sample techniques is indicated by a dot on the histogram. The bars represent the range

Table 3.3.12 Reproducibility of Replicate Tests using Method 29 in the Demonstration Test Facility

Element	Mean	Standard Deviation	*Margin of Error	*% Margin of Error
As	177.7	35.5	34.8	± 19.5
Be	164.9	31.8	31.2	± 18.9
Cd	2.6	0.8	0.8	± 30.0
Co	407.6	114.0	111.7	± 27.4
Hg	1.9	0.2	0.2	± 10.3
Mn	217.0	32.5	31.9	± 14.6
Ni	990.5	172.2	168.8	± 17.0
Pb	257.7	45.7	44.7	± 17.3
Sb	192.6	44.2	43.3	± 22.5
Se	68.3	13.3	13.0	± 19.1

* 95% Confidence Interval and n=4

of emission values calculated using the percent margin of error derived for each element. If the range in emission level for a particular element, measured by the two techniques for the Taggart seam fuel, is less than the margin of error obtained for the four Middle Kittanning seam runs, then there is no significant difference in the data measured by the two sampling techniques. The emission levels for cadmium and mercury in Figures 3.3.32 through 3.3.34 are given in the graph legends.

For all the elements measured at the demonstration boiler inlet (Figure 3.3.32) there is good agreement between the two techniques. At the outlet, the difference in the emission level of each element is within experimental error. The only exception is the range in cadmium and selenium, which exceeds the range of experimental error seen in the Middle Kittanning seam runs. The results at the research boiler inlet also show good agreement between the emissions measured by Method 29 and the PSU Method for the multielements, except for mercury. The percent margin of error for mercury was $\pm 10.3\%$ for the four samples at a 95% confidence interval. This compares well to Dunham et al. (1998) who reported $\pm 13\%$ for mercury for the same sample size ($n = 4$) and confidence interval (95%).

Ontario Hydro Method versus PSU Method

The Ontario Hydro train was run in conjunction with the Method 29 and PSU Method trains during emission tests of the Taggart fuel in the research boiler. Duplicate simultaneous samples were taken at the baghouse inlet and outlet during each test run. The results of the

preliminary tests for selected elements are shown in Figures 3.3.35 and 3.3.36. In general the Ontario Hydro train had consistently lower values, possibly due to the presence of only one H₂O₂ impinger in the sample train. Trace elements were present in the second H₂O₂ impinger in the PSU train. There was fair reproducibility for several elements:

As: 1s within $\pm 19\%$ of X (mean)

Be: 1s within $\pm 9\%$ of X (mean)

Pb: 1s within $\pm 9\%$ of X (mean)

Sb: 1s within $\pm 12\%$ of X (mean)

Se: 1s within $\pm 10\%$ of X (mean)

Method 29 values were within ± 1 -10% of the mean for the total sample set for each element.

Figure 3.3.35 EMISSIONS DATA OF Co and Ni MEASURED BY MULTIPLE METHODS

Figure 3.3.36 EMISSIONS DATA OF INORGANIC HAPS MEASURED BY MULTIPLE METHODS

3.3.3.4 Preliminary Test Results – Total Mercury

Method 29 versus PSU Method

The total, particulate (Hg), elemental (Hg^0) and oxidized (Hg^{+2}) mercury levels measured in the demonstration boiler test at the baghouse inlet and outlet are shown in Figure 3.3.37. The impinger solutions that were analyzed for mercury and mercury species for the two trains are indicated on Figure 3.3.30. It is acknowledged that Method 29 is not appropriate for detailed mercury speciation (Laudal et al., 1997). Therefore, the Modified Ontario Hydro method was adopted as the technique best suited to measuring the forms of mercury in the PSU Method.

Theoretically, the inclusion of an extra impinger containing $\text{HNO}_3/\text{H}_2\text{O}_2$ in the PSU train, as compared to the Modified Ontario Hydro train, should have no effect on the amount of mercury trapped. In addition, the interaction between mercury and the solid sample, i.e., ash and char, collected on the filter should be the same for both sample trains.

The total mercury measured by the two techniques in the demonstration boiler was within calculated experimental error. Both methods measured the amount of particulate mercury in

Figure 3.3.37 MERCURY SPECIES MEASURED BY METHOD 29 AND PSU METHOD AT THE DEMONSTRATION BOILER BAGHOUSE INLET AND OUTLET

both tests to within $\pm 10\%$ of an average emissions value. There was little agreement between the two techniques in the amount of Hg^{+2} present in the gas stream. Studies conducted at UNDEERC suggest that Method 29 tends to overestimate the mass percent of oxidized mercury by 10-30% (Zygarlicke and Pavlish, 1997). Interestingly the data presented in Figure 3.3.37 show the amount of Hg^{+2} measured by the PSU Method to be three times greater than that measured by Method 29. The percent of particulate Hg, Hg^0 and Hg^{+2} measured by the PSU Method at the demonstration boiler baghouse inlet was 60, 10, and 30, respectively, and was 8, 25 and 67, respectively, at the baghouse outlet. The reduction in particulate Hg was expected. The change, i.e., reduction in solid phase Hg and increase in Hg^{+2} , in the partitioning of mercury between the inlet and outlet, indicated by the two methods, was similar. There was no change in the level of Hg^0 emissions. It should be stressed that the mercury concentrations in the samples analyzed for Hg^{+2} were at or below instrument detection limits.

PSU Method versus Ontario Hydro Method

Tests were conducted to compare the total mercury measured by the Ontario Hydro Method and PSU Method with the new train configuration. Preliminary results with a small

sample size at the baghouse inlet (n=5, PSU Method; n=4, Ontario Hydro Method) and outlet (n=4, PSU Method; n=3, Ontario Hydro Method) are shown graphically in Figures 3.3.38 and 3.3.39.

Figure 3.3.38 PRELIMINARY STATISTICAL REPRESENTATION OF TOTAL MERCURY DATA OBTAINED VIA THE PSU METHOD AND ONTARIO HYDRO METHOD AT THE BAGHOUSE INLET

Surprisingly, the means for the two sample sets were within one standard deviation of each other in spite of the small sample sizes. This implies that the two sample sets are from the same population. In theory, the statistical distribution of the two sample sets should approximate one another as the number of tests is increased. The data was assumed to be distributed normally.

3.3.3.5 Further Test Results

The preliminary test results suggested that the PSU Method performed as well as Method 29 for multielements and the Ontario Hydro Method for mercury speciation. Therefore, additional tests were conducted to expand the statistical database to confirm the preliminary test results. Duplicate simultaneous tests were performed at the inlet and outlet of a baghouse using

Figure 3.3.39 PRELIMINARY STATISTICAL REPRESENTATION OF TOTAL MERCURY DATA OBTAINED VIA THE PSU METHOD AND ONTARIO HYDRO METHOD AT THE BAGHOUSE OUTLET

the PSU Method, EPA Method 29 and Ontario Hydro Mercury Speciation Method sample trains. A total of fourteen tests were conducted on the 2 million Btu/h research boiler firing deeply-cleaned Taggart seam (Virginia), hvA bituminous coal as MCWM. Similar combustion conditions were maintained during all test runs. Coal analysis and average flue gas composition are given in Table 3.3.13.

Laudel et al. studied the effects of flue gas constituents on mercury speciation (Laudal, Brown and Nott, 1998). They evaluated the presence of SO_2 , HCl, Cl_2 , NO, NO_2 , and fly ash individually and in different combinations at different concentrations on the oxidation of elemental mercury. The study concluded that Cl_2 present at 10ppm has a significant effect on mercury speciation as measured by the Ontario Hydro Mercury Speciation Method. However, it is not known if Cl_2 is present at such levels in coal flue gas. There is significant interaction, i.e., oxidation of elemental mercury, with fly ash, NO_x (NO/ NO_2 at 600/30ppm), HCl (50ppm) and SO_2 (1,500ppm) when present in various combinations resulting in over estimation of oxidized mercury. HCl alone has little effect on mercury speciation at typical baghouse and stack temperatures. The concentration of SO_2 and HCl in the Taggart flue gas was significantly lower

Table 3.3.13 Average Coal, Fuel and Flue Gas Composition for Test Runs

<u>Coal Composition</u> <u>(Wt. %)</u>	<u>Coal</u> <u>(dry basis)</u>	<u>MCWM</u> <u>(as fired)</u>
C	85.1	51.9
H	5.6	3.4
O	5.5	3.3
N	1.5	0.9
S	0.7	0.4
Ash	1.7	1.0
H ₂ O	0	39.0
<u>Trace Elemental Analysis</u>		
<u>MCWM (as-fired)</u>	<u>µg/g</u>	<u>lbs/10¹²Btu</u>
As	1.32	144.2
Be	0.87	95.0
Cd	<0.04	<4.26
Co	5.22	570.2
Hg	0.031	3.4
Ni	6.97	761.3
Pb	2.77	302.6
Sb	0.9	98.3
Se	1.31	143.1
Sr	17.42	1,902
Cl*	85.4	9,100
<u>Flue Gas Composition</u>		
<u>(Vol %, dry basis)</u>		
CO ₂	13.6	
H ₂ O	(14.8)	
SO ₂	0.0296	
NO _x	0.0464	
O ₂	3.61	
HCl**	0.001069	

*Cl analysis provided by AMAX Research and Development Center

**Theoretical calculation assuming all Cl in the fuel is converted to HCl. This assumption is based on reports in which Cl₂ generally makes up only 5 to 7% of the total chloride (Cl₂ + HCl).

than the concentrations cited and therefore should not compromise the Ontario Hydro Mercury Speciation Method. The concentration of NO_x (464ppm) in the Taggart flue gas approached the concentration of 600ppm used by Laudal et al. (1998) suggesting that it may have some effect in oxidizing elemental mercury in the presence of fly ash.

The level of unburned carbon in the fly ash collected on the inlet filter was also high which may have compromised the speciation data. However, it should be noted that the PSU Method and Ontario Hydro Mercury Speciation Method train configurations, recovery and analysis are the same with regards to mercury speciation; therefore, any bias in speciation measurements would be reflected equally in both methods.

DeVito and Rosenhoover (1997) reported emissions measured by Method 29 from coal-fired power plants using coals having chlorine concentrations ranging from 300 to 1400 $\mu\text{g/g}$. The Taggart fuel contained only 85.4ppm chlorine. The study by DeVito and Rosenhoover showed good closure for coals containing 0.11 to 0.16% (db) chlorine as compared to the Taggart fuel that contained 0.014% (db) chlorine. This suggests that the data obtained by Method 29 from the Taggart fuel would not be compromised with regards to the content of chlorine in the fuel.

Statistical Analysis

The data were analyzed using a statistical software package via a two sample *t*-test with unpaired, i.e., unequal, variances (MINITAB Release 10Xtra 1995). A standard *t*-test assumes that the two samples come from populations with the same standard deviation/variance. If the standard deviations in the two groups are markedly different, the above assumption does not hold. An approximate test, i.e., the unpaired two sample *t*-test, allows for statistical comparison of unequal standard deviations/variances. The test determines whether both sample sets represent the same distinct population by determining the difference between the means of the two sets of observations and the sample variance. This test was also chosen due to the limited degrees of freedom of some of the elemental sample sets. The sample sets are assumed to be independent and quantitative and plausibly normal (Statistics at Square One 2000). A probability (*P*) is calculated for $\mu_1 - \mu_2$, where μ represents the mean for each population and tests $H_0: \mu_1 = \mu_2$ versus $H_1: \mu_1 \neq \mu_2$. H_0 represents the null hypothesis in which the two data sets are not distinct and represent the same statistical population. Rejecting the null hypothesis, i.e.,

accepting H_1 , means that the two data sets are distinct and could not have been sampled from the same population. The calculated P is the probability that the two sample sets would be obtained by sampling the same population for a given element.

The standard deviation of $\bar{x}_1 - \bar{x}_2$ is estimated by:

$$s = \sqrt{\frac{s_1^2}{n_1} + \frac{s_2^2}{n_2}}$$

where $\bar{x}$ is the sample mean, s is the sample standard deviation and n is the sample size for each sample set. The test is based on the statistic that $t = (\bar{x}_1 - \bar{x}_2) / s$. This statistic has approximately a t-distribution with the degrees of freedom given by:

$$df = \frac{(VAR_1 + VAR_2)^2}{\left[\frac{(VAR_1)^2}{(n_1 - 1)} \right] + \left[\frac{(VAR_2)^2}{(n_2 - 1)} \right]}$$

where VAR is the sample variance for each sample set. The confidence interval is given by $(\bar{x}_1 - \bar{x}_2) - ts$ and $(\bar{x}_1 - \bar{x}_2) + ts$ where t is the value from the t -table corresponding to a specific confidence interval. In this case a 95% confidence interval was used. The probability (P) that a value exceeds the confidence interval is 0.05 at 1.96 standard deviations, marking the points within which 95% of the observations lie. This means that the null hypothesis ($H_0: \mu_1 = \mu_2$) is accepted if the calculated probability (P) is > 0.05 and rejected if P is ≤ 0.05 .

It was necessary to determine if the PSU Method produced data that were statistically similar or different than data produced by Method 29 and the Ontario Hydro Mercury Speciation Method for HAPs and mercury species. Before accepting the PSU Method, it is necessary to accept the null hypothesis ($H_0: \mu_1 = \mu_2$), when comparing the data sets obtained by Method 29 and the PSU Method for multielements and when comparing the sample sets obtained by the Ontario Hydro and PSU Methods for mercury species. Descriptive statistics were also performed to determine the reproducibility of each of the methodologies. Statistical analysis was performed on the inlet and outlet samples for all trace elements and mercury species.

Results and Discussion – Multielements and Total Mercury

Quality of Methodology

Average percent closure across the baghouse for the two methodologies is given in Table 3.3.14. The baghouse ash and fuel were analyzed in the same manner as the particulate samples collected by the sample trains. Average burnout values were calculated from a series of ash samples collected during a test and ranged from 96 to 99 wt%. UNDEERC reported excellent mercury mass balances at $\pm 15\%$ for measurements made at a series of power plants. DeVito et al. (1993) reported that total mercury measurements obtained by Method 29 from a pilot-scale combustor had an accuracy of 20-40%. DeVito and Rosenhoover (1997) later reported on the fate of mercury in coal-fired power plants equipped with wet scrubbers. They sampled flue gas using a Method 29 and an Ontario Hydro sampling train. The Method 29 train was modified to include two deionized water impingers prior to the nitric acid/peroxide impingers. The percent closure reported for mercury for three plants (eleven material balances total) ranged from 72 to 131% with an average of 100%. For each plant the average closure was 103%, 85% and 115%, respectively. The percent closures were within the QA project objective of $\pm 20\%$ closure for the average material balance and no test closure being greater than $\pm 30\%$. The material balances obtained by the PSU Method and Method 29 for mercury are within those obtained by DeVito et al. (1993), DeVito and Rosenhoover (1997) and UNDEER (University of North Dakota Energy and Environmental Research Center 1999).

Table 3.3.14 Average Percent Closure across the Baghouse

Element	PSU Method	Method 29
As	73	110
Be	113	110
Cd	73	120
Co	118	136
Hg	81	110
Ni	110	116
Pb	101	93
Sb	111	113
Se	96	113
Sr	134	52

The material balances of the other elements are within acceptable sampling and analytical error. Miller et al. (1996) reported intralaboratory repeatability for coal analysis averaged 14.6% for trace elements and the percent difference for individual elements ranged from 7.8% for Cr to 33% for Cd.

PSU Method Compared to EPA Method 29

Preliminary tests comparing the sampling and measurement of trace elements by Method 29 and the PSU Method Train were conducted on the Middle Kittanning seam coal (dry micronized coal) fired in Penn State's demonstration boiler. Figure 3.3.40 shows the system capture (lbs/10¹² Btu) measured for a series of elements as a function of method. There was good agreement between the methods. Method 29 tended to overestimate the total mercury as compared to the PSU Method. These preliminary results suggested that a more detailed investigation of the use of the modified Ontario-Hydro Method train as a multielement and mercury speciation method was appropriate.

Figure 3.3.40 SYSTEM CAPTURE MEASURED BY METHOD 29 AND PSU METHOD

Statistical analysis of the data collected at the inlet and outlet of the baghouse showed acceptance of the null hypothesis ($H_0: \mu_1 = \mu_2$) for all trace elements. The statistical analysis demonstrated that the two sample sets generated for each element, by the two methods at the inlet, were from the same population at a 95% confidence interval ($P > 0.05$). This was also true of the data sets generated at the baghouse outlet. A plot of P for each element measured at the baghouse inlet and outlet is shown in Figure 3.3.41. The PSU Method produced data for the multielements that were statistically indistinguishable from data produced by Method 29.

Figure 3.3.41 PROBABILITY DISTRIBUTION OF MULTIELEMENTS MEASURED AT THE BAGHOUSE INLET AND OUTLET AS CALCULATED BY THE TWO SAMPLE T-TEST. (DASHED LINE REPRESENTS $P=0.05$ ACCEPTANCE OF $H_0: \mu_1 = \mu_2$ AT THE 95% CONFIDENCE INTERVAL)

The P calculated at the inlet for all elements ranged from 0.23 for cadmium to 0.93 for arsenic. The P calculated at the outlet for all elements ranged from 0.21 for barium to 1.0 for mercury. Total mercury data had a $P=0.74$ and 1.0 at the inlet and outlet, respectively. Arsenic had a $P=0.93$ and 0.70 in the inlet and outlet sample sets, respectively.

In theory, all else being equal, the P values should be similar at the inlet and outlet, given that the same sampling and analytical methodologies were followed at both sites. However, studies have shown that there are issues of reproducibility for Method 29 (Falcone Miller et al., 1999a, 1999b; Zygarlicke and Pavlish, 1997). These variations are to be expected in the PSU Methodology as well. Therefore the calculated P values will vary reflecting the inherent error in the methodologies. It is noteworthy that the calculated P values for the cadmium data sets measured at the inlet and outlet by the two methods were significantly different, i.e., 0.23 and 0.99, respectively. The only difference between the sampling conducted at the inlet and outlet of the baghouse was the total mass collected of each element. In addition, two more data points for cadmium were available at the baghouse outlet than at the inlet due to operational problems during sampling. This resulted in a difference in the degrees of freedom for the statistical calculations performed on the inlet and outlet data, thereby affecting the t -value for rejecting the H_0 . In addition, the smaller sample size of the Method 29 cadmium data set makes the assumption that the data are representative of the total population less reliable. The mean ($\bar{x}$), standard deviation (s) and standard error of the mean (s_M) for the two sample sets at the baghouse inlet and outlet are given in Tables 3.3.15 and 3.3.16, respectively. The standard error of the mean is the standard deviation of the sampling distribution of the mean and is defined as:

$$s_M = \frac{s}{\sqrt{n}}$$

Ontario Hydro Mercury Speciation Method Compared to EPA Method 29

Statistical analysis of the data collected at the inlet of the baghouse by the Ontario Hydro Mercury Speciation Method was not possible for all the trace elements, as not all the Ontario Hydro test runs were analyzed for all elements. Therefore, there was an insufficient sample size to run a t -test. Statistical analysis of the mercury data generated by Method 29 and the Ontario Hydro Mercury Speciation Method was possible. The result was acceptance of the null hypothesis ($H_0: \mu_1 = \mu_2$) for total mercury at the inlet ($P=0.97$) and the outlet ($P=0.57$).

Table 3.3.15 Descriptive Statistics for Emissions Measured at the Inlet (Units in lb/10¹²Btu)

	METHODOLOGY					
	PSU Method			EPA Method 29		
Element	$(\bar{x})$	(s)	(s_M)	$(\bar{x})$	(s)	(s_M)
As	74.0	18.74	8.38	73.01	9.36	6.61
Be	69.6	6.22	2.78	81.7	19.00	13.5
Cd	1.23	0.54	0.24	2.86	0.79	0.56
Co	322.8	32.6	14.6	358.3	89.8	63.5
Hg	1.36	0.28	0.12	1.31	0.13	0.09
Ni	485.1	20.99	9.38	586.0	148.0	105.0
Pb	124.4	10.44	4.67	131.27	9.44	6.67
Sb	33.2	4.44	1.99	34.35	7.57	5.35
Se	33.6	3.37	1.51	35.58	2.86	2.02
Sr	660.7	163.9	73.3	755.2	106.4	75.2

Table 3.3.16 Descriptive Statistics for Emissions Measured at the Outlet (Units in lb/10¹²Btu)

	METHODOLOGY					
	PSU Method			EPA Method 29		
Element	$(\bar{x})$	(s)	(s_M)	$(\bar{x})$	(s)	(s_M)
As	3.51	3.51	1.75	2.58	2.82	1.75
Be	0.60	.013	0.06	2.66	2.62	1.31
Cd	0.22	0.28	0.14	0.23	0.22	0.11
Co	3.82	2.94	1.47	14.8	17.99	9.00
Hg	0.26*	0.07	0.03	0.26*	0.24	0.12
Ni	7.06	1.89	0.95	17.08	19.01	9.51
Pb	3.86	0.37	0.19	5.17	5.42	2.71
Sb	0.40	0.13	0.07	1.34	1.46	0.73
Se	2.38	1.86	0.93	1.15	1.49	0.74
Sr	9.88	4.77	2.39	38.2	52.1	26.1

*Equivalent to 0.28 μ g/m³

Results and Discussion – Mercury Speciation

PSU Method Compared to Ontario Hydro Mercury Speciation Method

The PSU and Ontario Hydro Mercury Speciation Methods measured the oxidized (Hg^+), elemental (Hg^0) and particulate mercury (Hg_p) at the baghouse inlet and outlet. Method 29 samples were analyzed only for total mercury as it has been shown that the method does not accurately measure mercury species (Zygarlicke and Pavlish, 1997; Laudal and Schultz, 1997). Actual measured emissions at the baghouse inlet are shown in Figure 3.3.42.

FIGURE 3.3.42 MERCURY SPECIES EMISSIONS MEASURED BY PSU METHOD ($n=4$) AND ONTARIO HYDRO MERCURY SPECIATION METHOD ($n=3$) AT THE BAGHOUSE INLET.

(Bars represent one standard deviation from the mean.)

Statistical analysis of the sample sets showed acceptance of the null hypothesis that the sample sets were equal and derived from the same population. The calculated P values for the different mercury species are shown in Figure 3.3.43. The mercury species data sets generated by the PSU Method and Ontario Hydro Mercury Speciation Method are not distinguishable at a 95% confidence interval and $n=4$. P values for elemental mercury were greatest at 0.77 and 0.73

at the inlet and outlet, respectively. Descriptive statistics for the PSU and Ontario Hydro Mercury Speciation Methods sample sets are given in Tables 3.3.17 and 3.3.18.

Figure 3.3.43 PROBABILITY (P) CALCULATED BY TWO SAMPLE T -TEST FOR MERCURY SPECIES MEASURED BY THE PSU METHOD AND ONTARIO HYDRO MERCURY SPECIATION METHOD.

(Dashed line represents $P=0.05$ acceptance of $H_0: \mu_1 = \mu_2$ at the 95% confidence interval.)

Table 3.3.17 Descriptive Statistics for Mercury Emissions Measured at the Baghouse Inlet (Units in lb/10¹²Btu)

Mercury Species	PSU Method			Ontario Hydro Mercury Speciation Method		
	$(\bar{x})$	(s)	(s_M)	$(\bar{x})$	(s)	(s_M)
Hg ⁺	0.27	0.09	0.05	0.22	0.11	0.07
Hg [°]	0.32	0.07	0.03	0.38	0.31	0.18
Hg _p	0.97	0.18	0.09	1.13	0.35	0.20

A significant percentage of the total mercury measured at the inlet is present within the particulate portion, i.e., 48 to 83%. This is not typical of power plant ash. The elevated level of particulate mercury is attributed to the significant amount of carbon present in the char, i.e., 47 to 84%, due to the low level of burnout attained in the test facility and the low ash content of the fuel ($\approx 1.4\%$). The calculated burnout in the baghouse ash averaged 96.1% for the PSU Method

Table 3.3.18 Descriptive Statistics for Mercury Emissions Measured at the Baghouse Outlet (Units in lb/10¹²Btu)

Mercury Species	PSU Method			Ontario Hydro Mercury Speciation Method		
	$(\bar{x})$	(s)	(s_M)	$(\bar{x})$	(s)	(s_M)
Hg ⁺	0.17	0.07	0.03	0.13	0.04	0.02
Hg ^o	0.21	0.08	0.04	0.23	0.08	0.04
Hg _P	0.03	0.01	0.00	0.05	0.02	0.01

tests and 96.7% for the Ontario Hydro Mercury Speciation Method tests. Interestingly, there is a correlation of 0.61 between the percent of carbon present in the char and the percent of particulate mercury measured at the inlet and outlet. This supports the work of others showing the interaction of mercury in the gas phase with char (Livengood et al., 1994; Senior et al., 1997).

The concentration of mercury in the fuel was quite low (0.031 µg/g or 3.31 lbs Hg/10¹² Btu). The coal was cleaned by oil agglomeration making it less than ideal from the standpoint that the total mass of mercury in the system during testing was quite low. The test results represent a worst-case scenario for the methodologies given the high percentage of carbon in the char and low mercury levels.

3.3.3.6 Comparison of PSU Method to ATS Unified Method

ATS conducted comparative sampling of air toxic emissions from DOE's 500 lb/h combustor at the National Energy Technology Laboratory located in Pittsburgh, Pennsylvania. ATS evaluated two sampling methodologies (ATS Methods 1 and 2) in conjunction with the Method 29 and Ontario-Hydro Method (Advanced Technology Systems, Inc 1996a, 1996b). Four tests were conducted during which two different methodologies were run simultaneously. The ATS Methods 1 and 2 are quite different from the PSU Method proposed here (Figure 3.3.44). ATS Method 1 combines the first four impingers (1 M KCl and acidified H₂O₂) of the Ontario-Hydro Method. ATS Method 2 replaces the KCl reagent with 0.5% BaCl₂ combined with acidified H₂O₂.

Figure 3.3.44 PSU AND ATS METHOD TRAINS

In both methodologies, multielements were analyzed in the KCl and BaCl₂ reagents by Inductively Coupled Argon Plasma Atomic Emission Spectroscopy (ICPAES). This analytical technique is appropriate for analysis for multielements in the H₂O₂ impinger solution in Method 29 as recommended by EPA SW 846 Method 6010 (U.S. EPA, 1988). The use of ICP and GFAAS for analysis of the combined KCl and H₂O₂ impinger solutions is not appropriate due to the effect of chlorine on sample viscosity and reduced detection levels. ATS has acknowledged that the use of an ICP and GFAAS for analyzing multielements is not appropriate for the combined KCl and H₂O₂ impinger (ATS Method 1) and the BaCl₂ impinger (ATS Method 2) (McManus, 1999). A significant percentage of the concentrations measured by the ICP and

GFAAS were below instrument detection level. This is not to say that the sample train set up is flawed but rather the analytical protocol used is inappropriate. Therefore, the multielemental data presented by ATS in their reports on “Comparative Method Evaluation for the Assessment of Mercury and Other Metal Air Toxic Emissions from a Pilot-Scale Combustion Unit” dated August and November 1996 is suspect (Advanced Technology Systems, Inc 1996a, 1996b). The ATS Method 1 may be appropriate if another analytical protocol is developed, however, it is our understanding that there is no future plans for additional testing by ATS of their proposed methodologies.

3.3.3.7 Conclusions

Statistical analysis of the data verified the use of the PSU Method for measuring trace elements and mercury species in combustion flue gas. Two sample sets generated for each element by the PSU Method and Method 29 at the inlet were from the same population at a 95% confidence interval ($P > 0.05$). This was also true of the data sets generated at the baghouse outlet. The PSU Method produced data for the multielements that were statistically indistinguishable from data produced by Method 29. The P calculated at the inlet for all elements ranged from 0.23 for cadmium to 0.93 for arsenic. The P calculated at the outlet for all the elements ranged from 0.21 for barium to 1.0 for mercury. Total mercury data had a $P=0.74$ and 1.0 at the inlet and outlet, respectively. Arsenic had a $P=0.93$ and 0.70 in the inlet and outlet sample sets, respectively. Statistical analysis of the sample sets also showed acceptance of the null hypothesis that the sample sets generated by the PSU Method and Ontario Hydro Mercury Speciation Method were equal and derived from the same population with regard to the measurement of mercury species in the flue gas.

The statistics reflect the sometimes high variance in the data obtained by both methodologies for particular elements. This variance is attributed to the issue of reproducibility due to procedural and analytical error. However, both methods produced similar means and levels of variance so as to be statistically indistinguishable. This new method represents a time and cost savings for the complete characterization of trace elements in combustion flue gas.

The study suggests that the PSU Method is capable of simultaneously sampling all inorganic trace elements listed as HAPs and mercury species, and produces data which are

statistically indistinguishable from data produced by Method 29 for multielements and the Ontario Hydro Mercury Speciation Method for mercury species.

3.4 CO₂ Removal

Existing energy utilization systems are largely based on the combustion of fossil fuels in stationary and mobile devices. The increased CO₂ concentration in the atmosphere due to emissions of CO₂ from fossil fuel combustion has caused concerns for global warming (Reuther, 1999). Improving the efficiency of energy utilization and increasing the use of low-carbon energy sources are considered to be potential ways to reduce CO₂ emissions (DOE, 1999a). CO₂ capture and sequestration have been receiving significant attention and are being recognized as a third option (DOE, 1998b). Also, enriched CO₂ streams can be an important starting material for synthetic clean fuels and chemicals (Song, 2002). For carbon sequestration, the costs for capture and separation are estimated to comprise about three-fourths of the total costs of ocean or geologic sequestration (DOE, 1998b). It is therefore important to explore new approaches for CO₂ separation.

There are chemical and physical methods for separation of CO₂ from gas mixtures. Adsorption is one of the promising methods that could be applicable for separating CO₂ from gas mixtures. Various adsorbents, such as zeolites, activated carbons, carbon molecular sieves, pillared clays, and metal oxides have been investigated. At lower temperatures (e.g., room temperature), the zeolite-based adsorbents and activated carbon have generally been found to show higher adsorption capacity. However, their selectivity to CO₂ in the presence of other gases (N₂, etc.) is still low and their adsorption capacities rapidly decline with increasing temperature above 30°C, becoming negligible at temperature in excess of 200°C. It has been reported that the CO₂ adsorption capacity for zeolite 13X, 4A and activated carbon is approximately 160, 135 and 110 mg/g-adsorbent, respectively, at 25°C and 1 atm CO₂ partial pressure (Ananad et al., 1995; Siriwardane et al., 2001). Van der Vaart et al. (1999) investigated the single and mixed gas adsorption equilibrium of CO₂/CH₄ on Norit RBI activated carbon. The CO₂ adsorption capacity was 108 mg/g-adsorbent at 21.5°C. With increasing temperature, the adsorption capacity decreased to 77 mg/g-adsorbent at 30°C, 56 mg/g-adsorbent at 56°C, and 40 mg/g-adsorbent at 75°C. Meanwhile, the activated carbon also adsorbed methane. The CO₂/CH₄ selectivity was around 2. Berlier et al. (1997) and Heuchel et al. (1999) also found that

CO₂ adsorption capacity of activated carbon decreased significantly when the temperature slightly increased and the CO₂/CH₄ selectivity was low. Under 0.1 MPa CO₂ partial pressure, the adsorption capacity was 126 mg/g-adsorbent at 15°C and 78 mg/g-adsorbent at 45°C. The best CO₂/CH₄ ratio (selectivity) was around 2. Since all the gases are physically adsorbed into/onto these carbon and zeolite adsorbents, the separation factors (e.g., CO₂/N₂ ratio) are low. In order to operate at relatively high temperature and reach high separation selectivity, chemical adsorption has been studied. Unfortunately, the adsorbents only show a low CO₂ adsorption capacity at high temperature although the selectivity for CO₂ is high. Anand et al. (1995) reported that MgO showed an adsorption capacity of 8.8 mg/g-adsorbent at 400°C. Ding et al. (2000) investigated the adsorption performance of hydrotalcite and this material showed a CO₂ adsorption capacity of 22 mg/g-adsorbent at 400°C and 0.2 atm CO₂ partial pressure. Both types of adsorbents need high temperature operation and have a low adsorption capacity, thus they are not suitable for practical use for CO₂ separation. For practical applications, selective adsorbents with high capacity are desired. Many of the separations should preferably be operated at elevated temperature, e.g., higher than room temperature and up to ≈150°C for the flue gas of power plants. Developing an adsorbent with high CO₂ selectivity and high CO₂ adsorption capacity, which can also be operated at elevated temperature, is desired for more efficient CO₂ separation by the adsorption approach.

A new concept called “molecular basket”, was investigated to develop a high-capacity, highly-selective CO₂ adsorbent (Xu, et al, 2002a,b; Xu et al., 2003). The proposed CO₂ “molecular basket” adsorbents can selectively “pack” CO₂ in condensed form in the mesoporous molecular sieve “basket” and therefore show high CO₂ adsorption capacity and high CO₂ selectivity. To make the “molecular basket” adsorbent and to capture a large amount of CO₂ gas, the adsorbent needs to have large-pore volume channels to serve as the “basket”. In order to make the “basket” be a CO₂ “molecular basket”, a substance with numerous CO₂-affinity sites is loaded into the pores of the support to increase the affinity between the adsorbent and CO₂ and, therefore, to increase the CO₂ adsorption selectivity and CO₂ adsorption capacity. In addition, the adsorption affinity to CO₂ by the CO₂-philic substance increased in the confined mesoporous environment and therefore the mesoporous molecular sieve can have a synergetic effect on the adsorption of CO₂ by the CO₂-philic substance. In the current investigation, the mesoporous molecular sieve MCM-41, which has a large pore volume, was selected as the “basket”. The

sterically-branched polymer polyethylenimine (PEI), which has branched chains with numerous CO₂-capturing amino groups, was immobilized into the channels of the mesoporous molecular sieve. Branched amines have lower heat of adsorption than that of the primary amines, and thus desorption can be easier and requires less energy. The novel CO₂ “molecular basket” adsorbents were applied in the separation of CO₂ from simulated flue gases and boiler flue gases.

3.4.1 Experimental

3.4.1.1 Preparation of the Adsorbents

Mesoporous molecular sieve MCM-41 was hydrothermally synthesized from a mixture with the following composition: $x\text{Al}_2\text{O}_3:50\text{SiO}_2:4.32\text{Na}_2\text{O}:2.19(\text{TMA})_2\text{O}:15.62(\text{CTMA})\text{Br}:3165\text{H}_2\text{O}$ ($x=0, 0.05, 0.25$). The synthesis procedure was established at Penn State (Reddy and Song, 1996), and is based on the method invented by researchers at Mobil (Beck et al., 1992). Cab-O-Sil fumed silica (Cabot Corporation), tetramethylammonium silicate solution (0.5 TMA/SiO₂, 10 wt.% SiO₂, Sachem Inc.), sodium silicate (containing 14 wt.% NaOH and 27 wt.% SiO₂, Aldrich), cetyltrimethyl-ammonium bromide [(CTMA)Br, Aldrich], aluminum isopropoxide (Aldrich) and deionized water were used as raw materials. The synthesis was carried out at 100°C for 40 h. After the synthesis, the solid product was recovered by filtration, washed several times with deionized water, dried at 100°C overnight and calcined at 550°C for 5 h to remove the template.

The polyethylenimine (PEI) modified MCM-41 was prepared by a wet impregnation method. In a typical preparation, the desired amount of PEI (viscous liquid) was dissolved in 8 g methanol while stirring for about 15 min, after which 2 g calcined MCM-41 was added to the PEI/methanol solution. The resultant slurry was continuously stirred for about 30 min. Then the slurry was dried at 70°C for 16 h under reduced pressure (700 mmHg). The as-prepared adsorbent was denoted as MCM-41-PEI-X, where X represents the loading of PEI as weight percentage in the sample. Different adsorbents were prepared by changing the preparation conditions, such as PEI loadings, preparation methods, preparation procedures, types of solvents, solvent/MCM-41 ratios, Si/Al ratios of MCM-41, and adding additives of polyethylene glycol (PEG). In one adsorbent, commercially available silica gel (SiO₂) (Merck, surface area 550 m²/g) was used as the support instead of MCM-41. Details on the adsorbents and their corresponding preparation conditions are summarized in Table 3.4.1.

Table 3.4.1 Preparation Conditions and the Adsorption/Desorption Performance of the Adsorbents

Sample Name	Si/Al of MCM-41	PEI loading (wt %)	Solvent	Solution/MCM-41 weight ratio	Adsorption Capacity ^a (mg CO ₂ /g-adsorbent)	Desorption capacity (%)
Si-MCM-41	Pure silica	---	---	---	8.6	103
Al-MCM-41-100	100	---	---	---	7.6	101
Al-MCM-41-500	500	---	---	---	7.5	100
Si-MCM-41-PEI-5	Pure silica	5	Methanol	4:1	7.7	105
Si-MCM-41-PEI-15	Pure silica	15	Methanol	4:1	19.4	103
Si-MCM-41-PEI-30	Pure silica	30	Methanol	4:1	68.7	98
Si-MCM-41-PEI-50	Pure silica	50	Methanol	4:1	112	99
Si-MCM-41-PEI-75	Pure silica	75	Methanol	4:1	133	101
Si-MCM-41-PEI-25-25 ^b	Pure silica	50	Methanol	4:1	96.8	97
Al-MCM-41-100-PEI-50	100	50	Methanol	4:1	127	100
Al-MCM-41-500-PEI-50	500	50	Methanol	4:1	121	99
Si-MCM-41-PEI-50-H ₂ O	Pure silica	50	Water	4:1	111	100
PEI-30-Si-MCM-41 ^c	Pure silica	30	Methanol	4:1	65.7	99
Si-MCM-41-PEI-50-M ^d	Pure silica	50	---	--	99	97
Si-MCM-41-PEI-50-S2	Pure silica	50	Methanol	2:1	97	98
Si-MCM-41-PEI-50-S8	Pure silica	50	Methanol	8:1	126	99
Si-MCM-41-PEI-30-PEG-20 ^e	Pure silica	30	Methanol	4:1	77.1	100
SilicaGel-PEI-50 ^f	---	50	Methanol	4:1	78.0	98
PEI	---	100	---	---	109	56

^a The adsorption capacity was measured by TGA under pure CO₂ atmosphere at a flow rate of 100 ml/min at 75°C.

^b The adsorbent was prepared by two-step impregnation method. In each step, 25 wt % PEI was loaded.

^c The adsorbent was prepared by adding the PEI/methanol solution to the MCM-41 powder.

^d The adsorbent was prepared by mechanical mixing method.

^e PEG was added to the adsorbent of MCM-41-PEI-30. The weight percentage of PEG was 20 wt%.

^f Silica Gel was used as support material, instead of MCM-41.

3.4.1.2 Characterization of the Adsorbents

The mesoporous molecular sieve MCM-41, before and after modification, and before and after adsorption separation, was characterized by X-ray diffraction (XRD) and N₂ adsorption/desorption. The XRD patterns were obtained on a Rigaku Geigerflex using Cu K α radiation. The N₂ adsorption/desorption was carried out on a Quantachrome Autosorb 1 automated adsorption apparatus, from which the BET surface area, the pore volume and the pore size were obtained. The sample was out-gassed at 75°C for 48 h on a high vacuum line prior to adsorption. The pore volume of the mesoporous molecular sieve was calculated from the adsorbed nitrogen after complete pore condensation ($P/P_0=0.995$) using the ratio of the densities of liquid and gaseous nitrogen. The pore size was calculated by using the BJH method. The thermal chemical and physical properties of MCM-41, PEI and MCM-41-PEI were characterized by thermal gravimetric analysis (TGA). The TGA was performed on a PE-TGA 7 thermal gravimeter. About 10 mg of the sample was heated at 10°C/min to 600°C in airflow (100 ml/min).

3.4.1.3 Measurement of the CO₂ Adsorption/Desorption Performance

The adsorption and desorption performance of the adsorbent was measured using a PE-TGA 7 analyzer. The weight change of the adsorbent was used to determine the adsorption and the desorption performance. In a typical adsorption/desorption process, about 10 mg of the adsorbent was placed in a small sample cell, heated up to 100°C in a N₂ atmosphere at a flow rate of 100 ml/min, and held at that temperature (about 30 minutes) until there was no loss in weight. Then the temperature was decreased to the desired temperature and the 99.8 % bone-dry CO₂ adsorbate was introduced at a flow rate of 100 ml/min. In one experiment CO₂/N₂ gas mixtures were also used. After the adsorption, the gas was switched to 99.995% pure N₂ at a flow rate of 100 ml/min to perform desorption at the same temperature. The time for both the adsorption and desorption was 150 minutes. Adsorption capacity in mg of adsorbate/g-adsorbent and desorption capacity in percentage were used to characterize the adsorbent and were calculated from the weight change of the sample in the adsorption/desorption process. Desorption capacity in percentage was defined as the ratio of the amount of the gas desorbed over the amount of gas adsorbed. The impact of adsorption/desorption temperature and the CO₂

concentration in the adsorbate on the adsorption and desorption performance of the adsorbent was studied.

3.4.1.4 Adsorption Separation of Simulated Flue Gas Mixture

A gas mixture containing 14.9% CO₂, 4.25% O₂ and 80.85% N₂ was used as the adsorbate. In some experiments, the gas mixture was mixed with 10% moisture and used as the adsorbate.

The adsorption separation was carried out in a mobile flow adsorption system as shown in Figure 3.4.1. In a typical adsorption/desorption run, about 2.0 g of the powder adsorbent was placed in a stainless steel adsorption column (O.D., 1/2'; I.D., 3/8'), heated to 100°C in a helium atmosphere at a flow of 50 ml/min and held at that temperature overnight to ensure that there was no carbon dioxide or moisture adsorbed. The temperature was then decreased to 75°C and the dry or moist CO₂+N₂+O₂ gas mixture was introduced at 10 ml/min. Generally, the adsorption was carried out for 120 minutes. After the adsorption, the gas was switched to pure helium at a flow rate of 50 ml/min to perform the desorption at the same temperature. The time for desorption was 300 minutes. The flow rate of the effluent gas was measured by a soap-film flowmeter. The concentration of the effluent gas was measured on-line using a SRI 8610 C Gas Chromatography (GC) with TCD detector. The analysis was carried out every 5 minutes when the dry CO₂+N₂+O₂ gas mixture was used as the adsorbate and every 15 minutes when the moist CO₂+N₂+O₂ gas mixture was used as the adsorbate. Adsorption capacity in ml (STP) of adsorbate/g adsorbent and desorption capacity in percentage were used to characterize the adsorbent. The adsorption/desorption capacity was calculated from mass balance before and after the adsorption. The separation factor was defined as the mole ratio of the gases adsorbed over the mole ratio of the gases in the feed.

3.4.1.5 Adsorption Separation of a Boiler Flue Gas Mixture

The adsorption separation of CO₂ from boiler flue gases was investigated. Both natural gas-fired and coal-fired flue gases were used as feed gas. The composition of the natural gas-fired flue gas mixture was 7.4-7.7% CO₂, ~ 4.45% O₂, 200-300 ppm CO, 60-70 ppm NO_x, 14.6% H₂O and 73-74% N₂. The composition of the coal-fired flue gas mixture was 12.5-12.8% CO₂,

Figure 3.4.1 SET-UP OF THE ON-LINE CO₂ MONITORING SYSTEM

~4.4% O₂, 50 ppm CO, 420 ppm NO_x, 420 ppm SO₂, 6.2% H₂O and 76-77% N₂. Analysis of the effluent gas composition was carried out after removing water vapor from the gas mixture.

In a typical adsorption/desorption process, about 30 g of the adsorbent was placed in the central part of a stainless steel adsorption column (O.D., 2'; I.D., 1.7'). The adsorbent was pressed to 18-35 mesh. The beginning and the end of the adsorption column were filled with alumina (≈170 g) to decrease the dead volume in the separation system. The adsorption column was heated to 100°C in a helium atmosphere at a flow of 5 l/min and held at that temperature until there was no carbon dioxide detected in the effluent gas. The temperature was then adjusted to 80±10°C and the flue gas mixture was introduced at 5-6 l/min. Generally, the adsorption was carried out for 300-600 seconds. After the adsorption, the gas was switched to helium at a flow rate of 5 l/min to perform the desorption at the same temperature. The time for desorption was 300-600 seconds. The flow rate of the effluent gas was measured by a rotameter. The concentration of O₂, CO, CO₂, SO₂ and NO_x in the effluent gas was measured on-line using a model NGA 2000 paramagnetic oxygen analyzer; a model NGA 2000 non-dispersive infrared CO analyzer; a model NGA 2000 non-dispersive infrared CO₂ analyzer; a model 890 ultraviolet

SO₂ analyzer; and a model NGA 2000 chemiluminescence NO_x analyzer (Rosemount Analytical, Inc.), respectively. The analysis was carried out every 5-6 seconds. Since the alumina also adsorbed the gases, a blank separation test with the column filled with alumina (≈ 210 g) was carried out. Therefore, the adsorption/desorption capacity of the “molecular basket” adsorbent was calculated as the difference between the adsorption/desorption capacity from the adsorption experiment and blank experiment. The adsorption/desorption capacity was calculated from mass balance before and after the adsorption. The separation factor was defined as the mole ratio of the gases adsorbed over the mole ratio of the gases in the feed.

3.4.2 Results and Discussion

3.4.2.1 Preparation of the CO₂ “Molecular Basket” Adsorbent

Preparation and Characterization of the CO₂ “Molecular Basket” Adsorbent

Si-MCM-41-PEI with different PEI loadings was prepared and characterized by XRD, N₂ adsorption/desorption and TGA. Figure 3.4.2 shows the XRD patterns of Si-MCM-41 and Si-MCM-41-PEI with different PEI loadings. From comparing the diffraction patterns of Si-MCM-41 with those of Si-MCM-41-PEI with different PEI loadings, the degrees of the Bragg diffraction angles were nearly identical, indicating that the structure of Si-MCM-41 was preserved after loading the PEI. However, the intensity of the diffraction patterns of Si-MCM-41 decreased after the PEI was loaded. With increasing PEI loadings, the intensity of the diffraction peaks decreased. The intensity of the diffraction peaks of Si-MCM-41-PEI-50 and Si-MCM-41-PEI-75 were only 11% of that of the Si-MCM-41 support. At the same time, the degree of the Bragg diffraction angle of the 100 plane slightly increased from 2.265 for Si-MCM-41 to 2.295~2.325 for Si-MCM-41-PEI. These changes were possibly caused by the pore filling effect of the MCM-41 channels and the PEI coating on the outer surface of the MCM-41 crystals. Reddy and Song (1996) reported that the XRD patterns of MCM-41, after removal of the template, exhibited peaks with increased intensity and a shift to lower Bragg diffraction angle compared to the template-containing MCM-41. Since the pore volume of the Si-MCM-41 support is 1.0 ml/g and the density of PEI is about 1.0 g/ml, the maximum PEI loading in the pores of Si-MCM-41 is 50 wt %. More PEI should be coated on the outer surface of Si-MCM-

Figure 3.4.2 XRD PATTERNS OF Si-MCM-41-PEI WITH DIFFERENT PEI LOADINGS AS SHOWN BY THE TRAILING DIGITS (The top three curves were shown at different Y-scales.)

41 crystals for Si-MCM-41-PEI-75 than for Si-MCM-41-PEI-50. However, the diffraction intensity of Si-MCM-41 for Si-MCM-41-PEI-50 and Si-MCM-41-PEI-75 was nearly the same, which indicated that the PEI coating on the outer surface of the crystals hardly influenced the diffraction intensity of the Si-MCM-41 support. Therefore, the decrease in the diffraction intensity and the shift of the Bragg diffraction angle of the 100 plane to a high degree can be ascribed mainly to the loading of PEI into the Si-MCM-41's pore channels.

The pore size, surface area and pore volume of Si-MCM-41 before and after loading the PEI were obtained from the nitrogen adsorption/desorption isotherms. Figure 3.4.3 shows the pore size distributions of Si-MCM-41-PEI with different PEI loadings and Figure 3.4.4 shows their surface areas and pore volumes. The pore size of the Si-MCM-41 support was 2.75 nm. After the PEI was loaded into its channels, the pore size decreased. The pore size of Si-MCM-41-PEI-5 was 2.47 nm, smaller than that of the Si-MCM-41 support, which confirmed that PEI was loaded into the Si-MCM-41 pore channels. With increasing PEI loading, the pore size

Figure 3.4.3 THE PORE SIZE DISTRIBUTION OF Si-MCM-41-PEI WITH DIFFERENT PEI LOADINGS

Figure 3.4.4 THE SURFACE AREA AND PORE VOLUME OF Si-MCM-41-PEI WITH DIFFERENT PEI LOADINGS

further decreased. The pore sizes were 2.19 nm and 1.68 nm for Si-MCM-41-PEI-15 and Si-MCM-41-PEI-30, respectively. When the PEI loading was further increased to 50 wt %, the mesoporous pore channels were completely filled with PEI, restricting the access of nitrogen into the pores at the liquid nitrogen temperature. Therefore, information on the pore size cannot be obtained from the N₂ adsorption/desorption isotherms for PEI loadings above 50 wt %. This is also consistent with the estimated maximum PEI loading of 50 wt % inside the pores of Si-MCM-41 (based on the pore volume of the Si-MCM-41 and the density of PEI).

The surface area and the pore volume of Si-MCM-41, after loading the PEI, exhibited the same trends as the pore size. The Si-MCM-41 support had a surface area of 1,486 m²/g and a pore volume of 1.0 ml/g. The surface area and the pore volume decreased after PEI was loaded into its channels. For example, when the PEI loading was 5.0 wt %, the adsorbent showed a pore volume of 0.79 ml/g, smaller than that of the Si-MCM-41 support, which confirmed that PEI was loaded into the Si-MCM-41's channels. With increasing PEI loadings, the surface area and the pore volume further decreased. When the PEI loading was higher than 30 wt %, the surface area decreased sharply, which indicated that some of the mesoporous pores were completely filled with PEI. The surface area was only 4.2 m²/g and the residual pore volume was only 0.011 ml/g for Si-MCM-41-PEI-50 at the liquid nitrogen temperature. These results correlate with the pore filling effect of the PEI, which was also evident from the XRD characterization, and further confirms that PEI was loaded into the pore channels of Si-MCM-41.

The thermal chemical and physical properties of Si-MCM-41, PEI and Si-MCM-41-PEI were measured by TGA. Figure 3.4.5 shows the TGA results and Figure 3.4.6 shows the differential thermal gravimetric analysis (DTGA) results. As expected, there was no weight loss for Si-MCM-41 below 600°C, which indicated that the hydrophobic Si-MCM-41 was free from adsorbed water or other gases at room temperature. The PEI lost 3.8% of its original mass at 100°C, which can be mainly ascribed to the desorption of CO₂ and moisture. This was confirmed by analyzing the effluent gas with a gas chromatograph (GC). This also indicated that PEI has a low vapor pressure and unlike the commercially used amines such as diethanolamine (DEA), makes PEI suitable for long-term use at relatively high temperature. The PEI began to decompose above 150°C and a sharp weight loss appeared at 205°C. When the temperature was increased above 225°C, the rate of weight loss decreased, indicating that a different decomposition process took place. At 600°C, the PEI was completely decomposed and removed

Figure 3.4.5 TGA PROFILE FOR Si-MCM-41-PEI WITH DIFFERENT LOADINGS

Figure 3.4.6 DTGA PROFILE FOR Si-MCM-41-PEI WITH DIFFERENT PEI LOADINGS

as volatiles. After the PEI was loaded into the Si-MCM-41's channels, the sharp weight loss of the PEI took place at a lower temperature, which indicated that the decomposition temperature of PEI decreased. The weight loss also took place over a narrower temperature range than that of the pure PEI. These phenomena can be explained by the uniform dispersion of PEI into the nano-porous support pores, since the melt or decomposition temperature will decrease when the particle size of a substance decreases (Zeng, et al., 1998). In addition, when the PEI loading was below 50 wt %, Si-MCM-41-PEI showed a ~10.5% weight loss at 100°C, which was higher than that of the pure PEI and can be ascribed also to desorption of CO₂ and moisture. When the PEI loading was 75 wt %, the weight loss was only ~3.5% at 100°C, similar with that of the pure PEI, which indicated that PEI was coated on the outer surface of the molecular sieve crystals and the composite material behaved like the pure PEI particles. The different weight losses observed at 100°C, between the Si-MCM-41-PEI-50 and Si-MCM-41-PEI-75, further confirmed that the PEI was loaded into the MCM-41's channels. The weight loss of Si-MCM-41-PEI also indicated that there was no PEI loss during the preparation process. For example, the total weight loss for Si-MCM-41-PEI-50 was ~56% at 600°C. If the adsorbed moisture or CO₂ is excluded from the total weight loss, then the PEI loading is calculated to be about 50 wt %, which is in accordance with the designed PEI loading.

Adsorption Performance of MCM-41-PEI

The Influence of Operating Temperature on CO₂ Adsorption

The adsorption and desorption performances of the MCM-41-PEI-50 at different temperatures in pure CO₂ atmosphere were measured and the results are shown in Figure 3.4.7. With increasing temperatures, the adsorption capacity of the MCM-41-PEI-50 increases and reaches its maximum of 112 mg/g adsorbent at 75°C. When the temperature was increased to 100°C, the adsorption capacity slightly decreased to 110 mg/g adsorbent. The desorption was not complete at low temperature. When the temperature was 50°C, about 20% of the CO₂ desorbed. The desorption was complete when the temperature reached 75°C. However, the desorption capacity decreased when the temperature was 100°C.

The adsorption of CO₂ into PEI or MCM-41 is an exothermic process (Kohl and Nielson, 1997). Accordingly, the adsorption capacity should decrease with increasing temperature. However, the adsorption capacity increased with increasing temperature in this study. At low

Figure 3.4.7 THE INFLUENCE OF OPERATING TEMPERATURE ON THE ADSORPTION AND DESORPTION PERFORMANCE OF Si-MCM-41-PEI-50

temperature, the PEI exists in the channels of MCM-41 like nano-sized particles. In this case, only the CO₂ affinity sites on the surface of the particles can readily react with CO₂. The affinity sites inside the nano-sized particles can only react with CO₂ when the CO₂ has diffused into the particles. This is a kinetically (diffusion)-controlled process, since the access of CO₂ to the affinity sites inside the PEI particles will be subject to the diffusion limitation, and CO₂ may reach more affinity sites if the diffusion time is sufficiently long. In this case, more affinity sites will be exposed to the CO₂ and thus the adsorption capacity will increase when using a short adsorption time.

In order to verify the above hypothesis, the following experiment was designed. First, the adsorption of CO₂ was carried out at 75°C in flowing CO₂ for 150 min, the temperature was then decreased to 25°C and held at that temperature for another 150 min. If the adsorption capacity increased at low temperature, it can be concluded that the low adsorption capacity at low temperature was caused by the kinetic limitation. If the adsorption capacity decreased, other reasons may cause the low adsorption capacity at low temperature. The experimental results are shown in Figure 3.4.8. The adsorption capacity was 110 mg/g-adsorbent at 75°C. When the temperature was decreased to 25°C and held for 150 min, the adsorption capacity increased to

Figure 3.4.8 TEMPERATURE PROGRAMMED ADSORPTION OF CO₂ BY Si-MCM-41-PEI-50 (Pure CO₂ atmosphere, CO₂ flow rate: 100 ml/min)

112 mg/g adsorbent, which means that the adsorption is kinetically controlled. The low adsorption capacity at low temperature is therefore a result from the low adsorption rate. The adsorption capacity at low temperature will be larger than that at high temperature if the adsorption time is long enough to ensure that equilibrium is attained.

The Influence of CO₂ Concentration on CO₂ Adsorption

Because the concentration of CO₂ varies in different gas mixtures, from about 13% in flue gas from coal-fired power plants to about 0.6% in space vehicles (Satyapal et al., 1973), the influence of CO₂ concentration in a CO₂/N₂ gas mixture on the adsorption and desorption performance of MCM-41-PEI-50 was investigated at 75°C. The results are shown in Figure 3.4.9. Since the adsorption of N₂ is negligible in the presence of CO₂, the adsorption amount measured for the CO₂/N₂ mixture was considered to be the CO₂ adsorption capacity. For pure CO₂, the adsorption capacity was 112 mg/g adsorbent. With a decrease in the CO₂ concentration, the adsorption capacity of the MCM-41-PEI-50 also decreased somewhat. When the CO₂ concentration in the feed decreased to 30%, the adsorption capacity decreased to 102

Figure 3.4.9 THE INFLUENCE OF CO₂ CONCENTRATION IN THE CO₂/N₂ MIXTURE ON THE ADSORPTION PERFORMANCE OF Si-MCM-41-PEI-50 (Operating temperature 75°C, flow rate 100 ml/min)

mg/g adsorbent and was about 9% lower than that in the pure CO₂ atmosphere. When the CO₂ concentration was further decreased to 8% and 2%, the adsorption capacities decreased to 90.4 and 67.5 mg/g respectively. Even when the CO₂ concentration was 0.5% (i.e., 3.8 mmHg), the MCM-41-PEI-50 still effectively captured CO₂ from the gas mixture. The adsorption capacity was 47.6 mg/g adsorbent at a CO₂ concentration of 0.5%. The desorption of the CO₂ was complete over the CO₂ concentration range investigated.

The main reaction responsible for CO₂ interaction with amine (chemical adsorption) is believed to be carbamate formation:

Since the experiment was carried out in a flow system, the amount of CO₂ was in significant excess. However, the adsorption capacity was related to the CO₂ concentration in the gas mixture at a given temperature, which implied that the reaction was partially thermodynamically controlled. In order to increase the adsorption capacity, the partial pressure of the CO₂ must be increased. But the experimental data showed that the adsorption capacity seems to be related to the logarithm of the CO₂ pressure. Increasing the CO₂ pressure will lead to less increase in the

adsorption capacity and may increase the operating cost. The other optional way is to make MCM-41-PEI materials in a membrane form. In this case, the reaction (chemical adsorption) will occur at the feed side and the inverse reaction (desorption after chemisorption) will occur at the permeate side. Thus, the CO₂ can be recovered. In addition, the separation can be operated continuously.

The Influence of Preparation Conditions on the Adsorption Performance of MCM-41-PEI

The Effect of PEI Loadings on CO₂ Adsorption

The influence of PEI loading on the CO₂ adsorption performance of Si-MCM-41-PEI was investigated in a pure CO₂ atmosphere at 75°C and the results are shown in Figure 3.4.10. Before the PEI was loaded, the Si-MCM-41 support alone showed a CO₂ adsorption capacity of 8.6 mg/g-adsorbent. The low adsorption capacity was caused by the weak interaction between CO₂ and Si-MCM-41 at relatively high temperature. In order to strengthen the interaction between CO₂ and Si-MCM-41, the branched polymeric substance PEI with numerous CO₂-capturing sites, was loaded into the Si-MCM-41 channels. Surprisingly, the PEI made no contribution to the CO₂ adsorption capacity at low PEI loading. When the PEI loading was 5.0 wt %, the CO₂ adsorption capacity was only 7.7 mg/g-adsorbent, nearly identical to that of the Si-MCM-41 support. When the PEI loading was further increased, the adsorption capacity increased. The adsorption capacities were 19 mg/g-adsorbent and 69 mg/g-adsorbent for PEI loadings of 15 wt % and 30 wt %, respectively. At 50 wt % PEI loading, the adsorption capacity was 112 mg/g-adsorbent, which is higher than that of the pure PEI of 110 mg/g-adsorbent. The mesoporous molecular sieve of Si-MCM-41 showed a synergetic effect on the adsorption of CO₂ by PEI. The highest adsorption capacity of 133 mg/g-adsorbent was obtained when the PEI loading was 75 wt %.

The desorption was complete for all the Si-MCM-41-PEI adsorbents as well as for the Si-MCM-41 support. However, the desorption for pure PEI was slow and was not complete in the desorption time of the Si-MCM-41-PEI adsorbents. The fast desorption of CO₂ from the Si-MCM-41-PEI adsorbents can be explained by the high dispersion of the PEI into the Si-MCM-41 channels as shown by the XRD, N₂ adsorption/desorption and TGA characterizations.

Figure 3.4.10 THE INFLUENCE OF PEI LOADINGS ON THE ADSORPTION AND DESORPTION PERFORMANCE OF Si-MCM-41-PEI

The Effect of Preparation Methods and Preparation Procedures on CO₂ Adsorption

Two methods, i.e., wet impregnation and mechanical mixing, were employed for the preparation of Si-MCM-41-PEI. For the mechanical mixing method, PEI was solidified, ground into powder and mixed uniformly with Si-MCM-41 at the liquid nitrogen temperature. The powder mixture was heated to 70°C and held at that temperature for 16 h under reduced pressure (700 mmHg). The PEI loading was 50 wt %. The adsorption performance of this adsorbent was compared with that of the adsorbent prepared by the wet impregnation method in Table 3.4.1. The adsorption capacity was 99 mg/g-adsorbent for the adsorbent prepared by the mechanical mixing method (Si-MCM-41-PEI-50-M) and 112 mg/g-adsorbent for the adsorbent prepared by the wet impregnation method (Si-MCM-41-PEI-50). The adsorption capacity of Si-MCM-41-PEI-50-M was lower than that of Si-MCM-41-PEI-50. The higher adsorption capacity for the wet impregnation method was possibly caused by the uniform dispersion of PEI into the Si-MCM-41 channels. However, if the PEI and the Si-MCM-41 were simply mechanically mixed in Si-MCM-41-PEI-50-M, the adsorption capacity of MCM-41-PEI-50-M should be the linear sum of the adsorption capacities contributed by the Si-MCM-41 and PEI, and was calculated to be 58.9 mg/g-adsorbent. The adsorption capacity of Si-MCM-41-PEI-50-M was higher than this

value, which indicated that part of the PEI may also be loaded into the Si-MCM-41 channels or be coated on the external surface of the support crystals during the thermal treatment. The desorption for both of the adsorbents was complete.

In addition, two different impregnation procedures were examined. In the first set of experiments, the adsorbent was prepared by adding the PEI-methanol solution to Si-MCM-41 (PEI-30-Si-MCM-41). Its adsorption performance was compared with that of the adsorbent prepared by adding Si-MCM-41 to the PEI-methanol solution (Si-MCM-41-PEI-30) in Table 3.4.1. The adsorption/desorption capacities of these two adsorbents were similar, which indicates that the impregnation procedure did not influence the dispersion of PEI into the Si-MCM-41 channels nor the adsorption performance of the adsorbent.

In the second set of experiments, the adsorbent was prepared by the following procedure: Si-MCM-41 was first impregnated with 25 wt % PEI, then dried under vacuum at 70°C, and another 25 wt % PEI was then loaded. The adsorption performance of this adsorbent (Si-MCM-41-PEI-25-25) was compared with that of the adsorbent loaded with 50 wt % PEI in one step (Si-MCM-41-PEI-50) in Table 3.4.1. The adsorption capacity of Si-MCM-41-PEI-50 was higher than that of Si-MCM-41-PEI-25-25, which indicates that one-step impregnation is a better method than two-step impregnation. Since PEI is hydrophilic in nature, the PEI impregnated in the first step may adsorb the methanol and prevent the PEI from entering the MCM-41 channels in the next step of impregnation. Therefore, PEI may coat the external surface of the support crystals in the second impregnation and the adsorption capacity is decreased.

The Effect of Solvent Type and Methanol/MCM-41 Weight Ratio on CO₂ Adsorption

Since the Si-MCM-41 that was used was pure silica and was hydrophobic in nature, the polarity of the solvent for the dissolution of PEI may influence the dispersion of PEI into the Si-MCM-41 channels. Two solvents, i.e., methanol and water, were used to examine the effect of solvent on adsorbent performance. The PEI loading for both adsorbents was 50 wt % and the adsorption results are shown in Table 3.4.1. The adsorption capacity was nearly the same for the two adsorbents, which indicated that these two solvents had limited influence on the adsorption performance of the resulting adsorbent. The results also indicate that PEI is more hydrophobic than both water and methanol, and was therefore preferentially adsorbed into the Si-MCM-41 channels from the solution. Because methanol is easier to volatilize than water and the Si-MCM-

41 structure is not stable under a steam atmosphere at high temperature, methanol was selected as the solvent for the dissolution of PEI in the subsequent study.

The influence of the methanol/Si-MCM-41 weight ratio on the adsorption performance of the resultant adsorbents was also examined and the adsorption results are shown in Table 3.4.1. From the adsorption results, it can be concluded that the adsorption capacity increased with increasing methanol/Si-MCM-41 weight ratio. The highest adsorption capacity of 126 mg/g-adsorbent was obtained when the methanol/Si-MCM-41 weight ratio was 8, which was 23% higher than that of the adsorbent prepared with a methanol/Si-MCM-41 weight ratio of 2. When the methanol/Si-MCM-41 weight ratio was 2, the Si-MCM-41 cannot be wetted well by the PEI/methanol solution. In this case, the PEI cannot be dispersed well and the adsorption capacity was nearly equal to that of the adsorbent prepared by the mechanical mixing method. When the methanol/Si-MCM-41 weight ratio was increased to 4, MCM-41 can be wetted by the PEI/methanol solution, allowing more PEI to be loaded into the Si-MCM-41 channels and thus increasing the adsorption capacity.

The Effect of Polyethyleneglycol (PEG) as an Additive on CO₂ Adsorption

The influence of PEG on the adsorption performance of Si-MCM-41-PEI was investigated. Figure 3.4.11 compares the adsorption and desorption performance of Si-MCM-41-PEI-30 and Si-MCM-41-PEI-30-PEG-20. The CO₂ adsorption capacity and the CO₂ adsorption/desorption rate for Si-MCM-41-PEI changed after adding the PEG to Si-MCM-41-PEI. The adsorption capacity of Si-MCM-41-PEI-30-PEG-20 was higher than that of Si-MCM-41-PEI-30. In addition, the adsorption/desorption rates for Si-MCM-41-PEI-30-PEG-20 were faster than those for Si-MCM-41-PEI-30. Satyapal et al. (2001) also found that adding PEG to the PEI/polymer adsorbent can accelerate the CO₂ adsorption and desorption rates. The authors suggested that the enhancement of the adsorption and desorption rates was related to the preponderance of OH⁻ ions from the PEG molecules. They also suggested that the PEG attracts more water to the adsorbent due to the hydroscopic nature of the chemical. Since the adsorbate was pure CO₂ in the current experiment and there was no moisture in the gas, the enhanced adsorption capacity in this study must emanate from the CO₂ alone.

Figure 3.4.11 THE EFFECT OF PEI ON THE CO₂ ADSORPTION/DESORPTION PERFORMANCE OF Si-MCM-41-PEI (PEI loading: 30 wt %, PEG loading: 20 wt %)

The Effect of the Si/Al Ratio on the MCM-41 Support on CO₂ Adsorption

The above results were derived using pure silica of MCM-41. To clarify whether aluminum incorporation into MCM-41 framework can affect the property of the MCM-41-PEI adsorbent, the influence of the Si/Al ratio of the MCM-41 support on the adsorption performance of the resultant adsorbents was examined. Figure 3.4.12 shows the adsorption capacity of MCM-41 and MCM-41-PEI-50 with different support Si/Al ratios. The adsorption capacity was nearly identical for the MCM-41 support with different Si/Al ratios. However, the adsorption capacity of MCM-41-PEI was influenced by the Si/Al ratio of the MCM-41 support. The lower the Si/Al ratio of the MCM-41 support, the higher the adsorption capacity of the resultant adsorbent. The adsorption capacity of Al-MCM-41-100-PEI-50 was 127 mg/g-adsorbent and was 12 % higher than that of Si-MCM-41-PEI-50.

The Effect of the “Molecular Basket” on CO₂ Adsorption

In order to reveal any synergetic effects between MCM-41 and PEI on CO₂ adsorption, the linear adsorption capacity and the synergetic adsorption gain were defined as follows:

Figure 3.4.12 THE INFLUENCE OF Si/Al RATIO OF THE MCM-41 SUPPORT ON THE CO₂ ADSORPTION PERFORMANCE OF MCM-41 and MCM-41-PEI (PEI loading: 50 wt %)

Linear adsorption capacity (mg adsorbate/g-adsorbent) = [(MCM-41 weight percentage in the adsorbent × Adsorption capacity of pure MCM-41) + (PEI weight percentage in the adsorbent × Adsorption capacity of pure PEI)]

Synergetic adsorption gain (mg adsorbate/g-adsorbent) = Adsorption capacity of the adsorbent – Linear adsorption capacity

The linear adsorption capacity and the synergetic adsorption gain were calculated and are shown in Figure 3.4.13. For Si-MCM-41-PEI-5 and Si-MCM-41-PEI-15, the synergetic adsorption gain was close to zero, indicating that the loading of PEI into the Si-MCM-41 channels had no effect on the increase in the CO₂ adsorption capacity at these low PEI loadings. When the PEI loading was higher than 30 wt %, a synergetic adsorption gain was observed, which indicated that PEI modified Si-MCM-41 began to act as a CO₂ “molecular basket”. The highest synergetic gain was obtained at a PEI loading of 50 wt %, while the synergetic adsorption gain decreased at the PEI loading of 75 wt %.

