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Technical Support Document:

CHEMICAL RECOVERY COMBUSTION SOURCES AT KRAFT AND SODA PULP MILLS

Emission Standards Division

U. S. ENVIRONMENTAL PROTECTION AGENCY
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LIST OF ACRONYMS AND UNITS OF MEASURE

a.c.	alternating current
acfm	actual cubic foot (feet) per minute
acmm	actual cubic meter(s) per minute
ADMBP/d	air-dried megagram(s) of bleached pulp per day
ADMP	air-dried megagram(s) ton of pulp
ADMUP/d	air-dried megagram(s) of unbleached pulp per day
ADTBP/d	air-dried ton(s) of bleached pulp per day
ADTP	air-dried ton(s) of pulp
ADTUP/d	air-dried ton(s) of unbleached pulp per day
APCD	air pollution control device
As	arsenic
Be	beryllium
BLO	black liquor oxidation
BLRBAC	Black Liquor Recovery Boiler Advisory Committee
Btu/lb	British thermal unit per pound
$Ca(OH)_2$	calcium hydroxide
CAA	Clean Air Act as amended in 1990
$CaCO_3$	calcium carbonate
CaO	calcium oxide
Cd	cadmium
CEM	continuous emission monitor
CFR	Code of Federal Regulations
cm	centimeter(s)
Co	cobalt
CO	carbon monoxide
CO_2	carbon dioxide
Cr	chromium
CRF	capital recovery factor
CS_2	carbon disulfide
d/yr	day(s) per year
d.c.	direct current
DAC	direct annual cost(s)
DAS	data acquisition system
DCE	direct contact evaporator
DMF	N,N-dimethylformamide
EMTIC	Emission Measurement Technical Information Center
EPA	U. S. Environmental Protection Agency
ESP	electrostatic precipitator
FID	flame ionization detector
ft	foot (feet)
ft^2/1,000 acfm	square foot (feet) per 1,000 actual cubic feet per minute
ft^2	square foot (feet)
FTIR	fourier transform infrared
g/L	gram(s) per liter
g/dscm	gram(s) per dry standard cubic meter
gal	gallon(s)
gal/acf	gallon(s) per actual cubic foot
gal/yr	gallon(s) per year
GC	gas chromatography
GFC	gas filter correlation
gpm	gallon(s) per minute
gr/dscf	grain(s) per dry standard cubic foot

H_2O	water
H_2S	hydrogen sulfide
HAP	hazardous air pollutant
HCl	hydrochloric acid
Hg	mercury
HOMER	hazardous organic mass emission rate
hp	horsepower
hr/d	hour(s) per day
HSCSST	heated summa canister source sampling train
HVLC	high-volume, low-concentration
HWI	hazardous waste incinerator
IAC	indirect annual cost(s)
IMS	ion mobility spectroscopy
in.	inch(es)
in. H_2O	inch(es) of water
ITAC	incremental total annual cost(s)
KCl	potassium chloride
kg BLS/d	kilogram(s) of black liquor solids per day
kg/ADMP	kilogram(s) per air-dried megagram of pulp
kg/Mg	kilogram(s) per megagram
kg PM/MWh	kilogram(s) of PM per megawatt-hour
kJ/kg	kilojoule(s) per kilogram
kW	kilowatt(s)
kWh	kilowatt-hour(s)
L/min	liter(s) per minute
L	liter(s)
L/yr	liter(s) per year
L/acm	liter(s) per actual cubic meter
L/G	liquid-to-gas
lb PM/MM Btu	pound(s) of PM per million Btu
lb BLS/d	pound(s) of black liquor solids per day
lb/gal	pound(s) per gallon
lb/ADTP	pound(s) per air-dried ton of pulp
lb/ton	pound(s) per ton
LOQ	limit of quantitation
LVHC	low-volume, high-concentration
m	meter(s)
$m^2/(m^3/sec)$	square meter(s) per cubic meter per second
m^2	square meter(s)
m^3	cubic meter(s)
m^3/sec	cubic meter(s) per second
MACT	maximum available control technology
MEE	multiple-effect evaporators
mg	milligram(s)
Mg/yr	megagram(s) per year
min	minute(s)
MJ/yr	megajoule(s) per year
ml	milliliter(s)
mm Hg	millimeter(s) of mercury
MM Btu/yr	million Btu per year
MM	million(s)
Mn	manganese
MSD	mass selective detector
MST	Methanol Sampling Train
MWI	medical waste incinerator
Na_2CO_3	sodium carbonate

Na_2S	sodium sulfide
$Na_2S_2O_3$	sodium thiosulfate
Na_2SO_4	sodium sulfate
NaCl	sodium chloride
NaOH	sodium hydroxide
NCASI	National Council of the Paper Industry for Air and Stream Improvement, Inc.
NCG	noncondensible gases
NDCE	nondirect contact evaporator
NESHAP	national emission standards for hazardous air pollutants
ng	nanogram(s)
Ni	nickel
NO_x	nitrogen oxides
NSPS	new source performance standards
NSSC	neutral sulfite semichemical
O_2	oxygen
OAQPS	Office of Air Quality Planning and Standards
Pb	lead
PCDD/PCDF	polychlorinated dibenzo-p-dioxins and dibenzofurans
PM	particulate matter
ppb	part(s) per billion
ppm	part(s) per million
ppmdv	part(s) per million dry volume
ppmv	part(s) per million by volume
QA/QC	quality assurance/quality control
RRF	relative response factor
Sb	antimony
SCA	specific collecting area
SDT	smelt dissolving tank
Se	selenium
sec	second(s)
SIE	specific ion electrodes
SO_2	sulfur dioxide
T-R	transformer-rectifier
TAC	total annual costs
TCI	total capital investment
TDS	total dissolved solids
THC	total hydrocarbon
Tl	thallium
ton/yr	ton(s) per year
TPIEC	Texas Paper Industry Environmental Committee
TRS	total reduced sulfur
TSD	technical support document
μm	micrometer(s)
UV	ultraviolet
VOC	volatile organic compound

1.0 INTRODUCTION

National emission standards for hazardous air pollutants (NESHAP) are under development for combustion processes in the chemical recovery area of kraft and soda pulp mills under authority of Section 112(d) of the Clean Air Act (CAA) as amended in 1990. This technical support document (TSD) provides technical information and analyses used in the development of this NESHAP. Effluent guideline limitations and a separate NESHAP covering pulp and paper manufacturing processes are being developed concurrently under the authority of the Clean Water Act and the CAA, respectively; the technical information and analyses for these regulations are found in separate documents. The U. S. Environmental Protection Agency (EPA) is coordinating these efforts to produce air and water regulations for the pulp and paper industry. The remainder of this chapter describes the scope and organization of this TSD.

1.1 SCOPE OF THE TECHNICAL SUPPORT DOCUMENT

This TSD presents the technical information and the analyses used in the development of the NESHAP for combustion processes in the chemical recovery area of kraft and soda pulp mills. The chemical recovery combustion processes included in this NESHAP are (1) recovery furnaces (including associated smelt dissolving tanks and black liquor oxidation systems) and (2) lime kilns.

1.2 ORGANIZATION OF THE TECHNICAL SUPPORT DOCUMENT

This TSD is organized into six chapters. Chapter 2.0 describes the processes and equipment associated with the chemical recovery cycle at kraft and soda pulp mills, identifies the hazardous air pollutants (HAP's) emitted and the emission

sources, and quantifies baseline HAP emissions. Chapter 3.0 describes the emission control techniques that can be used to reduce HAP emissions from chemical recovery combustion processes, including add-on control systems and equipment changes. Chapter 4.0 presents model plants and control and enhanced monitoring options for chemical recovery combustion processes. Chapter 5.0 discusses the environmental and energy impacts on the model process units from the application of the HAP control options presented in Chapter 4.0. Finally, Chapter 6.0 presents the model process unit control and enhanced monitoring costs for the application of the control options. Appendix A lists the evolution of this technical support document. Appendix B summarizes the available HAP emissions data and discusses the test methods and monitoring methods that could be used to demonstrate compliance with proposed standards for chemical recovery combustion sources at kraft and soda pulp mills.

2.0 COMBUSTION PROCESSES IN THE KRAFT PULP INDUSTRY

This chapter describes the combustion processes and equipment associated with the chemical recovery cycle at kraft and soda pulp mills, as well as the HAP's emitted from these processes and equipment. The chemical recovery process includes four sources of HAP emissions that are the focus of this document. These sources are (1) chemical recovery furnaces (including direct contact evaporator [DCE] recovery furnaces and nondirect contact evaporator [NDCE] recovery furnaces), (2) smelt dissolving tanks (SDT's), (3) black liquor oxidation (BLO) systems, and (4) lime kilns. Section 2.1 provides an overview of the kraft chemical recovery process. More detailed information on the combustion processes and equipment is included in Section 2.2. Section 2.3 presents existing Federal and State regulations affecting kraft pulp mills, process emission factors, and baseline HAP emissions. The references to this chapter are provided in Section 2.4.

2.1 OVERVIEW OF CHEMICAL RECOVERY PROCESSES

This section provides an overview of the chemical recovery cycle at kraft and soda pulp mills. References 1 and 2 listed in Section 2.4 provide more detailed descriptions of the chemical recovery cycle.[1,2]

There are approximately 122 kraft pulp mills currently operating in the United States.[3] There are only two soda mills currently operating in the United States.[4] One soda mill is located in Pennsylvania and the other is located in Tennessee.[4] Figure 2-1 shows the number of kraft and soda pulp mills located in each State. As shown in Figure 2-1, the majority (52 percent) of kraft pulp mills are located in the southeastern United

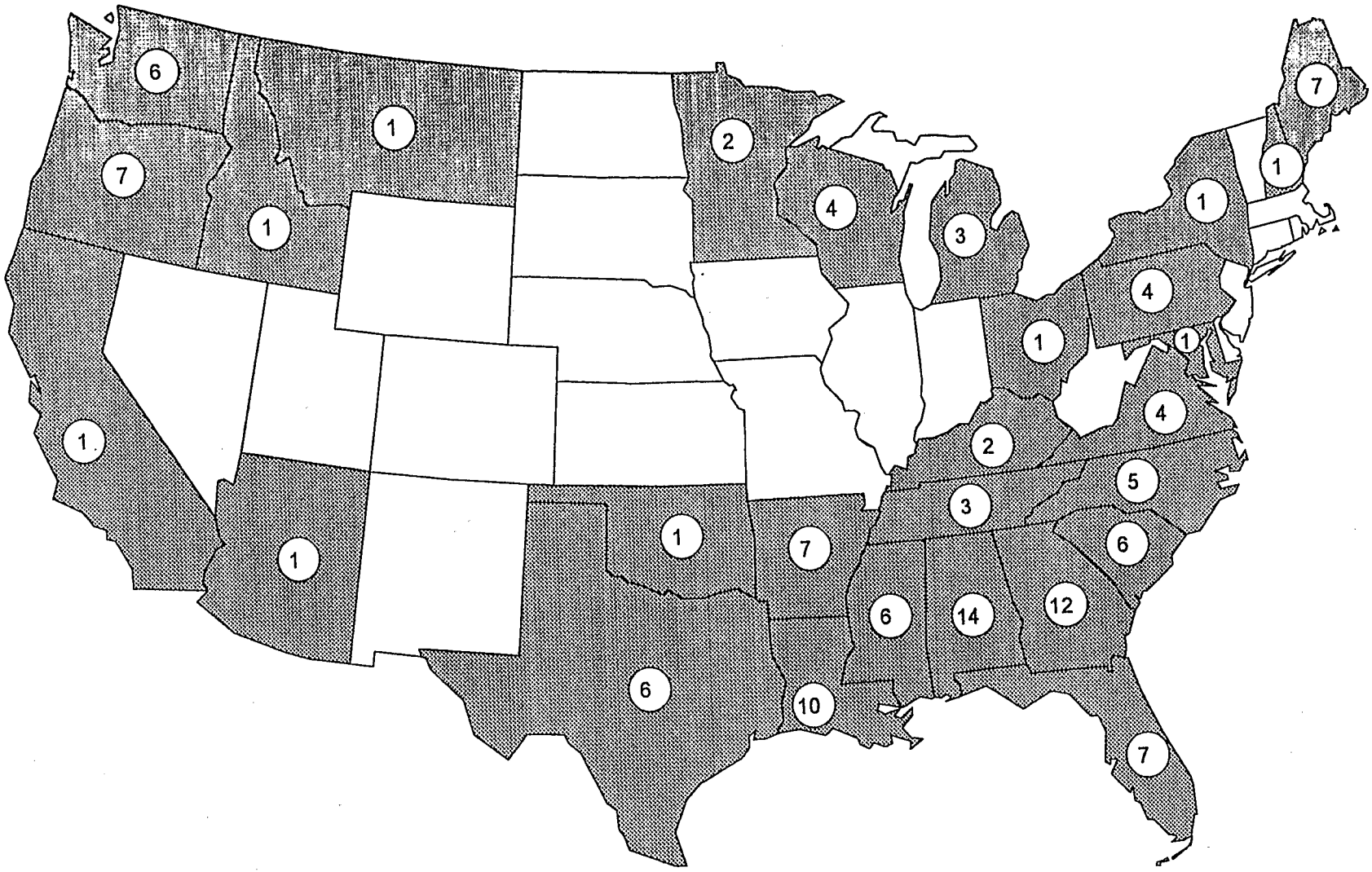

Figure 2-1. Distribution of U.S. kraft and soda pulp mills.

States. The kraft process is the dominant pulping process in the United States, accounting for approximately 85 percent of all domestic pulp production.[5] The soda pulping process is similar to the kraft process, except that soda pulping is a nonsulfur process. One reason why the kraft process dominates the paper industry is because of the ability of the kraft chemical recovery process to recover approximately 95 percent of the pulping chemicals and at the same time produce energy in the form of steam. Other reasons for the dominance of the kraft process include its ability to handle a wide variety of wood species and the superior strength of its pulp.

The production of kraft and soda paper products from wood can be divided into three process areas: (1) pulping of wood chips, (2) chemical recovery, and (3) product forming (includes bleaching). The relationship of the chemical recovery cycle to the pulping and product forming processes is shown in Figure 2-2. Process flow diagrams of the chemical recovery area at kraft and soda pulp mills are shown in Figures 2-3 and 2-4, respectively. Kraft and soda chemical recovery processes are similar, except that the soda process does not require BLO systems, which are installed to reduce total reduced sulfur (TRS) emissions from DCE recovery furnaces. Both kraft and soda mills have chemical recovery furnaces, SDT's, and lime kilns. This document focuses on the four primary emission sources in the chemical recovery area. The emission sources covered in this document are the BLO system (kraft process only), the recovery furnace, the SDT, and the lime kiln. These emission sources are shaded in both process flow diagrams. Air emissions from these sources are expected to be similar for both kraft and soda processes, with the exception of TRS emissions. The soda process does not utilize sulfur-containing compounds; therefore, no TRS emissions are generated by soda mills.

Because of the similarities between the kraft and soda process and the predominant use of the kraft process, the information presented in the remainder of this document focuses on the kraft chemical recovery process.

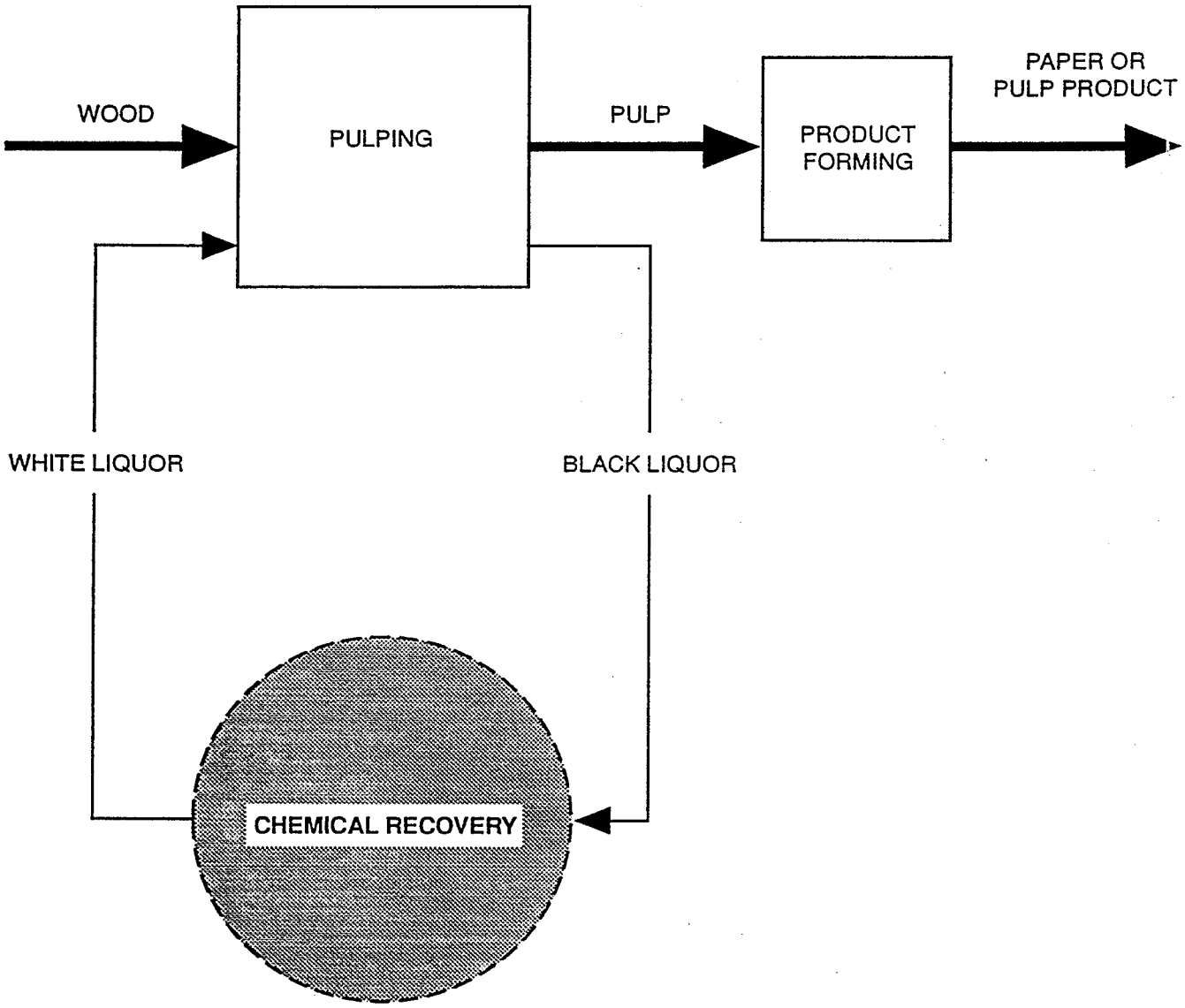

Figure 2-2. Relationship of the chemical recovery cycle to the pulping and product forming processes.

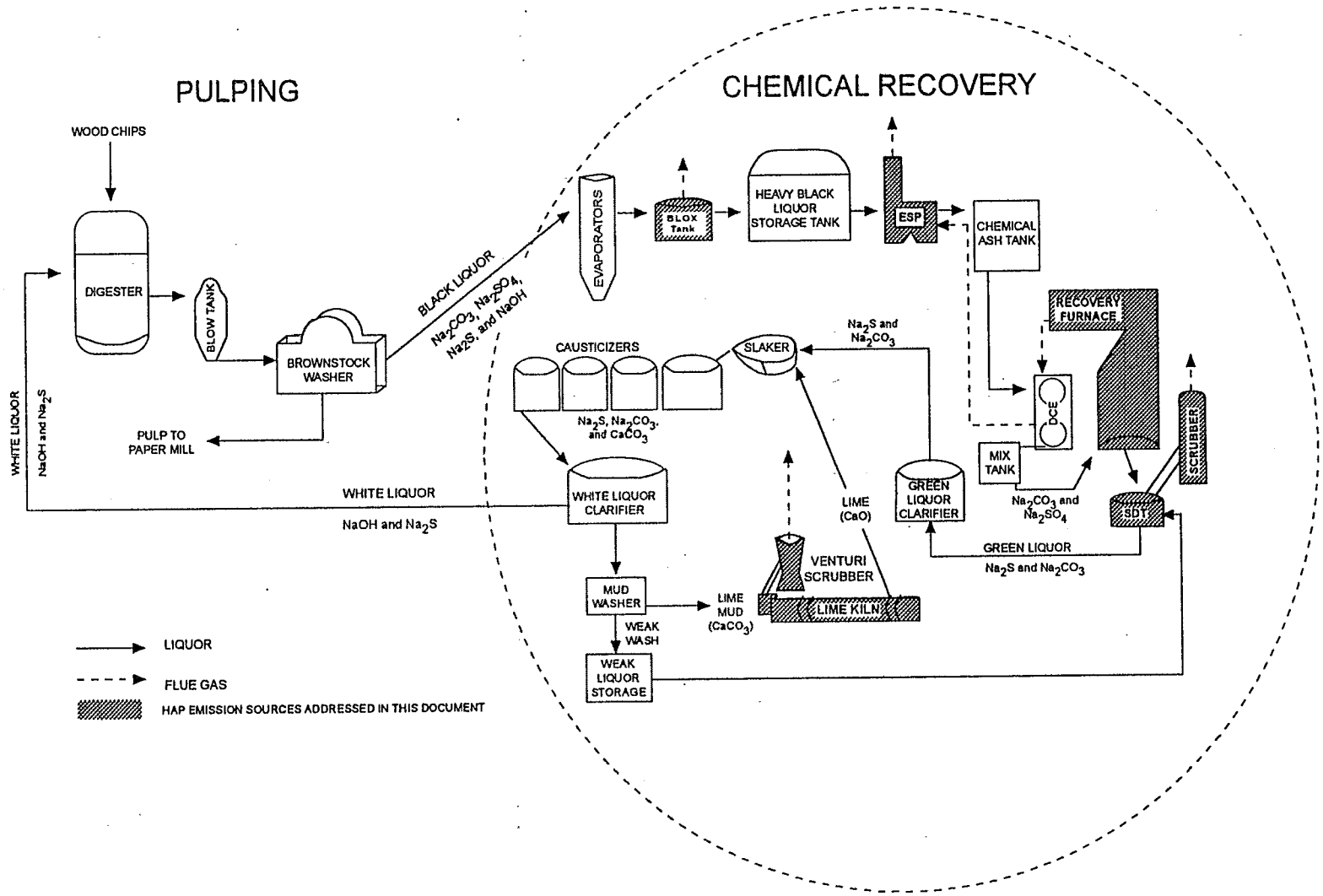

Figure 2-3. Kraft process--chemical recovery area (DCE recovery furnace).

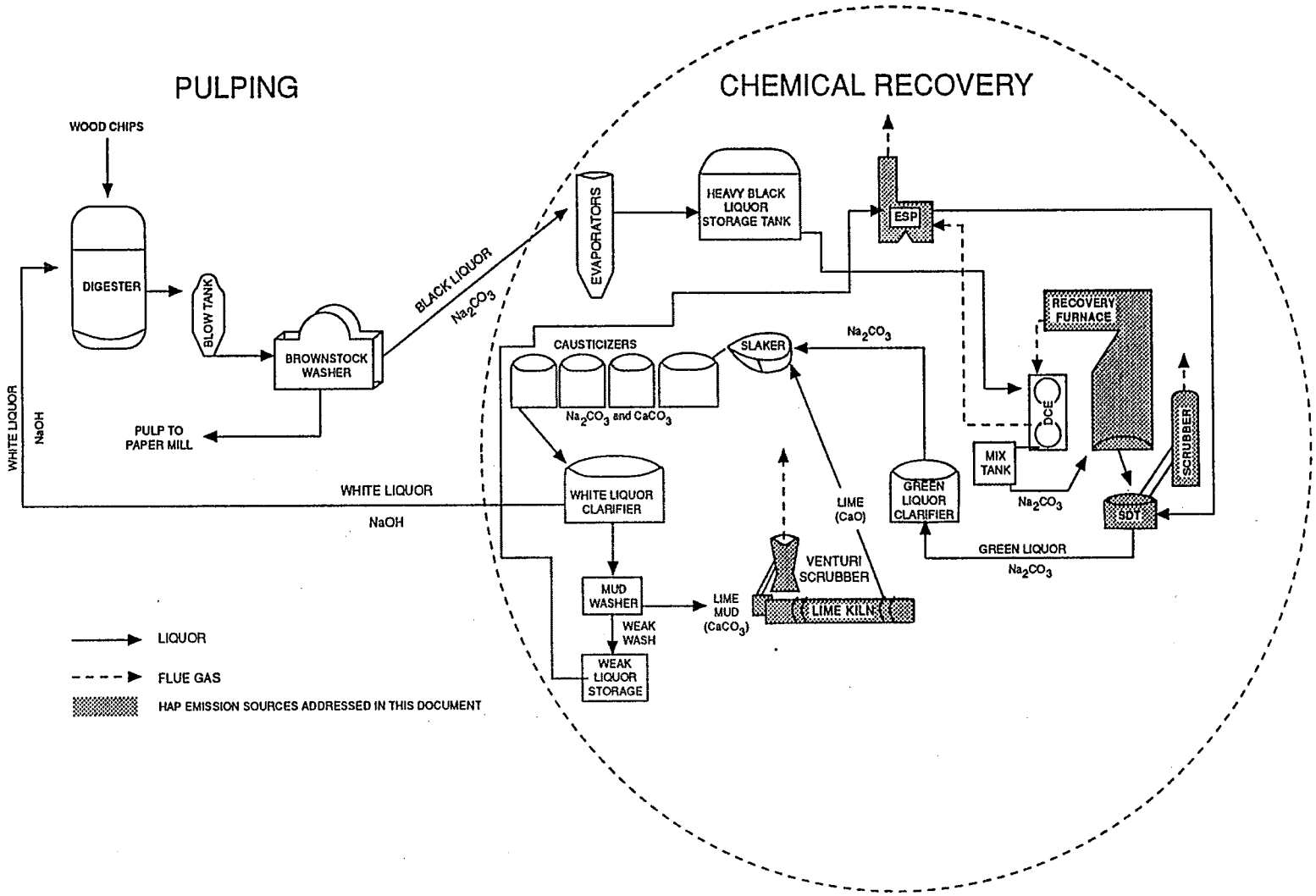

Figure 2-4. Soda process--chemical recovery area (DCE recovery furnace).

The purpose of the chemical recovery cycle is to recover cooking liquor chemicals from spent cooking liquor. The process involves concentrating black liquor, combusting organic compounds, reducing inorganic compounds, and reconstituting cooking liquor.

Cooking liquor, which is referred to as "white liquor," is an aqueous solution of sodium hydroxide (NaOH) and sodium sulfide (Na_2S) that is used in the pulping area of the mill. In the pulping process, white liquor is introduced with wood chips into digesters, where the wood chips are "cooked" under pressure. The contents of the digester are then discharged to a blow tank, where the softened chips are disintegrated into fibers or "pulp." The pulp and spent cooking liquor are subsequently separated in a series of brown stock washers. Spent cooking liquor, referred to as "weak black liquor," from the brown stock washers is routed to the chemical recovery area. Weak black liquor is a dilute solution (approximately 12 to 15 percent solids) of wood lignins, organic materials, oxidized inorganic compounds (sodium sulfate [Na_2SO_4], sodium carbonate [Na_2CO_3]), and white liquor (Na_2S and NaOH).

In the chemical recovery cycle, weak black liquor is first directed through a series of multiple-effect evaporators (MEE's) to increase the solids content to about 50 percent. The "strong" (or "heavy") black liquor from the MEE's is then either oxidized in the BLO system if it is further concentrated in a DCE or routed directly to a concentrator (NDCE). Oxidation of the black liquor prior to evaporation in a DCE reduces emissions of TRS compounds, which are stripped from the black liquor in the DCE when it contacts hot flue gases from the recovery furnace. The solids content of the black liquor following the final evaporator/concentrator typically averages 65 to 68 percent.

Concentrated black liquor is sprayed into the recovery furnace, where organic compounds are combusted, and the Na_2SO_4 is reduced to Na_2S. The black liquor burned in the recovery furnace has a high energy content (13,500 to 15,400 kilojoules per kilogram [kJ/kg] of dry solids [5,800 to 6,600 British thermal

units per pound {Btu/lb} of dry solids]), which is recovered as steam for process requirements, such as cooking wood chips, heating and evaporating black liquor, preheating combustion air, and drying the pulp or paper products. Particulate matter (PM) (primarily Na_2SO_4) exiting the furnace with the hot flue gases is collected in an electrostatic precipitator (ESP) and added to the black liquor to be fired in the recovery furnace. Additional makeup Na_2SO_4, or "saltcake," may also be added to the black liquor prior to firing.

Molten inorganic salts, referred to as "smelt," collect in a char bed at the bottom of the furnace. Smelt is drawn off and dissolved in weak wash water in the SDT to form a solution of carbonate salts called "green liquor," which is primarily Na_2S and Na_2CO_3. Green liquor also contains insoluble unburned carbon and inorganic impurities, called dregs, which are removed in a series of clarification tanks.

Decanted green liquor is transferred to the causticizing area, where the Na_2CO_3 is converted to NaOH by the addition of lime (calcium oxide [CaO]). The green liquor is first transferred to a slaker tank, where CaO from the lime kiln reacts with water to form calcium hydroxide ($Ca(OH)_2$). From the slaker, liquor flows through a series of agitated tanks, referred to as causticizers, that allow the causticizing reaction to go to completion (i.e., $Ca(OH)_2$ reacts with Na_2CO_3 to form NaOH and $CaCO_3$).

The causticizing product is then routed to the white liquor clarifier, which removes $CaCO_3$ precipitate, referred to as "lime mud." The lime mud, along with dregs from the green liquor clarifier, is washed in the mud washer to remove the last traces of sodium. The mud from the mud washer is then dried and calcined in a lime kiln to produce "reburned" lime, which is reintroduced to the slaker. The mud washer filtrate, known as weak wash, is used in the SDT to dissolve recovery furnace smelt. The white liquor (NaOH and Na_2S) from the clarifier is recycled to the digesters in the pulping area of the mill.

At about 7 percent of kraft mills, neutral sulfite semichemical (NSSC) pulping is also practiced.[6] The NSSC process involves pulping wood chips in a solution of sodium sulfite and sodium bicarbonate, followed by mechanical defibrating. The NSSC and kraft processes often overlap in the chemical recovery loop, when the spent NSSC liquor, referred to as "pink liquor," is mixed with kraft black liquor and burned in the recovery furnace. In such cases, the NSSC chemicals replace most or all of the makeup chemicals. For Federal regulatory purposes, if the weight percentage of pink liquor solids exceeds 7 percent of the total mixture of solids fired and the sulfidity of the resultant green liquor exceeds 28 percent, the recovery furnace is classified as a "cross-recovery furnace."[7,8] Because the pink liquor adds additional sulfur to the black liquor, TRS emissions from cross recovery furnaces tend to be higher than from straight kraft black liquor recovery furnaces.[9]

2.2 COMBUSTION PROCESSES AND EQUIPMENT

The following section describes the process and characterizes the equipment for: (1) recovery furnaces, (2) SDT's, (3) BLO systems, and (4) lime kilns.

The process descriptions presented in this section provide overviews of the various processes associated with the chemical recovery cycle. Much information is available on the kraft chemical recovery cycle and associated equipment; therefore, the process descriptions in this section do not cover every aspect of the chemical recovery cycle. Additional information about the chemical recovery cycle can be found in References 1, 2, and 10 through 13 listed in Section 2.4.[1,2,10-13]

2.2.1 Recovery Furnaces

2.2.1.1 Process Description. The purpose of the kraft recovery furnace is to (1) recover inorganic pulping chemicals (e.g., Na_2S) and (2) produce steam. Inputs to the furnace include concentrated black liquor, combustion air, and auxiliary fuel (usually, auxiliary fuel is only used during shutdown or startup). Outputs include molten smelt (primarily Na_2S and Na_2CO_3), flue gases, and steam. The smelt exits from the bottom

of the furnace into a SDT tank, where the recovery of cooking chemicals continues. Particulate matter (primarily Na_2SO_4 and Na_2CO_3) entrained in the flue gases is also recovered using an ESP and subsequently added to the concentrated black liquor. Steam produced by the recovery furnace is used in other processes around the mill.

Prior to being fired in the recovery furnace, the black liquor is concentrated using an NDCE or DCE. Figures 2-5 and 2-6 show the equipment associated with NDCE and DCE recovery furnaces, respectively. The NDCE is an indirect, steam-heated black liquor concentrator. Black liquor typically enters the NDCE at a solids concentration of 50 percent and exits at a concentration of 68 percent or higher. The DCE uses the hot combustion gases exiting the furnace to increase the solids content of the black liquor from about 50 percent to 65 percent.

