DRAFT

LIFAC DEMONSTRATION

AT
RICHMOND POWER AND LIGHT

WHITEWATER VALLEY UNIT NO. 2

FINAL REPORT VOLUME 2:
PROJECT PERFORMANCE AND ECONOMICS

Sponsored by:

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ABSTRACT

This publication discusses the demonstration of LIFAC sorbent injection technology at Richmond Power and Light's Whitewater Valley Unit No. 2, performed under the auspices of the U.S. Department of Energy's (DOE) Clean Coal Technology Program. LIFAC is a sorbent injection technology capable of removing 75 to 85 percent of a power plant's SO_2 emissions using limestone at calcium to sulfur molar ratios of between 2 and 2.5.

The site of the demonstration is a coal-fired electric utility power plant located in Richmond, Indiana. The project is being conducted by LIFAC North America (LIFAC NA), a joint venture partnership of Tampella Power Corporation and ICF Kaiser Engineers, in cooperation with DOE, RP&L, and the Electric Power Research Institute (EPRI), the State of Indiana, and Black Beauty Coal Company.

The purpose of *Final Report Volume 2: Project Performance and Economics* is to consolidate, for public use, information that provides a technical and economic overview of the LIFAC Process. The report has been prepared pursuant to Cooperative Agreement No. DE-FC22-90PC90548 between LIFAC NA and the U.S. Department of Energy.

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LIST OF ABBREVIATIONS, ACRONYMS, AND TERMS

$CaCO_3$	Calcium carbonate
CaO	Calcium oxide
$CaSO_3$	Calcium sulfite
$CaSO_4$	Calcium sulfate
$Ca(OH)_2$	Calcium hydroxide, hydrated lime
Ca/S	Calcium/sulfur molar ratio (moles of Ca in sorbent/moles of S in inlet flue gas stream)
CCT	Clean Coal Technology
DOE	U.S. Department of Energy
EER	Energy and Environmental Research Corporation
EPA	U.S. Environmental Protection Agency
EPRI	Electric Power Research Institute
ESP	Electrostatic precipitator
FGD	Flue gas desulfurization
ICF Kaiser	ICF Kaiser Engineers, Inc.
ID fan	Induced draft fan
LIFAC	Limestone Injection into the Furnace with Activation of unreacted Calcium oxide, registered trademark of Tampella desulfurization process
LIFAC NA	LIFAC - North America (partnership)
LIMB	Lime Injection to Multistage Burner (registered trademark of EPA desulfurization process)
LNCFS	Low-NO_x Concentric Firing System
$MgCO_3$	Magnesium carbonate
MW	Megawatt
NO	Nitric oxide
NO_x	Nitrogen oxides
O_2	Oxygen
PLC	Programmable logic controller
PMS	Process monitoring system
RP&L	Richmond Power & Light
SCA	Specific collection area
SO_2	Sulfur dioxide

SO_3	Sulfur trioxide
ΔT	Temperature difference between actual flue gas temperature and saturation temperature (approach to saturation)
Injection	Pulverized limestone injection into the furnace through several injection nozzles at the furnace walls
Activation	Activation of the remaining CaO particles with water for further sulfur dioxide capture
Recycling	Recycling of ash and sorbent from ESP hoppers to the reactor inlet ductwork.

0.0 EXECUTIVE SUMMARY

The Clean Coal Technology (CCT) Program has been recognized in the National Energy Strategy as a major initiative whereby coal will be able to reach its full potential as a source of energy for the nation and the international marketplace. Attainment of this goal depends upon the development of highly efficient, environmentally sound, competitive coal utilization technologies responsive to diverse energy markets and varied consumer needs. The CCT Program is an effort jointly funded by government and industry whereby the most promising of the advanced coal-based technologies are being moved into the marketplace through demonstration. The CCT Program is being implemented through a total of five competitive solicitations.

LIFAC North America, a joint venture partnership of ICF Kaiser Engineers, Inc., and Tampella Power Corporation, is currently demonstrating the LIFAC flue gas desulfurization technology developed by Tampella Power. This technology provides sulfur dioxide emission control for power plants, especially existing facilities with tight space limitations. Sulfur dioxide emissions are expected to be reduced by up to 85% by using limestone as a sorbent. The LIFAC technology is being demonstrated at Whitewater Valley Unit No. 2, a 60-MW coal-fired power plant owned and operated by Richmond Power and Light (RP&L) and located in Richmond, Indiana. The Whitewater plant consumes high-sulfur coals, with sulfur contents ranging from 2.0-2.9 %.

The project, co-funded by LIFAC North America and DOE, is being conducted with the participation of Richmond Power and Light, the State of Indiana, the Electric Power Research Institute (EPRI), and the Black Beauty Coal Company. The project has a total cost of $21.4 million and a duration of 48 months from the preliminary design phase through the testing program.

The sponsors of this project believe that LIFAC has the potential to be a new and important sulfur dioxide SO_2 control option for U.S. utilities subject to the Clean Air Act's acid rain regulations. To be considered as a commercially feasible option in this particular emissions control market, LIFAC must demonstrate a high SO_2 removal rate while remaining competitive with other options on a cost per ton of SO_2 removed basis.

The LIFAC system combines conventional limestone injection into the upper furnace region with a post-furnace humidification reactor located between the air preheater and the electrostatic

precipitator (ESP). The process produces a dry, stable waste product that is removed from both the bottom of the humidification reactor and the ESP.

Finely pulverized limestone (80% < 325 mesh) is pneumatically conveyed and injected into the upper region of the boiler where temperatures are approximately 1800 to 2200 ° Fahrenheit. At these temperatures the limestone ($CaCO_3$) thermally decomposes to form calcium oxide (CaO). As the CaO passes through the furnace, initial desulfurization reactions occur. A portion of the SO_2 reacts with the CaO to form calcium sulfite ($CaSO_3$) which oxidizes to calcium sulfate ($CaSO_4$). Essentially all of the sulfur trioxide (SO_3) reacts with CaO to form $CaSO_4$.

At an overall SO_2 removal efficiency of 75%, approximately 25% of the sulfur dioxide removal occurs in the boiler, with an additional 50% sulfur removal after the unreacted lime passes through the air preheater to the vertical elongation of ductwork known as the LIFAC activation reactor. There the flue gas is sprayed with atomized water that allows the unreacted lime to hydrate to $Ca(OH)_2$ which more readily reacts with sulfur dioxide and forms $CaSO_3$. A combination of the proper water droplet size and residence time allows for effective hydration of the lime and complete water evaporation to create a dry reactor bottom product.

After exiting the humidification reactor, the flue gas is reheated before entering the ESP. Forty percent of the LIFAC-produced spent sorbent and fly ash is collected in the humidification reactor with the remaining 59.9% collected by the ESP. The LIFAC system can be designed so that both the reactor and ESP ash may be recycled to a point ahead of the reactor to improve sorbent utilization and to improve the SO_2 removal efficiency of the system to the range of 75 to 85%.

LIFAC is similar to other current sorbent injection technologies but has unique advantages with its use of a patented vertical humidification reactor. And while LIFAC's sulfur dioxide removal efficiency is not as high as traditional wet flue gas desulfurization systems, its cost and simplicity of design, construction and operation offer other advantages over these alternative systems. In particular the advantages of the LIFAC system are:

- High SO_2 removal rates - Currently available sorbent injection systems have been unable to sustain high SO_2 removal rates with any consistency. LIFAC has proven in the past and has demonstrated during this project the ability to achieve and sustain high SO_2 removal rates of 75 to 85% over long operating periods.

- By-products - Wet lime and limestone scrubbing systems create a wet by-product ash that must be further treated before disposal. LIFAC produces a dry solid waste ash containing calcium sulfite, calcium sulfate and fly ash. This waste is easily disposed of under U.S. regulatory requirements, may be recycled to increase LIFAC's efficiency, and may have commercial applications in the construction material industry.

- Compatibility and adaptability - LIFAC has minimal impact on the host site and systems, primarily the boiler, ESP and ID fan. In addition, LIFAC requires little space and few utilities and, therefore, is easily installed even in small or cramped power plant sites.

Construction of the LIFAC system has occurred in two phases over a period of one and a half years. The first phase of construction was completed during a routine plant outage in March 1991. This period was utilized to install tie-ins to the host site's existing systems.

Ductwork and three dampers were installed between the air preheater and ESP to allow flue gas flow to the LIFAC activation reactor. Tie-ins were also made to the power plant's steam, condensate and river-water supplies. Medium-pressure steam is used to reheat the flue gas exiting the LIFAC reactor, and water is needed for flue gas humidification inside the reactor. Limestone injection ports were installed in the boiler walls about 10 feet above the nose elevation.

The second phase of construction began in the fall of 1991 with the driving of piles from the reactor and the installation of underground conduit runs. Work continued through to the summer of 1992, with no need for plant downtime other than normally scheduled outages. During this time, the limestone storage area was completed and the injection system was installed on Unit No. 2. The activation reactor was constructed and tested with both cold air during a scheduled Unit No. 2 outage, and hot flue gas during a low electricity demand period. Other power plant tie-ins, such as the steam and condensate system, were also tested during low demand periods in the evenings or on weekends.

All of the construction work associated with the LIFAC system was performed in close proximity to the exterior of the power plant or in cramped areas inside the plant. The ductwork tie-ins and new steel work required inside the plant are located in small, difficult-to-access work areas. The reactor structure is approximately ten feet from the power plant with the outside ductwork and piping crossing over offices and the plant maintenance area. All of these new structures and equipment were constructed with no interference to daily plant operations.

The schedule for the LIFAC demonstration program extends over a four-year period from the beginning of preliminary design in August 1990 through the testing program completed in early August 1994. The LIFAC system was originally scheduled to come on-line in June of 1992, but due to delays in receiving construction permits and some minor startup problems, this date was moved to March 1993. Testing was then scheduled to continue through the summer of 1994.

The test plan for the LIFAC demonstration was composed of five distinct phases, each with its own objective. The first of these phases consisted of the initial baseline testing portion of the project. Measurements were taken to characterize the operation of the host boiler and associated subsystems prior to the use of the LIFAC system. The results were used for comparison purposes with the LIFAC system in operation and with data collected at the end of the project to determine any changes in the host's systems.

The second, or parametric, phase of testing was performed to determine the best combination of LIFAC process variables for SO_2 removal. The variables studied included the limestone injection nozzles' angle and location, the Ca/S molar ratio, the need for supplemental injection air at the boiler, the water droplet size and injection nozzle arrangement in the reactor, the ash recycling ratio, and the approach to saturation temperature of the flue gas exiting the activation reactor. The best combination of these variables was chosen at the conclusion of this phase and used for the remainder of the test program.

Parametric tests were conducted to examine the effects of different coal and limestone feeds on the SO_2 capture rate. Coals with sulfur contents as high as 3.3% were tested to determine LIFAC's compatibility with high sulfur U.S. coals. Limestones with different compositions were also tested to determine the LIFAC system's adaptability to local sorbent sources.

Optimization and long-term testing was also performed to demonstrate LIFAC's performance under commercial conditions. The LIFAC system was in operation 24 hours per day for several weeks using the power plant's baseline coal, high calcium limestone, and the optimum combination of process variables. In addition to process performance measurements, during this phase, the operating and maintenance requirements of the system were examined.

The final phase of testing consisted of the post-LIFAC tests. The baseline tests were repeated to gather information on the condition of the boiler and its associated subsystems. Comparisons were

made to the original baseline data to identify any changes either caused by the LIFAC system or independent of its operation.

It has also been shown at RP&L and other LIFAC installations that the system can be installed without affecting normal power plant operations. The demonstration showed that the system can economically reduce SO_2 emission when compared with other flue gas desulfurization technologies.

1.0 PROJECT OVERVIEW

1.1 Purpose of the Project Performance and Economics Report

The purpose of this Project Performance and Economics Report is to provide a technical and economic overview of the LIFAC desulfurization project. The report serves as a reference for the demonstration technology and its future commercialization. Final Report Volume 2: Project Performance and Economics has been prepared pursuant to Cooperative Agreement No. DE-FC22-90PC90548 between LIFAC North America (LIFAC NA) and the U.S. Department of Energy (DOE) titled "LIFAC Demonstration at Richmond Power and Light (RP&L) Whitewater Valley Unit No. 2."

1.2 Description of the Project

1.2.1 Project History

In 1983, Finland enacted acid rain legislation which applied limits on sulfur dioxide (SO_2) emissions sufficient to require that flue gas desulfurization (FGD) systems have the capability to remove about 80% of the SO_2 in the flue gas. This level could be met by conventional wet limestone scrubbers but not by then available sorbent injection technology. Tampella, therefore, began developing an alternative sorbent injection system which resulted in the LIFAC process.

In 1986, the first full scale test was performed at Imatran Voima's Inkoo power plant using a 70 MW side-stream from a 250 MW boiler. A 76% SO_2 removal rate with 1.5% sulfur coal was reached. A second LIFAC activation reactor was constructed to handle an additional 125 MW side-stream. This newer reactor is achieving removal rates of 75 to 80% while using Ca/S molar ratios of between 2 and 2.5 to 1. Also in 1988, the first tests with high sulfur U.S. coals were run at the Neste Kullo Laboratory. A Pittsburgh No. 8 Seam coal containing 3% sulfur was tested and a SO_2 removal rate of 77% was achieved at a Ca/S molar ratio of 2 to 1.

DOE has emphasized the use and further development of coal as an energy source for utilities and the industrial sector. At the same time, environmental responsibility has been mandated by the passage of the Clean Air Act Amendments of 1990. This Act establishes new lower emission levels of SO_2 from utility power plants, with Phase I of the regulations having come into effect in January

1995, and the more stringent Phase II regulations beginning in January 2000. To realize full potential of coal as an energy source while still complying with the new air pollution regulations, the DOE initiated the Clean Coal Technology (CCT) Program.

The CCT Program is a jointly funded government-industry effort to select the most promising advanced coal-based technologies and, over the next decade, move them into the commercial marketplace through demonstration. These demonstrations are conducted at a scale large enough to generate the data from design, construction, and operation that is necessary for the private sector to judge commercial potential and to make informed and confident decisions on commercial readiness.

The goal of the program is to make available to the U.S. energy marketplace, particularly the industrial and utility sectors, a number of advanced, more efficient, and environmentally responsive coal technologies. These technologies will reduce and/or eliminate the economic and environmental impediments that limit the full consideration of coal as a future energy resource. The program is being implemented through a series of five competitive solicitations which are now completed. Selections for the fifth solicitation were made in May 1993. Federal funding of $2.75 billion is committed for the five rounds of the program. When the private sector cost share is included, total funding approaches $7 billion. When the program is completed, clean coal technology options that will reduce the uncertainties of subsequent commercial-scale applications.

The intent as well as the objective of the DOE, as related to coal, has been endorsed most recently in the language of the Energy Policy Act of 1992 (Public Law 102-486). This legislation identifies a number of energy goals which already are a key part of the CCT Program, including achieving greater efficiencies in the conversion of coal to useful energy; achieving control of sulfur oxides, oxides of nitrogen, air toxics, solid and liquid wastes, greenhouse gases, or other emissions resulting from coal use; and promoting the export and transfer of U.S. clean coal technologies and services to developing countries and countries making the transition to free market economies.

CCT projects seek to demonstrate the commercial feasibility of the most promising advanced coal technologies that have already reached the proof-of-concept stage. These projects are conducted under jointly funded cooperative agreements-not contracts-between government and industry. The industrial partner in each project contributes at least 50% of the total cost-in many cases, more-and the patent rights for inventions developed during the demonstrations are normally granted to the

participant. Each project involves a technology that the industrial partner believes has very real commercial potential. The program preserves incentives the industrial partner needs to subsequently bring the technology into the marketplace.

The emphasis in the program has evolved through the five rounds. Clean Coal I covered a broad range of advanced technologies. Clean Coal II focused on technologies to reduce acid rain precursors, especially those that can be applied to existing facilities using high-sulfur coal. Clean Coal III expanded the scope of the Clean Coal II solicitation to include coal-based technologies that help to meet future energy demands in an environmentally acceptable manner. Clean Coal IV included technologies to address similar needs in new as well as existing plants. The emphasis on high efficiency and high environmental performance has increased in each successive round of the program. Clean Coal V gave significant credit to projects that offer increased efficiency and environmental performance.

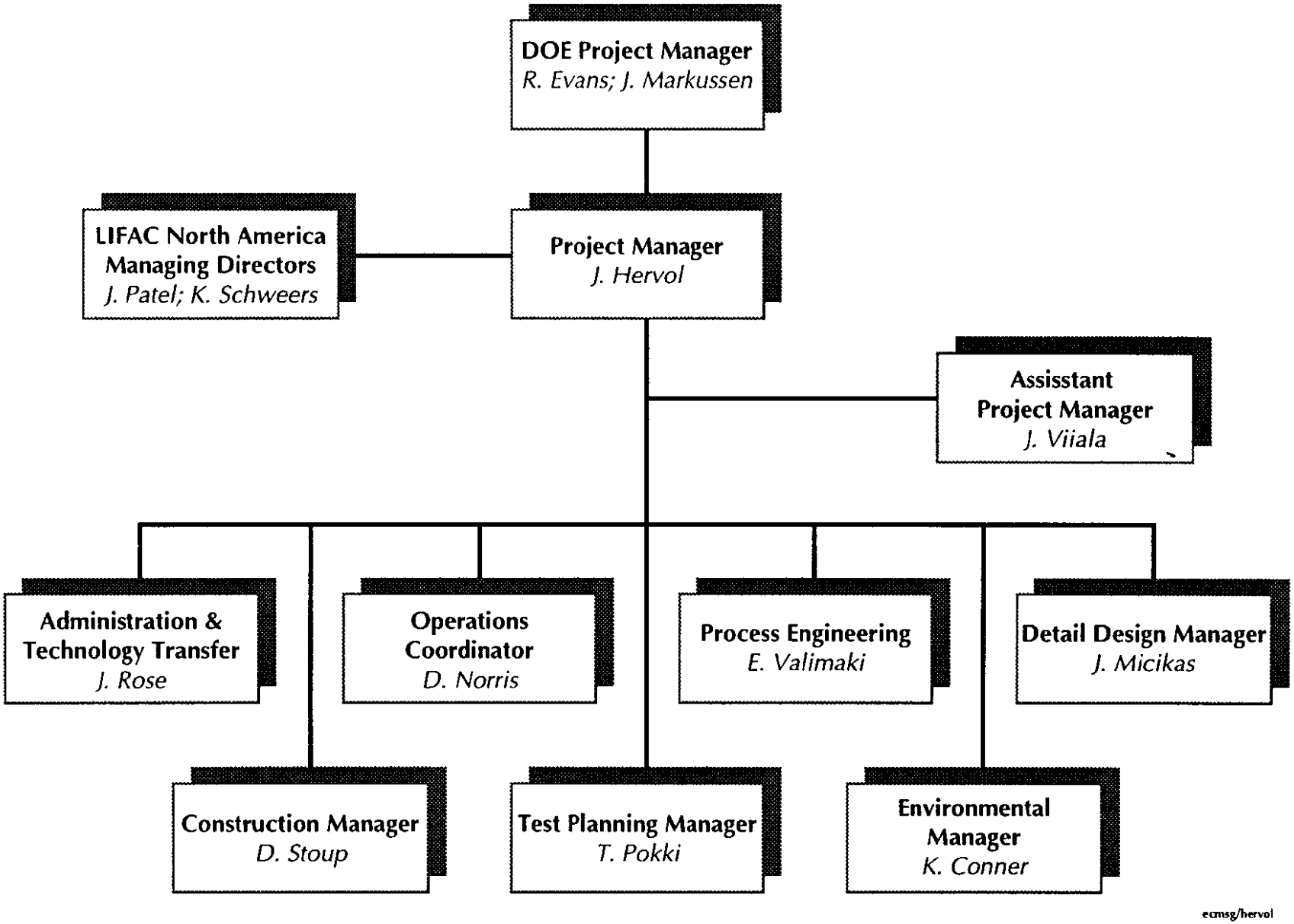
The LIFAC system was one of thirteen projects selected for funding under Round III of the CCT Program. A Cooperative Agreement between DOE and LIFAC NA was signed in November 1990. Due to scheduled outages at the host site, Richmond Power and Light's (RP&L's) Whitewater Valley Unit No. 2 in Richmond, Indiana, design and procurement of critical equipment began in August 1990, with DOE funding contingent on final signing of the Cooperative Agreement.

1.2.2 Project Organization

The LIFAC demonstration was conducted by LIFAC NA, a joint venture partnership between ICF Kaiser Engineers, Inc. (ICF Kaiser) and Tampella Power Corporation. ICF Kaiser is a U.S. company based in Oakland, California, and a subsidiary of ICF Kaiser International, Inc., based in Fairfax, Virginia. Tampella Power is a subsidiary of a large diversified international company, Tampella Corporation, which is based in Tampere, Finland, and the original developer of the LIFAC technology.

LIFAC NA is responsible for the overall administration of the project and for providing the 50% matching funds. With the exception of project administration, most of the actual work is being performed by the two parent firms under service agreements with LIFAC NA. Both parent firms work closely with RP&L and the other project team sponsors, including ICF Resources, EPRI, Indiana Corporation for Science and Technology (ICS&T), and Black Beauty Coal Company. ICF

Figure 1-1 Project Organization



Kaiser managed the demonstration project out of its Pittsburgh office, which provided excellent access to DOE representatives at the Pittsburgh Energy Technology Center. A project organization chart is provided in Figure 1-1.

1.2.3 Host Site

The project site for the LIFAC demonstration is RP&L's Whitewater Valley Unit No. 2 pulverized coal-fired power station (60 MW), located in Richmond, Indiana. Richmond is approximately 75 miles east of Indianapolis, Indiana, and nearly 40 miles west of Dayton, Ohio. Whitewater Valley Unit No. 2, which began service in 1971, is a Combustion Engineering, tangentially-fired boiler which uses high-sulfur bituminous coal from Western Indiana. Actual power produced by the unit approaches 65 MW.

Whitewater Valley 2 has several important qualities as a LIFAC demonstration site. One of these is that Whitewater Valley Unit No. 2 was the site of a prior demonstration of LIMB sorbent injection technology, jointly sponsored by EPRI and the U.S. Environmental Protection Agency (USEPA). Much of the sorbent injection equipment remained on-site and was used in the LIFAC demonstration. Another advantage of the site is that Whitewater Valley Unit No. 2 was a challenging candidate for a retrofit due to the cramped conditions at the site. The plant is thus typical of many U.S. power plants which are potential sites for application of LIFAC. In addition, the Whitewater Valley Unit No. 2 boiler is small relative to its capacity; hence, it has a higher temperature profile relative to other boilers. This situation requires sorbent injection at higher points in the furnace to minimize deadburning of the reagent, but it decreases residence times needed for sulfur removal. The demonstrate project was intended to show LIFAC's performance under operating conditions typical of U.S. power plants. The project demonstrated LIFAC on high-sulfur U.S. coals and was a logical extension of the Finnish demonstration work which is important for LIFAC's commercial success in the U.S.

1.2.4 Project Schedule

To demonstrate the technical viability of the LIFAC process to economically reduce sulfur emissions from the Whitewater Valley Unit No. 2, LIFAC NA conducted a three-phase project, as follows:

- Phase I: Design

- Phase IIA: Long Lead Procurement
- Phase IIB: Construction
- Phase III: Operations

Except for Phase IIA, each phase was comprised of three tasks: a management and administration task, a technical task, and an environmental task. The design phase began on August 8, 1990, and was scheduled to last six (6) months. Phase IIA, long lead procurement, overlaps the design phase and was expected to require about four (4) months to complete. The construction phase was to continue for another seven (7) months, while the operations phase was scheduled to last about twenty-six (26 months). Figure 1-2 shows the original estimated project schedule which is based on an August 8, 1990, start date and a planned outage of Whitewater Valley Unit No. 2 during March 1991.

It was during this outage that all the tie-ins and modifications to existing Unit No. 2 equipment were made. This required that the construction phase begin in early February 1991 and be completed by the end of August 1991. Operations and testing were to begin September 1991, and continue for 26 months. However, the project encountered delays in receiving its construction permit. These delays, along with some design changes and an approved expansion in project scope required that the Design Phase be extended by about eleven months. Therefore, construction was not completed until early June 1992. This represented an eight-month extension in the overall schedule.

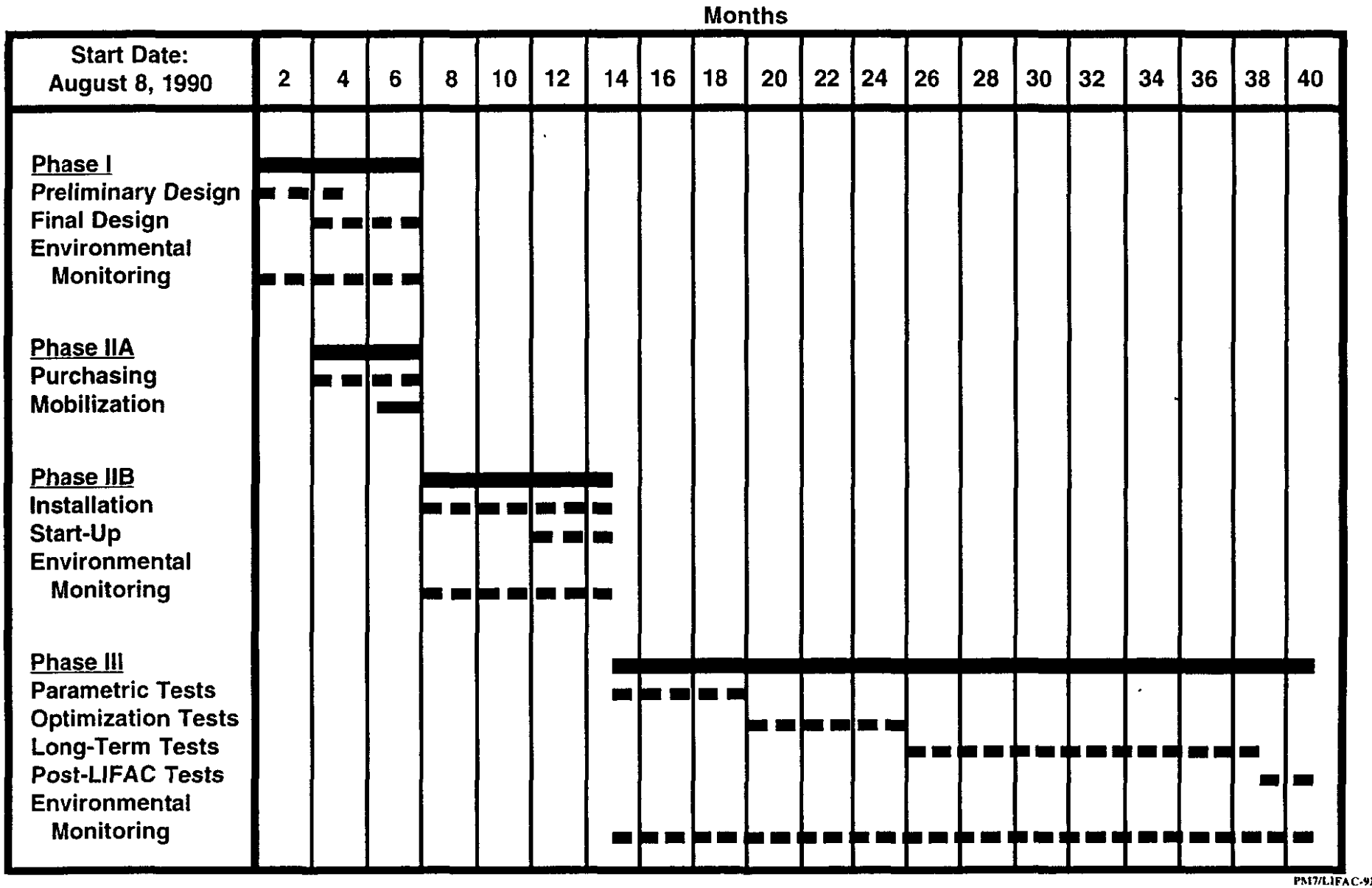
During the last half of 1992, problems were encountered during startup and commissioning of some of the LIFAC components and systems. These problems required the parametric tests to be delayed until the first quarter 1993, which subsequently required adjustments in the entire testing schedule. During the initial parametric tests, problems were encountered with increased opacity levels. These problems forced an extension in the parametric test schedule, and, consequently, an adjustment was made to the testing schedule as shown in Figure 1-3. These delays, however, did not impact the overall duration of the Phase III activities and the total project duration remained at the modified 48 months.

Figure 1-2

LIFAC

LIFAC Demonstration

Original Project Schedule



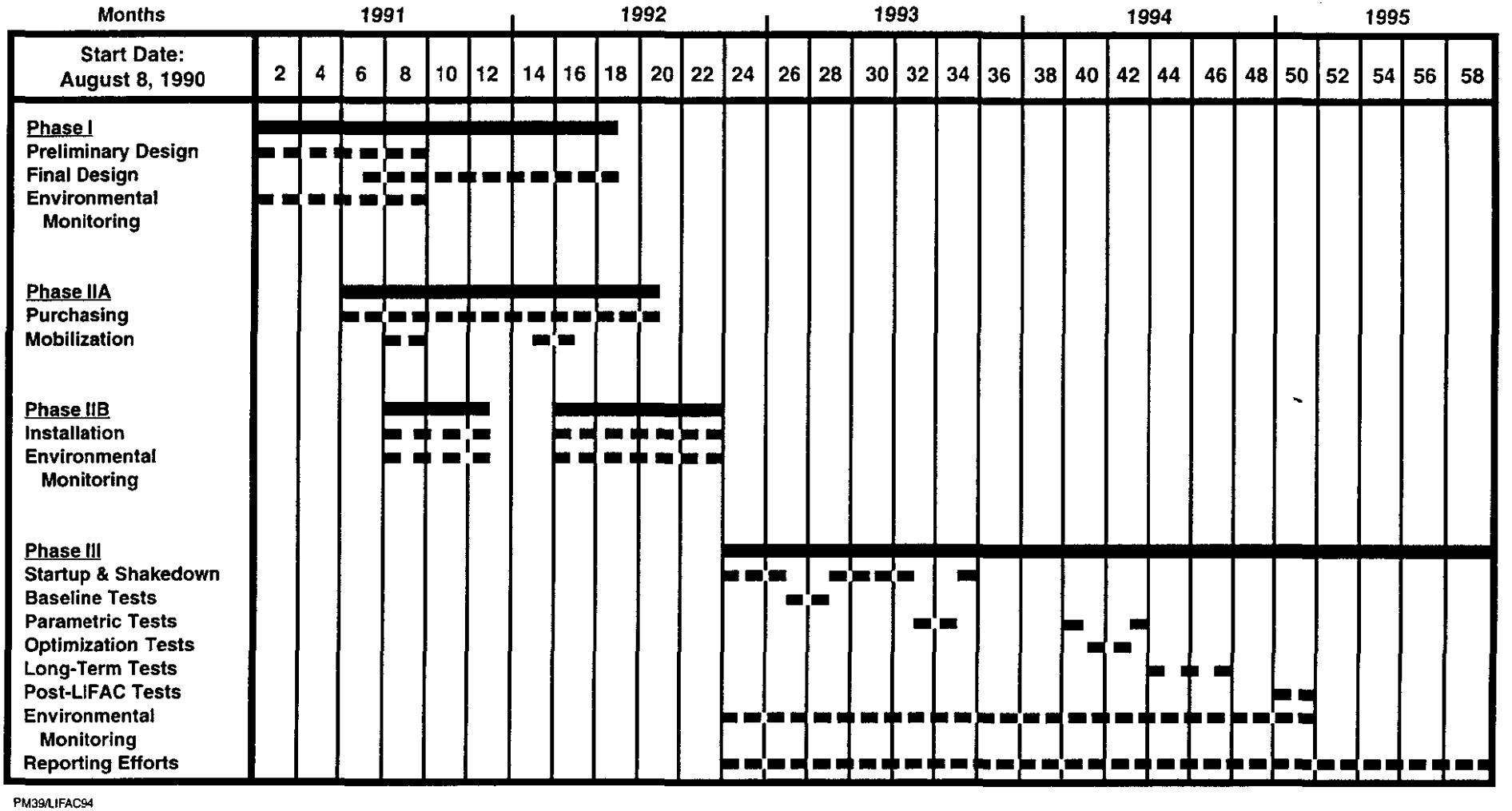
PM7/LIFAC-92

LIFAC Demonstration

Current Project Schedule

(Revised July 1994)

Figure 1-3



PM39/LIFAC94

1.3 Project Objectives

The sponsors of this project believe that LIFAC has the potential to be a new and important SO_2 control option for U.S. utilities subject to the Clean Air Act's acid rain regulations. To be considered as a commercially feasible option in this particular emission control market, LIFAC must demonstrate a high SO_2 removal rate while remaining competitive with other options on a cost per ton of SO_2 removed basis. To this end, the sponsors of this project designed the demonstration with the following objectives in mind:

- **Sustained High SO_2 Removal Rate** - Incorporated into the test plan were several periods of long-term testing which were intended to demonstrate LIFAC's SO_2 removal and reliability characteristics under normal operating conditions.

- **Cost** - LIFAC must compete with both low capital cost, low SO_2 removal rate options such as sorbent injection, and high capital cost, high SO_2 removal rate options such as wet scrubbing. This project was designed to demonstrate LIFAC's competitiveness on a cost per ton of SO_2 removed basis with currently available alternatives.

- **Retrofit Adaptability** - The host site chosen required a retrofit with tight construction conditions that would prove LIFAC's ability to be installed where other technologies might not be possible. Construction was intended to also demonstrate LIFAC's ability to be built and brought on-line with zero plant downtime other than scheduled outages.

- **System Compatibility** - A major concern of utilities is the degree of compatibility of SO_2 removal systems with their existing operations. This demonstration was to show LIFAC's minimal impact on the host site's boiler and associated subsystems.

1.4 Significant of the Project Commercialization and Process Advantages

The significance of this project was to show that the LIFAC technology could provide SO_2 removal at a significant rate and competitive costs. While utilizing U.S. coals, a successful demonstration would provide another option for power plants besides costly wet scrubbers, with minimal impacts to the host site.

Wet scrubbers are the most prevalent FGD technology and account for approximately 90% of U.S. scrubber systems. Wet FGD systems that use lime or limestone remove about 90% of the SO_2 and usually produce a sulfite/sulfate sludge waste product. Although the LIFAC process cannot match the high removal rates (90% or more) achieved by conventional wet scrubbers, the process does offer several advantages including:

- The technology can be more easily retrofit onto most power plants because the vertical activation chamber requires less space.

- The technology has lower capital costs which makes it especially attractive to existing plants that have fewer years to amortize capital investments as compared to long-lived power plants.

- The technology uses a widely available reagent, limestone, rather than more expensive sorbent materials, such as lime.

- The need for slurry preparation/handling equipment is eliminated.

- The waste product is dry and easy to handle. In comparison, conventional wet limestone scrubbers produce a wet sludge which requires special handling and treatment.

- The technology is typically compatible with other plant systems such as electrostatic precipitators (ESPs) and induced draft (ID) fans, thereby minimizing costly retrofit plant modifications in order to employ the technology.

The LIFAC system also has potential advantages over less conventional sorbent injection systems now being tested. These include:

- Use of limestone as opposed to lime or other more expensive sorbents.

- Removal rates of 75-85%, which exceed the removal rates of many dry sorbent injection systems.

- Improved control of wall deposition with humidification in vertical chamber as opposed to in-duct humidification.

The LIFAC technology's potential for commercialization is increased by its ability to remove 75-85% of the SO_2, its low space requirement, and its low retrofit costs.

1.5 DOE's Role in the Project

The DOE was responsible for monitoring all aspects of the project and for granting or denying all approvals required by the Cooperative Agreement. The DOE Contracting Officer is DOE's authorized representative for all matters related to the Cooperative Agreement.

The DOE Contracting Officer appointed a Contracting Officer's Technical Representative (COTR) who is the authorized representative for all technical matters and has the authority to issue "Technical Advice" which may:

- Suggest redirection of the Cooperative Agreement effort, recommend a shifting of work emphasis between work areas or tasks, and suggest pursuit of certain lines of inquiry which assist in accomplishing the Statement of Work.

- Approve those reports, plans, and items of technical information required to be delivered by the Participant to DOE under the Cooperative Agreement.

The DOE COTR does not have the authority to issue any technical advice which:

- Constitutes an assignment of additional work outside the Statement of Work.

- In any manner causes an increase or decrease in the total estimated cost or the time required for performance of the Cooperative Agreement.

- Interferes with the Participant's right to perform the terms and conditions of the Cooperative Agreement.

2.0 TECHNOLOGY DESCRIPTION

2.1 Chemical Process

The LIFAC (Limestone Injected into the Furnace with Activation of untreated Calcium oxide) technology combines upper-furnace limestone injection followed by post-furnace humidification in an activation reactor located between the air preheater and the ESP. The process produces a dry and stable by-product that is partially removed from the bottom of the activation reactor and partially removed at the ESP.

Finely pulverized limestone (80% < 325 mesh) is pneumatically conveyed and injected into the upper part of the furnace. Since the temperatures at the point of injection are in the range of 1800 - 2000 ° F, the limestone ($CaCO_3$) thermally decomposes to form calcium oxide (CaO). As the CaO passes through the furnace, initial desulfurization reactions occur. A portion of the sulfur dioxide (SO_2) reacts with the CaO to form calcium sulfite ($CaSO_3$), which oxidizes to calcium sulfate ($CaSO_4$). Essentially all of the sulfur trioxide (SO_3) reacts with CaO to form $CaSO_4$.

The flue gas, unreacted CaO, and ash exit the boiler and pass through the air preheater. On leaving the air preheater, the gas/CaO/ash mixture is directed to the LIFAC activation reactor. In the reactor, additional SO_2 capture occurs after the flue gas is humidified with a water spray. Humidification converts CaO to calcium hydroxide, $Ca(OH)_2$, which enhances further SO_2 removal. The primary reaction product in the activation reactor is calcium sulfite ($CaSO_3$). The activation reactor is designed to allow time for effective humidification of the flue gas, activation of the CaO, and reaction of SO_2 with the sorbent. All the water droplets evaporate before the flue gas leaves the activation reactor. The net effect is that at a Ca/S molar ratio in the range of 2:1 to 2.5:1, 75-85% of the SO_2 is removed from the flue gas.

The flue gas leaving the activation reactor enters the existing ESP, where the spent sorbent and fly ash are removed from the flue gas and sent to the disposal facilities. The solids collected by the ESP consist of fly ash, $CaCO_3$, CaO, $Ca(OH)_2$, $CaSO_4$, and $CaSO_3$. To improve utilization of the calcium and increase SO_2 removal, a portion of the spent sorbent collected in the ESP hoppers is recycled to the ductwork just ahead of the activation reactor.

Figure 2-1 is a simple flow diagram of the LIFAC process as designed at the RP&L host site. The major process areas are as follows, which are discussed in detail below:

- Limestone Storage and Handling Area
- Boiler Injection Area
- Activation Reactor Area
- ESP Ash Recycle Area
- Process Monitoring and Control

2.1.1 Limestone Storage and Handling Area

The majority of the equipment remaining from the previous lime injection demonstration was used for handling and storage of the hydrated lime. Because of the similarities between the physical characteristics of limestone and hydrated lime, the utilization of this equipment was a high priority in designing the limestone system. Inspection of this equipment allowed reuse of the following items:

- 125 ton Storage Silo (now a LIFAC feed silo)
- Limestone Feeding System
- Two Vent Baghouses

The LIFAC system was designed for operation on a continuous basis at a limestone injection rate of 200 lb/min, which necessitated the inclusion of additional limestone storage and handling equipment. A new 250 ton storage silo was designed to provide enough storage capacity for LIFAC to operate three days, such as over a weekend, without any limestone deliveries.

The pulverized limestone arrives at the plant via truck transport. A pneumatic transport line which can serve either the new or existing silo was designed with a maximum operating pressure of 16 psi and flow rate of 17 lb/min. The transport line has replaceable wear back fittings on all elbows along with Victaulic couplings. A manual diverter valve on the roof of the new silo directs the limestone to either of the two silos.

Limestone is transported pneumatically from the new storage silo to the existing silo for injection into the boiler. The new silo has air slides which fluidize the limestone and ensure an even, continuous flow of material by gravity to a rotary valve. The rotary valve feeds a conveying tee where transport

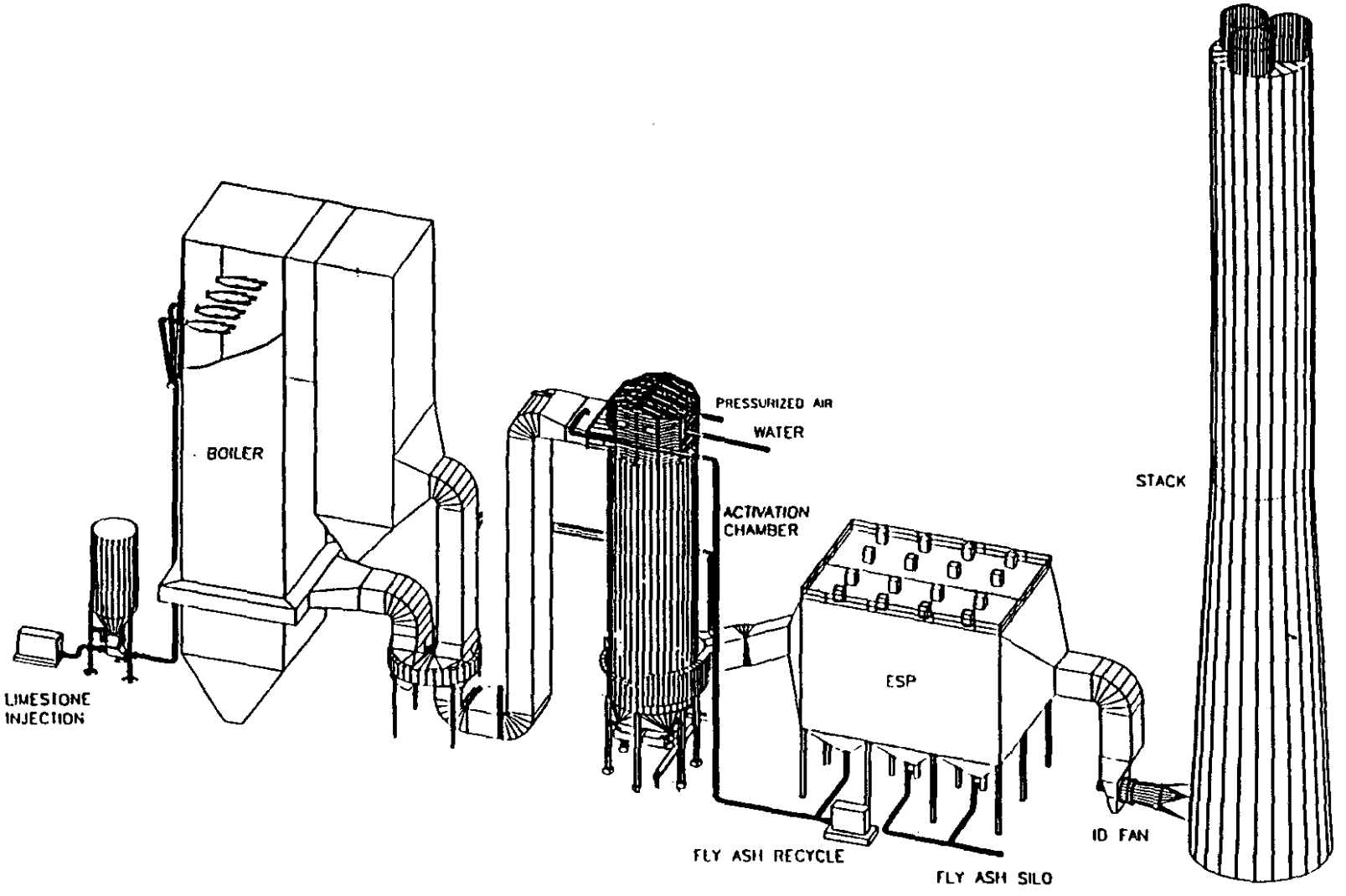


FIGURE 2-1: A Schematic of the LIFAC Process at Richmond Power & Light

air is introduced to carry the limestone to the top of the existing silo. The capacity of the transport pipe is 400 lb/min of limestone. The transport air is supplied by a new rotary lobe air blower with a maximum capacity of 1600 ACFM at 12 psig.

Limestone quantity in the silos is determined by measuring the weight of the silos with weight cells. Both the new and existing silo have a set of level indicators. The new silo has been equipped with low, high and high/high indicators, while the existing silo has only low and high levels indicators from the previous demonstration. On the top of the existing silo there are two vent baghouses to prevent dust emissions from the silos during truck unloading. The new storage silo has a pressure equalizing vent to the existing silo. Both silos have manually-operated knife gate valves above their rotary valves. The gate valve on the existing silo is used to isolate the silo material from the weigh feeding equipment.

Limestone injection into the host boiler may be performed only from the existing feed silo. The LIMB demonstration left behind the following equipment which was utilized in the LIFAC design:

- Gravimetric Weigh Feeder and Control System
- Fuller-Kinyon Pump
- Rotary Valve
- Flexible Lime Transport Pipe to Boiler
- Flow Control and Pressure Relief Valve

Limestone from the existing feed silo is fed by the rotary valve to the weigh feeder. The rotary valve is equipped with a variable speed DC-drive which receives an input control signal from the weigh feeder system's controller. The weigh feed consists of a feeder screw and a weigh screw. The feeder screw operates at a constant speed and transports the limestone evenly to the weigh screw. The weigh screw is mounted on load cells and the mass flow rate of limestone is determined by multiplying the weight of material on the screw by its rotation rate.

The limestone is dropped by the weigh screw feeder into the Fuller-Kinyon pump for transport to the boiler. The Fuller-Kinyon pump has a screw which feeds the limestone to a chamber where transport air from the new blower is introduced for pneumatic transport. The maximum design feed rate is 300 lb/min of limestone at 12 psig. The Fuller-Kinyon pump is equipped with an existing vent baghouse and it also serves as an airlock, using the material and check valve to isolate the transport

line from the silo. Attached to the pump is a section of flexible hose 8 inches in diameter, which is used to pneumatically convey the pulverized limestone into the boilerhouse and connects to steel piping with replaceable elbows.

2.1.2 Boiler Injection Area

Because of higher limestone flow rates and the large number of injection ports on the boiler, no boiler injection equipment from the LIMB demonstration could be utilized. A new primary splitter was designed which separates the incoming limestone from the Fuller-Kinyon pump into six streams. The primary splitter is equipped with two blowout connections for each of the six streams to clean any plugged material. Each of these six streams has another secondary splitter to achieve the needed twelve streams for every injection location on the boiler.

The limestone is moved from the secondary splitters to the boiler injection nozzles via a carbon steel pipe with flexible hose ends to allow for boiler expansion and contraction. There are six injection nozzles on two different levels of the boiler. Each level has four injection ports on the south boiler wall and one on both the east and west walls. These ports may be used in any combination to allow optimum SO_2 removal at different boiler loads.

The injection nozzles are made of stainless steel and include a sight glass to check for limestone pluggage. Mounted to the boilerhouse structural steel, these nozzles may be adjusted vertically to accommodate boiler expansion. The nozzles can also pivot $\pm$ 15° to achieve a range of injection angles into the boiler.

A secondary air fan has been provided at the injection location to ensure that the velocity needed for deep penetration and even dispersal of the limestone into the boiler is achieved. This constant speed fan is equipped with a remotely-controlled variable position damper and moves a maximum of 6,700 SCFM of air at 1.73 psig. The air is carried by ductwork to the south boiler wall, where 12 sections of flexible tubing are connected to each injection nozzle. Each section of flex tubing contains one manually-controlled fixed position damper and a second flow controlled damper. The mixing of the secondary air and limestone in the injection nozzle may be viewed through the sight glass.

2.1.3 Activation Reactor Area

The reactor area includes all the ductwork, the reactor vessel itself, and the associated systems that handle and humidify the flue gas after the boiler and before it enters the ESP. This proprietary aspect of the process is unique to the LIFAC process; and, therefore, the only equipment that could be reused were three Ingersoll Rand compressors for atomizing air supply.

The reactor vessel, is a vertical chamber in the ductwork where humidification of the flue gas occurs. The vessel is 133 ft. high and 28 ft. in diameter. It is designed with 3/8-inch thick stainless steel walls to prevent corrosion caused by the precipitation of acid in the humidified flue gas. The vessel has a maximum design pressure of negative 25 inches of water and a temperature of 400°F. Also included in its design are base slide plates and walkways, with one fixed and one free end to allow for maximum thermal growths of 4.5 inches vertical and 1 inch horizontal. There are also five levels of inspection doors, with six doors on each level, to permit access to the interior of the vessel for visual inspection, flue gas measurements, and repair work.

The bottom section of the reactor vessel interior contains a baffle cone which reduces the cross-sectional area and redirects the flue gas flow 180 degrees upward. This allows some of the ash to fall out of the gas stream for collection. At the top outside portion of the baffle cone are three equally spaced outlet duct openings for the exiting flue gas.

Instrumentation on the reactor vessel consists of thermocouples and differential pressure indicators. These instruments create a temperature profile and measure the pressure drop across the reactor vessel. Temperature readings determine the amount of reheat steam necessary to keep the flue gas safely above its saturation temperature.

Flue gas from the boiler is carried to the reactor vessel via a section of bypass ductwork that is tied to the host site's ductwork immediately following the air preheater. The ductwork has been sized according to a maximum gas flow rate of 282,100 ACFM with the Unit No. 2 load at 65 MW. Thermal expansion is compensated for by three expansion joints from 9 to 18 inches in width. Sliding base plates are also employed for lateral ductwork movement. The return ductwork is similar, although there are three outlet openings from the reactor vessel, each having its own expansion joint. The remaining ductwork has six expansion joints, each 12 inches in width. The design of the

ductwork includes both turning vanes in the bends to reduce turbulence in the gas flow and 6 inches of insulation to reduce heat loss.

Installed in the three outlet duct legs leaving the reactor are three stainless steel steam reheaters. These reheaters raise the outlet flue gas temperature enough to prevent acid precipitation in the Lodge Cottrell ESP units. The steam supply for the reheaters is from the plant's medium pressure steam system which is at 570°F and 246 psi. The maximum combined throughput of the three units at these operating conditions is 9000 lb/hr of steam. A steam-driven condensate collection system returns the condensate to the plant's system.

In both the inlet and outlet sections of ductwork to the reactor vessel are located a pair of analyzers that take SO_2, NO, and O_2 readings, thereby measuring the effectiveness of the reactor vessel in reducing emissions. Three flow elements measure the actual amount of flue gas being treated by the reactor, while three differential pressure indicators measure the flow exiting the reactor.

The humidification system requires water and a compressed air supply. Water is provided by a ground level centrifugal pump with maximum operating specifications of 115 GPM and 175 psig. The water supply may come from the plant's river water supply or its chilled water system. A duplex basket-type strainer, installed immediately after the pump, removes any large particulates. The water is then pumped vertically approximately 135 feet to the top of the reactor.

Three Ingersoll Rand helical screw air compressors provide the atomizing air for the humidification. Each compressor has a capacity of 870 ACFM at 125 psig and is located in the limestone storage area. At the top of the reactor in the penthouse, the air and water are combined in dual fluid nozzles. The air distributes the water evenly into the flue gas inside the top of the reactor. Each of the clusters of fluid nozzles has an air-operated scraper to remove ash deposits. The compressors supply the air for the scrapers as well as two sets of vibrators which prevent the buildup of ash on the inside of the reactor vessel.

Instrumentation on the water supply consist of an orifice plate to measure flow and a flow control valve. The flow rate of the water is determined by the temperature of the flue gas leaving the reactor vessel. The air line has a pressure control valve which can be set independent of the flue gas temperature.

The ash separated from flue gas in the bottom of the activation reactor is removed with two bottom drag conveyors. Both of these conveyors have a 22 ton/hr capacity and are furnished with crushers for larger accumulations of ash which may be dislodged by the reactor's vibrators. The motors on the drag chain and the crusher are reversible in case of jamming. Ash from the bottom drag conveyors is dropped onto a flight transfer conveyor, moving the ash to a pair of double-dump valves which empty into roll off containers for disposal. The double dump valves provide a constant seal to maintain the negative pressure of the flue gas stream through the reactor.

2.1.4 ESP Recycle Area

The ash separated from flue gas by the Lodge Cottrell ESP units is removed by gravity and pneumatically recycled directly into the flue gas ductwork immediately preceding the reactor. Two of the four ESP hoppers are equipped with variable speed rotary valves which feed two conveying tees where transport air is introduced. A rotary lobe blower provides a maximum of 890 ACFM of air at 7 psig. A manually operated diverter valve in the transport line determines whether the ash is directly recycled or sent to a 10 ton capacity ash surge tank in the reactor area. The surge bin was designed and installed for future use if needed.

2.1.5 Process Monitoring and Control

The operation of the LIFAC process is controlled by the PLC (programmable logic controller) using ICF Kaiser proprietary process control software. The process control system is used only for the LIFAC equipment. The operation of the boiler equipment and associated subsystems continue to be controlled by the power plant's original control system and is only monitored by LIFAC's software.

The process control system hardware consists of two identically installed IBM 756 Industrial computers with 19" VGA color monitors. Both computers are located in the boiler control room. One unit is used to control the entire LIFAC process. A printer for alarm reporting is connected to this computer. The other computer, the monitoring computer, is intended to monitor and collect data. It can also be used as a spare control computer, if necessary.

All the control commands from the control computer to the equipment and feedback data from the equipment and instruments to the computers are transported through three input/output (I/O) racks. Rack No. 1 is for the boilerhouse equipment and instruments; rack No. 2 is for the limestone area

equipment and instruments; and rack No. 3 is for the activation reactor area equipment and instruments. The system has a total of 333 input or output points: 29 RTD's, 51 analog inputs, 35 analog outputs, 160 digital inputs, and 58 digital outputs.

3.0 UPDATE OF THE PUBLIC DESIGN REPORT

3.1 Design and Equipment Changes

In order to improve operability and efficiency of the LIFAC process, several modification/changes to the system were made since the completion of construction and the completion of Final Report Volume 1 - Public Design[7]. This section includes a brief description of each of these modifications/ changes.

The variable frequency drive (VFD) failed several times during the demonstration program. The VFD installed as part of the project to increase the efficiency of the ID fan, which experiences higher loading when LIFAC is in operation. In the absence of VFD, the main bypass damper was opened 10-15% to lessen the system pressure drop. An unknown amount of unreacted lime bypassed through the damper. The VFD failures also contributed to the interruption of many tests and significant process downtime. The VFD was eventually deemed unreliable and placed out-of-service for most of the demonstration. However, it is now operational.

Proximity limit switches were installed on the double dump valves after the existing mechanical switches continued to fail and disrupt process operation. The new switches have no mechanical parts and are more suitable for a dirty environment.

A reheat bypass duct was installed between the area slightly above Unit No. 2 boiler's economizer and the activation reactor outlet ductwork. This reheating was designed to complement the existing steam reheat system. The duct is equipped with a damper which can be controlled automatically using the process monitoring system.

Ladder turning vanes were positioned in the ESP inlet, replacing the existing turning vanes. Also, perforated plates were installed on both the inlet and outlet ducts of the ESP. As a result, gas distribution through the ESP was improved significantly.

4.0 DEMONSTRATION PROGRAM

4.1 Test Plan

The LIFAC test program had two main objectives. The first was to showcase the LIFAC system by demonstrating the:

- LIFAC system for U.S. utilities.
- Economic feasibility and process suitability for different coals and limestones.
- Effectiveness at reducing SO_2 emissions when burning high sulfur coals.

The second objective was to perform an in-depth evaluation of LIFAC's effect on the host site's boiler and associated equipment. In meeting this objective, the test program provided data for the following:

- The effect on other boiler emissions, such as NO_x and particulates.
- Optimization of system performance at the host site.
- The impact of LIFAC on the host boiler and associated subsystems.

The test program at RP&L was divided into five sets of tests. The first set of tests was designed to gather baseline data on the boiler and other plant facilities. The LIFAC system was then optimized and tested with various coals and limestones, as well as long-term testing. Finally, post-LIFAC testing provided data for comparison to pre-LIFAC baseline data. The tests are described briefly below:

- Baseline Tests - Baseline testing was performed prior to startup of the LIFAC system. This testing provided data on the baseline operating conditions of the host boiler, used for comparison throughout the demonstration program.

- Parametric Tests - Parametric testing followed baseline testing and was designed to evaluate the effectiveness of various combinations of LIFAC process parameters. During this test period, five different types of sorbent were tested. The optimum combination of process parameters determined in the tests was used throughout the remaining demonstration.

- Optimization Tests - Optimization testing was performed after the parametric tests to determine the reliability and capability of the LIFAC process over short, continuous operation periods. These tests were performed continuously 24 hours a day at actual operation conditions, as opposed to the short 8- to 12-hour parametric tests performed previously. The baseline coal and limestone, along with the optimum parametric settings, were used during this phase of testing.

- Long-Term Tests - Long-term testing was performed after the optimization tests to determine the reliability and capability of the LIFAC process over long, continuous operating periods. These tests were performed continuously over several hundred hours. One coal quality was used during this test period. Following this phase and preceding the post-LIFAC tests, selected parametric and optimization tests were repeated.

- Post-LIFAC Tests - Post-LIFAC testing was similar to baseline testing. Data were gathered without the LIFAC system in operation to compared with the baseline data to determine the impact LIFAC had on the host site's boiler and associated subsystems.

The test program was staffed by a combination of personnel from RP&L, ICF Kaiser, and Tampella. The test program, including all five sets of tests, lasted approximately 2,800 hours over a 26-month period. The structure of the test plan at RP&L is shown in Figure 4-1.

4.2 Operating Procedures

4.2.1 Instrumentation and Data Acquisition

- **Process Control and Data Collection System**

The impact of the LIFAC process on gaseous and particulate emissions was determined by measuring the SO_2, sulfur trioxide (SO_3), oxygen (O_2), nitric oxide (NO), opacity and particulate contents of the flue gas. During the test program, SO_2, O_2, and opacity were measured continuously. Effects of the LIFAC process on NO and SO_3 emissions and on ash resistivity were determined by short, periodic tests. Particulate emission measurements were performed periodically in connection with ESP testing. The emission measurement locations are shown on Figure 6-1. Emission measurement

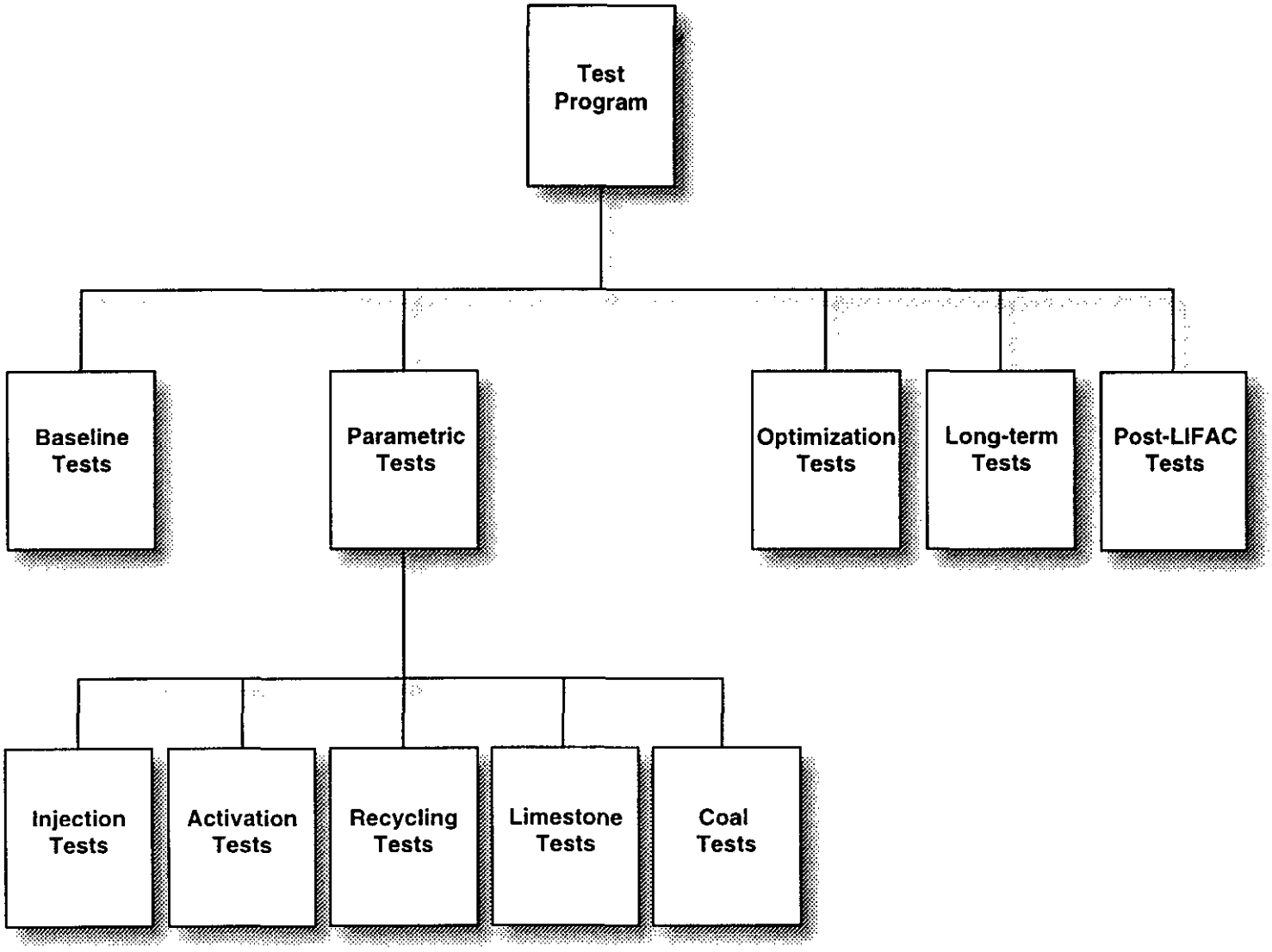


Figure 4-1
Structure of the LIFAC Demonstration Test Plan

procedures are summarized in Appendix II. See Section 2.1.5 for a detailed description of the Process Monitoring and Controls.

- **SO_2 Monitoring**

For continuous SO_2 monitoring, two in-situ Lear Siegler SM8100 microprocessor-based gas analyzers were used. As a backup analyzer, there was an extractive system Monitor Labs analyzer downstream of the reactor prior to the ESP. The Monitor Labs analyzer uses a diluting probe. One pair of SO_2 and O_2 analyzers was located after the air preheater before the activation reactor inlet. The second pair of analyzers was located after flue gas reheating and before the ESP inlet. The equipment utilizes a second derivative spectroscopic measurement technique, measuring the narrow band absorption of ultraviolet energy by SO_2 molecules. The measurement was performed on moist gas, and the output was corrected to standard temperature and pressure. The maximum measurement range of the equipment is 3,000 ppm and was calibrated for 0 to 3,000 ppm gas range monitoring. A detailed description of the SO_2 analyzer is presented in Appendix II.

Baseline flue gas SO_2 content ranged from 1,000 to 3,000 ppm, depending on coal quality. A typical value of 1,750 ppm corresponds to a 2.5% sulfur coal. With the LIFAC process in operation, the SO_2 content varied from approximately 260 to 900 ppm.

The physical conditions where SO_2 is monitored are demanding. In the sampling location before the activation reactor, the flue as contains a large quantity of solids, typically $1{,}800 \times 10^{-6}$ lb/SCF (29,000 mg/Nm^3). Downstream of the reactor, the relative humidity of the flue gas is typically 15%, and particulate loading is typically $2{,}500 \times 10^{-6}$ lb/SCF (40,000 mg/Nm^3). Loading after the reactor is higher due to limestone injection and recycle load.

The minimum temperature of the flue gas after the reactor was approximately 190°F. At these low temperatures, the Lear Siegler analyzer after the reactor was unreliable. Low temperatures cause ash buildup in the analyzer probe. The Monitor Labs SO_2 analyzer probe was equipped with a heater, and there was no ash buildup in that probe. Ash buildup would have partially absorbed SO_2, affecting the accuracy of the analyzer.

The ductwork was fitted with flanges for in-situ analyzers at both SO_2 measuring locations. Grid measurements for excess O_2 content in the flue gas were performed to verify the consistency of SO_2 and O_2 concentration patterns across the duct sections.

The reliability of the in-situ SO_2 analyzers were verified at times during the test program with a portable analyzer. The verification measurements were performed under baseline conditions at the beginning of baseline testing and during post-LIFAC tests. To verify the reliability of SO_2 measurements in the dusty conditions before the reactor and in the dusty and high moisture conditions after the reactor, verification measurements were also taken at the beginning of injection and activation testing. Verification was performed anytime during testing, if there was any doubt about the accuracy of the measurements. Indication of an unreliable reading was obtained, if the reading of the two stationary analyzers were not equal when the reactor was not in operation.

Throughout the test program, the analyzer readings were compared to the calculated sulfur dioxide content of the flue gas based on the sulfur content in the coal.

- **SO_3 Monitoring**

SO_3 emissions were monitored using EPA referenced Method 8. The description of this method is presented in Appendix II. SO_3 content was measured by subcontractors who performed ESP efficiency and performance measurements.

SO_3 emissions are typically estimated based on SO_2 emissions, since they are generally on the order of 0.5 to 1.0% of the flue gas SO_x content. In earlier studies at RP&L, baseline SO_3 emissions had been nearly 18 ppm during combustion of high sulfur coal (SO_2 concentration of 4,000 ppm).

Sorbent injection typically removes most or all of the SO_3 present in the flue gas, since SO_3 reacts rapidly with calcium oxide formed during the calcium carbonate decomposition reaction. SO_3 content was measured after the ESP. All SO_3 measurements were coupled with ESP performance testing.

- **O_2 Measurement**

O_2 measurements were performed using two in-situ Lear Siegler Dynatron 401 oxygen monitoring systems. The equipment uses a solid-state zirconium oxide fuel cell sensor to measure oxygen

content. Measurements were made on a moist, total sample basis using the Nernst relationship. The maximum measurement range of the equipment is from 0 to 25%. A detailed description of the O_2 analyzer is presented in Appendix II. O_2 analyzers were installed in the flue gas duct in the same locations as the SO_2 analyzers before and after the activation reactor.

Typical flue gas oxygen content after the boiler ranged from 2 to 5 volume %. The O_2 measurements were carried out under the same demanding conditions as the SO_2 measurements. However, a moist ash layer on the probe does not affect O_2 measurements as long as gas could penetrate the ash layer and filter.

The continuous oxygen measurement was used to correct the SO_2 measurements to a 6% O_2 content. Verification of the O_2 analyzers were used to measure concentration patterns of flue gas in selected monitoring locations.

- **NO Measurement**

For NO measurements, the same two in-situ Lear Siegler SM8100 analyzers were used. Duel exit slits on the monochromator of the analyzer allow two separate wave length (one corresponding to SO_2 and the other to NO) to impinge on the detector. Sequential time separation of the SO_2 and NO signals is accomplished with a small shutter in the monochromator. This shutter moves back and forth between the two beams at one minute intervals, providing one minute sequential SO_2 and NO concentration outputs.

Baseline NO content of the flue gas was measured during the baseline tests. During parametric testing, NO levels were measured by the subcontractor who performed stack testing as part of the Environmental Monitoring Program.

- **Particulate Emissions**

Particulate emissions were monitored using EPA reference Method 17 and Method 5. These methods are described in Appendix II.

The particulate load measuring locations were in the activation reactor and ESP inlet ducts. Emissions were measured in the duct breaching between the Unit No. 2 ID fan and the stack.

After the boiler and before the ESP, typical baseline particulate content of the flue gas was 650×10^{-6} lb/SCF. Limestone injection into the furnace increased this particulate load to between $1,300 \times 10^{-6}$ and $3,500 \times 10^{-6}$ lb/SCF. This variation depended mainly on the Ca/S molar ratio and limestone quality. In the activation reactor, the flue gas was humidified and a portion of the particulates are removed from the bottom of the reactor. After the reactor and before the ESP, the particulate content of the flue gas varied from 700×10^{-6} to $1,900 \times 10^{-6}$ lb/SCF without ash recycling. With ash recycling in operation, the particulate load increased to nearly $3,600 \times 10^{-6}$ lb/SCF.

The efficiency of the ESP increased with LIFAC in operation due to an increase in inlet solids loading with the same outlet emissions. During baseline operating conditions, the ESP efficiency varied from 98 to 98.4%. With the LIFAC process, it increased to 99.3%. Particulate emissions are typically 10-12 $\times 10^{-6}$ lb/SCF without LIFAC. Ash recycling may increase particulate load to 29×10^{-6} lb/SCF for short periods of time.

The purpose of the particulate measurements was to evaluate the efficiency of the ESP under different conditions and evaluate the ash separation capabilities of the activation reactor. A list of particulate measurements is shown in Table 6-H.

- **Opacity**

Opacity of the combined flue gas from Units No. 1 and No. 2 was measured by an in-situ instrument located in the stack. The opacity limit was 40% at the beginning of the project. During the demonstration, the RP&L received a new variance from IDEM which reduced the opacity to 30%. Opacity ranged from 8 to 25 throughout the demonstration (see Table 6-I) and was not a problem during the testing.

- **Moisture**

The measurement of flue gas moisture content was done simultaneously with particulate measurements. EPA reference Method 17 was used. This method is described in Appendix II.

- **ESP Performance**

ESP performance measurements were performed periodically during the test program. The final measurements were taken during long-term testing and reported in the long-term test report.

Ash resistivity was measured with an in-situ, point-to-plane resistivity probe system. The resistivity probe electrostatically collects a layer of ash, while flue gas conditions are maintained at constant values. Collection of the ash is accomplished by applying a high voltage across the point-to-plane electrodes in the probe. At a sufficient voltage, a highly compressed electric field or corona forms in the vicinity of the point (negative electrode) and ionizes gas molecules in this region. Dust particles in the gas are then charged by the negative ions formed in the corona and are attracted by electrical forces onto the collection plate (positive electrode). This method of ash collection is intended to simulate the behavior of ash in an actual ESP.

From earlier studies at RP&L, the average resistivity of fly ash was 5.9×10^{10} ohm-cm.

Ash resistivity measurements were performed during optimization tests, with and without the LIFAC process in operation. This testing was only done during 65-MW boiler loads. The measurement was taken in the ductwork before the ESP.

- **Boiler Performance**

To optimize limestone injection into the furnace and evaluate its impact on boiler performance, certain boiler operating conditions were measured. Measurements were performed using on-line instruments, which send information to the process control computer for handling and processing.

The coal feed rate measurement was used to calculate flue gas volume formed during coal combustion and the amount of sulfur entering the furnace. The coal feed rate was measured with a mechanical counter in the coal weigh feeder buckets. To verify the accuracy of these measurements, the weigh feeders were calibrated at the beginning of the baseline tests. During the tests, continuous verification was made by back-calculating the coal feed rate from the generated steam rate.

The quantity of combustion air and excess oxygen were measured for flue gas calculation purposes. The primary goal of the excess oxygen measurement was to attain optimum combustion conditions

in the furnace. Steam generation and spray water flow measurements were used as indicators of changes in boiler performance.

Two new thermocouples were installed in the furnace to measure injection level temperatures. During the baseline and demonstration tests, a suction pyrometer was used to verify these measurements.

Fouling in the economizer and air preheater was evaluated from pressure drop and temperature analyses at both locations. These measurements were continuously monitored during the demonstration.

The activation reactor also caused an increase in the system pressure drop. It was important to study this pressure drop effect on the ID fan's suction pressure, fan speed and motor current.

- **Temperature Measurement**

By measuring flue gas temperatures before, within, and after the activation reactor, pertinent conditions in the reactor were determined. Flue gas temperature before the reactor was measured by four temperature elements which accurately indicate any changes in the reactor inlet temperatures. Four temperature elements are installed at different elevations along the reactor. They were used as key indicators of humidification and evaporation within the reactor.

Reactor outlet temperature, measured at the bottom of the reactor, was used to control humidification water flow to the reactor. Optimum process efficiency is achieved when reactor outlet temperature is minimized and the amount of water evaporated is maximized. A handleable (dry) by-product is essential to the process.

Three units are installed in the reactor outlet ductwork to reheat the flue gas before the ESP. Each duct is furnished with a temperature element after these units to verify the efficiency of each individual reheating unit. The gas temperatures after the three outlet ducts are compared, and the lowest temperature is selected to control reheat steam flow to the heat exchangers.

- **Pressure Measurement**

Flue gas pressure was measured in the ductwork before the reactor, in the reactor, after each reheater, and before the ESP. The pressure measurements were used to determine LIFAC induced pressure changes in the system and predict ash buildups in the activation reactor. Individual pressure gauges measured the total pressure of the flue gas. Pressure differences were calculated by the PLCs.

- **Flow Measurement**

The average flue gas flow rate before the activation reactor was measured by pitot tube-type flow elements. Three flue gas traverse probes, consisting of multiple flow sensors mounted on self-averaging signal manifolds, are positioned in the duct on a equal area basis. The flow element system is furnished with an instrument air purge system to protect the internal mechanism of the transmitter from the migration of airborne particulates.

The flue gas flow measurement gives an early signal to the humidification water control system to increase or decrease the water flow to match an increase or decrease in the flue gas flow.

Humidification water flow was measured by an orifice plate and controlled by a motor-operated control valve. The quantity of humidification water sprayed into the activation reactor was controlled by flue gas temperature measurements after the reactor. Air flow to the dual fluid nozzles was controlled by regulating the air pressure with a motor-operated control valve.

Steam flow to the reheaters was controlled by the flue gas temperature ahead of the ESP. A motor-operated control valve controlled steam flow. The reheating steam flow was verified by calculations from the flue gas temperature increase measurement or by counting the discharge cycles of the condensate return tank.

- **Limestone and Recycle Ash Feed Measurements**

The limestone and recycle ash feed measurements were made in three locations for each material: silo measurements before the silo shut-off valves, feeder measurements from the silo discharge opening to pneumatic conveying line, and pneumatic conveying system measurements.

- **Silo Measurements**

The quantity of limestone in the feed and storage was determined by weigh cell measurements. Both the feed and storage silos were equipped with two weigh cells that measured the load bearing columns' compression and calculated the weight. The storage capacity of the feed silo was 270,000 lb, and that of the storage silo is 530,000 lb. The ESP ash surge tank has two weigh cells and a storage capacity of 20,000 lb.

The feed silo has two level switches for low and high levels. The limestone storage silo has three level switches for low, high, and high/high levels. The ESP ash surge tank has one high level switch.

- ***Feeder Measurements***

Limestone mass flow to the boiler injection nozzles was measured by a gravimetric weigh feeder system which consists of two screws and a rotary feeder arranged in series. The rotary feeder controlled the mass flow from the silo to the first screw feeder, and both screw feeders are connected to the same variable speed controller. The screw feeder is used to ensure smooth flow of limestone to the second weigh screw which measures the mass flow. The weight screw is mounted on load cells, operates at a constant speed, and determines the feed rate by measuring the weight of the screw multiplied by screw speed. The limestone flow measurement was verified by comparing it to limestone feed silo weight measurements.

The ESP ash recycling ratio was measured and adjusted by two ESP hopper rotary feeders. The rotary feeders are installed in the bottom of the first two parallel ESP hoppers and feed ESP ash to the pneumatic conveying system for recycle. Ash flow for recycle from the ESP ash surge tank was measured and controlled by a variable speed drive screw feeders under the tank. Reactor ash flow to the recycling system from the reactor bottom conveyor was measured and controlled by a variable speed drive screw feeder.

- ***Pneumatic Conveying Measurements***

The LIFAC process contains four pneumatic transport air blowers: two for the limestone injection into the furnace and two for ash recycling. In addition, the limestone injection system was furnished with a booster air blower to increase the initial velocity of the limestone particles injected into the

boiler. The limestone transport air, booster air, ESP ash recycling air, and recycling mix transport air blowers were operated at constant speeds.

Limestone transport air flow was measured using an orifice plate. The pressure and temperature of the air was also monitored. The transport air flow (1,060 SCFM) was controlled with a pneumatic control valve. Booster air flow rate was approximately 6,360 SCFM.

In addition, the air pressure from the ESP ash recycling air flow was measured. By measuring pressure changes in these flows, potential blockages in the piping and nozzles could be identified.

4.2.2 Sampling and Analysis

Coal, limestone, and ash samples were taken manually during the demonstration. Aqueous samples were also monitored as part of the Environmental Monitoring Plan (EMP). Sampling locations are shown in Figure 6-1, and solid sampling methods are presented in Table 4-A.

TABLE 4-A

SOLIDS SAMPLING FREQUENCY, METHOD AND LOCATION

SAMPLE	SAMPLING FREQUENCY	METHOD	LOCATION
Coal	Once per hour or three times per hour during special tests	Grab sample	Coal pulverizer feeder
Limestone	Occasionally	Grab sample	Feeding silo discharge
ESP ash	Occasionally	Grab sample	ESP hopper
Reactor ash	Occasionally	Grab sample	Truck hopper at reactor ash discharge
Economizer ash	During EMP sampling	Grab sample	Air preheater hoppers
Bottom Boiler ash	During EMP sampling	Grab sample	Bottom ash dewatering bin

Coal samples were taken from the feeder belt before each of the pulverizers. Coal was grabbed once per hour from each feeder, and a composite sample was formed for a three-hour period. Samples were shipped in double layer plastic bags to a laboratory for analysis. Special one-hour coal samples were collected during stack testing. One-hour samples were also taken during baseline and calibration periods.

Since the sulfur content of the feed coal dictates the limestone feed rate, it was the most significant coal quality parameter studied during the demonstration. Sulfur analyses were also used to calculate the baseline SO_2 content of the flue gas. During the demonstration, feed coal sulfur content ranged between 1.4 and 2.8%.