By correlating the adsorption performance and the pore structure of Si-MCM-41 and Si-MCM-41-PEI, a simple model was proposed in Figure 3.4.14 to account for the synergetic effect of Si-MCM-41 on CO₂ adsorption by PEI. Since the adsorption was operated at a relatively high temperature and the pore size of Si-MCM-41 is in the mesoporous region, the CO₂ adsorption

Figure 3.4.13 SYNERGETIC EFFECT OF Si-MCM-41 ON THE ADSORPTION OF PEI AS A FUNCTION OF PEI LOADING IN MCM-41-PEI

Figure 3.4.14 SCHEMATIC DIAGRAM OF PEI LOADED IN THE MESOPOROUS MOLECULAR SIEVE OF MCM-41 ((A) MCM-41 support; (B) Low PEI loading; (C) High PEI loading; (D) Extremely high PEI loading)

capacity of Si-MCM-41 alone is low. When the PEI, which has branched chains with numerous CO₂ adsorption sites of the amine group, is loaded into the Si-MCM-41 channels, both the physical adsorption by capillary condensation and the chemical adsorption by reaction with PEI will contribute to the adsorption capacity. It may be expected that the adsorption capacity of Si-MCM-41-PEI would increase with increasing amount of PEI loaded. However, only at high PEI loading did Si-MCM-41-PEI show a synergetic effect on CO₂ adsorption, indicating a strong interaction between Si-MCM-41 and PEI. When the PEI loading was low, the pore size of the adsorbent slightly decreased, from 2.75 nm for Si-MCM-41 to 2.47 nm for Si-MCM-41-PEI-5. In this case, the PEI was mostly adsorbed on the inner pore wall of the Si-MCM-41 support (Figure 3.4.14B). Both physical adsorption and chemical adsorption contribute to CO₂ uptake in this case. The adsorption capacity was even smaller than that of the linear adsorption capacity. With increasing PEI loadings, the pore size further decreased. At the same time, because more PEI was loaded into the channels, the chemical adsorption of CO₂ became much more dominant than that at low PEI loadings. The physical adsorption on the unmodified pore wall of Si-MCM-41 (and the capillary condensation in the mesopore) became negligible when compared to the chemical adsorption force. In addition, mesoporous molecular sieve Si-MCM-41 had a synergetic effect on the adsorption of CO₂ by PEI in the confined mesoporous pores (Xu et al., 2002a). Therefore, the CO₂ adsorption capacity increased. The highest synergetic adsorption gain (Figures 3.4.13 and 3.4.14C) was obtained when the mesoporous channels were completely filled with PEI, i.e., about 50 wt % of PEI loading. When the PEI loading further increased to 75 wt % (Figures 3.4.13 and 3.4.14D), the synergetic adsorption gain was smaller than that for the PEI loading of 50 wt %. Since the pore volume of Si-MCM-41 is about 1.0 ml/g and the density of PEI is about 1.0 g/ml, the largest amount of PEI that can be theoretically loaded into 1.0 g Si-MCM-41 is 1.0 ml, i.e., 50 wt % PEI loading. For the adsorbent with PEI loading of 75 wt %, theoretically, only one third of the PEI could be loaded into the Si-MCM-41 channels and two thirds of the PEI would be coated on the external surface of the molecular sieve particles. Interestingly, the highest adsorption gain (weighed on PEI) was also at the 50 wt % PEI loading, which indicated that only when the PEI was loaded into the mesoporous molecular sieve did the Si-MCM-41 and PEI show a strong synergetic effect on CO₂ adsorption. When the PEI was coated on the external surface of the molecular sieve crystals, the synergetic effect was limited. In order to distinguish the synergetic effect of the PEI in the mesoporous channels and on the

external surface of the molecular sieve crystals, the adsorption capacity of the PEI only in the “molecular basket” adsorbent was calculated as follows:

$$\text{PEI adsorption capacity (mg adsorbate/g-PEI)} = [\text{Adsorption capacity of the adsorbent} - (\text{MCM-41 weight percentage in the adsorbent} \times \text{Adsorption capacity of pure MCM-41})] / (\text{PEI weight percentage in the adsorbent})$$

The calculated PEI adsorption capacity is 215 mg/g-PEI for Si-MCM-41-PEI-50 and 174 mg/g-PEI for Si-MCM-41-PEI-75. Since the Si-MCM-41-PEI-75 was prepared with 25 wt % of Si-MCM-41 and 75 wt % of PEI, it was assumed that 25 wt % of the PEI is loaded into the pores of the Si-MCM-41 and the other 50 wt % of the PEI is coated on the external surface of the Si-MCM-41. If it is assumed that the PEI loaded into the Si-MCM-41 channel shows the same adsorption capacity as that of the Si-MCM-41-PEI-50, the adsorption capacity of the PEI that was coated on the external surface of Si-MCM-41 for Si-MCM-41-PEI-75 can be calculated and is found to be only 154 mg/g-PEI. The adsorption capacity for the PEI coated on the external surface of the crystal was much lower than that of the Si-MCM-41-PEI-50 (215 mg/g-PEI) when the PEI was fully filled in the channels of the Si-MCM-41, which confirmed that only when the PEI was loaded into the channels of the mesoporous molecular sieve did the Si-MCM-41 exhibit the highest synergetic effect and proved that the “molecular basket” concept resulted in the significant improvement on the CO₂ adsorption by PEI.

The adsorption capacity for the PEI coated on the external surface of the crystal was slightly greater than that of the pure PEI of 110 mg/g-PEI. When the PEI was coated on the crystal surface, more adsorption sites will be exposed to the adsorbate. Therefore, the adsorption capacity will be higher than that of the bulk PEI. This can also be verified by the CO₂ adsorption capacity of PEI coated on other high-surface-area materials such as silica gel. When the PEI was coated on a high-surface-area silica gel (550 m²/g) and the PEI loading was 50%, the CO₂ adsorption capacity was only 156 mg/g-PEI, slightly higher than that of the bulk PEI and much lower than that of the “molecular basket” adsorbent with the same PEI loading, which also verified that only when the PEI was loaded into the channels of the mesoporous molecular sieve did the Si-MCM-41 exhibit the highest synergetic effect on the adsorption of CO₂. The CO₂ adsorption capacity of Silica Gel-PEI was nearly identical to that of the calculated CO₂ adsorption capacity on the external surface of the MCM-41 crystals. Satyapal et al. (2001) also

observed the same phenomenon when they coated the PEI on the high-surface-area solid polymethyl methacrylate polymeric support.

By employing the “molecular basket” concept, the CO₂ adsorption and desorption kinetics can also be improved significantly. Figure 3.4.15 compares the CO₂ adsorption curves of Si-MCM-41-PEI-50, Si-MCM-41-PEI-75 and PEI in the first minute (60 seconds) of adsorption. The CO₂ adsorption rate decreased in the following order: Si-MCM-41-PEI-50 > Si-MCM-41-PEI-75 > PEI. If the linear part of the adsorption curve was selected to calculate the CO₂ adsorption rate, the CO₂ adsorption rate was 5.60, 3.23, and 2.67 mg/s.g-PEI for Si-MCM-41-PEI-50, Si-MCM-41-PEI-75 and PEI, respectively. The CO₂ adsorption rate for Si-MCM-41-PEI-50 was much faster than that of the Si-MCM-41-PEI-75 and PEI, which confirmed that only when the PEI was loaded into the channels of the mesoporous molecular sieve did the Si-MCM-41 exhibit the highest synergetic effect and proved that the “molecular basket” concept resulted in the significant improvement in CO₂ adsorption by PEI. Coating PEI on the high surface area materials only slightly increased the CO₂ adsorption rate. The CO₂ desorption rate showed the same trend as the adsorption. The CO₂ desorption rate decrease in the following order: Si-MCM-41-PEI-50 > Si-MCM-41-PEI-75 > PEI. The CO₂ desorption was complete for Si-MCM-41-PEI-50 and Si-MCM-41-PEI-75 in 150 minutes, whereas only 56% of the adsorbed CO₂ was desorbed when the same desorption time as the Si-MCM-41-PEI adsorbent was used.

Figure 3.4.15 COMPARISON OF THE CO₂ ADSORPTION KINETICS OF Si-MCM-41-PEI-50, Si-MCM-41-PEI-75, AND PEI

When compared with conventional adsorbents, the “molecular basket” showed better CO₂ adsorption performance at relatively high temperature. Table 3.4.2 lists the CO₂ adsorption performance of zeolite, activated carbon, PEI/polymer composite and the “molecular basket” adsorbent. The “molecular basket” adsorbent showed superior CO₂ selectivity in terms of very high CO₂/N₂ ratios (>1,000 for MCM-41-PEI-50 vs 2-6 for aluminosilicates and activated carbons). The high selectivity for CO₂ is particularly important when it was applied in the separation of CO₂ from flue gas.

3.4.2.2 Adsorption Separation of CO₂ from Simulated Flue Gas

Adsorption Separation of CO₂ from Simulated Flue Gas

Figure 3.4.16 shows the breakthrough curve of CO₂ during the adsorption separation of the CO₂, N₂ and O₂ gas mixture by MCM-41-PEI-50. At the beginning of the separation, CO₂ was completely adsorbed by the adsorbent and the concentration of CO₂ was below the detection limit of the gas chromatography, i.e. < 100 ppm. After 50 minutes of adsorption, CO₂ began to break through and was detected in the effluent gas. The amount of break through concentration, C , was followed as a fraction of the CO₂ concentration in the feed gas mixture, C_0 . At ≈ 55 minutes, the C/C_0 ratio was about 0.8, showing that the adsorbent still adsorbed 20% of the CO₂ in the feed. After 120 minutes of adsorption, the adsorbent can only adsorb 5% of the CO₂ from the feed.

In order to calculate the adsorption capacity, the volume of the adsorbed gases was calculated from the mass balance of the gas flow rate, the CO₂ concentration and the adsorption time. Figure 3.4.17 shows the normalized volume V_{ad}/V_0 (where V_{ad} equals to the volume of gases adsorbed and V_0 equals to the volume of gases feed into the system) of the adsorbed CO₂, N₂ and O₂ as a function of adsorption time. The CO₂ adsorption capacity was 45.4 ml (STP)/g adsorbent after 120 minutes. The adsorption capacity is comparable with that previously measured from the TGA as shown in Figure 3.4.18 at about 15% CO₂ concentration. If only the complete adsorption is considered, the adsorption capacity is 37.5 ml (STP)/g adsorbent. The adsorption of O₂ is much lower than that of the CO₂ and the adsorption capacity of O₂ is 0.07 ml (STP)/g adsorbent after 120 minutes. The separation factor for CO₂ and O₂ was calculated to be 185. The adsorbent hardly adsorbs any N₂. Therefore, the selective adsorption of CO₂ is

Table 3.4.2 Comparison of CO₂ Adsorption Performance of “Molecular Basket” Adsorbent and other Adsorbents

Adsorbents	Temp (°C)	Pressure (atm)	Adsorption Capacity (mg CO ₂ /g adsorbent)	CO ₂ /N ₂ or CO ₂ /CH ₄ selectivity	Reference
Si-MCM-41	25	1	27.3	--	this study
Si-MCM-41	75	1	8.6	--	this study
Si-MCM-41	75	0.149	6.3	2.9 (CO ₂ /N ₂)	this study
Al-MCM-41-100	75	1	7.6	--	this study
Al-MCM-41-500	75	1	7.5	--	this study
Si-MCM-41-PEI-50	25	1	32.9	--	this study
Si-MCM-41-PEI-50	75	1	112	--	this study
Si-MCM-41-PEI-50	75	0.149	89.2	>1000 (CO ₂ /N ₂)	this study
Al-MCM-41-100-PEI-50	75	1	127	--	this study
Al-MCM-41-500-PEI-50	75	1	121	--	this study
Zeolite 13 X	25	1	168	--	Siriwardane et al., 2001
Zeolite 4A	25	1	135	--	Siriwardane et al., 2001
Activated Carbon	25	1	110	--	Siriwardane et al., 2001
Norit RBI activated carbon	21.5	1	108	~ 2 (CO ₂ /CH ₄)	Van der Vaart, et al., 2000
Norit RBI activated carbon	75	1	40	~ 2 (CO ₂ /CH ₄)	Van der Vaart, et al., 2000
Activated carbon	20	1	88	~ 2 (CO ₂ /CH ₄)	Berlier aand Frere, 1997
Norit RBI activated carbon	25	1	140.8	~ 1.9 (CO ₂ /CH ₄)	Dreisbach, et al, 1999
PEI-silica gel	75	1	78.1	--	this study
PEI-polymer	~ 50	0.02	~ 40	--	Kohl and Nielsen, 1997

Figure 3.4.16 BREAKTHROUGH CURVE OF CO₂ WITH MCM-41-PEI-50 (Operating conditions: weight of adsorbent: 2.0 g; feed composition: 14.9% CO₂, 4.25% O₂ and 80.85% N₂; feed flow rate: 10 ml/min; temperature: 75°C)

Figure 3.4.17 NORMALIZED ADSORPTION VOLUME OF CO₂, O₂ AND N₂ (V_{ad}/V_0) AS A FUNCTION OF ADSORPTION TIME DURING THE ADSORPTION SEPARATION OF CO₂ FROM SIMULATED FLUE GAS MIXTURE BY “MOLECULAR BASKET” ADSORBENT (V_{ad} : volume of CO₂, O₂ or N₂ adsorbed; V_0 : volume of CO₂, O₂ or N₂ feed into the system)

Figure 3.4.18 THE ADSORPTION CAPACITY OF MCM-41-PEI-50 AS A FUNCTION OF FEED CO₂ CONCENTRATION MEASURED BY BOTH TGA AND THE FLOW ADSORPTION SYSTEM

realized by loading the PEI into the MCM-41 channels. For molecular sieves or zeolites from other studies, the separation factor of CO₂ over other gases was only around 2.

The Influence of Operating Conditions on the Separation of CO₂ from Simulated Flue Gas Mixtures

The Influence of Temperature on the Adsorption Separation of CO₂

The influence of operating temperature on the adsorption separation of CO₂, N₂ and O₂ gas mixture was investigated. Figure 3.4.19 shows the breakthrough curve of CO₂ at different temperatures. The higher the operating temperature, the more CO₂ was adsorbed. The adsorption capacities were 19.3, 28.8, 45.4, and 47.4 ml (STP)/g adsorbent for 25, 50, 75, and 100°C, respectively. The desorption of CO₂ at low temperature was very slow and not complete, which is in accordance with the results by TGA. The desorption capacity was only ≈10 and 15% at 25 and 50°C, respectively. The desorption was complete at 75°C.

Figure 3.3.19 THE INFLUENCE OF OPERATING TEMPERATURE ON THE BREAKTHROUGH CURVE OF CO₂ (Operating conditions: weight of adsorbent: 2.0 g; feed composition: 14.9% CO₂, 4.25% O₂ and 80.85% N₂; feed flow rate: 10 ml/min)

The Influence of Feed Flow Rate on the Adsorption Separation of CO₂

The influence of the feed flow rate on the adsorption capacity was investigated and the results are shown in Figure 3.4.20. The total adsorption capacities at 120 minutes were nearly identical among the three flow rates investigated. However, the adsorption capacity for the zero carbon dioxide emissions was different. The lower the feed flow rate, the higher the adsorption capacity for the zero CO₂ emissions. The adsorption capacities for the zero CO₂ emissions were 75, 74, and 67 ml (STP)/g adsorbent for the feed flow rate of 10, 20, and 30 ml/min, respectively.

The Influence of CO₂ Concentration in the Feed on the Adsorption Separation of CO₂

Since the CO₂ concentration in flue gas varies depending on the combustion processes and conditions used, the influence of CO₂ concentration in the feed on the adsorption performance was investigated. The CO₂ breakthrough curves are shown in Figure 3.4.21. The lower the CO₂ concentration was in the gas mixture, the longer the time it took to achieve zero CO₂ emissions. However, the total adsorption capacities increased with increasing CO₂ concentration and were 33.3, 35.5, 42.5, and 45.5 ml (STP)/g adsorbent for carbon dioxide concentrations of 0.652, 1.94, 8.09, and 14.9%, respectively. The adsorption capacities are

Figure 3.4.20 THE INFLUENCE OF FEED FLOWRATE ON THE BREAKTHROUGH CURVE OF CO₂ (Operating conditions: weight of adsorbent: 2.0 g; feed composition: 14.9% CO₂, 4.25% O₂ and 80.85% N₂; temperature: 75°C)

Figure 3.4.21 THE INFLUENCE OF FEED CO₂ CONCENTRATION ON THE BREAKTHROUGH CURVE OF CO₂ (Operating conditions: weight of adsorbent: 2.0 g; temperature: 75°C; feed flow rate: 30 ml/min)

compared with those previously obtained from TGA in Figure 3.4.18. At low CO₂ concentrations, the adsorption capacities from the flow system were somehow higher than those from the TGA measurements. The higher adsorption capacities for the flow system may be caused by the longer adsorption time of 20 h used for the flow system as compared to 2.5 h for the TGA. When the adsorption time for the flow system was shortened from 20 h to 6 h for the feed with a CO₂ concentration of 0.652%, the adsorption capacity decreased from 33.3 to 28.9 ml (STP)/g. The adsorption capacity was close to that of 25.5 ml (STP)/g adsorbent obtained by the TGA measurement. With increasing CO₂ concentration in the feed, the adsorption capacity in the flow system is lower than that from the TGA measurement. The slightly reduced adsorption capacity may be caused by the inhomogeneous temperature distribution along the adsorbent bed. The temperature in this experiment was only measured in the center part of the adsorbent bed. At the beginning or end of the adsorbent bed, the temperature could be 8°C lower than at center bed, which will result in a lower adsorption capacity.

The Influence of Moisture on the Adsorption Separation of CO₂

Since the flue gas contains about 8-20% moisture, the influence of moisture on the adsorption of CO₂ was investigated. Figure 3.4.22 compares the breakthrough curve of CO₂ for the feed gas without moisture and with 10% moisture. The adsorption of CO₂ increases when moisture is added to the feed. The adsorption capacity was increased from 47.4 ml (STP)/g adsorbent for dry feed to 63.7 ml (STP)/g adsorbent for moist feed. At the same time, the water was also adsorbed into the adsorbent. The adsorption capacity of water was 58.5 ml (STP)/g adsorbent.

The increase in the CO₂ capacity can be ascribed to reactions (3.4.1) and (3.4.2) for the dry condition, and to reactions (3.4.3), (3.4.4), and (3.4.5) for the moist condition:

Figure 3.4.22 COMPARISON OF THE CO₂ BREAKTHROUGH CURVE IN THE ABSENCE AND IN THE PRESENCE OF 10% MOISTURE (Operating conditions: weight of adsorbent: 2.0 g; temperature: 75°C; feed flow rate: 10 ml/min)

Reactions (3.4.1) and (3.4.2) show that 2 moles of amine are needed under dry conditions to adsorb one mole of carbon dioxide. When moisture was added to the feed, the carbamate ion formed in reactions (3.4.1) and (3.4.2) will further react with CO₂ and H₂O to form bicarbonate (reaction (3.4.3)). The amine itself can also react directly with CO₂ and H₂O (reaction (3.4.4) and (3.4.5)).

Therefore, in the presence of water, one mole of amine can adsorb one mole of carbon dioxide.

The effect of moisture on the enhancement of the adsorption capacity can further be verified by changing the moisture concentration in the feed. Table 3.4.3 shows the relationship between the adsorption capacity and the moisture concentration in the feed. At low moisture concentration, the adsorption capacity increased rapidly with increasing moisture concentration

in the feed. However, when the moisture concentration reached 16%, the increase in the adsorption capacity slowed. It is also noteworthy that the adsorption capacities increased even though the CO₂ concentration in the feed decreased. This is opposite of that observed for the dry condition. According to reactions (3.4.4) and (3.4.5), the formation of bicarbonate needs one mole of H₂O per mole of CO₂. When the moisture concentration is lower than that of the CO₂, the adsorption capacity of CO₂ increases rapidly with increase in moisture concentration. When the moisture concentration is larger than that of the CO₂, the CO₂ adsorption capacity can not be further increased by the excess water. The above phenomena indicated strongly that the presence of water has a synergetic effect on the CO₂ adsorption capacity.

Table 3.4.3 The Influence of Moisture Concentration on the Adsorption Separation of Carbon Dioxide from the Simulated Flue Gas

Feed concentration (%)				CO ₂ adsorption capacity (ml (STP)/g adsorbent)
CO ₂	N ₂	O ₂	H ₂ O	
14.90	80.85	4.25	---	45.5
13.91	75.71	3.95	6.43	54.0
13.55	72.72	3.86	9.87	63.7
12.61	68.25	3.56	15.59	66.8

The Influence of Sweep Gas Flow Rate on Desorption of CO₂

Three helium sweep gas flow rates of 10, 20, and 50 ml/min were selected to clarify the effect of the sweep gas flow rate on the desorption of carbon dioxide after an adsorption experiment. The results are shown in Figure 3.4.23. With increasing sweep gas flow rate, the desorption rate of the carbon dioxide increases. Desorption at 300 minutes was not complete when the sweep gas flow rates were 10 and 20 ml/min and the desorption capacities were 48 and 76%, respectively. When the sweep gas flow rate was increased to 50 ml/min, desorption was complete after 300 minutes. This indicated that the regeneration of the adsorbent can easily be optimized.

Stability of the “Molecular Basket” Adsorbent

Thermal Stability of the “Molecular Basket” Adsorbent

The stability of the adsorbent was investigated at 75 and 100°C. The successive adsorption and desorption cycles were carried out at 75 and 100°C, and the results are shown in

Figure 3.4.23 THE INFLUENCE OF SWEEP GAS FLOW RATE ON THE DESORPTION OF CO₂ (Operating conditions: weight of adsorbent: 2.0 g; temperature: 75°C; feed flow rate: 10 ml/min)

Figures 3.4.24 and 3.4.25, respectively. At 75°C, the adsorption capacity did not change between the successive measurements, indicating that the adsorption performance was stable. However, when the operating temperature was increased to 100°C, there was a decrease in the adsorption capacity in the second measurement. The adsorption capacity in the second measurement was 43.6 ml (STP)/g adsorbent, and was about 8% lower than that of the first measurement of 47.4 ml (STP)/g adsorbent. The deactivation of the adsorbent is caused by oxidation of the polymer at this relatively high temperature. Figure 3.4.26 shows that there is a significant increase in the O₂ consumption at 100°C, when compared to that at 75°C. The consumption of O₂ increases from 0.12 ml (STP)/g adsorbent at 75°C to 5.1 ml (STP)/g adsorbent at 100°C. Further, the amount of O₂ desorbed after the adsorption at 100°C is only 0.06 ml (STP)/g adsorbent, which represents only 1% of the consumed oxygen. In addition, the color of the adsorbent turned from white in its original form to pale yellow after the adsorption measurement at 100°C. Therefore, the increased consumption of O₂ with increasing temperature has been linked to the oxidation of the polymer, indicating that the adsorbent is not suitable for use at temperatures higher than 100°C.

Figure 3.4.24 STABILITY OF THE ADSORBENT AT 75°C. (Operating conditions: weight of adsorbent: 2.0 g; temperature: 75°C; feed flow rate: 10 ml/min)

Figure 3.4.25 STABILITY OF THE ADSORBENT AT 100°C. (Operating condition: weight of adsorbent: 2.0 g; temperature: 100°C; feed flow rate: 10 ml/min)

Figure 3.4.26 COMPARISON OF THE OXYGEN “ADSORPTION” AT DIFFERENT TEMPERATURES INDICATING THE DEGREE OF OXIDATION OF THE POLYMER (Operating conditions: weight of adsorbent: 2.0 g; feed flow rate: 10 ml/min)

Cyclic Adsorption/Desorption Stability

Cyclic Adsorption/Desorption Stability Under Dry Flue Gas Conditions. For practical applications, the adsorbent should not only possess high selectivity and high adsorption capacity, but must also exhibit stable adsorption and desorption performance for thousands of cycles. In this study, cyclical adsorption and desorption were carried out, and the results are shown in Figure 3.4.27. During ten cycles of operation, the CO₂ adsorption capacity was relatively unchanged, indicating the adsorbent is stable in the separation process. Desorption was also complete. The adsorption selectivity also did not change in the ten cycles of operation. The stable adsorption and desorption performance suggests that this adsorbent is promising for practical applications.

Stable adsorption and desorption performance is a key issue for a adsorbent using a chemical adsorption mechanism. Bakker et al. (1973) observed that CO₂ adsorption capacity decreased 40% for a CaO adsorbent after five cycles of operation at 866°C. The author suggested that the decrease in adsorption capacity was due to the formation of a nonporous CaCO₃ layer on the particle surface. This layer hindered the inward diffusion of CO₂. Cyclical experiments conducted at a carbonation temperature of 880°C and a calcination temperature of

Figure 3.4.27 STABILITY OF THE “MOLECULAR BASKET” ADSORBENT UNDER DRY FLUE GAS CONDITIONS (Operating conditions: weight of adsorbent: 2.0 g; feed composition: 14.9% CO₂, 4.25% O₂ and 80.85%; temperature: 75°C; Feed flow rate: 10 ml/min)

860°C led to a drop in conversion of CaO to CaCO₃ from 70% in the first carbonation to 38% in the 7th carbonation step. The “molecular basket” adsorbent was stable after ten cycles of operation. Since the active species of PEI was uniformly dispersed into the mesoporous molecular sieve of MCM-41 with a uniform pore size of around 3 nm, the diffusion of CO₂ to the active site was not hindered. Using the same reasoning, the adsorbent was easily regenerated. However, when bulk PEI was used as the adsorbent, a diffusion hindrance was observed (Xu et al., 2002b). The bulk PEI showed not only a low adsorption capacity, but also an incomplete desorption. The “molecular basket” adsorbent was also stable at relatively high temperature, e.g., 75°C. In addition, the structure of MCM-41 was well preserved after ten cycles of separation as evidenced in Figure 3.4.28 for the XRD characterization results before and after the adsorption separation. The stable structure of MCM-41 is of critical importance to the “molecular basket” effect of the adsorbent (Xu, et al., 2002a,b; Xu et al., 2003).

Cyclic Adsorption/Desorption Stability under Moist Flue Gas Conditions. In practical situations, water vapor is often contained in the gas mixture, e.g., ≈10% water vapor in flue gas. Adsorption separation under such high moisture conditions is not well characterized. In addition, zeolite and molecular sieves are not stable in moist atmospheres at high temperature.

Figure 3.4.28 XRD PATTERNS OF THE “MOLECULAR BASKET” ADSORBENT BEFORE AND AFTER CYCLIC OPERATION UNDER DRY FLUE GAS CONDITIONS AT 75°C

In this study, cyclical adsorption and desorption were carried out under moist conditions, and the results are shown in Figure 3.4.29. During ten cycles of operation, the CO₂ adsorption capacity remained relatively unchanged, indicating the adsorbent was stable in the separation process. The CO₂ adsorption capacity varied between 69.0 to 72.4 ml (STP)/g adsorbent. Note that the CO₂ and H₂O concentration also varied slightly in the feed during the ten cycles of operation. Desorption was also complete. The adsorption selectivity also did not change in the ten cycles of operation.

The structure of MCM-41 and MCM-41-PEI before and after adsorption separation was characterized by XRD. Figure 3.4.30 shows the XRD patterns of MCM-41 and Figure 3.4.31 shows the XRD patterns of the “molecular basket” adsorbent. The structure of MCM-41 collapsed only after 1 cycle operation under the 10% moisture condition. However, the structure of MCM-41 was preserved even after 10 cycles of operation, which indicated that loading PEI into the channels of MCM-41 protected the structure of MCM-41. It is well known that MCM-

Figure 3.4.29 STABILITY OF THE “MOLECULAR BASKET” ADSORBENT UNDER MOIST FLUE GAS CONDITIONS (Operating conditions: weight of adsorbent: 2.0 g; feed composition: 13.4 % CO₂, 3.8% O₂, 10.4% H₂O and 72.4 %; temperature: 75°C; feed flow rate: 10 ml/min)

Figure 3.4.30 XRD PATTERNS OF MCM-41 BEFORE AND AFTER OPERATING UNDER MOIST SIMULATED FLUE GAS CONDITIONS AT 75°C

Figure 3.4.31 XRD PATTERNS OF THE “MOLECULAR BASKET” ADSORBENT BEFORE AND AFTER OPERATING UNDER MOIST SIMULATED FLUE GAS CONDITIONS AT 75°C.

41 is unstable in the presence of moisture, even at medium temperatures. The collapse of the MCM-41 structure was caused by the reaction between MCM-41 and water. Since the PEI is more hydrophilic than MCM-41, the adsorption of water by PEI is stronger than for the MCM-41. Therefore, the structure of MCM-41 was protected after the PEI was loaded into the MCM-41. Preservation of the MCM-41 structure is critically important for the adsorption separation performance of this novel “molecular basket” adsorbent.

3.4.2.3 Adsorption Separation of Boiler Flue Gas

Adsorption Separation of Natural Gas-Fired Flue Gas

Typical CO₂ breakthrough curves for the ‘molecular basket’ adsorbent and the blank “alumina” are shown in Figure 3.4.32 for the natural gas-fired flue gas. Clearly, both the alumina and the “molecular basket” adsorbent adsorbed CO₂. However, the adsorption performance of the “molecular basket” adsorbent was much better than that of the alumina. The lowest CO₂ emission concentration was $\approx 0.8\%$ for alumina, while it was less than 0.1% for the “molecular basket” adsorbent. The CO₂ adsorption capacity was 26 ml (STP)/g adsorbent for the “molecular basket” adsorbent and 1.1 ml (STP)/g adsorbent for the alumina. In addition, the “molecular basket” adsorbent showed better selectivity. The “molecular basket” adsorbent did not adsorb O₂, N₂ and CO while the CO₂/O₂ selectivity was 3.5 for alumina. The adsorption

Figure 3.4.32 BREAKTHROUGH CURVE OF CO₂ FOR FLUE GAS FROM NATURAL GAS-FIRED BOILER

capacity of NO_x for the “molecular basket” adsorbent was higher than that for alumina, as shown in Figure 3.4.33. The NO_x adsorption capacity was 0.0505 ml (STP)/g adsorbent for the “molecular basket” adsorbent and 7.5×10^{-4} ml (STP)/g adsorbent for the alumina. The selectivity of CO₂/NO_x was 0.45 for the “molecular basket” adsorbent. However, very little NO_x was absorbed before CO₂ breakthrough. The CO₂ and NO_x adsorption capacities before CO₂ breakthrough were 21 ml (STP)/g adsorbent and 7.1×10^{-3} ml (STP)/g adsorbent, respectively. Therefore, the selectivity of CO₂/NO_x was 2.5. Further, while the desorption of CO₂ was complete, very little NO_x desorbed.

Adsorption Separation of Coal-Fired Flue Gas

The adsorption separation of CO₂ from coal-fired flue gas was also investigated. In the coal fired flue gas, there was a significant concentration of SO₂. The adsorption separation results are shown in Figures 3.4.34 to 3.4.36. Similar trends were observed for the coal-fired flue gas as for the natural gas-fired flue gas. The adsorptions of N₂ and O₂ by the “molecular basket” adsorbent were below the detect limit of the apparatus. The adsorption capacities for CO₂, SO₂ and NO_x were 36, 0.11, and 0.21 ml (STP)/g adsorbent, respectively. The separation selectivities for CO₂/SO₂ and CO₂/NO_x were 1.07 and 0.57, respectively. The adsorption capacities for CO₂, SO₂ and NO_x before CO₂ breakthrough were 24, 0.0074, and 0.028 ml (STP)/g adsorbent, respectively. Therefore, the separation selectivities for CO₂/SO₂ and

Figure 3.4.33 NO_x CONCENTRATION CHANGE DURING THE SEPARATION OF FLUE GAS FROM A NATURAL GAS-FIRED BOILER

Figure 3.4.34 BREAKTHROUGH CURVE OF CO₂ FOR FLUE GAS FROM A COAL-FIRED BOILER

Figure 3.4.35 SO_2 CONCENTRATION CHANGE DURING THE SEPARATION OF A COAL-FIRED FLUE GAS MIXTURE

Figure 3.4.36 NO_x CONCENTRATION CHANGE DURING THE SEPARATION OF A COAL-FIRED FLUE GAS MIXTURE

CO₂/NO_x were 10.7 and 2.86, respectively, before CO₂ breakthrough. Again, while desorption of CO₂ was complete, very little NO_x and SO₂ desorbed.

Compared with the adsorption capacity in the simulated flue gas, the adsorption capacity in the boiler flue gas was about 30% lower at similar feed compositions. The adsorption capacity for the coal-fired flue gas was 36 ml (STP)/g adsorbent at 80±10°C for the feed composition of 12.5-12.8% CO₂, ≈4.4% O₂, 50 ppm CO, 420 ppm NO_x, 420 ppm SO₂, 6.2% H₂O and 76-77% N₂. The adsorption capacity for the simulated flue gas mixture was 53 ml (STP)/g adsorbent at 75°C for the feed composition of 13.9% CO₂, 3.95% O₂, 6.43% H₂O and 75.7% N₂. There are several explanations for the difference in adsorption capacity. First, the CO₂ concentration in the coal-fired flue gas mixture was lower than that in the simulated flue gas mixture. At low CO₂ concentration, the CO₂ adsorption capacity will be low. Second, different physical forms of the adsorbent will have different CO₂ adsorption capacities. A comparison study on the CO₂ adsorption of a powder adsorbent and a pellet adsorbent was carried out and the results are shown in Figure 3.4.37. It is evident in Figure 3.4.37 that the pellet adsorbent had

Figure 3.4.37 COMPARISON OF THE CO₂ BREAKTHROUGH CURVE OF THE “MOLECULAR BASKET” ADSORBENT IN PELLET AND IN POWDER FORM (Operating conditions: weight of adsorbent: 2.0 g; feed composition: 14.9% CO₂, 4.25% O₂ and 80.85%; temperature: 75°C; feed flow rate: 10 ml/min)

a lower adsorption capacity than the powder adsorbent. The adsorption capacity of the pellet adsorbent was 42 ml (STP)/g adsorbent and was 10% lower than that of the powder adsorbent. Third, the higher flow rate in the boiler flue gas separation contributed to the lower adsorption capacity. The feed flow rate was 167 ml/min/g adsorbent in the boiler flue gas separation and was 5 ml/min/g adsorbent in the simulated flue gas separation. Lastly, the minor gas components, i.e., NO_x and SO_2 , influenced the adsorption of CO_2 . Since NO_x and SO_2 are acid gases and the adsorbent is basic in nature, the NO_x and SO_2 are competitive adsorbates for CO_2 . These experimental results showed that adsorption of NO_x by the “molecular basket” adsorbent was even stronger than that of CO_2 under some experimental conditions. The pre-removal of NO_x and SO_2 from the flue gas mixture is therefore required for the adsorption separation of CO_2 from the flue gas mixture by this “molecular basket” adsorbent.

3.4.3 Conclusions

Novel CO_2 “molecular basket” adsorbents based on a polyethylenimine (PEI)-modified mesoporous molecular sieve of MCM-41 type (MCM-41-PEI) were successfully developed. The highest CO_2 adsorption capacity obtained was 246 mg/g-PEI was obtained, which is 30 times higher than that of the MCM-41 and more than twice that of the pure PEI. The experimental results showed that, although the dispersion of PEI on high surface area materials could increase the adsorption capacity, it was the synergetic effect between PEI and MCM-41 pore channels (CO_2 “molecular basket”) that contributed to major increases in the CO_2 adsorption capacity. The uniform dispersion of the PEI into the pore channels of the MCM-41 support is critical for the preparation of this novel “molecular basket” adsorbent. Proper incorporation of Al into the MCM-41 framework in the synthesis step can significantly enhance the performance of the final MCM-41-PEI adsorbents. The relatively lower Si/Al ratio of the MCM-41 support and a relatively high methanol/MCM-41 weight ratio in adsorbent preparation can lead to higher CO_2 adsorption capacity for MCM-41-PEI. Adding polyethyleneglycol (PEG) into the MCM-41-PEI adsorbent can increase not only the CO_2 adsorption and desorption rates, but also the CO_2 adsorption capacity.

Carbon dioxide was successfully separated from a simulated flue gas mixture containing CO_2 , O_2 , N_2 and moisture by using the novel “molecular basket” adsorbent. The novel adsorbent showed a CO_2 adsorption capacity of 45.5 ml (STP)/g adsorbent and separation factors of above

1,000 and 182 for CO₂/N₂ and CO₂/O₂, respectively, at a feed composition of 14.9% CO₂, 4.25% O₂ and 80.85% N₂. The adsorption/desorption mechanisms were favored at high temperature. However, the PEI polymer oxidized when the temperature was higher than 100°C. The best separation temperature was 75°C. The presence of moisture in the simulated flue gas mixture promoted the adsorption of CO₂. The adsorption capacity in the presence of 10% moisture was 63.7 ml (STP)/g adsorbent, and was 1.34 times higher than that for a dry gas mixture, showing a clear synergetic effect between the PEI and moisture on CO₂ adsorption. Cyclic adsorption/desorption experiments showed that the “molecular basket” adsorbent was stable under both dry and moist feed conditions.

Carbon dioxide can be selectively separated from boiler flue gases by using the novel “molecular basket” adsorbent. The adsorbent adsorbs little N₂ and O₂. The selectivity of CO₂/NO_x was 2.5 for natural gas-fired flue gas and the separation selectivities for CO₂/SO₂ and CO₂/NO_x were 10.7 and 2.86, respectively, for coal-fired flue gas. The desorption of CO₂ was complete. However, very little NO_x and SO₂ desorbed after adsorption indicating the need for pre-removal of NO_x and SO₂ from the flue gas mixture before capture of CO₂ by the PEI based “molecular basket” adsorbent.

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4.0 ECONOMIC EVALUATION

The objectives of this activity were to determine cost and market penetration, selection of incentives, and regional economic impacts of coal-based fuel technologies. In addition, DOD's fuel mix was determined and a national energy portfolio constructed that minimizes energy price shock effects. Each of these activities is discussed in detail in the following sections.

4.1 Cost and Market Penetration of Coal-Based Fuel Technologies

4.1.1 Introduction

The United States is awash in coal and a plethora of technologies have been advanced to promote coal utilization as a means to avoid a dependence upon a foreign-controlled fuel source. The main problem is that, in many applications, coal-based fuels are simply not as easy and economical to use as oil-based fuels. The result is that the potential of these technologies has remained only a promise. Beginning in the 1970s, the prospect of greatly increased coal utilization seemed to be close to realization with expectations of oil prices rising to the margin of coal-based fuel technology viability. When oil prices did not achieve or maintain these levels, many technology development efforts refocused their attention on attaining economic viability through lowering costs. This task focuses upon those technologies that utilize coal-water mixtures (CWMs), with an emphasis upon micronized coal-water mixtures (MCWMs).

A motivation for developing boiler retrofit technologies stems from U.S. policy makers' desire to decrease dependence on foreign sources of energy. Dependence upon imported energy sources makes the U.S. vulnerable to an "oil price shock," such as that experienced from 1973 to 1974 and later during 1979 to 1980. Another motivation is to improve the international competitive position of the U.S. through developing technologies that utilize the country's coal resources at a cost lower than current domestic and imported energy supply sources.

A market penetration model was formulated that is based upon the optimal mix of a boiler retrofit technology adoption among a sample population of active boilers in the Pennsylvania counties of Cambria and Indiana. It was found that six of eighteen boilers are candidates for adopting the retrofit technology resulting in a total cost savings of \$1.89 million annually. It was also found that the study region could likely be expanded to other adjacent counties since the actual mileage-dependent portion of the MCWM costs is small compared to all of the other costs.

An overview of boiler retrofit technology, background on the motivation for retrofitting, a brief survey of other technologies developed for utilization of CWMs, and factors affecting the economic feasibility of the technology has been presented in the Phase II Final Report (Miller et al., 2000). Reviews of oil price trends and the effects of divergent cost estimates on CWM economics are also presented in the Phase II Final Report (Miller et al., 2000). This information was instrumental in implementing the market penetration model presented here.

Section 4.1.2 defines and discusses the market penetration model approach and structure utilized in the study. The section begins with a review of the range of market penetration models available to provide the context and justification for selection of the nonlinear programming framework. The section concludes with a description of the model and its major components such as fuel production costs, transportation costs, retrofit costs, potential production sites, and the inventory of boilers that are retrofit candidates. Results and conclusions of the boiler retrofit market penetration model are presented in Section 4.1.3. The base case model results are presented followed by a sensitivity analysis of alternative siting and less than optimal number of boilers being retrofitted. Finally, conclusions are made regarding the impact of boiler retrofits in terms of increased coal sales to the industrial sector in Pennsylvania with extensions by analogy to the other coal-producing and industrial states of Ohio, Indiana, and Illinois.

4.1.2 Market Penetration Model Approach and Structure

Market penetration models have been ascribed to a wide set of methods directed toward predicting the likely future adoption or emplacement of products, services, or technologies within an economy (Bodington and Quinn, 1982). The models that have been developed and used in market penetration studies are tailored to fulfilling the objective of forecasting market share over time or estimating the potential, or equilibrium, market share. Models that forecast market share over time often must be explicitly defined with an estimate of the equilibrium market share toward which adoption proceeds. The types of models include:

- Diffusion models;
- Time series models;
- Econometric models;
- Historical analogy; and
- Optimization models.

A detailed description of these models can be found in Schaal (1995). The diffusion, time series, econometric, and historical analogy models rely upon first obtaining some estimate of the maximum market penetration of the technology. This crucial estimate of the maximum market share, or equilibrium optimal mix, then provides a 'force' toward which adoption would proceed. The method used in this study to predict market penetration of the retrofit technology is an optimization model solvable by nonlinear programming methods. The results derived from this model are then subjected to a sensitivity analysis to assess how the technology adoption fares when acceptance starts, or ends, at a level lower than the optimal level. The specific optimization model used is described in the next section.

4.1.2.1 Optimization Model

The type of market penetration model formulated in this study was based upon a partial equilibrium structure toward which the market share of the new technology (boiler retrofits) would trend. In this model, it is assumed that producers minimize their input costs to maximize profits. Within this context, those producers who own boilers seek to minimize their energy expenditures through minimizing their cost of steam production. The boiler retrofit technology represents an opportunity to further reduce costs by displacing fuel oil or natural gas with lower cost MCWM through the expenditure of a capital investment to allow use of the alternate fuel. Producers are assumed to adopt the boiler retrofit technology if the capital and variable costs of retrofitting are less than the current variable costs that would be displaced.

An illustration of a few of the relevant factors that would enter into a single boiler owner's decision to forgo or accept adoption of the boiler retrofit technology are shown in Figure 4.1.1. In this figure, the line that includes the segment **OB** represents the variable costs, mostly of the fuel being displaced that would be avoided if the boiler were retrofitted. However, if the boiler were retrofitted, fixed costs (line segment **OA**) and variable costs (line that includes segment **AB**) would be incurred. The slopes of **OA** and **OB** are composed primarily of the \$/MM Btu price of MCWM and oil or gas respectively. Utilization is the ratio of actual boiler output, or load, to the maximum output that the boiler is capable of over a specified period of time, expressed as a percentage. In this analysis, utilization and costs are annualized.

Figure 4.1.1 OPPORTUNITY COSTS OF RETROFITTING VERSUS BOILER UTILIZATION

If the combined fixed cost displacement and slope of retrofitting opportunity cost lines result in an intersection with the non-retrofit opportunity cost line, such as occurs at point **B** in Figure 4.1.1, then it is possible that the boiler owner's costs would be minimized by adopting the retrofit technology. The utilization at point **C** defines the breakeven point at which the boiler owner would be indifferent to adopting the retrofit technology on the basis of cost alone. If the boiler utilization is greater than **C** then the annualized costs of retrofitting are less than not retrofitting and the technology is adopted. If boiler utilization is less than **C** then the technology is not adopted.

Figure 4.1.1 is useful for demonstrating the decision rule of whether or not to adopt the technology on the basis of the single parameter of boiler utilization but it does not take into account other factors that enter into the model. For example, fixed costs consist primarily of the annualized capital charges necessary to retrofit the boiler. The magnitude of these retrofit costs vary non-linearly with the size of the boiler being considered. The slope of the retrofit cost line reflects the \$/MM Btu per percent utilization for producing the MCWM and disposing of the ash products. The cost of MCWM is a variable since MCWM production is subject to economies of scale. Therefore, MCWM costs are dependent upon the total demands of all boiler owners in the region. This cost is further complicated by considering that multiple possible MCWM supply

points occur throughout the region that could potentially supply the boilers that are dispersed geographically throughout the region. Supply points would therefore be chosen such to minimize both the costs of MCWM production and transportation.

The model considered here is a partial equilibrium analysis since it is assumed that there are no substitution effects on oil or natural gas prices due to their displacement by MCWM. Another assumption is that there would be no increase or decrease of the boiler utilization due to displacing the utilization of other types of boilers that may be in use at a site. A producer may own several different types of boilers at a single site and choose to operate the ones that provides the greatest competitive advantage at a particular point in time.

It was assumed that the specific costs that boiler owners consider when deciding whether or not to adopt the retrofit technology are:

- Total capital requirement (TCR) to effect the boiler retrofit;
- MCWM cost, f.o.b. mine (preparation plant location);
- Transportation costs for delivery of MCWM to the boiler;
- Incremental Operation & Maintenance (O&M) costs due to retrofitting; and
- Ash disposal costs.

4.1.2.2 Market Penetration Model Formulation

The model may be succinctly expressed as an optimization problem where the objective to minimize costs is subject to constraints that describe the interrelationship among variables and parameters. The model thus defined in a standardized form may then be solved using optimization software, such as GAMS (General Algebraic Modeling System), which utilizes standardized solution algorithms to determine the optimal objective function value (maximum or minimum) and the variable levels necessary to achieve this optimal state. The programming language chosen to compile the nonlinear programming model was GAMS release 2.5 (Brooke, et al., 1992).

An important constraint in the model is that each boiler owner annually demands a specific quantity of fuel to carry out his production process. The costs introduced in the previous section are dependent upon whether MCWM is selected for a particular boiler, the flows of fuel from

each source to each boiler, and the aggregate demand for fuel at each source. The model is defined mathematically as:

$$\min z = \sum_{i=1}^m \sum_{j=1}^n FuelCost_{ij} x_{ij} + \sum_{i=1}^m \sum_{j=1}^n TCost_{ij} x_{ij} \quad (4.1.1)$$

Subject to:

$$\sum_{j=1}^n x_{ij} \leq FS_i \quad \text{for all } i \quad (4.1.2)$$

$$\sum_{i=1}^m x_{ij} = BQ_j \quad \text{for all } j \quad (4.1.3)$$

$$PMCWSF_i = PMCWSF_i \left(K, \sum_{j=1}^n x_{ij} \right) \quad \text{for all } i \text{ not oil or gas} \quad (4.1.4)$$

$$FuelCost_{ij} = PMCWSF_i + FuelFC_{ij} \quad \text{for all } i \text{ and } j \quad (4.1.5)$$

Where:

z	= total annual fuel costs for all boilers in the region (objective), \$
i	= potential fuel supply points
j	= watertube boilers
$FuelCost_{ij}$	= total fuel cost from i th source to j th boiler less transportation cost, \$/MM Btu
x_{ij}	= annual flow of fuel from i th source to j th boiler, MM Btu
$TCost_{ij}$	= transportation cost from i th source to j th boiler, \$/MM Btu
FS_i	= total supply of fuel annually available at the i th source, MM Btu
BQ_j	= total fuel demanded by the j th boiler, MM Btu
$PMCWM_i$	= price of fuel at the i th source, \$/MM Btu
K	= MCWM cost curve constant
$FuelFC_{ij}$	= the annual fixed costs to supply the j th boiler from the i th source, \$/MM Btu

The objective function to be minimized, that of fuel costs to boiler owners, is defined by equation 4.1.1. The objective function is split into two parts in order to highlight the transportation cost framework that drives the selection of supply sites due to their geographical location relative to the boilers in the region. The variable $FuelCost_{ij}$ includes both fixed and variable costs on a \$/MM Btu measure. This is accomplished by assuming that boiler demands for fuel are the same during the year regardless of any other considerations.

The constraints are described in equations 4.1.2 through 4.1.5. The fuel supply constraint is defined in equation 4.1.2 which specifies that the total demand of all the boilers for fuel from a particular fuel source may not exceed the maximum amount available from that supply source. In keeping with the partial equilibrium structure of the model the fuel currently being fired by the boilers is set to an arbitrarily high level. Equation 4.1.3 specifies that the annual aggregate flow of fuel from each supply source to a particular boiler must equal that boiler's annual demand for fuel. Equation 4.1.4 specifies that the costs of producing MCWM at a particular supply source and is a function of a constant and the total demand for fuel from that particular supply source. Equation 4.1.5 provides an accounting of fuel costs.

A listing of the GAMS program that implements the market penetration model in detail is given in Appendix 4A.

4.1.2.3 Fuel Supply Costs

The production of fuels at each supply point is subject to economies of scale. It is assumed that MCWM production can recover coal fines from existing waste streams and prepare it for use by the retrofitted boilers. An overview of the MCWM production process is shown in Figure 4.1.2.

The cost model basis for MCWM fuel supply using this production arrangement has been estimated by Schaal and Gordon (1994). The MCWM Supply Cost Model is presented in Appendix 4B. Using the results from this work, the cost curve shown in Figure 4.1.3 was constructed.

Figure 4.1.2 MCWM PRODUCTION AND UTILIZATION BLOCK FLOW DIAGRAM

Figure 4.1.3 MCWM COST CURVE FOR A SINGLE SUPPLY POINT

A power function regression was estimated for the cost curve for Figure 4.1.3 with the resulting equation 4.1.6. Equation 4.1.6 is defined only for those supply points that are not oil or gas. The cost of supply for oil and gas was taken as the average fuel cost paid by the Pennsylvania industrial sector (Miller et al., 1994) or the actual fuel cost reported by the producer, if available.

$$PMCWSF_i = 0.6727 + 211561 \left(\sum_{j=1}^n x_{ij} \right)^{-0.969} \quad \text{for all } i \text{ not oil or gas} \quad (4.1.6)$$

4.1.2.4 Transportation Costs

The transportation costs for MCWM were determined through consultation with a local hauler of mine wastes using Department of Transportation (DOT) approved tank trucks (Punalla, 1994). The cost model derived from this conversation is given in equation 4.1.7.

$$T = \$0.168 / MMBtu + \$0.00373 / MMBtu / mile \quad (4.1.7)$$

The fixed \$/MM Btu cost is related to the fixed amount of time both the driver and tanker truck are required to spend to complete unloading of the fuel at the site. This fixed cost is included with the other fixed costs for operating the boiler with MCWM. The variable costs are expressed as \$/MM Btu per mile. In the model, this variable cost is multiplied by the distance from the i th supply site to the j th boiler to obtain the transportation cost matrix.

4.1.2.5 Retrofit Technology Costs

The capital cost of the retrofit technology was estimated by EER for the Crane site's 25.2 MM Btu/h boiler (Miller et al., 2000). The standard method for scaling capital costs is to use the power factor method as shown in equation 4.1.8 (see Addy and Considine, 1994, for example). In this equation, C_2 is the desired cost at the new output level Q_2 , and C_1 is the known cost at output level Q_1 . The power factor, PF , reflects the economies of scale in construction which for boiler plants is usually taken as 0.75.

$$\frac{C_2}{C_1} = \left(\frac{Q_2}{Q_1} \right)^{PF} \quad (4.1.8)$$

Anecdotal evidence exists that a portion of the cost estimate for converting the 25.2 MM Btu/h Crane boiler is not subject to scaling for boilers as large as 100 MM Btu/h (Miller et al., 2000). Given this information, a cost model was initially postulated that assumed that the 25.2 MM Btu/h capital cost to retrofit is subject to a graduated increasing percentage of costs subject to scaling. This approach yielded a linear cost curve. Refinements to this approach were made by incorporating more detailed information concerning how the individual cost components scale with increasing boiler size. This analysis is presented in Appendix 4C. Figure 4.1.4 shows the result of this work in estimating the total capital requirement (TCR) necessary to implement the boiler retrofit. The function that best describes this result is shown in equation 4.1.9. It can be observed from the TCR curve shown in the figure and equation, that the retrofit TCR increases at nearly a linear rate through much of the range of boiler sizes.

Figure 4.1.4 TOTAL CAPITAL REQUIREMENT VERSUS BOILER SIZE

$$TCR_j = 0.441BoilerSize_j^{0.576} \quad \text{for all } j \quad (4.1.9)$$

Equation 4.1.9 provides the TCR needed retrofit each boiler in million dollars given the size of the boiler in MM Btu/h. This result is multiplied by the factor \$73.582 per thousand dollars of TCR to provide an annualized capital cost given an interest rate of 4% amortized over 20 years. This annualized capital cost is then accounted for in the model by considering it as part of the fixed costs of operating a retrofitted boiler.

A consequence of investment in any new, or additional, technology is increased operating and maintenance (O&M) costs in excess of fuel costs. Standard cost estimation methodology calls for annually budgeting 4% of the initial capital investment to cover operating and maintenance costs that would occur over the life of a boiler. Two percent is a more likely factor in this case since: 1) retrofitting implies that a portion of the existing boiler plant will be removed, at some cost, resulting in reduced plant equipment requiring O&M expenditures; and 2) a large percentage of costs are attributable to civil and structural work, piping, tanks, etc., that are not particularly 'high tech' in nature.

4.1.2.6 Potential MCWM Supply Sites

A database of active Pennsylvania coal mines was compiled (Miller et al, 1994). His data is very useful in that it includes location, production levels, and coal quality. A review of Cambria and Indiana County coal mines showed that as many as eight could conceivably produce a MCWM of acceptable ash and sulfur levels. These mines and the pertinent data, are shown in Table 4.1.1. The data shown in this table can be used to help define the spatial distribution of potential MCWM supply sites relative to candidate boiler retrofits in the region.

4.1.2.7 Boiler Inventory

The chief problem with developing a market penetration model is that the population of boilers numbers is in the tens of thousands. Gathering and analyzing detailed information on this many boilers would be a formidable task. The approach taken then is narrowing the model space to include only those regions that have both a large industrial base and a significant coal industry already in place. The observation that the initial market for boiler retrofits is likely to be concentrated in a small area to allow economies of scale in fuel supply to be realized substantiates this approach.

Table 4.1.1 Cambria and Indiana County Premium Bituminous Coal Mines

Mine No.	Mine Name	Company	Lat. ° ' "	Long. ° ' "	Sulfur (%)	Ash (%)	Annual Output (tons)
1	Cambria Slope Mine No. 33	Beth Energy Mines Inc.	40 27 54	78 41 51	0.88	6.5	1,686,875
2	Greenwich 2-580 Portal	Div. of Pa. Mines	40 40 30	78 53 0	1.70	14.0	878,230
3	Greenwich 2-S. Portal	Div. of Pa. Mines	40 42 0	78 52 0	1.70	14.0	870,403
4	Westover Mine	K & J Coal Co., Inc.	40 41 45	78 38 30	1.00	7.0	258,030
5	Permit # 11803038	Cooney Bros. Coal Co.	40 14 50	78 45 31	1.40	12.5	188,760
6	No. 1 Portal	Tunnelton Mining	40 27 20	79 22 34	1.50	12.9	865,017
7	Tanoma Mine	Tanoma Mining Co.	40 46 0	79 2 45	0.85	6.5	485,980
8	Heshbon Mine	Florence Mining Co.	40 27 30	79 6 35	2.10	11.3	282,732

Source: Eduardo, 1994.

The states that best meet these conditions are Indiana, Illinois, Ohio, and Pennsylvania (Miller et al., 1994). These four states comprise the second tier, or rank, of states with large industrial sector total fuel use. The first tier states, consisting of Texas, Louisiana, and California, are of lesser interest, according to Gordon (see Miller et al., 1994), since the last two have no significant coal reserves and Texas' lignite deposits would be problematic for the boiler retrofit application. Among the second tier states, Pennsylvania has the advantage of being well known to the researchers. In addition, the sulfur content of Pennsylvania coal (and that in West Virginia that can be barged to many Pennsylvania customers) is generally lower, while the heating value is higher, than in the other three states.

4.1.2.8 Pennsylvania Inventory of Active Boilers

The effort to define the model space thus began by obtaining a current census database of industrial and commercial boilers located in Pennsylvania. The census was based upon data collected by the Pennsylvania Department of Labor & Industry (DL&I) and includes only those boilers that have been in active use within the past two years.

A review of the data revealed that the population of boilers in Pennsylvania is quite large, with a total of 66,262 boilers in active service throughout the state. The boiler database itself required about 13 megabytes (MB) of memory. The useable information on each boiler in the DL&I database included:

- Location;
- Equipment type;
- Manufacturer;
- Maximum allowable working pressure;
- Pennsylvania DL&I serial number; and
- National Board Number.

While the DL&I database represented the most authoritative source of information known regarding active boilers in the state, it lacked details such as fuel type(s), annual fuel consumption, size, utilization, and other data that were required to conduct a meaningful market

penetration study. The approach, then, was to identify a concentration of possible boiler retrofit candidates located adjacent to, or co-located with, a concentration of useable coal reserves. The data provided by the DL&I were classified as to the particular type of boiler, or boilers, in use at each location. This allowed for a 'first-pass' sifting of the database to isolate a subset of boilers for further analysis that included only watertube boilers. A summary of the boiler census by boiler type is presented below in Table 4.1.2. The table shows that 6,936 boilers are of watertube design.

Table 4.1.2 Summary of Pennsylvania Boilers by Type

Boiler Type	Count	%
Cast Iron	30,760	46.4
Firetube	12,093	18.3
Watertube	6,936	10.5
Electric	4,228	6.4
Other	12,245	18.5
Total	66,262	100.0

The summary presented above should be treated with caution since the equipment type data field contained various non-standardized designations to denote the equipment type. Through a laborious process, the various indicators of boiler type were identified and standardized. In addition, the count includes boilers of all sizes and using all types of fuels.

4.1.2.9 Focus on Cambria and Indiana Counties

Two regions, corresponding to the Pittsburgh and Philadelphia metropolitan areas, encompass over 1,000 boilers each. The major drawbacks to retrofits in these areas are that the boilers in these areas would likely face greater physical space constraints and that the communities would likely provide a greater resistance to retrofit projects than elsewhere in the state. In addition, most of the bituminous coal is mined in the middle to western part of the state, with coal quality generally increasing towards the southwest.

The highest concentration of the remaining areas centers around the Cambria county region with a population of 311 boilers. A slightly larger study area was constructed by including the 151 boilers of Indiana county. The economy in this region historically has been dependent upon coal and has undergone a steady decline over the past two decades. Therefore, it

is anticipated that the residents are more likely to accept, or even endorse, technologies that utilize coal-based fuels in an environmentally acceptable manner.

Another advantage of selecting the Cambria and Indiana counties areas is that the Pennsylvania Electric Company (Penelec) is currently engaged in a long-term test program to co-fire CWM with coal in a 32 MW (107 MM Btu/h) utility boiler. Battista (1994) reports some encouraging results from this project. The presence of a slurry preparation facility already configured to supply the coarser CWM for the utility boiler market should make it easier and more economical to supply MCWM for the industrial and commercial boiler market. Economies of scale would prevail and slurry demand would be leveled out by the utility markets' year round demand for fuel.

After narrowing the population of boilers down to the Cambria and Indiana county region the next step was to contact each of the boiler owners and operators to obtain data regarding the size and average utilization of the boilers. During the early stages of this portion of the data collection effort, it became apparent that most of the 462 boilers in the region were too small to be retrofitted. Part of the reason is the economies of scale considerations for the retrofit TCR as described above in Section 4.1.2.5. The number of boilers was reduced to fewer than 35.

The result of canvassing the boilers in the Cambria and Indiana region yielded the 18 candidates for retrofitting as shown in Table 4.1.3. This table includes only those watertube boilers of sufficient size that are not already using coal as a fuel. Most of the boilers shown in this table had the capability of firing either natural gas or No. 2 fuel oil but all were currently using natural gas due to cost advantages. Details as to the size and operating characteristics of these 18 boilers are shown in Table 4.1.4.

The distances between the mines identified in Table 4.1.1 and the boilers identified in this section were determined. This final piece of data for the model is presented in Table 4.1.5. The table shows the driving distances in miles between each mine and the four cities in which the eighteen boilers are located. The program shown in Appendix 4A takes the mileage, as shown in Table 4.1.5, and constructs the distance matrix between each mine and boiler. This result is then multiplied by the mileage dependent portion of the transportation cost Equation 4.1.7 to obtain the costs to deliver MCWM from each mine to each boiler.

Table 4.1.3 Candidate Retrofit Boilers - Location and AWP

Boiler No.	City	Zip	AWP ^a psig	Year Built
1	Indiana	15701	300	UNK ^b
2	Johnstown	15901	350	1982
3	Johnstown	15901	350	1982
4	Johnstown	15901	350	1982
5	Johnstown	15901	350	1983
6	Indiana	15701	300	UNK
7	Johnstown	15901	270	1966
8	Johnstown	15901	270	1966
9	Johnstown	15905	250	1954
10	Johnstown	15905	250	1955
11	Cresson	16330	250	1971
12	Cresson	16330	250	1971
13	Cresson	16630	250	1971
14	Johnstown	15901	200	1957
15	Johnstown	15901	200	1957
16	Johnstown	15901	200	1957
17	Johnstown	15902	200	1953
18	Ebensburg	15931	200	1990

^a "AWP" is maximum allowable working pressure.

^b "UNK" indicates that the year boiler was built is unknown.

Table 4.1.4 Candidate Retrofit Boilers Operating Characteristics

Boiler No.	Size MM Btu/h	Avg. Annual Utilization	Fuel Cost \$/MM Btu	Annual Fuel Cost
1	68.0	60%	3.62	\$ 1,294,703
2	50.4	53%	3.62	\$ 840,078
3	50.4	53%	3.62	\$ 840,078
4	50.4	53%	3.62	\$ 840,078
5	50.4	53%	3.62	\$ 840,078
6	68.0	60%	3.62	\$ 1,294,703
7	19.4	50%	3.62	\$ 307,029
8	19.4	50%	3.62	\$ 307,029
9	11.6	5%	4.5	\$ 22,858
10	11.6	5%	4.5	\$ 22,858
11	27.7	30%	3.62	\$ 266,692
12	27.7	30%	3.62	\$ 266,692
13	27.7	30%	3.62	\$ 266,692
14	11.6	30%	3.62	\$ 110,329
15	11.6	30%	3.62	\$ 110,329
16	11.6	30%	3.62	\$ 110,329
17	53.2	1%	3.62	\$ 16,872
18	94.5	5%	3.62	\$ 150,014

Table 4.1.5 Driving Mileage From Mine to Boiler Zip Code

Mine No.	Boiler Zip Code			
	15701	15901	16330	15931
1	30	23	9	3
2	22	35	32	20
3	20	40	33	25
4	40	44	32	22
5	45	15	35	29
6	30	42	57	51
7	14	44	43	36
8	23	19	33	26

4.1.3 Results and Conclusions

4.1.3.1 Market Penetration Model Results

The results of the equilibrium market penetration model are summarized in Table 4.1.6. The results indicate that six of the eighteen candidate retrofit boilers could be profitably retrofitted. While these six boilers constitute only one-third of the retrofit candidates, they are responsible for 75.5% of the annual fuel demand. The total annual savings due to retrofitting is \$3.58 million per year as compared to the six boilers continuing to fire natural gas. The total capital requirement (TCR) to retrofit all six boilers is \$26.9 million. The combined annual savings and capital required result in an overall rate of return of 11.9%.

The optimal mine location was found to be the Heshbon mine (Table 4.1.1). The total fuel savings results in approximately 69,000 tons per year (tpy) of additional coal demand. This amount of coal would be equivalent to 24.4% of total annual production for that mine.

Table 4.1.7 shows the optimal retrofit results broken out by the individual boilers. The table shows that there are two types of boilers that are amenable to retrofitting. Boiler nos. 1 and 6, termed “large” boilers, each require \$5.01million TCR in order to generate a \$788,000 annual savings per boiler resulting in a 14.7% rate of return. Boiler nos. 2, 3, 4, and 5 are “small” boilers that each require \$4.22 million TCR in order to generate a \$501,000 annual savings per boiler resulting in a 10.1% rate of return.

The fuel costs (f.o.b. boiler) shown in Table 4.1.7 for the “large” and “small” boilers are \$1.42 per MM Btu and \$1.48 per MM Btu, respectively. These costs include all costs necessary to fire MCWM except annualized capital charge in order for the rate of returns to be calculated. Adding in the annualized capital charges results in MCWM costs of \$2.45 per MM Btu and \$2.81 per MM Btu respectively.

Figures 4.1.5 and 4.1.6 show the relative MCWM costs for “large” and “small” boilers, respectively. Capital charges represent the largest cost component associated with retrofitting the boiler, followed by costs associated with producing MCWM at the mine. Transportation and Operations & Maintenance (O&M) combined to contribute about 20% to costs. Ash disposal charges constitute only 0.4% of costs.

The optimal mix of retrofitted boilers and non-retrofitted as presented above, was determined by solving the nonlinear program as described in Section 4.1.3. Solving this problem presented certain challenges that deserve further mention.

Table 4.1.6 Summary of Optimal Retrofit Results

Number of candidate boiler retrofits	18
Annual total boiler fuel demand	2,188 thousand MM Btu/ year
Number of boilers that are optimal to retrofit	6
Retrofitted boiler's annual fuel demand	1,652 thousand MM Btu/ year
Percentage of total annual fuel demand by retrofitted boilers	75.5 %
Retrofitted boilers' estimated additional coal consumption	69,332 tons/ year
Total Capital Requirement (TCR) to effect boiler retrofits	26.9 million \$
Total annual savings due to retrofitting	3.58 million \$/ year
Retrofit project aggregate rate of return	11.90 %

Notes:

1. Optimal mine location for base case is the Heshbon Mine (mine no. 8).
2. Annual additional boiler coal consumption estimated by assuming coal higher heating value of 12,810 Btu/ lb and a moisture content of 7%.
3. Total annual savings are calculated without annualized capital charges.
4. Retrofit rate of return calculated as annual savings realized over a 20 year boiler life.

Table 4.1.7 Detailed Optimal Retrofit Results

Boiler No.	Optimal Fuel Choice	Fuel Costs f.o.b. Boiler \$/MM Btu	TCR M\$	Fuel Demand thousand MM Btu/y	Annual Fuel Cost M\$/y	Annual Retrofit Savings M\$/y	Retrofit Rate of Return
1	MCWM	1.416	5.01	357.7	0.506	0.788	14.7%
2	MCWM	1.482	4.22	234.2	0.347	0.501	10.1%
3	MCWM	1.482	4.22	234.2	0.347	0.501	10.1%
4	MCWM	1.482	4.22	234.2	0.347	0.501	10.1%
5	MCWM	1.482	4.22	234.2	0.347	0.501	10.1%
6	MCWM	1.416	5.01	357.7	0.506	0.788	14.7%
7	N.G.	3.62	N.R.	85.0	0.308	-	-
8	N.G.	3.62	N.R.	85.0	0.308	-	-
9	N.G.	4.50	N.R.	5.1	0.023	-	-
10	N.G.	4.50	N.R.	5.1	0.023	-	-
11	N.G.	3.62	N.R.	72.8	0.264	-	-
12	N.G.	3.62	N.R.	72.8	0.264	-	-
13	N.G.	3.62	N.R.	72.8	0.264	-	-
14	N.G.	3.62	N.R.	30.5	0.110	-	-
15	N.G.	3.62	N.R.	30.5	0.110	-	-
16	N.G.	3.62	N.R.	30.5	0.110	-	-
17	N.G.	3.62	N.R.	4.7	0.017	-	-
18	N.G.	3.62	N.R.	41.4	0.150	-	-
TOTALS			26.90	2,188.3	4.352	3.516	11.6%

Notes:

1. Optimal mine location for base case is the Heshbon Mine (Sequence No. 8).
2. N.G. is Natural Gas.
3. N.R. indicates that retrofitting is not performed on the boiler.
4. Fuel Costs (f.o.b. boiler) does not include annualized capital charges.
5. Retrofit rate of return calculated as annual savings realized over 20 year boiler life.

Figure 4.1.5 DISTRIBUTION OF "LARGE" BOILER MCWM COSTS

Figure 4.1.6 DISTRIBUTION OF "SMALL" BOILER MCWM COSTS

The programming language chosen to compile the nonlinear programming model was GAMS release 2.5 (Brooke et al., 1992). The specific solver used was the MINOS 5.3 developed by Gill et al., (1990). The solver uses an “heuristic” method to find an optimal solution. The solution of nonlinear problems is difficult, particularly in models where the nonlinearity, in this case the MCWM production costs, is one of the constraints. There are two issues to consider. The first issue involves finding an optimal solution. The second issue is that a particular solution may only be “locally” optimal. That is, there may exist other, perhaps better, solutions with notably different conditions.

In this light, the difficulty of solving the nonlinear problem presented in Section 4.1.3 becomes clearer. The main difficulty involves the economies of scale inherent in MCWM production needing to be considered at eight different potential locations. Locally optimal solutions were found that called for MCWM production to be split up among two or more coal mines. However, limiting production to one coal field produces the best solution. The MINOS 5.3 solver could find this optimal solution through the judicious selection of initial conditions of flows of coal and natural gas from the supply points to the boilers. The same result can be obtained by adding a constraint that all MCWM production must occur at only one of the eight mines. Even with this limiting constraint in place, however, it was necessary to run the solver multiple times to produce the optimal solution. The results from one, non-optimal, run would serve as initial conditions for the next until an acceptable optimal solution was found.

4.1.3.2 Sensitivity Analysis

Sensitivity to MCWM Supply Location

A consideration in the model results is the sensitivity of MCWM supply location on the optimal results. Since the nonlinear programming model minimizes costs, of which transportation costs represent one component, then the solution to this model minimizes the transportation costs given the spatial distribution of potential MCWM production plant locations and of candidate retrofit boilers.

One method of showing the sensitivity of MCWM production plant location upon the optimal results is to find solutions to the nonlinear problem that force MCWM production to occur at each location, in turn, and compare the results. Table 4.1.8 shows the results of this analysis. This table shows the marginal costs of MCWM supply to each boiler, on a \$/MM Btu

basis, that is conditional upon a particular location being selected to provide all MCWM to the boilers that are optimal to retrofit. The table then gives the additional costs to MCWM supply that arise due to the increased transportation costs associated with locating the production plant at other than the optimal location (mine no. 8). For instance, locating the MCWM production plant at mine no. 6 results in an additional \$0.03/MM Btu for boiler nos. 1 and 6, and an additional \$0.09/MM Btu for boiler nos. 2 through 5. A negative marginal cost (e.g., mine no. 7 supplying boiler no. 1) indicates that MCWM could be delivered to that particular mine at lower than the optimal case results. This result occurs because other boilers in the group are at such a cost disadvantage given that mine location. Boiler nos. 7 through 18 are not optimal to retrofit. Therefore, the marginal costs shown for these boilers shows the marginal cost to supply that boiler with MCWM if retrofitting of that boiler is insisted upon.

Table 4.1.8 also shows the sensitivity of the results to changes in the price level of natural gas. For instance, the price of natural gas would have to fall \$0.81/MM Btu to \$2.81/MM Btu before boiler nos. 2 to 5 would no longer be economical to retrofit. Likewise, long term increases in natural gas prices would increase the number of boilers that would be economical to retrofit. For example, with MCWM production located at mine no. 8 higher natural gas prices of approximately \$0.18/MM Btu would result in boiler nos. 7 and 8 becoming economical to retrofit.

It is clear from the results shown in Table 4.1.8 that MCWM siting does not greatly effect the optimal solution. Boilers that were economical to retrofit under optimal conditions would still be economical to retrofit even if the MCWM production location were to be sited at any of the other seven mines.

Another way of demonstrating this result is to show how the aggregate annual retrofit savings and rate of return varies with selecting a different mine location. Such an analysis is presented in Table 4.1.9. As can be seen by inspecting the table, the annual retrofit savings vary between \$3.4 to \$3.6 million resulting in a variation of rate of returns of 11.3% to 11.9%. With such a small variation in return, other factors, such as coal quality or technical expertise, may be considered when deciding where to locate the MCWM production plant in the two-county region.

Table 4.1.8 Marginal Cost of Fuel Supply Conditional Upon MCWM Siting (\$/MM Btu)

Boiler No.	Fuel Origin								
	N.G.	1	2	3	4	5	6	7	8
1	1.17	0.03	0.00	-0.01	0.06	0.08	0.03	-0.03	-
2	0.81	0.02	0.06	0.08	0.09	-0.01	0.09	0.09	-
3	0.81	0.02	0.06	0.08	0.09	-0.01	0.09	0.09	-
4	0.81	0.02	0.06	0.08	0.09	-0.01	0.09	0.09	-
5	0.81	0.02	0.06	0.08	0.09	-0.01	0.09	0.09	-
6	1.17	0.03	0.00	-0.01	0.06	0.08	0.03	-0.03	-
7	-	0.19	0.24	0.25	0.27	0.16	0.26	0.27	0.18
8	-	0.19	0.24	0.25	0.27	0.16	0.26	0.27	0.18
9	-	29.95	30.00	30.02	30.03	29.92	30.02	30.03	29.94
10	-	29.95	30.00	30.02	30.03	29.92	30.02	30.03	29.94
11	-	1.31	1.39	1.40	1.39	1.40	1.48	1.43	1.40
12	-	1.31	1.39	1.40	1.39	1.40	1.48	1.43	1.40
13	-	1.31	1.39	1.40	1.39	1.40	1.48	1.43	1.40
14	-	3.07	3.11	3.13	3.15	3.04	3.14	3.15	3.05
15	-	3.07	3.11	3.13	3.15	3.04	3.14	3.15	3.05
16	-	3.07	3.11	3.13	3.15	3.04	3.14	3.15	3.05
17	-	84.81	84.85	84.87	84.89	84.78	84.88	84.89	84.79
18	-	11.13	11.20	11.22	11.20	11.23	11.31	11.26	11.22

Notes:

1. Optimal MCWM fuel supply siting for boiler nos. 1 to 6 is coal mine no. 8, the remaining boilers are optimally supplied by natural gas.
2. Values shown are marginal costs as compared to the optimal case conditional upon the mine (column) supplying all the fuel for boiler nos. 1 to 6.
3. N.G. is Natural Gas.
4. Fuel prices (f.o.b. boiler) include annualized capital charges.