Direct contact evaporators may be of the cascade or cyclone design. Figures 2-7 and 2-8 present detailed diagrams of the cascade and cyclone designs, respectively. The cascade evaporator consists of a rotating assembly of tubes that are alternately submerged in black liquor and exposed to hot flue gases. A cyclone evaporator is a cylindrical vessel with a conical bottom. Black liquor is sprayed into the side of the evaporator, where it contacts the hot combustion gases that are introduced tangentially, creating a "cyclone" effect. The flue gases exit from the top of the evaporator, and the concentrated black liquor drains down to and exits from the bottom of the evaporator.[2]

To minimize the stripping of TRS compounds when the hot flue gases contact the black liquor in the DCE, most DCE's are preceded by BLO units, which stabilize the sulfur compounds in the black liquor. Black liquor that is concentrated in NDCE's does not contact the hot flue gases, and, therefore, does not require oxidation. Because the NDCE recovery furnace typically has lower TRS emissions than does the DCE recovery furnace, the NDCE recovery furnace is also referred to as the "low-odor"

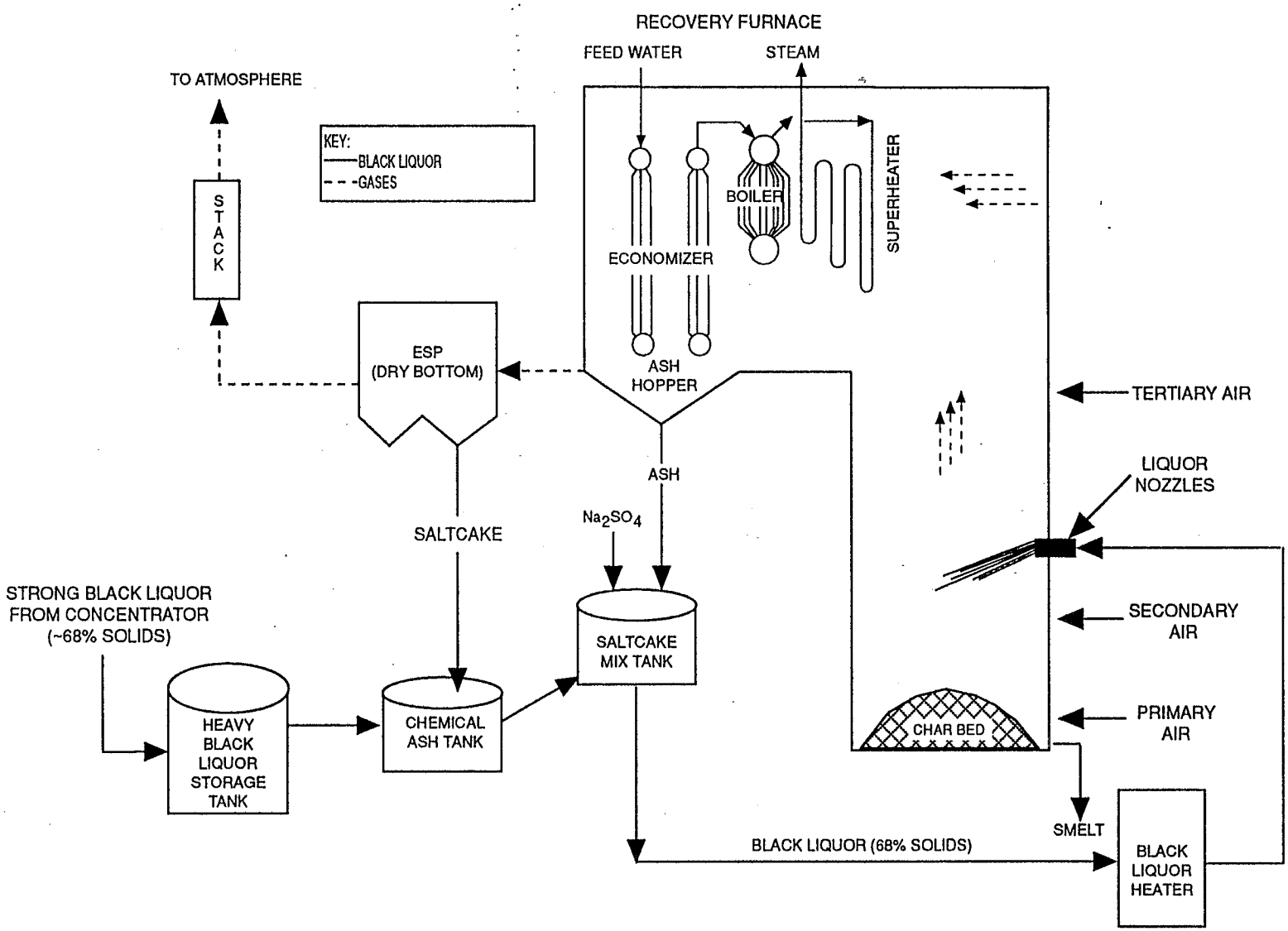

Figure 2-5. Schematic of NDCE recovery furnace and associated equipment.

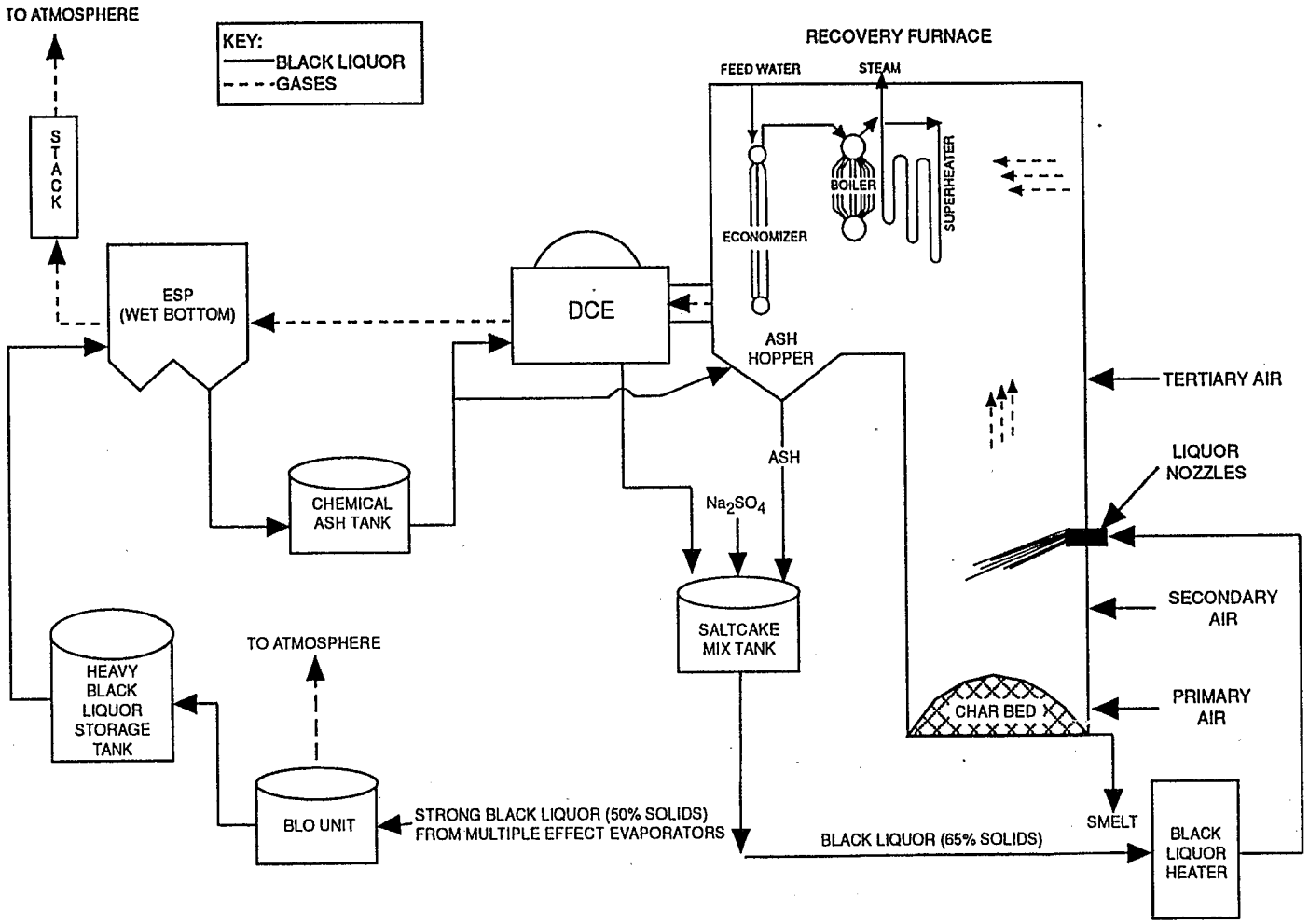

Figure 2-6. Schematic of DCE recovery furnace and associated equipment.

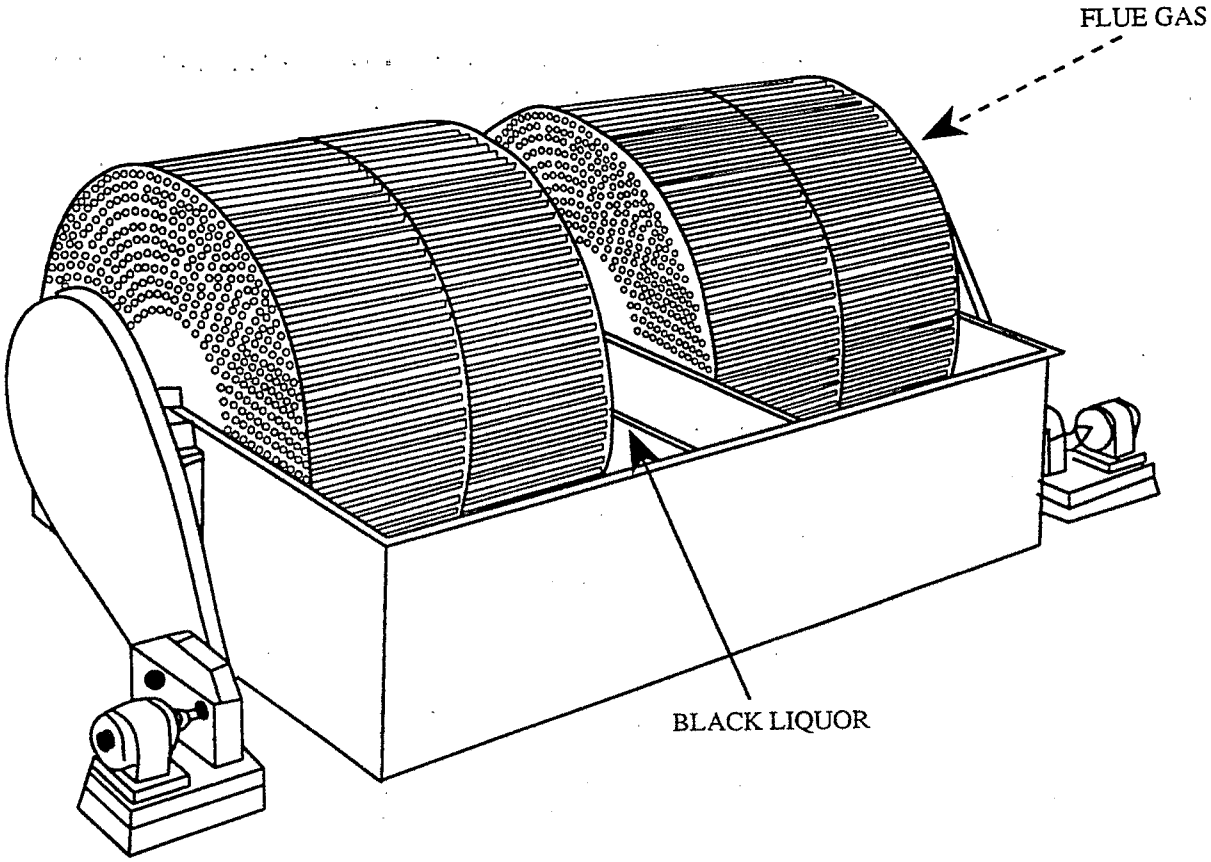

Figure 2-7. Cascade design evaporator.

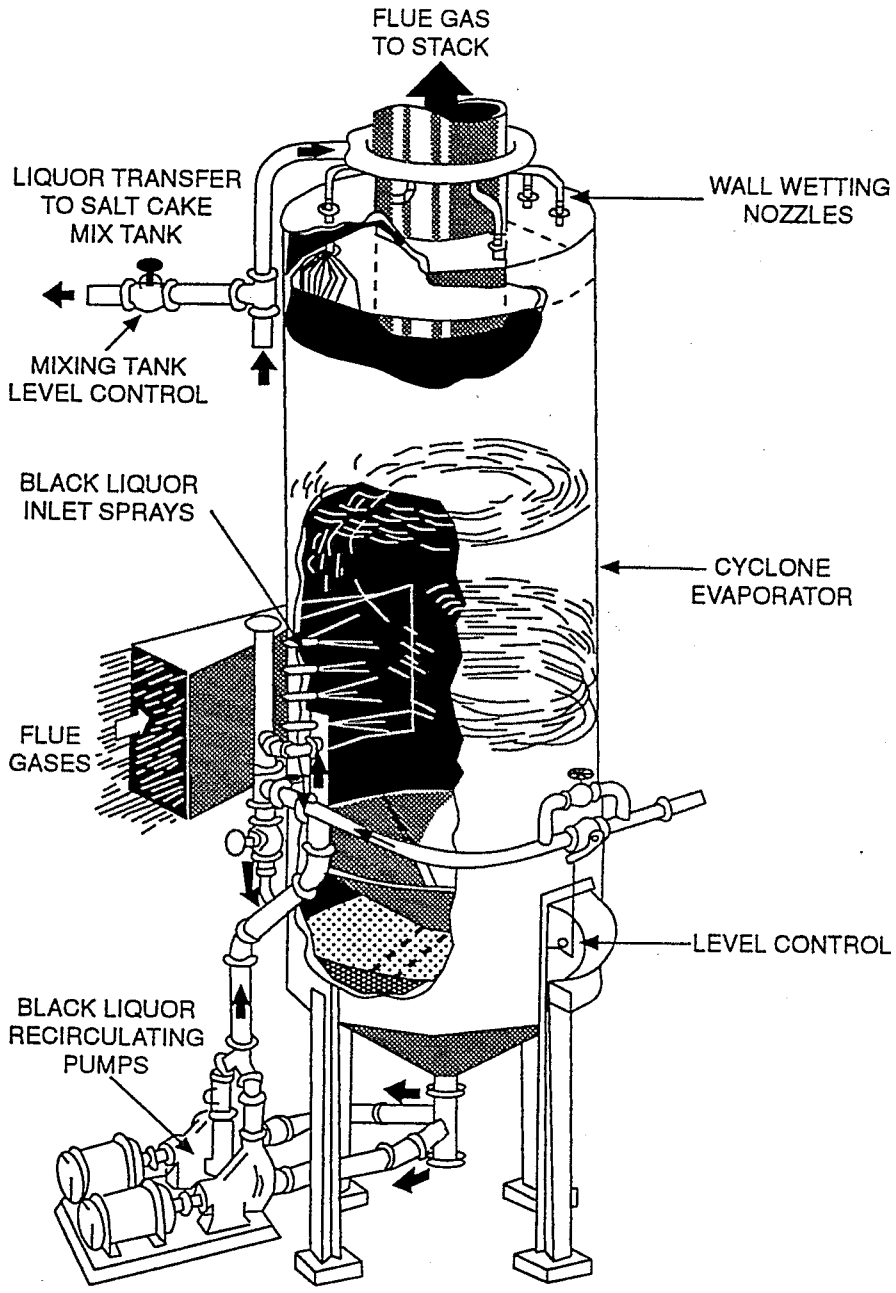

Figure 2-8. Cyclone design evaporator.

design. Since the 1970's, most new recovery furnaces have been designed with NDCE's.

Regardless of how the black liquor is concentrated, the chemical reactions that take place inside an NDCE- or DCE-type furnace are the same. The concentrated black liquor is sprayed into the furnace through fixed or oscillating nozzles or "guns" mounted in the walls of the furnace. Depending upon the design and operation of the recovery furnace, the sprayed black liquor may hit the opposing wall, where it dries and burns before falling to the hearth, or may fall short of the opposing wall and dry and burn in suspension. Combustion air is generally supplied to the furnace at three levels, with two levels located below the black liquor nozzles and one above. Some furnaces may have only two combustion air levels (i.e., one above and one below the liquor nozzles).

The inorganic chemicals in the black liquor are recovered in three distinct zones inside the recovery furnace. Figure 2-9 shows where each zone is located in the furnace and the chemical reactions that take place in each zone. The area of the furnace extending from the black liquor spray to just above the molten smelt at the bottom of the furnace is referred to as the "drying zone." The purpose of the drying zone is to evaporate the water from the liquor droplets. The reducing zone is just below the drying zone and includes the char bed. The reduction of Na_2SO_4 to Na_2S takes place in the reducing zone. Volatile gases that are released in the drying and reducing zones of the furnace travel to the highly turbulent upper section of the furnace, referred to as the oxidizing zone, where the gases are combusted.[2] The heat generated from the combustion of these gases is then used to generate steam as the combustion gases are drawn through the heat exchanger section of the furnace (i.e., superheater, boiler bank, and economizer).

For NDCE recovery furnaces, the design economizer exit gas temperature ranges from 177° to 190°C (350° to 375°F).[1] For DCE recovery furnaces, the heat from the recovery furnace is used to evaporate the black liquor. Thus, the required economizer exit

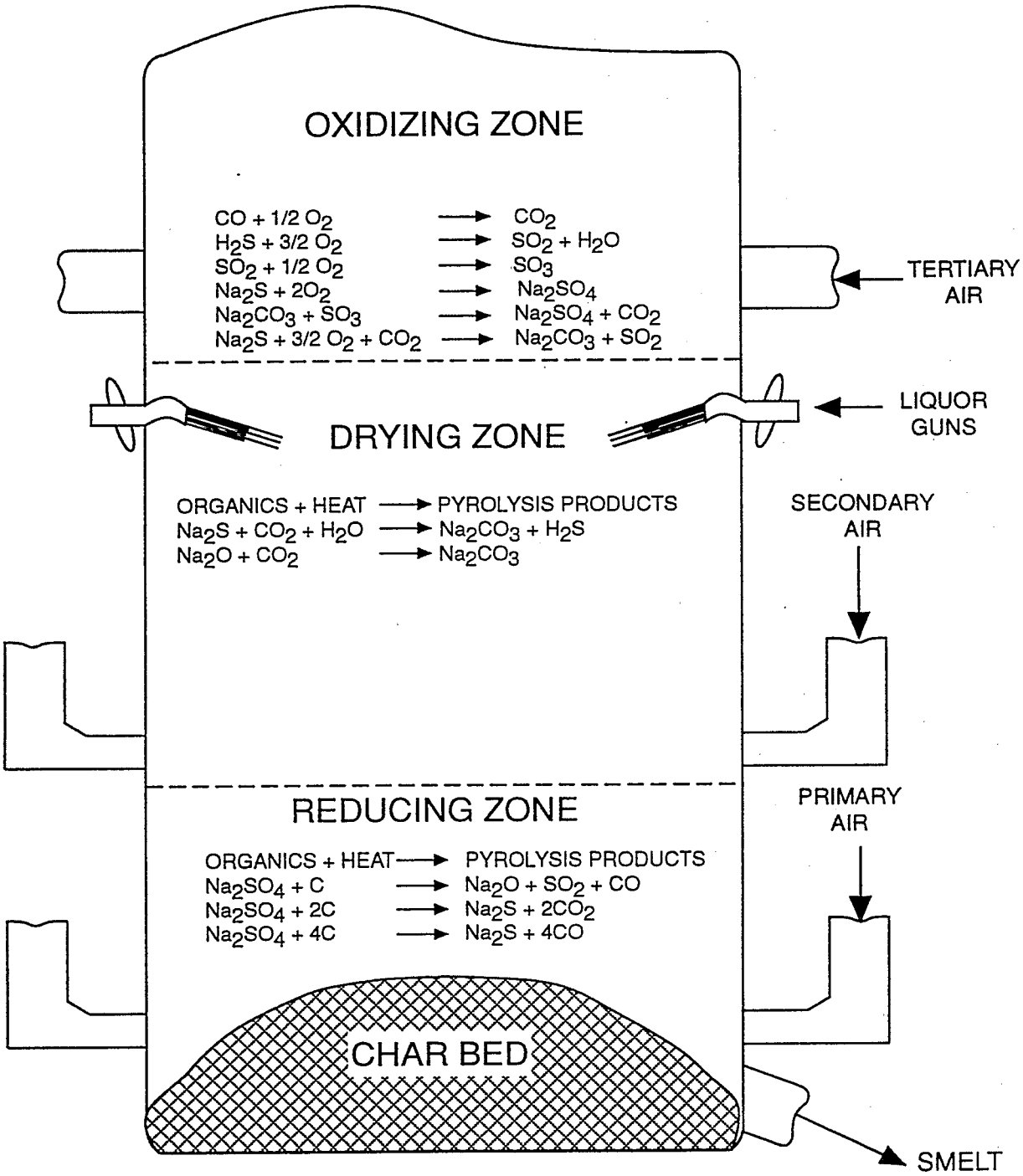

Figure 2-9. Recovery furnace zones and air stages.

gas temperature for DCE recovery furnaces, 370° to 430°C (700° to 800°F), is much higher than that for NDCE recovery furnaces. Because a large portion of the combustion heat from DCE recovery furnaces is required for black liquor evaporation, less combustion heat is available to produce steam.

Flue gases exiting the economizer are routed either directly to a PM control device (i.e., for NDCE recovery furnaces) or to a DCE followed by a PM control device (i.e., for DCE recovery furnaces). The recovered PM (primarily Na_2SO_4 [saltcake] and Na_2CO_3) is subsequently added to the concentrated black liquor, and the cleaned flue gas exits through the stack. Approximately 95 percent of the Na_2SO_4 is recovered; additional makeup saltcake is added to the concentrated black liquor as needed.

The inorganic chemical in the black liquor, Na_2SO_4, is reduced to Na_2S, a cooking liquor chemical, in the reducing zone (lower section) of the furnace. The Na_2S and other inorganic chemicals, predominantly Na_2CO_3, drain as molten smelt from the furnace bottom to the SDT, where reprocessing into cooking liquor continues.

2.2.1.2 Equipment Characterization. An estimated 211 recovery furnaces operate at U.S. kraft pulp mills.[3] Detailed information is available on about 85 percent of the estimated number of furnaces.[14] Based on the available data, 61 percent of these recovery furnaces are NDCE recovery furnaces and 39 percent are DCE recovery furnaces.[3] Eight percent of the recovery furnaces (76 percent of which are NDCE and 24 percent of which are DCE recovery furnaces) are located at kraft pulp mills that also practice NSSC pulping.[6] The number of recovery furnaces at these mills that can be classified as cross-recovery furnaces is uncertain.

The nationwide distribution of DCE and NDCE recovery furnaces by age (i.e., original installation date) is provided in Figures 2-10a and 2-10b, respectively.[3] As shown in Figures 2-10a and 2-10b, most recovery furnaces installed since the 1970's have been NDCE recovery furnaces. The first installation of a new NDCE recovery furnace took place in 1969.[15]

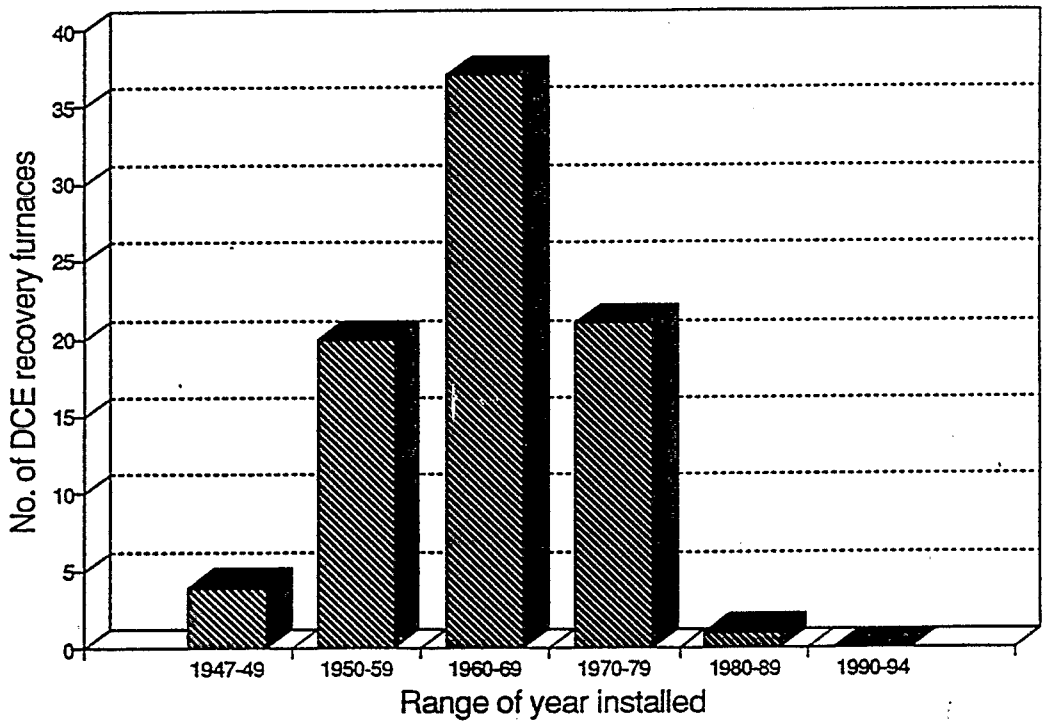

Figure 2-10a. DCE recovery furnace age distribution.

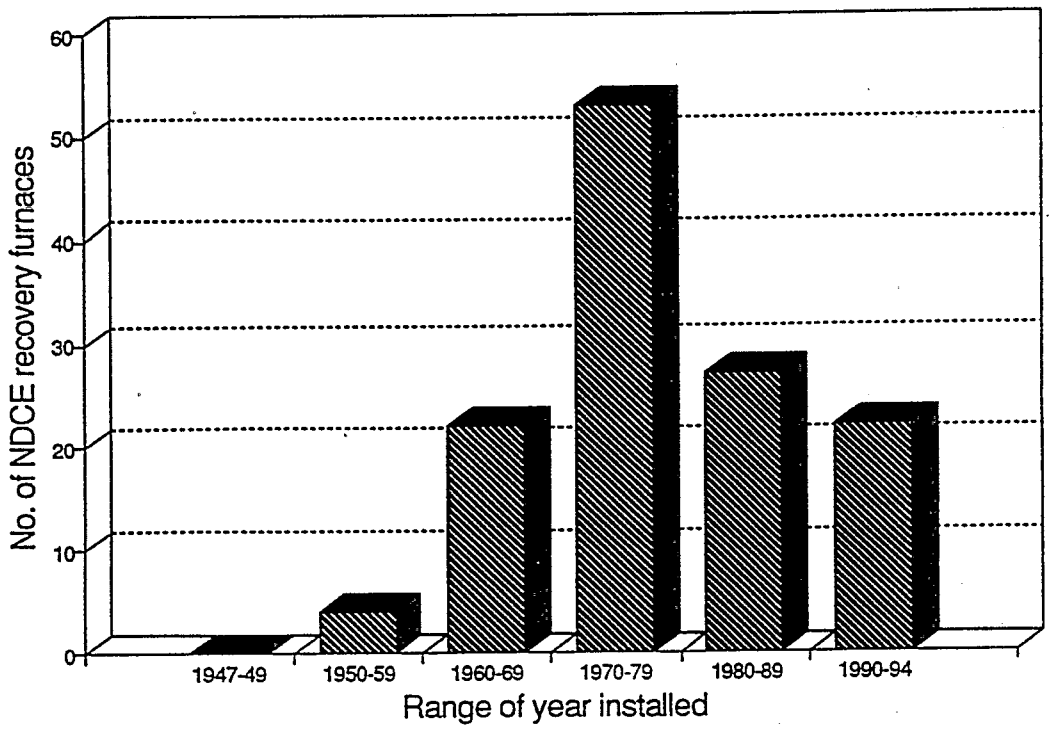

Figure 2-10b. NDCE recovery furnace age distribution.

Although Figure 2-10b shows that approximately 22 NDCE recovery furnaces currently in operation were installed in or before 1969, most, if not all of these furnaces were originally installed as DCE recovery furnaces and later converted to NDCE recovery furnaces. The most recent installation of a new DCE recovery furnace occurred in 1988; however, prior to 1988, the last installation of a new DCE recovery furnace was in 1979.[3]

The nationwide distribution of DCE and NDCE recovery furnaces by size (i.e., black liquor solids [BLS] feed rate) is provided in Figures 2-11a and 2-11b, respectively.[3] The DCE recovery furnaces are not only typically older than NDCE recovery furnaces, but, as shown in Figures 2-11a and 2-11b, DCE recovery furnaces are also typically smaller. The majority of DCE recovery furnaces have a firing rate less than or equal to 900,000 kilograms of BLS per day (kg BLS/d) (2 million pounds of BLS/d [lb BLS/d]); whereas, the majority of NDCE recovery furnaces have firing rates greater than 900,000 kg BLS/d (2 million lb BLS/d).

The two furnace types also differ in the types of ESP's that are used to control particulate emissions from the furnace. For example, the ESP's that control PM emissions from DCE recovery furnaces tend to be wet-bottom ESP's (i.e., the PM catch falls directly into a pool of black liquor at the bottom of the ESP), whereas ESP's on NDCE recovery furnaces tend to be dry-bottom ESP's. Other differences between DCE and NDCE recovery furnaces include the following: (1) the inlet loading of PM to the ESP tends to be lower for DCE recovery furnaces than for NDCE recovery furnaces due to the recovery of some PM as the flue gases pass through the DCE; (2) emissions of TRS compounds tend to be higher from DCE recovery furnaces; and (3) NDCE recovery furnaces do not require black liquor oxidation, which eliminates one source of emissions.

Due to State and Federal regulations regarding PM emissions and the economic benefits of recycling PM captured from the recovery furnace flue gases, all kraft recovery furnaces are equipped with add-on PM control devices. Particulate matter

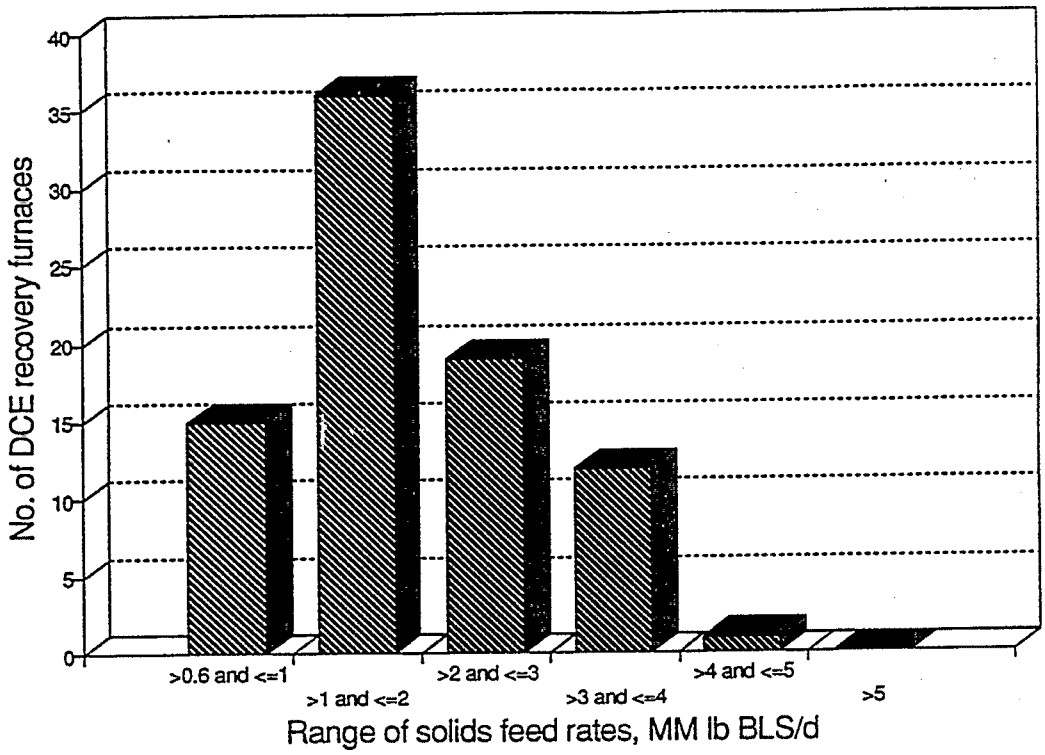

Figure 2-11a. DCE recovery furnace size distribution.

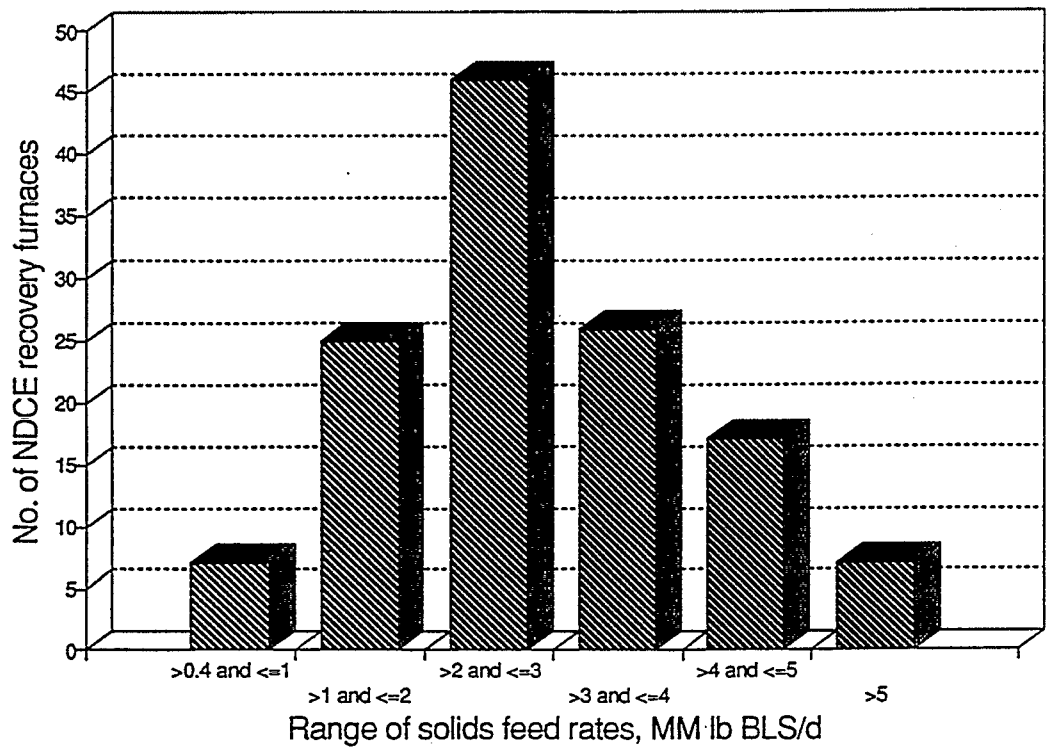

Figure 2-11b. NDCE recovery furnace size distribution.

emissions from approximately 95 percent of all kraft recovery furnaces are controlled with an ESP alone.[3] The remaining furnaces are controlled with ESP's followed by wet scrubbers (4 percent) or with wet scrubbers alone (1 percent).[3] Additional information on recovery furnace air pollution control devices (APCD's) and the effectiveness of these APCD's in controlling HAP emissions is provided in Chapter 3.

In general, recovery furnaces are not used to incinerate waste streams generated in other parts of the mill, with the exception of a few recovery furnaces (approximately 3 percent) that receive high-volume, low-concentration (HVLC) noncondensible gases (NCG's) from the pulping area of the mill.[14] Because of the importance of the recovery furnace to the chemical recovery cycle and the potential for catastrophic explosion (due to water entering the furnace during operation), the industry's Black Liquor Recovery Boiler Advisory Committee (BLRBAC) recommends that the recovery furnace not be used for NCG incineration.[1]

2.2.2 Smelt Dissolving Tanks

2.2.2.1 Process Description. Molten smelt is one of the main products from the combustion of black liquor. Smelt, which is predominantly Na_2S and Na_2CO_3, is formed in the bottom of the furnace. Smelt, at approximately 1040° to 1150°C (1900° to 2100°F), filters through the char bed and is continuously discharged through water-cooled smelt spouts into the SDT. In the SDT, smelt is mixed with weak wash water from the recausticizing area to form green liquor, an aqueous solution of Na_2CO_3 and Na_2S in about a three to one ratio. The green liquor is subsequently transferred to the recausticizing area for reprocessing into cooking liquor (i.e., white liquor). A schematic of an SDT with a wet scrubber is provided in Figure 2-12.