Most analyses were performed by one laboratory. However, other labs were often utilized to ensure reliable results. The scope of coal analyses included short proximate, ultimate, and mineral analyses. See Table 6-9 for a summary of the typical coal analysis results. (Actual coal analysis are provided in Final Report Volume 1: Public Design.[7])

Mineral analysis included the components listed below:

- Silicon dioxide (SiO_2)
- Aluminum oxide (Al_2O_3)
- Titanium dioxide (TiO_2)
- Calcium oxide (CaO)
- Potassium oxide (K_2O)
- Magnesium oxide (MgO)
- Sodium oxide (Na_2O)
- Phosphorous pentoxide (P_2O_5)
- Ferric oxide (Fe_2O_3)
- Sulfur trioxide (SO_3)
- Base/Acid ratio
- Slag viscosity
- Fouling index
- Slagging index
- Silica value
- % alkali as (Na_2O)

The limestone suppler was obligated to monitor quality once per hour, and the data were made available to the LIFAC test team. Therefore, only periodic verification samples and analyses were performed. The most significant limestone parameters were the calcium carbonate ($CaCO_3$) content and particle size. Additional information was also provided by the supplier, as shown in Table 4-B.

TABLE 4-B

LIMESTONE ANALYSIS COMPONENTS
OF THE FINE QUALITY (80%<325 MESH) LIMESTONE

COMPONENT	TYPICAL CONTENT, WEIGHT %
Na_2O	0.12
MgO	1.50
Al_2O_3	0.68
SiO_2	2.87
SO_3	0.14
K_2O	0.16
CaO	51.86
TiO_2	0.05
Fe_2O_3	0.35
LOI	42.27
TOTAL	**100.00**
Calcium and Magnesium Carbonates	
$CaCO_3$	92.6
$MgCO_3$	3.15

Ash samples were grabbed from four different locations during the demonstration:

- Economizer Ash - Collected from the economizer hoppers.
- ESP Ash - Collected from the ESP hoppers; EPS ash samples were taken from the first two parallel hoppers.
- Reactor Ash - Collected from the activation reactor bottom conveyor.
- Boiler Bottom Ash - Collected from dewatering bin (Unit No. 2 only).

During baseline and post-LIFAC tests, fly ash samples were collected every eight hours. During parametric and long-term test periods, ESP and reactor ash samples were collected during each individual test.

During the tests in which particulate measurements were performed, all ash samples were analyzed for calcium compounds, elemental analysis, and particle size distribution. A resistivity analysis of ESP fly ash was performed on samples collected during on-line ash resistivity measurements.

4.2.3 Data Analysis Methodology

Test data were acquired in many different ways, depending on the type of data and required procedures. Test data for LIFAC were, for the most part, recorded automatically by the process control computer. Necessary additional information was recorded manually in a logbook and log sheets, either in the control room or in the plant.

Data were recorded from the boiler, activation reactor, electrostatic precipitator, and all input and output streams. A test coordinator kept the log book where all pertinent test information was recorded, such as injection settings and recycling methods.

Particulate emission measurement data were recorded on forms (see EPA Method 17 in Appendix II). All other manually measured data were recorded on specific forms. Samples requiring laboratory analysis were recorded on laboratory supplied forms.

The process control system with process monitoring system (PMS) software was used to collect and handle the automatically measured data. The process computers read values from the instrumentation every three seconds, typically calculated three-minute averages, and saved the data on the computer's hard drive. Data from the PMS were converted into a format that could be imported to a Lotus 1-2-3- spreadsheet. Data were transferred via disk to a personal computer to perform on-site data processing.

The Ca/S molar ratio is a factor which reflects the efficiency and economy of the LIFAC desulfurization process. It is calculated using equation (1). In equation (1) limestone mass flow is measured by the limestone feeder. The $CaCO_3$ content of the limestone is obtained from the limestone analyses and is constant if the limestone comes from the same location. Coal flow is measured at the coal feeder, and the coal sulfur content is obtained from the coal analysis. Estimated values of these parameters were stored in the process monitoring computer, but the corrected values were used for adjustments during data handling procedures on a spreadsheet program.

Ca/S molar ratio:

$$Ca/S = \frac{m(ls) \times p(ls) \times 100.09}{m(coal) \times s \times 32.09} \quad (1)$$

where:

Ca/S	=	the ratio of the number of moles of calcium injected into the furnace to the number of moles of sulfur burned in the furnace
m(ls)	=	limestone mass flow (lb/min)
p(ls)	=	limestone $CaCO_3$ content (weight %)
100.09	=	molecular weight of $CaCO_3$
m(coal)	=	coal flow (lb/min)
s(coal)	=	sulfur content of coal (lbs/lbs)
32.09	=	atomic weight of sulfur

Variations of boiler operating conditions caused variations in flue gas composition, such as SO_2 and O_2 levels. Use of different SO_2 analyzers also cause variations in the measured results. In order to compare different SO_2 measurements, the measured values were corrected to a constant O_2 content and moisture level. Sulfur dioxide in the flue gas is corrected to 6% O_2 content. The equations are shown below; equation (2) corrects for O_2, and equation (3) corrects for moisture.

$$SO_2(6\%\ O_2) = SO_2(m)\ \frac{21 - 6}{221 - O_2(m)} \quad (2)$$

where:

$SO_2(6\%\ O_2)$	=	the SO_2 content in the flue gas corrected to 6% O_2 content, ppm
$SO_2(m)$	=	the SO_2 content of the flue gas at the actual flue gas O_2 content, ppm
21	=	0_2 content in the air, %
$O_2(m)$	=	measured O_2 content in the flue gas, %

Corrections for moisture were performed, because SO_2 before and after the reactor were measured in moist flue gas conditions.

$$SO_2(cor) = SO_2(6\%\ O_2)\ \frac{[100 - C(wb)]}{[100 - C(wr)]} \quad (3)$$

Where

SO_2(cor)	=	O_2 content corrected to the same moisture as in the boiler, ppm
SO_2 (6% O_2)	=	SO_2 content of the flue gas, corrected to 6%O_2 content, ppm
C(wb)	=	flue gas moisture in the boiler (estimated to be 8.5 vol %, because no measurement is available)
C(wr)	=	flue gas moisture in the reactor (estimated to be 15 vol %, because no measurement is available)

Total SO_2 reduction is calculated from equation (4), which requires baseline SO_2 content of the flue gas. Since this value was not measured continuously, it was calculated with a computer program. Occasionally, LIFAC was shut down for short periods for baseline SO_2 measurements.

SO_2 reduction calculation:

$$XSO_2(tot) = \frac{100\,[C(base) - C(r)]}{C(base)} \tag{4}$$

Where

XSO_2(tot)	=	total SO_2 reduction, %
C(base)	=	baseline SO_2 level, ppm
C(r)	=	SO_2 content after the reactor, ppm

4.3 Operability and Reliability

4.3.1 Critical Component Failure Analysis

A detailed discussion of critical component failures which occurred during operation is important to the commercialization of the LIFAC process. Since this was a demonstration project and utilized equipment from an earlier demonstration, component failures were to be expected. Critical component failures are those which upset normal process operation, causing prolonged downtime or decrease process efficiency. The incidents which occurred during this demonstration are thoroughly addressed in this section. Many failures can be avoided in future installations through an understanding of these experiences. Component failure analyses are broken down into the following five process areas:

- Limestone Storage and Handling Area

- Boiler Injection Area
- Activation Reactor Area
- ESP Ash Recycle Area
- Process Monitoring and Control System

Limestone Handling and Storage Area

The limestone feeding silo used by EER did not have sufficient capacity for the LIFAC demonstration. However, the project utilized the silo and its limestone feeding equipment in conjunction with a newly installed storage silo. The limestone screw feeder was part of the remaining EER equipment. During the long-term test period, the screw feeder below the feeding silo stopped, causing process downtime. The manufacturer's representative identified three problems with the feeder: 1) a shorted silicon controlled rectifier (SCR) in the west motor control center; 2) a faulty tachometer near the feeder; and 3) a sticking rotor cuff under the weigh feeder. These problems were corrected on-site, and no more incidents occurred with the limestone feed system.

The feeding silo fluidizers were not effective when the limestone in the feeding silo dropped below 50,000 lb. These fluidizers were also part of the remaining EER equipment. The poor condition of those fluidizers results in unstable limestone flow. In most cases, limestone flow was restored by physically hammering the bottom of the silo. The problem was avoided by maintaining a weight greater than 50,000 lb in the feeding silo. To solve the problem, the fluidizing elements would need to be replaced.

Boiler Injection Area

Process components involved in injection limestone into the boiler consisted of a two-stage splitter, injection nozzles, and piping. No critical component failures were experienced in this area. Rubber flex-hose was used between the splitters and the injection nozzles. These hoses needed to be changed periodically due to the abrasiveness of the sorbent. No downtime was experienced during replacement.

ID Fan Speed Control System

Several problems were experienced with the variable frequency drive (VFD) which was installed as part of the LIFAC system. Since the induced draft (ID) fan operated at full capacity without LIFAC, the VFD interfaced with the fan and increased its capacity. A VFD failure caused the boiler to trip. Unit No. 2 tripped five times as a result of VFD difficulties during the LIFAC demonstration. Consequently many hours of process testing were interrupted. Manufacturer's representative(s) visited the site after each incident, isolated deficiencies, corrected them, and tested the unit prior to each startup. The unit is currently operational, however, it was out of services for most of the demonstration.

Activation Reactor Area

Some problems occurred with the proprietary spray nozzle assemblies located in the humidification nozzle headers. The impact bolts which are situated within the nozzle sheared during operation. These specially designed bolts are composed of two dissimilar metals (stainless steel and tungsten carbide) and are designed to have increased hardness for their application. However, after prolonged exposure to high-pressure water, the tungsten carbide tip separated from the bolt and hindered the atomization effect of the nozzle. After three impact bolts sheared in succession, the test team decided to replace them all with stainless steel bolts. The river water used for humidification at RP&L is not abrasive enough to require a tungsten carbide tip. The nozzle supplier admitted that this failure was due to a manufacturing problem.

The original humidification water control valve failed to accurately respond to control signals from the process monitoring system. The actuator motor and V-ball valve were insufficient for this application. The unit was replaced by a globe valve coupled to a pneumatic actuator.

The mechanical limit switches installed on the ash disposal double dump valves failed to perform after hours of operation in the dirty environment. If a switch failed, an alarm signal was sent to the PMS, and the conveyors stopped. Periodically, the switches needed to be disassembled, cleaned, and reassembled. The switches were eventually replaced by proximity limit switches, which have no mechanical parts and are more suitable for the dusty environment.

The three air compressors, which were also remnants of EER's LIMB project, began to overheat and trip during LIFAC operation when the weather became hot. The compressors were originally designed for outdoor applications and became excessively hot after the LIFAC project enclosed them in the new limestone storage and handling building. The compressor intercoolers were also more susceptible to fouling, since the building became dusty at times. Manufacturer's representatives came to the site to clean the intercoolers and remove the covers which were necessary for outdoor use only. The compressors operated at normal temperatures for the remainder of the project.

ESP Ash Recycle Area

The recycle rotary feeder located under ESP hopper No. 7 ceased operating during parametric testing. After running a series of mechanical diagnostics on the feeder, it was determined that the problem existed in its variable speed controller. A replacement transformer inverter was ordered after consultation with a manufacturer's representative. However, this repair limited the testing of the recycle rate parameter, because the rotary feeder would only operate at a manufacturer for repair. Some surface mount resistors were replaced at the factory, and the unit was sent back to be reinstalled. No further problems were encountered with the controller.

Process Monitoring and Control System

A grounding problem was discovered by a manufacturer's representative in the east motor control center I/O panel. The fault was recognized by LIFAC personnel when the process values, which are transmitted and controlled by this I/O rack, remained unchanged for prolonged periods. The I/O rack needed to be reset. Prior to resetting the rack, the test team had to bring the entire process down, since vital equipment (i.e, conveyors, water pump) was affected when the unit was reset.

4.3.2 Impact of LIFAC on Plant Systems

Power plants are often concerned about how a pollution control system will affect existing equipment and procedures at their facility. Since the LIFAC system is installed between the air preheater and the ESP, it does have an impact on operations.

Boiler

Limestone is injected into the upper region of the furnace prior to the superheater. This injection location creates a potential for increased superheater and wall tube fouling with certain limestone qualities and injection rates. Fine limestone is typically injected at a rate of 130 lb/min when Unit No. 2 load is 60 MW. This high rate of injection is attributed to the 950 lb/min of high sulfur coal which is consumed by Unit No. 2 at this load. Under these conditions, sootblowing frequency needed to be increased from every six hours to every 4.5 hours, since this higher rate cause the sootblowing sequence to be initiated sooner when limestone particles coated the superheater tubes and insulated them from desuperheat spray water effects. No fouling effect attributed to limestone injection were discovered on the boiler walls, in the economizer, or air preheater. The economizer is not equipped with sootblowing, and the frequency in the air preheater remained unchanged. Also, the boiler's wet bottom ash pulling system continued to operate normally.

The LIFAC flue gas reheat system utilizes medium pressure steam from the host facility. This steam is extracted from a water heater in the boiler house and condensate is returned to a deaerator tank. The steam reheat system did not have any noticeable effect on normal operations. Additional reheating was implemented during the course of the demonstration. A small reheat duct was installed between the economizer inlet and reactor outlet ductwork. This bypass contributed an additional 30°F of reheat to the exiting flue gas The flue gas reheat duct bypassed approximately 5% of the total gas volume generated in the boiler and had no noticeable effect on plant operations.

ID Fan

The LIFAC process introduces additional air and water into the power plant systems downstream of the reactor, causing occasional fluctuations in flue gas density, pressure and particulate loading. These variations have the potential to cause excessive stress on the ID fan. Pressure drop through the LIFAC reactor was 4.5 in. wg. If the steam reheaters were replaced with a hot flue gas reheat system, the pressure drop is predicted to be close to 2.0 in. wg.

ESP

The low temperature, high humidity flue gas conditions created by the process caused opacity excursions during the transient periods associated with startup and shutdown. These excursions were

directly related to the emission of particles from the ESP. Excursions were observed even when only humidification was engaged to drop the gas temperature with no change in ESP inlet mass or particle size distribution. The maximum opacity limit was 40% during the demonstration, and startup of the LIFAC process had caused opacity levels exceed acceptable limits. Southern Research Institute (SRI) was contracted under EPRI to assess the causes and suggest solutions to the opacity problems.

SRI determined that the electrical operating conditions of the ESP were satisfactory and there was no indication that high opacity occurred because the fundamental collection performance of the ESP was degraded. The source of the excursions appeared to be a release of residual dust from the ESP collection plates. At any given time there are tens of thousands of pounds of fly ash in the collected dust layer, and the release of only a minute quantity can create the observed excursions.

The low temperature, high humidity flue gas conditions generated by the LIFAC process cause re-entrainment of the fly ash on the ESP collection plates. Reentrainment results from a reduction in resistivity and tensile strength of the fly ash. The electrical clamping force of the ESP is a function of resistivity and current density. When the flue gas is humidified, current density increases only slightly and ash resistivity significantly decreases resulting in a 2 to 3 order of magnitude drop in the electrical clamping force. SRI concluded that the most likely cause of the high opacity excursions is a combination of reduced holding force and poor gas distribution in the ESP.

The following recommendations were made by the SRI/EPRI evaluation in regards to improving the operation of the Whitewater Valley Unit No. 2 ESP during LIFAC operations:

- Increase the amount of reheat to decrease the relative humidity of the flue gas until acceptable ESP performance is achieved.

- Upgrade the ESP so that the gas flow distribution is uniform.

- Allow sufficient time for transient conditions to pass by gradually decreasing ESP inlet temperature until design operating conditions are reached.

5.0 TECHNICAL PERFORMANCE

5.1 Effects of Operating Parameters on Results

Parametric testing results summarize data collected during the numerous test periods between January and December of 1993. Many preliminary, short duration tests were performed between October 1992 and January 1993. The majority of this data can be found in the Parametric Test Report which is provided in Appendix IV. It provided valuable preliminary information to assess the effects of parameters on process performance and their impacts on the host facility. Parametric test results encompass data from several different test periods. Data points were grouped according to corresponding operating conditions for ease of comparison and evaluation.

Since many conditions and equipment functions affected SO_2 reduction, some test results were deemed unreliable. Most tests were performed during constant boiler load. However, load was unsteady at times due to fluctuating power demands. Coal quality varied slightly during operations, although an attempt was made to burn constant quality coal for most of the demonstration. The sootblowing interval decreased from 6 to 4.5 hours with LIFAC engaged. SO_2 analyzer readings were unreliable during sootblowing.

Deviations from the original test plan were necessary, since the project experienced several operational interruptions. Unacceptable opacity excursions occurred during LIFAC startup periods. Hence, more time was spent on ESP evaluations than originally planned. Equipment and instrument malfunctions occurred periodically, and SO_2 analyzers failed to give reliable readings at times. Process parameters and their respective test ranges are shown in Table 5-A.

Limestone Injection Nozzle Settings

There are twelve limestone injection nozzles at two elevations on the boiler. The six upper nozzles were used during most of the testing. Several nozzle combinations were tested with the full range of boiler loads. Most tests were performed to find the optimum injection locations to maximize SO_2 reduction in the furnace. During most of the injection tests, the main flue gas bypass damper was 10 to 15% open, because the ID fan could not handle full gas flow through the reactor with the VFD not in operation.

TABLE 5-A

PARAMETERS AND RANGES TESTED

PARAMETER	TEST RANGE
Coal Quality	Sulfur content varied from 1.4% to 2.8%. The average sulfur content was approximately 2.25%.
Limestone Particle Size	Two particle sizes were tested. Coarse (85% <200 mesh) quality provide to be less effective, and fine (80% <325 mesh) limestone was selected as the optimum sorbent.
Boiler Load	Boiler load ranged from 40 MW to 62 MW (nameplate size is 60 MW).
Ca/S	Molar ratios between 1.0 to 3.0 were evaluated.
High Level Injection	Ports 1, 2, 3, 4, 5, 6 in use
Lower Level Injection	Ports 7, 8, 9, 10, 11, 12 in use
Right Side Injection	Ports 4, 5, 6, 10, 11, 12 in use
Left Side Injection	Ports 1, 2, 3, 7, 8, 9 in use
Middle Position Injection	Ports 2, 3, 4, 5, 10, 11 in use
Injection Nozzle Angle	Although injection nozzles can be adjusted ± 15° from horizontal, but only the horizontal position was tested.
Booster Air	Not tested.
ΔT	The temperature difference between the reactor flue gas outlet temperature and the adiabatic saturation temperature of flue gas. Temperatures between 4 and 11°F above adiabatic saturation were tested.
Water Droplet Size and Nozzle Position	Droplet size and nozzle position were tested during the early stages of parametric testing. Droplet size is propriertay information.
Ash Recycle Ratio	The recycle ratio is the mass flow rate of recycled ash versus the mass flow rate of boiler ash. Ash recycling ratio varied from zero to 2. The maximum ratio equates to nearly 4-5 tons per hour.

Only slight changes in process efficiency were observed during the testing of various injection nozzle combinations. With low boiler loads (40-50 MW), the lower six nozzles performed nearly 5 percentage points better than the high level nozzles. If limestone is injected in the lower level at low loads, there is ample residence time for limestone reactions at an ideal temperature. The upper injection level proved to be 2 to 3 percentage points more efficient during high loads. This is probably due to extreme temperatures at the lower level. Figure 5-2 shows the results of injection nozzle combination testing. Although Figure 5-2 does not show it, the data were interpreted that SO_2 reduction rates seemed to improve when the upper injection level was utilized. Figure 5-3 shows SO_2 reduction in the furnace only.

Injection nozzle heads can be tilted 15° vertically. However, all tests were performed with the nozzles in a horizontal position. This was based on recommendation by Tampella per their experience on previous projects.

Booster air for limestone injection was tested only briefly because it increased loading on the ID fan, which was already operating at full capacity. The secondary air fan also exceeded noise levels in the boilerhouse. Additional noise protection or soundproofing would have been required to continue to operate the fan. Therefore, no results of booster air's effect on SO_2 capture were obtained

Limestone Particle Size

Two different particle sizes of limestone were tested. One was a coarse ground (85% < 200 mesh) and the other was a fine ground (85% < 325 mesh). Both had high calcium carbonate content, ranging from 90 to 95%. These sedimentary limestones were shipped from two different quarries. The coarse limestone was shipped from southwest Indiana, and the fine came from northern Kentucky.

Fine limestone proved to significantly improve process efficiency as compared to the coarse limestone. Figures 5-4 and 5-5 show the results of limestone particle size testing at both high and low boiler loads. Fine limestone increases process efficiency 15 to 20 percentage points and was injected during most of the parametric and long-term tests. Despite the scatter of results, the trend is evident; smaller particle size provides more active surface area for absorption reactions to take place.

Figure 5-2 Total SO_2 Removal at High and Low Injection Levels

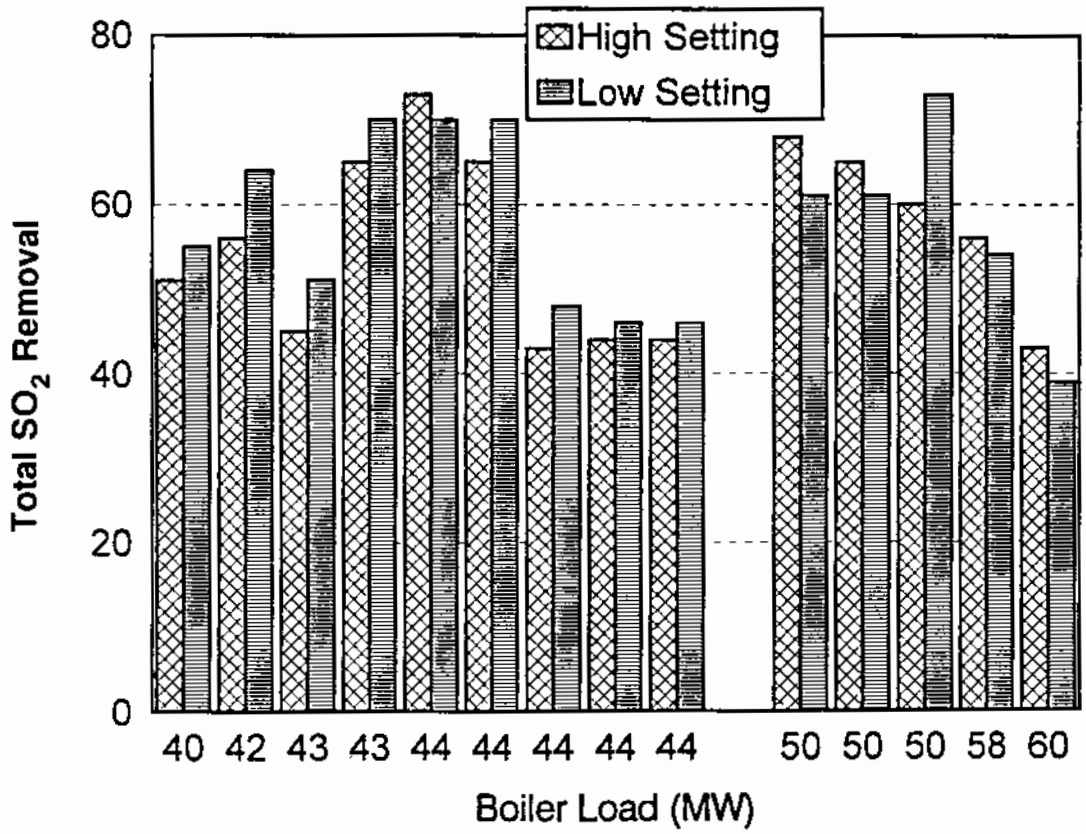


Figure 5-3 Effects of Limestone Injection Level on Furnace SO_2 Removal

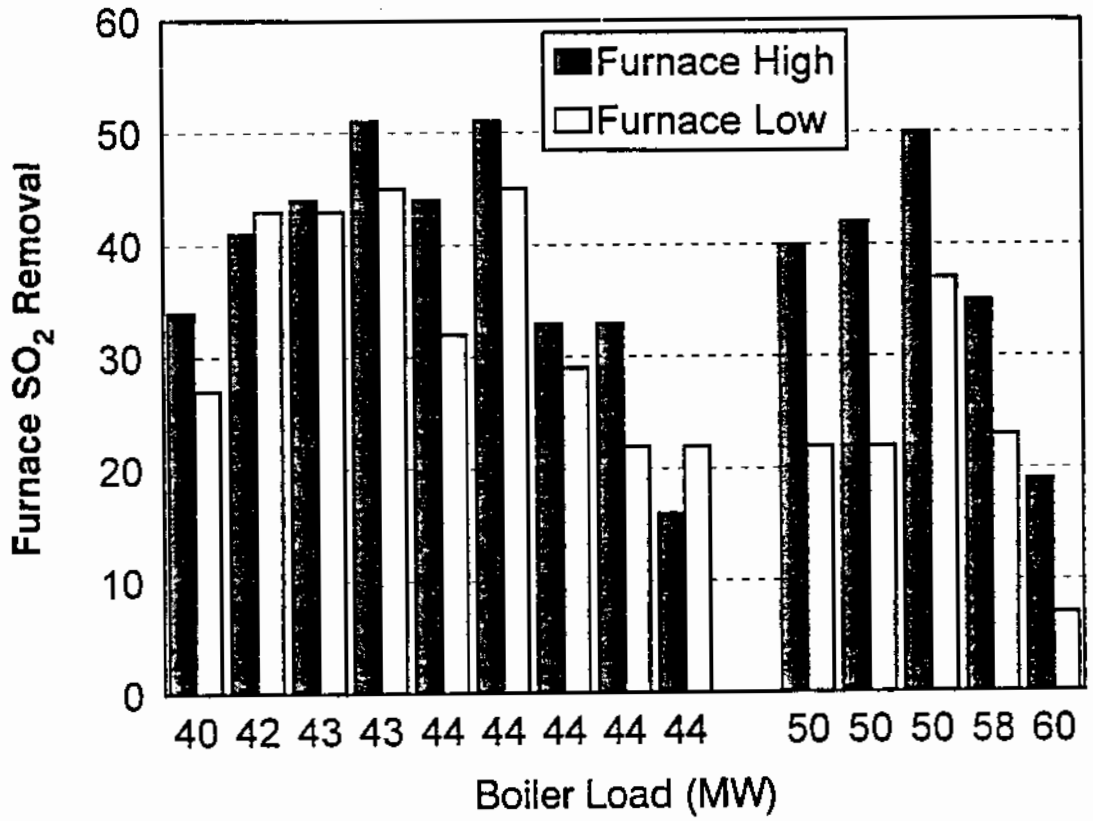


Figure 5-4 *Effect of Limestone Particle Size on Total SO_2 Removal During High Boiler Load (59-61 MW)*

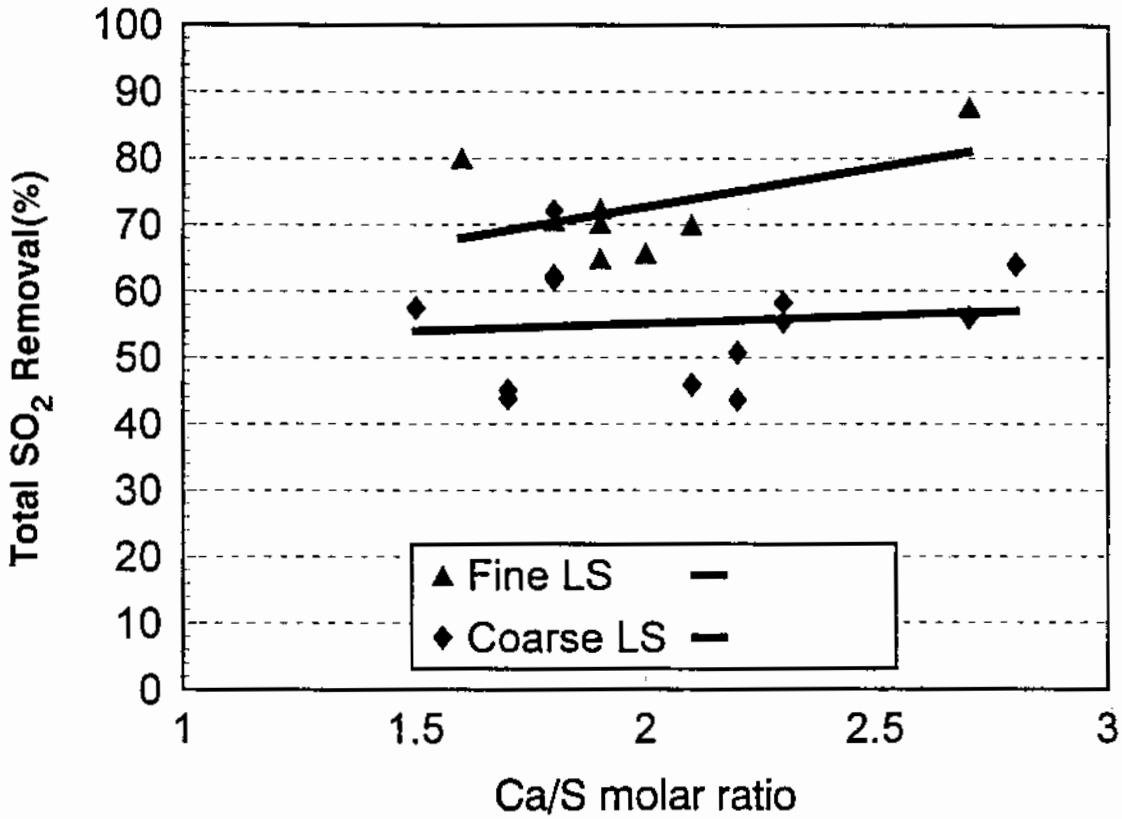
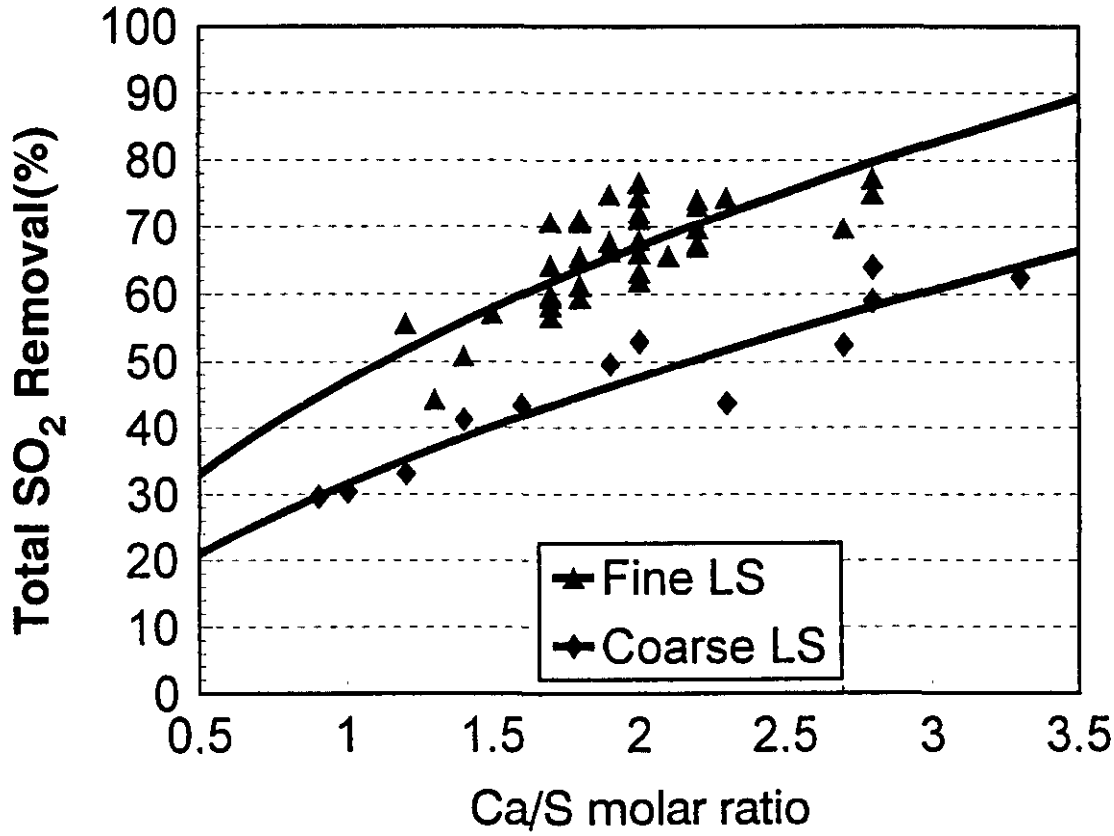


Figure 5-5 *Effect of Limestone Particle Size on SO_2 Removal During Lower Boiler Load (41-44 MW)*



Boiler Load

The nameplate load of Unit No. 2 is 60 MW. However, the unit can be operated as high as 65 MW. During parametric testing, the load ranged from 40 to 62 MW. Most parametric tests were performed at a high boiler load due to high power demands.

Boiler load was decreased during weekend and overnight. Due to the unavailability of low load results at similar operating conditions, a comparison of boiler load effects on SO_2 was not possible. Load was held constant during parameter measurements by adjusting the non-LIFAC boiler (Unit No. 1) to suit power demand.

Boiler load effect on SO_2 reduction was not as significant as expected. Unit No. 2 is small for its capacity and burns hot. Low loads should have improved efficiency. Instead, the difference between high and low load efficiency is unclear. SO_2 capture in furnace and reactor at a range of boiler loads are shown in Figures 5-6 and 5-7.

Coal Quality

Test coal was shipped from three locations in three different states. Small amounts of low sulfur coal came from Kentucky and Ohio. The majority of the test coal (high sulfur) was shipped from southwest Indiana.

Sulfur content of the coal often changed during testing and cause slight inaccuracies in the results. Coal samples were grabbed once per hour from each of the three feeders to get representative samples of the test coal.

Sulfur content varied between 1.4 and 2.8 wt%. It was not possible to evaluate the effect of coal sulfur content on SO_2 capture, since other conditions were not stable during testing. Figure 5-8 shows the sulfur content of the coal during the parametric tests.

Ca/S Molar Ratio

Limestone flow rate to the furnace is set by the desired Ca/S molar ratio. The ratio is calculated from the coal sulfur content and limestone quality, along with their respective flow rates. The

Figure 5-6 SO_2 Removal in the Furnace Versus Boiler Load

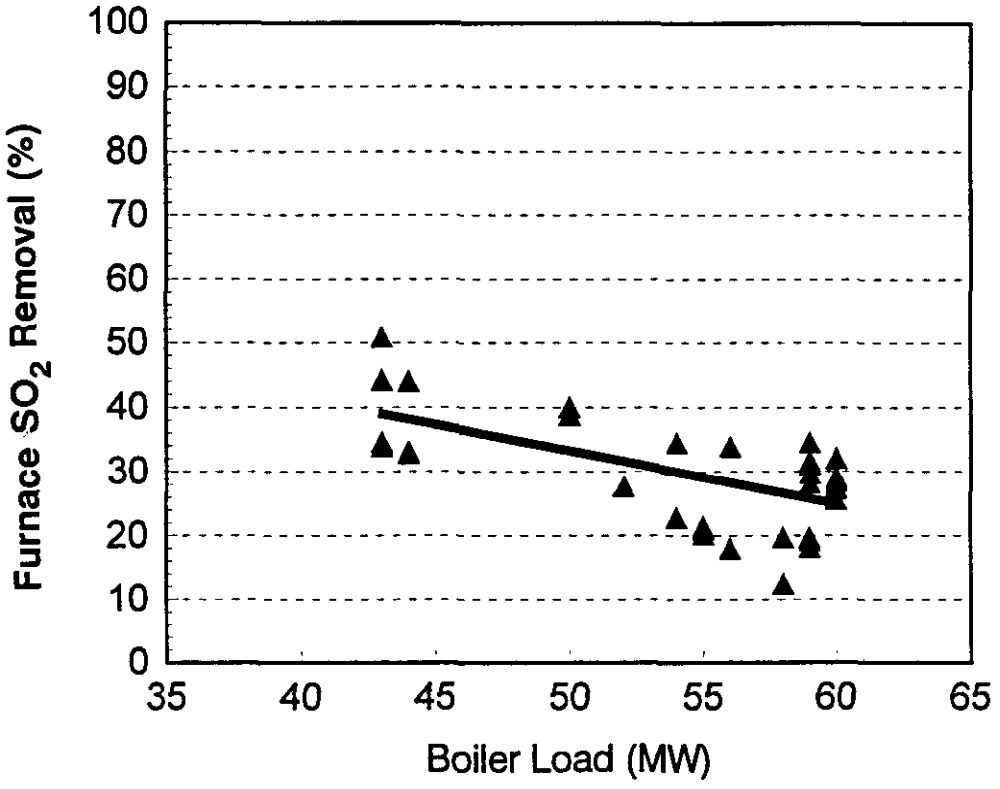


Figure 5-7 Total SO_2 Removal Versus Boiler Load

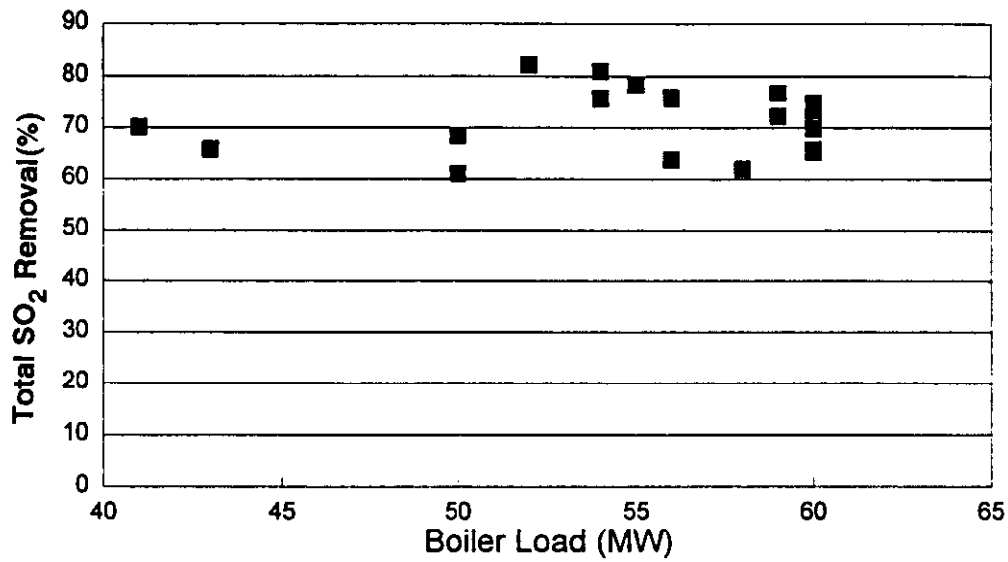
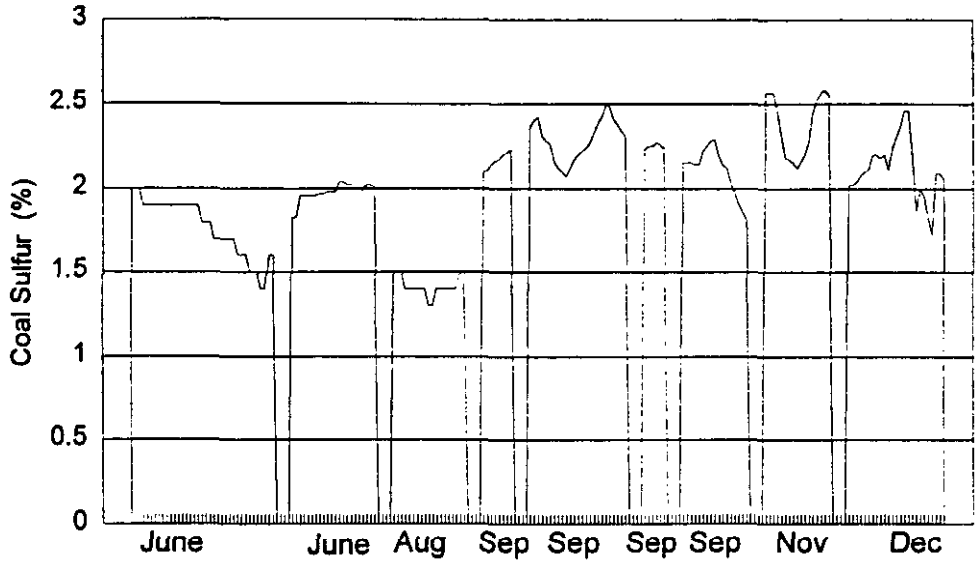


Figure 5-8 Sulfur Content of the Combustion Coal During Parametric Testing



calculation is described in Section 4.2.3. Ca/S ratios ranged from 1.0 to 3.0 during the parametric test period. The Ca/S molar ratio was maintained at 2.0 for most of the testing. Baseline coal sulfur content was determined from the SO_2 concentration in the flue gas. This coal sulfur estimate was entered into the PMS and remained unchanged until the next baseline adjustment and calibration period. The actual ratio varied according to boiler load and was, at times, inaccurate if the sulfur content of the coal changed during operation. The actual ratio was later calculated from the lab analyses of the coal and incorporated into data analysis.

Ca/S molar ratio parameter testing was performed with a range of boiler loads, two different limestone particle sizes, and with and without recycle. The effects of limestone particle size and ash recycle are reported in the respective sections. Under all test conditions, SO_2 capture increased as the Ca/S molar ratio increased. Figure 5-9 illustrates the effect of the Ca/S ratio on overall SO_2 removal at reduced boiler loads. Previous LIFAC testing showed that Ca/S ratios higher than 3.0 slightly improved process efficiency but increased costs. The optimum ratio was determined to be between 1.5 and 2.5, which resulted in 60 to 75% SO_2 capture. Figure 5-10 shows the effect of Ca/S molar ratio on total SO_2 removal at higher boiler loads, using two different limestone particle sizes.

Activation Reactor Outlet Temperature

Total SO_2 capture improves as reactor bottom temperature approaches the flue gas saturation temperature. It is necessary to keep the reactor outlet temperature above the flue gas saturation temperature (~126°F). Otherwise, moisture content of the reactor bottom ash will be too high for the ash handling system, which is designed to transport only dry ash. With the improved reactor design at RP&L, a closer approach to saturation was possible. The design value was 11°F above saturation. Parametric tests were performed with temperatures as low as 3°F above saturation. Even at these low temperatures the reactor bottom ash was dry and easy to handle. Typical bottom ash had a moisture content of between 5 and 20%. Figure 5-11 shows the effect of reactor bottom temperature on process efficiency.

Humidification Nozzle Arrangement

It is possible to move each of the seven nozzle headers up and down. In the beginning of the tests, the headers were at their lowest position. During the tests, the headers were lifted to a higher position. There was a danger that ash would accumulate on the reactor's internal sealing if the

Figure 5-9 Total SO_2 Removal Versus Ca/S Molar Ratio During Low Boiler Load

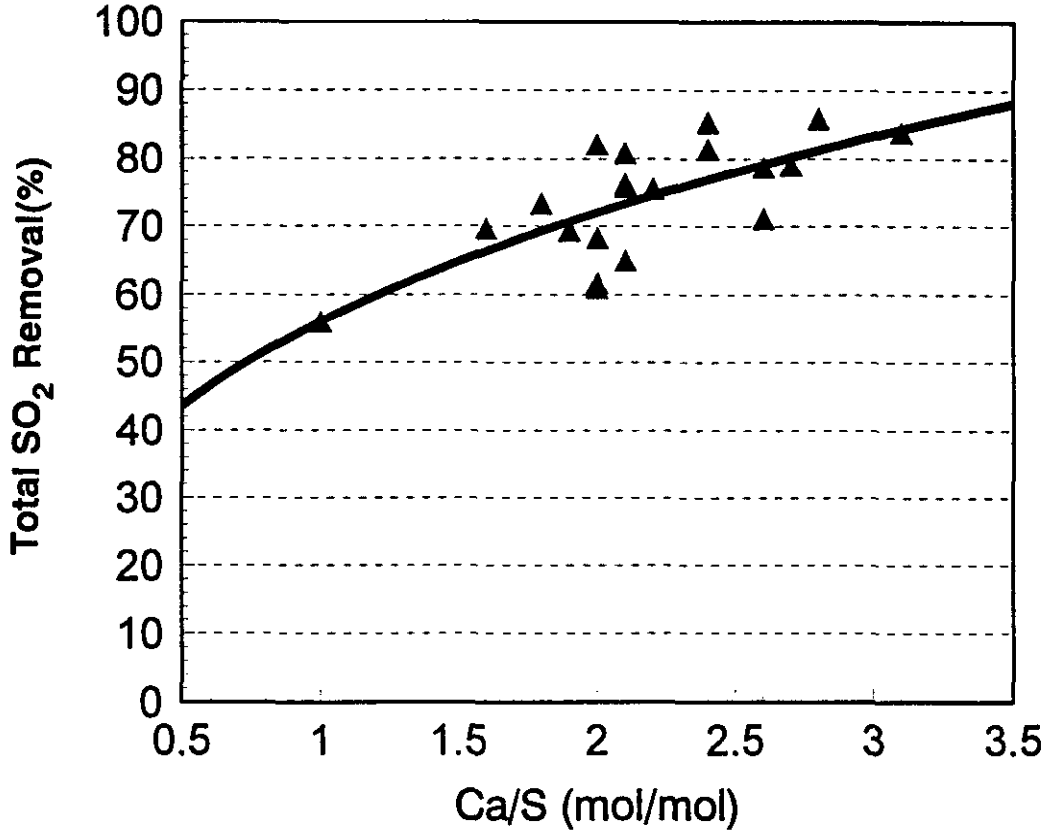


Figure 5-10 Total SO_2 Removal Versus Ca/S Molar Ratio During High Boiler Load

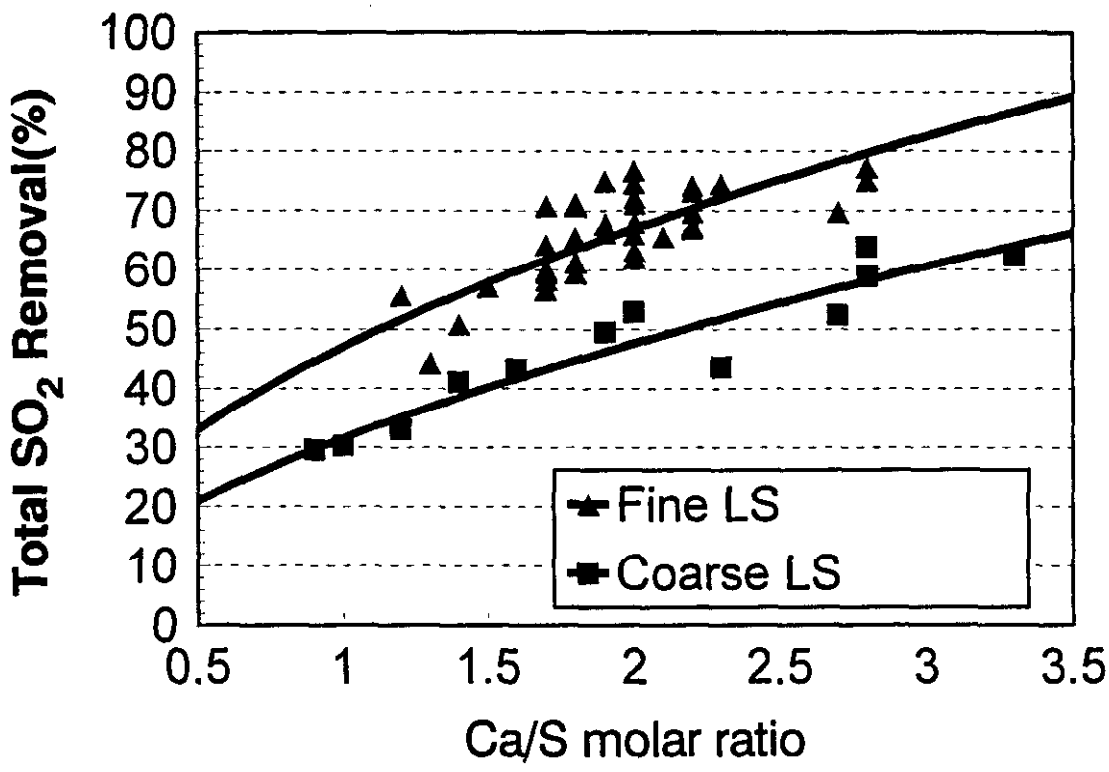
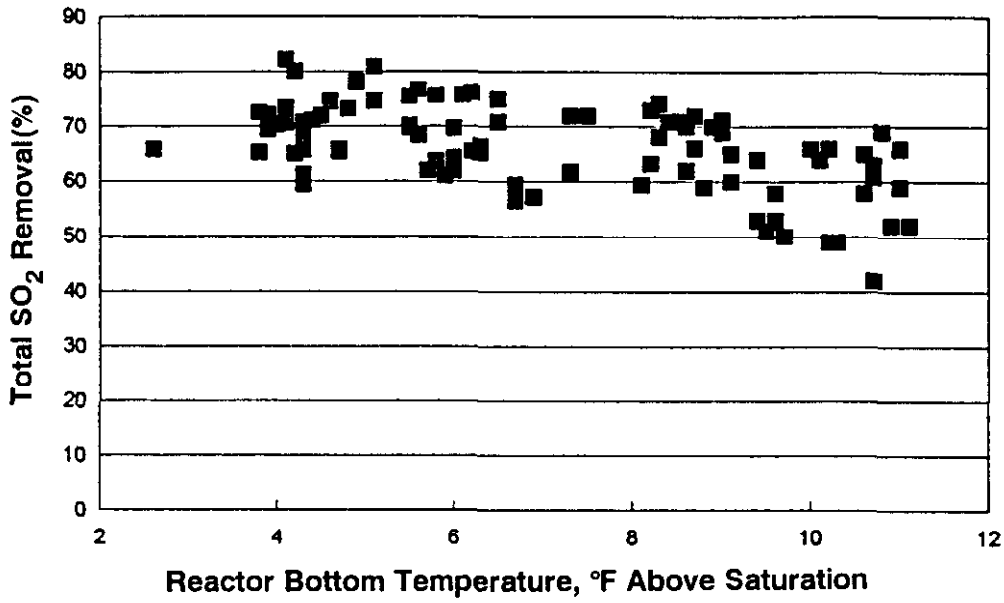


Figure 5-11 Reactor Bottom Approach to Saturation Temperature and Its Effect on Process Efficiency



nozzles were too high. The effects of the nozzle positions on sulfur capture were not documented. It is expected that a higher header position is beneficial because the mixing of the flue gas and water should improve and result in better SO_2 capture.

Water Droplet Size

Atomizing air pressure controls humidification water droplet size into the reactor. Air pressures between 38 and 65 psig were tested. The goal is to efficiently evaporate the droplets during their residence time in the reactor. With low air pressure forming large water droplets, complete evaporation may not take place. This would create a wet ash by-product. Conversely, smaller water droplets may evaporate too quickly and decrease SO_2 capture efficiency.

Ash Recycling

Fly ash containing unreacted calcium is recycled from two ESP hoppers to the reactor inlet duct. The quantity of the recycled material was limited, because the reactor separated more ash than expected. All ash collected in the bottom of the reactor was disposed of at a landfill.

The quantity of ash recycled was limited by the availability of fly ash collected in the ESP hoppers. Ash from the first two ESP hoppers was recycled back to the reactor. If the quantity of ash recycle was increased, the hoppers emptied, and the process became unstable.

Ash recycling improved SO_2 removal by 15 to 20 percentage points. An increased level of recycling would improve SO_2 capture, but the equipment arrangement and performance permitted only small quantities of ash to be recycled. Ash recycling results with high and low boiler loads are shown in Figures 5-12 and 5-13.

5.2 Data Selection

Data were carefully selected from periods which represented stable process operating conditions and provided reliable results. Data representative of certain parameter combinations and operating conditions were selected from a summary sheet by eliminating non-eligible test points. Trends and curves were generated using these selected data points.

Figure 5-12 Unspent Sorbent Recycling Effect on Total SO_2 Capture During High Boiler Load

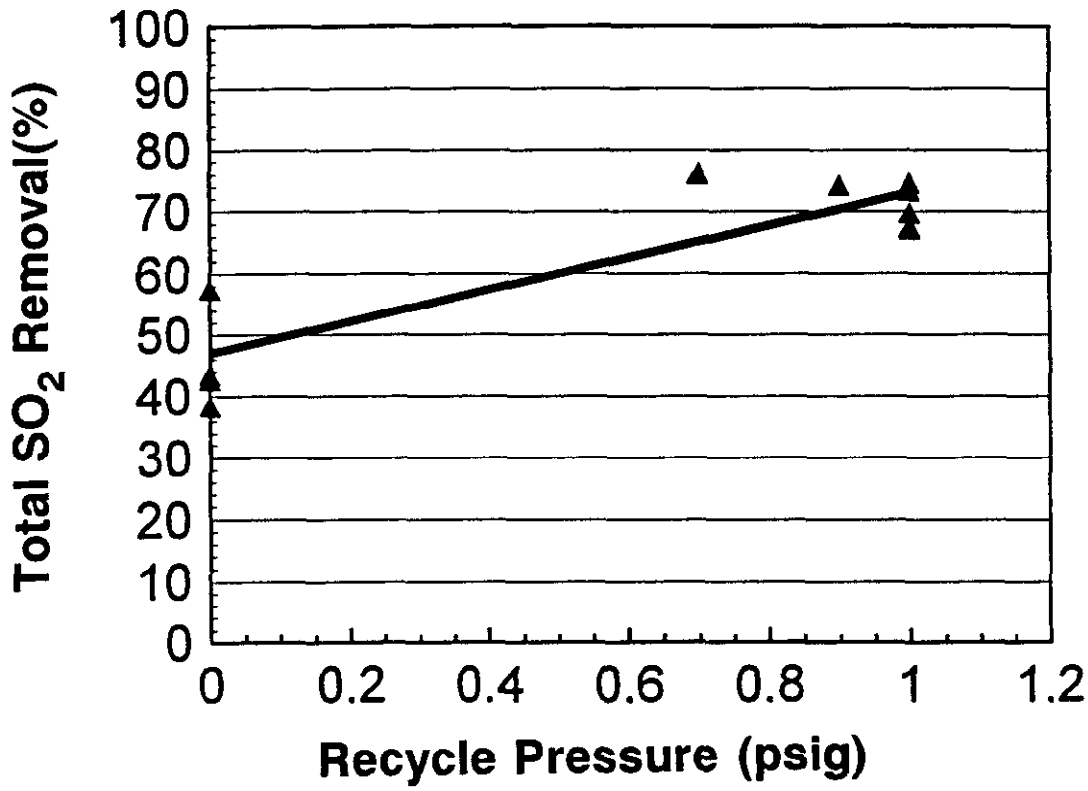
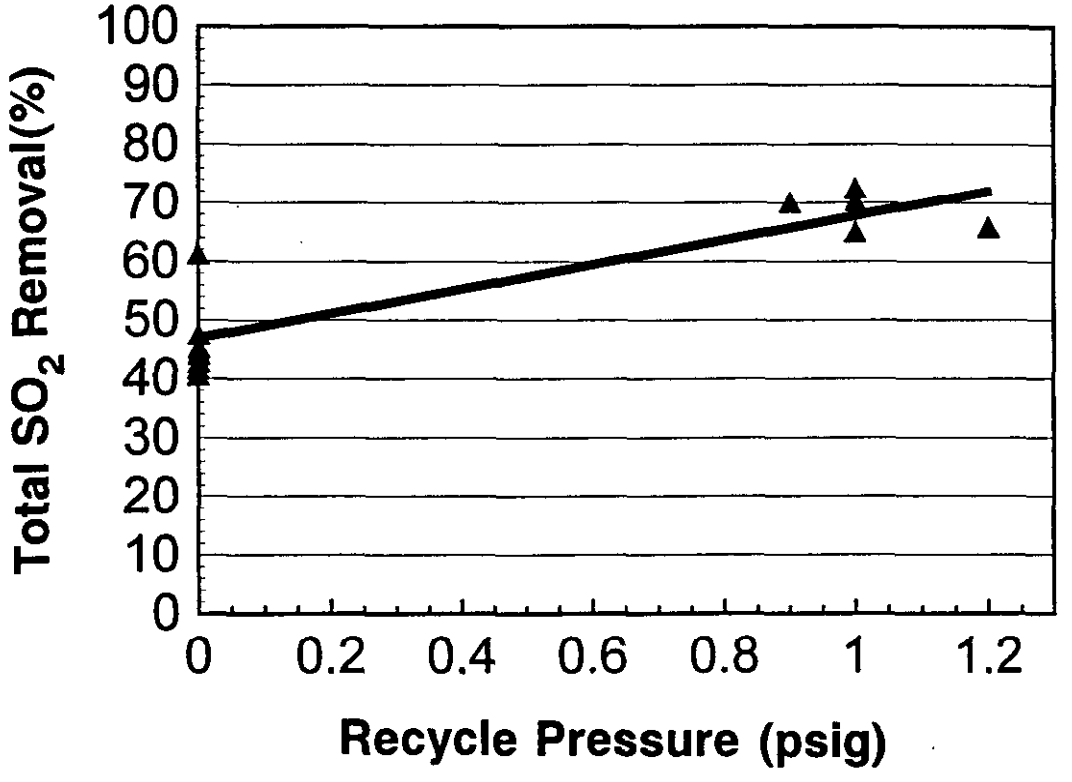


Figure 5-13 Unspent Sorbent Recycling Effect on Total SO_2 Capture During Low Boiler Load



Several factors impacted the reliability of the test results and the planned procedures. Measurements were acquired during normal boiler operation, which were not always favorable for the test program. However, RP&L employs two separate boilers, and load demand was met with the non-LIFAC (Unit No. 1) boiler, when possible. Attempts were made to inject constant quality coal into the Unit No. 2 Boiler during testing. Sootblowing frequency on Unit No. 2 was 4.5 and 6.0 hours as discussed under Section 4.3.2. SO_2 measurements were deemed unreliable during these soot-blowing periods.

Due to the increase in pressure drop caused by LIFAC and the limited capacity of the ID fan motor, the bypass damper was kept 15% open during parametric testing. This bypassed flue gas contained an unknown amount of unreacted lime. This situation was eventually corrected, and long-term tests were performed with the bypass damper fully closed.

5.3 Optimum Settings for Long-Term Operation

Parametric test results were used to establish process settings for the remainder of the test program. Selection criteria were based on optimum SO_2 capture and overall system performance.

Coal burned during long-term testing had a sulfur content of nearly 2.25%, which is the quality that RP&L has been burning during the last few years. (A higher coal sulfur content would require a higher limestone feed rate if the Ca/S ratio is maintained at 2.0.)

A Ca/S molar ratio of 2.0 was selected as the optimum limestone flow setting. This ratio yields the highest sulfur capture without interfering with process or boiler operability. Higher Ca/S molar ratios would increase boiler superheater fouling, thus requiring more frequent sootblowing.

Two different limestone qualities were tested. Fine quality (80% < 325 mesh) was significantly better than the coarse quality (80% > 200 mesh) and was used during long-term testing. Sulfur capture was 20 percentage points higher when fine quality limestone was employed. Sootblowing frequency was greater during injection of fine limestone. Coarse limestone was more abrasive on limestone transport hoses. Coarse limestone is less expensive partly because of the lower grinding cost.

The most important parameter for effective sulfur capture in the reactor is the reactor bottom temperature. During long-term tests, the reactor bottom temperature was kept as low as possible. Lower temperatures result in a higher sulfur capture in the reactor. The temperature setpoint was

limited to 135°F, approximately 10°F above saturation temperature, to prevent reheat steam coils in the exiting ductwork from becoming plugged. Without ash recycle, the reactor temperature setpoint was limited to 140°F, 14°F above the saturation temperature.

Humidification nozzles were situated at their second highest position. In this position the internal sealing remains intact and the humidification effect and gas mixing gave the best results.

Atomizing air pressure was maintained between 55 and 60 psig. This pressure generates an optimum humidification water droplet size. If droplets are too large, they do not completely evaporate in the reactor, resulting in too high a moisture content in the ash exiting the reactor. Smaller droplets may evaporate too fast and reduce sulfur capture in the reactor.

Flue gas leaving the reactor was reheated with both steam heat exchanger coils and with hot gas by-pass. During initial ESP tests, it was discovered that ESP performance improved if the ESP temperature was above 195°F. Both reheater systems were needed to maintain this temperature.

Ash recycling should be as high as possible. In this demonstration the quantity of recycled ash was limited by the availability from the ESP hoppers. Ash from the first two ESP hoppers was recycled to the reactor. If the ash quantity would have been increased, the hoppers may have emptied and the process would have been unstable.

5.4 Optimization and Long-Term Test Implementation and Results

The results of parametric data analyses enabled the test team to determine the optimum process parameter values which would maximize the efficiency of the demonstration system. In order to perform a controlled study, a fine limestone (85% passing 325 mesh, 93% $CaCO_3$) and a consistent quality coal (2.25% sulfur) were employed. The process parameters that were maintained during optimization testing were:

Ca/S Molar Ratio	2.0
Recycle Pressure	1.0 psig
Reactor Bottom Temperature	135°F
Atomizing Air Pressure	55-60 psig
ESP Inlet Temperature	200°F
Bypass Damper	0% Open

Since the optimization tests were coupled with long-term testing, test procedures were similar. A detailed discussion of these procedures is provided below.

The process was in operation continuously for several days during each test period. Normal boiler operating procedures were followed by operators at RP&L. The LIFAC Test Team consisted of one test coordinator and one field operator who were employed 24 hours per day during operations. The test coordinator operated the process, coordinated sampling, and regulated the process according to host facility fluctuations. Most coordination tasks were implemented in RP&L's control room via the PMS, grab samples, and coordinate by-product shipping and limestone receiving.

Limestone feed rate, which is determined by a Ca/S molar ratio, automatically followed coal flow oscillations. Constant quality coal was specially blended and burned during long-term testing. Coal sulfur content of 2.25 wt% was preset in the process control system (PCS).

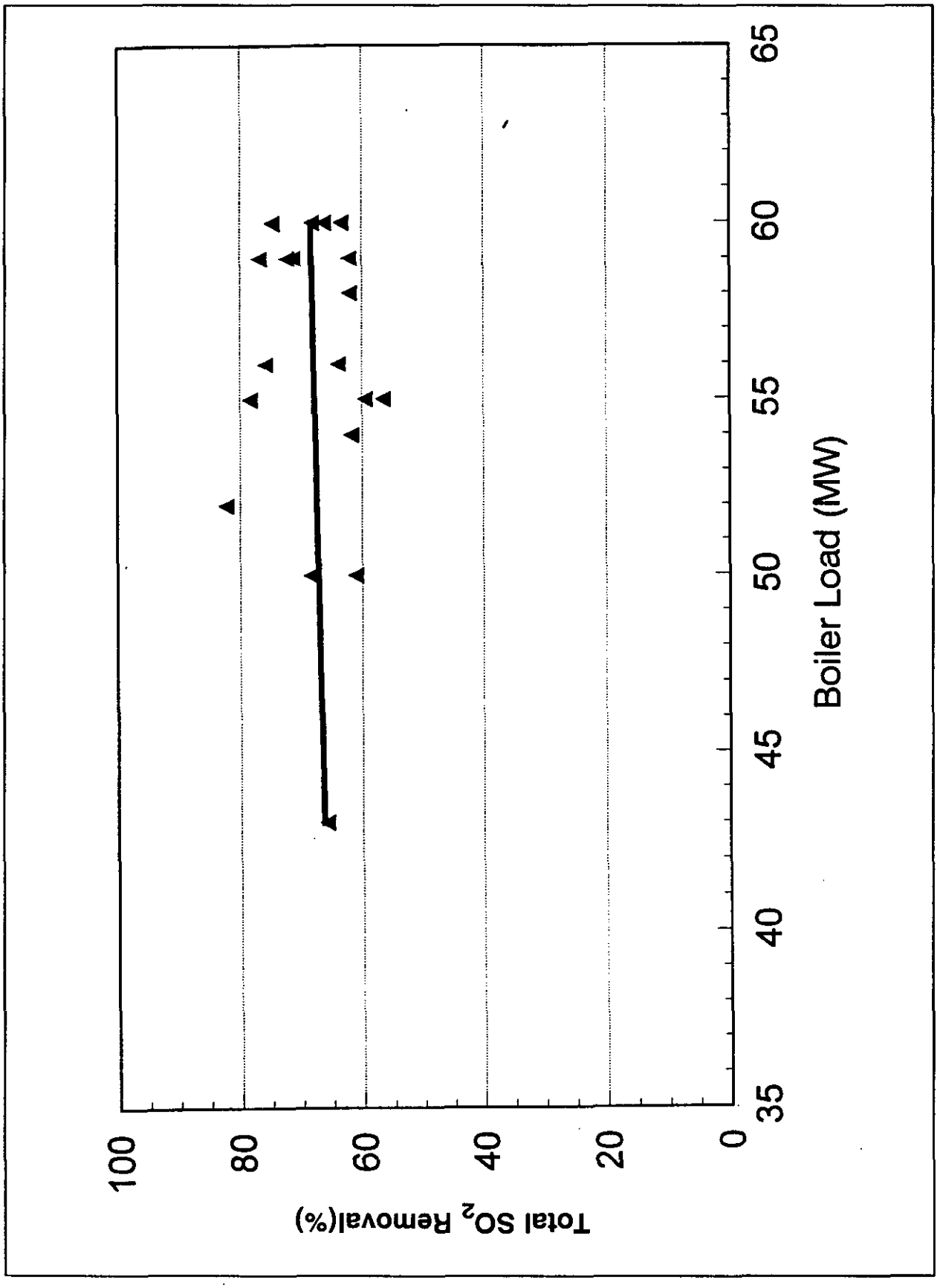
Humidification water flow pressure changed according to the reactor bottom temperature setting. During long-term testing, a reactor bottom temperature of approximately 10°F above saturation temperatures was maintained. This outlet temperature was not optimum, but was necessary to eliminate the possibility of plugging the steam reheaters in the exiting ductwork.

Optimization and long-term tests were performed to evaluate process performance, economics, and operability over long, continuous operations periods. Any impact on the host facility as a result of LIFAC operations was also studied. Extensive ESP performance evaluations were conducted during these tests.

Many process performance results were not reliable due to variations in both boiler and LIFAC operating conditions. The boiler operated normally and fluctuated according to power demand. The operability of the LIFAC system was evaluated during long-term tests, and the results are reported in Section 4.3 of this report.

SO_2 capture in the furnace was calculated based on an estimated sulfur dioxide level resulting from coal combustion compared to the measured SO_2 level leaving the furnace. During steady operation conditions, SO_2 removal across the process varied from 53% to 82%. Boiler load was high (59-60 MW) for most of the test period. Load was slightly lower overnight and on weekends. Figure 5-14 shows total SO_2 removal versus boiler load during long-term testing.

Figure 5-14 Long-Term SO_2 Removal Versus Boiler Load



Limestone Injection and Boiler Operation

The six upper level limestone injection nozzles were used during optimization and long-term testing. Flue gas temperature was slightly cooler (~100°F) near these upper injection nozzles. A higher SO_2 removal rate was observed in the upper level during parametric testing. SO_2 capture in the furnace was only calculated using the analyzer situated between the furnace and the activation reactor. The nozzle may be tilted up to 15° in the vertical direction. However, all tests were performed with the nozzles in a straight horizontal position. Based on Tampella's past experience, tilting effects on SO_2 capture would be minimal as compared to nozzle combination variations which are also minimal. With Ca/S ratios between 1.7 and 2.2, the average total SO_2 removal ranged from 55% to 77% as shown on Figure 5-15.

The booster or secondary air fan was not operated during long-term or optimization tests for three reasons: with the VFD inoperable, the ID fan operated at capacity, and additional flue gas volume would have been detrimental; second, the booster air fan was extremely loud; lastly, no improvement was seen in SO_2 removal in the furnace that was not achievable with proper injection combinations.

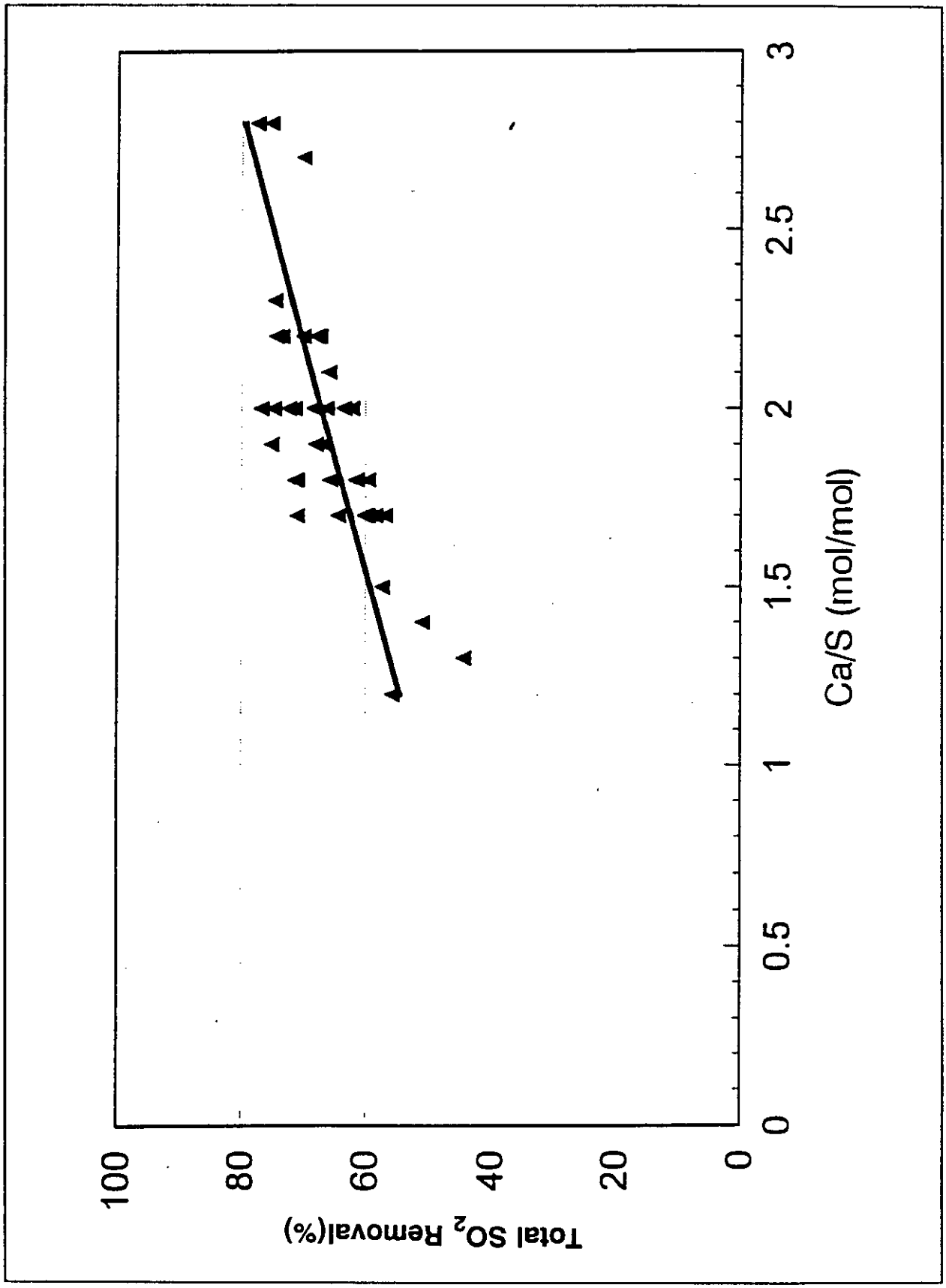
Limestone flow rate to the furnace was determined by the desired Ca/S molar ratio. This ratio was held constant at 2.0 during optimization and long-term tests. In order to maintain this ratio, the limestone feed rate must follow coal quality and flow variations. A 2.25 wt% sulfur content of the feed coal was estimated. If coal quality changed during operation, a correction was made upon receipt of laboratory analyses. Parametric test results indicated that a 2.0 Ca/S molar ratio yields approximately 70% SO_2 reduction.

Limestone quality was consistent at all times. A high calcium (90-95% $CaCO_3$), fine grind (85% < 325 mesh) limestone was injected for long-term evaluations. Smaller particle size limestone proved to be more efficient during parametric testing.

Activation

The bypass damper was 15% open during parametric tests in order to maintain a gas temperature of 200°F through the ESP. The reheat system was modified for long-term evaluations, and only a small amount of hot gas (850°F) was bypassed during this long-term test period.

Figure 5-15 Long-Term SO_2 Removal Versus Ca/S Molar Ratio



The adiabatic saturation temperature of the flue gas is estimated to be approximately 126°F. Reactor bottom temperature was held approximately 10°F above saturation temperature during long-term tests. A lower reactor bottom temperature would have resulted in an improved SO_2 capture; however, the possibility of plugging the steam reheaters constrained the process to these higher temperatures. Steam reheater cleaning is costly and requires days of downtime. At these reactor temperatures, the ash separated in the reactor contained 5 to 15% moisture.

Humidification water droplet size decreases as atomizing air pressure increases. Air pressures between 55 and 60 psig were maintained for this test period. These pressures create droplets small enough to maximize evaporation effects during its residence time in the reactor. Droplets too small may evaporate too quickly, decreasing process efficiency, while large droplets may not completely evaporate and cause the ash to moisten. Humidification nozzle headers were fixed in the second highest vertical position. Figure 5-16 shows the correlation between reactor bottom temperature and process efficiency.

Humidification nozzle scrapers were employed to keep ash from building up on the headers. Each header is scraped six times per hour.

Ash Recycling

A mixture of fly ash and unreacted calcium compounds collected in the ESP is partially recycled to the reactor inlet ductwork. The ratio between the recycled material and material coming from the reactor is called the recycle ratio.

Fly ash flow from the ESP hoppers was controlled with variable speed rotary feeders located under the hoppers. The feeders were programmed to operate at 30% speed, as determined from parametric testing. This is the maximum attainable speed, since the ESP hoppers will empty if it is increased. The reactor bottom ash recycling system was not completed for the demonstration. Ash recycling from the first two ESP hoppers improved SO_2 reduction between 15 to 20 percentage points as shown in Figure 5-17.

Figure 5-16 Long-Term Total SO_2 Removal Versus Activation Temperature

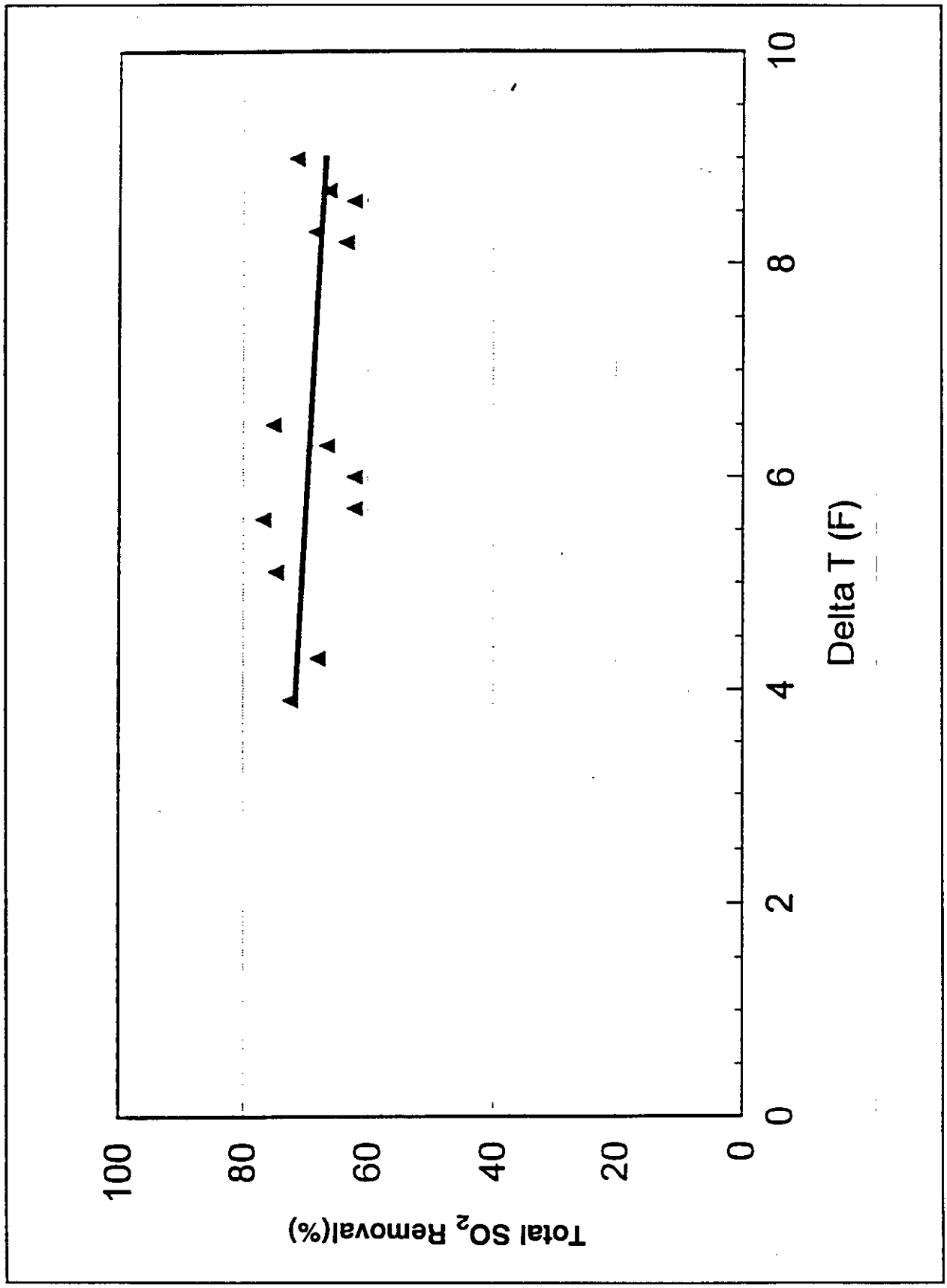
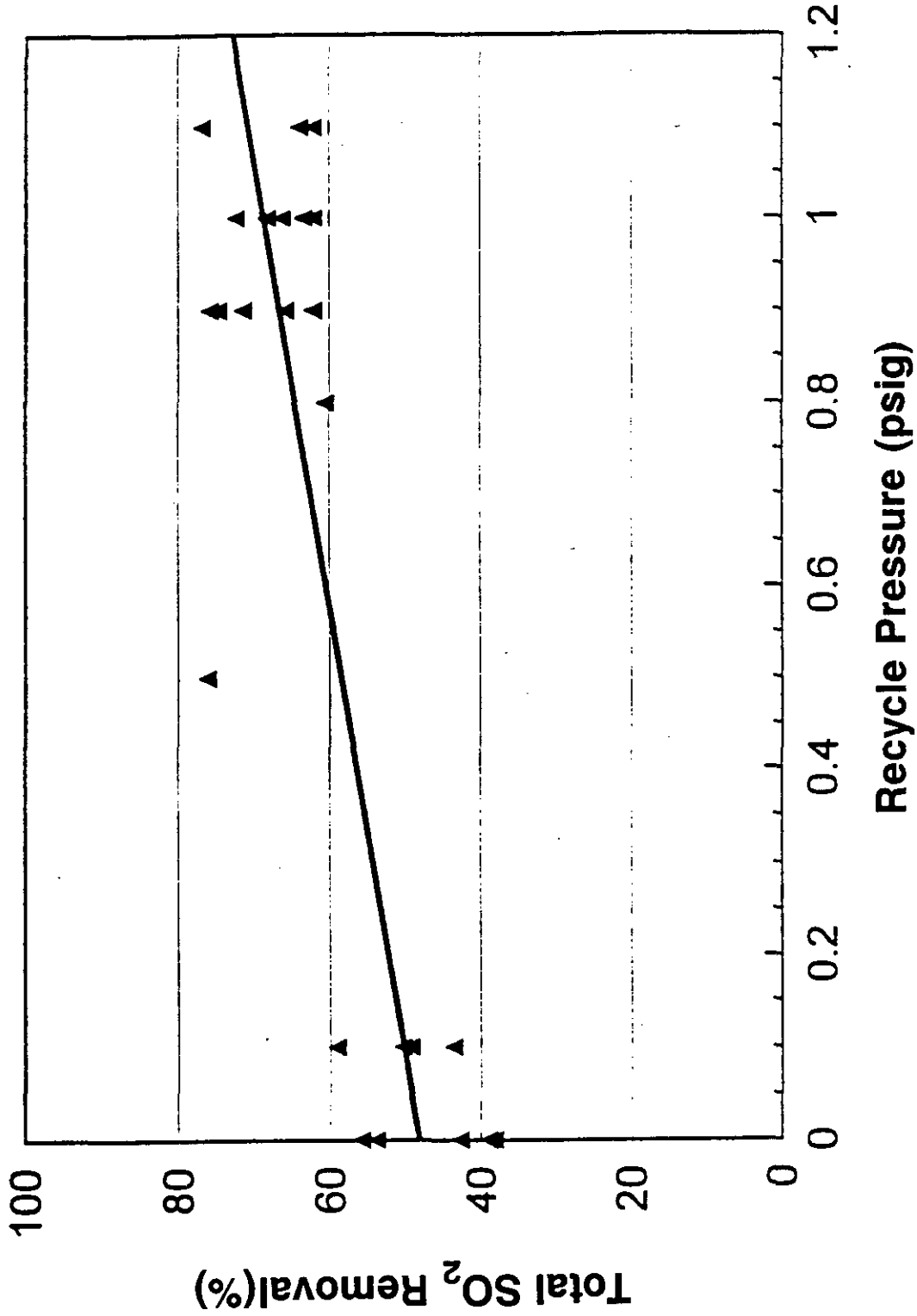


Figure 5-17 Long-Term Total SO_2 Removal Versus Ash Recycling Pressure



ESP Performance Test

ESP performance parameters were recorded throughout the demonstration. EPRI and the Southern Research Institute conducted an extensive evaluation of Unit No. 2 ESP in two phases. The first phase of tests were performed during parametric testing in April of 1993. The second, and final, phase occurred during long-term tests in June, 1994 and the results are presented in Appendix I.