Table 4.1.9 Sensitivity of Aggregate Project Results to MCWM Plant Siting

Mine No.	Aggregate TCR M\$	Annual Retrofit Savings M\$/y	Aggregate Retrofit Rate of Return
1	26.9	3.546	11.8%
2	26.9	3.526	11.7%
3	26.9	3.514	11.6%
4	26.9	3.447	11.3%
5	26.9	3.534	11.7%
6	26.9	3.480	11.5%
7	26.9	3.516	11.6%
8	26.9	3.579	11.9%

Sensitivity to Number of Retrofits Performed

The equilibrium results found here assume that all boiler owners instantaneously retrofit their boilers if there exists an economic benefit to doing so. As explained in Section 4.1.3, however, the adoption of the retrofit technology may be slow or not happen at all.

A sensitivity analysis of the optimal results investigated the effects of fewer than six boilers being retrofitted. The results of this analysis are shown in Table 4.1.10. As expected, lowering the number of retrofits lowers both the total TCR necessary and the aggregate annual retrofit savings. The aggregate retrofit rate of return does not follow a smooth decreasing rate. This is mainly due to the “lumpy” form of capital investment. For instance, when one of the “small” boilers is excluded, this concentrates the capital investment, as a percent, with the larger boilers that provided a larger rate of return. The other effect at work is the increased MCWM costs associated with decreasing aggregate MCWM demand in the region. This effect is responsible for lowering the retrofit rate of return but is not a particularly significant effect until only one of the “small” boilers is retrofitted.

Breakeven Radius of MCWM Supply

Another analysis considered is determination of the effective radius that a retrofitted boiler could be economically served from a single MCWM production plant. This radius is determined by

Table 4.1.10 Sensitivity of Aggregate Project Results to Number of Retrofits

Number of "Large" Boilers	Number of "Small" Boilers	Aggregate MCWM Demand thousand MM Btu/y	Fuel Price f.o.b. Prep Plant \$/MM Btu	Total TCR M\$	Aggregate Annual Retrofit Savings M\$/y	Aggregate Retrofit Rate of Return
2	4	1,652	0.872	26.9	3.58	11.9%
2	3	1,418	0.904	22.7	3.03	12.0%
1	4	1,294	0.926	21.9	2.72	10.9%
2	2	1,184	0.948	18.5	2.49	12.1%
1	3	1,060	0.980	17.7	2.18	10.7%
0	4	937	1.019	16.9	1.87	9.1%
1	2	826	1.063	13.5	1.63	10.5%
1	1	592	1.212	9.2	1.09	10.0%
1	0	358	1.552	5.0	0.55	8.9%
0	1	234	1.998	4.2	0.24	1.1%

varying the distance of one of the "large" boilers from its current location out to a distance that would result in retrofitting to be a breakeven proposition as compared to continuing to fire natural gas. This type of analysis is sensitive to the interest rates when considering whether to expend the capital needed to effect the retrofit.

The breakeven retrofit radius analysis that considers both distance and interest rates is presented in Table 4.1.11. This table shows that at an interest rate of 4% the maximum distance a "large" boiler can be located from the mine for retrofitting to remain economical is 338 miles. At an interest rate of 10% that breakeven radius drops to 173 miles. At just over a 14% interest rate the breakeven retrofit radius drops to zero miles.

4.1.4 Conclusions

A market penetration model was formulated to find the equilibrium optimal mix of boiler retrofit technology adoption among a sample population of 462 water tube boilers located in the Pennsylvania counties of Cambria and Indiana. This region was selected so that detailed information could be collected to allow the analysis to be carried out. It was found that six of

Table 4.1.11 Breakeven Retrofit Radius for "Large" Boiler

Interest Rate %	Annual Capital Charges \$	Annual Retrofit Savings Less Transportation \$	Annual Boiler Fuel Demand thousand MM Btu	Breakeven Retrofit Radius miles
2	306,395	512,572	357.7	384
4	368,645	450,322	357.7	338
6	436,795	382,172	357.7	286
8	510,280	308,687	357.7	231
10	588,473	230,494	357.7	173
12	670,733	148,234	357.7	111
14	756,440	62,527	357.7	47
16	845,022	-26,055	357.7	-

Notes:

1. Analysis considers boiler no. 1 requiring \$5.01M TCR.
2. Annual capital charges are determined by amortizing TCR over 20 years.
3. Annual retrofit savings are \$819,000 (f.o.b. mine).
4. Mileage dependent transportation rate is \$0.00373/MM Btu/mile.

these boilers would benefit economically from adopting the retrofit technology. Retrofitting these six boilers would result in a total cost savings of \$3.58 million annually for an aggregate rate of return of 11.9%. This equilibrium result leads to 69,000 tons per year (tpy) additional coal consumption in the region.

It was also found that the study region could be expanded to other portions of the state out to 173 miles assuming that industrial boiler owners use 10% interest rates when considering capital investments. Considering the distribution of boilers throughout Pennsylvania, as described in Section 4.1.2.8, it can be concluded that the boilers located in the eastern part of the state could not be served from the coal producing regions of the west central to south western part of the state. With this consideration in mind, it is estimated that only 40% of Pennsylvania's 6,936 water tube boilers are located in areas that could consider retrofitting as a viable option. It is estimated that 36 Pennsylvania boilers would benefit from adopting the retrofit technology. This estimate assumes that the ratio of boiler retrofits to total water tube boilers in the study

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Pennsylvania is not the only major potential market for the boiler retrofit technology. Ohio, Indiana, and Illinois are also states that have the combination of a large industrial base near extensive deposits of coal. An estimated total number of retrofits for each of these states is given in Table 4.1.12. The estimates for Ohio, Indiana, and Illinois use the ratio of industrial sector energy consumption in each state, as compared to Pennsylvania, as a proxy for the relative number of boilers in each state. In addition, the other states are likely to suffer less from geographic isolation between boilers and coal fields than is the case for Pennsylvania. It was assumed that 60% of the boilers in each of these states would be near enough to coal fields to justify considering adoption of the retrofit technology.

The estimates presented in Table 4.1.12 indicate that the total market for the boiler retrofit technology is some 186 boilers comprising 3,400 tpy of coal consumption. Again, these are rough estimates only. For example, it may be the case that no Ohio coal is of sufficient quality to allow its use in a retrofitted boiler. It can be concluded, however, that the total market potential for the retrofit technology in Pennsylvania, Ohio, Indiana, and Illinois is likely to be at least 90 boilers but less than 200 boilers. This range of retrofitted boilers would constitute about 1,700 to 3,700 tpy of coal consumption.

4.2 Selection of Incentives for Commercialization of the Coal-Using Technology

4.2.1 Motivation and Aim of the Study

The results of the Phase II work indicated that there are social, economic and political benefits of substituting clean burning coal-based fuels for oil and gas (Miller et al., 2000). A

Table 4.1.12 Estimated Number of Boiler Retrofits and Coal Consumption

State	1992 Industrial Sector				Estimated Number of Retrofits	Estimated Retrofit Fuel Consumption thousand MM Btu	Estimated Retrofit Coal Consumption thousand tons
	1992 Industrial Sector Energy Consumption trillion Btu	Bituminous & Lignite Coal Consumption thousand tons					
Pennsylvania	1,357	13,859		36	15,779		662
Ohio	1,377	50,358		55	24,017		1,008
Indiana	1,154	11,416		46	20,128		845
Illinois	1,237	6,050		49	21,575		906
TOTALS	5,125	81,683		186	81,499		3,420

Notes:

1. Consumption data is from U.S. DOE/EIA "State Energy Data Report, 1992: Consumption Estimates."
2. Coal assumed to have a higher heating value of 12,810 Btu/lb and 7% moisture content.

heavy reliance on imported oil makes the United States vulnerable to price shocks, i.e., periods of increased price volatility, and disruptions in the supply of oil. Clearly, price shocks and disruptions in supply may have significant economic and political effects in the U.S. As previous studies document, the infamous oil embargo of 1973-1974 by OPEC resulted in approximately a 2% rise in unemployment and a 2% increase in the rate of inflation.

The world has been witnessing a downtrend in oil prices, on average, since 1989; nevertheless, due to the concentrated political and highly volatile nature of the oil market, it is not an easy task to predict how oil prices would fare over the next decades. On the other hand, coal prices have been exhibiting one of the smallest price fluctuations in energy markets over the last one and a half decades. This consistent and statistically significant difference in the volatilities of the two price series, oil and coal prices, may be viewed as indicating that there are advantages to switching to a coal-based technology from a oil-based one for risk management purposes.

The Phase II Final Report addressed the question of estimating the net costs of this technology switch at the firm level (Miller et al., 2000). Specifically, the report addressed the estimate of the magnitude of the gross costs and benefits associated with the suggested retrofitting. As it was remarked, the “success (of the new technology) will most likely be achieved through those technologies that can achieve the margin of economic viability by reducing capital costs and not through ‘serendipitous’ changes in oil price alone.” (Miller et al., 2000). This finding clearly shows that even though there may be significant advantages in substituting clean-burning coal-based fuels for oil and gas at the macroeconomic level, at the microeconomic level, e.g., firm level, the costs of retrofitting are considerably high for individual firms, so that managers may be tempted to decide against it. Consequently, in the conclusions of the Phase II Final Report, it is suggested that “those technologies that can reach that marginal level of viability could make their market entry by highlighting fuel flexibility as quantified by option pricing methods.” Even though there exist undeniable benefits -in terms of options- by adopting the new technology, it is not possible to come up with an estimate of an option value for this “technological investment” due to the complex nature of the problem at hand and lack of data. Thus, the net cost figures were provided that are based on the discounted cash flow technique (Miller et al., 2000).

The initial premise of the study was that the new coal-using technologies offer significant macroeconomic benefits in terms of fuel price differentials and managerial, economic and political options (some of which are to be quantified by a further study) and that the sum of positive externalities outweigh the negative externalities, which are associated with the differential environmental damage and so forth. The previous research showed that the adoption of the coal-based technology, i.e., the investment in the new technology, had a significant cost component attached to it, which may induce the managers at the firm level to be hesitant or even unwilling to undertake it (Miller et al., 2000). Recognizing the net positive externalities, the government may decide that it is appropriate to offer incentives for firms, which have a potential use for it, to facilitate the use and commercialization of the new technology.

Economic and financial incentives in general, and those provided to manufacturing enterprises in particular, have become accepted tools for the implementation of preferred government policies. These incentives not only affect the size of investments, but also alter the basic parameters of design and operation of industrial firms. However, the combined result of implemented incentives and particularly, the quantitative decision about the choice of the incentives have not been analyzed in depth in the literature.

Typical examples of government incentives that are employed as policy tools are: investment tax credits, reduced or no-interest loans, capital subsidies and tax incentives through accelerated depreciation. These policy tools have been subject to academic research on several occasions, recently, for instance, the discussion on the effectiveness of tax rates on business investment by Chirinko (1985), the debate on the impact of investment tax credit on investment behavior and value of the firm by Auerbach and Kotlikoff (1982), Auerbach (1986) and Lyon (1988) can be mentioned as examples of research involving assessment of effectiveness of these tools in achieving objectives set forth by the government.

The studies conducted on the effectiveness of the policy tools mentioned above remain highly macroeconomics oriented and have little to say about the consequences of them in a given sector, or sub-sector in a given economy. This deficiency was addressed in two sector-oriented studies: Cone et al. (1978) and more recently, Rose and Mor (1993). Nevertheless, all previous studies attempted to estimate the impact of incentives by using aggregate market data and have ignored the response generated by the economic agents operating in those markets, i.e., the firms. This study, therefore, is aimed at filling that gap in the literature and focuses on the

microeconomic responses generated by the firms to the incentives. The study examines firm-specific data on firms that are potential customers of the new coal-using technology where the optimal incentive or combination of incentives is determined based on the empirically estimated reactions of firms.

A careful inspection of the government incentives mentioned above reveals that they are designed at fostering projects with different characteristics. In other words, one incentive may be optimal with projects with characteristic “A”, whereas another may be of choice if the project has characteristic “B”. For example, tax incentives, such as direct tax cuts, accelerated depreciation and so forth, can be thought to be more effective on projects that require a big initial capital outlay. On the other hand, reduced or no-interest loans are more useful on projects which are highly sensitive to interest rate fluctuations in the market, e.g., due to the nature of their cash flow stream.

Modern finance theory has shown that, under realistic assumptions, project selection and project financing decisions are not independent from the current financial attributes of the firm. In other words, risk exposure of firms becomes crucial in project selection and financing decisions, e.g., the same project can be regarded as profitable and desirable by one firm and not so by another.

The switch to a new technology can be thought as an investment in technology, and thus an “investment project” by itself. The importance of this is self-explanatory: if firms are given incentives, regardless of their financial and operational risk exposures, the result may be suboptimal or even off the policy target set forth by the government in some cases. The distortion of prices and the market mechanism may even produce undesirable outcomes in those industries: for instance generous capital subsidies may induce the firms to take projects with an - otherwise- unacceptable levels of risk exposure.

In sum, devising a government incentive scheme that aims at widespread commercialization of the new coal-using technology is a complex task. On the one hand, a miscalculation or negligence of firm characteristics and behavior may lead to suboptimal or even unwanted outcomes. On the other hand, calculation of the optimal mix of incentives presents another challenge for the policy makers. This task addresses both of these issues by taking individual firm characteristics into consideration in assessing the optimal strategy that should be implemented by the government.

4.2.2 Data

A list of 6,823 watertube boiler locations in the Commonwealth of Pennsylvania was generated from a database obtained from the Pennsylvania Department of Labor and Industry (PDL&I). These locations were then cross referenced against the names of publicly traded corporations or their subsidiaries for the entire United States. A total of 128 corporations or their subsidiaries were identified as having watertube boiler locations in the Commonwealth of Pennsylvania.

The latest and past one, three and five year annual financial statements, i.e., income statements and balance sheets, and key financial and operating ratios for the aforementioned 128 corporations were obtained from a CD-ROM provided by the Compact Disclosure Database Company. A closer inspection of the individual characteristics of these boilers revealed that only 57 firms (and subsidiaries), some with multiple boiler locations, fit into the category of boilers for which the new technology is developed. A list of firms included in this study is presented in Table 4.2.1.

Based on the balance sheets and income statements, the following financial ratios are calculated according to their standard definitions as follows: quick ratio (acid test), i.e., $(\text{current assets} - \text{inventory}) / \text{current liabilities}$; current ratio, i.e., $\text{current assets} / \text{current liabilities}$; net sales/cash; net sales/working capital; net sales/current assets; net sales/assets; total liabilities/total assets (D/A); liabilities/equity (D/E); total net income/net sales; measure of operating leverage (MOL), i.e., percentage change in EBIT per 1% change in sales; measure of financial leverage (MFL), i.e., percentage change in net income per 1% change in EBIT; and finally, measure of total leverage (MTL), i.e., MOL multiplied by MFL.

The data about the new coal-based technology costs are taken from the Phase II Final Report (Miller et al., 2000). Similarly, boiler-specific costs and benefits are estimated using the same algorithm that is utilized in the aforementioned sections. The data were then tabulated and organized into a convenient format to facilitate quantitative analysis of the impacts of various government incentives for commercialization of the new technology. For given levels of boiler capacity the switch in technology is treated as a real investment and the net return on investment is calculated. Tables 4.2.2 and 4.2.3 display some selected balance sheet items and calculated financial ratios for the firms in the sample.

Table 4.2.1 List of Firms Included in the Sample

- ACF INDUSTRIES INC
- ALCAN ALUMINUM LTD
- ALCOA INTERNATIONAL HOLDINGS COMPANY
- ALLEGHENY LUDLUM CORP
- ALLIED SIGNAL INC
- ALUMINUM CO OF AMERICA
- AMERICAN HOME PRODUCTS CORP
- ANGELICA CORP
- ARCO CHEMICAL CO
- ARMCO INC
- ASHLAND OIL INC
- AT&T CORP
- BEATRICE FOODS INC
- BETHLEHEM STEEL CORP
- BETZ LABORATORIES INC
- BORDEN INC
- CABOT CORP
- CARBIDE GRAPHITE GROUP INC
- CATERPILLAR INC
- CHEVRON CORP
- CONSOLIDATED CIGAR CORP NEW JERSEY
- EXXON CORP
- GENCORP INC
- GENERAL ELECTRIC CO
- GENERAL SIGNAL CORP
- GUILFORD MILLS INC
- H J HEINZ CO
- HANOVER FOODS CORP
- HERCULES INC
- INDSPEC CHEMICAL CORP
- J&L SPECIALTY STEEL INC
- KRAFT GENERAL FOODS INC
- LTV STEEL CO INC
- LUKENS INC
- MASLAND CORP
- MERCK & CO INC
- MINNESOTA MINING & MANUFACTURING
- NATIONAL GYPSUM CO
- OCCIDENTAL PETROLEUM CORP
- PPG INDUSTRIES INC
- PROCTER & GAMBLE CO
- RHONE POULENC SA
- ROHM & HAAS CO
- SEARS ROEBUCK & CO
- SMITHKLINE BEECHAM PLC
- SONOCO PRODUCTS CO
- SPS TECHNOLOGIES INC
- ST JOE PAPER CO
- TEMPLE INLAND INC
- USX CORP
- VALSPAR CORP
- WARNER LAMBERT CO
- WEST PENN POWER CO
- WESTINGHOUSE ELECTRIC CORP
- WESTVACO CORP
- WITCO CORP
- YORK INTERNATIONAL CORP

Table 4.2.2 Selected Items from the Balance Sheets of Included Firms

COMPANY NAME	TOT CUR ASSETS	TOTAL ASSETS	INCOME TAXES	CURRENT LIABILITY	TOTAL LIABILITY	COMMON STOCK	RETAINED EARNINGS	TOT LIAB-NET WORTH
ACF INDUSTRIES INC	640,603.00	1,706,454.00	8,336.00	469,213.00	1,388,925.00	76,573.00	(18,453.00)	1,706,454.00
ALCAN ALUMINIUM LTD	2,402,000.00	9,810,000.00	16,000.00	1,335,000.00	5,291,000.00	1,183,000.00	2,813,000.00	9,810,000.00
ALCOA INTL HOLDINGS COMPANY	1,111,800.00	3,872,600.00	181,900.00	571,700.00	1,243,500.00	0.00	1,827,800.00	3,872,600.00
ALLEGHENY LUDLUM CORP	469,788.00	1,174,049.00	20,634.00	210,877.00	770,627.00	7,288.00	152,258.00	1,174,049.00
ALLIED SIGNAL INC	4,567,000.00	10,829,000.00	0.00	3,489,000.00	8,439,000.00	358,000.00	1,023,000.00	10,829,000.00
ALUMINUM CO OF AMERICA	3,702,500.00	11,596,900.00	0.00	2,092,900.00	6,623,900.00	88,800.00	2,946,100.00	11,596,900.00
AMERICAN HOME PRODUCTS CORP	4,807,684.00	7,687,353.00	171,404.00	1,584,411.00	3,612,235.00	103,442.00	2,884,244.00	7,687,353.00
ANGELICA CORP	210,255.00	332,861.00	5,530.00	53,067.00	140,868.00	9,448.00	190,301.00	332,861.00
ARCO CHEMICAL CO	943,000.00	3,502,000.00	28,000.00	487,000.00	1,803,000.00	100,000.00	703,000.00	3,502,000.00
ARMCO INC	625,400.00	1,904,700.00	0.00	353,000.00	2,208,100.00	1,000.00	(1,450,300.00)	1,904,700.00
ASHLAND OIL INC	1,973,001.00	5,551,817.00	41,560.00	1,618,913.00	4,097,023.00	60,022.00	1,008,264.00	5,551,817.00
AT&T CORP	29,738,000.00	60,766,000.00	0.00	25,334,000.00	46,334,000.00	1,352,000.00	857,000.00	60,766,000.00
BEATRICE FOODS INC	137,062.00	655,641.00	4,157.00	113,772.00	491,942.00	155,140.00	4,226.00	655,641.00
BETHLEHEM STEEL CORP	1,591,100.00	5,876,700.00	0.00	914,200.00	5,180,100.00	93,400.00	(939,900.00)	5,876,700.00
BETZ LABORATORIES INC	208,635.00	521,129.00	6,838.00	92,041.00	221,810.00	3,365.00	394,726.00	521,129.00
BORDEN INC	1,290,200.00	3,871,700.00	56,500.00	1,371,500.00	3,117,000.00	121,900.00	835,100.00	3,871,700.00
CABOT CORP	544,206.00	1,489,473.00	26,314.00	354,221.00	1,047,200.00	33,887.00	861,803.00	1,489,473.00
CARBIDE GRAPHITE GROUP INC	102,693.00	171,870.00	213.00	32,665.00	115,314.00	70.00	42,869.00	171,870.00
CATERPILLAR INC	6,071,000.00	14,807,000.00	111,000.00	4,671,000.00	12,608,000.00	835,000.00	1,234,000.00	14,807,000.00
CHEVRON CORP	8,682,000.00	34,736,000.00	782,000.00	10,606,000.00	20,739,000.00	1,069,000.00	13,955,000.00	34,736,000.00
CONSOLIDATED CIGAR CORP NEW JERSEY	49,748.00	205,906.00	0.00	15,771.00	173,027.00	1.00	2,879.00	205,906.00
EXXON CORP	14,859,000.00	84,145,000.00	2,359,000.00	18,590,000.00	46,958,000.00	2,822,000.00	49,365,000.00	84,145,000.00
GENCORP INC	430,000.00	1,164,000.00	14,000.00	341,000.00	929,000.00	3,000.00	229,000.00	1,164,000.00
GENERAL ELECTRIC CO	195,240,000.00	251,506,000.00	0.00	178,638,000.00	224,026,000.00	584,000.00	28,613,000.00	251,506,000.00
GENERAL SIGNAL CORP	594,545.00	1,224,841.00	7,385.00	325,848.00	699,655.00	77,082.00	583,099.00	1,224,841.00
GUILFORD MILLS INC	248,638.00	506,742.00	0.00	96,644.00	287,003.00	393.00	244,066.00	506,742.00
H J HEINZ CO	2,291,530.00	6,381,146.00	130,535.00	1,692,362.00	4,042,595.00	71,850.00	3,633,385.00	6,381,146.00

HANOVER FOODS CORP	73,013.00	124,646.00	1,092.00	50,734.00	81,656.00	21,042.00	26,371.00	124,646.00
HERCULES INC	1,226,523.00	3,161,961.00	0.00	884,211.00	1,793,754.00	31,198.00	1,955,005.00	3,161,961.00
INDSPEC CHEMICAL CORP	32,269.00	237,125.00	0.00	23,571.00	230,874.00	1.00	8.00	237,125.00
J&L SPECIALTY STEEL INC	209,384.00	626,038.00	165.00	99,120.00	358,522.00	387.00	(36,959.00)	626,038.00
KRAFT GENERAL FOODS INC	6,982,000.00	32,669,000.00	428,000.00	6,577,000.00	18,602,000.00	0.00	1,529,000.00	32,669,000.00
LTV STEEL CO INC	1,350,900.00	4,584,100.00	0.00	716,600.00	8,399,500.00	100.00	(5,239,100.00)	4,584,100.00
LUKENS INC	307,739.00	817,178.00	0.00	161,705.00	550,424.00	158.00	193,977.00	817,178.00
MASLAND CORP	101,924.00	203,774.00	0.00	79,629.00	129,503.00	132.00	34,755.00	203,774.00
MERCK & CO INC	5,734,600.00	19,927,500.00	1,430,400.00	5,895,700.00	8,761,400.00	4,576,500.00	9,393,200.00	19,927,500.00
MINNESOTA MINING & MANUFACTU RING	6,363,000.00	12,197,000.00	290,000.00	3,282,000.00	5,685,000.00	6,512,000.00	0.00	12,197,000.00
NATIONAL GYPSUM CO	196,480.00	774,340.00	10,868.00	67,144.00	1,428,135.00	1.00	(744,195.00)	774,340.00
OCCIDENTAL PETROLEUM CORP	1,934,000.00	17,123,000.00	110,000.00	2,048,000.00	13,152,000.00	61,000.00	(1,883,000.00)	17,123,000.00
PPG INDUSTRIES INC	2,025,900.00	5,651,500.00	4,700.00	1,281,000.00	3,126,500.00	242,100.00	3,436,800.00	5,651,500.00
PROCTER & GAMBLE CO	9,988,000	25,535,000	0	8,040,000	16,703,000	684,000	7,496,000	25,535,000
RHONE POULENC SA	41,813,000.00	114,481,000.00	0.00	31,492,000.00	64,730,000.00	6,271,000.00	13,155,000.00	114,481,000.00
ROHM & HAAS CO	1,200,000.00	3,524,000.00	3,000.00	701,000.00	2,012,000.00	197,000.00	1,444,000.00	3,524,000.00
SEARS ROEBUCK & CO	25,549,800.00	90,807,800.00	0.00	57,290,200.00	76,809,700.00	293,800.00	8,162,800.00	90,807,800.00
SMITHKLINE BEECHAM PLC	3,393,000.00	5,438,000.00	0.00	2,178,000.00	3,608,000.00	335,000.00	831,000.00	5,438,000.00
SONOCO PRODUCTS CO	513,110.00	1,707,125.00	3,071.00	303,178.00	918,761.00	7,175.00	623,500.00	1,707,125.00
SPS TECHNOLOGI ES INC	161,010.00	285,979.00	646.00	66,527.00	183,152.00	6,362.00	60,516.00	285,979.00
ST JOE PAPER CO	283,856.00	1,491,271.00	2,737.00	93,399.00	348,940.00	8,714.00	851,511.00	1,491,271.00
TARKETT INTERNATION AL GMBH	13,048.00	13,048.00	0.00	12,998.00	12,998.00	50.00	0.00	13,048.00
TEMPLE INLAND INC	4,671,243.00	11,959,260.00	0.00	9,021,256.00	10,259,080.00	61,390.00	1,482,093.00	11,959,260.00
USX CORP	3,180,000.00	17,374,000.00	0.00	3,334,000.00	13,510,000.00	366,000.00	(831,000.00)	17,374,000.00
VALSPAR CORP	197,480.00	336,798.00	11,412.00	113,481.00	140,280.00	13,330.00	223,483.00	336,798.00
WARNER LAMBERT CO	2,218,700.00	4,828,100.00	180,300.00	2,015,900.00	3,438,500.00	160,300.00	2,287,700.00	4,828,100.00
WEST PENN POWER CO	229,283.00	2,544,763.00	11,533.00	184,109.00	1,501,086.00	425,994.00	412,288.00	2,544,763.00
WESTINGHO USE ELECTRIC CORP	4,774,000.00	10,553,000.00	0.00	3,925,000.00	9,474,000.00	393,000.00	1,401,000.00	10,553,000.00

WESTVACO CORP	609,284.00	3,927,837.00	15,574.00	365,325.00	2,103,849.00	545,166.00	1,294,130.00	3,927,837.00
WITCO CORP	792,573.00	1,838,998.00	0.00	341,338.00	1,125,583.00	254,089.00	488,241.00	1,838,998.00
YORK INTERNATION AL CORP	702,775.00	1,335,181.00	35,072.00	521,699.00	878,214.00	188.00	36,227.00	1,335,181.00

Table 4.2.3 Selected Financial Ratios for the Firms in the Sample

COMPANY NAME	QUICK RATIO	CURRENT RATIO	LIABILITY/E QUITY	INCOME/SAL ES	INCOME/ASS ETS	INCOME/EQ UITY
ACF INDUSTRIES INC	1.24	1.37	4.37	(0.05)	(0.01)	(0.06)
ALCAN ALUMINIUM LTD	0.86	1.80	1.31	(0.01)	(0.01)	(0.03)
ALCOA INTERNATIO NAL HOLDINGS COMPANY	1.33	1.94	1.22	0.07	0.05	0.18
ALLEGHENY LUDLUM CORP	0.99	2.23	1.91	0.06	0.06	0.18
ALLIED SIGNAL INC	0.64	1.31	3.53	0.03	0.04	0.17
ALUMINUM CO OF AMERICA	0.90	1.77	3.10	0.00	0.00	0.00
AMERICAN HOME PRODUCTS CORP	2.28	3.03	0.98	0.18	0.19	0.40
ANGELICA CORP	1.32	3.96	0.73	0.03	0.03	0.06
ARCO CHEMICAL CO	1.01	1.94	1.24	0.07	0.06	0.15
ARMCO INC	0.99	1.77	(4.34)	(0.39)	(0.34)	1.26
ASHLAND OIL INC	0.75	1.22	3.53	0.01	0.03	0.12
AT&T CORP	0.94	1.17	3.49	(0.06)	(0.06)	(0.29)
BEATRICE FOODS INC	0.80	1.20	3.01	0.00	0.00	0.01
BETHLEHEM STEEL CORP	0.80	1.74	7.59	(0.06)	(0.05)	(0.39)
BETZ LABORATORI ES INC	1.59	2.27	0.75	0.10	0.13	0.22
BORDEN INC	0.32	0.94	(11.86)	(0.11)	(0.16)	2.40
CABOT CORP	0.84	1.54	2.83	0.01	0.01	0.03
CARBIDE GRAPHITE GROUP INC	0.00	3.14	2.24	0.02	0.03	0.11
CATERPILLA R INC	0.79	1.30	5.73	0.06	0.04	0.30
CHEVRON CORP	0.55	0.82	1.48	0.03	0.04	0.09
CONSOLIDAT ED CIGAR CORP NEW JERSEY	0.78	3.15	5.26	0.03	0.01	0.09
EXXON CORP	0.46	0.80	1.48	0.05	0.06	0.17
GENCORP INC	0.55	1.26	3.95	0.02	0.04	0.18

GENERAL ELECTRIC CO	1.07	1.09	9.27	0.07	0.02	0.18
GENERAL SIGNAL CORP	0.79	1.82	1.33	0.02	0.03	0.07
GUILFORD MILLS INC	1.56	2.57	1.31	0.04	0.06	0.13
H J HEINZ CO	0.56	1.35	1.73	0.09	0.09	0.26
HANOVER FOODS CORP	0.49	1.44	1.94	0.03	0.05	0.15
HERCULES INC	0.83	1.39	1.31	(0.01)	(0.01)	(0.02)
INDSPEC CHEMICAL CORP	0.71	1.37	36.93	0.00	0.00	0.00
J&L SPECIALTY STEEL INC	0.68	2.11	1.34	0.03	0.03	0.07
KRAFT GENERAL FOODS INC	0.54	1.06	1.32	0.02	0.02	0.05
LTV STEEL CO INC	0.00	1.89	(2.20)	(0.08)	(0.06)	0.07
LUKENS INC	0.78	1.90	2.35	(0.06)	(0.06)	(0.21)
MASLAND CORP	0.84	1.28	2.06	0.05	0.10	0.33
MERCK & CO INC	0.62	0.97	0.99	0.21	0.11	0.24
MINNESOTA MINING & MANUFACTURING	1.00	1.94	0.87	0.09	0.10	0.19
NATIONAL GYPSUM CO	1.99	2.93	(2.18)	(0.15)	(0.09)	0.11
OCCIDENTAL PETROLEUM CORP	0.34	0.94	3.90	0.03	0.02	0.08
PPG INDUSTRIES INC	0.87	1.58	1.29	0.00	0.00	0.01
PROCTER & GAMBLE CO	0.72	1.24	2.42	0.07	0.09	0.32
RHONE POULENC SA	0.60	1.33	3.19	0.01	0.01	0.05
ROHM & HAAS CO	0.91	1.71	1.63	0.03	0.03	0.09
SEARS ROEBUCK & CO	0.38	0.45	9.89	0.05	0.03	0.31
SMITHKLINE BEECHAM PLC	1.04	1.56	2.08	0.13	0.15	0.47
SONOCO PRODUCTS CO	0.93	1.69	1.49	0.06	0.07	0.19
SPS TECHNOLOGIES INC	0.84	2.42	1.78	(0.10)	(0.11)	(0.30)
ST JOE PAPER CO	2.02	3.04	0.53	0.02	0.01	0.02
TARKETT INTERNATIONAL GMBH	0.00	1.00	259.96	0.00	0.00	0.00
TEMPLE INLAND INC	0.35	0.52	6.03	0.04	0.01	0.07
USX CORP	0.36	0.95	3.60	(0.01)	(0.01)	(0.07)
VALSPAR CORP	0.94	1.74	0.71	0.06	0.12	0.20

WARNER LAMBERT CO	0.66	1.10	2.47	0.06	0.07	0.24
WEST PENN POWER CO	0.64	1.25	1.68	0.09	0.04	0.11
WESTINGHO USE ELECTRIC CORP	0.51	1.22	9.45	(0.04)	(0.03)	(0.33)
WESTVACO CORP	0.77	1.67	1.15	0.04	0.03	0.06
WITCO CORP	1.53	2.32	1.58	0.01	0.01	0.03
YORK INTERNATIO NAL CORP	0.64	1.35	1.92	0.00	0.00	0.01

4.2.3 Methodology

In order to assess firms' responsiveness to alternative incentives and financial ratios, the following regression equation is estimated:

$$y_i = \alpha + \sum_j \beta_j Z_{ji} + \varepsilon_i \quad (4.2.1)$$

where y_i stands for the net income of the i -th firm, α is a constant, Z_j is the j -th vector of the Z -matrix which includes the explanatory variables, and finally β_j is the estimated coefficient of Z_j and ε_i is a white-noise error term. The Z_j matrix includes variables such as, cost of goods sold, research and development expenses, fixed costs, depreciation, interest expenses, taxes, measures of financial, operational and total leverages, and other aforementioned financial ratios, e.g., debt-equity ratio, current ratio, quick ratio, etc.

In order to determine the explanatory variables in the model, in the regression equation, a stepwise regression procedure is applied. Stepwise regression can adopt a forward selection criterion, where variables are added to the model sequentially until none of the remaining would have t -statistics with a P -value (significance level) smaller than a threshold value. Alternatively, it can also adopt a backward criterion, where starting from the full set of regressors, variables are deleted sequentially as long as their t -statistics produce a P -value larger than a threshold value. In this study, variables are added to the model sequentially; at each stage in this forward selection procedure, the backward selection algorithm is run to delete variables that now have small t -statistics.

It is known that ordinary least squares (OLS) provides a consistent estimator for β in the regression model $Y = X\beta + u$ in a large number of settings where the standard assumption that

the residuals satisfy: $V = E(uu') = \sigma^2 I$. If this assumption is violated and the form of V is known, it may be possible to obtain a more efficient estimator by some form of generalized least squares (GLS). However, in certain cases, GLS for serially correlated residuals produces inconsistent parameter estimates (See Hayashi and Sims 1983). Moreover, in the case of heteroscedasticity, it may not always be clear what form V should take. Hansen (1982) demonstrated that it is possible to compute consistent estimators for the covariance matrix of estimators in a wide range of situations using a procedure that imposes little structure upon matrix V . An alternative method for calculating consistent covariance matrices for the estimated coefficients is provided by Newey and West (1987). Hence, to assure the reliability of the reported test statistics of estimated coefficients the regressions are performed with the Newey and West method.

Once the model is determined and estimated the coefficients can be translated to elasticity measures (at the averages), such that the sensitivity of firms' income with respect to 1% change in any of the explanatory variables can be determined. For example, the coefficient of the tax variable, once converted into an elasticity, will reveal how firms income will respond to a 1% change in taxes.

The net cost of the technological investment is estimated for each particular boiler in the sample; the framework developed in Sections 4.1 and 4.2 of the Phase II Final Report is utilized for this task (Miller et al., 2000). The net cost is divided by the current income of the firm in order to express the necessary increase in income in percentage points.

Consequently, the ratio of the percentage increase in income, which is necessary to induce the firms to adopt the new technology, to the elasticity of income to explanatory variables (incentives) reveals the amount of percentage change in incentive variables. Next, these percentage increases necessary to induce the desired change are converted into dollar amounts. The relationship between the induced change in income and the required amount of incentives is expressed as the "rate of return" of the incentive. Similarly, the variance-covariance matrix of estimated coefficients is employed to calculate the "standard deviations" and "correlation matrix" of incentives.

The financial theory of investments is based on the assumption that the essential characteristics of individual investment opportunities and portfolios are captured by information about their expected rate of return and the standard deviation of the return, i.e., the first two

moments of the rate of return on the investment. Accordingly, any individual project or combination of projects, i.e., portfolios, can be represented in the “mean-variance (or standard deviation)” space. Modern portfolio theory suggests that individual investment opportunities (or portfolios) can be combined into further portfolios. Depending on the risk-return characteristics of the original investment opportunities, i.e., portfolios, and their correlation structure, the investors will end up with a set of portfolios that cannot be dominated by any other combination of portfolios. This set of portfolios is termed as “the efficient frontier” (see Figure 4.2.1). In the absence of a risk-free investment opportunity, the optimal portfolio of choice for the investors will be determined by their “risk tolerance” and it will be a point that is located on the efficient frontier. If, on the other hand, a risk-free investment alternative is allowed in the model, then portfolio theory suggests that, irrespective of their attitude towards risk, it is in the best interest of all investors to hold a combination of the “market portfolio” (M in Figure 4.2.1) and the risk-free asset. The market portfolio is simply the point of tangency of a ray to the efficient frontier which originates from the point of risk-free rate of return (RF in Figure 4.2.1) on the y-axis which measures the return.

Figure 4.2.1 EFFICIENT FRONTIER AND THE MARKET PORTFOLIO

Once the incentives can be expressed in their first two moments, i.e., mean and variance, the standard Markowitz mean-variance portfolio analysis can be applied to the present context. Accordingly, one is able to determine: (i) the combination of incentives, which would require the least amount of risk-taking possible at the expense of a low return the minimum-variance

portfolio, i.e., the minimum-variance portfolio (MVP in Figure 4.2.1); (ii) the efficient frontier, i.e., the set of combinations of portfolios which stochastically dominate others in terms of their moments; and finally, (iii) given the rate of return on a risk-free investment opportunity (real or financial), one can also calculate the optimal combination of incentives to induce the desired outcome, i.e., point M in Figure 4.2.1. In other words, the portfolio theory provides a useful decision-making tool in determining the optimal mix of incentives.

The risk-free rate of return in the present context can be understood as the ratio of the increase in firms' revenues (in terms of dollars) per dollar of incentives offered. For example, if the government were to pay *directly* for the costs of technological switch, the amount of increase in firms' revenues (or reduction in their costs) would be exactly equal to the amount of transfer. Thus, in such an event, the risk-free rate of return is simply equal to zero. Clearly, based on familiar welfare- or profit-maximization motives, one may suggest that the risk-free rate of return, in our context, should be set at a positive rate by the government. On the other hand, if there are significant positive externalities attached to the project, even moderate negative returns may be acceptable from the point of welfare maximization. In this study, a conservative approach was taken and consider non-negative risk-free rates of return, e.g., 0%, 3% and 5% were considered.

4.2.4 Empirical Results

As suggested in the previous section, a stepwise regression equation is performed as a first step in order to determine the correct model to be estimated. For this purpose *all* of the available RHS (right-hand-side) candidates are entered into the model and are held subject to sequential elimination where P (significance level) is set at 0.20. Out of the entire set of variables the following succeeded to remain in the model: cost of goods sold (COGS), research and development expenses (RDEXP), fixed costs (FIXED), depreciation allowances (DEPR), taxes (TAX), and measure of operating leverage (MOL).

After determining the estimation model, the regression equation is estimated by the method developed by Newey and West (1987). The results of the estimation and the calculated elasticities are reported in Table 4.2.4. The estimated equation has an R^2 of 0.80 and an adjusted R^2 of 0.76.

Table 4.2.4 Results of the Regression Equation

Variables	Coefficients	Elasticity	Absolute t-statistics
COGS	-0.0638	-1.1731	3.08
RDEXP	-1.9589	-0.8002	4.14
FIXED	0.0368	0.35265	5.09
DEPR	0.3344	0.20273	2.58
TAX	-2.4695	-1.9847	4.88
MOL	-33923.0837	-	3.06
Constant	273296.7198	-	3.02

Next, the magnitude of costs is calculated which is necessary to induce the desired technological change for each of the boilers in the sample (some firms have multi-boiler sites) using the framework that was reported the Phase II Final Report (Miller et al., 2000). The results about boiler-size, required capital costs and operating and maintenance costs are summarized in Table 4.2.5.

Table 4.2.5 Boiler-Specific Costs

Company Name	Boiler Size	Capital Costs	O&M Costs	TOTAL
ACF INDUSTRIES INC	50.0	1,677,521	(175,516)	1,502,005
ALCAN ALUMINUM LTD	5.0	298,310	(12,329)	285,981
ALCOA INTERNATIONAL HOLDINGS COMPANY	60.0	1,923,330	(214,207)	1,709,123
ALCOA INTERNATIONAL HOLDINGS COMPANY	60.0	1,923,330	(214,207)	1,709,123
ALCOA INTERNATIONAL HOLDINGS COMPANY	40.0	1,419,010	(137,333)	1,281,677
ALCOA INTERNATIONAL HOLDINGS COMPANY	40.0	1,419,010	(137,333)	1,281,677
ALLEGHENY LUDLUM CORP	50.0	1,677,521	(175,516)	1,502,005
ALLEGHENY LUDLUM CORP	50.0	1,677,521	(175,516)	1,502,005
ALLEGHENY LUDLUM CORP	36.0	1,311,195	(122,236)	1,188,959
ALLIED SIGNAL INC	60.0	1,923,330	(214,207)	1,709,123
AMERICAN HOME PRODUCTS CORP	20.0	843,748	(63,297)	780,452
AMERICAN HOME PRODUCTS CORP	27.6	1,074,292	(90,953)	983,339
AMERICAN HOME PRODUCTS CORP	125.0	3,335,208	(473,134)	2,862,074
AMERICAN HOME PRODUCTS CORP	125.0	3,335,208	(473,134)	2,862,074
ANGELICA CORP	20.0	843,748	(63,297)	780,452
ARCO CHEMICAL COMPANY	27.6	1,074,292	(90,953)	983,339
ARCO CHEMICAL COMPANY	27.6	1,074,292	(90,953)	983,339
ARMCO INC	30.0	1,143,619	(99,825)	1,043,794
ARMCO INC	30.0	1,143,619	(99,825)	1,043,794
ASHLAND OIL INC	17.0	746,925	(52,613)	694,313
ASHLAND OIL INC	23.0	936,993	(74,124)	862,869
AT&T CORP	60.0	1,923,330	(214,207)	1,709,123
AT&T CORP	60.0	1,923,330	(214,207)	1,709,123
AT&T CORP	30.0	1,143,619	(99,825)	1,043,794
AT&T CORP	60.0	1,923,330	(214,207)	1,709,123
BEATRICE FOODS INC	24.0	967,384	(77,761)	889,623
BEATRICE FOODS INC	24.0	967,384	(77,761)	889,623

BETHLEHEM STEEL CORP	16.0	713,724	(49,088)	664,636
BETHLEHEM STEEL CORP	16.0	713,724	(49,088)	664,636
BETHLEHEM STEEL CORP	16.0	713,724	(49,088)	664,636
BETHLEHEM STEEL CORP	30.0	1,143,619	(99,825)	1,043,794
BETHLEHEM STEEL CORP	30.0	1,143,619	(99,825)	1,043,794
BETHLEHEM STEEL CORP	100.0	2,821,243	(372,384)	2,448,859
BETZ LABORATORIES INC	40.0	1,419,010	(137,333)	1,281,677
BORDEN INC	10.0	501,696	(28,456)	473,240
BORDEN INC	10.0	501,696	(28,456)	473,240
BORDEN INC	25.0	997,460	(81,410)	916,050
CABOT CORP	30.0	1,143,619	(99,825)	1,043,794
CABOT CORP	30.0	1,143,619	(99,825)	1,043,794
CARBON GRAPHITE GROUP INC	50.0	1,677,521	(175,516)	1,502,005
CARBON GRAPHITE GROUP INC	50.0	1,677,521	(175,516)	1,502,005
CATERPILLAR INC	40.0	1,419,010	(137,333)	1,281,677
CHEVRON CORP	25.0	997,460	(81,410)	916,050
CHEVRON CORP	25.0	997,460	(81,410)	916,050
CONSOLIDATED CIGAR CORP NEW JERSEY	150.0	3,823,921	(574,894)	3,249,027
CONSOLIDATED CIGAR CORP NEW JERSEY	150.0	3,823,921	(574,894)	3,249,027
EXXON CORP	35.0	1,283,783	(118,481)	1,165,302
EXXON CORP	35.0	1,283,783	(118,481)	1,165,302
GENCORP INC	60.0	1,923,330	(214,207)	1,709,123
GENCORP INC	60.0	1,923,330	(214,207)	1,709,123
GENERAL ELECTRIC COMPANY	8.5	444,124	(23,480)	420,645
GENERAL ELECTRIC COMPANY	150.0	3,823,921	(574,894)	3,249,027
GENERAL ELECTRIC COMPANY	150.0	3,823,921	(574,894)	3,249,027
GENERAL SIGNAL CORP	30.0	1,143,619	(99,825)	1,043,794
GENERAL SIGNAL CORP	30.0	1,143,619	(99,825)	1,043,794
GUILFORD MILLS INC	5.0	298,650	(12,353)	286,297
GUILFORD MILLS INC	12.0	575,210	(35,220)	539,990
GUILFORD MILLS INC	35.0	1,283,783	(118,481)	1,165,302
H J HEINZ COMPANY	44.0	1,524,158	(152,537)	1,371,621
H J HEINZ COMPANY	44.0	1,524,158	(152,537)	1,371,621
H J HEINZ COMPANY	40.0	1,419,010	(137,333)	1,281,677
H J HEINZ COMPANY	150.0	3,823,921	(574,894)	3,249,027
HANOVER FOODS CORP	34.0	1,256,174	(114,733)	1,141,441
HANOVER FOODS CORP	34.0	1,256,174	(114,733)	1,141,441
HERCULES INC	12.0	575,210	(35,220)	539,990
HERCULES INC	12.0	575,210	(35,220)	539,990
INDSPEC CHEMICAL CORP	40.0	1,419,010	(137,333)	1,281,677
INDSPEC CHEMICAL CORP	40.0	1,419,010	(137,333)	1,281,677
INDSPEC CHEMICAL CORP	50.0	1,677,521	(175,516)	1,502,005
INDSPEC CHEMICAL CORP	50.0	1,677,521	(175,516)	1,502,005
INDSPEC CHEMICAL CORP	100.0	2,821,243	(372,384)	2,448,859
INDSPEC CHEMICAL CORP	150.0	3,823,921	(574,894)	3,249,027
J&L SPECIALTY STEEL INC	40.0	1,419,010	(137,333)	1,281,677
KRAFT GENERAL FOODS INC	50.0	1,677,521	(175,516)	1,502,005
LTV STEEL CO INC	60.0	1,923,330	(214,207)	1,709,123
LTV STEEL CO INC	77.5	2,330,326	(282,843)	2,047,483
LTV STEEL CO INC	77.5	2,330,326	(282,843)	2,047,483
LTV STEEL CO INC	77.5	2,330,326	(282,843)	2,047,483
LTV STEEL CO INC	80.0	2,386,481	(292,728)	2,093,753
LTV STEEL CO INC	80.0	2,386,481	(292,728)	2,093,753
LUKENS INC	18.0	779,641	(56,156)	723,485
LUKENS INC	18.0	779,641	(56,156)	723,485
LUKENS INC	47.5	1,614,212	(165,918)	1,448,295

LUKENS INC	47.5	1,614,212	(165,918)	1,448,295
MACK TRUCKS	55.0	1,801,824	(194,806)	1,607,019
MACK TRUCKS INC	55.0	1,801,824	(194,806)	1,607,019
MACK TRUCKS INC	55.0	1,801,824	(194,806)	1,607,019
MACK TRUCKS INC	55.0	1,801,824	(194,806)	1,607,019
MASLAND CORP	75.0	2,273,717	(272,977)	2,000,740
MASLAND CORP	60.0	1,923,330	(214,207)	1,709,123
MERCK & COMPANY INC	80.0	2,386,481	(292,728)	2,093,753
MERCK & COMPANY INC	80.0	2,386,481	(292,728)	2,093,753
MERCK & COMPANY INC	80.0	2,386,481	(292,728)	2,093,753
MINNESOTA MINING & MANUFACTURING	45.0	1,550,065	(156,353)	1,393,712
NATIONAL GYPSUM CO	53.0	1,752,457	(187,076)	1,565,381
OCCIDENTAL PETROLEUM CORP	60.0	1,923,330	(214,207)	1,709,123
OCCIDENTAL PETROLEUM CORP	100.0	2,821,243	(372,384)	2,448,859
OCCIDENTAL PETROLEUM CORP	100.0	2,821,243	(372,384)	2,448,859
OCCIDENTAL PETROLEUM CORP	100.0	2,821,243	(372,384)	2,448,859
PPG INDUSTRIES INC	40.0	1,419,010	(137,333)	1,281,677
PPG INDUSTRIES INC	50.0	1,677,521	(175,516)	1,502,005
PPG INDUSTRIES INC	50.0	1,677,521	(175,516)	1,502,005
PPG INDUSTRIES INC	125.0	3,335,208	(473,134)	2,862,074
PROCTER & GAMBLE CO	99.0	2,800,057	(368,379)	2,431,678
PROCTER & GAMBLE CO	99.0	2,800,057	(368,379)	2,431,678
RHONE POULENC SA	50.0	1,677,521	(175,516)	1,502,005
RHONE POULENC SA	50.0	1,677,521	(175,516)	1,502,005
RHONE POULENC SA	50.0	1,677,521	(175,516)	1,502,005
RHONE POULENC SA	50.0	1,677,521	(175,516)	1,502,005
RHONE POULENC SA	50.0	1,677,521	(175,516)	1,502,005
RHONE POULENC SA	50.0	1,677,521	(175,516)	1,502,005
ROHM & HAAS COMPANY	30.0	1,143,619	(99,825)	1,043,794
ROHM & HAAS COMPANY	30.0	1,143,619	(99,825)	1,043,794
ROHM & HAAS COMPANY	30.0	1,143,619	(99,825)	1,043,794
ROHM & HAAS COMPANY	110.0	3,030,296	(412,545)	2,617,750
ROHM & HAAS COMPANY	110.0	3,030,296	(412,545)	2,617,750
ROHM & HAAS COMPANY	125.0	3,335,208	(473,134)	2,862,074
ROHM & HAAS COMPANY	125.0	3,335,208	(473,134)	2,862,074
ROHM & HAAS COMPANY	125.0	3,335,208	(473,134)	2,862,074
ROHM & HAAS COMPANY	125.0	3,335,208	(473,134)	2,862,074
ROHM & HAAS COMPANY	125.0	3,335,208	(473,134)	2,862,074
SEARS ROEBUCK & CO	40.0	1,419,010	(137,333)	1,281,677
SEARS ROEBUCK & CO	40.0	1,419,010	(137,333)	1,281,677
SEARS ROEBUCK & CO	40.0	1,419,010	(137,333)	1,281,677
SEARS ROEBUCK & CO	40.0	1,419,010	(137,333)	1,281,677
SMITHKLINE BEECHAM PLC	40.0	1,419,010	(137,333)	1,281,677
SONOCO PRODUCTS COMPANY	80.0	2,386,481	(292,728)	2,093,753
SPS TECHNOLOGIES INC	55.0	1,801,824	(194,806)	1,607,019
SPS TECHNOLOGIES INC	55.0	1,801,824	(194,806)	1,607,019
ST JOE PAPER CO	40.0	1,419,010	(137,333)	1,281,677
ST JOE PAPER CO	40.0	1,419,010	(137,333)	1,281,677
TARKETT INTERNATIONAL GMBH	75.0	2,273,717	(272,977)	2,000,740
TEMPLE INLAND INC	40.0	1,419,010	(137,333)	1,281,677
USX CORP	8.6	449,014	(23,891)	425,123
USX CORP	50.0	1,677,521	(175,516)	1,502,005
USX CORP	50.0	1,677,521	(175,516)	1,502,005
USX CORP	56.0	1,826,339	(198,677)	1,627,662
USX CORP	56.0	1,826,339	(198,677)	1,627,662
USX CORP	60.0	1,923,330	(214,207)	1,709,123
VALSPAR CORP	20.7	865,801	(65,811)	799,990

WARNER LAMBERT COMPANY	27.0	1,056,728	(88,744)	967,984
WARNER LAMBERT COMPANY	30.0	1,143,619	(99,825)	1,043,794
WARNER LAMBERT COMPANY	30.0	1,143,619	(99,825)	1,043,794
WEST PENN POWER COMPANY	32.0	1,200,336	(107,261)	1,093,075
WEST PENN POWER COMPANY	32.0	1,200,336	(107,261)	1,093,075
WESTINGHOUSE ELECTRIC CORP	10.3	514,809	(29,629)	485,179
WESTINGHOUSE ELECTRIC CORP	80.0	2,386,481	(292,728)	2,093,753
WESTINGHOUSE ELECTRIC CORP	20.0	843,748	(63,297)	780,452
WESTINGHOUSE ELECTRIC CORP	20.0	843,748	(63,297)	780,452
WESTINGHOUSE ELECTRIC CORP	20.0	843,748	(63,297)	780,452
WESTVACO CORP	85.0	2,497,495	(312,549)	2,184,946
WESTVACO CORP	85.0	2,497,495	(312,549)	2,184,946
WITCO CORP	50.0	1,677,521	(175,516)	1,502,005
WITCO CORP	50.0	1,677,521	(175,516)	1,502,005
WITCO CORP	60.0	1,923,330	(214,207)	1,709,123
WITCO CORP	60.0	1,923,330	(214,207)	1,709,123
WITCO CORP	60.0	1,923,330	(214,207)	1,709,123
WITCO CORP	65.0	2,042,328	(233,709)	1,808,619
WITCO CORP	65.0	2,042,328	(233,709)	1,808,619
WITCO CORP	75.0	2,273,717	(272,977)	2,000,740
WITCO CORP	90.0	2,606,888	(332,435)	2,274,453
YORK INTERNATIONAL CORP	30.0	1,143,619	(99,825)	1,043,794
YORK INTERNATIONAL CORP	46.0	1,575,828	(160,174)	1,415,654
			SUM =	\$ 249,600,082
			AVERAGE =	\$ 1,531,289

The firms in the sample have boilers of an average size of 54.4 million Btu/h and total size of 8,868 million Btu/h. The capital costs of retrofitting for all boilers is estimated to be \$281.4 million; the operation and maintenance cost(+) / benefit(-) ratio is estimated to be -\$31.8 million, which is due to fuel savings. Thus, the total cost of retrofitting all of the boilers in the sample amounts to \$249.6 million.

The boiler-specific cost figures are expressed as percentages of firms' net income. This represents the amount that needs to be induced by means of government incentives. The desired increase in firms' income combined with the calculated elasticities (taken from Table 4.2.4) - which measure the sensitivity of firms' net income to a 1% change in the value of incentives- to obtain the dollar amounts of each of the incentives required to generate the desired outcome. The findings are presented in Table 4.2.6.

Table 4.2.6 Factors and their Effectiveness

Incentive	Amount Needed	Standard Deviation	Rate of Return
Cogs (VAT)	\$ 3,912,226,000	2.1%	-93.62%
R&D Subsidy	\$ 127,418,400	47.4%	95.9%
Depr.Allowance	\$ 746,411,720	13.0%	-66.6%
Tax Cut	\$ 101,073,100	50.5%	+146.95%

The results of the stepwise regression estimation (given in Tables 4.2.4 and 4.2.6) show that the relationship between interest payments and firms' net incomes are not statistically significant. This result indicates that interest rate-related government incentives, such a reduced or zero-interest loan, will not necessarily induce the desired increase in net income needed for technology adoption.

On the other hand, it is observed that the net income of firms in the sample are responsive to fluctuations in the cost of goods sold, the amount paid for research and development expenses, depreciation allowances, and taxes. Table 4.2.6 presents the estimated dollar amounts of change needed to induce the desired increase in firms' income, i.e., \$249.6 million, to induce them to undertake the technological investment at each and every boiler site in the sample. Accordingly, declines in the cost of goods sold, research and development expenses, and tax burdens by 3,912.2, 127.4 and 101.1 million dollars, respectively, and an increase in their depreciation allowances by \$746.4 million will induce an increase in their profits by the targeted amount.

In Table 4.2.6, the necessary changes are also expressed in terms of "rate of return" in order to obtain an understanding about their effectiveness. The "rate of return" is achieved by taking the ratio of the target change in income to the required change in the aforementioned factors and subtracting one from it. Clearly, this approach assumes that there are no externalities (one way or another) attached with any of the incentives under consideration and hence, the "return" of the incentive is measured as the monetary increase in firms' profits. The results are striking. It turns out that only tax cuts and reductions in research and development expenses provide positive returns. Magnitudes of required changes in cost of goods sold and depreciation allowances exceed the 'benefits', i.e., the increase in firms' profits by a significant amount.

If one were to choose only one investment incentive, the best candidates appear to be tax cuts or subsidies that are specifically aimed at covering research and development expenses. The empirical findings indicate that interest-related incentives, such as reduced or no-interest loans, will not generate the desired outcome for the firms studied. On the other hand, an incentive that will reduce the cost of goods sold or to increase the depreciation allowance is statistically significant on firms' profits, and thus present feasible alternatives. However, analyzing their effectiveness yields the result that they are inferior when compared with the other two alternatives: tax cuts and subsidies extended for research and development.

Nevertheless, someone who is familiar with the benefits of portfolio analysis may wonder whether a combination of these alternatives can yield a superior result. To explore possible benefits from portfolio analysis the reported rates of return (see Table 4.2.7), their volatilities (which are based on the standard errors of estimated coefficients), and the correlation matrix of estimated parameters are utilized to perform a standard, Markowitz type, portfolio analysis.

Table 4.2.7 The Correlation Matrix of Estimated Coefficients

	RDEXP	DEPR	TAX
VAT(Subsidy)	0.08	-0.28	0.85
R&D Subsidy	1.00	-0.22	0.22
Depr. Allowance	-0.22	1.00	0.08
Tax Cut	0.22	0.08	1.00

In the case of depreciation allowance and taxes, it is relatively obvious how they are linked to a government incentive scheme. An increase in depreciation allowances or a decrease in tax rates works as an incentive for firms. The link of cost of goods sold and research and development taxes to government incentives of is less obvious. The government can affect the cost of goods sold by means of a subsidy/tax that depends on the sales volume. A good example for this scheme is the value-added tax (VAT). An incentive, in the current context, can simply be thought as a negative VAT. Similarly, an incentive through research and development expenses can be imagined as a specific government subsidy that is meant to cover a portion or the entire amount of firm's research and development expenses.

One may argue that introduction of a (negative) VAT-like schedule would require institutional changes, and thus, in the short- to middle-run its feasibility is questionable. Recognizing this potential criticism, the portfolios were stimulated with and without the COGS (cost of goods sold) variable to account for both possible states of the world.

Portfolio return and volatility calculations are performed assuming a wide range of rate of return that varies between 0 and 100%. In addition, the characteristics of the minimum variance portfolio are calculated for both sets of portfolios. Furthermore, the tangent of the straight line (ray) originating from the locus of the risk-free return, i.e., the market rate of return, is calculated, as well. According to the theory this is the portfolio that investors of all types of risk

aversion would hold to maximize their expected utility. The results of the portfolio simulations with COGS variable and without it are presented in Tables 4.2.8 and 4.2.9, respectively.

Table 4.2.8 Risk and Return of Various Portfolios (with COGS)

MVP	1	2	3	4	5	6	7	8	9	10	11
VAT	96.2	7	4.2	2.4	-2.1	-11.1	-20.1	-29.1	-38.2	-60.7	-83.3
R&D	1	19.8	20.3	20.7	21.7	23.6	25.5	27.4	29.3	34	38.8
DEPR	6.5	56.3	57.8	58.8	61.3	66.3	71.3	76.4	81.4	94	106.5
TAX	-3.7	17	17.7	18.1	19.1	21.2	23.3	25.4	27.5	32.7	38
Return	-98.9	0	3	5	10	20	30	40	50	75	100
Risk	0.6	15.2	15.7	16	16.8	18.3	19.8	21.4	22.9	26.8	30.6

Table 4.2.9 Risk and Return of Various Portfolios (without COGS)

MVP	1	2	3	4	5	6	7	8	9	10	11
R&D	11.1	19.7	20.3	20.7	21.6	23.5	25.4	27.3	29.1	33.8	38.5
DEPR	87.8	64.1	62.6	61.5	59.0	53.8	48.7	43.6	38.4	25.6	12.8
TAX	1.1	16.1	17.1	17.8	19.4	22.7	25.9	29.2	32.4	40.6	48.7
Return	-46.1	0	3	5	10	20	30	40	50	75	100
Risk	11.5	15.3	15.7	16	16.7	18.4	20.1	21.9	23.8	28.6	33.6

Tables 4.2.8 and 4.2.9 display the analysis of eleven portfolios based on their return and risk characteristics. The MVP portfolio, in both tables, refers to the minimum-variance portfolio, i.e., the combination of risky assets that generate the portfolio with lowest possible level of risk. Portfolios 1-10 stand for portfolios with expected rates of returns varying between 0% (Portfolio 1) and 100% (Portfolio 10). All of the entries are reported as percentages and the figures next to the variables simply indicate their portfolio weights. For example, according to Table 4.2.8, in order to establish an incentive portfolio which generates a rate of return of 0% (Portfolio 1), the value added tax (subsidy) should have a weight of 7%, research and development subsidies should have a weight of 19.8%, depreciation allowances 56.3% and finally tax incentives 17%. This represents a particular mix of risky incentives, which are combined into a portfolio which offers exactly a 0% rate of return, just as a direct subsidy would which is offered to the firm which reimburses all of the costs (of the switch) directly. A comparison of the results presented in the two tables yields that the risk-return relationship for portfolios 1-10 is not significantly different from each other, i.e., for a given rate of return the portfolio volatility seems to be very similar.

In Table 4.2.8, it is observed that the weights of research and development subsidies and taxes increase as the resulting portfolios allow for more risk. The risk is controlled by the combination of VAT (subsidy) and depreciation allowances. It is interesting to note that after a certain level of risk the weight of VAT (subsidy) becomes negative, i.e., it suggests that after a certain level of risk it is beneficial from a portfolio analysis point of view for the government to introduce a positive value-added tax, which is to be compensated by the increase of depreciation allowances.

In Table 4.2.9, which excludes the cost of goods (COGS) variable, it is observed that as the expected rate of return increases the research and development subsidies and tax incentives move in opposite directions, i.e., the weight of taxes increases and the other one decreases. Depreciation allowances, which counterbalances the other two in the degree of risk, loses its importance in the portfolio greater risk is allowed, whereas, its weight becomes larger for portfolios which are exposed to a lesser degree of uncertainty.

Having calculated the risk-return relationships and the weights of each incentive in different portfolio mixes, the next relevant question to focus on is the location of the “market portfolio”. In the presence of a risk-free rate of return, the risk-free rate and the market portfolio span the so-called “capital-market line”, which is the locus of efficient and dominant portfolios in the market; accordingly any position on the capital-market line can be achieved by combining the risky market portfolio and the risk-free security.

Under the presumption that the government can with 100% probability achieve its goal of spreading the commercial usage of the new technology by extending direct subsidies, which cover all of the required switching costs to firms which are willing to undertake the necessary technological investment, the direct subsidy incentive can be regarded as a no-risk, i.e., *risk-free* incentive alternative. Hence, direct subsidies are introduced to the analysis as the risk-free security and solve for the point of tangency to find the market portfolio. In both scenarios, i.e., with and without the COGS variable (Tables 4.2.8 and 4.2.9, respectively), the optimal tangency portfolio coincides with a portfolio which is formed 100% by the tax incentive.

Furthermore, in order to account for values which are not accounted for by the standard discounted cash flow analysis, e.g., externalities and option values, the rate of return for the risk-free incentive for -3% and -5% is set and the optimal market portfolio is solved under these

assumptions. The results, once again depict the 100% tax incentive portfolio as the optimal market portfolio.

In other words, provided that the direct subsidies can be viewed as risk-free, i.e., the government is 100% confident that it can initiate any marginal technological investment behavior in the industry by undertaking all of the necessary cost by its own if it chooses to do so, it is in government's, and all taxpayers' best interest to offer government incentives in the form of tax cuts, only. Furthermore, even if the risk-free incentive, i.e., direct subsidies generate a subzero rate of return, due to externalities, or some other reasons, the result of the analysis remain unaffected and the 100% tax incentive emerges as the optimal portfolio incentive. In sum, the preceding analysis clearly shows that the tax incentives are the most effective incentives the government can offer to induce increases in firms' profits and thus to induce them to adopt the desired technological changes.

4.2.5 Conclusion

The initial premise of the study was that there are social, political and economic benefits (in market and non-market value) that can be gained by the widespread adoption and commercialization of the new coal-based technology. The adoption of the new technology will most likely be achieved through measures that reduce capital costs and not through serendipitous changes in oil price alone. Thus, the findings can be interpreted as if to indicate that even though the adoption of the new technology may have macroeconomic advantages for the economy, additional incentives may be needed to induce such an adoption at the microeconomic level, i.e., at the firm level.

Economic and financial incentives have become accepted tools for the implementation for preferred government policies. Typical examples range from investment tax credits and reduced or zero-interest loans to capital subsidies and direct and indirect (such as accelerated depreciation) tax incentives. The existing studies on the effectiveness of policy incentives, on average, remain highly macroeconomics oriented and have little to say about the consequences in a given sector in the economy.

This study focused on the microeconomic responses generated by the firms to the incentives. Firm-specific data were examined that have a potential use for the new technology. The optimal incentive and policy mixes determined are based on empirical estimation of firms'

reactions to various factors. It needs to be emphasized that one incentive may be optimal with projects with characteristic “A”, whereas another may be more suitable if the project has characteristic “B”. For example, tax incentives can be thought to be more effective on projects which require a big initial capital outlay, whereas, reduced or zero-loan incentives may prove to be more useful in the case of projects which are highly sensitive to interest rate fluctuations in the market.

The adoption of the new technology can be viewed as an “investment” in the new technology and thus, tools of modern finance and portfolio theory can be applied to the problem at hand. First, the sensitivity of firms’ net income to various factors is estimated. The required increases in a firms’ income is estimated by adopting a framework in the Phase II Final Report. It turns out that the cost of technology adoption (net of fuel savings benefits) is \$249.6 million for all 163 boilers in the sample (or \$1.5 million, on average). Consequently, based on the sensitivities (expressed in terms of percentage responses) to induce a 1% change in income, and the required change in income, the dollar amounts a set of incentives are calculated which are needed to generate the desired outcome. Accordingly, a decline of \$3,912.2 million in the cost of goods sold, \$127.4 million in research and development expenses, \$101.1 million in firms’ tax burden, or an increase of \$746.4 million in firms’ depreciation allowances, induce the desired increase in the net income of the firms in the sample.

Recognizing that incentives, on average, do not induce the desired outcome with 100% certainty, the benefits of offering the incentives as a portfolio rather than individually are investigated. The effectiveness of incentives is expressed in terms of their rate of return and risk. Consequently, the composition of optimal portfolios is calculated for several rates of return ranging from 0-100% and for the minimum risk portfolio. The results are estimated for two different cases: with and without cost-of-goods (COGS) as an incentive variable. Noting that the government can affect the cost of goods sold by imposing a value-added tax or subsidy, one may argue that COGS can be viewed as a possible incentive variable. Table 4.2.8 presents the optimal portfolios for this case. On the other hand, the introduction of a value-added tax/subsidy scheme can be viewed as technically difficult in the short to medium-term. Thus, alternatively, another set of optimal portfolios is calculated for the same set of returns, excluding the COGS variable (Table 4.2.9).

It appears that the risk-return relationship is not significantly altered by the inclusion/exclusion of that variable. Tables 4.2.8 and 4.2.9 show that for low risk-low return portfolios depreciation incentive has a significant weight in both portfolios and that its weight decreases as the portfolios become riskier. In contrast, tax and research and development expense incentives appear to have lower weight for low-risk portfolios, which increases gradually as portfolios are aimed at higher return. In addition, an interesting finding of the analysis is that the firms in the sample are not sensitive to interest-rate based incentives, such as a reduced or zero-rate loan.

Finally, given a risk-free investment opportunity in the economy the ‘market portfolio’ can be estimated. In theory this is the only risky portfolio, which will be held by all investors, irrespective of their attitude towards risk. If one makes the assumption that direct capital subsidies that cover the full cost of the technological investment are literally risk-free in achieving the desired changes, it can be incorporated in the model to estimate the market portfolio that dominates all other risky portfolios in the existence of a risk-free investment opportunity. The direct capital subsidies, by definition, offer a rate of return of 0%. However, to account for externalities and so forth which are not incorporated in the analysis directly, two negative rates of return of -3% and -5% are allowed for on the risk-free alternative. Notably, the point of tangency in all three cases happens to be a portfolio that consists of only the tax incentive, i.e., the weight of the tax incentive is 100%. This result shows that in the existence of a no-risk opportunity, the tax incentive is the best alternative to offer to the industry to induce an increase in their profits.

It should be added that a precise answer about the *best* incentive portfolio varies depending on the shape of government’s indifference curves, which depict information about how the risk-return tradeoff is viewed by the government at various levels of returns. The *best* portfolio is determined simply at the point of tangency between the indifference curves and the efficient frontier that is formed by the combination of the risk-free investment alternative and the market portfolio (or just by the locus of efficient risky portfolios in the absence of a risk-free alternative). Thus, for the selection of the best suitable incentive mix, the findings and results provided by this study will provide valuable insights to the policy makers.

4.3 Community Sensitivity to Coal Fuel Usage: Economic Valuation of Risk Perceptions: Measuring Public Perceptions and Welfare Impacts of Electric Power Facilities

4.3.1 Introduction

This activity developed methods that integrate economic valuation with the techniques used in psychology to characterize risk perceptions to value the welfare impacts due to the presence of energy production facilities. A contingent valuation survey, designed with cognitive survey design methods, was administered to elicit quantitative information regarding individuals' perceptions of the risks associated with fossil fuel-based electric power facilities and, the individuals' willingness to pay to prevent or change risk exposure levels. The quantitative measures of risk perceptions are related to the willingness to pay values using maximum likelihood estimation.

The underlying conceptual rationale for valuing changes in perceived risk combines findings from the risk perception literature with expected utility theory. Using an economic model of individual willingness to pay to avoid risks, this study identifies factors that contribute to individual willingness to prevent energy production facilities. Specific focus is placed on developing quantitative measures of perceived risk that can be utilized to derive welfare changes induced by such facilities.

This research measures the individual's *ex ante* marginal willingness to pay to prevent or change their (perceived) risk exposure level from an electric power facility, determined, in part, by the perceived risk attributes. Obtaining the value of individual preferences of risk levels can assist in facility siting decisions by measuring how much individuals will pay to influence decisions or to what extent they will willingly bear the costs of a more expensive, but more desirable fuel.