The SDT is a large, covered vessel located below the recovery furnace. Green liquor is maintained in the tank at a level of about half the depth of the tank. As the smelt exits the water-cooled smelt spouts and falls several feet into the SDT, it is shattered by high-pressure steam or shatter sprays of

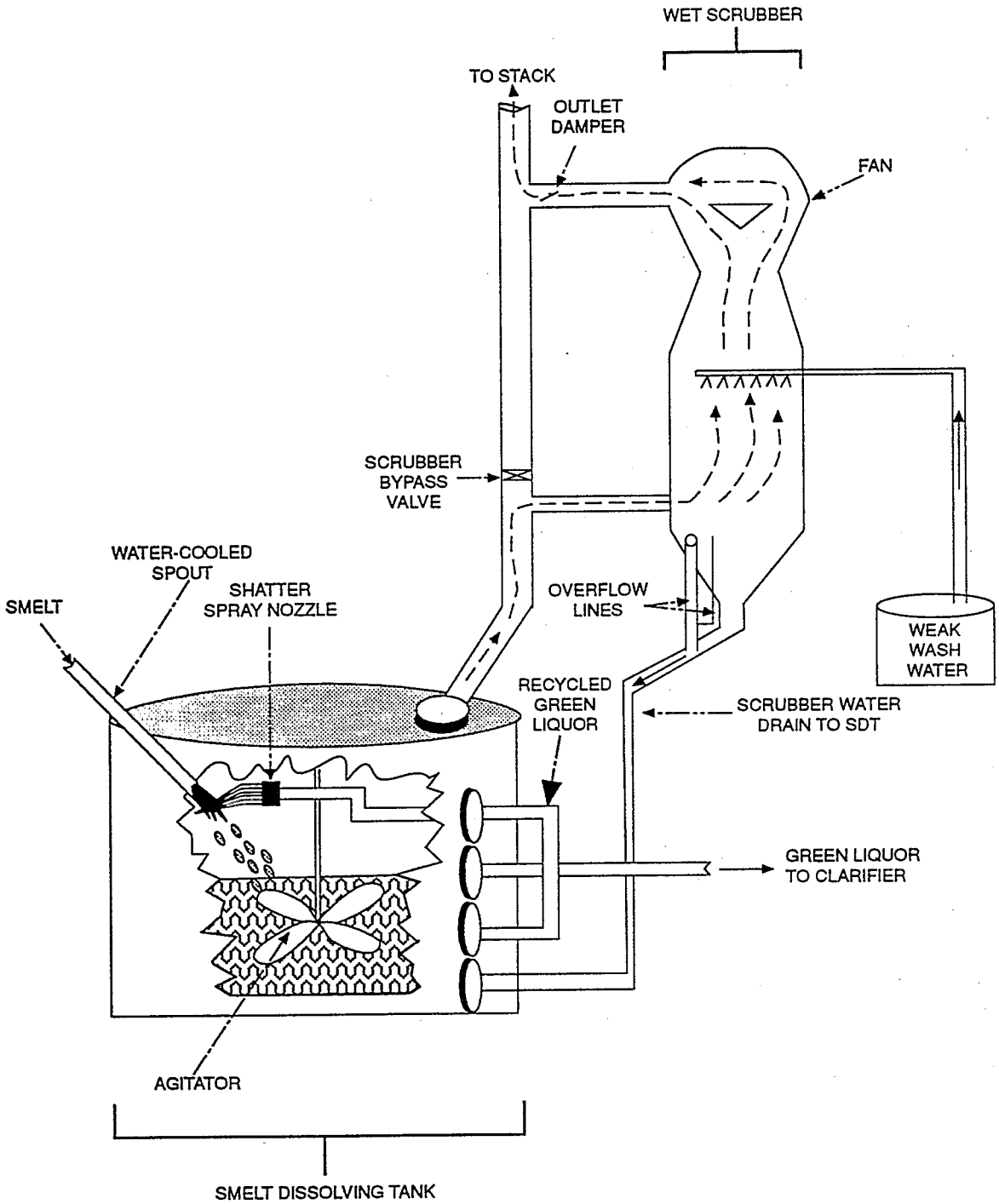

Figure 2-12. Smelt dissolving tank and wet scrubber.

recirculated green liquor. The steam or shatter sprays break the smelt flow into small droplets and cool the smelt before it falls into and reacts with the liquid in the SDT to form green liquor. Large volumes of steam are generated when the molten smelt and liquid mix. The vapor space above the liquid level provides an opportunity for water vapor and PM resulting from the quenching of smelt to settle out of suspension into the green liquor. An induced-draft fan constantly draws the vapor and entrained PM through a PM control device, generally a wet scrubber. Scrubber water is sprayed into the scrubber and allowed to drain directly into the SDT, where it reacts with smelt to form green liquor.

The SDT is constantly agitated to prevent formation of hot spots on the surface of the liquor and to keep solids from accumulating in the bottom of the tank. Surface liquor hot spots can contribute to the formation of explosive hydrogen gas from the dissociation of water reacting with the hot smelt.

Green liquor formed in the SDT is sent to the green liquor clarifier in the causticizing area. Green liquor is converted to white liquor (i.e., NaOH and Na_2S) in the causticizing area.

2.2.2.2 Equipment Characterization. An estimated 227 SDT's operate at U.S. kraft pulp mills.[3] Detailed information is available on approximately 83 percent of the estimated number of SDT's.[6] Smelt dissolving tanks are basically large, covered tanks. Some recovery furnaces have two SDT's, which explains why the estimated number of SDT's is higher than the estimated number of recovery furnaces. When there are two SDT's under one furnace, the flue gases from these SDT's may be combined or treated separately.

Due to State and Federal regulations regarding PM emissions, all but two kraft SDT's are equipped with add-on PM control devices. Particulate matter emissions from approximately 87 percent of kraft SDT's are controlled with wet scrubbers.[14] Particulate matter emissions from most of the remaining SDT's are controlled with mist eliminators alone.[14] Additional information on SDT APCD's and the effectiveness of these APCD's in controlling HAP emissions is provided in Chapter 3.

2.2.3 Black Liquor Oxidation Systems

2.2.3.1 Process Description. The BLO system reduces malodorous TRS emissions from DCE recovery furnaces at kraft pulp mills. Total reduced sulfur compounds (primarily hydrogen sulfide [H_2S]) are stripped from black liquor when hot flue gases from the recovery furnace contact the black liquor in the DCE. Hydrogen sulfide is stripped by the reaction between residual Na_2S in the black liquor and carbon dioxide (CO_2) and water (H_2O) in the recovery furnace flue gas, as follows:

$$\underset{\text{sodium sulfide}}{Na_2S} + CO_2 + H_2O \longrightarrow \underset{\text{sodium carbonate}}{Na_2CO_3} + \underset{\text{hydrogen sulfide}}{H_2S}$$

The BLO system minimizes the stripping of TRS compounds in the DCE by stabilizing the sulfur compounds in the black liquor prior to evaporation in the DCE. The main reaction that takes place in the BLO system is the oxidation of Na_2S to nonvolatile sodium thiosulfate ($Na_2S_2O_3$), as follows:

$$\underset{\text{sodium sulfide}}{2Na_2S} + 2O_2 + H_2O \longrightarrow \underset{\text{sodium thiosulfate}}{Na_2S_2O_3} + \underset{\text{sodium hydroxide}}{2NaOH}$$

The oxidation efficiency of the BLO process is measured by the percent conversion of Na_2S to $Na_2S_2O_3$ in the black liquor on a gram per liter (g/L) (pound per gallon [lb/gal]) basis. Oxidation efficiencies greater than 99 percent are common. High oxidation efficiencies require good mixing and the avoidance of excessive oxygen (O_2) consumption to limit side reactions. Sulfidity levels of the black liquor entering the DCE are targeted to less than 0.2 g/L (0.002 lb/gal) in order to meet TRS emission limits for DCE recovery furnaces.[1] Oxidizing black liquor results in a slight increase (2 to 3 percent) in the solids content of the black liquor and reduces its heating value by 2 to 5 percent.[16]

The BLO system is typically located after the MEE, a process referred to as "strong" black liquor oxidation. A small number of mills oxidize black liquor prior to evaporation in the MEE, which is referred to as "weak" black liquor oxidation. Other options are to oxidize both weak and strong black liquor or to oxidize the black liquor between effects of the MEE. To avoid excessive foaming caused by tall oil components in weak black liquor, mills in the south that pulp pine oxidize strong black liquor. With strong black liquor, the soaps and tall oil are removed during the soap-skimming process that follows black liquor evaporation in the MEE. With weak BLO systems, the effects of partial reversing of the oxidation reaction (i.e., oxidized sulfur compounds reducing to H_2S) that occurs in the MEE can be minimized by adding a second oxidation step, such as oxygen polishing of strong black liquor.

Black liquor can be oxidized using either air or pure (molecular) O_2. Because of economic considerations, the majority of BLO systems use air as the oxidant. Air BLO systems have higher capital costs than oxygen systems, but their operating costs are usually much lower.

Air-sparging units are the predominant type of BLO equipment used at kraft pulp mills. Figure 2-13 shows a schematic of an air-sparging BLO system. Air-sparging units operate by bubbling air, which is sometimes preheated, through the black liquor using multiple diffuser nozzles. Air systems require residence times of several hours or more to obtain high oxidation efficiencies. Because of this relatively long residence time, large oxidation tanks--on the order of 9 meters (m) (30 feet [ft]) in diameter and 12 m [40 ft] high) must be used. Air-sparging units have one to three tanks (or stages) that operate in series. Air-sparging units are equipped with mechanical foam breakers for foam control; chemical defoamers (e.g., diesel oil, turpentine, and kerosene) may also be used.[1,12] Each oxidation tank is vented. Gases exiting the BLO system flow through cyclone separators to have entrained water droplets removed prior to being vented to

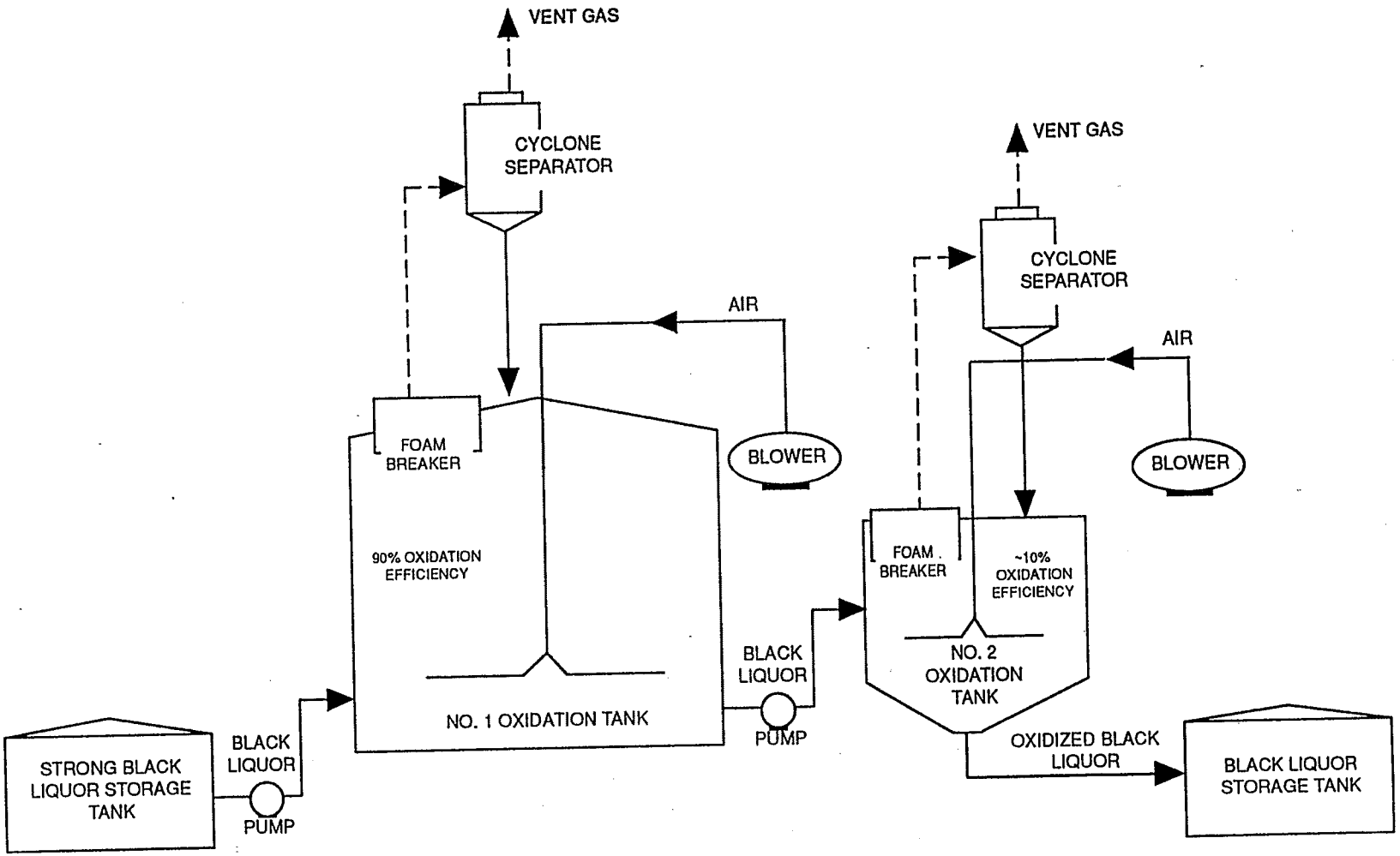

Figure 2-13. Two-stage air-sparging black liquor oxidation system.

the atmosphere. Add-on APCD's typically are not used with air oxidation units.

Oxygen BLO systems require only 30 seconds (sec) to 5 minutes (min) residence time, which enables the use of in-line reactors. Since all the gas (i.e., pure O_2) added to the system is consumed in the oxidation reaction, a system vent is not needed.[16] The oxygen consumption for weak black liquor oxygen systems is 25 to 65 kilograms per megagram (kg/Mg) of pulp (50 to 130 pounds per ton [lb/ton] of pulp); strong black liquor oxygen systems use 35 to 75 kg O_2/Mg pulp (70 to 150 lb O_2/ton pulp.)[17]

Oxygen polishing is sometimes used as a supplement to air oxidation systems to address (1) stricter TRS standards, (2) an overloaded BLO system resulting from production increases, or (3) peaks in TRS emissions resulting from process upsets or temporary production increases. In-line O_2 polishing systems are used to oxidize strong liquor and may follow either a weak liquor air system or a strong liquor air system. Polishing of strong black liquor requires approximately 2.5 to 13 kg O_2/Mg pulp (5 to 25 lb O_2/ton pulp) to achieve the desired incremental increases in oxidation efficiency.[17]

2.2.3.2 Equipment Characterization. Of the 47 U.S. kraft pulp mills that operate DCE recovery furnaces, only 1 (with 1 DCE recovery furnace) is known to operate without a BLO system. Therefore, the nationwide number of BLO systems is estimated to be 46.[3] Detailed information is available on approximately 91 percent of the estimated U.S. population.[14,18] The type of equipment operated and the type of liquor oxidized for the BLO systems is summarized in Table 2-1.[3] Oxidation of strong black liquor using air-sparging units is the predominant method of black liquor oxidation and accounts for an estimated 89 percent of the U.S. population.[3] Seven percent of the mills operating air-sparging units also have added O_2 polishing systems. Of the air-sparging BLO systems, 50 percent are two-stage systems, 31 percent are single-stage systems, and 18 percent have three oxidation tanks.[14]

TABLE 2-1. BLACK LIQUOR OXIDATION SYSTEMS: EQUIPMENT TYPES[3]

Equipment type	Liquor type	Estimated No. in U.S.	Percent
Air sparging	Strong	38	83
Air sparging with O_2 polishing	Strong	3	7
Air, sieve tray tower	Weak	1	2
Molecular O_2	Strong	2	4
Molecular O_2	Weak	1	2
Molecular O_2	Weak and strong	1	2
Total	--	46	100

Approximately 8.5 percent of BLO systems are molecular O_2 systems.[3] Strong black liquor is oxidized with O_2 after the second effect of the MEE (26 percent solids) at two of the four mills; weak black liquor (i.e., prior to the MEE) is oxidized at another mill; and both weak and strong liquor are oxidized with O_2 at the fourth mill.[3,14] One mill with a BLO system located after the second effect of the MEE reported an oxygen usage rate of 32 Mg/d (35 ton/d) (13 kg/Mg pulp [25 lb/ton pulp]).[14]

Emissions from most air-based BLO systems (95 percent) are uncontrolled.[3] Based on available information, TRS emissions from 5 percent of air-based BLO systems are currently controlled by using a condenser or mist eliminator to remove the water vapor and then venting the gas stream to a power boiler for incineration.[14] Molecular O_2 systems do not require system vents and, thus, have no emissions directly associated with the BLO unit.[13]

2.2.4 Lime Kilns

2.2.4.1 Process Description. The lime kiln is part of the causticizing process, in which green liquor from the SDT is converted to white liquor. The function of the lime kiln is to

convert lime mud (calcium carbonate [$CaCO_3$]) to reburned lime (calcium oxide [CaO]), a process known as calcining:

$$CaCO_3 + \text{heat} \longrightarrow CaO + CO_2 \quad \text{[lime kiln]}$$

lime mud/ calcium carbonate — lime/ calcium oxide

Lime mud that is burned in the lime kiln is a final product of the following causticizing reactions:

$$CaO + H_2O \longrightarrow Ca(OH)_2 \quad \text{[slaker]}$$

lime/ calcium oxide — calcium hydroxide

$$Ca(OH)_2 + Na_2CO_3 \longrightarrow NaOH + CaCO_3 \quad \text{[causticizer]}$$

calcium hydroxide — sodium carbonate — sodium hydroxide — lime mud/ calcium carbonate

Lime that is produced in the kiln is used in the causticizing reaction that takes place in the green liquor slaker and causticizers. The lime kiln typically produces about 95 percent of the lime needed for the causticizing reaction. Either make-up lime or limestone is purchased to account for losses.

Prior to calcining, lime mud from the causticizing tanks is washed and dewatered. Lime mud washers reduce the sodium and sulfide content of the lime mud, which lowers TRS emissions from the lime kiln. The lime mud is typically dewatered to about 70 percent solids using a rotary vacuum precoat filter. The precoat filter also helps reduce TRS emissions by oxidizing any remaining sodium sulfide to sodium thiosulfate using air that is pulled through the filter.

Rotary lime kilns, as shown in Figure 2-14, are typically used at kraft pulp mills. In a rotary kiln, lime mud from the precoat filter is introduced at the feed end (cold end) and flows downward towards the discharge end (hot end). The chemical composition of the lime mud is presented in Table 2-2. Natural gas or fuel oil are the fuels typically used to fire the kiln.

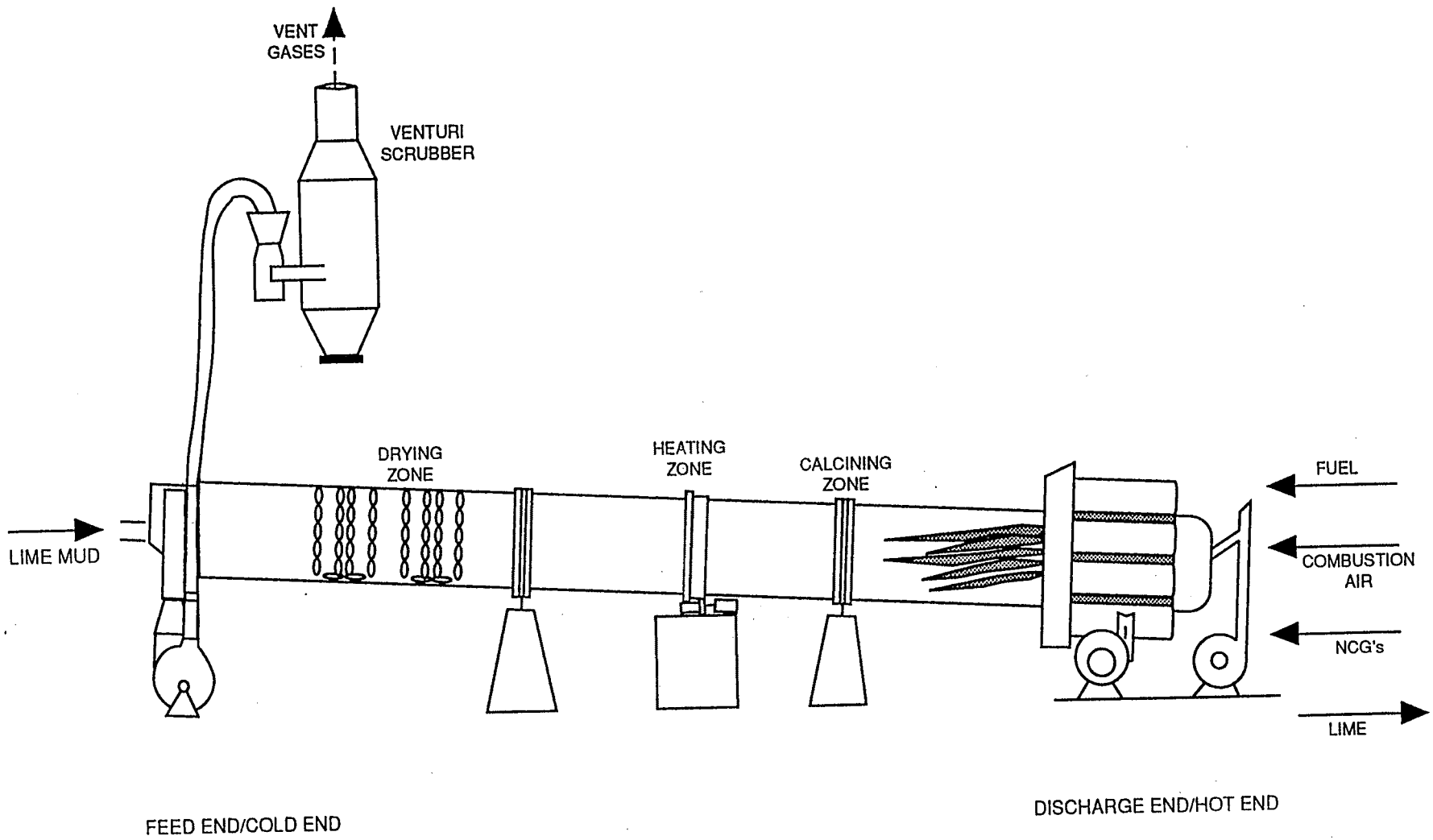

Figure 2-14. Schematic of a lime kiln used at kraft pulp mills.

TABLE 2-2. CHEMICAL COMPOSITION OF LIME MUD FEED[20]

Component	Composition (%)
Calcium carbonate	>95
Alumina	<0.5
Iron oxide	<0.5
Calcium oxide	<0.5
Sodium oxide	<0.2
Silica	<0.2
Sodium sulfide	<0.01

Primary combustion air is introduced through a concentric tube around the fuel pipe, and preheated secondary combustion air is introduced through the bottom of the firing hood.

The majority of lime kilns at kraft pulp mills also burn NCG's from various process vents, such as digester and evaporator vents. The NCG streams may be introduced into the kiln through a dedicated nozzle or combined and fed with either primary or secondary air. The components of the NCG stream include TRS compounds, turpentine, methanol, acetone, alpha-pinene, water vapor, nitrogen, and oxygen.[1,19] To avoid excessive sulfur dioxide (SO_2) formation in the lime kiln, the NCG gas stream may be scrubbed prior to incineration to remove TRS compounds. Packed columns with white liquor as the scrubbing fluid are commonly used. The TRS removal efficiency is typically about 70 percent.[1]

Rotary kilns have three internal zones: (1) the drying zone, (2) the heating zone, and (3) the calcining zone. In the drying zone, water is evaporated from the lime mud as it passes through metal chains that are suspended from the kiln shell in a curtain or garland arrangement. The lime mud is dried to about 95 percent or greater solids as it passes through the chains that are heated by the hot flue gases that flow countercurrent to the lime mud. In addition to providing additional heat transfer area for drying the lime mud, the chains also help reduce the amount

of lime dust exiting with flue gases. In the heating zone, other heat transfer devices, such as tumblers and lifters, are used to heat the dried mud uniformly in preparation for calcining. The calcining reaction, which takes place in the calcining zone, requires a minimum temperature of 815°C (1500°F); temperatures greater than about 1150°C (2100°F) can cause overburning, which leads to a less-reactive lime product.[1] Ideally, calcination produces lime pellets that are about 2 centimeters (cm) (0.75 inches [in.]) in diameter; however, if the lime mud is improperly dried and heated, large lime balls may be produced.[1] The hot lime product is cooled by incoming secondary air as it passes under the burner towards the discharge end of the kiln. Newer kilns use integral tube coolers to preheat secondary combustion air while cooling the discharged lime pellets. In this heat exchange process, the air is heated to about 315°C (600°F) and the lime is cooled to about 190°C (375°F). The reburned lime product from the integral tube coolers, or from the kiln discharge hoppers in the absence of coolers, is transported to the lime storage bin and subsequently introduced into the green liquor slaker.

Combustion gases exit the lime mud feed end of the kiln at temperatures of approximately 150° to 200°C (300° to 400°F). The exhaust gases consist of combustion products, CO_2 released during calcination, water vapor evaporated from the mud, and entrained lime dust. Particulate matter in the exhaust gas is mainly sodium salts, $CaCO_3$, and CaO. Add-on PM control devices are required to meet Federal and State PM standards. Venturi scrubbers are the most commonly used control device, and water is typically used as the scrubbing fluid. The exhaust stream may be scrubbed with a caustic solution with the added benefit of lowering TRS and SO_2 emissions. However, Federal and State TRS standards can be met through good lime mud washing practices, which reduces the sulfide content of the lime mud feed. If a wet scrubber is used, a cyclone separator may be installed upstream. The dust collected by the separator is returned directly to the lime kiln. In recent years, the use of ESP's has been more

prevalent. As with the cyclone separator, the PM catch from the ESP is returned directly to the lime kiln.

2.2.4.2 Equipment Characterization. An estimated 192 lime kilns operate at U.S. kraft pulp mills.[3] Detailed information is available on about 85 percent of the estimated number of lime kilns.[14] Rotary kilns are the most commonly used type of lime kiln at kraft pulp mills, accounting for about 98 percent of the kilns.[6,21] Fluidized-bed calciners are also used by the kraft pulp industry. Industry representatives estimated the number of fluidized-bed calciners at two to five, accounting for the remaining 2 percent of lime kilns at kraft pulp mills.[6,21] Based on available data, the oldest operating lime kiln was installed in 1940.[14] New lime kilns were installed at a rate of approximately three kilns per year from 1981 to 1991.[14]

The predominant fuel used by the industry is natural gas, accounting for about 68 percent of lime kilns.[14] Fuel oil is the primary fuel for about 29 percent of kilns.[14] Petroleum coke is fired in about 3 percent of the kilns as the primary fuel.

Due to State and Federal regulations regarding PM emissions, all lime kilns are equipped with add-on PM control devices. Particulate matter emissions from the majority (90 percent) of lime kilns at kraft pulp mills are controlled with wet scrubbers. Two percent of these scrubbers are operated in series with a second scrubber. Venturi scrubbers are the most commonly used type of wet scrubber.[14] Water is the typical scrubbing fluid, but caustic and weak wash are also used.[14] The scrubbing fluid is recirculated, and the scrubber blowdown is recycled to the lime mud washer. Particulate matter emissions from the remaining 10 percent of lime kilns are controlled by single-chamber ESP's (9 percent) or the combination of an ESP and scrubber (1 percent).[14] Installing ESP's to control PM from lime kilns has been more widespread in recent years; about half of the APCD installations since 1990 have been ESP's.

An estimated 66 percent of lime kilns are either primary or backup incineration devices for NCG streams.[14] Of the lime kilns that incinerate NCG's, 57 percent incinerate low-volume, high-

concentration (LVHC) NCG streams; 28 percent incinerate HVLC NCG streams; and 24 percent incinerate both LVHC and HVLC NCG streams. Approximately 33 percent of the lime kilns in which NCG's are incinerated use an NCG scrubber prior to incineration.[14] The majority of these NCG scrubbers use white liquor as the scrubbing fluid; a small percentage use water.

2.3 BASELINE EMISSIONS

2.3.1 Federal and State Regulations Affecting Kraft Pulp Mill Combustion Sources

Federal regulations that affect the pulp and paper industry include new source performance standards (NSPS) for kraft pulp mills. The NSPS does not include soda mills. The NSPS specifies PM and TRS emission limits for various pulp mill equipment that are constructed, modified, or reconstructed after September 24, 1976.[7,8] Recovery furnaces, SDT's, and lime kilns are regulated under the kraft pulp mill NSPS.

The percentage of recovery furnaces, SDT's, and lime kilns that are subject to the NSPS were estimated using installation dates. These estimates, therefore, do not include those recovery furnaces, SDT's, and lime kilns that are subject to the NSPS through the modification/reconstruction provisions. Approximately 4 percent (i.e., 3 of 83) of the DCE recovery furnaces and 39 percent (50 of 128) of the NDCE recovery furnaces are subject to the NSPS.[3] Based on recovery furnace installation dates, approximately 29 percent of the SDT's are subject to the NSPS.[3] Approximately 28 percent of the lime kilns are subject to the NSPS.[14] Nine new BLO systems have been installed since the 1976 NSPS proposal for kraft pulp mills; eight of these new BLO systems are still in operation, making up 17 percent of the existing population of operating BLO systems.[18]

Table 2-3 lists the NSPS PM and TRS emission limits. The EPA TRS emission limit guidelines for sources not subject to the NSPS are also listed in Table 2-3. The TRS guidelines are the same as the NSPS emission limits for NDCE and cross-recovery furnaces. The NSPS emission limit is 5 parts per million dry volume (ppmdv) at 8 percent O_2. For DCE recovery furnaces, the

TABLE 2-3. NSPS AND EMISSION GUIDELINES FOR KRAFT PULP MILL COMBUSTION SOURCES[a]

Source	Particulate matter--NSPS	Total reduced sulfur	
		NSPS	Guidelines[b]
NDCE recovery furnace	0.10 g/dscm (0.044 gr/dscf) @ 8% O_2	5 ppmdv @ 8% O_2	5 ppmdv @ 8% O_2
DCE recovery furnace	0.10 g/dscm (0.044 gr/dscf) @ 8% O_2	5 ppmdv @ 8% O_2	20 ppmdv @ 8% O_2
Cross recovery furnace[c]	0.10 g/dscm (0.044 gr/dscf) @ 8% O_2	25 ppmdv @ 8% O_2	25 ppmdv @ 8% O_2
Smelt dissolving tank[d]	0.10 kg/Mg (0.20 lb/ton) BLS	0.016 kg/Mg (0.033 lb/ton) BLS	---
Lime kiln (gas-fired)	0.15 g/dscm (0.067 gr/dscf) @ 10% O_2	8 ppmdv @ 10% O_2	20 ppmdv @ 10% O_2
Lime kiln (liquid-fired)	0.30 g/dscm (0.13 gr/dscf) @ 10% O_2	8 ppmdv @ 10% O_2	20 ppmdv @ 10% O_2
Black liquor oxidation[e]	--	No control	No control

[a]g/dscm = grams per dry standard cubic meter
gr/dscf = grains per dry standard cubic foot
kg/Mg BLS = kilograms per megagram of black liquor solids
lb/ton BLS = pounds per ton of black liquor solids
ppmdv = parts per million dry volume

[b]EPA guidelines for TRS emissions from kraft pulp mills not subject to the NSPS; ppmdv = ppm dry on volume basis.

[c]If the weight percentage of pink liquor solids exceeds 7 percent of the total mixture of solids fired and the sulfidity of the resultant green liquor exceeds 28 percent, the recovery furnace is classified as a "cross-recovery furnace."[7,8]

[d]BLS = black liquor solids.

[e]EPA originally promulgated a TRS emission limit for BLO units of 5 ppm; this requirement was rescinded during the NSPS review.

TRS guidelines are 20 ppmdv at 8 percent O_2, which are higher than the NSPS limit. Similarly, the TRS guidelines for lime kilns (20 ppmdv at 10 percent O_2) are higher than the NSPS limit (8 ppmdv at 10 percent O_2). The TRS compounds are hydrogen sulfide, methyl mercaptan, dimethyl sulfide, and dimethyl disulfide. None of these TRS compounds are currently listed in the Clean Air Act for regulation under Section 112. However, PM emitted from kraft pulp mill sources does contain small amounts of HAP compounds (e.g., HAP metals).

State regulations for PM and TRS emissions for new, modified, and reconstructed recovery furnaces, lime kilns, and SDT's are the same as Federal regulations, with a few exceptions.[22] The State of Georgia has promulgated process-weighted standards for PM, which, for some Georgia mills, are slightly more stringent than the NSPS PM limits. Also, the State of Alabama requires new sources to use the "highest and best practicable controls for PM."[22] In addition, several States regulate toxic air emissions from pulp mills. These regulations, however, tend to limit the concentration of air toxics at the pulp mill fence line, and thus do not directly regulate emissions from specific equipment in the chemical recovery area of the mill. For existing recovery furnaces that are not subject to the NSPS, most States have emission standards equivalent to or more stringent than the EPA TRS emission limit guidelines.[22,23] For existing lime kilns that are not subject to the NSPS, approximately half of the States with kraft mills have emission standards equivalent to or more stringent than the EPA TRS emission limit guidelines.[22,23]

2.3.2 Baseline Emission Estimates

Section 2.3.2.1 presents the baseline gaseous HAP emission estimates from combustion sources at kraft and soda pulp and paper mills. Baseline gaseous HAP emissions were developed for DCE recovery furnaces, NDCE recovery furnaces, and BLO systems. Baseline gaseous HAP emissions were not developed for lime kilns and SDT's because the gaseous HAP emissions from these units result from the use of contaminated condensates in the makeup

waters. The regulation of contaminated condensates is part of the noncombustion source category for kraft and soda pulp and paper mills. Section 2.3.2.2 presents baseline PM and TRS emissions from combustion sources at kraft and soda pulp and paper mills.

2.3.2.1 Baseline Gaseous HAP Emissions. Table 2-4 presents gaseous HAP process emission factors for the two types of recovery furnaces (DCE and NDCE) used in the kraft chemical recovery loop. The process emission factors were calculated based on available emission test data and process emission rates for each of the recovery furnaces.[24] In the case of DCE recovery furnace systems, emission points include the recovery furnace/DCE stack and the BLO vent. The only emission point associated with NDCE recovery furnaces is the recovery furnace stack. As shown in Table 2-4, the gaseous HAP compounds emitted in the largest quantities from DCE and NDCE recovery furnaces are hydrochloric acid (HCl) and methanol. The following sections discuss the potential mechanisms by which HCl and gaseous organic HAP's such as methanol are emitted, as well as the nationwide emissions of these pollutants.