6.0 ENVIRONMENTAL PERFORMANCE

Environmental impacts of the technology were monitored to demonstrate the benefits of the technology such as the reduction of emissions of SO_2 and oxides of nitrogen (NO_x), and also to evaluate potential impacts to any of the waste streams.

6.1 Impact on the Environment

6.1.1 Summary of the Environmental Monitoring Plan (EMP)

Environmental monitoring was conducted in accordance with the required Environmental Monitoring Plan, which was developed for the following purpose:

1. To identify and characterize potential environmental and health impacts of the demonstration project, both on-site and off-site, and

2. To develop an information base for the assessment and mitigation of impacts associated with the replication of CCT projects.

Environmental monitoring was necessary to quantify the project-specific and site-specific environmental mitigation measures implemented as part of the project.

The EMP contained compliance and supplemental monitoring as defined below:

- **Compliance Monitoring:** Compliance monitoring is the monitoring required by agencies of Federal, state and local governments to satisfy statutes, regulations and terms of leases, permits, grants, and other requirements.

- **Supplemental Monitoring:** Supplemental monitoring is monitoring required in addition to compliance monitoring to identify and characterize potential environmental and health impacts of the project, both on-site and off-site.

A variance to operate Unit No. 2 for the testing performed during the different project stages was obtained from the Indiana Department of Environmental Management (IDEM). The adjusted

compliance limits based on this variance are noted in Table 6-A. The variance required specific compliance tests to be conducted at milestones identified in the variance request. In addition to the compliance testing, the project also required supplemental monitoring to be performed in accordance with DOE requirements. This supplemental monitoring focused on the measurement of coal feedstocks, boiler and generator operating conditions, gaseous and particulate discharges, and solid and aqueous process by-products. Measurement of coal feedstocks included coal analysis and feed rate to boiler. Boiler and generator operating conditions were recorded on operator log sheets. Measured discharges included total particulate matter, NO_x, SO_2, carbon monoxide (CO), carbon dioxide (CO_2), and oxygen (O_2). Determinations of alkalinity, pH, sulfate content, and any quantities of specified metals, organic compounds and volatile organic compounds (VOC) present in the process liquid and solid wastes were made. Tables 6-A and 6-B provide a summary of the required compliance and supplemental monitoring required under the EMP, while Table 6-B1 summarizes the actual versus planned monitoring conducted.

A schematic of the system is included as Figure 6-1, which identifies locations where environmental monitoring was performed.

The majority of the compliance and supplemental monitoring centered around the testing of particulate matter and gaseous parameters. The test programs utilized methodologies promulgated in 40 CFR 60, Appendix A, for the particulate matter and duct gas determinations necessary for the test program. EPA Method 1, 2, 3A, and 4 were used for the determination of the sampling point locations, gas velocities and volumetric flow rates, gas molecular weight on a dry basis, and percent moisture content, respectively. Particulate matter was determined using EPA Method 5. Gas concentrations of NO_x, SO_2, and CO were determined utilizing EPA Methods 7e, 6C, and 10, respectively. PM_{10} was determined using EPA Method 201A. Also, verification tests for absence of cyclonic flow was performed each test day at the breach sampling location, in accordance with EPA Method 1, Section 2.5.

Analyses of the solid and aqueous LIFAC waste followed EPA methods and ASTM extraction methods. The specific EPA methods which were referenced are cited in the footnotes on the actual analytical results provided by Antech Ltd. in the Baseline Report[1], Parametric and Optimization Report[6], and the Operations and Post-LIFAC Report[5]. Total alkalinity of both the aqueous and solid samples were determined by EPA Method 310.2. Before the alkalinity of the solid samples could be measured, they were extracted using ASTM extraction methods. The remaining analyses

TABLE 6-A

COMPLIANCE MONITORING
EMP

Parameter	Sample	Sample Frequency and Duration	Limit[1]	Sample Location (see Figure 6-1)
GASEOUS EMISSIONS				
SO_2[2]	Calculation	Daily	6.0 lb/MBtu	9 - Breeching-duct downstream of the I.D. fan*
Total Suspended Particulates (TSP)	EPA Method 5[3]	2/Year	0.22 lb/MBtu	9 - Breeching-duct downstream of the I.D. fan*
Opacity	CEM[4]	Continuous	40%/30%[5]	9 - Breeching-duct downstream of the I.D. fan*
AQUEOUS EFFLUENT (RP&L's NPDES Discharge Permit Requirements)				
Flow	Instantaneous	3 × Weekly	Report	6 - Weir at ponds system outfall
Total Suspended Solids (TSS)	Grab	2 × Weekly	70 mg/l	6 - Weir at ponds system outfall
Oil and Grease	Grab	2 × Weekly	15 mg/l	6 - Weir at ponds system outfall
Total Residual Oxidants (Chlorine)[6]	Grab	2 × Weekly	0.02 mg/l	6 - Weir at ponds system outfall
Temperature	Instantaneous	2 × Weekly	Report	6 - Weir at ponds system outfall
Ammonia (NH_3)	Grab	2 × Monthly	1.8 mg/l (summer) 2.6 (winter)	6 - Weir at ponds system outfall
Copper	Grab	2 × Monthly	0.05 mg/l	6 - Weir at ponds system outfall
Lead	Grab	2 × Monthly	0.02 mg/l	6 - Weir at ponds system outfall
Zinc	Grab	2 × Monthly	0.46 mg/l	6 - Weir at ponds system outfall
Aluminum	Grab	2 × Monthly	0.40 mg/l	6 - Weir at ponds system outfall
Iron	Grab	2 × Monthly	Report	6 - Weir at ponds system outfall
Total Polychlorinated Biphenols (PCBs)	Grab	2 × Monthly	Report	6 - Weir at ponds system outfall

1 Limits for air emissions are based on the LIFAC Variance to the draft boiler Operating Permits. Limits for water discharges are daily limits.
2 Complete coal analysis is performed as part of the Test Program.
3 40 CFR Part 60.
4 Continuous Emissions Monitor.
5 Opacity Limit revised to 30% as per RP&L's variance from IDEM.
6 Chlorine sampling was conducted consistently with RP&L.

* Emissions tests can be conducted at the 250′ level on the stack when Unit #1 is not in operation.

TABLE 6-B

SUPPLEMENTAL MONITORING
EMP

Parameter	Sample	Sample Frequency and Duration	Limit[1]	Sample Location (see Figure 6-1)
GASEOUS EMISSIONS				
NO_x	EPA Method 7 D&E[1]	Continuous or 3/condition	01- lb/MBtu	9 - Breeching-duct* downstream of the I.D. fan
SO_2[2]	EPA Method 6, 6C or 8	Continuous or 3/condition		9 - Breeching-duct* downstream of the I.D. fan
TSP	EPA Method 1-5	Continuous or 3/condition	0-1 lb/MBtu	9 - Breeching-duct* downstream of the I.D. fan
PM_{10}	EPA Method 201A	Continuous or 3/condition	01-lb/MBtu	9 - Breeching-duct* downstream of the I.D. fan
CO_2	EPA Method 3	Continuous or 3/condition	0-500 lb/MBtu	9 - Breeching-duct* downstream of the I.D. fan
CO	EPA Method 10	Continuous or 3/condition	0-500 ppmv	9 - Breeching-duct* downstream of the I.D. fan
AQUEOUS EFFLUENT				
Alkalinity	EPA Method 310.2	1/Condition	mg $CaCO_3$	6 - Weir at ponds system outfall
SOLID WASTE AND BY-PRODUCT				
TCLP	EPA Method 1311 TCLP Metals, Methods 9240 and 8270 for Organics	1/Condition	Various per parameter	3 - Economizer Hopper 4 - LIFAC Ash 5 - ESP Hoppers 10 - Bottom Ash
Sulfates	ASTM Extraction[3] EPA Method 9035	1/Condition	0-1,000 mg/l	3 - Economizer Hopper 4 - LIFAC Ash 5 - ESP Hoppers 10 - Bottom Ash
Alkalinity	ASTM Extraction Method EPA Method 310.2	1/Condition	mg $CaCO_3$	3 - Economizer Hopper 4 - LIFAC Ash 5 - ESP Hoppers 10 - Bottom Ash
Organics	EPA Method 8240	1/Condition	1-1,000 mg/l	3 - Economizer Hopper 4 - LIFAC Ash 5 - ESP Hoppers 10 - Bottom Ash
pH	EPA Method 9045	1/Condition	0-14 H+	3 - Economizer Hopper 4 - LIFAC Ash 5 - ESP Hoppers 10 - Bottom Ash

1 40 CFR Part 60 EMP.
2 SO_2 monitoring will be performed as part of process monitoring.
3 Annual book of ASTM standards.

* Emissions tests can be conducted at the 250′ level on the stack when Unit #1 is not in operation.

TABLE 6-B1

PLANNED vs. ACTUAL ENVIRONMENTAL MONITORING

	Baseline			Parametric[1]		Optimization		Long-Term Operations	Post-LIFAC
	9/2/92	12/18/92	9/22/93	9/23/93	9/24/93	12/7/93	12/8/93	5-17-94	8-16-94
Compliance									
Gaseous Emissions	No Test	TSP only	No test	X	X	X	X	TSP and Opacity only	X
Aqueous Effluent	X	No Test	X (and feed water)	X (and boiler bottom ash and feed water)	X (and boiler bottom ash and feed water)	X (and boiler bottom ash and feed water)	X (and boiler bottom ash and feed water)	X (and boiler bottom ash and feed water)	X (and boiler bottom ash)
Supplemental									
Gaseous Emissions	X	No Test	No Test	X	X	X	X	No Tests	X
Aqueous Effluent	X	No Test	No Test	X	X	X	X	X	X
Solid Waste and By-Product	X	No Test	X	X	X	X	X	X	X

X - All required analysis conducted per the EMP.

[1] Emissions monitoring conducted at Stack 250' level, all others conducted at duct breeching located downstream of the I.D. fan.

LOCATION	MEDIA	TEST PROGRAM	MONITORING PURPOSE EMP	
			SUPPLEMENTAL	COMPLIANCE
1	SOLID	X		
2	SOLID	X		
3	SOLID	X	X	
4	SOLID	X	X	
5	SOLID	X	X	
6	AQUEOUS			X
7	GASEOUS	X		
8	GASEOUS	X	X	
9	GASEOUS			X
10	SOLID	X	X	

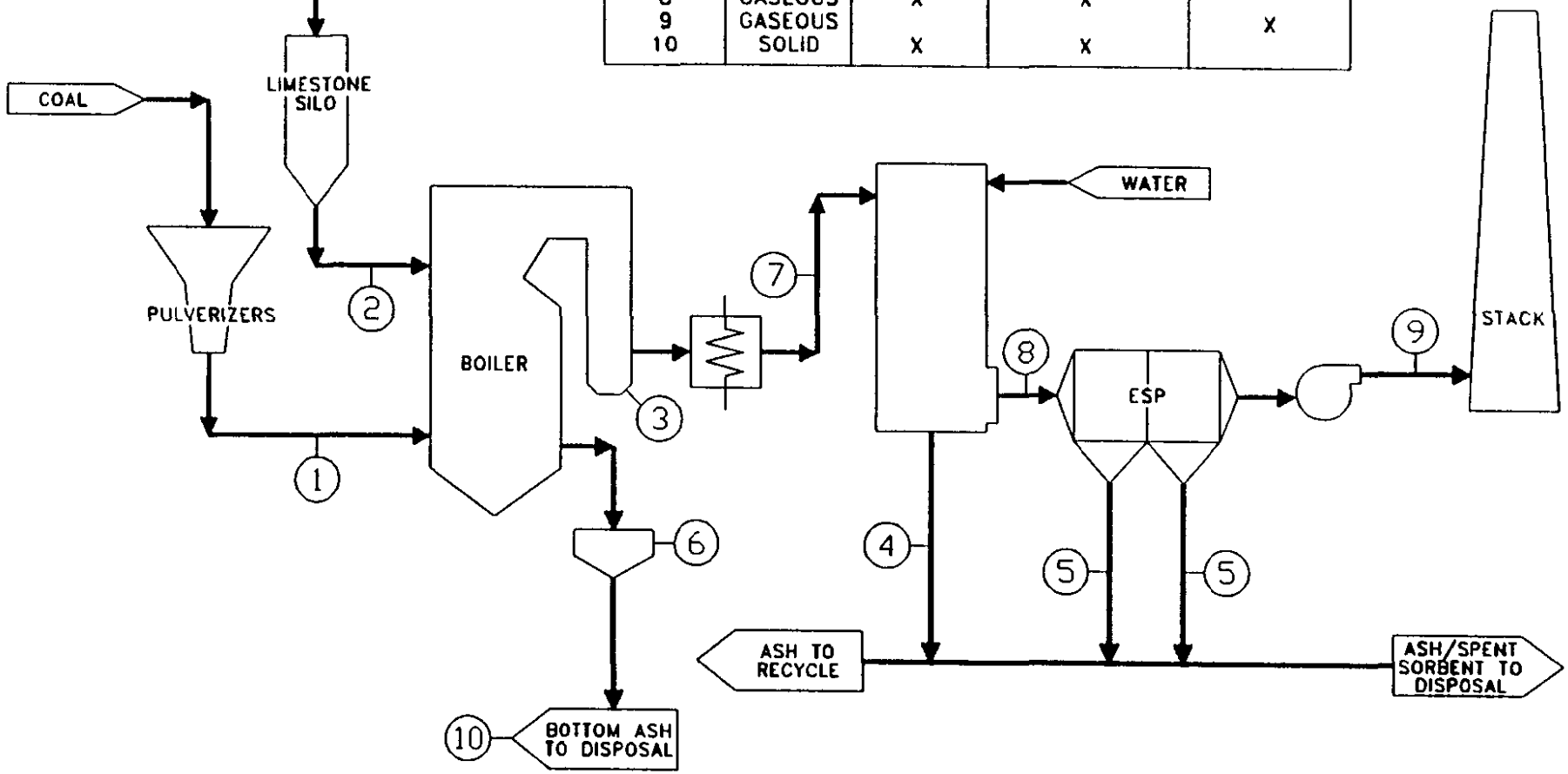


Figure 6-1
Whitewater Valley Unit No.2
LIFAC Demonstration Sampling Locations

of the solid wastes were performed using the following EPA Methods: pH, 9045; sulfate, 9038 (using ASTM extraction method); metals, Toxicity Characteristics Leaching Procedures (TCLP) 1311, 6101, 7470, and 7740; organic materials, TCLP 8260 and 8270; and VOC, TCLP 8260,

6.1.2 LIFAC Test Parameters

Compliance and supplemental monitoring in accordance with the EMP was conducted during the following LIFAC operations tests.

- **Baseline Testing** (conducting September 2, 1992, December 18, 1992 and September 22, 1993): As defined in the Test Plan, Baseline Operations were conducted during the post-construction phase and before any on-line application of hardware or limestone absorbent that would normally be utilized by the LIFAC process. Baseline testing was conducted to develop a information base for normal plant operations to be utilized to evaluate technology and environmental impacts.

Compliance and supplemental tests were conducted each day including emissions testing except for September 22, 1993. Also, no gaseous supplemental monitoring was conducted December 18, 1992. Since Units No. 1 and No. 2 were both in operation during the September 1992 and December 1992 testing, emissions monitoring was conducted at the test portals located in the duct breeching located downstream of the I.D. fan. The September 1993 tests were conducted at the 250 foot level of the stack, since Unit No. 1 was off-line.

Ash samples were collected at the economizer hoppers, EPS inlet and outlet hoppers, and the boiler bottom ash disposal bin.

The significant variation in the two baseline particulate concentrations was due to an error in the September 1992 test. The December results were more in line with anticipated results and prior testing.

LIFAC test parameters during the August 16, 1994 environmental monitoring are summarized in Table 6-C:

TABLE 6-C

BASELINE TESTING - TEST PARAMETERS

	September 2, 1992	December 18, 1992	September 22, 1993
Boiler Load (MW)	65.3	65.8	61.0
Coal Quality - % S	2.20	2.3	2.25
- SO_2 lb/MBtu	4.05	4.0	3.98

- **Parametric Testing** (conducted September 23 and 24, 1993): As per the Test Plan, parametric testing was designed to evaluate the different possible combinations of LIFAC process parametric and their effectiveness. This set of tests was to determine the optimum combination and settings of process parameters which were then used throughout the remaining demonstration tests. Process parameters included sulfur content of coal, limestone quality, limestone calcium to sulfur ratio, injection ports combinations, injection nozzle angles, booster air, reactor outlet temperature, and ash recycling ratio.

  Monitoring included compliance and supplemental emissions testing being conducted for each day. Since Unit No. 1 was off-line during the sampling event, emissions were monitored at the 250-foot level of the stack and not at the duct breeching as done during the 1993 baseline testing. Sootblowing was performed during one of the three PM tests for each of the two days.

  Under supplemental monitoring, ash samples were collected at the economizer hoppers, LIFAC bottom hoppers, EPS inlet and outlet hoppers, and boiler bottom ash disposal bin. Monitoring of the plant feedwater was also conducted since its incoming quality was believed to be impacting compliance with effluent criteria in accordance with RP&L's National Pollutant Discharge Elimination System (NPDES) Permit. Feedwater for the plant is either pumped from the river, or during dry seasons, is pumped from the local sanitary treatment plant's effluent discharge. Feedwater during the September 1993 tests was coming from the treatment plant effluent.

  The boiler bottom ash discharge water was also sampled since it is believed that immediate impacts will not be seen at the pond outfall as a result of LIFAC operations due to the long residence time through the pond system. The team plans to utilize RP&L's pond discharge

monitoring analyses as required under their NPDES permit to identify any long-term trends after LIFAC has been running.

LIFAC test parameters during the September 23 and 24, 1993 environmental monitoring are summarized in Table 6-D:

TABLE 6-D

PARAMETRIC TESTING - TEST PARAMETERS

	September 23, 1993	September 24, 1993
Boiler Load (MW)	60	60
Coal Quality - % S - SO_2 lb/MMBtu	2.23 4.09	2.18 3.87
Limestone Quality	85% - 200 Coarse	85% - 325 Fine
Ca/S Molar Ratio	1.8	2.0
Reactor Outlet Temperature (°F) (approach to saturation)	6	4
% Bypass[1]	30-35%	30-35%
Recycling[2]	No	Yes

(1) % Bypass is the amount of flue gas that is bypassed around the reactor.
(2) Recycle is when spent sorbent from the inlet ESP hoppers is recycled through LIFAC to improve sorbent utilization and increase SO_2 reduction.

- **Optimization Testing** (conducted December 7 and 8, 1993): As defined in the Test Plan, optimization testing was performed to determine the reliability and capability of the LIFAC process over short, continuous operation periods. This set of tests was performed continuously (24 hours a day to simulate actual operating conditions, as opposed to the short 8- to 12-hour parametric tests performed previously). The baseline coal and limestone along with the optimum parametric setting were used during this phase of testing.

Compliance and supplemental tests were conducted each day including emissions testing. Since Units No. 1 and No. 2 were both in operation, emissions monitoring was conducted at the test portals located in the duct breeching located downstream of the I.D. fan. Sootblowing was performed during one of the three PM tests for each of the two days.

Ash samples were collected at the economizer hoppers, LIFAC bottom hoppers, ESP inlet and outlet hoppers and the boiler bottom ash disposal bin. The sampling team also continued

to monitor the feedwater to the plant which, at this time, came from the river, as well as the boiler bottom ash discharge water.

LIFAC test parameters during the December 7 and 8, 1993, environmental monitoring area summarized in Table 6-E:

TABLE 6-E

OPTIMIZATION TESTING - TEST PARAMETERS

	December 7, 1993	December 8, 1993
Boiler Load (MW)	63.3	58.5
Coal Quality - % S - SO_2 lb/MMBtu	2.24 3.93	2.14 3.75
Limestone Quality	85% < 325; Fine	85% < 325; Fine
Ca/S Molar Ratio	2.0	2.0
Reactor Outlet Temperature (°F) (approach to saturation)	8	7-8
% Bypass	40%	40-42%
Recycling	Yes	No

(1) % Bypass is the amount of flue gas that is bypassed around the reactor.
(2) Recycle is when spent sorbent from the inlet ESP hoppers is recycled through LIFAC to improve sorbent utilization and increase SO_2 reduction.

- **Operation (Long-Term) Testing** (conducted May 17, 1994): As per the Test Plan, operations (long-term) testing was performed after the optimization tests and the second set of parametric tests to determine the reliability and capability of the LIFAC process overlong, continuous operating periods. This set of tests was performed continuously over 500 to 800 hours. Although three different coals were to be tested during this period to evaluate LIFAC's suitability for high sulfur Indiana coals, environmental monitoring was only conducted for coal No. 1, which is the primary coal containing an average 2.5% by weight of sulfur.

Monitoring included compliance and supplemental emissions testing. With both units in operation, sampling was performed at the test portals located in the duct breeching, the point just prior to where exhaust from Unit No. 2 joins with Unit No. 1 exhaust, downstream of the I.D. fan. Sootblowing was performed during one of the three PM tests conducted that day.

Under supplemental monitoring, ash samples were collected at the economizer hoppers, LIFAC bottom hoppers, ESP inlet and outlet hoppers, and boiler bottom ash disposal bin. Monitoring of the plant's feedwater was also conducted. Feedwater during the May test came from the river. The boiler bottom ash discharge water was also sampled.

LIFAC test parameters during the May 17, 1996 environmental monitoring are summarized in Table 6-F:

TABLE 6-F

OPERATIONS TESTING - TEST PARAMETERS

	May 17, 1994
Boiler Load (MW)	58
Coal Quality - % S - SO_2 lb/MMBtu	2.25 3.98
Limestone Quality	80% < 325; Fine
Ca/S Molar Ratio	2.0
Reactor Outlet Temperature (°F) (approach to saturation)	10
% Bypass[1]	0
Recycling[2]	Yes

(1) % Bypass is the amount of flue gas that is bypassed around the reactor.
(2) Recycle is when spent sorbent from the inlet ESP hoppers is recycled through LIFAC to improve sorbent utilization and increase SO_2 reduction.

- **Post-LIFAC Testing** (conducted August 16, 1994): As defined in the Test Plan, post-LIFAC testing was similar to the baseline testing. Data were to be gathered without the LIFAC system operating which, when compared with the baseline data, yielded evidence of the impacts LIFAC may have had on the host site's boiler and associated subsystem.

Compliance and supplemental tests were conducted, including emissions testing. Since Units No. 1 and No. 2 were both in operation, emissions monitoring was conducted at the test portals located in the duct breeching located downstream of the I.D. fan. Sootblowing was performed during one of the three PM tests for that day.

Ash samples were collected at the economizer hoppers, ESP inlet and outlet hoppers and the boiler bottom ash disposal bin. The sampling team also continued to monitor the feedwater

to the plant which, at this time, was coming from the treatment plant's effluent discharge, as well as the boiler bottom ash discharge water.

LIFAC test parameters during the August 16, 1994 environmental monitoring are summarized in Table 6-G:

TABLE 6-G

POST-LIFAC TESTING - TEST PARAMETERS

	August 16, 1994
Boiler Load (MW)	64.0
Coal Quality - % S	2.20
- SO_2 lb/MBtu	4.05

6.1.3 Environmental Monitoring Results

A summary of results for compliance and supplemental monitoring for particulate matter testing for LIFAC's baseline, parametric, optimization, operations and post-LIFAC testing are presented in Tables 6-H and 6-I. Table 6-H, Summary of Particulate Matter Test Data, includes pertinent process conditions such as power generation in units of megawatts (MW), coal feed rates (lb/hr) and firing rates (MBtu/hr). It also includes correction to the particulate matter loading to account for sootblowing which IDEM required to be done during one of the three PM tests conducted each test day. (Note, this requirement was not implemented until after the baseline testing was complete.) Table 6-I provides a summary of gaseous emissions as required under supplemental monitoring. (Note, the opacity presented in Table 6-H is an average value, while in Table 6-I it is presented as a range provided by RP&L). No opacity readings were obtained under the baseline conditions. This table indicates a reduction in SO_2 emissions was not measured during the operations testing. However, an overall SO_2 removal was seen during continuous operations as discussed in detail in Section 5. Also note that NO_x and CO_2 stayed within the expected ranges. For detailed particulate matter test data, actual gaseous emission results and backup detailed particulate matter test data, actual gaseous emission results and backup calculations refer to the Baseline Monitoring Report[1] and Operations and Post-LIFAC Report[5]. Coal data for the different testing events are summarized in Table 6-I, with actual analysis provided in the above referenced reports.

TABLE 6-H

SUMMARY OF ENVIRONMENTAL MONITORING
PARTICULATE MATTER TEST DATA

	Units	Baseline		Parametric		Optimization		Operations	Post-LIFAC
		09/02/92	12/18/92	09/23/93	09/24/93	12/07/93	12/08/93	05/17/94	08/18/94
		Duct Breeching		Stack 250' Level		Duct Breeching		Duct Breeching	Duct Breeching
Process Conditions									
Power Generation	MW	65.1	65.8	62.3	60.6	63.3	58.5	58.0	64
Coal Feed Rate	lb/hr	61220	59677	58860	54360	56800	53400	52817	57050
Heat Input	MBtu/hr	691.1	672.3	665.0	614.2	644.9	609.6	606.7	656.2
Particulate Matter Loadings									
Concentration	gr/dscf	0.0135	0.0912	0.0425	0.0672	0.1335	0.1109	0.0511	0.0627
Mass Flow Rate	lb/hr	17.28	120.30	55.36	84.83	193.10	155.60	72.55	96.07
Mass Emission/Heat Input (1)	lb/MBtu	0.025	0.180	0.085	0.131	0.295	0.250	0.121	0.146
Mass Emission/Heat Input (2)	lb/MBtu	0.024	0.174	0.074	0.140	0.257	0.209	0.103	0.116
Mass Emission/Heat Input (2*)	lb/MBtu	--	--	0.072	0.128	0.253	0.206	0.103	0.116
Duct Conditions									
Average Flow Rate	acfm	247167	265939	236622	229237	264200	257800	245528	304100
Dry Standard Flow Rate	dscfm	149267	155968	152089	147086	168900	163600	165867	179367
Temperature	deg F	338	342	225	225	250	254	235	337
Percent Moisture	%	9.9	8.6	13.2	14.0	12.5	12.5	9.5	9.3
Percent O_2	%	4.4	5.7	4.1	6.8	5.7	5.4	6.4	5.1
Percent CO_2	%	14.8	14.2	14.0	13.0	13.3	13.8	13.3	13.8
Isokinetics	%	102.8	108.1	105.7	104.7	102.7	101.7	99	101
Opacity	%	--	--	12.6	12.1	19.7	15.7	12.3	8.0

(1) Based on measured coal feed rates and heating values.
(2) Based on EPA Method 19 (F factor method).
(2*) Corrected for sootblowing.

TABLE 6-I

SUMMARY OF ENVIRONMENTAL MONITORING
GASEOUS EMISSIONS

Parameters (Units)	Limits	Baseline		Parametric		Optimization		Operations	Post-LIFAC
		09/02/92	12/18/92	09/23/93	09/24/93	12/07/93	12/08/93	05/17/94	08/16/94
Sample Location		Duct Breeching		Stack 250' Level		Duct Breeching		Duct Breeching	Duct Breeching
TSP (lb/MBtu)	0.22-0.40*	0.024	0.174	0.074	0.140	0.253	0.206	0.103	0.116
Opacity	30%	--	--	8-25%	10-18%	19-21%	14-18%	11-14%	7.8-8.5%
SO_2 lb/MMbtu (ppmv)	6.0	4.63	--	3.32	2.90	2.80	2.93	--	3.9
NOx lb/MBtu (ppmv)	0-1 (0-588)	(375.6)	--	(379.0)	(352.7)	(337.2)	(350.3)	--	0.6
PM-10 lb/MBtu (%) (1)	0-1 (100%)	--	--	(77.1 >)	(51.1>)	(55.27 >)	(50.66>)	--	0.02
				(22.9 <)	(48.9 <)	(44.73 <)	(49.34 <)		
CO_2 lb/MBTU (%)	0-500 (0-50)	(14.8)	--	(14.36)	(13.95)	(13.35)	(13.83)	--	(13.8)
CO (ppmv)	0-500	19.9	--	66.57	216.23	4.1	6.5	--	6.0

* Per IDEM Variance

(1) PM-10 presented in % greater or less than 10 microns in diameter.

The particulate matter mass loadings summarized in Table 6-H are expressed in grains per dry standard cubic feet (gr/dscf), pounds per hour (lb/hr), and mass per heat input, expressed as pounds per million Btu (lb/MBtu). Mass emission per heat input was calculated using two different methods. The first method was based on actual process parameters such as coal feed rate to the boiler and coal heating value. The second method utilized EPA Method 19, also know as the F factor method. The following equation, extracted from the method, calculates a theoretical mass emission per heat input based on F_d, which is the ratio of the gas volume of the products of combustion to the heat contents of the fuel (dependent on the type of coal), measured pollutant concentration, and the measured oxygen contact, with all variables being on a dry basis.

$$F = E + F_d * C_d * [20.9/(20.9 - \%O_{2d})]$$

where:

E	=	Emission rate (lb/MBtu)
F_d	=	Oxygen based F factor (dscf/MBtu), dry basis (9780 dscf/MBtu for bituminous coal)
C_d	=	Pollutant concentration (lb/dscf)
$\%O_{2d}$	=	Percent oxygen (%), dry basis

6.2 Waste Streams and Their Disposal

Under the EMP all waste streams impacted by the technology have been monitored before, during and after LIFAC operations. These waste streams include ash from the economizer, LIFAC reactor bottom hopper, ESP hoppers (front and back) and the boiler bottom ash disposal bin. All but the LIFAC ash was and still is disposed by RP&L by transporting it by truck from the ash disposal bin to an approved landfill. LIFAC ash was trucked off-site separately. It was collected at the bottom of the reactor in dumpsters then hauled to an approved landfill by the LIFAC partnership.

Another waste stream is the water from the boiler bottom which is discharged to the ash disposal bin, and then goes to the power plants pond system. This waste stream was monitored by RP&L at the pond outfall under their NPDES Discharge permit. However, as previously discussed, we also monitored this discharge at the ash disposal bin due to the concern of a long residence time through the pond system and inability to verify any impacts. No impact was evident based on the time frame our monitoring was conducted.

Results of all monitoring of ash, feed water, and discharge water are summarized in Tables 6-J through 6R, provided in Appendix III. Table 6-R provides a summary of detects identified from the solid waste sampling.

6.3 Potential Environmental Concerns

The presence of arsenic in the LIFAC ash is an environmental concern particularly to the workers exposed to the dust generated during its discharge into the dumpsters and eventual disposal. ICF Kaiser conducted an industrial hygiene test to determine if levels of exposure were dangerous. Although the results determined that they were not, it was recommended that steps be taken to control the dust and protect workers from exposure, such as wearing surgical masks. The arsenic levels in the ash are still acceptable to the landfill where the ash was disposed; however, in other locations or in the future, this could cause a disposal problem. Another safety concern is the noise level of the blowers in the Limestone building. It was determined that hearing protection was warranted.

Finally, the detection of methylene chloride in several ash samples is a concern, since its origin is unknown. Methylene chloride was detected in economizer ash during the baseline testing, ESP hopper ash (outlet) during parametric and post-LIFAC testing, and in boiler bottom ash during parametric, optimization and post-LIFAC testing. Although there is no current regulatory limit for methylene chloride, further evaluation may be required for future facilities.

7.0 ECONOMICS

7.1 Project Capital Cost

The capital cost breakdown of the LIFAC installation at RP&L is unique, since this installation was a retrofit to an existing facility and installed at the site of an earlier FGD demonstration. It is necessary to present a detailed structure of capital spending. Figure 7-1 shows the total capital cost of the LIFAC unit as it was installed. Engineering reactor fabrication and erection, and structural/ mechanical/piping constituted over half of the capital. The total capital cost of the LIFAC demonstration at RP&L was approximately $8,101,000.

Since RP&L was the site of an earlier FGD demonstration, capital cost savings were experienced. The equipment from the previous demonstration was donated to LIFAC. Each piece of equipment was selected to be utilized based on its operability and reliability for the LIFAC demonstration. This equipment can easily be replaced when LIFAC advances into a commercial operation mode at RP&L. The reactor ash disposal system was built for demonstration purposes only. Additional capital will need to be spent on ash conveyors and a storage silo before commercial operation. Also, no back-up systems were installed as part of the demonstration. A commercial unit would require some redundancy systems such as a limestone feeder bypass option, a backup water pump, and spare I/O cards for the process control system.

The demonstrative nature of the project required additional capital spending. The LIFAC test program incorporated many testing and measuring techniques. For this reasons, additional instrumentation and data collection devices needed to be purchase and utilized. Supplementary sampling ports and man doors were also required for the demonstration in order to inspect and assess the process' impact on the reactor and within the ductwork. Five limestone injection ports were installed on the boiler's walls to test several combinations of injection settings.

Additional capital costs also ensued as a result of retrofitting the LIFAC process to an existing facility. Extra engineering, construction, and equipment costs were realized. The following is a breakdown of additional costs assumed as a result of retrofitting:

Capital Cost	
Equipment & Materials :	
Limestone Handling & Storage	$160,000
Activation Reactor & Ductwork	$669,000
Sorbent Recycle System	$67,000
Electrical / Instrumentation	$272,000
I.D. Fan Upgrade	$255,000
Total Equip. & Materials	**$1,423,000**
Subcontracts :	
Foundations	$324,000
Reactor Fab. & Erection	$1,670,000
Structural / Mechanical / Piping	$1,569,000
Electrical / Instrumentation	$574,000
Insulation & Cladding	$268,000
Misc. Fabrication	$573,000
Total Subcontracts	**$4,978,000**
Engineering :	$1,200,000
Management & Administration :	$300,000
Construction Supervision :	$200,000
Total Capital Cost	**$8,101,000**

Figure 7–1 Total Capital Cost of the LIFAC Demonstration as it was Installed at Richmond Power & Light

Table 7-A *Procurement Cost and Budgets - Equipment, Materials and Subcontracts - Budget Period II Commitments*

Activity No	Description	P.O. Number	Supplier	P.O. Amount	Freight	Current Commitment
C9006	Transformers	91001-1012	Richmond Power & Light	($0.00)	$0.00	($0.00)
C9003	Flue Gas Analyzer Parts	91001-1026A	Lear Siegler	$1,169.30	$33.75	$1,203.05
C4004	Expansion Joints	91001-1027B	Dynex	$3,564.00	$77.50	$3,641.50
C4004	Isolation Damper	91001-1027C	Damper Design Inc.	$6,980.00	$697.70	$7,677.70
C4004	Actuator Parts and Installation	91001-1027D	Damper Design Inc.	$2,000.95	$0.00	$2,000.95
R0012	ESP Inspection	91001-1028	Lodge-Cottrell	$4,800.00	$0.00	$4,800.00
R0013	Particulate Emission Modeling	91001-1033	Meteorological Evaluation Ser.	$4,975.00	$0.00	$4,975.00
C9014	Temperature Transmitters	91001-1035A	Andrews Industrial Controls	$320.00	$16.10	$336.10
C9012	Temperature Elements	91001-1038	Ram Sensors	$2,517.66	$35.16	$2,552.82
C9013	Pressure Indicators	91001-1043A	Thermoflo Equipment Co.	$241.78	$11.23	$253.01
C9027	Flow Element - Flue Gas	91001-1045A	Measurement Instruments East	$922.00	$0.00	$922.00
R0014	Stack Gas Sampling	91001-1046	Keystone Environmental	$16,200.00	$0.00	$16,200.00
R0017	Sutorbuilt Blower Inspection	91001-1047	Indiana Bulk & Pneumatics	$520.00	$0.00	$520.00
R0018	Acrison Weigh Feeder Inspection	91001-1048	Acrison Inc	$2,204.35	$0.00	$2,204.35
R0019	Compressor Inspection	91001-1049	Ingersoll Rand Corp	$1,099.50	$0.00	$1,099.50
R0019	Compressor Inspection	91001-1049B	Ingersoll Rand Corp	$1,205.20	$0.00	$1,205.20
R0019	Compressor Repairs	91001-1049C	Ingersoll Rand Corp	$1,490.26	$0.00	$1,490.26
R0019	Compressor Repairs	91001-1049D	Ingersoll Rand Corp	$213.35	$0.00	$213.35
R0019	Compressor Repairs	91001-1049E	Ingersoll Rand Corp	$1,989.36	$0.00	$1,989.36
R0020	Fuller Kinyon Pump Inspection	91001-1050	Fuller Company	$0.00	$0.00	$0.00
R0015	Fuller Kinyon Pump Parts	91001-1051	Fuller Company	$1,566.00	$35.35	$1,601.35
R0016	Conveyor Chain Lubrication	91001-1052	Tribology Tech-Lube	$2,410.20	$199.38	$2,609.58
R0018	Air Compressor Parts and Oil	91001-1053	Ingersoll Rand Corp	$5,266.90	$9.33	$5,276.23
C2008	Root's Air Blower	91001-1055	Airtek Inc.	$13,925.00	$0.00	$13,925.00
R0021	Flue Gas Probes	91001-1056	Napp, Inc	$4,823.00	$147.58	$4,970.58
R0022	Calibration Gases	91001-1057	Beck's Welding Supply	$6,000.00	($487.80)	$5,512.20
R0023	Signal Isolators	91001-1058	Instrument Services Inc	$1,638.00	$18.50	$1,656.50
R0024	CO2 and O2 Analyzers	91001-1059	VWR Scientific	$852.50	$71.10	$923.60
R0025	Span Potentiometer	91001-1060	Ametek Pressure Measurement	$60.00	$23.50	$83.50
R0026	Analog Input Module	91001-1061	Tri-State Supply Co.	$916.50	$19.25	$935.75
R0026	Analog Input Module Repair	91001-1061A	Tri-State Supply Co.	$1,365.00	$19.25	$1,384.25
R0027	Zero Speed Switch	91001-1062	Hobden Schilbe & Smith	$505.00	$26.55	$531.55
R0028	Water & Solid Waste Analysis	91001-1063	Antech Ltd.	$35,000.00	($1,002.62)	$33,997.38
R0029	Stack Gas Sampling Equipment	91001-1064	Keystone Environmental	$10,000.00	($7,971.98)	$2,028.02
R0030	Pilot Operated Solenoid Valve	91001-1065	T. F. Cambell Co.	$470.00	$31.45	$501.45
R0031	Ashcroft Pressure Gage	91001-1066	M. S. Jacobs & Associates	$36.00	$13.70	$49.70
R0032	Calibration Adapter Plug	91001-1067	Equipment and Controls	$51.00	$13.50	$64.50
R0034	Oxygen Monitor Rental	91001-1068	Response Rentals	$3,000.00	($2,668.23)	$331.77
R0035	ASCO Replacement Diaphragm	91001-1069	T. F. Campbell	$33.50	$14.25	$47.75
R0036	Sample Containers	91001-1070	Warehouse Paint Centers	$2,000.00	($1,532.00)	$468.00
R0037	Control Valve Parts	91001-1071	Equipment & Controls	$277.71	$0.00	$277.71
R0038	Crushed Limestone	91001-1072	Rodgers Group	$20,903.23	$0.00	$20,903.23
R0039	Portable Radios	91001-1073	Communications Unlimited	$1,054.00	$12.00	$1,066.00
R0040	I/P Replacement Parts	91001-1074	Equipment and Controls	$161.93	$0.00	$161.93
R0041	Coal & Ash Analyses	91001-1075	Standard Laboratories	$20,000.00	($5,273.84)	$14,726.16
R0042	Switchgear Relays	91001-1076	Verhill Associates	$184.00	$0.00	$184.00
R0043	Electrical Parts	91001-1077	Grant Industrial Controls	$178.01	$41.00	$219.01
R0044	Analyzer Repairs	91001-1078	Lear Siegler	$822.00	$130.70	$952.70
R0045	Hoist and Chain Fall	91001-1079	Mazzella	$646.60	$15.95	$662.55
R0046	Transmitter - ID Fan Speed	91001-1080	Rosemount	$474.00	$19.25	$493.25
R0047	Analyzer Fittings	91001-1081	Indianapolis Valve & Fitting Co.	$109.80	$0.00	$109.80
R0048	Sample Port Installation	91001-1082	H.J. Osterfeld	$3,717.00	$0.00	$3,717.00
R0048	Injection Piping Installation	91001-1082A	H.J. Osterfeld	$12,932.00	$0.00	$12,932.00
R0048	Control Valve Piping Installation	91001-1082B	H.J. Osterfeld	$1,245.00	$0.00	$1,245.00
R0049	Chart Paper & Cartridge	91001-1083	Instrument Sales & Service	$293.85	$3.70	$297.55
R0050	Chart Paper & Pens	91001-1084	Ray Knitter Associates	$225.75	$5.26	$231.01

Table 7-A, page 2

Activity No	Description	P.O. Number	Supplier	P.O. Amount	Freight	Current Commitment
R0051	LIFAC Ash Handling	91001-1085	Rumpke	$150,000.00	($41,072.35)	$108,927.65
R0052	Electrical Relays	91001-1086	Grant Industrial Controls	$61.60	$2.24	$63.84
R0053	Analyzer Calibration	91001-1087	Lear Siegler	$2,051.00	$0.00	$2,051.00
R0054	Certified Smoke Reader	91001-1088	Crawfordsville EL & P	$525.73	$0.00	$525.73
R0055	Opacity Monitoring & Consulting	91001-1089	Mostardi-Platt Associates, Inc	$18,104.76	$0.00	$18,104.76
R0055	Opacity Monitoring & Consulting	91001-1089C	Mostardi-Platt Associates, Inc	$18,142.00	$0.00	$18,142.00
R0055	Opacity Monitoring & Consulting	91001-1089D	Mostardi-Platt Associates, Inc	$103,940.00	($64,684.60)	$39,255.40
R0056	Sorbent Injection Hose	91001-1090	Specialty Hose Corp	$2,504.70	$117.35	$2,622.05
R0057	MOHR INC	91001-1091	Crane Rental	$1,240.00	($322.00)	$918.00
R0058	Ross Hills Controls	91001-1092	VFD Service	$1,500.00	($1,500.00)	$0.00
R0059	Limestone Supply	91001-1093	Cosmos Cement	$160,000.00	($19,495.50)	$140,504.50
R0060	Repair Air Conditioning System	91001-1094	Fuller Engineering Services	$361.05	$0.00	$361.05
R0060	Repair Air Conditioning System	91001-1094A	Fuller Engineering Services	$1,192.50	$0.00	$1,192.50
R0060	Repair Air Conditioning System	91001-1094B	Fuller Engineering Services	$500.00	($500.00)	$0.00
R0061	Repair Pressure Transmitter	91001-1095	Rosemount, Inc	$800.00	($682.17)	$117.83
R0062	Water Control Valve	91001-1096	Equipment & Controls	$3,363.00	$0.00	$3,363.00
R0063	Signal Isolators	91001-1097	EIL Instruments	$732.00	$11.98	$743.98
R0064	Signal Isolator/Convertor	91001-1098	Gilson Engineering	$205.00	$3.58	$208.58
R0065	Pneumatic Tugger Rental	91001-1099	Oberle and Associates	$4,000.00	($2,758.84)	$1,241.16
R0066	Electro-Pneumatic Transducer	91001-1100	M.S. Jacobs & Assoc. Inc.	$407.00	$4.29	$411.29
R0067	Roof Repairs	91001-1101	Mid-Miami Roofing	$3,196.50	$0.00	$3,196.50
R0068	Flue Gas Reheat Study	91001-1102	Black & Veatch	$2,920.00	$0.00	$2,920.00
R0069	Filter Cartridges	91001-1103	Dynamic Air Conv. Systems	$672.00	$42.08	$714.08
R0070	Pneumatic Tugger Rental	91001-1104	F & M Supply	$3,000.00	$1,712.25	$4,712.25
R0072	RTD Temperature Detectors	91001-1105	The Ginger Bonfili Co	$282.00	$68.90	$350.90
R0073	Particle Size Analyses	91001-1106	Penn State University	$15,000.00	($12,220.80)	$2,779.20
R0075	Proximity Limit Switches	91001-1107	Denko Engineering	$864.00	$22.00	$886.00
R0076	Waste Disposal	91001-1108	Chemical Waste Management	$50,000.00	($19,718.26)	$30,281.74
R0077	High Pressure Cleaning	91001-1109	Miami Valley Services	$6,185.50	$3,592.50	$9,778.00
R0078	Solenoid Valves	91001-1110	TriState Supply	$328.00	$15.00	$343.00
R0079	460V Transformer	91001-1111	Horner Electric	$48.75	$3.27	$52.02
R0080	Inverter Repair	91001-1112	Horner Electric	$856.10	$39.45	$895.55
R0081	Weigh Feeder Repairs	91001-1113	Acrison	$31.26	$27.35	$58.61
R0081	Weigh Feeder Repairs	91001-1113A	Acrison	$2,971.10	$0.00	$2,971.10
R0071	Project Computer	91001-16073	Gateway 2000	$1,995.00	$425.22	$2,420.22
R0002	Steel, Mechanical, Piping Inst.	91001-3002-6	Enerfab Inc	$19,627.64	$0.00	$19,627.64
R0002	Steel, Mechanical, Piping Inst.	91001-3002-7	Enerfab Inc	$14,026.23	$0.00	$14,026.23
R0002	Steel, Mechanical, Piping Inst.	91001-3002-8	Enerfab Inc	$518.74	$0.00	$518.74
R0002	Steel, Mechanical, Piping Inst.	91001-3002-9	Enerfab Inc	$1,640.94	$0.00	$1,640.94
R0003	Mechanical Repairs	91001-3004	Oberle & Associates	$8,509.08	($5,639.00)	$2,870.08
R0001	Electrical & Instrumentation	91001-3005	Cummins Electric	$8,785.61	$0.00	$8,785.61
R0001	Electrical & Instrumentation	91001-3005	Cummins Electric	$14,927.18	$0.00	$14,927.18
R0007	Insulation and Lagging	91001-3006	Kramig Co.	$2,380.58	$0.00	$2,380.58
R0033	Stack Sampling	91001-3007	Keystone Environmental Res.	$18,497.31	$0.00	$18,497.31
R0074	Reheat Duct Installation	91001-3008	Kennedy Tank & Mfg. Co.	$12,153.00	$0.00	$12,153.00
			TOTAL PERIOD II	$864,603.05	($179,700.54)	$684,902.51

Indicates a New or Revised Commitment in the Current Month

LAYOUT

- Erected a building around existing RP&L limestone storage and handling equipment
- Provided two motor control centers
- Modified ESP ductwork with vanes and baffle plates
- Erected additional stairs in boilerhouse
- Extensively routed ductwork to/from reactor

CONSTRUCTION

- Some work was performed only during boiler outages
- Rerouted some existing plant piping
- Rerouted some existing plant electric cables
- Cleanup after construction (paving, grass, painting, etc.)

EQUIPMENT

- Repaired, replaced, and calibrated existing RP&L limestone storage and handling equipment
- Installed additional limestone silo for increased capacity
- Installed a "stand-alone" process control system
- Installed a VFD to increase ID fan efficiency
- Added an instrument air compressor for LIFAC instrumentation
- Installed moisturizing screw conveyor on ESP ash disposal silo
- Changed ESP fly ash removal from hydroveyor to blower-operated vacuum system

Fabrication of LIFAC's process elements was performed both in the shop and on-site. Prefabrication was initiated by the subcontractors at their respective facilities prior to shipment. Most of the structural steel applications were prefabricated, including reactor support, stair tower, reactor building, and the limestone building. All steel structures were provided with bolted connections. The steel was pre-painted; only touchup painting was required at the site. It was necessary for certain sections of ductwork to be assembled in the shop. Ducts utilized flanges for easy bolt connections. No on-site painting of the ductwork was necessary.

On-site fabrication was performed on the grounds of RP&L before erection. Most of the activation reactor vessel was assembled at RP&L. The humidification (top) and discharge (bottom) sections of the LIFAC reactor were prefabricated, cut in half, shipped to the site, then reassembled before installation. Each circular section of the reactor was welded on-site with three arcs, each one-third

of the total circumference. The reactor and ductwork were fitted with insulation and cladding prior to erection. The remaining process equipment was shipped directly to the site for installation.

Table 7-B presents estimated costs of the existing equipment which was incorporated into the LIFAC Project. (Note that these costs represent what it would cost to purchase these items as part of the project, if they were not existing).

TABLE 7-B

ESTIMATED COSTS FOR EXISTING EQUIPMENT INCORPORATED INTO THE LIFAC PROJECT

Existing equipment		
125-ton Storage Silo	$	106,400
Limestone Feeding System	$	694,600
Two Vent Baghouses	$	20,400
TOTAL COSTS	$	821,400

Also, the additional cost incurred because this was a demonstration project is estimated at $534,100. This basically includes 75% of the environmental costs which is included under engineering in Figure 7-1, with the required monitoring, and also approximately 25% of project management/administration.

7.2 Projected Operating Cost

7.2.1 Fixed Operating Cost

During the two year demonstration period, the project required a surplus of operating and test personnel. Labor costs were greater during testing. Two additional operators per shift were required for the test phase of the project, while a unit in commercial operation would require a limited number of operators. Minor maintenance tasks were performed by the test team and the RP&L maintenance crew. Major repairs or modifications were executed by subcontractors. The LIFAC process was demonstrated over short operation periods, from one day to three weeks of testing. Except for the onset of unexpected repairs, maintenance costs were expected to be low during the 2800 hours of the demonstration period. Table 7-C shows the estimated annual fixed operating and

maintenance cost of the RP&L LIFAC unit in continuous operation. The total annual fixed O&M cost is approximately $581,200.

TABLE 7-C

ESTIMATED ANNUAL FIXED OPERATING AND MAINTENANCE COSTS

ANNUAL FIXED OPERATING COSTS, CONTINUOUS		
Operating Labor Cost Details:		
Number of Operators Per Shift	2	
Number of Shifts Per Week	4.2	
Operating Pay Rate Per Hour	$25	
		Cost, $/year
Total Annual Operating Labor Cost		499,200
Total Annual Maintenance Labor Cost		25,000
Total Annual Maintenance Material Cost		50,000
Total Annual Administrative and Support Labor Cost		7,000
Total Annual Fixed O & M Cost		581,200

7.2.2 Variable Operating Cost

Variable operating cost includes all the commodities necessary for process operation. The major variable costs, which are presented in detail in Table 7-D, consist of limestone, waste disposal, energy, and water. These values are based on operating the power plant at full load (65 MW). Limestone delivery contracts for the demonstration are short-term with the two suppliers located 150 and 250 miles from Richmond. This type of contract and distance from the site induced higher limestone prices. However, local limestone suppliers do not have the capacity or required quality for the demonstration. The price of limestone ranged from $26/ton to $37/ton. The average commercial value of limestone is about $15/ton. Ash disposal cost varied during testing. Disposal cost is dependent upon the waste management company and landfill used. The price ranged between $11/ton and $35/ton.

TABLE 7-D

VARIABLE OPERATING COSTS OF LIFAC SYSTEM AT RP&L

VARIABLE OPERATING COST (65 MW, NORMAL FLOW)				
Commodity	Unit	$/Unit	Quan./hr	$/hr
Limestone	tons	35.00	6	210.00
Reheat Steam	lbs	0.003	6180	18.50
Water	gals	0.00	6180	0.00
Aux. Power	kWh	0.02	360	7.20
Ash Removal	tons	17.00	5.4	91.80
Total Variable Operating Cost				327.50

Energy costs involve auxiliary power consumed by the process equipment and the reheating of exiting flue gas using medium pressure steam. The total connected horsepower for LIFAC is 736 kW. However, some equipment was not used continuously for the demonstration and the estimated average consumption of auxiliary power was approximately 362 kW. Steam energy is used to reheat the flue gas prior to entering the ESP. Nearly 100 lb/min. of medium pressure steam will be needed to increase the gas temperature 35°F. Water was provided by RP&L from the Whitewater River or the plant recirculation system.

7.3 Startup and Checkout Costs

The startup and checkout period was initiated shortly after baseline testing in September of 1992. Since the process is easy to start up and shut down, all checkouts were performed by operating the process for short periods of time. Most shakedown activities were accomplished by LIFAC personnel. Some equipment, especially the remaining EER units (from the LIMB tests), needed the attention of manufacturer representatives for calibration or repair. RP&L maintenance also helped to expedite the check out process. The approximate startup cost of the LIFAC process is shown in Table 7-E. The startup cost of the demonstration was about $200,000.

TABLE 7-E

ESTIMATED STARTUP COST OF THE LIFAC PROCESS
BASE YEAR: 1993

Startup Cost Element	Cost, $
Operating Labor Cost	$110,000
Maintenance and Materials Cost	38,500
Administrative and Support Cost	38,500
Commodity Cost:	
Limestone	6,250
Reheat Steam	5,400
Water	0
Power	1,600
Ash Removal	4,250
TOTAL	$204,500

Length of Startup Period, months	2

An extra set of gas analyzers were rented for calibration and verification. The limestone feeding system was calibrated by continuously filling a large bucket with limestone and observing the change in silo weight. It was discovered during the startup phase that some items needed modifications or replacement parts. Major modifications were made to the following items due to various operational problems. The ID fan's variable frequency drive failed several times; the flue gas dampers were sticking; the steam reheat condensate return system was not operational; and the water control valve and ESP recycle rotary feeders were the wrong type.

The process control system was calibrated via the process computers in the RP&L control room while LIFAC was operating at reduced flows.

Several training classes were held for RP&L personnel to educate them on the various principles and operating procedures of the LIFAC process. More detailed training was given during the course of the demonstration.

8.0 COMMERCIALIZATION PLANS

The LIFAC flue gas desulfurization (FGD) system has a relatively low overall cost due to simplicity of design and low operating costs. The limestone sorbent used with LIFAC is cheaper and material handling equivalent and possibly easier to handle than other FGD processes. Although LIFAC has been installed with new boilers, it is primarily a retrofit application. Existing power plants and industrial facilities do not have the rigid emission limits which are associated with new facilities. Also, existing plants have limited space available for FGD systems. A LIFAC unit can be installed with minimum space requirements and remove up to 85% of the sulfur dioxide from a plant's emissions.

LIFAC systems have been designed for coal-fired boilers ranging from 25 to 300 MW. The number of units needed is dependent upon the amount of flue gas generated and the size of the ESP at the facility. Often, a 300 MW boiler will require two LIFAC units to treat two exiting gas streams (150 MW each). Based on Tampella's experience 150-160 MW LIFAC unit is the maximum size unit feasible, larger facilities develop engineering problems when trying to evenly distribute the gas volume flow in the humidification zone. The LIFAC process is easy to operate from a process monitoring and control system located in the plant's control room. Additional controls (e.g., dampers) may be incorporated into control room panels or isolated from plant operations.

Limestone injection into the furnace may affect certain boiler operations, such as sootblowing frequency. However, its impact on the host is minimal. Sorbent injection rate is dependent on the sulfur content of the coal.

The ash by-product generated in the desulfurization process does not require additional treatment before dumping at a landfill. Thus, waste handling and disposal tends to be less expensive. The by-product is dry and may need to be moisturized to quell fugitive dust. The ash from some commercial installations has been used for concrete block production and in the mining industry.

There are currently several LIFAC units in operation on boilers burning various types of coal, from low lignite to high bituminous coal. The first full-size LIFAC installation treating high sulfur (2.5%) coal emissions is the RP&L Demonstration Facility. A listing of LIFAC installations worldwide is presented in Table 8-A.

TABLE 8-A

LIFAC INSTALLATIONS WORLDWIDE

Client - Plant	Country	Delivery Year	Boiler Size	Amount of Flue Gas to LIFAC Reactors	SO_2 % Removed	Ca/S Molar Ratio
Imatran Voima, Inkoo 4/1 Pilot	Finland	1986	250 MW	20%	76%	--
Nestre, Kulloo	Finland	1986	3 MW	100%	77%	2
Imatran Voima, Inkoo 4/2	Finland	1988	250 MW	50%	75-80%	2 and 2.5
Prommashimport, Baikal	Russia	1988	45 MW	100%	--	--
Imatran Voima, Inkoo 4/1	Finalnd	1989	250 MW	50%	--	--
Saskatchewan Power, Poplar River	Canada	1990	300 MW	50%	--	--
Richmond Power & Light, Whitewater Valley	USA	1992	60 MW	100%	55-77%	1.7 - 2.2
Saskatchewan Power, Shand 1	Canada	1992	330 MW	50%	--	--
Kostamuksha, Mining Combine	Russia	1994	230 MW	100%	--	--
Xiaguan Power Station	China	1995	125 MW	100%	--	--

Utilities are concerned about spending capital funds on FGD systems. LIFAC capital costs are significantly less than those of conventional wet lime and limestone scrubbers. The LIFAC system is relatively simple, employing a single patented vertical humidification reactor, utilizing a dry reagent, and producing a dry, safe, easy to handle by-product. Table 8-B shows a breakdown of total installed costs of the two most recent LIFAC installations: RP&L (65 MW) and the Shand Power Station (300 MW) in Canada. The Shand station has a 300 MW boiler that was fitted with only one activation reactor (150 MW); however, the costs are presented for both a single reactor and for two reactors, had it been necessary for compliance purposes. Also shown on the table is the estimated cost of a 300 MW wet scrubber for the Shand station. The capital costs for LIFAc range from about $100/kW for a 65 MW system down to about $66/kW for a 300 MW System. This reduction in cost is due to economies of scale. In comparison, the cost of a wet scrubber for a 300 MW unit is on the order of $205/kW. Also, estimated costs for a 150 MW LIFAC system are provided in Table 8-C.

At this point, it has been shown at RP&L and other LIFAC installations that system can be installed economically and operated without affecting normal power plant operations.

It is believed that LIFAC units become a less competitive option for power plant >500 MW. Based on an internal marketing study conducted by LIFAC North America, there are approximately 850 (less than 500 MW) potential units where the LIFAC technology could be applicable. It could also be noted that as long as emissions credits are around $200 per ton of SO_2 removed, the LIFAC units are not economically feasible. Emissions credits need to be in the $400/ton range to make LIFAC marketable.

TABLE 8-B

LIFAC INSTALLED COST COMPARISON

		Shand Station		RP & L
	Wet Scrubber (300 MW)	Two LIFAC Reactors (300 MW)	One LIFAC Reactor (150 MW)	One LIFAC Reactor (65 MW)
CAPITAL COSTS ($ Millions)				
Reactor and Auxiliaries	61.6	16.0	8.7	4.5
Sorbent Injection	incl.	0.7	0.7	0.4
Sorbent Processing	incl.	0.0	0.0	0.0
Delay ESP Duct Erection	incl.	0.4	0.4	0.0
Electrical Installation/Controls	incl.	0.7	0.7	1.1
Misc. Mechanical Installation	incl.	0.1	0.1	0.1
Air Compressor	incl.	0.2	0.2	incl.
Foundations	incl.	0.3	0.3	0.3
Ash System	incl.	0.2	0.2	incl.
Total	61.6	18.6	11.3	6.4
Cost ($/kW)	$205	$66	$76	$99

TABLE 8-C

ESTIMATED COSTS FOR 150 MW LIFAC SYSTEM FOR COMMERCIAL APPLICATION

Capital Costs ($85/kW)	1.5 Mills/kW
Limestone ($15-20/ton (2.0 Ca/S))	1.1-1.25 Mills/kW
Disposal ($15/ton)	0.9 Mills/kW
Other O&M	0.8-0.9 Mills/kW
Total	4.3-4.55 Mills/kW
SO_2 Removal (coal @ 4.0/lb/MBtu)	70
Amount Removed	2.8 lb/MBtu
$ / ton SO_2 Removed	$307-325

9.0 REFERENCES

1. *Baseline Environmental Monitoring Report* - LIFAC Sorbent Injection Desulfurization Demonstration Project presented to the U.S. Department of Energy, Pittsburgh Energy Technology Center, Pittsburgh, PA, October 13, 1993.

2. *LIFAC Demonstration at Richmond Power and Light Whitewater Valley Unit No. 2* - U.S. DOE Contract No. DE-FC22-90PC90548, March 1993.

3. *LIFAC Flue Gas Desulfurization* - Rose, J.; Hervol, J.; Viiala, J.; Koskinen, J.; Patel, J.; and Huffman, I., March 1992.

4. *LIFAC Process Evaluation Test Plan* - LIFAC Sorbent Injection Desulfurization Demonstration Project presented to U.S. Department of Energy, Pittsburgh Energy Technology Center, Pittsburgh, PA, August 1992.

5. *Operations and Post-LIFAC Tests Environmental Monitoring Report* - LIFAC Sorbent Injection Desulfurization Demonstration Project presented to the U.S. Department of Energy, Pittsburgh Energy Technology Center, Pittsburgh, PA, October 31, 1994.

6. *Parametric and Optimization Tests Environmental Monitoring Report* - LIFAC Sorbent Injection Desulfurization Demonstration Project presented to the U.S. Department of Energy, Pittsburgh Energy Technology Center, Pittsburgh, PA, April 1, 1994.

7. *Final Report Volume 1: Public Design* - LIFAC Sorbent Injection Desulfurization Demonstration Project presented to U.S. Department of Energy, Pittsburgh Energy Technology Center, Pittsburgh, PA, April 1997.

APPENDIX I

THE EFFECTS OF THE LIFAC SO_2 CONTROL PROCESS ON THE PERFORMANCE OF THE WHITEWATER VALLEY UNIT NO. 2 ESP

The Effects of the LIFAC SO_2 Control Process on the Performance of the Whitewater Valley Unit 2 ESP

Topical Report on EPRI Project 3005-1
January 1995

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1

INTRODUCTION

Among the options available to utilities for controlling SO_2 emissions are the so-called "dry" SO_2 control processes which produce dry waste products composed of fly ash combined with reacted and unreacted sorbent products (calcium or sodium hydroxides, sulfites, and sulfates). These processes all involve introduction of large quantities of calcium or sodium sorbents which must be collected in the downstream particle control device along with the fly ash. The change in dust characteristics will influence the performance of particulate control systems. When such systems were initially introduced, complex and contradictory effects were sometimes observed, particularly on electrostatic precipitators (ESPs). Accordingly, the Electric Power Research Institute (EPRI) established research project RP3005-1 to investigate these effects and to develop guidelines for the use of particulate control downstream from SO_2 control processes.

One SO_2 control system investigated during this project was the Tampella LIFAC process (Limestone Injection into the Furnace with downstream Activation of Calcium). The system was installed at Whitewater Valley Unit 2 of Richmond Power and Light in Richmond, Indiana. The LIFAC demonstration was conducted under the Department of Energy Clean Coal Technology Program. EPRI funded the detailed evaluation of the effects of LIFAC on the downstream ESP, which is the subject of this report.

The LIFAC process uses calcium-based sorbent (limestone) injection into the furnace with a reactivation zone downstream of the air heater to combine the effects of both low and high temperature reactions between the calcium and flue gas SO_2. The high temperature reaction occurs at furnace exit temperatures (1800-2200°F) similar to simple furnace sorbent injection, which has also been evaluated during this project.[1] The reactivation occurs through cooling the flue gas to temperatures close to adiabatic saturation using water sprays into a large vertical reaction chamber. Reheat is used downstream of the reactivation chamber to provide temperatures of 170 to 200°F to the particle control device. Fly ash and partially-sulfated sorbent are recycled from the ESP hoppers to the activation chamber to improve sorbent utilization. Because of the low temperatures which result from the high degree of humidification, the effect of LIFAC on ESP performance is similar to other low-temperature processes, such as spray dryers.[1]

Over the span of this project, two separate test programs were conducted on the LIFAC/ESP system. During the initial phases of the work at Whitewater Valley, high opacity during LIFAC startup impeded operation at the desired temperatures and sorbent addition rates. A limited test program was conducted in April 1993 to evaluate the cause of the high opacity and to

recommend solutions. Those results and recommendations were contained in a previous report.[2] The conclusions from that effort were:

- Electrical reentrainment of previously collected fly ash particles from the ESP was responsible for the transient opacity excursions during startup. Reduction of gas temperature should be done gradually over several hours or days to minimize the transient effect.

- Mixtures of sorbent and ash produced by the LIFAC process were found to have higher tensile strength than the Whitewater Valley fly ash alone. From this, it was inferred that long-term LIFAC operation might result in better ESP performance than that observed during the transient periods of short-term parametric tests.

- It was concluded that the reentrainment effect was dependent on temperature and that using additional reheat to maintain ESP temperature above 200°F should provide acceptable long-term performance. The use of a hot-gas bypass from the air heater inlet to the ESP inlet was suggested to obtain the additional reheat.

- Adjustment of the gas flow distribution in the ESP to reduce the very high gas velocities at the bottom of the ESP plates would also help to minimize the detrimental effects of particle reentrainment.

Following implementation of these recommendations, it was possible to operate the LIFAC system under the conditions necessary for good SO_2 removal and still achieve acceptable ESP performance. After a period of extended, continuous LIFAC operation to season the ESP, a comprehensive test of ESP performance was conducted in June of 1994. The results of this comprehensive test are the subject of this report.

Section 2 of this report will provide a discussion of the significant results from the comprehensive test program, although some comparison to previous data will be made. Section 3 will discuss the results of computer modeling and analysis of the ESP during LIFAC, while Section 4 will provide a summary of the significant findings. Cited literature references are listed in Section 5.

2

TEST RESULTS

The operation and performance of an ESP can be significantly affected as a result of the upstream use of a low-temperature, SO_2 removal process. Among the changes with which an ESP must cope are:

- Increased inlet particle mass loading.
- Decreased flue gas temperature, gas volume flow, and gas viscosity.
- Modified particle size distribution.
- Reduced dust resistivity.
- Increased non-rapping reentrainment of dust from the ESP collecting plates.
- Increased rate of accumulation of discharge electrode deposits.

Most of these changes will act to increase particle emissions from the ESP, although some are beneficial and others can go either way depending on the magnitude of the change. Measurements were made during this program to quantify all of these effects. The specific mechanisms that affect ESP performance associated with these changes will be discussed in detail in the subsections covering the measurements that illustrate them.

Process Conditions

During the three-day test program, the LIFAC process and the other plant systems were operated, to the extent possible, under constant conditions during testing. The LIFAC system used an average fresh limestone Ca/S stoichiometric ratio of 2.0 along with recycle from the ESP hoppers at a rate approximately equal to fresh limestone addition. Boiler load was held close to 60 MWe during testing. One change was purposefully made on the last day of testing when the amount of reheat was lowered and the ESP gas temperature reduced by 12°F. Table 1 provides a summary of the significant operating conditions for each test day.

The coal supply was restricted to a single mine for the test program. However, the coal analyses shown in Table 2 indicate that the coal sulfur content varied significantly, decreasing from a high of 2.1% on the first day to below 1.6% by the end of the tests. Despite this change, since the

LIFAC process operated consistently, adding essentially the same amount of sorbent and recycle, the effect of the sulfur variation on the ESP should not be significant.

Particle Mass Concentration and ESP Collection Efficiency

EPA Method 17 was used to quantify the total mass collection performance of the ESP during LIFAC operation, in addition to collecting a number of important pieces of information on the mechanisms that control ESP performance. The results of measurements that were made simultaneously at the inlet and outlet of the ESP on two of the test days are shown in Table 3. The table contains particle mass loadings, gas volume flows, gas temperature, gas moisture and gas oxygen content.

The mass loading measured at the ESP inlet with LIFAC was 17.15 lb/10^6 Btu. This is approximately a factor of 2 higher than the value measured with LIFAC during the 1993 test program, which is consistent with the lack of recycle and lower limestone addition rate (92 lb/min) during that earlier program. Based on the average ash content of the coal (8.8%), the EPRI data base[7] predicts a fly ash mass loading for a bituminous coal of 4.9 lb/10^6 Btu, indicating that LIFAC increased inlet mass slightly more than a factor of 3 over that expected with fly ash alone. The one measurement made with soot blowers in service gave a time-averaged value approximately 10% higher than the other inlet results. The correlations published in the EPRI SO_2/particulate control guidelines[1] calculate an inlet mass loading of 18.3 lb/10^6 Btu for the operating conditions in Table 1. This prediction is in reasonable agreement (<7% error) with the measured value.

There is a discrepancy between the gas temperatures measured during LIFAC with the ESP outlet hotter than the inlet by an average by 17°F. Extensive calibration and direct comparison of the two measurement systems did not resolve the difference. Some of the temperature difference might be attributed to continuation of the exothermic reaction between the flue gas SO_2 (or CO_2) and the unreacted lime in the ESP, impaction of damp particles or unevaporated water droplets on the thermocouples at the ESP inlet, or to compressive heating in the ID fan. During the 1993 test, a similar temperature increase was observed from inlet to outlet during all water injection conditions (some, but not all, including sorbent addition), but not under baseline conditions. This suggests that most of the difference is caused by thermocouple effects at the ESP inlet and that the outlet value is the most appropriate temperature to correlate with ESP performance. Some gas flow differences were also observed between inlet and outlet that are probably related to the relatively poor sampling location at the ESP inlet. Since the outlet sampling location is much more conducive to collection of representative samples than is the inlet location, the outlet values will be considered correct and used for ESP analysis.

The gas temperature in the ESP was somewhat higher during this test than is typically expected downstream of a low-temperature SO_2 control process. Most particle control devices in such systems operate in the range of 165°F to 180°F rather than the 197°F to 210°F at Whitewater Valley. The higher temperature is a result of the additional reheat added to reduce particle reentrainment from the ESP. The particle reentrainment is thought to occur because of the low

electrical resistivity of the dust (resulting in low electrical clamping force in the ESP) and low tensile strength of the dust layer.[1] Both of these effects occur because of the low gas temperature and high relative humidity resulting from the low-temperature SO_2 control process. The effect of this reentrainment on ESP performance will be considered further in subsequent sections.

Table 4 shows a number of ESP performance indicators calculated from the inlet and outlet mass data of Table 3. The average collection efficiency was 99.3% with a particle emission rate of 0.119 lb/10^6 Btu. As indicated by the high values of the Matts-Ohlnfeldt particle migration velocity (ω_k), this level of performance is quite good for an ESP with a modest specific collection area (SCA) of 198 ft^2/1000 acfm. Despite the 12°F lower ESP temperature on 6/9/94, there was no obvious effect on ESP emissions or collection efficiency. Apparently, even 198°F was sufficiently high to limit particle reentrainment to reasonable levels.

Note that the opacity value shown in the table is measured in a stack which is common to both Units 1 and 2. This value is useful in assessing compliance with environmental regulations, but is only a gross indicator of Unit 2 ESP performance with LIFAC.

Table 5 shows the ESP performance measured during the 1993 LIFAC test program. A value of ω_k similar to the 1994 LIFAC test was measured during the 1993 baseline test, although the emissions and efficiency were different. The similarity of these values indicates that the ESP was charging and collecting particles with the same proficiency during both programs. Therefore, the difference in efficiency (99.05% baseline vs. 99.3% LIFAC) is consistent with the change in SCA from 181 to 198 ft^2/1000 acfm. The change in SCA was caused by the reduction in gas volume flow resulting from the lower temperature with LIFAC. Despite the higher collection efficiency with LIFAC, the greatly increased inlet mass loading resulted in increased particle emissions. The ESP collection performance with LIFAC will be further analyzed in the subsequent section on ESP modeling.

Particle Size Distribution and Fractional Efficiency

Particle size distributions entering and exiting the ESP during LIFAC were measured on the second day of the test program using cascade impactors. The size distributions measured at the ESP inlet during both the most recent test and the 1993 program are shown on a cumulative mass basis in Figure 1. The open symbols represent the baseline fly ash distribution and the solid symbols show the LIFAC data. The error bars associated with the symbols are 90% confidence intervals to the mean distribution.

The two LIFAC size distributions are essentially identical with two exceptions: 1) a lower concentration of particles smaller than 0.5 μm was measured during the 1994 test, and 2) a higher total mass concentration was measured during the 1994 test. The lower concentration of submicron particles appears to be related to differences in the coal since the 1993 test indicated that the LIFAC and baseline distributions were identical in this range, and by the fact that the 1994 LIFAC distribution falls below the baseline data in this range. Collection of particles this small in the activation chamber is unlikely to be responsible for the lower concentration. As

discussed previously, the higher total mass loading during the 1994 test resulted from the addition of recycle material to the system and a higher limestone addition rate. The size distributions indicate that the recycle mass was mainly associated with particles or particle agglomerates larger than 10 μm. These large particles should be relatively easy to collect in the ESP and will contribute to higher overall collection efficiency.

Comparison of the LIFAC and baseline data indicate that the LIFAC process contributed only particles larger than 0.8-1.0 μm to the gas stream and that only 2% of the mass was smaller than 2 μm. Figure 2 shows the cumulative mass distribution for sorbent only calculated from the difference between the LIFAC and baseline distributions. This distribution is somewhat finer (more small particles) than measured for spray dryers, but somewhat coarser than for furnace injection of hydrated lime at Edgewater which contained significant numbers of particles down to 0.2 μm.[1] Particles in the 0.5 to 2.0 μm range contribute more to opacity per unit mass than do larger particles. Therefore, for constant collection efficiency, an increase in these particle sizes would be expected to increase opacity.

The small particles contributed by the LIFAC process will affect ESP performance by increasing the total suspended space charge in the ESP electrode space. The increased space charge can have both positive and negative effects on the ESP. Moderate increases in space charge will change the electric field profile in the electrode space resulting in increased electric field adjacent to the collecting plate and increased particle collection. However, excessive increases in particle loading can result in corona current quenching and degraded particle collection performance. The concentrations of small particles contributed by the LIFAC process were probably not high enough to cause corona quenching, but this will be evaluated further in a later discussion of ESP electrical operation.

Figure 3 shows the ESP collection efficiency as a function of particle size calculated from the inlet and outlet particle size distributions. These data indicate that the lowest collection occurred for particle sizes of 0.2 μm with increasing efficiency for larger particles. This behavior is normal and is consistent with ESP theory. The particle agglomerates reentrained during electrode rapping typically have a mean size of 6 μm,[3] producing a dip in the efficiency curve in that region. Although there is a slight dip in the efficiency curve, there is no evidence in these data of excessive reentrainment. However, because of the low tensile strength of the dust at low temperature, the particle agglomerates could decrepitate thus reducing efficiency over a wide size range. If this were the case, the effect of reentrainment might not be discernable in the size distribution data.

Dust Chemistry

Samples of the dust produced by the LIFAC process were collected from the ESP hoppers on each day of the test program. Chemical compositions of the samples are shown in Table 6. The samples collected from the inlet and outlet hoppers were analyzed separately for one test day, while on the other days proportional blends were prepared. The proportions from each hopper used in the blends were calculated from ESP theory to reproduce the dust composition in the

inlet gas stream.[3] Also shown in the table is the average baseline fly ash composition from the 1993 test.