Results show that welfare impacts, as measured by option price, depend on an individual's perceptions of the health, environmental, aesthetic, and economic impacts as well as their socio-demographic characteristics. Individuals seemed to have difficulty distinguishing between the probability and the severity of a risk in the manner suggested by the definition of risk, although the survey instrument may have been unable to capture the difference. Perceived environmental, health, and aesthetic impacts play a larger role than potential economic impacts

in determining option prices, explaining in part why residents may oppose a facility even when it will likely bring economic benefits to an area.

Risk has long been an important focus in psychology. According to Slovic (1987), a professor of psychology, "[t]he ability to sense and avoid harmful environmental conditions is necessary for the survival of all living organisms." Economists also recognize that individuals place an implicit value on risk, but have only recently begun to incorporate such values in the public decision-making process. This study integrates the economics of welfare measurement with the psychological characterization of risk perceptions. Using methods from cognitive psychology (Jabine, et al., 1984), a contingent valuation survey is developed to elicit quantitative information about individuals' perceptions of the risks associated with locally sited fossil-fuel electric power facilities, including utility generators, and independent power producers. Psychometric scales are used to quantify risk perceptions, which are then related to the option price for eliminating or reducing risks. This research extends prior work from the psychology literature by focusing on a specific hazard scenario rather than general "risk" and by examining perceptions of different components making up a risk. The study estimates individuals' marginal willingness to pay (WTP) as a function of these risk perceptions as well as sociodemographic characteristics conventionally modeled as determinants of WTP. Empirically, the model considers multiple perceived risk components identified through focus groups and cognitive interviews. An implicit factor analysis identified four risk components: health risks, aesthetic risks, environmental risks, and beneficial or detrimental economic changes resulting from facility construction and operation.

The field of risk assessment has only recently been extended into economic valuation, so that values can be assigned to the welfare changes resulting from changes in risk exposure. An industrial facility such as an electric power plant can expose a community to a wide variety of hazards, including environmental and health hazards, amenity or aesthetic effects, and changes in traffic patterns and congestion. In addition, their construction and operation can induce economic impacts in the host community such as changes in employment, price levels, and tax revenues. An associated probability distribution characterizes the likelihood that each consequence due to a facility will actually occur. Individuals make behavioral decisions based upon their perceptions of the probability and severity of impacts from such facilities. In this manner, welfare is based upon each individual's perception of risk.

Regulatory changes in the 1980s and 1990s have prompted a significant increase in the location of small energy generation facilities close to residential communities. The Energy Information Agency (U.S. Department of Energy, 1993) predicts a 15% increase in new coal-fired generation capability (170 250-MW plants) and a 31% increase in oil and natural gas-fired generation capability (925 100-MW facilities) by the year 2010. Some proposed facilities have been vehemently opposed by local residents, regardless of experts' attempts to communicate that the benefits could significantly outweigh any associated risks. Such opposition to otherwise economically efficient projects may arise, in part, because layperson or non-expert risk perceptions related to energy generation facilities differ from those of "experts."

Risk assessments by experts are likely to differ from risk assessments by ordinary citizens (Fisher 1991). Laypeople often react to potential hazards or undesirable facilities in a way that is considered disproportionate to the risks involved. Experts have made substantial progress in identifying and measuring "objective" risks but understand less well how individuals perceive risks and how they value changes in risk. Because an individual's behavior and values are based on risk perceptions, a better understanding of risk perceptions and how they influence individual values needs to be developed. Improved understanding of risk perceptions will provide a basis for understanding and anticipating responses to hazards and undesirable facilities and aid in designing risk communication programs to increase understanding of such facilities.

Results from this study show that individuals' willingness to pay for risk prevention or reduction is a function of their perceptions of the health, environmental, aesthetic, and economic impacts as well as their socio-demographic characteristics. Individuals appear unable to distinguish between the probability and the severity of a risk in the manner suggested by the definition of the risk. Perceived environmental, health, and aesthetic impacts play a larger role in determining option prices than potential economic impacts, explaining in part why residents may oppose a facility even when it will likely bring economic benefits to an area.

4.3.2 Background And Previous Research

4.3.2.1 Risk Perceptions

Associated with any hazard are a variety of consequences. Risk is a quantitative measure of the likelihood and severity of those consequences, usually expressed in terms of conditional probabilities or other technical and quantitative measures. Scientists and engineers must assign

probabilities to the occurrence of hazardous events in order for risk mitigation policy decisions to be made. Expert risk assessments are based upon quantitative and technical data that are often not readily understood by the general public. Perceived risk, though, is a function of the actual impacts of a hazard in addition to many unquantifiable and unique cognitive dimensions that are derived from personal experiences and preferences (Bostrom et al., 1992).

Risk perception research entails understanding, from a layperson's point of view, what is known about a hazard, what is thought to cause it, and its perceived impacts. The objective is to learn what people know about a particular hazard and how they incorporate risk information into their personal knowledge set or "mental model." Risk perceptions are then defined as a function of the "true" risk and the differential between layperson and expert mental models. Past research has revealed that laypersons rank as most serious risks that are catastrophic, involuntary, unfamiliar, dreadful, uncontrollable, or having an uncertain and inequitable distribution of consequences. In general, laypersons tend to have greater concern for and overestimate "small" risks and less concern for and underestimate "large" risks (Slovic et al., 1985; Slovic 1987; Covello et al., 1993).

Slovic (1987) mapped the perceived risk of 81 hazards over the factors unknown risk and dread risk. Non-nuclear power generation, coal combustion, and fossil fuels appear in the upper-right quadrant of the perceived risk mapping, that includes risks characterized as "unobservable," "unknown to those exposed," "delayed effect," "new risk unknown to science," "uncontrollable," "catastrophic," "inequitable," "not easily reduced," "of high risk to future generations," and "involuntary." Risks located in this quadrant can be thought of as those most difficult to mitigate through regulatory channels and presenting the greatest challenges to risk communicators and mitigators.

Cognitive dissonance explains systematic differences in the interpretation of information, as well as in individuals' receptivity to new information according to preferences and beliefs (Akerlof and Dickens, 1982). Cognitive dissonance theory has also been useful in explaining why a differential between lay and expert risk judgments persists and why individuals may not believe risk information provided by experts. Studies have revealed that risk communication is more effective and more likely to be understood by the targeted public when lay risk perceptions are identified *a priori* (Fisher et al., 1991). Effective risk communication requires that the

relevant risk information be presented so that it is most likely understood by the public in the manner in which it was intended.

The study and characterization of risk perceptions are an established precursor to developing effective risk communication. More recently, the field of risk assessment has been extended into economic valuation of the welfare changes resulting from changes in risk exposure. Determining the welfare impacts of undesirable land uses should include the value individuals place on perceived changes in risk. This study measures laypersons' *ex ante* WTP (or option price) to decrease or completely eliminate a particular risk exposure level, determined in part by the perceived attributes of a risk.

4.3.2.2 Valuation and Perceived Risks

Slovic's (1987) characterizations of perceived risk according to the degree of voluntariness, immediacy of effect, familiarity, controllability, likelihood of catastrophic consequences, dread, and severity of consequences have provided a solid foundation for many subsequent studies. Slovic et al. (1985) found the following: (i) different groups, including laypersons and experts, have very different attitudes towards risks, (ii) experts tend to rate risks according to annual fatalities or other technical and quantitative measures, (iii) laypersons rate risks on different criteria than experts, and (iv) laypersons tend to want stricter regulation of the hazards they perceive as most risky.

Empirical studies have tended to value welfare shifts induced by changes in risk for a range of hazards. A few studies have developed a methodology to evaluate changes in the risk of a specific hazard. Although the studies differ in many regards, all combine a quantitative component, such as Slovic's psychometric scales, with a contingent valuation instrument (CVM) to elicit willingness to pay for changes in the risk levels of one or more hazards. The conceptual framework underlying these studies assumes that an individual's utility is a function of socioeconomic characteristics and some variant of risk attributes or risk levels. Applying the theory of cognitive dissonance to economic theory, Akerlof and Dickens (1982) incorporated a subjective assessment of risk into an economic valuation of hazardous jobs in the labor market. Römer and Pommerehne (1994) developed a contingent valuation instrument to elicit WTP for the reduction of hazardous waste risk in West Berlin; they consider private averting activities and strategic behavior. Savage (1993) evaluated risk judgments and their influence on relative WTP

values to reduce the exposure levels of four "cognitively different" risks. The study found that people are most likely to have different values for reducing different types of risks.

McDaniels et al. (1992) model a household's option price for decreased risk exposure ("safety") as a function of the household's socioeconomic characteristics and perceived attributes of each of ten common risks. The perceived characteristics component consists of a household's familiarity with the risk, perceived exposure, and the degree of dread associated with each risk, which are essentially the factor composites Slovic derived in his 1987 study (Slovic 1987). A CVM instrument using psychometric scales and open-ended valuation questions elicited respondent's willingness to pay for decreases in risk exposure. Their analysis suggests that perceived risk characteristics, perceived exposure levels, age, and income all significantly affect an individual's valuation for a reduction in risk across a range of hazards. In their model with only well-defined risks, personal exposure to the risk was an important determinant, while dread and severity were important factors for the less-defined risks. McDaniels et al. (1992) consider general risks such as air safety rather than specific risks, and do not separate risk into components such as health and environment.

Smith and Desvousges (1987) consider hazardous wastes to explore how marginal valuations of risk changes vary with the size of the baseline risk and the direction of the risk change (i.e., either a decrease or increase in risk levels). Individual risk judgments are related to WTP for changes in perceived risk exposure, income, and other socio-economic characteristics. Marginal valuations to *avoid risk increases* declined with increases in the risk level; the mean valuations to *reduce baseline risk* were greater than the valuation to *avoid risk increases*. The results could imply that valuations to reduce the risk imply a different property right than valuations to avoid a risk increase. Valuations to reduce risk and to avoid a risk increase are essentially measuring two different types of changes in utility.

Psychological methods to characterize perceived risk have contributed to improved risk communication methods and provided a solid foundation for risk valuation studies. Studies based upon cognitive psychology have found that a differential exists between expert and non-expert risk judgments and that risk attributes influence risk perceptions. Economic models that incorporate perceived risk have shown that income and perceived risk are strong determinants of WTP to reduce/change risk exposure. In addition, research has found that the stated baseline risk level may be a determinant of WTP values, and that cognitively different risks are likely to

generate different values for reducing risk exposure. However, nearly all studies in this area exclude non-health risks and focus on how hazards influence values to change personal exposure to risks. This study extends previous studies to include risks to the environment, the economy, and aesthetics, in addition to human health.

4.3.3 Theory

Expected utility theory is based upon “objective” or true probabilities and the certainty of future states of the world (von Neumann and Morgenstern, 1944). Savage (1954) and Anscombe and Aumann (1963) extended expected utility theory by incorporating subjective probabilities into the expected utility model. Subjective probability is extensively discussed in the economics literature, though little is said about the perceptions of the magnitude or severity of risk consequences. This research assumes that individuals make choices within an expected utility framework according to their perceptions of the probabilities and the consequences; the expected utility model is naturally inclusive of this concept.

The contingent valuation survey used for this analysis elicits individuals’ WTP to avoid or reduce exposure risks due to a facility that burns either coal or natural gas to generate electricity. Individuals are uncertain with regard to the likelihood and severity of risk consequences as a result of a particular facility or fuel type. The risks of such a facility can be disaggregated into probability and severity characteristics to emphasize their individual effects on option price.

Let EN measure the environmental consequences of an adverse event and X the bundle of consumption goods over which the individual maximizes utility. Utility (U) is a state-dependent function of environmental quality and consumption,

$$U = U(X, EN), \quad \text{where} \quad \frac{\partial U}{\partial X} > 0, \quad \frac{\partial U}{\partial EN} < 0. \quad (4.3.1)$$

Maximizing utility subject to a budget constraint, with I denoting income and P denoting a price vector, results in the *ex post* indirect utility function, $u = v(I, P, EN)$.

Consistent with a framework where policy decisions are made and associated welfare impacts measured, individuals are assumed to make *ex ante* decisions when facing a risk. An individual may be willing to pay, *ex ante*, to reduce or eliminate a risk, regardless of which state

of the world is realized, *ex post*. An *ex ante* payment made independent of the future outcome is referred to as a state-independent payment or an option price (*OP*).

Let p represent the probability of an adverse event occurring and $1-p$ the probability of the event not occurring. If $EN = 0$, the event has no negative environmental consequence. The level of a negative environmental impact associated with the event is measured by $EN = EN^*$. The corresponding option price for reducing EN^* to zero is the solution to

$$\begin{aligned} (\pi)v(I, P, EN^*) + (1 - \pi)v(I, P, 0) = \\ (\pi)v(I - OP, P, 0) + (1 - \pi)v(I - OP, P, 0) = v(I - OP, P, 0). \end{aligned} \quad (4.3.2)$$

If $p=0$, the event does not occur. The corresponding option price for reducing p to zero is the solution to

$$\begin{aligned} (\pi)v(I, P, EN^*) + (1 - \pi)v(I, P, 0) = \\ (0)v(I, P, EN^*) + (1 - 0)v(I - OP, P, 0) = v(I - OP, P, 0). \end{aligned} \quad (4.3.3)$$

The option price defined in equation 4.3.2 reduces the severity of the adverse impact to zero. Freeman (1993) defines this as *risk reduction*. In equation 4.3.3, the probability of the risky event occurring (π) is reduced to zero, which is termed *risk prevention*. In theory, the option price is identical for either a reduction of EN or π to zero. As the utility of expected value and expected utility generally differ, marginal changes in severity are likely valued differently than marginal probability changes which result in equivalent changes in expected value. For this reason, decomposing risk into severity and probability is worthwhile for measuring welfare impacts of risk changes.

Assume individuals facing the construction of an electric power producing plant perceive n possible impacts, EN_i , and associated probabilities, p_i , where $i = 1, \dots, n$. Individuals do not know, *ex ante*, the true severity or the true probability of risk consequences due to the plant and must rely upon their perceptions of future states of the world to make utility maximizing decisions. Expected utility takes the form

$$E[U] = \sum_{i=1}^n (\pi_i | y) U[X, EN_i | y] \quad \text{where} \quad \sum_{i=1}^n \pi_i = 1 \quad (4.3.4)$$

Individuals may conceivably state perceived probabilities that sum to greater than or less than one. This is a relevant topic in dealing with perceptions based on incomplete information, but is not considered here. Judgments of the probabilities, $\pi_i|y$, and the severity of the risk, $EN_i|y$, are likely to be conditional upon y , the individual's experience, familiarity, and knowledge of the risk and its consequences. Previous studies have found an individual more familiar with hazard consequences is more likely to perceive a higher probability of being exposed to the hazard (Slovic 1987). This does not imply that individuals with pre-existing knowledge or familiarity will always perceive risk to be greater than individuals without previous experience. Perceived severity may be less for individuals with prior experience than those without. The influence of prior experience and knowledge on perceived risk and valuation of risk changes is not explicitly addressed in this study, although such issues were addressed in a preliminary manner in a debriefing questionnaire.

Using the indirect utility representation, $\bar{V}$ is the maximum expected utility given market prices, income, and perceived probability and severity of the risks created by the power plant:

$$V = \sum_{i=1}^n (\pi_i|y) v[P_x, I, EN_i|y]. \quad (4.3.5)$$

The option price (or *ex ante* WTP) to prevent the power plant from being built or to reduce its environmental impacts to zero, even when operating, will be

$$\bar{V} = \sum_{i=1}^n (\pi_i|y) v[P_x, I, EN_i|y] = v(P_x, I - OP, 0). \quad (4.3.6)$$

Solving for OP yields the individual's value for the perceived level of environmental risk of the power plant, holding prices and income constant. The individual's option price, or willingness to pay as elicited from a contingent valuation survey, to change or eliminate perceived risks due to an electric power producer will be a function of prices P_x , income I , and the individual's perceptions of the probabilities and severity of the risk consequences, conditional on the individual's knowledge and experience with the risk:

$$OP = OP(P_x, I, \pi_i|y, EN_i|y). \quad (4.3.7)$$

In estimation, the model will be expanded to include multiple perceived risk components, including health, aesthetic, and environmental risks, and economic impacts resulting from facility construction and operation.

4.3.4 Survey Design

Deriving the values individuals place on risk level changes requires the integration of methods for characterizing perceived risk with contingent valuation methods. Advances in cognitive psychology, survey design, contingent valuation, and econometric techniques provide an opportunity to analyze the relationship between an individual's risk perceptions and the value corresponding to changes in risk.

Contingent valuation is considered the most appropriate method for non-market valuation of public goods (Mitchell and Carson, 1989) as well as the only method available for measuring non-use values (Freeman 1993). Contingent valuation enables researchers to create a surrogate market, where subjects reveal their values for incremental increases or decreases in the provision of a non-market good. Contingent Valuation Method (CVM) estimates depend upon the researchers' representation of a hypothetical market and are vulnerable to sources of measurement error and the survey's reliability to elicit valid responses. These derived values depend on the entire process of designing, implementing, and analyzing CVM survey instruments.

Many aspects of CVM surveys create cognitive challenges. For example, the survey might involve fairly technical information beyond the respondents' understanding, leading to problems at the comprehension stage. Cognitive psychology methods, including focus groups and verbal protocols, provide a potential solution to some of the shortcomings of CVM survey design. Complete elimination of all measurement error is impossible, but a systematic approach to survey design with a goal of minimizing measurement error helps. This systematic approach is applicable regardless of the mode of survey administration or the anticipated method of data analysis.

The CVM survey used in this analysis was designed to elicit quantitative measures of perceived risk and the payments that individuals would be willing to make to change the perceived risk levels of an electric power facility. To explore whether perceived risk can be

defined as the product of perceived severity and perceived probability (as risk is defined as severity times probability), measures of both perceived severity and perceived probability were necessary. Other measures of perceived risk are also considered. A copy of the survey is given in Appendix 4D.

Ten focus groups and twenty-one cognitive interviews were conducted from June 1995 through April 1996 with subjects recruited through local newspapers. These sessions were intended for survey development, not primary data collection, and the sample participants were not considered representative of the population. The objectives were as follows: (i) to gather insights into how individuals think about the relationships between the combustion of fossil fuels, electric power facilities, their community, their health, and the environment; (ii) to identify lay terminology for technical aspects of energy production; (iii) to identify lay perceptions of both the risks and benefits of energy production and the presence of an energy production facility; and (iv) to determine what information individuals use to form their perceptions of such facilities.

Discussion during the exploratory sessions revealed that individuals may not associate their own use of electricity with the demand for fossil fuels by electric power facilities. Scenario rejection is likely if individuals do not understand the need for and intended purpose of a facility. A hypothetical facility within three miles of a residential area raised numerous concerns, including health, environmental, economic, aesthetic, land use, intergenerational, and equity concerns. Concern about potential health impacts and air pollution from a facility seemed to be greater than concern about other impacts such as noise, traffic, and economic impacts.

Four primary survey design issues were addressed throughout survey development: implicit factor analysis to determine risk categories, scale format, survey vocabulary, and overcoming survey bias. A two-step process served as an implicit factor analysis to derive composite risk categories. During focus groups and interviews, participants were asked what they thought the future risks might be from a proposed electric power facility three to five miles from their house. Participants then categorized the list into four main groups, as shown in Table 4.3.1. The subjects labeled the resulting composite risk categories as human health risks, environmental risks, aesthetic risks, and economic impacts.

Scales are used to elicit quantitative measures of qualitative variables. The scale questions were designed to quantify perceptions of the four risk categories. Scale range and

Table 4.3.1 Specific Concerns and Composite Risk Categories

Specific Concerns	Composite Risk Category
Cancer Safety (of workers in the facility) Respiratory problems Transmission lines	Human Health Risks
Ozone depletion Mining—strip mining, water impacts Water quality—mine seepage Air quality, pollution Wildlife impacts Disposal of facility waste such as ash, waste heat, and steam Land use—location of the facility, land requirements Coal economically important to PA	Environmental Risks
Cost of pollution control technology Property values Cost of electricity (retail) Taxes Need for additional capacity Cost of fuel Funding of project development Creation of local jobs	Economic Impacts
Odor—diesel trucks Appearance of the facility Noise Transportation of fuel to the facility Truck traffic Dirty emissions Transmission lines	Aesthetic/ Amenity Risks
Concern for future generations Equity—distribution of costs and benefits Regulatory compliance and enforcement Full disclosure of information Fuel choice and characteristics Coal—dirty, polluting, environmental damage from coal mining, dust, trucks, acid rain, Natural Gas—clean burning, gas leaks and explosions	Affecting Multiple Categories

units can be varied and depend upon the desired quantitative accuracy and the subject's cognitive ability to distinguish between choices.

Vocabulary refers to the wording of survey text, questions, and scale labels. Focus groups and interview discussions revealed that subjects anchored their interpretation of scale questions on the scale labels and format more than on the actual question. Survey text, question wording, and scale labels were iteratively revised until the subjects' interpretations of the questions concurred with their intended interpretation.

Finally, survey bias refers to the extent to which subjects felt the survey was encouraging them to respond in a particular way. Scales and scenario descriptions were iteratively revised to minimize the extent of bias in the survey. Debriefing questions were also used to check for bias. The final survey is composed of five sections (see Appendix 4D). The introduction to the survey contains a brief description of the nature of the survey and "warm-up" questions to assess the participant's awareness of their own consumption and expenses for electricity, their proximity to an energy generating facility, and how they rate the risk of an electric power facility relative to other "industrial" hazards. The next two sections provide a hypothetical description of the respondent's community, the growing need for electricity in the community, general features of the proposed power plant (size, land requirements, life of project), and regulatory compliance. One section proposes a coal-fired plant and the other a natural gas-fired plant (U.S. Department of Energy, 1993). A set of four scale questions for each risk category follows each scenario to quantify the respondents' perception of the possible risks and impacts resulting from the proposed facility. Following the scale questions, subjects rate their concern for each of the four risk components.

The valuation sections asks respondents to vote in favor of or against the proposed facility, and the maximum amount extra they would pay each month on their electricity bill to *prevent* the proposed facilities in their community. The question specifically states why the respondents need to pay their utility company extra to prevent construction of the proposed power plant: *"If the supplier is not able to build this power plant, it will have to increase the price you pay for electricity because the much needed electricity must be purchased at a higher cost from other power producers elsewhere in the state."* Subjects were reminded of their budget constraint prior to answering the valuation question.

The final section of the survey retrieves socio-demographic information about each respondent. The survey was followed by a debriefing questionnaire to enable researchers to gain additional insight into the respondents' answers and to test sample questions. The final survey contains 63 questions, eight of which are demographic.

Data were collected from written surveys in Harrisburg and State College, Pennsylvania. Subjects were recruited via random digit dialing; the adult with the last birthday was asked to participate. Two hundred and twenty surveys were administered. The sample is evenly split by gender, with a mean age of 44 years, a mean education level of 15.6 years, and a mean household income of \$46,300.

4.3.5 Estimation And Results

The mean values of the perceived risk attributes for the coal and natural-gas fired plants are presented in Table 4.3.2. Participants responded to these questions after reading the hypothetical proposals for each plant. In general, a majority of respondents believed both plants would create risks, coal more so than natural gas. Their perceptions of severity and probability of health, environmental, and aesthetic impacts were also more negative for coal than for natural gas. The coal facility generated greater feelings of dread or fear, as well. The differences in perceptions between health, environmental, and aesthetic risks from coal and natural gas were significant at the 1% level for all the measures. The difference in perceptions of the significance or the probability of economic impacts arising from a coal or natural gas plant was not significant, although if economic impacts did occur, participants expected the impacts to be greater for natural gas. Comparisons between perceptions of health, environmental, and aesthetic risks or economic impacts within one fuel are not possible. Given the unique nature of these risk components, the scales are not likely to be comparable, and thus mean measures between factors do not have any ordinal meaning.

The seriousness, probability, and dread variables are all significantly correlated with each other within and between each facility type. Severity and probability are positively correlated with dread and positively correlated with each other. Correlations are generally less prominent for the natural gas-fired plant than the coal-fired plant. The perceived severity, probability, and dread measures for the coal-fired facility are positively correlated with respective measures for the natural gas-fired facility.

Table 4.3.2 Mean Values of Perceived Risk Attributes

	Believe Plant Will Create Risk		Perceived Impact		Perceived Severity		Perceived Probability		Dread	
	<u>Coal</u>	<u>Gas</u>	<u>Coal</u>	<u>Gas</u>	<u>Coal</u>	<u>Gas</u>	<u>Coal</u>	<u>Gas</u>	<u>Coal</u>	<u>Gas</u>
Health	82.3	48.2*	-1.07	-0.19*	3.89	2.45*	50.45	29.41*	-1.19	0.03*
Environmental	94.5	76.3*	-1.57	-0.68*	4.57	3.09*	63.73	44.36*	-1.72	-0.59*
Aesthetic	92.3	76.4*	-1.71	-0.80*	4.42	3.11*	62.27	48.86*	-1.61	-0.34*
Economic	90.4	91.3	0.62	1.54*	4.40	4.38	62.15	59.95	0.08	0.86*
Scale	0-100%		very negative = -4 no impact = 0 very positive = +4		not serious = 1 very serious = 7		0-100%		afraid = -4 neutral = 0 comfort = +4	

*The difference between the coal and natural gas figure is significant at 1%.

A measure of perceived risk can be calculated by multiplying perceived severity by perceived probability. Table 4.3.3 reports the mean and standard deviation of this perceived risk measure for each of the four risk categories. Average perceived risk is higher for the coal-fired plant than the natural gas plant in all four categories. Health risks from a natural gas plant are perceived as almost nonexistent.

The dependent variables are the respondent's WTP to prevent a coal-fired and natural gas-fired electric power plant. The values range from \$0.00 to \$120.00 for coal with a mean of \$16.98 and from \$0.00 to \$70.00 with a mean of \$13.37 for natural gas. The difference in means is significant at the 1% level. Summary statistics for all variables are found in Tables 4.3.3 and 4.3.4. The sociodemographic variables included in all models are income, education, age, gender, and an interaction term between income and education. The interaction term implies that the effect of education on WTP depends on the level of income, that is the relationship between WTP and education depends on income.

The actual model estimated is

$$WTP_j = \beta_0 + \beta_1 INC_j + \beta_2 EDUC_j + \beta_3 INC_j * EDUC_j + \beta_4 GENDER_j + \beta_5 AGE_j + \beta_6 VERSION_j + \beta_7 HEALTH_j + \beta_8 ECON_j + \beta_9 ENVIR_j + \beta_{10} AESTH_j + \varepsilon_j \quad (4.3.8)$$

where j indexes individuals, VERSION is a dummy variable referring to whether the coal or natural gas section came first, and HEALTH, ECON, ENVIR, and AESTH are measures of the four perceived risk categories considered. Three different measures were used for the risk attributes. In Model 1, risk is measured as the *perceived seriousness* (e.g., CHLTSERI) of each of the four impacts, which is a measure of the severity of the consequences. Model 2 uses the constructed *perceived risk* (e.g., CHLTPRSK) measure of perceived severity times perceived probability, which is closest to the standard definition of risk. Model 3 uses a *weighted measure of perceived seriousness* (e.g., CHLTWSER), constructed by multiplying perceived seriousness CHLTWSER), constructed by multiplying perceived seriousness times a value from 0 to 1 that reflects the relative importance of each impact to the respondent. Tobit estimation is the appropriate estimation choice because the dependent variables, WTP, are censored at zero (27% of the coal valuations and 45% of the natural gas valuations were zero). Ordinary least squares estimates would be biased upward and inconsistent.

Table 4.3.3 Perceptions Measures

PERCEPTION SCALE	FUEL TYPE	VARIABLE NAME	MEAN	STD. DEV.
Seriousness of health impacts	COAL	CHLTSERI	3.89	1.75
	GAS	GHLTSERI	2.45	1.45
Significance of economic impacts	COAL	CECNSERI	4.40	1.34
	GAS	GECNSERI	4.38	1.39
Seriousness of environmental impacts	COAL	CENVSERI	4.57	1.59
	GAS	GENVSERI	3.09	1.59
Seriousness of aesthetic impacts	COAL	CAESSERI	4.42	1.59
	GAS	GAESSERI	3.11	1.57
Probability of health impacts	COAL	CHLTPROB	0.50	0.28
	GAS	GHLTPROB	0.29	0.24
Probability of economic impacts	COAL	CECNPROB	0.62	0.24
	GAS	GECNPROB	0.59	0.24
Probability of environmental impacts	COAL	CENVPROB	0.63	0.27
	GAS	GENVPROB	0.44	0.28
Probability of aesthetic impacts	COAL	CAESPROB	0.62	0.26
	GAS	GAESPROB	0.49	0.28
Health risk = SERIOUSNESS*PROBABILITY	COAL	CHLTPRSK	2.36	1.87
	GAS	GHLTPRSK	0.99	1.23
Economic risk = SIGNIFICANCE *PROBABILITY	COAL	CECNPRSK	2.28	1.56
	GAS	GECNPRSK	1.34	1.23
Environmental risk = SERIOUSNESS*PROBABILITY	COAL	CENVPRSK	3.22	1.98
	GAS	GENVPRSK	1.68	1.63
Aesthetic risk = SERIOUSNESS*PROBABILITY	COAL	CAESPRSK	3.04	1.92
	GAS	GAESPRSK	1.79	1.56
Weighted seriousness of health impacts	COAL	CHLTWSER	1.08	0.56
	GAS	GHLTWSER	0.67	0.44
Weighted seriousness of economic impacts	COAL	CECNWSER	-0.33	1.11
	GAS	GECNWSER	-0.72	0.86
Weighted seriousness of environmental impacts	COAL	CENVWSER	1.28	0.53
	GAS	GENVWSER	0.87	0.49
Weighted seriousness of aesthetic impacts	COAL	CAESWSER	1.03	0.54
	GAS	GAESWSER	0.73	0.47

Table 4.3.4 WTP and Socio-Demographic Summary Statistics

VARIABLE NAME	VARIABLE DESCRIPTION	MEAN	STD. DEV.	MINIMUM	MAXIMUM
CWTP	Willingness to pay to prevent coal plant (\$0.00 bids-26.8%)	\$13.72	16.98	\$0.00	\$120.00
GWTP	Willingness to pay to prevent gas plant (\$0.00 bids-45.0%)	\$8.41	13.37	\$0.00	\$70.00
INCOME	Midpoint of income range	\$46,300	31,026	\$4,999.50	\$189,999.50
GENDER	Dummy variable for gender, 1 = Male	0.5091	0.5011	0	1
EDUC	Years of education completed	15.64	3.12	9	31
AGE	Age of respondent (years)	44.03	17.07	18	87
VERSION	Dummy variable for survey version, 1 = ???	0.53	0.50	0	1

The regression results are listed in Table 4.3.5 for all six models estimated. Income is positive and significant in all models, consistent with the hypothesis that individuals with more disposable income are willing to pay more for normal goods of this type. Education is also positive for all models, a common result in nonmarket valuation studies. Age and gender do not seem to play a role in determining an individual's WTP. Version is insignificant in all models, indicating that an individual's stated WTP is independent of the order in which the coal and natural gas valuation questions were presented. Some have suggested that a second WTP question will be invalid, but these results indicate otherwise. It further provides a measure of the reliability of the survey instrument.

When risk perceptions are measured by *seriousness* in Model 1, economic impacts and environmental risks significantly impact the respondents' WTP to prevent a coal-fired facility. All four risk components significantly affect the WTP to prevent a natural gas-fired plant. The results are less appealing when risk is measured by the constructed *risk perception* variable. In Model 2, only environmental risk for coal and health and aesthetic risks for natural gas are significant. Although this *risk perception* variable most closely resembles the accepted definition of risk, it may be difficult for respondents to differentiate between severity and probability. Alternatively, the survey may not have been able to adequately elicit the difference.

In Model 3, risk is measured by a constructed measure which weights perceived seriousness by the importance of each impact to the respondent. The parameter estimates are larger than in Model 1 because the size of the independent variables has been reduced by multiplying them by a number between zero and one. The results for natural gas are similar to Model 1. For coal, aesthetic and health risks become significant but economic impacts become insignificant.

The negative coefficient on economic impacts indicates individuals are less willing to pay to prevent the plant the larger the economic impacts are perceived. This result suggests that separately identifying positive impacts (economic benefits) from negative impacts (health, environment, or aesthetics) can improve interpretation of individuals' responses to CVM scenarios. It may also provide more detailed information to planning officials and improve risk communication methods.

Table 4.3.5 Tobit Regression Results (Standard Errors in Parentheses)

	<u>Model 1</u>		<u>Model 2</u>		<u>Model 3</u>	
	Coal	Gas	Coal	Gas	Coal	Gas
CONSTANT	-67.07* (15.73)	-60.7642* (14.5966)	-51.087* (14.7202)	-44.0057* (14.0162)	-60.0099* (15.3419)	-58.7228* (14.7563)
INCOME	0.00068* (0.00023)	0.0005317* (0.000215)	0.00064* (0.00023)	0.00051* (0.000214)	0.000608* (0.000232)	0.00052* (0.000217)
EDUC	2.5328* (0.8674)	1.8537* (0.8117)	2.333* (0.856)	1.6029* (0.8164)	2.2007* (0.8652)	1.7198* (0.8226)
INC* EDUC	-0.00004* (0.000014)	-0.0000283* (0.000013)	-0.0000356* (0.000014)	-0.0000261* (0.000013)	-0.0000335* (0.000014)	-0.0000266* (0.000013)
AGE	0.0715 (0.0872)	0.0737 (0.08336)	0.0999 (0.0882)	0.0808 (0.08413)	0.0699 (0.0869)	0.0976 (0.0844)
GENDER	-3.8579 (2.9098)	-3.2056 (2.8561)	-4.612 (2.8758)	-3.2451 (2.8976)	-3.1938 (2.9077)	-3.0391 (2.8505)
VERSION	3.1739 (2.9123)	3.0252 (2.6994)	4.0784 (2.9719)	3.6096 (2.7691)	4.1425 (2.8416)	2.8529 (2.7178)
HLTSERI	1.5892 (1.2835)	2.6205* (1.1761)				
ECNSERI	-0.5388* (0.3222)	-0.7591* (0.3983)				
ENVSERI	3.6424* (1.4159)	2.0495* (1.2174)				
AESSERI	1.7048 (1.1954)	3.5338* (1.0934)				
HLTPRSK			1.1753 (1.22)	4.2369* (1.4411)		
ECNPRSK			-0.6964 (0.4365)	-0.7092 (0.5374)		
ENVPRSK			2.7631* (1.206)	0.4883 (1.2977)		
AESPRSK			1.4865 (1.0549)	3.3587* (1.1137)		
WHLTSER					7.0531* (3.1846)	8.5007* (3.7048)

WECNSER	-1.2322 (1.3437)	-2.6873* (1.6251)
WENVSER	10.2932* (3.3579)	11.6373* (3.4963)
WAESSER	6.9441* (2.6495)	9.2607* (3.0972)

*Significant at 5%.

Overall, regardless of model specification, the evidence is strong that the perception of environmental risks influences a respondent's willingness to pay to avoid a coal-fired electric utility in their community. The results are less strong but suggest health, aesthetic, and economic impacts also influence WTP. For the natural gas-fired facility, health and aesthetic risks influence WTP in all models. The evidence is also strong that environmental risks and economic impacts influence WTP.

4.3.6 Conclusions

Results show that welfare impacts, as measured by option price, depend on an individual's perceptions of the health, environmental, aesthetic, and economic impacts as well as their socio-demographic characteristics. Individuals seemed to have difficulty distinguishing between the probability and the severity of a risk in the manner suggested by the definition of risk, although the survey instrument may have been unable to capture the difference. Perceived environmental, health, and aesthetic impacts play a larger role than potential economic impacts in determining option prices, explaining in part why residents may oppose a facility even when it will likely bring economic benefits to an area. Although this study focused on fossil-fuel electric power facilities, the methodology developed here is transferable to a multitude of hazards imposing welfare impacts which are a function of perceived risk.

The results confirm that CVM values are sensitive to the information set and perceptions of a participant (see Fischhoff and Furby, 1988 for additional evidence). While the general approach in contingent valuation methodology has been to mold or correct those perceptions in line with "expert" information, this study has explicitly measured those perceptions and related them to willingness to pay. A correct measure of welfare impacts requires consideration of the participants' perceptions rather than the researcher's perceptions of the commodity.

The potential exists to gather comparable data from "experts" to define levels of "true" risk. These could then be compared to laypersons' perceptions to estimate the welfare impacts from a divergence of expert and layperson risk judgements associated with a facility. The survey could also be administered at different phases of a project's life, from proposal through operation or defeat, to examine changes in perceptions and welfare impacts over time. A further extension could examine the determination of an individual's perceptions as a function of prior experience, y. This approach is implicit in programs to identify the impact or value of public information efforts and is of considerable importance in risk communication programs.

4.4 Regional/National Economic Impacts of New Coal Utilization Technologies

4.4.1 Introduction

Energy conservation is almost universally considered a prime strategy for mitigating greenhouse gases. At present, 97.9 and 70.6% of the CO₂ emitted from industrial and developing countries, respectively, stems from fossil fuel combustion (World Resources Institute 1994). Of course, this anthropogenic emission is only a small fraction of the carbon exchange between the atmosphere and terrestrial ecosystems. Efforts to alter atmospheric concentrations through climate engineering are still only on the drawing board. With any potential major shift to renewables many years away, outright reduction in the utilization of coal, oil, and natural gas is an obvious strategy. It appears even more appealing when one considers that a good deal of conservation can be attained at a cost-savings when less energy is used outright, or at a zero net cost when, for example, energy-saving equipment must be installed. These factors have led to energy conservation being placed in the category of "no regrets" strategies, which refers to measures that do not incur added costs even if projected warming trends are not forthcoming (Cline 1992)

Clearly, production cost-savings and preservation of energy resources are pluses. However, to date, very few studies have focused on the potential down-side. For example, there are jobs and profits at stake in the energy industries. Moreover, declines in fossil fuel sectors will lead to declines in output in successive rounds of upstream suppliers (e.g., mining equipment, fuel service companies), as well as some downstream customers (e.g., railroads, electric utilities). It is not clear whether these negative effects will be offset by the increased

efficiency of the economy, various factor substitutions, purchasing power improvements for consumers, or any multiplier effects stemming from increased production of energy-saving equipment.

The purpose of this review is to estimate the effects on the U.S. economy and its energy sectors of conservation strategies to reduce CO₂ emissions. The analysis is undertaken with a 20-sector computable general equilibrium (CGE) model by simulating various responses to command and control, carbon tax, and carbon emission permit policies.

The results indicate that the characterization of energy conservation as a "no regrets" strategy is too strong. In all of the simulations, energy sectors stand to lose, though, in some cases, not anywhere near as much as would be expected. Each of the simulations of mandated conservation also leads to a decline in output and employment for the U.S. economy. In contrast, some of the price-induced conservation response strategies also simulated have a neutral impact on the overall economy.

This review is divided into five sections. In the first section, basic features of the conservation strategy and some overlooked issues are discussed. Then the model used in the simulations is summarized. Next, basic results are presented, as well as some sensitivity tests. The review is concluded with a summary and a discussion of policy implications.

4.4.2 Basic Features of the Conservation Strategy

Many proposals have been put forth to combat potential global warming. One that has received considerable attention calls for a 20% reduction in current greenhouse gas (GHG) emission levels for industrialized countries and a stabilization of developing country emissions at current levels. Several policy instruments are available to implement this reduction, the carbon tax being the most widely supported and most thoroughly analyzed (Manne and Richels, 1991; Nordhaus 1993). In the U.S., however, strong support exists for a marketable permits approach (Winer 1991).

Actually, the conservation response to a carbon tax and a marketable permit system can be modeled in the same manner. First, the optimal carbon tax rate would be equivalent to the equilibrium permit price (Weitzman 1974; Pezzey 1992). Second, the two instruments would result in the same efficient response, in which each polluter equates its marginal cost of abatement to the tax rate or permit price. Third, even though polluters must pay for each unit of

GHG emissions under a tax regime, these emissions are usually free (entitlements) under a permit system and this does not affect the response in the short-run. That is, the tax payments or permit revenues/expenditures affect a firm's average cost but not its marginal cost and thus only bear upon long-run considerations such as exit and entry. Moreover, these conclusions pertain to the application of these instruments at both the international level in relation to total GHG reductions and within national boundaries.

In a recent study, Rose and Stevens (1993) estimated an equilibrium permit price of \$38.35 that would be associated with an agreement to limit *global* CO₂ emissions at 20% below year 2000 levels. The 20% reduction stems from an oft-espoused policy pronouncement that would contribute to the stabilization of CO₂ concentrations. Emissions for the year 2000 are projected based on population and economic growth factors. The year 2000 was chosen as a base rather than, for example, 1990, since the former is likely to be closer in time to when an agreement on CO₂ mitigation might actually be reached. The optimal response of the U.S. to this price is a reduction of 12.8% of baseline emissions. Moreover, given the uniqueness of the outcome of the Coase Theorem, this abatement level is not sensitive to how the permits are initially distributed across countries. Also, within the U.S., there would be a unique optimal response though control levels would vary across polluters (e.g., economic sectors) given differences in marginal abatement costs.

The various tactics that can be applied to the mitigation of CO₂ are depicted in Figure 4.4.1, utilizing a step function to highlight their usual relative marginal cost positions. The first step of the cost function refers to *no regrets* (costless or even cost-saving) conservation. This could stem from either a technological innovation or a move toward the efficiency frontier as a result of correcting a misallocation (e.g., eliminating energy-wasting practices). There is considerable disagreement in the literature about the extent of conservation potential. Manne and Richels (1992) refer to it as *autonomous energy efficiency improvement* (AEEI), which they estimate to be on the order of 0.5% to 1.0% per year. More optimistic estimates of costless conservation in the range of 20% to 30% total for the near term are offered by OTA (no date), NAS (no date), Lovins and Lovins (1991) and Jaccard et al., (1993). This holds open the possibility that the optimal U.S. CO₂ emission reduction could be met entirely by this tactic.

Another major form of conservation is *price-induced*, e.g., decreasing energy use in response to a change in the price of energy relative to the price of other inputs, as would be

caused by a carbon tax or permit regime. There are several types of input substitutions that could take place and we categorize them under the headings below following Cline (1992):

Figure 4.4.1 CO₂ COST FUNCTION

OFS—Other factor substitution

IFFS—Inter-fossil fuel substitution

NFFS—Non-fossil fuel substitution

PMS—Product mix substitution

Thus, if the tax is based on carbon content of fuels, there are optimal substitution responses within the class of fossil fuels (coal emits approximately 1.26 times as much carbon per unit as oil and 1.86 times as much as natural gas) and between the class of fossil fuels and other sources (e.g., nuclear, hydro, solar). All of these responses incur some costs unless the elasticity of substitution is infinite.

The final category of mitigation tactics shown in Figure 4.4.1, though limited in the near-term, is "end-of-pipe" abatement, such as CO₂ scrubbers. Of course other measures, such as climate engineering and carbon absorption through tree planting, might be used but are beyond the scope of this study.

The discussion above pertains only to a first set of adjustments for any decision-maker. If the price of energy inputs decreases because of improved efficiency, ironically energy then becomes more attractive, and there will be some offsetting increase through substitution of energy for other inputs. There is also the question of whether cost-savings will be passed along to industrial and/or final consumers or whether they will increase the returns to labor and/or capital. Similar possibilities arise with respect to responses that incur positive costs, though in terms of price increases and decreases in profits and wages.

The presentation thus far has been limited to partial equilibrium analysis, but a host of general equilibrium effects could potentially further enhance or offset these effects. First, if energy conservation savings were passed on to other industrial customers, there would be the possibility of further rounds of price reductions. This could potentially change the mix of material inputs in favor of those that are energy-intensive. The overall price reduction would increase the purchasing power of consumers and provide a stimulus to the entire economy. This would also take place, but to a different degree, if price decreases were foregone in favor of increasing wages or profits.

At the same time, the reduction in energy use would lower production levels in the coal, oil, gas, refined petroleum, and electric and gas utility industries. This will touch off a chain of negative multiplier effects to upstream suppliers, such as mining equipment, field service, and finance industries, as well as downstream customers, such as railroads, pipeline companies, and electric utilities. These negative impacts would be reinforced by declines in wages and profits in all of these sectors as well. Interestingly, all of these negative effects engender additional energy conservation, though through reduction in economic growth rather than an improvement in energy efficiency. On the positive side, the resources released from the energy industries would stimulate economic activity elsewhere, though it is unlikely that they would be fully employed. Also, any increased demand for energy-saving equipment will have positive multiplier effects analogous to those mentioned in the previous paragraph. At the same time, this increased economic activity will result in increased energy use, partially offsetting conservation efforts.

Obviously, there are a sizable number of expansionary and contractionary influences. It is impossible to ascertain the net outcome a priori, and hence the need for empirical analysis based on a general equilibrium model.

4.4.3 The U.S. CGE Model

An updated version of a 20-sector CGE model, developed by Lin (1991) and similar in nature to most SAM-based CGE models (Dervis et al., 1982; Shoven and Whalley, 1992), was utilized. A brief summary of the model is presented. Domestic producers, being profit-maximizers, produce goods and services using two primary factors, as inputs i.e., labor and capital, and intermediate goods. Intermediate goods are either produced domestically or imported, and are assumed to be qualitatively different (the Armington assumption). The utilization of inputs follows a two-stage decision process in which intermediate goods are modeled as a nested function of aggregates and components. Specifically, the energy aggregate consists of individual fuels (both primary and secondary energy), while the materials aggregate consists of goods such as plastics, glass, metals, etc., and the remaining input aggregate consists of all other intermediate goods. Inter-fuel and inter-material substitutions are allowed within their respective aggregates, which is a feature that is fundamentally required in evaluating environmental quality regulations (Hazilla and Kopp, 1990). The two-stage decision involves finding the optimum combination of components within energy and material aggregates, and then optimizing the levels of capital, labor, energy, and materials.

To take account of inter-fuel and inter-material substitution and substitution among aggregates, flexible functional form cost functions are used to represent the technology of production sectors and it is assumed that these functions are homothetically weakly separable. The relationships between aggregates, and within the energy and material aggregates with flexible functional forms, and the relationship between other intermediate inputs in terms of fixed proportions are specified. For the current application, the Generalized Leontief (GL) functional form is used for all the flexible cost functions.

The demand component of the model includes both intermediate and final demands. The intermediate demand is determined by the cost-minimizing process discussed above. Final demand includes private consumption, government expenditures, and investment.

The modeling framework is general enough to incorporate several alternative views of equilibrium. In one version of the model the total employment is exogenously given so full employment is achieved. Furthermore, the investment level is determined by savings, with savings rates being fixed. These specifications would have the model belong to the "classical" category. However, in the analysis below, an alternative (Keynesian) formulation of the labor

market is invoked, in which the wage rate is fixed and labor supply adjusts to possibly less than a full-employment equilibrium. In addition, other components of the flow-of-funds account, such as the current account balance and government deficit, are considered in conjunction with the above for policy analysis purposes.

The core of the base year equilibrium data set of the model in this study is a SAM of the U.S. economy for 1987, constructed by Hanson and Robinson (1987). This data set contains both *make* and *use* versions of the U.S. Input-Output Table to capture the production and utilization of commodities. The SAM, in addition to providing the basic data set for model calibration, also provides initial values for endogenous variables and levels for exogenous variables. In this study, a 20-sector version of the model is used, including five mining sectors and two energy utility sectors (see Appendix 4E).

Some key parameters of the empirical model—Allen elasticities of substitution between aggregate inputs—are presented in Table 4.4.1. They represent a synthesis of estimates available in the literature. Sensitivity tests were performed on these values.

Table 4.4.1 Allen Elasticities of Substitution for Selected Sectors

Sector	σ_{KE}	σ_{KL}	σ_{KM}	σ_{LE}	σ_{LM}	σ_{EM}
Construction	.70	.20	.55	-.25	.89	.77
Petroleum Refining	.25	.90	.35	.20	.50	.05
Manufacturing	.70	.20	.55	-.25	.89	.77
Transportation	.70	.90	.35	.25	.10	.05
Utilities	.70	.90	.02	-.10	.01	.01

^aThe symmetry properties of our production function require $\sigma_{ij} = \sigma_{ji}$.

4.4.4 Simulation Results

In this section, simulation results are presented for both autonomous and price-induced conservation strategies to be implemented in the Year 2000. The first set of simulations represents best estimates of the effects of these responses on major economic indicators for the

economy in general and the energy sectors in particular. Other simulations examine the sensitivity of the results to key parameters and behavioral considerations.

Energy policy responses are simulated in two ways. First, to analyze *autonomous* conservation, or the *mandated* response to a command-and-control policy, the energy use parameters were reduced in the model by 12.8% across-the-board (recall the U.S. response to a global policy regime discussed above). The model then analyzes sectoral (partial equilibrium) and multi-sectoral (general equilibrium) responses. Also, sensitivity tests were performed to ascertain the possible offsetting effects of energy-saving equipment needed to implement the policy goal. Note that in actuality, abatement cost curves will vary across sectors and the least-cost no regrets level will thus vary. Due to the lack of data on sectoral conservation potential, a uniform level was simulated. Given the fact that command and control policies are typically applied across the board, this exercise may not in fact represent too much of a departure from reality.

The direct application of a carbon tax or permit trading to examine *price-induced* conservation is simulated. This is modeled as a price increase in primary energy (tax on fossil fuel production), as indicated in Table 4.4.2. Note that in this case the optimal response calls forth differential reductions in energy use and differential levels of interfuel substitution across sectors.

Note that, in effect, the optimal response to CO₂ mitigation policies is likely to be a combination of autonomous conservation and various types of substitution, which have been separated to isolate unique features of each. The overall outcome will thus be some weighted average of the two strategies. Referring to Figure 4.4.1, the exact combinations would be determined by the intersection of the carbon tax level (or equilibrium permit price) and the marginal cost of abatement.

4.4.4.1 Basic Results

The prime simulations are presented in Table 4.4.3 in terms of their impacts on five major economic indicators. The basic conservation case (line 1 of Table 4.4.3) is estimated to yield an overall decrease in GNP of 1.44% and a reduction in employment of 2.05%. Economywide investment decreases by 4.41%, with exports actually increasing by 2.75% and imports declining by 3.63%. The difference in the GNP and Employment results indicates that there is proportionally greater substitution away from labor-intensive goods

Table 4.4.2 Transformation of a Carbon Tax to an Ad Valorem Tax

	Fuel		
	Coal (ton)	Oil (bbl)	Gas (tcf)
Heat Content (million Btu per unit)	21.94	5.80	1.03
Emission Rate (kg per million Btu)	26.90	21.40	14.50
Conversion Factor (tons per unit)	0.59	0.12	0.02
Carbon Tax (1990 \$ per unit carbon)	38.35	38.35	38.35
Ad Valorem Tax (1990 \$ per physical unit)	22.63	4.76	0.57
Fuel Price, Year 2000 (1990 \$ per physical unit)	26.64	26.40	2.72
Price Increase (percentage change)	85.12	18.02	20.83

Specifically, the following four cases were simulated:

1. Simple conservation—a 12.8% cutback in purchases of all fossil fuel energy, including utilities, in all intermediate and final demand sectors;
2. Conservation, but with the entire cost-savings in each sector being assigned to increased purchases of energy-saving equipment;
3. Price-induced substitution, but with the carbon tax being imposed only on domestic use of domestically produced primary energy; and
4. Price-induced substitution, but with the carbon tax being imposed on domestic use of both domestic and imported primary energy.

than is the case for other inputs. Note also that the impact of this case (and all others) in terms of a welfare measure such as compensating variation, are also negative, though by a smaller percentage than the macro indicators.

Aggregate energy use declines as do CO₂ emissions, but startlingly less than expected. As expected, coal and electric utilities suffer the largest declines. Even though direct effects are

Table 4.4.3 Economywide Impacts of CO₂ Mitigation, Year 2000: Base Cases (percentage change from baseline)

Case	Real GNP	Employment	Investment	Exports	Imports	Welfare ^a	Energy Use	CO ₂ Emissions
1. Conservation (100% cost-saving)	-1.44	-2.05	-4.41	2.75	-3.63	-0.84	-3.32	-3.44
2. Conservation (100% equipment offset)	-3.65	-3.94	-4.11	0.84	-3.02	-2.25	-9.16	-8.96
3. Interfuel Substitution (no tax on imported oil)	-1.56	-2.02	-3.59	-1.79	-1.93	-1.39	-24.17	-26.07
4. Interfuel Substitution (tax on all oil)	-1.77	-2.28	-3.37	-2.60	-3.38	-1.64	-25.49	-27.22

^aAs measured by compensating variation.

proportional, general equilibrium effects allow for substitution away from these sectors. At first one would expect declines of greater than 12.8% in each energy sector given the direct response and subsequent multiplier effects. However, it appears that the price decrease for each fuel causes energy to be more attractive to the point where there is a significantly offsetting substitution effect toward it. This has been pointed out in a number of studies that warn of the unintended side-effects of autonomous or mandated conservation (Khazzoom, 1980). Other general equilibrium effects are operative as well but are too difficult to sort out without further experimentation (see below). Overall, the downside effects are not overcome by stimuli from increased purchasing power, international competitiveness, or multiplier effects. The results for the energy sectors are presented in Table 4.4.4.

Table 4.4.4 Energy Sector Impacts of the Conservation (100% Savings) Response, Year 2000:
Base Case (percentage change from baseline)

Energy Type	Gross Output	Employment	Exports	Imports
Coal	-3.85	-4.64	0.39	-6.52
Oil/Gas Extraction	-1.03	-2.75	4.36	-4.43
Petroleum Refining	-2.28	-2.84	2.61	-5.16
Electric Utility	-4.21	-5.36	1.99	-7.18
Gas Utility	-2.49	-3.07	3.37	-5.33

Note the irony of this policy response. In effect, the initial willingness of industry to decrease energy consumption by 12.8% results in offsetting factors that would not enable the U.S. to meet its CO₂ reduction target. That is, the overall average decline of energy use of 3.32% would mean that CO₂ emission reductions would only be one-fourth of those intended. This is somewhat disconcerting and indicates that energy users on the average might need to undertake several times the amount of initial energy conservation to yield the intended overall 12.8% reduction.

The results of a modified conservation response is presented in Case 2 of Table 4.4.3. When the entire cost savings is offset by increased costs of energy-saving equipment, the negative impacts are even greater than in the base case. Table 4.4.3 shows that GNP declines by 3.65% and employment by 3.94%. The key to understanding this decline is the simultaneous reduction in economywide investment. It would appear that the *crowding-out* effect of investment in energy-conserving equipment is substantial and has a dampening effect on the economy (see also Jorgenson and Wilcoxon, 1993). However, there is some bias in the model and its application. The investment equations are specified for more general cases. The model thus views the earmarking of investment funds for energy conservation as sub-optimal (as do the vast majority of the models in the literature). On the other hand, if the rate of return on this investment specifically reflected the gains that could be brought about, it might very well be that energy-saving equipment would be one of the better uses of investment funds. The otherwise expansionary effect of this increased investment might cause overall investment and output declines to be lower than Case 1. To be positive overall, however, would require considerable impetus. The 12.8% decrease in energy costs in most sectors translates into a 0.1 to 1.0% cost savings in each sector. This presents a relatively minor advantage to investment in energy-saving equipment over other alternatives.

The effect of Case 2 on individual energy sectors is presented in Table 4.4.5. The reductions come close to the 12.8% due in great part to investment considerations, which represent a type of forced substitution of capital for energy, in addition to other responses. Whereas in Case 1, energy intensity declines were below 0.2% for all sectors, here several sectors wind up with declines of greater than 2.0%.

In Cases 3 and 4 (Table 4.4.3), the response to a carbon tax or permit trading in terms of direct inter-fossil fuel substitution (IFFS) and other factor substitution (OFS) is simulated. In Cases 3 and 4, the increased cost of the energy aggregate leads to substitution between it and other aggregate input categories. These responses are further affected by various other general equilibrium interactions that take place, including product mix substitution (PMS). In Case 3, the tax applies only to domestic use of energy produced in the U.S., while in Case 4 it applies to imported energy (primarily oil) as well.

Case 3 also yields a negative impact on the economy in terms of GNP and employment. The economic impacts of this case are quantitatively and qualitatively similar to Cases 1 and 2,

Table 4.4.5 Energy Sector Impacts of the Conservation (100% Equipment Offset) Response, Year 2000: Base Case (percentage change from baseline)

Energy Type	Gross Output	Employment	Exports	Imports
Coal	-8.05	-9.12	-13.01	-4.73
Oil/Gas Extraction	-9.63	-12.07	-1.34	-13.66
Petroleum Refining	-11.35	-12.10	-34.42	3.47
Electric Utility	-6.92	-9.30	-12.06	-4.24
Gas Utility	-10.05	-11.32	-30.66	2.48

except that exports decrease. The negative impacts on the economy are more pronounced in Case 4 because more inputs (i.e., imported oil) suffer price increases. These impacts are not offset by relatively more favorable terms of trade in Case 4 versus Case 3.

The effect of Case 3 on individual fuels is presented in Table 4.4.6. As would be expected from Table 4.4.2, coal bears the brunt of the carbon tax, as reflected in a reduction in the sector's gross output of over 44%. Domestic crude oil production declines by 6.77% and imports decline by 8.77%. Sectoral results for Case 4 are very similar to Case 3, except that crude oil import reductions are much greater in the latter.

Perhaps the major differences between the mandated and incentive-based conservation responses are the energy and CO₂ reductions. In Case 3 (Table 4.4.2) aggregate energy use decreased by 24.17% and total CO₂ emissions decreased by 26.07% (both figures are slightly higher for Case 4). The nearly 2.0% differential between energy use and CO₂ emission reduction reflects a significant amount of fuel switching. The majority of the emission reduction, however, stems from a relatively much greater decrease in energy use in Cases 3/4 vs. Cases 1/2 (compare also Tables 4.4.4 to 4.4.6). Some energy intensive sectors are the hardest hit by the general equilibrium effects, e.g., steel, stone (cement), and transportation.

Table 4.4.6 Energy Sector Impacts of the Interfuel Substitution Response to Tax on Domestically Produced Energy, Year 2000: Base Case (percentage change from baseline)

Energy Type	Gross Output	Employment	Exports	Imports
Coal	-44.37	-37.43	-65.44	-19.31
Oil/Gas Extraction	-6.77	-8.76	-3.35	-8.77
Petroleum Refining	-7.72	-6.39	-32.05	8.74
Electric Utility	-7.85	-10.11	-34.49	9.32
Gas Utility	-1.69	-2.90	-18.94	8.34

Apparently declines in the energy sectors do contribute to the overall negative effect on the economy in the various simulations, but there are some offsetting effects in the carbon tax/permits cases, since decreases in GNP are only slightly higher despite a 7-8 fold decrease in energy production vis-a-vis Case 1. The relatively greater labor intensity of non-energy industries is one part of the explanation (compare the economywide employment decreases between Cases 1 and 3). Other explanations include the relatively lower negative impact on investment and the spending impetus of carbon tax revenues (though only to a slight degree as will be shown below).

4.4.4.2 Sensitivity Analysis

It is acknowledged that the CGE model is based on a calibration method with less than ideal statistical properties, and that several facile assumptions have been invoked. Therefore, it is not unreasonable to question the robustness of the results. This is tested by examining the effects of utilizing alternative estimates of capital-energy substitution elasticities and by invoking alternative CGE closure rules.

Recall the K-E elasticities of substitution for major sectors are presented in Table 4.4.1. They range from 0.250 to 0.700, which means capital and energy are considered substitutes. It is

possible that these elasticities overstate the degree of the substitutability relationship. Moreover, it is also possible that capital and energy are complements.

Note that in addition to the elasticity values in Table 4.4.1, a further substitution relationship between capital and energy in Cases 2 and 3 of the previous subsection has been modeled. In effect, a direct decrease in energy and an increase in capital by an equivalent amount and by half the amount, respectively, has been inserted essentially increasing the K-E elasticities by these quantities. Indirect effects are still modeled with the ordinary K-E substitution elasticities.

Two sets of sensitivity tests on elasticities were performed. The first reduced the K-E elasticity values by half. The overall results (not shown) yielded imperceptible differences for Cases 1 and 2 and only minor differences for Cases 3 and 4. The second set of sensitivity tests utilized elasticities for capital and energy that exhibit complementary relationships. This required recalibration of the production functions and mainly minor adjustments in other elasticities. The results of the application of these new parameters also had little affect on the results and are not presented here. In effect, the results are thus more sensitive to substitution levels across input aggregates, than within them, even for the case of interfuel substitution.

The third set of sensitivity tests is based on alternative assumptions about the labor market, factor mobility, and fiscal balances. In order to achieve equilibrium in constructing the CGE model, one cannot over- or under-constrain any of the markets. For example, in the case of labor supply, both the labor supply and wage rate fixed cannot be held fixed, nor can both of them vary. The specification of which aspect is fixed and which is variable in this and in other markets is known as a *closure rule*.

In the simulations above, a fixed real wage rate was assumed. Next the implications of an alternative closure rule that sets labor supply fixed and allows the wage rate to adjust is examined. This forces full employment of resources directly and indirectly released by the decrease in energy production. In addition, the implications of the assumption about the perfect mobility of capital are examined. Finally, the sensitivity of the result to whether carbon tax revenues are used to expand government expenditures or used for deficit reduction (with government expenditures being fixed) are examined. The conditions underlying the various sensitivity tests are presented in Table 4.4.7.

Table 4.4.7 Definitions of Subcases of Simulation 5

Sub-cases	Government Expenditure	Total Labor Supply	Sectoral Labor Demand	Total Capital Stocks	Sectoral Capital Stocks	Characterization
5.1	NF	NF	M	F	M	UNEM
5.2	NF	F	M	F	M	FULEM
5.3	NF	NF	M	F	F	S-R
5.4	F	NF	M	F	M	UNEM
5.5	F	F	M	F	M	FULEM
5.6	F	NF	M	F	F	S-R

Note:	F	=	Fixed	UNEM	=	Underemployment equilibrium
	NF	=	Not Fixed	FULEM	=	Full employment equilibrium
	M	=	Mobile	S-R	=	Short-run case

The results of these sensitivity tests applied to Case 4 (Interfuel Substitution/Tax on All Oil) are presented in Table 4.4.8 (note that Case 4.1 is the same as Case 4 listed in Table 4.4.3). As in the original simulations, the impacts on real GNP are negative in all cases, though they are negligibly so in Cases 4.2 and 4.5 because of the employment forcing. In only two of the cases (4.1 and 4.4) are the negative impacts on GNP and employment greater than in Case 1, and then only marginally so. It appears that using carbon tax revenues for deficit reduction, rather than expanding government expenditures, does enhance negative impacts but only slightly. It also appears that a fixed capital stock assumption reduces the negative impact somewhat (Case 4.3). The explanation would appear to be that capital is prevented from leaving the U.S. (also compare the export and import figures for Cases 4.1 and 4.3).

Table 4.4.8 Interfuel Substitution Impacts, Year 2000: Sensitivity Tests (percentage change from baseline)

Case	Real GNP	Employment	Investment	Exports	Imports	Welfare ^a	Energy Use	CO ₂ Emissions
4.1	-1.77	-2.28	-3.37	-2.60	-3.38	-1.64	-25.49	-27.22
4.2	-0.14	—	-2.24	-1.36	-1.60	-0.65	-24.09	-25.82
4.3	-1.34	-1.75	-3.71	-1.64	-3.20	-1.39	-22.45	-23.93
4.4	-1.79	-2.34	-1.83	-2.53	-3.25	-1.66	-25.55	-27.27
4.4	-0.08	—	2.97	-1.03	-0.99	-0.63	-24.13	-25.86
4.6	-1.50	-2.00	-1.67	-1.62	-3.17	-1.48	-22.59	-24.08

Results for individual energy sectors (not shown) are reasonably similar for all cases. The only significant difference is that crude oil imports decrease slightly in Cases 4.1 and 4.6.

4.4.5 Conclusions

The general equilibrium impacts of a conservation strategy to reduce CO₂ emissions results in a negative impact on GNP, employment, and other macroeconomic indicators. Not surprisingly, it was also found that the impact on the energy industries was also strongly negative. Potentially positive ramifications of conservation, such as cost savings, increased consumer purchasing power, and multiplier effects of investment in energy-saving equipment, were not able to offset the partial and general equilibrium downside effects of decreased energy use. Moreover, the results were robust to alternative assumptions on the degree of cost savings associated with conservation, the ease of substitution between energy and other inputs, and various macroeconomic closure rules. Finally, the results are in the ballpark of estimates undertaken by others, such as Manne and Richels (1991, 1992) and Jorgenson and Wilcoxon (1993), though these studies placed a lesser emphasis on conservation. Thus, it has been concluded that conservation should not be characterized as a "no regrets" strategy.

This does not, of course, mean that conservation is a poor strategy, but simply that it should not be oversold as costless. In addition, the scope of this review has been limited to one side of the ledger, and the existence of reasonable estimates of net benefits of reducing CO₂ emissions in the tens of billions of dollars per year for the U.S. alone (Nordhaus 1993; Boyd et al, 1995) is acknowledged.

However, it should be pointed out that a 3% decline in GNP in the Year 2000 translates into nearly \$200 billion per year of opportunity costs.

Finally, it must be noted that the results pertain only to the short term. This, however, is the time period for which "no regrets" strategies apply, i.e., measures that can be taken until uncertainties about global warming are resolved. If predictions about the onset of warming are verified, a strategy need not be costless to be viable. Moreover, conservation may be able to play an expanded role in the longer term. Rather than the limited range of short-term options analyzed, in this study some researchers have proposed more sweeping "eco-restructuring" for both industrialized and developing

economies (Ayres and Simonis, 1994). It is likely that these strategies would not bode well for traditional energy industries. However, their goal is to integrate innovative approaches to resource utilization into the economy so as to establish a path of sustainable development. Thus, further research is warranted to examine the long-term impacts of conservation.

4.5 Economic Analysis of the Defense Department's Fuel Mix

This activity examines U.S. military fuel consumption and describes alternative technologies that could be currently or potentially applied to the military in order to comply with U.S. government directives to lower emissions, lower dependence on imported oil and to become more energy efficient.

4.5.1 Introduction

DOD is responsible for approximately 2% of U.S. energy consumption. Of this 2%, DOD consumed approximately 34% of all aviation fuel and 33% of all distillate used for shipping in 1990 (Energy Information Agency (EIA) 1994) making DOD an important market player with respect to specific fuels. To provide perspective, Table 4.5.1 contains the breakdown of U.S. military fuel consumption with U.S. government usage and total consumption for the U.S. as a whole. It is clear that defense consumption makes up the vast majority of U.S. government use, although it only accounts for 2% of the total U.S. consumption in all of the major source fuel markets (i.e., petroleum, electricity, natural gas and coal).

EIA has made forecasts regarding certain types of fuel consumption for the years 2000, 2005 and 2010. Table 4.5.2 shows the breakdown of usage of various fuels in the military and their relative importance in terms of U.S. consumption. The military is a major user of shipping and aviation fuels that are predominantly petroleum based.

DOD energy use is divided into two major user types for the purposes of this study: 1) power and heat generation users at military installations representing fairly large expenditures on energy and; 2) transport users which represent smaller absolute amounts but larger relative amounts in terms of percentage of U.S. usage. These two categories will be used as the basis for organization of this report.

Table 4.5.1 Consumption of Energy (1992) (trillion Btu)

Fuel/Energy Type	Defense	US Government	USA
Motor Gasoline	12	35	13980
Distillate/Residual Fuel	214	236	8870
Jet Fuel	765	775	3080
Other	2	5	7540
Total Petroleum Based	993	1050	33470
%	3%	3%	100%
Electricity	117	190	9440
%	1%	2%	100%
Natural Gas	109	154	20320
%	1%	1%	100%
Coal & Other	51	65	19140
%	0.0%	0.0%	100%
Total	1269	1460	82360
% of U.S. Consumption	2%	2%	100%

Source : EIA. 1992. Annual Energy Review.

Table 4.5.2 Military Fuel Use Forecasts

Transportation Sector Energy Use by Mode and Type - Military Use
(trillion Btu per year)

Year	1990	2000	2005	2010	Annual Growth 1990-2010	1990 % of total market use
Total Military Use	902.1	656.2	660.3	670.4	-1.50%	
Aviation fuel	795	570.1	573.7	582.5	-1.50%	34.05
Residual Fuel Use	15.6	12.7	12.7	12.9	-0.90%	1.54
Distillate Fuel Use	91.5	73.4	73.9	75	-1.00%	33.11

Transportation Sector Energy Use by Mode and Type Within a Mode
(trillion Btu per year)

International Shipping

Distillate (diesel)	62.2	81.74	92.27	102.06	2.50%
Residual oil	920.24	1209.7	1356.11	1509.33	2.50%

Domestic Shipping

Distillate (diesel)	214.19	231.25	246.36	262.96	1.00%
Residual oil	94.37	101.92	108.55	115.82	1.00%

Air Transport

Jet Fuel	2334.55	3147.5	3509.52	3856.3	2.50%
Aviation Gasoline	45.36	42.56	42.26	42.14	-0.40%

Military Use

Jet Fuel (Kerosene)	451.55	334.13	336.23	341.44	-1.40%
Jet Fuel (Naphtha)	343.4	235.95	237.43	241.11	-1.80%
Residual Fuel	15.58	12.66	12.74	12.94	-0.90%
Distillate	91.52	73.41	73.87	75.02	-1.00%

Source: EIA 1994.

4.5.2 Power Generation Technologies that could Lead to Changes in Fuel Consumption Patterns in the Military

A large percentage of military fuel consumption is due to the needs of military bases in terms of space heating and electricity generation for other purposes. The U.S. military has around 275 bases divided between the Army, Navy and Air Force. The fuel needs for these are divided between coal, gas and petroleum products such as distillate and heavy fuel oils.

Alternative methods to reduce fuel cost decrease dependence on imported oil and to improve environmental quality of emissions are discussed below with respect to improvements in energy management, new technologies and alternative fuels.

4.5.2.1 Energy Management Initiatives

The DOD uses approximately \$2.9 billion annually in energy at military installations. The U.S. government's "Energy 2005" initiative calls for a 10% reduction in energy use at military installations by FY 1995 and a 20% reduction by 2005 (1985 is the base level). Industrial energy use is also targeted to reduce by 20% of 1985 levels in 2005.

Since the late 1970's the army has been installing Energy Monitoring and Control Systems (EMCS) at its installations. These are computer systems designed to keep track of and adjust if necessary, energy usage in order to minimize energy costs. Their

development is evolutionary and implementation on large scale of the newer technologies in this field would bring significant savings in energy use due to advances in control technologies and software. MacDonald and Gettings (1991) have studied military use of such systems and concluded that "dramatic changes (in the technology) are possible in the next decade".

4.5.2.2 Flue Gas Desulfurization : “e-scrub technology”

The SERDP (Strategic Environmental Research and Development Program 1992) allocated \$160 million for FY 1994 for the development for military and commercial use of the “e-scrub” flue gas desulfurization device adapted from other military technology; according to DOD (Department of Energy 1994a) this device “may be able to efficiently turn dirty, high sulfur coal emissions into a fertilizer, resulting in cleaner water, air and a commercial product.” This is but one of many flue-gas desulfurization technologies now available.