2.3.2.1.1 HCl emissions. Typically, HCl emissions from combustion processes are generated from oxidation reactions involving chlorinated organic compounds in the fuel, and all of the organic chlorine in the fuel is presumed to be emitted as HCl. However, chlorine in black liquor is expected to be in the form of inorganic chlorine. Consequently, only a small fraction (0 to 10 percent) of the black liquor chlorine is believed to be emitted from the recovery furnace as HCl.[25] A possible mechanism for HCl formation in the recovery furnace is gas phase reactions of volatilized alkali chlorides with SO_2 in the presence of water vapor and oxygen.[25,26] The proposed reactions are as follows:[25,26]

$$2NaCl(s) + SO_2(g) + 1/2\ O_2(g) + H_2O(g) \rightarrow 2HCl(g) + Na_2SO_4(s) \quad (1)$$

$$2KCl(s) + SO_2(g) + 1/2\ O_2(g) + H_2O(g) \rightarrow 2HCl(g) + K_2SO_4(s) \quad (2)$$

TABLE 2-4. HAP EMISSION FACTORS

HAP	Emission factors, mass/mass BLS fired				
	DCE recovery furnace systems		NDCE recovery furnaces		
	Air-sparging BLO system[a]	DCE recovery furnace	Dry ESP system[b]	Wet ESP system[c]	Average[d]
Acetaldehyde	1.48×10^{-5}	2.99×10^{-5}	1.52×10^{-5}	7.50×10^{-6}	1.03×10^{-5}
Benzene	1.21×10^{-6}	1.45×10^{-5}	1.73×10^{-6}	9.73×10^{-6}	7.97×10^{-6}
Formaldehyde	2.89×10^{-7}	2.82×10^{-6}	1.93×10^{-6}	2.98×10^{-6}	9.20×10^{-6}
Methyl ethyl ketone	4.24×10^{-6}	8.53×10^{-6}	1.80×10^{-6}	7.46×10^{-6}	6.59×10^{-6}
Methyl isobutyl ketone	1.77×10^{-6}	1.21×10^{-5}	2.27×10^{-6}	9.23×10^{-6}	8.17×10^{-6}
Methanol	1.73×10^{-4}	1.70×10^{-4}	4.93×10^{-6}	5.04×10^{-5}	3.71×10^{-5}
Phenol	1.93×10^{-6}	3.95×10^{-5}	2.05×10^{-6}	1.50×10^{-5}	1.62×10^{-5}
Toluene	1.44×10^{-6}	9.73×10^{-6}	2.03×10^{-6}	8.37×10^{-6}	7.41×10^{-6}
Xylenes	4.28×10^{-6}	2.47×10^{-5}	4.76×10^{-6}	1.92×10^{-5}	1.70×10^{-5}
Hydrochloric acid	N/A	1.20×10^{-4}	1.20×10^{-4}	1.20×10^{-4}	1.20×10^{-4}
Total HAP	2.03×10^{-4}	4.32×10^{-4}	1.57×10^{-4}	2.50×10^{-4}	2.40×10^{-4}

[a]For the BLO systems, the emission factors are based on equivalent BLS firing rates.
[b]Dry ESP systems include a dry-bottom ESP and a dry PM return system.
[c]Wet ESP systems include wet-bottom ESP's and those dry-bottom ESP's that use black liquor or HAP-contaminated water in the PM return system.
[d]The average represents the average emission factor calculated based on all available data for NDCE recovery furnaces.

These reactions are believed to occur at temperatures below about 900°C (1650°F) in the furnace post-combustion zone and in higher temperature areas in the ductwork downstream from the furnace.[25,26] Assuming the above reactions are the correct mechanism for HCl formation, HCl emissions will be higher for well-designed and operated combustion systems. Combustion system parameters that could potentially affect the amount of HCl formed include higher combustion zone temperatures leading to greater emissions of volatile sodium chloride (NaCl) and potassium chloride (KCl), furnace SO_2 generation rates, and good mixing of combustion gases with excess air in the post-combustion zone.

A review of the available emissions data indicates that HCl emissions from recovery furnaces are highly variable. A recent study of kraft recovery furnace HCl emissions conducted by the National Council of the Paper Industry and Air and Steam Improvement, Inc. (NCASI) examined the relationships between several recovery furnace operating parameters and HCl emissions.[27] Some earlier studies had suggested that there may be strong correlations between stack SO_2 emissions and HCl emissions and between stack O_2 concentration and HCl emissions.[25,26] The NCASI study, which was based on tests of 14 recovery furnaces conducted specifically for this study, examined the relationship between HCl emissions and each of the following parameters: stack SO_2 concentration, as-fired black liquor chloride content, stack O_2 levels, chloride content of precipitated ash, and furnace load or liquor firing rate.[27] Based only on the tests conducted in this study, NCASI found a strong correlation using a Langmuir adsorption-type relationship between HCl emissions and SO_2 emissions for NDCE recovery furnaces and found no clear correlation between these parameters for DCE recovery furnaces.[27] The study found no clear correlations between HCl emissions and other recovery furnace operating parameters.[27]

As shown in Table 2-4, the HCl emission factor presented for both NDCE and DCE recovery furnace systems is the same. Based on a statistical analysis of the available HCl emissions data, it

was concluded that the NDCE and DCE recovery furnace HCl emission factors should be averaged together. The range of HCl emission factors for NDCE recovery furnaces extends above and below the range of factors for DCE recovery furnaces. Both the 90 and 95 percent confidence intervals of the difference between the DCE and NDCE recovery furnace means include zero. Stated another way, the hypothesis that the means are equal cannot be rejected at the 0.1 and 0.05 levels of significance.[28]

2.3.2.1.2 Gaseous organic HAP emissions. Methanol and other gaseous organic HAP's are emitted from process units throughout the chemical recovery area of the pulp mill. The primary mechanism for gaseous organic HAP emissions from recovery furnace stacks is the stripping of gaseous organic HAP's from black liquor, which occurs when hot flue gases directly contact the black liquor. Gaseous organic HAP's are also emitted from the BLO vent as a result of stripping of these HAP's from the black liquor as the oxidizing air is forced up through the liquor.

The emission potential for DCE recovery furnace systems (which include the DCE recovery furnace and the BLO unit) is significantly higher than for NDCE recovery furnaces because of the increased opportunity for stripping of HAP compounds from black liquor in process equipment. In the DCE recovery furnace system, opportunities for stripping of HAP's exist in (1) air-based BLO systems, (2) DCE's, and (3) ESP systems (i.e., the ESP and the PM return system associated with the ESP). However, with NDCE recovery furnaces, the ESP system is the only process equipment where stripping of HAP's may occur.

Methanol emission test data are available for DCE recovery furnaces equipped with air-sparging BLO systems. These data show that the total amount of methanol emitted from the BLO vents and DCE recovery furnace stack (which includes any methanol stripped in the DCE and ESP) at a given mill will vary, as will the quantity of methanol emitted from each source. Tests conducted at one mill showed that methanol emissions from the two sources were related (i.e., if methanol emissions from the BLO vents were

high, then methanol emissions from the DCE recovery furnace stack were low).[29]

No emission data are available from DCE recovery furnaces with molecular oxygen BLO systems for comparison with DCE recovery furnaces with air-based BLO systems. Therefore, the effect of molecular oxygen BLO systems on total emissions from the DCE recovery furnace system is uncertain. Unlike air-based systems, molecular oxygen systems use pure oxygen, and thus, no diluents are introduced that could strip organic compounds from the black liquor; consequently, organic compounds not released from the black liquor during the oxidation process could be subsequently stripped, in theory, from the oxidized black liquor when the black liquor enters the ESP or direct contact evaporator.

Most DCE recovery furnaces and some older NDCE recovery furnaces are equipped with wet ESP systems. Either black liquor or HAP-contaminated process water may be present in the bottom of these ESP's and in the system used to transport the collected PM to the saltcake mix tank. Also, for wet ESP systems, the PM may be collected as a dry product in the bottom of the ESP (i.e., a dry-bottom ESP) but black liquor or HAP-contaminated water is used in the PM return system. Because there is contact between the hot recovery furnace flue gases and the black liquor or HAP-contaminated process water in wet ESP systems, gaseous organic HAP's are stripped from the liquor and emitted from the recovery furnace stack. Most newer NDCE recovery furnaces (i.e., those constructed after 1989) emit lesser quantities of methanol and other gaseous organic HAP's because they tend to be equipped with dry ESP systems (i.e., dry-bottom ESP's and dry PM return systems).[30]

Data are not available to determine the effects of converting a wet ESP system to a dry ESP system on overall gaseous HAP emissions for a DCE recovery furnace system. Gaseous organic HAP's that would have been stripped from the black liquor in the wet ESP system potentially could be stripped from the black liquor when it enters the DCE. Also, based on emission

data for NDCE recovery furnaces with wet ESP systems, the ESP is a less significant source of gaseous organic HAP emissions than the BLO or DCE. Data are available for NDCE recovery furnaces with both wet ESP systems and dry ESP systems such that the impacts that different ESP systems have on gaseous organic HAP emissions from NDCE recovery furnaces can be determined. These impacts can be more readily determined since the ESP represents the only opportunity where stripping can occur for NDCE recovery furnaces.

2.3.2.1.3 Nationwide gaseous HAP emissions. Table 2-5 presents the nationwide baseline gaseous HAP emissions from recovery furnaces. The nationwide baseline total gaseous HAP emissions from the estimated 128 NDCE recovery furnaces are estimated to be 13,700 megagrams per year (Mg/yr) (15,100 tons per year [ton/yr]). The nationwide baseline total gaseous HAP emissions from the estimated 83 DCE recovery furnace systems (which includes BLO systems) are estimated to be 15,700 Mg/yr (17,300 ton/yr). The baseline emissions shown in Table 2-5 were calculated assuming no control of gaseous HAP's (with the exception of the two BLO systems with emissions controlled via incineration) because gaseous HAP emissions from recovery furnaces are highly variable and largely uncontrolled. The baseline emissions were calculated by multiplying the gaseous HAP emission factors by the process emission rate in kg BLS/hr (lb BLS/hr) for each mill for which data were available and then extrapolating based on the available data to account for those mills where data were unavailable. The baseline emissions were also calculated based on the assumption that each emission source operates 24 hours per day (hr/d) for 351 days per year (d/yr).[24]

2.3.2.2 Baseline PM and TRS Emissions. Table 2-6 presents nationwide total PM, PM HAP, and TRS baseline emissions from each combustion source. The HAP portion of total PM consists of HAP metals and accounts for less than 1 percent of the total PM emitted on average; therefore, only the total estimated quantity of HAP metals are presented in Table 2-6, instead of presenting emission estimates for each individual HAP metal.[31] The baseline

TABLE 2-5a (Metric). NATIONWIDE BASELINE GASEOUS HAP EMISSION ESTIMATES[24]

Gaseous HAP	Nationwide emissions, Mg/yr[a]		
	DCE recovery furnace systems		NDCE recovery furnaces
	Air-sparging BLO units	DCE recovery furnaces	
Acetaldehyde	333	769	588
Benzene	27.3	373	454
Formaldehyde	6.5	72.6	525
Methyl ethyl ketone	95.3	220	376
Methyl isobutyl ketone	39.6	312	466
Methanol	3,890	4,390	2,120
Phenol	43.3	1,020	924
Toluene	32.3	251	422
Xylenes	96.1	637	968
Hydrochloric acid	N/A	3,090	6,840
Total gaseous HAP's	4,560	11,100	13,700

[a]The emissions in this table were calculated by multiplying the emissions in Table 2-5b by 0.9072.

TABLE 2-5b (English). NATIONWIDE BASELINE GASEOUS HAP EMISSION ESTIMATES[24]

Gaseous HAP	Nationwide emissions, ton/yr[a]		
	DCE recovery furnace systems		NDCE recovery furnaces
	Air-sparging BLO units[b]	DCE recovery furnaces	
Acetaldehyde	367	848	648
Benzene	30.0	411	501
Formaldehyde	7.2	80.0	578
Methyl ethyl ketone	105	242	414
Methyl isobutyl ketone	43.7	343	513
Methanol	4,290	4,840	2,330
Phenol	47.8	1,120	1,020
Toluene	35.6	276	465
Xylenes	106	702	1,070
Hydrochloric acid	N/A	3,410	7,540
Total gaseous HAP's[c]	5,030	12,300	15,100

[a]The nationwide emissions were calculated using the following equation:
Emissions, ton/yr = (emission factor, lb/lb BLS) x (nationwide BLS firing rate, lb/hr) x (24 hr/d) x (351 d/yr) x (1 ton/2,000 lb).
[b]Emissions from the two BLO units that are controlled via incineration were multiplied by 0.02 to account for an estimated 98 percent emission reduction.
[c]Numbers may not add exactly due to rounding.

PM and TRS emissions were calculated based on the sum of the PM or TRS emission rates calculated for each combustion source at each mill operating at the U.S. The baseline emissions were also calculated based on the assumption that each combustion source operates 24 hr/d for 351 d/yr.[32]

TABLE 2-6. NATIONWIDE BASELINE PM, PM HAP, AND TRS EMISSIONS

Combustion source	Nationwide emissions, Mg/yr (ton/yr)		
	Total PM	PM HAP's[a]	TRS[b]
NDCE recovery furnaces	29,400 (32,400)	54 (65)	658 (725)
DCE recovery furnaces	16,000 (17,600)	31 (35)	2,720 (3,000)
Smelt dissolving tanks	7,680 (8,470)	5 (5)	7,350 (8,100)
BLO units	--	--	658 (725)
Lime kilns	8,830 (9,730)	124 (136)	648 (714)
Total[c]	61,900 (68,200)	219 (241)	12,000 (13,300)

[a]On average, PM HAP's account for 0.2 percent of total PM emitted from recovery furnaces, 0.06 percent of total PM emitted from SDT's, and 1.4 percent of total PM emitted from lime kilns.[31]
[b]As H_2S.
[c]Numbers may not add exactly due to rounding.

2.4 REFERENCES FOR CHAPTER 2

1. Green, R. and G. Hough (eds.). Chemical Recovery in the Alkaline Pulping Processes. 3rd Edition. Prepared by the Alkaline Pulping Committee of the Pulp Manufacture Division. Atlanta, GA, TAPPI Press. 1992. 196 p.

2. Smook, G. Handbook for Pulp and Paper Technologists. Montreal, Quebec, Canada. Canadian Pulp and Paper Association. Atlanta, GA, TAPPI Press. 1987.

3. Memorandum from Nicholson, R., MRI, to Telander, J., EPA/MICG. June 13, 1996. Addendum to Summary of Responses to the 1992 NCASI "MACT" Survey.

4. Lockwood Post's Directory of the Pulp, Paper and Allied Trades, 1993. San Francisco, Miller Freeman Publications. 1993. 966 p.

5. Pulp, Paper, and Paperboard Industry--Background Information for Proposed Air Emission Standards, Manufacturing Processes at Kraft, Sulfite, Soda, and Semi-Chemical Mills. U. S. Environmental Protection Agency, Research Triangle Park, NC. Publication No. EPA-453/R-93-050a. October 1993.

6. Memorandum from Nicholson, R., MRI, to the project file. September 20, 1996. Semichemical Pulping Operations Co-Located at Kraft Pulp Mills.

7. U. S. Environmental Protection Agency. New Source Performance Standards for Kraft Pulp Mills. 43 FR 7568. Washington, DC. U.S. Government Printing Office. February 23, 1978.

8. U. S. Environmental Protection Agency. New Source Performance Standards for Kraft Pulp Mills. 51 FR 18538. Washington, D.C. U.S. Government Printing Office. May 20, 1986.

9. Review of New Source Performance Standards for Kraft Pulp Mills. U. S. Environmental Protection Agency. Research Triangle Park, NC. Publication No. EPA-450/3-83-017. September 1983. p. 4-7.

10. Stultz, S. and J. Kitto (eds.). Steam: Its Generation and Use. 40th Edition. Babcock & Wilcox. New York. 1992. Chapter 26.

11. Someshwar, A. and J. Pinkerton. Wood Processing Industry. In: Air Pollution Engineering Manual, Air & Waste Management Association. Buonicore, A. and W. Davis (eds.). New York, Van Norstrand Reinhold. 1992. pp. 835-849.

12. Proposed Standards of Performance for Kraft Pulp Mills. In: Standards Support and Environmental Impact Statement. Volume 1. U. S. Environmental Protection Agency. Research Triangle Park, NC. Publication No. EPA-450/2-76-014-a. September 1976.

13. Environmental Pollution Control: Pulp and Paper Industry, Part I, Air. U. S. Environmental Protection Agency. Cincinnati, OH. Publication No. EPA-625/7-76-001. October 1976.

14. Memorandum from Soltis, V., Nicholson, R., and Holloway, T., MRI, to Telander, J., EPA/MICG. July 29, 1994. Summary of Responses to the NCASI "MACT" Survey--Kraft and Soda Pulp Mills.

15. Barsin, J. Process Recovery Boiler Design Considerations. Babcock and Wilcox. (Presented at the Kraft Recovery Operations Seminar. Orlando, FL. February 10-14, 1986.)

16. Kirby, M. Economic and Process Considerations in the Use of Oxygen for Black Liquor Oxidation. Union Carbide Canada Limited. Ontario, Canada. (Presented at the 21st Pulp and Paper ABCP Annual Meeting. Sao Paulo, Brazil. November 21-25, 1988.) 10. p.

17. Cooper, H. TRS Control Implications Reviewed, Part II. Southern Pulp and Paper. April 1981. p. 42-42.

18. Memorandum from Soltis, V., MRI, to the project file. April 3, 1995. Kraft and Soda Pulp Mill Combustion Sources Data Base.

19. Walther, J. and H. Amberg. Crown Zellerbach Corp. A Positive Air Quality Control Program at a New Pulp Mill. Journal of the Air Pollution Control Association. 20(1):9-18. January 1970.

20. Reference 1, p. 156.

21. Memorandum from Ramsey, M., MRI, to Telander, J., EPA/ISB. January 27, 1992. Meeting minutes from the August 1, 1991 meeting between National Council of the Paper Industry for Air and Stream Improvement, Inc., American Paper Institute, the U. S. Environmental Protection Agency, and Midwest Research Institute.

22. Memorandum from Subramarian, S. and Nicholson, R., MRI, to the project file. September 13, 1993. State Regulations on Particulate Matter Emissions from Kraft Lime Kilns, Recovery Furnaces, and Smelt Dissolving Tanks.

23. Memorandum from Gideon, L. and Watkins, S., Radian Corp., to Lassiter, P., EPA/CPB. January 22, 1992. Summary of Regulatory Survey for the Pulp and Paper NESHAP.

24. Memorandum from Randall, D., Jones, R., and Nicholson, R., MRI, to Telander, J., EPA/MICG. April 25, 1995. Nationwide Baseline Emissions for Combustion Sources of Kraft and Soda Pulp Mills.

25. Someshwar, A. and A. Jain. Emissions of Hydrochloric Acid from Kraft Recovery Furnaces. National Council of the Paper Industry for Air and Stream Improvement, Inc. Gainesville, FL. (Presented at the 1992 TAPPI International Chemical Recovery Conference. Seattle, WA. June 18, 1992.) 14 p.

26. Warnqvist, B. and H. Norrstrom. Chlorides in the Recovery Boiler and Mechanism for Chloride Removal. TAPPI. 59(11):89. 1976.

27. Someshwar, A. A Study of Kraft Recovery Furnace Hydrochloric Acid Emissions. National Council of the Paper Industry and Air and Steam Improvement, Inc. New York. Technical Bulletin No. 674. August 1994.

28. Memorandum from Randall, D., MRI. to the project file. November 8, 1994. Comparison of EPA and NCASI HCl Emission Factors for Recovery Furnaces.

29. Telecon. Ramsey, M., MRI, with Morris, T., Westvaco Corp., Covington, VA. June 17 and 25, 1993. Supplemental information for Westvaco emission test reports.

30. Memorandum from Nicholson, R., MRI to the project file. March 31, 1995. Methanol Emissions from Kraft NDCE Recovery Furnaces.

31. Memorandum from Holloway, T., MRI, to the project file. June 14, 1996. Summary of PM and HAP Metals Data.

32. Memorandum from Holloway, T., MRI, to the project file. May 20, 1996. Nationwide Baseline PM and TRS Emissions.

3.0 EMISSION CONTROL TECHNIQUES

This chapter discusses emission control techniques for reducing HAP emissions from recovery furnaces, BLO systems, SDT's, and lime kilns located at kraft pulp and paper mills. Recovery furnaces include DCE recovery furnaces and NDCE recovery furnaces. Add-on controls and equipment changes/modifications are discussed in Sections 3.1 and 3.2, respectively. Both PM and gaseous HAP control techniques are included in this discussion. Section 3.3 contains the references cited in this chapter.

Tables 3-1 and 3-2 present applicable control techniques for reducing PM and gaseous HAP emissions, respectively, from the aforementioned combustion sources located at kraft pulp and paper mills. These control techniques and their effectiveness in reducing PM and gaseous HAP emissions are discussed below.

3.1 ADD-ON CONTROLS

This section describes the various add-on air pollution control equipment that can be applied to the combustion sources located in the chemical recovery areas at kraft pulp and paper mills. Sections 3.1.1, 3.1.2, and 3.1.3 describe ESP's, wet scrubbers, and BLO control, respectively; each section presents a description of the control equipment and describes the performance of the equipment in controlling HAP (PM and/or gaseous) emissions from the applicable combustion sources.

3.1.1 Electrostatic Precipitators

Electrostatic precipitators are a demonstrated control technique for reducing PM emissions from kraft recovery furnaces and lime kilns. The PM emissions from over 98 percent of kraft recovery furnaces and approximately 10 percent of kraft lime kilns are controlled with ESP's.[1] A general description of the

TABLE 3-1. PARTICULATE MATTER HAP CONTROL TECHNIQUES

Source	Control technique	Percent PM reduction
Recovery furnace	ESP	99+
Lime kiln	ESP	99+
	Venturi scrubber	99+
SDT	Wet scrubber	99+

TABLE 3-2. GASEOUS HAP CONTROL TECHNIQUES

Source	Control technique	Percent reduction	
		HCl	Other HAP's
NDCE recovery furnace systems	Packed-bed scrubber	Up to 99% or 5 ppmv[a]	--[b]
	Elimination of black liquor from ESP system	--	72%
DCE recovery furnace system (includes air-sparging BLO system)	Packed-bed scrubber	Up to 99% or 5 ppmv[a]	--[b]
	Conversion to NDCE recovery furnace with dry- ESP system[c]	--[d]	93%
	Incineration of BLO vent gases	--[e]	38%[f]
Lime kiln	Control makeup water quality in causticizing area	--[g]	
SDT	Control makeup water quality in causticizing area	--[g]	

[a]Control efficiency depends on inlet HCl concentration. Based on manufacturers' guarantees, scrubbers can achieve either a 99 percent control efficiency or an outlet concentration of 5 ppmv.
[b]Insufficient data are available to determine the emission reductions of other HAP's.
[c]Conversion includes replacement of the DCE recovery furnace with an NDCE recovery furnace and elimination of the BLO unit.
[d]HCl emissions are highly variable. Available data indicate no overall change in HCl emissions is expected from converting a DCE recovery furnace system to an NDCE recovery furnace.
[e]HCl is not emitted from BLO systems.
[f]The overall percent reduction is based on average HAP emission factors for DCE recovery furnaces and air-sparging BLO systems, and on a control efficiency of 98 percent for HAP's emitted from the BLO system.
[g]Wastewater control option performance is addressed in a separate wastewater regulatory development program.

types of ESP's used to control PM emissions from kraft recovery furnaces and lime kilns is provided in Section 3.1.1.1, below. The performance of these ESP's is discussed in Section 3.1.1.2.

3.1.1.1 Description. The ESP's used to control PM emissions from kraft recovery furnaces and lime kilns are generally classified as plate-wire ESP's. In plate-wire ESP's, the flue gas flows between parallel sheet metal plates and high-voltage electrodes. The flue gas passes between collecting plates into a field of ions that have been negatively charged by the high-voltage electrodes located between the plates.[2] Each paired set of electrodes and plates forms a separate electrostatic field within the ESP. Electrostatic precipitators used to control PM emissions from kraft recovery furnaces typically have two parallel precipitator chambers (i.e., flue gas passages) with three or four electrostatic fields per chamber. Lime kiln ESP's typically have one chamber with two or three electrostatic fields.[1]

As the flue gas passes through each electrostatic field, the particles suspended in the flue gas are bombarded by the ions, imparting a negative charge to the particles. The negatively charged particles then migrate towards the positively charged or grounded "collecting" plates, where the particles transfer a portion of their charge, depending upon their resistivity. The particles are kept on the collecting plates by the electrostatic field and the remaining charge.[3] At periodic intervals, the collection plates are knocked ("rapped"), and the accumulated PM falls into the bottom of the ESP. The recovered PM is subsequently recycled to the black liquor in recovery furnace applications or, in lime kiln applications, fed back to the kiln. The ESP's used on recovery furnaces may be designed with either a wet or dry bottom. In wet-bottom ESP's, the collected PM falls directly into a pool of liquid, which may be black liquor or water, in the bottom of the ESP. In dry-bottom ESP's, the collected PM falls to the (dry) bottom of the ESP and is transferred from the ESP bottom to a mix tank (containing black liquor) via drag-chain or screw conveyors. Black liquor is

sometimes used to transport the dry collected PM to the mix tank. More recent ESP installations employ a dry PM return system to transport the PM to the mix tank. Because the PM removed by the ESP is recycled to the black liquor in the mix tank, the ESP is an integral part of the chemical recovery loop as well as an air pollution control device.

The design of the plate-wire ESP's used to control PM emissions from recovery furnaces and lime kilns may include either weighted-wire electrodes cr rigid electrodes. With the weighted-wire design, the wire electrodes are suspended inside the ESP, and weights are attached to the wires to maintain tension. In the rigid-electrode design, the discharge electrodes are rigid tubes with pointed corona emitters welded to the surface; each tube replaces two weighted wires. Although the weighted-wire design has been available for more than 50 years, rigid-electrode ESP's have only been available since the late 1970's. The rigid-electrode ESP represents the current stage of development in ESP technology and offers the following advantages over the weighted-wire design: a higher tolerance of in-service abuse (no wires to break), better collection efficiencies, and better cleaning characteristics.[4] According to one manufacturer, all new ESP installations in the pulp and paper industry since 1990 have rigid-electrode designs.[5] Additionally, since 1981, 64 of the 66 (97 percent) recovery furnace ESP installations supplied by this manufacturer have been of the rigid-electrode design, and approximately 80 percent of the lime kiln ESP applications use rigid-electrode designs.[5] Figure 3-1 is a schematic of a rigid-electrode, plate-wire ESP.

The average lifetime of an ESP in service on a kraft recovery furnace varies depending upon the type of ESP bottom (i.e., wet vs. dry), the inlet temperature of the gases, and maintenance practices. The lifetime of ESP's used to control PM emissions from recovery furnaces with NDCE's, which tend to operate with dry-bottom ESP's, typically ranges from 12 to 15 years.[4] After that point, major repairs or a rebuild may be required. Recovery furnaces with DCE's tend to have cooler inlet

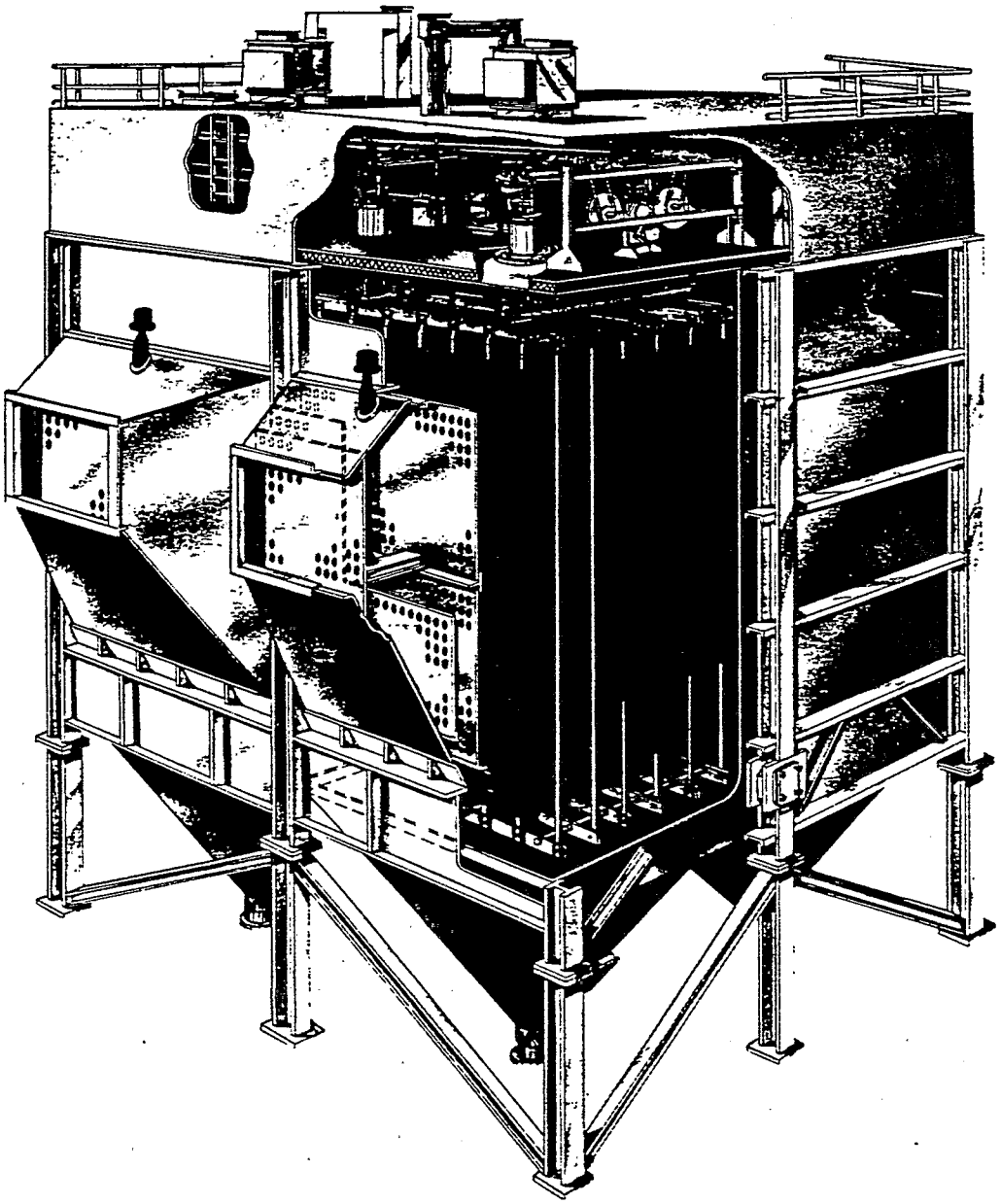

Figure 3-1. Rigid-electrode ESP.

gases and wet-bottom ESP's; these two factors promote corrosion through condensation of acid gases and shorten the life of the ESP. The lifetime of an ESP on a DCE recovery furnace typically ranges from 7 to 10 years.[4]

Lime kiln ESP's operate in a "milder" environment (i.e., hotter temperatures prevent any acid gases from condensing and corroding the ESP, and the primary constituent of the PM collected is lime, which creates an alkaline environment that further protects the ESP from acid gas corrosion). Therefore, lime kiln ESP's have fewer corrosion problems than do recovery furnace ESP's. The expected lifetime of a lime kiln ESP is typically more than 15 years.[6]

The size of the ESP is often expressed in terms of the specific collecting area (SCA), which is defined as the total collecting plate surface area divided by the flue gas flow rate. Specific collecting areas of ESP's used to control PM emissions from kraft recovery furnaces and lime kilns typically range from about 39 to 160 square meters per cubic meter per second ($m^2/[m^3/sec]$) (200 to 800 square feet per 1,000 actual cubic feet per minute [ft^2/1,000 acfm]).[1] The SCA's of ESP's used to control emissions from DCE recovery furnaces tend to be somewhat lower than those associated with NDCE recovery furnaces. The average SCA for ESP's used to control emissions from DCE recovery furnaces is approximately 15 percent lower than for NDCE recovery furnaces.[1] The primary reason for the difference is that the DCE removes 20 to 40 percent of the PM before the ESP; therefore, the inlet loading of PM to the ESP operating on a DCE recovery furnace is lower than that of an NDCE recovery furnace.[7] The lower SCA's for DCE recovery furnaces may also be because DCE recovery furnaces tend to be older than NDCE recovery furnaces, and most are not subject to the NSPS for Kraft Pulp Mills.

3.1.1.2 Performance. Electrostatic precipitators can control PM HAP emissions but provide no control of gaseous HAP emissions. Properly designed and operated ESP's used on kraft recovery furnaces and lime kilns routinely achieve PM removal efficiencies of 99 percent or greater. Although emission test

data from recovery furnace ESP's on PM HAP performance are limited, data collected from other combustion sources on the relative performance of ESP's for PM and metals indicate that those systems that achieve the greatest PM removal also provide the best performance for the HAP portion of the PM.[8] Therefore, total PM performance can be used as a surrogate for PM HAP's.

Because performance guarantees made by ESP manufacturers and permit limits are usually in the form of outlet PM emission concentrations rather than in the form of achievable control efficiencies, outlet concentrations will be the focus of the discussion of the performance of ESP's on kraft recovery furnaces and lime kilns in the following sections. Following the format of the NSPS, all PM concentrations for recovery furnace emissions are corrected to 8 percent O_2, and those for lime kiln emissions are corrected to 10 percent O_2. The discussion below also addresses the factors that affect ESP performance.

3.1.1.2.1 Recovery furnaces. The PM emitted from the recovery furnace and subsequently recovered by the ESP is primarily composed of sodium sulfate (i.e., "saltcake") and sodium carbonate, neither of which are HAP's. Particulate matter HAP's, which are primarily metals (although small quantities of high-boiling-point semivolatile organic compounds also may be emitted in particle form), are emitted from the ESP in small quantities. For any single metal, the quantity emitted from either NDCE or DCE recovery furnaces is typically less than 140 kg/yr (300 lb/yr).[9] The paragraphs below first discuss the outlet concentrations achievable by ESP's based on information compiled as a part of the previous NSPS development effort and on long-term performance data compiled by the State of Washington. Then factors that can affect ESP performance are described.

The NSPS for kraft recovery furnaces requires that PM emissions from recovery furnaces constructed, reconstructed, or modified after September 24, 1976 be less than or equal to 0.10 grams per dry standard cubic meter (g/dscm) (0.044 grains per dry standard cubic foot [gr/dscf] of flue gas).[10,11] Based on installation dates, approximately 39 percent of NDCE recovery

furnaces (i.e., 50 NDCE recovery furnaces) and 4 percent of DCE recovery furnaces (i.e., 3 DCE recovery furnaces) are subject to the NSPS.[12] In addition, some NDCE and DCE recovery furnaces not subject to the NSPS are subject to State PM permit limits that are less than or equal to the NSPS limit.[13] Therefore, about 25 percent of the 211 existing recovery furnaces nationwide are required by regulation to emit PM in concentrations less than or equal to 0.10 g/dscm (0.044 gr/dscf).