The most striking feature of the dust chemistry is the high level of calcium (~40%) contributed by the LIFAC process. Although no samples were obtained during the most recent tests, the baseline fly ash in 1993 had a CaO content of only 2.5%. Comparison of the inlet and outlet hopper samples indicates that a higher fraction of the sorbent particles are collected in the inlet field of the ESP. The drop in CaO from 41% to 26% from inlet to outlet is consistent with the relatively larger particle size distribution of the sorbent compared to the fly ash. The effect of the chemical composition on resistivity and ESP performance will be discussed in the following subsection.

One of the most significant degrading effects of low-temperature SO_2 control systems on ESP performance results from the reentrainment of dust from the ESP electrodes. This is thought to be a function of both low resistivity and low dust layer tensile strength. High chloride content in the dust has been associated with reducing the reentrainment problem and improving ESP performance, presumably by making the dust more adhesive. However, the chloride content of the samples shown in Table 6 are probably not sufficiently high to produce this effect. Chloride content over 1% has been required at other sites to improve performance.[4] The use of chloride compounds to improve performance is not generally desirable because of the increased corrosion which can occur.

Dust Resistivity

One effect generally expected of a low-temperature SO_2 control process is low dust resistivity that allows excellent ESP electrical operation. When such a process is applied to a fly ash that has an inherently high resistivity under baseline conditions, ESP performance can actually be improved. Dust resistivity data collected in-situ with a point-plane resistivity probe are shown in Table 7. The resistivity values obtained from both methods of measurement with the probe (spark and V-I) are included in the table. The average resistivity measured with the spark method was 2×10^{10} ohm-cm, while most values obtained with the V-I method were below the lower limit of resolution of the technique at ~5×10^{9} ohm-cm. Both techniques have strengths and weaknesses, but the V-I method may be more valid in this case because of the characteristics of this dust. The dust cakes collected in the probe were fluffy and compressible, potentially causing poor contact with the movable electrode during the spark measurement. Thus the higher value with the spark measurement may actually be a result of high resistance in the interface between the dust layer and movable electrode. Regardless of which technique is considered, the resistivity is low enough not to significantly limit ESP operation.

The temperature measured with the resistivity probe was somewhat lower than those measured at the ESP inlet possibly because of incomplete mixing of the reheat gas and the discharge from the reactivation chamber or due to cooling of the probe by damp particles or residual water droplets. The resistivity measurements were made only a few feet downstream of the reheat gas entry point.

Figures 4 and 5 show the results of laboratory resistivity measurements made on the hopper samples from the test program. Figure 4 was measured in ascending-temperature mode, while the data in Figure 5 were collected in descending-temperature mode. As frequently seen with mixtures of ash and sorbent, the descending-temperature data indicate higher resistivity than the ascending data. Obviously some chemical changes occur in the sample at higher temperature, but it is not clear which condition is most representative of the conditions which exist in the ESP. As one might expect from the dust chemistry, only the outlet hopper sample is significantly different from the other samples. However, at the temperature of the Whitewater Valley ESP there is no significant difference between any of the samples.

If the temperature of the resistivity probe (173°F) accurately described the conditions of the dust layer in the resistivity probe, the resistivity could be considerably higher than the in-situ measurements indicated at the ESP outlet temperature (200°F). Figure 6 compares the in-situ and laboratory measurements for the 6/9/94 proportional blend. If both the spark and V-I methods are considered, the in-situ measurements, which should be most representative of the conditions experienced by the ESP, can be considered to agree with either of the lab measurements. If the lab measurements are used to estimate resistivity for the ESP outlet temperature, quite different results are obtained for the two curves. At 200°F, the descending data indicate a resistivity value of 5×10^{11} ohm-cm while the ascending curve gives a value of 1.5×10^{10} ohm-cm. The lower value would result in excellent ESP performance, but the higher resistivity would significantly degrade ESP performance. In the following subsection, the ESP electrical conditions will be used to assist in this analysis.

Also shown on Figure 6 are resistivity predictions made by two techniques developed by Roy Bickelhaupt.[5,6] As expected for ash/sorbent mixtures, there is poor agreement between the resistivity vs. temperature measured in the laboratory and that which is predicted. This is despite the fact that the resistivity predictions are based on correlations between lab measurements and ash chemistry. Our experience on this project has indicated that both resistivity models do a poor job on ash-sorbent mixtures. The very optimistic estimates of resistivity of ash-sorbent mixtures, particularly with high-temperature processes, makes the use of these models dangerous when estimating ESP performance effects.

ESP Electrical Operation

High dust electrical resistivity affects ESP performance primarily by limiting the useful power input to the ESP. The reduced voltage and current that occur with high resistivity result in poorer charging and collection of particles and increased emissions. The average electrical operating conditions of the ESP with LIFAC are shown by field in Table 8. As expected, the voltage is highest and current lowest in the inlet field because of the space charge associated with the high particle loading. Voltage is reduced and current increased toward the outlet of the ESP as the particles are collected and the space charge effect is reduced. This effect is typical of all ESPs and does not indicate a serious corona quenching problem resulting from the LIFAC process. The current density of 18 nA/cm^2 observed in the inlet field is typical of fly ash ESPs and will provide acceptable performance.

No sparking was occurring in any of the fields indicating that they were at some artificial limit and not controlled by the dust layer characteristics. The outlet field was at voltage limit, while the other fields were at the current limit of the power supplies. Normally, the current densities measured during LIFAC (20-25 nA/cm^2) would be considered to represent a slight resistivity limitation, since most ESPs have current limits in the range of 50 nA/cm^2. However, a current limit of 25 nA/cm^2 is typical of the Lodge-Cottrell ESP design used at Whitewater Valley.

Even though relatively high values of voltage and current were observed in the ESP, back corona caused by high resistivity could still degrade ESP performance. Figure 7 shows voltage-current curves measured on the ESP on the last day of the test program. These curves show no evidence of back corona or other anomalous electrical behavior. Back corona would be indicated by infinite or negative slope in the V-I curve. Therefore, the average electrical conditions in Table 8 are completely useful for charging and collecting particles and should be representative of ESP performance.

Figure 8 compares the current density values during LIFAC with a correlation between resistivity and current density developed from the EPRI data base.[7] The correlation generally confirms that the electrical conditions, at least up to the point where current limit is reached, are consistent with the in-situ measurements of dust resistivity. The high resistivity indicated by the descending-temperature lab resistivity data at 200°F is not consistent with the ESP operation.

One potential detrimental effect of low-temperature processes on ESP operation is the accumulation of excessive dust deposits on the discharge electrodes. Such buildups could quench corona current causing reduce charging effectiveness and degrading collection efficiency. Although the rapid accumulation of deposits has been observed with low-temperature SO_2 control on a pilot-scale ESP, [9] problems have not been reported with full-scale installations. The lack of reported problems may imply that the deposits are removed by normal electrode rapping. As discussed above in the context of space charge corona suppression, the electrical conditions at Whitewater Valley do not indicate any such problems specific to the LIFAC process. There is no evidence of excessive electrode deposits during LIFAC operation.

3

ESP PERFORMANCE MODELING

Version 3 of the SRI Mathematical Model of ESP Performance[6] was used to simulate the Whitewater Valley Unit 2 ESP under both baseline and LIFAC conditions. Modeling of the performance of the ESP under baseline conditions was done to calibrate of the model and to determine if the performance of the ESP was reasonable in the absence of the complicating effects of LIFAC. Modeling with LIFAC illustrates the mechanisms through which LIFAC affects ESP performance.

For modeling the baseline condition, two sets of model non-ideal conditions were used to estimate a range of expected performance. Two non-ideal conditions (s, σ_g) are used in the model to account for the combined effects of sneakage and reentrainment (s) and for the uniformity of the gas flow distribution in the ESP (σ_g). Sneakage describes the particles which bypass each electrified section of the ESP by flowing through hoppers and over the tops of plates, while reentrainment accounts for particles which are collected, but subsequently reentrain into the gas stream. The σ_g parameter is the coefficient of variation of the gas velocity profile in the ESP. For ESPs collecting fly ash, the performance range of typical ESPs can be represented by modeling with a lower limit of s and σ_g of 0.05 and 0.15 and an upper limit of 0.10 and 0.25.[7]

During the previous work in 1993, the baseline ESP model calculations could not be made to match the measured baseline performance. At that time, we speculated that incorrect voltage values displayed on the ESP power supply controls were probably responsible for the disagreement. Those voltages (labeled as "original" in Table 9), were considered too low for the current densities and electrode spacing of the Unit 2 ESP. During the recent test, the indicated voltage values were compared to a precision high-voltage probe and it was established that three of the four fields were indicating values that were low by up to 6.9 kV. The "corrected voltage" values in Table 9 are based on the assumption that the same voltage error was present in the 1993 data. The corrected baseline voltage values were used in the ESP model. The ESP control panel readouts were calibrated to the corrected values.

The hardware specifications of the Unit 2 ESP are shown in the top portion of Table 10. Other model inputs specific to the baseline operating conditions (documented in the previous report) are included in the lower portion of Table 10. In addition to the measured gas volume flow, the ESP was also modeled with 110% and 60% of the measured gas flow to simulate a range of performance for a typical range of boiler loads. The baseline model calculations are shown in Table 11 and plotted as particle emissions vs. SCA in Figure 9. The lower and upper limits of the shaded region on the figure represent model calculations with the better and poorer sets of typical non-ideal conditions described above. The ESP emission rate measured during the 1993

baseline test is indicated on the figure by the solid circle. In contrast to the poor agreement during the 1993 modeling, the model does a good job of predicting baseline performance with the corrected electrical conditions. Our ability to accurately predict baseline performance provides extra confidence in the LIFAC predictions.

Because of the increased reentrainment associated with low-temperature SO_2 control processes, it is generally been necessary to use higher values of the ESP model's non-ideal conditions to accurately simulate actual performance.[1] The range of values established by the EPRI guidelines are a lower limit of $s = 0.25$, $\sigma_g = 0.25$ and an upper limit of $s = 0.40$, $\sigma_g = 0.25$. The ESP performance with LIFAC was modeled under both the non-ideal conditions typical of fly ash and these poorer values from the guidelines. The model results are shown in Table 12 and Figure 10 for a typical range of gas flow values. The measured emissions rates from both the 1993 and 1994 tests are also shown on the figure. The measured values fall between the two sets of non-ideal conditions indicating higher emissions than expected for fly ash, but lower than for most other low-temperature SO_2 control processes. This suggests that some increased reentrainment was occurring in the Unit 2 ESP with LIFAC, but that it is not as severe as expected. The improved performance with LIFAC at Whitewater Valley is attributed to the higher flue gas temperature resulting from the additional reheat.

Although it could not be demonstrated at this site because of the need to comply with opacity regulations, we expect that if the ESP temperature were reduced to 170°F the performance of the ESP would degrade substantially. The lower level of reentrainment at Whitewater Valley should not be taken to mean that the LIFAC process has a significantly different effect on ESP operation than do other low-temperature SO_2 control processes. We expect that reduced reentrainment would also be observed for other low-temperature processes if additional reheat were used to increase ESP temperature to 200°F. Therefore, if ESP temperatures will be significantly below 200°F, we recommend a conservative approach to estimating the effects of LIFAC on ESP performance by using the ESP model non-ideal conditions developed for spray dryers ($s = 0.25$, $\sigma_g = 0.25$ and $s = 0.40$, $\sigma_g = 0.25$) and documented in the EPRI guidelines.[1]

4

Summary and Conclusions

Comprehensive tests were conducted of the effects of the Tampella LIFAC SO_2 control process on the performance of the Whitewater Valley Unit 2 ESP. The following summary statements and conclusions can be draw from the data collected.

- The LIFAC system operated reliably throughout the test program and provided consistent ESP conditions for the performance evaluation. Although the coal sulfur content did vary from 2.1% to 1.6%, since the LIFAC system operation was consistent, ESP performance should not have been biased by the sulfur variation.

- The LIFAC process produced a particle mass loading entering the ESP of 17.15 lb/10^6 Btu. This loading is a factor of 3.5 higher than expected for fly ash alone, and is in reasonable agreement with the predictions of the EPRI guidelines.

- The sorbent particles contributed to the ESP inlet gas stream were generally larger than 0.8 μm in diameter. The LIFAC distribution contained fewer submicron particles than were produced by hydrated lime injection into the Edgewater furnace, but more than for a typical spray dryer. There were insufficient submicron particles to cause corona quenching problems, but the particles in the 0.5 - 2.0 μm range will contribute to opacity independently of mass emissions.

- ESP temperature was in the range of 198-209°F during the tests, which is 20-40°F higher than typical of low-temperature SO_2 control processes. The higher temperatures were thought necessary to reduce reentrainment of particles from the ESP to acceptable levels. Although reduction in temperature from 209°F to 198°F did not measurably increase emissions, it was generally believed that further temperature decreases would degrade ESP performance. This hypothesis was not tested because of concerns over opacity limit violations.

- ESP performance with LIFAC was quite good with particle collection efficiency of 99.3% and a particle emission rate of 0.119 lb/10^6 Btu. Baseline measurements in 1993 indicated 99.05% efficiency and 0.049 lb/10^6 Btu. This is quite impressive performance for an ESP with a modest SCA below 200 ft^2/1000 acfm. Despite the improved collection efficiency with LIFAC, the increased inlet mass resulted in increased outlet emissions.

- Fractional efficiency data did not indicate a distinct dip in efficiency in the size region that is typical of particle reentrainment. However, decrepitation of the low-tensile-strength particle

agglomerates could prevent the effect of reentrainment from being detected by this technique.

- Dust resistivity data were scattered and confusing as we have come to expect for these processes. The results of the V-I method of the in-situ measurement are probably the most appropriate values and generally indicated low resistivity. The ascending-temperature lab measurement also predicted low resistivity at the higher temperature of the ESP outlet. At the ESP outlet temperature, the descending-temperature lab measurement predicted high resistivity that was not consistent with ESP electrical operation.

- ESP electrical operation was good, consistent with low resistivity values. There were no indications of high resistivity limitations or of corona quenching from excessive space charge or excessive discharge electrode deposits. Although current densities were lower than typical for the outlet fields, this is attributed to the relatively low current limit of Lodge-Cottrell ESPs.

- Calibration of the ESP power supply voltage meters indicated that the readings were low in three of the four fields by up to 6.9 kV. The low indicated values are consistent with our inability to model the performance of the ESP under baseline conditions in 1993. The control readouts were calibrated during the 1994 test.

- With corrected voltage values, the ESP model predicted a range of baseline performance that bracketed the 1993 baseline measurements. This indicates that the ESP is operating as expected for its hardware and operating conditions and that no anomalous conditions or hidden problems exist.

- Because of the increased reentrainment with low-temperature SO_2 control processes, higher values of the non-ideal parameters in the ESP model are generally required to accurately simulate ESP performance. However, under the conditions tested at Whitewater Valley, the measured ESP performance fell between the predictions for fly ash only and those made with the poorer non-ideal conditions typical of low-temperature operation. This suggests that some increased reentrainment was occurring, but that it was not as severe as expected. The improved performance with LIFAC is attributed to an ESP operating temperature that was 20-40°F higher than typical of low-temperature SO_2 control processes. The higher temperature was used at Whitewater Valley specifically to limit the degradation from reentrainment and to maintain compliance with opacity limits. The recommendation made in the EPRI guidelines to design for high levels of reentrainment in ESPs installed on low-temperature processes is still valid for operation at temperatures of 180°F and below.

5

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THE EFFECTS OF THE LIFAC SO_2 CONTROL PROCESS ON THE PERFORMANCE OF THE WHITEWATER VALLEY UNIT 2 ESP

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Table 1
LIFAC Process Operating Conditions

Parameter	6/7/94	6/8/94	6/9/94	Avg.
Boiler Load, MWe	60	61	61	61
Calcium/Sulfur Stoichiometric Ratio[a]	2.2	1.8	1.9	2.0
Coal Sulfur Content, %	2.0	2.2	2.3	2.2
Coal Feed Rate, lb/min	943	952	912	936
Limestone Feed Rate, lb/min	141	122	132	132
Recycle Feed Rate, lb/min	133	133	133	133
Reactor Water Flow, gpm	64	63	68	65
Average Reactor Inlet Temperature, °F	341	337	346	341
Minimum Reactor Outlet Temperature, °F	136	136	136	136
Minimum Steam Reheater Outlet Temp, °F	166	165	164	165
Average ESP Inlet Temperature, °F	187	185	178	183
Reactor Outlet SO_2 Concentration, ppm[b]	560	743	488	597

a. Assumes constant coal sulfur content of 2.25%.
b. Corrected to 6% O_2 and boiler outlet H_2O content.

Table 2
Coal Chemical Composition

Component	6/7/94	6/8/94	6/9/94	Average
Sulfur Content, %	2.10	1.70	1.57	1.79
Ash Content, %	8.91	9.79	7.65	8.78
Moisture, %	10.97	10.17	11.98	11.04
Btu/lb Received	11602	11625	11659	11629

Table 3
Method 17 Mass Measurement Summary

Date	Mass Loading		Gas Flow		Gas Temp., °F	Gas H_2O, %	Gas O_2, %
	gr/acf	lb/10^6 Btu	acfm	dscfm			
ESP Inlet							
6/7/94	5.39	15.42	223,000	143,000	188	15.2	5.0
	5.71	16.37	223,000	143,000	190	15.6	4.9
	5.54	16.09	228,000	145,000	190	15.7	5.1
6/9/94	5.42	15.60	230,000	152,000	180	14.5	5.5
	7.59[a]	22.25[a]	235,000	153,000	182	15.8	5.5
	4.87	13.74	225,000	148,000	181	14.7	5.1
Average	5.93	17.15	228,000	147,000	186	15.4	5.2
ESP Outlet							
6/7/94	0.0312	0.092	212,000	136,000	210	15.6	5.3
	0.0397	0.114	217,000	139,000	209	14.7	4.9
	0.0443	0.133	212,000	135,000	209	14.9	5.6
6/9/94	0.0410	0.119	214,000	143,000	196	14.2	5.9
	0.0471[a]	0.138[a]	225,000	149,000	198	14.8	5.8
	0.0376	0.107	244,000	164,000	197	13.7	5.6
Average	0.0407	0.119	216,000	140,000	203	14.8	5.5

a. Measured with Soot Blowers in Service.

Table 4
ESP Performance Summary

Date	Collection Efficiency, %	Particle Penetration, %	ω_k,[b] cm/sec	ESP SCA, ft^2/kacfm	Particle Emissions, lb/10^6 Btu	Stack[c] Opacity, %
6/7/94	99.403	0.597	66	202	0.092	10.2
	99.305	0.695	62	201	0.114	10.6
	99.174	0.826	59	197	0.133	10.6
6/9/94	99.235	0.765	62	195	0.119	10.6
	99.380[a]	0.620[a]	69[a]	191[a]	0.138[a]	13.0[a]
	99.220	0.780	60	200	0.107	10.1
Average	99.299	0.701	64	198	0.119	10.9

a. Measured with soot blowers in service.
b. Effective particle migration velocity from Matts-Ohlnfeldt equation with k = 0.5.
c. Stack opacity measurement is common to Units 1 and 2. Relative indication only.

Table 5
ESP Performance Summary From 1993 Test

Condition	ESP Efficiency, %	ESP Penetration, %	ESP[a] ω_k, cm/sec	ESP SCA, ft^2/kacfm	Particle Emissions, lb/10^6 Btu	Stack[b] Opacity, %
Baseline	99.050	0.950	62	182	0.0490	10
Water Excursion	95.625	4.375	26	195	0.2192	22
LIFAC Excursion	96.056	3.944	27	198	0.3621	24
LIFAC Post-Exc.	99.195	0.805	58	208	0.0701	13

a. Effective Migration Velocity from Matts-Ohlnfeldt Equation with k=0.5
b. Measured in Stack with Both Units On-Line.

Table 6
Chemical Composition of ESP Hopper Dust

Oxide	6/7/94		6/8/94 Proportional Blend	6/9/94 Proportional Blend	Average 1993 Fly Ash
	Inlet Hopper	Outlet Hopper			
Li_2O	0.01	0.01	0.01	0.01	0.04
Na_2O	0.53	0.39	0.58	0.50	0.71
K_2O	0.74	1.19	0.83	0.60	2.4
MgO	1.60	1.31	1.64	1.77	0.93
CaO	41.07	25.54	39.11	40.79	2.5
Fe_2O_3	8.10	12.62	8.06	8.69	21.2
Al_2O_3	10.11	16.32	11.55	9.72	22.8
SiO_2	21.00	31.10	22.21	20.22	48.3
TiO_2	0.44	0.70	0.46	0.39	1.2
P_2O_5	0.28	0.34	0.34	0.29	0.48
SO_3	15.94	10.25	15.02	16.65	0.55
LOI	4.81	7.11	5.30	5.68	4.6
Cl^{-2}	0.16	0.52	0.14	0.13	--

Table 7
In-Situ Dust Resistivity Measurements

Date	Temp-erature, °F	Layer Thickness, mm	Resistivity, ohm-cm	
			Spark	V-I
6/7/94	175	1.03	3.8×10^{10}	4.6×10^{10}
	175	1.00	1.0×10^{10}	2.6×10^{10}
	175	1.07	3.5×10^{10}	$<5 \times 10^{9}$
6/8/94	176	1.57	1.8×10^{10}	$<5 \times 10^{9}$
	175	2.06	2.9×10^{10}	$<5 \times 10^{9}$
	177	2.02	[a]4.1×10^{11}	$<5 \times 10^{9}$
6/9/94	168	2.54	8.1×10^{9}	$<5 \times 10^{9}$
	172	1.51	8.6×10^{9}	$<5 \times 10^{9}$
	167	1.42	8.6×10^{9}	$<5 \times 10^{9}$
	170	1.52	1.6×10^{10}	$<5 \times 10^{9}$
Average	173		1.9×10^{10}	$<5 \times 10^{9}$

a. Outlier - excluded from average.

Table 8
ESP Electrical Operating Conditions

Date	ESP Field	ESP Voltage, kV	ESP Current, mA	Current Density, nA/cm^2
6/7/94	1	55.3	190	18.2
	2	50.9	220	21.1
	3	47.9	230	22.1
	4	47.8	250	24.0
6/8/94	1	55.7	197	18.9
	2	51.5	223	21.4
	3	48.6	237	22.8
	4	48.8	250	24.0
6/9/94	1	55.6	187	18.0
	2	51.9	217	20.8
	3	49.4	227	21.8
	4	49.2	247	23.7
Average	1	55.5	191	18.4
	2	51.4	220	21.1
	3	48.6	231	22.2
	4	48.6	249	23.9

Table 9
Baseline ESP Electrical Conditions

	ESP Field	ESP Voltage, kV	ESP Current, mA	Current Density, nA/cm^2
Original	1	43.4	180	17.3
	2	42.7	193	18.5
	3	47.1	203	19.5
	4	49.7	240	23.0
Corrected Voltage	1	47.7	180	17.3
	2	50.3	193	18.5
	3	50.9	203	19.5
	4	49.7	240	23.0

Table 10
ESP Model Input Data

ESP Hardware Specifications		
Fields in Direction of Gas Flow	4	
Collection Area per Field, ft^2	11,250	
Collecting Plate Spacing, in	11	
Collecting Plate Height, ft	30	
Collecting Plate Length, ft	30	
Number of Gas Passages	25	
Wire-to-Wire Spacing, in	9	
Number of Wires in Flow Direction	40	
Number of Baffled Sections	4	
Condition-Specific Model Input		
	Baseline	LIFAC
Inlet Mass Loading, gr/acf	1.61	5.93
Gas Volume Flow, acfm	248,000	216,000
Gas Temperature, °F	307	207
Gas Velocity, ft/sec	6.6	5.3
Gas Moisture Content, %	7.9	14.8
Gas Oxygen Content, %	5.6	5.5
Gas Viscosity, kg/(m*sec)	2.26×10^{-5}	1.95×10^{-5}

Table 11
Revised Baseline ESP Model Results

Data Source	ESP SCA, ft^2/kacfm	Model Non-Ideal Conditions	Collection Efficiency, %	Particle Emissions, lb/10^6 Btu	Stack Opacity, %
Modeled with increased Gas Flow	165	0.05, 0.15 0.10, 0.25	98.87 98.27	0.055 0.084	13 17
Modeled with Measured Gas Flow	181	0.05, 0.15 0.10, 0.25	99.13 98.61	0.042 0.068	11 14
Modeled with Decreased Gas Flow	299	0.05, 0.15 0.10, 0.25	99.84 99.66	0.008 0.017	2 4
Measured	182	N.A.	99.05	0.049	10*

* - Measured in common stack with Unit 1.

Table 12
ESP Model Results with LIFAC

Data Source	ESP SCA, ft^2/kacfm	Model Non-Ideal Conditions	Collection Efficiency, %	Particle Emissions, lb/10^6 Btu	Stack Opacity, %
Modeled with Increased Gas Flow	166	0.05, 0.15 0.10, 0.25 0.25, 0.25 0.40, 0.25	99.58 99.23 97.69 94.05	0.059 0.109 0.327 0.839	12 18 36 60
Modeled with Measured Gas Flow	207	0.05, 0.15 0.10, 0.25 0.25, 0.25 0.40, 0.25	99.80 99.56 98.37 95.38	0.029 0.062 0.230 0.652	6 11 27 51
Modeled with Decreased Gas Flow	310	0.05, 0.15 0.10, 0.25 0.25, 0.25 0.40, 0.25	99.96 99.86 99.26 97.46	0.006 0.019 0.105 0.358	1 3 13 35
Measured	207	N.A.	99.30	0.119	11*

* - Measured in common stack with Unit 1. Some dilution of the Unit 2 emissions.

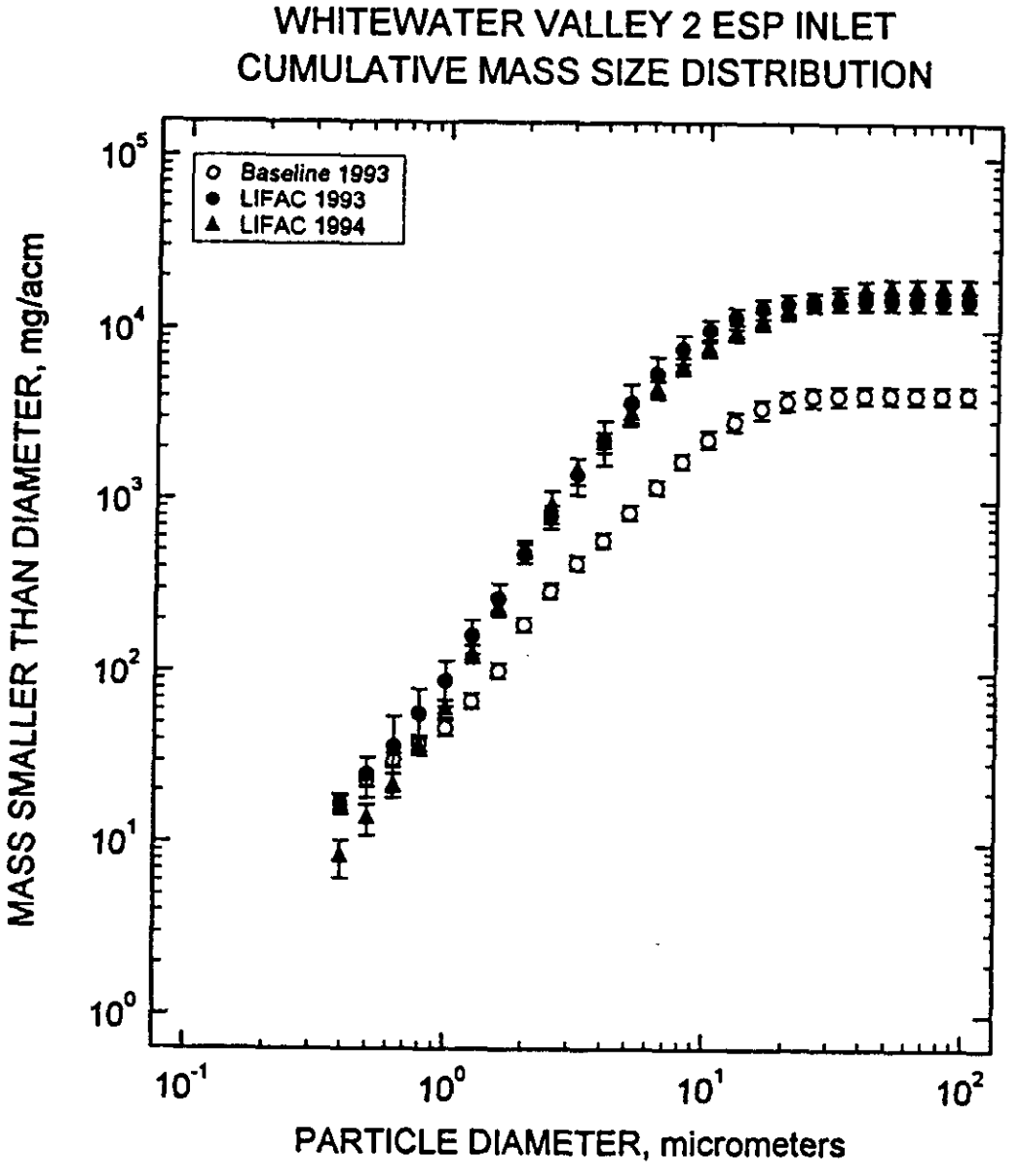

Figure 1
ESP Inlet Cumulative Mass Particle Size Distributions

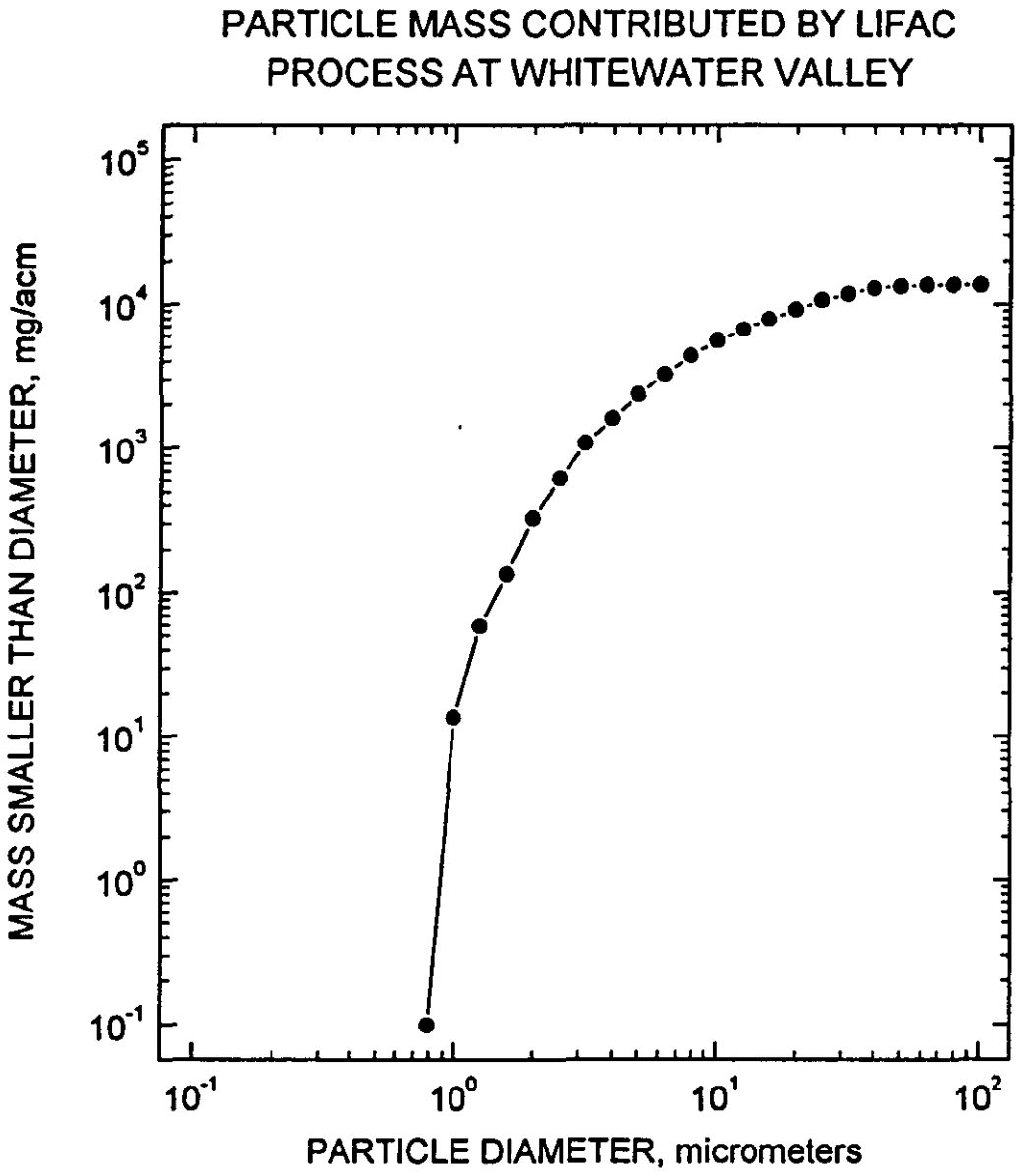

Figure 2
ESP Inlet Particle Mass Contributed by the LIFAC Process

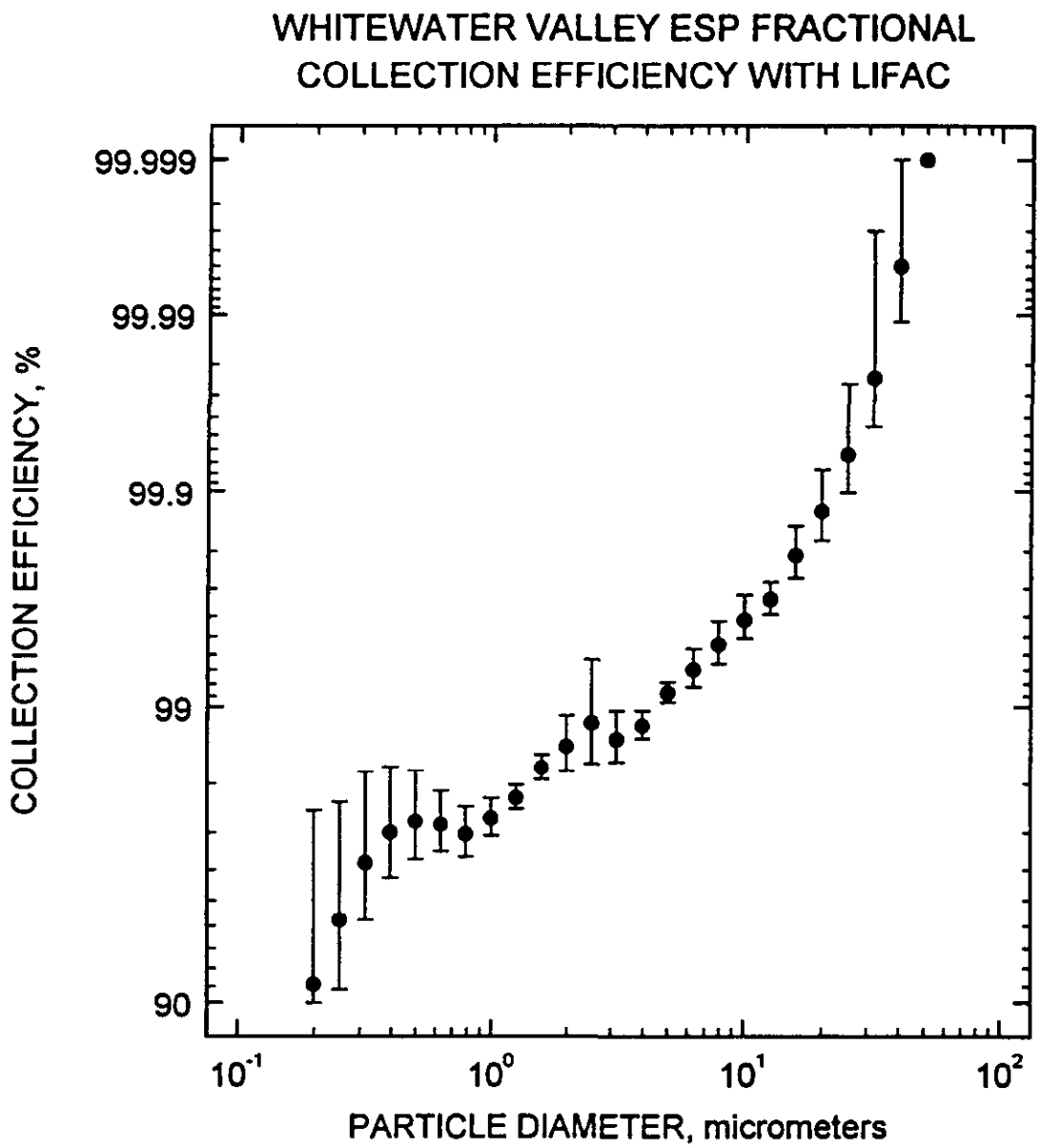

Figure 3
ESP Collection Efficiency as a Function of Particle Size During LIFAC

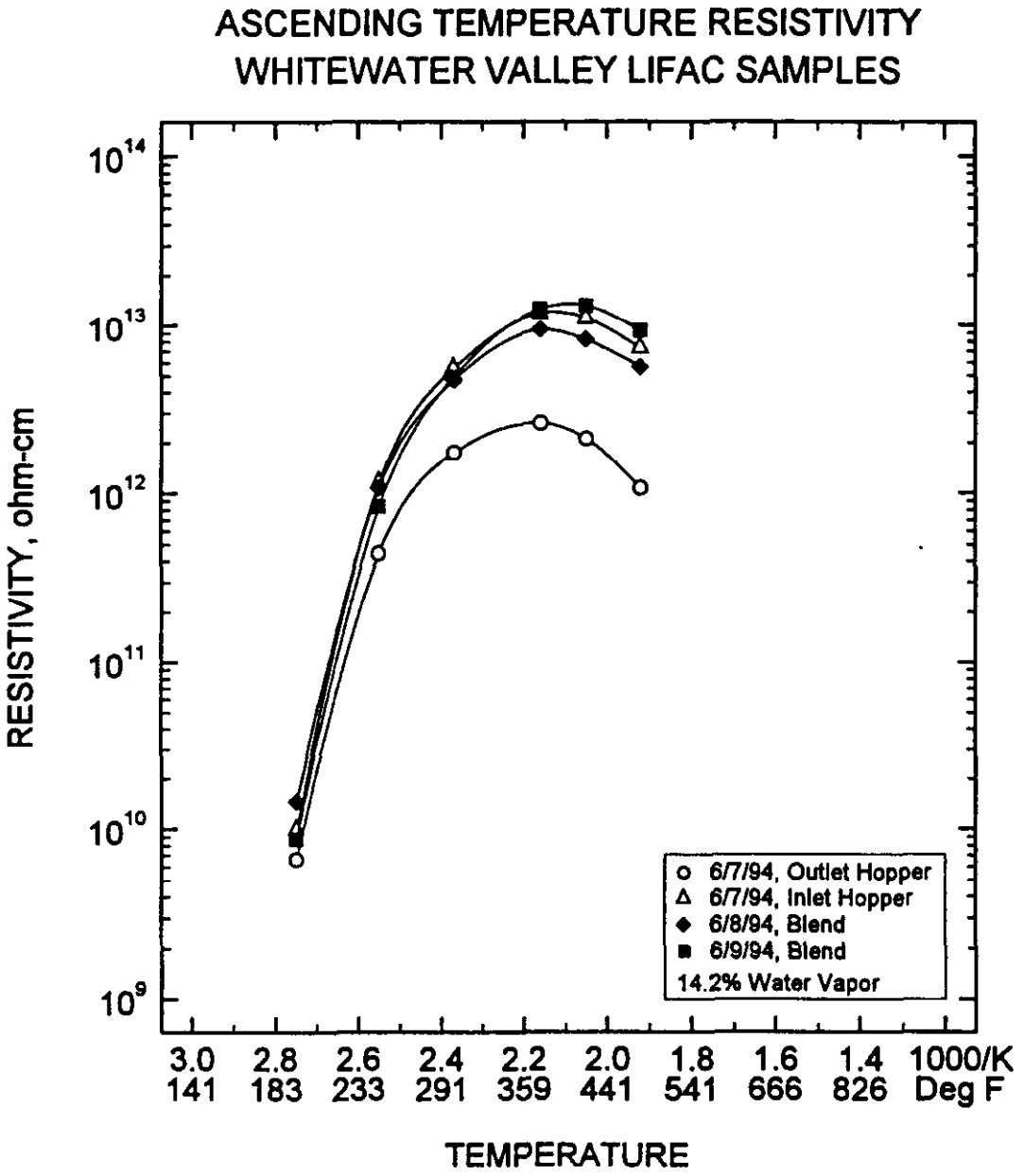

Figure 4
Ascending-Temperature Resistivity Measurement on LIFAC Samples

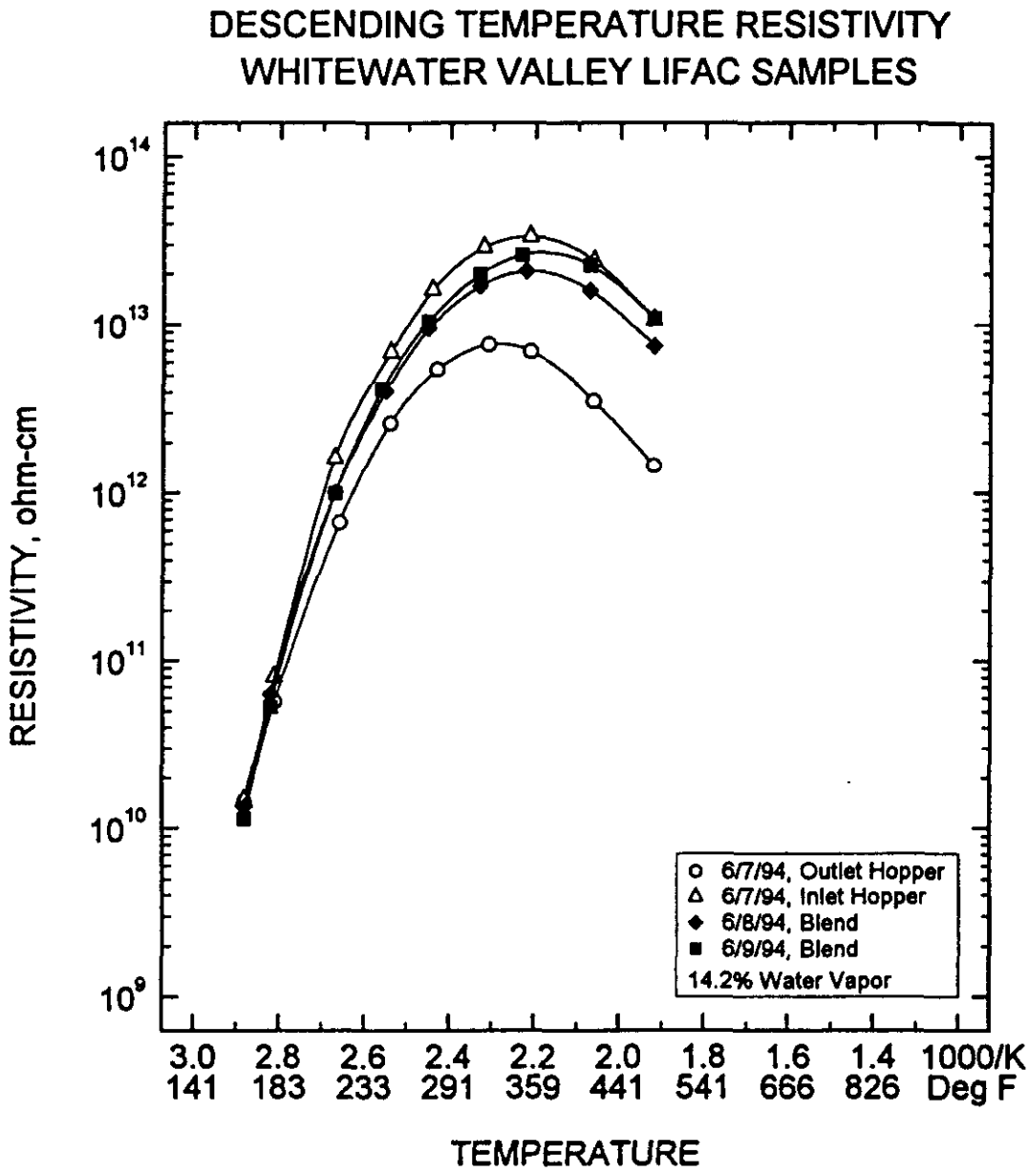

Figure 5
Descending-Temperature Resistivity Measurement on LIFAC Samples

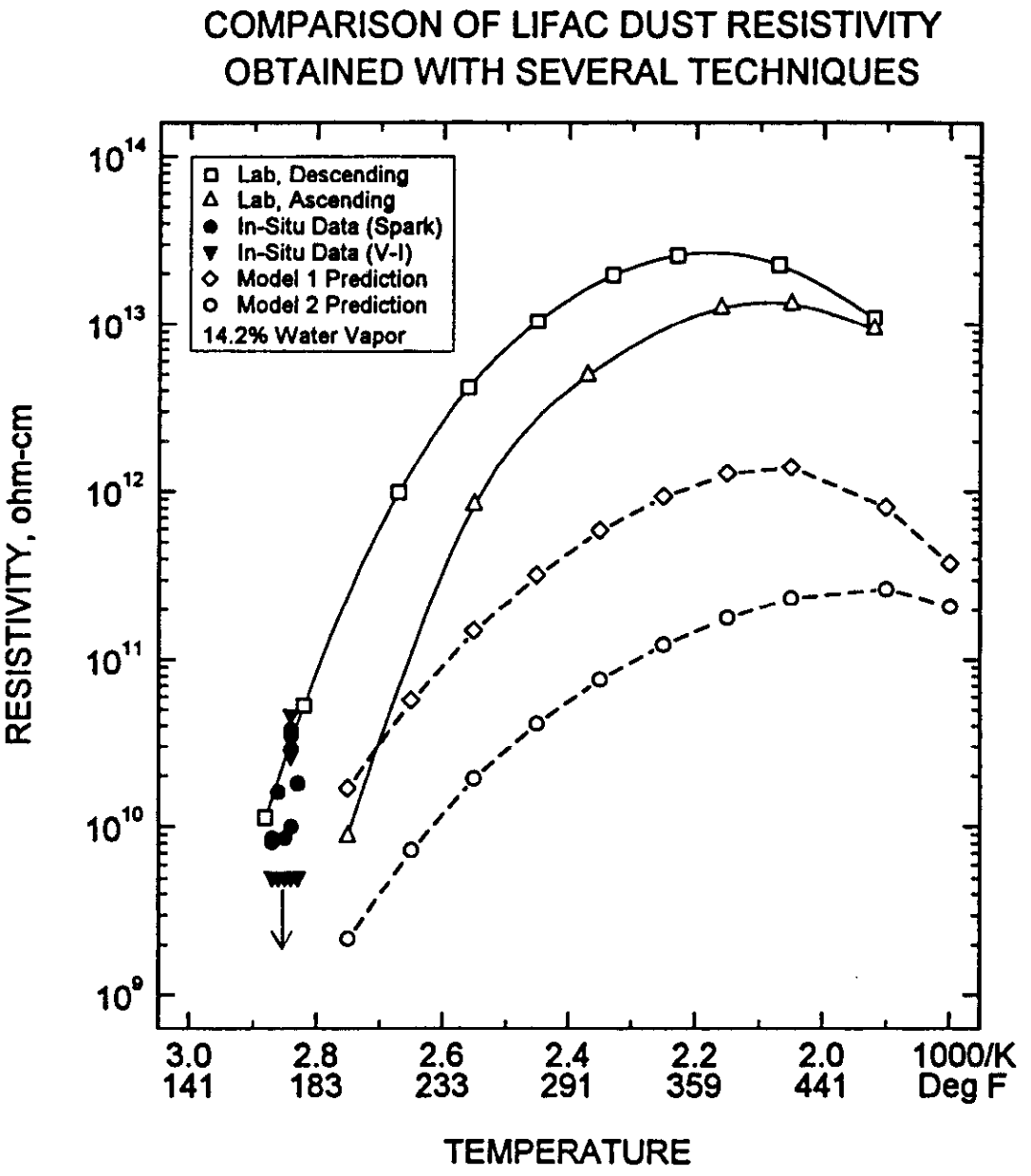

Figure 6
Comparison of LIFAC Dust Resistivity Obtained with Several Techniques

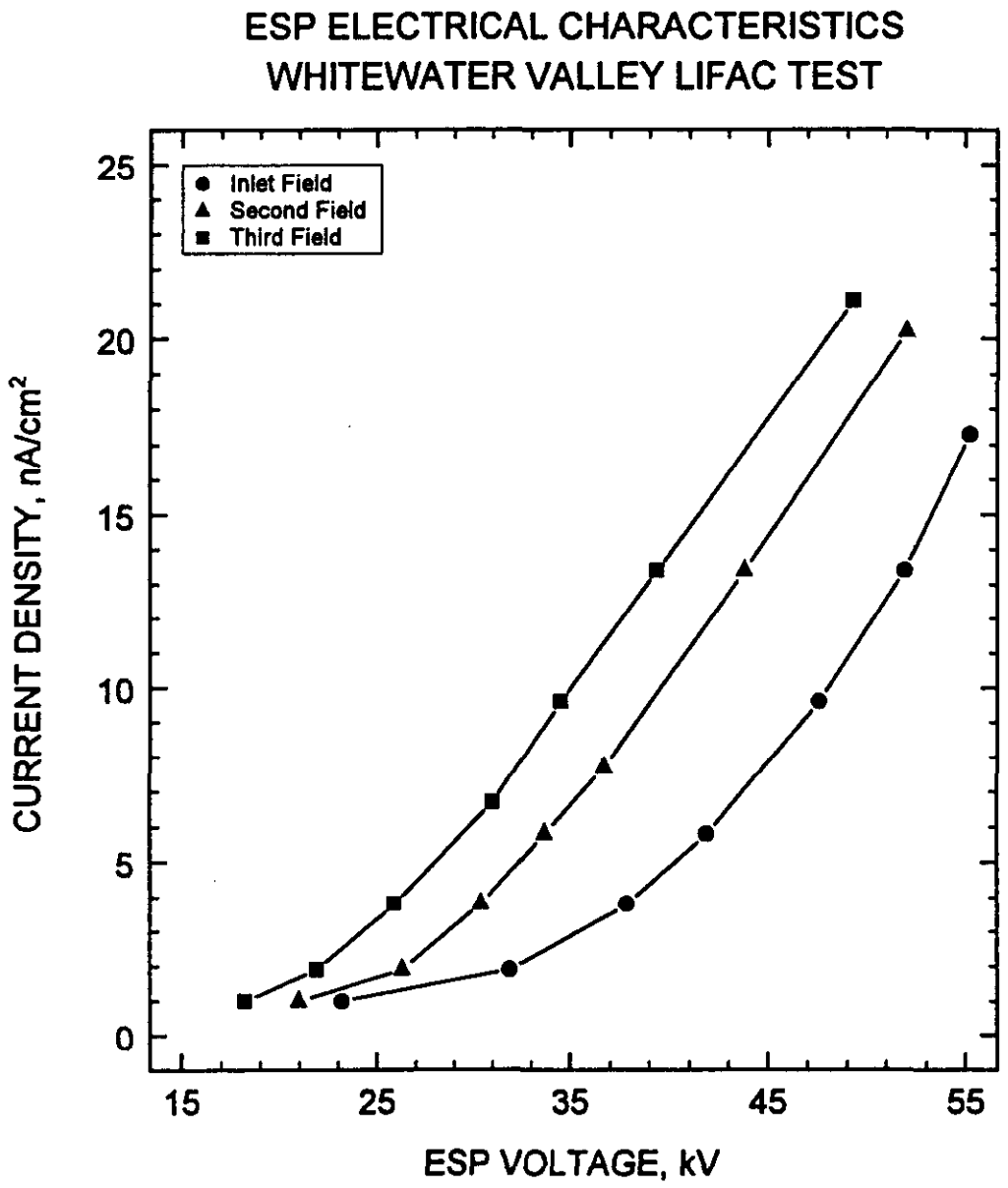

Figure 7
Whitewater Valley Unit 2 ESP Electrical Characteristics During LIFAC

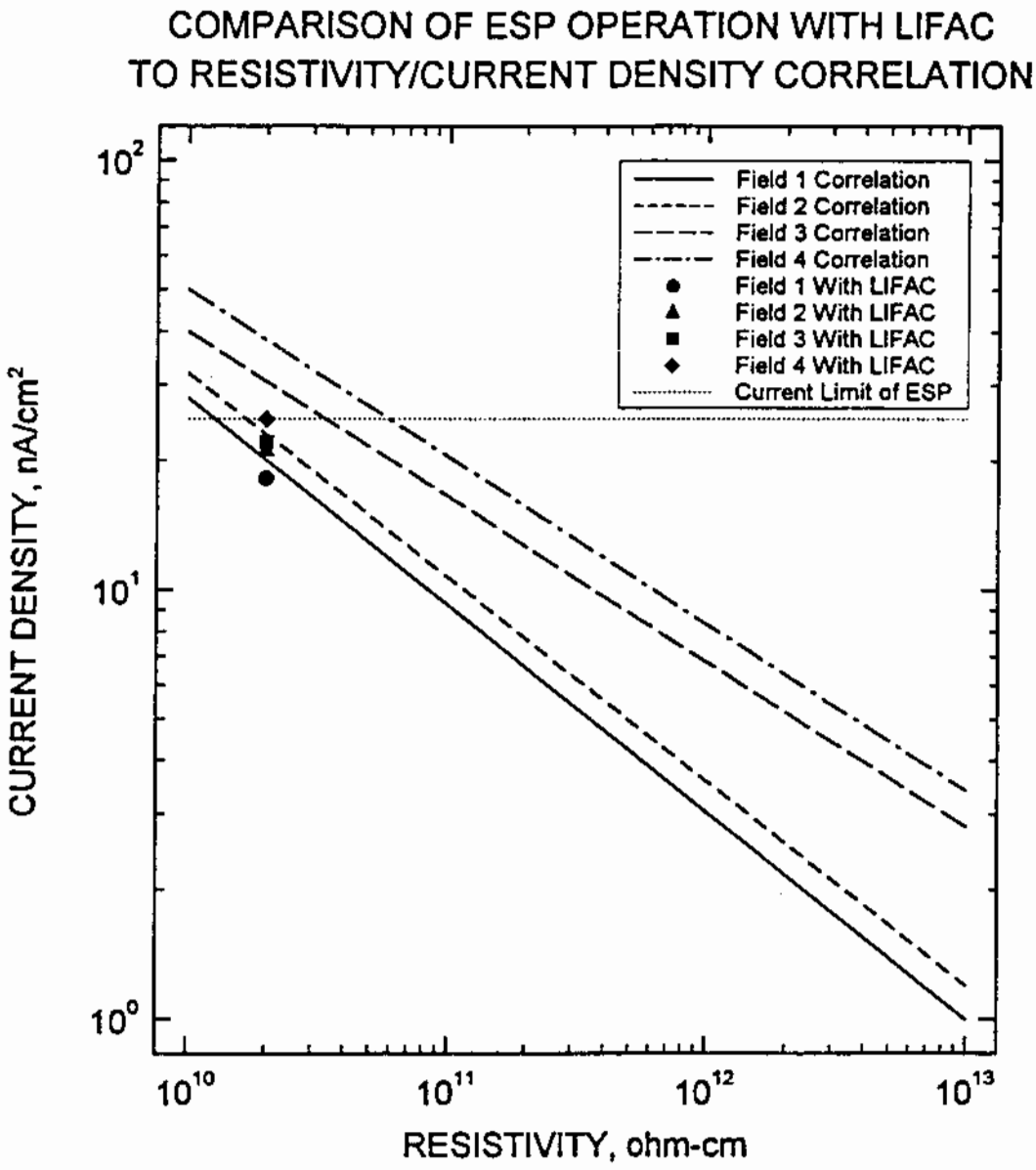

Figure 8
Comparison of ESP Operation with LIFAC to Typical Correlation Between Resistivity and Current Density

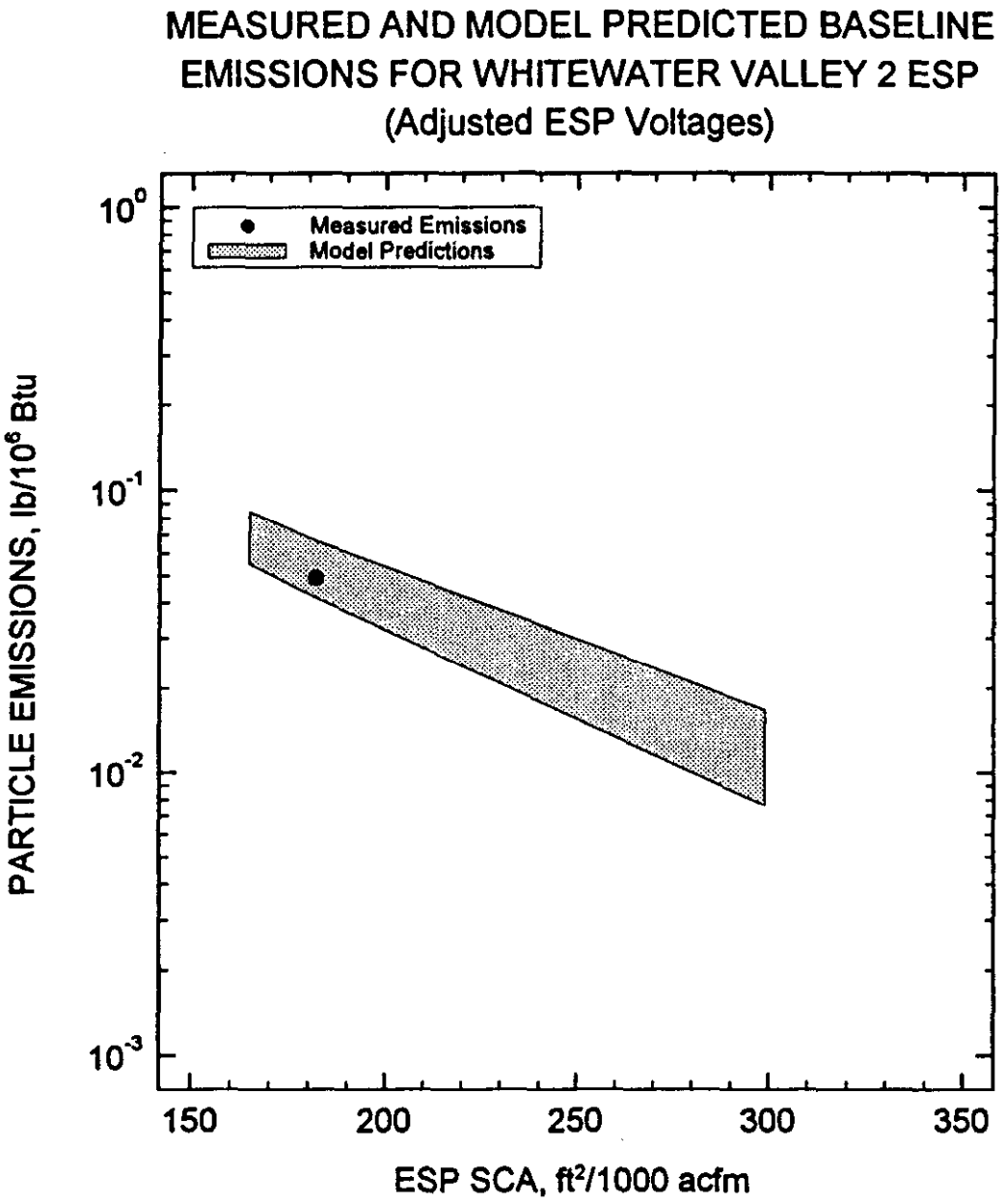

Figure 9
Comparison of Measured and Modeled Baseline ESP Emissions

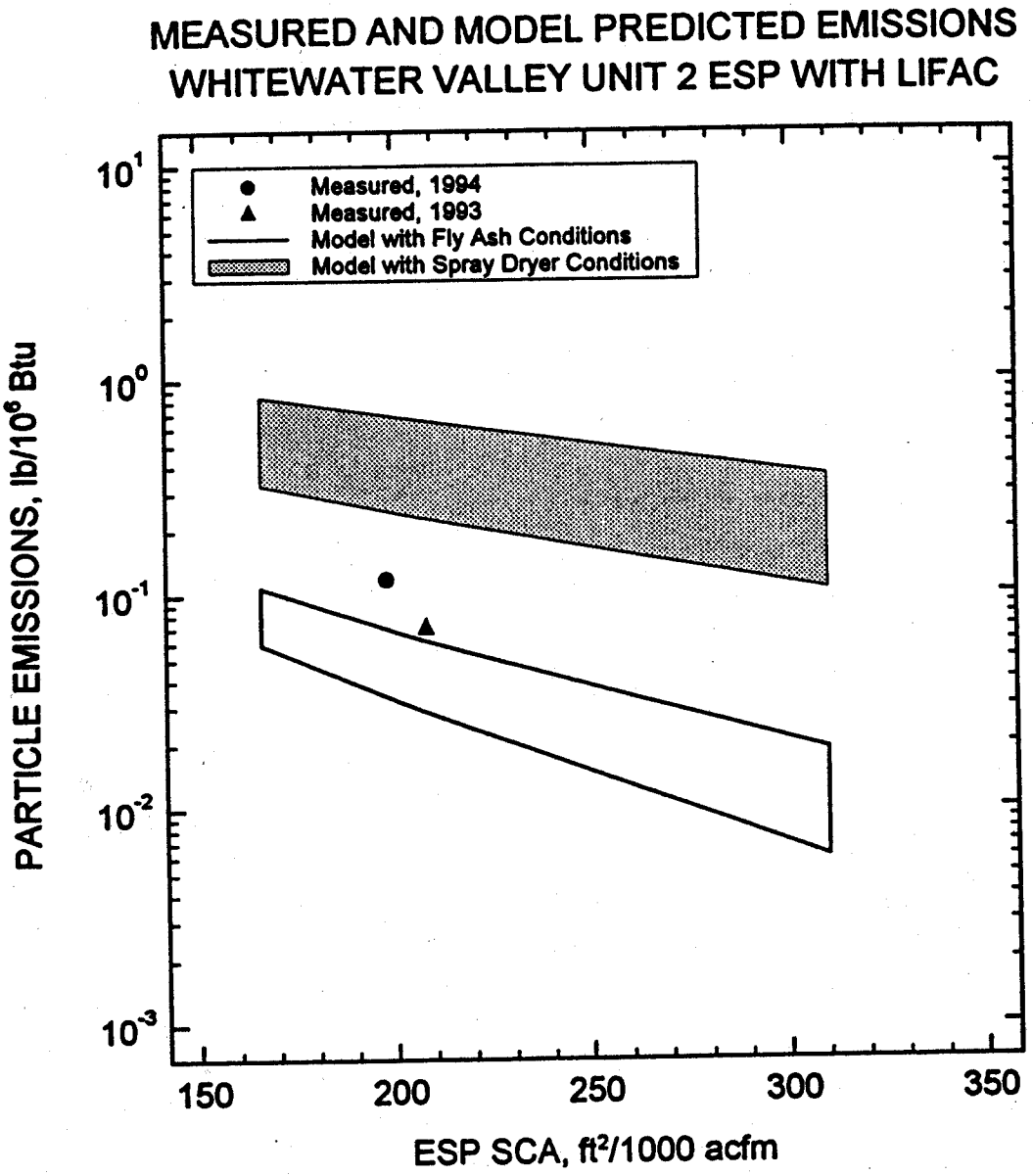

Figure 10
Comparison of Measured and Modeled ESP Emissions During LIFAC

APPENDIX II

EMISSION MEASUREMENT SPECIFICATIONS AND PROCEDURES

Method 5 - Determination of Particulate Emissions from Stationary Sources

1. PRINCIPLE AND APPLICABILITY

1.1 Principle. Particulate matter (PM) is withdrawn isokinetically from the source and collected on a glass fiber filter maintained at a temperature in the range of 120 ± 14°C (248 ± 25°F) or such other temperature as specified by an applicable subpart of the standards or approved by the Administrator, U.S. Environmental Protection Agency, for a particular application. The PM mass, which includes any material that condenses at or above the filtration temperature, is determined gravimetrically after removal of uncombined water.

1.2 Applicability. This method is applicable for the determination of PM emissions from stationary sources.

2. APPARATUS

2.1 *Sampling Train.* *A schematic of the sampling train used in this method is shown in Figure 5-1.* Complete construction details are given in APTD-0581 (Citation 2 in the Bibliography); commercial models of this train are also available. For changes from APTD-0581 and for allowable modifications of the train shown in Figure 5-1, see the following subsections.

The operating and maintenance procedures for the sampling train are described in APTD-0576 (Citation 3 in the Bibliography). Since correct usage is important in obtaining valid results, all users should read APTD-0576 and adopt the operating and maintenance procedures outlined in it, unless otherwise specified herein. The sampling train consists of the following components:

2.1.1 Probe Nozzle. Stainless steel (316) or glass with sharp, tapered leading edge. The angle of taper shall be ≤30°, and the taper shall be on the outside to preserve a constant internal diameter. The probe nozzle shall be of the button-hook or elbow design, unless otherwise specified by the Administrator. If made of stainless steel, the nozzle shall be constructed from seamless tubing; other materials of construction may be used, subject to the approval of the Administrator.

A range of nozzle sizes suitable for isokinetic sampling should be available, e.g., 0.32 to 1.27 cm (1/8 to 1/2 in.)--or larger if higher volume sampling trains are used--inside diameter (ID) nozzles in increments of 0.16 cm (1/16 in.). Each nozzle shall be calibrated according to the procedures outlined in Section 5.

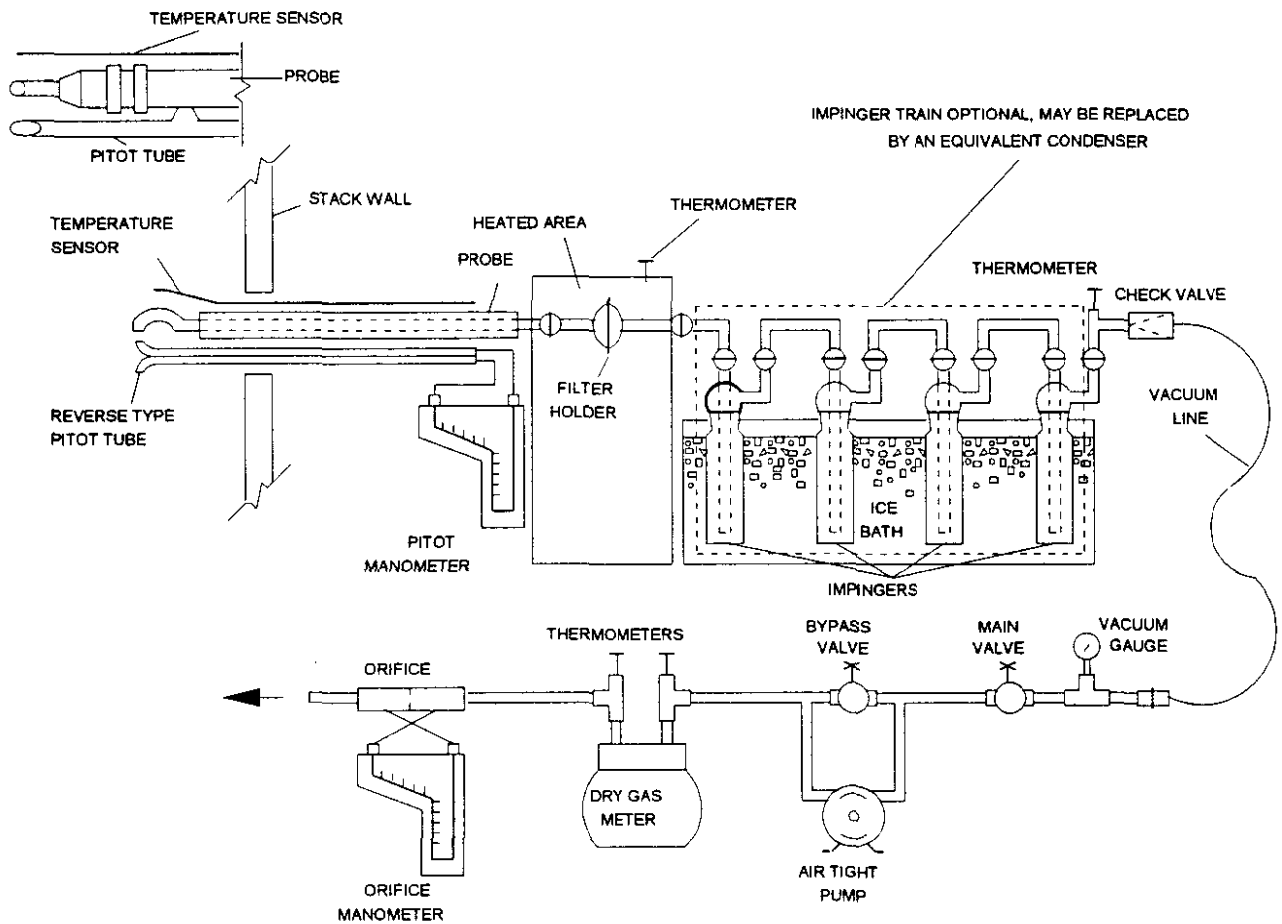

Figure 5-1. Particulate Sampling Train

2.1.2 Probe Liner. Borosilicate or quartz glass tubing with a heating system capable of maintaining a gas temperature at the exit end during sampling of 120 ± 14°C (248 ± 25°F), or such other temperature as specified by an applicable subpart of the standards or approved by the Administrator for a particular application. (The tester may opt to operate the equipment at a temperature lower than that specified.) Since the actual temperature at the outlet of the probe is not usually monitored during sampling, probes constructed according to APTD-0581 and utilizing the calibration curves of APTD-0576 (or calibrated according to the procedure outlined in APTD-0576) will be considered acceptable.

Either borosilicate or quartz glass probe liners may be used for stack temperatures up to about 480°C (900°F); quartz liners shall be used for temperatures between 480 and 900°C (900 and 1,650°F). Both types of liners may be used at higher temperatures than specified for short periods of time, subject to the approval of the Administrator. The softening temperature for borosilicate is 820°C (1,508°F), and for quartz it is 1,500°C (2,732°F).

Whenever practical, every effort should be made to use borosilicate or quartz glass probe liners. Alternatively, metal liners (e.g., 316 stainless steel, Incoloy 825 or other corrosion resistant metals) made of seamless tubing may be used, subject to the approval of the Administrator.

NOTE: Mention of trade names or specific products does not constitute endorsement of the Environmental Protection Agency.

2.1.3 Pitot Tube. Type S, as described in Section 2.1 of Method 2, or other device approved by the Administrator. The pitot tube shall be attached to the probe (as shown in Figure 5-1) to allow constant monitoring of the stack gas velocity. The impact (high pressure) opening plane of the pitot tube shall be even with or above the nozzle entry plane (see Method 2, Figure 2-6b) during sampling. The Type S pitot tube assembly shall have a known coefficient, determined as outlined in Section 4 of Method 2.

2.1.4 Differential Pressure Gauge. Inclined manometer or equivalent device (two), as described in Section 2.2 of Method 2. One manometer shall be used for velocity head (Δp) readings, and the other, for orifice differential pressure readings.

2.1.5 Filter Holder. Borosilicate glass, with a glass frit filter support and a silicone rubber gasket. Other materials of construction (e.g., stainless steel, Teflon, Viton) may be used, subject to the approval of the Administrator. The holder design shall provide a positive seal against leakage from the outside or around the filter. The holder shall be attached immediately at the outlet of the probe (or cyclone, if used).

2.1.6 Filter Heating System. Any heating system capable of maintaining a temperature around the filter holder during the sampling of 120 ± 14°C (248 ± 25°F), or such other temperature as specified by an applicable subpart of the standards or approved by the Administrator for a particular application. Alternatively, the tester may opt to operate the equipment at a temperature lower than that specified. A temperature gauge capable of measuring temperature to within 3°C (5.4°F) shall be installed so that the temperature around the filter holder can be regulated and monitored during sampling. Heating systems other than the one shown in APTD-0581 may be used.

2.1.7 Condenser. The following system shall be used to determine the stack gas moisture content: Four impingers connected in series with leak-free ground glass fittings or any similar leak-free noncontaminating fittings. The first, third, and fourth impingers shall be of the Greenburg-Smith design, modified by replacing the tip with a 1.3 cm (1/2 in.) ID glass tube extending to about 1.3 cm (1/2 in.) from the bottom of the flask. The second impinger shall be of the Greenburg-Smith design with the standard tip. Modifications (e.g., using a flexible connections between the impingers, using materials other than glass, or using flexible vacuum lines to connect the filter holder to the condenser) may be used, subject to the approval of the Administrator. The first and second impingers shall contain known quantities of water (Section 4.1.3), the third shall be empty, and the fourth shall contain a known weight of silica gel, or equivalent desiccant. A thermometer, capable of measuring temperature to within 1°C (2°F) shall be placed at the outlet of the fourth impinger for monitoring purposes.

Alternatively, any system that cools the sample gas stream and allows measurement of the water condensed and moisture leaving the condenser, each to within 1 ml or 1 g may be used, subject to the approval of the Administrator. Acceptable means are to measure the condensed water either gravimetrically or volumetrically and to measure the moisture leaving the condenser by: (1) monitoring the temperature and pressure at the exit of the condenser and using Dalton's law of partial pressures; or (2) passing the sample gas stream through a tared silica gel (or equivalent desiccant) trap with exit gases kept below 20°C (68°F) and determining the weight gain.

If means other than silica gel are used to determine the amount of moisture leaving the condenser, it is recommended that silica gel (or equivalent) still be used between the condenser system and pump to prevent moisture condensation in the pump and metering devices and to avoid the need to make corrections for moisture in the metered volume.

NOTE: If a determination of the PM collected in the impingers is desired in addition to moisture content, the impinger system described above shall be used, without modification. Individual States or control agencies requiring this information shall be contacted as to the sample recovery and analysis of the impinger contents.

2.1.8 Metering System. Vacuum gauge, leak-free pump, thermometers capable of measuring temperature to within 3°C (5.4°F), dry gas meter (DGM) capable of measuring volume to within 2 percent, and related equipment, as shown in Figure 5-1. Other metering systems capable of maintaining sampling rates within 10 percent of isokinetic and of determining sample volumes to within 2 percent may be used, subject to the approval of the Administrator. When the metering system is used in conjunction with a pitot tube, the system shall enable checks of isokinetic rates.

Sampling trains utilizing metering systems designed for higher flow rates than that described in APTD-0581 or APTD-0576 may be used provided that the specifications of this method are met.

2.1.9 Barometer. *Mercury, aneroid, or other barometer capable of measuring atmospheric pressure to within* 2.5 mm (0.1 in.) Hg.

NOTE: In many cases, the barometric reading may be obtained from a nearby National Weather Service station, in which case the station value (which is the absolute barometric pressure) shall be requested and an adjustment for elevation differences between the weather station and sampling point shall be applied at a rate of minus 2.5 mm (0.1 in.) Hg per 30 m (100 ft) elevation increase or vice versa for elevation decrease.

2.1.10 Gas Density Determination Equipment. Temperature sensor and pressure gauge, as described in Section 2.3 and 2.4 of Method 2, and gas analyzer, if necessary, as described in Method 3. The temperature sensor shall, preferably, be permanently attached to the pitot tube or sampling probe in a fixed configuration, such that the tip of the sensor extends beyond the leading edge of the probe sheath and does not touch any metal. Alternatively, the sensor may be attached just prior to use in the field. Note, however, that if the temperature sensor is attached in the field, the sensor must be placed in an interference-free arrangement with respect to the Type S pitot tube openings (see Method 2, Figure 2-7). As a second alternative, if a difference of not more than 1 percent in the average velocity measurement is to be introduced, the temperature gauge need not be attached to the probe or pitot tube. (This alternative is subject to the approval of the Administrator.)

2.2 Sample Recovery. The following items are needed:

2.2.1 Probe-Liner and Probe-Nozzle Brushes. Nylon bristle brushes with stainless steel wire handles. The probe brush shall have extensions (at least as long as the probe) of stainless steel, Nylon, Teflon, or similarly inert material The brushes shall be properly sized and shaped to brush out the probe liner and nozzle.

2.2.2 Wash Bottles. Two. Glass wash bottles are recommended; polyethylene wash bottles may be used at the option of the tester. It is recommended that acetone not be stored in polyethylene bottles for longer than a month.

2.2.3 Glass Sample Storage Containers. Chemically resistant, borosilicate glass bottles, for acetone washes, 500-ml or 1000-ml. Screw cap liners shall either be rubber-backed Teflon or shall be constructed so as to be leak-free and resistant to chemical attack by acetone. (Narrow mouth glass bottles have been found to be less prone to leakage.) *Alternatively, polyethylene bottles may be used.*

2.2.4 Petri Dishes. For filter samples, glass or polyethylene, unless otherwise specified by the Administrator.

2.2.5 Graduated Cylinder and/or Balance. To measure condensed water to within 1 ml or 1 g. Graduated cylinders shall have subdivisions no greater than 2 ml. Most laboratory balances are capable of weighing to the nearest 0.5 g or less. Any of these balances is suitable for use here and in Section 2.3.4.

2.2.6 Plastic Storage Containers. Air-tight containers to store silica gel.

2.2.7 Funnel and Rubber Policeman. To aid in transfer of silica gel to container; not necessary if silica

gel is weighed in the field.