4.5.2.3 Dry Micronized Coal and Coal/Water Mixture Technology - New Boiler and Handling Technologies

The Department of Defense Appropriations Act 1986 (Public Law (PL) 99-190, Section 8110) directed DOD to rehabilitate and convert its existing domestic power plants to burn a higher percentage of coal. By FY 1994 coal consumption by the military was to have increased to 1.6 million tons per year above the level of consumption in 1985. The FY 87 Defense Authorizations Act directs that the primary fuel source used in any new heating system installed be the most life-cycle cost effective. Thus the DOD cannot install new coal-fueled plants without making sure the plant is the most cost effective over the life of the plant. These laws, together with a directive canceling the restriction that oil or gas heating systems be built on a scale smaller than 50 million Btu/h, have provided impetus for the military to look at conversion of boilers at military installations to coal.

U.S. military coal consumption from 1985 - 1991 is listed in Table 4.5.3. It can be seen that the army is the largest user of coal with the navy second largest and air force is the smallest user. It was recommended by the U.S. Army Construction Engineering

Table 4.5.3 DOD Coal Consumption (short tons)

Year	Army	Air Force	Navy	DOD Total
1985	704190	452242	162095	1336261
1986	733490	491122	168689	1409446
1987	734457	472086	163891	1384984
1988	802218	478092	178227	1471240
1989	768962	448945	222888	1453444
1990	867690	439512	178490	1498740
1991	766725	435563	167232	1381155
(% of DOD)	56 %	32 %	12 %	100 %

Source: Adapted from Lin (1993)

Research Laboratories (USACERL) in 1991 that the Navy had the highest potential for increasing its coal consumption after a study in 1987 identified bases where conversion to coal would be most advantageous.

Work by USACERL (Lin 1994) indicates that there are several choices in achieving this goal once sites for increased coal use have been identified. These include the following choices: to build new plants to replace old ones; to retrofit oil and gas boilers for coal firing; or to build third party coal fired cogeneration plants in conjunction with commercial power companies. Various technologies are available for retrofit projects (Miller et al., 1994). Briefly these are coal/water mixtures, dry micronized coal, coal gasification, retrofit fluidized-bed combustors, and slagging combustors.

So far, USACERL work has concluded that the building of new coal-fired plants to replace old oil/gas fired plants is economic at only one site. Retrofitting heavy oil plants for coal firing was found to be an attractive strategy for 38 plants of the 88 studied.

4.5.2.4 Fuel Cell Technology

The underlying principle of fuel cells has been known about for over a hundred and fifty years but the technology has only been proposed recently for commercial use. The principle is similar to how a battery works, but, unlike a battery, it continues to provide power so long as fuel and oxidant are supplied. It works by chemically reacting air/oxygen with a fuel such as hydrogen processed, using a reformer, from natural gas. This produces a direct chemical transfer of energy as a DC current which is then

transformed to an AC current for use. The method of reaction means that energy efficiency is extremely high relative to thermal power plants. Newer plants are expected to produce efficiencies of 50-60% compared to 30-40% at thermal plants. Other advantages are flexibility due to their modular construction, quietness, reliability (allowing reduction in backup systems required), and environmental advantages such as extremely low NO_x, CO and unburned hydrocarbons emissions.

Fuel cells require a feedstock that enables the release of hydrogen. Alternatives include all fossil fuels. Coal gasification is among those methods currently being investigated for fuel production. Natural gas is probably the most convenient feedstock. Research and development is currently going on with respect to four types of technology based on the type of electrolyte used in the cell (see Table 4.5.4): phosphoric acid, molten carbonate, solid oxygen and proton exchange.

Canada's military has made a strategic decision to investigate fuel cell technology and has contracted Ballard Power Systems, Inc. to investigate polymer electrolyte fuel cells and their operation on air and reformat fuels. The results of the program are a demonstration program with power outputs up to 5 kW and hydrogen-oxygen and hydrogen-air low kW units have been delivered to customers. The Canadian department of defense is also sponsoring a program to demonstrate the use of 100 MWh and 300 kW Ballard fuel cells for submarine propulsion to replace diesel engines currently used. A 50 kW scaled version is being developed by Ballard that will operate on methanol and oxygen or methanol and air (on surface).

At present, units range from very low output (<10 kW) to 200 kW. In the very near future units will be available up to 10 MW. This range will probably find many uses in the military. U.S. military research has been active since the 1980's by participating in DOE, GRI (Gas Research Institute) and utility companies' field tests of 40 kW phosphoric acid fuel cell units produced by International Fuel Cells Corporation. In FY 1993, the DOD initiated a program to demonstrate and use commercially produced fuel cell power plants at sites located at military bases. DOD has also initiated a program to develop fuel cell power plants for defense applications; particularly in order to develop units fueled by standard logistic diesel or jet fuel.

Table 4.5.4 Summary of Worldwide Fuel Cell Development Progress

	Fuel Cell Status						
Country	Phosphoric Acid (PAFC)	Molten Carbonate (MCFC)	Solid Oxide (SOFC)	Proton Exchange (PEFC)	Operating Capacity (kW)	Reformer made ?	Approx. Funding (\$ U.S. Million/yr)
Austria	Dn ¹				200		-
Australia			R,Dt				5 Total
Canada	Dn ¹			R,Dt,Dn	325		Unknown
Denmark	Dn ¹	R,Dn ¹	R,Dt	R	200	Y	4 Total SOFC
Finland	Dn ¹			R	200		-
France		R	R	R			2 Total PEFC
Germany ²	Dn ⁴	R,Dt	R,Dt	R,Dt	800		6 Gov., No PAFC
Italy	R,Dn ¹	R,Dt	R	R,Dt	1500		4 Gov
Japan	R,Dt,Dn ⁴ ,C	R,Dt	R,Dn ⁵	R	+/- 40,000	Y	46 Gov.
Korea	R,Dn ⁴	R,Dt	R		200	Y	Unknown
Netherlands		R,Dt	R				10 Gov.
Norway			R,Dt				2
Spain	Dn ³	R,Dt			200		1 No PAFC
Sweden	Dn ¹	R			200		2 Gov., No PAFC.
Switzerland	Dn ¹		R,Dt		200		2 Gov., No PAFC
United States	R,Dt,Dn,C	R,Dt,Dn	R,Dt,Dn	R,Dt,Dn	+/- 4500	Y	104 Gov. & Institutes

R Research

Dt Development

Dn Demonstration

C Near Commercial or Commercial Sales

Notes:

1. U.S power plant(s)
2. Germany demonstrated a 100 kW alkaline fuel cell system in a submarine
3. Japanese power plant(s)
4. U.S and Japanese power plants
5. Westinghouse (U.S) SOFC modules

Belgium is demonstrating an alkaline fuel cell unit in a bus.

Source: After USDOE report prepared by Gilbert/Commonwealth Inc.(1993)

The technology seems to be rapidly developing. It has a wide variety of applications due to the different scales in which it can be applied. Environmentally, the technology provides extremely low emissions energy. It is deemed more reliable and flexible than conventional thermal power/boilers. It is also more fuel efficient, therefore, implementation would result in lower fuel consumption. Its major drawbacks are that it is still very high cost technology. Current capital costs for natural gas fueled plants are estimated at between \$600 – 1,000/kW, and for coal fueled plants are \$1,000 – 1,200/kW.

4.5.2.5 Combined Cycle Gas Turbine

The waste heat generated in burning oil or gas in a boiler is used to run a separate turbine to increase energy efficiency. The technology has been used for many years. As such it is proven technology and, with recent advances in the technology, represents an option for new plants because it is suited to utilization of gas (now more abundant at low cost). Coal can be adapted to use in this technology through gasification of the coal and subsequent burning of the gas in the turbine (Rose et al., 1991).

4.5.2.6 Other Clean Coal Technology Programs

The Office of Clean Coal Technology (part of the U.S. Department of Energy) sponsors, in partnership with industry, a wide range of clean coal technology programs which apply to research and demonstration at various levels in the processing of coal to energy such as preparation (e.g., coal washing process developed for Phase I, Miller et al., 1994), transformation (e.g. production of “Syncoal” by the Rosebud Syncoal Partnership (U.S. Department of Energy 1994a), combustion (Miller et al., 1994), post-combustion (the SNOX process, DOE 1994a) and other relevant areas.

The Federal Government has invested \$2.5 billion in the Clean Coal Technology Program; other contributions from industry and individual states have amounted to more than \$4 billion (U.S. Department of Energy 1994b). As a result, 45 new demonstration projects are underway (U.S. Department of Energy 1994c). Advanced technologies are being tested in emissions reduction using scrubbers for NO_x and SO_x gases, clean coal Fuels, clean coal fluidized-bed technologies, integrated combined cycle gasification projects and other innovations. Earlier demonstration results are now being used in the

design and construction of larger scale commercial facilities. It is expected that most of the 45 projects will have provided sufficient data for full scale commercialization decisions in the early years of the next century (i.e., in 5 - 15 years time) (U.S. Department of Energy 1994d).

The current drive to commercialize these technologies could affect military choice of fuel in favor of coal as the military may place more urgency on strategic considerations.

4.5.3 Transport: Technology that could Lead to Changes in Fuel Consumption Patterns in the Military

The consumption of various types of fuels could be affected by technological change in two main ways: 1) fuel economy changes and; 2) substitution to other fuels. Table 4.5.5 presents the types of transportation used in the military that are considered in this analysis. The currently used fuels are noted as well as possibilities for conversion either physically or by re-equipping the fleet with new engines/vehicles.

4.5.3.1 Alternative fuels

Currently, conventional fuels are those most widely in use including gasoline and diesel for land based vehicles and residual oil and distillate for ships and kerosene/naphtha for aircraft. DOD is now under pressure to have their consumption reduced for two main reasons: 1) to reduce reliance on imported oil (a strategic issue); and 2) to reduce harmful emissions caused by their use, especially those used in land-based applications, for environmental reasons.

Alternative fuels have characteristics that make them more attractive than conventional fuels on strategic and/or environmental grounds. Table 4.5.6 lists currently known alternative “fuels” in the broad sense (including nuclear and electric technologies), the conventional fuels that they replace and the technologies that are now being developed to exploit them in either the military or private/civilian sectors.

In terms of projects already under way the DOD has undertaken to convert a portion of its 200,000 administrative vehicles (assumed to include all buses, trucks, LDV's, automobiles and other such) to “alternative fuels”. By FY 1995 the DOD has undertaken to acquire 10,000 alternative fuel vehicles.

Table 4.5.5 Possibilities for Conversion of Vehicles to Alternative Fuels

Mode of transport	Main Fuel	Possibilities for Conversion
Light vehicles	diesel/gasoline	CNG,LNG,LPG,M-85,M-15,E-95,E-100,M-100, electricity, H-Fuel Cell
Buses	diesel/gasoline	CNG,LNG,LPG,M-85,M-15,E-95,E-100,M-100, electricity, H-Fuel Cell
Trucks	diesel/gasoline	CNG,LNG,LPG,M-85,M-15,E-95,E-100,M-100, electricity, H-Fuel Cell
Ships	distillate residual fuel - oil ? nuclear	Fuel cells, CNG,LNG,
Jet aircraft	jet fuel JP-8, JP-4, (kerosene, naphtha)	Coal gasification derived JP-8
Helicopters/prop aircraft	kerosene	Coal gasification derived JP-8
Tanks & other armored vehicles	diesel ?	

4.5.3.2 State and Federal Government Initiatives and Legislative Aspects

1970s oil price increases and current environmental concerns have lead to governmental sponsoring of and influence over efforts to introduce alternative fuel use. The Clean Air Act Amendments of 1990 set goals for improvement of air quality especially concerned with CO (carbon monoxide) and O₃ (ozone) emission reduction in the transport sector. The law is viewed as favorable towards alternative fuels such as chemically simpler fuels (e.g., methanol and methane).

The Energy Policy Act of 1992 (EPACT) was passed in the wake of the Persian Gulf War of 1990-1991. This act set out to decrease independence of the U.S. on imported oil. The act targets fleets of vehicles for increased use because of the costs of conversion, refueling and fuel storage facilities. EPACT mandates acquisition targets for alternative fueled vehicles for Federal and State agencies.

Table 4.5.7 illustrates how, under EPACT, the federal government (including the DOD) will be required to increase the numbers of alternative fuel vehicles rapidly in the next 5 years. 75% of all new federal vehicles acquired in 1999 will have to be alternative

fuel powered vehicles. New military vehicles will undoubtedly be affected by these regulations.

Table 4.5.6 Alternative Fuels and Technology Developments for Transportation

Alternative fuel	Conventional fuel replaced	Technologies available	Development stage
Compressed natural gas (CNG)	Gasoline Diesel	Light Duty vehicles: Buses and other heavy duty vehicles.	Demonstration
Liquefied natural gas (LNG)	Gasoline Diesel	Light Duty vehicles: Buses and other heavy duty vehicles.	Demonstration
Liquefied Petroleum gas (LPG) (Propane)	Gasoline Diesel		Demonstration
Methanol/gasoline (M-85)	Gasoline Diesel	Cars, trucks, heavy duty vehicles	Demonstration/ Commercial
Methanol (M-100)	Gasoline Diesel	Cars, trucks, heavy duty vehicles	Demonstration/Co mmercial
Ethanol (E-95, E-100)	Gasoline Diesel	Cars, trucks, heavy duty vehicles	Demonstration/Co mmercial
Ethanol/Gasoline (E-10 ¹ , E-15 ¹ , E-85)	Gasoline	Cars, trucks, heavy duty vehicles	Demonstration/Co mmercial
Electricity	Gasoline Diesel	Cars, trucks, heavy duty vehicles	Demonstration/Co mmercial
Reformulated gasoline ²	Gasoline Diesel	Cars, trucks, heavy duty vehicles	Demonstration/Co mmercial
Fuel cell	Gasoline Diesel	Methanol fueled bus - Mercedes, GM, Chrysler - also cars, trucks and plans for locomotives	Demonstration

Notes:

¹These fuels are statutory not defined as “alternative” in EPACT but are defined as “replacement”

²These fuels are aimed at lowering emissions regardless of source (and thus strategic influence).

Source: EIA. June 1994.

4.5.4 Conclusions

It is shown that for both power/heat generation and transport applications there is a wide variety of possible alternatives that could be applied for military use. Many new coal technologies such as dry micronized coal, fluidized bed, flue-gas desulfurization, coal gasification, combined cycle and advanced combustors are currently being implemented and designed (Miller et al., 1994).

Table 4.5.7 Alternative Fueled Vehicle Targets for the Federal Fleet as proposed in Two Pieces of Legislation.

Fiscal Year	EPACT Requirements	Executive Order 12844
1993	5,000	7,500
1994	7,500	11,250
1995	10,000	15,000
1996 ¹	25% of acquisitions	-
1997 ¹	33% of acquisitions	-
1998 ¹	50% of acquisitions	-
1999 ¹	75% of acquisitions	-

Notes:

¹Percentage of acquisitions in these years is based on light duty vehicle acquisitions. The Executive Order did not provide AFV targets or percentage acquisition for these years.

Source: EIA. June 1994.

Energy use in the military could also be significantly affected by new energy management systems and implementation of modified, more efficient, conventional technology.

Fuel cells appear to be a fast emerging technology for both generation and transport applications, with small scale power applications currently closest to proven commercial use. Numerous commercial and demonstration programs are underway regarding use of alternative fuels in transportation. Many of these, including fuels for fuel cell use, could use coal as a feed stock if processed by gasification or liquefaction either on site or at processing plants prior to distribution.

4.6 Constructing a National Energy Portfolio which Minimizes Energy Price Shock Effects

4.6.1 Introduction

Investors have long sought to reduce their exposure to market swings by holding a diverse mix of investment instruments, which can reduce investor risk dramatically. By examining the variance, covariance and expected return between a group of assets, Markowitz (1952) constructed an efficient portfolio that maximized expected return for a

given level of risk. The dual goal of maximizing return while minimizing variance produces a set of efficient portfolios an agent can choose from according to his or her personal preferences.

Portfolio theory and diversification have proved useful in areas other than personal and corporate investing. Adegbulugbe et al. (1989) examine the long-term optimal structure of the Nigerian energy supply mix. They use a multiperiod linear programming model of the total energy system to minimize direct fuel costs while maintaining balanced development. Bar-Lev and Katz (1976) examined fuel procurement in the electric utility industry and created a Markowitz efficient frontier of fuel mixes which minimizes expected fuel costs and risk. Electric utilities rely heavily on long-term contracts with price adjustment clauses for fuel supplies. The possibility of price increases (or decreases) creates an atmosphere of uncertainty which must be factored into the fuel procurement decision, not unlike the problem of selecting risky securities. Kroner and Claessens (1991) show how diversifying the composition of a country's foreign debt can act as a hedge against changes in exchange rates and commodity prices. The conditional covariance matrix of exchange rates and terms of trade, which changes through time, determines the optimal portfolio.

This section uses portfolio theory to demonstrate how the energy mix consumed in the United States could be chosen if the goal is to reduce the risks to the domestic macroeconomy of unanticipated energy price shocks. An efficient portfolio frontier of U.S. energy consumption is constructed using time-varying variances and covariances estimated with generalized autoregressive conditional heteroskedastic models. This allows the efficient portfolios to shift over time in response to price changes and past volatility. A trade-off between risk and cost exists. The set of efficient portfolios developed are intended to minimize the impact of price shocks, but are not the least cost energy consumption bundles. Although linear or quadratic programming techniques could also provide estimates of a unique optimal energy mix, the portfolio methodology is superior because it creates a set of efficient energy mixes, which vary in response to market events, and leaves the final choice to policy makers.

The United States government, through national energy policy, has long attempted to decrease the domestic economy's exposure to the impact of an energy shock.

Such attempts have included quotas on imported oil, greater self-sufficiency in energy, increased research into alternative fuel sources, the strategic petroleum reserve, and increased domestic coal consumption. Underlying this goal was the belief that decreasing the heavy dependence on oil, and foreign oil in particular, would limit the potential effects of a price shock from a single energy source. As recently as 1993, Secretary of Energy for the Clinton Administration Hazel O'Leary called for further diversification of U.S. energy consumption.

A negative correlation between oil prices and real output has been well established empirically. Fried and Trezise (1993) state that the 1973 oil shocks increased the general rate of inflation by two percentage points, led to a term of trade loss for oil importing countries equal to 1.5 to 2% of the GNP, and decreased economic output by two to three percent within OECD countries. Rasche and Tatom (1981) estimate a 7.0% average long-run reduction in U.S. GNP following the 1973-1976 oil shocks, Darby (1982) estimates a 2.5% decline, and Hamilton (1983) confirms the negative relationship between oil prices and real output for the U.S. both before and after the price increases of the 1970s. Mork et al. (1994) found this correlation persists in data through 1992, including the Persian Gulf War of 1991.

The goal of minimizing exposure to foreign oil price shocks by reducing oil imports ignores the possible covariance relationships between different energy sources. Shocks in the oil market can spill over to the coal and natural gas markets, and the relationship between domestic and foreign oil prices is strong and positive. Thus, establishing a goal of oil independence will not buffer the U.S. economy from foreign oil price shocks unless accompanied by the total isolation of the entire domestic energy market from the world market. As Heal (1992) notes, a choice of isolation would create artificially high energy prices domestically and be economically inefficient. A better solution is to look for those combinations of oil, natural gas, and coal where price volatility in one fuel is offset by opposing volatility in another fuel, as portfolio diversification and management are designed to do.

4.6.2 Methodology

The primary goals of a rational investor are to maximize expected return and minimize risk. Higher risk is usually compensated for by higher expected returns. By purchasing a combination of assets, an investor can generate the highest possible return to their investment dollars at the least possible risk. The expected return of such a portfolio depends on the expected returns of the individual assets and the relative percentage of funds invested in each. The risk of the portfolio depends on the degree of risk of the assets and the covariance or correlation of the assets. While the correlation has no impact on portfolio return, it plays an important role in determining the risk associated with the portfolio. The lower the correlation between two assets, the greater the risk reduction potential when combining them in a single portfolio.

The expected return for an n -asset risky portfolio is

$$E(R_p) = \sum_{i=1}^n w_i E(R_i) \quad (4.6.1)$$

where $E(R_i)$ is the expected return from asset i and w_i is the weight of asset i held in portfolio p . The corresponding risk of such a portfolio, as measured by its variance, is

$$\sigma_p^2 = \sum_{i=1}^n \sum_{j=1}^n w_i w_j \text{cov}(i, j) \quad (4.6.2)$$

where $\text{cov}(i, j)$ is the covariance between two risky assets i and j . By varying the weights of the assets, a set of potential portfolios can be generated. The efficient portfolio frontier is the subset of all dominant portfolios from the set of all possible portfolios. In a dominant portfolio, return cannot be increased while holding variance constant and variance can not be decreased while holding return constant. Assuming risk aversion in decision makers, rational choice should lead to investment decisions only on the efficient frontier.

The return and variance of any portfolio can be calculated and the efficient frontier generated as long as the expected return of each asset and the covariance matrix

is known (equations 4.6.1 and 4.6.2). In their analysis of the electric utility industry, Bar-Lev and Katz (1976) use a quadratic programming approach to estimate these values. However, such a methodology requires the covariance matrix to remain constant over time, regardless of any new information which enters the market. Allowing the covariance matrix to be systematically updated over time as new events occur provides more realistic and efficient estimates. The generalized autoregressive conditional heteroskedasticity process (GARCH) introduced by Engle (1982) and Bollerslev (1986) allows the error variance and covariance to respond to price shocks and changes in volatility. GARCH has been used extensively for modeling financial time series. Bera and Higgins (1993) and Bollerslev et al. (1992) provide surveys of the GARCH literature.

The GARCH(p,q) process is defined in terms of the properties of the distribution of the error terms in the model

$$Y_t = X_t' \vartheta + \varepsilon_t \quad \varepsilon_t \text{ i.i.d.}, \quad t=1, \dots, T \quad (4.6.3)$$

where X_t is a $k \times 1$ vector of exogenous variables which can contain lagged dependent variables, ϑ is a $k \times 1$ vector of regression coefficients, and ε_t is the error term. ε_t is assumed to have a traditional constant unconditional variance, but its conditional distribution depends upon the information set ψ_{t-1} , or

$$\varepsilon_t | \psi_{t-1} \sim N(0, h_t) \quad (4.6.4)$$

where

$$h_t = c_0 + \sum_{i=1}^q \alpha_i \varepsilon_{t-i}^2 + \sum_{j=1}^p \beta_j h_{t-j}. \quad (4.6.5)$$

The conditions ensuring non-negativity and stationarity can be found in Nelson and Cao (1992). The conditional variance h_t is a function of the information set ψ_{t-1} , and the underlying functional form effectively models the volatility clustering commonly seen in financial time series data. The generalized GARCH(p,q) model allows the present

conditional variance to depend on past conditional variances in addition to past squared innovations.

Computing time-varying covariances requires the simultaneous estimation of a multivariate system (Bollerslev et al., 1988). Generalizing a univariate GARCH model to a multivariate model requires allowing the entire covariance matrix to change with time, rather than just the variance. Thus, the elements of the covariance matrix would be linear functions of lagged squared errors, lagged cross-products of the errors, lagged variances, and lagged covariances. The traditional means of estimating such a system, similar to a vector autoregression, is cumbersome and sometimes impossible. Bollerslev (1990) has suggested a multivariate GARCH specification which holds the correlation matrix constant over time. Using the relationship between covariance and correlation,

$$h_{ij,t} = \rho \sqrt{h_{i,t} h_{j,t}} \quad (4.6.6)$$

each period's covariance can be calculated. This allows the variance and covariance to change over time while the fundamental relationship between assets remains unchanged. The constant correlations model has been used successfully with foreign exchange rate data (Bollerslev 1990) and interest rate data (Cecchetti et al., 1988). Giovannini and Jorion (1989) find the estimated variances from a constant correlation estimation are almost perfectly correlated with the estimated variances from the VAR-type estimation, and Baillie and Bollerslev (1990) show that the data does not usually reject the constant correlation assumption.

The analysis of energy portfolios differs slightly from investor portfolio analysis. Since every energy investment has risk, no risk-free energy asset is available. Another difference is the inability of energy investors to hold negative weights of, or "go short on," an energy type. The equivalent to shorting stock in energy investment would be using electricity to create natural gas and then selling the natural gas, but it is impossible to efficiently create natural gas from electricity.

4.6.3 Estimation and Results

The data set consists of monthly spot energy price series of oil, natural gas, and coal. Over the next ten years, 85% of new non-nuclear electric generation in the U.S. is projected to be from these three sources. Further investment in nuclear energy production has been stalled and much current production will be phased out over the next 20 years. The share of hydroelectricity and other non-conventional energy sources is not expected to grow over the next decade. Few new potential hydro power sites exist in the United States, and non-conventional energy sources are not currently competitive on a large scale and will likely not be so over the next ten years.

Two price series were obtained for each energy source, both nominal and not seasonally adjusted. Basic energy costs were derived from average monthly first nearby oil prices and natural gas prices from the NYMEX and average wellhead price for periods before NYMEX trading. The coal price series is the average spot price paid by electric utilities as reported to the Federal Energy Regulatory Commission. All prices are in terms of monthly million Btu/\$. The second set of price series adds electric utility non-fuel costs to basic energy costs to generate a total operating cost for electricity generation from each energy source. The data extend from January 1974 to August 1995 (260 observations) and are plotted in Figure 4.6.1.

Each series is tested for the presence of a unit root. A series with a unit root is nonstationary with an infinite unconditional variance. The results of Dickey-Fuller and augmented Dickey-Fuller tests are reported in Table 4.6.1. All series fail to reject the null hypothesis of a unit root at the 95% confidence level. Thus, to achieve stationarity in the data, all series were differenced once.

Table 4.6.2 presents the historical moments of each series. The spread between the average MM Btu/\$ (column 1) is smaller in the utility energy cost group than the raw energy cost group. This reflects the inverse relationship of fixed costs to variable costs. In general, oil has provided the least MM Btu/\$ for an investment while natural gas has provided the most. Examining the first differenced log for each series, the mean return is not significantly different from zero and all exhibit skewness and excess kurtosis. This is reflected in the high Bera-Jarque statistic, which rejects the null of a normal distribution at 0.05% in all cases.

Figure 4.6.1 HISTORICAL ENERGY RETURN (MM BTU/\$) FOR EACH ENERGY TYPE

Table 4.6.1 Unit Root Tests

	AR	DF	ADF	Number of
Parameter	Parameter	Statistic	Statistic	Significant Lags
Coal	0.9544	-2.748	-2.748	0
Natural Gas	0.9754	-2.148	-2.120	12
Crude Oil	0.9762	-1.788	-2.269	1

The 5% and 10% critical values for Dickey-Fuller tests are -2.84 and -2.57. Lags were considered significant at the 10% level.

Table 4.6.2 Historical Moments of Price Series

	Average MM Btu/\$	Differenced Log MM Btu/\$				
		Mean	Variance	Skewness	Kurtosis	Bera- Jarque
<u>Raw Energy Costs</u>						
Coal	0.76	-0.0021	0.0021	-0.98	6.1	146
Oil	0.38	-0.0036	0.0068	-0.84	12.6	1016
Natural Gas	0.95	-0.0077	0.0068	0.79	8.4	336
<u>Utility Energy Costs</u>						
Coal	0.43	-0.0010	0.0007	-0.64	4.3	36
Oil	0.28	-0.0026	0.0043	-0.68	11.7	842
Natural Gas	0.53	-0.0040	0.0035	0.89	9.0	425

Table 4.6.3 reports the correlation matrix between the price series. The correlations are generally small, with the relationship between coal and both oil and natural gas weak whether or not non-fuel costs are included. Natural gas and oil have a slightly larger though still small correlation coefficient. The relationship between natural gas and oil in the last five years of the data set (1990-1995) has become stronger, with a correlation coefficient of 0.40, probably because of the deregulation of natural gas.

Table 4.6.3 Correlation of Changes in Monthly Energy Return (MM Btu/\$) Series
1973-1995

	Basic Energy Costs			Electric Utility Operating Costs		
	Coal	Natural Gas	Oil	Coal	Natural Gas	Oil
Coal	1			1		
Natural Gas	0.02 (0.46)	1		0.02 (0.46)	1	
Oil	-0.01 (-0.16)	0.13 (3.61)	1	-0.01 (-0.18)	0.12 (3.47)	1

The GARCH parameter estimates for equations 4.4.3 and 4.6.5 are found in Table 4.6.4. The significance of $\hat{\alpha}$ and $\hat{\beta}$ in the variance equation (equation 4.6.5) supports the choice of a GARCH model in place of a homoskedastic system. Predicted volatility for each of the energy price series based on the parameter estimates is plotted in Figure 4.6.2. Coal prices are the most stable, with a relatively constant conditional variance and return. Oil shows significant volatility increases during each historic price shock (i.e., 1979, 1986, 1991) and an overall trend to increasing volatility in the 1990s. These price shocks affect oil volatility for approximately one year. Natural gas shows the most marked change. Returns to investment have declined significantly as prices have increased over the last twenty years. Natural gas prices have also been very volatile in recent years, reflecting the deregulation of the natural gas industry and the development of active natural gas futures markets.

Combining the parameter estimates and the correlation coefficient estimates according to equation 4.6.6 yields estimates of the time-varying covariances of the system. Energy portfolios are constructed using these covariances and expected returns in equations 4.6.1 and 4.6.2. Figure 4.6.3 depicts the efficient portfolio frontier in 1990 for both basic energy costs and electric utility operating costs. The addition of fixed costs to energy costs lowers returns to electric utilities and results in a lower curve. The additional fixed charges also mean that the annual volatility becomes smaller relative to the absolute size of the costs, shifting the curve to the left.

When viewing the actual energy portfolios of U.S. consumption for 1980, 1990, and 1995 along with an example of an efficient portfolio from the estimated portfolio frontier for each year, it is noted that the actual energy consumption portfolio is far from efficient. The efficient portfolio Sharpe ratio is two to three times the size of the actual portfolio ratio. In general, a more efficient consumption bundle would include a higher percentage of coal and less oil and natural gas in the overall economy. Increasing coal consumption would both decrease volatility and increase energy return, a movement to a dominant portfolio. It is acknowledged that the effective price of utilizing fuels differs from the market price somewhat. Given market imperfections, relatively higher costs of pollution mitigation associated with coal may not have properly been reflected, and similarly for uncertainties related to deregulation of natural gas.

Table 4.6.4 GARCH Estimation of Energy Return Series (T-statistics in parenthesis)

	Mean Equation Eqn. 9-3		Variance Equation Eqn. 9-5	
	δ	c	α	β
<u>Raw Energy Costs</u>				
Coal	0.001 (0.61)	0.0001 (2.81)	0.068 (2.94)	0.72 (9.58)
Natural Gas	-0.014 (-6.22)	0.00002 (2.40)	0.096 (4.57)	0.83 (26.87)
Oil	-0.005 (-1.75)	0.0008 (5.86)	0.183 (4.47)	0.40 (4.65)
<u>Electric Utility Operating Costs</u>				
Coal	0.001 (0.70)	0.00003 (2.42)	0.077 (3.02)	0.72 (9.41)
Natural Gas	-0.010 (-6.49)	0.00005 (4.17)	0.208 (5.08)	0.62 (11.70)
Oil	-0.004 (-1.63)	0.00047 (6.02)	0.194 (4.78)	0.41 (5.15)

Annual Energy Volatility

Figure 4.6.2 GARCH ESTIMATES OF ENERGY RETURN VOLATILITY

Figure 4.6.3 1990 EFFICIENT PORTFOLIO FRONTIERS FOR THE UTILITY INDUSTRY AND GENERAL U.S. CONSUMPTION

For the total costs faced by the electric utility industry (energy costs plus nonfuel costs), operation has moved towards more efficient portfolios since 1980. In 1990 and 1995 the industry operated very close to the efficient frontier. The industry also operated at a position of very low volatility. The high level of coal consumption by utilities suggests a relatively high level of risk aversion, in contrast to the results of Bar-Lev and Katz (1976), who found electric utilities pursuing high-risk consumption strategies. They argued that this was the result of regulatory policy that allowed electric producers to pass cost through to the consumer. However, the price shocks and volatility in the oil market since their study might have shown the limit of such regulatory largesse and have led to the observed increase in risk aversion seen by the utility consumer.

As the variance and covariance changes over time in response to price shocks and past volatility, the efficient portfolio set also changes. Some general observations can be made. Certain relationships remain unchanged. Coal is more stable than either oil or

natural gas. Natural gas tends to be more volatile than either coal or oil, with superior returns to the electric utility industry and lower returns in the overall energy portfolio.

4.6.4 Policy Recommendations

The actual energy portfolio selected should be determined by the tangency of the efficient portfolio frontier and the utility curve which reflects the country's preferences regarding risk and return. Some general recommendations can be offered. The volatility (but almost certainly not the level) of overall U.S. energy consumption costs could be lowered (and expected returns raised) by increasing the percentage of coal within the overall consumption mix. Lower volatility should lead to fewer disruptions in the domestic macroeconomy. The actual level of each energy source should be determined by feasibility and the utility curves of the country. Shifting as far as 70% coal consumption (Table 4.6.4, 1990) would be physically impossible since coal is best used in boilers, which account for far less than 70% of U.S. energy consumption. Choice is valuable because the ability to easily switch between fuels to adjust the consumption mix allows the fuel mix to move with the efficient frontier in response to market events. Discovering ways to move cheaply from one fuel to another would be desirable.

The recommendation to shift energy consumption towards coal rests on the premise that future coal price movements are similar to historic prices. This may not be the case. The deregulation of the electric utility industry and potential introduction of coal futures contracts may lead to an increase in coal price volatility similar to that seen in the natural gas industry. However, for the near future, coal prices should remain relatively constant.

Although transparent future markets exist in oil and natural gas and there exists an active coal market, the market price of energy does not reflect the true cost of energy consumption due to the presence of externalities. Environmental, national security, monopoly, and depletion costs may not be adequately included in the market price. These costs can be incorporated into the analysis by including an estimated externality cost in the total cost of an energy source. This creates a new set of variances, covariances, and expected returns, and thus a new set of efficient portfolios from which to choose an efficient portfolio. The portfolio framework also allows selection of

portfolios for sectors that may face different externality costs. The sector would construct an efficient portfolio frontier using their own externality costs and then choose from this frontier. Even with the use of portfolio analysis to select a consumption bundle, a remaining inefficiency may arise due to over-consumption of energy in general. Substitution away from energy should take place as its costs rise.

As an illustration, consider the environmental costs associated with energy consumption. Many estimates have been made of the environmental costs of energy consumption, optimal tax rates, and carbon taxes (e.g., Viscusi et al., 1989; Verleger 1993). Barbir and Veziroglu (1991) estimate that the total costs of energy consumption in 1990 were \$2.36 trillion dollars with estimated externality costs of \$9.31/MM Btu for coal, \$8.03/MM Btu for oil, and \$5.31/MM Btu for natural gas. Adding these cost estimates to the energy cost data creates a new set of efficient portfolios, illustrated in Figure 4.6.4 for 1990. Including environmental costs greatly decreases energy returns and gives natural gas a higher return than coal. Although the minimum variance portfolio is the same in both this and the previous analysis (64% coal, 19% oil, and 17% natural gas), every other point on the frontier is different. Ignoring environmental costs leads to a decision to increase coal consumption to reduce volatility. Including environmental costs shifts consumption away from coal towards natural gas in order to increase the portfolio's return. Because the analysis is highly sensitive to the estimates of externality costs used, reliable estimates are essential.

Of frequent political interest is the potential security premium representing the benefit of self-sufficiency in energy needs. Like the disagreement which exists over the size and actual impact of energy price shocks on the domestic economy, discrepancies also exist in the size and even existence of a security premium. Bohi and Toman (1996) provide an overview of security premia estimates from the Department of Energy which range from \$0.17/bbl of oil to \$10/bbl of oil. Heal (1992) estimates that between additional defense and strategic petroleum reserve spending, the national security expenditure in 1985 was approximately \$9/bbl of imported oil.

To illustrate the effect of a security premium on portfolio choice, \$10/bbl is added to the total oil price. Figure 4.6.5 illustrates the resulting efficient frontier. Again, the minimum variance portfolio remains unchanged, but for the same expected return a

Figure 4.6.4 1990 EFFICIENT PORTFOLIO AND EFFICIENT ENVIRONMENTAL PORTFOLIO

different portfolio composition arises. On average, the portfolios including a security premium include 5% less oil than the portfolios based solely on market costs. Groups such as the U.S. military, who may place a higher premium on the use of imported oil, would want to shift even more of their oil consumption towards coal.

4.6.5 Conclusions

This section has introduced a method for choosing efficient energy mixes which would reduce the risk to the domestic economy of energy price shocks. A frontier (range) of possible portfolios is generated from which policymakers must choose a desired portfolio based on their risk and expected returns preferences. The results indicate that the electric utility industry is operating very close to the minimum variance position with a risk aversion strategy. In contrast, overall energy consumption in the United States is far from an efficient mix. A shift towards coal consumption would reduce price volatility. With the inclusion of potential externality costs, the shift remains

Figure 4.6.5 1990 EFFICIENT PORTIFOLIO WITH AND WITHOUT A \$9.00/BBL SECURITY PREMIUM

away from oil but towards natural gas instead of coal. To achieve such shifts, policymakers could use regulation or tax incentives to industries to encourage the use of certain fuels. Of course, a minimum risk portfolio does not imply a minimum cost portfolio, and selecting a low-risk portfolio may lead to higher energy costs overall. The costs associated with an occasional energy price shock may be far less than the cost associated with energy independence or a dramatic shift towards coal, synfuels, and other alternative energy sources.

4.7 Proposed Research on the Coal Markets and their Impact on Coal- Based Fuel Technologies

4.7.1 Historical and Prospective Changes in Electric Power Coal Use 1947-1993

The following discussion is based primarily on data taken from the Statistical Yearbook for the Electric Utility Industry (Edison Electric Institute 1977), Trends in

Electric Utility Industry Experience (National Coal Association 1960), and Electric Power Annual (U.S. Department of Energy 1986a, 1986b, 1993a, 1993b, 1993c). Several factors in fuel markets as well as in government regulation affected the choice of fuels. Among the major fuels, the tradeoff is between fuel and capital costs. Coal and nuclear power have relatively low fuel costs but require large capital investments. Natural gas and oil have relatively more expensive fuel costs but have lower capital costs.

Three major factors affecting fuel choice are: first, fuel prices differ markedly in different regions of the United States therefore there are regional differences in fuel choice; second, intervention by the U.S. government via Federal power projects, general regulations of specific fuels, regulations affecting fuel use in electric utilities and air pollution regulations; and third, the changing status of the world oil market.

The choice of fuels in the electric utility industry has been based mainly on fuel cost and capital costs. Over time, regulatory pressures have increased capital and operating costs, particularly in nuclear and coal-powered power plants. Natural gas has been the dominant fuel only in the West South Central region and California. Historically, these regions used gas largely produced within the region. Oil-based generation saw a dramatic increase after the deregulation of imports of residual fuel oil in 1966 then a rapid decrease after the 1973 and 1978 OPEC oil shocks.

Hydroelectric generation has a declining share of generation. Capacity expansion failed to keep up with demand growth except in the Northwest region.

A principal regulatory impact on the choice of fuels in the electric utility industry is the Clean Air Act of 1963 and its subsequent amendments. The act is a disincentive to coal use and promotes natural gas in electricity generation.

4.7.1.1 Air Quality Legislation

The main areas initially affected by air quality legislation were the Northeast and Midwest area of the U.S. while other areas, such as the Western and Southern regions were not as affected. However, from the early 1970s to the 1990s, coal's share of generation grew dramatically in the West South Central region. In contrast, the Eastern region of the U.S. saw little or no growth in coal's share of generation. Texas became a large producer of lignite. This, and a move to Wyoming subbituminous coals meant a

shift away from gas and oil towards coal as a major fuel source. In the other West South Central states, the pattern was similar but not as dramatic.

California has instituted its own air quality regulation before the Clean Air Act of 1963 restrictions went into force due to severe air quality problems. These regulations and other policies in effect precluded California from using coal to generate electricity. New York state and New Jersey also instituted air quality legislation that made coal generation undesirable and thus favored less polluting fuels such as natural gas and oil (Sanders 1983).

4.7.1.2 Influence of Regulations on Fuel Use

A complete review of the influences of regulations and cost factors on fuel selection, i.e., natural gas, nuclear power, hydroelectric power and oil, in electric utilities is presented in an unpublished master's thesis (The Pennsylvania State University) titled "Fuel Use in United States Electric Utilities, Developments 1947-1993" by Aabakken (1995). The following discussion will pertain only to the use of coal in the electric power generation industry.

The economics of coal have been more stable compared to the competing fuels. Coal prices have not endured the kind of regulatory pressure that oil and gas have and the prices of coals have been less volatile. However, coal has had regulatory impacts that are present today and will last for decades.

One form of regulation on the producer side of coal has been for mine reclamation. Federal surface coal mine regulation legislation required that strip coal mines reclaim the land to return it back to its past level of usefulness as an integral part of the mining process. Extensive remediation measures also need to be installed to prevent acid mine drainage. Despite the added costs of all these measures, the coal industry's productivity gains seem to have negated any price increases. The price of coal has consistently been lower than oil or gas. As such, this legislation does not seem to have placed coal in a disadvantaged situation with its competitors.

The Clean Air Act of 1963 has had a significant influence as a disincentive for coal use. In 1978, Congress passed the Best Available Control Technology (BACT) requiring new coal burning plant to employ devices to reduce SO₂ emissions. Ultimately

this regulation resulted in a move away from coal as the favored fuel for expansion of capacity. In 1993, no coal plants were added to the electric utility industry. Natural gas became the favored fuel for capacity expansion.

4.7.1.3 Slower Generation Growth

Electric generation varied with economic activity in the period studied (1947-1993). From 1950 to 1960, generation grew at an average annual rate of 8.6%. Strong growth continued into the next decade and generation grew by 7.3% from 1960 to 1970. However, the 1970s signaled the first major trend in a slowdown in electricity generation. From 1970 to 1980, generation grew by only 4.1% annually. This is attributed to the 1973 and 1978 oil crises in the U.S. From 1980 to 1990, electric generation growth was only 2.1% per year. In 1992, generation growth was minus 0.98%.

The slow down in growth is attributed to oil price shocks leading to reduced energy consumption. The Ford, Carter and first Bush administrations also enacted conservation policies. As the economy in the U.S. and other industrial nations matured, economic growth slowed down. In addition, a move away from a manufacturing based economy to a service based economy reduced the derived demand for energy.

4.7.2 Fuel Use in the Electric Utility Industry 1947-1993

A complete review of the history of fuel use, i.e., natural gas, nuclear power, hydroelectric power and oil, in electric utilities is presented by Aabakken (1995). A summary of coal usage is presented here.

Coal has played the largest role in electric utility generation. Due to its abundance and low cost, coal has remained competitive in electricity generation since the birth of the industrial revolution. From 1947, coal increased in generation and maintained a fairly constant share of the total generation in all but a few recession years. In 1947, coal generated a total of 255,739 MWh, which was 53.6% of the total electricity generation capacity at that time. By 1950, coal's share had dropped to 46.9% due to gains in oil and natural gas generation. By the end of the decade, coal generation had more than doubled to 710,000 MWh and provided 53.3% of total electricity production. During the 1960's electric generation by coal again doubled but by 1969 coal generation

accounted for only 49% of the total electrical generation. The loss of share reflected mostly gains made by natural gas generation.

During the 1970s, the share of coal generation stayed below 50%, but by 1980 coal again made up half of the electricity generated in the U.S. Coal generation accounted for 2.2 million MWh.

The period from 1980 to 1989 shows a slow, steady growth for coal, and by 1989 the coal market share had increased to 54.6% of the total. This share persisted until 1993, when coal made 56.9% of the total electricity.

The fortunes of coal began to turn when concerns were raised about the emissions of sulfur and nitrous oxides from the burning of coal.

Future threats to coal are taxes that aim to limit carbon dioxide emissions. Any such tax would hurt coal disproportionately, as coal produces relatively more CO₂ per KWh of generation than oil or natural gas. A tax on carbon emissions would, therefore, increase the cost of coal generation more than the cost of gas or oil generation. These possible future liabilities have been among the disincentives to further investments in coal-fired generation plants.

4.7.2.1 Regional Electricity Fuel Use

The choice among fuels for electric generation depended to a large extent on the resources available with the regions. States such as Oregon, Washington, and Idaho that have ample hydroelectric resources, chose to utilize hydroelectric power for the majority of their electricity needs. Oil generation was relatively low until 1966 and then increased rapidly. However, oil generation was concentrated among costal states that had sea access for imported oil. After the 1970s oil crises, oil lost favor, and generation fell rapidly in both absolute and relative terms.

Natural gas was concentrated in the largest gas-producing states in the West South Central region. Shortages due to regulatory constraints made gas less attractive after the 1970s, especially for states outside the producing region.

Nuclear generation grew rapidly in the 1970s and 1980s. Most of the nuclear generation capacity was installed in the Eastern and Midwestern Regions, where demand for electricity was greatest.

Coal use was largely concentrated in the East in 1947, and as Western coal producers emerged, coal found new markets and quickly gained generation share where access to inexpensive coal was possible. This process is particularly visible in the Mountain and West South Central regions where coal production initially was limited. Overall, coal comprised roughly half of electricity generation throughout the period covered.

The New England Region rapidly developed during the Industrial Revolution and by 1947 was dependent upon coal for half of its electricity. This remained until 1966 when oil drastically took market share from coal and nuclear power began to emerge. By 1978, coal's share was down to 2.5% of electrical generation. In the 1980s, coal regained part of the share from oil. Nuclear power continued to grow and gas began to emerge as a power utility fuel. In 1993, nuclear made up 52.8% of the total electrical generation, oil 20.7%, and coal 17.9%.

The heavily industrialized Middle Atlantic Region (Pennsylvania, New Jersey and New York) relied heavily on coal for the majority of its electricity production in 1947 (77.5%). The remaining electrical generation came from oil and hydroelectric, 4.2 and 18.2%, respectively. In 1966, nuclear and oil began to replace some of coal's share of the electrical market. Nuclear grew rapidly until 1993 and there were significant variations in the share of electrical generation by nuclear fuel within the region. Electrical generation by nuclear in Pennsylvania, New Jersey, and New York was 35.7, 72.7, and 25.3%, respectively. Oil rose to a peak of 34.6% in 1973 declining to 6.5% in 1993. Gas had declined to almost nothing by 1978 but made a return and made up 7.5% in 1993. Hydro declined to 9.1% in 1993.

The East North Central Region (Illinois, Indiana, Michigan, Ohio, and Wisconsin) generated almost all its electricity by coal in 1947. Up to 1970, coal's share remained constant while gas made up some of hydro's share and nuclear began to become important in the area. Oil began to capture some of coal's market share in the late 1960s. By 1980, nuclear power made up 14.1% of total generation. By 1993, coal and nuclear were responsible for 73.3 and 25%, respectively, of the total electrical generation capacity. There is wide variation in the nuclear generating capacity in the region.

The West North Central Region (Iowa, Kansas, Minnesota, Missouri, Nebraska, North Dakota, and South Dakota) had a well-diversified mix of fuels for electricity generation in 1947. Coal made up 46.4%, oil 11.1%, gas 24.8% and hydro 17.6% of total electric utility generation. In 1960, coal began to capture market share at the expense of gas as western coal producers increased access to coal. Nuclear just emerged on the market and grew rapidly to 10% in 1974. During the 1980s, coal and nuclear continued to gain market share while the other fuels declined. In 1993, coal generated 75.9%, nuclear 17.5% and hydro 4.4% of the total market.

In 1947, coal made up more than half of the market in the South Atlantic Region (Delaware, District of Columbia, Florida, Georgia, Maryland, North and South Carolina, Virginia, and West Virginia). Hydroelectric made up one third of the market at that time. In 1967, oil began a growth period along with nuclear power. Nuclear power increased until 1993 when it was responsible for 27.5% market share. Oil peaked with a share of 26.3% in 1974 and then declined until it was at 7.5% in 1993. The region had extreme variations in fuel use. For example, Washington, D.C. had 99% coal generation from 1947 to 1966 then a rapid fall to zero percent in 1976 through 1992. Oil replaced coal and had 100% of the market share by 1976 through 1993. By contrast, coal makes up 95% of the market share in West Virginia. Nuclear power generation in Virginia peaked in 1985 at 53% and was at 43.5% in 1993. Nuclear power accounts for 61.1% of the market in South Carolina and 44.2% in Florida.

The East South Central Region (Alabama, Kentucky, Mississippi, and Tennessee) depended on hydroelectric power for over 80% of its electricity in 1947. Gas and coal made up 6.6 and 26.9% of the market share, respectively. By 1960, coal produced 74.5% of the electricity while hydroelectric has decreased to 20%. These share remained constant until the 1970s when oil began to take up a portion of the market share. However, the market share by oil vanished by the 1980s. Nuclear emerged in the 1990s and made up 10.6% of the market share by 1993. The delay of nuclear power in the region was due to the ongoing problems by the Tennessee Valley Authority to get their reactors online.

The West South Central Region (Arkansas, Louisiana, Oklahoma, and Texas) was dependent upon gas for the majority (86.8%) of its electrical generation in 1947. This is

due to the vast gas resources in the region. It was not until 1970 that coal, oil and nuclear use began to increase. Production of lignite in Texas made inexpensive coal available and Arkansas built four large coal power plants in the early 1980s. By 1993, coal, gas and nuclear were responsible for 49.4, 37.1, and 10.2% of the market share, respectively.

The Mountain Region (Arizona, Colorado, Idaho, Montana, Nevada, New Mexico, Utah, and Wyoming) relied heavily upon hydroelectric power (79.9%). Coal and oil combined accounted for 10% and gas made up the other 10% of the market. Coal use began to increase in 1960. By 1975, coal was responsible for half the electricity in the region. By 1993, coal made up 75.9% of the market and nuclear accounted for 8.6% of the electrical generation. Arizona and Colorado both had nuclear power. Nuclear power peaked in 1981 at 2.8%. The plant was shut down in 1990. Nuclear power generation peaked in Arizona in 1988 at 37.3% share. In 1993, nuclear accounted for 32.4% of the market in Arizona.

The Pacific Region (California, Oregon, and Washington) generated 75.9% of its electricity with hydro in 1947. Oil generated 16.7% while gas made up 6.6% of the market share. From 1970 to 1978, oil grew rapidly at the expense of hydro and gas but shrank rapidly to less than 2% by 1984. By 1992, nuclear power generated 18.6% of the electricity. Coal emerged into the market and accounted for 4.9% in 1993. Hydro still holds a share of 56.3% in the region. Hydro had a share of 99% through 1964 in Washington and Oregon. In Washington and Oregon, coal generation accounts for 10.5% in 1993. In 1993, nuclear power accounted for 8.5% in Washington and 0% in Oregon.

By contrast, nuclear power accounts for 25.1% of the market share in California. Coal use was nonexistent and oil had roughly 20% of the market until 1961. Oil use peaked in 1977 at 55.7%. In 1993, oil only accounted for 0.8% of the market share. Natural gas made up the market lost by oil and made up 37% of the market share in 1993. Hydroelectric power accounted for 30.3% of the market in California in 1993.

A further detailed discussion of the history of regional coal use in the U.S. is presented in Aabakken (1995).

4.7.3 Coal Consumption In Electric Utilities 1947 to 1993

The focus of this discussion is the use of coal in electric utilities since coal has been the single largest fuel source for electric generation since 1947. Coal generating power plants have provided roughly half the electricity produced by electric utilities in the U.S. Coal is the least cost fuel for electricity generation, however, new environmental regulations threaten to increase capital and operating costs.

Coal generation was concentrated in the Eastern and Midwestern states where coal resources were abundant. In the 1970s, Western coal producers emerged and coal increased in its generation share in those states. Within the U.S., the market share of coal for electric generation varies significantly as a function of regional natural resources and state environment regulations.

In 1947, coal generated 136,985 MWh of energy or 53.6% of total generation of utility electricity. A decade later coal generation was 631,507 MWh or an increase of 360% or a market share of 54.9%. In 1977, the coal share of generation had fallen to 46.4% but its generation in absolute terms had grown to 2,124,323 MWh (roughly double the generation of the previous decade). In 1987, coal generated 2,572,127 MWh, and increase of only 20.1% from the previous decade. This slowdown in increase coal consumption coincides with the 1973 and 1978 oil crisis and the overall reduction in the growth of electrical consumption. By 1990, coal's market share was 54.9% increasing slightly to 56.9% in 1993.

Coal's market share in electric generation has remained fairly constant (approximately 50%) since 1947; however, there have been significant regional variations in coal use during that time. The most marked development is the growth of western consumption of coal. In 1947, coal only accounted for 10% (8,297 MWh) of the electricity consumed in the Western U.S. In contrast, coal accounted for 73% (136,985 MWh) of the electricity consumed in the Eastern U.S. A decade later western coal generation had nearly doubled to 14,265 MWh, while in relative terms coal generation had fallen to 7% of total generation. In the East, coal generation had nearly tripled to 332,120 MWh with the share of coal generation increasing to 78%.

In the 1970s, coal experienced a rapid growth in the West. Generation grew to 216,287 MWh corresponding to a market share of 26%. The East also saw a rapid

growth in total generation 768,930 MWh but this corresponds to a drop in market share to 58% (from 78% in the 1950's). In the 1980s, coal use in electrical generation increased in both absolute and relative terms in both the West and East. By 1993, the market share by coal in electricity generation in the East and West had increased to 61.2 (1,019,751 MWh) and 50.2% (556,145 MWh), respectively. A detailed discussion of the regional coal use for electricity generation is presented by Aabakken (1995).

4.7.4 Trends in Coal Consuming Electric Utilities

Tables 4.7.1 and 4.7.2 list the 25 largest coal consuming utilities in 1969 and 1993. The physical receipts of coal to these utilities are used as a proxy for actual consumption in 1993. Aabakken (1995) gives additional data covering the years 1979 and 1989. The general trend is that the role of the "large consumers" has reduced with time. The percent of utility coal receipts accounted for by "large consumers" has decreased from three-quarters to one-half. This reflects the spread of coal use into regions where coal was previously not used for electricity generation.

4.7.5 Historical Patterns in Electric Utility Power Plant Additions, Developments 1947-1993

Capacity additions were growing until 1974 at a rate of 7% per year, after which, a general downward trend was observed. This is attributed to the 1974 oil crisis, which resulted in increased energy prices and a decrease in consumer demand. The basic pattern of capacity additions between 1947 and 1993 follow an inverted parabola. There were low levels of capacity additions in 1947 followed by a gradual increase. A peak was reached in 1974 in absolute terms.

In 1992, only 3,886 MW of new units were added. This is less than the 4,349 MW added in 1949 and is only a fraction of the 40,727 MW added at the peak in 1974. Only six years after the peak, capacity additions were roughly halved. The 1989 additions marked the first time since 1957 that additions were less than 10,000 MW. Coal, oil and gas are discussed in the following sections. More detailed discussion regarding these fuels, hydroelectric and nuclear power are given in Aabakken (1995).

Table 4.7.1 1969 Top 25 Coal Consumers 25 Largest Coal Consuming Utilities, 1969

Company	Thousand Tons	Percent of Total	Cumulative Percent of Total
Tennessee Valley Authority	30,890	9.95%	9.95%
American Electric Power Co.*	30,802	9.93%	19.88%
Commonwealth Edison Co.	18,905	6.09%	25.97%
The Southern Co.	17,409	5.61%	31.58%
Ohio Edison Co.	12,914	4.16%	35.74%
Duke Power Co.	11,845	3.82%	39.56%
Detroit Edison Co.	11,745	3.78%	43.35%
Centerior*	11,429	3.68%	47.03%
General Public Utilities	9,901	3.19%	50.22%
Allegheny Power Systems, Inc.	8,979	2.89%	53.11%
Consumers Power Co.	6,557	2.11%	55.23%
Virginia Electric & Power Co.	6,449	2.08%	57.31%
Carolina Power & Light Co.	5,880	1.89%	59.20%
Wisconsin Electric Power Co.	5,038	1.62%	60.82%
Public Service Co. of Indiana	5,008	1.61%	62.44%
Niagara Mohawk Power Corp.	4,712	1.52%	63.96%
Potomac Electric Power Co.	4,425	1.43%	65.38%
Duquesne Light Co.	4,230	1.36%	66.75%
Consolidated Edison Co.	3,754	1.21%	67.95%
Cincinnati Gas & Electric Co.	3,575	1.15%	69.11%
Electric Energy Inc.	3,511	1.13%	70.24%
Baltimore Gas & Electric Co.	3,451	1.11%	71.35%
Northern States Power Co.	3,402	1.10%	72.45%
Arizona Public Service Co.	3,175	1.02%	73.47%
Philadelphia Electric Co.	3,120	1.01%	74.48%
Total Above	231,106	74.48%	
Total Utility Receipts	310,312		

*As defined by its 1993 components

Sources: Details from National Coal Association, *Steam Electric Plant Factors*, 1970.

Total from Edison Electric Institute, *Statistical Yearbook*, 1970, as compiled in Richard L. Gordon, *U.S. Coal and the Electric Power Industry*, 1975.

Table 4.7.2 1993 Top 25 Coal Consumers 25 Largest Coal Consuming Utilities, 1993

Company	Thousand Tons	Percent of Total	Cumulative Percent of Total
Southern Company	46,970	6.11%	6.11%
AEP owned or managed	46,770	6.08%	12.19%
Tennessee Valley Authority	40,387	5.25%	17.44%
Texas Utilities	31,462	4.09%	21.53%
Pacificorp	29,687	3.86%	25.39%
Detroit Edison	18,261	2.37%	27.76%
Houston Lighting and Power	18,056	2.35%	30.11%
Central and Southwest	17,995	2.34%	32.45%
General Public Utilities	15,659	2.04%	34.49%
Basin Electric Cooperative	14,963	1.95%	36.43%
Allegheny Power System	14,048	1.83%	38.26%
Ohio Edison Total	13,506	1.76%	40.01%
Arizona Public Service	12,243	1.59%	41.61%
Northern States Power	12,012	1.56%	43.17%
Duke Power	11,953	1.55%	44.72%
Public Service Co. Indiana	11,461	1.49%	46.21%
Middle South Utilities	11,164	1.45%	47.66%
Commonwealth Edison, Total	10,596	1.38%	49.04%
Virginia Electric and Power	10,277	1.34%	50.38%
Salt River Project	10,176	1.32%	51.70%
Carolina Power and Light	9,840	1.28%	52.98%
Kansas City Power and Light	9,622	1.25%	54.23%
Cincinnati Gas and Electric	8,865	1.15%	55.38%
Western Energy	8,863	1.15%	56.53%
Public Service Co. Colorado	8,788	1.14%	57.68%
Total Above	443,624	57.68%	
Total Utility Receipts	769,152		

Source: U.S. Government, Energy Information Administration: *Cost and Quality of Fuels for Electric Utilities* 1994.

4.7.5.1 Coal Capacity Additions

The slowdown in coal capacity additions is attributed to the slowdown in total electrical generation growth and the impact of environmental regulations specifically aimed at coal-fired power plants. The slowdown in electrical generation growth reflected increasing fuel prices and the subsequent increase in conservation measures. Also important is the fact that there was excess capacity in some areas where growth was slow. The economics of scale is also important. Coal has traditionally been dependent on a relatively large scale of operation to be competitive. Even as the coal itself is relatively inexpensive compared with other fuels, it is voluminous and to get economies of scale a relatively large plant is necessary. The average size of a coal-burning unit seemed to increase up to 1971. This indicates that the optimum scale of a coal plant has changed over time. The economies of scale for coal units seem to be exhausted at roughly 600 MW (Joskow 1987). Interestingly, the average size in 1988, 1990, and 1992 were 300 MW or less.

The heating value of coal has been declining in general since 1960 (Joskow 1987). At the same time the average size of coal plant was increasing. This reflects the increased use of Western coals that have lower heating value than Eastern coals. Longer construction time results in higher capital costs from interest expenses and also exposes that electricity generator to greater risks associated with demand growth, fuel prices and regulatory changes (Dowlatabadi 1991).

Coal capacity additions reached its peak in absolute terms in 1973 when 16,440 MW of coal units were added. Coal reached its peak in market share of 81.2% in 1982. No coal plants were added in 1993 and none were planned through 1995.

4.7.5.2 Oil Capacity Additions

In 1947, oil was a large part of new unit additions (46%). However, by 1954, oil only accounted for 6.6% of new units. Oil became attractive as a fuel after 1966 when the import quotas on residual fuel oil were removed. This led to an increase in the use of oil in the electric power industry. In 1967, 11% of capacity additions were from oil. In 1975, a peak of 20.9% of new capacity was from oil. After 1978 OPEC oil crisis, oil lost

favor as a fuel. In 1980, only 6% of the added capacity was oil generated. The low point was in 1984 when oil additions were only 0.2% of the total.

4.7.5.3 Natural Gas Capacity Additions

Natural gas made up 29% of added units in 1947. Additions for gas were generally above 20% until 1973. A slowdown in gas additions in the 1970s was attributed to shortages and the 1978 Fuel Use Act that made it illegal to expand further into gas generation. By 1980, there was little gas capacity being added. Deregulation in the 1980s made gas more attractive. Natural gas combined cycle gas turbines have led to a movement of smaller scale units. The advantage of smaller scale units is their increased thermal efficiency. Two or more units operating a maximum capacity offer better thermal efficiency than a partially loaded larger unit. By 1987, gas accounted for only 0.46% of additions, however, by 1993, 49.4% of additions were from gas.

4.7.6 Market Structure and Reorganization of the U.S. Electric Utility Industry

The electric utility industry is unique in its economic considerations due to its organizational structure, related regulation problems, and the extreme capital intensity. Technically this results from technical diversity and complexity, the promise of emerging new technologies and environmental problems caused by traditional technologies. Politically, this is due to a long complex history of government involvement.

The electric utility industry has undergone significant changes in recent years: regulatory constraints have been reduced, new entities have been created and new technologies have been developed. From an economic standpoint these changes are heading toward market competition and efficiency improvement. However, there is no consensus of opinion regarding the manner in which the industry needs to restructure.

Restructuring the electric utility industry is an economic problem that cannot be separated from political and technological constraints. In fact, economic efficiency and equity-oriented political concerns seem to be in conflict. Many differences between the existing proposals lie in the efficiency-versus-equity dilemma.

The intent of this study is to consider economic, political and technological factors as compatible when evaluating a proposed model with the goal to improve economic efficiency. To achieve this objective the following topics were researched: the status quo of the electric utility industry; analyze the merits and deficiencies of existing restructuring proposals; and factors to be considered for any restructuring approach. This all leads up to a new proposal for restructuring of the electric utility industry. The new proposal is based on the premise that the transmission sector can become competitive. By eliminating the practice of exclusive franchising, transmission competition can occur automatically. With this breakthrough, changes such as divestiture and reorganization can occur as guided by market forces. As a result, both the generation and transmission sector will become competitive and regulation will become unnecessary. Regulation in the distribution sector may not be completely excluded.

The following discussion is a summary of a unpublished doctorate thesis by Li (1995) titled “Market Structure and Reorganization of the U.S. Electric Utility Industry: A New Proposal.”

4.7.6.1 Status Quo of the Electric Utility Industry

This section describes the U.S. electric utility industry in the 1990s. The status quo of the industry is the result of a century-long evolution which has been heavily affected by regulations. The rate of return (ROR) regulation, established at the very beginning of this industry, is one of the most important in shaping the industry. In 1935, the Public Utility Holding Company Act (PUCHA), together with other regulations, was enacted to reinforce ROR regulation by confining organizational structures for investor-owned electric utilities.

The deregulation process started in 1978 when the Public Utility Regulatory Policies Act (PURPA) was enacted. In 1992, the Energy Policies Act (EPACT), a deregulation effort, was passed to reduce regulatory constraints. Many other regulations, such as the Clean Air Acts and their amendments, affected the electric utility industry. A detailed review of the organizational structure of the electric utility industry, the regulations that shaped the industrial structure, PUCHA and ROR are presented in Li (1995).

Organizational Structure

A summary of utilities by ownership is given in Table 4.7.3. Sales to ultimate consumers are by investor-owned utilities (IOUs) account for more than 75% of total sales in terms of either dollars or megawatt hours (MWh). In terms of sales for resale, investor owned utilities count for only over 40% (23% of their sales to ultimate customers in MWh). The role of Federal power in the retail market is negligible. The sales for resale of Federal, cooperative, and other publicly owned utilities are similar in either terms of dollars, MWh or percentage. In terms of either sales to ultimate consumers or sales for resale, cooperatives and non-Federal publicly owned utilities have similar importance.

The organizational structure of investor-owned utilities is defined by PUCHA. Each utility has an integrated franchised service area and no utilities have overlapped service territories. Utilities are responsible for providing services to their franchised territories. In return they get regulated prices for their services. Since there is no common market there is little competition between utilities.

Publicly-owned utilities are non-profit entities organized as government agencies. More than 70% of these municipal utilities rely on others, mainly Federally-owned and partially investor-owned utilities, to supply bulk electricity for resale (U.S. Department of Energy, 1993d). Municipal utilities usually have lower rates than investor owned utilities due to favorable regulations, access to lower capital cost and exemption from paying taxes or dividends. They also have preferential access to electric energy produced at Federal facilities. Economists from an efficiency perspective have criticized publicly-owned utilities since none of these advantages are based upon market competition. The largest publicly-owned utilities are similar in size to the typical investor-owned utility. Sales to ultimate consumers by a typical publicly-owned utility are only 196,037 MWh or approximately only 2% that of the typical investor-owned utility.

Cooperatives were established in 1935 via the Rural Electrification Administration (REA). Legislation provided low interest loans for the construction of rural transmission and distribution networks. The cooperatives were formed to take advantage of this low cost financing by building facilities necessary to move purchased

Table 4.7.3 Selected Electric Utility Data by Ownership, 1991¹

	Investor-Owned	Publicly Owned	Federal	Cooperative	Total
Number of Electric Utilities	265	2,007	10	949	3,231
Electric Utilities (Percent)	8.2	62.1	0.003	29.4	100.0
Revenue from Sales to Ultimate Consumers (Thousand Dollars)	147,587,609	23,151,761	1,479,079	14,147,154	186,360,603
Revenue from Sales to Ultimate Consumers (Percent)	79.2	12.4	.8	7.6	100.0
Sales of Electricity to Ultimate Consumers (Thousand Megawatthours)t)	2,110,528	393,448	52,943	205,082	2,762,003
Sales of Electricity to Ultimate Consumers (Percent)	76.4	14.3	1.9	7.4	100.0
Revenue from Sales to Resale (Thousand Dollars)	17,459,163	7,308,471	6,341,529	9,784,621	40,893,784
Revenue from Sales to Resale (Percent)	42.7	17.9	15.5	23.9	100.0
Sales of Electricity Available for Resale (Thousand Megawatthours)	487,094	193,273	200,666	235,620	1,116,654
Sales of Electricity Available for Resale (Percent)	43.6	17.3	18.0	21.1	100.0

Source: U.S. Department of Energy, 1992. *Financial Statistics of Major Investor-Owned Electric Utilities*, 1991.

¹ This table overstates the number of independent private companies by including every entity that regulators treat as a separate unit. Several forms of subsidiary relationships prevail. The first is the holding company form discussed more fully below in which a single nonoperating corporation owns two or more operating subsidiaries. Alternatively, one operating company may have one or more operating subsidiaries. Independent companies may jointly own other companies. Several nuclear plants in New England and three coal-fired plants in the Middle West were built as separately incorporated joint ventures. Once all these relationships are considered, the number of large independent investor-owned companies is under 100.

electricity to rural customers. Cooperatives are owned by their customers and are non-profit and pay no taxes. The average sale of a cooperative is 222,531 MWh and the average number of customers is 12,889 (Table 4.7.3).

The dominant activity for Federally-owned power organizations is the sales for resale. Due to subsidies, the rate from Federally-owned powers is lower than that from investor-owned utilities. The Tennessee Valley Authority (TVA) and the Bonneville Power Administration (BPA) are the largest Federal entities. In 1990, the TVA and BPA accounted for 49 and 29%, respectively, of the total sales for resale (195,674 KWhs) by Federal electric utilities (U.S. Department of Energy 1992).

The TVA was created in 1933 primarily as a navigation and flood control project with hydro electrical power generation as a by-product. However the generation of electric power has become the main purpose of the TVA. In addition, the majoring of the power is generated via steam (fossil fuels and nuclear). The success of the TVA is disputed due to problems associated with pollution and environmental impacts and the expense and difficulties in managing nuclear power plants.

Given the demographics of electric power generation, the focus of restructuring will be on the dominant investor-owned utility sector. A detailed discussion of the following can be found in Li (1995): PUHCA and investor-owned utilities; ROR regulation; the Public Utility Regulatory Policies Act; the Energy Policy Act; mergers, power pools and larger scale cooperation; and research and development activities.

4.7.6.2 Existing Proposals for Restructuring Total Deregulation

Advocacy of regulation is based upon the belief that the electric utility industry is a natural monopoly since only one firm is needed to serve an area. The purpose of regulation, in theory, is to mimic the result of what a competitive market would produce. However, Demstet (1968) states that the asserted relationship between market concentration and competition can not be derived from existing theoretical considerations, and the fact that only one firm serves each market does not mean monopoly prices will be charged. Because competition exists to win a franchise and the principal has the option to grant the franchise to another firm at the end of each contract

period, Demsetz believes that the rivalry of the open market place disciplines more effectively than do the regulatory processes of public utility commissions.

To promote competition, Demsetz advocates public ownership of distribution facilities, and then private firms would bid for the right to operate these facilities. Given the capability of long distance transmission, many firms can serve any area and any firm can serve many areas.

Posner's (1969) conclusion is similar to Demsetz, however, Posner believes that it is plausible to assume that an unregulated monopolist will typically set prices and obtain profits that are excessive. Posner realizes the importance of technological developments. Posner argues that society should not be deeply concerned if a natural monopolist charges an excessive price. "Regulation is assumed by nearly all who work or write in the field, as by the public in general, to be fundamentally inevitable, wise, and necessary. However, ... public utility regulation is probably not a useful exertion of governmental powers; that its benefits cannot be shown to outweigh its costs; and that even in markets where efficiency dictates monopoly we might do better to allow natural economic forces to determine business conduct and performance subject only to the constraints of antitrust policy (Posner 1969, p. 549)."

Gordon (1982, 1990) states that generation probably is not a natural monopoly and should not be regulated. He argues that regulatory improvements in this sector is unlikely and stresses regulatory reform over restructuring. According to Gordon, the present structure of the electric power industry is far from optimal and the need for reorganization of the operating firms is a key reason for eliminating regulation. The key problem comes from the under-pricing of electricity.

Gordon believes in interfuel competition and trusts more the strength of interfuel competition than the effectiveness of regulation. He favors Weiss's (1975) proposal to increase competition by the breaking up of combined electric and gas companies. Gordon recommends the following changes to correct inefficiencies in the electric utility industry: increase the number of participants; allow participants' access to a larger number of customers; and allow more flexible pricing (Gordon 1990).

Full Regulation

A total deregulatory approach is based upon economic principal while the full regulatory advocate is based upon consideration of daily practice. Criticism of Demsetz's contract approach is given by Williamson (1976). Williamson states that the difficulties of specifying a long-term contract with uncertainties included lead to the contracting approach converging to regulation. Therefore, the competition of bidding to win a franchising contract is not enough to replace regulation.