The NSPS emission limit was established based on the performance of a weighted-wire ESP design, taking into consideration maintenance issues such as wire breakage, the corrosive and sticky nature of the PM removed, any long-term deterioration in ESP performance, and the inherent variability of ESP performance.[14] In considering these issues, EPA found that ESP's that formed the basis for the standard were installed with added insulation or heated shells to maintain the gas temperature throughout the precipitator to prevent corrosion.[14] Also, although some amount of wire breakage was noted, most occurred soon after start-up and lessened in frequency with more operating time. A loss of 5 to 10 percent of the wires was determined to have no noticeable effect on the performance of the ESP.[14] Finally, about long-term deterioration in ESP performance, manufacturers of ESP's were emphatic in stating that a properly maintained precipitator should not deteriorate over the expected life of the unit.[14] Based on a review of all available information collected from both the industry and control device manufacturers, ESP systems were determined to be capable of meeting an emission limit of 0.10 g/dscm (0.044 gr/dscf) on a continuous, long-term basis.[14]

More recent information obtained from ESP manufacturers indicates that since the NSPS was promulgated, the industry has demanded further improvements to ensure continued compliance with PM emission limits associated with the NSPS or State permit requirements. Mills often require that ESP manufacturers provide emission limit guarantees such that the furnace can meet the permitted limit (e.g., 0.10 g/dscm [0.044 gr/dscf]) with only one

chamber of the ESP in operation. Two ESP manufacturers commonly guarantee that the permitted emission limit will not be exceeded while the recovery furnace is operating at 70 percent capacity with one chamber of the ESP in operation.[4,6] Requests by mills for these types of guarantees began in the late 1980's and have resulted in the installation of ESP's that are sized to ensure that these emission limits are achieved continuously. This design provides mills with the advantage of being able to make ESP repairs to one chamber of the ESP without disrupting the operation of the furnace, and thus avoiding production losses while maintaining the integrity of the ESP. In addition, ESP's that meet these design criteria routinely achieve PM outlet emissions at levels less than or equal to half the NSPS (i.e., 0.050 g/dscm [0.022 gr/dscf]) under normal operating conditions.[4]

The information supplied by manufacturers is supported by long-term PM emission test data for recovery furnace ESP's supplied by the State of Washington.[13] Figures 3-2 through 3-4 show monthly PM emission data for two kraft NDCE recovery furnaces with older ESP's over a 6-year period from 1988 to 1994 and one kraft NDCE recovery furnace with a newer ESP over a 5½-year period from 1990 to 1995. The figures provide an indication of concentrations achievable by older systems designed to meet the NSPS limits and by newer ESP's with improved design. The data also demonstrate the variability in ESP performance.

The data in Figure 3-2 were generated by an ESP of weighted-wire design that was designed to comply with a permit limit of 0.10 g/dscm (0.044 gr/dscf). The ESP was installed in 1974 and has a design and operating SCA of about 92 $m^2/(m^3/sec)$ (467 ft^2/1,000 acfm).[15]

The data in Figure 3-3 were generated by another ESP of weighted-wire design that was designed to comply with a permit limit of 0.10 g/dscm (0.044 gr/dscf). The recovery furnace was originally installed as a DCE recovery furnace in 1975 and converted to an NDCE recovery furnace in 1982.[15] The ESP was last upgraded in 1982 and has a design and operating SCA of about 75 $m^2/(m^3/sec)$ (383 ft^2/1,000 acfm).[15] The control system is

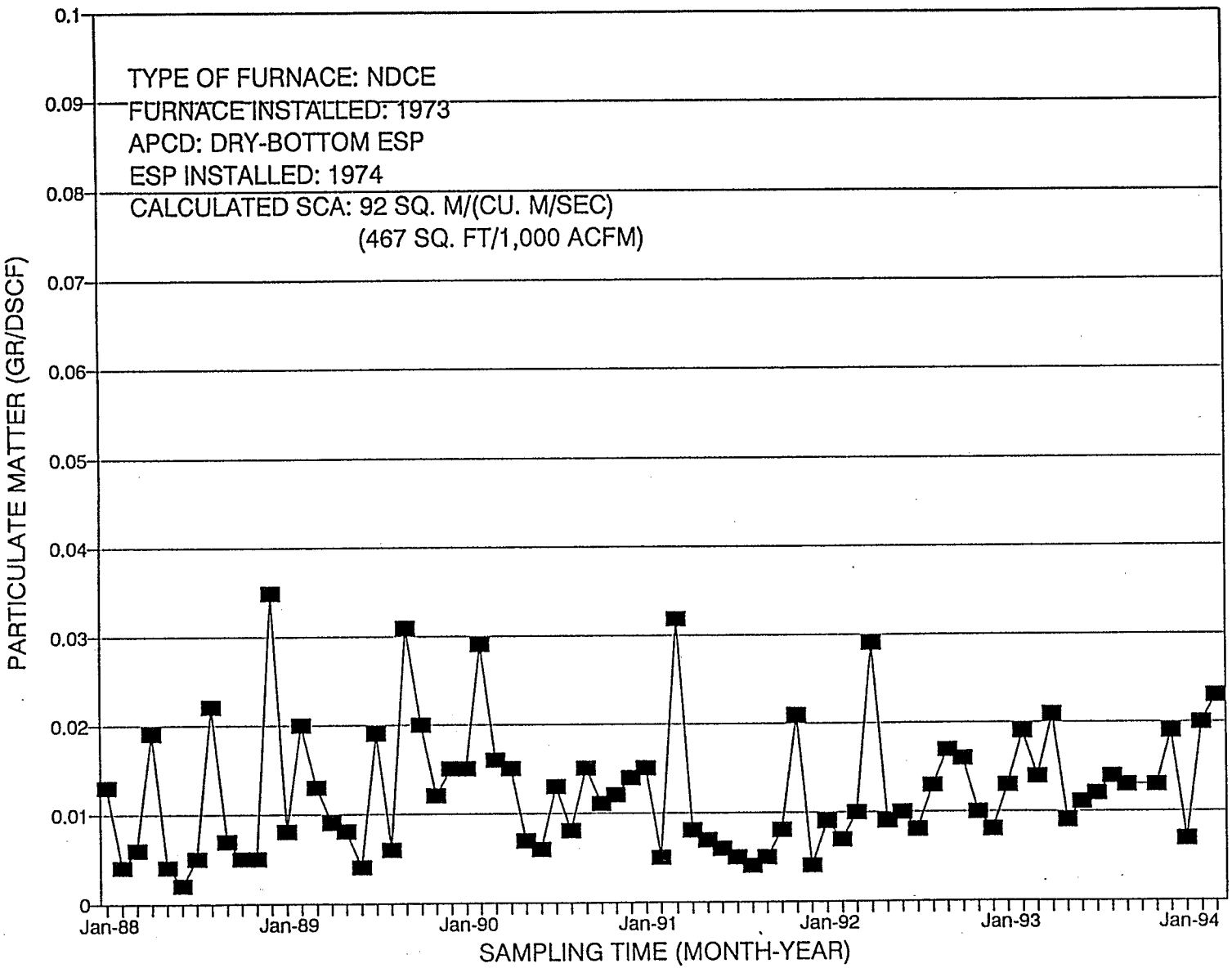

Figure 3-2. PM emission data for recovery furnace No. 3 at Mill A.

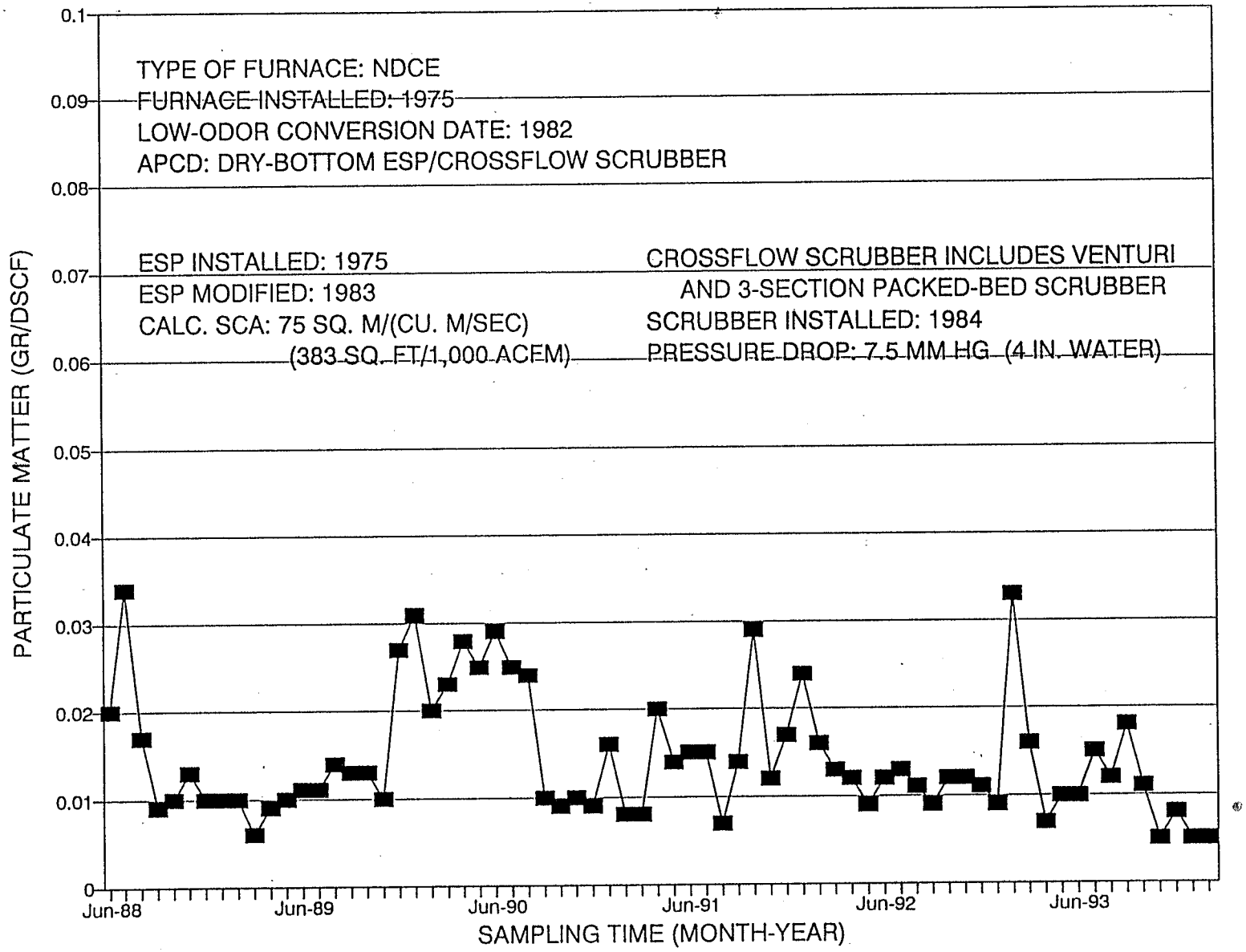

Figure 3-3. PM emission data for recovery furnace No. 4 at Mill B.

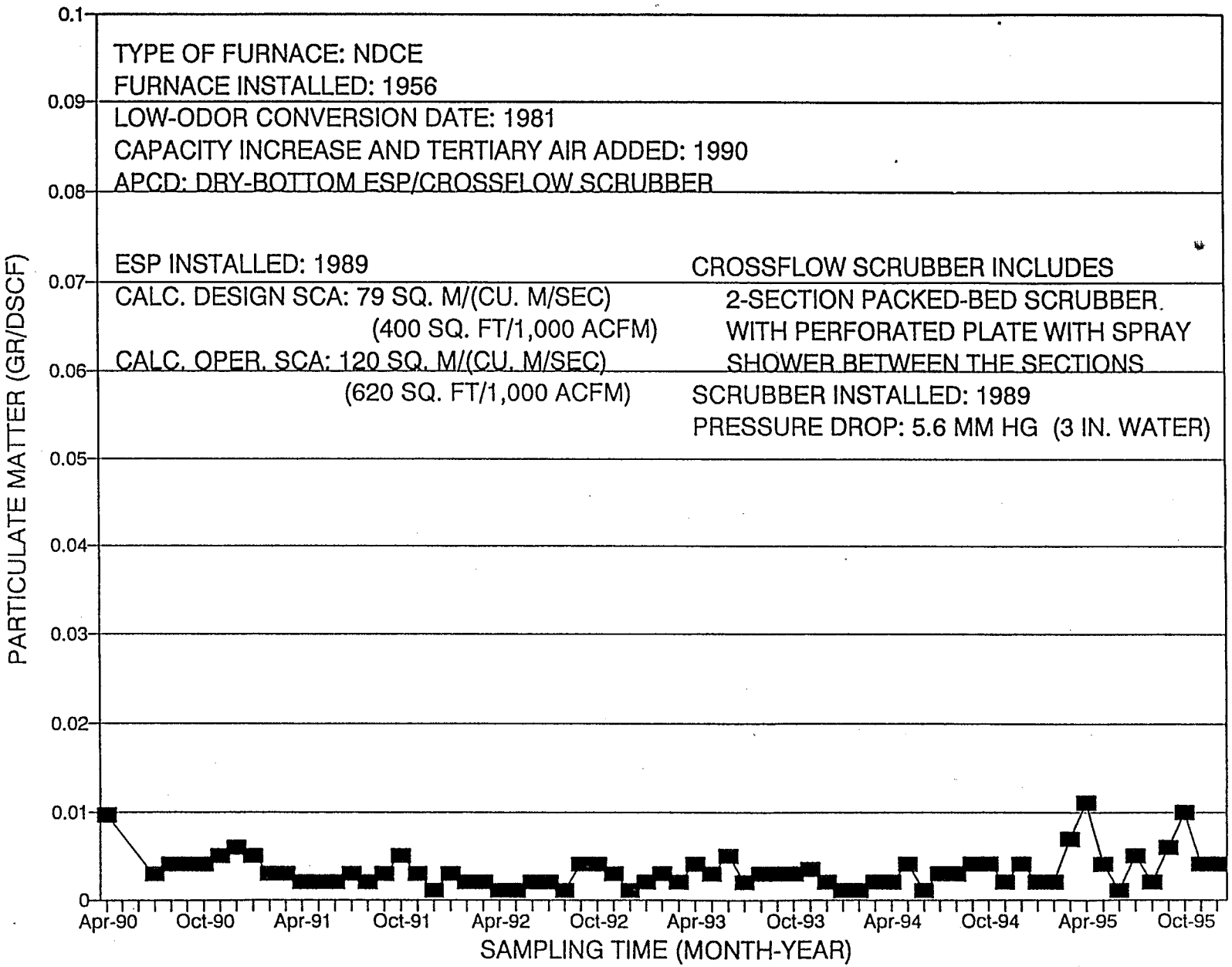

Figure 3-4. PM emission data for recovery furnace No. 3 at Mill B.

augmented with a combination low-energy venturi scrubber and packed-bed scrubber in series installed in 1984.[15] The scrubbers provide some added PM removal and act as a heat-transfer unit to provide hot water.[16]

The data in Figures 3-2 and 3-3 indicate that although emissions are quite variable, these units can readily meet the NSPS limit of 0.10 g/dscm (0.044 gr/dscf) on a long-term, continuous basis.

Outlet PM emission concentrations at least as low as 0.002 g/dscm (0.001 gr/dscf) for recovery furnaces with newer ESP's have been documented in the Washington data as well as in the data provided by individual pulp mills.[13,15] However, the ability of an ESP to maintain emissions on a continuous basis at levels substantially below 0.034 g/dscm (0.015 gr/dscf) has not been demonstrated.[13] However, the data in Figure 3-4 do demonstrate that a system can readily meet a level of 0.034 g/dscm (0.015 gr/dscf) on a long-term, continuous basis. These data were generated by an NDCE recovery furnace with a rigid electrode ESP followed by a packed-bed scrubber. The recovery furnace was originally installed as a DCE recovery furnace in 1956 and converted to an NDCE recovery furnace in 1981.[15] Both the ESP and the scrubber were installed in 1989.[15]

While the data in Figure 3-4 for recovery furnace No. 3 at Mill B are somewhat variable, they show considerably less variability and consistently lower emissions (less than 0.034 g/dscm [0.015 gr/dscf]) than the data associated with the two older ESP's. The ESP on recovery furnace No. 3 at Mill B was designed to achieve an outlet concentration of 0.069 g/dscm (0.033 gr/dscf). Furthermore, although the system was designed with an SCA of 79 $m^2/(m^3/sec)$ (400 $ft^2/1,000$ acfm), the furnace has operated at about 60 to 70 percent capacity over its life.[16] Consequently, the operating SCA of the unit is estimated to be in the range of 110 to 130 $m^2/(m^3/sec)$ (570 to 670 $ft^2/1,000$ acfm) or 120 $m^2/(m^3/sec)$ (620 $ft^2/1,000$ acfm), on average. These data demonstrate that systems comprising a recent ESP with an SCA of about 120 $m^2/(m^3/sec)$ (620 $ft^2/1,000$ acfm) followed by a packed-

bed scrubber can achieve a maximum outlet concentration of 0.034 g/dscm (0.015 gr/dscf) on a long-term, continuous basis.

No single parameter in and of itself is a good predictor of ESP performance. Both the design of the ESP and operating variables affect the performance of the ESP. Although the SCA may provide some indication of the expected efficiency of the ESP, other design features, such as the location of the rappers and plate spacing (which differs from one ESP manufacturer to another), also affect the ESP's performance. For example, differences in plate spacing from one manufacturer to another have resulted in different manufacturers requiring different SCA's for the same outlet PM concentration.[2,4] Another factor that complicates assessment of the effect of SCA on performance is that the PM emission concentrations for newer ESP's may be measured based on gas flow through one chamber with the recovery furnace operating at about 70 percent capacity.[4] The mechanism used to report emissions and calculate the SCA (i.e., based on gas flow at 70 or 100 percent capacity) can mask any relationship between the SCA and PM emissions.[4] Consequently, available data for recovery furnaces show only a limited relationship between outlet concentration and SCA.

The power input to the ESP can be a useful parameter in monitoring ESP performance. The value of the power input for each field and for the total ESP indicates how much work is being done to collect the PM. The power supply to the ESP consists of four basic components: (1) a step-up transformer (to step up the line voltage), (2) a high-voltage rectifier (to convert alternating current [a.c.] voltage to direct current [d.c.]), (3) a control element, and (4) a control system sensor.[17] Most ESP's installed since the 1980's are equipped with primary voltage and current meters on the low-voltage (a.c.) side of the transformer and secondary voltage and current meters on the high-voltage rectifier (d.c.) side of the transformer.[17] The power input for each transformer-rectifier (T-R) set is calculated by multiplying the voltage and current values of either the primary or secondary side of the transformer. The ESP control system is

designed to provide the highest possible voltage without causing sustained sparking between the discharge electrode and the collection surface. Spark meters can be used to set the maximum voltage/current levels for efficient operation. When the resistivity of the particulate is normal to moderately high, ESP performance improves as the total power input increases. However, when particulate resistivity is either low or very high or when spark rates are very high, improving the performance of the ESP may require decreasing the power input.[17] In addition, newer, high-efficiency ESP's that are generously sized and sectionalized are relatively unaffected by changes in power input. The normal power input of some of these newer and larger ESP's may be reduced by one-half to two-thirds without substantially affecting performance.[17] Therefore, power input cannot be used as a sole indicator of ESP performance.

Other operating variables that reportedly impact the performance of the ESP include temperature of the flue gas and maintenance practices. When flue gas temperatures exceed 260°C (500°F), the ESP's PM collection efficiency starts to decline. At low temperatures (less than 150°C [300°F]), the ESP may perform as desired, but corrosion will be accelerated (due to condensation of acid gases).[4] To obtain optimal control on a continuous basis, the system should be operated within this temperature envelope.

Poor operating and maintenance practices may also result in a decline in ESP performance, and differences in operating and maintenance practices are expected to contribute to the variability in emissions from mill to mill and over time at the same mill. Repairs (e.g., replacing broken wires) are often delayed, and routine on-line maintenance is sometimes neglected. Mills commonly solicit bids for maintenance and repair work from ESP manufacturers and contracting firms that provide ESP maintenance services.[4]

3.1.1.2.2 Lime kilns. Although venturi scrubbers have traditionally been the most common PM control device used on kraft lime kilns, the use of ESP's to control PM emissions from

lime kilns has steadily increased since about 1980. The ESP is generally mounted on top of the lime kiln feed building, and the captured dry PM is rerouted to the kiln by gravity feed.[18] The trend towards ESP's as PM control devices at new lime kiln installations and as replacement control devices for older scrubbers is primarily related to the lower energy costs, reduced maintenance, and increased reliability of the ESP's in comparison to venturi scrubbers that provide equivalent control. An added benefit is that lime kiln ESP's produce a dry product that can be recycled directly to the kiln. The wastewater produced by the venturi scrubber is typically recycled to the mud washers before the kiln to recover the lime particulate in the spent scrubbing fluid. Additional energy is needed to remove the excess water in the lime mud filter and to complete evaporation in the kiln. (Additional information about venturi scrubbers is provided in Section 3.1.2.2.)

The NSPS for kraft lime kilns requires that PM emissions from gas-fired and oil-fired lime kilns constructed, reconstructed, or modified after September 24, 1976 be less than or equal to 0.15 and 0.30 g/dscm (0.067 and 0.13 gr/dscf), respectively.[10,11] Based on installation dates, approximately 28 percent of existing lime kilns are subject to the NSPS.[1]

The PM emissions reported by the industry for lime kiln ESP's show less variability month to month and kiln to kiln than those associated with recovery furnace ESP's. Point-in-time PM data were reported for nine facilities with ESP's controlling lime kilns subject to the NSPS; these PM data showed outlet concentrations ranging from 0.0057 to 0.059 g/dscm (0.0025 to 0.026 gr/dscf).[15]

These reported data are confirmed by monthly PM emission data from the State of Washington for two gas-fired lime kilns with ESP's at Mills C and D.[13] Both kilns are subject to the NSPS; the ESP's at Mills C and D were installed in 1982 and 1985, respectively.[15]

As shown in Figure 3-5, the PM emission data from the kiln at Mill C vary from 0.0025 to 0.0519 g/dscm (0.0011 to

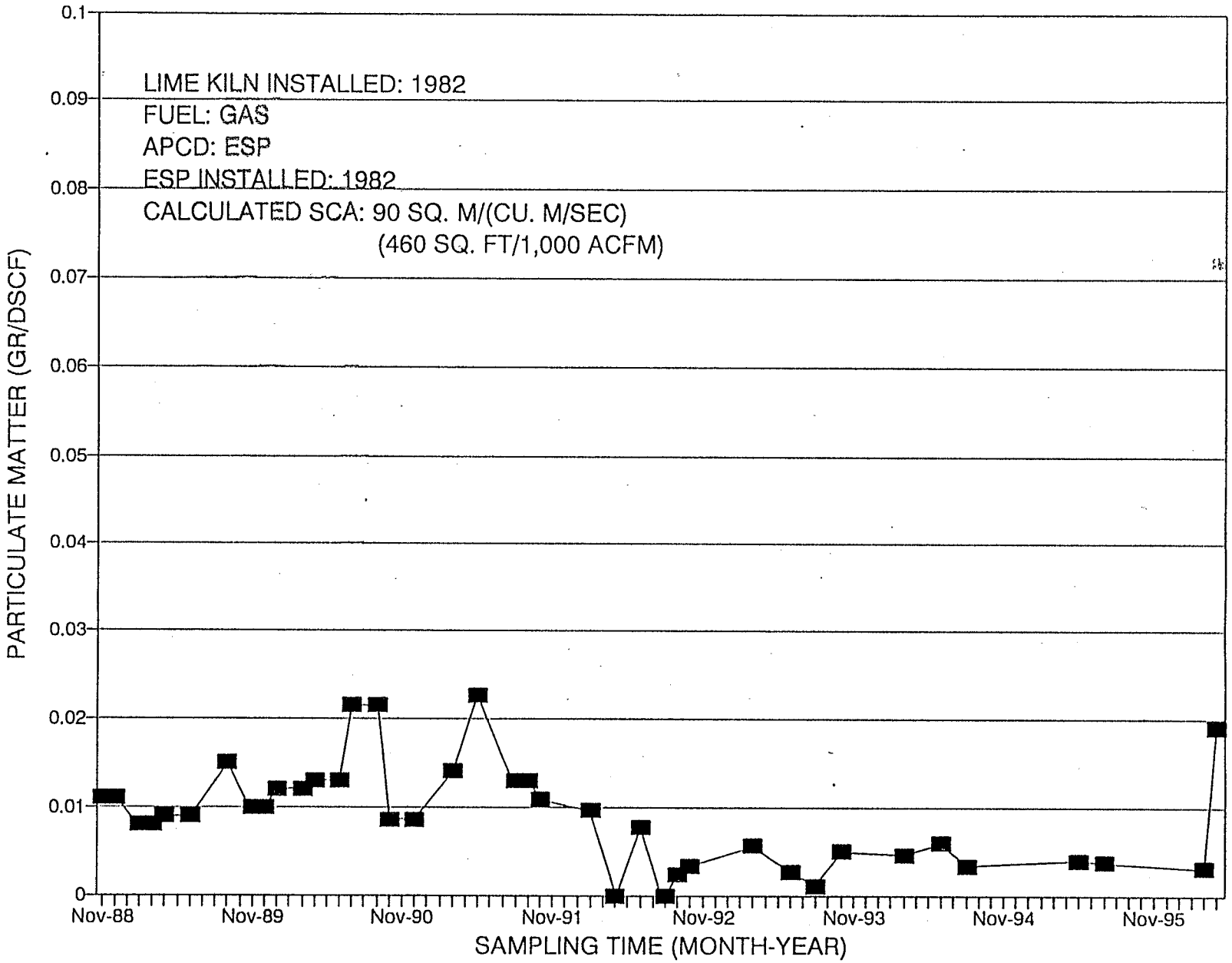

Figure 3-5. PM emission data for lime kiln No. 5 at Mill C.

0.0227 gr/dscf) during the period from November 1988 to April 1996. The SCA of the ESP on this lime kiln is 90 $m^2/(m^3/sec)$ (460 ft^2/1,000 acfm).[15] The data for Mill C confirm that ESP's on existing lime kilns can readily meet the NSPS limit of 0.15 g/dscm (0.067 gr/dscf) on a long-term, continuous basis.

The lime kiln at Mill D (shown in Figure 3-6) consistently emitted PM at or below 0.018 g/dscm (0.008 gr/dscf) during the 7-year period from March 1989 to April 1996. The SCA for the lime kiln ESP at Mill D is 220 $m^2/(m^3/sec)$ (1,120 ft^2/ 1,000 acfm).[15] The monthly PM emission data from Mill D demonstrate that new ESP's on lime kilns can achieve a maximum outlet concentration of 0.023 g/dscm (0.010 gr/dscf) on a long-term, continuous basis.

3.1.2 Wet Scrubbers

The following discussion focuses on the control of gaseous HAP emissions using packed-bed scrubbers and PM HAP emissions using venturi scrubbers. Section 3.1.2.1 discusses packed-bed scrubbers as the primary mechanism of HCl control; the control of gaseous organic HAP's by packed-bed systems will also be discussed. Section 3.1.2.2 discusses venturi scrubbers as a means of controlling PM emissions. In addition to venturi scrubbers, the pulp and paper industry uses other types of wet scrubbers, such as impingement plate, cyclone, flooded disc, and packed-bed, to control PM emissions from lime kilns and SDT's. However, venturi scrubbers are the most prevalent type of wet scrubber used to control PM from these sources, and the other scrubber types do not control PM emissions more effectively than venturi scrubbers.

3.1.2.1 Packed-Bed Scrubbers. Test data indicate that both NDCE and DCE kraft recovery furnaces emit HCl in the furnace exhaust gas at concentrations as high as 100 parts per million by volume (ppmv), corrected to 8 percent O_2.[19,20] These exhaust streams also contain SO_2 at concentrations as high as 979 ppmv, corrected to 8 percent O_2.[1,19] Typically, facilities that use add-on control measures for acid gas emissions use packed-bed wet

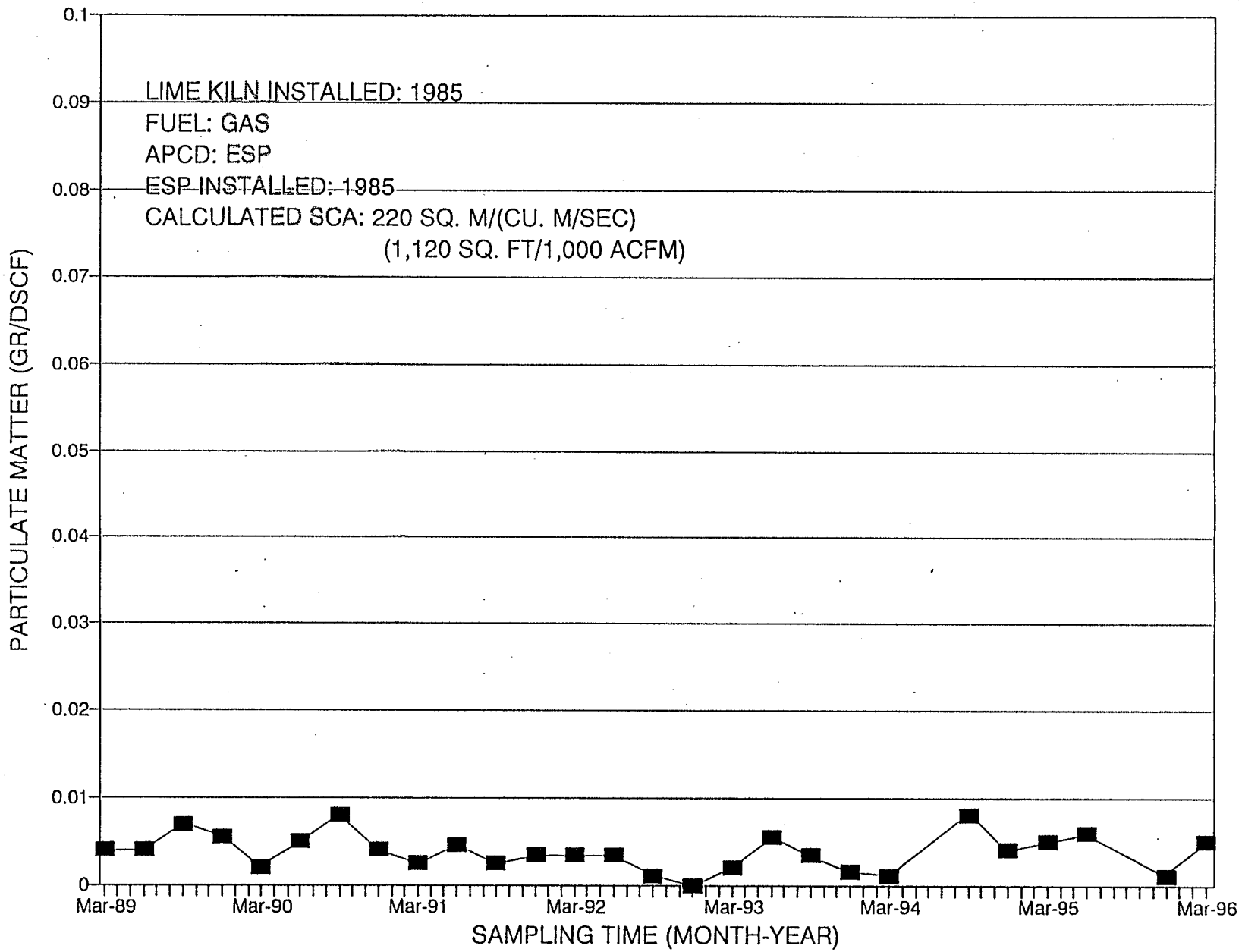

Figure 3-6. PM emission data for lime kiln No. 1 at Mill D.

scrubbers. Although the packed-bed scrubbers on kraft recovery furnaces are not specifically designed to reduce HCl emissions, HCl control can be achieved as an ancillary benefit.

This section focuses on control of gaseous HCl emissions with packed-bed scrubbers. Packed-bed systems have been installed to a limited extent on kraft recovery furnaces for a variety of reasons, including acid gas (i.e., SO_2) control and hot-water generation (i.e., to recover heat from the flue gas).[21,22] They have also been demonstrated to achieve high levels of HCl control on other combustion systems. The remainder of the section describes packed-bed scrubbing systems, presents a brief summary of the basic operating principles of packed-bed systems and their theoretical performance for both HCl and SO_2 emission reductions, examines packed-bed scrubber performance data for kraft recovery furnaces and for other combustion sources, and discusses existing packed-bed scrubber applications for recovery furnaces.