2.2.8 Funnel. Glass or polyethylene, to aid in sample recovery.

2.3 Analysis. For analysis, the following equipment is needed:

2.3.1 Glass Weighing Dishes.

2.3.2 Desiccator.

2.3.3 Analytical Balance. To measure to within 0.1 mg.

2.3.4 Balance. To measure to within 0.5 g.

2.3.5 Beakers. 250-ml.

2.3.6 Hygrometer. To measure the relative humidity of the laboratory environment.

2.3.7 Temperature Gauge. To measure the temperature of the laboratory environment.

3. REAGENTS

3.1 Sampling. The reagents used in sampling are as follows:

3.1.1 Filters. Glass fiber filters, without organic binder, exhibiting at least 99.95 percent efficiency (<0.05 percent penetration) on 0.3-micron dioctyl phthalate smoke particles. The filter efficiency test shall be conducted in accordance with ASTM Method D 2986-71 (Reapproved 1978)(incorporated by reference--see §60.17). Test data from the supplier's quality control program are sufficient for this purpose. In sources containing SO_2 or SO_3, the filter material must be of a type that is unreactive to SO_2 or SO_3. Citation 10 in the Bibliography may be used to select the appropriate filter.

3.1.2 Silica Gel. Indicating type, 6- to 16-mesh. If previously used, dry at 175°C (350°F) for 2 hours. New silica gel may be used as received. Alternatively, other types of desiccants (equivalent or better) may be used, subject to the approval of the Administrator.

3.1.3 Water. When analysis of the material caught in the impingers is required, deionized distilled water shall be used. Run blanks prior to field use to eliminate a high blank on test samples.

3.1.4 Crushed Ice.

3.1.5 Stopcock Grease. Acetone-insoluble, heat-stable silicone grease. This is not necessary if screw-on connectors with Teflon sleeves, or similar, are used. Alternatively, other types of stopcock grease may be used, subject to the approval of the Administrator.

3.2 Sample Recovery. Acetone--reagent grade, ≤0.001 percent residue, in glass bottles--is required. Acetone from metal containers generally has a high residue blank and should not be used. Sometimes, suppliers transfer acetone to glass bottles from metal containers; thus, acetone blanks shall be run prior to field use and only acetone with low blank values (≤0.001 percent) shall be used. In no case shall a blank value of greater than 0.001 percent of the weight of acetone used be subtracted from the sample weight.

3.3 Analysis. Two reagents are required for the analysis:

3.3.1 Acetone. Same as in Section 3.2.

3.3.2 Desiccant. Anhydrous calcium sulfate, indicating type. Alternatively, other types of desiccants may be used, subject to the approval of the Administrator.

4. PROCEDURE

4.1 Sampling. The complexity of this method is such that, in order to obtain reliable results, testers should be trained and experienced with the test procedures.

4.1.1 Pretest Preparation. It is suggested that sampling equipment be maintained according to the procedures described in APTD-0576.

Weigh several 200- to 300-g portions of silica gel in air-tight containers to the nearest 0.5 g. Record the total weight of the silica gel plus container, on each container. As an alternative, the silica gel need not be preweighed, but may be weighed directly in its impinger or sampling holder just prior to train assembly.

Check filters visually against light for irregularities and flaws or pinhole leaks. Label filters of the proper diameter on the back side near the edge using numbering machine ink. As an alternative, label the shipping containers (glass or plastic petri dishes), and keep the filters in these containers at all times except during sampling and weighing.

Desiccate the filters at 20 ± 5.6°C (68 ± 10°F) and ambient pressure for at least 24 hours, and weigh at intervals of at least 6 hours to a constant weight, i.e., ≤0.5-mg change from previous weighing; record results to the nearest 0.1 mg. During each weighing the filter must not be exposed to the laboratory atmosphere for a period greater than 2 minutes and a relative humidity above 50 percent. Alternatively (unless otherwise specified by the Administrator), the filters may be oven dried at 105°C (220°F) for 2 to 3 hours, desiccated for 2 hours, and weighed. Procedures other than those described, which account for relative humidity effects, may be used, subject to the approval of the Administrator.

4.1.2 Preliminary Determinations. Select the sampling site and the minimum number of sampling points according to Method 1 or as specified by the Administrator. Determine the stack pressure, temperature, and the range of velocity heads using Method 2; it is recommended that a leak-check of the pitot lines (see Method 2, Section 3.1) be performed. Determine the moisture content using Approximation Method 4 or its alternatives for the purpose of making isokinetic sampling rate settings. Determine the stack gas dry molecular weight, as described in Method 2, Section 3.6; if integrated Method 3 sampling is used for molecular weight determination, the integrated bag sample shall be taken simultaneously with, and for the *same total length of time as, the particulate sample run.*

Select a nozzle size based on the range of velocity heads, such that it is not necessary to change the nozzle size in order to maintain isokinetic sampling rates. During the run, do not change the nozzle size. Ensure that the proper differential pressure gauge is chosen for the range of velocity heads encountered (see Section 2.2 of Method 2).

Select a suitable probe liner and probe length such that all traverse points can be sampled. For large stacks, consider sampling from opposite sides of the stack to reduce the length of probes.

Select a total sampling time greater than or equal to the minimum total sampling time specified in the test procedures for the specific industry such that (1) the sampling time per point is not less than 2 minutes (or some greater time interval as specified by the Administrator), and (2) the sample volume taken (corrected to standard conditions) will exceed the required minimum total gas sample volume. The latter is based on an approximate average sampling rate.

The sampling time at each point shall be the same. It is recommended that the number of minutes sampled at each point be an integer or an integer plus one-half minute, in order to avoid timekeeping errors.

In some circumstances, e.g., batch cycles, it may be necessary to sample for shorter times at the traverse points and to obtain smaller gas sample volumes. In these cases, the Administrator's approval must first be obtained.

4.1.3 Preparation of Collection Train. During preparation and assembly of the sampling train, keep all openings where contamination can occur covered until just prior to assembly or until sampling is about to begin.

Place 100 ml of water in each of the first two impingers, leave the third impinger empty, and transfer approximately 200 to 300 g of preweighed silica gel from its container to the fourth impinger. More silica gel may be used, but care should be taken to ensure that it is not entrained and carried out from the impinger during sampling. Place the container in a clean place for later use in the sample recovery. Alternatively, the weight of the silica gel plus impinger may be determined to the nearest 0.5 g and recorded.

Using a tweezer or clean disposable surgical gloves, place a labeled (identified) and weighed filter in the filter holder. Be sure that the filter is properly centered and the gasket properly placed so as to prevent the sample gas stream from circumventing the filter. Check the filter for tears after assembly is

completed.

When glass liners are used, install the selected nozzle using a Viton A O-ring when stack temperatures are *less than 260°C (500°F) and an asbestos string gasket when temperatures are higher. See APTD-0576 for* details. Other connecting systems using either 316 stainless steel or Teflon ferrules may be used. When metal liners are used, install the nozzle as above or by a leak-free direct mechanical connection. Mark the probe with heat resistant tape or by some other method to denote the proper distance into the stack or duct for each sampling point.

Set up the train as in Figure 5-1, using (if necessary) a very light coat of silicone grease on all ground glass joints, greasing only the outer portion (see APTD-0576) to avoid possibility of contamination by the silicone grease. Subject to the approval of the Administrator, a glass cyclone may be used between the probe and filter holder when the total particulate catch is expected to exceed 100 mg or when water droplets are present in the stack gas.

Place crushed ice around the impingers.

4.1.4 Leak-Check Procedures.

4.1.4.1 Pretest Leak-Check. A pretest leak-check is recommended, but not required. If the tester opts to conduct the pretest leak-check, the following procedure shall be used.

After the sampling train has been assembled, turn on and set the filter and probe heating systems at the desired operating temperatures. Allow time for the temperatures to stabilize. If a Viton A O-ring or other leak-free connection is used in assembling the probe nozzle to the probe liner, leak-check the train at the sampling site by plugging the nozzle and pulling a 380 mm (15 in.) Hg vacuum.

NOTE: A lower vacuum may be used, provided that it is not exceeded during the test.

If an asbestos string is used, do not connect the probe to the train during the leak-check. Instead, leak-check the train by first plugging the inlet to the filter holder (cyclone, if applicable) and pulling a 380 mm (15 in.) Hg vacuum
(see NOTE immediately above). Then connect the probe to the train, and leak-check at about 25 mm (1 in.) Hg vacuum; alternatively, the probe may be leak-checked with the rest of the sampling train, in one step, at 380 mm (15 in.) Hg vacuum. Leakage rates in excess of 4 percent of the average sampling rate or 0.00057 m^3/min (0.02 cfm), whichever is less, are unacceptable.

The following leak-check instructions for the sampling train described in APTD-0576 and APTD-0581 may be helpful. Start the pump with bypass valve fully open and coarse adjust valve completely closed. Partially open the coarse adjust valve, and slowly close the bypass valve until the desired vacuum is reached. Do not reverse direction of bypass valve; this will cause water to back up into the filter holder. If the desired vacuum is exceeded, either leak-check at this higher vacuum or end the leak-check as shown below, and start over.

When the leak-check is completed, first slowly remove the plug from the inlet to the probe, filter holder, or cyclone (if applicable), and immediately turn off the vacuum pump. This prevents the water in the impingers from being forced backward into the filter holder and silica gel from being entrained backward into the third impinger.

4.1.4.2 Leak-Checks During Sample Run. If, during the sampling run, a component (e.g., filter assembly or impinger) change becomes necessary, a leakcheck shall be conducted immediately before the change is made. The leak-check shall be done according to the procedure outlined in Section 4.1.4.1 above, except that it shall be done at a vacuum equal to or greater than the maximum value recorded up to that point in the test. If the leakage rate is found to be no greater than 0.00057 m^3/min (0.02 cfm) or 4 percent of the average sampling rate (whichever is less), the results are acceptable, and no correction will need to be applied to the total volume of dry gas metered; if, however, a higher leakage rate is obtained, the tester shall either record the leakage rate and plan to correct the sample volume as shown in Section 6.3 of this method, or shall void the sample run.

Immediately after component changes, leak-checks are optional; if such leak-checks are done, the procedure outlined in Section 4.1.4.1 above shall be used.

4.1.4.3 Post-Test Leak-Check. A leak-check is mandatory at the conclusion of each sampling run. The leak-check shall be done in accordance with the procedures outlined in Section 4.1.4.1, except that it shall be conducted at a vacuum equal to or greater than the maximum value reached during the sampling run. If the

leakage rate is found to be no greater than 0.00057 m^3/min (0.02 cfm) or 4 percent of the average sampling rate (whichever is less), the results are acceptable, and no correction need be applied to the total volume of dry gas metered. If, however, a higher leakage rate is obtained, the tester shall either record the *leakage rate and correct the sample volume as shown in Section 6.3 of this method, or shall void the* sampling run.

4.1.5 Particulate Train Operation. During the sampling run, maintain an isokinetic sampling rate (within 10 percent of true isokinetic unless otherwise specified by the Administrator) and a temperature around the filter of 120 ± 14°C (248 ± 25°F), or such other temperature as specified by an applicable subpart of the standards or approved by the Administrator.

For each run, record the data required on a data sheet such as the one shown in Figure 5-2. Be sure to record the initial dry gas meter (DGM) reading. Record the DGM readings at the beginning and end of each sampling time increment, when changes in flow rates are made, before and after each leak-check, and when sampling is halted. Take other readings required by Figure 5-2 at least once at each sample point during each time increment and additional readings when significant changes (20 percent variation in velocity head readings) necessitate additional adjustments in flow rate. Level and zero the manometer. Because the manometer level and zero may drift due to vibrations and temperature changes, make periodic checks during the traverse.

Clean the portholes prior to the test run to minimize the chance of sampling deposited material. To begin sampling, remove the nozzle cap, verify that the filter and probe heating systems are up to temperature, and that the pitot tube and probe are properly positioned. Position the nozzle at the first traverse point with the tip pointing directly into the gas stream. Immediately start the pump, and adjust the flow to isokinetic conditions. Nomographs are available, which aid in the rapid adjustment of the isokinetic sampling rate without excessive computations. These nomographs are designed for use when the Type S pitot tube coefficient is 0.85 ± 0.02, and the stack gas equivalent density (dry molecular weight) is equal to 29 ± 4. APTD-0576 details the procedure for using the nomographs. If C_p and M_d are outside the above stated *ranges, do not use the nomographs unless appropriate steps (see Citation 7 in Bibliography) are taken to* compensate for the deviations.

When the stack is under significant negative pressure (height of impinger stem), take care to close the coarse adjust valve before inserting the probe into the stack to prevent water from backing into the filter holder. If necessary, the pump may be turned on with the coarse adjust valve closed.

When the probe is in position, block off the openings around the probe and porthole to prevent unrepresentative dilution of the gas stream.

Traverse the stack cross-section, as required by Method 1 or as specified by the Administrator, being careful not to bump the probe nozzle into the stack walls when sampling near the walls or when removing or inserting the probe through the portholes; this minimizes the chance of extracting deposited material.

During the test run, make periodic adjustments to keep the temperature around the filter holder at the proper level; add more ice and, if necessary, salt to maintain a temperature of less than 20°C (68°F) at the condenser/silica gel outlet. Also, periodically check the level and zero of the manometer.

If the pressure drop across the filter becomes too high, making isokinetic sampling difficult to maintain, the filter may be replaced in the midst of the sample run. It is recommended that another complete filter assembly be used rather than attempting to change the filter itself. Before a new filter assembly is installed, conduct a leak-check (see Section 4.1.4.2). The total PM weight shall include the summation of the filter assembly catches.

A single train shall be used for the entire sample run, except in cases where simultaneous sampling is required in two or more separate ducts or at two or more different locations within the same duct, or, in cases where equipment failure necessitates a change of trains. In all other situations, the use of two or more trains will be subject to the approval of the Administrator.

Note that when two or more trains are used, separate analyses of the front—half and (if applicable) impinger catches from each train shall be performed, unless identical nozzle sizes were used on all trains, in which case, the front—half catches from the individual trains may be combined (as may the impinger catches) and one analysis of front-half catch and one analysis of impinger catch may be performed. Consult *with the Administrator for details concerning the calculation of results when two or more trains are used.*

At the end of the sample run, turn off the coarse adjust valve, remove the probe and nozzle from the stack, turn off the pump, record the final DGM meter reading, and conduct a post-test leak-check, as outlined in

Section 4.1.4.3. Also, leak-check the pitot lines as described in Method 2, Section 3.1; the lines must pass this leak-check, in order to validate the velocity head data.

4.1.6 *Calculation of Percent Isokinetic. Calculate percent isokinetic (see Calculations, Section 6) to* determine whether the run was valid or another test run should be made. If there was difficulty in maintaining isokinetic rates because of source conditions, consult with the Administrator for possible variance on the isokinetic rates.

4.2 Sample Recovery. Proper cleanup procedure begins as soon as the probe is removed from the stack at the end of the sampling period. Allow the probe to cool.

When the probe can be safely handled, wipe off all external PM near the tip of the probe nozzle, and place a cap over it to prevent losing or gaining PM. Do not cap off the probe tip tightly while the sampling train is cooling down as this would create a vacuum in the filter holder, thus drawing water from the impingers into the filter holder.

Before moving the sample train to the cleanup site, remove the probe from the sample train, wipe off the silicone grease, and cap the open outlet of the probe. Be careful not to lose any condensate that might be present. Wipe off the silicone grease from the filter inlet where the probe was fastened, and cap it. Remove the umbilical cord from the last impinger, and cap the impinger. If a flexible line is used between the first impinger or condenser and the filter holder, disconnect the line at the filter holder, and let any condensed water or liquid drain into the impingers or condenser. After wiping off the silicone grease, cap off the filter holder outlet and impinger inlet. Either ground-glass stoppers, plastic caps, or serum caps may be used to close these openings.

Transfer the probe and filter-impinger assembly to the cleanup area. This area should be clean and protected from the wind so that the chances of contaminating or losing the sample will be minimized.

Save a portion of the acetone used for cleanup as a blank. Take 200 ml of this acetone directly from the wash bottle being used, and place it in a glass sample container labeled "acetone blank."

Inspect the train prior to and during disassembly, and note any abnormal conditions. Treat the samples as follows:

Container No. 1. Carefully remove the filter from the filter holder, and place it in its identified petri dish container. Use a pair of tweezers and/or clean disposable surgical gloves to handle the filter. If it is necessary to fold the filter, do so such that the PM cake is inside the fold. Carefully transfer to the petri dish any PM and/or filter fibers that adhere to the filter holder gasket, by using a dry Nylon bristle brush and/or a sharp-edged blade. Seal the container.

Container No. 2. Taking care to see that dust on the outside of the probe or other exterior surfaces does not get into the sample, quantitatively recover PM or any condensate from the probe nozzle, probe fitting, probe liner, and front half of the filter holder by washing these components with acetone and placing the wash in a glass container. Deionized distilled water may be used instead of acetone when approved by the Administrator and shall be used when specified by the Administrator; in these cases, save a water blank, and follow the Administrator's directions on analysis. Perform the acetone rinse as follows:

Carefully remove the probe nozzle, and clean the inside surface by rinsing with acetone from a wash bottle and brushing with a Nylon bristle brush. Brush until the acetone rinse shows no visible particles, after which make a final rinse of the inside surface with acetone.

Brush and rinse the inside parts of the Swagelok fitting with acetone in a similar way until no visible particles remain.

Rinse the probe liner with acetone by tilting and rotating the probe while squirting acetone into its upper end so that all inside surfaces will be wetted with acetone. Let the acetone drain from the lower end into the sample container. A funnel (glass or polyethylene) may be used to aid in transferring liquid washes to the container. Follow the acetone rinse with a probe brush. Hold the probe in an inclined position, squirt acetone into the upper end as the probe brush is being pushed with a twisting action through the probe; hold a sample container underneath the lower end of the probe, and catch any acetone and particulate matter that is brushed from the probe. Run the brush through the probe three times or more until no visible PM is *carried out with the acetone or until none remains in the probe liner on visual inspection. With stainless* steel or other metal probes, run the brush through in the above prescribed manner at least six times since metal probes have small crevices in which particulate matter can be entrapped. Rinse the brush with acetone, and

quantitatively collect these washings in the sample container. After the brushing, make a final acetone rinse of the probe as described above.

It is recommended that two people clean the probe to minimize sample losses. Between sampling runs, keep brushes clean and protected from contamination.

After ensuring that all joints have been wiped clean of silicone grease, clean the inside of the front half of the filter holder by rubbing the surfaces with a Nylon bristle brush and rinsing with acetone. Rinse each surface three times or more if needed to remove visible particulate. Make a final rinse of the brush and filter holder. Carefully rinse out the glass cyclone, also (if applicable). After all acetone washings and particulate matter have been collected in the sample container, tighten the lid on the sample container so that acetone will not leak out when it is shipped to the laboratory. Mark the height of the fluid level to determine whether leakage occurred during transport. Label the container to identify clearly its contents.

Container No. 3. Note the color of the indicating silica gel to determine whether it has been completely spent, and make a notation of its condition. Transfer the silica gel from the fourth impinger to its original container, and seal. A funnel may make it easier to pour the silica gel without spilling. A rubber policeman may be used as an aid in removing the silica gel from the impinger. It is not necessary to remove the small amount of dust particles that may adhere to the impinger wall and are difficult to remove. Since the gain in weight is to be used for moisture calculations, do not use any water or other liquids to transfer the silica gel. If a balance is available in the field, follow the procedure for Container No. 3 in Section 4.3.

Impinger Water. Treat the impingers as follows: Make a notation of any color or film in the liquid catch. Measure the liquid that is in the first three impingers to within 1 ml by using a graduated cylinder or by weighing it to within 0.5 g by using a balance (if one is available). Record the volume or weight of liquid present. This information is required to calculate the moisture content of the effluent gas.

Discard the liquid after measuring and recording the volume or weight, unless analysis of the impinger catch is required (see NOTE, Section 2.1.7).

If a different type of condenser is used, measure the amount of moisture condensed either volumetrically or gravimetrically.

Whenever possible, containers should be shipped in such a way that they remain upright at all times.

4.3 Analysis. Record the data required on a sheet such as the one shown in Figure 5-3. Handle each sample container as follows:

Container No. 1. Leave the contents in the shipping container or transfer the filter and any loose PM from the sample container to a tared glass weighing dish. Desiccate for 24 hours in a desiccator containing anhydrous calcium sulfate. Weigh to a constant weight, and report the results to the nearest 0.1 mg. For purposes of this Section, 4.3, the term "constant weight" means a difference of no more than 0.5 mg or 1 percent of total weight less tare weight, whichever is greater, between two consecutive weighings, with no less than 6 hours of desiccation time between weighings.

Alternatively, the sample may be oven dried at 105°C (220°F) for 2 to 3 hours, cooled in the desiccator, and weighed to a constant weight, unless otherwise specified by the Administrator. The tester may also opt to oven dry the sample at 105°C (220°F) for 2 to 3 hours, weigh the sample, and use this weight as a final weight.

Container No. 2. Note the level of liquid in the container, and confirm on the analysis sheet whether leakage occurred during transport. If a noticeable amount of leakage has occurred, either void the sample or use methods, subject to the approval of the Administrator, to correct the final results. Measure the liquid in this container either volumetrically to ±1 ml or gravimetrically to ±0.5 g. Transfer the contents to a tared 250-ml beaker, and evaporate to dryness at ambient temperature and pressure. Desiccate for 24 hours, and weigh to a constant weight. Report the results to the nearest 0.1 mg.

Container No. 3. Weigh the spent silica gel (or silica gel plus impinger) to the nearest 0.5 g using a balance. This step may be conducted in the field.

"Acetone Blank" Container. Measure the acetone in this container either volumetrically or gravimetrically. Transfer the acetone to a tared 250-ml beaker, and evaporate to dryness at ambient temperature and pressure. Desiccate for 24 hours, and weigh to a constant weight. Report the results to the nearest 0.1 mg.

NOTE: At the option of the tester, the contents of Container No. 2 as well as the acetone blank container may be evaporated at temperatures higher than ambient. If evaporation is done at an elevated temperature, the temperature must be below the boiling point of the solvent; also, to prevent "bumping," the evaporation process must be closely supervised, and the contents of the beaker must be swirled occasionally to maintain an even temperature. Use extreme care, as acetone is highly flammable and has a low flash point.

4.4 Quality Control Procedures. The following quality control procedures are suggested to check the volume metering system calibration values at the field test site prior to sample collection. These procedures are optional for the tester.

4.4.1 Meter Orifice Check. Using the calibration data obtained during the calibration procedure described in Section 5.3, determine the $\Delta H_@$ for the metering system orifice. The $\Delta H_@$ is the orifice pressure differential in units of in. H_2O that correlates to 0.75 cfm of air at 528°R and 29.92 in. Hg. The $\Delta H_@$ is calculated as follows:

$$\Delta H_@ = 0.0319 \Delta H \frac{T_m}{P_{bar}} \frac{\Theta^2}{Y^2 V_m^2} \qquad \text{Eq. 5-9}$$

where:

ΔH = Average pressure differential across the orifice meter, in. H_2O.

T_m = Absolute average DGM temperature, °R.

P_{bar} = Barometric pressure, in. Hg.

Θ = Total sampling time, min.

Y = DGM calibration factor, dimensionless.

V_m = Volume of gas sample as measured by DGM, dcf.

0.0319 = (0.0567 in. Hg/°R)(0.75 cfm)^2

Before beginning the field test (a set of three runs usually constitutes a field test), operate the metering system (i.e., pump, volume meter, and orifice) at the $\Delta H_@$ pressure differential for 10 minutes. Record the volume collected, the DGM temperature, and the barometric pressure. Calculate a DGM calibration check value, Y_c, as follows:

$$Y_c = \frac{10}{V_m} \left[0.0319 \frac{T_m}{P_{bar}} \right]^{1/2} \qquad \text{Eq. 5-10}$$

where:

Y_c = DGM calibration check value, dimensionless.

10 = Run time, min.

Compare the Y_c value with the dry gas meter calibration factor Y to determine that: $0.97Y < Y_c < 1.03Y$. If the Y_c value is not within this range, the volume metering system should be investigated before beginning the test.

4.4.2 Calibrated Critical Orifice. A calibrated critical orifice, calibrated against a wet test meter or spirometer and designed to be inserted at the inlet of the sampling meter box, may be used as a quality control check by following the procedure of Section 7.2.

5. CALIBRATION

Maintain a laboratory log of all calibrations.

5.1 Probe Nozzle. Probe nozzles shall be calibrated before their initial use in the field. Using a micrometer, measure the inside diameter of the nozzle to the nearest 0.025 mm (0.001 in.). Make three separate measurements using different diameters each time, and obtain the average of the measurements. The difference between the high and low numbers shall not exceed 0.1 mm (0.004 in.). When nozzles become nicked, dented, or corroded, they shall be reshaped, sharpened, and recalibrated before use. Each nozzle shall be permanently and uniquely identified.

5.2 Pitot Tube. The Type S pitot tube assembly shall be calibrated according to the procedure outlined in Section 4 of Method 2.

5.3 Metering System.

5.3.1 Calibration Prior to Use. Before its initial use in the field, the metering system shall be calibrated as follows: Connect the metering system inlet to the outlet of a wet test meter that is accurate to within 1 percent. Refer to Figure 5.5. The wet test meter should have a capacity of 30 liters/rev (1 ft^3/rev). A spirometer of 400 liters (14 ft^3) or more capacity, or equivalent, may be used for this calibration, although a wet test meter is usually more practical. The wet test meter should be periodically calibrated with a spirometer or a liquid displacement meter to ensure the accuracy of the wet test meter. Spirometers or wet test meters of other sizes may be used, provided that the specified accuracies of the procedure are maintained. Run the metering system pump for about 15 minutes with the orifice manometer indicating a median reading as expected in field use to allow the pump to warm up and to permit the interior surface of the wet test meter to be thoroughly wetted. Then, at each of a minimum of three orifice manometer settings, pass an exact quantity of gas through the wet test meter, and note the gas volume indicated by the DGM. Also note the barometric pressure, and the temperatures of the wet test meter, the inlet of the DGM, and the outlet of the DGM. Select the highest and lowest orifice settings to bracket the expected field operating range of the orifice. Use a minimum volume of 0.15 m^3 (5 cf) at all orifice settings. Record all the data on a form similar to Figure 5.6, and calculate Y, the DGM calibration factor, and $\Delta H_@$, the orifice calibration factor, at each orifice setting as shown on Figure 5.6. Allowable tolerances for individual Y and $\Delta H_@$ values are given in Figure 5.6. Use the average of the Y values in the calculations in Section 6.

Before calibrating the metering system, it is suggested that a leak-check be conducted. For metering systems having diaphragm pumps, the normal leak-check procedure will not detect leakages within the pump. For these cases the following leak-check procedure is suggested: make a 10-minute calibration run at 0.00057 m^3/min (0.02 cfm); at the end of the run, take the difference of the measured wet test meter and DGM volumes; divide the difference by 10, to get the leak rate. The leak rate should not exceed 0.00057 m^3/min (0.02 cfm).

5.3.2 Calibration After Use. After each field use, the calibration of the metering system shall be checked by performing three calibration runs at a single, intermediate orifice setting (based on the previous field test), with the vacuum set at the maximum value reached during the test series. To adjust the vacuum, insert a valve between the wet test meter and the inlet of the metering system. Calculate the average value of the DGM calibration factor. If the value has changed by more than 5 percent, recalibrate the meter over the full range of orifice settings, as previously detailed.

Alternative procedures, e.g., rechecking the the orifice meter coefficient, may be used, subject to the approval of the Administrator.

5.3.3 Acceptable Variation in Calibration. If the DGM coefficient values obtained before and after a test series differ by more than 5 percent, the test series shall either be voided, or calculations for the test series shall be performed using whichever meter coefficient value (i.e., before or after) gives the lower value of total sample volume.

5.4 Probe Heater Calibration. The probe heating system shall be calibrated before its initial use in the field.

Use a heat source to generate air heated to selected temperatures that approximate those expected to occur in the sources to be sampled. Pass this air through the probe at a typical sample flow rate while measuring the probe inlet and outlet temperatures at various probe heater settings. For each air temperature generated, construct a graph of probe heating system setting versus probe outlet temperature. The procedure outlined in APTD-0576 can also be used. Probes constructed according to APTD-0581 need not be calibrated if

the calibration curves in APTD-0576 are used. Also, probes with outlet temperatures monitoring capabilities do not require calibration.

5.5 Temperature Gauges. Use the procedure in Section 4.3 of Method 2 to calibrate in-stack temperature gauges. Dial thermometers, such as are used for the DGM and condenser outlet, shall be calibrated against mercury-in-glass thermometers.

5.6 Leak Check of Metering System Shown in Figure 5-1. That portion of the sampling train from the pump to the orifice meter should be leak checked prior to initial use and after each shipment. Leakage after the pump will result in less volume being recorded than is actually sampled. The following procedure is suggested (see Figure 5-4): Close the main valve on the meter box. Insert a one-hole rubber stopper with rubber tubing attached into the orifice exhaust pipe. Disconnect and vent the low side of the orifice manometer. Close off the low side orifice tap. Pressurize the system to 13 to 18 cm (5 to 7 in.) water column by blowing into the rubber tubing. Pinch off the tubing, and observe the manometer for one minute. A loss of pressure on the manometer indicates a leak in the meter box; leaks, if present, must be corrected.

5.7 Barometer. Calibrate against a mercury barometer.

6. CALCULATIONS

Carry out calculations, retaining at least one extra decimal figure beyond that of the acquired data. Round off figures after the final calculation. Other forms of the equations may be used as long as they give equivalent results.

6.1 Nomenclature.

A_n = Cross-sectional area of nozzle, m^2 (ft^2).

B_{ws} = Water vapor in the gas stream, proportion by volume.

C_a = Acetone blank residue concentration, mg/g.

c_s = Concentration of particulate matter in stack gas, dry basis, corrected to standard conditions, g/dscm (g/dscf).

I = Percent of isokinetic sampling.

L_a = Maximum acceptable leakage rate for either a pretest leak check or for a leak check following a component change; equal to 0.00057 m^3/min (0.02 cfm) or 4 percent of the average sampling rate, whichever is less.

L_i = Individual leakage rate observed during the leak check conducted prior to the "i^{th}" component change (i = 1, 2, 3...n), m^3/min (cfm).

L_p = Leakage rate observed during the post-test leak check, m^3/min (cfm).

m_a = Mass of residue of acetone after evaporation, mg.

m_n = Total amount of particulate matter collected, mg.

M_w = Molecular weight of water, 18.0 g/g-mole (18.0 lb/lb-mole).

P_{bar} = Barometric pressure at the sampling site, mm Hg (in. Hg).

P_s = Absolute stack gas pressure, mm Hg (in. Hg).

P_{std} = Standard absolute pressure, 760 mm Hg (29.92 in. Hg).

R = Ideal gas constant, 0.06236 [(mm Hg)(m^3)]/[(°K)(g-mole)] {21.85 [(in. Hg)(ft^3)]/[(°R)(lb-mole)]}.

T_m = Absolute average DGM temperature (see Figure 5-2), °K (°R).

T_s = Absolute average stack gas temperature (see Figure 5-2), °K (°R).

T_{std} = Standard absolute temperature, 293°K (528°R).

V_a = Volume of acetone blank, ml.

V_{aw} = Volume of acetone used in wash, ml.

V_{lc} = Total volume liquid collected in impingers and silica gel (see Figure 5-3), ml.

V_m = Volume of gas sample as measured by dry gas meter, dcm (dcf).

$V_{m(std)}$ = Volume of gas sample measured by the dry gas meter, corrected to standard conditions, dscm (dscf).

$V_{w(std)}$ = Volume of water vapor in the gas sample, corrected to standard conditions, scm (scf).

v_s = Stack gas velocity, calculated by Method 2, Equation 2-9, using data obtained from Method 5, m/sec (ft/sec).

W_a = Weight of residue in acetone wash, mg.

Y = Dry gas meter calibration factor.

ΔH = Average pressure differential across the orifice meter (see Figure 5-2), mm H_2O (in. H_2O).

ρ_a = Density of acetone, mg/ml (see label on bottle).

ρ_w = Density of water, 0.9982 g/ml (0.002201 lb/ml).

θ = Total sampling time, min.

θ_1 = Sampling time interval, from the beginning of a run until the first component change, min.

θ_i = Sampling time interval, between two successive component changes, beginning with the interval between the first and second changes, min.

θ_p = Sampling time interval, from the final (n^{th}) component change until the end of the sampling run, min.

13.6 = Specific gravity of mercury.

60 = Sec/min.

100 = Conversion to percent.

6.2 Average Dry Gas Meter Temperature and Average Orifice Pressure Drop. See data sheet (Figure 5-2).

6.3 Dry Gas Volume. Correct the sample volume measured by the dry gas meter to standard conditions (20°C, 760 mm Hg or 68°F, 29.92 in. Hg) by using Equation 5-1.

$$V_{m(std)} = V_m Y \left(\frac{T_{std}}{T_m} \right) \left[\frac{P_{bar} + \frac{\Delta H}{13.6}}{P_{std}} \right] = K_1 V_m Y \frac{P_{bar} + \left(\frac{\Delta H}{13.6} \right)}{T_m} \qquad \text{Eq. 5-1}$$

where:

K_1 = 0.3858 °K/mm Hg for metric units,

= 17.64 °R/in. Hg for English units.

NOTE: Equation 5-1 can be used as written unless leakage rate observed during any of the mandatory leak checks (i.e., the post-test leak check or leak checks conducted prior to component changes) exceeds L_a. If L_p or L_i exceeds L_a, Equation 5-1 must be modified as follows:

(a) Case I. No component changes made during sampling run. In this case, replace V_m in Equation 5-1 with the expression:

$$[V_m - (L_p - L_a)\,\theta]$$

(b) Case II. One or more component changes made during the sampling run. In this case, replace V_m in Equation 5-1 by the expression:

$$\left[V_m - (L_1 - L_a)\,\theta_1 - \sum_{i=2}^{n} (L_i - L_a)\,\theta_i - (L_p - L_a)\,\theta_p\right]$$

and substitute only for those leakage rates (L_i or L_p) which exceed L_a.

6.4 Volume of Water Vapor.

$$V_{w(std)} = \frac{V_{lc}\rho_w R\, T_{std}}{M_w\, P_{std}} = K_2\, V_{lc} \qquad \text{Eq. 5-2}$$

where:

K_2 = 0.001333 m^3/ml for metric units,

= 0.04707 ft^3/ml for English units.

6.5 Moisture Content.

$$B_{ws} = \frac{V_{w(std)}}{V_{m(std)} + V_{w(std)}} \qquad \text{Eq. 5-3}$$

NOTE: In saturated or water droplet-laden gas streams, two calculations of the moisture content of the stack gas shall be made, one from the impinger analysis (Equation 5-3), and a second from the assumption of saturated conditions. The lower of the two values of B_{ws} shall be considered correct. The procedure for determining the moisture content based upon assumption of saturated conditions is given in the NOTE of Section 1.2 of Method 4. For the purposes of this method, the average stack gas temperature from Figure 5-2 may be used to make this determination, provided that the accuracy of the in-stack temperature sensor is ±1°C (2°F).

6.6 Acetone Blank Concentration.

$$C_a = \frac{m_a}{V_a \rho_a} \qquad \text{Eq. 5-4}$$

6.7 Acetone Wash Blank.

$$W_a = C_a V_{aw} \rho_a \qquad \text{Eq. 5-5}$$

6.8 Total Particulate Weight. Determine the total particulate matter catch from the sum of the weights obtained from Containers 1 and 2 less the acetone blank (see Figure 5-3).

NOTE: Refer to Section 4.1.5 to assist in calculation of results involving two or more filter assemblies or two or more sampling trains.

6.9 Particulate Concentration.

$$C_s = (0.001\,g/mg)\,(m_n/V_{m(std)}) \qquad \text{Eq. 5-6}$$

6.10 Conversion Factors:

From	To	Multiply by
scf	m^3	0.02832
g/ft^3	gr/ft^3	15.43
g/ft^3	lb/ft^3	2.205×10^{-3}
g/ft^3	g/m^3	35.31

6.11 Isokinetic Variation.

6.11.1 Calculation from Raw Data.

$$I = \frac{100\,T_s \left[K_3 V_{1c} + (V_m Y/T_m)\,(P_{bar} + \Delta H/13.6)\right]}{60\,\theta v_s P_s A_n} \qquad \text{Eq. 5-7}$$

where:

K_3 = 0.003454 [(mm Hg)(m^3)]/[(ml)(°K)] for metric units,

= 0.002669 [(in. Hg)(ft^3)]/[(ml)(°R)] for English units.

6.11.2 Calculation from Intermediate Values.

$$I = \frac{100\,T_s V_{m(std)} P_{std}}{60\,T_{std} v_s \theta A_n P_s (1 - B_{ws})} = \frac{K_4 T_s V_{m(std)}}{P_s v_s A_n \theta (1 - B_{ws})} \qquad \text{Eq.5-8}$$

where:

K_4 = 4.320 for metric units,

= 0.09450 for English units.

6.12 Acceptable Results. If 90 percent ≤ I ≤ 110 percent, the results are acceptable. If the PM results are low in comparison to the standard, and "I" is over 110 percent or less than 90 percent, the Administrator may opt to accept the results. Citation 4 in the Bibliography may be used to make acceptability judgments. If "I" is judged to unacceptable, reject the results, and repeat the test.

6.13 Stack Gas Velocity and Volumetric Flow Rate. Calculate the average stack gas velocity and volumetric *flow rate, if needed, using data obtained in this method and the equations in Sections 5.2 and 5.3 of Method 2.*

7. ALTERNATIVE PROCEDURES

7.1 Dry Gas Meter as a Calibration Standard. A DGM may be used as a calibration standard for volume measurements in place of the wet test meter specified in Section 5.3, provided that it is calibrated initially and recalibrated periodically as follows:

7.1.1 Standard Dry Gas Meter Calibration.

7.1.1.1. The DGM to be calibrated and used as a secondary reference meter should be of high quality and have an appropriately sized capacity, e.g., 3 liters/rev (0.1 ft^3/rev). A spirometer (400 liters or more capacity), or equivalent, may be used for this calibration, although a wet test meter is usually more practical. The wet test meter should have a capacity of 30 liters/rev (1 ft^3/rev) and capable of measuring volume to within 1.0 percent; wet test meters should be checked against a spirometer or a liquid displacement meter to ensure the accuracy of the wet test meter. Spirometers or wet test meters of other sizes may be used, provided that the specified accuracies of the procedure are maintained.

7.1.1.2 Set up the components as shown in Figure 5.7. A spirometer, or equivalent, may be used in place of the wet test meter in the system. Run the pump for at least 5 minutes at a flow rate of about 10 liters/min (0.35 cfm) to condition the interior surface of the wet test meter. The pressure drop indicated by the manometer at the inlet side of the DGM should be minimized [no greater than 100 mm H_2O (4 in. H_2O) at a flow rate of 30 liters/min (1 cfm)]. This can be accomplished by using large diameter tubing connections and straight pipe fittings.

7.1.1.3 Collect the data as shown in the example data sheet (see Figure 5-8). Make triplicate runs at each of the flow rates and at no less than five different flow rates. The range of flow rates should be between 10 and 34 liters/min (0.35 and 1.2 cfm) or over the expected operating range.

7.1.1.4 Calculate flow rate, Q, for each run using the wet test meter volume, V_w, and the run time, θ. Calculate the DGM coefficient, Y_{ds}, for each run. These calculations are as follows:

$$Q = K_1 \frac{P_{bar}}{t_w + t_{std}} \frac{V_w}{\theta}$$

$$Y_{ds} = \frac{V_w}{V_{ds}} \frac{\left(t_{ds} + t_{std}\right)}{\left(t_w + t_{std}\right)} \frac{P_{bar}}{\left(P_{bar} + \frac{\Delta p}{13.6}\right)}$$

where:

K_1 = 0.3858 for international system of units (SI); 17.64 for English units.

V_w = Wet test meter volume, liter (ft^3).

V_{ds} = Dry gas meter volume, liter (ft^3).

t_{ds} = Average dry gas meter temperature, °C (°F).

t_{std} = 273°C for SI units; 460°F for English units.

t_w = Average wet test meter temperature, °C (°F)

P_{bar} = Barometric pressure, mm Hg (in. Hg).

Δp = Dry gas meter inlet differential pressure, mm H_2O (in. H_2O).

θ = Run time, min.

7.1.1.5 Compare the three Y_{ds} values at each of the flow rates and determine the maximum and minimum values. The difference between the maximum and minimum values at each flow rate should be no greater than 0.030. Extra sets of triplicate runs may be made in order to complete this requirement. In addition, the meter coefficients should be between 0.95 and 1.05. If these specifications cannot be met in three sets of successive triplicate runs, the meter is not suitable as a calibration standard and should not be used as such. If these specifications are met, average the three Y_{ds} values at each flow rate resulting in five average meter coefficients, Y_{ds}.

7.1.1.6 Prepare a curve of meter coefficient, Y_{ds}, versus flow rate, Q, for the DGM. This curve shall be used as a reference when the meter is used to calibrate other DGM's and to determine whether recalibration is required.

7.1.2 Standard Dry Gas Meter Recalibration.

7.1.2.1 Recalibrate the standard DGM against a wet test meter or spirometer annually or after every 200 hours of operation, whichever comes first. This requirement is valid provided the standard DGM is kept in a laboratory and, if transported, cared for as any other laboratory instrument. Abuse to the standard meter may cause a change in the calibration and will require more frequent recalibrations.

7.1.2.2 As an alternative to full recalibration, a two-point calibration check may be made. Follow the same procedure and equipment arrangement as for a full recalibration, but run the meter at only two flow rates [suggested rates are 14 and 28 liters/min (0.5 and 1.0 cfm)]. Calculate the meter coefficients for these two points, and compare the values with the meter calibration curve. If the two coefficients are within 1.5 percent of the calibration curve values at the same flow rates, the meter need not be recalibrated until the next date for a recalibration check.

7.2 Critical Orifices As Calibration Standards. Critical orifices may be used as calibration standards in place of the wet test meter specified in Section 5.3, provided that they are selected, calibrated, and used as follows:

7.2.1 Selection of Critical Orifices.

7.2.1.1 The procedure that follows describes the use of hypodermic needles or stainless steel needle tubings which have been found suitable for use as critical orifices. Other materials and critical orifice designs may be used provided the orifices act as true critical orifices; i.e., a critical vacuum can be obtained, as described in Section 7.2.2.2.3. Select five critical orifices that are appropriately sized to cover the range of flow rates between 10 and
34 liters/min or the expected operating range. Two of the critical orifices should bracket the expected operating range.

A minimum of three critical orifices will be needed to calibrate a Method 5 DGM; the other two critical orifices can serve as spares and provide better selection for bracketing the range of operating flow rates. The needle sizes and tubing lengths shown below give the following approximate flow rates:

Gauge/cm	Flow rate, liters/min	Gauge/cm	Flow rate, liters/min
12/7.6	32.56	14/2.5	19.54
12/10.2	30.02	14/5.1	17.27
13/2.5	25.77	14/7.6	16.14
13/5.1	23.50	15/3.2	14.16
13/7.6	22.37	15/7.6	11.61
13/10.2	20.67	15/10.2	10.48

7.2.1.2 These needles can be adapted to a Method 5 type sampling train as follows: Insert a serum bottle stopper, 13- by 20-mm sleeve type, into a 1/2—inch Swagelok quick connect. Insert the needle into the stopper as shown in Figure 5-9.

7.2.2 **Critical Orifice Calibration.** The procedure described in this section uses the Method 5 meter box configuration with a DGM as described in Section 2.1.8 to calibrate the critical orifices. Other schemes may be used, subject to the approval of the Administrator.

7.2.2.1 **Calibration of Meter Box.** The critical orifices must be calibrated in the same configuration as they will be used; i.e., there should be no connections to the inlet of the orifice.

7.2.2.1.1 Before calibrating the meter box, leak check the system as follows: Fully open the coarse adjust valve, and completely close the by-pass valve. Plug the inlet. Then turn on the pump, and determine whether there is any leakage. The leakage rate shall be zero; i.e., no detectable movement of the DGM dial shall be seen for 1 minute.

7.2.2.1.2 Check also for leakages in that portion of the sampling train between the pump and the orifice meter. See Section 5.6 for the procedure; make any corrections, if necessary. If leakage is detected, check for cracked gaskets, loose fittings, worn O-rings, etc., and make the necessary repairs.

7.2.2.1.3 After determining that the meter box is leakless, calibrate the meter box according to the procedure given in Section 5.3. Make sure that the wet test meter meets the requirements stated in Section 7.1.1.1. Check the water level in the wet test meter. Record the DGM calibration factor, Y.

7.2.2.2 **Calibration of Critical Orifices.** Set up the apparatus as shown in Figure 5-10.

7.2.2.2.1 Allow a warm-up time of 15 minutes. This step is important to equilibrate the temperature conditions through the DGM.

7.2.2.2.2 Leak check the system as in Section 7.2.2.1.1. The leakage rate shall be zero.

7.2.2.2.3 Before calibrating the critical orifice, determine its suitability and the appropriate operating vacuum as follows: Turn on the pump, fully open the coarse adjust valve, and adjust the by-pass valve to give a vacuum reading corresponding to about half of atmospheric pressure. Observe the meter box orifice manometer reading, ΔH. Slowly increase the vacuum reading until a stable reading is obtained on the meter box orifice manometer. Record the critical vacuum for each orifice. Orifices that do not reach a critical value shall not be used.

7.2.2.2.4 Obtain the barometric pressure using a barometer as described in Section 2.1.9. Record the barometric pressure, P_{bar}, in mm Hg (in. Hg).

7.2.2.2.5 Conduct duplicate runs at a vacuum of 25 to 50 mm Hg (1 to 2 in. Hg) above the critical vacuum. The runs shall be at least 5 minutes each. The DGM volume readings shall be in increments of complete revolutions of the DGM. As a guideline, the times should not differ by more than 3.0 seconds (this includes allowance for changes in the DGM temperatures) to achieve ±0.5 percent in K'. Record the information listed in Figure 5-11.

7.2.2.2.6 Calculate K' using Equation 5-9.

$$K' = \frac{K_1 V_m Y \left(P_{bar} + \frac{\Delta H}{13.6} \right) \sqrt{T_{amb}}}{P_{bar} T_m \theta} \qquad \text{Eq. 5-9}$$

where:

$$K' = \text{Critical orifice coefficient}, \frac{m^3 K^{1/2}}{(mmHg)(min)} \left[\frac{ft^3 R^{1/2}}{(in.Hg)(min)} \right]$$

T_{amb} = Absolute ambient temperature, °K (°R).

Average the K' values. The individual K' values should not differ by more than ±0.5 percent from the average.

7.2.3 Using the Critical Orifices as Calibration Standards.

7.2.3.1 Record the barometric pressure.

7.2.3.2 Calibrate the metering system according to the procedure outlined in Sections 7.2.2.2.1 to 7.2.2.2.5. Record the information listed in Figure 5—12.

7.2.3.3 Calculate the standard volumes of air passed through the DGM and the critical orifices, and calculate the DGM calibration factor, Y, using the equations below:

$$V_{m(std)} = K_1 V_m \frac{P_{bar} + \frac{\Delta H}{13.6}}{T_m} \qquad \text{Eq. 5-10}$$

$$V_{cr(std)} = K' \frac{P_{bar} \theta}{\sqrt{T_{amb}}} \qquad \text{Eq. 5-11}$$

$$Y = \frac{V_{cr(std)}}{V_{m(std)}} \qquad \text{Eq. 5-12}$$

where:

$V_{cr(std)}$ = Volume of gas sample passed through the critical orifice, corrected to standard conditions, dscm (dscf).

K' = 0.3858 °K/mm Hg for metric units
= 17.64 °R/in. Hg for English units.

7.2.3.4 Average the DGM calibration values for each of the flow rates. The calibration factor, Y, at each of the flow rates should not differ by more than ±2 percent from the average.

7.2.3.5 To determine the need for recalibrating the critical orifices, compare the DGM Y factors obtained from two adjacent orifices each time a DGM is calibrated; for example, when checking orifice 13/2.5, use orifices 12/10.2 and 13/5.1. If any critical orifice yields a DGM Y factor differing by more than 2 percent

from the others, recalibrate the critical orifice according to Section 7.2.2.2.

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Plant _____
Location _____
Operator _____
Date _____
Run No. _____
Sample box No. _____
Meter box No. _____
Meter $\Delta H@$ _____
C factor _____
Pitot tube coefficient, Cp _____

SCHEMATIC OF STACK CROSS SECTION

Ambient temperature _____
Barometric pressure _____
Assumed moisture, % _____
Probe length, (ft.) _____
Nozzle identification No. _____
Average calibrated nozzle diameter, (in.) _____
Probe heater setting _____
Leak rate, (cfm) _____
Static pressure, (in. Hg) _____
Filter No. _____

Traverse point number	Sampling time	Vacuum	Stack temperature	Velocity head	Pressure differential across orifice meter	Gas meter reading	Gas sample temperature at dry gas meter		Filter temperature	Temperature of gas leaving condenser or last impinger
							Inlet	Outlet		
	min.	(in.Hg)	(T_s) (°F)	(ΔP_s) (in. H_2O)	(in. H_2O)	(ft^3)	(°F)	(°F)	(°F)	(°F)
Total							Avg.	Avg.		
Average							Avg.			

Figure 5-2. Particulate Field Data.

Plant _____
Date _____
Run No. _____
Filter No. _____
Amount liquid lost during transport _____
Acetone blank volume, ml _____
Acetone blank concentration, mg/mg (Equation 5-4) _____
Acetone wash blank, mg (Equation 5-5) _____

Container number	Weight of particulate collected, mg		
	Final weight	Tare weight	Weight gain
1.			
2.			
Total			
Less acetone blank			
Weight of particulate matter			

	Volume of liquid water collected	
	Impinger volume, ml	Silica gel weight, g
Final.....		
Initial.....		
Liquid collected		
Total volume collected....		g* ml

*Convert weight of water to volume by dividing total weight increase by density of water (1 g/ml).

$$\frac{\text{Increase, g}}{(1\text{g/ml})} = \text{Volume water, ml}$$

Figure 5-3. Analytical Data Sheet.

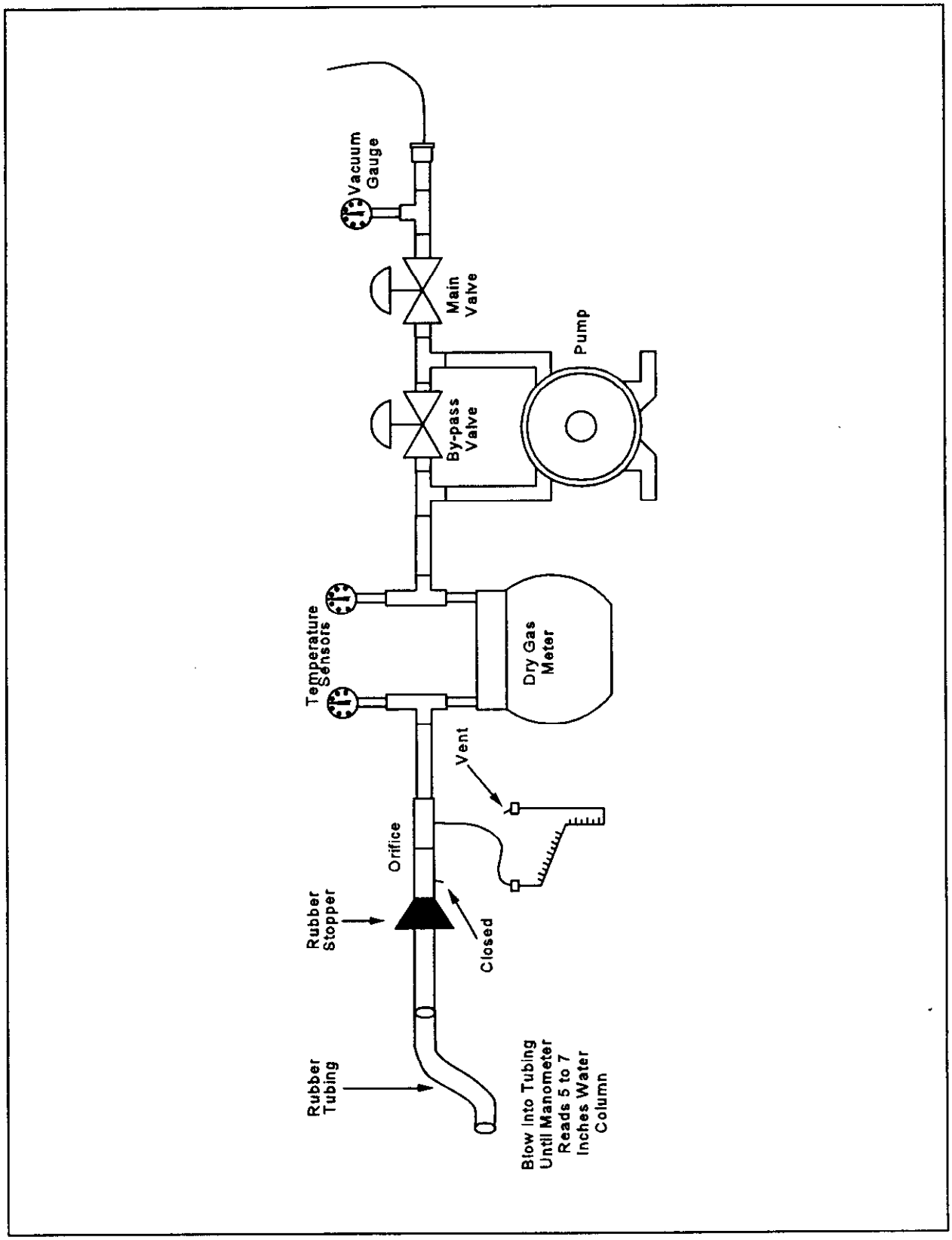

Figure 5-4. Leak Check of Meter Box.

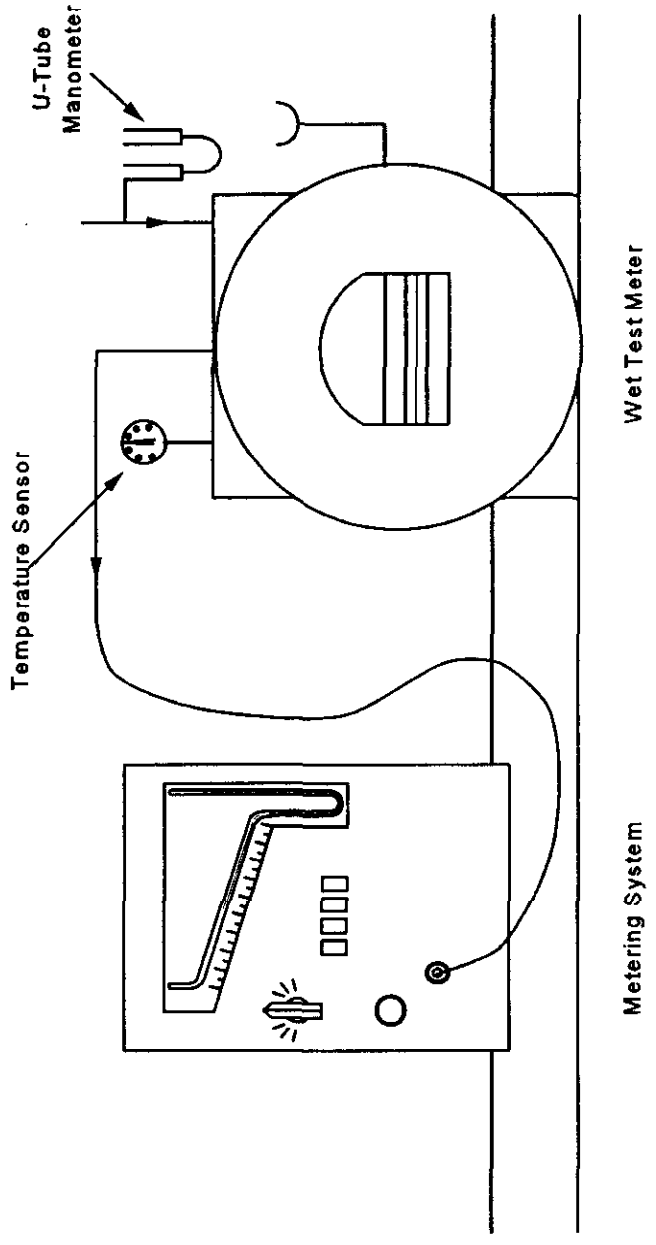

Figure 5-5. Equipment arrangement for metering system calibration.

Date: _____ Metering System Identification: _____

Barometric pressure, P_b = _____ in. Hg

Orifice manometer setting ΔH in. H_2O	Spirometer (wet meter) gas volume, V_w ft^3	Dry gas meter volume, V_m, ft^3	Temperatures				Time Θ min
			Spirometer (wet meter) T_w °F	Dry Gas Meter			
				Inlet T_i °F	Outlet T_o °F	Average T_m °F	

Calculations

ΔH in H_2O	Y				ΔH@		
Average							

Y = Ratio of reading of wet test meter to dry test meter; tolerance for individual values ± 0.02 from average.

$\Delta H_@$ = Orifice pressure differential that equates to 0.75 cfm of air @ 68°F and 29.92 inches of mercury, in. H_2O; tolerance for individual values ± 0.20 from average.

Figure 5-6. Example Data Sheet for Calibration of Metering System (English Units).

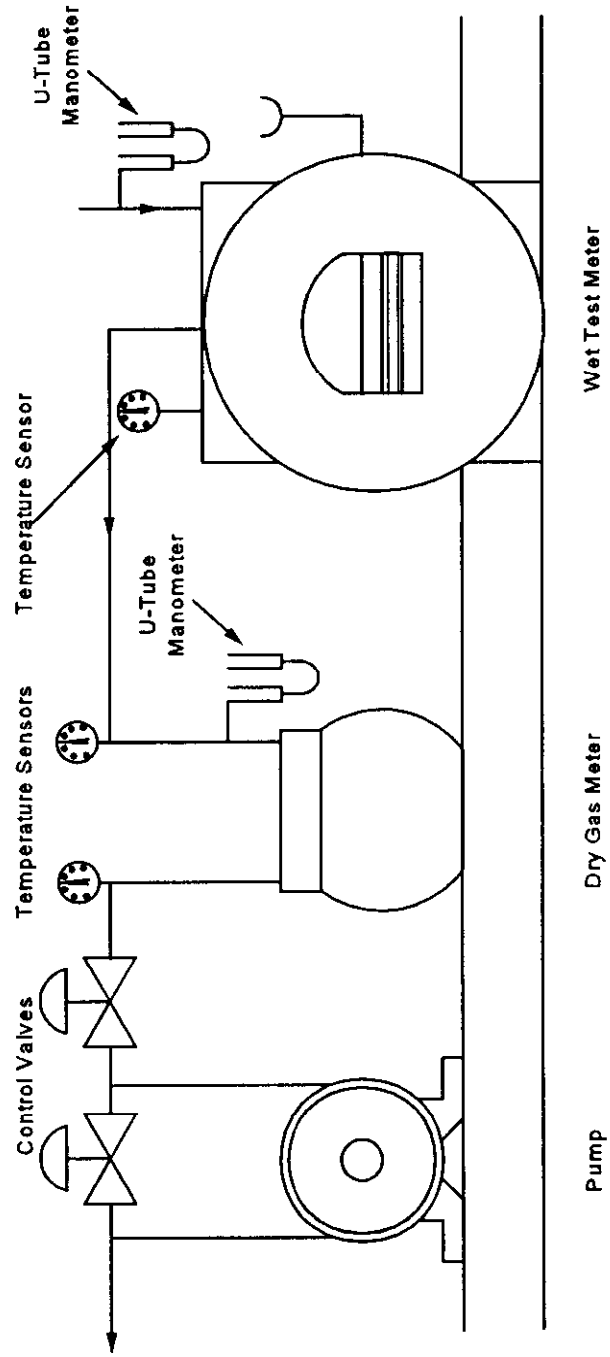

Figure 5-7. Equipment Arrangement for Dry Gas Meter Calibration.

Date: _____

Dry Gas Meter Identification: _____

Barometric Pressure (P): _____ in. Hg

Approximate Flow Rate (Q) cfm	Spirometer (Wet Meter) Gas Volume (V_s) ft^3	Dry Gas Meter Volume (V_m) ft^3	Temperatures				Dry Gas Meter Pressure (ΔP) in. H_2O	Time (θ) ft^3	Flow Rate (Q) cfm	Meter Coefficient (Y_m)	Average Meter Coefficient ($\bar{Y}_m$)
			Spirometer (Wet Meter) (t_s) °F	Dry Gas Meter							
				Inlet (t_i) °F	Outlet (t_o) °F	Average ($\bar{t}_m$) °F					
0.40											
0.60											
0.80											
1.00											
1.20											

Figure 5-8. Example Data Sheet for Calibration of a Standard Dry Gas Meter for Method 5 Sampling Equipment (English units).

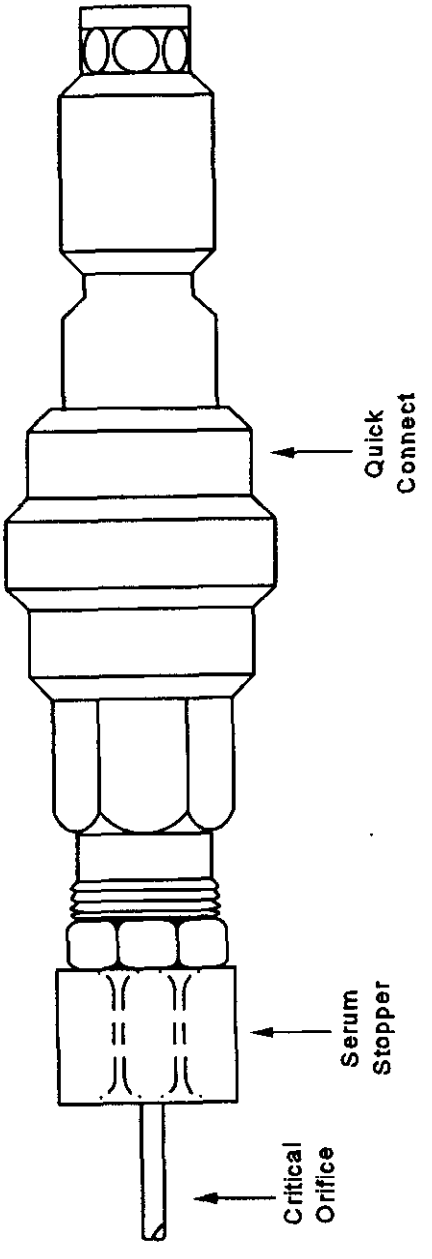

Figure 5-9. Critical Orifice Adaptation to Method 5 Metering System.

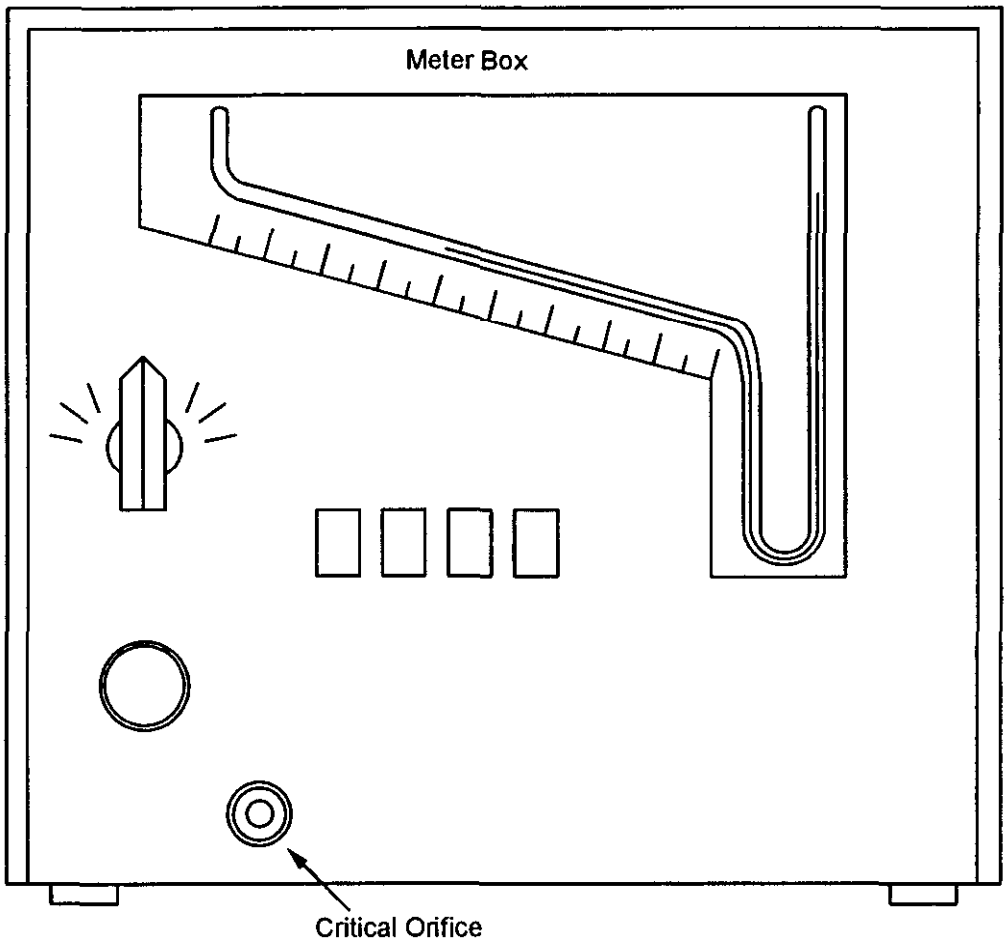

Figure 5-10. Apparatus Setup.

Date _____
Train ID _____
DGM cal. factor _____
Critical orifice ID _____

Dry gas meter		Run number	
		1	2
Final reading.....	m^3 (ft^3).....	.....	.....
Initial reading.....	m^3 (ft^3).....	.....	.....
Difference, V_m.....	m^3 (ft^3).....	.....	.....
Inlet/Outlet temperatures:			
Initial.....	°C (°F).....	/	/
Final.....	°C (°F).....	/	/
Avg. Temperature, t_m	°C (°F).....	.....	.....
Time, Θ.....	min/sec.....	/	/
	min.....	.....	.....
Orifice man. rdg., ΔH..	mm (in.) H_2O..	.....	.....
Bar. pressure, P_{bar}....	mm (in.) Hg...	.....	.....
Ambient temperature, t_{amb}	°C (°F).....	.....	.....
Pump vacuum.....	mm(in.) Hg...		
K' factor.....	.....	.....	.....
Average.....	.....	.....	.....
		.....	.....

Figure 5-11. Data sheet for determining K' factor.

Date _____
Train ID_____
Critical orifice ID _____
Critical orifice K' factor _____

Dry gas meter		Run number	
		1	2
Final reading	m^3 (ft^3)	.....	.....
Initial reading	m^3 (ft^3)	.....	.....
Difference, V_m	m^3 (ft^3)	.....	.....
Inlet/outlet temperatures:			
Initial	°C (°F)	/	/
Final	°C (°F)	/	/
Avg. Temperature, t_m	°C (°F)	.....	.....
Time, Θ	min/sec	/	/
	min	.....	.....
Orifice man. rdg., ΔH . .	min	.....	.....
Bar. pressure, P_{bar} . . .	mm (in.) H_2O . .	.....	.....
	mm (in.) Hg . . .	.....	.....
Ambient temperature, t_{amb}	°C (°F)	.....	.....
Pump vacuum	mm (in.) Hg . . .	.....	.....
$V_{m(std)}$	m^3 (ft^3)	.....	.....
$V_{cr(std)}$	m^3 (ft^3)	.....	.....
DGM cal. factor, Y . . .	. . . . .	.....	.....

Figure 5-12. Data sheet for determining DGM Y Factor.

Method 8 - Determination of Sulfuric Acid Mist and Sulfur Dioxide Emissions from Stationary Sources

1. PRINCIPLE AND APPLICABILITY

1.1 Principle. A gas sample is extracted isokinetically from the stack. The sulfuric acid mist (including sulfur trioxide) and the sulfur dioxide are separated, and both fractions are measured separately by the barium-thorin titration method.

1.2 Applicability. This method is applicable for the determination of sulfuric acid mist [including sulfur trioxide (SO_3) in the absence of other particulate matter] and sulfur dioxide (SO_2) emissions from stationary sources. Collaborative tests have shown that the minimum detectable limits of the method are 0.05 mg/m^3 (0.03 x 10^{-7} lb/ft^3) for SO_3 and 1.2 mg/m^3 (0.74 x 10^{-7} lb/ft^3) for SO_2. No upper limits have been established. Based on theoretical calculations for 200 ml of 3 percent hydrogen peroxide solution, the upper concentration limit for SO_2 in a 1.0 m^3 (35.3 ft^3) gas sample is about 12,500 mg/m^3 (7.7 x 10^{-4} lb/ft^3). The upper limit can be extended by increasing the quantity of peroxide solution in the impingers.

Possible interfering agents of this method are fluorides, free ammonia, and dimethyl aniline. If any of these interfering agents are present (this can be determined by knowledge of the process), alternative methods, subject to the approval of the Administrator, U.S. Environmental Protection Agency, are required.

Filterable particulate matter may be determined along with SO_3 and SO_2 (subject to the approval of the Administrator) by inserting a heated glass fiber filter between the probe and isopropanol impinger (see Section 2.1 of Method 6). If this option is chosen, particulate analysis is gravimetric only; H_2SO_4 acid mist is not determined separately.

2. APPARATUS

2.1 Sampling. A schematic of the sampling train used in this method is shown in Figure 8-1; it is similar to the Method 5 train, except that the filter position is different, and the filter holder does not have to be heated. Commercial models of this train are available. For those who desire to build their own, however, complete construction details are described in APTD-0581. Changes from the APTD-0581 document and allowable modifications to figure 8-1 are discussed in the following subsections.

The operating and maintenance procedures for the sampling train are described in APTD-0576. Since correct usage is important in obtaining valid results, all users should read the APTD-0576 document and adopt the operating and maintenance procedures outlined in it, unless otherwise specified herein. Further details and guidelines on operation and maintenance are given in Method 5 and should be read and followed whenever they are applicable.

2.1.1 Probe Nozzle. Same as Method 5, Section 2.1.1.

2.1.2 Probe Liner. Borosilicate or quartz glass, with a heating system to prevent visible condensation during sampling. Do not use metal probe liners.

2.1.3 Pitot Tube. Same as Method 5, Section 2.1.3.

2.1.4 Differential Pressure Gauge. Same as Method 5, Section 2.1.4.

2.1.5 Filter Holder. Borosilicate glass, with a glass frit filter support and a silicone rubber gasket. Other gasket materials, e.g., Teflon or Viton, may be used subject to the approval of the Administrator. The holder design shall provide a positive seal against leakage from the outside or around the filter. The filter holder shall be placed between the first and second impingers. Note: Do not heat the filter holder.

2.1.6 Impingers. Four, as shown in Figure 8-1. The first and third shall be of Greenburg-Smith design with standard tips. The second and fourth shall be of Greenburg-Smith design modified by replacing the insert with an approximately 13-mm (0.5-in.) ID glass tube, having an unconstricted tip located 13 mm (0.5 in.) from the bottom of the flask. Similar collection systems, subject to the approval of the Administrator, may be used.

2.1.7 Metering System. Same as Method 5, Section 2.1.8.

2.1.8 Barometer. Same as Method 5, Section 2.1.9.

2.1.9 Gas Density Determination Equipment. Same as Method 5, Section 2.1.10.

2.1.10 Temperature Gauge. Thermometer, or equivalent, to measure the temperature of the gas leaving the impinger train to within 1°C (2°F).

2.2 Sample Recovery.

2.2.1 Wash Bottles. Polyethylene or glass, 500-ml (two).

2.2.2 Graduated Cylinders. 250-ml, 1-liter. (Volumetric flasks may also be used.)

2.2.3 Storage Bottles. Leak-free polyethylene bottles, 1-liter size (two for each sampling run).

2.2.4 Trip Balance. 500-g capacity, to measure to ±0.5 g (necessary only if a moisture content analysis is to be done).

2.3 Analysis.

2.3.1 Pipettes. Volumetric 25-ml, 100-ml.

2.3.2 Burette. 50-ml.

2.3.3 Erlenmeyer Flask. 250-ml (one for each sample, blank, and standard).

2.3.4 Graduated Cylinder. 100-ml.

2.3.5 Trip Balance. 500 g capacity, to measure to ± 0.5 g.

2.3.6 Dropping Bottle. To add indicator solution, 125-ml size.

3. REAGENTS

Unless otherwise indicated, all reagents are to conform to the specifications established by the Committee on Analytical Reagents of the American Chemical Society, where such specifications are available. Otherwise, use the best available grade.

3.1 Sampling.

3.1.1 Filters. Same as in Method 5, Sections 3.1.1.

3.1.2 Silica Gel. Same as Method 5, Section 3.1.2.

3.1.3 Water. Deionized, distilled to conform to ASTM Specification D1193-77, Type 3 (incorporated by reference - see §60.17). At the option of the analyst, the $KMnO_4$ test for oxidizable organic matter may be omitted when high concentrations of organic matter are not expected to be present.

3.1.4 Isopropanol, 80 Percent. Mix 800 ml of isopropanol with 200 ml of deionized, distilled water.

Note: Experience has shown that only A.C.S. grade isopropanol is satisfactory. Tests have shown that isopropanol obtained from commercial sources occasionally has peroxide impurities that will cause erroneously high sulfuric acid mist measurement. Use the following test for detecting perioxides in each lot of isopropanol: Shake 10 ml of the isopropanol with 10 ml of freshly prepared 10 percent potassium iodide solution. Prepare a blank by similarly treating 10 ml of distilled water. After 1 minute, read the absorbance on a spectrophotometer at 352 nanometers. If the absorbance exceeds 0.1, the isopropanol shall not be used. Peroxides may be removed from isopropanol by redistilling, or by passage through a column of activated alumina. However, reagent grade isopropanol with suitably low peroxide levels is readily available from commercial sources; therefore, rejection of contaminated lots may be more efficient than following the peroxide removal procedure.

3.1.5 Hydrogen Peroxide, 3 Percent. Dilute 100 ml of 30 percent hydrogen peroxide to 1 liter with deionized, distilled water. Prepare fresh daily.

3.1.6 Crushed Ice.

3.2 Sample Recovery.

3.2.1 Water. Same as in Section 3.1.3.

3.2.2 Isopropanol, 80 Percent. Same as in Section 3.1.4.

3.3 Analysis.

3.3.1 Water. Same as 3.1.3.

3.3.2 Isopropanol, 100 percent.

3.3.3 Thorin Indicator. 1-(o-arsonophenylazo) 2-naphthol-3, 6-disulfonic acid, disodium salt, or equivalent. Dissolve 0.20 g in 100 ml of deionized, distilled water.

3.3.4 Barium Perchlorate (0.0100 Normal). Dissolve 1.95 g of barium perchlorate trihydrate ($Ba(ClO_4)_2 \cdot 3H_2O$) in 200 ml deionized, distilled water, and dilute to 1 liter with isopropanol; 1.22 g of barium chloride dihydrate ($Ba(ClO_4)_2 \cdot 2H_2O$) may be used instead of the barium perchlorate. Standardize with sulfuric acid as in Section 5.2. This solution must be protected against evaporation at all times.

3.3.5 Sulfuric Acid Standard (0.0100 N). Purchase or standardize to ± 0.0002 N against 0.0100 N NaOH that has previously been standardized against primary standard potassium acid phthalate.

3.3.6 Quality Assurance Audit Samples. Same as in Method 6, Section 3.3.6.

4. PROCEDURE

4.1 *Sampling.*

4.1.1 Pretest Preparation. Follow the procedure in Method 5, Section 4.1.1; filters should be inspected but need not be desiccated, weighed, or identified. If the effluent gas can be considered dry, i.e., moisture free, the silica gel need not be weighed.

4.1.2 Preliminary Determinations. Follow the procedure in Method 5, Section 4.1.2.

4.1.3 Preparation of Collection Train. Follow the procedure in Method 5, Section 4.1.3 (except for the second paragraph and other obviously inapplicable parts), and use Figure 8-1 instead of Figure 5-1. Replace the second paragraph with: Place 100 ml of 80 percent isopropanol in the first impinger, 100 ml of 3 percent hydrogen peroxide in both the second and third impingers; retain a portion of each reagent for use as a blank solution. Place about 200 g of silica gel in the fourth impinger. (Note: If moisture content is to be determined by impinger analysis, weigh each of the first three impingers (plus absorbing solution) to the nearest 0.5 g, and record these weights. Weigh also the silica gel (or silica gel plus container) to the nearest 0.5 g, and record.)

4.1.4 Pretest Leak-Check Procedure. Follow the basic procedure in Method 5, Section 4.1.4.1, noting that the probe heater shall be adjusted to the minimum temperature required to prevent condensation, and also that verbage such as "....plugging the inlet to the filter holder...." shall be replaced by "...plugging the inlet to the first impinger...." The pretest leak-check is optional.

4.1.5 Train Operation.

Follow the basic procedures in Method 5, Section 4.1.5, in conjunction with the following special instructions. Record the data on a sheet similar to the one in Figure 8-2.(or use Figure 5-2 in Method 5) The sampling rate shall not exceed 0.030 m^3/min (1.0 cfm) during the run. Periodically during the test, observe the connecting line between the probe and first impinger for signs of condensation. If it does occur, adjust the probe heater setting upward to the minimum temperature required to prevent condensation. If component changes become necessary during a run, a leak-check shall be done immediately before each change, according to the procedure outlined in Section 4.1.4.2 of Method 5 (with appropriate modifications, as mentioned in Section 4.1.4 of this method); record all leak rates. If the leakage rate(s) exceeds the specified rate, the tester shall either void the run or shall plan to correct the sample volume as outlined in Section 6.3 of Method 5. Immediately after component changes, leak-checks are optional. If these leak-checks are done, the procedure in Section 4.1.4.1 of Method 5 (with appropriate modifications) shall be used.

After turning off the pump and recording the final readings at the conclusion of each run, remove the probe from the stack. Conduct a post-test (mandatory) leak-check as in Section 4.1.4.3 of Method 5 (with appropriate modifications), and record the leak rate. If the post-test leakage rate exceeds the specified acceptable rate, the tester shall either correct the sample volume, as in Section 6.3 of Method 5, or shall void the run.

Drain the ice bath and, with the probe disconnected, purge the remaining part of

the train by drawing clean ambient air through the system for 15 minutes at the average flow rate used for sampling. (Note: Clean ambient air can be provided by passing air through a charcoal filter. At the option of the tester, ambient air (without cleaning) may be used.)