Marshall (1982) feels that the nuts and bolts of industry have not received enough attention and that too much attention has been placed on broad economic theories. Marshall feels that the theories fail to recognize the forces and effects of the traditional responsibility of a utility to provide services. He is arguing that without regulation electric utility markets will not satisfy the economic definition of well-functioning markets. Market stability is also a requirement. Marshall also feels that it is impossible to separate the generation, transmission, and distribution sectors, and treat the transmission sector as a common carrier.

David and Burton (1982) feel that the electric utility industry cannot be vertically disintegrated on a purely voluntary basis. They seek solution by modifying the existing regulations instead of changing the organizational structure of the industry. Bushnell (1990) states that none of the failures of regulation, the failures of deregulation, the promotion of bypass and the profit motivation from Wall Street provide compelling reasons for restructuring. Bushnell considers the "path to restructuring and competition to be somewhat perilous (Bushnell 1990, p. 593)" due to the concerns of reliability, equity, obligations to serve and the risk of being wrong. As long as economic power is concentrated, the potential for monopoly abuse will persist (Bushnell 1990). There is also concern regarding equity. There is a risk that a two-class society will develop as market forces (geographic and demographic disparities) are not equal in rural and metropolitan areas. Bushnell suggests the use of inefficient cross subsidies to solve the equity issue. While it is well known that regulation is not the best way to promote equity and that cross subsidy makes utilities vulnerable, it may not be practical to ignore the equity issues.

Weiss's Proposal

According to Weiss (1975), the economics of vertical integration between generation-transmission companies and distribution firms seems unlikely. Competition should be favored as the benefit of competition outweighs the cost it incurs. To promote competition, Weiss proposes the separation of generation-transmission companies from distribution companies, while at the same time maintaining the integration of generation and transmission systems. The generation sector will be governed by competition, including sales to distributors and large industrial customers, as well as bulk power sales. With regards to transmission, monopoly power still exists and the necessity of regulation is justified. Transmission should become a common carrier and the function of regulation is to guarantee proper interconnections and wheeling charges. Weiss does not see much potential for competition in the transmission part but recognizes its important role.

Weiss also proposes the dissolution of combination utilities that own both gas and electric business to increase competition. This proposal has been supported by Gordon (1990).

Cohen's Model

Cohen's (1979) model is to rely on Congress to enact legislation to provide for the chartering of regional bulk-power dispatching corporations (RDCs). "Each such corporation would be directed to acquire all of the high-voltage transmission capacity within its region, to lease generating plants from producers and to dispatch electricity for resale to independent distribution companies. Utilities would continue to own generating plants and/or distribution systems, but the authority to dispatch bulk power facilities to transmit electricity at high voltage, and to sell bulk power at wholesale would be vested exclusively in the RDC (Cohen 1979, p. 1539)."

"Each RDC would be an investor-owned corporation, run by share-holders for their own benefit, subject to statutory restrictions. Diversification would be forbidden and RDCs would be required to serve all distributors in the region. Each RDC would enjoy a monopoly over transmission in its region. The sacrifice of competition at the transmission stage is a necessary concession to the physical properties of electric power,

which made transmission a natural monopoly. Thus that part of the industry would continue to require rate regulation, ... (Cohen 1979, p. 1540-1541)."

The advantages of the new structure are as follows: 1) the new structure would create a regional bulk power system large enough to capture all feasible economies of scale; 2) competition is enhanced and innovation would be spurred; 3) RDCs would have no incentive to discriminate against any producer or any type of technology; 4) competition would be increased at the distribution stage; and 5) disparity would be eliminated by charging the same rates for bulk power throughout the region.

Cohen acknowledges some problems with the new structure. The rate of return regulation could encourage RDCs to over invest in the transmission grid. The reliability of electricity supply would be affected and there could be an effect on long-term planning and development of new power plants. Given the long lead time, the huge investment, and many unforeseen factors, Cohen acknowledges that no prudent firm would bind itself to deliver capacity from a new plant at a specified price without first running the regulatory gamut.

Berry's Model

According to Berry's Model (Berry 1982), the current vertically-integrated system would be replaced by separating generating and distribution companies linked by energy brokers. The generating companies would be free of regulation, whereas, the brokers and distribution companies would still be regulated. The advantage of the Berry Model is the efficiency gain of increased coordination. Currently there are 152 separate centers controlling the use of the nation's generating capacity (NERC 1992). Berry's model is the only model that addresses the addition of new capacity among all the restructuring proposals. Under Berry's model, the financing and development of new capacity is completely determined by market force. Berry is also the only author who clearly describes alternatives a distribution can have in a competitive environment. Berry envisions a dozen or so electric regions each of which is operated by an energy broker, i.e., a large organization. Berry also proposes the creation of secondary markets and even possibly future markets to trade long-term contracts. He also envisions the need for short-term and spot markets. Weiss's, Cohen's and Berry's models all agree that the

generation sector is competitive and the distribution sector needs to be regulated. They differ in how to solve the transmissions problem.

Other models summarized by Li (1995) but not discussed here include: direct competition among integrated utilities as proposed by Primeaux (1975); Houston (1983/84); O'Connor (1985); Joskow and Schmalensee Scenarios (1983); Scenarios of Office of Technology Assessment (Blair 1990); and spot pricing by Schweppe et al. (1988) and Hogan (1993).

Summary

The proposals presented and reviewed by Li (1995) are quite diverse. However, the following considerations are critical to each of the models.

1. Economics of coordination. The potential of efficiency gains by large-scale coordination has not been fully realized. Most authors clearly realize this point and treat it as an important objective of reform. The efficiency gains can be realized by RDCs, energy brokers, regional dispatching or by mergers and power pools, but at the same time these approaches will result in market power concentration. An ideal solution should realize the efficiency gains from large-scale coordination and at the same time avoid market power concentration.

2. Reliability. Most proposals consider this factor explicitly. Maintaining reliability is costly and the optimal level changes with the distribution area. Maintaining optimal reliability levels for different regions should be one of the objectives of restructuring.

3. Planning. Restructuring the existing industry seems easier than the development of new power plants since the physical assets are already present. However, no proposal can be adequate without addressing the capacity addition problem.

4. Transmission bottleneck. Most proposals realize the decisiveness of the transmission sector to the whole industry. It is clear the organizational structure of the transmission sector will determine if market forces can work.

5. Regulation. Every proposal recognizes the problem of regulation. Most proposals advocate regulation of the distribution sector and deregulation of the generation sector. Solutions to the transmission sector are quite diversified.

Several authors (e.g., Weiss (1975) and Gordon (1982, 1990)), propose the disintegration between transmission and distribution. Houston proposes direct competition at the retail level. However, direct competition for electricity consumption at this level seems unlikely. So the practical competition only happens between transmission and distribution, and between generation and transmission. Another important efficiency concern is whether the industry can accomplish the transition in a timely and efficient way that reduces total cost (production and adjustment costs). The restructuring process itself is also very costly. Therefore any reform can only go forward slowly but provide enough flexibility so future adjustments will occur incrementally and timely in response to any possible future changes.

The overall objective of restructuring is to minimize the total cost of providing electricity services by minimizing the cost of electrical supply and consumption. The objective of the new proposal developed by Li (1995) is to create a competitive environment by solving the transmission bottleneck problem and is summarize below.

4.7.6.3 Factors to be Considered for Restructuring

Factors to be considered for restructuring include the following: status of the existing generation, transmission and distribution system; threshold of transmission and distribution competition; transmission and market structure; market differentiation and competition; reliability and competition; uncertainties; flexibility and equity issues. A detailed discussion regarding how these factors impact restructuring considerations is given in Li (1995).

4.7.6.4 A New Proposal

Technologies of transmission have made it possible for the transmission sector to become competitive. As long as this condition exists, the generation sector will automatically become competitive regardless of its economies of scale. After market differentiation, all retail customers that can be potentially involved by market competition will be included in the transmission competition part. Market competition definitely cannot happen in the remaining distribution sector. Based upon these results, this new restructuring approach focuses on the transmission sector.

Reconfiguring a Transmission Grid

To create alternatives for distribution companies (discos) and minimize the cost of transmission competition, this model proposes breaking up the geographically isolated ownership structure of the existing transmission system. Once this artificial boundary is eliminated, the whole nation becomes the potential service area for each transmission company (transco). By slightly reconfiguring the physical structure and at the same time moderately modifying the existing ownership structure, this system guarantees that each small area can be served by more than one transco. In this way, competition is introduced into the transmission sector and all generating companies (genecos) and discos will have alternative for transmission services. It is argued that the existing lines can enjoy substantial Ricardian efficiency rents due to their superior locations. Since this restructuring approach does not need massive change of the physical transmission system, the total amount of efficiency rents on the existing lines will not be significantly affected. A detailed discussion as to how competition can be introduced into the transmission sector via reconfiguration is given in Li (1995).

Configuration of the Existing Grid

Reconfiguration of the existing transmission structure depends on the density and the geographic layout of the transmission lines. The density of the transmission lines affects competition in many ways. If the density is higher then the cost of a disco's switching from one transco to another will be lower and vice versa. The lower switching cost will intensify transmission competition, and the higher cost will inhibit competition. Secondly, if the density of transmission lines is lower, fewer substations will be needed, which means that the mutual restriction mentioned in the above section will be weaker. On the other hand, with the lower density of transmission lines, the potential benefit of line extension might be higher. Thirdly, if the density is higher, it will be easier for each transco to form an interconnected independent system. Also it will be easier to have more transcos serving one area.

Given the density of the transmission lines, the actual layout of the transmission lines is another determinant of competition. In addition, the number of lines linked to one node is also important. If several lines are linked to one node, that means a disco

located there can get access to several transcos through the node and the competition at that point will be strong. The density and layout of lines varies geographically.

Theoretically, there are an unlimited number of transmission line combinations to form transcos. Therefore, it is impossible to create an exact reconfiguration plant and only some general considerations can be provided. The technology of electricity transmission has already made it possible to form a nationwide competitive transmission market. By reconfiguring the transmission system, large transcos can be formed with each transco serving a very large area, possibly even the entire nation. As a result not many transcos are needed. Due to regional differences in transmission line configuration and electricity demand, the strength of the competitive force will vary geographically. Although it is possible that, for some area, the market competition approach may not be superior to regulation, in general market competition is clearly preferred.

The Cost of Reconfiguration

Physically, the reconfiguration involves only changes in the transmission sector. Due to the involvement of asset reallocation, which includes not only the transmission sector but also the generation and distribution sectors, the cost of reconfiguration involves much more than the incremental cost generated by the physical changes in the transmission system. Since little is known about the asset reallocation, this section only estimates the cost resulting from the physical changes in the transmission sector.

From 1987 to 1991, the total structure miles of transmission line changed 719, -1,656, 3,568 and 4,167 miles per year, respectively (this represents a change of 0.2, -0.5, 1.0, and 1.2% of transmission line addition respectively). Assuming the transmission system extends 10% due to the restructuring, then it requires at least \$5.5 billion dollars of new investment. This represents only about 1% of the \$476.6 billion dollars in total assets of the entire system in 1991. This 10% extension only accounts for 3% the total revenue in 1991 of \$167 billion dollars. It should be mentioned that an increase cost of communication would occur with reconfiguration due to the larger service area.

There are also effects on the organizational structure. In general, the new structure is highly flexible. The reconfiguration only requires breaking up the geographical concentration of transmission ownership. No other structural restrictions

are imposed. The critical issue of restructuring is whether the benefit of vertical disintegration between generation and transmission outweighs the related cost. For this restructuring proposal, the answer for this issue is left to the market. If the benefit of integration is bigger then vertical integration will prevail. Otherwise a disintegrated structure will result. It is quite likely that for some of the power plants the benefit of vertical integration is greater while for others the benefit of disintegration is greater.

Another organizational alternative is a contract arrangement. The organizational structure can also be affected by population density. Further details are discussed in detail by Li (1995).

Capacity Addition

If the financial market is reasonably efficient then investment can always be financed as long as returns compensate for related risks. The question is whether the restructuring makes the risk of new power plants so high that the cost of power plant financing becomes so expensive that the price of electricity becomes too high to be acceptable. Investment can be divided into two parts: systematic risk and investment risk of each individual power plant. Systematic risk defines the overall average risk of investment. It is affected by many factors, e.g., the overall future economic development, future electricity demand, future fuel prices, environmental regulation, and technological progress. The accuracy of forecasting these variables depends on three factors: quality of input information for forecasting; capability of forecasting models; and the future time interval of forecasting. The data quality used for forecasting is not going to change with the structural change. The forecasting model will not be improved by structural change. With the structural change the regulatory processes for power plant investment will be simplified. As a result, the lead-time of plant construction will be shortened. The reduced time interval of forecasting will increase the accuracy of forecasting.

Resistance to restructuring includes concerns about who pays for what are termed 'stranded' investments. These are the investments by utility and nonutility investors in facilities that cannot compete in a deregulated environment. Critics of deregulation express concerns that customers (residential and commercial) will bear an excess share of the burden of covering these costs. The argument assumes that whatever control

mechanism is in place will allow utilities significantly to increase rates to these inflexible customers.

Capacity addition can take three forms: vertical integration; contract arrangement; and independent initiatives. For each individual power plant, the investment risk depends on final demand uncertainties. Since the demand from different discos is independent, the more discos are involved, the lower the variance of the final demand. Under the new system, a transco can cover a much larger geographic area and link many more discos together. Electricity generated by the integrated power plant can be sold through the whole transco system. If one disco has lower than forecast demand, another disco might have demand higher than forecast. As a result, the demand uncertainties may be lower under the new system than under the current system. Under the new system, no demand guarantee is given to the discos. However, the demand discos face comes from final consumer. No guarantee from the final consumers exists under the current or new system.

For contract agreements, a new genco needs only to contract with a transco directly and does not have to rely on one transco to provide transmission services. In addition, the genco does not need a contract that covers the whole life of a new power plant. Depending upon a geneco's capability and willingness to take risk, a contract can be just long enough to reduce the investment risk to the geneco's acceptable level. As a result, contract arrangements will be easier and the transaction cost of contracting will be reduced. Further discussion relating to contract arrangements between genecos and transcos and independent initiatives is given by Li (1995).

Each of these alternatives has advantages and disadvantages. The one that is most appealing can only be determined by specific circumstances. In general, it might be expected that larger plants would be inclined towards vertical integration and smaller plants would encourage independent initiatives. A larger plant requires a greater amount of investment, requires a longer lead time and involves more uncertainties; therefore, a vertical integration approach for adding capacity would minimize risk.

For the new system, the justification for new capacity addition is market demand. As with the current system, this is no guarantee for return, however, under the new system the variance of return on power plant investments is widened. In general, the total

societal risk of restructuring will not increase but the way that risk is allocated will change.

The cost of investment is also related to the size of the firm that will own a new power plant. If the amount of investment is too large then a small firm may find it too difficult to handle. Under the new structure the geographic concentration of transmission lines will not exist and the number of genecos can also be greatly reduced from the present level. Even if there are only ten generation firms they will compete in every region. In essence, each region will have ten competing genecos that may make the market force work reasonably well. The effect of PURPA and EPACT and independent power production on added capacity is discussed in detail by Li (1995).

Regulation

With the new structure, the generation and transmission sectors will be very competitive such that no franchising and rate of return regulations are necessary. With the competition among transcos, besides line extension and modification there can also be some exchanges of existing transmission lines, even after the restructuring is completed. So antitrust scrutiny might be needed. For the distribution sector, regulation may still be needed depending on the structure of each disco. If a disco is independent, then only retail rates may need to be regulated. If a subsidiary of a company that owns a transco, regulation may be needed to make sure that the transco does not get any offer from its own disco that excludes other transcos based upon noncompetitive factors. This integrated structure is similar to the current structure. The difference is under the new system each firm has more choices as to how they operate.

If a firm owns several genecos and discos together with a transco, the same regulatory approach applies as the firm makes a special arrangement with a transco to transmit electricity from its own genecos to its own discos. The same regulatory approach may be needed to guarantee that the geneco does not get any special treatment. If the firm owns genecos and discos but no special arrangement is made between them and all transactions are based upon competition, then no other regulation is needed besides rate of return regulation on the disco.

Given the potential of improving efficiency on the demand side, discos may also want to take measures to improve the efficiency of electricity consumption. The new structure does not propose universal elimination of regulation. The assumption is that whether regulation is inferior to competition depends on how strong the competitive force is and which of the two approaches is more efficient cannot be resolved by theoretical debate. The new structure is designed to provide an opportunity for these two approaches to compete.

Another related issue is whether the responsibility for regulation might be decentralized or centralized. Each municipality might decide to regulate its local discos. The advantage would be greater familiarity with local situations. The drawbacks include the possible losses of economies of scale and the danger that regulation would be even more politicized than under state commissions.

If the government wants to participate in the electric utility industry for the purpose of equity considerations, it only needs to do so through discos. Subsidized discos can go out to shop for the cheapest electricity supplier. Since no regulatory distortion exists in the generation and transmission sectors, government involvement would not result in inefficiency in these two sectors. Only the distribution sector would be distorted by government interference. The restructuring model presented here does not advocate government participation for equity purposes.

How to Get There

To bring market competition into this industry requires the break up of the geographical concentration of existing transmission line ownership. This can be achieved by completely relying on market force. As long as the exclusive franchising is eliminated, transmission competition can arise automatically among neighboring utilities along their boundaries. The primary targets at the initial stage for this kind of competition would be large industrial and commercial customers currently located along a utility border but in a neighbor utility's territory. To attract these customers, utilities would need to extend their transmission lines into each other's territory.

Given the existing integrated ownership structure, it would be very common for a firm to initially own a transco and several genecos and discos. Although the transmission

line extension would be the basic form of market competition, firms' strategies would change as winners and losers emerge and the differences of firms' strengths become obvious. Some firms would want to become more concentrated in the transmission sector and others would want to focus on the generation and distribution sectors. Buyers and sellers of transmission lines would be created and much divestiture would take place. At the same time, divestiture of generation plants and distribution sectors would also occur. Since all these changes are market driven, there would be no need to worry about windfalls and losses. Also there would be no need to worry about how to compensate present owners, since these exchanges are just normal market transactions.

The final structure of this industry would be driven by efficiency. For example, if owning both genecos and discos without owning transcos does not improve efficiency then the separation of the two parts by divestiture is also inevitable.

No constraints exist for the addition of new power plants. New power plants would have a complex pattern of geographical distribution. Over time, the geographical concentration of power plant ownership resulting from the franchising practice would disappear.

Structure reform can start from a special area where the density of transmission lines is high and the potential of market competition is strong. At the beginning, only several IOUs need to be involved. If the restructuring is successful locally it can be extended nationally.

Advantages and Disadvantages

The new proposal provides a solution to the transmission bottleneck. Once this problem is solved, market forces can work to solve other issues. In general, the system must be able to respond quickly to changing electric demand. Since a transco is an independent managerial entity, no time-consuming contract arrangements will be needed to respond to demand changes. Since fewer transcos will be needed, many genecos and discos will be linked with one transco to form a system that is larger than any of the existing utility systems. Economies of coordination will be realized. The new system would possess all the advantages of a power pool but with less transaction cost.

Competition should not happen within one transmission system since the flow of electricity within a transmission grid is determined by its physical nature. Although gencos and transcos are different entities, reliability concern and market competition force these two sectors to take reliability into account. Due to these reasons, reliability will not be weakened.

Capacity addition will also be driven by market competition and the lead-time of plant construction will be reduced due to simplified regulatory procedures. The risk of individual investment can be high since firms are solely responsible for their investments.

Technological advancement will be facilitated by that fact that gencos and transcos can be very large and able to support R&D. With technological and economic development, the market structure can adjust continuously due to the reduction in regulation.

For the new system, government needs only to provide subsidy or tax relief (driven by equity concerns due to political factors) through discos. The generation and transmission sectors are not affected thereby reducing inefficiency due to equity concerns.

With this new system, the regulatory burden will be reduced however the complexity of regulation will be increased since the organizational structure will be much more diversified than it is now. Given that gencos and transcos will operate at national or at least regional scale, realignment of jurisdiction between FERC and state utility commissions may be needed. Given that discos can be affiliated with firms in other industries, the rate of return regulation may be more difficult and less effective.

Li (1995) provides a detailed comparison of the new proposed structure with the following existing proposals: total deregulation approach, full regulation approach, Weiss's proposal, Cohen's model, Berry's model, direct competition approach, O'Connor's model, Joskow and Schmalensee scenarios, scenarios of The Office of Technology Assessment and the spot pricing approach.

4.7.6.5 Conclusion

The essential feature of the new proposal is to recognize the critical role of the transmission sector in determining the market structure of the entire industry, identify the

technological feasibility of market competition in this sector, consider both the costs and benefits of making this sector competitive, and provide schemes to make this sector competitive

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5.0 EVALUATION OF DEEPLY-CLEANED COAL AS BOILER FUELS

The objective of this activity was to evaluate deeply-cleaned coals produced in DOE's Premium Fuel Program as industrial boiler fuels. Cleaned coals from Cyprus/AMAX Research & Development Center's and CQ Inc.'s programs were received for this activity. The CQ Inc. coals, which were found not to be deeply cleaned, were tested as pulverized fuels with the emphasis on trace elements emissions. CQ Inc. provided coals from the Pittsburgh, Kittanning, and Upper Freeport seams (all from Pennsylvania) that were cleaned down to ≈ 10 wt.% ash using a combination of heavy media cyclones, water spirals, and froth flotation. Since the CQ, Inc. coals were not deeply cleaned, they are not discussed in Section 5.0 but the results are provided in the trace elements discussion in Section 3.0. Consequently, the work for this activity focused on using deeply-cleaned filter cakes from AMAX where the handling characteristics, combustion performance, and emissions were determined.

The objectives of DOE's Premium Fuel Program were to: 1) produce an alternative fuel which will produce less toxic emissions (by cleaning); 2) move advanced cleaning technologies from the laboratory to a fine-coal cleaning plant; and 3) produce a highly-loaded micronized coal-water mixtures (MCWM), which is to be competitive with fuel oil. Penn State complemented the Premium Fuel Program by performing combustion and emissions testing on the coals cleaned and provided by AMAX. AMAX provided coals from the Taggart seam (Virginia), Indiana VII (also referred to as 7) seam (Indiana), and the Hiawatha seam (Utah), cleaned using column flotation and selective oil agglomeration. A comprehensive discussion of AMAX's coal cleaning DOE-funded program is given elsewhere (Jha et al., 1997).

To accomplish the overall objective, Penn State's MCWM circuit was modified to handle filter cake to produce MCWMs, the MCWMs' atomization performance, stability, and rheological characteristics were determined, the handling characteristics of the filter cakes were evaluated through bench-scale evaluations, and pilot-scale and demonstration-scale combustion tests were conducted to evaluate their combustion performance and emissions, specifically trace elements.

5.1 Modification of the MCWM Preparation Circuit

Penn State's MCWM production circuit was modified to enable the preparation of highly-loaded MCWMs from advanced coal cleaning filter cakes. The MCWM production

facility is extremely versatile in that it is capable of producing highly-loaded MCWMs from both coarse beneficiated clean coals (e.g., 2 inch x 0) as well as deeply-cleaned fine coal filter cakes (e.g., -100 mesh). Installing the filter cake reentrainment circuit included the following: 1) circuit design, 2) ribbon mixer procurement, 3) installation of electrical, water, and air to the ribbon mixer, 4) replumbing of MCWM transfer piping, and 5) installation of a low-profile platform scale. A generalized schematic of Penn State's modified MCWM production facility is presented in Figure 5.1.1.

Figure 5.1.1 SCHEMATIC DIAGRAM OF THE HYBRID MCWM DOUBLE-STAGE GRINDING / MIXING CIRCUIT LOCATED IN THE FUEL PREPARATION FACILITY

The Munson Model HD-36-MS mixer is skid mounted with a capacity of 40 ft³. The mixer has a 40 HP motor, which can be operated at variable speed, and a flanged double helical ribbon style agitator.

The system was engineered to produce 1 ton/h of MCWM. The MCWM preparation procedure that was developed included the following steps. The supersacks were weighed on a 2,500 lb capacity, Aegis low profile 4 x 4 ft platform scale to weigh the amount of coal fines to

the nearest 0.5 lb. The supersacks were then hoisted above the ribbon mixer and allowed to discharge into the mixer's 4 x 5 ft receiving hopper. The supersacks had a pull bottom, spouted discharge chute. After the filter cake was discharged into the ribbon mixer, it was sampled for moisture content. The moisture content and the weight of the cake were entered into a spreadsheet to calculate the amount of water and dispersant needed for the formulation.

Process water was introduced using an internal spray manifold. The MCWM was mixed and pumped to a 1,000 gallon, baffled process tank, which was equipped with a five horsepower, center mounted Lightnin mixer. The Lightnin mixer had a lower 18-inch six bladed R-100 impeller and an upper 30-inch three bladed A-305 impeller. The process tank was used in addition to the Munson ribbon mixer to thoroughly reentrain the filter cake and at this point the stabilizer was added.

5.2 Fuels Characterization

Fuels characterization consisted of analyzing the filtercakes, determining their handling performance, formulating them into MCWMs, preparing MCWMs for the demonstration and pilot-scale testing, and evaluating atomization performance of the MCWMs.

5.2.1 AMAX Filter Cakes

Six filter cakes were received from Amax Research & Development Center of Golden, Colorado. AMAX shipped 206 supersacks of deeply-cleaned filter cake from the Taggart, Indiana VII, and Hiawatha seams that were cleaned by two cleaning methods – column flotation and selective oil agglomeration. AMAX shipped 50 and 31 supersacks of deeply-cleaned Taggart seam filter cake cleaned by column flotation and oil agglomeration, respectively, one supersack of Indiana VII seam column flotation filter cake and seven supersacks of Indiana VII seam oil agglomeration filter cake, and 50 and 66 supersacks of Hiawatha seam filter cake cleaned by column flotation and oil agglomeration, respectively. A limited quantity of Indiana VII was received since its moisture content was greater than ~ 50 wt.%. It was decided not to test this filter cakes as a MCWM because previous experience at Penn State had indicated that a solids loading of at least 55 wt.% is preferred.

5.2.2 MCWM Preparation – Shakedown

Prior to beginning pilot-scale production, the MCWM circuit was operated in a batch mode to gain general operating experience aimed at identifying and resolving any technical and/or logistical hurdles, which may affect the MCWM production. The selective oil agglomeration cake of the Hiawatha seam coal was used for the initial shakedown effort. This cake varied from a nominal 30 to 35 wt.% moisture content.

Supersacks were lifted by inserting forklift forks through the supersack lifting straps and then were weighed using a platform scale. After the supersack was weighed, the supersack was lifted above the receiving hopper on the ribbon mixer. Although the supersack had a centered, bottom discharge chute, the bottom of the bag had to be cut to allow the coal fines to discharge. After the supersack bottom was cut, the filter cake discharged as discrete chunks ranging from a less than one inch to several inches in diameter. The filter cake was then “chopped” to allow it to flow through the grizzly screen into the ribbon mixer chamber. After several bags of filter cake were converted into MCWM, the grizzly opening was increased to 3.5 x 3.5 inches to allow the filter cake to more easily flow into the ribbon mixer. This enlargement significantly improved the ability for the filter cake to be fed into the ribbon mixer. As the filter cake was discharged onto the grizzly screen, representative samples were taken to determine the moisture content. Moisture was determined using a CEM AVC-80 microwave moisture analyzer. Two moisture contents were determined for each supersack. The moisture content was entered into a spreadsheet along with the sample weight. The spreadsheet had prespecified solids loading and dispersant concentration targets. The spreadsheet automatically calculated the amount of water and dispersant needed to convert the filter cake into a MCWM that met the predetermined formulation. Approximately 50% of the water and dispersant was added to the ribbon mixer prior to dumping the filter cake into the hopper. The water volume was controlled using a water meter. Dispersant was weighed out and mixed with the water prior to dumping the filter cake. The remaining portion of the water and dispersant was added after the majority of the filter cake had been transferred into the hopper.

The MCWM was allowed to mix for approximately 15 to 20 minutes in the mixer. The MCWM was then pumped across a Liquidtix screen to remove the plus 20 mesh oversize particles. After screening the MCWM was thoroughly mixed in a series of process and storage tanks prior to its use.

5.2.3 MCWM Formulation

A series of filter cakes was formulated prior to full-scale production. Baseline formulations were performed on the Taggart (selective agglomeration), Indiana VII (column flotation) and Hiawatha (column flotation) seam filter cakes. Baseline formulations were geared to obtain a relative indication of the slurryability of these feedstocks. The MCWM were formulated without the use of dispersant or stabilizer to determine the feedstock's inherent slurryability. The analyses and slurryability behavior of these feedstocks are presented in Table 5.2.1 and Figures 5.2.1 through 5.2.3, respectively.

The Hiawatha seam selective agglomeration filter cake was formulated with and without viscosity and stability modifiers. The Hiawatha seam selective agglomeration filter cake represented the first premium coal filter cake supplied by AMAX that was converted to a highly-loaded MCWM using Penn State's MCWM preparation facility. The Hiawatha seam selective agglomeration filter cake was formulated using a bench-scale variable speed mixer. The Hiawatha seam supersacks were sampled and a 55 gallon drum composite sample prepared. The Hiawatha seam filter cake was systemically diluted with water to determine the effect of solids loading on apparent viscosity for MCWMs not having a viscosity modifier (dispersant). In addition to the baseline testing, the Hiawatha seam filter cake was converted into a MCWM using 1.0, 0.75, 0.5, and 0.35 wt.% A-23-M dispersant. Figure 5.2.4 provides a comparison between the Hiawatha seam MCWM prepared with and without viscosity modifiers. Figures 5.2.5 and 5.2.6 show the MCWM characteristics when prepared using varying quantities of the A-23-M dispersant. Despite the dispersant concentration, the solids loading could only be adjusted to a nominal 54 to 55 wt.% to meet a 200 cp apparent viscosity target. The effect of stabilizer concentration on the MCWM's apparent viscosity and static stability was determined by varying the stabilizer concentration from 0 ppm to 200, 400, and 800 ppm while holding the solids loading and dispersant constant at 54.0 and 0.5 wt.%, respectively. Flocon 4800C was the stabilizer used throughout the stability testing. Figure 5.2.7 illustrates the relationship between the concentration of Flocon and the apparent viscosity of the MCWM. Apparent viscosity increased significantly as the Flocon concentration was increased with minimal improvement in

Table 5.2.1 Analysis of the Taggart, Indiana VII, and Hiawatha Seam Filter Cakes

	Taggart Seam		Indiana VII		Hiawatha	
	Column Flotation	Selective Agglomeratio n	Column Flotation	Selective Agglomeratio n	Column Flotation	Selective Agglomeratio n
~ # Bags	50	31	0.5	7	50	66
Grinding Required	Yes	No	No	No	Yes	No
Proximate Analysis^a						
Volatile Matter	-	34.9	36.6	-	42.7	-
Fixed Carbon	-	63.4	60.0	-	53.5	-
Ash	-	1.7	3.4	-	3.8	-
Ultimate Analysis^b						
Carbon	-	84.0	76.8	-	77.0	-
Hydrogen	-	5.4	5.3	-	5.8	-
Nitrogen	-	1.5	1.6	-	1.3	-
Sulfur	-	0.6	0.8	-	0.8	-
Oxygen	-	8.5	12.1	-	11.3	-
HHV, Btu/lb^a	-	14,975	13,653	-	13,633	-
Particle Size Distribution						
(μm)^b						
Top Size	315	106	97.8	164	219	132
D (v, 0.9)	96.0	48.3	34.5	68.9	78.3	53.2
D (v, 0.5)	33.4	13.0	13.3	22.1	31.0	16.4
D (v, 0.1)	7.5	4.6	4.1	5.2	7.1	5.2
Slurry pH	-	5.6	6.9	-	7.8	-

^a Reported on dry basis^b The D(v, 0.9), D(v, 0.5), and D(v, 0.1) values are the particle sizes where, respectively, 90, 50, and 10 of the particles, by volume, are less than the indicated size.

Figure 5.2.1 APPARENT VISCOSITY VS SOLIDS LOADING FOR THE TAGGART SEAM SELECTIVE AGGLOMERATION FILTER CAKE

Figure 5.2.2 APPARENT VISCOSITY VS SOLIDS LOADING FOR THE HIAWATHA SEAM COLUMN FLOTATION FILTER CAKE

Figure 5.2.3 APPARENT VISCOSITY VS SOLIDS LOADING FOR THE INDIANA VII SEAM COLUMN FLOTATION FILTER CAKE

Figure 5.2.4 APPARENT VISCOSITY VS SOLIDS LOADING FOR THE HIAWATHA SEAM MCWM PREPARED WITH AND WITHOUT DISPERSANT

Figure 5.2.5 DISPERSANT CONCENTRATION VS APPARENT VISCOSITY OF THE HIAWATHA SEAM SELECTIVE AGGLOMERATION FILTER CAKE (a: 1.0 wt.% dispersant; b: 0.75 wt.% dispersant)

Figure 5.2.6 DISPERSANT CONCENTRATION VS APPARENT VISCOSITY OF THE HIAWATHA SEAM SELECTIVE AGGLOMERATION FILTER CAKE (a: 0.5 wt.% dispersant; b: 0.35 wt.% dispersant)

Figure 5.2.7 FLOCON CONCENTRATION VS APPARENT VISCOSITY OF THE HIAWATHA SEAM SELECTIVE AGGLOMERATION FILTER CAKE

the MCWM's static stability. The static stability of the Hiawatha seam MCWMs was very poor and hardpack was noted for each sample. In light of the poor stability and apparent minimal impact of dispersant concentration on apparent viscosity, the formulation targeted for the demonstration-scale CWSF production efforts were: 1) 54 to 55 wt.% solids, 2) 0.35 wt.% dispersant, 3) no stabilizer, and 4) use immediately after production.

5.2.4 MCWM Full-Scale Production

Demonstration-scale production of ton quantities of MCWM was performed using four of the AMAX filter cakes sent to Penn State. The MCWMs were prepared to characterize their slurryability, combustion performance, and trace element emissions. The as-received filter cakes were converted into highly-loaded MCWMs by using a viscosity modifier (dispersant) marketed under the tradename A-23M. Stabilizers were not used since the MCWMs were tested immediately. Particle size distributions were not modified because the objective of the testing was to determine if these filter cakes produced by advanced deep-cleaning technologies can be directly used as a MCWM after the filter cake has been entrained. Ton quantities of MCWM

were prepared from the following filter cakes: 1) Hiawatha seam column flotation cake, 2) Hiawatha seam selective agglomeration cake, 3) Taggart seam column flotation cake, and 4) Taggart seam selective agglomeration cake. The production summary is given in Table 5.2.2.

The MCWMs were prepared using the procedure discussed in Section 5.2.2 with some slight modifications. The number of moisture contents that were determined for each supersack during emptying into the hopper was increased from two up to four. This was done because some supersacks exhibited more heterogeneity than others. In addition, a pH modifier, ammonium hydroxide, was used on the Taggart seam column flotation filter cake to raise the pH above 7.

The MCWMs were allowed to mix for approximately 15 to 20 minutes in the ribbon mixer. The MCWM was then pumped across a Liquidtix screen to remove the plus 20 mesh oversize particles. One of the major components in the oversize particles is degraded supersack fragments. Small pieces (e.g., 1 to 3 mm in size) of supersack littered the MCWM. These particles were removed to ensure that no plugging of the in line strainers and/or in the atomizer tip occurred. In addition, two supersacks contained debris (e.g., rope, extraneous metal) from AMAX's coal cleaning plant. After screening, the MCWM was thoroughly mixed in a series of process and storage tanks before transfer to the day-tank located in the boilerhouse.

5.2.5 MCWM Atomization Performance

The atomization characteristics of the MCWMs produced from the AMAX filter cakes were determined in support of the combustion performance and trace element emissions testing. The atomization testing was performed using the equipment shown in Figures 5.2.8 and 5.2.9. The fuel gun was removed from the boiler and the MCWM was sprayed into a modified fly ash hopper. A Malvern 2600C Droplet and Particle Size Analyzer was used to determine the droplet PSD (Figure 5.2.8). PVC piping was installed from the top of the hopper to the inlet of the baghouse in order to create a negative pressure on the chamber, using the induced draft fan, to eliminate back misting (Figure 5.2.9). The atomization tests were conducted after the combustion tests were performed to ensure that the same solids loading and atomizing air-to-fuel (A/F) ratio were used.

Figure 5.2.10, which is typical of the atomization results, shows the atomization performance using the Hiawatha advanced column flotation (CF) MCWM. The tests were

Table 5.2.2 Production Summary of the Amax Filter Cakes

	Hiawatha		Taggart	
	Column Flotation	Selective Agglomeration	Column Flotation	Selective Agglomeration
MCWM Targets				
Solids Loading (wt.%)	55	55	60	60
Dispersant	0.35	0.35	0.30	0.30
pH Modifier Used	No	No	Yes	No
Production Summary				
Production Dates	8/12/98 – 8/19/98	4/6/98 – 6/19/98	6/24/98 – 8/10/98	6/4/98 – 8/10/98
Moisture Content	33 – 43 wt.%	22 – 41 wt.%	20 – 43 wt.%	18 – 31 wt.%
Nominal lbs/ bag	~ 1800	~ 2300	~ 1300	~ 2100
Estimated lbs of cake entrained	36,192	106,400	42,344	54,113
Particle Size Distribution (µm)				
Top Size	219	132	315	106
D (v, 0.9)	78.3	53.2	96.0	48.3
D (v, 0.5)	31.0	16.4	33.4	13.0
D (v, 0.1)	7.1	5.2	7.5	4.6

Figure 5.2.8 FRONT VIEW OF THE ATOMIZATION TEST CHAMBER

Figure 5.2.9 SIDE VIEW OF ATOMIZATION CHAMBER

Figure 5.2.10 ATOMIZATION RESULTS FOR THE MCWM PRODUCED FROM THE HIAWATHA SEAM ADVANCED COLUMN FLOTATION FILTER CAKE

conducted by varying the atomizing air pressure (i.e., the atomizing air flow) for three flow rates of MCWM. This set of curves brackets the fuel flow rates and A/F ratios of the Hiawatha-CF MCWM tests. The atomization tests for the other three MCWMs were conducted similarly.

The atomization results for the four filter cake MCWMs are summarized in Figure 5.2.11, which illustrates the relationship between A/F ratio and the volume median diameter (D_{50}). The atomization characteristics of the Hiawatha-CF, Hiawatha selective agglomeration (SA), and Taggart-CF MCWMs were similar; however, the Taggart-SA MCWM behaved quite differently. The droplet size distribution of the Taggart-SA MCWM was much coarser. This was due to the narrow coal PSD, which results in a higher apparent viscosity (see Table 5.2.3) and dilatant rheology (i.e., viscosity increases as shear increases) (Miller et al., 1997b). Note that Table 5.2.3 contains MCWMs produced during other test programs at Penn State and are included as part of the combustion performance discussion in Section 5.4.

Table 5.2.3 Coal and MCWM Analyses

Coal Seam Circuit Mode	Hiawatha-SA ^a	Hiawatha-CF ^b	Taggart-SA ^c	Taggart-CF ^d	Coalberg		Middle Kittanning		
	FCR ^e	FCR	FCR	FCR	Single-Stage	Double-Stage	Single-Stage ^f	Double-Stage ^f	Single-Stage ^g
Coal Properties									
<i>Proximate Analysis</i> (wt. %, dry basis)									
Volatile Matter	43.8	41.7	33.3	33.3	35.2	35.2	31.0	31.0	31.5
Ash	2.2	2.5	1.5	1.7	3.1	3.1	4.5	4.5	4.5
Fixed Carbon	54.0	55.8	65.2	65.0	61.8	61.8	64.5	64.5	64.0
<i>Ultimate Analysis</i> (wt. %, dry basis)									
Carbon	80.0	83.2	87.3	88.2	80.5	80.5	81.9	81.9	85.4
Hydrogen	5.8	6.1	6.4	5.7	5.2	5.2	5.1	5.1	5.5
Nitrogen	1.5	1.5	1.4	1.5	1.5	1.5	1.6	1.6	1.4
Sulfur	0.5	0.7	0.8	0.7	0.7	0.7	0.7	0.7	0.9
Oxygen	10.0	6.0	2.6	2.2	9.0	9.0	6.3	6.3	2.3
Higher Heating Value (Btu/lb, dry)	14,130	14,255	15,012	14,975	13,612	13,612	14,049	14,049	14,488
HGI	44 ^h	44 ^h	49 ^h	49 ^h	44 ⁱ	44 ⁱ	57 ^j	57 ^j	Not Det.
CWSF Properties									
Dispersant Conc.(wt.%)	0.35	0.35	0.30	0.30	0.70	0.70	0.80	0.80	0.80
Stabilizer Conc. (wt.%)	None	None	None	None	0.04	0.04	0.04	0.04	0.04
pH	7.2	7.1	5.7	6.2	8.5 to 9.5	8.5 to 9.5	8.5 to 9.5	8.5 to 9.5	9.4
Solids Loading (wt.%)	54.3	55.2	59.5	57.7	61.5	60.1	63.4	65.0	60.6
Apparent Viscosity (cp, @ 100 sec ⁻¹)	372	384	1,088	291	454	592	483	518	295
Particle Size									
Distribution (µm)									
Top Size	207	189	119	250	273	273	190	114	274
D(v,0.9)	70.7	73.2	57.7	101.0	147.1	82.7	87.6	47.8	107.1
D(v,0.5)	21.5	28.4	15.0	34.0	34.4	24.7	22.4	13.7	24.2
D(v,0.1)	5.5	6.1	4.6	6.1	5.6	5.4	4.6	2.7	5.0

^a Hiawatha Seam Selective Agglomeration; test on June 23, 1998^b Hiawatha Seam Column Flotation; test on August 17, 1998^c Taggart Seam Selective Agglomeration; test on June 16, 1998^d Taggart Seam Column Flotation; test on August 11, 1998^e Filter Cake Re-entrainment^f Testing performed during a previous project (Miller *et al.*, 1997b)^g Test performed during this project to compare with filter cakes^h From Jha *et al.*, 1997ⁱ Average of December 1995 to February 1996 Samples^j Average of August to November 1995 Samples

Figure 5.2.11 SUMMARY OF ATOMIZATION RESULTS

5.2.6 Fine Coal Handleability

Deep cleaning of coal usually involves grinding to fine sizes to ensure liberation of sulfur and ash-forming minerals. Such material, especially when wet, can present serious problems in handling and storage. This study has been concerned, primarily, with investigation of the flow characteristics of fine coal “filter-cake” materials and their influence on handling and storage behavior. Particular emphasis has been placed on the role of particle size distribution and moisture content and on the effects of blending with coarser material.

5.2.6.1 Background

The flow characteristics of powders such as fine coal are governed by friction between adjacent particles – *internal friction* – and against confining surfaces – *wall friction* – and by attractive forces between particles – *cohesion*. The latter is especially important for very fine material and is usually further enhanced by the presence of moisture. Flow of a powder occurs in response to applied stress – shear stresses generally promote flow, while normal stresses tend to increase frictional opposition to flow. In order to initiate flow in a bed of powder, the applied

shear stress must be sufficient to overcome both the frictional forces and cohesion. Typically, this condition can be expressed in terms of the Mohr-Coulomb relationship:

$$\tau = \sigma \tan \phi + c \quad (5.2.1)$$

where τ is the shear stress needed to initiate flow, σ is the normal stress acting on the powder, ϕ is the angle of internal friction and c is the cohesion. A standardized procedure for evaluating and characterizing powder flow behavior was developed by Jenike (1964; 1970) based on measurements obtained in a simple shear cell.

The Jenike shear cell consists of a short cylinder, split across the axis and mounted with its axis vertical. The lower half of the cell is fixed, while the upper half is free to move in response to an applied horizontal force. The cell is filled with powder and pre-consolidated by a standard procedure. A normal force is applied to the top of the bed of powder and the bed is sheared by applying the horizontal force. The test results are used to establish a set of flow criteria as defined by Jenike.

Powder Flow Characteristics

The Effective Yield Locus

While an initial or consolidation load is applied to the pre-consolidated powder, a force is applied in a horizontal direction to promote shearing of the bed. Eventually, the shear force reaches a constant level and is a measure of the stress needed to maintain steady motion. This value, together with the applied normal stress, defines one point on the *Effective Yield Locus*, EYL, for the bulk solid. Tests over a range of consolidation loads typically result in a roughly linear plot of shear stress *versus* normal stress. The slope of the line gives the *Effective Angle of Internal Friction*, δ (slope = $\tan \delta$). An example is given in Figure 5.2.12. Mohr's Circles, drawn through the individual consolidation points, tangent to the effective yield locus and centered on the normal stress axis, define the major principal stresses, σ_{1c} and s_{3c} , needed to maintain steady flow at a particular consolidation level. The values can be obtained graphically from the plot or they can be calculated using the following procedure. First t_{c0} is determined from the plot. This is the point estimated from the graph where the EYL intercepts the shear stress axis at zero consolidation stress. Next, the value where the shear stress is zero is calculated using

Figure 5.2.12 EXAMPLE OF AN EFFECTIVE YIELD LOCUS (EYL) FOR THE TAGGART COAL SHOWING THE MAJOR AND MINOR PRINCIPAL CONSOLIDATION STRESSES, σ_{1c} AND σ_{3c}

$$\sigma_{c0} = -\frac{\tau_{c0}}{\tan \delta} \quad (5.2.2)$$

which in turn is applied to

$$\sigma = \sigma_c (1 + \tan^2 \delta) - \sigma_{c0} \tan^2 \delta \quad (5.2.3)$$

to determine the center point, s , of each Mohr's Circle on the EYL plot. Now, the radius of the Mohr's Circle can be derived from

$$r = (\sigma - \sigma_{c0}) \sin \delta \quad (5.2.4)$$

Finally, the major and minor principal stresses, s_{1c} and s_{3c} , can be determined from:

$$\sigma_{1c} = \sigma + r \quad (5.2.5)$$

and

$$\sigma_{3c} = \sigma - r \quad (5.2.6)$$

The Yield Locus

The *Yield Locus* (YL) represents the change in shear strength of the material with the extent of consolidation. After each consolidation shear, the bulk material is again sheared but with a smaller normal stress than that used for consolidation. This shear stress needed is also normally less than that used in the consolidation stage. Another difference is that the force peaks and drops off rather than remaining constant as was evident in the consolidation shear. For a set of constant consolidations with differing shear loads, a plot of the shear stress *versus* the normal stress gives the yield locus. The slope of the YL gives the *Static Angle of Internal Friction*, ϕ , and the intercept at zero normal stress is the *Cohesion*, c . An example is shown in Figure 5.2.13.

A Mohr's Circle, tangent to the YL, centered on the normal stress axis and passing through the origin, defines the stresses acting at on a plane in the powder where the minor

Figure 5.2.13 YIELD LOCUS (YL) FOR TAGGART COAL CONSOLIDATED AT 6.3 KN/M² SHOWING THE UNCONFINED YIELD STRESS, F_c

principal stress is zero and the major principal stress is just sufficient to initiate flow. Using the values of the cohesion c and the angle of internal friction ϕ estimated from the YL, the *Unconfined Yield Stress*, f_c , can be evaluated graphically as shown in Figure 5.2.13 or by:

$$f_c = \frac{2c \cos \phi}{1 - \sin \phi} \quad (5.2.7)$$

This *Unconfined Yield Stress* represents the major principal stress necessary to cause failure of a dome or arch in the material at that particular degree of consolidation.

The Flow Function

The *Flow Function*, FF of a powder describes the relationship between unconfined yield stress, f_c , and the degree of consolidation. Specifically, it is represented by a plot of the unconfined yield stress, f_c , determined from the yield loci *versus* the major consolidation stress, σ_{1c} , obtained from the effective yield loci. The flow function, illustrated in Figure 5.2.14,

Figure 5.2.14 FLOW FUNCTION (FF) FOR TAGGART COAL WITH LINE CORRESPONDING TO A FLOW FACTOR (ff) OF 1.7

provides a basis for predicting the flow behavior of the powder in storage devices, particularly the tendency for blockages (stable arches) to form. For a free-flowing powder ($c = 0$), the flow function is zero for any degree of consolidation. The more cohesive the material, the higher the unconfined yield stress for a given degree of consolidation.

Wall Friction

As a material flows, it not only comes in contact with the other bulk material, it also contacts solid surfaces such as the walls of a bin or hopper. Testing is, therefore, required for estimating the effects on flow of the relative roughness of the wall material. Jenike shear tests can be performed in much the same manner as those used to determine internal friction. In this case, however, the bottom cell (cup) is replaced by a sample of the wall material. Typically, wall friction tests lead to an expression analogous to equation 6.2.1, i.e.,

$$\tau = a + \sigma \tan \phi_w \quad (5.2.8)$$

where a is the *adhesion* and ϕ_w is the *angle of wall friction*. In many cases, the adhesion is negligibly small, but it may be significant for wet, sticky materials.

Bulk Density

The bulk density of a powder plays an important role in gravity flow. Since bulk density depends on consolidation, it is useful to include its measurement as part of the shear testing procedure. A simple way to determine the bulk density in the Jenike Shear Cell is to weigh the cell, empty and after performing the test, and measure the final depth of the powder. The bulk density is then determined from the mass of powder and the relationship between cell height and volume.

Application to Bin Design

Mass Flow/Funnel Flow

Flow through a bin generally falls between two extremes. *Mass flow* occurs primarily by sliding of the material at the walls, while *funnel flow* occurs in a channel formed within the material itself. Mass flow is usually the preferred mode since it tends to be more reliable, does

not lead to “dead zones” and is less likely to cause segregation. The type of flow occurring in a bin depends on the flow characteristics of the powder and the bin geometry.

The Flow Factor

A major concern in the design of storage devices such as bins or silos is to ensure that blockages, due to the formation of stable arches or domes in the powder, do not occur. The weight of powder in a bin leads to the stresses that promote flow. At the same time, these stresses cause consolidation of the powder, increasing its resistance to flow. If, at some location in the bin, the strength of the consolidated powder exceeds the stresses that would act in an arch, a stable arch will form, leading to an obstruction to flow. Based on an analysis of stress distributions in bins, the relationship between consolidation stress and the stress in an arch has been determined by Jenike for a variety of material types and hopper geometries.

The results are presented in terms of a *flow factor* ff , defined by:

$$ff = \frac{\sigma_{1c}}{\sigma_1} \quad (5.2.9)$$

where σ_{1c} is the major principal *consolidation* stress and σ_1 is the major principal stress that *would* act in an arch under the same, but static, conditions. Flow factors are assumed to be constant for a given material/geometry. Flow factor charts are available for various bin geometries and material characteristics (Jenike, 1970). Examples (schematic), for conical and plane-flow hoppers, are given in Figure 5.2.15. For conical hoppers, mass flow is possible only in the region where the ff contours are shown. The mass-flow region is generally more extensive for plane-flow hoppers. However, Jenike (1970), recommends that wall angles be selected so as to fall to the left of the dashed line shown on the chart.

Obstructions to flow are more likely to occur if the flow factor is large. The limiting conditions for ensuring that obstructions do not form can be determined by superimposing on the plot of the flow function FF a line through the origin with slope equal to $1/ff$ (see Figure 5.2.14). Obstructions can be expected to form if conditions exist where the flow function lies above the $1/ff$ line. If the two lines intersect, the intersection point defines a critical stress σ_{cr} that must be exceeded to prevent blockage. Since the stresses generally decrease towards the apex of a

Figure 5.2.15 ESTIMATED FLOW FACTOR CONTOURS FOR $\delta = 35^\circ$ BASED ON INTERPOLATION FROM CHARTS PUBLISHED BY JENIKE (1970) (A) CONICAL; (B) PLANE FLOW

converging hopper, the critical stress can be used to establish a minimum outlet dimension – ensuring that stresses less than σ_{cr} do not occur. The minimum outlet dimension can be obtained from:

$$B = \frac{\sigma_{cr} H(\theta)}{g \rho_b} \quad (5.2.10)$$

where g is the acceleration due to gravity, ρ_b is the bulk density of the solid and $H(\theta)$ is a factor that depends on outlet shape (circular, square or rectangular) and the wall angle θ . Values of $H(\theta)$ are presented by Jenike (1970) and typically range from 2.0 – 2.5 for circular outlets, from 1.8 – 2.2 for square outlets and from 1.0 – 1.2 for rectangular outlets with length-to-width ratios greater than 3.0. The dimension B is defined as the diameter of a circular outlet, the side of a square or the width of a rectangular opening.

Similar conditions apply to the development of obstructions such as stable pipes or ratholes in funnel flow. No-piping flow factors ff_p are available (Jenike, 1970) expressed as functions of the effective angle of internal friction δ and the static angle ϕ . These can be used to establish critical stresses and minimum outlet dimensions as before. In this case, the minimum outlet dimension is given by

$$D = \frac{\sigma_{cr} G(\phi)}{g \rho_b} \quad (5.2.11)$$

where the factor $G(\phi)$, analogous to $H(\theta)$, depends only on the angle of internal friction ϕ which determines the shape of the internal flow channel. Values of $G(\phi)$ range from about 2 – 10 and are available graphically from Jenike (1970). D is the diameter of a circular outlet or the diagonal of square or rectangular openings.

5.2.6.2 Experimental

Samples

Initial tests were conducted using filter-cake coal from the Taggart seam from Southwestern Virginia; however, this coal was dropped from the project due to a lack of

availability of coarse coal from the same seam that could be blended with the filter cake. Attempts were made to acquire more material by contacting regional mining companies, but they were unable to supply a duplicate coarse material. Practical training and initial tests were therefore conducted using Taggart filter-cake coal, but Indiana-7 coal was chosen for blending experiments due to an abundance of both coarse, labeled as *Indiana-7 Parent* and fine particles, labeled merely as *Indiana-7*. Tests were run for *Indiana-7* at: *Indiana-7 filter-cake* coal, and blends of 1, 5, 10, 20, 80, 90, and 100% *Indiana-7 Parent* (by weight). Also, in later experiments, the percent moisture was increased for the 20 and 80% *Indiana-7 Parent* blends from the moisture of approximately 17 to 25%. This was done to simulate the more realistic scenario in which the filter cake coal would be air-dried in large piles after being dredged from the refuse ponds with the use of heavy equipment.

Sample Preparation

Initial samples of the filter-cake material were contained in two large 55-gallon barrels, which were, in turn, kept at a sample storage facility. Due to the samples' excess moisture content and small particle size, they were removed from the barrels and spread on a clean concrete floor where they were occasionally mixed with a rake to ensure surface exposure. This is a common drying procedure that produces a consistent moisture content throughout the sample. After two to four days of drying, the material from the two barrels was reduced to one 55-gallon barrel, a 25-gallon barrel, and a number of 5-gallon buckets that would ease transport and use. Initially, riffing was considered as the method for the sample size reduction, but the idea was quickly dropped in favor of "long-piling", a method of random sampling, which is primarily used when particle sizes are small enough to cause loss from air-borne particles. It is a process in which the coal is spread into a pile that is approximately 1 foot wide, a few inches high, and as long as necessary to include the entire sample. The sampler shovels the material into the first barrel, skips one shovel's width of the spread material to where a second sample is taken, which is placed in the next barrel and so on until the desired number of samples is collected. At this time the entire process is repeated starting with the first barrel and continues until the entire pile is sampled.

Samples of the parent (coarse) coal were prepared by the same procedure as described above except that, in the absence of significant amounts of fines, the simpler, riffing process was used for the splitting operation.

Jenike shear testing is typically considered unreliable when particles with a top size larger than 16 US Mesh are used, due to data repeatability inconsistencies. Larger particles tend to shear erratically, thus causing large fluctuations in the resulting data obtained from the tests. Also, larger particles tend to have less natural cohesiveness and therefore play little or no role in preventing flow. In order to remove the coarse particles, approximately 100 grams of the oven-dried coal was randomly selected and wet screened at 325 US Mesh (44 mm). Size distributions of the -325 mesh material were determined by laser diffraction using a Leeds and Northrup *Microtrac X-100™* Size Analyzer. Size distributions of the +325 mesh material were determined by sieving at standard sieve sizes. After determining the size characteristics, and knowing that the top size of the material should not exceed the aforementioned 16 US Mesh, all samples were screened at this size. The oversize material was discarded while the undersize (approximately 80% of the raw material screened) was stored in one-gallon zip-top bags for later use.

Sample Analysis

Proximate and Ultimate analyses were performed on each coal type. The proximate analysis determines the moisture, ash, volatile matter, and fixed carbon as weight percent while the ultimate analysis determines weight percent the carbon, hydrogen, nitrogen, and sulfur in the coal.

Blending

One of the goals of this project was to determine what amounts, if any, of coarse coal added to filter-cake would be required to reduce bin failures or, alternatively, how much filter-cake material could be added to coarse coal without causing such failures. For the blending experiments, a coarser material from the same coal seam was introduced to the blend at different percentages of the entire sample weight. A volume of approximately 300 ml of powdered coal is needed for each Jenike test. Because of substantial variations among the bulk densities of different samples, the corresponding weights depend on the sample being tested. Empirically it

was found that about 160 g of filter cake material was needed to fill the cell, while only 130 g of the coarser, or 16 US Mesh x 0, material was needed. When compiling samples of the filter cake, repeated tests determined that varying the sampling method had no effect on the shear cell results and, therefore, it was decided that random sampling would be adequate for the fine coal sample selection. However, this method was not acceptable for the coarse material sample preparation due to natural segregation. In this case, the sample was riffled down to the weight needed for each test. To create blended combinations, the fines were weighed and placed aside, and then coarse coal was riffled to the desired weight, added to the fines, and then thoroughly mixed and sealed in a zip-top bag until required.

In cases where the moisture content was adjusted, individual amounts of distilled water were measured and added to each separate sample, blended thoroughly to ensure proper mixing, and the samples were sealed in zip-top bags for later use. Most samples were prepared the day the testing was scheduled to occur in an attempt to eliminate inconsistencies.

Shear Cell Testing

Equipment Modification

The test equipment used in prior tests at Penn State included a pneumatically driven pressure sensor attached to a circular chart recorder. Although this was a reliable system, it limited the range over which the tests could be performed. Also, the chart was difficult to interpret reliably, thereby introducing human errors in the data. This was corrected by installing an electronic pressure transducer with a range of 4-20 milliamps attached via a screw drive to an electric motor. A data acquisition card that, after being calibrated using known constants, could transmit the milliamp signal provided comprehensive data that could later be converted by computer to a force basis. Computer programming also provided the ability to record each individual test thus removing the need for interpolation of charts, which inevitably leads to guessing. Human error is attributed to many of the problems associated with Jenike shear tests and through the “modernization” of the existing equipment many of these problems were eliminated.

Calibration

The data or raw output sent from the force transducer is a signal (in milliamps) of pressure that represents the resistance interpreted by the computer data acquisition card. By hanging known weights from a custom lever, the LabView™ program can be mathematically set to read the current and convert this reading to usable data in the form of kilograms. The data must be converted to a stress σ with units of kN/m^2 , and this is done by:

$$\sigma = \frac{Wg}{A} \quad (5.2.12)$$

where W is the weight in kilograms, g is the acceleration due to gravity and A is the cross-sectional area of the cell (0.0071 m^2 for the 9.5 cm cell). Thus, the stress is given by:

$$\sigma = 1.38W \quad (5.2.13)$$

In order to simplify the calculations and setup procedure, the consolidation and test loads were standardized so that the weights used for each sample were identical. The weights include the addition of the hanging bracket (500 g) and the test lid (174.4 g), which must be accounted for as they are part of the load applied to the sample.

Test Procedure

One of the main concerns with the use of the Jenike Shear Cell system for determining the flow characteristics of powders is the difficulty of reproducing test results for identical conditions unless the operator is well trained in the operation of the equipment. Other test cells such as the Peschl Rotational Cell are available (Peschl and Colijn, 1976) and claim more ease of use and reliable results. However, the Jenike cell is widely accepted for industrial applications. The procedure recommended by Singhal (1984) was the basis for the method used on all the tests performed for this thesis. Due to the difficulty in repeating tests with consistent results, in the first few months after updating the Jenike equipment and prior to actual testing, numerous tests were performed and repeated in an attempt to produce reliable results from the procedure. This ensured that every test was performed accurately and with confidence. Each test, however, was duplicated to ensure validity.

When using material containing less than 50% of the coarse coal, about 160-180 gram samples (total weight) were individually prepared and stored in zip-top bags. When using more than 50% coarse material, about 140-g samples were required to fill the cell completely. The

coal was scooped out of the bag at about 20 grams per scoop. It was not poured from the bag, as that may have caused segregation within the cell. The cell itself (see Figure 5.2.16) consists of:

- a cup-like container with an inside diameter of 9.5 cm (3.75 in), which rests on the base;
- a ring-like cylinder of equal inside diameter, which sits on the cup but is slightly closer to the stem; and
- a fill ring, or the mold, which sits on the ring.

Figure 5.2.16 SCHEMATIC DIAGRAM OF THE JENIKE SHEAR CELL

The fill ring allows the operator to place extra material in the cell in order to compensate for consolidation and ensures an equal height of the sample. The weight chosen was sufficient to fill the cell without overflowing. After filling the cell, the material was gently prodded with a spike fifty times. This created an air passage that aids in compaction by allowing excess air to exit the sample. The material was scraped flush with the top of the mold and a consolidation lid, or twisting ring, was temporarily placed on the material within the cell. This lid differs from the lid used in the tests in that it has a smooth surface that sits on the coal in the cell, and has a pointed center pin with square edges surrounding the pin with which a wrench-like tool can be used to rotate the lid while a consolidation load, s_c , is applied from a hanging bracket. The lid was rotated $\pm 30^\circ$ approximately 30 times to ensure even consolidation within the cell. The consolidation load was removed from the pin, the consolidation lid and the fill cell were removed, and the excess coal was again scraped off flush with the top surface of the cell using a straightedge. The test lid is then placed on the cell and the hanger and the same consolidation load was reapplied. The test was labeled in the LabView™ program on the computer along with the identification of the test weight applied during the second stage of the test.

The first stage of the test is consolidation. This represents the stress required for the material to overcome its own cohesive strength and begin to flow. As is evident in Figure 5.2.17, the initial forces increased rapidly until steady flow of the material was achieved, at which time the pressure did not increase any further. After the consolidation force leveled off, implying that the material was in a state of steady flow, the drive motor was reversed to remove all pressure on the transducer, and a new, lighter weight was applied to the hanger. The lighter weights used for the test load stage of the experiments were selected at intervals as near as possible to 1/5 of the consolidation weight. This new weight represented a normal stress applied to the consolidated material. The motor was once again placed in the “forward” direction, and as the stem contacted the cell, the amount of force needed to begin the motion was again recorded. The force reached a maximum as indicated, where the line peaked and then dropped off indicating that the specimen had “failed” i.e., that the strength of the consolidated powder had been overcome. In Figure 5.2.17 the force returns to zero due mainly to the fact that the test weight had been removed and the transducer was no longer in contact with the cell. LabView™, when told, automatically recorded the resulting data as a “Data File”, which can easily be opened in a Microsoft Excel™ spreadsheet file. Because data points were recorded 8 times per second, they provided a very detailed description of the applied shear forces. The data were averaged in an attempt to smooth over the sensitivity of the pressure transducer. By taking the average of a set of five consecutive samples, the roughness of the curve was eliminated so that the resulting data better represented what actually occurred and the small fluctuations in the curve were ignored.

Measurements of wall friction were carried out using a flat piece of 16-gage steel as a standard reference material. The steel plate was placed upon a spacer to ensure that the overall height of the steel and spacer did not exceed the height of the cell used in previous experiments. In an attempt to prevent the plate from slipping on the base, an adhesive tape was applied to the bottom side creating a solid, immovable surface. The ring and retaining ring were then placed upon the steel plate and the filling and consolidation steps continued as usual. When it came to the actual test, however, the consolidation weight was added as a stack of smaller weights to the top of the lid. As each failure occurred, the next weight was removed until only the lid remained. Testing for wall friction was only performed on 1, 80 and 90% parent coal blends.

Figure 5.2.17 EXAMPLE OF A SHEAR FORCE/TIME PROFILE

5.2.6.3 Shear Cell Test Results

Sample Characteristics

Particle size distributions for the Indiana-7 filter-cake material and the parent coal (after screening at 16 US mesh) are shown in Figure 5.2.18. The filter cake material is considerably finer (70% passing 325 US mesh) than the parent coal (6% passing 325 US mesh). The results of the proximate and ultimate analyses are listed in Tables 5.2.4 and 5.2.5.

Shear Cell Test Data

Effective Yield Loci

The Effective Yield Locus EYL defines the stresses needed to maintain a continuous flow through a channel or hopper. A Mohr's circle, tangent to the EYL at some point σ , τ and centered on the σ -axis gives the major and minor principal consolidation stresses: s_{1c} and s_{3c} , respectively. The slope of the line gives the effective angle of internal friction δ . As seen in Figure 5.2.19, δ increases slightly as the percentage of parent coal increases from 0 to 100%,

Figure 5.2.18 SIZE DISTRIBUTIONS OF INDIANA 7 COAL SAMPLES

Table 5.2.4 Proximate Analyses of Coal Samples

Sample	Moisture	Ash	Volatile Matter	Fixed Carbon
Indiana 7 Parent	9.885	33.305	10.0175	56.6775
Indiana 7 Filter cake	1.4775	36.08	4.065	59.8525
Taggart	2.745	33.4275	1.58	64.9925

Table 5.2.5 Ultimate Analyses of Coal Samples

Sample	Carbon	Hydrogen	Nitrogen	Sulfur
Indiana 7 Parent	63.26	5.89	1.47	0.31
Indiana 7 Filter cake	59.31	6.82	1.31	0.45
Taggart	76.78	6.07	1.31	0.55

while the shear stress needed to maintain flow also increases. These results imply that, as the material becomes coarser, the force needed to maintain a steady flow increases due to increased internal friction.

Figure 5.2.19 EFFECTIVE YIELD LOCI FOR INDIANA 7 PARENT COAL, FILTER CAKE AND A BLEND OF THE FILTER CAKE MATERIAL WITH 20% OF THE PARENT COAL

The effects of moisture content are illustrated in Figure 5.2.20, using a blend of 20% parent coal as an example. With the increase of moisture, the consolidation shear stresses decrease slightly. This may be due to the lubrication effect of the liquid.

Yield Loci

The yield locus, YL, is a measure of the stresses required to initiate flow in a consolidated powder. Yield loci for the filter-cake material, the parent coal, and several blends were determined over a range of consolidation levels. A typical set for a blend containing 80% of the parent coal is shown in Figure 5.2.21. It can be seen that the angle of internal friction ϕ is essentially constant, at about 35° , independent of consolidation. The cohesion c increases slightly as the material becomes more consolidated. The effects of size consist on the yield loci at a constant consolidation stress σ_c of about 4 kN/m^2 are illustrated in Figure 5.2.22. As would be expected, the filter-cake material is more cohesive than the coarser parent coal. However, the blends, especially those containing 80% of the coarse (parent) fraction, are generally more cohesive than either of the single components. This may be a consequence of denser packing in the coarse/fine mixtures used and may have important consequences in practical applications.

Figure 5.2.20 EFFECTIVE YIELD LOCI FOR 20% INDIANA-7 PARENT COAL BLEND AT 5% AND 25% MOISTURE

Figure 5.2.21 TYPICAL YIELD LOCI PLOTS AT VARYING CONSOLIDATION LOADS FOR A 20% INDIANA-7 PARENT BLEND

Figure 5.2.22 THE EFFECT OF COMPOSITION ON YIELD LOCI FOR BLENDS OF INDIANA 7 COAL CONSOLIDATED UNDER A NORMAL STRESS OF 4 KN/M²

Blending of small amounts of filter cake material into a power-plant feed could potentially lead to increased materials-handling problems.

The effects of moisture content on the yield loci for the 80% blend are shown in Figure 5.2.23. It can be seen that additional moisture increases the cohesion but decreases the angle of internal friction. These effects could be attributed to enhancement of capillary bonding between particles combined with lubrication by the liquid.

Bulk Density

Examples of the effect of consolidation on the bulk density of test specimens are given in Table 5.2.6. The finer materials (1% and 20% parent) show a slight increase in bulk density with increased consolidation stress, while the coarser sample is essentially incompressible. The effects of particle size consist are shown in Table 5.2.7. As expected, the broad, bimodal size distributions of the blends, especially that containing 80% of the coarse, parent coal, lead to enhanced packing and higher bulk density.

Figure 5.2.23 EFFECT OF MOISTURE CONTENT ON THE YIELD LOCUS FOR A BLEND CONTAINING 80% PARENT COAL CONSOLIDATED AT 4 KN/M²

Table 5.2.6 Effect of Consolidation Stress on the Bulk Density of Filter Cake/Parent Coal Blends

Consolidation Stress, kN/m ²	Average Bulk Density, kg/m ³		
	1% Parent	20% Parent	80% Parent
4.1	614	625	883
7.2	623	631	878
10.4	634	651	883
13.5	626	645	896

Flow Functions

The flow function FF provides an indication of the ability of a consolidated powder to support stable blockages such as domes or ratholes. Flow functions for the Indiana 7 coal system are presented in Figure 5.2.24. The trends essentially parallel those seen in Figure 5.2.22 for the yield loci with the coarse/fine mixtures exhibiting lower flowability than either

Table 5.2.7 Effect of Particle Size Consist (blend composition) on the Bulk Density of Consolidated Specimens

Material	Average Bulk Density, kg/m ³
Parent Coal	844
80% Parent	885
20% Parent	638
Filter Cake	624

Figure 5.2.24 FLOW FUNCTIONS FOR BLENDS OF INDIANA 7 COAL

the parent coal or the filter-cake material. The flow function for the parent coal is essentially flat and close to the origin of the plot. This is to be expected for a material containing a relatively small amount of fines. The flow function for the filter cake material is also quite flat but shifted to higher values of the unconfined yield stress f_c than that for the parent coal. Because of the low compressibility of this material (see Table 5.2.6), cohesion is relatively

unaffected by consolidation stresses. In the case of the 20% and 80% blends, internal friction plays a more important role (see Figure 5.2.22) leading to somewhat steeper flow functions. Again, these results imply that blending the filter-cake material with a relatively coarse coal feed could lead to problems in handling and storage.

The effects of moisture on the flow functions for the blend containing 80% of the parent coal are illustrated in Figure 5.2.25. The results indicate that additional moisture reduces the flowability of the coal. Again, this can probably be attributed to enhanced capillary bonding between particles.

Table 5.2.6 Effect of Consolidation Stress on the Bulk Density of Filter Cake/Parent Coal Blends

Consolidation Stress, kN/m^2	Average Bulk Density, kg/m^3		
	1% Parent	20% Parent	80% Parent
4.1	614	625	883
7.2	623	631	878
10.4	634	651	883
13.5	626	645	896

Figure 5.2.25

THE EFFECT OF MOISTURE CONTENT ON THE FLOW FUNCTION FOR 80% INDIANA 7 PARENT COAL

Wall Friction

Friction between the flowing material and the walls of a hopper is an important factor in determining both the type of flow (mass or funnel) that can be expected and the likelihood of blockages occurring. Examples of wall yield loci WYL for blends of Indiana 7 filter cake and the parent coal in contact with the standard 16-gage steel reference material are given in Figure 5.2.26. The curves for mixtures of widely different composition show only minor differences, with slightly higher friction for the finer material, especially at the higher normal stresses. In both cases, the angle of wall friction is about 27° . There appears to be a small degree of adhesion at the wall as indicated by the positive intercept at zero normal stress.