3.1.2.1.1 Description. As illustrated in Figure 3-7, the recovery furnace exhaust gases are ducted to the acid gas control system after PM has been removed in the ESP. Acid gas scrubbers control HCl and SO_2 emissions by mass transfer of these pollutants from the exhaust gas stream to a liquid scrubbing solution, typically either water or a caustic solution with water. Packed-bed scrubbers are the most frequently used absorption device because they provide excellent liquid-gas contact, leading to efficient mass transfer.[23,24]

Schematics of two standard types of packed-bed scrubbers, the counter-flow design and the cross-flow design, are shown in Figures 3-8 and 3-9. Typically, the gases pass through a quench section before entering the scrubber to saturate the gas stream. The gases then enter the inlet plenum, which is designed to distribute the gases uniformly across the bed cross-section. As the gases flow through the bed, they contact a liquid film created by a flowing liquid that is distributed evenly throughout the bed. After the gas stream exits the bed, it passes through a mist eliminator to remove entrained droplets that may contain

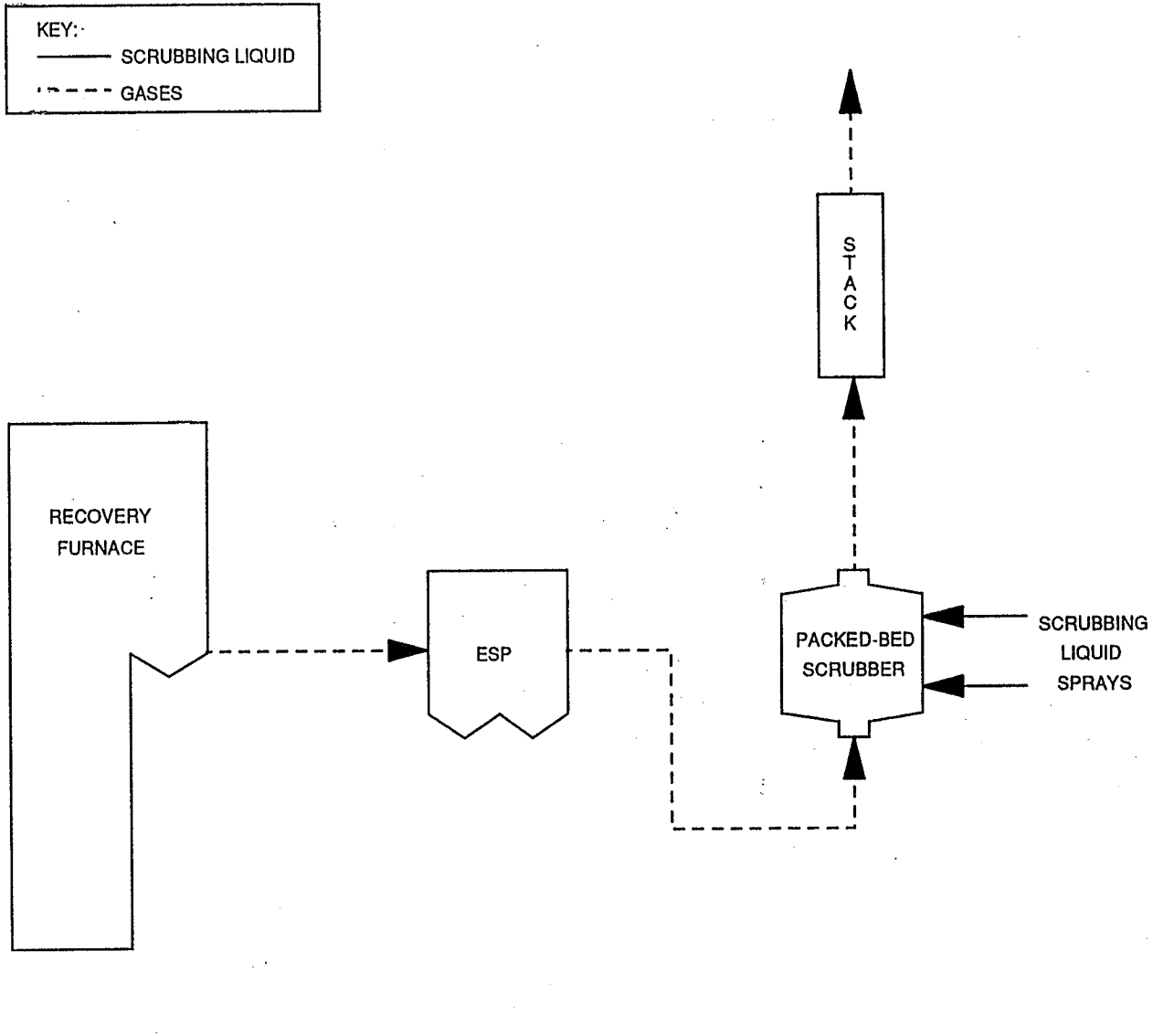

Figure 3-7. Acid gas control system configuration on recovery furnace exhaust gases.

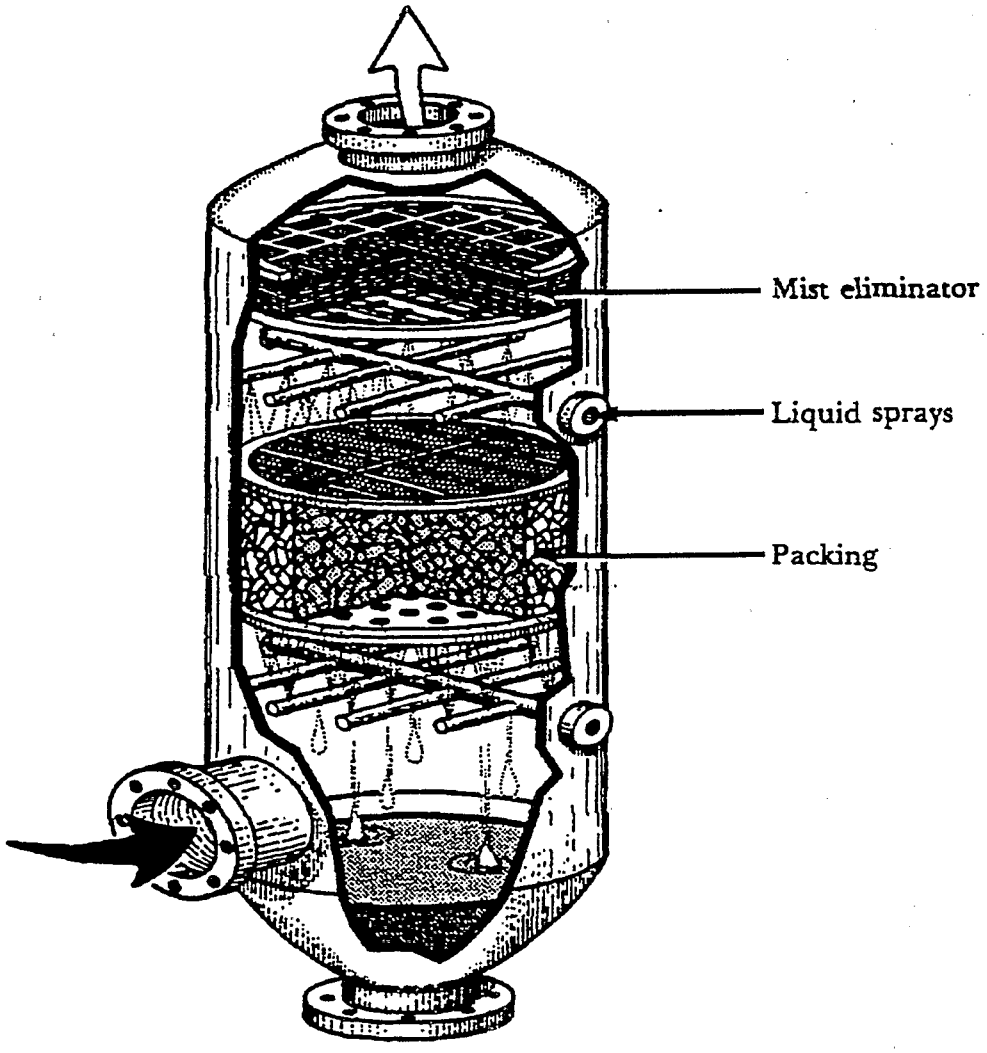

Figure 3-8. Schematic of a counterflow packed-bed scrubber.

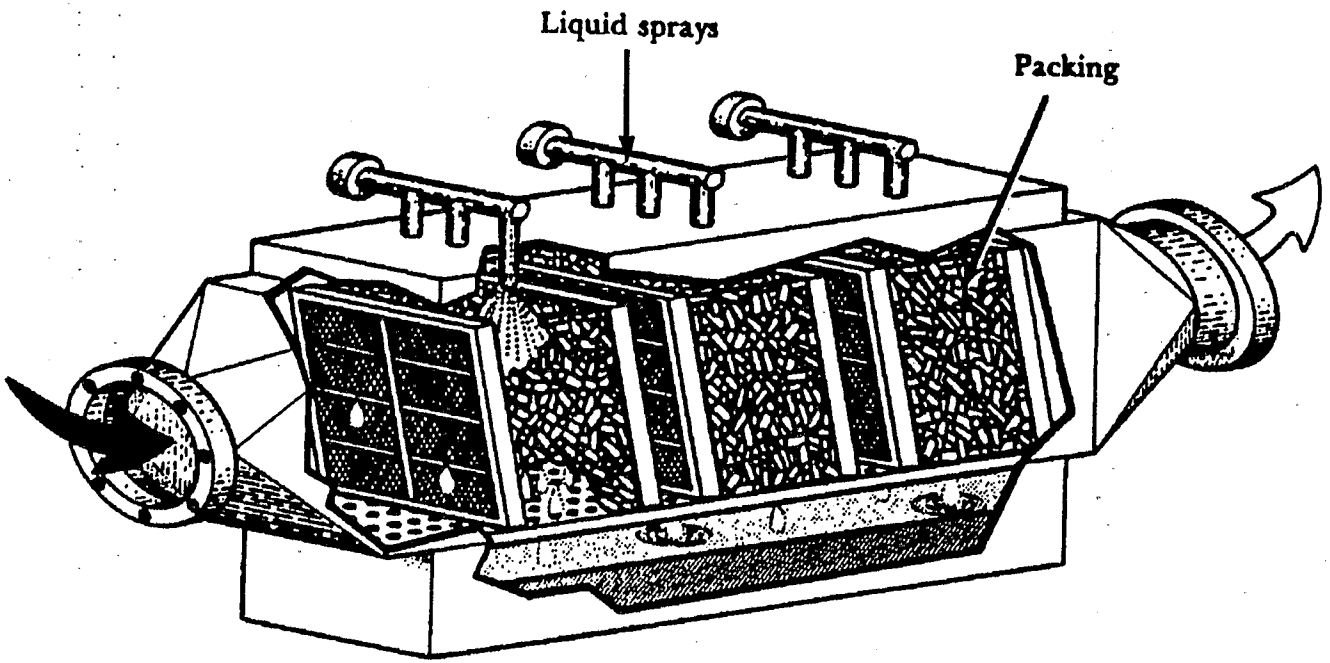

Figure 3-9. Schematic of a cross-flow packed-bed scrubber.

alkaline salts (generated as the acid gases are scrubbed from the exhaust gas), residual absorbed acids, and PM that was not removed by the ESP or the packed bed. The paragraphs below describe the role of the packing and the scrubbing liquid in more detail.

A packed-bed scrubber is filled with packing material that is used to produce a large liquid film surface area for contact between the gas stream and scrubbing liquid. Packing materials are available in a variety of forms and materials, each having specific characteristics with respect to surface area, pressure drop, weight, corrosion resistance, and cost.[25] Specific packing materials are selected by the manufacturer to conform with the gas and liquid stream properties of a particular application. The gas stream moves through the packed bed, contacting the liquid film on the surface of the packing. In the counterflow configuration, the gas moves upward, while the liquid moves downward through the packed-bed section. In a cross-flow configuration, the gas moves in a horizontal direction, perpendicular to the liquid stream, which moves downward through the packing. The countercurrent system is more efficient for acid gas removal, but height limitations or particle loadings that can lead to plugging sometimes dictate use of the cross-flow configuration.[26]

An aqueous scrubbing solution, which may be either water or an alkaline solution, is used in acid gas scrubbers. Because HCl is highly soluble in water, adequate control can often be achieved with water alone. However, because the exhaust from kraft recovery furnaces also contains SO_2, alkaline solutions are expected to be used for this source. In addition to providing substantially better SO_2 control and somewhat better HCl control, an alkaline solution typically is used to maintain a constant scrubber liquid pH at or slightly below 7.0 to prevent both scale buildup and corrosion of scrubber components. The scrubbing liquid that is expected to be used for acid gas control is caustic solution (sodium hydroxide [NaOH]), although sodium carbonate (Na_2CO_3) and calcium hydroxide ($Ca[OH]_2$) (slaked lime)

also can be used. When the acid gases are absorbed into the scrubbing solution, they react with alkaline compounds to produce neutral salts. The rate of absorption of the acid gases is dependent on the solubility of the acid gases in the scrubbing liquid. The neutralization reactions are essentially stoichiometric, i.e., the stoichiometric ratio of alkaline compounds added to the system to that required for complete neutralization of the acid absorbed into the scrubber solution is essentially 1:1 in packed-bed scrubbers.[26]

3.1.2.1.2 Theoretical performance of packed-bed scrubbers. Packed-bed scrubbers are chemical engineering unit operations that provide mass transfer between a soluble gas component (in the case of kraft recovery furnaces, HCl, SO_2, and possibly some organic HAP's) and a solvent liquid scrubbing solution (water or an alkaline solution). The driving force for absorption is the difference between the partial pressure of the soluble compound in the gas mixture and the vapor pressure of the solute gas in the liquid film in contact with the gas.[27] The performance of the system is a function of the mass transfer rate of the pollutant from the gas to the liquid solute and of the contact time between the gas and the liquid. These properties form the basis for the design equations that are typically used to characterize the theoretical performance of gas absorption systems like packed-bed scrubbers.

Because the pollutant concentration in the gas stream is very dilute, several assumptions that simplify these design equations considerably are reasonable. The waste gas is assumed to operate as an ideal gas; the scrubbing solution is assumed to be an ideal solution; heat effects associated with absorption are assumed to be negligible; and the reaction of the HCl and SO_2 with the alkaline material in the scrubbing solution is assumed to be fast compared to the rate-limiting absorption step. Also, because equilibrium characteristics of a counterflow scrubber are much less complex mathematically than the characteristics of a cross-flow tower, the theoretical analyses focus on the counterflow tower.

The design and performance equations for counterflow packed-bed scrubbers are based on the Whitman two-file theory, which is described in detail in reference 28.[28] The key element of this theory for the analysis of kraft recovery furnace packed-bed scrubbers is that the system is considered to be gas-film controlling or liquid-film controlling, depending on the pollutant to be controlled and the scrubbing liquid. For the recovery furnace exhaust stream, HCl absorption is liquid-film controlling, while both the liquid film and the gas film provide appreciable resistance to SO_2 absorption.[28] These system properties simplify the theoretical assessment of the performance of the system for HCl in that only gas-phase transfer needs to be considered. They also partially explain the difference in performance of packed-bed scrubbers for SO_2 and HCl.

An estimate of the theoretical performance of a packed-bed scrubber for HCl emissions can be developed by manipulating the system design equations.[29-31] For a specific set of inlet conditions, these design equations can be modified to obtain the following expression for the efficiency of a packed-bed scrubber:

$$\begin{aligned} Eff &= 1 - y_o/y_i \\ &= 1 - \exp-(Z/H_{OG}) \end{aligned}$$

where y_o is the outlet gas mole fraction, y_i is the inlet gas mole fraction, Z is the packing height, and H_{OG} is the height of a gas transfer unit.

These theoretical analyses have two important implications for the performance of packed-bed scrubbers in controlling HCl emissions from kraft recovery furnaces. First, the theoretical efficiency of the system is limited by the concentration that can be achieved in the exhaust gas from the scrubber when it is in equilibrium with the scrubber liquor feed. At the concentrations found in kraft furnace recovery exhaust gases and scrubbers, the Henry's Law coefficient is of the order of 10^{-3} or less.[32] If the scrubber liquor feed is maintained at a pH of 6.5 or greater, the molar concentration of HCl in the feed is on the order of

6×10^{-9}. Theoretically, the limiting HCl concentration in the exhaust gas is well below 1 ppm. However, this theoretical limit is based only on ideal mass transfer considerations and does not address nonuniformities in gas and liquid flows that may be found in operating systems.

Second, if for a given system, the height of a transfer unit changes between pollutants by a factor of k, the penetration (i.e, 1 - efficiency) for the two pollutants changes exponentially with 1/k. The reported range of manufacturer's H_{OG} for plastic packings is 0.18 to 0.33 meters (m) (0.6 to 1.1 feet [ft]) for water as a scrubbing liquid and 0.15 to 0.31 m (0.5 to 0.7 ft) for caustic as a scrubbing liquid.[29] Note that the same reference reports a range of 0.15 to 0.61 m (0.7 to 2.0 ft) for scrubbing SO_2 with caustic. These data indicate that the height of a transfer unit for SO_2 is 2 to 3 times that for HCl, so that the penetration for SO_2 will be the square root or the cube root of that for HCl or, equivalently, that the penetration for HCl will be the square or the cube of the penetration for SO_2. For example, if a system achieves 99 percent control for HCl, (i.e., the penetration is 0.01), the control efficiency for SO_2 would be expected to be in the range of 78 to 90 percent.

3.1.2.1.3 <u>Performance of HCl scrubbers for kraft recovery furnaces</u>. To provide a basis for subsequent analyses of HCl performance potential, a typical sample of emission test data on HCl concentrations and kraft recovery furnace combustion gas characteristics is compiled in Table 3-3. The emission test data are taken from tests conducted in 1992-93 by the National Council of the Paper Industry for Air and Stream Improvement, Inc. (NCASI).[19] A complete compilation of available HCl emission test data for kraft recovery furnaces is presented in a separate memorandum.[33] The data in Table 3-3 reflect two factors that complicate the design, operation, and characterization of performance of packed-bed scrubbing systems--the relatively large flue gas flow rates generated by these systems and the relatively low and variable concentrations of both HCl and SO_2 generated by these systems. Because of the large flow rates, either large-

TABLE 3-3. SUMMARY OF DATA ON KRAFT RECOVERY FURNACE COMBUSTION GAS CHARACTERISTICS[19]

Mill code	Furnace no.	Furnace type	BLS fired, million kg/d (million lb/d)	%Cl in BLS fired	Flue gas flow rate, dscmm (dscfm)	Stack O_2, %	Flue gas concentrations[a]		
							ppm HCl @ 8% O_2	ppm SO_2 @ 8% O_2	ppm CO @ 8% O_2
AA	1	NDCE	1.51 (3.32)	0.43	3,766 (132,994)	3.1	44.0	224.1	41.8
AA	9	DCE	1.18 (2.61)	0.36	4,352 (153,705)	6.6	5.9	60.0	40.9
BB	3	DCE	0.87 (1.92)	1.35	2,575 (90,970)	5.2	30.2	115.9	NM
BB	4	NDCE	1.32 (2.90)	1.29	4,479 (158,180)	5.6	0.5	0.5	NM
CC	2	NDCE	0.37 (0.82)	0.45	1,349 (47,652)	10.5	36.2	273.2	42.4
CC	3	NDCE	1.32 (2.90)	0.44	4,269 (150,772)	7.3	48.4	322.2	38.6
DD	4	NDCE	1.76 (3.89)	1.05	4,378 (154,610)	2.8	0.8	0.1	NM
EE	6	DCE	0.84 (1.86)	1.04	3,024 (106,785)	8.3	19.9	66.5	NM
EE	7	DCE	0.70 (1.55)	1.09	2,621 (92,569)	7.5	8.1	13.1	NM
FF	2	NDCE	0.68 (1.51)	1.06	1,872 (66,110)	8.8	4.3	7.1	NM
FF	3	NDCE	0.80 (1.77)	0.96	1,988 (70,204)	7.6	0.2	ND	NM
FF	4	NDCE	2.01 (4.43)	1.10	4,102 (144,847)	4.7	0.1	ND	NM
GG	10	NDCE	0.89 (1.97)	0.61	3,338 (117,869)	5.4	24.1	29.7	112.7
HH	3	NDCE	1.08 (2.38)	0.63	4,027 (142,222)	11.0	40.8	61.5	NM

[a]In all cases emissions given are average values from two to six consecutive tests.
BLS - black liquor solids.
dscmm - dry standard cubic meters per minute.
dscfm - dry standard cubic feet per minute.
NM - not measured.
ND - not detected.

diameter scrubbers or parallel systems are needed to maintain proper flow conditions in the scrubber. The low HCl concentrations such as those in Table 3-3 are reported to affect the ability of the scrubber to achieve high removal efficiencies because of limitations on the outlet concentrations that can be achieved with the systems.[23,34]

Only limited test data on acid gas control with packed-bed scrubbers are available for kraft recovery furnaces. However, packed-bed scrubbers are used widely on waste combustion systems, and a substantial body of information is available on their performance for these systems. These data show that these scrubbers can achieve removal efficiencies of 99 percent or greater with relatively high concentrations of HCl. Efficiencies of this order are confirmed by scrubber manufacturer representatives, who indicate that properly designed and operated packed-bed scrubbers can achieve HCl emission reductions of at least 99 percent or an HCl outlet concentration not exceeding 5 ppmv for recovery furnace exhaust streams.[23,34]

Available data on scrubber performance are presented in Table 3-4 for seven kraft DCE and NDCE recovery furnaces and five medical waste incinerators (MWI's).[15,21,22,35,36] With one exception, the data in Table 3-4 indicate that packed-bed scrubbers on waste combustion systems can achieve HCl removal efficiencies of 98.8 percent or greater. The one exception is a cross-flow scrubber, which manufacturers indicate are less efficient than counterflow systems. The high HCl removal efficiencies with packed-bed scrubbers are supported by a summary of trial burn data for hazardous waste incinerators (HWI's).[37] This study contains limited data on HCl control efficiencies for 10 HWI's based on feed rates of organic chlorine to the system and outlet HCl measurements.[37] Because the reported HWI data are insufficient to characterize inlet concentrations or scrubber characteristics, they were excluded from Table 3-4. However, efficiencies for all 10 systems are greater than 98 percent, and 6 of the 10 systems showed efficiencies of substantially greater than 99 percent.[37] More detailed data on two of these systems

TABLE 3-4. SUMMARY OF ACID GAS SCRUBBER PERFORMANCE DATA[15,21,22,35,36]

Type of facility	Wet scrubber description	Inlet gas flow, acmm (acfm)	L/G ratio, L/acm (gal/10^3 acf)	HCl performance data			SO_2 performance data		
				Inlet conc. (ppmv)[a]	Outlet conc. (ppmv)[a]	Efficiency (%)	Inlet conc. (ppmv)[a]	Outlet conc. (ppmv)[a]	Efficiency (%)
NDCE recovery furnace	ESP followed by packed-bed scrubber with 2 cross-flow packed beds separated by spray chamber	3,540 (125,000)	2.1-2.8 (16-21)	NA	NA	NA	150[b]	2.5[b]	98
				NA	NA	NA	NA	0.08	NA
NDCE recovery furnace	ESP followed by venturi scrubber and cross-flow packed-bed scrubber with 3 sections	4,960 (175,000)	1.7-2.3 (13-17)	NA	NA	NA	NA	0.22	NA
DCE recovery furnace	ESP followed by packed-bed scrubber with process water	7,360 (260,000)	1.5-2.0 (12-15)	NA	NA	NA	NA	10-20	NA
3 DCE recovery furnaces	2 ESP's serving 3 recovery furnaces followed by 1 spray tower/packed-bed scrubber with caustic/water	14,800 (522,049)	NA	NA	NA	NA	NA	<1	NA
DCE recovery furnace	ESP followed by packed-bed scrubber with sodium hydroxide/water	9,200 (325,000)	2.9 (22)	NA	NA	NA	NA	56-64	NA
MWI	Venturi scrubber followed by 2 packed-bed scrubbers	270 (9,600)	3.7-4.0 (28-30)	1,160	1.38	99.9	1.5	0.6	77.5
MWI	Baghouse followed by cross-flow packed-bed scrubber	65 (2,300)	3.2 (24)	198	44	85.8[c]	21.5	2.2	80
MWI	Venturi scrubber followed by packed-bed scrubber	180 (6,400)	0.75 (5.6)	1,160	6.3	99.3	NA	NA	NA
MWI	Venturi scrubber followed by packed-bed scrubber	63 (2,300)	3.7 (28)	1,280	15.1	98.8	7.3	≈0.3	>90
MWI	Venturi scrubber followed by packed-bed scrubber	91 (3,200)	2.1 (16)	1,356	0.7	99.9	53.7	<0.4	>99.9

[a]MWI data corrected to 7 percent O_2; recovery furnace data corrected to 8 percent O_2.
[b]Design criteria; actual outlet SO_2 equals 0.08 ppm.
[c]This efficiency may be biased low due to plugged nozzles during the test.
acmm = actual cubic meters per minute.
acfm = actual cubic feet per minute.
L/acm = liters per actual cubic meter.
gal/10^3 acf = gallons per thousand actual cubic feet.
ppmv = parts per million volume.

that used a caustic scrubbing solution showed outlet concentrations of 1.3 and 13 ppmv, with removal efficiencies of >99.9 percent and 98.2 percent, respectively.[38] Taken together, the data in Table 3-4 and the HWI data show that a properly designed and operated packed-bed scrubber can achieve HCl removal efficiencies in the range of 99 percent.

For two MWI systems in Table 3-4, outlet HCl concentrations of less than 1.5 ppmv were obtained with HCl inlet concentrations of greater than 1,000 ppmv. This same type of reduction appears to have been achieved by the one HWI system with an outlet concentration of 1.3 ppmv. These MWI and HWI data support the results from the theoretical analyses on the ability of these systems to achieve low outlet concentrations. Also, systems with caustic scrubbing solution are able to achieve outlet concentrations not exceeding 5 ppmv.[23,34]

3.1.2.1.4 Performance data--gaseous organic HAP's. Data were presented in Chapter 2 on levels of gaseous organic HAP emissions expected from recovery furnaces. While there are no available gaseous organic HAP emission test data that would allow the determination of the performance of packed-bed scrubbers in controlling HAP emissions from recovery furnaces, packed-bed scrubbers designed primarily for HCl control would be expected to control emissions of the water-soluble gaseous organic HAP's (such as methanol and formaldehyde) to a limited degree. However, because of the low concentrations of gaseous organic HAP's expected in the furnace exhaust streams, absorption of these pollutants may be limited in packed-bed scrubbers designed for HCl control due to the small concentration gradient (driving force) between the gas and scrubbing liquid.

In providing design and cost information for packed-bed scrubbers for HCl control, vendors were asked to provide estimates of the degree to which the HAP's methanol, formaldehyde, and acetaldehyde would be controlled with the same equipment. One vendor estimated removal efficiencies of 50 percent for methanol and formaldehyde and 20 percent for acetaldehyde but would not offer a performance guarantee for the

three pollutants.[34] While the packed-bed scrubber may remove some gaseous organic HAP's from the gas stream, a portion of the pollutants absorbed into the scrubbing fluid may volatilize back into the atmosphere from the wastewater treatment process, which would decrease the overall system removal efficiencies. Currently, data are insufficient to determine whether the blowdown of the scrubber liquor is sufficient to maintain low levels of HAP's in the scrubber water to maintain the concentration gradient and to prevent stripping.

3.1.2.1.5 Existing applications of packed-bed scrubbers. Currently, two NDCE recovery furnaces, both at the same kraft pulp mill, are each equipped with a cross-flow packed-bed scrubber that operates downstream from an ESP. For other recovery furnace applications, the packed-bed scrubber also would be installed downstream from the PM control device, which would remove over 99 percent of the PM from the emission stream before scrubbing.

One of the packed-bed scrubbers currently in operation on an NDCE recovery furnace was designed for SO_2 control and hot-water generation (i.e., to recover heat from the flue gas).[21,22] Fresh water is used as the scrubbing medium, with some added caustic. Although no HCl emission data are available for the recovery furnaces at this mill, SO_2 scrubber design criteria are available for this recovery furnace. As shown in Table 3-4, the design inlet SO_2 concentration was 150 ppmv, while the design outlet SO_2 concentration was 2.5 ppmv, for a design SO_2 emission reduction of 98 percent.[35] Because HCl is more soluble in water than SO_2 and based on the theoretical discussion presented earlier, HCl emission reductions greater than 98 percent would be expected for this packed-bed scrubber.

While only one facility has installed packed-bed scrubbers on NDCE recovery furnaces, three other facilities have reported that packed-tower systems are being used to control emissions from five DCE recovery furnaces.[15] Although no information is available on the control of HCl achieved by these systems, the available data indicate that they achieve SO_2 outlet

concentrations ranging from <1 to about 60 ppmv.[15] Furthermore, the data indicate that the three packed-tower systems handle flow rates in the range of 7,360 to 14,800 actual cubic meters per minute (acmm) (260,000 to 522,000 acfm).[15] These systems show that counterflow packed-tower systems are compatible with both DCE and NDCE recovery furnace exhaust streams and provide a viable technology for control of HCl emissions from these furnaces.

3.1.2.2 Venturi Scrubbers. Venturi scrubbers are a demonstrated control technology for reducing PM emissions and, therefore, PM HAP emissions from lime kilns and SDT's. Particulate matter emissions from approximately 86 percent of lime kilns and 35 percent of SDT's are controlled using venturi scrubbers.[1] However, venturi scrubbers are used on only one existing recovery furnace (an NDCE recovery furnace) as the sole PM control device.[12] Venturi scrubbers may also provide, to a more limited extent, control of gaseous HAP emissions. This section briefly describes venturi scrubbers and presents PM HAP performance data.

3.1.2.2.1 Description. Venturi scrubbers are designed to remove PM primarily by impaction through high-energy contact between the scrubbing liquid and suspended PM in the gas stream.[26] A schematic of a venturi scrubbing system is shown in Figure 3-10. A venturi scrubbing system typically consists of a venturi scrubbing vessel and cyclonic separator. The venturi scrubbing vessel has a converging section, a throat section, and diverging section.

The exhaust gas enters the converging section and, as the cross-sectional area of the scrubber vessel decreases, gas velocity increases. Scrubbing liquid is introduced either at the entrance to the converging section or at the throat.[39] The exhaust gas, which is pulled through the venturi vessel by the system's induced draft fan and forced to move at extremely high velocities in the throat, shears the scrubbing liquid from the walls, atomizing the liquid.[26,39] Particulate matter and gaseous pollutants are transferred from the gas stream to the liquid

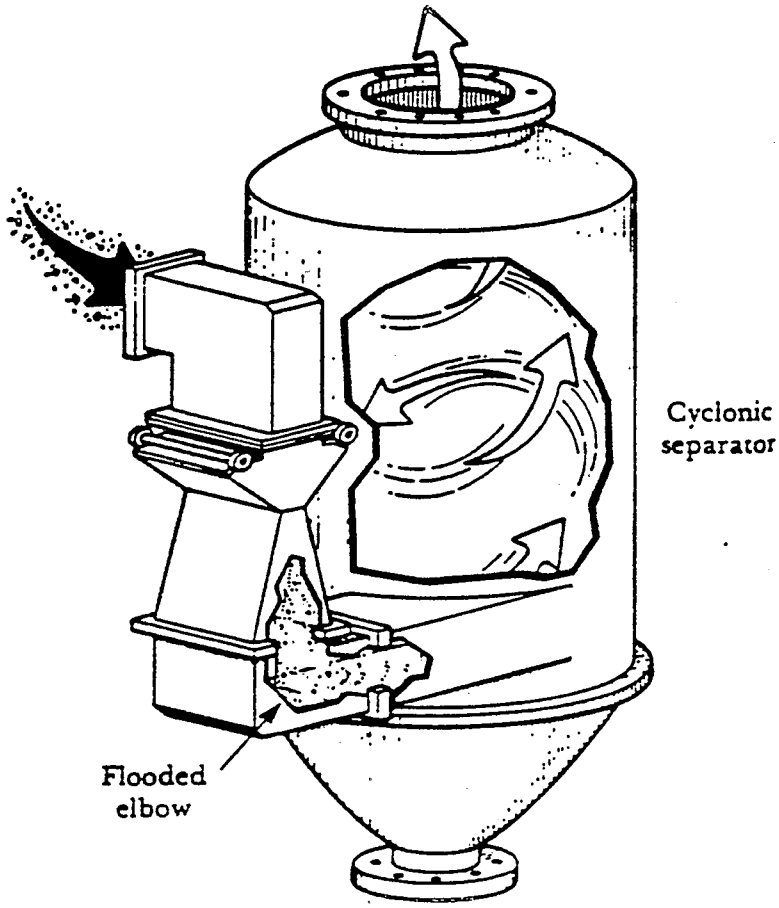

Figure 3-10. Schematic of a venturi scrubber.

droplets via impaction and diffusive mass transfer in the throat section as the exhaust stream turbulently mixes with the atomized scrubbing liquid. The throat section may be constructed so that the size of the throat opening is adjustable. With an adjustable-throat (or variable-throat) venturi, the gas velocity across the throat can be maintained at a constant speed as the gas flow rate changes, thereby maintaining the desired PM collection efficiency. From the throat section, the exhaust stream passes through the diverging section of the venturi scrubbing vessel, where the velocity decreases. Diffusion, which is an effective collection mechanism for fine particles and also the primary mechanism of gaseous pollutants transfer, usually occurs in the diverging section, where the velocities of the gas stream and liquid droplets are almost equal. Collection of fine particles by the liquid droplets is possible because the path of fine particles is influenced primarily by Brownian motion rather than by the path of the gas stream.[40]

An entrainment separator, typically a cyclonic separator, is needed to collect the PM entrained in the droplets because the high velocity of the exhaust stream from the venturi scrubbing vessel would have a tendency to exhaust the droplets.[39]

3.1.2.2.2 Performance data--PM HAP's. Available information indicates that control systems that achieve the greatest total PM collection efficiencies also achieve the best performance for PM HAP emissions.[8] Thus, although the following information is presented in terms of total PM, the relative performance of the different systems is applicable to that portion of the total PM that is comprised of PM HAP's.

The performance of the scrubber in terms of PM collection is strongly affected by the pressure drop across the scrubber throat, the liquid-to-gas (L/G) ratio, and the particle size distribution. Particulate matter collection efficiency generally increases as the throat velocity and turbulence of the gas stream increase, as indicated by an increased pressure drop across the scrubber.

Typical venturi scrubber L/G ratios for PM control range from 0.4 to 1.3 liters per actual cubic meter (L/acm) (3 to 10 gallons per 1,000 actual cubic feet [gal/10^3 acf]). While L/G ratios up to 2.7 L/acm (20 gal/10^3 acf) can be used, increasing the L/G ratio beyond 1.3 L/acm (10 gal/10^3 acf) usually does not significantly improve PM collection efficiency. However, venturi scrubbers with L/G ratios ranging from 2.7 to 5.3 L/acm (20 to 40 gal/10^3 acf) are used where absorption of gaseous pollutants in addition to PM control is desired. Liquid-to-gas ratios below 0.4 L/acm (3 gal/10^3 acf) are usually not sufficient to cover the throat section.[39]

For lime kiln applications, PM collection efficiencies for venturi scrubbers average 99 percent.[39] These results are supported by data reported for four lime kiln scrubbers.[15,41] One scrubber provided a removal efficiency of 98 percent with a pressure drop of 26 millimeters of mercury (mm Hg) (14 inches of water [in. H_2O]).[41] The other three scrubbers provided removal efficiencies of greater than 99.5 percent with scrubber pressure drops in the range of 37 to 45 mm Hg (20 to 24 in. H_2O).[15]

Variable-throat venturi scrubbers with pressure drops that range from 19 to 56 mm Hg (10 to 30 in. H_2O) and average 37 mm Hg (20 in. H_2O) are typically used for controlling PM emissions from lime kilns.[1,40] Typical L/G ratios are between 1.3 and 2.7 L/acm (10 and 20 gal/10^3 acf).[40] Particle size distribution data, based on limited lime kiln test data, show that approximately 17 percent of the particles entering the scrubber are less than 10 micrometers (μm) in diameter, while 98 percent of particles exiting the scrubber are less than 10 μm in diameter.[42] Lime mud with a high soda content, which results from inefficient washing of the lime mud, can produce a large quantity of PM in the 0.1 to 1.0 μm size range.[40] Particles in this size range are the most difficult to remove and may require increased scrubber pressure drops or process modifications to improve lime mud washing.[40]

The NSPS for kraft lime kilns requires that PM emissions from gas-fired and oil-fired lime kilns constructed, reconstructed, or modified after September 24, 1976 be less than

or equal to 0.15 and 0.30 g/dscm (0.067 and 0.13 gr/dscf), respectively.[10,11] Based on installation dates, approximately 28 percent of existing lime kilns are subject to the NSPS.[1]

Long-term PM data are available for two gas-fired lime kilns with venturi scrubbers that are subject to the NSPS PM limit of 0.15 g/dscm (0.067 gr/dscf). Figures 3-11 and 3-12 present monthly PM emission data from the State of Washington for lime kilns at Mills A and B over a 6-year period from 1988 to 1994.[13] The lime kilns at Mills A and B were installed in 1980 and 1979, respectively.[15] The data demonstrate that existing venturi scrubbers on lime kilns can readily meet the NSPS PM level of 0.15 g/dscm (0.067 gr/dscf) on a long-term, continuous basis.