4.1.6 Calculation of Percent Isokinetic. Follow the procedure in Method 5, Section 4.1.6.

4.2 Sample Recovery.

4.2.1 Container No. 1. If a moisture content analysis is to be done, weigh the first impinger plus contents to the nearest 0.5 g, and record this weight.

Transfer the contents of the first impinger to a 250-ml graduated cylinder. Rinse the probe, first impinger, all connecting glassware before the filter, and the front half of the filter holder with 80-percent isopropanol. Add the filter to the solution, mix, and transfer to the storage container. Protect the solution against evaporation. Mark the level of liquid on the container, and identify the sample container.

4.2.2 Container No. 2. If a moisture content analysis is to be done, weigh the second and third impingers (plus contents) to the nearest 0.5 g, and record the weights. Also, weigh the spent silica gel (or silica gel plus impinger) to the nearest 0.5 g, and record.

Transfer the solutions from the second and third impingers to a 1-liter graduated cylinder. Rinse all connecting glassware (including back half of filter holder) between the filter and silica gel impinger with water, and add this rinse water to the cylinder. Dilute to 1 liter with water. Transfer the solution to a storage container. Mark the level of liquid on the container. Seal and identify the sample container.

4.3 Analysis. Note the level of the liquid in Containers No. 1 and 2, and confirm whether any sample was lost during shipment; note this on the analytical data sheet. If a noticeable amount of leakage has occurred, either void the sample or use methods, subject to the approval of the Administrator, to correct the final results.

4.3.1 Container No. 1. Shake the container holding the isopropanol solution and the filter. If the filter breaks up, allow the fragments to settle for a few minutes before removing a sample. Pipette a 100-ml aliquot of this solution into a 250-ml Erlenmeyer flask, add 2 to 4 drops of thorin indicator, and titrate to a pink endpoint using 0.0100 N barium perchlorate. Repeat the titration with a second aliquot of sample, and average the titration values. Replicate titrations must agree within 1 percent or 0.2 ml, whichever is greater.

4.3.2 Container No. 2. Thoroughly mix the solution in the container holding the contents of the second and third impingers. Pipette a 10-ml aliquot of sample into a 250-ml Erlenmeyer flask. Add 40 ml of isopropanol, 2 to 4 drops of thorin indicator, and titrate to a pink endpoint using 0.0100 N barium perchlorate. Repeat the titration with a second aliquot of sample, and average the titration values. Replicate titrations must agree within 1 percent or 0.2 ml, whichever is greater.

4.3.3 Blanks. Prepare blanks by adding 2 to 4 drops of thorin indicator to 100 ml of 80 percent isopropanol. Titrate the blanks in the same manner as the samples.

4.4 Quality Control Procedures. Same as in Method 5, Section 4.4.

4.5 Audit Sample Analysis. Same as in Method 6, Section 4.4.

5. CALIBRATION

5.1 Calibrate equipment using the procedures specified in the following sections of Method 5: Section 5.3 (metering system), Section 5.5 (temperature gauges), and Section 5.7 (barometer). Note that the recommended leak-check of the metering system, described in Section 5.6 of Method 5, also applies to this method.

5.2 Standardize the barium perchlorate solution with 25 ml of standard sulfuric acid, to which 100 ml of 100 percent isopropanol has been added.

6. CALCULATIONS

NOTE: Carry out calculations retaining at least one extra decimal figure beyond that of the acquired data. Round off figures after final calculation.

6.1 Nomenclature.

A_n	=	Cross-sectional area of nozzle, m^2 (ft^2).
B_{ws}	=	Water vapor in the gas stream, proportion by volume.
C_{H2SO4}	=	Sulfuric acid (including SO_3) concentration, g/dscm (lb/dscf).
C_{SO2}	=	Sulfur dioxide concentration, g/dscm (lb/dscf).
I	=	Percent of isokinetic sampling.
N	=	Normality of barium perchlorate titrant, g-equivalents/liter.
P_{bar}	=	Barometric pressure at the sampling site, mm Hg (in. Hg).
P_s	=	Absolute stack gas pressure, mm Hg (in. Hg).
P_{std}	=	Standard absolute pressure, 760 mm Hg (29.92 in. Hg).
T_m	=	Average absolute dry gas meter temperature, (see Figure 8-2), °K (°R).
T_s	=	Average absolute stack gas temperature, (see Figure 8-2), °K (°R).
T_{std}	=	Standard absolute temperature, 293°K (528°R).
V_a	=	Volume of sample aliquot titrated, 100 ml for H_2SO_4 and 10 ml for SO_2.
V_{lc}	=	Total volume of liquid collected in impingers and silica gel, ml.
V_m	=	Volume of gas sample as measured by dry gas meter, dcm (dcf).
$V_{m(std)}$	=	Volume of gas sample measured by the dry gas meter corrected to standard conditions, dscm (dscf).
v_s	=	Average stack gas velocity, calculated by method 2, equation 2-9, using data obtained from method 8, m/sec (ft/sec).
V_{soln}	=	Total volume of solution in which the H_2SO_4 or SO_2 sample is contained, 250 ml or 1000 ml, respectively.
V_t	=	Volume of barium standard solution titrant used for the sample, ml.
V_{tb}	=	Volume of barium standard solution titrant used for the blank, ml.
Y	=	Dry gas meter calibration factor.
ΔH	=	Average pressure drop across orifice meter, mm Hg (in. Hg).
θ	=	Total sampling time, min.
13.6	=	Specific gravity of Mercury.
60	=	sec/min.
100	=	Conversion to percent.

6.2 Average Dry Gas Meter Temperature and Average Orifice Pressure Drop. See data sheet (Figure 8-2).

6.3 Dry Gas Volume. Correct the sample volume measured by the dry gas meter to standard conditions (20°C and 760 mm Hg or 68°F and 29.92 in. Hg) by using Equation 8-1.

$$V_{m(std)} = V_m Y \left(\frac{T_{std}}{T_m} \right) \frac{P_{bar} + \left(\frac{\Delta H}{13.6} \right)}{P_{std}} = K_1 V_m Y \frac{P_{bar} + \left(\frac{\Delta H}{13.6} \right)}{T_m} \qquad \text{Eq. 8-1}$$

Where:

K_1 = 0.3858 °K/mm Hg for metric units.
= 17.64 °R/in. Hg for English units.

Note: If the leak rate observed during any mandatory leak-checks exceeds the specified acceptable rate, the tester shall either correct the value of V_m in Equation 8-1 (as described in Section 6.3 of Method 5), or shall invalidate the test run.

6.4 Volume of Water Vapor and Moisture Content. Calculate the volume of water vapor using Equation 5-2 of Method 5; the weight of water collected in the impingers and silica gel can be converted directly to milliliters (the specific gravity of water is 1 g/ml). Calculate the moisture content of the stack gas, using Equation 5-3 of Method 5. The Note in Section 6.5 of Method 5 also applies to this method. Note that if the effluent gas stream can be considered dry, the volume of water vapor and moisture content need not be calculated.

6.5 Sulfuric Acid Mist (Including SO_3) Concentration.

$$C_{H_2SO_4} = K_2 \frac{N (V_t - V_{tb}) \left(\frac{V_{soln}}{V_a} \right)}{V_{m(std)}} \qquad \text{Eq. 8-2}$$

Where:

K_2 = 0.04904 g/milliequivalent for metric units,
= 1.081×10^{-4} lb/meq for English units.

6.6 Sulfur Dioxide Concentration.

$$C_{SO2} = K_3 \frac{N (V_t - V_{tb}) \left(\frac{V_{soln}}{V_a} \right)}{V_{m(std)}} \qquad \text{Eq. 8-3}$$

Where:

K_3 = 0.03203 g/meq for metric units,
= 7.061×10^{-5} lb/meq for English units.

6.7 Isokinetic Variation.

6.7.1 Calculation from raw data.

$$I = \frac{100 T_s \left(K_4 V_{lc} + \left(\frac{V_m Y}{T_m} \right) \left(P_{bar} + \frac{\Delta H}{13.6} \right) \right)}{60 \theta v_s P_s A_n} \qquad \text{Eq. 8-4}$$

Where:
K_4 = 0.003464 mm Hg- m^3/ml-°K for metric units.
= 0.002676 in. Hg-ft^3/ml-°R for English units.

6.7.2 Calculation from intermediate values.

$$I = \frac{T_s V_{m(std)} P_{std} 100}{T_{std} v_s \theta A_n P_s 60 (1 - B_{ws})} = K_5 \frac{T_s V_{m(std)}}{P_s v_s A_n \theta (1 - B_{ws})} \qquad \text{Eq. 8-5}$$

Where:
K_5 = 4.320 for metric units.
= 0.09450 for English units.

6.8 Acceptable Results. If 90 percent < I < 110 percent, the results are acceptable. If the results are low in comparison to the standards and I is beyond the acceptable range, the Administrator may opt to accept the results. Use Citation 4 in the Bibliography of Method 5 to make judgements. Otherwise, reject the results and repeat the test.

6.9 Stack Gas Velocity and Volumetric Flow Rate. Calculate the average stack gas velocity and volumetric flow rate, if needed, using data obtained in this method and equations in Sections 5.2 and 5.3 of Method 2.

6.10 Relative Error (RE) for QA Audit Samples. Same as in Method 6, Section 6.4.

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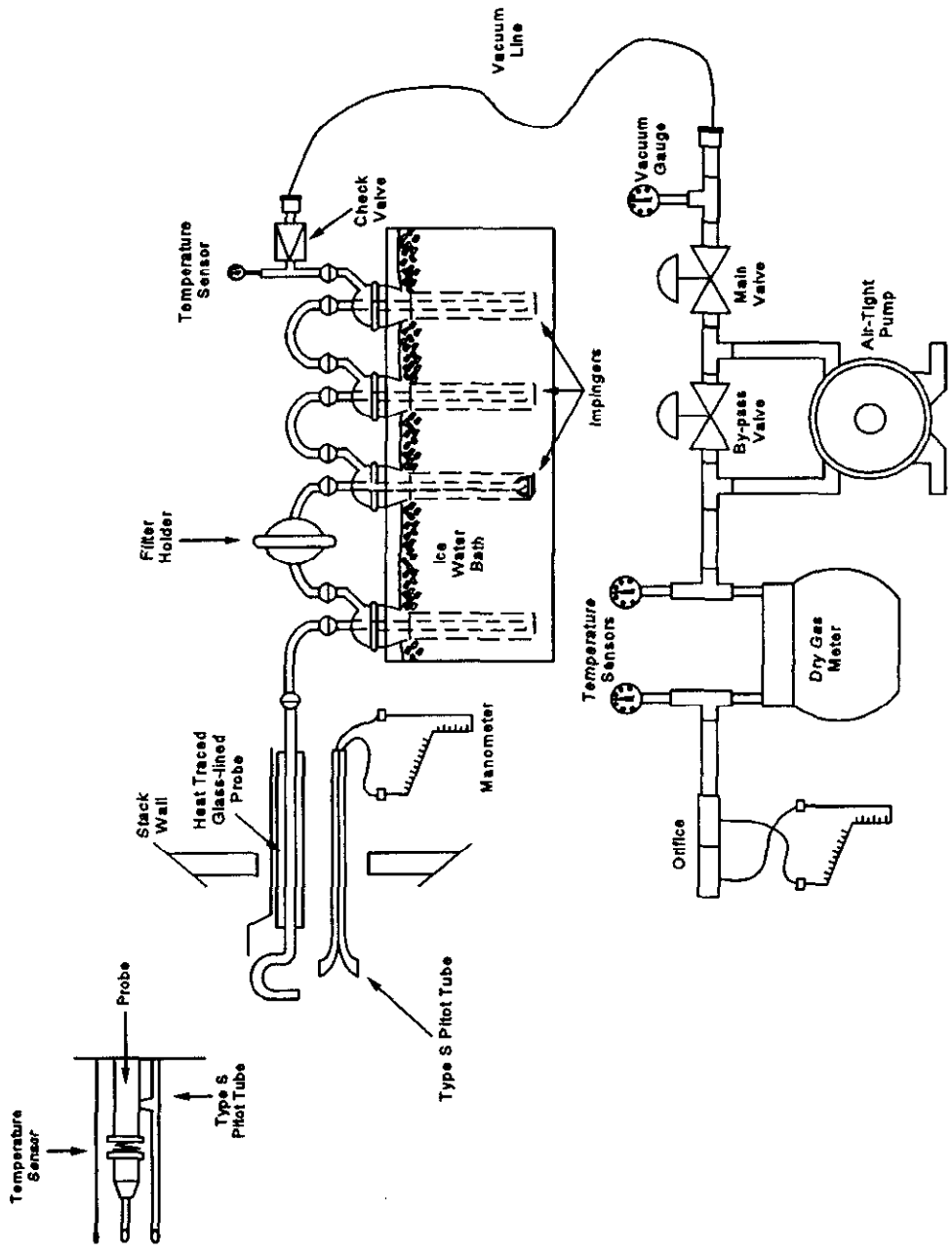

Figure 8-1. Sulfuric Acid Mist Sampling Train.

Plant _____
Location _____
Operator _____
Date _____
Run No. _____
Sample box No. _____
Meter box No. _____
Meter $\Delta H@$ _____
C factor _____
Pitot tube coefficient, C_p _____

SCHEMATIC OF STACK CROSS SECTION

Ambient temperature _____
Barometric pressure _____
Assumed moisture, % _____
Probe length, (ft.) _____
Nozzle identification No. _____
Average calibrated nozzle diameter, (in.) _____
Probe heater setting _____
Leak rate, (cfm) _____
Static pressure, (in. Hg) _____
Probe liner material _____
Filter No. _____

Traverse point number	Sampling time	Vacuum	Stack temperature	Velocity head	Pressure differential across orifice meter	Gas meter reading	Gas sample temperature at dry gas meter		Filter temperature	Temperature of gas leaving condenser or last impinger
							Inlet	Outlet		
	min.	(in.Hg)	(T_s) (°F)	(ΔP_s) (in. H_2O)	(in. H_2O)	(ft^3)	(°F)	(°F)	(°F)	(°F)
Total							Avg.	Avg.		
Average							Avg.			

Figure 8-2. Data Sheet.

METHOD 17-DETERMINATION OF PARTICULATE EMISSIONS FROM STATIONARY SOURCES (IN-STACK FILTRATION METHOD)

INTRODUCTION

Particulate matter is not an absolute quantity; rather, it is a function of temperature and pressure. Therefore, to prevent variability in particulate matter emission regulations and/or associated test methods, the temperature and pressure at which particulate matter is to be measured must be carefully defined. Of the two variables (i.e., temperature and pressure), temperature has the greater effect upon the amount of particulate matter in an effluent gas stream; in most stationary source categories, the effect of pressure appears to be negligible.

In Method 5, 250° F is established as a nominal reference temperature. Thus, where Method 5 is specified in an applicable subpart of the standards, particulate matter is defined with respect to temperature. In order to maintain a collection temperature of 250° F, Method 5 employs a heated glass sample probe and a heated filter holder. This equipment is somewhat cumbersome and requires care in its operation. Therefore, where particulate matter concentrations (over the normal range of temperature associated with a specified source category) are known to be independent of temperature, it is desirable to eliminate the glass probe and heating systems, and sample at stack temperature.

This method describes an in-stack sampling system and sampling procedures for use in such cases. It is intended to be used only when specified by an applicable subpart of the standards, and only within the applicable temperature limits (if specified), or when otherwise approved by the Administrator.

1. Principle and Applicability

1.1 Principle. Particulate matter is withdrawn isokinetically from the source and collected on a glass fiber filter maintained at stack temperature. The particulate mass is determined gravimetrically after removal of uncombined water.

1.2 Applicability. This method applies to the determination of particulate emissions from stationary sources for determining compliance with new source performance standards, only when specifically provided for in an applicable subpart of the standards. This method is not applicable to stacks that contain liquid droplets or are saturated with water vapor. In addition, this method shall not be used as written if the projected cross-sectional area of the probe extension-filter holder assembly covers more than 5 percent of the stack cross-sectional area (see Section 4.1.2).

2. Apparatus

2.1 Sampling Train. A schematic of the sampling train used in this method is shown in Figure 17-1. Construction details for many, but not all, of

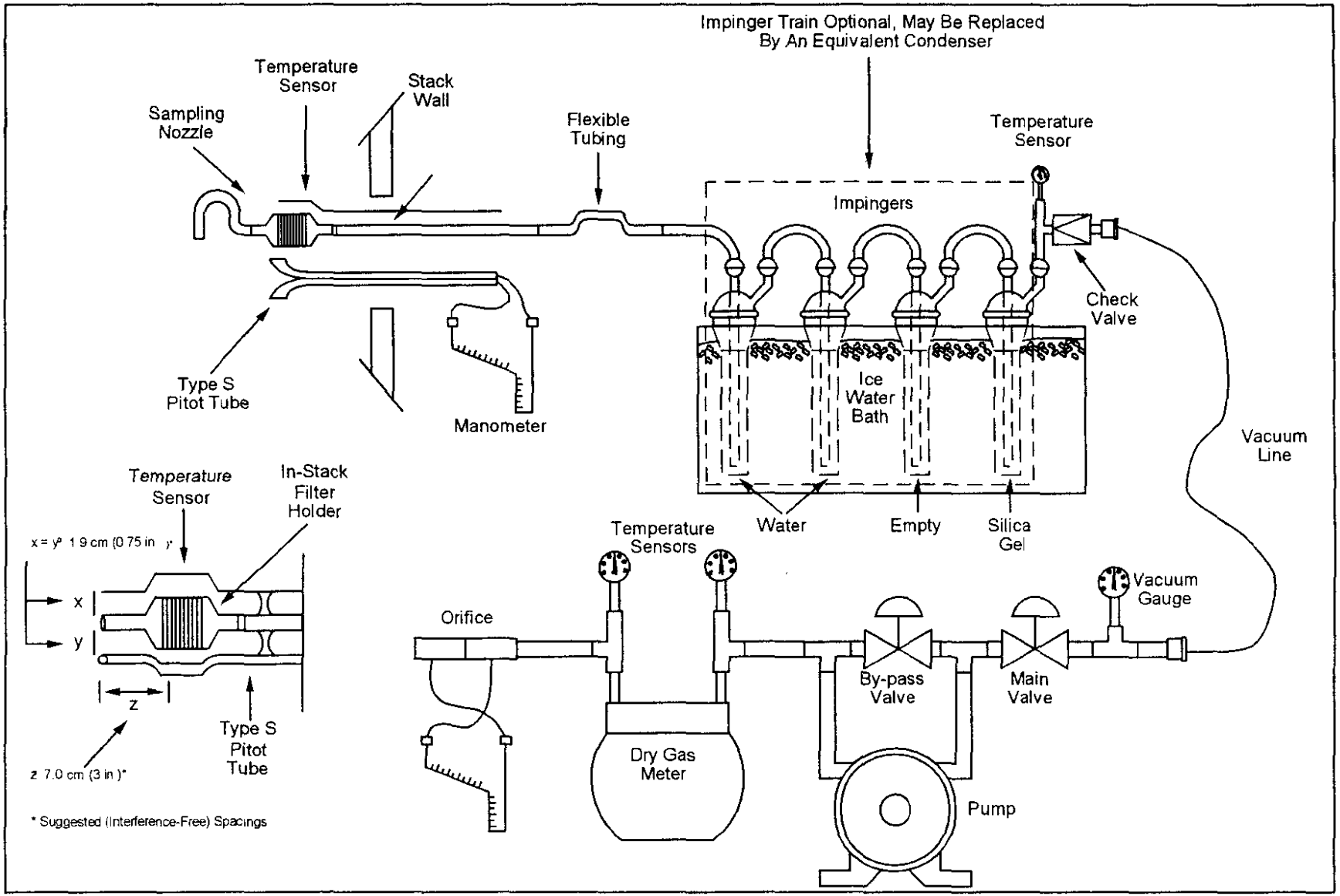

Figure 17-1. Particulate sampling train, equipped with in-stack filter.

the train components are given in APTD-0581 (Citation 2 in Bibliography); for changes from the APTD-0581 document and for allowable modifications to Figure 17-1, consult with the Administrator.

The operating and maintenance procedures for many of the sampling train components are described in APTD-0576 (Citation 3 in Bibliography). Since correct usage is important in obtaining valid results, all users should read the APTD-0576 document and adopt the operating and maintenance procedures outlined in it, unless otherwise specified herein. The sampling train consists of the following components:

2.1.1 Probe Nozzle. Stainless steel (316) or glass, with sharp, tapered leading edge. The angle of taper shall be 30° and the taper shall be on the outside to preserve a constant internal diameter. The probe nozzle shall be of the button-hook or elbow design, unless otherwise specified by the Administrator. If made of stainless steel, the nozzle shall be constructed from seamless tubing. Other materials of construction may be used subject to the approval of the Administrator.

A range of sizes suitable for isokinetic sampling should be available, e.g., 0.32 to 1.27 cm (1/8 to 1/2 in.)-or larger if higher volume sampling trains are used-inside diameter (ID) nozzles in increments of 0.16 cm (1/16 in.). Each nozzle shall be calibrated according to the procedures outlined in Section 5.1.

2.1.2 Filter Holder. The in-stack filter holder shall be constructed of borosilicate or quartz glass, or stainless steel; if a gasket is used, it shall be made of silicone rubber, Teflon, or stainless steel. Other holder and gasket materials may be used subject to the approval of the Administrator. The filter holder shall be designed to provide a positive seal against leakage from the outside or around the filter.

2.1.3 Probe Extension. Any suitable rigid probe extension may be used after the filter holder.

2.1.4 Pitot Tube. Type S, as described in Section 2.1 of Method 2, or other device approved by the Administrator; the pitot tube shall be attached to the probe extension to allow constant monitoring of the stack gas velocity (see Figure 17-1). The impact (high pressure) opening plane of the pitot tube shall be even with or above the nozzle entry plane during sampling (see Method 2, Figure 2-6b). It is recommended: (1) that the pitot tube have a known baseline coefficient, determined as outlined in Section 4 of Method 2; and (2) that this known coefficient be preserved by placing the pitot tube in an interference-free arrangement with respect to the sampling nozzle, filter holder, and temperature sensor (see Figure 17-1). Note that the 1.9 cm (0.75 in.) free-space between the nozzle and pitot tube shown in Figure 17-1, is based on a 1.3 cm (0.5 in.) ID nozzle. If the sampling train is designed for sampling at higher flow rates than that described in APTD-0581, thus necessitating the use of larger sized nozzles, the free-space shall be 1.9 cm (0.75 in.) with the largest sized nozzle in place.

Source-sampling assemblies that do not meet the minimum spacing requirements of Figure 17-1 (or the equivalent of these requirements, e.g., Figure 2-7 of Method 2) may be used; however, the pitot tube coefficients of such assemblies shall be determined by calibration, using methods subject to the approval of the Administrator.

2.1.5 Differential Pressure Gauge. Inclined manometer or equivalent device (two), as described in Section 2.2 of Method 2. One manometer shall be used for velocity head (ˆp) readings, and the other, for orifice differential pressure readings.

2.1.6 Condenser. It is recommended that the impinger system described in Method 5 be used to determine the moisture content of the stack gas. Alternatively, any system that allows measurement of both the water condensed and the moisture leaving the condenser, each to within 1 ml or 1 g, may be used. The moisture leaving the condenser can be measured either by: (1) monitoring the temperature and pressure at the exit of the condenser and using Dalton's law of partial pressures; or (2) passing the sample gas stream through a silica gel trap with exit gases kept below 20° C (68° F) and determining the weight gain.

Flexible tubing may be used between the probe extension and condenser. If means other than silica gel are used to determine the amount of moisture leaving the condenser, it is recommended that silica gel still be used between the condenser system and pump to prevent moisture condensation in the pump and metering devices and to avoid the need to make corrections for moisture in the metered volume.

2.1.7 Metering System. Vacuum gauge, leak-free pump, thermometers capable of measuring temperature to within 3° C (5.4° F), dry gas meter capable of measuring volume to within 2 percent, and related equipment, as shown in Figure 17-1. Other metering systems capable of maintaining sampling rates within 10 percent of isokinetic and of determining sample volumes to within 2 percent may be used, subject to the approval of the Administrator. When the metering system is used in conjunction with a pitot tube, the system shall enable checks of isokinetic rates.

Sampling trains utilizing metering systems designed for higher flow rates than that described in APTD-0581 or APTD-0576 may be used provided that the specifications of this method are met.

2.1.8 Barometer. Mercury, aneroid, or other barometer capable of measuring atmospheric pressure to within 2.5 mm Hg (0.1 in.
Hg). In many cases, the barometric reading may be obtained from a nearby National Weather Service station, in which case the station value (which is the absolute barometric pressure) shall be requested and an adjustment for elevation differences between the weather station and sampling point shall be applied at a rate of minus 2.5 mm Hg (0.1 in. Hg) per 30 m (100 ft) elevation increase or vice versa for elevation decrease.

2.1.9 Gas Density Determination Equipment. Temperature sensor and pressure gauge, as described in Sections 2.3 and 2.4 of Method 2, and gas analyzer, if necessary, as described in Method 3.

The temperature sensor shall be attached to either the pitot tube or to the probe extension, in a fixed configuration. If the temperature sensor is attached in the field, the sensor shall be placed in an interference-free arrangement with respect to the Type S pitot tube openings (as shown in Figure 17-1 or in Figure 2-7 of Method 2). Alternatively, the temperature sensor need not be attached to either the probe extension or pitot tube during sampling, provided that a difference of not more than 1 percent in the average velocity measurement is introduced. This alternative is subject to the approval of the Administrator.

2.2 Sample Recovery.

2.2.1 Probe Nozzle Brush. Nylon bristle brush with stainless steel wire handle. The brush shall be properly sized and shaped to brush out the probe nozzle.

2.2.2 Wash Bottles-Two. Glass wash bottles are recommended; polyethylene wash bottles may be used at the option of the tester. It is recommended that acetone not be stored in polyethylene bottles for longer than a month.

2.2.3 Glass Sample Storage Containers. Chemically resistant, borosilicate glass bottles, for acetone washes, 500 ml or 1000 ml. Screw cap liners shall either be rubber-backed Teflon or shall be constructed so as to be leak-free and resistant to chemical attack by acetone. (Narrow mouth glass bottles have been found to be less prone to leakage.) Alternatively, polyethylene bottles may be used.

2.2.4 Petri Dishes. For filter samples; glass or polyethylene, unless otherwise specified by the Administrator.

2.2.5 Graduated Cylinder and/or Balance. To measure condensed water to within 1 ml or 1 g. Graduated cylinders shall have subdivisions no greater than 2 ml. Most laboratory balances are capable of weighing to the nearest 0.5 g or less. Any of these balances is suitable for use here and in Section 2.3.4.

2.2.6 Plastic Storage Containers. Air tight containers to store silica gel.

2.2.7 Funnel and Rubber Policeman. To aid in transfer of silica gel to container; not necessary if silica gel is weighed in the field.

2.2.8 Funnel. Glass or polyethylene, to aid in sample recovery.

2.3 Analysis.

2.3.1 Glass Weighing Dishes.

2.3.2 Desiccator.

2.3.3 Analytical Balance. To measure to within 0.1 mg.

2.3.4 Balance. To measure to within 0.5 mg.

2.3.5 Beakers. 250 ml.

2.3.6 Hygrometer. To measure the relative humidity of the laboratory environment.

2.3.7 Temperature Gauge. To measure the temperature of the laboratory environment.

3. Reagents

3.1 Sampling.

3.1.1 Filters. The in-stack filters shall be glass mats or thimble fiber filters, without organic binders, and shall exhibit at least 99.95 percent efficiency (0.05 percent penetration) on 0.3 micron dioctyl phthalate smoke particles. The filter efficiency tests shall be conducted in accordance with ASTM Standard Method D2986 - 71 (Reapproved 1978) (incorporated by reference-see § 60.17). Test data from the supplier's quality control program are sufficient for this purpose.

3.1.2 Silica Gel. Indicating type, 6- to 16-mesh. If previously used, dry at 175° C (350° F) for 2 hours. New silica gel may be used as received. Alternatively, other types of desiccants (equivalent or better) may be used, subject to the approval of the Administrator.

3.1.3 Crushed Ice.

3.1.4 Stopcock Grease. Acetone-insoluble, heat-stable silicone grease. This is not necessary if screw-on connectors with Teflon sleeves, or similar, are used. Alternatively, other types of stopcock grease may be used, subject to the approval of the Administrator.

3.1.5 Water. Same as in Method 5, section 3.1.3.

3.2 Sample Recovery. Acetone, reagent grade, 0.001 percent residue, in glass bottles. Acetone from metal containers generally has a high residue blank and should not be used. Sometimes, suppliers transfer acetone to glass bottles from metal containers. Thus, acetone blanks shall be run prior to field use and only acetone with low blank values (0.001 percent) shall be used. In no case shall a blank value of greater than 0.001 percent of the weight of acetone used be subtracted from the sample weight.

3.3 Analysis.

3.3.1 Acetone. Same as 3.2.

3.3.2 Desiccant. Anhydrous calcium sulfate, indicating type. Alternatively, other types of desiccants may be used, subject to the approval of the Administrator.

4. Procedure

4.1 Sampling. The complexity of this method is such that, in order to obtain reliable results, testers should be trained and experienced with the test procedures.

4.1.1 Pretest Preparation. All components shall be maintained and calibrated according to the procedure described in APTD-0576, unless otherwise specified herein.

Weigh several 200 to 300 g portions of silica gel in air-tight containers to the nearest 0.5 g. Record the total weight of the silica gel plus container, on each container. As an alternative, the silica gel need not be preweighed, but may be weighed directly in its impinger or sampling holder just prior to train assembly.

Check filters visually against light for irregularities and flaws or pinhole leaks. Label filters of the proper size on the back side near the edge using numbering machine ink. As an alternative, label the shipping containers (glass or plastic petri dishes) and keep the filters in these containers at all times except during sampling and weighing.

Desiccate the filters at 20±5.6° C (68±10° F) and ambient pressure for at least 24 hours and weigh at intervals of at least 6 hours to a constant weight, i.e., 0.5 mg change from previous weighing; record results to the nearest 0.1 mg. During each weighing the filter must not be exposed to the laboratory atmosphere for a period greater than 2 minutes and a relative humidity above 50 percent. Alternatively (unless otherwise specified by the Administrator), the filters may be oven dried at 105° C (220° F) for 2 to 3 hours, desiccated for 2 hours, and weighed. Procedures other than those described, which account for relative humidity effects, may be used, subject to the approval of the Administrator.

4.1.2 Preliminary Determinations. Select the sampling site and the minimum number of sampling points according to Method 1 or as specified by the Administrator. Make a projected-area model of the probe extension-filter holder assembly, with the pitot tube face openings positioned along the centerline of the stack, as shown in Figure 17-2. Calculate the estimated cross-section blockage, as shown in Figure 17-2. If the blockage exceeds 5 percent of the duct cross sectional area, the tester has the following options: (1) a suitable out-of-stack filtration method may be used instead of in-stack filtration; or (2) a special in-stack arrangement, in which the sampling and velocity measurement sites are separate, may be used; for details concerning this approach, consult with the Administrator (see also Citation 10 in Bibliography). Determine the stack pressure, temperature, and the range of velocity heads using Method 2; it is recommended that a leak-check of the pitot lines (see Method 2, Section 3.1) be performed. Determine the moisture

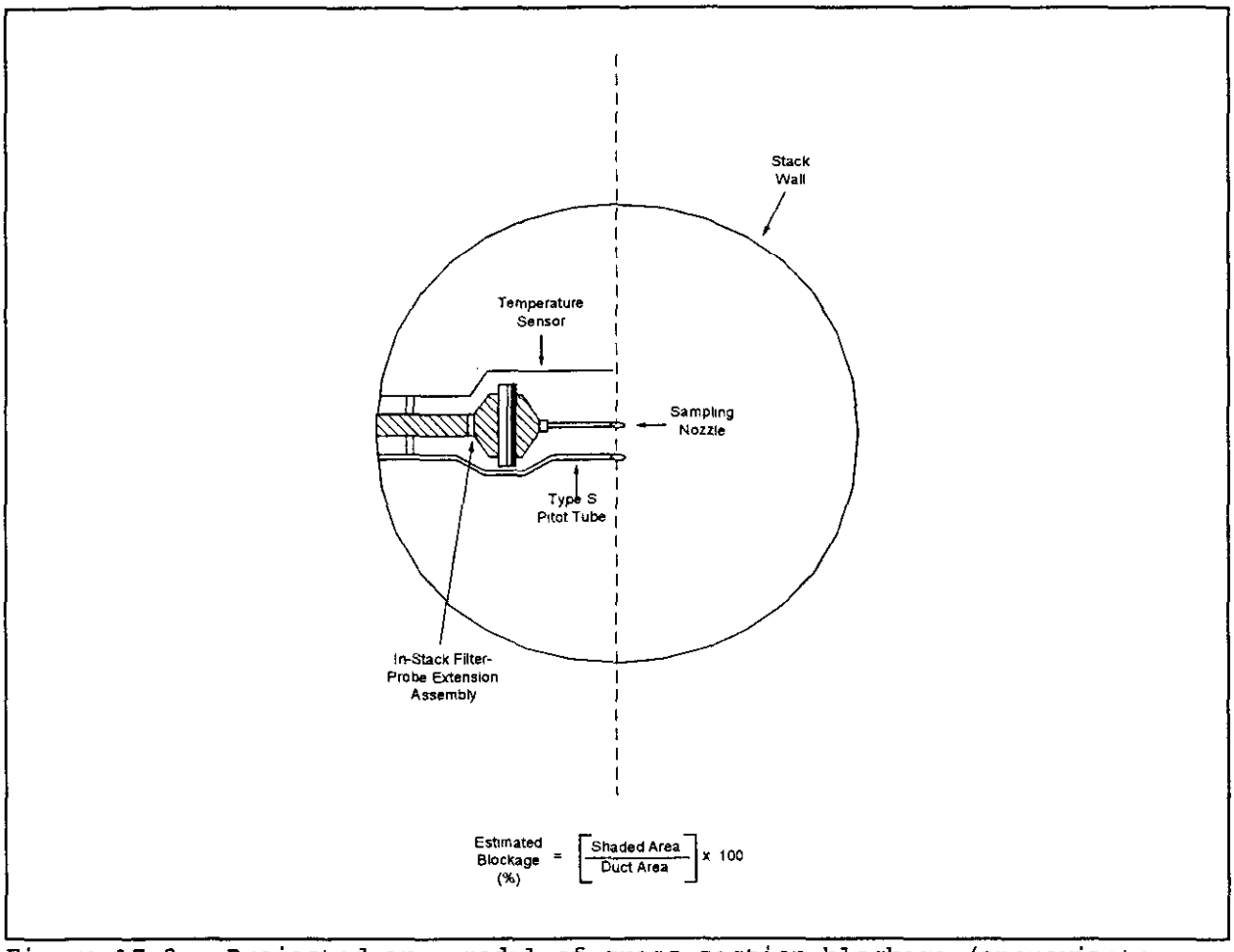

Figure 17-2. Projected-area model of cross-section blockage (approximate average for a sample traverse) caused by an in-stack filter holder-probe extension assembly.

content using Approximation Method 4 or its alternatives for the purpose of making isokinetic sampling rate settings. Determine the stack gas dry molecular weight, as described in Method 2, Section 3.6; if integrated Method 3 sampling is used for molecular weight determination, the integrated bag sample shall be taken simultaneously with, and for the same total length of time as, the particulate sample run.

Select a nozzle size based on the range of velocity heads, such that it is not necessary to change the nozzle size in order to maintain isokinetic sampling rates. During the run, do not change the nozzle size. Ensure that the proper differential pressure gauge is chosen for the range of velocity heads encountered (see Section 2.2 of Method 2).

Select a probe extension length such that all traverse points can be sampled. For large stacks, consider sampling from opposite sides of the stack to reduce the length of probes.

Select a total sampling time greater than or equal to the minimum total sampling time specified in the test procedures for the specific industry such that (1) the sampling time per point is not less than 2 minutes (or some greater time interval if specified by the Administrator), and (2) the sample volume taken (corrected to standard conditions) will exceed the required minimum total gas sample volume. The latter is based on an approximate average sampling rate.

It is recommended that the number of minutes sampled at each point be an integer or an integer plus one-half minute, in order to avoid timekeeping errors.

In some circumstances, e.g., batch cycles, it may be necessary to sample for shorter times at the traverse points and to obtain smaller gas sample volumes. In these cases, the Administrator's approval must first be obtained.

4.1.3 Preparation of Collection Train. During preparation and assembly of the sampling train, keep all openings where contamination can occur covered until just prior to assembly or until sampling is about to begin.

If impingers are used to condense stack gas moisture, prepare them as follows: place 100 ml of water in each of the first two impingers, leave the third impinger empty, and transfer approximately 200 to 300 g of preweighed silica gel from its container to the fourth impinger. More silica gel may be used, but care should be taken to ensure that it is not entrained and carried out from the impinger during sampling. Place the container in a clean place for later use in the sample recovery. Alternatively, the weight of the silica gel plus impinger may be determined to the nearest 0.5 g and recorded.

If some means other than impingers is used to condense moisture, prepare the condenser (and, if appropriate, silica gel for condenser outlet) for use.

Using a tweezer or clean disposable surgical gloves, place a labeled (identified) and weighed filter in the filter holder. Be sure that the filter is properly centered and the gasket properly placed so as not to allow the sample gas stream to circumvent the filter. Check filter for tears after assembly is completed. Mark the probe extension with heat resistant tape or by some other method to denote the proper distance into the stack or duct for each sampling point.

Assemble the train as in Figure 17-1, using a very light coat of silicone grease on all ground glass joints and greasing only the outer portion (see APTD-0576) to avoid possibility of contamination by the silicone grease. Place crushed ice around the impingers.

4.1.4 Leak Check Procedures.

4.1.4.1 Pretest Leak-Check. A pretest leak-check is recommended, but not required. If the tester opts to conduct the pretest leak-check, the following procedure shall be used.

After the sampling train has been assembled, plug the inlet to the probe nozzle with a material that will be able to withstand the stack temperature. Insert the filter holder into the stack and wait approximately 5 minutes (or longer, if necessary) to allow the system to come to equilibrium with the temperature of the stack gas stream. Turn on the pump and draw a vacuum of at least 380 mm Hg (15 in. Hg); note that a lower vacuum may be used, provided that it is not exceeded during the test. Determine the leakage rate. A leakage rate in excess of 4 percent of the average sampling rate or 0.00057 m3/min. (0.02 cfm), whichever is less, is unacceptable.

The following leak-check instructions for the sampling train described in APTD-0576 and APTD-0581 may be helpful. Start the pump with by-pass valve fully open and coarse adjust valve completely closed. Partially open the coarse adjust valve and slowly close the by-pass valve until the desired vacuum is reached. Do not reverse direction of by-pass valve. If the desired vacuum is exceeded, either leak-check at *this higher vacuum or end the leak-check as shown below and start over.*

When the leak-check is completed, first slowly remove the plug from the inlet to the probe nozzle and immediately turn off the vacuum pump. This prevents water from being forced backward and keeps silica gel from being entrained backward.

4.1.4.2 Leak-Checks During Sample Run. If, during the sampling run, a component (e.g., filter assembly or impinger) change becomes necessary, a leak-check shall be conducted immediately before the change is made. The leak-check shall be done according to the procedure outlined in Section 4.1.4.1 above, except that it shall be done at a vacuum equal to or greater than the maximum value recorded up to that point in the test. If the leakage rate is found to be no greater than 0.00057 m3/min (0.02 cfm) or 4 percent of the average sampling rate (whichever is less), the results are acceptable, and no correction will need to be applied to the total volume of dry gas metered; if, however, a higher leakage rate is obtained, the tester shall either record the leakage rate and plan to correct the sample volume as shown in Section 6.3 of this method, or shall void the sampling run.

Immediately after component changes, leak-checks are optional; if such leak-checks are done, the procedure outlined in Section 4.1.4.1 above shall be used.

4.1.4.3 Post-Test Leak-Check. A leak-check is mandatory at the conclusion of each sampling run. The leak-check shall be done in accordance with the procedures outlined in Section 4.1.4.1, except that it shall be conducted at a vacuum equal to or greater than the maximum value reached during the sampling run. If the leakage rate is found to be no greater than 0.00057 m3/min (0.02 cfm) or 4 percent of the average sampling rate (whichever is less), the results are acceptable, and no correction need be applied to the total volume of dry gas metered. If, however, a higher leakage rate is obtained, the tester shall either record the leakage rate and correct the sample volume as shown in Section 6.3 of this method, or shall void the sampling run.

4.1.5 Particulate Train Operation. During the sampling run, maintain a sampling rate such that *sampling is within 10 percent of true isokinetic, unless otherwise specified by the Administrator.*

For each run, record the data required on the example data sheet shown in Figure 17-3. Be sure to record the initial dry gas meter reading. Record the dry gas meter readings at the beginning and end of each sampling time increment, when changes in flow rates are made, before and after each leak check, and when sampling is halted. Take other readings required by Figure 17-3 at least once at each sample point during each time increment and additional readings when significant changes (20 percent variation in velocity head readings) necessitate additional adjustments in flow rate. Level and zero the manometer. Because the

manometer level and zero may drift due to vibrations and temperature changes, make periodic checks during the traverse.

Clean the portholes prior to the test run to minimize the chance of sampling the deposited material. *To begin sampling, remove the nozzle cap and verify that the pitot tube and probe extension are properly* positioned. Position the nozzle at the first traverse point with the tip pointing directly into the gas stream. Immediately start the pump and adjust the flow to isokinetic conditions. Nomographs are available, which aid in the rapid

Plant _____
Location _____
Operator _____
Date _____
Run No. _____
Sample box No. _____
Meter box No. _____
Meter DH@ _____
C factor _____
Pitot tube coefficient, Cp _____

SCHEMATIC OF STACK CROSS SECTION

Barometric pressure _____
Assumed moisture, % _____
Probe length, (ft.) _____
Nozzle identification No. _____
Average calibrated nozzle diameter, (in.) _____
Filter No. _____
Leak rate, (cfm) _____
Static pressure, (in. Hg) _____

Traverse point number	Sampling time	Vacuum	Stack temperature	Velocity head	Pressure differential across *orifice meter* DH	Gas meter reading	Gas sample temperature at dry gas meter		Filter temperature	Temperature of gas leaving condenser or last impinger
						V_m	Inlet	Outlet		
	min.	(in.Hg)	(T_s) (°F)	(DP_s) (in. H_2O)	(in. H_2O)	(ft^3)	(°F)	(°F)	(°F)	(°F)
Total							Avg.	Avg.		
Average							(T_m)			

Figure 17-3. Particulate Field Data Sheet.

adjustment to the isokinetic sampling rate without excessive computations. These nomographs are designed for use when the Type S pitot tube coefficient is 0.85±0.02, and the stack gas equivalent density (dry molecular weight) is equal to 29±4. APTD-0576 details the procedure for using the nomographs. If C_p and M_d are outside the above stated ranges, do not use the nomographs unless appropriate steps (see Citation 7 in Bibliography) are taken to compensate for the deviations.

When the stack is under significant negative pressure (height of impinger stem), take care to close the coarse adjust valve before inserting the probe extension assembly into the stack to prevent water from being forced backward. If necessary, the pump may be turned on with the coarse adjust valve closed.

When the probe is in position, block off the openings around the probe and porthole to prevent unrepresentative dilution of the gas stream.

Traverse the stack cross section, as required by Method 1 or as specified by the Administrator, being careful not to bump the probe nozzle into the stack walls when sampling near the walls or when removing or inserting the probe extension through the portholes, to minimize chance of extracting deposited material.

During the test run, take appropriate steps (e.g., adding crushed ice to the impinger ice bath) to maintain a temperature of less than 20° C (68° F) at the condenser outlet; this will prevent excessive moisture losses. Also, periodically check the level and zero of the manometer.

If the pressure drop across the filter becomes too high, making isokinetic sampling difficult to maintain, the filter may be replaced in the midst of a sample run. It is recommended that another complete filter holder assembly be used rather than attempting to change the filter itself. Before a new filter holder is installed, conduct a leak check, as outlined in Section 4.1.4.2. The total particulate weight shall include the summation of all filter assembly catches.

A single train shall be used for the entire sample run, except in cases where simultaneous sampling is required in two or more separate ducts or at two or more different locations within the same duct, or, in cases where equipment failure necessitates a change of trains. In all other situations, the use of two or more trains will be subject to the approval of the Administrator. Note that when two or more trains are used, a separate analysis of the collected particulate from each train shall be performed, unless identical nozzle sizes were used on all trains, in which case the particulate catches from the individual trains may be combined and a single analysis performed.

At the end of the sample run, turn off the pump, remove the probe extension assembly from the stack, and record the final dry gas meter reading. Perform a leak-check, as outlined in Section 4.1.4.3. Also, leak-check the pitot lines as described in Section 3.1 of Method 2; the lines must pass this leak-check, in order to validate the velocity head data.

4.1.6 Calculation of Percent Isokinetic. Calculate percent isokinetic (see Section 6.11) to determine whether another test run should be made. If there is difficulty in maintaining isokinetic rates due to source conditions, consult with the Administrator for possible variance on the isokinetic rates.

4.2 Sample Recovery. Proper cleanup procedure begins as soon as the probe extension assembly is removed from the stack at the end of the sampling period. Allow the assembly to cool.

When the assembly can be safely handled, wipe off all external particulate matter near the tip of the probe nozzle and place a cap over it to prevent losing or gaining particulate matter. Do not cap off the probe tip tightly while the sampling train is cooling down as this would create a vacuum in the filter holder, forcing condenser water backward.

Before moving the sample train to the cleanup site, disconnect the filter holder-probe nozzle assembly from the probe extension; cap the open inlet of the probe extension. Be careful not to lose any condensate, if present. Remove the umbilical cord from the condenser outlet and cap the outlet. If a flexible line is used between the first impinger (or condenser) and the probe extension, disconnect the line at the probe extension and let any condensed water or liquid drain into the impingers or condenser. Disconnect the probe extension from the condenser; cap the probe extension outlet. After wiping off the silicone grease, cap off the condenser inlet. Ground glass stoppers, plastic caps, or serum caps (whichever are appropriate) may be used to close these openings.

Transfer both the filter holder-probe nozzle assembly and the condenser to the cleanup area. This area should be clean and protected from the wind so that the chances of contaminating or losing the sample will be minimized.

Save a portion of the acetone used for cleanup as a blank. Take 200 ml of this acetone directly from the wash bottle being used and place it in a glass sample container labeled "acetone blank." Inspect the train prior to and during disassembly and note any abnormal conditions. Treat the samples as follows:

Container No. 1. Carefully remove the filter from the filter holder and place it in its identified petri dish container. Use a pair of tweezers and/or clean disposable surgical gloves to handle the filter. If it is necessary to fold the filter, do so such that the particulate cake is inside the fold. Carefully transfer to the petri dish any particulate matter and/or filter fibers which adhere to the filter holder gasket, by using a dry Nylon bristle brush and/or a sharp-edged blade. Seal the container.

Container No. 2. Taking care to see that dust on the outside of the probe nozzle or other exterior surfaces does not get into the sample, quantitatively recover particulate matter or any condensate from the probe nozzle, fitting, and front half of the filter holder by washing these components with acetone and placing the wash in a glass container. Distilled water may be used instead of acetone when approved by the

Administrator and shall be used when specified by the Administrator; in these cases, save a water blank and follow Administrator's directions on analysis. Perform the acetone rinses as follows:
Carefully remove the probe nozzle and clean the inside surface by rinsing with acetone from a wash bottle and brushing with a Nylon bristle brush. Brush until acetone rinse shows no visible particles, after which make a final rinse of the inside surface with acetone.

Brush and rinse with acetone the inside parts of the fitting in a similar way until no visible particles remain. A funnel (glass or polyethylene) may be used to aid in transferring liquid washes to the container. Rinse the brush with acetone and quantitatively collect these washings in the sample container. Between sampling runs, keep brushes clean and protected from contamination.

After ensuring that all joints are wiped clean of silicone grease (if applicable), clean the inside of the front half of the filter holder by rubbing the surfaces with a Nylon bristle brush and rinsing with acetone. Rinse each surface three times or more if needed to remove visible particulate. Make final rinse of the brush and filter holder. After all acetone washings and particulate matter are collected in the sample container, tighten the lid on the sample container so that acetone will not leak out when it is shipped to the laboratory. Mark the height of the fluid level to determine whether or not leakage occurred during transport. Label the container to clearly identify its contents.

Container No. 3. If silica gel is used in the condenser system for mositure content determination, note the color of the gel to determine if it has been completely spent; make a notation of its condition. Transfer the silica gel back to its original container and seal. A funnel may make it easier to pour the silica gel without spilling, and a rubber policeman may be used as an aid in removing the silica gel. It is not necessary to remove the small amount of dust particles that may adhere to the walls and are difficult to remove. Since the gain in weight is to be used for moisture calculations, do not use any water or other liquids to transfer the silica gel. If a balance is available in the field, follow the procedure for Container No. 3 under "Analysis."

Condenser Water. Treat the condenser or impinger water as follows: make a notation of any color or film in the liquid catch. Measure the liquid volume to within ±1 ml by using a graduated cylinder or, if a balance is available, determine the liquid weight to within ±0.5 g. Record the total volume or weight of liquid present. This information is required to calculate the moisture content of the effluent gas. Discard the liquid after measuring and recording the volume or weight.

4.3 Analysis. Record the data required on the example sheet shown in Figure 17-4. Handle each sample container as follows:

Container No. 1. Leave the contents in the shipping container or transfer the filter and any loose particulate from the sample container to a tared glass weighing dish. Desiccate for 24 hours in a desiccator containing anhydrous calcium sulfate. Weigh to a constant weight and report the results to the nearest 0.1 mg. For purposes of this Section, 4.3, the term "constant weight" means a difference of no more than 0.5 mg or 1 percent of total weight less tare weight, whichever is greater, between two consecutive weighings, with no less than 6 hours of desiccation time between weighings.

Alternatively, the sample may be oven dried at the average stack temperature or 105° C (220° F), whichever is less, for 2 to 3 hours, cooled in the desiccator, and weighed to a constant weight, unless otherwise specified by the Administrator. The tester may also opt to oven dry the sample at the average stack temperature or 105° C (220° F), whichever is less, for 2 to 3 hours, weigh the sample, and use this weight as a final weight.

Container No. 2. Note the level of liquid in the container and confirm on the analysis sheet whether or not leakage occurred during transport. If a noticeable amount of leakage has occurred, either void the sample or use methods, subject to the approval of the Administrator, to correct the final results. Measure the liquid in this container either volumetrically to ±1 ml or gravimetrically to ±0.5 g. Transfer the contents to a tared 250-ml beaker and evaporate to dryness at ambient temperature and pressure. Desiccate for 24 hours and weigh to a constant weight. Report the results to the nearest 0.1 mg.

Container No. 3. This step may be conducted in the field. Weigh the spent silica gel (or silica gel plus impinger) to the nearest 0.5 g using a balance.

FIGURE 17-4 - ANALYTICAL DATA

Plant _____
Date _____
Run No. _____
Filter No. _____
Amount liquid lost during transport _____
Acetone blank volume, ml _____
Acetone blank concentration, mg/mg (Equation 17-4) _____
Acetone wash blank, mg (Equation 17-5) _____

Container number	Weight of particulate collected, mg		
	Final weight	Tare weight	Weight gain
1.			
2.			
Total			
Less			
acetone blank			
Weight of particulate matter			

	Volume of liquid water collected	
	Impinger volume, ml	Silica gel weight, g
Final.....		
Initial.....		
Liquid collected		
Total volume collected....		g* ml

*Convert weight of water to volume by dividing total weight increase by density of water (1 g/ml).

$$\frac{\text{Increase, g}}{(1\text{g/ml})} = \text{Volume water, ml}$$

"Acetone Blank" Container. Measure acetone in this container either volumetrically or gravimetrically. Transfer the acetone to a tared 250-ml beaker and evaporate to dryness at ambient temperature and pressure. Desiccate for 24 hours and weigh to a constant weight. Report the results to the nearest 0.1 mg.

Note: At the option of the tester, the contents of Container No. 2 as well as the acetone blank container may be evaporated at temperatures higher than ambient. If evaporation is done at an elevated temperature, the temperature must be below the boiling point of the solvent; also, to prevent "bumping," the evaporation process must be closely supervised, and the contents of the beaker must be swirled occasionally to maintain an even temperature. Use extreme care, as acetone is highly flammable and has a low flash point.

5. Calibration

Maintain a laboratory log of all calibrations.

5.1 Probe Nozzle. Probe nozzles shall be calibrated before their initial use in the field. Using a micrometer, measure the inside diameter of the nozzle to the nearest 0.025 mm (0.001 in.). Make three separate measurements using different diameters each time, and obtain the average of the measurements. The difference between the high and low numbers shall not exceed 0.1 mm (0.004 in.). When nozzles become nicked, dented, or corroded, they shall be reshaped, sharpened, and recalibrated before use. Each nozzle shall be permanently and uniquely identified.

5.2 Pitot Tube. If the pitot tube is placed in an interference-free arrangement with respect to the other probe assembly components, its baseline (isolated tube) coefficient shall be determined as outlined in Section 4 of Method 2. If the probe assembly is not interference-free, the pitot tube assembly coefficient shall be determined by calibration, using methods subject to the approval of the Administrator.

5.3 Metering System. Before its initial use in the field, the metering system shall be calibrated according to the procedure outlined in APTD-0576. Instead of physically adjusting the dry gas meter dial readings to correspond to the wet test meter readings, calibration factors may be used to mathematically correct the gas meter dial readings to the proper values.

Before calibrating the metering system, it is suggested that a leak-check be conducted. For metering systems having diaphragm pumps, the normal leak-check procedure will not detect leakages within the pump. For these cases the following leak-check procedure is suggested: make a 10-minute calibration run at 0.00057 m3/min (0.02 cfm); at the end of the run, take the difference of the measured wet test meter and dry gas meter volumes; divide the difference by 10, to get the leak rate. The leak rate should not exceed 0.00057 m3/min (0.02 cfm).

After each field use, the calibration of the metering system shall be checked by performing three calibration runs at a single, intermediate orifice setting (based on the previous field test), with the vacuum set at the maximum value reached during the test series. To adjust the vacuum, insert a valve between the wet test meter and the inlet of the metering system. Calculate the average value of the calibration factor. If the calibration has changed by more than 5 percent, recalibrate the meter over the full range of orifice settings, as outlined in APTD-0576.

Alternative procedures, e.g., using the orifice meter coefficients, may

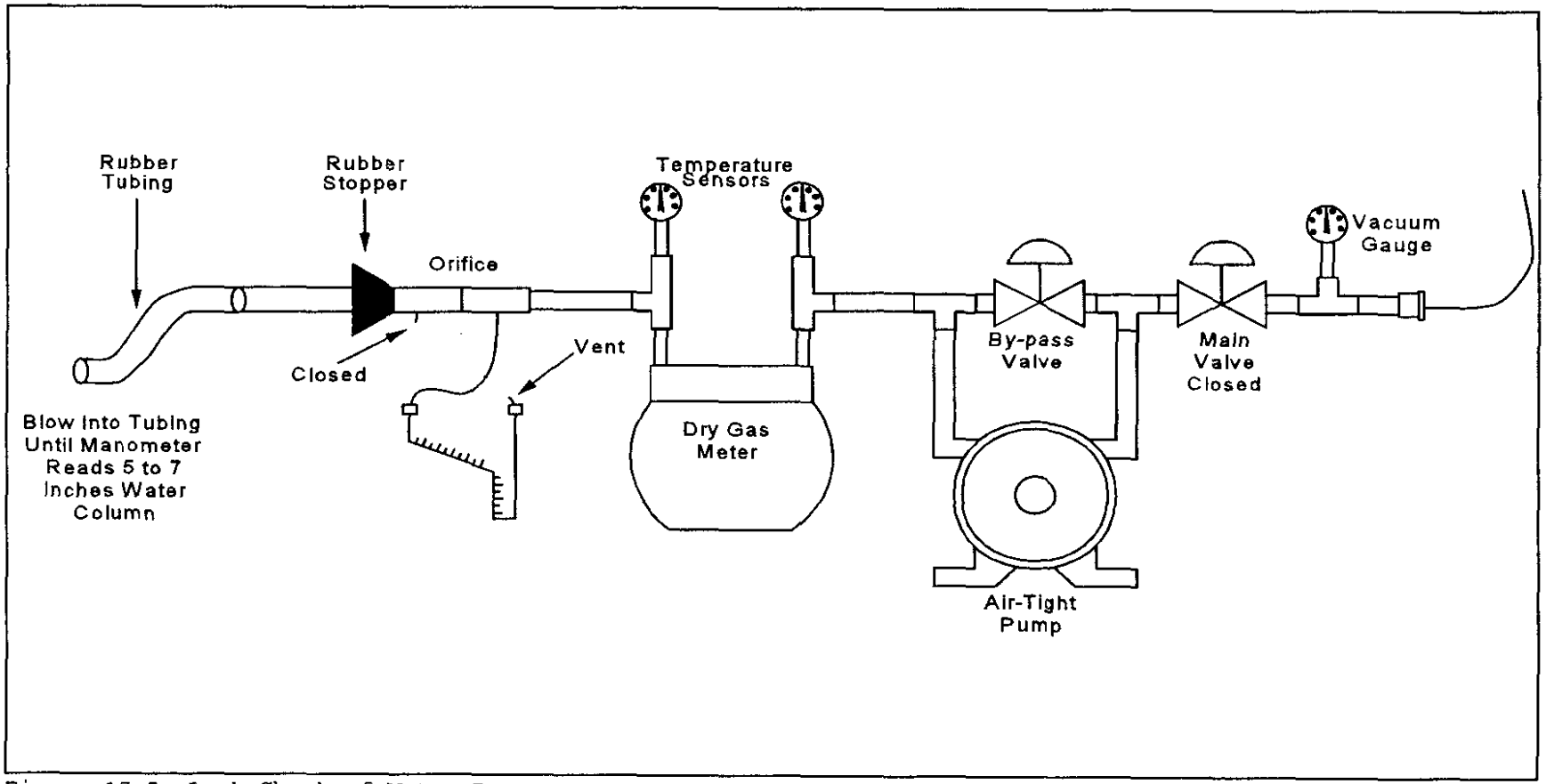

Figure 17-5. Leak Check of Meter Box.

be used, subject to the approval of the Administrator.

Note: If the dry gas meter coefficient values obtained before and after atest series differ by more than 5 percent, the test series shall either be voided, or calculations for the test series shall be performed using whichever meter coefficient value (i.e., before or after) gives the lower value of total sample volume.

5.4 Temperature Gauges. Use the procedure in Section 4.3 of Method 2 to calibrate in-stack temperature gauges. Dial thermometers, such as are used for the dry gas meter and condenser outlet, shall be calibrated against mercury-in-glass thermometers.

5.5 Leak Check of Metering System Shown in Figure 17-1. That portion of the sampling train from the pump to the orifice meter should be leak checked prior to initial use and after each shipment. Leakage after the pump will result in less volume being recorded than is actually sampled. The following procedure is suggested (see Figure 17-5). Close the main valve on the meter box. Insert a one-hole rubber stopper with rubber tubing attached into the orifice exhaust pipe. Disconnect and vent the low side of the orifice manometer. Close off the low side orifice tap. Pressurize the system to 13 to 18 cm (5 to 7 in.) water column by blowing into the rubber tubing. Pinch off the tubing and observe the manometer for one minute. A loss of pressure on the manometer indicates a leak in the meter box; leaks, if present, must be corrected.

5.6 Barometer. Calibrate against a mercury barometer.

6. Calculations

Carry out calculations, retaining at least one extra decimal figure beyond that of the acquired data. Round off figures after the final calculation. Other forms of the equations may be used as long as they give equivalent results.

6.1 Nomenclature.

A_n = Cross-sectional area of nozzle, m^2 (ft^2).
B_{ws} = Water vapor in the gas stream, proportion by volume.
C_a = Acetone blank residue concentration, mg/mg.
c_s = Concentration of particulate matter in stack gas, dry basis, corrected to standard conditions, g/dscm (g/dscf).
I = Percent of isokinetic sampling.
L_a = Maximum acceptable leakage rate for either a pretest leak check or for a leak check following a component change; equal to 0.00057 m^3/min (0.02 cfm) or 4 percent of the average sampling rate, whichever is less.
L_i = *Individual leakage rate observed during the leak check conducted prior to the "ith"* component change (i=1, 2, 3 . . . n), m^3/min (cfm).
L_p = Leakage rate observed during the post-test leak check, m^3/min (cfm).
m_a = Mass of residue of acetone after evaporation, mg.
m_n = Total amount of particulate matter collected, mg.
M_w = Molecular weight of water, 18.0 g/g-mole (18.0 lb/lb-mole).
P_{bar} = Barometric pressure at the sampling site, mm Hg (in. Hg).
P_s = Absolute stack gas pressure, mm Hg (in. Hg).
P_{std} = Standard absolute pressure, 760 mm Hg (29.92 in. Hg).
R = Ideal gas constant, 0.06236 mm Hg-m3/°K-g-mole (21.85 in. Hg-ft^3/°R-lb-mole).
T_m = Absolute average dry gas meter temperature (see Figure 17 3), °K (°R).
T_s = Absolute average stack gas temperature (see Figure 17-3), °K (°R).
T_{std} = Standard absolute temperature, 293°K (528°R).
V_a = Volume of acetone blank, ml.
V_{aw} = Volume of acetone used in wash, ml.
V_{lc} = Total volume of liquid collected in impingers and silica gel (see Figure 17-4), ml.
V_m = Volume of gas sample as measured by dry gas meter, dcm (dcf).
$V_{m(std)}$ =Volume of gas sample measured by the dry gas meter, corrected to standard conditions, dscm (dscf).
$V_{w(std)}$ =Volume of water vapor in the gas sample, corrected to standard conditions, scm (scf).
v_s = Stack gas velocity, calculated by Method 2, Equation 2-9, using data obtained from Method 17, m/sec (ft/sec).
W_a = Weight of residue in acetone wash, mg.
Y = Dry gas meter calibration coefficient.
ΔH = Average pressure differential across the orifice meter (see Figure 17-3), mm H2O (in. H2O).
ρ_a = Density of acetone, mg/ml (see label on bottle).
ρ_w = Density of water, 0.9982 g/ml (0.002201 lb/ml).
θ = Total sampling time, min.
θ_i = Sampling time interval, from the beginning of a run until the first component change, min.

θ_i = Sampling time interval, between two successive component changes, beginning with the interval between the first and second changes, min.
θ_p = Sampling time interval, from the final (nth) component change until the end of the sampling run, min.
13.6 = Specific gravity of mercury.
60 = Sec/min.
100 = Conversion to percent.

6.2 Average Dry Gas Meter Temperature and Average Orifice Pressure Drop. See data sheet (Figure 17-3).

6.3 Dry Gas Volume. Correct the sample volume measured by the dry gas meter to standard conditions (20 °C, 760 mm Hg or 68 °F, 29.92 in. Hg) by using Equation 17-1.

$$V_{m(std)} = V_m Y \frac{T_{std} \left(P_{bar} + \frac{\Delta H}{13.6}\right)}{T_m P_{std}}$$

$$= K_1 V_m Y \frac{P_{bar} + \left(\frac{\Delta H}{13.6}\right)}{T_m} \qquad \text{Eq. 17-1}$$

where:
K_1 = 0.3858 °K/mm Hg for metric units,
= 17.64 °R/in. Hg for English units.

NOTE: Equation 17-1 can be used as written unless the leakage rate observed during any of the *mandatory leak checks (i.e., the post-test leak check or leak checks conducted prior to component changes)* exceeds L_a. If L_p or L_i exceeds L_a, Equation 17-1 must be modified as follows:

(a) Case I. No component changes made during sampling run. In this case, replace V_m in Equation 17-1 with the expression:

$$[V_m - (L_p - L_a)\,\theta]$$

(b) Case II. One or more component changes made during the sampling run. In this case, replace V_m in Equation 17-1 by the expression:

$$[V_m - (L_1 - L_a)\,\theta_1 - \sum_{i=2}^{n} (L_i - L_a)\,\theta_i - (L_p - L_a)\,\theta_p]$$

and substitute only for those leakage rates (L_i or L_p) which exceed L_a.

6.4 Volume of Water Vapor.

$$V_{w(std)} = V_{lc} \frac{\rho_w R T_{std}}{M_w P_{std}}$$

$$= K_2 V_{lc} \qquad \text{Eq. 17-2}$$

where:
K_2 = 0.001333 m^3/ml for metric units,
= 0.04707 ft^3/ml for English units.

6.5 Moisture Content.

$$B_{ws} = \frac{V_{w(std)}}{V_{m(std)} + V_{w(std)}} \qquad \text{Eq. 17-3}$$

6.6 Acetone Blank Concentration.

$$C_a = \frac{m_a}{V_a \rho_a} \qquad \text{Eq. 17-4}$$

6.7 Acetone Wash Blank.

$$W_a = C_a V_{aw} \rho_a \qquad \text{Eq. 17-5}$$

6.8 Total Particulate Weight. Determine the total particulate matter catch from the sum of the weights obtained from Containers 1 and 2 less the acetone blank (see Figure 17-4).

NOTE: Refer to Section 4.1.5 to assist in calculation of results involving two or more filter *assemblies or two or more sampling trains.*

6.9 Particulate Concentration.

$$c_s = (0.001 \text{ g/mg})(m_n/V_{m(std)}) \qquad \text{Eq. 17-6}$$

6.10 Conversion Factors:

From	To	Multiply by
scf	m^3	0.02832
g/ft^3	gr/ft^3	15.43
g/ft^3	lb/ft^3	2.205×10^{-3}
g/ft^3	g/m^3	35.31

6.11 Isokinetic Variation.
6.11.1 Calculation from Raw Data.

$$I = \frac{100\, T_s \left[K_3 V_{lc} + \frac{(V_m Y)}{T_m} \left(P_{bar} + \frac{\Delta H}{13.6}\right) \right]}{60\, \theta\, v_s P_s A_n} \qquad \text{Eq. 17-7}$$

where:
K_3 = 0.003454 $[(mm\ Hg)(m^3)]/[(ml)(°K)]$ for metric units,
= 0.002669 $[(in.\ Hg)(ft^3)]/[(ml)(°R)]$ for English units.

6.11.2 Calculation from Intermediate Values.

$$I = \frac{T_s V_{m(std)} P_{std} 100}{T_{std} v_s \theta A_n P_s 60 (1-B_{ws})}$$

$$= K_4 \frac{T_s V_{m(std)}}{P_s v_s A_n \theta (1 - B_{ws})} \qquad \text{Eq. 17-8}$$

where:
K_4 = 4.320 for metric units,
= 0.09450 for English units.

6.11.3 Acceptable Results. If 90 percent ≤ I ≤ 110 percent, the results are acceptable. If the PM results are low in comparison to the standard, and "I" is over 110 percent or less than 90 percent, the Administrator may opt to accept the results. Use citation 4 in Bibliography to make acceptability judgments. Otherwise, reject the results and repeat the test.

7. Bibliography

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APPENDIX III

ENVIRONMENTAL MONITORING SUMMARY TABLES

TABLE 6-J
SUMMARY OF
ENVIRONMENTAL MONITORING
COAL DATA

PARAMETERS	Baseline		Parametric		Optimization		Operations	Post-LIFAC
	9-2-92	9-22-93	9-23-93	9-24-93	12-7-93	12-8-93	(5-17-94)	(8-16-94)
Btu/lb. (Rec'd)	11355	11307	11250	11253	11390	11417	11487	11513
Btu/lb. (Dry)	13084	12893	12967	12983	13016	13094	13274	13247
Moisture % (Rec'd)	13.22	12.30	13.24	13.33	12.49	12.81	13.46	13.09
Ash (Rec'd)	9.00	10.53	9.98	9.86	9.79	9.53	8.27	8.62
Ash (Dry)	10.37	12.00	11.50	11.37	11.19	10.93	9.56	9.91
Sulfur % (Rec'd)	2.46	2.25	2.20	2.18	2.24	2.14	2.12	2.34
Sulfur % (Dry)	2.83	2.57	2.54	2.51	2.56	2.45	2.45	2.69

TABLE 6 – K
SUMMARY OF
ENVIRONMENTAL MONITORING
SOLID WASTE AND BY PRODUCT
ECONOMIZER ASH

SAMPLE LOCATION : ECONOMIZER HOPPER (3)

PARAMETERS	REGULATORY LIMIT (mg/l)* (if established)	UNITS	BASELINE (9-2-92)	BASELINE (9-22-93)	PARAMETRIC (9-23-93)	PARAMETRIC (9-24-93)	OPTIMIZATION (12-7-93)	OPTIMIZATION (12-8-93)	OPERATIONS (5-17-94)	POST-LIFAC (8-16-94)
Alkalinity	–	mg/l CaCO3	87.0	417	2540	2760	2090	2130	2430	858
pH	–	pH units	11.90	12.48	12.55	12.44	12.55	12.52	12.70	12.78
Sulfate	–	mg/l	100	170	1200	1200	1600	1700	1700	310
Percent Carbon	–	%	–	5.6	4.5	4.4	6.5	5.3	4.4	4.7
Total Metals:										
Arsenic (Total)	–	mg/kg	–	–	–	26	47	35	23	34
TCLP (5) Metals :										
Silver	5.0	mg/l	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10
Arsenic	5.0	mg/l	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10
Barium	100.0	mg/l	<10	<10	<10	<10	<10	<10	<10	<10
Cadmium	1.0	mg/l	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10
Chromium	5.0	mg/l	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10
Mercury	0.2	mg/l	<0.010	<0.010	<0.010	<0.010	<0.010	<0.010	<0.010	<0.010
Lead	5.0	mg/l	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10
Selenium	1.0	mg/l	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10
TCLP Extraction Fluid Data:										
Extraction Fluid	–		No.1	No.1	No.2	No.2	No.2	No.2	No.2	No.1
pH w/ Deionized Water	–	pH Units	11.05	10.70	11.75	11.75	12.69	12.71	12.06	11.04
pH after Addition of 1 Normal HCL	–	pH Units	1.55	3.21	11.17	11.40	11.95	11.78	11.84	2.54
pH of TCLP Extract	–	pH Units	5.15	5.33	12.31	12.37	12.30	12.59	12.16	6.03
Amount of Sample Extracted	–	g	100	100	100	100	100	100	100	80
TCLP Volatile Organic Analysis : (8260)										
Benzene	0.5	ug/l	<50	<50	<50	<50	<50	<50	<50	<50
2-Butanone	200.0	ug/l	<5000	<5000	<5000	<5000	<5000	<5000	<5000	<5000
Carbon Tetrachloride	0.5	ug/l	<50	<50	<50	<50	<50	<50	<50	<50
Chlorobenzene	100.0	ug/l	<1000	<1000	<1000	<1000	<1000	<1000	<1000	<1000
Chloroform	6.0	ug/l	<500	<500	<500	<500	<500	<500	<500	<500
1,2-Dichloroethane	0.5	ug/l	<50	<50	<50	<50	<50	<50	<50	<50
1,1-Dichloroethene	0.7	ug/l	<50	<50	<50	<50	<50	<50	<50	<50
Tetrachloroethene	0.7	ug/l	<50	<50	<50	<50	<50	<50	<50	<50
Trichloroethene	0.5	ug/l	<50	<50	<50	<50	<50	<50	<50	<50
Vinyl Chloride	0.2	ug/l	<50	<50	<50	<50	<50	<50	<50	<25
TCLP Base/Neutral Extractables : (8270)										
1,4-Dichlorobenzene	7.5	ug/l	<500	<500	<500	<500	<500	<500	<500	<500
2,4-Dinitrotoluene	0.13	ug/l	<50	<50	<50	<50	<50	<50	<50	<50
Hexachlorobutadiene	0.5	ug/l	<50	<50	<50	<50	<50	<50	<50	<50
Hexachlorobenzene	0.13	ug/l	<100	<100	<100	<100	<100	<100	<100	<100
Hexachloroethane	3.0	ug/l	<500	<500	<500	<500	<500	<500	<500	<500
Nitrobenzene	2.0	ug/l	<100	<100	<100	<100	<100	<100	<100	<100
Pyridine	5.0	ug/l	<500	<500	<500	<500	<500	<500	<500	<500

TABLE 6– K (continued)
SUMMARY OF
ENVIRONMENTAL MONITORING
SOLID WASTE AND BY PRODUCT
ECONOMIZER ASH

SAMPLE LOCATION : ECONOMIZER HOPPER (3)

PARAMETERS	REGULATORY LIMIT (mg/l)* (if established)	UNITS	BASELINE (9–2–92)	BASELINE (9–22–93)	PARAMETRIC (9–23–93)	PARAMETRIC (9–24–93)	OPTIMIZATION (12–7–93)	OPTIMIZATION (12–8–93)	OPERATIONS (5–17–94)	POST–LIFAC (8–16–94)
0										
Total Cresol	200.0	ug/l	<5000	<5000	<5000	<5000	<5000	<5000	<5000	<5000
Pentachlorophenol	100.0	ug/l	<5000	<5000	<5000	<5000	<5000	<5000	<5000	<5000
2,4,5–Trichlorophenol	400.0	ug/l	<5000	<5000	<5000	<5000	<5000	<5000	<5000	<5000
2,4,6–Trichlorophenol	2.0	ug/l	<100	<100	<100	<100	<100	<100	<100	<100
Acetone	–	ug/kg	<500	<13000	<13000	<13000	<13000	<13000	<500	<500
Benzene	0.5	ug/kg	<25	<630	<630	<630	<630	<630	<25	<25
Bromodichloromethane	–	ug/kg	<25	<630	<630	<630	<630	<630	<25	<25
Bromoform	–	ug/kg	<25	<630	<630	<630	<630	<630	<25	<25
Bromomethane	–	ug/kg	<50	<1300	<1300	<1300	<1300	<1300	<50	<50
2–Butanone	200.0	ug/kg	<500	<13000	<13000	<13000	<13000	<13000	<50	<50
Carbon Disulfide	–	ug/kg	<25	<630	<630	<630	<630	<630	<25	<25
Carbon Tetrachloride	0.5	ug/kg	<25	<630	<630	<630	<630	<630	<25	<25
Chlorobenzene	100.0	ug/kg	<25	<630	<630	<630	<630	<630	<25	<25
Chlorodibromomethane	–	ug/kg	<25	<630	<630	<630	<630	<630	<25	<25
Chloroethane	–	ug/kg	<50	<1300	<1300	<1300	<1300	<1300	<50	<50
Chloromethane	–	ug/kg	<50	<1300	<1300	<1300	<1300	<1300	<50	<50
Chloroform	6.0	ug/kg	<25	<630	<630	<630	<630	<630	<25	<25
1,1–Dichloroethane	–	ug/kg	<25	<630	<630	<630	<630	<630	<25	<25
1,2–Dichloroethane	0.5	ug/kg	<25	<630	<630	<630	<630	<630	<25	<25
1,1–Dichloroethene	0.7	ug/kg	<25	<630	<630	<630	<630	<630	<25	<25
cis–1,2–Dichloroethene	–	ug/kg	<25	<630	<630	<630	<630	<630	<25	<25
trans–1,2–Dichloroethene	–	ug/kg	<25	<630	<630	<630	<630	<630	<25	<25
1,2–Dichloropropane	–	ug/kg	<25	<630	<630	<630	<630	<630	<25	<25
cis–1,3–Dichloropropene	–	ug/kg	<25	<630	<630	<630	<630	<630	<25	<25
trans–1,3–Dichloropropene	–	ug/kg	<25	<630	<630	<630	<630	<630	<25	<25
Ethylbenzene	–	ug/kg	<25	<630	<630	<630	<630	<630	<25	<25
2–Hexanone	–	ug/kg	<250	<6300	<6300	<6300	<6300	<6300	<250	<250
Methylene Chloride	–	ug/kg	46	<630	<630	<630	<630	<630	<25	<25
4–Methyl–2–Pentanone (MIBK)	–	ug/kg	<250	<6300	<6300	<6300	<6300	<6300	<250	<250
Styrene	–	ug/kg	<25	<630	<630	<630	<630	<630	<25	<25
1,1,2,2–Tetrachloroethane	–	ug/kg	<25	<630	<630	<630	<630	<630	<25	<25
Tetrachloroethene	0.7	ug/kg	<25	<630	<630	<630	<630	<630	<25	<25
Toluene	–	ug/kg	<25	<630	<630	<630	<630	<630	<25	<25
1,1,1–Trichloroethane	–	ug/kg	<25	<630	<630	<630	<630	<630	<25	<25
1,1,2–Trichloroethane	–	ug/kg	<25	<630	<630	<630	<630	<630	<25	<25
Trichloroethene	0.5	ug/kg	<25	<630	<630	<630	<630	<630	<25	<25
Vinyl Chloride	0.2	ug/kg	<50	<1300	<1300	<1300	<1300	<1300	<50	<50
Xylenes (Total)	–	ug/kg	<25	<630	<630	<630	<630	<630	<25	<25

* Regulatory limit units may differ from measured value units.