FIGURE 5.2.26 WALL YIELD LOCI FOR 1% AND 80% INDIANA 7 PARENT COAL

5.2.6.4 Design Criteria

The shear-cell test data provide a basis for the design of storage systems for these fine coal mixtures. In particular, it is possible to establish limiting hopper wall angles to ensure mass flow and minimum outlet dimensions to prevent the formation of obstructions to flow.

Flow Type

Mass flow is preferred for most applications. The conditions for mass flow through a hopper are determined by the effective angle of internal friction δ , the angle of wall friction ϕ_w , the shape of the hopper (conical, planar, pyramidal, etc.) and the slope of the hopper walls (angle θ from the vertical). The effective angle of internal friction for the materials studied varied between 35 and 40°, while the angle of wall friction was about 27°. Referring to Figure 5.2.15 and the equivalent charts for $\delta = 40^\circ$ presented by Jenike (1970), mass flow for these materials would require hopper angles of less than 15° for conical hoppers or less than about 25° for plane-flow hoppers.

Selecting hopper angles of 12 and 20° for conical and plane-flow hoppers, respectively, the flow factors, from Figure 5.2.15, would be about 1.6 in both cases. The corresponding line is superimposed on the flow function plot in Figure 5.2.27. Best estimates of the critical stresses, obtained by linear regression and extrapolation of the flow functions, are presented in Table 5.2.8, together with the appropriate bulk densities and minimum outlet dimensions calculated from equations 5.2.10 and 5.2.11. The required opening sizes ranged from about 22 cm for the parent coal in a plane-flow channel to more than one meter for the filter cake in a conical hopper. The use of smoother walls (smaller ϕ_w) would allow less-steep walls to be used – increasing bin capacity for a given height – but would have little effect on outlet size requirements. It is interesting to note that, despite the apparently greater cohesion for the blends indicated in Figure 5.2.22, the filter-cake material requires the largest outlet openings. This is a consequence of the flat form of the flow function for that material.

It is clear from Table 5.2.8 that the addition of the filter-cake material to coarse coal can significantly increase the likelihood of obstructions forming in a bin. The use of such blends in existing systems, designed for coarse coal, could lead to serious materials handling problems.

The need to extrapolate the flow functions in order to estimate critical stresses is a common and often serious problem in the use of the Jenike test procedure for bin design. The situation arises because the equipment is insufficiently sensitive to measure shear stresses at very low normal-stress values. For example, to extend the data in Figure 5.2.27 close to the line corresponding to the flow factor of 1.6 would require consolidation of a sample at stresses as low as 2 kN/m², followed by shear testing at less than 1 kN/m². The latter value would correspond approximately to the weight of the hanging bracket, the cell lid and the coal itself.

Measurements of the shear force under such small normal loads tend to be highly erratic and unreliable.

Figure 5.2.27 FLOW FUNCTIONS AND FLOW FACTOR FOR INDIANA 7 BLENDS

Table 5.2.8 Critical Stresses and Minimum Outlet Dimensions for Storage of Indiana 7 Blends in Conical and Plane-Flow Hoppers

Material	Critical Stress, σ_{cr} (kN/m ²)	Bulk Density, ρ_b (kg/m ³)	H(θ)		Outlet Dimension (m)	
			Circular	Rectangular	Circular	Rectangular
Parent Coal	1.5	844	2.2	1.2	0.399	0.218
Filter Cake	2.9	624	2.2	1.2	1.043	0.569
Blend (20% Parent)	1.8	645	2.2	1.2	0.626	0.342
Blend (80% parent)	2.5	885	2.2	1.2	0.634	0.346

5.2.6.5 Bin Testing

Some limited testing was carried out using a specially designed, adjustable bin. The primary objective was to provide a practical validation of the design criteria obtained from shear-cell testing.

Bin Design and Construction

The design of the bin needed to address a few, very important conditions. First, the angles for the sloped sides needed to be fully adjustable along with the bin opening at the bottom, and the design had to match the geometries indicated by the shear tests. Secondly, the bin had to remain a batch-type device, meaning that the bin would be filled with material and then lifted with the aid of a fork-truck, at which time all the bulk solids should exit the bin. One of the problems associated this approach is the possibility that vibrations caused by the lifting of the bin may aid in the release of the material from a stopped or stored position. Another potential problem is the inability to produce constant consolidation of the bulk material. As the bulk material exits a steady state bin, it moves vertically through the bin and more material is replaced on top of it, keeping the compression load constant. In the batch system, the material at the top does not have this compression force exerted upon it, therefore, the only realistic representation of the bin discharge would be the first material to exit the bin. This method was chosen mainly due to the difficulty of re-circulating the material from the opening at the bottom to the feed side. This could be done using a beltline to transport the material from the feeder back to the hopper but, for this project, the simpler batch approach was adopted.

Napotnic Welding Co. from Armaugh, Pennsylvania, aided in the design of the intended adjustability of the bin and constructed the final design. The design maintained the geometry needed enabling the bottom part of the hopper to be fully adjustable by hinging it to the vertical part of the bin. Screws attached to the base allowed the angle to be altered while a sliding mechanism constricted the upper portion at 90°. One major concern that had to be resolved prior to construction was the ability of the angles to be adjusted without changing the bin opening width or location on a flat ground surface. This was solved by fitting slides to the top screws while securely fastening the bottom screws to the sloping side of the bin. Thus the angles and the bottom opening width could be adjusted independently.

Bin Testing Procedure

Because of its ready availability and minimal tendency to segregate, the bin tests were performed on the Indiana-7 “filter-cake” material. Bin tests were carried out under the conditions given in Table 5.2.9. For practical purposes, tests were not performed on any blends. It was felt that the amount of material needed to conduct these experiments would exceed the

amount of coal available and the ability to reliably blend the material at the laboratory scale would prove too difficult.

The general operational procedure for the tests was as follows:

1. The predetermined width of the opening was set using a standard tape measure;
2. An “angle finder” was attached to the bottom portion of the hopper and the sides were adjusted to the desired slope;
3. “Filter-cake” material was hand-shoveled into the bin;
4. The material was compacted in an attempt to simulate the pressure that would occur under normal load conditions;
5. Using a forklift, the bin was slowly raised at a constant velocity to avoid any jerking motion that may aid in flow. As the bin was raised, flow was observed and noted; and
6. Tests were repeated at a variety of angles and opening widths. The tests were performed in attempts to induce bin failure and to determine minimum requirements for flow.

Table 5.2.9 Bin Settings and Test Results

Test Number	Bin Angle θ (degrees)	Opening Width (Inches)	Type of Flow
1	30	34.00	Funnel
2	30	34.00	Funnel
3	30	34.12	Mass
4	30	34.12	Mass
5	40	32.75	Funnel
6	40	32.75	Funnel
7	35	32.75	Funnel
8	35	32.75	Funnel
9	25	32.75	Funnel
10	25	32.75	Funnel
11	25	33.38	Mass
12	25	33.38	Mass

Results

Due to constraints of time and material availability in the final stage of this project, only a limited amount of testing with the pilot-scale bin was possible. As noted above, the primary objective of the bin-test work was to validate the use of the Jenike test procedure for predicting flow type and the occurrence of bin failures for fine, deep-cleaned coal. The conditions corresponding to each test are indicated on the flow-factor chart for plane-flow hoppers with $\delta = 35^\circ$ in Figure 5.2.28. It can be seen that the observed flow type is in general agreement with the predictions. For the steeper hopper angles (tests 1 – 4 and 9 – 12), conditions correspond to points close to or slightly to the right of the dashed line representing the mass flow/funnel flow transition. The results given in Table 5.2.9 show such transitions with small changes in bin geometry, leading to a switch from mass flow to funnel flow. In the case of the less steep angles (tests 5 – 8) conditions lie far to the right of the dashed line and funnel flow is observed in all cases.

Figure 5.2.28 BIN TEST CONDITIONS INDICATED ON THE FLOW-FACTOR CHART FOR PLANE FLOW WITH $\delta = 35^\circ$
(A: TESTS 9 – 12; B: TESTS 1 – 4; C: TESTS 7 & 8; D: TESTS 5 & 6)

Because of the problems noted above regarding appropriate consolidation of the material during flow through the hopper, reliable data on the minimum outlet dimensions required to prevent blockage could not be obtained and this aspect of the study was inconclusive. Nevertheless, with further refinement, this general approach can provide a useful means of confirming the results of laboratory-scale tests.

5.2.6.6 Conclusions

The objective of this research was to evaluate the flow characteristics of fine, filter-cake coal using the Jenike shear-testing procedure and to investigate the use of a pilot-scale test bin for validation of the results. Revamping the Jenike shear tester through the use of an electronic force transducer and a digital data acquisition system was found to increase the sensitivity and range of the results. This eliminated one of the major weaknesses of the Jenike equipment as the data were numerically recorded rather than directly interpreted from data recording charts. The following conclusions can be drawn from the results of shear-cell testing of a fine, filter-cake material, the corresponding parent coal and various blends of the two:

- The effective yield locus (EYL) – a measure of the stresses required to maintain steady flow in a consolidated powder – appears to be relatively insensitive to mixture composition and moisture content. Increasing coarseness due to the addition of the coarser, parent coal causes a slight increase in the frictional resistance to flow while moisture addition reduces friction, probably by a lubrication effect;
- The effects of mixture composition, i.e., size consist, on the (static) yield locus (YL) – a measure of the stresses needed to initiate flow in a consolidated powder – reveal that blends of the filter-cake material with the parent coal tend to be more cohesive than either of the separate components. This rather surprising result is attributed to enhanced packing in the blends and indicates potential materials-handling problems if coal fines are added to the normal feed coal in an existing system. Increased moisture content increases cohesion – probably through capillary forces – but reduces internal friction, presumably by lubrication;
- Because of their bimodal size distributions, the coarse/fine blends generally pack to higher bulk density than the separate components. The filter-cake material is

slightly compressible – bulk density increases with applied stress – while the parent coal is essentially incompressible;

- Flow functions for these materials indicate that while the flowabilities of the filter cake and parent coal are relatively insensitive to consolidation, those for the blends tend to decrease as consolidation stresses are increased. It is postulated that consolidation stresses applied to the blends affect the fine component disproportionately, leading to enhanced cohesion despite very little change in overall bulk density; and
- Measurements of wall friction against a steel reference material show very little variation with mixture composition and lead consistently to angles of wall friction of about 27° .

Application of the shear-cell test data to the development of bin-design criteria lead to the following general conclusions regarding potential materials handling problems in industrial use of fine coal, such as the filter cake evaluated here:

- Mass flow can be ensured for the filter cake, parent coal or any of the blends provided sufficiently steep hoppers are used. For hoppers constructed from the standard reference steel angles less than about 15° for conical and 25° for plane-flow would be required;
- Because they involve convergence of the material in two rather than three dimensions, plane-flow hoppers can generally operate with smaller outlet dimensions than conical types;
- The filter-cake material is most prone to blockage by arching across the hopper. Blending with coarse coal can reduce this tendency. On the other hand, addition of the filter-cake material generally increases the likelihood of arch formation in storage of coarse coal; and
- Reducing wall friction, using smooth liners for example, can permit mass flow with less steep hopper walls but will not reduce outlet size requirements.

Shear-cell testing using the Jenike procedure can provide a sound basis for bin design or for fault diagnosis in existing units. However, this approach suffers from several limitations,

particularly particle size restrictions on test samples, test reproducibility, the need to extrapolate test data (i.e., flow functions) in order to obtain important design parameters (critical stresses) and the inherent problems of scale-up from laboratory to industrial sizes. In practice, these problems are generally circumvented by the conservative nature of the design procedure, with built-in safety factors, etc. Typically, the result is over-designed systems that work successfully but at a cost in efficiency and expense. The availability of a low-cost, pilot-scale test procedure involving an actual bin would provide a useful extension to the laboratory-scale tests.

A rectangular bin with independently adjustable wall angles and outlet width was designed and constructed during this project. The system was successfully tested in terms of mass flow/funnel flow predictions using the Indiana 7 filter-cake material. However, further refinements of the system and test procedures are necessary for reliable evaluation of the outlet sizes needed to prevent arching in the bin.

5.3 Pilot-Scale Combustion Tests

Pilot-scale combustion tests were performed using the AMAX filter cakes and cleaned coals from CQ, Inc. The tests were performed in Penn State's research boiler as part of the trace element emissions study; therefore, the results are not presented here but are provided in Section 3.0.

5.4 Demonstration-Scale Combustion Tests

5.4.1 Demonstration Boiler System

The combustion testing was conducted using Penn State's demonstration boiler, which is shown schematically in Figure 5.4.1. The watertube boiler is of D-type design, manufactured by Tampella Power Corporation, and rated for 15,000 lb/h saturated steam (@300 psig). It is integrated into the University's steam distribution system. The boiler was designed to fire fuel oil but the overall system has been modified to fire coal-based fuels, i.e., MCWM and dry, micronized coal.

Modifications were made to the original system to provide combustion air and MCWM preheating, fly ash removal from the flue gas using either a conventional fabric filter baghouse

Figure 5.4.1 GENERALIZED DIAGRAM OF PENN STATE'S COAL-WATER SLURRY FUEL-FIRED DEMONSTRATOR BOILER SYSTEM WITH A PULSE-JET BAGHOUSE (Note: Coal handling and coal-water slurry fuel preparation equipment are not shown)

a ceramic membrane filter chamber, coal unloading, storage, and crushing, and either MCWM preparation or dry coal micronization. Fuel preparation and handling facilities are not shown in Figure 5.4.1. The combustion chamber and convective pass have not been modified. The furnace is approximately 8 ft long by 6 ft wide, with a volume of 370 ft³. The bulk gas residence time in the furnace is less than one second. Discussions of the equipment, program history, and a comparison between results when firing micronized coal and MCWM can be found elsewhere (Miller and Scaroni, 1990; Jennings et al., 1994; Miller et al., 1995).

When firing CWSF, it is pumped from the storage tank in the MCWM preparation facility to a 2,000-gallon surge tank located in the boiler room. From the surge tank, the MCWM is pumped through a preheater (tube and shell heat exchanger) to the burner.

The burner used during this test program was ABB Combustion Engineering's RSFC burner and it is shown in Figure 5.4.2. A detailed description of the burner can be found elsewhere (Miller et al., 1997b). This burner has been used in the demonstration boiler when firing natural gas, No. 6 fuel oil, micronized coal, and MCWM.

The atomizer that was used during the AMAX filter cake testing was a modified Faber oil nozzle. The nozzle, an XHC75° (extra-high capacity with a 75° spray angle) was found to produce the best results among several tested in the demonstration boiler and has been used for much of the reported testing. Details of the Faber atomizer and results from other atomizer testing can be found elsewhere (Miller et al., 1995; 1997a,b).

Figure 5.4.2 SCHEMATIC DIAGRAM OF THE RSFC BURNER AND ATOMIZER

5.4.2 Combustion Performance

This section discusses the combustion performance of the MCWMs produced from the AMAX filter cakes and compares them to previous testing that was performed on the demonstration boiler. Table 5.4.1 contains selected combustion results from the thirteen tests

using the AMAX filter cakes along with results from six Middle Kittanning seam and Coalberg seam MCWMs.

Table 5.4.1. Selected Combustion Results^a

Coal	Solids Loading (%)	Gas Cofire (%)	Atomization Pressure (psig)	Combustion Efficiency (%)
Hiawatha – SA ^b	55.5	30	135	98.7
	56.4	26	135	98.4
	54.3	30	135	98.2
	56.0	23	135	97.7
Hiawatha – CF ^c	55.0	30	135	97.5
	54.5	26	135	97.4
Taggart – SA	59.5	30	135	97.7
	58.5	30	80	95.1
	60.3	23	135	96.5
	59.9	30	80	93.0
Taggart – CF	58.6	30	135	93.7
	57.7	30	80	90.9
	59.3	30	135	92.5
<hr/>				
Middle Kittanning – SS ^d	60.6	30	135	93.6
Coalberg – SS	61.8	38	90	79.7
Coalberg – DS ^e	57.5	27	150	79.9
Middle Kittanning – SS	60.9	30	150	82.7
Middle Kittanning – DS	60.9	29	150	85.8
Middle Kittanning – SS	60.2	10	120	85.4

^a All combustion tests were performed using the RSFC burner. The XHC-75° atomizer was used for all the tests except for the Coalberg seam and two Middle Kittanning seam tests noted with asterisks, where an ABB 1-hole Tangential 30° atomizer was used.

^b Selective Agglomeration

^c Column Flotation

^d Single-stage grinding circuit

^e Double-stage grinding circuit

Many coals have been prepared into MCWMs in Penn State's hybrid MCWM double-state grinding/mixing circuit and tested in the demonstration boiler. These MCWMs were produced for projects in which the viability of retrofitting fuel oil-designed boilers to fire MCWM was determined (Miller et al., 1997a,b). The properties of five MCWMs and their parent coals, which have been selected for combustion performance comparison with the AMAX

filter cakes, are given in Table 5.2.3. These MCWMs were produced using a Middle Kittanning seam coal from Pennsylvania and a Coalberg seam coal from Kentucky and fired in the demonstration boiler using the same burner. They have similar particle size distributions (PSDs) and hence similar slurryability characteristics to the AMAX filter cake MCWMs.

The Middle Kittanning and Coalberg seam coals were prepared into MCWMs using the double-stage grinding circuit. With this scheme, run of mine coal (≈ 2 inch x 0) is crushed and conveyed into a 4x8 ft ball mill (first stage), where it is ground with dispersant, water, and a pH modifier. The ball mill discharge is pumped across the vibratory screen to the 1,000-gallon process tank. When single-stage grinding is preferred, the stabilizer is added in the process tank and the MCWM is pumped to the blend tank and subsequently to the boilerhouse. When double-stage grinding is desired, all or a portion of the MCWM is pumped from the process tank (before stabilizer is added) to the stirred ball mill, surge tank, and blend tank. The blend tank is used to homogenize the MCWM when only a portion of the MCWM from the process tank is passed through the stirred ball mill, and the stabilizer is added at this point. Finally, the MCWM is pumped to the 2,000-gallon surge tank located in the boilerhouse.

5.4.2.1 AMAX Filter Cake MCWMs

As previously mentioned, the objectives of the AMAX filter cake MCWM testing were to evaluate the filter cakes as potential fuels for industrial boilers and to evaluate the effect of deep cleaning on trace element emissions. In order to determine the effect of combustion efficiency (i.e., char content and surface area) on trace element emissions, tests were conducted where the atomization pressure and level of natural gas cofire (as a percent of total thermal input) were varied. The trace element results are presented in Section 3.0; however, the combustion performance as a result of these changes will be discussed in this section. Previous results have indicated that the combustion efficiency tends to increase with increasing solids content and level of natural gas cofire (Miller et al., 1997a,b).

Prior to beginning the suite of tests, preliminary operation indicated that $\approx 30\%$ natural gas cofire was necessary to maintain a stable flame. This was based on previous operating experience and by visually observing the flame character, steam generation rate, and gaseous emissions when firing Hiawatha-SA MCWM during preliminary operation. This level of natural gas cofire was chosen as the baseline level for the AMAX filter cake testing. Tests were then

conducted where the level was decreased until the flame became too unstable for operation. The lowest level of natural gas cofire achievable was 23% when firing the Hiawatha-SA and Taggart-SA MCWMs and 26% when testing the Hiawatha-CF MCWM. When firing the Taggart-CF MCWM, the level of natural gas cofire could not be reduced below 30% without the flame becoming too unstable

Two atomizing air pressures were used during the testing. The University air compressors were used to deliver 80 psig atomizing air (low atomizing air pressure) while a rented air compressor delivered 135 psig atomizing air (high atomizing air pressure).

Hiawatha-SA and -CF MCWM Tests

Four tests were conducted firing Hiawatha-SA MCWM at three levels of natural gas cofire and one atomizing air pressure. The 30% natural gas cofire repeat tests resulted in a combustion efficiency difference of 0.5%, which indicates that the combustion efficiency of the 30 and 26% natural gas cofire cases are essentially the same and may be slightly higher than for the 23% natural gas cofire test. Likewise, two tests were conducted with the Hiawatha-CF MCWM and the combustion efficiency was not affected when decreasing the level of natural gas cofire from 30 to 26%. The combustion efficiencies for the six tests ranged from 97.4 to 98.7%.

Taggart-SA and -CF MCWM Tests

The Taggart fuels behaved differently than the Hiawatha fuels. The Taggart-SA MCWM test conducted at 30% natural gas cofire and high atomizing air pressure resulted in a combustion efficiency of 97.7%. However, the combustion performance decreased as the level of natural gas cofire decreased. In addition, the combustion performance decreased as the atomizing air pressure was decreased. Similarly, the combustion performance decreased as the atomizing air pressure was decreased when firing the Taggart-CF MCWM. Unlike any of the other three MCWMs produced from the AMAX filter cakes, the Taggart-CF MCWM could not be fired with less than 30% natural gas cofire. The reason for this is not clear as the atomization performance of the Taggart-CF MCWM was similar to the Hiawatha MCWMs, which performed well; however, the coal PSD of the Taggart MCWM was the coarsest of the four AMAX MCWMs. Interestingly, the Taggart-SA MCWM exhibited the worst atomization performance

but its combustion performance, at 30% natural gas cofire and 135 psig atomizing air pressure, was still 97.7%.

General Conclusions

The MCWMs prepared from the selective agglomeration filter cakes exhibited higher combustion efficiency than the MCWMs produced from the column flotation filter cakes, especially for the Taggart seam filter cakes. A relative ranking, based on combustion efficiency, is that the Hiawatha-SA MCWM performed slightly better than the Hiawatha-CF MCWM, which performed similarly to the Taggart-SA MCWM, which performed much better than the Taggart-CF MCWM. Combustion efficiency decreased with decreasing atomizing air pressure, and the combustion efficiencies at 30 and 26% natural gas cofire were approximately the same but were higher than those at 23% natural gas cofire.

5.4.2.2 Comparison of the AMAX Filter Cake MCWMs with other MWMs Tested at Penn State

A direct comparison of the AMAX filter cake MCWMs with other MCWMs tested at Penn State is difficult because many MCWMs have been produced and tested using different burners, atomizers, and burner throat configurations, and each of these variations in hardware has an effect on the combustion performance. Several MCWMs are listed in Table 5.2.3, along with the AMAX MCWMs, and will be used for comparison. The MCWMs listed in Table 5.2.3 were conducted using the RSFC burner.

One MCWM test using Middle Kittanning seam coal was conducted during the AMAX MCWM test program. The MCWM was produced using the single-stage grinding circuit and the combustion efficiency was 93.6% when fired with 30% natural gas cofire and 135 psig atomizing air pressure. This combustion efficiency is similar to the Taggart-CF MCWM, which was much poorer than the Hiawatha seam CWSFs and the Taggart-SA MCWM.

The remaining five tests listed in Table 5.4.1 were selected from a previous program (Miller et al., 1997a), in which the effect of coal PSD on combustion efficiency was determined. Unfortunately, the Faber atomizer was only used during the last test listed, while the other tests used an ABB 1-hole tangential 30° atomizer.

As can be seen from Table 5.4.1, the combustion performance was poor when using the ABB atomizer as the combustion efficiencies ranged from ≈ 80 to 86%. What can be seen from these results though is the positive influence of fine grinding on combustion efficiency. Tests firing MCWMs produced from the double-stage grinding circuit resulted in higher combustion efficiencies. When firing the Coalberg seam MCWMs, the boiler could be operated with much less natural gas cofire even though the solids content was significantly less. It is acknowledged that the atomizing air pressure was higher for the MCWM produced from the double-stage grinding circuit; however, the level of natural gas cofire and the solids loading were much lower for the MCWM produced from the double-stage grinding circuit than for the MCWM produced from the single-stage grinding circuit.

The last test listed, the Middle Kittanning seam MCWM produced via single-stage grinding, was conducted using the Faber atomizer and 10% natural gas cofire, which was the lowest level of natural gas cofire achievable during the test program using the RSFC burner (Miller et al., 1997b). The combustion efficiency was low, compared to the AMAX MCWMs, but was as high as the best combustion efficiency obtained with an ABB atomizer.

5.4.3 Concluding Remarks

Penn State evaluated the use of deeply-cleaned coals, cleaned by AMAX Research & Development Center using advanced coal cleaning techniques, as industrial boiler fuels. Specifically, the combustion performance of filter cakes prepared into MCWMs was determined in Penn State's demonstration boiler.

The filter cakes were prepared into MCWMs that exhibited satisfactory handling and short-term storage properties (the MCWMs were prepared without stabilizers or particle size manipulation) for use as industrial boiler fuels. The combustion performance of the MCWMs was good, with the exception of the Taggart-CF MCWM, and combustion efficiencies ranged from ≈ 97 to 98%. The filter cake MCWMs performed as well or better than MCWMs prepared in the single- and double-stage grinding circuit. The level of natural gas cofire that was required to achieve these combustion efficiencies was 23 to 30%.

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Appendix 3A
Summary Sheets for Phase III MCWM and DMC Testing

Summary Sheet

Phase III MCWM Testing

TEST/DESCRIPTION:

Date	4/28/98	4/29/98	5/21/98	5/28/98
Filtering device	Fabric	Fabric	Ceramic	Ceramic
Coal seam	Hiawatha	Hiawatha	Hiawatha	Hiawatha
Cleaning tech.	Oil Agglomeration	Oil Agglomeration	Oil Agglomeration	Oil Agglomeration
Natural gas support	31%	30%	30%	30%

WATER/STEAM SIDE

Steam flow rate; lb/h	12,171	12,431	11,871	11,542
Water temperature into boiler; ° F	230	232	218	219
Drum pressure; psig	227	226	228	228
Calorimeter temperature; ° F	315	318	318	318
Steam temperature; ° F	393	397	397	397
Steam quality; %	99.9	100.1	100.1	100.1
Blowdown rate; lb/h	3,329	3,344	3,332	3,155

Air, Fuel, Flue Gas Side

Total firing rate; MM Btu/h	15.6	15.8	15.0	14.9
Natural gas flow rate; lb/h; MM Btu/h	206; 4.8	194; 4.5	193; 4.	193; 4.5
CWSF flow rate; lb/h; MM Btu/h	1,478;10.5	1,560;10.5	1,322;10.5	1,289;10.5
Gas support (%)	30.8	28.8	30.1	30.2
Air temperature entering air heater; ° F	163	165	175	179
Air temperature leaving air heater; ° F	400	396	406	404
Air temperature into boiler; ° F	385	385	397	395
Furnace outlet temperature; ° F	596	591	594	596
Gas temperature leaving air heater; ° F	385	383	384	387
Bagfilter inlet temperature; ° F	389	385	381	386
Bagfilter outlet temperature; ° F	344	338	354	367
Ash content of fly ash;%	48.99	41.66	54.62	55.5
Combustion air flow; acfm	3,870	3,877	3,973	4,048
Boiler draft; in. W.C.	-0.06	-0.07	-0.06	-0.06
Boiler efficiency; %	76.71	76.74	79.57	79.39
Atomizing air pressure; psig	135	135	135	135
Atomizing air flow rate; lbs/h	756	657	756	720
A/F ratio; lb/lb	0.51	0.42	0.57	0.56
CWSF temperature; ° F	137	117	141	138
Natural gas temperature; ° F	90	90	90	90
Coal combustion efficiency; %	97.0	96.6	98.2	97.9
Overall combustion efficiency; %	97.3	97.0	98.4	98.2

EMISSIONS (Dry Basis)

O2; %	3.8	3.5	3.5	3.8
CO; ppm	15	44	N/A	N/A
CO2; %	13.0	12.8	14.0	13.9
SO2; ppm	496	610	N/A	N/A
NOx ppm	586	540	691	453

FUEL ANALYSIS DATA

Solids Content; %	51.5%	50.3%	55.5%	57.0%
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Summary Sheet

Phase III MCWM Testing

TEST/DESCRIPTION:

Date	6/1/98	6/3/98	6/9/98	6/16/98
Filtering device	Fabric	Fabric	Fabric	Fabric
Coal seam	Hiawatha	Hiawatha	Taggart	Taggart
Cleaning tech.	Oil Agglomeration	Oil Agglomeration	Oil Agglomeration	Oil Agglomeration
Natural gas support	30%	26%	30%	30%

WATER/STEAM SIDE

Steam flow rate; lb/h	11,996	13,110	11,645	11,302
Water temperature into boiler; ° F	231	232	226	231
Drum pressure; psig	232	234	231	229
Calorimeter temperature; ° F	318	319	318	318
Steam temperature; ° F	397	398	397	397
Steam quality; %	100.1	100.1	100.1	100.1
Blowdown rate; lb/h	3,339	3,357	3,330	3,320

Air, Fuel, Flue Gas Side

Total firing rate; MM Btu/h	14.8	14.6	14.7	14.8
Natural gas flow rate; lb/h; MM Btu/h	192; 4.5	168; 3.9	192; 4.5	194; 4.5
CWSF flow rate; lb/h; MM Btu/h	1,332; 10.5	1,428; 11.0	1,178; 10.5	1,152; 10.5
Gas support (%)	30.4	26.7	30.7	30.5
Air temperature entering air heater; ° F	173	158	171	177
Air temperature leaving air heater; ° F	413	410	415	415
Air temperature into boiler; ° F	389	386	391	391
Furnace outlet temperature; ° F	600	611	597	591
Gas temperature leaving air heater; ° F	377	379	370	360
Bagfilter inlet temperature; ° F	390	395	381	375
Bagfilter outlet temperature; ° F	363	371	354	345
Ash content of fly ash;%	61.16	58.20	37.61	24.02
Combustion air flow; acfm	4,259	4,438	3,732	3,835
Boiler draft; in. W.C.	-0.05	-0.04	-0.05	-0.05
Boiler efficiency; %	79.97	80.16	80.11	79.41
Atomizing air pressure; psig	135	135	135	80
Atomizing air flow rate; lbs/h	754	768	781	389
A/F ratio; lb/lb	0.57	0.54	0.66	0.34
CWSF temperature; ° F	141	147	141	155
Natural gas temperature; ° F	89	82	86	93
Coal combustion efficiency; %	98.7	98.4	97.7	95.1
Overall combustion efficiency; %	98.9	98.6	98.0	95.8

EMISSIONS (Dry Basis)

O ₂ ; %	3.7	3.2	3.7	3.4
CO; ppm	N/M	N/M	N/M	212.8
CO ₂ ; %	14.1	15.0	14.4	15.9
SO ₂ ; ppm	669	828	599	368
NO _x ppm	242	224	188	488

FUEL ANALYSIS DATA

Solids Content; %	55.5%	56.4%	59.5%	59.9%
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Summary Sheet

Phase III MCWM Testing

TEST/DESCRIPTION:

Date	6/18/98	6/23/98	6/24/98	7/21/98
Filtering device	Fabric	Ceramic	Fabric	Fabric
Coal seam	Taggart	Hiawatha	Hiawatha	Taggart
Cleaning tech.	Oil Agglomeration	Oil Agglomeration	Oil Agglomeration	Froth Floatation
Natural gas support	23%	30%	23%	30%

WATER/STEAM SIDE

Steam flow rate; lb/h	11,143	11,476	9,446	10,073
Water temperature into boiler; ° F	231	231	231	229
Drum pressure; psig	230	230	229	231
Calorimeter temperature; ° F	318	318	316	318
Steam temperature; ° F	397	397	397	398
Steam quality; %	100.1	100.1	100.0	100.1
Blowdown rate; lb/h	3,322	3,327	3,320	3,332

Air, Fuel, Flue Gas Side

Total firing rate; MM Btu/h	15.0	14.7	14.0	14.5
Natural gas flow rate; lb/h; MM Btu/h	150; 3.5	195; 4.5	145; 3.5	190; 4.5
CWSF flow rate; lb/h; MM Btu/h	1,272; 11.5	1,353; 10.5	1,428; 11.6	1,200; 10.5
Gas support (%)	23.4	30.6	25.0	31.0
Air temperature entering air heater; ° F	179	180	184	186
Air temperature leaving air heater; ° F	424	423	402	415
Air temperature into boiler; ° F	398	399	379	391
Furnace outlet temperature; ° F	597	600	559	582
Gas temperature leaving air heater; ° F	372	378	359	370
Bagfilter inlet temperature; ° F	381	388	368	442
Bagfilter outlet temperature; ° F	353	366	344	347
Ash content of fly ash; %	29.00	52.27	48.30	20.84
Combustion air flow; acfm	3,774	3,859	3,806	3,657
Boiler draft; in. W.C.	-0.05	-0.05	-0.05	-0.05
Boiler efficiency; %	79.13	77.86	76.95	76.75
Atomizing air pressure; psig	135	135	135	135
Atomizing air flow rate; lbs/h	761	728	573	628
A/F ratio; lb/lb	0.60	0.54	0.40	0.52
CWSF temperature; ° F	156	140	147	130
Natural gas temperature; ° F	98	97	95	97
Coal combustion efficiency; %	96.4	96.5	97.7	93.7
Overall combustion efficiency; %	96.8	97.0	97.9	94.6

EMISSIONS (Dry Basis)

O ₂ ; %	3.8	3.6	4.7	4.2
CO; ppm	N/M	N/M	N/M	N/M
CO ₂ ; %	13.9	13.9	13.4	12.6
SO ₂ ; ppm	204	N/M	N/M	250
NO _x ppm	584	610.2	525	227

FUEL ANALYSIS DATA

Solids Content; %	60.3%	54.3%	56.0%	59.1%
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Summary Sheet

Phase III MCWM Testing

TEST/DESCRIPTION:

Date	8/11/98	8/12/98	8/17/98	8/19/98	8/26/98
Filtering device	Fabric	Fabric	Fabric	Fabric	Fabric
Coal seam	Taggart	Taggart	Hiawatha	Hiawatha	Middle Kittanning
Cleaning tech.	Froth Floatation	Froth Floatation	Froth Floatation	Froth Floatation	None
Natural gas support	30%	30%	30%	27%	30%

WATER/STEAM SIDE

Steam flow rate; lb/h	11,342	10,855	11,582	9,949	10,633
Water temperature into boiler; ° F	229	229	234	233	233
Drum pressure; psig	N/M	237	233	232	231
Calorimeter temperature; ° F	319	317	319	317	318
Steam temperature; ° F	398	398	398	397	397
Steam quality; %	N/M	100.0	100.1	100.0	100.1
Blowdown rate; lb/h	3,443	3,372	3,347	3,339	3,328

Air, Fuel, Flue Gas Side

Total firing rate; MM Btu/h	15.1	14.9	15.3	14.6	15
Natural gas flow rate; lb/h; MM Btu/h	198; 4.6	192; 4.5	195; 4.5	174; 4.0	187; 4.5
CWSF flow rate; lb/h; MM Btu/h	1,210; 10.5	1,182; 10.5	1,392; 10.5	1,440; 11.0	1,176; 10.5
Gas support (%)	29.8	30.2	29.4	27.3	30.0
Air temperature entering air heater; ° F	186	184	180	177	182
Air temperature leaving air heater; ° F	421	412	419	410	419
Air temperature into boiler; ° F	398	387	396	386	396
Furnace outlet temperature; ° F	598	576	599	577	592
Gas temperature leaving air heater; ° F	373	365	379	365	379
Bagfilter inlet temperature; ° F	380	369	386	369	376
Bagfilter outlet temperature; ° F	366	343	364	348	353
Ash content of fly ash;%	16.06	14.92	50.72	53.74	42.50
Combustion air flow; acfm	3,775	3,610	3,979	3,943	3,885
Boiler draft; in. W.C.	-0.05	-0.05	-0.05	-0.06	-0.05
Boiler efficiency; %	75.58	77.45	78.59	77.52	77.41
Atomizing air pressure; psig	80	135	135	135	135
Atomizing air flow rate; lbs/h	453	681	706	632	690
A/F ratio; lb/lb	0.38	0.58	0.51	0.44	0.59
CWSF temperature; ° F	158	152	141	110	152
Natural gas temperature; ° F	98	94	96	89	99
Coal combustion efficiency; %	90.9	92.5	97.5	97.4	93.6
Overall combustion efficiency; %	92.1	93.6	97.8	97.7	94.5

EMISSIONS (Dry Basis)

O2; %	3.6	3.6	3.6	4.4	4.0
CO; ppm	N/M	N/M	N/M	N/M	N/M
CO2; %	14.1	14.2	13.9	12.6	13.2
SO2; ppm	437	417	211	442	290
NOx ppm	276	282	734	475	N/M

FUEL ANALYSIS DATA

Solids Content; %	57.7%	59.3%	55.0%	54.5%	60.6%
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Summary Sheet

Phase III MCWM Testing

TEST/DESCRIPTION:

Date	9/9/98
Filtering device	Fabric
Coal seam	Taggart
Cleaning tech.	Oil Agglomeration
Natural gas support	30%

WATER/STEAM SIDE

Steam flow rate; lb/h	10,474
Water temperature into boiler; ° F	234
Drum pressure; psig	233
Calorimeter temperature; ° F	317
Steam temperature; ° F	397
Steam quality; %	100.0
Blowdown rate; lb/h	3,348

Air, Fuel, Flue Gas Side

Total firing rate; MM Btu/h	14.8
Natural gas flow rate; lb/h; MM Btu/h	186; 4.5
CWSF flow rate; lb/h; MM Btu/h	1,170; 10.3
Gas support (%)	30.5
Air temperature entering air heater; ° F	168
Air temperature leaving air heater; ° F	406
Air temperature into boiler; ° F	383
Furnace outlet temperature; ° F	570
Gas temperature leaving air heater; ° F	359
Bagfilter inlet temperature; ° F	354
Bagfilter outlet temperature; ° F	339
Ash content of fly ash; %	16.40
Combustion air flow; acfm	3,706
Boiler draft; in. W.C.	-0.05
Boiler efficiency; %	77.99
Atomizing air pressure; psig	80
Atomizing air flow rate; lbs/h	370
A/F ratio; lb/lb	0.32
CWSF temperature; ° F	158
Natural gas temperature; ° F	86
Coal combustion efficiency; %	93.0
Overall combustion efficiency; %	94.0

EMISSIONS (Dry Basis)

O ₂ ; %	4.0
CO; ppm	N/M
CO ₂ ; %	N/M
SO ₂ ; ppm	150
NO _x ppm	490

FUEL ANALYSIS DATA

Solids Content; %	59.9%
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Summary Sheet

DMC Testing

TEST/DESCRIPTION:	5/5/98	7/8/98	7/13/98	7/15/98	9/27/99	9/28/99	9/29/99	9/30/99	10/21/99	10/25/99	10/26/99	10/27/99
Filtering Device:					Sleeve 0.5	Sleeve 3.5	Sleeve 3.5	Sleeve 4.5		Sleeve 3.5	Sleeve 3.5	Sleeve 3.5
Burner Settings:	Prim 0	Prim 0	Prim 0	Prim 0	Center 1	Center 0	Center 0	Center 0		Center 2	Outer 7.5	Center 0
	Sec 100	Sec 100	Sec 100	Sec 100	Inner Max	Inner 6.5	Inner 6.5	Inner Max		Inner Max	Inner 6.5	Inner 6.5
	Tert 50	Tert 50	Tert 50	Tert 50	Outer Max	Outer	Outer 7.5	Outer Max		Outer Max	Outer 7.5	Outer 7.5
	Coal Gun +2	Coal Gun +2	Coal Gun +2	Coal Gun +2	Coal Gun -7	Coal Gun -7	Coal Gun -7	Coal Gun -7	Coal Gun	Coal Gun -6	Coal Gun -6	Coal Gun -6
	Gas Gun +2	Gas Gun +2	Gas Gun +2	Gas Gun +2	Gas Gun -2	Gas Gun -2	Gas Gun -2	Gas Gun -2	Gas Gun	Gas Gun -2	Gas Gun -2	Gas Gun -2
WATER/STEAM SIDE												
Steam flow rate; lb/h	11,700	11,505	11,066	11,360	10,867	10,492	11,107	10,496	10,809	10,875	10,714	10,067
Water temperature into boiler; °F	233	231	230	230	236	236	236	236	236	236	236	236
Drum pressure; psig	226	229	227	229	246	244	244	245	246	246	244	243
Calorimeter temperature; °F	318	318	317	318	319	319	319	319	319	319	319	320
Steam temperature; °F	397	398	398	398	402	402	402	401	402	402	402	401
Steam quality; %	100.1	100.1	100.0	100.1	100.3	100.4	100.3	100.3	100.3	100.3	100.3	100.4
Blowdown rate; lb/h	3,332	3,345	3,379	3,337	3,439	3,425	3,423	3,429	3,440	3,438	3,429	3,415
AIR,FUEL, FLUE GAS SIDE												
Coal Flow rate; lb/h	1,080	1,080	1,080	1,080	1,050	1,050	1,044	1,044	1,080	1,050	1,050	1,050
Coal flow rate; MMBtu/h	14.9	15.5	15.4	15.5	15.0	15.0	14.9	14.9	15.5	15.1	15.3	14.9
Boiler outlet temperature; °F	614	605	604	612	612	587	602	595	586	602	603	604
Gas temperature leaving air heater; °F	384	382	380	386	369	363	367	365	360	364	366	358
Air temperature entering air heater; °F	173	182	179	186	166	183	178	170	171	166	174	137
Air temperature leaving air heater; °F	417	417	411	421	421	428	427	420	412	422	430	416
Air temperature into boiler; °F	403	405	397	409	400	566	545	400	390	400	409	395
Ash content of particulate; %	44.70	38.94	45.16	39.41	47.36	47.66	58.53	47.84	48.47	46.79	45.23	54.48
Coal combustion efficiency; %	94.6	92.4	95.6	94.1	95.2	94.6	96.6	94.7	96.0	95.7	95.6	95.5
Combustion air flow; acfm	3,208	3,227	3,234	3,120	3,082	2,887	3,126	2,809	3,029	2,956	2,901	3,022
Boiler draft; inches W.C.	-0.06	-0.06	-0.09	-0.09	-0.08	-0.07	-0.07	-0.07	-0.07	-0.06	-0.06	-0.06
Boiler efficiency; %	83.4	81.6	83.8	83.0	84.9	85.0	86.6	84.6	84.8	85.2	84.9	84.4
Mill air flow rate; acfm	401	405	396	400	397	409	403	361	402	403	408	382
Mill outlet temperature; °F	243	251	322	239	257	239	235	220	231	235	235	216
Filter type	Fabric	Fabric	Ceramic	Ceramic	Fabric	Fabric	Fabric	Fabric	Ceramic	Ceramic	Ceramic	Ceramic
Pressure drop; in W.C.	4.2	3.9	13.1	14.0	1.8	1.5	1.7	1.6	1.5	1.8	2.1	2.1
Filter inlet; ° F	381	383	380	387	384	377	385	384	383	381	389	374
Filter outlet; ° F	330	338	354	384	455	461	438	460	360	355	387	345
EMISSIONS												
O2 %	3.4	3.9	4.0	3.4	4.0	3.4	3.9	3.9	3.2	3.3	3.7	4.5
CO ppm	40	79	26	69	169	163	228	137	214	348	595	384
CO2 %	14.5	15.7	14.9	15.6	14.5	15.0	15.1	14.3	15.5	15.4	15.3	14.0
SO2 ppm	N/A	504	494	506	440	362	455	357	469	456	475	427
NOx ppm	413	355	375	397	192	402	411	351	399	336	236	306
MISCELLANEOUS DATA												
Maximum load												
(based on 14,700 lb steam/h); %	79.6	78.3	75.3	77.3	73.9	71.4	75.6	71.4	73.5	74.0	72.9	68.5
NOTES												
Flue gas temperature leaving air heater and boiler efficiencies from 9/27/99 through 4/26/00 are incorrect. The expansion joint between the air heater and outlet ducting developed a leak and ambient air was drawn into the system. First noticeable leak was observed starting 9/27/99, second and more serious leak was observed starting 3/15/00, and expansion joint was replaced between the 4/27/00 and 5/15/00 testing.												

Summary Sheet

DMC Testing

TEST/DESCRIPTION:		10/28/99	10/29/99	11/1/99	11/2/99	11/3/99	11/4/99	3/15/00	3/16/00	3/17/00	3/20/00	3/21/00	4/5/00
Filtering Device:		Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 5.5	Sleeve 3.5	Sleeve 3.5
Burner Settings:		Center 0	Center 0	Center 0	Center 0	Center 0	Center 0	Center 1	Center 1	Center 1	Center 0	Center 1	Center 1
		Inner 6.5	Inner 6.5	Inner Max	Inner Max	Inner Max	Inner Max	Inner 6.5	Inner 6.5	Inner 6.5	Inner Max	Inner 6.5	Inner 6.5
		Outer 7.5	Outer 7.5	Outer Max	Outer Max	Outer Max	Outer Max	Outer 7.5	Outer 7.5	Outer 7.5	Outer Max	Outer 7.5	Outer 7.5
		Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6
		Gas Gun -2	Gas Gun -2	Gas Gun -2	Gas Gun -2	Gas Gun -2	Gas Gun -2	Gas Gun -2	Gas Gun -2	Gas Gun -6	Gas Gun -2	Gas Gun -2	Gas Gun -2
WATER/STEAM SIDE													
Steam flow rate; lb/h		10,630	10,766	10,879	10,688	10,833	10,739	10,203	10,572	10,400	10,417	10,484	10,922
Water temperature into boiler; °F		236	235	235	235	235	236	236	235	236	235	236	234
Drum pressure; psig		241	242	242	241	238	240	253	251	250	249	248	231
Calorimeter temperature; °F		319	319	320	319	320	319	320	320	318	319	319	318
Steam temperature; °F		401	401	401	401	400	401	404	404	403	403	402	398
Steam quality; %		100.3	100.3	100.4	100.3	100.4	100.3	100.4	100.4	100.3	100.3	100.3	100.3
Blowdown rate; lb/h		3,402	3,409	3,411	3,407	3,382	3,397	3,491	3,477	3,465	3,459	3,453	3,330
AIR,FUEL, FLUE GAS SIDE													
Coal flow rate; lb/h		1,050	1,050	1,050	1,050	1,050	1,050	1,055	1,055	1,055	1,055	1,055	1,055
Coal flow rate; MMBtu/h		15.0	14.9	15.0	14.7	14.7	14.8	14.8	14.8	14.8	15.1	14.7	13.5
Boiler outlet temperature; °F		605	611	611	612	618	609	588	623	637	646	593	604
Gas temperature leaving air heater; °F		367	369	367	357	359	361	340	344	337	347	316	313
Air temperature entering air heater; °F		160	171	176	167	147	159	179	175	150	157	152	157
Air temperature leaving air heater; °F		421	431	430	429	424	420	415	439	443	452	407	420
Air temperature into boiler; °F		398	409	408	408	404	398	390	414	417	424	381	398
Ash content of particulate; %		48.78	53.17	51.19	47.48	56.75	64.42	39.26	37.08	31.61	31.52	42.48	44.11
Coal combustion efficiency; %		95.3	95.7	95.8	95.9	96.8	97.6	93.0	91.9	91.1	91.1	93.2	85.6
Combustion air flow; acfm		3,163	2,679	3,133	3,012	2,987	3,125	3,011	3,047	2,929	2,946	3,056	3,132
Boiler draft; inches W.C.		-0.06	-0.04	-0.04	-0.11	-0.05	-0.08	-0.07	-0.07	-0.06	-0.08	-0.07	-0.06
Boiler efficiency; %		84.3	84.6	84.4	84.4	85.4	85.8	83.4	82.8	82.0	81.5	83.4	79.2
Mill air flow rate; acfm		406	404	404	412	407	406	403	403	392	407	324	405
Mill outlet temperature; °F		218	230	238	223	221	209	220	238	220	230	170	223
Filter type		Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic
Pressure drop; in W.C.		4.6	4.6	4.8	5.0	5.0	5.5	4.7	5.4	5.6	5.9	5.6	5.9
Filter inlet; ° F		377	386	388	378	372	375	358	374	356	389	346	342
Filter outlet; ° F		343	375	374	362	385	351	419	396	369	378	315	360
EMISSIONS													
O2 %		3.3	3.4	3.3	3.8	3.7	3.2	3.5	3.5	4.0	3.2	3.9	3.9
CO ppm		268	305	291	432	225	152	319	511	1,068	1,045	367	515
CO2 %		14.9	15.2	15.1	15.1	15.4	15.0	15.1	15.7	15.4	15.8	13.6	15.4
SO2 ppm		469	478	461	465	470	462	463	471	445	488	312	88
NOx ppm		342	325	353	500	338	371	367	355	342	373	317	343
MISCELLANEOUS DATA													
Maximum load													
(based on 14,700 lb steam/h); %		72.3	73.2	74.0	72.7	73.7	73.1	69.4	71.9	70.7	70.9	71.3	74.3
NOTES													
Flue gas temperature leaving air heater and b													
expansion joint between the air heater and ou													
system. First noticeable leak was observed st													
3/15/00, and expansion joint was replaced be													

Summary Sheet

DMC Testing

TEST/DESCRIPTION:	4/6/00	4/7/00	4/11/00	4/12/00	4/18/00	4/19/00	4/20/00	4/21/00	4/24/00	4/25/00	4/26/00	7/5/00
Filtering Device:	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5
Burner Settings:	Center 1	Center 1	Center 1	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2
	Inner Max	Inner Max	Inner 6.5	Inner 6.5	Inner 6.5	Inner 6.5	Inner 6.5	Inner 6.5	Inner Max	Inner 6.5	Inner 6.5	Inner 6.5
	Outer Max	Outer Max	Outer 7.5	Outer 7.5	Outer 7.5	Outer 7.5	Outer 7.5	Outer 7.5	Outer Max	Outer 7.5	Outer 7.5	Outer 7.5
	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6
	Gas Gun -2	Gas Gun -2	Gas Gun -6	Gas Gun -6	Gas Gun -2	Gas Gun -6	Gas Gun -6	Gas Gun -6	Gas Gun -2	Gas Gun -6	Gas Gun -6	Gas Gun -6
					Corning Slipstream	Corning Slipstream	Corning Slipstream	Corning Slipstream	Corning Slipstream	Corning Slipstream	Corning Slipstream	Corning Slipstream
WATER/STEAM SIDE												
Steam flow rate; lb/h	11,137	11,397	11,468	11,540	11,456	11,863	11,709	11,698	11,387	11,383	11,671	10,477
Water temperature into boiler; °F	233	233	234	234	234	233	233	233	233	233	233	231
Drum pressure; psig	231	232	226	226	222	221	219	219	225	225	222	226
Calorimeter temperature; °F	318	318	317	317	317	318	319	318	318	319	319	318
Steam temperature; °F	398	396	396	396	395	394	394	394	396	396	395	396
Steam quality; %	100.3	100.3	100.2	100.2	100.2	100.3	100.3	100.3	100.3	100.3	100.3	100.3
Blowdown rate; lb/h	3,329	3,339	3,294	3,296	3,264	3,255	3,241	3,239	3,283	3,289	3,267	3,293
AIR,FUEL, FLUE GAS SIDE												
Coal Flow rate; lb/h	1,055	1,055	1,055	1,055	1,055	1,055	1,055	1,055	1,055	1,055	1,055	1,080
Coal flow rate; MMBtu/h	14.3	14.6	14.7	14.4	14.5	14.5	14.5	14.5	14.4	14.4	14.4	14.5
Boiler outlet temperature; °F	613	619	626	625	607	618	608	635	608	626	627	602
Gas temperature leaving air heater; °F	320	284	266	257	263	272	274	260	277	281	287	396
Air temperature entering air heater; °F	178	166	160	157	161	167	168	169	175	171	158	191
Air temperature leaving air heater; °F	431	430	430	429	404	415	421	423	414	426	422	424
Air temperature into boiler; °F	409	408	406	407	383	395	400	404	393	406	402	404
Ash content of particulate; %	49.63	50.71	61.75	68.90	50.10	61.02	76.25	78.38	53.74	56.27	55.93	61.89
Coal combustion efficiency; %	84.8	94.8	96.5	97.6	94.7	96.6	99.0	98.6	95.5	95.6	96.3	96.5
Combustion air flow; acfm	3,051	3,012	3,149	2,868	3,147	3,172	3,253	3,243	3,203	3,267	3,306	2,777
Boiler draft; inches W.C.	-0.07	-0.07	-0.07	-0.06	-0.07	-0.06	-0.07	-0.06	-0.07	-0.06	-0.07	-0.07
Boiler efficiency; %	78.5	87.0	88.8	90.1	87.1	88.3	90.0	90.1	87.1	87.1	87.3	83.7
Mill air flow rate; acfm	403	405	404	401	401	401	408	404	407	404	401	317
Mill outlet temperature; °F	222	224	219	221	222	228	217	223	216	220	224	224
Filter type	Ceramic	Ceramic	Ceramic	Ceramic	Fabric	Fabric	Fabric	Fabric	Fabric	Fabric	Fabric	Ceramic
Pressure drop; in W.C.	6.2	6.7	7.2	7.3	2.8	1.4	1.4	1.7	1.2	1.2	1.2	4.3
Filter inlet; °F	352	343	340	335	323	328	343	335	332	339	343	296
Filter outlet; °F	354	334	327	323	287	295	308	299	299	304	306	270
EMISSIONS												
O2 %	3.7	3.6	3.8	3.8	4.0	3.6	3.8	3.8	4.0	3.9	3.7	4.7
CO ppm	228	236	179	100	156	179	133	116	252	306	227	849
CO2 %	15.3	15.4	15.2	15.3	15.1	15.6	15.5	15.3	15.1	15.2	15.4	12.4
SO2 ppm	90	88	88	253	341	426	472	526	412	442	444	30
NOx ppm	325	339	397	419	368	425	469	469	392	436	466	374
MISCELLANEOUS DATA												
Maximum load												
(based on 14,700 lb steam/h); %	75.8	77.5	78.0	78.5	77.9	80.7	79.7	79.6	77.5	77.4	79.4	71.3
NOTES												
Flue gas temperature leaving air heater and boiler expansion joint between the air heater and boiler system. First noticeable leak was observed September 3/15/00, and expansion joint was replaced by September 3/15/00.												

Summary Sheet DMC Testing

TEST/DESCRIPTION:	7/6/00	7/7/00	7/10/00	7/11/00	7/13/00	7/14/00	7/17/00	7/18/00	7/19/00	7/20/00	7/21/00	7/24/00
Filtering Device:	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5
Burner Settings:	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2
	Inner 6.5	Inner Max	Inner Max	Inner Max	Inner Max	Inner Max	Inner Max	Inner Max	Inner Max	Inner Max	Inner Max	Inner Max
	Outer 7.5	Outer Max	Outer Max	Outer Max	Outer Max	Outer Max	Outer Max	Outer Max	Outer Max	Outer Max	Outer Max	Outer Max
	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6
	Gas Gun -6	Gas Gun -2	Gas Gun -6	Gas Gun -2	Gas Gun -2	Gas Gun -6	Gas Gun -2	Gas Gun -2	Gas Gun -6	Gas Gun -2	Gas Gun -2	Gas Gun -6
	Corning	Corning	Corning	Corning	Corning	Corning	Corning	Corning	Corning	Corning	Corning	Corning
WATER/STEAM SIDE	Slipstream	Slipstream	Slipstream	Slipstream	Slipstream	Slipstream	Slipstream	Slipstream	Slipstream	Slipstream	Slipstream	Slipstream
Steam flow rate; lb/h	11,278	11,069	11,360	11,255	7,849	11,303	11,613	11,448	11,179	11,955	11,605	11,616
Water temperature into boiler; °F	231	231	231	231	222	231	231	231	231	231	231	232
Drum pressure; psig	226	225	228	225	216	218	217	217	217	217	214	214
Calorimeter temperature; °F	306	318	317	318	317	317	317	317	316	318	317	317
Steam temperature; °F	396	396	397	396	393	394	394	394	393	393	393	392
Steam quality; %	99.6	100.3	100.3	100.3	100.2	100.2	100.2	100.2	100.2	100.3	100.2	100.2
Blowdown rate; lb/h	3,295	3,285	3,309	3,289	3,222	3,235	3,230	3,228	3,228	3,205	3,197	3,202
AIR,FUEL, FLUE GAS SIDE												
Coal Flow rate; lb/h	1,080	1,080	1,080	1,110	1,110	1,110	1,110	1,080	1,101	1,101	1,101	1,091
Coal flow rate; MMBtu/h	14.7	14.6	14.5	15.0	15.1	15.0	15.1	14.6	14.9	15.1	15.1	14.4
Boiler outlet temperature; °F	617	617	625	600	548	603	614	585	589	602	613	606
Gas temperature leaving air heater; °F	404	406	413	396	360	392	398	383	382	391	397	392
Air temperature entering air heater; °F	186	182	188	190	202	191	191	187	182	188	188	189
Air temperature leaving air heater; °F	429	427	432	423	412	431	438	416	419	433	438	437
Air temperature into boiler; °F	407	406	411	403	388	408	414	394	396	411	416	413
Ash content of particulate; %	62.03	59.45	52.39	47.33	39.28	42.69	42.52	41.79	34.88	36.00	38.98	58.77
Coal combustion efficiency; %	96.6	96.2	94.6	94.2	91.7	92.7	92.2	92.8	89.4	90.4	92.0	94.7
Combustion air flow; acfm	3,241	3,160	3,329	3,186	2,515	3,060	3,114	3,133	3,228	3,111	3,136	2,983
Boiler draft; inches W.C.	-0.06	-0.07	-0.07	-0.02	-0.08	-0.07	-0.07	-0.05	-0.05	-0.07	-0.07	-0.05
Boiler efficiency; %	85.1	83.7	80.2	83.3	79.4	81.8	81.6	82.2	79.7	80.3	81.4	83.8
Mill air flow rate; acfm	400	403	402	405	294	416	418	411	414	425	422	410
Mill outlet temperature; °F	242	243	257	191	125	137	140	141	141	144	143	142
Filter type	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic
Pressure drop; in W.C.	6.5	6.7	6.6	5.9	3.9	5.3	5.4	5.4	5.4	5.7	6.3	5.8
Filter inlet; °F	383	386	395	377	340	372	378	363	361	375	377	372
Filter outlet; °F	353	357	363	347	321	340	345	331	329	345	347	342
EMISSIONS												
O2 %	3.9	4.8	8.5	4.5	8.9	5.2	4.2	4.5	4.5	4.1	4.5	3.3
CO ppm	412	811	345	275	424	424	405	273	537	786	666	463
CO2 %	15.5	13.9	10.0	14.8	10.1	14.1	15.0	14.4	14.5	15.0	14.6	15.7
SO2 ppm	50	94	59	105	71	98	89	88	99	91	91	97
NOx ppm	578	330	210	417	288	409	435	419	517	499	506	620
MISCELLANEOUS DATA												
Maximum load												
(based on 14,700 lb steam/h); %	76.7	75.3	77.3	76.6	53.4	76.9	79.0	77.9	76.0	81.3	78.9	79.0
NOTES												
Flue gas temperature leaving air heater and boiler expansion joint between the air heater and boiler system. First noticeable leak was observed at 3/15/00, and expansion joint was replaced by												

Summary Sheet

DMC Testing

TEST/DESCRIPTION:	8/21/00	8/22/00	8/23/00	8/24/00	8/25/00	8/29/00	8/30/00	8/31/00	9/1/00	9/5/00	9/6/00	9/7/00
Filtering Device:	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5
Burner Settings:	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2
	Inner Max	Inner Max	Inner 6.5	Inner Max	Inner Max	Inner 6.5	Inner 6.5	Inner 6.5	Inner 6.5	Inner 6.5	Inner 6.5	Inner 6.5
	Outer Max	Outer Max	Outer 7.5	Outer Max	Outer Max	Outer 7.5	Outer 7.5	Outer 7.5	Outer 7.5	Outer 7.5	Outer 7.5	Outer 7.5
	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6
	Gas Gun -2	Gas Gun -2	Gas Gun -6	Gas Gun -2	Gas Gun -2	Gas Gun -6	Gas Gun -6	Gas Gun -6	Gas Gun -6	Gas Gun -6	Gas Gun -6	Gas Gun -6
	Corning	Corning	Corning	Corning	Corning	Corning	Corning	Corning	Corning	Corning	Corning	Corning
WATER/STEAM SIDE	Slipstream	Slipstream	Slipstream	Slipstream	Slipstream	Slipstream	Slipstream	Slipstream	Slipstream	Slipstream	Slipstream	Slipstream
Steam flow rate; lb/h	11,940	12,110	11,881	12,371	12,470	12,370	12,463	12,606	12,189	11,448	11,912	11,747
Water temperature into boiler; °F	232	232	231	231	230	230	230	230	230	231	231	231
Drum pressure; psig	213	212	213	212	211	211	212	211	211	211	213	213
Calorimeter temperature; °F	318	316	318	317	313	319	319	318	318	317	317	318
Steam temperature; °F	393	392	392	392	392	392	392	392	392	391	392	392
Steam quality; %	100.3	100.2	100.3	100.2	100.0	100.3	100.3	100.3	100.3	100.2	100.3	100.3
Blowdown rate; lb/h	1,244	3,191	3,195	3,190	3,182	3,186	1,245	3,182	1,243	3,182	3,198	1,247
AIR,FUEL, FLUE GAS SIDE												
Coal Flow rate; lb/h	1,091	1,091	1,091	1,091	1,091	1,091	1,091	1,091	1,091	1,091	1,091	1,091
Coal flow rate; MMBtu/h	14.4	14.5	14.4	14.5	14.7	15.1	15.0	15.1	14.1	14.2	14.1	14.1
Boiler outlet temperature; °F	613	621	588	611	600	598	609	616	609	597	610	618
Gas temperature leaving air heater; °F	404	411	388	407	399	402	408	409	402	392	406	411
Air temperature entering air heater; °F	187	182	185	184	184	183	182	183	183	183	177	178
Air temperature leaving air heater; °F	450	447	432	443	432	435	438	440	436	432	436	441
Air temperature into boiler; °F	410	408	390	404	394	397	401	404	398	393	397	403
Ash content of particulate; %	70.27	79.94	60.26	77.22	75.00	71.79	81.36	80.69	77.59	69.98	90.61	92.73
Coal combustion efficiency; %	96.4	98.2	95.5	98.1	98.0	98.5	98.9	98.7	98.4	95.7	98.9	99.3
Combustion air flow; acfm	3,102	3,172	3,103	3,215	3,228	3,260	3,286	3,270	3,206	3,106	3,213	3,185
Boiler draft; inches W.C.	-0.09	-0.09	-0.07	-0.09	-0.10	-0.07	-0.08	-0.09	-0.09	-0.10	-0.09	-0.08
Boiler efficiency; %	85.1	86.1	84.6	90.9	86.3	86.8	86.9	86.8	86.6	84.7	86.8	87.2
Mill air flow rate; acfm	405	408	408	411	405	405	410	414	406	391	404	408
Mill outlet temperature; °F	217	212	207	212	207	214	214	210	208	208	219	212
Filter type	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic
Pressure drop; in W.C.	6.4	6.9	6.3	7.0	7.0	7.6	8.0	8.1	7.5	6.4	7.5	7.7
Filter inlet; ° F	386	392	370	389	382	388	390	392	385	375	387.2	392
Filter outlet; ° F	364	389	373	410	368	367	374	376	364	351	363	367
EMISSIONS												
O2 %	3.1	3.5	3.5	3.3	3.5	3.4	3.5	3.5	3.3	4.1	3.8	3.7
CO ppm	368	261	198	134	98	115	104	164	201	148	98	86
CO2 %	16.1	15.7	15.5	74.2	15.6	15.7	15.5	15.7	15.6	15.2	15.4	15.8
SO2 ppm	453	409	509	422	221	402	414	424	415	404	387	384
NOx ppm	545	541	441	569	561	613	666	677	590	461	615	670
MISCELLANEOUS DATA												
Maximum load (based on 14,700 lb steam/h); %	81.2	82.4	80.8	84.2	84.8	84.1	84.8	85.8	82.9	77.9	81.0	79.9
NOTES												
Flue gas temperature leaving air heater and boiler expansion joint between the air heater and boiler system. First noticeable leak was observed at 3/15/00. and expansion joint was replaced by												

Summary Sheet

DMC Testing

TEST/DESCRIPTION:		9/8/00	9/11/00	9/12/00	9/13/00	9/14/00	9/15/00	9/18/00	9/19/00	9/20/00	9/21/00	9/22/00	9/25/00
Filtering Device:		Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5
Burner Settings:		Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2
		Inner 6.5	Inner 6.5	Inner Max	Inner Max	Inner Max	Inner Max	Inner 6.5	Inner 6.5	Inner 6.5	Inner 6.5	Inner 6.5	Inner Max
		Outer 7.5	Outer 7.5	Outer Max	Outer Max	Outer Max	Outer Max	Outer 7.5	Outer 7.5	Outer 7.5	Outer 7.5	Outer 7.5	Outer Max
		Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6
		Gas Gun -6	Gas Gun -6	Gas Gun -2	Gas Gun -2	Gas Gun -2	Gas Gun -2	Gas Gun -6	Gas Gun -6	Gas Gun -6	Gas Gun -6	Gas Gun -6	Gas Gun -2
		Corning	Corning	Corning	Corning	Corning	Corning	Corning	Corning	Corning	Corning	Corning	Corning
WATER/STEAM SIDE		Slipstream	Slipstream	Slipstream	Slipstream	Slipstream	Slipstream	Slipstream	Slipstream	Slipstream	Slipstream	Slipstream	Slipstream
Steam flow rate; lb/h		11,861	11,544	11,934	12,048	12,303	12,044	11,479	11,981	12,099	11,975	11,793	11,345
Water temperature into boiler; °F		231	231	230	230	231	230	231	230	230	230	201	231
Drum pressure; psig		213	211	209	211	219	214	213	212	211	212	201	211
Calorimeter temperature; °F		318	318	318	318	318	318	318	317	318	317	314	317
Steam temperature; °F		392	392	392	392	392	392	392	391	391	391	391	423
Steam quality; %		100.3	100.3	100.3	100.3	100.3	100.3	100.3	100.2	100.3	100.2	100.0	100.2
Blowdown rate; lb/h		1,246	3,184	3,168	3,184	3,197	3,206	3,199	3,191	3,181	3,190	3,082	3,186
AIR,FUEL, FLUE GAS SIDE													
Coal Flow rate; lb/h		1,091	1,091	1,091	1,091	1,091	1,091	1,091	1,091	1,091	1,091	1,091	1,091
Coal flow rate; MMBtu/h		14.1	14.3	14.5	14.5	14.6	14.6	14.4	14.5	14.4	14.5	14.4	14.5
Boiler outlet temperature; °F		618	587	606	614	622	615	589	606	610	617	583	592
Gas temperature leaving air heater; °F		409	388	406	408	410	404	386	401	406	407	389	385
Air temperature entering air heater; °F		182	182	187	185	182	181	186	183	183	179	187	173
Air temperature leaving air heater; °F		442	429	441	443	449	444	431	440	440	440	430	424
Air temperature into boiler; °F		404	388	401	407	412	407	389	403	403	403	397	381
Ash content of particulate; %		91.71	82.20	86.63	80.50	76.37	72.51	64.41	79.33	86.93	89.99	89.66	72.01
Coal combustion efficiency; %		99.3	97.8	98.5	98.3	98.5	97.9	95.6	98.0	98.9	99.2	99.3	96.9
Combustion air flow; acfm		3,189	3,032	3,150	3,183	3,232	3,239	3,047	3,125	3,217	3,219	2,840	3,071
Boiler draft; inches W.C.		-0.09	-0.13	-0.10	-0.09	-0.09	-0.08	-0.08	-0.08	-0.09	-0.09	-0.15	-0.08
Boiler efficiency; %		87.1	86.0	86.7	86.2	86.2	85.8	84.5	86.1	86.7	86.8	86.0	85.3
Mill air flow rate; acfm		403	400	410	414	414	406	404	407	409	414	354	404
Mill outlet temperature; °F		213	197	174	147	142	142	172	177	174	165	151	158
Filter type		Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic
Pressure drop; in W.C.		7.9	6.0	7.0	7.1	6.8	7.1	6.4	6.5	7.4	8.2	6.8	5.7
Filter inlet; °F		391	368	386	388	390	383	369	385	390	388	388	367
Filter outlet; °F		366	342	362	361	360	352	346	357	377	384	385	330
EMISSIONS													
O2 %		3.6	4.6	3.5	3.6	3.1	3.7	3.9	3.3	3.5	3.9	4.8	4.2
CO ppm		9079	186	95	150	197	411	235	219	116	87	82	174
CO2 %		15.5	14.6	15.6	15.7	16.1	15.5	15.3	15.9	15.7	15.6	13.3	15.0
SO2 ppm		375	380	421	416	428	402	434	412	401	455	353	377
NOx ppm		698	543	662	674	720	690	485	648	717	765	683	385
MISCELLANEOUS DATA													
Maximum load (based on 14,700 lb steam/h); %		80.7	78.5	81.2	82.0	83.7	81.9	78.1	81.5	82.3	81.5	80.2	77.2
NOTES													
Flue gas temperature leaving air heater and boiler expansion joint between the air heater and our system. First noticeable leak was observed September 3/15/00, and expansion joint was replaced by													

Summary Sheet

DMC Testing

TEST/DESCRIPTION:	9/26/00	9/27/00	9/28/00	4/17/01	4/19/01	4/20/01	4/23/01	4/24/01	4/25/01	4/26/01	4/27/01	4/30/01
Filtering Device:	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5
Burner Settings:	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2
	Inner 6.5	Inner 6.5	Inner 6.5	Inner 6.5	Inner 6.5	Inner 6.5	Inner 6.5	Inner 6.5	Inner 6.5	Inner 6.5	Inner max	Inner 6.5
	Outer 7.5	Outer 7.5	Outer 7.5	Outer 7.5	Outer 7.5	Outer 7.5	Outer 7.5	Outer 7.5	Outer 7.5	Outer 7.5	Outer max	Outer 7.5
	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6.5	Coal Gun -6.5	Coal Gun -6.5	Coal Gun -6
	Gas Gun -6	Gas Gun -6	Gas Gun -6	Gas Gun -6	Gas Gun -6	Gas Gun -6	Gas Gun -6	Gas Gun -6	Gas Gun -6	Gas Gun -6	Gas Gun -6	Gas Gun -6
	Corning Slipstream	Corning Slipstream	Corning Slipstream									
WATER/STEAM SIDE												
Steam flow rate; lb/h	11,916	11,780	11,944	11,241	12,205	12,625	12,437	11,597	12,259	12,884	12,993	12,546
Water temperature into boiler; °F	231	231	231	231	230	230	230	230	231	231	230	231
Drum pressure; psig	210	210	210	213	215	214	216	217	215	215	215	213
Calorimeter temperature; °F	317	317	317	316	317	317	318	317	316	316	317	317
Steam temperature; °F	324	463	400	391	392	392	393	393	391	391	392	391
Steam quality; %	100.2	100.2	100.2	100.0	100.1	100.1	100.1	100.1	100.0	100.0	100.1	100.1
Blowdown rate; lb/h	3,177	3,177	3,176	3,196	3,217	3,207	3,218	3,229	3,217	3,215	3,213	3,200
AIR,FUEL, FLUE GAS SIDE												
Coal Flow rate; lb/h	1,091	1,091	1,091	1,091	1,140	1,164	1,158	1,158	1,158	1,158	1,158	1,158
Coal flow rate; MMBtu/h	14.6	14.5	14.6	14.3	14.6	15	15.0	15.2	15.3	15.5	15.7	15.5
Boiler outlet temperature; °F	609	617	620	588	605	602	600	587	603	604	605	591
Gas temperature leaving air heater; °F	398	401	402	379	395	396	403	391	393	397	406	399
Air temperature entering air heater; °F	175	122	176	157	164	160	189	174	161	164	179	179
Air temperature leaving air heater; °F	435	438	441	421	431	428	439	432	424	425	434	431
Air temperature into boiler; °F	395	401	176	378	389	388	397	388	387	387	395	391
Ash content of particulate; %	83.84	84.54	83.58	64.36	69.34	72.29	70.54	48.73	62.89	72.78	66.77	75.61
Coal combustion efficiency; %	94.8	98.7	98.6	95.5	96.1	96.9	96.8	91.5	94.5	97.5	96.6	97.1
Combustion air flow; acfm	3,140	3,153	3,148	3,023	3,202	3,273	3,300	3,222	3,284	3,310	3,288	3,213
Boiler draft; inches W.C.	-0.09	-0.09	-0.09	-0.08	-0.1	-0.09	-0.08	-0.09	-0.07	-0.07	-0.07	-0.07
Boiler efficiency; %	83.7	86.2	86.3	84.3	80.6	84.8	85.0	81.7	83.2	85.4	84.7	85.2
Mill air flow rate; acfm	396	392	411	406	412	411	415	404	403	405	397	404
Mill outlet temperature; °F	172	141	149	133	124	125	132	133	167	164	177	180
Filter type	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic
Pressure drop; in W.C.	6.8	7.2	7.5	9.2	7.8	8.2	8.2	10.2	9.9	11.5	12	11.9
Filter inlet; ° F	381	384	386	561	403	434	386	371	372	377	476	385
Filter outlet; ° F	356	367	368	321	355	341	356	335	347	339	354	346
EMISSIONS												
O2 %	3.6	3.7	3.6	3.8	3.5	3.4	3.2	3.6	3.6	3.6	3.2	3.5
CO ppm	112	255	140	105	69	82	75	N.M.	335	68	N.M.	54
CO2 %	15.6	15.0	15.6	15.4	15.8	15.8	16.0	15.4	15.5	15.7	15.6	15.6
SO2 ppm	391	364	396	461	448	460	434	456	458	457	452	486
NOx ppm	646	628	721	571	545	634	630	508	591	607	559	627
MISCELLANEOUS DATA												
Maximum load (based on 14,700 lb steam/h); %	81.1	80.1	81.3	76.5	83.0	85.9	84.6	78.9	83.4	87.6	88.4	85.3
NOTES												
Flue gas temperature leaving air heater and b expansion joint between the air heater and ou system. First noticeable leak was observed st 3/15/00, and expansion joint was replaced by												