For SDT applications, reported PM removal efficiencies for venturi scrubbers range from 97 to greater than 99 percent.[40,42,43] The average pressure drop for SDT venturi scrubbers is 15 mm Hg (8 in. H_2O).[1] Liquid-to-gas ratios range from 1.0 to 1.3 L/acm (8 to 10 gal/10^3 acf).[40] Weak wash (from the lime mud washer) is the scrubbing fluid for the majority of SDT venturi scrubbers.[1]

The NSPS for kraft pulp mills require that PM emissions from SDT's that are constructed, modified, or reconstructed after September 24, 1976 be less than 0.10 kg/Mg (0.20 lb/ton) of BLS fired in the recovery furnace.[10,11] Based on installation dates, approximately 29 percent of existing SDT's are subject to the NSPS.[12]

No long-term PM data are available for SDT's with venturi scrubbers that are subject to the NSPS limit of 0.10 kg/Mg (0.20 lb/ton) BLS. However, long-term data are available for two SDT's equipped with packed-bed scrubbers designed to meet a more stringent PM permit limit of 0.06 kg/Mg (0.12 lb/ton) BLS; high-efficiency venturi scrubbers are expected to achieve equivalent control. Figures 3-13 and 3-14 present monthly PM emission data from the State of Washington for two SDT's with packed-bed scrubber systems at Mill B over a 6-year period from 1988 to 1994.[13] Figure 3-15 presents monthly PM emission data from the State of Washington for a new SDT with a combination venturi and

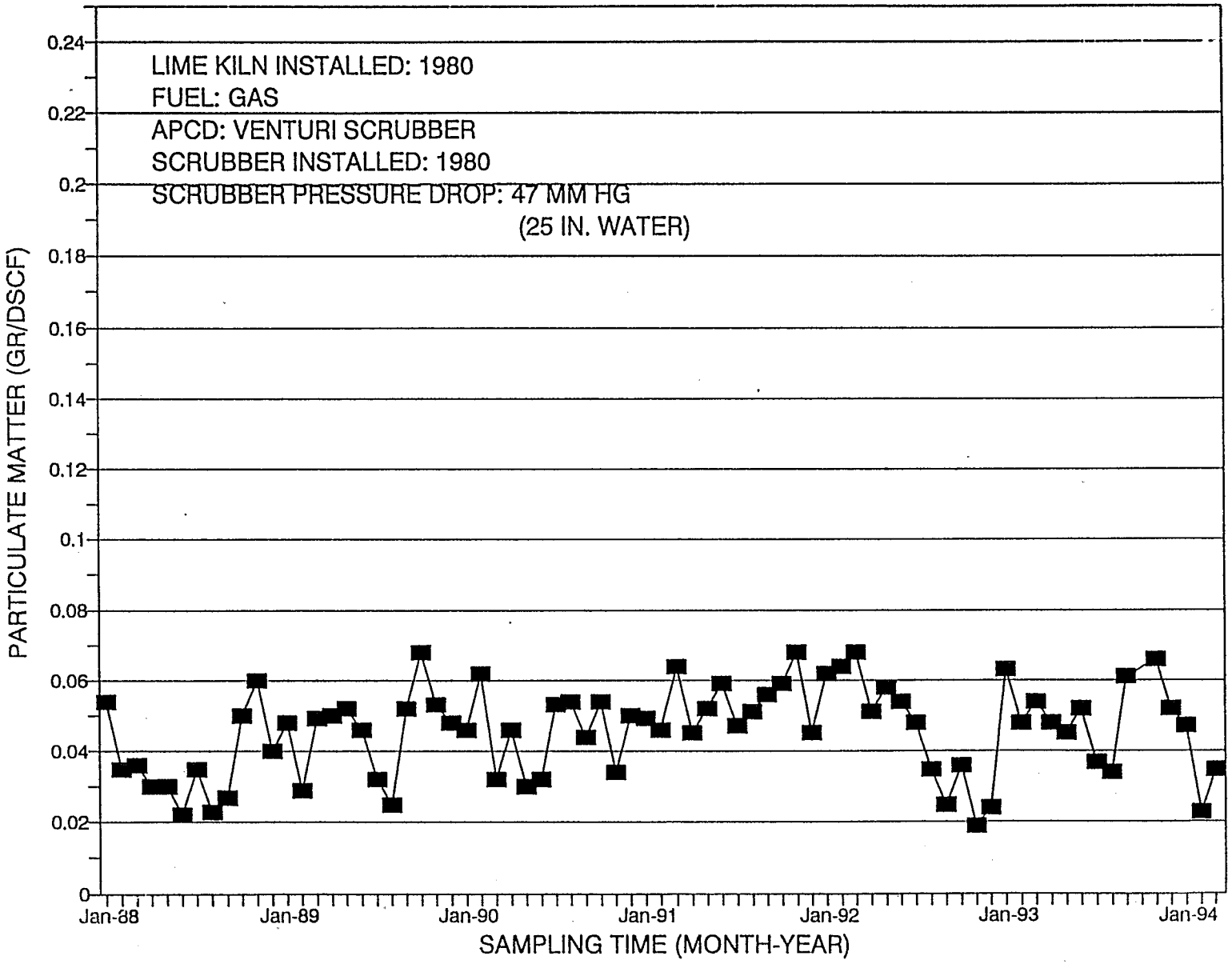

Figure 3-11. PM emission data for lime kiln at Mill A.

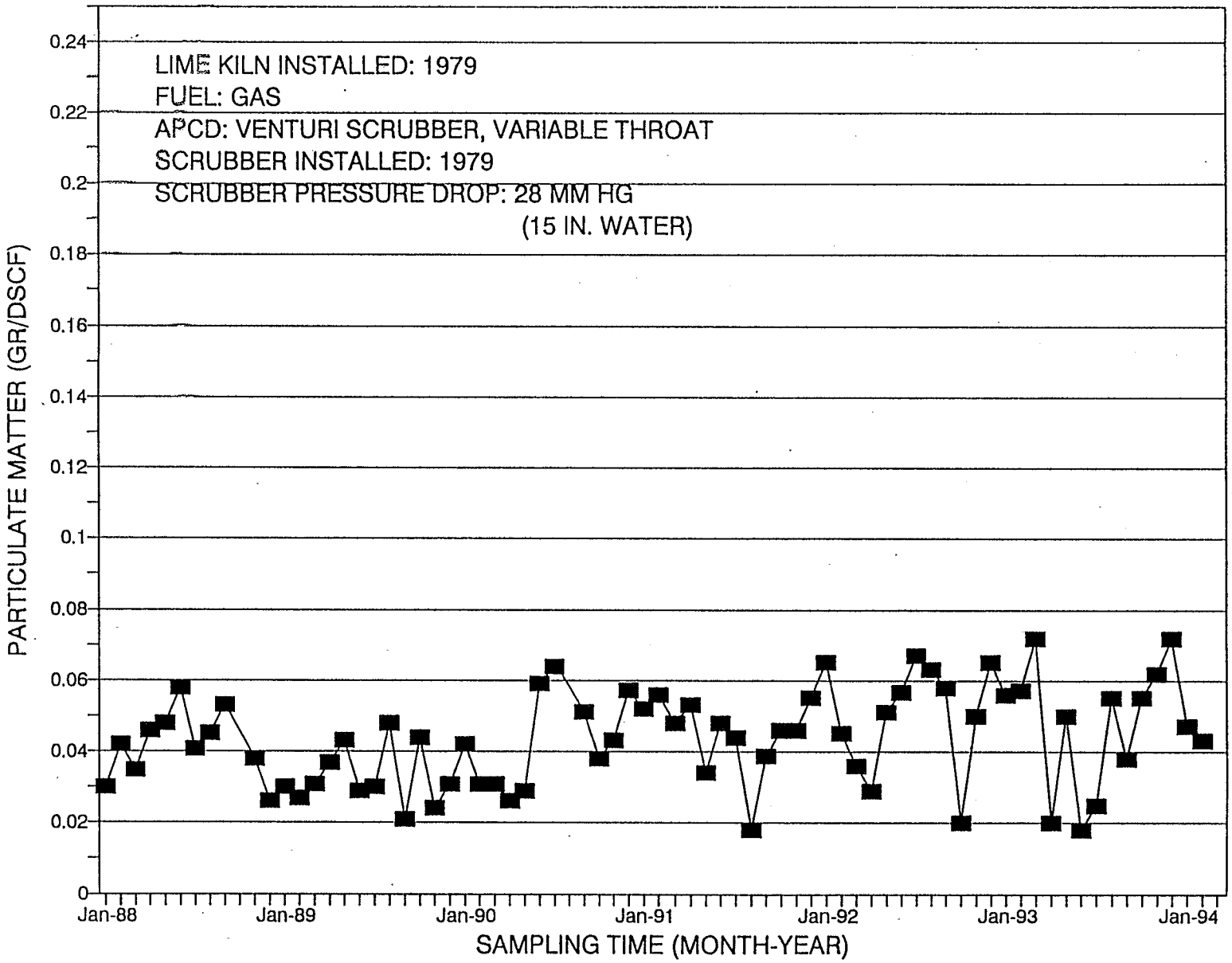

Figure 3-12. PM emission data for lime kiln No. 4 at Mill B.

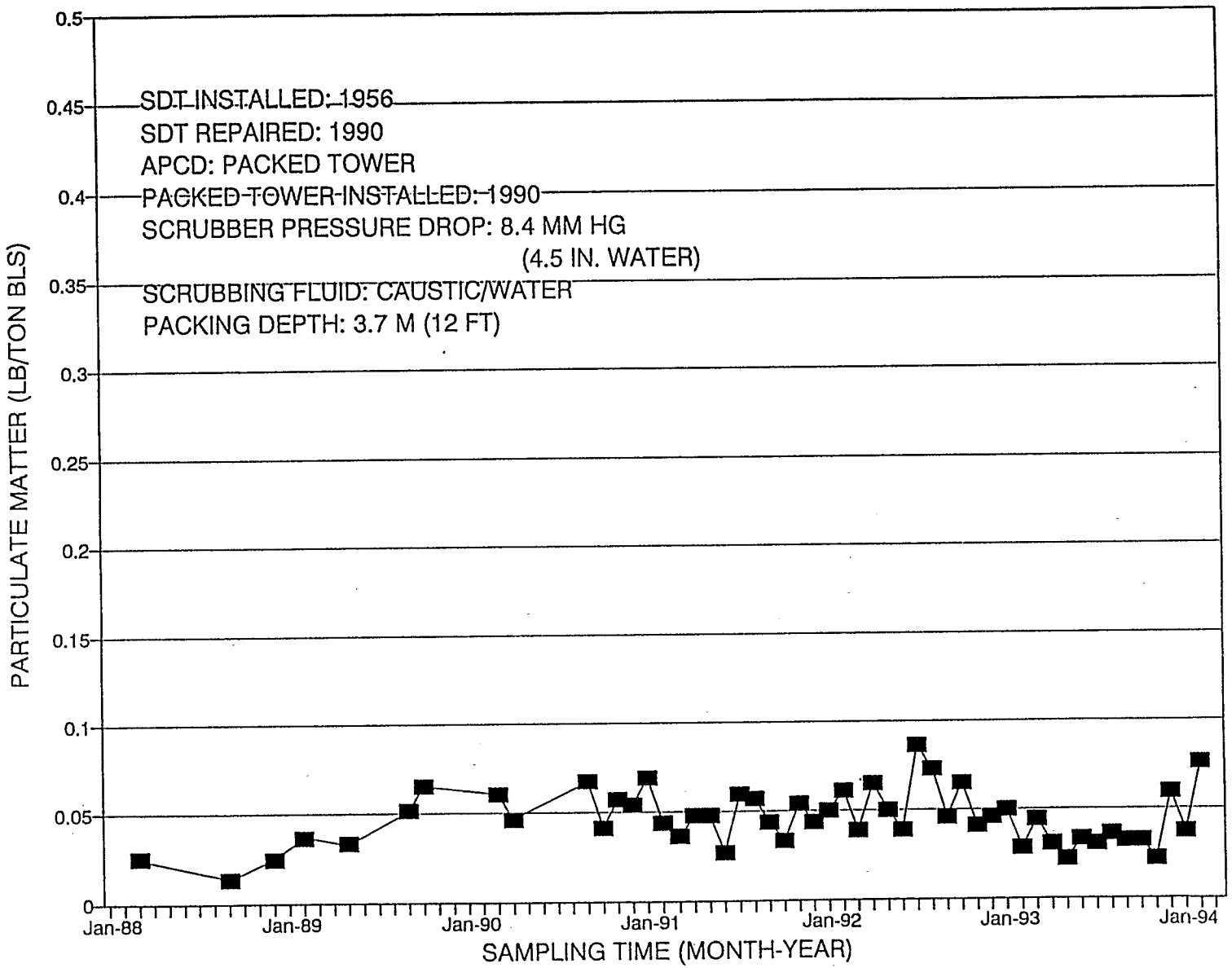

Figure 3-13. PM emission data for SDT No. 3 at Mill B.

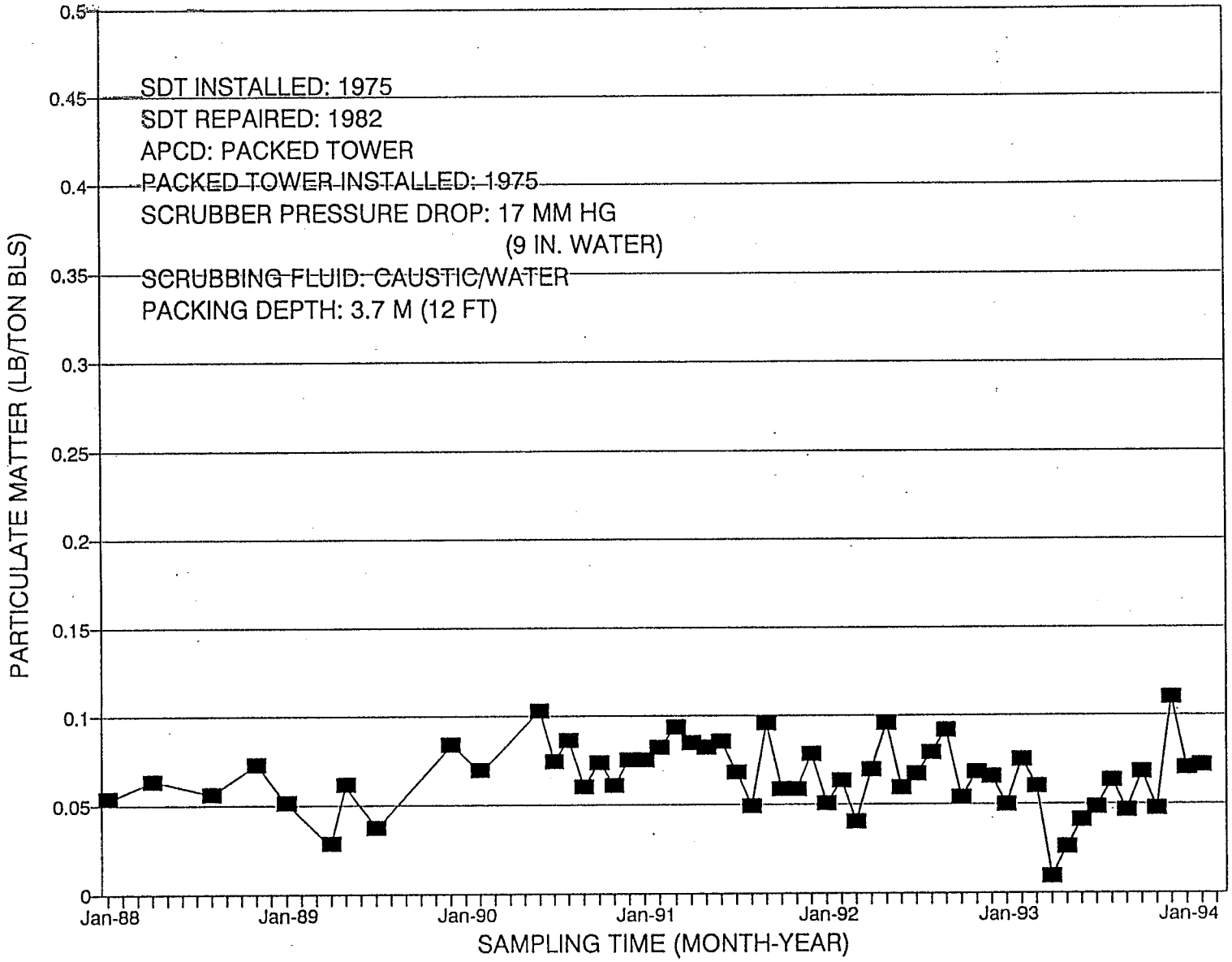

Figure 3-14. PM emission data for SDT No. 4 at Mill B.

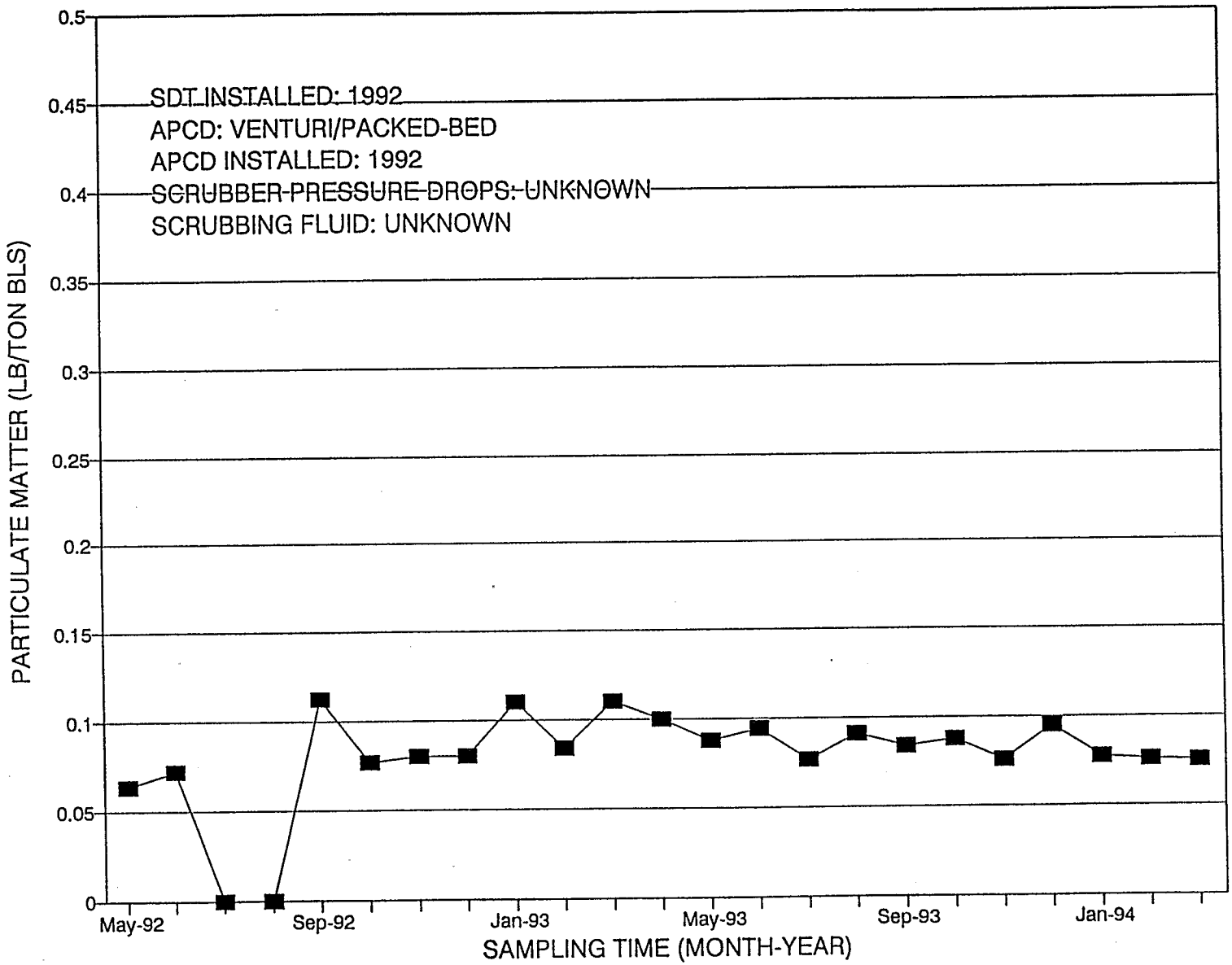

Figure 3-15. PM emission data for SDT No. 22 at Mill C.

packed-bed scrubber system at Mill C over a 2-year period from 1992 to 1994.[13] These data demonstrate that high-efficiency wet scrubbers on SDT's can achieve a maximum outlet PM level of 0.06 kg/Mg (0.12 lb/ton) BLS on a long-term, continuous basis.

3.1.3 BLO Control

As explained in Chapter 2, BLO systems that use pure oxygen instead of air to oxidize black liquor do not need a system vent because all of the oxygen is consumed in the oxidation reaction. As a result, there are no emission points for the molecular O_2 BLO unit. Black liquor oxidation systems that use air to oxidize black liquor require vents for each oxidation tank. Add-on controls currently used to reduce gaseous organic HAP emissions from air-sparging BLO units collect and incinerate exhaust gases from the BLO vents. This section describes the add-on controls and presents the reductions in gaseous organic HAP emissions that can be achieved with the add-on controls.

3.1.3.1 Description. Of the estimated 42 kraft mills with air-based BLO systems, two mills reduce gaseous organic HAP emissions from air-sparging BLO vents with vent gas collection/ incineration systems. The BLO vent gas collection/incineration method used by one mill involves ducting the vent gases to two condensers in series (to remove water from the gases), reheating the gases, and then ducting the dry gases to a power boiler for incineration.[44,45] The control method used by the second mill is identical except that a mesh-pad mist eliminator is used in place of the condensers to remove water from the vent gas.[46,47] Both mills incinerate their BLO vent gases in order to reduce TRS emissions. Because gaseous organic HAP's such as methanol, acetaldehyde, and methyl ethyl ketone are also emitted from BLO vents, emissions of these HAP's would also be reduced by the incineration of BLO vent gases. Particulate HAP's and acid gases (e.g., HCl) are not emitted from BLO vents.

Before the BLO vent gases can be incinerated, they must first be captured and transported to the incineration device. Vent gases from a BLO unit can be hard-piped to the combustion control device. The type of duct material that can be used is

determined by the characteristics of the gas in the vent stream.[48] Two materials commonly used in the construction of vent ducts in the pulp and paper industry are fiberglass and stainless steel. Fiberglass ducts have the advantages of relatively low cost, light weight, and corrosion resistance, but are unable to be electrically grounded to prevent the buildup of static charge and the absorption of hydrocarbons in its fiberglass resin.[48] Stainless steel is the preferred material of construction for NCG cransport systems.[49] Stainless steel has the advantage of resisting corrosion by water and sulfur compounds but is susceptible to corrosion by chlorides that may be present in a vent gas control system that includes other vents besides the BLO vent (e.g., bleach plant vent).[50]

The BLO vent gas may be conditioned to alter the moisture content and/or the temperature of the stream before it is vented to the combustion control device. Condensers, knockout drums, or entrainment separators may be used to condition the gas in the gas transport system.[48]

Vent gases are preheated only when emissions are controlled with a combustion device. Preheating is generally only practiced on high-volume, low-concentration (HVLC) streams, where the risk of explosion is sufficiently low.[51] The vent gases from BLO units can be classified as HVLC streams. Power boilers and stand-alone thermal oxidizers are generally the preferred combustion devices for incinerating HVLC streams. Power boilers, which include coal, natural gas, oil, wood waste, or combination fuel-fired boilers, are designed to produce heat, steam, and electricity for pulp mill operations.[48] A thermal oxidizer is a refractory-lined chamber containing a burner or burners used to oxidize vent streams containing volatile organic compounds (VOC's).[48] Both combustion devices have been demonstrated in the pulp and paper industry as control devices for vent emissions.[48]

Combustion control devices, such as power boilers and thermal oxidizers, destroy the chemical structure of organic compounds by oxidation at elevated temperatures. These devices operate on the principle that any VOC heated to a high enough

temperature in the presence of sufficient oxygen will oxidize to predominantly CO_2 and H_2O.[52] Combustion devices such as power boilers and thermal oxidizers can achieve VOC/HAP destruction efficiencies of 98 percent or greater.[53,54] Therefore, a net removal efficiency of 98 percent was assumed for all HAP's for the BLO incineration control system.

3.1.3.2 Performance data--gaseous organic HAP's. Because neither of the two vent gas collection/incineration systems currently existing has undergone emissions testing, the gaseous organic HAP emission reductions associated with a BLO vent gas control system were estimated based on gaseous organic HAP emission data available from uncontrolled air-sparging BLO systems, assuming a 98 percent destruction efficiency.[55] The gaseous organic HAP emission reductions resulting from BLO control are presented in Table 3-5. For a BLO unit with a BLS firing rate of 700,000 kilograms of BLS per day (kg BLS/d) (1.5 million pounds of BLS per day [lb BLS/d]) and 351 operating days per year (d/yr), controlling emissions from the BLO vent could reduce average gaseous organic HAP emissions by about 47 Mg/yr (54 ton/yr), of which methanol accounts for about 85 percent.

3.2 EQUIPMENT CHANGES/MODIFICATIONS

This section describes equipment changes/modifications that could be applied to recovery furnaces at kraft pulp and paper mills to reduce gaseous organic HAP emissions. The two equipment changes/modifications are elimination of black liquor or HAP-contaminated process water from ESP control systems and conversion from a DCE recovery furnace system to an NDCE recovery furnace, which are discussed in Sections 3.2.1 and 3.2.2, respectively.

3.2.1 Elimination of Black Liquor used in NDCE Recovery Furnace ESP Control Systems

When hot recovery furnace flue gases come in contact with ESP control systems, gaseous organic HAP's (primarily methanol) may be stripped from any black liquor or HAP-contaminated process water present in the system and emitted to the atmosphere.

TABLE 3-5. HAP EMISSIONS FOR BLO UNITS[55]

Hazardous air pollutant	Uncontrolled BLO unit		Controlled BLO unit	
	Emission factor, mass/mass BLS	Mass rate, kg/yr (lb/yr)[a]	Emission factor, mass/mass BLS	Mass rate, kg/yr (lb/yr)[a]
Total organic HAP's[b]	2.03×10^{-4}	48,500 (107,000)	4.06×10^{-6}	969 (2,140)
Acetaldehyde	1.48×10^{-5}	3,530 (7,790)	2.96×10^{-7}	70.7 (156)
Benzene	1.21×10^{-6}	289 (637)	2.42×10^{-8}	5.78 (12.7)
Formaldehyde	2.89×10^{-7}	69 (152)	5.78×10^{-9}	1.38 (3.04)
Methanol	1.73×10^{-4}	41,300 (91,100)	3.46×10^{-6}	826 (1,820)
Methyl ethyl ketone	4.24×10^{-6}	1,010 (2,230)	8.48×10^{-8}	20.3 (44.6)
Methyl isobutyl ketone	1.77×10^{-6}	423 (932)	3.54×10^{-8}	8.45 (18.6)
Phenol	1.93×10^{-6}	461 (1,020)	3.86×10^{-8}	9.22 (20.3)
Toluene	1.44×10^{-6}	344 (758)	2.88×10^{-8}	6.88 (15.2)
Xylenes	4.28×10^{-6}	1,020 (2,250)	8.56×10^{-8}	20.4 (45.1)

[a]Mass rates are based on a BLS firing rate of 700,000 kg BLS/d (1.5 million lb BLS/d) and 351 operating d/yr.

[b]Total organic HAP's include all gaseous HAP's in this table. Numbers may not add exactly due to rounding.

Electrostatic precipitator control systems include the ESP plus the PM return system associated with the ESP. One method of controlling gaseous organic HAP emissions from ESP systems is to prevent their generation by eliminating the black liquor or contaminated process water from the system. This equipment modification for ESP's used with NDCE recovery furnaces and the associated HAP emission reductions are discussed below. (This equipment modification is not discussed as an option for reducing gaseous organic HAP's from DCE recovery furnaces because, as discussed in Chapter 2, no data are available to determine if eliminating the wet bottom ESP's that are typically used with DCE recovery furnaces would actually decrease overall gaseous organic HAP emissions.)

3.2.1.1 Description. Particulate matter emissions from approximately 23 percent of all NDCE recovery furnaces are controlled with wet-bottom ESP's that use either unoxidized black liquor or process water in the ESP bottom. The majority (77 percent) of NDCE recovery furnaces, however, are controlled with dry-bottom ESP's.[1] Unoxidized black liquor is used in about 87 percent of the wet-bottom ESP's on NDCE recovery furnaces, and water is used in the remaining 13 percent.

Since the 1980's, the industry trend has been toward the use of dry-bottom ESP's. Some mills that installed wet-bottom ESP's on NDCE recovery furnaces in the late 1970's and early 1980's experienced difficulties in meeting the NSPS TRS emission limit of 5 ppm. These mills attempted to comply with the TRS emission limit through the use of one of the following techniques: (1) modifying the bottom of the ESP to limit contact between the flue gases and the black liquor (e.g., using baffles), (2) switching from the use of black liquor to using water in the ESP bottom, or (3) converting from a wet-bottom ESP to a dry-bottom ESP. In general, technique Nos. 2 and 3 proved to be the most viable options for meeting the TRS emission limit on a continuous basis.[56,57] Because the mechanism for generating TRS emissions in the wet-bottom ESP is the same mechanism for generating HAP emissions, technique Nos. 2 and 3 would also

prevent methanol and any other gaseous organic HAP's from being generated in the ESP, provided the water used in technique No. 2 is not contaminated with substantial quantities of organic compounds.

The major modifications involved in technique No. 3 (i.e., converting a wet-bottom ESP to the dry-bottom design) include (1) removing the existing agitator paddles and liquor piping, (2) installing a perpendicular drag scraper system, shallow fallout hoppers, drag chain conveyors, rotary valves, ash mixing tank, agitator, and associated instrumentation, and (3) making piping modifications.[58,59]

The older design dry-bottom ESP control systems sometimes use black liquor to sluice and transport the PM captured by dry-bottom ESP's to the saltcake mix tank. As a result, gaseous organic HAP's may be stripped from the black liquor as the hot recovery furnace flue gases are pulled through the ESP by the induced draft fan. The gaseous organic HAP emissions could be controlled by converting to a dry PM return system. Dry PM return systems eliminate the use of black liquor in the PM return system, so that the captured PM does not contact any black liquor until it reaches the mix tank. These newer PM return systems also are equipped with rotary valves (through which the dry captured PM passes) that provide an air lock between the ESP and the remainder of the PM return system and the mix tank (which contains black liquor).[60] According to one ESP manufacturer, systems that included dry return of the captured PM were initially installed around 1985.[60]

3.2.1.2 <u>HAP Emission Reduction Potential</u>. As shown in Table 3-6, the elimination of black liquor from ESP control systems results in substantial reductions for the majority of the gaseous organic HAP compounds. (Note: Although HCl emissions are included in Table 3-6, HCl emissions are unaffected by the presence or absence of black liquor in the ESP system.) Although the available data indicate that acetaldehyde emissions increase with the elimination of black liquor from ESP control systems,

TABLE 3-6. HAP EMISSIONS FOR NDCE RECOVERY FURNACES[55]

Hazardous air pollutant	ESP control system with black liquor		ESP control system without black liquor		Percent reduction
	Emission factor, mass/mass BLS[a,b]	Mass rate, kg/yr (lb/yr)[c]	Emission factor, mass/mass BLS[b,d]	Mass rate, kg/yr (lb/yr)[c]	
Total organic HAP's[e]	1.30×10^{-4}	31,000 (68,400)	3.67×10^{-5}	8,760 (19,300)	72
Acetaldehyde	7.50×10^{-6}	1,790 (3,950)	1.52×10^{-5}	3,630 (8,000)	-103
Benzene	9.73×10^{-6}	2,320 (5,120)	1.73×10^{-6}	413 (911)	82
Formaldehyde	2.98×10^{-6}	712 (1,570)	1.93×10^{-6}	461 (1,020)	35
Methanol	5.04×10^{-5}	12,000 (26,500)	4.93×10^{-6}	1,180 (2,600)	90
Methyl ethyl ketone	7.46×10^{-6}	1,780 (3,930)	1.80×10^{-6}	430 (948)	76
Methyl isobutyl ketone	9.23×10^{-6}	2,200 (4,860)	2.27×10^{-6}	542 (1,200)	75
Phenol	1.50×10^{-5}	3,580 (7,900)	2.05×10^{-6}	490 (1,080)	86
Toluene	8.37×10^{-6}	2,000 (4,410)	2.03×10^{-6}	485 (1,070)	76
Xylenes	1.92×10^{-5}	4,590 (10,100)	4.76×10^{-6}	1,140 (2,510)	75
HCl	1.20×10^{-4}	28,700 (63,200)	1.20×10^{-4}	28,700 (63,200)	—

[a]Emission factors for gaseous organic HAP's are based on the average emission factors for NDCE recovery furnaces equipped with wet-bottom ESP's or with dry-bottom ESP's with wet PM return systems.

[b]Emission factor for HCl is based on the average emission factor for all NDCE recovery furnaces.

[c]Mass rates are based on a furnace size of 700,000 kg BLS/d (1.5 million lb BLS/d) and 351 operating d/yr.

[d]Emission factors for gaseous organic HAP's are based on the average emission factors for NDCE recovery furnaces equipped with dry-bottom ESP's and dry PM return systems.

[e]Total organic HAP's include all gaseous HAP's in this table except HCl. Numbers may not add exactly due to rounding.

the difference in the acetaldehyde levels may be an anomaly of the limited data on acetaldehyde emissions.[55]

Emission data are available for 10 NDCE recovery furnaces with black liquor in the ESP control system (i.e., 6 NDCE recovery furnaces with wet-bottom ESP's and 4 NDCE recovery furnaces with dry-bottom ESP's and wet PM return systems).[55] Emission data are also available for three NDCE recovery furnaces with dry-bottom ESP's that have dry return of the captured PM.[55] Figure 3-16 illustrates the differences in methanol emissions between the two data sets. The methanol emission data from the ten NDCE recovery furnaces with black liquor in the ESP control system are significantly higher than those from the three NDCE recovery furnaces with dry-bottom ESP's and dry PM return systems.