TABLE 6-L
SUMMARY OF
ENVIRONMENTAL MONITORING
SOLID WASTE AND BY PRODUCT
LIFAC ASH

SAMPLE LOCATION : REACTOR BOTTOM ASH (4)

PARAMETERS	REGULATORY LIMIT (mg/l)* (if established)	UNITS	PARAMETRIC (9-23-93)	PARAMETRIC (9-24-93)	OPTIMIZATION (12-7-93)	OPTIMIZATION (12-8-93)	OPERATIONS (5-17-94)	POST-LIFAC (8-16-94)
Alkalinity		mg/l CaCO3	2220	2440	1050	1850	2220	---
pH		pH units	12.33	12.48	12.42	12.41	12.72	---
Sulfate		mg/l	850	950	1500	990	1600	---
Percent Carbon		%	3.9	2.9	3.3	3.5	3.6	---
Total Metals:								
Arsenic (total)	5.0	mg/kg	-	25	32	23	29	---
TCLP Metals :								
Silver	5.0	mg/l	<0.10	<0.10	<0.10	<0.10	<0.10	---
Arsenic	5.0	mg/l	<0.10	<0.10	<0.10	<0.10	<0.10	---
Barium	100.0	mg/l	<10	<10	<10	<10	<10	---
Cadmium	1.0	mg/l	<0.10	<0.10	<0.10	<0.10	<0.10	---
Chromium	5.0	mg/l	<0.10	<0.10	<0.10	<0.10	<0.10	---
Mercury	0.2	mg/l	<0.010	<0.010	<0.010	<0.010	<0.010	---
Lead	5.0	mg/l	<0.10	<0.10	<0.10	<0.10	<0.10	---
Selenium	1.0	mg/l	<0.10	<0.10	<0.10	<0.10	<0.10	---
TCLP Extraction Fluid Data:								
Extraction Fluid			No.2	No.2	No.2	No.2	No.2	---
pH w/ Deionized Water		pH Units	11.65	11.60	12.46	12.52	11.99	---
pH after Addition of 1 Normal HCL		pH Units	11.09	10.65	11.23	11.88	11.77	---
pH of TCLP Extract		pH Units	12.02	12.11	12.02	6.74	11.94	---
Amount of Sample Extracted		g	100	100	100	100	100	---
TCLP Volatile Organic Analysis : (8260)								
Benzene	0.5	ug/l	<50	<50	<50	<50	<50	---
2-Butanone	200.0	ug/l	<5000	<5000	<5000	<5000	<5000	---
Carbon Tetrachloride	0.5	ug/l	<50	<50	<50	<50	<50	---
Chlorobenzene	100.0	ug/l	<1000	<1000	<1000	<1000	<1000	---
Chloroform	6.0	ug/l	<500	<500	<500	<500	<500	---
1,2-Dichloroethane	0.5	ug/l	<50	<50	<50	<50	<50	---
1,1-Dichloroethene	0.7	ug/l	<50	<50	<50	<50	<50	---
Tetrachloroethene	0.7	ug/l	<50	<50	<50	<50	<50	---
Trichloroethene	0.5	ug/l	<50	<50	<50	<50	<50	---
Vinyl Chloride	0.2	ug/l	<50	<50	<50	<50	<50	---
TCLP Base/Neutral Extractables : (8270)								
1,4-Dichlorobenzene	7.5	ug/l	<500	<500	<500	<500	<500	---
2,4-Dinitrotoluene	0.13	ug/l	<50	<50	<50	<50	<50	---
Hexachlorobutadiene	0.5	ug/l	<50	<50	<50	<50	<50	---
Hexachlorobenzene	0.13	ug/l	<100	<100	<100	<100	<100	---
Hexachloroethane	3.0	ug/l	<500	<500	<500	<500	<500	---
Nitrobenzene	2.0	ug/l	<100	<100	<100	<100	<100	---
Pyridine	5.0	ug/l	<500	<500	<500	<500	<500	---

TABLE 6– L (continued)
SUMMARY OF
ENVIRONMENTAL MONITORING
SOLID WASTE AND BY PRODUCT
LIFAC ASH

SAMPLE LOCATION : REACTOR BOTTOM ASH (4)

PARAMETERS	REGULATORY LIMIT (mg/l)* (if established)	UNITS	PARAMETRIC (9–23–93)	PARAMETRIC (9–24–93)	OPTIMIZATION (12–7–93)	OPTIMIZATION (12–8–93)	OPERATIONS (5–17–94)	POST–LIFAC (8–16–94)
TCLP Acid Extractables : (8270)								
Total Cresol	200.0	ug/l	<5000	<5000	<5000	<5000	<5000	---
Pentachlorophenol	100.0	ug/l	<5000	<5000	<5000	<5000	<5000	---
2,4,5–Trichlorophenol	400.0	ug/l	<5000	<5000	<5000	<5000	<5000	---
2,4,6–Trichlorophenol	2.0	ug/l	<100	<100	<100	<100	<100	---
Acetone		ug/kg	<13000	<13000	<13000	<13000	<500	---
Benzene	0.5	ug/kg	<630	<630	<630	<630	<25	---
Bromodichloromethane		ug/kg	<630	<630	<630	<630	<25	---
Bromoform		ug/kg	<630	<630	<630	<630	<25	---
Bromomethane		ug/kg	<1300	<1300	<1300	<1300	<50	---
2–Butanone	200.0	ug/kg	<13000	<13000	<13000	<13000	<50	---
Carbon Disulfide		ug/kg	<630	<630	<630	<630	<25	---
Carbon Tetrachloride	0.5	ug/kg	<630	<630	<630	<630	<25	---
Chlorobenzene	100.0	ug/kg	<630	<630	<630	<630	<25	---
Chlorodibromomethane		ug/kg	<630	<630	<630	<630	<25	---
Chloroethane		ug/kg	<1300	<1300	<1300	<1300	<50	---
Chloromethane		ug/kg	<1300	<1300	<1300	<1300	<50	---
Chloroform	6.0	ug/kg	<630	<630	<630	<630	<25	---
1,1–Dichloroethane		ug/kg	<630	<630	<630	<630	<25	---
1,2–Dichloroethane	0.5	ug/kg	<630	<630	<630	<630	<25	---
1,1–Dichloroethene	0.7	ug/kg	<630	<630	<630	<630	<25	---
cis–1,2–Dichloroethene		ug/kg	<630	<630	<630	<630	<25	---
trans–1,2–Dichloroethene		ug/kg	<630	<630	<630	<630	<25	---
1,2–Dichloropropane		ug/kg	<630	<630	<630	<630	<25	---
cis–1,3–Dichloropropene		ug/kg	<630	<630	<630	<630	<25	---
trans–1,3–Dichloropropene		ug/kg	<630	<630	<630	<630	<25	---
Ethylbenzene		ug/kg	<630	<630	<630	<630	<25	---
2–Hexanone		ug/kg	<6300	<6300	<6300	<6300	<250	---
Methylene Chloride		ug/kg	<630	<630	<630	<630	<25	---
4–Methyl–2–Pentanone (MIBK)		ug/kg	<6300	<6300	<6300	<6300	<250	---
Styrene		ug/kg	<630	<630	<630	<630	<25	---
1,1,2,2–Tetrachloroethane		ug/kg	<630	<630	<630	<630	<25	---
Tetrachloroethene	0.7	ug/kg	<630	<630	<630	<630	<25	---
Toluene		ug/kg	<630	<630	<630	<630	<25	---
1,1,1–Trichloroethane		ug/kg	<630	<630	<630	<630	<25	---
1,1,2–Trichloroethane		ug/kg	<630	<630	<630	<630	<25	---
Trichloroethene	0.5	ug/kg	<630	<630	<630	<630	<25	---
Vinyl Chloride	0.2	ug/kg	<1300	<1300	<1300	<1300	<50	---
Xylenes (Total)		ug/kg	<630	<630	<630	<630	<25	---

* Regulatory limit units may differ from measured value units.

TABLE 6 – M
SUMMARY OF
ENVIRONMENTAL MONITORING
SOLID WASTE AND BY PRODUCT
ESP INLET HOPPER

SAMPLE LOCATION : ESP INLET HOPPERS #7 & #8 (5A)

PARAMETERS	REGULATORY LIMIT (mg/l)* (if established)	UNITS	BASELINE (9-2-92)	BASELINE (9-22-93)	PARAMETRIC (9-23-93)	PARAMETRIC (9-24-93) (5A-1)	PARAMETRIC (9-24-93) (5A-2)	OPTIMIZATION (12-7-93)	OPTIMIZATION (12-8-93)	OPERATIONS (5-17-94)	POST-LIFAC (6-18-94)
Alkalinity	–	mg/l CaCO3	299	2070	2920	3060	2780	2250	2400	2670	370
pH	–	pH units	11.60	12.65	12.51	12.44	12.50	12.42	12.60	12.69	11.83
Sulfate	–	mg/l	260	810.0	1400	1200	1100	1700	1600	1800	320
Percent Carbon	–	%	–	3.6	3.3	2.6	2.3	3.3	3.1	2.9	4.5
Total Metals:											
Arsenic (Total)	5.0	mg/kg	–	340	–	69	36	60	50	41	9.8
TCLP Metals :											
Silver	5.0	mg/l	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10
Arsenic	5.0	mg/l	1.3	0.14	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10
Barium	100.0	mg/l	<10	<10	<10	<10	<10	<10	<10	<10	<10
Cadmium	1.0	mg/l	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	0.13
Chromium	5.0	mg/l	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10
Mercury	0.2	mg/l	<0.010	<0.010	<0.010	<0.010	<0.010	<0.010	<0.010	<0.010	<0.010
Lead	5.0	mg/l	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10
Selenium	1.0	mg/l	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10
TCLP Extraction Fluid Data:											
Extraction Fluid	–		No.1	No.2	No.2	No.2	No.2	No.2	No.2	No.2	No.1
pH w/ Deionized Water	–	pH Units	11.50	11.18	11.74	11.76	11.76	12.61	12.71	12.21	10.88
pH after Addition of 1 Normal HCL	–	pH Units	1.90	10.39	11.60	11.37	11.59	11.48	12.20	11.95	2.07
pH of TCLP Extract	–	pH Units	5.20	6.31	12.44	12.55	12.45	12.36	12.66	12.15	5.33
Amount of Sample Extracted	–	g	100	100	100	100	100	100	100	100	80
TCLP Volatile Organic Analysis : (8260)											
Benzene	0.5	ug/l	<50	<50	<50	<50	<50	<50	<50	<50	<50
2-Butanone	200.0	ug/l	<5000	<5000	<5000	<5000	<5000	<5000	<5000	<5000	<5000
Carbon Tetrachloride	0.5	ug/l	<50	<50	<50	<50	<50	<50	<50	<50	<50
Chlorobenzene	100.0	ug/l	<1000	<1000	<1000	<1000	<1000	<1000	<1000	<1000	<1000
Chloroform	6.0	ug/l	<500	<500	<500	<500	<500	<500	<500	<500	<500
1,2-Dichloroethane	0.5	ug/l	<50	<50	<50	<50	<50	<50	<50	<50	<50
1,1-Dichloroethene	0.7	ug/l	<50	<50	<50	<50	<50	<50	<50	<50	<50
Tetrachloroethene	0.7	ug/l	<50	<50	<50	<50	<50	<50	<50	<50	<50
Trichloroethene	0.5	ug/l	<50	<50	<50	<50	<50	<50	<50	<50	<50
Vinyl Chloride	0.2	ug/l	<50	<50	<50	<50	<50	<50	<50	<50	<25
TCLP Base/Neutral Extractables : (8270)											
1,4-Dichlorobenzene	7.5	ug/l	<500	<500	<500	<500	<500	<500	<500	<500	<500
2,4-Dinitrotoluene	0.13	ug/l	<50	<50	<50	<50	<50	<50	<50	<50	<50
Hexachlorobutadiene	0.5	ug/l	<50	<50	<50	<50	<50	<50	<50	<50	<50
Hexachlorobenzene	0.13	ug/l	<100	<100	<100	<100	<100	<100	<100	<100	<100
Hexachloroethane	3.0	ug/l	<500	<500	<500	<500	<500	<500	<500	<500	<500
Nitrobenzene	2.0	ug/l	<100	<100	<100	<100	<100	<100	<100	<100	<100
Pyridine	5.0	ug/l	<500	<500	<500	<500	<500	<500	<500	<500	<500

TABLE 6-M (Continued)
SUMMARY OF
ENVIRONMENTAL MONITORING
SOLID WASTE AND BY PRODUCT
ESP INLET HOPPER

SAMPLE LOCATION : ESP INLET HOPPERS #7 & #8 (5A)

PARAMETERS	REGULATORY LIMIT (mg/l)* (if established)	UNITS	BASELINE (9-2-92)	BASELINE (9-22-93)	PARAMETRIC (9-23-93)	PARAMETRIC (9-24-93)	PARAMETRIC (9-24-93)	OPTIMIZATION (12-7-93)	OPTIMIZATION (12-8-93)	OPERATIONS (5-17-94)	POST-LIFAC (6-18-94)
TCLP Acid Extractables : (8270)											
Total Cresol	200.0	ug/l	<5000	<5000	<5000	<5000	<5000	<5000	<5000	<5000	<5000
Pentachlorophenol	100.0	ug/l	<5000	<5000	<5000	<5000	<5000	<5000	<5000	<5000	<5000
2,4,5-Trichlorophenol	400.0	ug/l	<5000	<5000	<5000	<5000	<5000	<5000	<5000	<5000	<5000
2,4,6-Trichlorophenol	2.0	ug/l	<100	<100	<100	<100	<100	<100	<100	<100	<100
Acetone	–	ug/kg	<13000	<13000	<13000	<13000	<13000	<13000	<13000	<500	<500
Benzene	0.5	ug/kg	<630	<630	<630	<630	<630	<630	<630	<25	<25
Bromodichloromethane	–	ug/kg	<630	<630	<630	<630	<630	<630	<630	<25	<25
Bromoform	–	ug/kg	<630	<630	<630	<630	<630	<630	<630	<25	<25
Bromomethane	–	ug/kg	<1300	<1300	<1300	<1300	<1300	<1300	<1300	<50	<50
2-Butanone	200.0	ug/kg	<13000	<13000	<13000	<13000	<13000	<13000	<13000	<50	<50
Carbon Disulfide	–	ug/kg	730	<630	<630	<630	<630	<630	<630	<25	<25
Carbon Tetrachloride	0.5	ug/kg	<630	<630	<630	<630	<630	<630	<630	<25	<25
Chlorobenzene	100.0	ug/kg	<630	<630	<630	<630	<630	<630	<630	<25	<25
Chlorodibromomethane	–	ug/kg	<630	<630	<630	<630	<630	<630	<630	<25	<25
Chloroethane	–	ug/kg	<1300	<1300	<1300	<1300	<1300	<1300	<1300	<50	<50
Chloromethane	–	ug/kg	<1300	<1300	<1300	<1300	<1300	<1300	<1300	<50	<50
Chloroform	6.0	ug/kg	<630	<630	<630	<630	<630	<630	<630	<25	<25
1,1-Dichloroethane	–	ug/kg	<630	<630	<630	<630	<630	<630	<630	<25	<25
1,2-Dichloroethane	0.5	ug/kg	<630	<630	<630	<630	<630	<630	<630	<25	<25
1,1-Dichloroethene	0.7	ug/kg	<630	<630	<630	<630	<630	<630	<630	<25	<25
cis-1,2-Dichloroethene	–	ug/kg	<630	<630	<630	<630	<630	<630	<630	<25	<25
trans-1,2-Dichloroethene	–	ug/kg	<630	<630	<630	<630	<630	<630	<630	<25	<25
1,2-Dichloropropane	–	ug/kg	<630	<630	<630	<630	<630	<630	<630	<25	<25
cis-1,3-Dichloropropene	–	ug/kg	<630	<630	<630	<630	<630	<630	<630	<25	<25
trans-1,3-Dichloropropene	–	ug/kg	<630	<630	<630	<630	<630	<630	<630	<25	<25
Ethylbenzene	–	ug/kg	<630	<630	<630	<630	<630	<630	<630	<25	<25
2-Hexanone	–	ug/kg	<6300	<6300	<6300	<6300	<6300	<6300	<6300	<250	<250
Methylene Chloride	–	ug/kg	<630	<630	<630	<630	<630	<630	<630	<25	<25
4-Methyl-2-Pentanone (MIBK)	–	ug/kg	<6300	<6300	<6300	<6300	<6300	<6300	<6300	<250	<250
Styrene	–	ug/kg	<630	<630	<630	<630	<630	<630	<630	<25	<25
1,1,2,2-Tetrachloroethane	–	ug/kg	<630	<630	<630	<630	<630	<630	<630	<25	<25
Tetrachloroethene	0.7	ug/kg	<630	<630	<630	<630	<630	<630	<630	<25	<25
Toluene	–	ug/kg	2200	<630	<630	<630	<630	<630	<630	<25	<25
1,1,1-Trichloroethane	–	ug/kg	<630	<630	<630	<630	<630	<630	<630	<25	<25
1,1,2-Trichloroethane	–	ug/kg	<630	<630	<630	<630	<630	<630	<630	<25	<25
Trichloroethene	0.5	ug/kg	<630	<630	<630	<630	<630	<630	<630	<25	<25
Vinyl Chloride	0.2	ug/kg	<1300	<1300	<1300	<1300	<1300	<1300	<1300	<50	<50
Xylenes (Total)	–	ug/kg	<630	<630	<630	<630	<630	<630	<630	<25	<25

* Regulatory limit units may differ from measured value units.

TABLE 6 – N
SUMMARY OF
ENVIRONMENTAL MONITORING
SOLID WASTE AND BY PRODUCT
ESP OUTLET HOPPER ASH

SAMPLE LOCATION : ESP OUTLET HOPPERS #5 & #6 (5B)

PARAMETERS	REGULATORY LIMIT (mg/l)* (if established)	UNITS	PARAMETRIC (9–23–93)	PARAMETRIC (9–24–93)	OPTIMIZATION (12–7–93)	OPTIMIZATION (12–8–93)	OPERATIONS (5–17–94)	POST–LIFAC (8–16–94)
Alkalinity	–	mg/l CaCO3	2840	2990	2560	2400	2740	42.0
pH	–	pH units	12.47	12.46	12.52	12.52	12.70	8.21
Sulfate	–	mg/l	1700	1200	1600	1600	1700	720
Percent Carbon	–	%	NA**	4.1	2.7	2.8	2.9	6.0
Total Metals:								
Arsenic (total)	5.0	mg/kg	–	95	67	68	52	21
TCLP Metals :								
Silver	5.0	mg/l	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10
Arsenic	5.0	mg/l	<0.10	<0.10	<0.10	<0.10	<0.10	0.29
Barium	100.0	mg/l	<10	<10	<10	<10	<10	<10
Cadmium	1.0	mg/l	<0.10	<0.10	<0.10	<0.10	<0.10	0.21
Chromium	5.0	mg/l	1.1	<0.10	<0.10	<0.10	<0.10	0 27
Mercury	0.2	mg/l	<0.010	<0.010	<0.010	<0.010	<0.010	<0.010
Lead	5.0	mg/l	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10
Selenium	1.0	mg/l	<0.10	<0.10	<0.10	<0.10	<0.10	<0 10
TCLP Extraction Fluid Data:								
Extraction Fluid			No.2	No.2	No.2	No.2	No.2	No.1
pH w/ Deionized Water	–	pH Units	12.17	11.69	12.69	12.69	12.11	8 91
pH after Addition of 1 Normal HCL	–	pH Units	12.05	11.51	12.03	12.14	11.90	1.77
pH of TCLP Extract	–	pH Units	12.02	12.51	12.13	12.66	12.13	4 99
Amount of Sample Extracted	–	g	50.0	100	100	100	100	80
TCLP Volatile Organic Analysis : (8260) (3)								
Benzene	0.5	ug/l	<50	<50	<50	<50	<50	<50
2–Butanone	200.0	ug/l	<5000	<5000	<5000	<5000	<5000	<5000
Carbon Tetrachloride	0.5	ug/l	<50	<50	<50	<50	<50	<50
Chlorobenzene	100.0	ug/l	<1000	<1000	<1000	<1000	<1000	<1000
Chloroform	6.0	ug/l	<500	<500	<500	<500	<500	<500
1,2–Dichloroethane	0.5	ug/l	<50	<50	<50	<50	<50	<50
1,1–Dichloroethene	0.7	ug/l	<50	<50	<50	<50	<50	<50
Tetrachloroethene	0.7	ug/l	<50	<50	<50	<50	<50	<50
Trichloroethene	0.5	ug/l	<50	<50	<50	<50	<50	<50
Vinyl Chloride	0.2	ug/l	<50	<50	<50	<50	<50	<5.0
TCLP Base/Neutral Extractables : (8270) (3)								
1,4–Dichlorobenzene	7.5	ug/l	<500	<500	<500	<500	<500	<500
2,4–Dinitrotoluene	0.13	ug/l	<50	<50	<50	<50	<50	<50
Hexachlorobutadiene	0.5	ug/l	<50	<50	<50	<50	<50	<50
Hexachlorobenzene	0.13	ug/l	<100	<100	<100	<100	<100	<100
Hexachloroethane	3.0	ug/l	<500	<500	<500	<500	<500	<500
Nitrobenzene	2.0	ug/l	<100	<100	<100	<100	<100	<100
Pyridine	5.0	ug/l	<500	<500	<500	<500	<500	<500

TABLE 6 – N (Continued)
SUMMARY OF
ENVIRONMENTAL MONITORING
SOLID WASTE AND BY PRODUCT
ESP OUTLET HOPPER ASH

SAMPLE LOCATION : ESP OUTLET HOPPERS #5 & #6 (5B)

PARAMETERS	REGULATORY LIMIT (mg/l)* (if established)	UNITS	PARAMETRIC (9-23-93)	PARAMETRIC (9-24-93)	OPTIMIZATION (12-7-93)	OPTIMIZATION (12-8-93)	OPERATIONS (5-17-94)	POST-LIFAC (8-16-94)
TCLP Acid Extractables : (8270) (3)								
Total Cresol	200.0	ug/l	<5000	<5000	<5000	<5000	<5000	<5000
Pentachlorophenol	100.0	ug/l	<5000	<5000	<5000	<5000	<5000	<5000
2,4,5-Trichlorophenol	400.0	ug/l	<5000	<5000	<5000	<5000	<5000	<5000
2,4,6-Trichlorophenol	2.0	ug/l	<100	<100	<100	<100	<100	<100
Acetone	–	ug/kg	<500	<13000	<13000	<13000	<100	<100
Benzene	0.5	ug/kg	<25	<630	<630	<630	<54	<5.0
Bromodichloromethane	–	ug/kg	<25	<630	<630	<630	<5.0	<5.0
Bromoform	–	ug/kg	<25	<630	<630	<630	<5.0	<5.0
Bromomethane	–	ug/kg	<50	<1300	<1300	<1300	<10	<10
2-Butanone	200.0	ug/kg	<500	<13000	<13000	<13000	<10	<10
Carbon Disulfide	–	ug/kg	<25	<630	<630	<630	<5.0	<5.0
Carbon Tetrachloride	0.5	ug/kg	<25	<630	<630	<630	<5.0	<5.0
Chlorobenzene	100.0	ug/kg	<25	<630	<630	<630	<5.0	<5.0
Chlorodibromomethane	–	ug/kg	<25	<630	<630	<630	<5.0	<5.0
Chloroethane	–	ug/kg	<50	<1300	<1300	<1300	<10	<10
Chloromethane	–	ug/kg	<50	<1300	<1300	<1300	<10	<10
Chloroform	6.0	ug/kg	<25	<630	<630	<630	<5.0	<5.0
1,1-Dichloroethane	–	ug/kg	<25	<630	<630	<630	<5.0	<5.0
1,2-Dichloroethane	0.5	ug/kg	<25	<630	<630	<630	<5.0	<5.0
1,1-Dichloroethene	0.7	ug/kg	<25	<630	<630	<630	<5.0	<5.0
cis-1,2-Dichloroethene	–	ug/kg	<25	<630	<630	<630	<5.0	<5.0
trans-1,2-Dichloroethene	–	ug/kg	<25	<630	<630	<630	<5.0	<5.0
1,2-Dichloropropane	–	ug/kg	<25	<630	<630	<630	<5.0	<5.0
cis-1,3-Dichloropropene	–	ug/kg	<25	<630	<630	<630	<5.0	<5.0
trans-1,3-Dichloropropene	–	ug/kg	<25	<630	<630	<630	<5.0	<5.0
Ethylbenzene	–	ug/kg	<25	<630	<630	<630	<5.0	<5.0
2-Hexanone	–	ug/kg	<250	<6300	<6300	<6300	<50	<50
Methylene Chloride	–	ug/kg	900	<630	<630	<630	<5.0	30
4-Methyl-2-Pentanone (MIBK)	–	ug/kg	<250	<6300	<6300	<6300	<50	<50
Styrene	–	ug/kg	<25	<630	<630	<630	<5.0	<5.0
1,1,2,2-Tetrachloroethane	–	ug/kg	<25	<630	<630	<630	<5.0	<5.0
Tetrachloroethene	0.7	ug/kg	<25	<630	<630	<630	<5.0	<5.0
Toluene	–	ug/kg	40	<630	<630	<630	5.0	<5.0
1,1,1-Trichloroethane	–	ug/kg	<25	<630	<630	<630	<5.0	<5.0
1,1,2-Trichloroethane	–	ug/kg	<25	<630	<630	<630	<5.0	<5.0
Trichloroethene	0.5	ug/kg	<25	<630	<630	<630	<5.0	<5.0
Vinyl Chloride	0.2	ug/kg	<50	<1300	<1300	<1300	<10	<10
Xylenes (Total)	–	ug/kg	<25	<630	<630	<630	<5.0	<5.0

TABLE 6-O
SUMMARY OF
ENVIRONMENTAL MONITORING
SOLID WASTE AND BY PRODUCT
BOILER BOTTOM ASH (SOLID)

SAMPLE LOCATION : BOILER BOTTOM ASH BIN (6B)

PARAMETERS	REGULATORY LIMIT (mg/l)* (if established)	UNITS	BASELINE (9-22-93)	PARAMETRIC (9-23-93) 6B-1	(9-23-93) 6B-2	(9-24-93)	OPTIMIZATION (12-7-93)	(12-8-93)	OPERATIONS (5-17-94)	POST-LIFAC (8-16-94)
Alkalinity	–	mg/l CaCO3	17	20	51	123	87	168	448	7.0
pH	–	pH units	8.57	10.47	10.98	11.14	11 21	11.86	12.70	7.96
Sulfate	–	mg/l	8.6	49	80	240	110	920	570	16
Percent Carbon	–	%	2.5	0.54	0.32	1.5	0.5	1.1	1.9	46.0
Total Metals:										
Arsenic (total)	5.0	mg/kg	10	–	–	9.6	5.4	8.7	8.4	0.56
TCLP (5) Metals :										
Silver	5.0	mg/l	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10
Arsenic	5.0	mg/l	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10
Barium	100.0	mg/l	<10	<10	<10	<10	<10	<10	<10	<10
Cadmium	1.0	mg/l	<0.10	<0.10	<0.10	<0.10	<0 10	<0.10	<0.10	<0.10
Chromium	5.0	mg/l	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10
Mercury	0.2	mg/l	<0.010	<0.010	<0.010	<0.010	<0.010	<0.010	<0.010	<0.010
Lead	5.0	mg/l	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10
Selenium	1.0	mg/l	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10
TCLP Extraction Fluid Data:										
Extraction Fluid			No.1	No.1	No.1	No.1	No.1	No.2	No.1	No.1
pH w/ Deionized Water	–	pH Units	9.70	10.00	9.57	9.48	10.61	11.16	11.26	8 46
pH after Addition of 1 Normal HCL	–	pH Units	2.93	1.62	1.68	2.02	1.35	8.69	2.08	1.38
pH of TCLP Extract	–	pH Units	5.21	4.97	3.76	5.69	5.02	4.78	6.89	4.92
Amount of Sample Extracted	–	g	100	100	100	100	100	100	100	80
TCLP Volatile Organic Analysis : (8260)										
Benzene	0.5	ug/l	<50	<50	<50	<50	<50	<50	<50	<50
2-Butanone	200.0	ug/l	<5000	<5000	<5000	<5000	<5000	<5000	<5000	<5000
Carbon Tetrachloride	0.5	ug/l	<50	<50	<50	<50	<50	<50	<50	<50
Chlorobenzene	100.0	ug/l	<1000	<1000	<1000	<1000	<1000	<1000	<1000	<1000
Chloroform	6.0	ug/l	<500	<500	<500	<500	<500	<500	<500	<500
1,2-Dichloroethane	0.5	ug/l	<50	<50	<50	<50	<50	<50	<50	<50
1,1-Dichloroethene	0.7	ug/l	<50	<50	<50	<50	<50	<50	<50	<50
Tetrachloroethene	0.7	ug/l	<50	<50	<50	<50	<50	<50	<50	<50
Trichloroethene	0.5	ug/l	<50	<50	<50	<50	<50	<50	<50	<50
Vinyl Chloride	0.2	ug/l	<50	<50	<50	<50	<50	<50	<50	<5.0
TCLP Base/Neutral Extractables : (8270)										
1,4-Dichlorobenzene	7.5	ug/l	<500	<500	<500	<500	<500	<500	<500	<500
2,4-Dinitrotoluene	0.13	ug/l	<50	<50	<50	<50	<50	<50	<50	<50
Hexachlorobutadiene	0.5	ug/l	<50	<50	<50	<50	<50	<50	<50	<50
Hexachlorobenzene	0.13	ug/l	<100	<100	<100	<100	<100	<100	<100	<100
Hexachloroethane	3.0	ug/l	<500	<500	<500	<500	<500	<500	<500	<500
Nitrobenzene	2.0	ug/l	<100	<100	<100	<100	<100	<100	<100	<100
Pyridine	5.0	ug/l	<500	<500	<500	<500	<500	<500	<500	<500

TABLE 6-O (Continued)
SUMMARY OF
ENVIRONMENTAL MONITORING
SOLID WASTE AND BY PRODUCT
BOILER BOTTOM ASH (SOLID)

SAMPLE LOCATION : BOILER BOTTOM ASH BIN (6B)

PARAMETERS	REGULATORY LIMIT (mg/l)*	UNITS	BASELINE (9-22-93)	PARAMETRIC (9-23-93) 6B-1	(9-23-93) 6B-2	(9-24-93)	OPTIMIZATION (12-7-93)	(12-8-93)	OPERATIONS (5-17-94)	POST-LIFAC (8-16-94)
TCLP Acid Extractables : (8270)										
Total Cresol	200.0	ug/l	<5000	<5000	<5000	<5000	<5000	<5000	<5000	<5000
Pentachlorophenol	100.0	ug/l	<5000	<5000	<5000	<5000	<5000	<5000	<5000	<5000
2,4,5-Trichlorophenol	400.0	ug/l	<5000	<5000	<5000	<5000	<5000	<5000	<5000	<5000
2,4,6-Trichlorophenol	2.0	ug/l	<100	<100	<100	<100	<100	<100	<100	<100
Acetone	–	ug/kg	<100	<100	<100	<13000	<100	<13000	<500	<100
Benzene	0.5	ug/kg	<5.0	<5.0	<5.0	<630	<5.0	<630	<25	<5.0
Bromodichloromethane	–	ug/kg	<5.0	<5.0	<5.0	<630	<5.0	<630	<25	<5.0
Bromoform	–	ug/kg	<5.0	<5.0	<5.0	<630	<5.0	<630	<25	<5.0
Bromomethane	–	ug/kg	<10	<10	<10	<1300	<10	<1300	<50	<10
2-Butanone	200.0	ug/kg	<100	<100	<100	<13000	<100	<13000	<50	<10
Carbon Disulfide	–	ug/kg	<5.0	<5.0	<5.0	<630	<5.0	<630	<25	<5.0
Carbon Tetrachloride	0.5	ug/kg	<5.0	<5.0	<5.0	<630	<5.0	<630	<25	<5.0
Chlorobenzene	100.0	ug/kg	<5.0	<5.0	<5.0	<630	<5.0	<630	<25	<5.0
Chlorodibromomethane	–	ug/kg	<5.0	<5.0	<5.0	<630	<5.0	<630	<25	<5.0
Chloroethane	–	ug/kg	<10	<10	<10	<1300	<10	<1300	<50	<10
Chloromethane	–	ug/kg	<10	<10	<10	<1300	<10	<1300	<50	<10
Chloroform	6.0	ug/kg	<5.0	<5.0	<5.0	<630	<5.0	<630	<25	<5.0
1,1-Dichloroethane	–	ug/kg	<5.0	<5.0	<5.0	<630	<5.0	<630	<25	<5.0
1,2-Dichloroethane	0.5	ug/kg	<5.0	<5.0	<5.0	<630	<5.0	<630	<25	<5.0
1,1-Dichloroethene	0.7	ug/kg	<5.0	<5.0	<5.0	<630	<5.0	<630	<25	<5.0
cis-1,2-Dichloroethene	–	ug/kg	<5.0	<5.0	<5.0	<630	<5.0	<630	<25	<5.0
trans-1,2-Dichloroethene	–	ug/kg	<5.0	<5.0	<5.0	<630	<5.0	<630	<25	<5.0
1,2-Dichloropropane	–	ug/kg	<5.0	<5.0	<5.0	<630	<5.0	<630	<25	<5.0
cis-1,3-Dichloropropene	–	ug/kg	<5.0	<5.0	<5.0	<630	<5.0	<630	<25	<5.0
trans-1,3-Dichloropropene	–	ug/kg	<5.0	<5.0	<5.0	<630	<5.0	<630	<25	<5.0
Ethylbenzene	–	ug/kg	<5.0	<5.0	<5.0	<630	<5.0	<630	<25	<5.0
2-Hexanone	–	ug/kg	<50	<50	<50	<6300	<50	<6300	<250	<50
Methylene Chloride	–	ug/kg	<5.0	8.0	<5.0	<630	13	<630	<25	12
4-Methyl-2-Pentanone (MIBK)	–	ug/kg	<50	<50	<50	<6300	<50	<6300	<250	<50
Styrene	–	ug/kg	<5.0	<5.0	<5.0	<630	<5.0	<630	<25	<5.0
1,1,2,2-Tetrachloroethane	–	ug/kg	<5.0	<5.0	<5.0	<630	<5.0	<630	<25	<5.0
Tetrachloroethene	0.7	ug/kg	<5.0	<5.0	<5.0	<630	<5.0	<630	<25	<5.0
Toluene	–	ug/kg	<5.0	<5.0	<5.0	<630	<5.0	<630	<25	<5.0
1,1,1-Trichloroethane	–	ug/kg	<5.0	<5.0	<5.0	<630	<5.0	<630	<25	<5.0
1,1,2-Trichloroethane	–	ug/kg	<5.0	<5.0	<5.0	<630	<5.0	<630	<25	<5.0
Trichloroethene	0.5	ug/kg	<5.0	<5.0	<5.0	<630	<5.0	<630	<25	<5.0
Vinyl Chloride	0.2	ug/kg	<10	<10	<10	<1300	<10	<1300	<50	<10
Xylenes (Total)	–	ug/kg	<5.0	<5.0	<5.0	<630	<5.0	<630	<25	<5.0

* Regulatory limit units may differ from measured value units.

TABLE 6– P
SUMMARY OF
ENVIRONMENTAL MONITORING
BOILER BOTTOM ASH (AQUEOUS)

SAMPLE LOCATION : BOILER BOTTOM ASH (AQUEOUS) (6A)

PARAMETERS	LIMITS (RP&L's NPDES Permit)	BASELINE (9–2–92)	PARAMETRIC (9–23–93) 6A–1	PARAMETRIC (9–23–93) 6A–2	PARAMETRIC (9–24–93)	OPTIMIZATION (12–7–93)	OPTIMIZATION (12–8–93)	OPERATIONS (5–17–94)	POST–LIFAC (8–16–94)
FLOW	REPORT	–	–	–	–	–	–	–	–
TSS (mg/l)	70	–	17	1	9	30	38	13	31
OIL & GREASE (mg/l)	15	–	<1.0	<1.0	<1.0	<1.0	<1.0	<1.0	<1.0
TOT. RESID. OXIDANTS (mg/l)	0.02	–	<1.0	<1.0	<.10	<1.0	<1.0	<0.050	<0.050
TEMPERATURE (F)	REPORT	–	–	–	–	–	–	–	–
AMMONIA (NH3) (mg/l)	1.8 (summer) 2.6 (winter)	–	0.11	0.13	0.15	<0.10	0.37	<0.10	<0.10
COPPER (mg/l)	0.05	–	0.014	<0.010	0.011	<0.010	0.012	<0.010	0.012
LEAD (mg/l)	0.02	–	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10
ZINC (mg/l)	0.46	–	0.026	0.023	<0.005	0.011	0.020	0.012	0.046
ALUMINUM (mg/l)	0.40	–	0.59	0.60	0.25	0.46	0.53	0.11	0.34
IRON (mg/l)	REPORT	–	0.20	0.18	0.061	0.16	0.11	<0.030	0.33
PCBS (mg/l)	REPORT	–	<1.0	<1.0	<1.0	<1.0	<1.0	<1.0	<1.0
ALKALINITY (mg CaCO3)	(per EMP)	111	182	200	65	31.1	40	254	186

TABLE 6– Q
SUMMARY OF
ENVIRONMENTAL MONITORING
RP&L's NPDES DISCHARGE AND FEEDWATER
(1993)

		RP&L MONTHLY MONITORING (1993)*						FEEDWATER SAMPLE LOCATION (Z)				
PARAMETERS	LIMITS	JULY	AUG.	SEPT.	OCT.	NOV.	DEC.	(9-22-93)	(9-23-93)	(9-24-93)	(12-7-93)	(12-8-93)
FLOW	REPORT	–	–	–	–	–	–	–	–	–	–	–
TSS (mg/l)	70	17.9	20.4	19.1	13.1	19.5	15.4	2	73	<1	<1	9
OIL & GREASE (mg/l)	15	<1.0	<1.0	<1.1	<2.72	<1.82	<1.34	1.0	1.8	<1.0	<1.0	<1.0
TOT. RESID. OXIDANTS (mg/l)	0.02	<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	<0.10	<1.0	0.12	<1.0	<1.0
TEMPERATURE (F)	REPORT	88.2	85.2	77.1	64.7	61.2	61.3	–	–	–	–	–
AMMONIA (NH3) (mg/l)	1.8 (SUM) 2.6 (WIN)	0.16	0.09	0.110	0.18	0.03	0.062	<0.10	0.14	0.14	<0.10	<0.10
COPPER (mg/l)	0.05	0.033	0.038	0.057	0.046	0.064	0.020	0.035	<0.01	0.013	<0.010	<0.010
LEAD (mg/l)	0.02	0.007	0.007	0.006	0.006	0.004	0.002	<0.10	<0.10	<0.10	<0.10	<0.10
ZINC (mg/l)	0.46	0.053	0.046	0.054	0.071	0.098	0.040	0.035	0.026	0.023	0.024	0.019
ALUMINUM (mg/l)	0.40	1.094	1.045	1.066	0.824	0.528	0.880	0.63	0.31	0.29	0.36	0.82
IRON (mg/l)	REPORT	1.281	0.612	0.936	1.040	1.747	0.280	0.056	0.20	0.099	0.057	0.73
pH (pH units)	6.0 MIN/9.0 MAX	7.5	7.5	7.7	7.4	7.7	7.8	–	–	–	–	–
PCBS (mg/l)	(EMP)	–	–	–	–	–	–	<1.0	<1.0	<1.0	<1.0	<1.0
ALKALINITY (mg CaCO3)	(EMP)	–	–	–	–	–	–	232	61	205	262	248

TABLE 6 – Q
SUMMARY OF
ENVIRONMENTAL MONITORING
RP&L's NPDES DISCHARGE AND FEEDWATER
(1994)

		RP&L MONTHLY MONITORING (1994)*								FEEDWATER SAMPLE LOCATION (Z)	
PARAMETERS	LIMITS	JAN	FEB	MAR	APR	MAY	JUNE	JULY	AUG	(5-17-94)	(8-16-94)
FLOW	REPORT	–	–	–	–	–	–	–	–	–	–
TSS (mg/l)	70	16.9	11.6	9.4	14.8	20.9	18.6	15.4	11.4	1	3
OIL & GREASE (mg/l)	15	<1.4	<1.09	2.31	1.78	<1.25	<1.31	<1.29	<1.1	1.1	<1.0
TOT. RESID. OXIDANTS (mg/l)	0.02	<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	<0.05
TEMPERATURE (F)	REPORT	59.5	60.3	64.2	66.2	75.6	84.4	–	–	–	–
AMMONIA (NH3) (mg/l)	1.8 (SUM) 2.6 (WIN)	0.25	0.183	0.140	0.164	0.226	0.182	–	–	<0.10	<0.10
COPPER (mg/l)	0.05	0.028	0.026	0.030	0.015	0.045	0.044	0.053	0.054	<0.010	<0.010
LEAD (mg/l)	0.02	0.004	0.004	0.008	0.004	0.009	0.010	0.009	0.006	<0.10	<0.10
ZINC (mg/l)	0.46	0.060	0.112	0.106	0.086	0.078	0.097	0.230	0.102	0.040	0.041
ALUMINUM (mg/l)	0.40	<0.003	1.243	0.834	1.440	1.149	0.960	0.580	0.400	0.14	<0.10
IRON (mg/l)	REPORT	0.606	0.566	0.438	1.840	0.700	0.646	0.714	0.447	0.045	0.051
pH (pH units)	6.0 MIN/9.0 MAX	7.6	7.7	8.0	7.6	7.7	7.8	7.5	7.7	–	–
PCBS (mg/l)	(EMP)	–	–	–	–	–	–	–	–	<1.0	<1.0
ALKALINITY (mg CaCO3)	(EMP)	–	–	–	–	–	–	–	–	264	209

TABLE 6- R

SOLID WASTE AND BY-PRODUCT EXCEEDING DETECTION LIMITS

	SAMPLE LOCATION									
PARAMETER	ECON. ASH 9/2/92	FRONT ESP HOPPER 9/2/92	FRONT ESP HOPPER 9/22/93	FRONT ESP HOPPER 8/16/94	BACK ESP HOPPER 9/23/94	BACK ESP HOPPER 5/17/94	BACK ESP HOPPER 8/16/94	BOTTOM ASH 9/24/94	BOTTOM ASH 12/7/93	BOTTOM ASH 8/16/94
TCLP METALS (mg/l)	BDL	BDL EXCEPT ARSENIC 1.3 mg/l	BDL EXCEPT ARSENIC 0.14 mg/l	BDL EXCEPT CADMIUM 0.13 mg/l	BDL EXCEPT CHROMIUM 1.1 mg/l	BDL	BDL EXCEPT ARSENIC 0.29 mg/l CADMIUM 0.21 mg/l CHROMIUM 0.27 mg/l	BDL	BDL	BDL
TCLP VOLATILES ORGANICS (mg/l)	BDL	BDL	BDL	BDL	BDL	BDL	BDL	BDL	BDL	BDL
TCLP BASE/NEUTRAL EXTRACTABLES (mg/l)	BDL	BDL	BDL	BDL	BDL	BDL	BDL	BDL	BDL	BDL
TCLP ACID EXTRACTABLES (mg/l)	BDL	BDL	BDL	BDL	BDL	BDL	BDL	BDL	BDL	BDL
VOLATILE ORGANICS ug/kg(VOC)	BDL EXCEPT METHYLENE CHLORIDE 46 µg/kg	BDL EXCEPT CARBON DISULFIDE 730 µg/kg TOLUENE 2200 µg/kg	BDL	BDL	BDL EXCEPT METHYLENE CHLORIDE 900 ug/kg	BDL EXCEPT TOLUENE 5.0 µg/kg	BDL EXCEPT METHYLENE CHLORIDE 30 µg/kg	BDL EXCEPT METHYLENE CHLORIDE 8.0 ug/kg	BDL EXCEPT METHYLENE CHLORIDE 13.0 µg/kg	BDL EXCEPT METHYLENE CHLORIDE 12 µg/kg

(BDL - BELOW DETECTABLE LIMITS)

APPENDIX IV

DRAFT PARAMETRIC TEST REPORT

Contract No. DE-FC22-90

Parametric Test Report

LIFAC Sorbent Injection Desulfurization Demonstration Project

DRAFT

Presented By

LIFAC North America, Inc.

A Joint Venture Between

ICF KAISER ENGINEERS

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Tampella power

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DRAFT

Presented To

DRAFT

U.S. Department of Energy

Pittsburgh Energy Technology Center
Pittsburgh, Pennsylvania 15236

LIFAC DEMONSTRATION

AT
RICHMOND POWER AND LIGHT

WHITEWATER VALLEY UNIT #2

PARAMETRIC TEST REPORT

Sponsored by:

U.S. Department of Energy
LIFAC North America
Richmond Power and Light
State of Indiana
Electric Power Research Institute
Black Beauty Coal Company

Cooperative Agreement No. DE-FC22-90PC90548

DRAFT

Submitted by:

LIFAC North America

Prepared by:

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DRAFT

ABSTRACT

This publication discusses the demonstration of the LIFAC sorbent injection technology at Richmond Power and Light's Whitewater Valley Unit No. 2, performed under the auspices of the U.S. Department of Energy's (DOE) Clean Coal Technology Program. LIFAC is a sorbent injection technology capable of removing 75 to 85 percent of a powerplant's SO_2 emissions using limestone at calcium to sulfur molar ratios of between 2 and 2.5 to 1. The site of the demonstration is a coal-fired electric utility powerplant located in Richmond, Indiana. The project is being conducted by LIFAC North America (LIFAC NA), a joint venture partnership of Tampella Power Corporation and ICF Kaiser Engineers, in cooperation with DOE, RP&L, and several other organizations including the Electric Power Research Institute (EPRI), the State of Indiana, and Black Beauty Coal Company. The purpose of this parametric test report is to explain the implementation of the tests and present all findings. The report has been prepared pursuant to the Co-operative Agreement No. DE-FC22-90PC90548 between LIFAC NA and the U.S. Department of Energy.

DRAFT

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PARAMETRIC TEST REPORT

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LIST OF ABBREVIATIONS, ACRONYMS AND TERMS

$CaCO_3$	Calcium carbonate
CaO	Calcium oxide
$CaSO_3$	Calcium sulfite
$CaSO_4$	Calcium sulfate
$Ca(OH)_2$	Calcium hydroxide, hydrated lime
Ca/S	Calcium/sulfur molar ratio
CCT	Clean Coal Technology
DOE	Department of Energy
EER	Energy and Environmental Research Center
EPA	Environmental Protection Agency
EPRI	Electric Power Research Institute
ESP	Electrostatic precipitator
FGD	Flue Gas Desulfurization
ICF KE	ICF Kaiser Engineers
ID fan	Induced draft fan
LIFAC	Limestone Injection into the Furnace with Activation of unreacted Calcium oxide, registered trademark of Tampella desulfurization process
LIFAC-NA	LIFAC - North America partnership
LIMB	Lime Injection to Multistage Burner, registered trademark of EPA desulfurization process
LNCFS	Low-NO_x Concentric Firing System
$MgCO_3$	Magnesium carbonate
NO	Nitrogen oxide
NO_x	Nitrogen oxides
O_2	Oxygen
PLC	Programmable Logic Control
PMS	Process Monitoring System
RP&L	Richmond Power & Light
SCA	Specific collecting area
SO_2	Sulfur dioxide
SO_3	Sulfur trioxide
dT	Temperature difference between actual flue gas temperature and saturation temperature (a.k.a. Approach to saturation)

Injection	Pulverized limestone injection into the furnace through several injection nozzles at the furnace walls
Activation	Activation of the remaining CaO particles with water for further sulfur capture
Recycling	Recycling of ash and unspent sorbent from ESP hoppers back to the reactor inlet

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0.0 EXECUTIVE SUMMARY

The Clean Coal Technology Program (CCT) has been recognized in the National Energy Strategy as a major initiative whereby coal will be able to reach its full potential as a source of energy for the nation and the international marketplace. Attainment of this goal depends upon the development of highly efficient, environmentally sound, competitive coal utilization technologies responsive to diverse energy markets and varied consumer needs. The CCT Program is an effort jointly funded by government and industry whereby the most promising of the advanced coal-based technologies are being moved into the marketplace through demonstration. The CCT Program is being implemented through a total of five competitive solicitations.

LIFAC North America, a joint venture partnership of ICF Kaiser Engineers, Inc. and Tampella Power Corporation, is currently demonstrating the LIFAC flue gas desulfurization technology developed by Tampella Power. This technology provides sulfur dioxide emission control for powerplants, especially existing facilities with tight space limitations. Sulfur dioxide emissions are expected to be reduced by up to 85% by using limestone as a sorbent. The LIFAC technology is being demonstrated at Whitewater Valley Unit No. 2, a 60-MWe coal-fired powerplant owned and operated by Richmond Power and Light (RP&L) and located in Richmond, Indiana. The Whitewater plant consumes high-sulfur coals with sulfur contents ranging from 2.0 - 2.9 percent.

The project, co-funded by LIFAC North America and DOE, is being conducted with the participation of Richmond Power and Light, the State of Indiana, the Electric Power Research Institute, and the Black Beauty Coal Company. The project has a total cost of 21.4 million dollars and a duration of 48 months from the preliminary design phase through the testing program.

The sponsors of this project believe that LIFAC has the potential to be a new and important SO_2 control option for U.S. utilities subject to the Clean Air Act's acid rain regulations. To be considered as a commercially feasible option in this particular emissions control market, LIFAC must demonstrate a high SO_2 removal rate while remaining competitive with other options on a cost per ton of SO_2 removed basis.

The LIFAC system combines conventional limestone injection into the upper furnace region with a post-furnace humidification reactor located between the air preheater and the ESP. The process produces a dry, stable waste product that is removed from both the bottom of the humidification reactor and the ESP.

Finely pulverized limestone is pneumatically conveyed and injected into the upper region of the boiler where temperatures are approximately 1800 to 2200 degrees Fahrenheit. At these temperatures the limestone ($CaCO_3$) calcines to form lime (CaO) which readily reacts with the SO_2 to form calcium sulfate ($CaSO_4$). All of the sulfur trioxide (SO_3) reacts with the CaO to form $CaSO_4$.

Approximately 25 percent of the sulfur dioxide removal occurs in the boiler with the remaining 75 percent and the unreacted lime passing through the air preheater to the vertical elongation of ductwork known as

the LIFAC activation reactor. There the flue gas is sprayed with atomized water that allows the unreacted lime to hydrate to $Ca(OH)_2$ which more readily reacts with the sulfur dioxide and forms $CaSO_3$. A combination of the proper water droplet size and residence time allows for effective hydration of the lime and complete water evaporation to create a dry reactor bottom product.

After exiting the humidification reactor, the flue gas is reheated before entering the ESP. Sixty percent of the LIFAC-produced spent sorbent and fly ash is collected by the ESP with the other 40 percent collected by the humidification reactor. Both the reactor and ESP ash may be recycled to a point ahead of the reactor to improve sorbent utilization and to improve the SO_2 removal efficiency of the system to the range of 75 to 85 percent.

LIFAC is similar to other current sorbent injection technologies but has unique advantages with its use of a patented vertical humidification reactor. And while LIFAC's sulfur dioxide removal efficiency is not as high as traditional wet flue gas desulfurization systems, its cost and simplicity of design, construction and operation offer other advantages over these alternative systems. In particular the advantages of the LIFAC system are:

- **High SO_2 removal rates** - Currently available sorbent injection systems have been unable to sustain high SO_2 removal rates with any consistency. LIFAC has proven in the past and is demonstrating during this project the ability to achieve and sustain high SO_2 removal rates of 75 to 85 percent over long operating periods.

- **By-products** - Wet lime and limestone scrubbing systems create a wet byproduct ash that must be further treated before disposal. LIFAC produces a dry solid waste ash containing calcium sulfite, calcium sulfate and fly ash. This waste is easily disposed of under U.S. regulatory requirements, may be recycled to increase LIFAC's efficiency and may have commercial applications in the cement industry.

- **Compatibility and Adaptability** - LIFAC has minimal impact on the host's site and systems, primarily the boiler, ESP and ID fan. In addition, LIFAC requires little space and few utilities and therefore is easily installed even in small or cramped powerplant sites.

Construction of the LIFAC system occurred in two phases over a period of one and a half years. The first phase of construction was completed during a routine plant outage in March, 1991. This period was utilized to install tie-ins to the host site's existing systems.

Ductwork and three dampers were installed between the air preheater and ESP to allow flue gas flow to the LIFAC activation reactor. Tie-ins were also made to the powerplant's high-pressure steam, condensate and river-water supplies. The high-pressure steam is required to reheat the flue gas exiting the LIFAC reactor and the water is needed for flue gas humidification inside the reactor. Injection ports were also installed in the boiler walls about 10 feet above the nose elevation.

The second phase of construction began in the Fall of 1991 with the driving of reactor piling and the installation of underground conduit runs. Work continued through to the Summer of 1992 with no need for plant downtime other than normally scheduled outages. During this time the limestone storage area was completed and the injection system was installed on Unit #2. The activation reactor was constructed and then tested with both cold air during a scheduled Unit #2 outage and hot flue gas during a low electricity demand period. Other powerplant tie-ins such as the steam and condensate system were also tested during low demand periods in the evenings or on weekends.

All of the construction work associated with the LIFAC system was performed in close proximity to the exterior of the powerplant or in cramped areas inside the plant. The ductwork tie-ins and new steelwork required inside the plant are located in small, difficult to access work areas. The reactor structure is approximately ten feet from the powerplant with the outside ductwork and piping crossing overtop of offices and the plant maintenance area. All of these new structures and equipment were constructed with no interference to daily plant operations.

The current schedule for the LIFAC demonstration program extends over a four year period from the beginning of preliminary design in August 1990 through the testing program to be completed in early August 1994. The LIFAC system was originally scheduled to come on-line in June of 1992 but due to delays in receiving construction permits and some minor startup problems, this date was moved to March 1993. Testing is scheduled through the summer of 1994.

The test plan for the LIFAC demonstration is composed of five distinct phases, each with its own objective. The first of these phases consists of the initial baseline testing portion of the project. Measurements will be taken to characterize the operation of the host's boiler and associated subsystems prior to the use of the LIFAC system. The results will be used for comparison purposes with the LIFAC system in operation and with data collected at the end of the project to determine any changes in the host's systems.

The second, or parametric, phase of testing will be performed to determine the best combination of LIFAC process variables for SO_2 removal. The variables to be studied include the limestone injection nozzles' angle and location, the Ca/S molar ratio, the need for supplemental injection air at the boiler, the water droplet size and injection nozzle arrangement in the reactor, the ash recycling ratio and the approach to saturation temperature of the flue gas exiting the activation reactor. The best combination of these variables will be chosen at the conclusion of this phase and used for the remainder of the test program.

Optimization tests will be conducted to examine the effects of different coal and limestone feeds on the SO_2 capture rate. Coals with sulfur contents as high as 3.3 percent will be tested to determine LIFAC's compatibility with high sulfur U.S. coals. Limestones with different compositions will also be tested to determine the LIFAC system's adaptability to local sorbent sources.

Long term testing will also be performed to demonstrate LIFAC's

performance under commercial conditions. The LIFAC system will be in operation 24 hours per day for several weeks using the powerplant's baseline coal, high calcium limestone and the optimum combination of process variables. In addition to process performance measurements, during this phase the operation and maintenance requirements of the system will be examined. Long term (two to three weeks) tests will also be conducted with two other coals; one lower sulfur coal (1.5%) and one higher sulfur content coal (3.3%).

The final phase of testing is composed of the post-LIFAC tests. The baseline tests will be repeated to gather information on the condition of the boiler and its associated subsystems. Comparisons will be made to the original baseline data to identify any changes either caused by the LIFAC system or independent of its operation.

At this point it has been shown at RP&L and other LIFAC installations that the system can be installed without affecting normal powerplant operations. The demonstration will show that the system can economically reduce SO_2 emissions when compared with other flue gas desulfurization technologies.

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1.0 PROJECT OVERVIEW

1.1 Purpose of the Parametric Test Report

The purpose of this report is to provide a description of parametric testing and results of the tests performed during the demonstration at Richmond Power & Light. Results will be used to assess the impacts of all process parameters on LIFAC's efficiency and its impact on the host facility. Parameters will then be optimized and used for long-term process testing. The Parametric Test Report has been prepared pursuant to the Co-operative Agreement No. DE-FC22-90PC90548 between LIFAC NA and DOE titled *"LIFAC Sorbent Injection Desulfurization Project at Richmond Power and Light Whitewater Valley Unit #2."*

1.2 Description of the Project

1.2.1 Project History

In 1983, Finland enacted acid rain legislation which applied limits on SO_2 emissions sufficient to require that flue gas desulfurization systems have the capability to remove about 80 percent of the sulfur dioxide in the flue gas. This level could be met by conventional wet limestone scrubbers but not by then available sorbent injection technology. Tampella, therefore, began developing an alternative sorbent injection system which resulted in the LIFAC process.

In 1986, the first full scale test was performed at Imatran Voima's Inkoo powerplant using a 70 megawatt side-stream from a 250 megawatt boiler. A 76 percent SO_2 removal rate with 1.5% sulfur coal was reached. A second LIFAC activation reactor was constructed to handle an additional 125 megawatt sidestream. This newer reactor is achieving removal rates of 75 to 80 percent while using Ca/S molar ratios of between 2 and 2.5 to 1. Also, in 1988 the first tests with high sulfur U.S. coals were run at the Neste Kullo Laboratory. A Pittsburgh No. 8 Seam coal containing 3 percent sulfur was tested and an SO_2 removal rate of 77 percent was achieved at a Ca/S molar ratio of 2 to 1.

The U.S. Department of Energy has emphasized the use and further development of coal as an energy source for utilities and the industrial sector. At the same time environmental responsibility has been mandated by the passage of the Clean Air Act Amendments of 1990. This Act establishes new lower emission levels of sulfur dioxide from utility powerplants with Phase One of the regulations coming into effect in January 1995 and the more stringent Phase Two regulations beginning in January 2000. To realize full potential of coal as an energy source while still complying with the new air pollution regulations, the Department of Energy

initiated the Clean Coal Technology (CCT) Program.

The CCT Program is an effort jointly funded by DOE and private industry whereby the most promising clean coal technologies are moved to the marketplace through demonstration. Five rounds of competitive solicitations comprise the CCT Program with four of these rounds having been completed. The LIFAC system was one of thirteen projects selected for funding under Round III of the CCT Program and a Cooperative Agreement between DOE and LIFAC-NA was signed in November, 1990. Due to scheduled outages at the host site, Richmond Power & Light's (RP&L) Whitewater Valley Unit #2 in Richmond, Indiana, design and procurement of critical equipment began in August 1990, with DOE funding contingent on final signing of the Cooperative Agreement.

1.2.2 Project Organization

The LIFAC demonstration is being conducted by LIFAC North America, a joint venture partnership between Tampella Power Corporation and ICF Kaiser Engineers. Tampella Power Corp. is a subsidiary of a large diversified international company, Tampella Corporation, based in Tampere, Finland and the original developer of the LIFAC technology. ICF KE is a U.S. company based in Oakland, California, and a subsidiary of ICF Kaiser International, Inc. based in Fairfax, Virginia.

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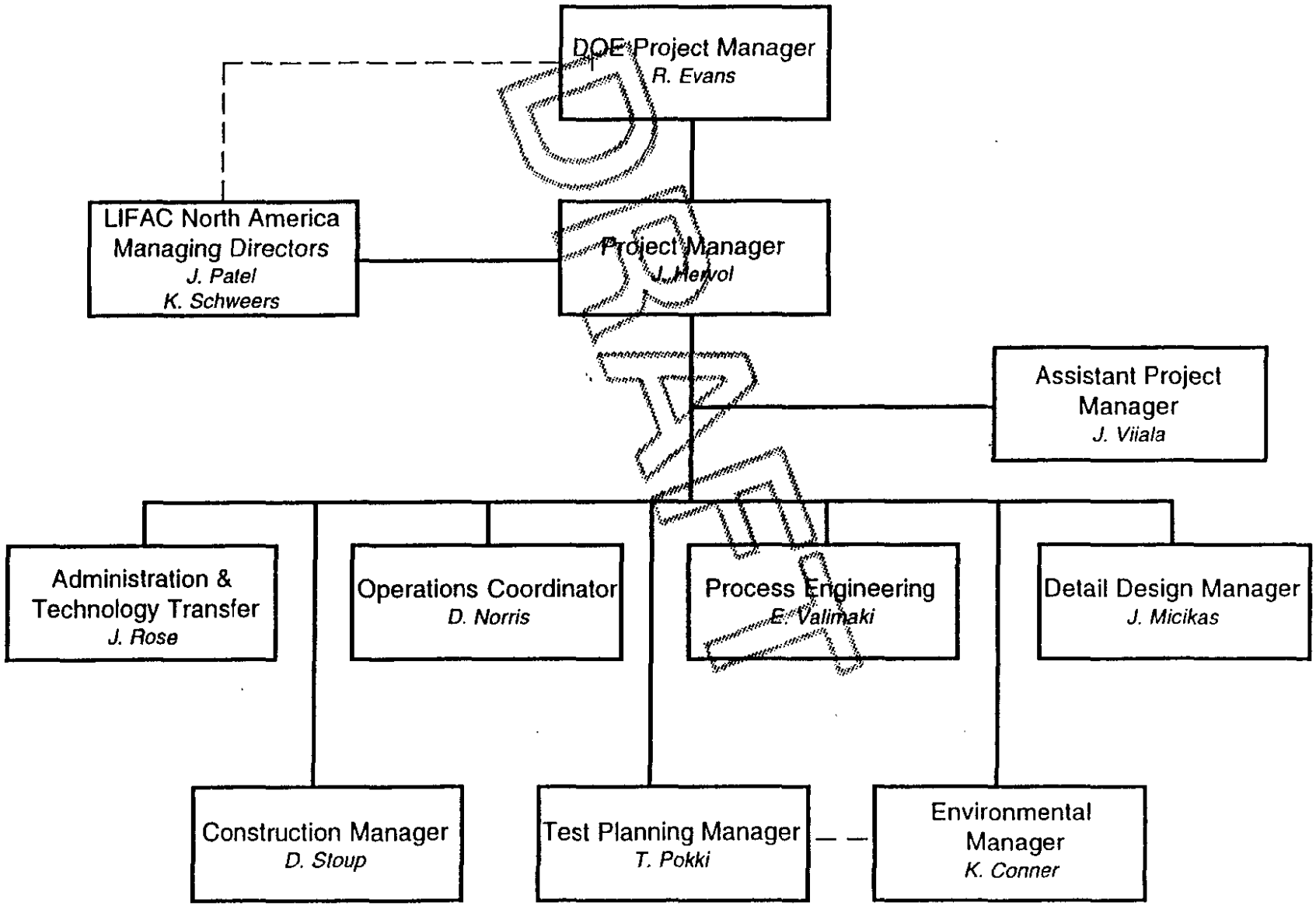
Figure 1-1 Project Organization



- **Cost** - LIFAC must compete with both low capital cost, low SO_2 removal rate options such as sorbent injection and high capital cost, high SO_2 removal rate options such as wet scrubbing. This project will demonstrate LIFAC's competitiveness on a cost per ton of SO_2 removed basis with these currently available alternatives.

- **Retrofit Adaptability** - The host site chosen required a retrofit with tight construction conditions that will prove LIFAC's ability to be installed where other technologies might not be possible. Construction will also demonstrate LIFAC's ability to be built and brought on-line with zero plant down time other than scheduled outages.

- **System Compatibility** - A major concern of utilities is the degree of compatibility of SO_2 removal systems with their existing operations. This demonstration will show LIFAC's minimal impact on the host site's boiler and associated subsystems.

1.4 Significance of the Project

1.4.1 Potential for Commercialization

Wet Scrubbers are the most prevalent scrubber technology and account for approximately 90% of U.S. scrubber systems. Wet FGD systems that use lime or limestone remove about 90% of the SO_2 and usually produce a sulfite/sulfate sludge waste product. Although the LIFAC process cannot match the high removal rates (90%) achieved by conventional wet scrubbers, the process does offer several advantages including:

- The need for slurry preparation/handling equipment is eliminated.

- The technology can be more easily retrofit onto most powerplants because the vertical activation chamber requires less space.

- The technology has lower capital costs which is especially attractive to existing plants that have fewer years to amortize capital investments as compared to new long-lived powerplants.

- The technology uses a widely available reagent material, limestone, rather than more expensive materials such as lime.

- The waste product is dry and easy to handle. In comparison, conventional wet limestone scrubbers produce a wet sludge which requires special handling and treatment.

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- The technology is typically compatible with other plant systems such as ESPs and ID fans, thereby minimizing costly retrofit plant modifications in order to employ the technology.

- The LIFAC system also has potential advantages over less conventional sorbent injection systems now being tested. These include:

- Use of limestone as opposed to lime or other more expensive sorbents.

- Removal rates of 75-85% which exceed the removal rates of many dry sorbent injection systems.

- Improved control of slagging and fouling associated with humidification in a vertical chamber as opposed to in-duct humidification.

- The LIFAC technology's potential for commercialization is increased by its ability to remove 75-85% of the SO_2, its low space requirement, and its low retrofit costs.

1.4.2 Process Advantages

The LIFAC process offers many advantages compared with other FGD systems. These technical and economic advantages are expected to gain industry acceptance of LIFAC as a viable SO_2 removal alternative. LIFAC's favorable attributes include:

- The LIFAC process has lower initial capital requirements and is less expensive to operate than other FGD systems, in particular, wet scrubbers.

- The system is suitable for retrofit applications since it requires less space than other FGD systems.

- The system is capable of removing more sulfur (75-85%) than other technologies competing with wet scrubbers. This level of removal makes LIFAC suitable for use with high-sulfur coal, unlike other dry sorbent injection processes and spray dryers.

- The LIFAC Process uses limestone, which is relatively inexpensive, as the reagent.

- The selection of the RP&L Whitewater Valley Unit 2 is particularly appropriate for demonstrating the LIFAC process. The host boiler has been designed to be very

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compact, making retrofit particularly difficult. This boiler burns high-sulfur coal and has a high utilization rate which will demonstrate the effectiveness and reliability of the process under U.S. operating conditions.

- The site also offers several advantages not related to the technical merits of the LIFAC process. The site was /used previously to demonstrate another upper-furnace sorbent injection technology and the equipment for sorbent handling and injection is available to the LIFAC demonstration, thus holding down project costs. The site is also located close to the nation's leading high-sulfur coal areas and has good transportation access to those areas.

1.5 DOE's Role in the Project

The DOE shall be responsible for monitoring all aspects of the project and for granting or denying all approvals required by the Cooperative Agreement. The DOE Contracting Officer is DOE's authorized representative for all matters related to the Cooperative Agreement.

The DOE Contracting Officer will appoint a Contracting Officer's Technical Representative (COTR) who is the authorized representative for all technical matters and has the authority to issue "Technical Advice" which may:

- Suggest redirection of the Cooperative Agreement effort, recommend a shifting of work emphasis between work areas or tasks, and suggest pursuit of certain lines of inquiry which assist in accomplishing the Statement of Work.

- Approve those reports, plans, and items of technical information required to be delivered by the Participant to DOE under the Cooperative Agreement.

- All technical advice shall be issued in writing by the DOE COTR.

The DOE COTR does not have the authority to issue any technical advice which:

- Constitutes an assignment of additional work outside the Statement of Work.

- In any manner causes an increase or decrease in the total estimated cost or the time required for performance of the Cooperative Agreement.

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- Interferes with the Participant's right to perform the terms and conditions of the Cooperative Agreement.

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2.0 TECHNOLOGY DESCRIPTION

2.1 Chemical Theory

The LIFAC (Limestone Injected into the Furnace with Activation of unreacted Calcium oxide) technology combines upper-furnace limestone injection followed by post-furnace humidification in an activation reactor located between the air preheater and the ESP. The process produces a dry and stable waste product that is partially removed from the bottom of the activation reactor and partially removed at the ESP. A Schematic of the entire process is shown in Figure 2-1.

Finely pulverized limestone is pneumatically conveyed and injected into the upper part of the boiler. Since the temperatures at the point of injection are in the range of 1800 - 2200° F, the limestone ($CaCO_3$) thermally decomposes to form calcium oxide (CaO). As the CaO passes through the furnace, initial desulfurization reactions occur. A portion of the sulfur dioxide (SO_2) reacts with CaO to form calcium sulfate ($CaSO_4$). Essentially all of the sulfur trioxide (SO_3) reacts with CaO to form $CaSO_4$.

The flue gas, unreacted CaO and ash exit the boiler and pass through the air preheater. On leaving the air preheater, the gas/CaO/ash mixture is directed to the LIFAC activation reactor. In the reactor, additional SO_2 capture occurs after the flue gas is humidified with a water spray. Humidification converts CaO to calcium hydroxide ($Ca(OH)_2$), which enhances further SO_2 removal. The primary reaction product in the activation reactor is calcium sulfite ($CaSO_3$). The activation reactor is designed to allow time for effective humidification of the flue gas, activation of the CaO, and reaction of the SO_2 with the sorbent. All the water droplets evaporate before the flue gas leaves the activation reactor. The net effect: at a Ca/S ratio in the range of 2:1 to 2.5:1, 75-85% of the SO_2 is removed from the flue gas.

The flue gas leaving the activation reactor then enters the existing ESP where the spent sorbent and fly ash are removed from the flue gas and sent to the disposal facilities. The solids collected by ESP consist of fly ash, $CaCO_3$, CaO, $Ca(OH)_2$, $CaSO_4$, and $CaSO_3$. To improve utilization of the calcium, and increase SO_2 reduction a portion of the spent sorbent collected in the bottom of the ESP hoppers is recycled back into the ductwork just ahead of the activation reactor.

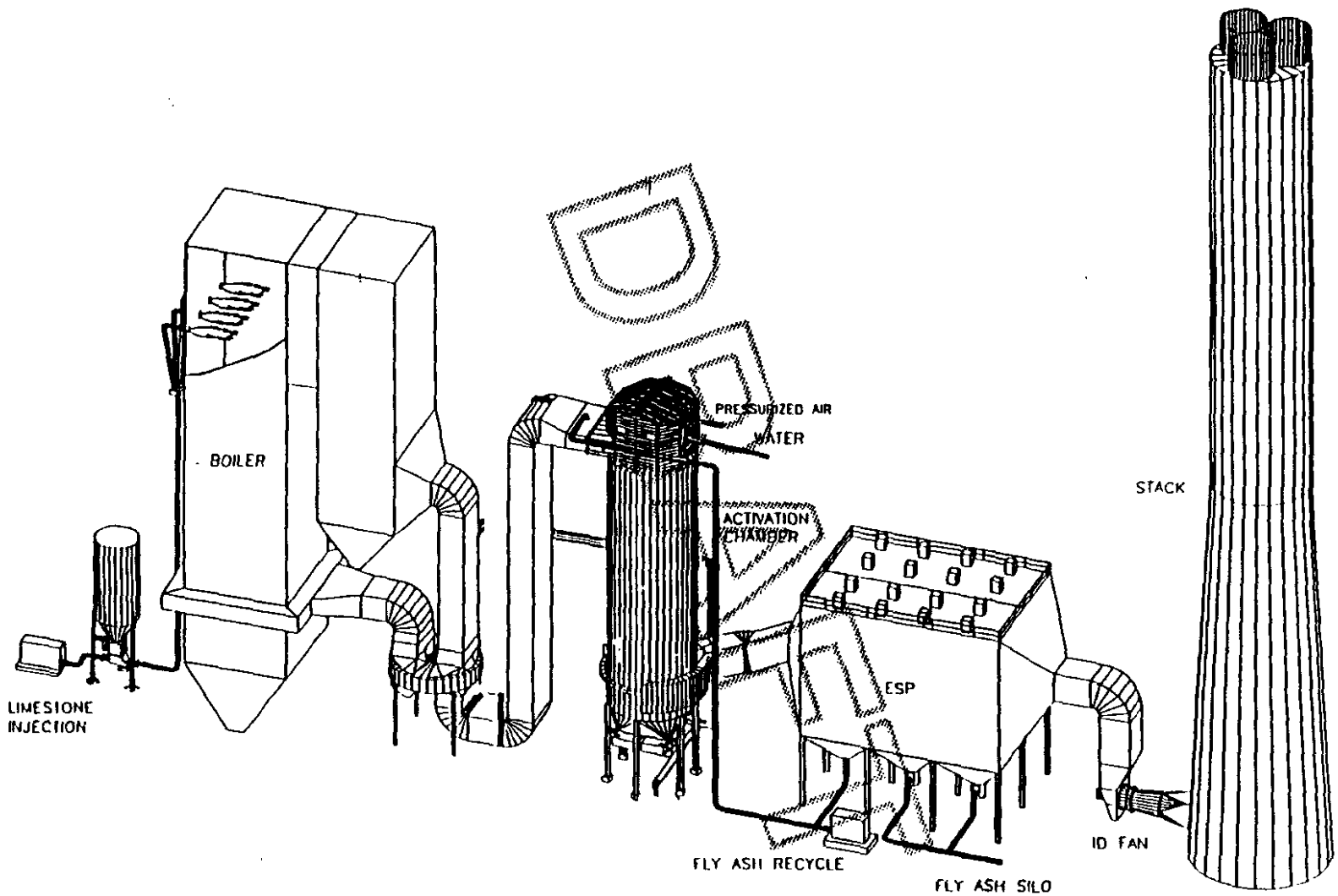

Figure 2-1

Schematic of LIFAC Desulfurization System

3.0 TEST PROGRAM DESCRIPTION

3.1 Objectives

The LIFAC test program has two main objectives. The first is to showcase the LIFAC system by demonstrating the:

- LIFAC system for U.S. utilities
- Economic feasibility of the system suitability of the process for different coals and limestones
- System's effectiveness at reducing SO_2 emissions when burning high sulfur coals.

The second objective is to perform an in-depth evaluation of the LIFAC system's effect on the host site's boiler and associated equipment. In meeting this objective, the test program will provide the data for the following activities:

- Study the effects on other boiler emissions such as NO and particulates.
- Optimization of the system's performance in the host boiler's environment.
- Evaluate the impacts of LIFAC on the host boiler and its associated subsystems.
- Evaluate any changes in boiler performance during and after LIFAC testing that are or are not related to the LIFAC system.

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3.2 Test plan

The test program at Richmond was divided into a series of five separate sets of tests. The first set of tests is designed to gather baseline data of the boiler and other plant operation conditions. The LIFAC system will then be optimized and tested with various coal and limestone feeds. Finally, post-LIFAC testing will provide data for comparison to pre-LIFAC baseline data. The series of tests with brief descriptions are in chronological order as follows:

1. Baseline Tests - Baseline testing was performed prior to start-up of the LIFAC system. This testing provided data on the current operating conditions of the host boiler which will be used for comparison throughout the demonstration program.

2. Parametric Tests - Parametric testing followed baseline testing and is designed to evaluate the different possible combinations of LIFAC process parameters and their effectiveness. Also during this test period, five different types of sorbent was to be tested. This set

of tests will determine the optimum combination and settings of process parameters which will then be used throughout the remaining demonstration tests.

3. Optimization Tests - Optimization testing will be performed after the parametric tests to determine the reliability and capability of the LIFAC process over short, continuous operation periods. This set of tests will be performed continuously 24 hours a day in actual operating conditions, as opposed to the short 8- to 12-hour parametric tests performed previously. The baseline coal and limestone along with the optimum parametric settings will be used during this phase of testing.

4. Long-Term Tests - Long-term testing will be performed after the optimization tests to determine the reliability and capability of the LIFAC process over long, continuous operation periods. This set of tests will be performed continuously over several hundreds of hours. One coal quality will be selected for this test period. Also, following this phase and preceding the post-LIFAC tests, selected parametric and optimization tests will be repeated.

5. Post-LIFAC Tests - Post-LIFAC testing will be similar to baseline testing. Data will be gathered without the LIFAC system operating which, when compared with the baseline data, will yield evidence of the impacts LIFAC may have had on the host site's boiler and associated subsystem.

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The test program will be staffed by a combination of personnel from RP&L, ICF KE, and Tampella. Requirements for each company will vary depending on the type of testing being performed and its duration. The test program, including all five sets of tests, is calculated to last approximately 2,800 hours over a 26-month period. The structure of the test plan at Richmond Power & Light is displayed in Figure 3-1.

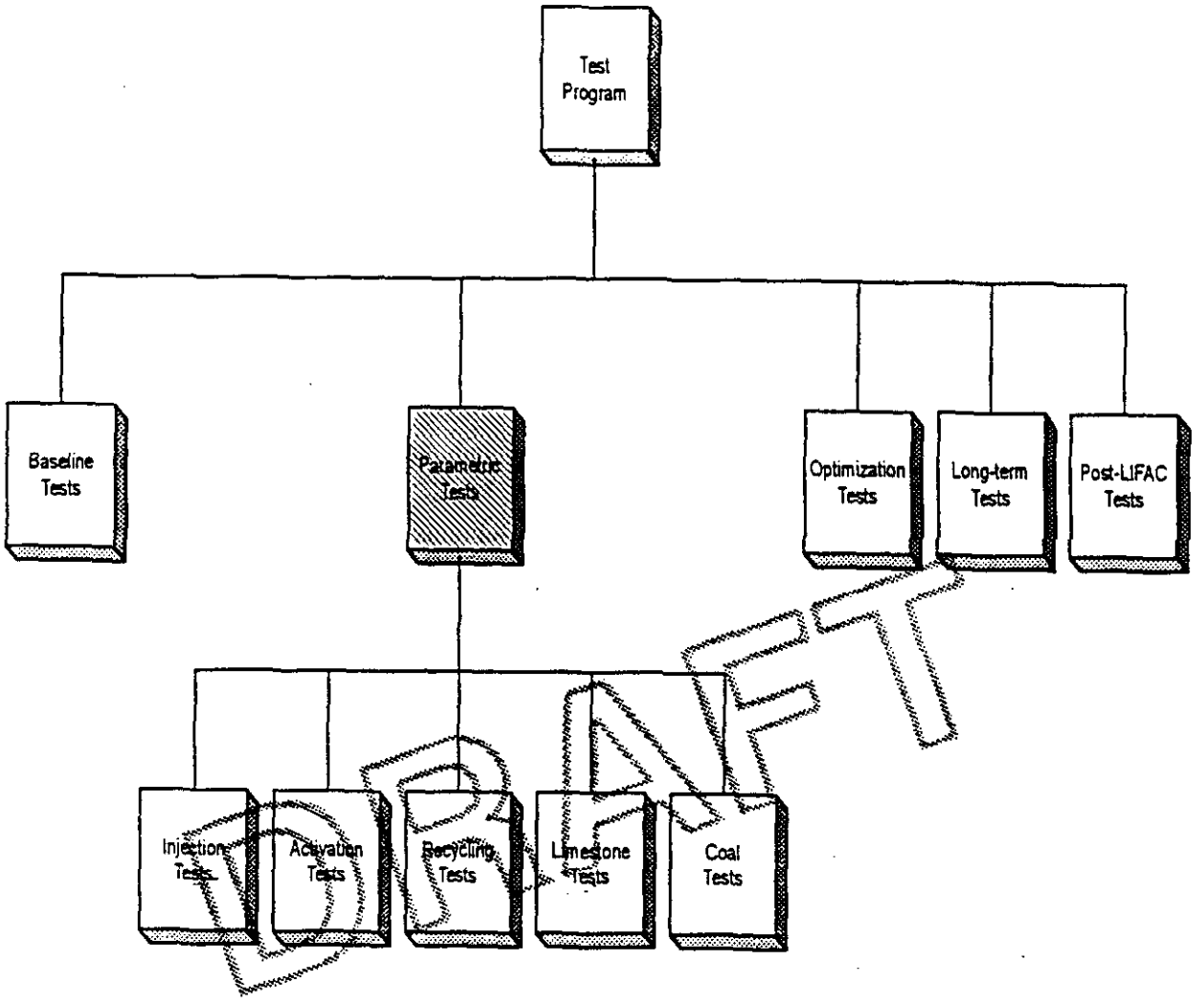

Figure 3-1

Structure of the LIFAC Demonstration Test Plan

3.3 Test Schedule

The total duration of all five sets of tests will be 26 months. Testing began in June 1992 and will be completed in August 1994. Actual test hours during this period will total 2,800.