Summary Sheet

DMC Testing

TEST/DESCRIPTION:		5/1/01	5/3/01	5/9/01	5/15/01	5/17/01	5/22/01	5/24/01	5/31/01	6/5/01	6/12/01	4/26/02	4/29/02
Filtering Device:		Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5
Burner Settings:		Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2
		Inner 6.5	Inner 6.5	Inner 6.5	Inner 6.5	Inner 6.5	Inner 6.5	Inner 6.5	Inner 6.5	Inner 6.5	Inner 6.5	Inner 6.5	Inner 6.5
		Outer 7.5	Outer 7.5	Outer 7.5	Outer 7.5	Outer 7.5	Outer 7.5	Outer 7.5	Outer 7.5	Outer 7.5	Outer 7.5	Outer 7.5	Outer 7.5
		Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6
		Gas Gun -6	Gas Gun -6	Gas Gun -6	Gas Gun -6	Gas Gun -6	Coal Gun -6	Gas Gun -6	Coal Gun -6	Gas Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6
WATER/STEAM SIDE													
Steam flow rate; lb/h		12,320	11,967	11,735	12,186	12,019	9,107	9,158	11,870	9,063	11,950	12,025	12,442
Water temperature into boiler; °F		230	231	230	231	230	231	231	231	231	230	234	234
Drum pressure; psig		210	210	211	210	210	214	214	215	212	211	234	229
Calorimeter temperature; °F		316	317	316	316	316	316	316	317	316	317	319	320
Steam temperature; °F		391	391	391	390	390	392	392	392	391	392	397.7	395.9
Steam quality; %		100.1	100.1	100.1	100.1	100.1	100.0	100.0	100.1	100.0	100.1	100.0	100.0
Blowdown rate; lb/h		3,176	3,177	3,180	3,176	3,178	3,208	3,208	3,211	3,190	3,183	3,351	3,317
AIR,FUEL, FLUE GAS SIDE													
Coal Flow rate; lb/h		1,091	1,091	1,091	1,091	1,091	830	829.8	1,091	830	1,091	1,158	1,158
Coal flow rate; MMBtu/h		14.6	13.6	14.5	14.1	14.2	10.5	10.7	14.3	10.7	13.9	15.3	15.4
Boiler outlet temperature; °F		599	592	587	598	589	591	575	602	569	615	598	596
Gas temperature leaving air heater; °F		409	404	398	400	397	395	387	409	383	414	393	388
Air temperature entering air heater; °F		189	190	187	179	174	194	199	170	197	196	168	148
Air temperature leaving air heater; °F		442	433	429	435	428	442	439	428	434	448	421	412
Air temperature into boiler; °F		401	395	387	393	388	396	394	389	390	409	381	374
Ash content of particulate; %		75.78	77.61	N.M.	77.20	79.96	80.82	80.71	83.38	80.90	76.03	74.30	65.00
Coal combustion efficiency; %		96.9	96.1	N.M.	97.1	97.6	97.3	97.8	98.2	97.5	96.6	97.2	96.1
Combustion air flow; acfm		3,179	3,128	3,133	3,123	3,202	2,707	2,475	3,462	2,456	3,118	3,145	3,300
Boiler draft; inches W.C.		-0.08	-0.09	-0.1	-0.09	-0.09	-0.09	-0.09	-0.11	-0.09	-0.09	-0.09	-0.09
Boiler efficiency; %		84.9	84.2	N.D.	85.3	85.6	85.0	85.9	85.8	85.8	83.5	85.3	84.2
Mill air flow rate; acfm		406	404	402	403	391	393	355	406	403	402	404	403
Mill outlet temperature; °F		171	163	159	161	163	165	170	179	172	175	184	185
Filter type		Ceramic	Ceramic	Ceramic	Ceramic	Fabric	Fabric	Fabric	Fabric	Fabric	Ceramic	Ceramic	Ceramic
Pressure drop; in W.C.		11.5	10.9	10.9	11.4	2.6	2	1.8	2.9	1.8	9.5	10.07	10.6
Filter inlet; ° F		390	387	380	382	380	378	370	394	366	397	370	364
Filter outlet; ° F		359	352	338	344	324	319	311	344	308	356	333.8	327
EMISSIONS													
O2 %		3.8	3.8	3.6	3.4	3.9	4.8	3.3	5.0	3.7	6.5	3.3	3.5
CO ppm		72	62	74	92	104	95	59	97	60	78	59	109
CO2 %		15.4	15.5	15.5	15.6	15.2	14.7	15.8	14.4	15.4	12.4	14.7	14.4
SO2 ppm		417	447	421	407	391	359	484	331	405	335	N/A	N/A
NOx ppm		611	664	N.M.	585	622	598	640	628	670	431	N/A	N/A
MISCELLANEOUS DATA													
Maximum load (based on 14,700 lb steam/h); %		83.8	81.4	79.8	82.9	81.8	62.0	62.3	80.7	61.7	81.3	81.8	84.6
NOTES													
Flue gas temperature leaving air heater and b													
expansion joint between the air heater and ou													
system. First noticeable leak was observed st													
3/15/00, and expansion joint was replaced be													

Summary Sheet

DMC Testing

		8/28/02	8/30/02	9/5/02	9/26/02	10/7/02	10/14/02	5/15/03	6/17/03
TEST/DESCRIPTION:									
Filtering Device:		Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5	Sleeve 3.5
Burner Settings:		Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2	Center 2
		Inner 6.5	Inner 6.5	Inner 6.5	Inner 6.5	Inner 6.5	Inner 6.5	Inner 6.5	Inner 6.5
		Outer 7.5	Outer 7.5	Outer 7.5	Outer 7.5	Outer 7.5	Outer 7.5	Outer 7.5	Outer 7.5
		Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6
		Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6	Coal Gun -6
WATER/STEAM SIDE									
Steam flow rate; lb/h		12,721	12,663	12,644	11,059	11,962	11,757	12,160	12,338
Water temperature into boiler; °F		230	232	231	235	235	232	232	222
Drum pressure; psig		225	223	221	223	231	229	228	223
Calorimeter temperature; °F		317	319	314	318	319	318	319	319
Steam temperature; °F		395.8	395	395	399	397	396	397	395
Steam quality; %		100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0
Blowdown rate; lb/h		3,279	3,271	3,262	3,388	3,334	3,317	3,309	3,274
AIR,FUEL, FLUE GAS SIDE									
Coal Flow rate; lb/h		1,164	1,164	1,164	1,164	1,152	1,152	1,152	1,152
Coal flow rate; MMBtu/h		15.5	15.7	15.2	15.2	15.4	15.4	15.2	15.0
Boiler outlet temperature; °F		606	620	608	584	613	616	630	618
Gas temperature leaving air heater; °F		395	404	412	378	404	401	419	409
Air temperature entering air heater; °F		177	166	182	157	169	150	177	168
Air temperature leaving air heater; °F		418	436	439	407	429	426	447	432
Air temperature into boiler; °F		376	396	402	366	391	388	406	391
Ash content of particulate; %		64.01	62.70	73.30	60.21	68.20	65.40	69.50	70.90
Coal combustion efficiency; %		95.2	95.7	96.8	94.3	95.9	94.7	96.6	96.5
Combustion air flow; acfm		3,443	3,589	3,250	3,273	3,227	3,731	3,139	3,240
Boiler draft; inches W.C.		-0.09	-0.09	-0.09	-0.09	-0.09	-0.09	-0.07	-0.09
Boiler efficiency; %		83.0	83.0	84.8	83.2	83.9	82.9	83.4	84.0
Mill air flow rate; acfm		399	400	406	397	404	402	401	406
Mill outlet temperature; °F		188	180.2	182.8	164	174	163	172	189
Filter type		Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic	Ceramic
Pressure drop; in W.C.		10.3	9.4	9.8	9.8	9.8	9.7	10.8	11.2
Filter inlet; ° F		389	388	392	357	394	N/A	396	393
Filter outlet; ° F		357	355	362	315	348	341	364	360
EMISSIONS									
O2 %		4.3	4.7	3.3	4.0	3.5	3.7	3.3	3.5
CO ppm		37	53	27	143	50	47	70	47
CO2 %		14.1	13.4	15.2	14.4	14.5	14.7	14.7	14.4
SO2 ppm		431	384	484	436	416	470	465	476
NOx ppm		459	444	514	358	523	483	581	527
MISCELLANEOUS DATA									
Maximum load									
(based on 14,700 lb steam/h); %		86.5	86.1	86.0	75.2	81.4	80.0	82.7	83.9
NOTES									
Flue gas temperature leaving air heater and boiler expansion joint between the air heater and oil system. First noticeable leak was observed at 3/15/00, and expansion joint was replaced by									

Appendix 3B

Ceramic Filter Chamber Pressure Drop as a Function of Time Firing Coal

Figure 3B-1 CERAMIC FILTER PRESSURE DROP AS A FUNCTION OF TIME FIRING COAL FOR RUN HOURS 0 TO 50

Figure 3B-2 CERAMIC FILTER PRESSURE DROP AS A FUNCTION OF TIME FIRING COAL FOR RUN HOURS 50 TO 100

Figure 3B-3 CERAMIC FILTER PRESSURE DROP AS A FUNCTION OF TIME FIRING COAL FOR RUN HOURS 100 TO 150

Figure 3B-4 CERAMIC FILTER PRESSURE DROP AS A FUNCTION OF TIME FIRING COAL FOR RUN HOURS 150 TO 200

Figure 3B-5 CERAMIC FILTER PRESSURE DROP AS A FUNCTION OF TIME FIRING COAL FOR RUN HOURS 200 TO 250

Figure 3B-6 CERAMIC FILTER PRESSURE DROP AS A FUNCTION OF TIME FIRING COAL FOR RUN HOURS 250 TO 300

Figure 3B-7 CERAMIC FILTER PRESSURE DROP AS A FUNCTION OF TIME FIRING COAL FOR RUN HOURS 300 TO 350

Figure 3B-8 CERAMIC FILTER PRESSURE DROP AS A FUNCTION OF TIME FIRING COAL FOR RUN HOURS 350 TO 400

Figure 3B-9 CERAMIC FILTER PRESSURE DROP AS A FUNCTION OF TIME FIRING COAL FOR RUN HOURS 400 TO 450

Figure 3B-11 CERAMIC FILTER PRESSURE DROP AS A FUNCTION OF TIME FIRING COAL FOR RUN HOURS 500 TO 550

Figure 3B-12 CERAMIC FILTER PRESSURE DROP AS A FUNCTION OF TIME FIRING COAL FOR RUN HOURS 550 TO 600

Figure 3B-13 CERAMIC FILTER PRESSURE DROP AS A FUNCTION OF TIME FIRING COAL FOR RUN HOURS 600 TO 650

Figure 3B-14 CERAMIC FILTER PRESSURE DROP AS A FUNCTION OF TIME FIRING COAL FOR RUN HOURS 650 TO 700

Figure 3B-15 CERAMIC FILTER PRESSURE DROP AS A FUNCTION OF TIME FIRING COAL FOR RUN HOURS 700 TO 750

Figure 3B-17 CERAMIC FILTER PRESSURE DROP AS A FUNCTION OF TIME FIRING COAL FOR RUN HOURS 800 TO 850

Figure 3B-19 CERAMIC FILTER PRESSURE DROP AS A FUNCTION OF TIME FIRING COAL FOR RUN HOURS 900 TO 950

Appendix 4A

Market Penetration Model Source Code Listing

\$TITLE OPTIMAL ALLOCATION OF BOILER RETROFITS

\$OFFUPPER

* This model determines the optimal allocation of retrofitted boilers
 * in the Indiana and Cambria county region among the installed base
 * of water tube boilers. The approach is to use a transportation model
 * that considers the costs to produce the boiler retrofit fuel, the costs
 * to transport MCWM to the retrofitted boilers, and the capital and
 * O&M costs of retrofitting. The transportation model framework minimizes
 * the costs to boiler owners given the location of each boiler, and its
 * individual characteristics, and the location of each possible source of
 * MCWM. The decision to retrofit is assumed to be made if the total costs
 * of retrofitting are less than the costs to continue firing oil or gas.

*

* AUTHOR: A. Michael Schaal

* DATE: March 3, 1995

* Revised: April 17, 1995 - Updated MCWM and retrofit cost curve

*

SETS

i fuel supply points / GAS, COAL1 * COAL8 /
 j boilers / BLR1 * BLR18 /
 k mine characteristics / ASH, SULFUR, OUTPUT /
 l boiler locations / 15701, 15901, 16330, 15931 /

LABELS for boiler plant data

/ BSIZE boiler size in MMBtu per hour
 UTIL average annual boiler utilization in %
 PGAS boiler's price of gas
 BLOC boiler location's zip code
 TCR total capital requirement to retrofit boiler /

;

***** PARAMETER DECLARATION *****

SCALARS

MFACT	% additional mine output due to MCWM	/	100.0	/
CMOIST	coal moisture content in %	/	0.0	/
PCOAL	price of chunk coal f.o.b. mine	/	0.0	/
CHHV	coal higher heating value in Btu per lb	/	12810	/
CWMHV	MCWM higher heating value in Btu per lb	/	14000	/
STHV	heat content of steam in Btu per lb	/	1190	/
ASHCST	coal ash disposal cost in \$ per ton	/	5.60	/
FXUNLD	fixed cost to unload MCWM in \$ per MMBtu	/	0.168	/
VARUNLD	transport cost in \$ per MMBtu per mile	/	0.00373	/
ASHPER	dry basis weight percent ash in MCWM	/	5.00	/
PIFACT	annual P+I on \$1000 at 4% and 20 years	/	73.582	/
FXOMFACT	fixed O&M maint. cost as a % of capital	/	2.0	/

;

PARAMETERS

ASHBCST ash disposal cost in \$ per MMBtu of coal burned


```

MINE(i)          annual mine coal output in tons
FS(i)            annual fuel available from each site in MMBtu
DIST(i,l)        distance from ith fuel source to lth town in miles
BDIST(i,j)       distance from ith fuel source to jth boiler in miles
BOILER(j,LABELS) boiler plant data
BQ(j)            annual boiler demand for fuel in MMBtu
ATCR(j)          annualized TCR to retrofit jth boiler
FUELC(i,j)       non-transport costs less production costs in $ per MMBtu
PCOALM           price of chunk coal in $ per MMBtu
TCOST(i,j)       MCWM transport cost in $ per MMBtu
;

```

```

***** PARAMETER ASSIGNMENT *****

```

```

PARAMETER MINE(i)
*          tpy
  / COAL1 1686875
    COAL2 878230
    COAL3 870403
    COAL4 258030
    COAL5 188760
    COAL6 865017
    COAL7 485980
    COAL8 282732 /

```

```

TABLE DIST(i,l)
          15701      15901      16330      15931
GAS              0          0          0          0
COAL1             30         23          9          3
COAL2             22         35         32         20
COAL3             20         40         33         25
COAL4             40         44         32         22
COAL5             45         15         35         29
COAL6             30         42         57         51
COAL7             14         44         43         36
COAL8             23         19         33         26

```

```

TABLE BOILER(j,LABELS)
*          BSIZE      UTIL      PGAS      BLOC      TCR
          MMBtu/hr      %      $/MMBtu      zip      $M
BLR1         68.0       60       3.62     15701     5.01
BLR2         50.4       53       3.62     15901     4.22
BLR3         50.4       53       3.62     15901     4.22
BLR4         50.4       53       3.62     15901     4.22
BLR5         50.4       53       3.62     15901     4.22
BLR6         68.0       60       3.62     15701     5.01
BLR7         19.4       50       3.62     15901     2.43
BLR8         19.4       50       3.62     15901     2.43
BLR9         11.6        5       4.50     15901     1.81
BLR10        11.6        5       4.50     15901     1.81
BLR11        27.7       30       3.62     16330     2.99
BLR12        27.7       30       3.62     16330     2.99
BLR13        27.7       30       3.62     16330     2.99
BLR14        11.6       30       3.62     15901     1.81
BLR15        11.6       30       3.62     15901     1.81
BLR16        11.6       30       3.62     15901     1.81
BLR17        53.2        1       3.62     15901     4.35
BLR18        94.5        5       3.62     15931     6.06
;

```

```

BDIST(i,j) $ (BOILER(j,"BLOC") EQ 15701) = DIST(i,"15701") ;
BDIST(i,j) $ (BOILER(j,"BLOC") EQ 15901) = DIST(i,"15901") ;

```

```

BDIST(i,j) $ (BOILER(j,"BLOC") EQ 16330) = DIST(i,"16330") ;
BDIST(i,j) $ (BOILER(j,"BLOC") EQ 15931) = DIST(i,"15931") ;
ASHBCST    = (10000 * ASHPER * ASHCST) / (2000 * CWMHV) ;
FS(i)      = (2000 * (MFACT/100) * MINE(i) * CHHV * (1-CMOIST/100)) / 1e6 ;
FS('GAS')  = 1e9 ;
BQ(j)      = BOILER(j,"BSIZE") * (BOILER(j,"UTIL")/100) * 24*365.25 ;
ATCR(j)     = PIFACT * BOILER(j,"TCR") * 1000 ;
FUELC(i,j) = ASHBCST + FXUNLD +
              ( ATCR(j) + (1e6*BOILER(j,"TCR")*FXOMFACT/100) )/BQ(j) ;
FUELC('GAS',j) = BOILER(j,'PGAS') ;
PCOALM      = 1e6*(PCOAL/(2000*CHHV)) ;
TCOST(i,j)  = VARUNLD * BDIST(i,j) ;

OPTION BQ:0:0:1 ;
OPTION FS:0:0:1 ;
OPTION ATCR:4:0:1 ;
DISPLAY PCOALM, FS, BOILER, BDIST, TCOST, BQ, ASHBCST, ATCR, FUELC ;

```

```

***** VARIABLE DECLARATION *****

```

VARIABLES

```

PMCWM(i)          MCWM production cost at supply point i
QFUEL(i)          fuel demanded of zero marginal cost coal
FUELCOST(i,j)     fuel costs less transport costs
X(i,j)            annual flow of fuel from supply point i to boiler j
Z                 total annual boiler costs
;

```

```

POSITIVE VARIABLES X, QFUEL, FUELCOST, PMCWM
;

```

```

***** VARIABLE INITIALIZATION *****

```

```

PMCWM.L('GAS')    = 0.0 ;
X.L(i,j)          = 0.0 ;
X.L('COAL8',j)    = BQ(j) ;
X.L('GAS',j)      = BQ(j) ;

```

```

***** EQUATION DECLARATION *****

```

EQUATIONS

```

COST              define objective function
FUELSPLY(i)       determine fuel demand for each supplier
SUPPLY(i)         observe supply limits
DEMAND(j)         satisfy boiler fuel demands
SPLYCST(i)        total fuel preparation cost at each mine
FUEL(i,j)         fuel cost less transportation
FUELG(i,j)        gas price
COALFX            coal mine fixed constraint
;

```

```

##### EQUATION DEFINITION #####

COST ..          Z =E= SUM((i,j), FUELCOST(i,j)*X(i,j) + TCOST(i,j)*X(i,j) ) ;

FUELSPLY(i) ..    QFUEL(i) =E= SUM(j, X(i,j)) ;
SUPPLY(i) ..      QFUEL(i) =L= FS(i) ;
DEMAND(j) ..      SUM(i, X(i,j)) =E= BQ(j) ;
SPLYCST(i)$MINE(i) .. PMCWM(i)*( (QFUEL(i)+1)**(0.969) ) -
                   0.6727*( (QFUEL(i)+1)**(0.969) ) =E= 211561 ;
FUEL(i,j)$MINE(i) .. FUELCOST(i,j) =E= PMCWM(i) + FUELFC(i,j) ;
FUELG('GAS',j) ..  FUELCOST('GAS',j) =E= FUELFC('GAS',j) ;
COALFX ..          SUM( i, QFUEL(i)$MINE(i) ) - QFUEL('COAL8') =E= 0

MODEL RETROFIT /ALL/ ;

OPTION NLP = MINOS5 ;
OPTION ITERLIM = 4000 ;

SOLVE RETROFIT USING NLP MINIMIZING Z ;
SOLVE RETROFIT USING NLP MINIMIZING Z ;

DISPLAY X.L, PMCWM.L, QFUEL

```

Appendix 4B

MCWM Supply Cost Model

Introduction

This appendix describes basic economics for the supply of coal-based fuels for an industrial boiler that has been converted from firing fuel oil to firing a premium coal fuel. The boiler conversion technologies were studied by Penn State for DOE and DOD. The focus of Phase I activities were the development, design, and evaluation of two candidate technologies for boiler conversion. The boiler that was considered for conversion was a 25.2 MM Btu/h unit located at the Naval Surface Warfare Center at Crane, Indiana. The conversion technologies considered were to fire either, a micronized coal water mixture (MCWM) or a dry, micronized coal (DMC).

MCWM and DMC Conversion Technology Overview

The MCWM and DMC technologies utilize the particular form of the feed coal to minimize solids handling problems and minimize boiler derating. Both technologies require premium quality coal fuels, one difference between the two being the physical form of the fuel as delivered to the boiler. DMC is delivered to the boiler as a chunk coal (nominal 2" x 0) and reduced to the necessary micron size at the boiler site. MCWM, on the other hand, is transported, stored, handled, and fired in the boiler in slurry form. Both technologies use coal with the same basic qualities as listed in Table 4B-1.

The first three coal characteristics were determined by Penn State as minimums acceptable for either technology. The coal higher heating value (HHV) is an estimate for a coal of this quality. The estimated HHV is probably on the conservative side considering other DOE researchers use a value closer to 15,000 Btu/lb coal.

Table 4B-1 MCWM and DMC Coal Quality Specifications

Ash Content	≤ 5	wt.% (dry basis, d.b.)
Sulfur Content	≤ 1	wt. % (d.b.)
Volatile Matter	> 30	wt. % (d.b.)
Higher Heating Value	≈ 14,000 Btu/lb coal	(d.b.)

Sufficient reserves of acceptable coal sources have been identified by Penn State to exist in the immediate area of Crane (within a radius of 40 miles, or closer) for either technology.

Crane Boiler Characteristics

The Crane boiler that was evaluated for retrofit to fire coal is one of three, roughly equivalent, boilers located within the same boiler house. These boilers are presently used

for building heat, with occasional demands for process steam. The Phase I case study considered only six month operation of the retrofitted boiler corresponding to the heating season. The basic boiler retrofit requirements are listed in Table 4B-2.

Table 4B-2 Crane Boiler Specifications

Boiler Rated Capacity	25.2	MM Btu/h
Annual Hours of Operation	4,383	Hours
Boiler Load During Operation	100	%
Coal Requirement (Operating)	0.9	tons per hour (tph) coal (d.b.)
Annual Coal Requirement	3,945	tons per year (tpy) coal (d.b.)

The annual operating hours assumes that the retrofitted boiler is used during a typical six-month heating season at full rated capacity. Additional steam requirements for DOD users are met by adding boilers to the system. This arrangement of smaller, modular, boilers follows a normal DOD facilities planning philosophy that emphasizes maximization of system reliability under adverse conditions and standardized designs. Industrial- and commercial-scale boilers tend to be larger in order to minimize per unit cost of steam.

MCWM Supply

The MCWM fuel form, as a slurry, has some potential advantages as compared to the DMC fuel. First, since the fuel is handled as a slurry it can be easily pumped into the system, stored, and sent to the burner for atomization into the boiler. Also, the slurried fines pose little problem in the generation of airborne dust (as long as splashing and spillage is avoided). Supply of the slurry may, potentially, come from a variety of relatively low cost sources as compared to other coal-forms. Since the slurry is delivered to the boiler at the proper size, the relatively capital and O&M cost intensive grinding mills for each boiler that is converted to fire the coal fuel are not necessary. Finally, cost advantages could be derived from the installation of a single slurry preparation system to supply several converted boilers.

One disadvantage of MCWM is that boiler derating tends to be more severe with slurry fuels, *ceteris paribus*, due to; 1) fuel atomization forming droplets consisting of several coal particles requiring longer residence times for burnout, 2) ash deposition and erosion on boiler tubes, and 3) heat losses (latent and sensible) associated with the higher moisture content of the fuel. Storage of the slurry can also be problematic due to a tendency of the solids in the slurry to settle and form a 'hard' or 'soft' pack. This settlement problem requires the development of specialized slurry preparation techniques and chemical additive packages. Minimizing boiler derating and improving the handleability and storage of the slurry were goals of the technology development effort by Penn State.

A basic difference governing the economics of fuel supply between MCWM and DMC technologies is that the form of the fuel delivered to the boiler is not readily

available from an existing market source. Coal is not generally produced for delivery to customers in slurry form. In fact, state-of-the-art production of a MCWM with the requisite handleability, storage, and atomizing qualities is limited in terms of total quantity produced annually and in the number of potential suppliers. In addition, there can be substantial variation in the quality of slurry produced by the different suppliers. For that reason, the MCWM retrofit option must consider the costs of production and delivery of slurry to the Crane boiler.

MCWM Factored Cost Estimate

A factored cost estimate for the production and transportation of MCWM to the Crane boiler was developed to aid in comparing this fuel to DMC. At the present time, no specific coal source has been identified. The production and transportation analysis presented here is loosely based on using AMAX Coal Company's Minnehaha operation (located about 40 miles from the Crane boiler) to site the MCWM preparation plant.

Creation of a market for MCWM potentially addresses a problem faced by nearly every coal preparation plant operation, namely the disposition of coal fines. Coal fines are difficult to clean and dry to a level of acceptable moisture content. Even when recovered and dried to acceptable levels, conventional handling methods tend to cause fines to become easily airborne, representing both a loss of product and the creation of a potential health hazard. Both the above problems can be addressed by producing and transporting the fines in a slurried fuel. Bearing this in mind, the feed to a MCWM production process can come from a variety of sources including; 1) existing coal impoundment, 2) existing operational coal preparation plant, and 3) conventionally produced chunk coal.

Considerable uncertainty surrounds the recovery of coal from existing coal impoundments due to the difficulties of locating appropriate impoundments, the adequate characterization of the distribution of coal (and potential contaminants), and the higher expected costs associated with reclamation/dredging of the coal. An impoundment must be able to provide an inexpensive product possessing the rather stringent quality specifications. Mixtures of the desired coal with other coals and contaminants in the impoundment complicates the beneficiation and slurry production process.

The surest method of providing the MCWM fuel is from coal obtained from the market in chunk form that already possesses the desired quality specifications. The size reduction and slurring required can then be carried out in a straight-forward manner. The disadvantages of this method are the expected higher cost of coal, as compared to the alternatives, and the extra equipment and effort required to reduce the +2" x 0 coal the size required for MCWM.

A third option exists for producing MCWMs from slurry refuse streams in an existing coal preparation plant. Here, the characteristics of individual refuse streams in the conventional preparation plant can be utilized with little variation expected in the stream compositions over time (when cleaning the same coal). Since coal is recovered from process streams that would normally be discarded, the marginal cost of feed coal and refuse disposal for the slurry production plant would be zero. In fact, an economic benefit can be derived for reducing quantities of solids requiring disposal.

To date a particular site has not been selected, hence characterizing a preparation plant sufficiently to carry out a detailed design of the slurry preparation plant is not

possible. The MCWM production plant speculated here, however, makes use of this last scenario by providing what may be considered typical slurry preparation plant equipment. A MCWM production plant for the Crane boiler may be sized as described in Table 4B-3.

Table 4B-3 MCWM Production Plant Characteristics

MCWM Plant Production Rate	2	tph (d.b.)
Slurry Plant Annual Operating Hours	1,972	
Annual Production (Crane)	3,945	tons coal (d.b.)
Annual Maximum Rated Capacity	14,026	tons slurry

As can be seen from Table 4B-3, the slurry plant is oversized relative to the boiler requirement. The MCWM sizing reflects, 1) difficulties associated with the problems in differentiating between processes of a capacity in the range of one to three tph, 2) a built-in margin for both the large uncertainty in the capital equipment requirement, and 3) a production rate sufficient to provide for preparation plant down time while the boiler operates 24 hours per day, seven days a week. The MCWM production plant maximum annual capacity assumes a plant availability of 80 percent and the existence of a market for all of the production.

A simple factored capital cost estimate is provided in \$90.6/ton coal (d.b.) or \$3.23/MM Btu. This cost includes all capital, operation and maintenance charges.

Table 4B-4. The equipment list provides allowances for a single mill, conventional flotation cells, spirals (or the equivalent in other cleaning equipment), pumps, and a 25,000 gallon slurry storage tank sized for a three day supply of Crane, with an additional 10% allowance for miscellaneous equipment items. The electrical load of the equipment was estimated at 94 horsepower. The estimated equipment costs and electrical loads were generated after reviewing several sources and applying standard cost estimation methodologies to account for inflation, at 4%, and scale. The slurry plant described here purposely avoids the use of advanced fine coal cleaning methods reflecting an emphasis upon locating refuse streams that require minimum efforts to recover coal fines of the required quality.

The total capital requirement is estimated to be \$993 thousand dollars. As can be seen from the equipment list, by far the largest capital requirement is represented by the mill. This item also represents the greatest proportion of the estimated horsepower due to the demanding grinding regime expected. The equipment list, and associated costs, is considered speculative due to the incomplete knowledge regarding potential feed stream characteristics.

The plant cost factor used for determining the installed plant cost was generated using the Chilton method. The selection of this method was due to ease of use and consideration of the fact that the MCWM plant is an addition to an existing coal preparation facility. That advantage, along with the simple to implement approach, is somewhat offset by the relatively small size of the plant and the speculative nature of

estimate. The factor derived and presented above of 3.39 is less than the simpler Lang factor of 3.63 for solid-fluid plants.

Annualized capital and operation & maintenance (O&M) costs are presented in the Table 4B-5. The MCWM cost, f.o.b. slurry preparation plant, is estimated to be \$90.6/ton coal (d.b.) or \$3.23/MM Btu. This cost includes all capital, operation and maintenance charges.

Table 4B-4 MCWM Factored Capital Costs (\$K '93)

Mill	176
Flotation Cells	30
Spirals	14
Pumps	13
Tankage	33
Miscellaneous Items (@ 10%)	27
Equipment Cost (Delivered)	<u>\$293</u>
Cost Factor (Chilton Method)	3.39
Total Installed Plant Cost	<u><u>\$993K</u></u>

Table 4B-5 MCWM Annualized Capital and O&M Costs (\$ '93)

	Annual Cost \$	\$/ton Clean Coal (d.b.)	\$/MM Btu
FIXED O&M COSTS:			
O&M Labor	39,447	10.00	0.36
Maintenance Materials & Supplies	39,708	10.07	0.36
VARIABLE O&M COSTS:			
Feed Coal	-	-	-
Refuse Disposal	-	-	-
Grinding Media	1,341	0.34	0.01
Flotation Reagents Allowance	2,209	0.56	0.02
Additive Package	30,453	7.72	0.28
Electrical Power	5,509	1.40	0.05
Capital Charges:	238,606	60.49	2.16
TOTAL O&M AND CAPITAL CHARGES	357,273	90.57	3.23

O&M labor is assumed to be marginal to the existing coal preparation plant operation. An allowance of one person at \$20/h is provided. Maintenance materials and supplies are estimated at four percent of total plant cost. Both feed and refuse disposal costs are considered as marginal to the existing preparation plant operation and therefore equal to zero (no benefit assumed to be derived from avoided disposal to an impoundment pond). Consumption of grinding media is estimated to be 0.68 lb media per ton coal (d.b.) at \$0.50/lb. An allowance is made for flotation reagents at \$0.16/ton flocculant and 2 lb fuel oil per ton of coal at \$0.20/lb. Electrical power consumption for the state of Indiana is taken at 4 cents/KWh. Capital charges are calculated at a 15% cost of capital with a payback period of 7 years. At this level of the analysis, the use of government funds to build the MCWM plant (and a 4% discount rate) is not assumed in order to provide an additional margin, and to allow for sizing the plant for multiples of Crane (as discussed below).

The additive package costs primarily consists of a dispersant to avoid the formation of agglomerates that pose problems for storage and atomization of the slurry by the burner. The dispersant is estimated as 0.6 wt.% to coal at a cost of \$0.55/lb (active ingredient). A base is also used in conjunction with the dispersant in order to correct slurry pH. The costs for using the base are estimated as 0.4 wt.% to coal at a cost of \$0.14/lb. It is recognized that several additive packages exist which may provide acceptable results. Physical slurry preparation techniques may also substantially reduce the amount of dispersant required.

A sensitivity analysis of MCWM costs, f.o.b. slurry preparation plant, was conducted using a methodology similar to that accomplished for the Crane base case. It was shown that the \$/MM Btu cost of coal decreases substantially with an increasing number of oil-designed boilers converted over to fire MCWM. At a coal requirement equivalent to 10 boilers the size of Crane, the MCWM cost is \$1.03/MM Btu and drops to \$0.67/MM Btu at 50 Crane boiler equivalents. The analysis presented here does not imply that this many boiler conversions are undertaken. Indeed, only a few industrial or commercial-sized boilers would need to be converted to create an equivalent level of demand.

It should be noted that throughout the analysis of MCWM costs the capital charges are not amortized over a significant fraction of maximum rated slurry plant capacity. Since the demand is assumed to be only for a six month heating season, the slurry plant with all its capital equipment, sits idle for half of the year, regardless of the type of fuel supply. An interesting analysis might be to size the preparation plant for the anticipated annual demand and compare the benefit of a reduced plant capital cost to the added storage costs of slurry.

Another consideration of using MCWM as a boiler fuel is the relatively high costs of transport that range from 0.87 to 1.45 \$/MM Btu. A significant factor in the slurry transportation costs is the weight and bulk penalty of transporting up to 40 wt.% of water with coal. Any slurry fuel system that reduces weight being transported from the source to the point of use would reduce slurry transportation charges. Consideration might be given to transporting MCWM with the highest solids loading possible and then adding the necessary quantity of water at the boiler site. This fuel form also tends to higher charges due the necessity of using tanker versus dry bulk trucks. Extending the previous

concept, the MCWM might be transported in covered trucks as filter cake and then slurried at the boiler site.

Appendix 4C

MCWM Retrofit Technology Cost Model

Table 4C-1 MCWM Cost Estimate to Retrofit a 25.2 MM Btu/h Boiler

Ref. No	Item Description	Material Cost	Installation Hours
General Equipment			
116	FD FAN w/vfd	\$14,000	60
117	Air Preheater 'steam'	\$4,715	40
118	Air Heater (Q-Tube)	\$85,000	80
119	Air Heater Sootblower	\$2,700	24
120	Bag House	\$49,734	160
121	ID Fan w/vfd	\$18,750	80
122	Boiler Mods	\$0	0
123	Stack	\$45,000	100
124	Floor Air Blast System	\$4,200	160
125	Air Dryer	\$8,000	24
126	Oil/Water Separator Equip	\$4,000	24
127	Air Receiver Tanks	\$1,500	24
128	Pitot Grid-Comb. Air	\$3,000	16
129	Underground Building	\$20,000	1000
130	Compressor Building	\$4,000	160
131	Retaining Wall	\$4,500	430
132	Excavation	\$1,000	120
133	Concrete Slab	\$11,500	600
134	Air and FG Ductwork	\$28,500	560
135	A.H. Support Steel	\$4,500	40
136	Valves and Instrumentation	\$80,000	800
137	Controls	\$100,000	
138	Electrical Work	\$70,000	1200
139	Run-Off Drain	\$400	24
140	Concrete Pipeway	\$650	100
141	A.H. Sootblower Piping	\$2,200	220
	Subtotal		
Ash Handling Equipment			
142	Pneumatic Conveying System	\$7,390	120
143	Ash Silo	\$18,034	
144	Ash Silo Bin Vent Filter	\$3,508	40
145	Rotary Valve	\$4,382	24
146	Ash Unloading Screw Conveyor	\$16,614	120

Ref. No	Item Description	Material Cost	Installation Hours
Slurry Feed Equipment			
147	MCWSF Transfer Pump (2)	\$4,000	24
148	MCWSF Storage Tank	\$52,000	120
149	MCWSF Mixer	\$34,000	
150	Strainer (3 Sets) @ 1500 each	\$4,500	32
151	MCWSF Injection Pump (2)	\$13,400	32
152	MCWSF Heater	\$16,000	60
153	Flush Water Booster Pump	\$1,200	24
154	Waste Processing Tank	\$5,200	24
155	MCWSF Burner	\$66,000	80
156	Air Compressor	\$33,000	80
157	Slurry Building	\$7,500	400
158	Slurry & Flush Water Piping	\$26,000	600
159	Waste Processing Equipment	\$27,500	80
160	Power Station	\$21,500	64
		<u>\$929,577</u>	<u>7970</u>
Direct Construction Costs			
	Total Materials	\$929,577	
	Total Installation Hours Cost (@ \$40/hr)	\$318,800	
Construction Indirect Costs			
	Field Supervision (15% of Labor hours)	\$47,820	
	Construction Overhead and Fee (20%)	\$249,675	
Other Project Costs			
	Freight (6%)	\$55,775	
	Taxes, Permits (7%)	\$65,070	
	Subtotal Subcontractor and Equipment	\$1,666,717	
	Estimating Accuracy Allowance (10%)	\$166,672	
	Total Subcontractor and Equipment	<u>\$1,833,389</u>	
	Engineering (15% of project)	\$376,724	
	Project Management (4% of project)	\$100,460	
	Construction Management (6% of project)	\$150,690	
	Start up (2% of project)	\$50,230	
	Subtotal Project	<u>\$2,511,492</u>	
	Contingencies (about 15%)	\$374,212	
	Total Project Cost w/contingency	<u>\$2,885,704</u>	

Appendix 4D

ELECTRIC POWER IN YOUR COMMUNITY

In this survey we are interested in how you feel an electric power plant could affect you and your community. Because we are asking how you feel, there are no right or wrong answers. You do not need to know anything about electricity or how it is produced to answer the questions in this survey. Just answer the questions according to what you know, your own personal experiences, and attitudes. Please answer every question.

Throughout this survey, think of a risk as something that may or may not happen and an impact as the result that actually happens. An impact can be either good or bad.

- Q1.** Below is a list of technologies and types of facilities common in today's society. With a 1 meaning "not at all risky" and a 5 meaning "very risky," indicate how risky you feel each item is to you and your family.
(Circle appropriate number for each item.)

	NOT AT ALL RISKY		SOMEWHAT RISKY		VERY RISKY
PESTICIDES	1	2	3	4	5
X-RAYS	1	2	3	4	5
NUCLEAR POWER PLANT	1	2	3	4	5
ASBESTOS	1	2	3	4	5
TRAFFIC	1	2	3	4	5
QUARRY OR MINE	1	2	3	4	5
SEWAGE TREATMENT PLANT	1	2	3	4	5
NATURAL GAS-FIRED POWER PLANT	1	2	3	4	5
MANUFACTURING PLANT	1	2	3	4	5
LANDFILL	1	2	3	4	5
COAL-FIRED POWER PLANT	1	2	3	4	5
GROUNDWATER CONTAMINATION	1	2	3	4	5

Many human activities, including the production of electricity, have both good and bad effects on your life and on the community in which you live. We want to learn how individuals like you feel about power plants that burn either coal or natural gas to produce electricity and how you feel electricity production could affect the quality of your life.

Q2. Are you aware of any facilities near your home or in your community that produce electricity?
(Circle appropriate number.)

1. NO
2. YES (GO TO Q2A)
3. NOT SURE

Q2A. Approximately how far away from your home is the nearest power plant?

_____ MILE(S)

Q3. Do you know what fuel is burned to produce the electricity that supplies your community with its electricity needs? (Circle appropriate number.)

1. NO
2. YES - PLEASE SPECIFY: _____
3. NOT SURE

Q4. Do you feel that the amount of electricity that your household consumes is less than, about equal to, or greater than the amount of electricity consumed by other households in your community?
(Circle appropriate number.)

1. LESS THAN AVERAGE
2. ABOUT EQUAL TO THE AVERAGE
3. GREATER THAN AVERAGE

Q5. What is the average amount that you pay for your electricity each month?
If you are not the person in your household who pays the electricity bill, please indicate about how much you think your electricity bill is each month.
(Write in dollar amount.)

\$ _____

A Coal-Fired Power Plant in Your Community

As the population in the United States grows, more electricity will be needed for the extra homes, businesses, and industries. We are interested in how you feel you will be affected if a coal-burning power plant were to be built and operate near your home. We would like you to consider a hypothetical situation in which a coal-burning power plant has been proposed for your community.

In the following pages, you will be asked how you feel the proposed coal-fired power plant will create risks to your health, the environment, and the aesthetic quality of your community. In addition, we will ask how you feel the proposed power plant could affect your community's economy.

4.1 CONSIDER THE FOLLOWING PROPOSAL:

YOU AND YOUR COMMUNITY	<ul style="list-style-type: none"> You have lived in your community for 5 years and plan to live there permanently. You rely on electricity for all of the heating, lighting, and cooking in your home. In other words, your household does not have a woodstove, an oil furnace, or a hook-up to a natural gas pipeline.
NEED FOR ELECTRICITY	<ul style="list-style-type: none"> Because of the growth in the area in which you live, the public utility that supplies your community with its electricity has announced that the proposed power plant is greatly needed. The price you pay for your electricity will remain the same after the new coal power plant is in operation.
GENERAL FEATURES OF THE POWER PLANT	<ul style="list-style-type: none"> The coal power plant will produce 180 megawatts of electricity per year. This will be enough power to supply the growth in your area for the next 30 years. The power plant will be built on a 40-acre plot of land near a local river. The site for the proposed coal-fired power plant is 3 miles away from your home. The life span of the power plant is projected to be at least 30 years.
REGULATORY COMPLIANCE	<ul style="list-style-type: none"> All local, state, and federal regulations will be obeyed during the construction and operation of the plant.

ANSWER THE FOLLOWING QUESTIONS
AS IF YOU WERE TRULY FACING THIS SITUATION.

The Coal-Fired Power Plant and Your Health

Human health impacts are due to changes in the air you breath and the water you drink. To answer the following questions, consider how you feel the proposed coal-fired power plant could affect your health and the health of others that live in your community. Please think only of the potential health impacts as you answer the questions on this page.

Q6. Do you feel that there might be health impacts as a result of the coal-fired power plant described on **page 4D-3?** (Circle appropriate number.)

1. NO
2. YES

Q7. Indicate below how positive or negative you feel that the impact(s) to your health could be from the proposed coal-fired power plant. (Circle appropriate number.)

VERY NEGATIVE IMPACT				NO IMPACT		VERY POSITIVE IMPACT			
-4	-3	-2	-1	0	1	2	3	4	

Q8. How serious do you feel the risks to your health could be from the proposed coal-fired power plant? (Circle appropriate number.)

NOT AT ALL SERIOUS			SOMEWHAT SERIOUS			VERY SERIOUS	
1	2	3	4	5	6	7	

Q9. What do you feel is the chance or likelihood that the coal plant will affect your health some time in the future? (Circle percentage.)

NO CHANCE HEALTH WILL BE AFFECTED				MAY OR MAY NOT AFFECT HEALTH				DEFINITELY WILL AFFECT HEALTH		
0%	10%	20%	30%	40%	50%	60%	70%	80%	90%	100%

Q10. How would you describe your feelings when you think about the health risks from the proposed power plant? (Circle appropriate number.)

I FEEL AFRAID AND DREADFUL				MY FEELINGS ARE NEUTRAL			I FEEL HAPPINESS AND COMFORT		
-4	-3	-2	-1	0	1	2	3	4	

The Coal-Fired Power Plant and Your Local Economy

Economic impacts may be caused by changes in the local employment level, tax revenues and tax payments, property values, prices, and overall changes in the business activity in your community. To answer the following questions, consider how you feel the proposed coal-fired power plant could affect your local economy. Please think only of the potential impacts on your community's economy as you answer the questions on this page.

Q11. Do you feel that there might be economic impacts as a result of the coal-fired power plant described on **page 4D-3**? (Circle appropriate number.)

1. NO
2. YES

Q12. Indicate below how positive or negative you feel the impact(s) on your local economy could be from the proposed coal-fired power plant.
(Circle appropriate number.)

VERY NEGATIVE IMPACT	NO IMPACT					VERY POSITIVE IMPACT			
-4	-3	-2	-1	0	1	2	3	4	

Q13. How significant do you feel the impacts on your local economy could be from the proposed coal-fired power plant? (Circle appropriate number.)

NOT AT ALL SIGNIFICANT	SOMEWHAT SIGNIFICANT			VERY SIGNIFICANT		
1	2	3	4	5	6	7

Q14. What do you feel is the chance or likelihood that the coal plant will affect your local economy some time in the future? (Circle percentage.)

NO CHANCE ECONOMY WILL BE AFFECTED				MAY OR MAY NOT AFFECT ECONOMY				DEFINITELY WILL AFFECT ECONOMY			
0%	10%	20%	30%	40%	50%	60%	70%	80%	90%	100%	

Q15. How would you describe your feelings when you think about the economic impacts associated with the proposed coal plant?
(Circle appropriate number.)

I FEEL AFRAID AND DREADFUL	MY FEELINGS ARE NEUTRAL					I FEEL HAPPINESS AND COMFORT			
-4	-3	-2	-1	0	1	2	3	4	

The Coal-Fired Power Plant and Your Environment

Environmental impacts could be caused by changes in the air quality, water quality, plants and wildlife, and how the land in your community is used. To answer the following questions, consider how you feel the proposed power plant may be a risk to the environment, and how serious you feel the actual environmental impact(s) could be from the coal-fired power plant. Please think only of the potential impacts on environmental quality as you answer the questions on this page.

Q16. Do you feel that there might be environmental impacts as a result of the coal-fired power plant described on **page 4D-3**? (Circle appropriate number.)

1. NO
2. YES

Q17. Indicate below how positive or negative you feel that the impact(s) to the environment could be from the proposed coal-fired power plant.
(Circle appropriate number.)

VERY NEGATIVE IMPACT				NO IMPACT		VERY POSITIVE IMPACT			
-4	-3	-2	-1	0	1	2	3	4	

Q18. How serious do you feel the risks to the environment could be from the proposed coal-fired power plant? (Circle appropriate number.)

NOT AT ALL SERIOUS			SOMEWHAT SERIOUS			VERY SERIOUS	
1	2	3	4	5	6	7	

Q19. What do you feel is the chance or likelihood that the coal plant will affect the environment some time in the future? (Circle percentage.)

NO CHANCE THE ENVIRONMENT WILL BE AFFECTED				MAY OR MAY NOT AFFECT THE ENVIRONMENT				DEFINITELY WILL WILL AFFECT THE ENVIRONMENT		
0%	10%	20%	30%	40%	50%	60%	70%	80%	90%	100%

Q20. How would you describe your feelings when you think about the environmental risks from the proposed power plant? (Circle appropriate number.)

I FEEL AFRAID AND DREADFUL			MY FEELINGS ARE NEUTRAL				I FEEL HAPPINESS AND COMFORT		
-4	-3	-2	-1	0	1	2	3	4	

The Coal-Fired Power Plant and Aesthetic Quality

The aesthetic quality of your community is how pleasant your community is to live in. Many people think of the noise, sights, and smells around their community when they think of its aesthetic quality. To answer the following questions, consider how you feel the proposed coal-fired power plant could be a risk to the aesthetic quality of your community. Please think only of the potential impacts on the aesthetic quality as you answer the questions on this page.

Q21. Do you feel that there might be aesthetic impacts as a result of the coal-fired power plant described on **page 4D-3**? (Circle appropriate number.)

1. NO
2. YES

Q22. Indicate below how positive or negative you feel that the impact(s) on the aesthetic quality of your community could be from the proposed coal-fired power plant. (Circle appropriate number.)

VERY NEGATIVE IMPACT			NO IMPACT			VERY POSITIVE IMPACT		
-4	-3	-2	-1	0	1	2	3	4

Q23. How serious do you feel the aesthetic risks could be from the proposed coal-fired power plant? (Circle appropriate number.)

NOT AT ALL SERIOUS			SOMEWHAT SERIOUS			VERY SERIOUS		
1	2	3	4	5	6	7		

Q24. What do you feel is the chance or likelihood that the coal plant will affect your community's aesthetic quality some time in the future? (Circle percentage.)

NO CHANCE AESTHETIC QUALITY WLL BE AFFECTED				MAY OR MAY NOT AFFECT AESTHETIC QUALITY				DEFINITELY WILL AFFECT AESTHETIC QUALITY		
0%	10%	20%	30%	40%	50%	60%	70%	80%	90%	100%

Q25. How would you describe your feelings when you think about the aesthetic risks from the proposed power plant? (Circle appropriate number.)

I FEEL AFRAID AND DREADFUL			MY FEELINGS ARE NEUTRAL			I FEEL HAPPINESS AND COMFORT		
-4	-3	-2	-1	0	1	2	3	4

Consider the “overall concern” you have about the proposed coal-burning power plant. We are interested in how much of your concern is due to each of the four types of impacts covered in the last four pages.

Q26. In the spaces below, write in the percentage of the overall concern that you have for each type of potential impact from the proposed coal power plant. Please make sure your answers add up to 100%.

POTENTIAL HEALTH IMPACTS	_____ %
POTENTIAL ENVIRONMENTAL IMPACTS	_____ %
POTENTIAL ECONOMIC IMPACTS	_____ %
POTENTIAL AESTHETIC IMPACTS	_____ %
OTHER POTENTIAL IMPACTS	_____ %
(PLEASE DESCRIBE: _____)	
_____	TOTAL = 100%

Consider how you feel the proposed coal-burning power plant will affect your health, your local economy, the environment and the aesthetic quality of your community.

Q27. Would you vote in favor of or against the coal-burning power plant proposed for your community?

1. I WOULD VOTE IN FAVOR OF THE COAL POWER PLANT
2. I WOULD VOTE AGAINST THE COAL POWER PLANT
3. DON'T KNOW
4. I WOULDN'T VOTE - PLEASE EXPLAIN: _____

Suppose that your electricity supplier proposed to build the coal-fired power plant described on **page 4D-3**. We would like you to imagine that you live in the community in which the coal power plant is being proposed. If the supplier is not able to build this power plant, it will have to increase the price you pay for electricity because the much needed electricity must be purchased at a higher cost from other power producers elsewhere in the state.

Think of how much more money you would be willing to pay on your monthly electricity bill in order to prevent the coal-burning power plant from being built in your community. When answering the following question, please consider your household income, your average monthly electricity bill, and how much extra you can actually afford to pay to prevent the coal-fired power plant.

Q28. Circle the dollar value closest to the largest amount you would be willing to pay extra each month on your electricity bill to prevent the coal-fired power plant from being built in your community.

\$0.00	\$0.10	\$0.25	\$0.50	\$0.75	\$1.00
\$1.25	\$1.50	\$2.00	\$3.00	\$4.00	\$5.00
\$6.00	\$8.00	\$11.00	\$14.00	\$19.00	\$22.00
\$26.00	\$30.00	\$35.00	\$40.00	\$45.00	\$50.00
\$55.00	\$60.00	\$70.00	\$80.00	\$90.00	\$100.00
\$110.00	\$120.00	\$140.00	\$160.00	\$220.00	\$300.00

OTHER: \$ _____

If You Answered \$0.00 in Question 28, Go To Page 4D-11.

If You Answered \$0.00 in Question 28, Go To Page 4D-11.

- Q29.** Think of the amount you would pay to prevent the coal-burning power plant from being built in your community (**Q28.**) About what percent of that amount would be to ensure
(Write in percentage.)

THAT MY HOUSEHOLD WILL NOT BE AFFECTED
AT ALL BY THE POTENTIAL IMPACTS FROM THE
COAL -FIRED POWER PLANT. _____ %

THAT OTHER HOUSEHOLDS WILL NOT BE
AFFECTED AT ALL BY THE POTENTIAL IMPACTS
FROM THE COAL-FIRED POWER PLANT. _____ %

THAT FUTURE GENERATIONS WILL NOT BE
AFFECTED AT ALL BY THE POTENTIAL IMPACTS
FROM THE COAL-FIRED POWER PLANT. _____ %

THAT COAL-FIRED POWER PLANTS SHOULD BE
PREVENTED EVEN IF NO ONE WILL HAVE TO LIVE
NEAR THEM. _____ %

REASONS OTHER THAN THESE
(PLEASE DESCRIBE: _____
_____) _____ %

TOTAL = 100%

PLEASE MAKE SURE YOUR ANSWERS ADD UP TO 100%.

A Natural Gas-Fired Power Plant in Your Community

As the population in the United States grows, more electricity will be needed for the extra homes, businesses, and industries. We are interested in how you feel you will be affected if a natural gas-burning power plant were to be built and operate near your home. We would like you to consider a hypothetical situation in which a natural gas-burning power plant has been proposed for your community.

In the following pages, you will be asked how you feel the proposed natural gas-fired power plant will create risks to your health, the environment, and the aesthetic quality of your community. In addition, we will ask how you feel the proposed power plant could affect your community's economy.

CONSIDER THE FOLLOWING PROPOSAL:

YOU AND YOUR COMMUNITY	<ul style="list-style-type: none"> You have lived in your community for 5 years and plan to live there permanently. You rely on electricity for all of the heating, lighting, and cooking in your home. In other words, your household does not have a woodstove, an oil furnace, or a hook-up to a natural gas pipeline.
NEED FOR ELECTRICITY	<ul style="list-style-type: none"> Because of the growth in the area in which you live, the public utility that supplies your community with its electricity has announced that the proposed power plant is greatly needed. The price you pay for your electricity will remain the same after the new natural gas power plant is in operation.
GENERAL FEATURES OF THE POWER PLANT	<ul style="list-style-type: none"> The natural gas power plant will produce 180 megawatts of electricity per year. This will be enough power to supply the growth in your area for the next 30 years. The power plant will be built on a 40-acre plot of land near a local river. The site for the proposed natural gas-fired power plant is 3 miles away from your home. The life span of the plant is projected to be at least 30 years.
REGULATORY COMPLIANCE	<ul style="list-style-type: none"> All local, state, and federal regulations will be obeyed during the construction and operation of the plant.

ANSWER THE FOLLOWING QUESTIONS
AS IF YOU WERE TRULY FACING THIS SITUATION.

The Natural Gas-Fired Power Plant and Your Health

Human health impacts are due to changes in the air you breath and the water you drink. To answer the following questions, consider how you feel the proposed natural gas-fired power plant could have an affect on your health and the health of others that live in your community. Please think only of the potential health impacts as you answer the questions on this page.

Q30. Do you feel that there might be health impacts as a result of the natural gas -fired power plant described on **page 4D-11**? (Circle appropriate number.)

1. NO
2. YES

Q31. Indicate below how positive or negative you feel that the impact(s) to your health could be from the proposed natural gas-fired power plant.
(Circle appropriate number.)

VERY NEGATIVE IMPACT				NO IMPACT		VERY POSITIVE IMPACT		
-4	-3	-2	-1	0	1	2	3	4

Q32. How serious do you feel the risks to your health could be from the proposed natural gas-fired power plant? (Circle appropriate number.)

NOT AT ALL SERIOUS			SOMEWHAT SERIOUS			VERY SERIOUS	
1	2	3	4	5	6	7	

Q33. What do you feel is the chance or likelihood that the natural gas plant will affect your health some time in the future? (Circle percentage.)

NO CHANCE HEALTH WILL BE AFFECTED				MAY OR MAY NOT AFFECT HEALTH				DEFINITELY WILL AFFECT HEALTH		
0%	10%	20%	30%	40%	50%	60%	70%	80%	90%	100%

Q34. How would you describe your feelings when you think about the health risks from the proposed power plant? (Circle appropriate number.)

I FEEL AFRAID AND DREADFUL			MY FEELINGS ARE NEUTRAL				I FEEL HAPPINESS AND COMFORT		
-4	-3	-2	-1	0	1	2	3	4	

The Natural Gas-Fired Power Plant and Your Local Economy

Economic impacts may be caused by changes in the local employment level, tax revenues and tax payments, property values, prices, and overall changes in the business activity in your community. To answer the following questions, consider how you feel the proposed natural gas-fired power plant could affect your local economy. Please think only of the potential impacts on your community's economy as you answer the questions on this page.

Q35. Do you feel that there might be economic impacts as a result of the natural gas -fired power plant described on **page 4D-11**? (Circle appropriate number.)

1. NO
2. YES

Q36. Indicate below how positive or negative you feel the impact(s) on your local economy could be from the proposed natural gas-fired power plant.
(Circle appropriate number.)

VERY NEGATIVE IMPACT			NO IMPACT			VERY POSITIVE IMPACT		
-4	-3	-2	-1	0	1	2	3	4

Q37. How significant do you feel the impacts on your local economy could be from the proposed natural gas-fired power plant? (Circle appropriate number.)

NOT AT ALL SIGNIFICANT		SOMEWHAT SIGNIFICANT			VERY SIGNIFICANT	
1	2	3	4	5	6	7

Q38. What do you feel is the chance or likelihood that the natural gas plant will affect your local economy some time in the future? (Circle percentage.)

NO CHANCE ECONOMY WILL BE AFFECTED				MAY OR MAY NOT AFFECT ECONOMY				DEFINITELY WILL AFFECT ECONOMY		
0%	10%	20%	30%	40%	50%	60%	70%	80%	90%	100%

Q39. How would you describe your feelings when you think about the economic impacts associated with the proposed natural gas plant?
(Circle appropriate number.)

I FEEL AFRAID AND DREADFUL			MY FEELINGS ARE NEUTRAL			I FEEL HAPPINESS AND COMFORT		
-4	-3	-2	-1	0	1	2	3	4

The Natural Gas-Fired Power Plant and Your Environment

Environmental impacts could be caused by changes in the air quality, water quality, plants and wildlife, and how the land in your community is used. To answer the following questions, consider how you feel the proposed power plant may be a risk to the environment, and how serious you feel the actual environmental impact(s) could be from the natural gas-fired power plant. Please think only of the potential impacts on environmental quality as you answer the questions on this page.

Q40. Do you feel that there might be environmental impacts as a result of the natural gas -fired power plant described on **page 4D-11**? (Circle appropriate number.)

1. NO
2. YES

Q41. Indicate below how positive or negative you feel that the impact(s) to the environment could be from the proposed natural gas-fired power plant.
(Circle appropriate number.)

VERY NEGATIVE IMPACT			NO IMPACT			VERY POSITIVE IMPACT		
-4	-3	-2	-1	0	1	2	3	4

Q42. How serious do you feel the risks to the environment could be from the proposed natural gas-fired power plant? (Circle appropriate number.)

NOT AT ALL SERIOUS		SOMEWHAT SERIOUS			VERY SERIOUS	
1	2	3	4	5	6	7

Q43. What do you feel is the chance or likelihood that the natural gas plant will affect the environment some time in the future? (Circle percentage.)

NO CHANCE THE ENVIRONMENT WILL BE AFFECTED				MAY OR MAY NOT AFFECT THE ENVIRONMENT				DEFINITELY WILL AFFECT THE ENVIRONMENT		
0%	10%	20%	30%	40%	50%	60%	70%	80%	90%	100%

Q44. How would you describe your feelings when you think about the environmental risks from the proposed power plant? (Circle appropriate number.)

I FEEL AFRAID AND DREADFUL			MY FEELINGS ARE NEUTRAL			I FEEL HAPPINESS AND COMFORT		
-4	-3	-2	-1	0	1	2	3	4

The Natural Gas-Fired Power Plant and Aesthetic Quality

The aesthetic quality of your community is how pleasant your community is to live in. Many people think of the noise, sights, and smells around their community when they think of its aesthetic quality. To answer the following questions, consider how you feel the proposed natural gas-fired power plant could be a risk to the aesthetic quality of your community. Please think only of the potential impacts on the aesthetic quality as you answer the questions on this page.

Q45. Do you feel that there might be aesthetic impacts as a result of the natural gas -fired power plant described on **page 4D-11**? (Circle appropriate number.)

1. NO
2. YES

Q46. Indicate below how positive or negative you feel that the impact(s) on the aesthetic quality of your community could be from the proposed natural gas-fired power plant. (Circle appropriate number.)

VERY NEGATIVE IMPACT				NO IMPACT		VERY POSITIVE IMPACT			
-4	-3	-2	-1	0	1	2	3	4	

Q47. How serious do you feel the aesthetic risks could be from the proposed natural gas-fired power plant? (Circle appropriate number.)

NOT AT ALL SERIOUS			SOMEWHAT SERIOUS			VERY SERIOUS	
1	2	3	4	5	6	7	

Q48. What do you feel is the chance or likelihood that the natural gas plant will affect your community's aesthetic quality some time in the future? (Circle percentage.)

NO CHANCE AESTHETIC QUALITY WLL BE AFFECTED				MAY OR MAY NOT AFFECT AESTHETIC QUALITY				DEFINITELY WILL AFFECT AESTHETIC QUALITY		
0%	10%	20%	30%	40%	50%	60%	70%	80%	90%	100%

Q49. How would you describe your feelings when you think about the aesthetic risks from the proposed power plant? (Circle appropriate number.)

I FEEL AFRAID AND DREADFUL				MY FEELINGS ARE NEUTRAL				I FEEL HAPPINESS AND COMFORT	
-4	-3	-2	-1	0	1	2	3	4	

Consider the “overall concern” you have about the proposed natural gas-burning power plant. We are interested in how much of your concern is due to each of the four types of impacts covered in the last four pages.

Q50. In the spaces below, write in the percentage of the overall concern that you have for each type of potential impact from the proposed natural gas power plant. Please make sure your answers add up to 100%.

POTENTIAL HEALTH IMPACTS	_____ %
POTENTIAL ENVIRONMENTAL IMPACTS	_____ %
POTENTIAL ECONOMIC IMPACTS	_____ %
POTENTIAL AESTHETIC IMPACTS	_____ %
OTHER POTENTIAL IMPACTS	_____ %
(PLEASE DESCRIBE: _____)	
	TOTAL = 100%

Consider how you feel the proposed natural gas-burning power plant will affect your health, your local economy, the environment and the aesthetic quality of your community.

Q51. Would you vote in favor of or against the natural gas-burning power plant proposed for your community?

1. I WOULD VOTE IN FAVOR OF THE NATURAL GAS POWER PLANT
2. I WOULD VOTE AGAINST THE NATURAL GAS POWER PLANT
3. DON'T KNOW
4. I WOULDN'T VOTE - PLEASE EXPLAIN: _____

Suppose that your electricity supplier proposed to build the natural gas-fired power plant described on **page 4D-11**. We would like you to imagine that you live in the community in which the natural gas power plant is being proposed. If the supplier is not able to build this power plant, it will have to increase the price you pay for electricity because the much needed electricity must be purchased at a higher cost from other power producers elsewhere in the state.

Think of how much more money you would be willing to pay on your monthly electricity bill in order to prevent the natural gas-burning power plant from being built in your community. When answering the following question, please consider your household income, your average monthly electricity bill, and how much extra you can actually afford to pay to prevent the natural gas-fired power plant.

Q52. Circle the dollar value closest to the largest amount you would be willing to pay extra each month on your electricity bill to prevent the natural gas-fired power plant from being built in your community.

\$0.00	\$0.10	\$0.25	\$0.50	\$0.75	\$1.00
\$1.25	\$1.50	\$2.00	\$3.00	\$4.00	\$5.00
\$6.00	\$8.00	\$11.00	\$14.00	\$19.00	\$22.00
\$26.00	\$30.00	\$35.00	\$40.00	\$45.00	\$50.00
\$55.00	\$60.00	\$70.00	\$80.00	\$90.00	\$100.00
\$110.00	\$120.00	\$140.00	\$160.00	\$220.00	\$300.00

OTHER: \$_____

If You Answered \$0.00 in Question 52, Go To Page 19.

If You Answered \$0.00 in Question 52, Go To Page 19.

- Q53.** Think of the amount you would pay to prevent the natural gas-burning power plant from being built in your community (**Q52.**) About what percent of that amount would be to ensure Please make sure your answers add up to 100%.
(Write in percentage.)

THAT MY HOUSEHOLD WILL NOT BE AFFECTED
AT ALL BY THE POTENTIAL IMPACTS FROM THE
NATURAL GAS-FIRED POWER PLANT. _____ %

THAT OTHER HOUSEHOLDS WILL NOT BE
AFFECTED AT ALL BY THE POTENTIAL IMPACTS
FROM THE NATURAL GAS-FIRED POWER PLANT. _____ %

THAT FUTURE GENERATIONS WILL NOT BE
AFFECTED AT ALL BY THE POTENTIAL IMPACTS
FROM THE NATURAL GAS-FIRED POWER PLANT. _____ %

THAT NATURAL GAS-FIRED POWER PLANTS
SHOULD BE PREVENTED EVEN IF NO ONE WILL
HAVE TO LIVE NEAR THEM. _____ %

REASONS OTHER THAN THESE
(PLEASE DESCRIBE: _____
_____) _____ %

TOTAL = 100%

***Your Preferences for the Use of Either Coal or Natural Gas
in the Proposed Power Plant***

Suppose that your electricity supplier has already received local, state, and federal approval to build a power plant for your community. Assume that a power plant will be built in your community, but the type of fuel that it will burn has not yet been determined.

Q54. Based upon your current knowledge, feelings, and opinions, what type of fuel would you prefer that the plant burn to produce electricity - coal or natural gas?

1. COAL
2. NATURAL GAS

Think of how much more money you would be willing to pay on your monthly electricity bill to make sure that the power plant uses the fuel type that you circled above in **Q54**. When answering the following question, please consider your household income, your average monthly electricity bill, and how much extra you can actually afford to pay so the plant burns the fuel that you prefer.

Q55. How much would you be willing to pay each month, in addition to your current electricity bill, to ensure that the plant burns the fuel that you prefer (**Q54**)?

Circle the dollar amount closest to the amount you would be willing to pay extra each month.

\$0.00	\$0.10	\$0.25	\$0.50	\$0.75	\$1.00
\$1.25	\$1.50	\$2.00	\$3.00	\$4.00	\$5.00
\$6.00	\$8.00	\$11.00	\$14.00	\$19.00	\$22.00
\$26.00	\$30.00	\$35.00	\$40.00	\$45.00	\$50.00
\$55.00	\$60.00	\$70.00	\$80.00	\$90.00	\$100.00
\$110.00	\$120.00	\$140.00	\$160.00	\$220.00	\$300.00

OTHER: \$_____

ABOUT YOU AND YOUR HOUSEHOLD

**PLEASE REMEMBER THAT ALL INFORMATION
IS ENTIRELY CONFIDENTIAL**

H1. Your gender:

1. FEMALE
2. MALE

H2. Your age:

_____ YEARS

H3. Including yourself, how many members in your household are in each age group? (If none, write "0".)

_____ UNDER 18 YEARS OF AGE

_____ 18-35

_____ 35-60

_____ 60 AND OVER

H5. How would you describe your racial or ethnic background?
(Circle one.)

1. WHITE OR CAUCASIAN
2. BLACK OR AFRICAN AMERICAN
3. HISPANIC OR MEXICAN AMERICAN
4. ASIAN OR PACIFIC ISLANDER
5. NATIVE AMERICAN INDIAN
6. OTHER - PLEASE SPECIFY: _____

H4. How many years of education have you completed?
A high school education = 12 years.
(Circle number of years.)

6	7	8	9	10	11	12
13	14	15	16	17	18	19
20	21	22	23	24	25	25+

H6. What is your present employment status?
(Circle the best answer.)

1. LABORER	7. SERVICE EMPLOYEE
2. TEACHER	8. SELF-EMPLOYEED
3. FULL TIME HOMEMAKER	9. STUDENT
4. PROFESSIONAL	10. RETIRED
5. GOVERNMENT EMPLOYEE	11. UNEMPLOYED
6. SECRETARY/CLERICAL	12. OTHER - PLEASE SPECIFY:
-

H7. What is your total annual household income before taxes and other deductions?
(Circle best answer.)

1. UNDER \$9,999	9. \$80,000 - 89,999
2. \$10,000 - 19,999	10. \$90,000 - 99,999
3. \$20,000 - 29,999	11. \$100,000 - 119,999
4. \$30,000 - 39,999	12. \$120,000 - 139,999
5. \$40,000 - 49,999	13. \$140,000 - 159,999
6. \$50,000 - 59,999	14. \$160,000 - 179,000
7. \$60,000 - 69,999	15. \$180,000 - 199,999
8. \$70,000 - 79,999	16. \$200,000 AND OVER

H8. How active have you been active in community related issues during the past two years?
(Circle appropriate number.)

NOT AT ALL ACTIVE		SOMEWHAT ACTIVE		VERY ACTIVE	
1	2	3	4	5	

Appendix 4E
Sector Definition

Sector	Name	BEA Code
1	Agriculture	1-4
2	Iron Mining	5
3	Non-Iron Mining	6
4	Other Mining	9,10
5	Coal Mining	7
6	Petroleum Mining	8
7	Construction	11,12
8	Manufacturing	13-30,33,34,39-64
9	Petroleum Refining	41
10	Plastics	32
11	Glass	35
12	Stone	36
13	Steel	37
14	Metal Manufacturing	38
15	Transportation	65
16	Communication	66,67
17	Electric Utilities	68.01
18	Gas Utilities	68.02
19	Trade & Finance	69,70
20	Services (including Water & Sanitary Services)	71-79,68.03

Notes: Sectors 5, 6, 9, 17, and 18 from the energy aggregate.