The baseline tests were performed during the summer of 1992. Parametric tests were performed in several test sets during 1993. The long-term tests will be performed during the winter and spring of 1994. Post-LIFAC testing will be performed during the summer 1994. Selected tests will be repeated before post-LIFAC testing is initiated.

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4.0 PROCESS PARAMETERS

4.1 General

Process parameters are variables of the LIFAC system which can have diverse effects on its performance and operation. The parameters are of two types: material characteristics and operation conditions. Different combinations of these variables can cause a wide range of sulfur removal efficiencies and may also affect the equipment, and other processes, integral to the host boiler and its ancillaries.

Material characteristic parameters include the limestone sorbent and combustion coal. These variables are dependent on their sources and/or preparation. During the test program, different types of coal were consumed as a fuel source and different sorbent qualities were injected for evaluation.

The operating condition parameters involve: boiler load, Ca/S molar ratio, injection setting, injection nozzle angle, booster air, activation reactor outlet temperature, and ash recycling ratio.

Boiler load is dependent on local power demands and varies accordingly throughout a typical day. During the test program, Unit No.2 operated at its normal range of loads. For some parametric tests there was a need to run specific boiler loads for a short time.

The three variables expected to most significantly effect desulfurization efficiency were Ca/S molar ratio, activation reactor outlet temperature, and the recycling ratio. The remaining parameters were also evaluated in an attempt to maximize the efficiency, operability, and economy of the LIFAC process. The ranges of each parameter are shown in Table 4.1.

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Table 4.1 Process Parameters and Ranges

PROCESS PARAMETER	UNIT OF MEASURE	TEST RANGE PLAN	TEST RANGE ACTUAL
Coal Quality	Weigh % of Sulfur as received	1.5 to 3.3	1.6 to 2.8
Limestone particle size & sorbent quality	Mesh & quality	85% < 325 I 85% < 325 II 85% < 200 dolomitic hydrated lime	85% < 325 Not tested 85% < 200 Not tested Not tested
Boiler load	MW	40 to 65	40 to 61
Ca/S molar ratio	mol/mol	1.0 to 3.0	1.0 to 2.8
Injection port combination	high, low, left, right	4 settings	4 settings
Injection nozzle angle	Degrees from horizontal	-15, 0, +15	0
Booster air	On & Off	On & Off	Not tested
Reactor outlet temperature	F above the saturation temperature	11 or more	2 to 14
Water droplet size and nozzle position	um	not included	50 to 100
Ash recycle ratio	lb/lb	0, 1 & 2	from 0 to 1

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4.2 Coal Quality

An important objective of this test program was to demonstrate the suitability of the LIFAC process to commonly used U.S. coals. Therefore, three coals with varying sulfur contents were tested. The primary coal (Test Coal #1) used during testing was a bituminous coal containing an average of 2.5% by weight of sulfur. Test Coal #2 had approximately 3.3% sulfur, while Test Coal #3 contained about 1.5% sulfur. The other properties of these coals such as heating value, ash, and moisture content remained essentially unchanged. A typical analysis of coal is shown in Appendix 7

4.3 Limestone Quality

Limestone quality is characterized by its calcium and magnesium contents, particle size, and its formation type. High calcium limestone contains more than 90 w% calcium carbonate ($CaCO_3$). Dolomitic limestone typically has 50 w% magnesium carbonate ($MgCO_3$) and 50 w% calcium carbonate content. At typical furnace temperatures, only calcium forms stable compounds with sulfur. Under these conditions the magnesium content of dolomites are considered an impurity of limestone. One advantage that dolomitic limestone has over high-calcium limestones is the high porosity and large specific surface area of the particles. This provides more active surface area for the sulfation reaction to occur.

Limestones were also classified according to particle size. The smaller the limestone particles, the more active surface area is available for the sulfation reaction, which increases the sulfur dioxide removal rate. However, the smaller particle size increases the cost of grinding, which adds to the operation cost of the LIFAC system.

Two types of limestone formations, which were to be tested, were sedimentary and crystalline. Crystalline limestone has been subjected to metamorphic geologic forces which do not contribute to the formation of sedimentary limestone. The resulting differences in physical structure may react differently in the furnace.

According to previous LIFAC tests, the best commercial limestone quality has been high calcium limestone with a particle size distribution of 85% passing 325 mesh.

During testing, five different sorbents were to be tested. Limestone #1 is a high calcium limestone (90% $CaCo_3$) with 85% passing 325 mesh and is of sedimentary origin. Limestone #2 is also a high calcium limestone and from sedimentary origin, but with 85% passing only 200 mesh. The coarser grind limestone was tested for economic reasons due to its lower cost. Limestone #3 is a dolomitic limestone with 50% $MgCO_3$ and a grind size of 85% passing 325 mesh. Limestone #4 will be a high calcium limestone of crystalline structure and 85%

passing 325 mesh. Finally, for comparison with earlier studies, hydrated lime will be tested as Limestone #5.

4.4 Boiler Load

Boiler load has a major impact on the sulfur removal efficiency in the LIFAC process. Different boiler loads create a wide range of gas temperatures and velocities within the furnace to which the LIFAC system is sensitive.

The SO_2 removal process begins with the calcination of limestone in the furnace. The two most influencing factors in this initial reaction are flue gas temperature and residence time. Calcination occurs most effectively in the 1800° - 2200°F temperature range and with a boiler residence time of 1 to 1.5 seconds. At lower temperatures and/or shorter residence times, the calcination reaction does not take place rapidly enough. At higher temperatures, the calcium particles' porous structure shrinks, leaving little active surface area for SO_2 removal in what is termed "dead burn."

The boiler at the host site is a 60 MW unit but can be operated past the design load to approximately 65 MW. At loads above 60 MW, more coal is fed to the boiler which increases the amount of flue gas produced. The extra flue gas lowers the heat transfer efficiency of the boiler walls and thereby raises the temperature inside the boiler. At a 65-MW load, temperatures in the boiler nose region range from 2200° to 2300°F. Also, with the increased quantity of flue gas comes a higher gas velocity and subsequent lowering of residence time. Both of these conditions lower the SO_2 removal efficiency in the boiler. An increased in gas flow may also effect SO_2 reduction in the reactor.

The corner-fired host boiler creates different gas flow patterns and temperatures in different regions of the boiler. These varying temperatures can lead to limestone particles "dead burning" in certain zones and only achieving partial calcination in others. The different gas velocities also cause large variations in residence times.

Injecting limestone into the boiler may also increase the required frequency of sootblowing which will causes temperature variations of more than 200°F within the boiler, all of which impact SO_2 removal.

To optimize the SO_2 removal process, three different boiler loads will be tested. For each load, the limestone injection system will be operated in different configurations to achieve the highest possible SO_2 removal efficiency for each load.

4.5 Calcium/Sulfur Molar Ratio (Ca/S)

The calcium/sulfur (Ca/S) ratio is the ratio of moles of calcium injected into the boiler versus the moles of sulfur

present in the boiler. The Ca/S molar ratio is calculated from operating data and chemical analyses of the limestone and coal feeds to the boiler, using equation (1) in Section 8.3

Calcium and sulfur in the forms of calcium oxide (CaO) and sulfur dioxide (SO_2) react in a stoichiometric molar ratio of 1:1. Increasing this ratio increases the completeness of the SO_2 removal until a ratio of approximately 3:1 is reached. At higher ratios, the additional increase in SO_2 removal is negligible and the process also becomes uneconomical. Also, at higher molar ratios, fouling of the furnace walls may increase, thus lowering the heat exchange efficiency and increasing the need for sootblowing.

Past experience shows that ratios of 2:1 to 2.5:1 are the most effective. Ratios between 1 and 3 were planed to be evaluated for comparison during the tests.

4.6 Limestone Injection Port combination

There are two objectives in evaluating the different injection port combinations in the boiler: first to identify which ports are best for maximum SO_2 removal efficiency, and second, to determine which combination(s) have the least affect on normal boiler operation.

Within the furnace environment the limestone particle's physical structure is influenced by the temperature of the flue gas and radiation of the flame. Its route through the upper part of the furnace is influenced by flue gas velocity, turbulence, and buoyancy force. The limestone must be injected into the boiler at the proper location where the temperature is in the desired 1800 - 2200°F range and the particles have the necessary 1 to 1.5 seconds of residence time. The dispersion pattern and penetration into the cross-section of the furnace must also be as uniform as possible.

To achieve these objectives under different boiler conditions, the furnace has twelve injection nozzle ports. Six nozzles are at the nose level and six more are eleven feet higher. Four nozzles at each level are in the front wall with one nozzle on each side wall. Any combination of nozzles can be in operation at the same time.

In determining the optimum injection locations for calcination and sulfation of limestone, five different injection port combinations were used. The results of these tests yielded the best limestone injection port combinations for the range of boiler loads.

4.7 Injection Nozzle Angle

In addition to injection port combinations, the angle of each injection nozzle is individually adjustable in a vertical direction. The angle of the nozzles can be changed ± 15

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degrees from the horizontal position allowing limestone at each injection port location to be injected into different temperature zones. Three different positions: 15° up, horizontal, and 15° down were to be tested.

4.8 Booster Air

To improve limestone injection penetration and distribution in the furnace, the LIFAC process employs a booster, or secondary, air system. Booster air increases the injection velocity of the limestone particles and protects the particles against dead burning in high furnace temperatures.

The effects of booster air for SO_2 removal efficiency were to be studied by toggling the booster air on and off.

4.9 Activation Reactor Outlet Temperature

At temperatures below 1800· F, calcium oxides formed in the boiler no longer react with SO_2. Typically, flue gas inlet temperature to the activation reactor is nearly 325°F. To activate this unreacted calcium, water droplets are sprayed into the flue gas and collide with the calcium oxide particles. Upon collision, the water forms a liquid film around the solid particle. Both gaseous sulfur dioxide and solid calcium are soluble in water. $CaSO_3$ forms in the liquid phase as an ion reaction. The $CaSO_3$ then precipitates onto the surface of the solid particle as the water evaporates.

One objective of the activation tests is to evaluate the sulfur dioxide removal efficiency of the activation reactor at different reactor outlet temperatures. The reactor outlet temperature is a function of the volume and temperature of the flue gas to the reactor and the humidification water flow rate. The LIFAC process controls only the water flow rate. The other two variables are carefully monitored to maintain a constant reactor outlet temperature. Three reactor outlet temperatures were slated to be evaluated 11, 18, and 25°F above the flue gas saturation temperature.

4.10 Water Droplet Size and Nozzle Position

The reaction between sulfur dioxide and the unreacted sorbent are strongly dependent on the number of collisions between water droplets and sorbent particles. To maximize the number of collisions, the optimum droplet size needs to be identified. The size of water droplets also dictates the time required for them to evaporate. Droplets that are too small cause water evaporation before collision and droplets that are too large do not have enough time to evaporate which causes the ash in the bottom of the reactor to moisten.

Water droplet size is not a test parameter but it has an effect to process operability and SO_2 removal rate, if the droplet size is not optimized. Tampella has performed many

tests to find the best possible nozzle and water droplet size. The purpose of these tests were to evaluate the effects of water droplet size on reactor operability.

Water droplet size is controlled by changing the air and water pressure to the dual fluid nozzles inside the reactor. To minimize fouling on the nozzle headers in the humidification section, the flue gas flow pattern is optimized by changing the elevation of the nozzle header. The sulfur dioxide content of the flue gas and the moisture content of the reactor bottom ash, along with the reactor outlet temperature, were used to determine the lowest operational temperature.

There are seven nozzle headers and each header has four nozzles. Each header can be moved up and down approximately three feet. Nozzles can be opened or closed as needed. Normal operating procedure is that all nozzle headers are continuously in operation. Nozzle headers are adjustable in order to control ash build-up in the reactor. Normal operating procedure was to keep the headers as high as possible. This yielded the most efficient humidification.

4.11 Ash Recycling Ratio

A mixture of fly ash and unreacted calcium compounds collected in the ESP is partially recycled back to the reactor inlet ductwork. The ratio between the recycled material and material coming from the boiler to the reactor is called the recycle ratio.

Ash collected from the ESP and reactor bottom contains about 10 to 20% unreacted Calcium oxide (CaO). By recycling unreacted calcium back to the reactor, the Ca/S molar ratio rises without additional sorbent injection to the furnace. Hence, ash recycling improves SO_2 removal rate and the calcium utilization level.

Ash recycle also increases the solid content of the flue gas. This allows for increased humidification and the ability to operate at lower temperatures in the reactor. Lower reactor temperature improves SO_2 removal rate.

Ash recycle may effect the process' operability. An increased ash loading to the reactor may change the amount of ash removed from the bottom of the reactor. It may also result in increased particulate loading to the ESP. Recycling ratios of 0, 1 and 2 were tested and all of the advantages and disadvantages were evaluated.

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5.0 PARAMETRIC TEST IMPLEMENTATION

5.1 System startup and shakedown period

Process startup and shakedown was accomplished over several operation periods. The LIFAC system is easy to startup and shutdown. The process was operated from one to several hours during the shakedown period.

In order to assess the effects of higher reactor temperatures, hot flue gas was diverted through the reactor and associated ductwork. Thermal expansions were measured and the reactor structure was observed. The entire LIFAC system is under negative pressure, so outward gas leaks were not a concern.

All mechanical equipment and components were tested during the startup period and the early stages of parametric testing. Some critical process equipment needed immediate attention before operations could commence. Minor problems were experienced with the dampers, the water control valve, and the condensate return system. Process control and instrumentation calibration was performed by project personnel, RP&L personnel, and subcontractors. The process monitoring system was programmed in an office but required some field modifications and interlocking adjustments.

5.2 Operation Procedures

The objective of parametric testing was to determine the desulfurization capability of the LIFAC process under several operating conditions and arrangements by testing a wide range of process parameters. Boiler and LIFAC process operation conditions effect the process performance, economy, and operability. All important process variables were evaluated during the parametric tests. Process values and settings were briefly tested at the beginning of this test phase. A typical test would last from thirty to ninety minutes. The reason for these short tests was to gather initial data on the effects of each parameter and later concentrate on testing the most significant parameters. As parametric testing progressed, settings were tested for several hours. Some parameters effected SO_2 capture within the reactor, and others had an impact in the furnace and reactor.

Boiler operation parameters which influenced efficiency both in the furnace and reactor are:

- Boiler Load
- Coal Quality
- Ca/S Molar Ratio
- Limestone Quality
- Limestone Injection Setting

LIFAC process parameters which effect only reactor efficiency:

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- Reactor Temperature
- ESP Ash Recycling Rate
- Humidification Nozzle Setting

Process operating conditions were stabilized as long as possible during each parametric evaluation. Sootblowing occurrences disturbed measurements for one and a half hours. Gradual changes in boiler load occurred as the power demand fluctuated in the morning and evenings. These adjustments lasted between one and two hours typically. Some of these changes in the host boiler had an immediate impact on LIFAC process efficiency, while others did not.

The operability of the process and its impacts on boiler operation will be studied and reported during the long-term operation phase. However, some preliminary evaluations of operability were recorded during the short parametric tests.

Assessments of ESP performance and operation were conducted during parametric testing. Both transient and steady-state conditions were observed. ESP evaluations will continue throughout the demonstration and reported, in detail, in the long-term test report.

Parametric Test settings or ranges

The actual order of test implementation was determined according to the available boiler load and other plant conditions. The list of the actual parameters and the test ranges is shown in Table 5.1.

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Table 5.1 Parameters and ranges tested

PARAMETER	TEST RANGE
Coal quality	Sulfur content varied from 1.4% to 2.8%. The average sulfur content was approximately 2.25%.
Limestone quality	Two qualities were tested. Coarse quality proved to be less effective and fine limestone was selected as the optimum sorbent.
Boiler Load	Boiler load ranged from 40MW to 62MW, the unit's nameplate size is 60MW.
Ca/S	Ca/S molar ratios between 1.0 and 3.0 were evaluated based on efficiency and economy.
High level injection setting Low level injection setting Right side injection setting Left side injection setting Middle position injection setting	Ports 1, 2, 3, 4, 5, 6 in use Ports 7, 8, 9, 10, 11, 12 in use Ports 4, 5, 6, 10, 11, 12 in use Ports 1, 2, 3,7, 8, 9 in use Ports 2, 3, 4, 5, 10, 11 in use (See Appendix 2 for Port Locations)
Injection Nozzle Angle	The angle of injection nozzles can be directed up to +/- 15° from horizontal, but only 0 was tested.
Booster Air	Booster air tests were not done.
dT	The temperature difference between the reactor flue gas outlet temperature and the saturation temperature of flue gas. Temperatures between 4 and 11 °F above saturation were tested.
Water droplet size and nozzle position	Droplet size and nozzle position were tested during the early stages of parametric testing. Droplet size is proprietary information.
Ash recycling Ratio	The recycling ratio is the mass flow rate of recycled ash versus the mass flow rate of boiler ash. Ash recycling ratio varied from zero to maximum. The maximum ratio equates to nearly 4-5 tons per hour.

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5.3 Injection Tests

5.3.1 Objective

The objective of injection testing was to find the most effective method for injecting limestone into the furnace under various boiler operating conditions. Another important goal of this testing was to assess the impact of limestone injection into the host boiler. After injection tests were performed, the optimum injection settings were selected for each boiler operating condition.

5.3.2 Implementation

The order of injection test implementation was determined according to the available boiler load and other plant conditions. The LIFAC process was in operation twenty-four hours per day.

Flue gas humidification was engaged during most of the limestone injection tests. Since limestone injection increases inlet solids loading to the ESP, it is important to humidify during injection.

The following were evaluated during limestone injection tests:

- Impact on gaseous emissions
- Measurement verification
- SO_2 after boiler/reactor
- O_2 after boiler/reactor
- Impact on boiler performance
- Steam generation, pressure, and temperature
- Superheater spray water flow
- Sootblowing frequency
- Furnace temperature at injection level
- Temperature before and after economizer
- Pressure drop across the economizer
- Temperature before and after air preheater
- Pressure drop across the air preheater
- ID fan motor current and speed

5.4 Activation Tests

5.4.1 Objective

Activation tests were performed to study the desulfurization efficiency and operability of the activation reactor at a range of boiler loads, Ca/S molar ratios, injection settings, flue gas outlet temperatures and ash recycling ratios.

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5.4.2 Implementation

The order of implementation was determined according to the available boiler load and other plant conditions. Activation reactor parameters take more time to stabilize than injection parameters. Also, reactor SO_2 readings have longer adjustment periods than boiler SO_2 reduction readings.

Activation testing provided an assessment of the following:

- Impacts on gaseous emissions
- Measurement verifications
- SO_2 after boiler/reactor
- O_2 after boiler/reactor
- Fouling of the activation reactor internals
- Operability and efficiency of the steam reheat system
- Pressure and temperature measurements before and after the reactor and reheaters
- ID fan motor current and speed

5.5 Ash Recycling Tests

5.5.1 Objective

The objective of recycling testing was to find the recycling ratio which optimized desulfurization efficiency, ESP operation, and recycling equipment performance.

5.5.2 Implementation

The mass flow rate of unreacted sorbent can be varied by changing the speed of the rotary feeders under the ESP hoppers. If the feed rate is too high, the hoppers will quickly empty and cause recycle to be unsteady. Tests were performed to determine the maximum possible speed of the rotary feeders. The recycle rate achieved by this maximum feeder speed was less than half of the desired recycle rate. Richmond Power & Light emptied the ESP hoppers twice per day to avoid high ash levels and to be sure that the system is operational. Recycle ash flow was unsteady for short periods during ESP ash pulling.

The order of implementation was determined according to the available boiler load and other plant conditions. The mass flow ratio between recycled ash and ash generated in the boiler is called recycle ratio. Two recycling ratios were

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tested. Zero and maximum recycle. The maximum ash recycle ratio was limited due to ESP hopper size and quantity of fly ash collected by the ESP. According to results based on the parametric tests, one recycling ratio will be used for the remainder of the test program.

Recycle tests were performed to evaluate:

- Impact on SO_2 reduction
- Operability of the reactor and recycle system
- Impact on the ESP, particulate emissions and opacity

5.6 Limestone tests

The main objective of testing different qualities of limestone was to evaluate their effects on SO_2 removal efficiency, the operability of the LIFAC system, and impacts on the host plant. Another objective was to evaluate the economics of different sorbent types. Process economy will be reported in the long-term test report. The order of test implementation was determined according to the available boiler load and other plant conditions.

Baseline limestone was a finely pulverized (85 % < 325 mesh, 44 μm), sedimentary, high-calcium limestone. The second was a coarse (85 % < 200 mesh, 74 μm), sedimentary type, high-calcium limestone.

Limestone tests were performed to evaluate:

- SO_2 removal efficiency in the furnace
- Evaluation of the ESP performance and operability
- Particulate measurements before/after the activation reactor and after the ESP
- Effects of limestone quality on plant operation

5.7 Coal tests

The plant's combustion coal came from two different sources. The majority of the coal (contract coal) came from southwest Indiana. This coal has a high sulfur content ranging from 2 to 3 w%. The low sulfur coal (spot coal) was shipped from Kentucky and Ohio and contains 1.3 to 2.0 w% sulfur. Normally this low sulfur coal is burned during the weekends.

Low sulfur coal testing was performed whenever low sulfur coal was shipped to the plant. Coal was constant (2.2 to 2.3 w%) for most of the testing. This constant quality coal was a special blend of several qualities which had the desired sulfur content. The necessary amount of constant quality coal was piled both at the mine and at the powerplant. During

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short period tests, special arrangements were not made for coal blending if the coal source and sulfur content were known.

5.8 Deviation from the original parametric test plan

All tests were performed according to available boiler and powerplant conditions. Since the powerplant has two boiler units, it was possible to keep Unit No.2 at a constant load during a test. Additional power is continually purchased to keep up with the city's demand. Both boilers operate at capacity during the daytime, Monday through Friday. Low load periods occurred at night and on weekends.

Most of the process parameters were evaluated at high boiler loads. The ID fan on Unit No.2 operates at maximum rpm during peak (65MW) load. A variable frequency drive was installed as part of the LIFAC process to increase the efficiency of this fan. With the VFD in operation, LIFAC can be engaged during high boiler loads. However, the VFD failed several times, tripping Unit No.2 on each occurrence. The VFD was in service for only five days of the LIFAC demonstration. All tests were performed with the VFD off-line and the reactor by-pass damper 15% open to reduce the system pressure drop.

It was also discovered during parametric testing that ESP performance was deficient if the inlet flue gas temperature was less than 200°F. Since the LIFAC steam reheat system could not maintain exiting flue gas temperatures greater than 200°F, the by-pass damper was adjusted accordingly. It was possible to operate the LIFAC system under normal conditions with the bypass damper slightly open. Also, SO_2 reduction measurements are still accurate since it is measured before the bypass point. However, this bypassed flue gas contains an unknown amount of unreacted lime, thus changing the actual Ca/S molar ratio in the reactor.

Most ESP evaluations were conducted by EPRI and SRI during the course of the demonstration. Due to unexpected ESP performance during startup and shutdown, these evaluations were more extensive than originally planned.

It was originally planned to test five different sorbents qualities. However, only two limestone qualities were evaluated. A fine (85% < 325M) and coarse (85% < 200M) were injected during the demonstration. These sorbents were shipped from quarries close to RP&L. Dolomitic limestone, which contains 40 to 50% $MgCO_3$, was not tested due to poor accessibility and a high required mass flow rate of injection to the furnace. Crystalline limestone is unavailable in this part of the country. Hydrated lime was previously tested as part of the LIMB project at RP&L. During these tests, extensive fouling was experienced and continuous sootblowing was required. There is information available on the effects of injecting hydrated lime at RP&L's Unit No.2.

Booster air was not tested for three reasons. With the VFD inoperational, the ID fan operated at capacity and additional flue gas volume may have been detrimental. Second, the booster air fan was extremely loud. Lastly, injection SO_2 reduction was already improved with the addition of the upper injection ports and nozzles.

A Ca/S molar ratio of 2.0 was used for most of the demonstration. It was originally planned to use a ratio closer to 2.5. However, a high limestone flow rate causes increased particulate emissions and boiler fouling. Since a target of greater than 70% reduction was established, this reduced Ca/S ratio was sufficient and was employed for most of the testing. Sorbent cost is approximately half of the total operating cost of the system. Hence, decreased limestone flow improves process economy.

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6.0 INSTRUMENTATION AND MEASUREMENTS

6.1 Process Control and Data Collection System

The operation of the LIFAC process is controlled by the PLC (Programmable Logic Controller) system. The process control system is used only for the LIFAC equipment. The operation of the boiler equipment and associated subsystems is controlled by the power plant's original control system and some of the measurements are only monitored by LIFAC's monitoring computers.

The process control system hardware consists of two identically installed IBM 756 Industrial computers with 19" VGA color monitors. Both computers are located in the boiler control room. One unit is needed to control the entire LIFAC process. The second unit is identical and serves as a spare unit. A printer for alarm reporting is connected to this computer. Both computers are intended to monitor and collect measured data.

All the commands from the control computer to the equipment, feedback information, and measured data from the instruments are transported through three input/output (I/O) racks. Rack #1 is for the boiler house equipment and instruments, rack #2 is for the limestone area equipment and instruments, and rack #3 is for the activation reactor area equipment and instruments. The system has a total of 329 input or output points: 29 RTD's, 51 analog inputs, 34 analog outputs, 160 digital inputs and 58 digital outputs.

6.2 Procedure

The impact of the LIFAC process on gaseous and particulate emissions was determined by measuring sulfur dioxide (SO_2), sulfur trioxide (SO_3), oxygen (O_2), nitrogen oxide (NO), opacity and particulate content of flue gas. During the test program, SO_2 and O_2 content of flue gas and opacity were measured continuously. Effects of the LIFAC process on NO and SO_3 emissions and on ash resistivity were determined by short periodic tests. Particulate emission measurements were performed periodically in connection with ESP testing. The emission measurement locations are shown on in Figure 6.1. The emission measurement procedures are then summarized in Appendix 6.1.

6.3 SO_2 Measurements

For continuous SO_2 monitoring two in-situ Lear Siegler SM8100 microprocessor-based gas analyzers were used. As a back-up analyzer there was an extractive system Monitor-Labs analyzer down stream of the reactor prior to the ESP. The Monitor Labs analyzer uses a diluting probe. One pair of SO_2 and O_2 analyzers were located after the air preheater before the activation reactor inlet. The second pair of analyzers were

located after flue gas reheating and before the ESP inlet. The equipment utilizes a second derivative spectroscopic measurement technique, measuring the narrow band absorption of ultraviolet energy by SO_2 molecules. The measurement was performed on moist gas and the output is corrected to standard temperature and pressure. The maximum measurement range of the equipment is 3000 ppm and was calibrated for 0 to 3000 ppm gas range monitoring. A detailed description of the SO_2 analyzer is presented in Appendix 6.1.

Baseline flue gas sulfur dioxide content ranged from 1000 to 3000 ppm, depending on the coal quality. A typical value of 1750 ppm corresponds to a 2.5% sulfur coal. With the LIFAC process in operation, the SO_2 content varied from approximately 260 to 900 ppm.

The physical conditions where SO_2 is monitored are demanding. In the sampling location before the activation reactor, the flue gas contains a large quantity of solids, typically 1,800 x 10^{-6} lb/SCF (29,000 mg/Nm^3). Down stream of the reactor, the humidity of flue gas is typically 15% and particulate loading is typically 2,500 x 10^{-6} lb/SCF (40,000 mg/Nm^3). Loading after the reactor is higher due to the recycle load.

The minimum temperature of the flue gas after the reactor was approximately 190°F. At these low temperatures the Lear Siegler analyzer after the reactor was unreliable. Low temperatures cause ash buildup in the analyzer probe. The Monitor Labs SO_2 analyzer probe was equipped with a heater and there was no ash buildup in that probe. Ash buildup will partially absorb SO_2, hampering the output of the analyzer.

The ductwork is fitted with flanges for in-situ analyzers at both SO_2 measuring locations. Grid measurements for excess O_2 content in the flue gas will be performed to verify the consistency of SO_2 and O_2 concentration patterns across the duct sections.

The reliability of the in-situ SO_2 analyzers will be verified at times during the test program with a portable analyzer. The verification measurements were performed under baseline conditions at the beginning of baseline testing and during post-LIFAC tests. To verify the reliability of SO_2 measurements in the dusty conditions before the reactor, and in the dusty and high moisture conditions after the reactor, verification measurements were also taken at the beginning of injection and activation testing. Verification was performed anytime during testing if there was any doubt about the accuracy of the measurements. Indication of an unreliable reading was obtained if the reading of the two stationary analyzers were not equal when the reactor was not in operation.

Throughout the test program, the analyzer readings were compared to the calculated sulfur dioxide content of flue gas

based on the sulfur content in the coal.

6.4 SO_3 Measurements

SO_3 emissions were monitored using EPA reference Method 8. The description of this method is presented in Appendix II. SO_3 content was measured by subcontractors who performed ESP efficiency and performance measurements.

SO_3 emissions are typically based on SO_2 emissions since they are generally on the order of 0.5 to 1.0% of the flue gas SO_x content. In earlier studies at RP&L, baseline SO_3 emissions have been nearly 18 ppm during combustion of high sulfur coal (SO_2 4000 ppm).

Sorbent injection is expected to remove all or most of the SO_3 normally present in the flue gas, since SO_3 reacts rapidly with calcium oxide formed during the calcium carbonate decomposition reaction. SO_3 content was measured after the ESP. All SO_3 measurements were coupled with ESP performance testing.

6.5 O_2 Measurements

O_2 measurements were performed using two in-situ Lear Siegler Dynatron 401 oxygen monitoring systems. The equipment uses a solid-state zirconium oxide fuel cell sensor to measure oxygen content of flue gas. Measurements were made on a moist, total sample basis using the Nernst relationship. The maximum measurement range of the equipment is from 0 to 25%. A detailed description of the O_2 analyzer is presented in Appendix 6.2. O_2 analyzers were installed in the flue gas duct in the same locations as the SO_2 analyzers before and after the activation reactor.

Typical flue gas oxygen content after the boiler was from 2 to 5 weight %. The O_2 measurements will be carried out in the same demanding conditions as the SO_2 measurements. However, the moist ash layer on the probe does not effect to O_2 measurement as long as the gas can penetrate the ash layer and filter.

The continuous oxygen measurement reading was used for correcting the results of SO_2 measurements to a 6% O_2 content. Verification of the O_2 analyzers were used to measure concentration patterns of flue gas in selected monitoring locations.

6.6 NO Measurements

For NO measurements, the same two in-situ Lear Siegler analyzers were used. Dual exit slits on the monochromator of the analyzer allow two separate wave lengths: one corresponding to SO_2 and the other to NO, to impinge on the detector. Sequential time separation of the SO_2 and NO

signals is accomplished with a small shutter in the monochromator. This shutter moves back and forth between the two beams at one minute intervals, providing one minute sequential SO_2 and NO concentration outputs.

Baseline NO content of flue gas was measured during the baseline tests. During parametric testing, NO levels were measured by the subcontractor who performed stack testing as part of the Environmental Monitoring Program.

6.7 Particulate Measurements

Particulate emissions were monitored using EPA reference Method 17 and Method 5. The description of these methods are described in Appendix II.

The particulate load measuring locations are in the activation reactor and ESP inlet ducts. Emissions were measured in the duct breaching between the Unit No.2 ID fan and the stack.

After the boiler and before the ESP, typical baseline particulate content of flue gas is 650 x 10^{-6} lb/SCF. Limestone injection into the furnace increases this particulate load to between 1,300 x 10^{-6} and 3,500 x 10^{-6} lb/SCF. The variation depends mainly on the Ca/S molar ratio and the limestone quality used for injection. In the activation reactor the flue gas is humidified and a portion of the particulates are removed from the bottom of the reactor. After the reactor and before the ESP, the particulate content of the flue gas varies from 700 x 10^{-6} to 1,900 x 10^{-6} lb/SCF without ash recycling. With ash recycling in operation, the particulate load may increase to nearly 3,600 x 10^{-6} lb/SCF.

The efficiency of the ESP increases with LIFAC in operation due to an increase in inlet solids loading with the same outlet emissions. During baseline operating conditions, the ESP efficiency varies from 98 to 98.4%. With the LIFAC process, it was estimated to increase to 99.3%. Particulate emissions are typically 10-12 x 10^{-6} lb/SCF without LIFAC. Ash recycling may increase particulate load to 29 x 10^{-6} lb/SCF for short periods of time.

The purpose of the particulate measurements was to evaluate the efficiency of the ESP under different conditions and evaluate the ash separation capabilities of the activation reactor. A list of particulate measurements are shown in Table 1. All particulate measurements were collected in accordance with EPA Method 17.

6.8 Opacity

Opacity of the combined flue gas from both Unit 1 and Unit 2 was measured by an in-situ instrument located in the stack. The opacity limit was 40% in the beginning of the project. During the demonstration, the plant received a new 30% opacity

limit.

6.9 Moisture

The measurement of flue gas moisture content was done simultaneously with particulate measurements. EPA reference Method 17 was used to collect this data. The description of this method is shown in Appendix II.

6.10 ESP Performance

ESP performance measurements were performed periodically during the entire test program. The final measurements will be taken during long-term testing and reported in the long-term test report.

Ash resistivity was measured with an in-situ, point-to-plane resistivity probe system. The resistivity probe electrostatically collects a layer of ash while flue gas conditions are maintained at constant values. Collection of the ash is accomplished by applying a high voltage across the point-to-plane electrodes in the probe. At a sufficient voltage, a highly compressed electric field or corona forms in the vicinity of the point (negative electrode) and ionizes gas molecules in this region. Dust particles in the gas are then charged by the negative ions formed in the corona and are attracted by electrical forces onto the collection plate (positive electrode). This method of ash collection is intended to simulate the behavior of ash in an actual ESP.

From earlier studies at RP&L, the average resistivity of fly ash was 5.9×10^{10} ohm-cm, as determined by the V-I method.

Ash resistivity measurements were performed during optimization tests, with and without the LIFAC process in operation. This testing will only be done during 65-MW boiler loads. The measurement is taken in the ductwork before the ESP.

6.11 Boiler Performance

To optimize limestone injection into the furnace and evaluate its impact on boiler performance, certain boiler operating conditions were measured. These conditions are listed in Table 1. Measurements were performed using on-line instruments which send information to the process control computer for handling and processing.

The coal feed rate measurement was used to calculate flue gas volume formed during coal combustion and the amount of sulfur entering the furnace. The coal feed rate was measured with a mechanical counter in the coal weigh feeder buckets. To verify the accuracy of these measurements, the weigh feeders were calibrated at the beginning of the baseline tests. During the tests, continuous verification will be made by

reactor, in the reactor, after each reheater, and before the ESP. The pressure measurements are used to determine LIFAC induced pressure changes in the system and predict ash buildups in the activation reactor. Individual pressure gauges measure the total pressure of the flue gas. Pressure differences are calculated by the PLCs.

6.14 Flow Measurements

The average flue gas flow rate before the activation reactor was measured by pitot tube-type flow elements. Three flue gas traverse probes consisting of multiple flow sensors mounted on self-averaging signal manifolds are positioned on an equal area basis to the duct. The flow element system is furnished with an instrument air purge system to protect the internal mechanism of the transmitter from the migration of airborne particulates.

The flue gas flow measurement gives an early signal to the humidification water control system to increase or decrease the water flow to match an increase or decrease in the flue gas flow.

Humidification water flow is measured by an orifice plate and controlled by a motor-operated control valve. The quantity of humidification water sprayed into the activation reactor is controlled by flue gas temperature measurements after the reactor. Air flow to the dual fluid nozzles is controlled by controlling the air pressure with a motor-operated control valve.

Steam flow to the reheaters is controlled by the flue gas temperature ahead of the ESP. A motor-operated control valve controls steam flow. The reheating steam is verified by calculations from the flue gas temperature increase measurement or by counting the discharge cycles of the condensate return tank.

6.15 Limestone and Recycle Ash Feed Measurements

The limestone and recycle ash feed measurements are divided into three categories: silo measurements before the silo shut-off valve; feeder measurements from the silo discharge opening to pneumatic conveying line; and pneumatic conveying system measurements. The feeders and silo were used to measure the limestone feed rate into the furnace.

The limestone quantity in the feed and storage silos were determined by weigh cell measurements. Both the feed and storage silos are equipped with two weigh cells that measure the load bearing columns' compression and calculate the weight. The storage capacity of the feed silo is 270,000 lbs. and the storage silo is 530,000 lbs. The ESP ash surge tank has two weigh cells and a storage capacity of 20,000 lb.

The feed silo has two level switches for low and high levels. The limestone storage silo has three level switches for low, high, and high/high levels. The ESP ash surge tank has one high level switch.

Limestone mass flow to the boiler injection nozzles is measured by a gravimetric weigh feeder system which consists of two screws and a rotary feeder arranged in series. The rotary feeder controls the mass flow from the silo to the first screw feeder, and both are connected to the same variable speed controller. The screw feeder is used to ensure smooth flow of limestone to the second weigh screw which measures the mass flow. The weigh screw is mounted on load cells, operates at a constant speed, and determines the feed rate by measuring the weight of the screw multiplied by screw speed. The limestone flow measurement will be verified by comparing it to limestone feed silo weight measurements.

The ESP ash recycling ratio is measured and adjusted by two ESP hopper rotary feeders. The rotary feeders are installed in the bottom of the first two parallel ESP hoppers and feed ESP ash to the pneumatic conveying system for recycling. Ash flow for recycling from the ESP ash surge tank is measured and controlled by variable speed drive rotary feeders under the tank. Reactor ash flow to the recycling system from the reactor bottom conveyor is measured and controlled by a variable speed drive screw feeder. The ESP ash flow measurements were to be verified by comparing them with the ESP ash surge tank weight measurements.

The LIFAC process contains four pneumatic transport air blowers: two for the limestone injection into the furnace and two for ash recycling. In addition, the limestone injection system is furnished with a booster air blower to increase the initial velocity of the limestone particles injected into the boiler. All of the limestone transport air, booster air, ESP ash recycling air, and recycling mix transport air blowers are operated at constant speeds.

Limestone transport air flow is measured using an orifice plate. The pressure and temperature of the air was also monitored. The transport air flow (1,060 SCFM) is controlled with a pneumatic control valve. Booster air flow rate was approximately 6,360 SCFM.

In addition, pressures of the air from the ESP ash recycling air flow were measured. By measuring pressure changes in these flows, potential blockages in the piping and nozzles could be identified.

7.0 SAMPLING AND ANALYSIS

7.1 General

Coal, limestone, and ash samples were taken manually during the demonstration. Aqueous samples were also monitored as part of the Environmental Monitoring Plan (EMP). Sampling locations are shown in Figure 7.1. While, sampling methods are presented in Table 7.2.

Table 7.1 Sampling Frequency, Method and Location

SAMPLE	SAMPLING FREQUENCY	METHOD	LOCATION
Coal	Once per hour or three times per hour during special tests	Grab sample	Coal pulverizer feeder
Limestone	Periodic	Grab sample	Feeding silo discharge
ESP ash	Periodic	Grab sample	ESP hopper
Reactor ash	Periodic	Grab sample	Truck hopper at reactor ash discharge
Air preheater ash	During EMP sampling	Grab sample	Air preheater hoppers

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7.2 Coal

Coal samples were taken from the feeder belt before each of the pulverizers. Coal was grabbed once per hour from each feeder and a composite sample was formed for a three hour period. Samples were shipped in double layer plastic bags to a laboratory for analysis.

Special one-hour coal samples were collected during stack testing. One hour samples were also taken during baseline and calibration periods.

Coal Analyses

Since the sulfur content of the feed coal dictates the limestone feed rate, it was the most significant coal quality parameter studied during the demonstration. Sulfur analyses were also used to calculate the baseline SO_2 content of the flue gas. During the demonstration, feed coal sulfur content ranged between 1.6 and 2.8%.

Most analyses were performed by one laboratory. However, other labs were often utilized to ensure reliable results. The scope of coal analyses included: short proximate, ultimate, and mineral analysis.

Mineral analysis included components as listed below:
Also see Appendix 7.1 for typical coal analysis.

Silicon dioxide (SiO2)	Base/Acid ratio
Aluminium oxide (Al2O3)	Slag viscosity
Titanium dioxide (TiO2)	Fouling index
Calcium oxide (CaO)	Slagging index
Potassium oxide (K2O)	
Magnesium oxide (MgO)	
Sodium oxide (NaO2)	
Phosphorous pentoxide (P2O5)	
Ferric oxide (Fe2O3)	
Sulfur trioxide (SO3)	

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7.3 Limestone Sampling

The limestone supplier was obligated to monitor their product's quality once per hour and the data was made available to the LIFAC test team. Therefore, only periodic verification samples and analyses were performed.

Limestone Analyses

The most significant limestone parameters are the calcium carbonate ($CaCO_3$) content and particle size. Additional information was also provided by the supplier and is shown in Table 7.3.1.

Table 7.3.1Limestone analysis components of the fine quality (80%<325 mesh) limestone

COMPONENT	TYPICAL CONTENT IN WEIGH %
Na2O	0.12
MgO	1.50
Al2O3	0.68
SiO2	2.87
SO3	0.14
K2O	0.16
CaO	51.86
TiO2	0.05
Fe2O3	0.35
LOI	42.27
TOTAL	**100.00**
Calcium and Magnesium Oxides	
CaCO3	**92.6**
MgCO3	3.15

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7.4 Ash Sampling

Ash samples were grabbed from four different locations during the demonstration and are listed below:

Economizer Ash - collected from the economizer hoppers;

ESP Ash - collected from the ESP hoppers; ESP ash samples were taken from the first two parallel hoppers;

Reactor Ash - collected from the activation reactor bottom conveyor;

Boiler Bottom Ash - collected from dewatering bin (Unit No.2 only);

During baseline and post-LIFAC tests, fly ash samples were

collected every eight hours. During parametric and long-term test periods, ESP and reactor ash samples were collected during each individual test.

Ash Analyses

During the tests in which particulate measurements were performed, all ash samples were analyzed for calcium compounds, elemental analysis, and particle size distribution. A resistivity analysis of ESP fly-ash was performed on samples collected during on-line ash resistivity measurements.

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8.0 DATA COLLECTION AND HANDLING

8.1 Data Collection

Test data was acquired in many different ways, depending on the type of data and required procedures. Test data for LIFAC was for the most part, recorded automatically by the process control computer. Necessary additional information was recorded manually to a logbook and log sheets either in the control room or in the plant.

Data was recorded from the boiler, activation reactor, electrostatic precipitator, and all input and output streams.

A test coordinator kept the log book where all pertinent information of the testing was recorded, such as injection settings and recycling methods. See Appendices 8.1 to 8.3 for typical manual data collection sheets, Log book page, and ESP data sheet.

Particulate emission measurement data was recorded on forms shown in the EPA Method 17 description in Appendix II. All other manually measured data was recorded on specific forms. Samples requiring laboratory analysis were recorded on laboratory supplied forms.

The process control system (PLC - Programmable Logic Controller) with Process Monitoring System (PMS) software was used to collect and handle the automatically measured data. The process computers read measured values from the instrumentation every third second, typically calculates three-minute averages, and saves them for further use in the computer's hard drive. Data from the PMS was converted into a format that can be imported to a Lotus 1-2-3 spreadsheet. Data will be transferred via disk to a personal computer on-site to perform on-site data processing.

8.2 Calculations

The Ca/S molar ratio is a factor which reflects the efficiency and economy of the LIFAC desulfurization process. It is calculated using equation (1). In equation (1) limestone mass flow m(ls) is measured by the limestone feeder. $CaCO_3$ content of the limestone is found in the limestone analyses and is constant if limestone comes from the same location. Coal flow is measured at the coal feeders. Coal sulfur content S is also found through analyses. Estimated values of these parameters were stored in the process monitoring computer, but the corrected values were used for adjustments during data handling procedures on a spread sheet program.

Ca/S molar ratio

$$Ca/S = \frac{m(ls) \times p(ls) \times 100.09}{m(coal) \times s \times 32.09} \qquad (1)$$

where

Ca/Sis a ratio of number of moles of calcium injected in to the furnace vs. number of moles of sulfur burned in the furnace

m(ls)is limestone mass flow (#/min)
p(ls)is limestone $CaCO_3$ content (weight %)
100.09is molecular weigh of $CaCO_3$
m(coal)is coal flow (#/min)
32.09is molecular weigh of sulfur

Variations of boiler reactor operation conditions cause variations in flue gas composition such as SO_2 and O_2 levels. Different measuring techniques of in-situ and extractive type of verification SO_2 analyzers cause variations in the measured results as well. In order to compare different SO_2 measurements, the measured values were corrected by the constant O_2 content and the moisture level of flue gas. Sulfur dioxide content in the flue gas is corrected to a 6% O_2 content. Correction equations are shown below. O_2 correction is in equation 2, moisture correction is in equation (3).

$$SO_2(6\%) = SO_2(m) \frac{21 - 6}{21 - O_2(m)} \qquad (2)$$

where

$SO_2(6\%)$is a sulfur dioxide content in the flue gas corrected to 6% O_2 content (ppm = parts per million)

$SO_2(m)$is a sulfur dioxide content of the flue gas in actual flue gas O_2 content (ppm = parts per million)

2121% O_2 content in the air

$O_2(m)$is measured O_2 content at the same place where SO_2 is measured

Flue gas moisture correction calculations were performed because SO_2 before and after the reactor is measured in moist flue gas conditions.

window. The upper injection level proved to be 2 to 3% points more efficient during high loads. This is probably due to extreme temperatures at the lower level. Figure 9.2.2 shows the results of injection nozzle combination testing. Overall, SO_2 reduction rates seemed to improve when the upper injection level was utilized. Figure 9.2.3 shows SO_2 reduction in the furnace only.

Injection nozzle heads can be tilted 15° vertically. However, all tests were performed with the nozzles in a horizontal position.

Booster air for limestone injection was tested only briefly because it increased loading on the ID fan which was already operating at full capacity. The secondary air fan also exceeded noise levels in the boilerhouse. Additional noise protection or soundproofing would have been required to continue to operate the fan. Therefore, no results of booster air's effect on SO_2 capture were obtained.

9.3 Limestone Quality

Two different qualities of limestone were tested. One was a coarse ground 85% < 200 mesh and the other was a fine ground 85% < 325 mesh. Both had high calcium carbonate content ranging from 90 to 95 %. These sedimentary limestones were shipped from two different quarries. The coarse limestone was shipped from southwest Indiana and the fine came from northern Kentucky.

Fine limestone proved to significantly improve process efficiency as compared to the coarse limestone. Figures 9.3.1 and 9.3.2 show the results of limestone quality testing at both high and low boiler loads. Fine limestone increases process efficiency 15 to 20 % points. This quality was injected during most of the parametric and long-term tests. Despite of large variation of results the trend is evident: smaller particle size provides more active surface area for reduction reactions to take place.

9.4 Boiler Load

The nameplate load of Unit No.2 is 60 MW. However, the unit can be operated as high as 65 MW. During parametric testing, the load ranged from 40 to 61 MW. Most parametric tests were performed at a high boiler load due to high power demands. Boiler load was decreased during weekends and overnight. Due to the unavailability of low load results at similar operating conditions, a comparison of boiler load effects on SO_2 was not possible. Load was held constant during parameter measurements by adjusting the non-LIFAC boiler (Unit No.1) to suit power demand.

Boiler load effect on SO_2 reduction was not as significant as expected. Unit No.2 is small for its capacity and burns hot.

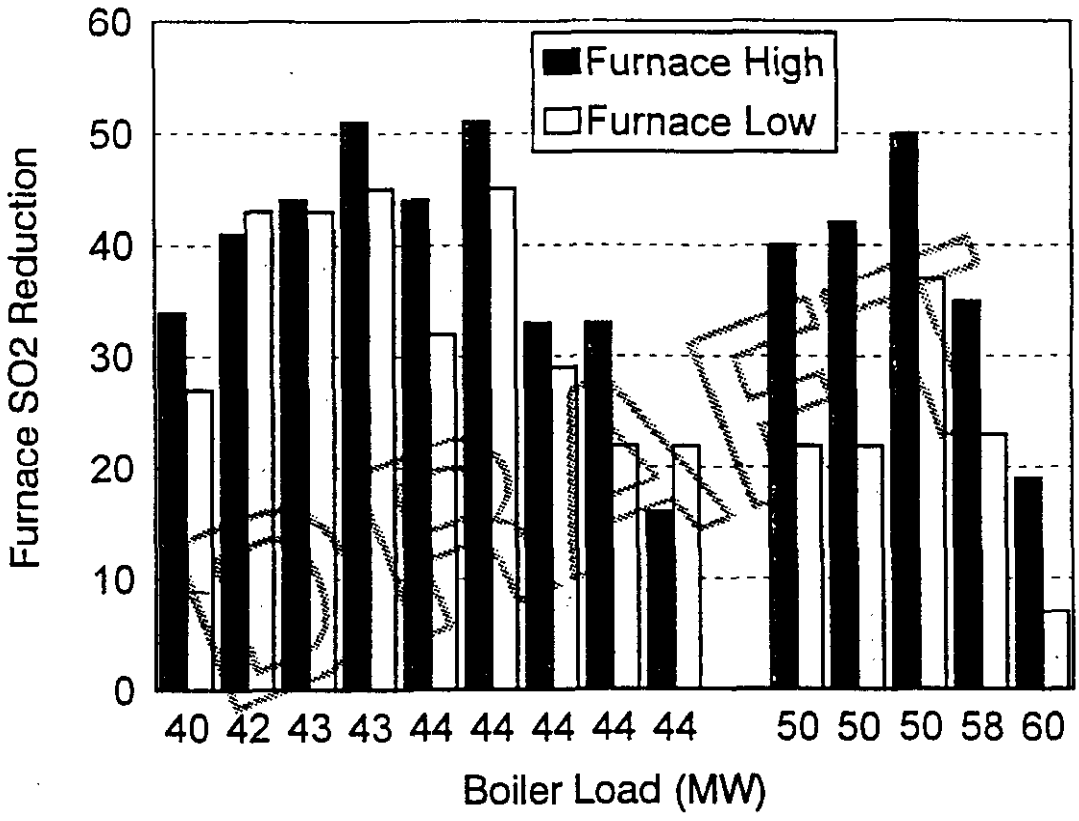

Figure 9-2-3

Effects of Limestone Injection Setting on Furnace SO_2 Reduction

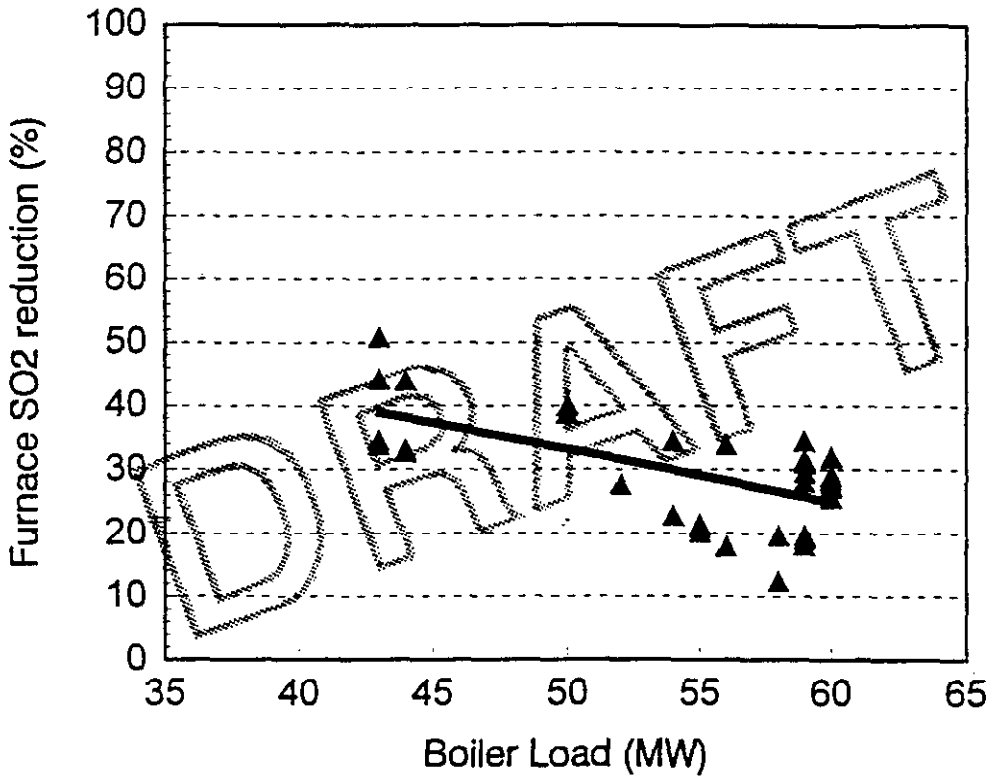

Figure 9-4-1

SO_2 Reduction in the Furnace Versus Boiler Load

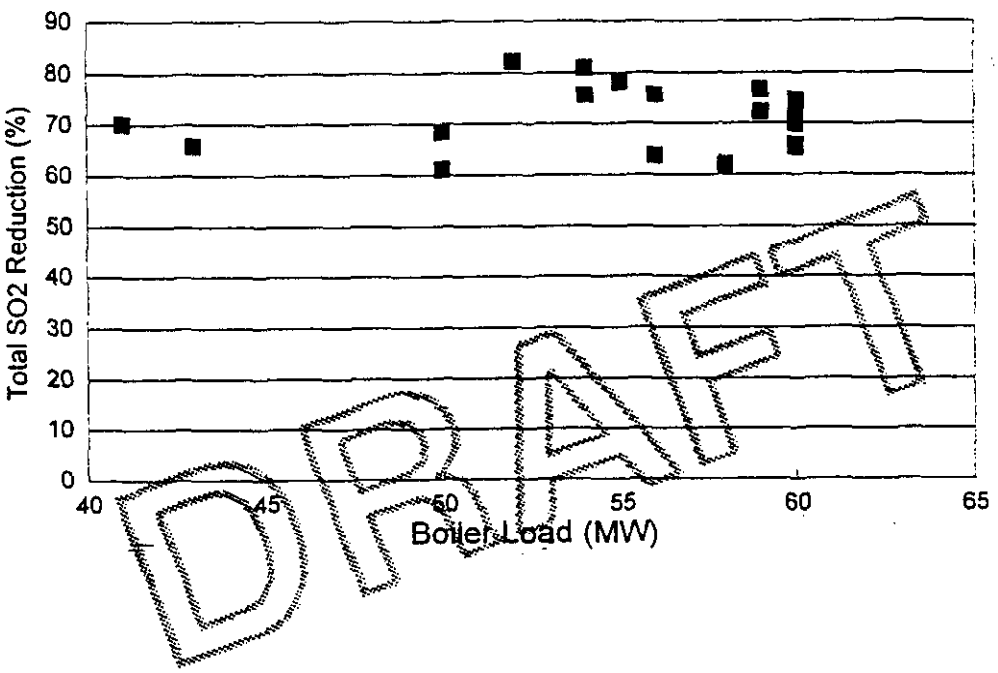

Figure 9-4-2

Total SO_2 Reduction Versus Boiler Load

Low loads should have improved efficiency. Instead, the difference between high and low load efficiency is unclear. The effect of boiler load will be studied again during long-term testing if low load periods are available. SO_2 capture in furnace and reactor at a range of boiler loads are shown in Figures 9.4.1 and 9.4.2.

9.5 Coal Quality

Test coal was shipped from three locations in three different states. Small amounts of low sulfur coal came from Kentucky and Ohio. The majority of the test coal (high sulfur) was shipped from southwest Indiana.

Sulfur content of the coal often changed during testing and caused slight inaccuracies in the results. Coal samples were grabbed once per hour from each of three feeders to get a representative samples of the test coal quality.

Sulfur content varied between 1.4 and 2.8 wt% sulfur. It was not possible to evaluate the effects of coal sulfur content on SO_2 capture since other conditions were not stable during testing. Figure 9.5.1. shows the sulfur content of the coal during the parametric tests.

9.6 Ca/S Molar Ratio

Limestone flow rate to the furnace is determined by a calcium/sulfur molar ratio. The ratio is calculated from the coal sulfur content and limestone quality, along with their respective flow rates. The calculation is described in Section 8.3. Ca/S ratios ranged between 1.0 and 3.0 during the parametric test period. The Ca/S molar ratio setpoint was maintained at 2.0 for most of the testing. Baseline coal sulfur content was determined from the sulfur dioxide concentration in the flue gas. This coal sulfur setpoint was entered into the PMS and remained unchanged until the next baseline adjustment and calibration period. The actual ratio varied according to boiler load and was, at times, inaccurate if the sulfur content of the combustion coal changed during operation. The actual ratio was later calculated from the lab analyses of the coal and incorporated into data analysis.

Ca/S molar ratio parameter testing was performed with a range of boiler loads, two different limestone qualities, and with recycle on and off. The effects of limestone quality and ash recycling are reported in the respective sections. Under all test conditions, sulfur dioxide capture increases as the Ca/S molar ratio increases. Figure 9.6.1 illustrates the effect of the Ca/S ratio with reduced boiler loads. According to previous LIFAC testing, Ca/S ratios higher than 3.0 slightly improve process efficiency, but degrade the economy of the system. The optimum settings were determined to be between 1.5 and 2.5 which result in a 60 to 75% SO_2 capture. Figure 9.6.2 shows the effects of Ca/S molar ratios at higher boiler loads.

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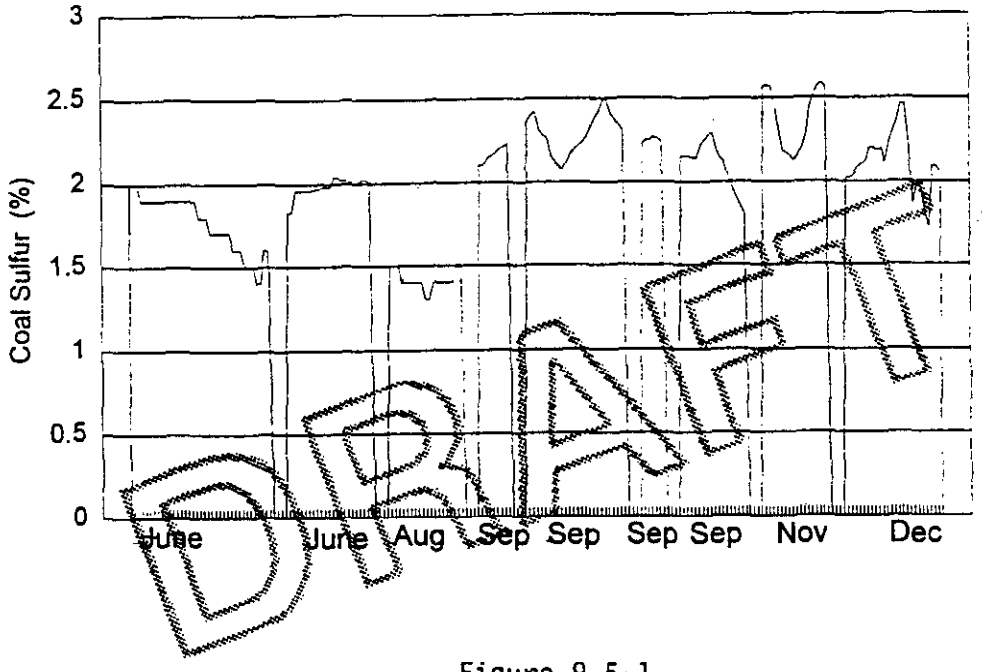

Figure 9-5-1

Sulfur Content of the Combustion Coal
During Parametric Testing

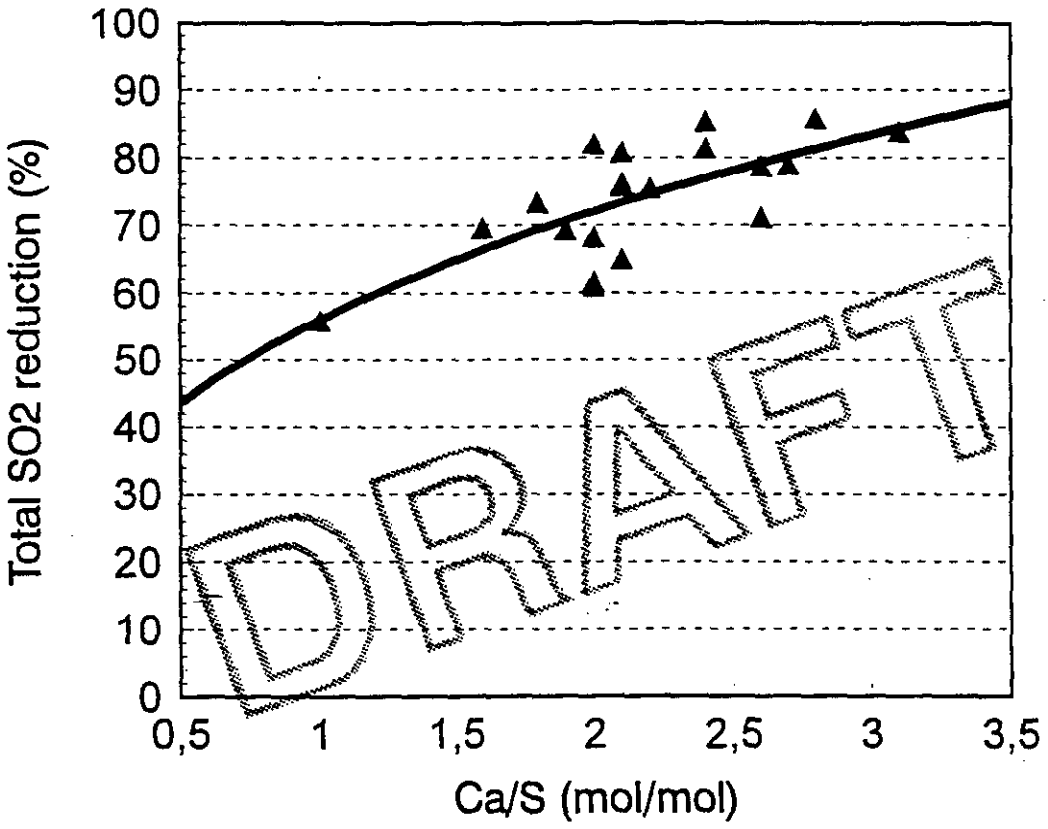

Figure 9-6-1
Total SO2 Reduction Versus Ca/S Molar Ratio
During Low Boiler Load

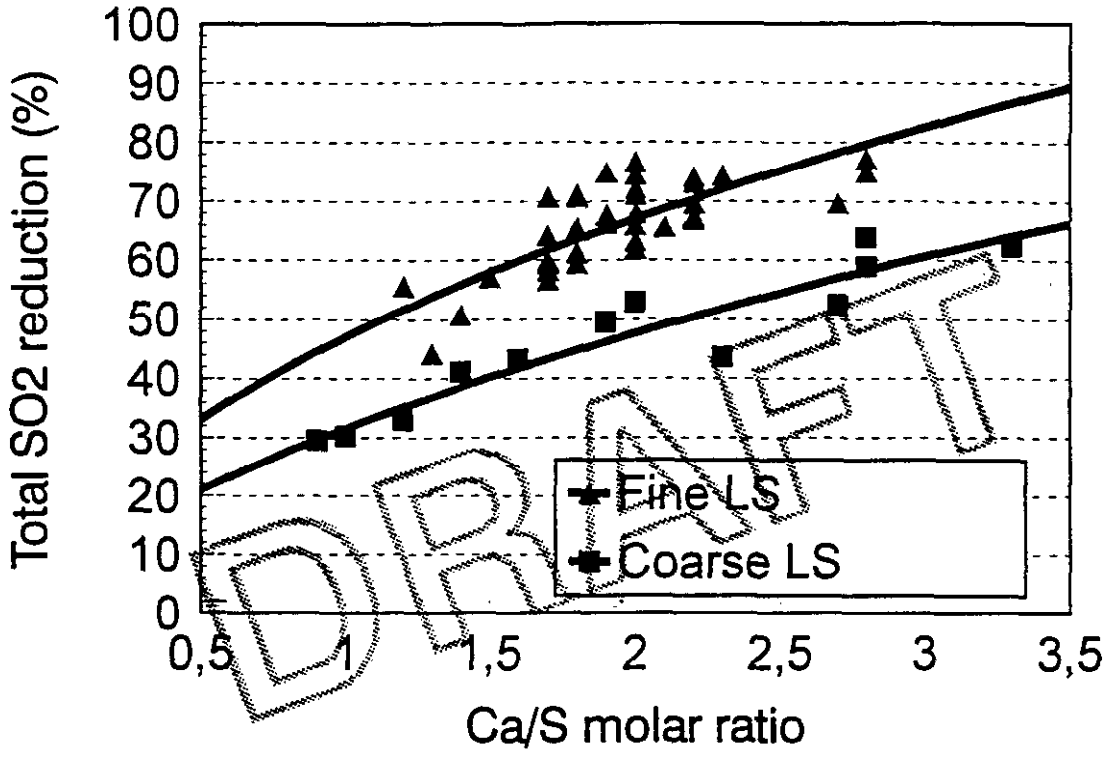

Figure 9-6-2
Total SO2 Reduction Versus Ca/S Molar Ratio
During High Boiler Load

9.7 Activation Reactor Outlet Temperature

Total SO_2 capture improves as reactor bottom temperature approaches the flue gas saturation temperature. It is necessary to keep the reactor outlet temperature above the flue gas saturation temperature (~126°F). If the exiting temperature is too low, the reactor bottom ash will become too moist. The ash handling system is designed to transport only dry ash. With the improved reactor design at Richmond Power & Light, a closer approach to saturation was possible. The design value was 11°F above saturation. Parametric tests were performed with temperatures as low as 3°F. Even at these low temperatures the reactor bottom ash was dry and easy to handle and dispose at a landfill. Typical bottom ash consisted of between 5 and 20% moisture content.

9.8 Humidification Nozzle Arrangement

It is possible to move each of the seven nozzle headers up and down. In the beginning of the tests the headers were at their lowest position. During the tests the headers were lifted to a higher position. There was a danger that ash would accumulate on the reactor's internal sealing if the nozzles were too high. The effects of the nozzle positions on sulfur capture was not documented. It is expected that a higher header position is beneficial because the mixing of the flue gas and water should improve and result in better SO2 capture.

9.9 Water droplet size

Atomizing air pressure controls humidification water droplet size in the reactor. Air pressures between 38 and 65 psig were tested. The goal is to efficiently evaporate the droplets during their residence in the reactor. With low air pressure forming large water droplets, complete evaporation may not take place. This will create a wet ash byproduct. Conversely, smaller water droplets may evaporate too quickly and decrease SO_2 capture efficiency. Figure __ shows the effects of atomizing air pressure, or water droplet size.

9.10 Ash Recycling

Fly ash containing unreacted calcium is recycled from two ESP hoppers to the reactor inlet duct. The quantity of the recycled material was limited because the reactor separated more ash than expected. All ash collected in the bottom of the reactor was disposed at a landfill.

The quantity of ash recycled was limited by the availability of fly ash collected in the ESP hoppers. Ash from the first two ESP hoppers was recycled back to reactor. If ash recycle quantity was increased, the hoppers emptied and the process became unsteady.

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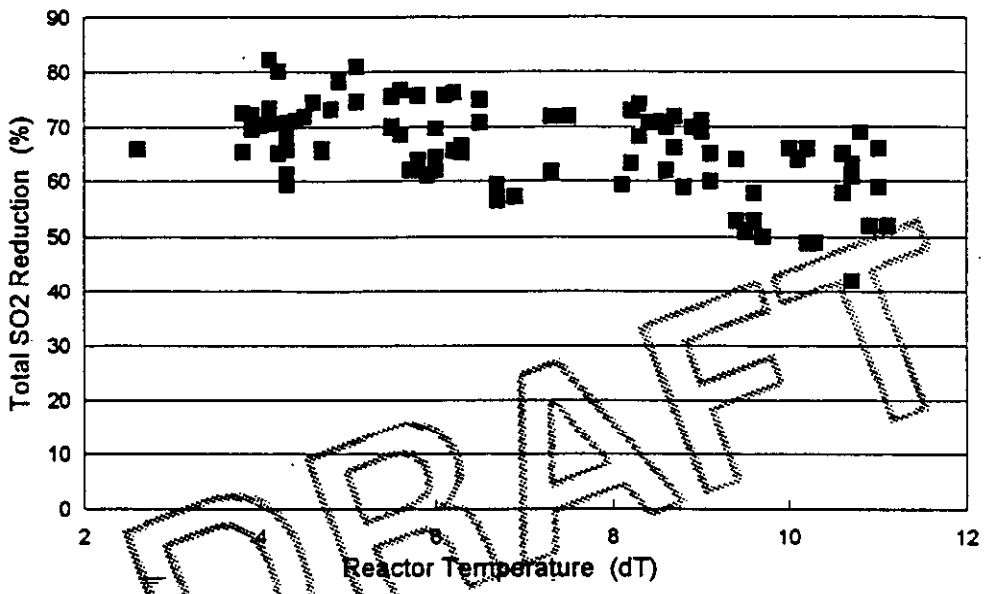

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Figure 9-7-1

Reactor Bottom Approach to Saturation Temperature and Its Effect on Process Efficiency

Ash recycling improved SO_2 reduction between 15 and 20% points. Increased recycling would foster improved SO_2 reduction. However, with the current installation arrangement, only small quantities of ash could be recycled. Ash recycling results, with high and low boiler loads, are shown in Figures 9.10.1 and 9.10.2.

9.11 Data Summary

Data was carefully selected from periods which represented stable process operation conditions and provided reliable results. Data representing certain parameter combinations and operating conditions were selected from a summary sheet by eliminating non eligible test points. Trends and curves were generated using these selected data points. An example of summary and parameter test data sheets are shown in Appendices __ and __.

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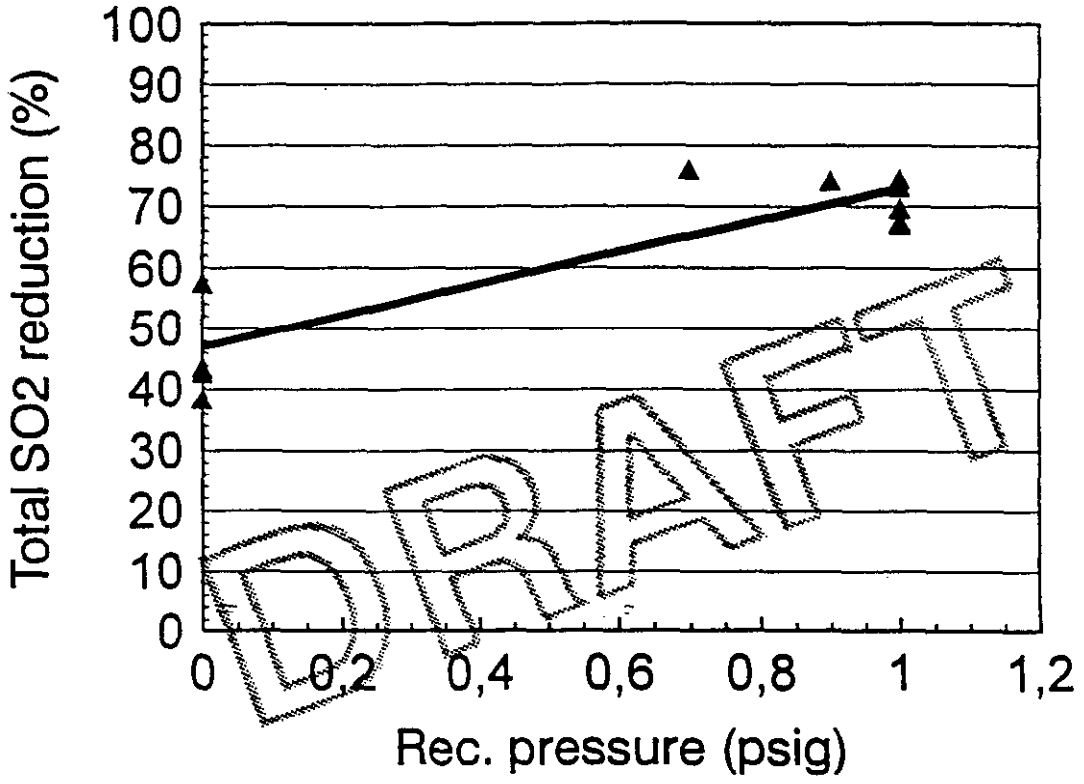

Figure 9-10-1
Unspent Sorbent Recycling Effect on Total SO_2 Capture
During High Boiler Load

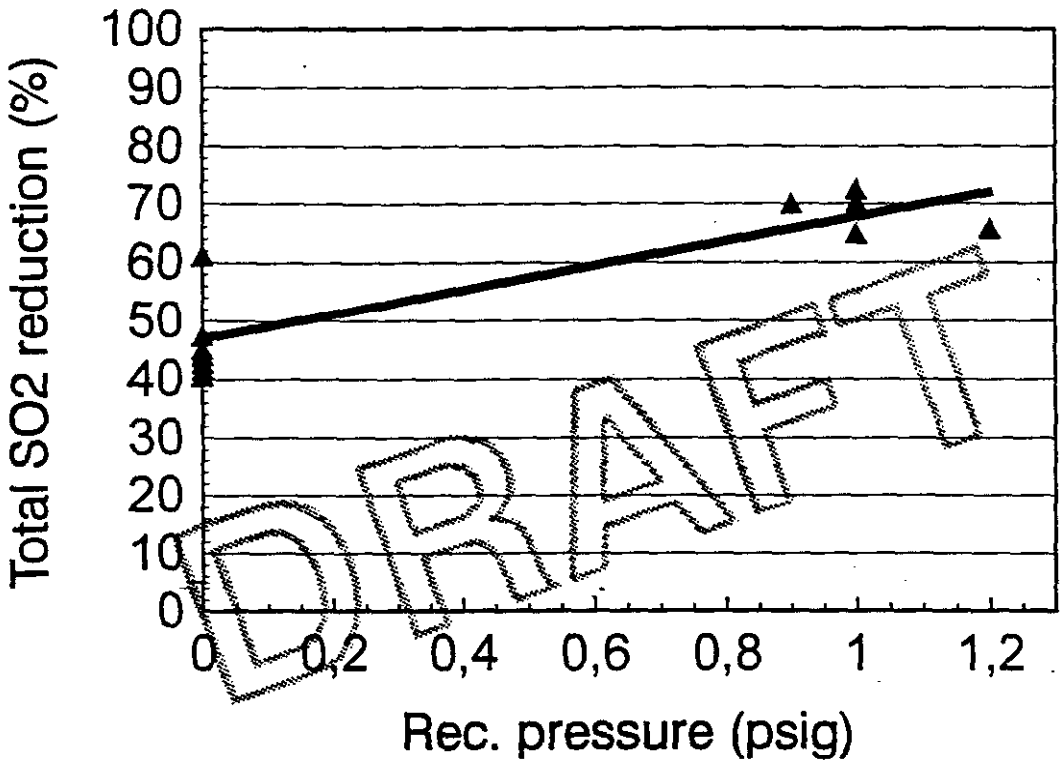

Figure 9-10-2
Unspent Sorbent Recycling Effect on Total SO_2 Capture
During High Boiler Load

10 CONCLUSIONS

10.1 Settings for Optimization and Long term tests

Parametric test results were used to establish process settings for the remainder of the test program. Selection criteria was based on optimum sulfur dioxide capture and overall system performance.

Combustion coal burned during long-term testing will have a sulfur content of nearly 2.25%. This is the quality that RP&L has been burning during the last few years. A higher coal sulfur content would require a higher limestone feedrate if the Ca/S ratio is maintained at 2.0.

A Ca/S molar ratio of 2.0 was selected as the optimum limestone flow setting. This ratio yields the highest sulfur capture without interfering with process or boiler operability. Higher Ca/S molar ratios would increase boiler superheater fouling, thus the sootblowing sequence would need to be increased. Smaller Ca/S molar ratios may be more economical. Process economy will be studied during long-term testing.

Two different limestone qualities were tested. Fine quality (80%< 325 mesh) was significantly better than the coarse quality (80%> 200 mesh) and will be used during long-term testing. Sulfur capture was 20% points higher when fine quality limestone was employed. Sootblowing frequency was greater during injection of fine limestone. Coarse limestone was more abrasive on limestone transport hoses. Coarse quality was less expensive partly because of the grinding cost.

The most important parameter for effective sulfur capture in the reactor is the reactor bottom temperature. During long-term tests the reactor bottom temperature needs to be as low as possible. Lower temperatures induce a higher sulfur capture in the reactor. The temperature setpoint was limited to 9°F above saturation temperature because the reheat steam coils in the exiting ductwork may become plugged. Without ash recycling, the reactor temperature setpoint was limited to 14°F above the saturation temperature.

Humidification nozzles will be situated at their second highest position. In this position the internal sealing remains intact and the humidification effect and gas mixing is ideal.

Atomizing air pressure will be maintained at 50 psig. This pressure generates an optimum humidification water droplet size. If droplets are too large, they will not completely evaporate during their residence in the reactor. Hence, the ash exiting the reactor will be moist. Smaller droplets may evaporate too fast and reduce sulfur capture in the reactor.

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Flue gas leaving the reactor will be reheated with both the steam heat exchanger coils and with the hot gas by-pass system. During initial ESP tests, it was discovered that ESP performance improved if the ESP temperature was above 195°F. Both reheater systems are needed to maintain this temperature. Further ESP tests will be performed during long-term testing with lower operating temperatures.

Ash recycling should be as high as possible. The quantity of recycled ash is limited by the availability from the ESP hoppers. Ash from the first two ESP hoppers is recycled back to reactor. If the ash quantity is increased, the hoppers may empty and the process will not be stable.

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