Based on the available emission test data, eliminating the black liquor from the ESP control system reduces total gaseous organic HAP emissions by approximately 72 percent.[55] In addition to their lower HAP emission potential, NDCE recovery furnaces with dry-bottom ESP's and dry PM return also have lower TRS emissions than NDCE recovery furnaces with black liquor in the ESP control system.[15]

For an NDCE recovery furnace with a furnace size of 700,000 kg BLS per day (1.5 million lb BLS per day) and 351 operating d/yr, eliminating black liquor from the ESP control system has the potential to reduce average gaseous organic HAP emissions by about 22 Mg/yr (24 ton/yr). This emission reduction represents the difference in emissions between NDCE recovery furnaces with black liquor in the ESP control system (i.e., wet-bottom ESP's or dry-bottom ESP's with wet PM return systems) and NDCE recovery furnaces with dry-bottom ESP's and dry PM return systems. Methanol emission reductions account for most of the estimated HAP emission reduction.

3.2.2 Conversion from a DCE Recovery Furnace System to an NDCE Recovery Furnace

The conversion from a DCE recovery furnace system to an NDCE recovery furnace (often referred to as a "low-odor conversion")

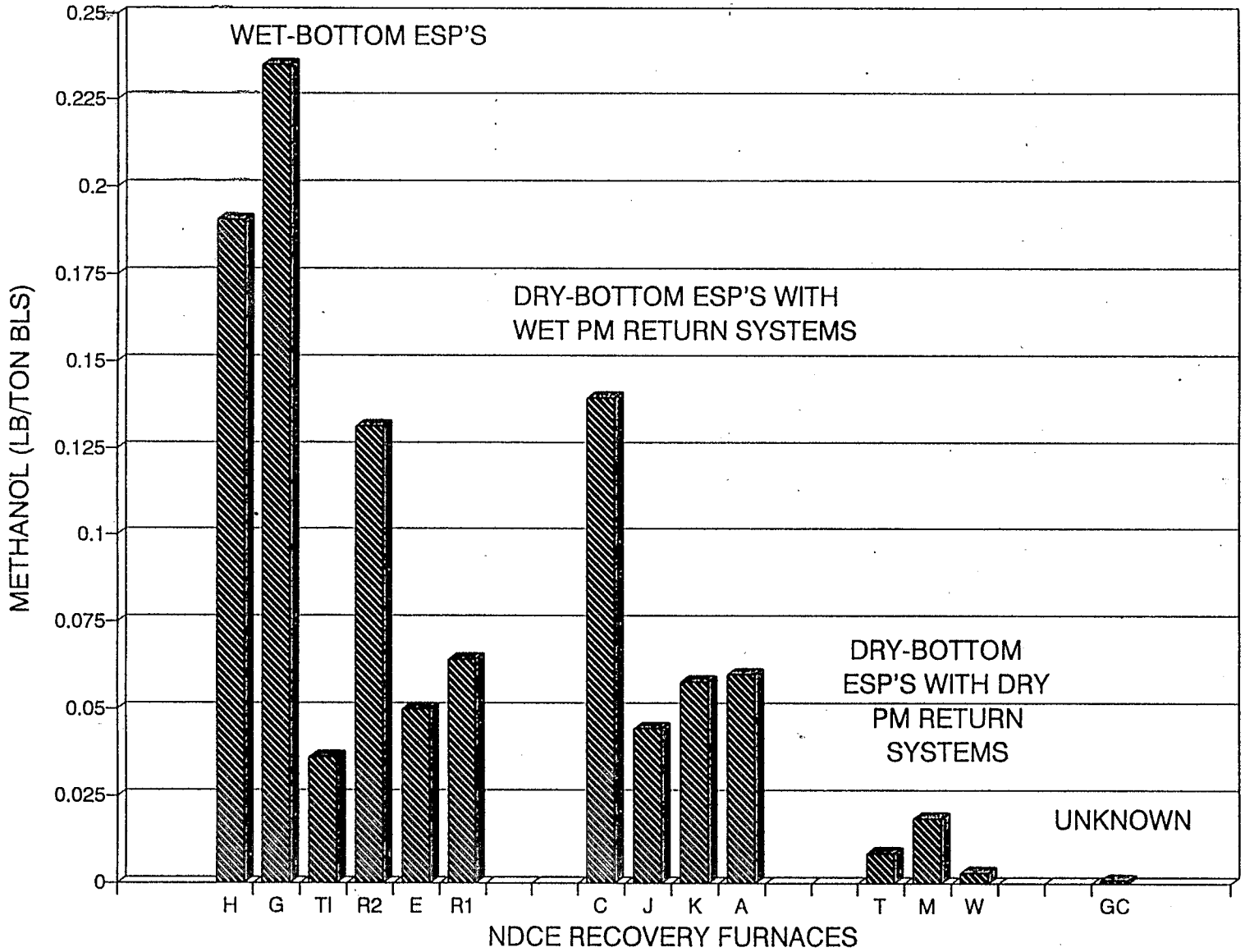

Figure 3-16. Methanol emissions for NDCE recovery furnaces.[52]

offers significant emission reduction and operational benefits. This conversion is an effective measure for reducing HAP emissions from a DCE recovery furnace system because it eliminates two sources of HAP emissions, the DCE and BLO unit. Conversion from a DCE recovery furnace system to an NDCE recovery furnace is a common modification; an estimated 24 percent of existing NDCE recovery furnaces were originally installed as DCE recovery furnaces.[1,61,62] These conversions have been performed for several reasons, including compliance with Federal and State TRS emission standards and increased energy efficiency. Other factors influencing the decision to convert a DCE recovery furnace system to an NDCE design include the age of the existing DCE recovery furnace, the condition of the system, and whether additional capacity is needed at the mill. The following sections describe the major modifications involved in converting a DCE recovery furnace system to an NDCE recovery furnace, including operational benefits, and present the potential HAP emission reductions.

3.2.2.1 Description. The major modifications involved in converting a DCE recovery furnace system to an NDCE recovery furnace are (1) replacing the DCE with a concentrator and associated equipment, (2) extending or replacing the economizer, (3) rebuilding or replacing the ESP, and (4) removing the BLO unit. Removal of the DCE is the driving force behind the latter three major modifications.

The DCE is replaced with a concentrator, which can achieve the desired black liquor solids content without direct contact between recovery furnace exhaust gases and black liquor. Eliminating contact between the hot exhaust gases and black liquor is desirable because the emissions that result from stripping are also eliminated. Operational benefits of replacing the DCE with a low-pressure, steam-driven concentrator include the higher solids content achievable, improved energy utilization, and reduced energy costs.[63] The concentrator, which may be one of several types, including falling film and forced circulation, is installed as part of the evaporator plant. With

current concentrator technologies, the final solids content of the black liquor before firing is generally limited to 75 percent; however, concentrator technologies that can achieve or exceed a solids content of 80 percent are currently in operation.[64] For existing DCE recovery furnace systems, the average solids content of the fired black liquor is 65 percent; for existing NDCE recovery furnaces, the average solids content of the fired black liquor is 68 percent.[1] Increasing the solids content of the fired black liquor is advantageous because it reduces the amount of energy required to evaporate residual moisture from the black liquor solids.

The economizer is modified to recover additional heat from the flue gas that has become available with removal of the DCE. The additional heat recovered in the economizer can be used to produce more low-cost, high-pressure steam.[63] To produce additional steam, the recovery furnace economizer is either expanded with the addition of modules or replaced (i.e., with a long-flow economizer). The economizer is the last section of the recovery furnace that absorbs heat from the flue gas for steam production. The economizer optimizes the thermal efficiency of the recovery furnace by lowering the exit temperature of the exhaust gas as much as possible while minimizing cold-end corrosion. Typical design exit temperatures for an NDCE recovery furnace range from 176° to 190°C (350° to 375°F). For a DCE recovery furnace system, an exit flue gas temperature of 370°C to 430°C (700°F to 800°F) is required to operate the DCE.[65] The thermal efficiency of the recovery furnace increases by 1 percent for every 22°C (40°F) drop in exit gas temperature.[64] Direct contact evaporator recovery furnaces typically operate at thermal efficiencies of 53 to 58 percent, whereas NDCE recovery furnaces typically operate at thermal efficiencies of 63 to 68 percent.[65] Auxiliary equipment that is required when expanding an economizer includes additional soot blowers, hoppers, and conveyors, which collect and transport the additional ash from the economizer.

The ESP is modified to handle the greater PM loading that results from the removal of the DCE. The DCE acts as a PM

control device, collecting 20 to 40 percent of PM emissions from the recovery furnace.[7] Thus, removal of the DCE, without upgrading the ESP, would likely result in increased PM emissions from the recovery furnace stack. Because of the increased PM emissions and a possible change in gas flow rate, the ESP may need to be either upgraded or replaced. As discussed in Section 3.1.1.1, the characteristics of the gas stream for NDCE recovery furnaces contribute to reduced ESP corrosion rates in comparison to DCE recovery furnaces.

With the removal of the DCE, the BLO unit is no longer needed; therefore, the emissions associated with the air-sparging BLO units are eliminated. The operational benefits associated with removing an air-sparging BLO unit are the elimination of the high costs associated with operating this unit and the increased heating value of the black liquor.

3.2.2.2 HAP Emission Reduction Potential. As shown in Table 3-7, the conversion from a DCE recovery furnace system to an NDCE recovery furnace results in substantial reductions for all of the gaseous organic HAP compounds. Based on the available emission test data, the conversion from a DCE recovery furnace system to an NDCE recovery furnace reduces total gaseous organic HAP emissions by approximately 93 percent.[55] In addition to a lower HAP emission potential, NDCE recovery furnaces also emit lower TRS emissions.[15]

Figure 3-17 presents methanol emission data for DCE recovery furnace systems and for NDCE recovery furnaces equipped with dry-bottom ESP's and dry PM return systems.[54] For the DCE recovery furnace systems presented in Figure 3-17, only the methanol emission data from mills that measured emissions from both the DCE recovery furnace and the BLO unit were included. There are three cases in Figure 3-17 where more than one DCE recovery furnace is present at a mill (MO1 and MO4, MP1/2 and MP3, and PB3 and PB4). In those cases, a portion of the total BLO methanol emissions was attributed to each recovery furnace system based on the BLS firing rates measured for the DCE recovery furnaces during the emission tests. (Note: Emissions for MP1/2 are

TABLE 3-7. HAP EMISSIONS FOR DCE RECOVERY FURNACE SYSTEMS AND NDCE RECOVERY FURNACES[55]

Hazardous air pollutant	DCE recovery furnace system		NDCE recovery furnace		Percent reduction
	Emission factor, mass/mass BLS[a,b]	Mass rate, kg/yr (lb/yr)[c]	Emission factor, mass/mass BLS[b,d]	Mass rate, kg/yr (lb/yr)[c]	
Total organic HAP's[e]	5.15×10^{-4}	123,000 (271,000)	3.67×10^{-5}	8,760 (19,300)	93
Acetaldehyde	4.47×10^{-5}	10,700 (23,500)	1.52×10^{-5}	3,630 (8,000)	66
Benzene	1.57×10^{-5}	3,750 (8,270)	1.73×10^{-6}	413 (911)	89
Formaldehyde	3.11×10^{-6}	742 (1,640)	1.93×10^{-6}	461 (1,020)	38
Methanol	3.43×10^{-4}	81,900 (181,000)	4.93×10^{-6}	1,180 (2,600)	99
Methyl ethyl ketone	1.28×10^{-5}	3,050 (6,720)	1.80×10^{-6}	430 (948)	86
Methyl isobutyl ketone	1.39×10^{-5}	3,310 (7,300)	2.27×10^{-6}	542 (1,200)	84
Phenol	4.14×10^{-5}	9,890 (21,800)	2.05×10^{-6}	490 (1,080)	95
Toluene	1.12×10^{-5}	2,670 (5,880)	2.03×10^{-6}	485 (1,070)	82
Xylenes	2.90×10^{-5}	6,920 (15,300)	4.76×10^{-6}	1,140 (2,510)	84
HCl	1.20×10^{-4}	28,700 (63,200)	1.20×10^{-6}	28,700 (63,200)	—

[a]Emission factors for gaseous organic HAP's are based on the sum of the average emission factors for DCE recovery furnaces and BLO units.

[b]Emission factor for HCl is based on the average emission factor for all recovery furnaces since no overall change in HCl emissions is expected from converting a DCE recovery furnace system to an NDCE recovery furnace.

[c]Mass rates are based on a furnace size of 700,000 kg BLS/d (1.5 million lb BLS/d) and 351 operating d/yr.

[d]Emission factors for gaseous organic HAP's are based on the average emission factors for NDCE recovery furnaces equipped with dry-bottom ESP's and dry PM return systems.

[e]Total organic HAP's include all gaseous HAP's in this table except HCl. Numbers may not add exactly due to rounding.

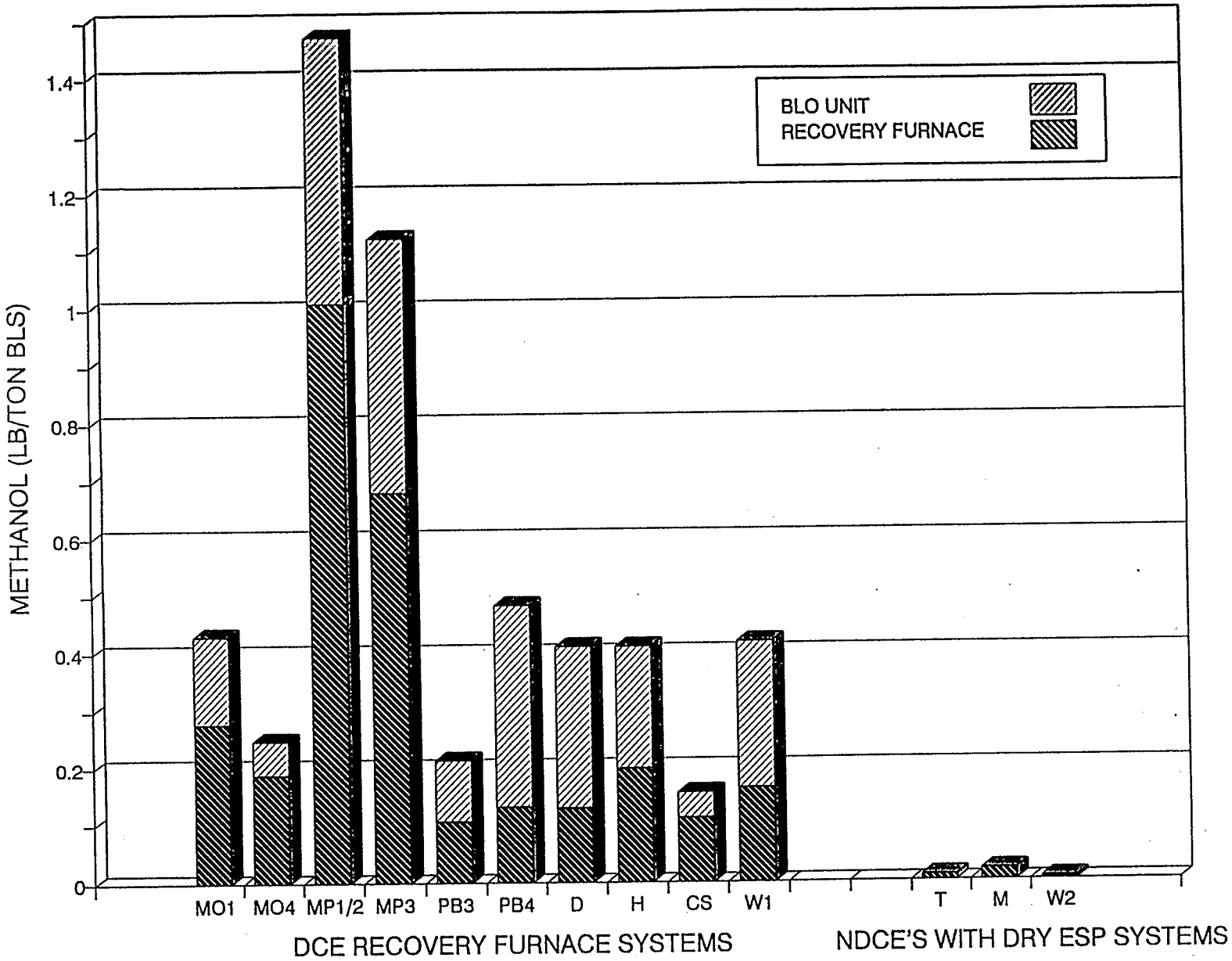

Figure 3-17. Methanol emissions for DCE recovery furnace systems and NDCE recovery furnaces.[52]

actually the combined emissions for DCE recovery furnaces MP1 and MP2, which have a combined stack.)

As shown in Figures 3-18 and 3-19, DCE recovery furnace systems have more gaseous organic HAP emission points than NDCE recovery furnaces (HAP emission points are shaded in the figures). In the DCE system, gaseous organic HAP's such as methanol can be stripped off in the DCE and wet-bottom ESP by contact between the hot flue gases and the black liquor. The BLO vents are also an emission source for methanol and other gaseous organic HAP's because gaseous organic HAP's can be stripped from the black liquor and vented as the oxidizing air is forced up through the black liquor. As described in Section 3.3.1., the wet-bottom ESP (commonly associated with DCE recovery furnaces) can be converted to a dry-bottom ESP with a dry PM return system, thereby eliminating the ESP system as a source of methanol emissions.

For a DCE recovery furnace system with a furnace size of 700,000 kg BLS per day (1.5 million lb BLS per day) and 351 operating d/yr, converting the furnace system to an NDCE design has the potential to reduce average gaseous organic HAP emissions by about 114 Mg/yr (126 ton/yr). This emission reduction represents the difference in emissions between DCE recovery furnace systems (including the BLO unit) and NDCE recovery furnaces (with dry-bottom ESP's and dry PM return systems). Methanol emissions reductions account for most of the estimated gaseous organic HAP emission reduction.

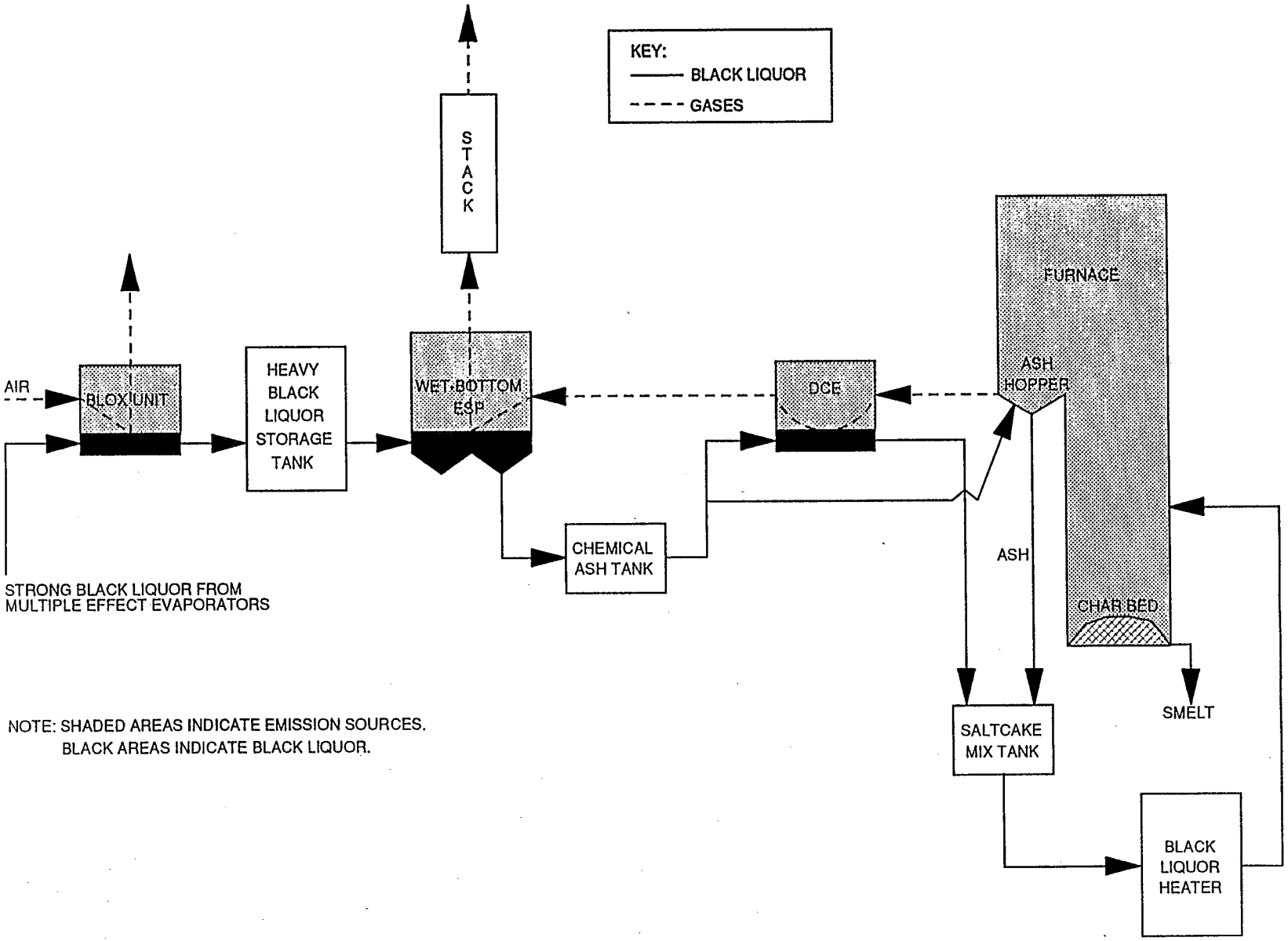

Figure 3-18. Emission sources for a DCE recovery furnace system.

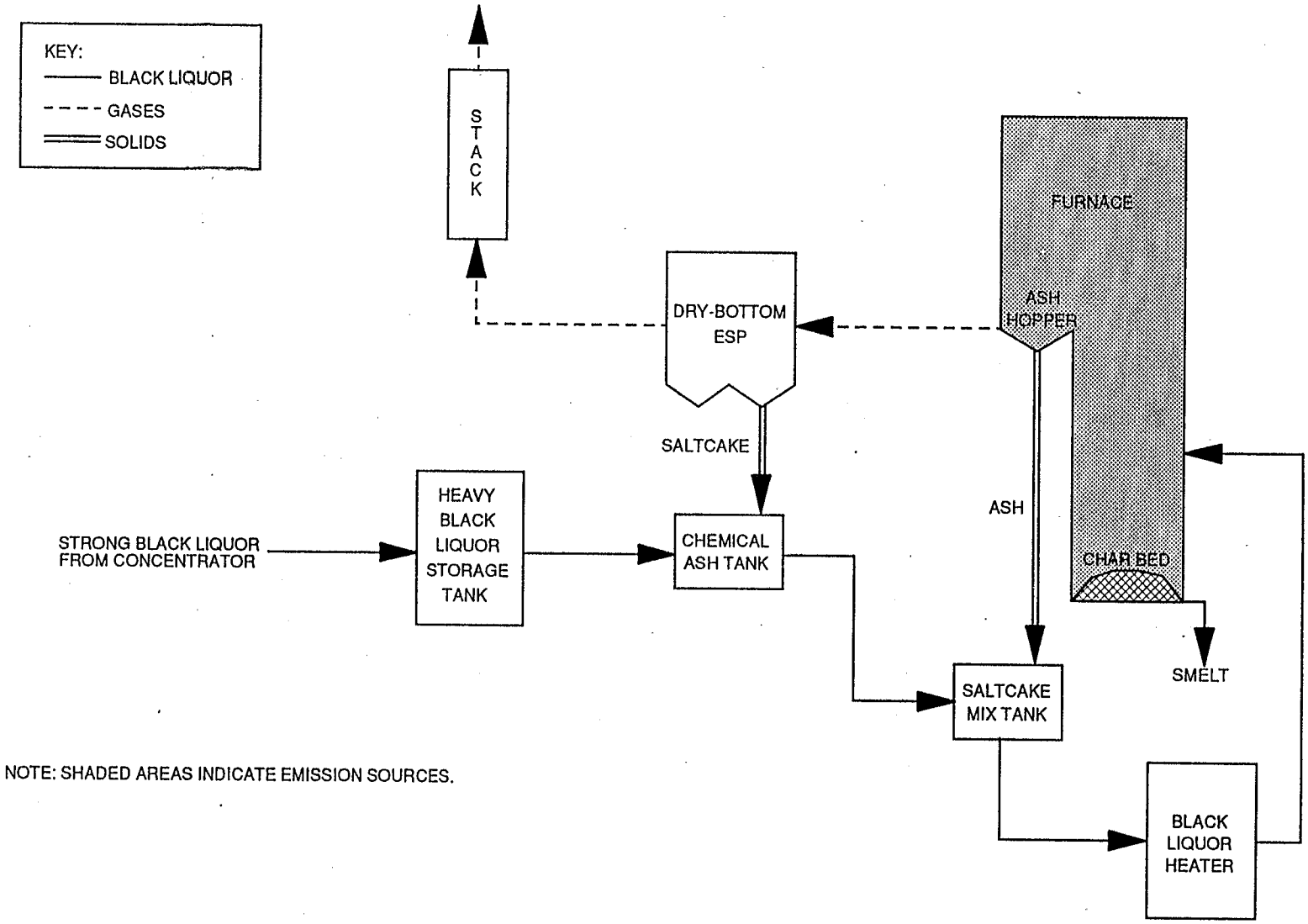

Figure 3-19. Emission sources for an NDCE recovery furnace.

3.3 REFERENCES FOR CHAPTER 3

1. Memorandum from Soltis, V., Nicholson, R., and Holloway, T., MRI, to Telander, J., EPA/MICG. July 29, 1994. Summary of Responses to the NCASI "MACT" Survey--Kraft and Soda Pulp Mills.

2. APTI Course SI:412B: Electrostatic Precipitator Plan Review, Self Instructional Guidebook. U. S. Environmental Protection Agency. Research Triangle Park, NC. Publication No. EPA-450/2-82-019. July 1983.

3. Environmental Pollution Control: Pulp and Paper Industry, Part I, Air. U. S. Environmental Protection Agency. Cincinnati, OH. Publication No. EPA-625/7-76-001. October 1976.

4. Memorandum from Nicholson, R., MRI, to Telander, J., EPA/ISB. May 28, 1993. Meeting minutes from the March 30, 1993 meeting between Environmental Elements Corp. and Midwest Research Institute.

5. Telecon. Bullock, K., MRI, with Bringman, L., Environmental Elements Corp. December 16, 1993. Rigid-electrode ESP's.

6. Memorandum from Nicholson, R., MRI, to Telander, J., EPA/MICG. June 15, 1995. Meeting minutes from the April 12, 1993 meeting between Research-Cottrell and Midwest Research Institute.

7. Review of New Source Performance Standards for Kraft Pulp Mills. U. S. Environmental Protection Agency. Research Triangle Park, NC. Publication No. EPA-450/3-83-017. September 1983.

8. McIlvane, R. Removal of Heavy Metals and Other Utility Air Toxics. (Presented at the EPRI Hazardous Air Pollutant Conference. 1993.)

9. Someshwar, A. Compilation of "Air Toxic" Emission Data for Boilers, Pulp Mills, and Bleach Plants. National Council of the Paper Industry for Air and Stream Improvement, Inc. New York. Technical Bulletin No. 650. June 1993. 128 p.

10. U. S. Environmental Protection Agency. New Source Performance Standards for Kraft Pulp Mills. 43 FR 7568. Washington, DC. U.S. Government Printing Office. February 23, 1978.

11. U. S. Environmental Protection Agency. New Source Performance Standards for Kraft Pulp Mills. 51 FR 18538. Washington, DC. U.S. Government Printing Office. May 20, 1986.

12. Memorandum from Nicholson, R., MRI, to Telander, J., EPA/MICG. June 13, 1996. Addendum to Summary of Responses to the 1992 NCASI MACT Survey.

13. Memorandum from Holloway, T., MRI, to the project file. July 16, 1996. State of Washington PM Data for Kraft Recovery Furnaces, Smelt Dissolving Tanks, and Lime Kilns.

14. Proposed Standards of Performance for Kraft Pulp Mills. In: Standards Support and Environmental Impact Statement. Volume 1. U. S. Environmental Protection Agency. Research Triangle Park, NC. Publication No. EPA-450/2-76-014a. September 1976. pp. 4-2 through 4-10.

15. Memorandum from Soltis, V., MRI, to the project file. April 3, 1995. Kraft and Soda Pulp Mill Combustion Sources Data Base.

16. Telecon. Wallace, D., MRI, with Prouty, A., James River Corp., Camas, WA. March 24, 1994. Performance of the James River recovery furnace ESP's.

17. Operation and Maintenance Manual for Electrostatic Precipitators. U. S. Environmental Protection Agency. Research Triangle Park, NC. Publication No. EPA-625/1-85-017. September 1985.

18. Reference 4, Attachment 9: Sludge lime kiln air pollution control equipment.

19. Someshwar, A. A Study of Kraft Recovery Furnace Hydrochloric Acid Emissions. National Council of the Paper Industry for Air and Stream Improvement, Inc. New York. Technical Bulletin No. 674. August 1994. 100 p.

20. Someshwar, A. and A. Jain. Emissions of Hydrochloric Acid from Kraft Recovery Furnaces. National Council of the Paper Industry for Air and Stream Improvement, Inc. (Presented at the 1992 TAPPI International Chemical Recovery Conference. Seattle, WA. June 18, 1992.) 14 p.

21. Fact Sheet for Prevention of Significant Deterioration and Notice of Construction: James River Camas Mill Energy and Recovery Modernization. Washington Department of Ecology. February 9, 1989.

22. Telecon. Soltis, V., MRI, with Bruno, J., Airpol Inc. October 29, 1993. Information of scrubbers used to control HCl emissions from recovery furnaces at pulp and paper mills.

23. Telecon. Soltis, V., MRI, with Bruno, J., AirPol, Inc. July 2, 1993. Information about scrubbers used to control HCl emissions.

24. Calvert, S. and H. Englund (eds.). Handbook of Air Pollution Technology. New York, John Wiley & Sons, Inc. 1984. p. 137.

25. OAQPS Control Cost Manual. 4th Edition. U. S. Environmental Protection Agency. Research Triangle Park, NC. Publication No. EPA-450/3-90-006. January 1990. p. 9-9.

26. Anderson 2000, Inc. Emission Control Systems for Incinerators. Bulletin TR 89-900239. Peachtree City, GA. February 1989.

27. Reference 24, p. 136.

28. Reference 24, pp. 147-149.

29. Buonicore, A. Absorption. In: Air Pollution Engineering Manual, Air and Waste Management Association. Buonicore, A. and W. Davis (eds.). New York, Van Norstrand Reinhold. 1992. pp. 23-24.

30. Reference 25, pp. 9-28, 9-60, and 9-61.

31. Memorandum from Randall, D. and Holloway, T., MRI, to the project file. October 4, 1994. Absorber Design and Cost Algorithm.

32. Reference 25, p. 9-45.

33. Memorandum from Randall, D., MRI, to the project file. November 8, 1994. Comparison of EPA and NCASI HCl Emission Factors for Recovery Furnaces.

34. Letter and attachments from Sanders, D., Andersen 2000, Inc., to Soltis, V., MRI. November 9, 1993. Preliminary proposal for three separate quench and packed-tower absorption systems to control hydrochloric acid gas emissions from recovery furnaces at kraft pulp and paper mills.

35. Telecon. Soltis, V., MRI, with Bruno, J., AirPol, Inc. October 29, 1993. Information about scrubbers used to control HCl emissions from recovery furnaces at pulp and paper mills.

36. Memorandum from Holloway, T., MRI, to the project file. July 2, 1993. Data to Support HCl Control Technology Transfer from MWI's to Kraft Recovery Furnaces.

37. Handbook: Permit Writer's Guide to Test Burn Data: Hazardous Waste Incineration. U. S. Environmental Protection Agency. Cincinnati, OH. Publication No. EPA-625/6-86-012. September 1986.

38. Trenholm, A., P. Gorman, and G. Junglcaus. Incinerator Performance Evaluation Results. In: Performance Evaluation of Full-Scale Hazardous Waste Incinerators, Volume II. U. S. Environmental Protection Agency. Cincinnati, OH. Publication No. EPA-600/2-84-181b. November 1984.

39. APTI Course ST:412C: Wet Scrubber Plan Review, Self Instructional Guidebook. U. S. Environmental Protection Agency. Research Triangle Park, NC. Publication No. EPA-450/2-82-020. March 1984.

40. Kraft Pulp Mill Inspection Guide. U. S. Environmental Protection Agency. Research Triangle Park, NC. Publication No. EPA-340/1-83-017. January 1983.

41. Roy F. Weston, Inc. Emissions Testing of Combustion Processes in a Pulp and Paper Facility: Champion International Corp., Roanoke Rapids, North Carolina. Prepared for U. S. Environmental Protection Agency--Emission Measurement Branch. Research Triangle Park, NC. EMB Report No. 92-KPM-27. October 1992. Volume I.

42. Someshwar, A. and J. Pinkerton. Wood Processing Industry, In: Air Pollution Engineering Manual, Air and Waste Management Association. Buonicore, A. and W. Davis (eds.). New York, Van Norstrand Reinhold. 1992. p. 840.

43. Modetz, H. and M. Murtiff. Kraft Pulp Mill Industry Particulate Emissions: Source Category Report. U. S. Environmental Protection Agency. Research Triangle Park, NC. Publication No. EPA-600/7-87-006. February 1987.

44. Letter and attachments from Black, C., Champion International Corp., Courtland, AL, to Crowder, J., EPA/ISB. June 10, 1993. Response to request for information on costs for the BLO vent gas control system at the Champion facility in Courtland, AL.

45. Memorandum from Soltis, V. and March, D., MRI, to Telander, J., EPA/ISB. May 17, 1993. Trip Report for Champion International Corp., Courtland, AL.

46. Telecon. Nicholson, R., MRI, with Porritt, T., S.D. Warren Co. May 27, 1993. The black liquor oxidation air pollution control system in use at S.D. Warren.

47. Memorandum from Nicholson, R. and Holloway, T., MRI, to Telander, J., EPA/MICG. September 6, 1995. Trip Report for S.D. Warren Co., Muskegon, MI.

48. Pulp, Paper, and Paperboard Industry--Background Information for Proposed Air Emission Standards, Manufacturing Processes at Kraft, Sulfite, Soda, and Semi-Chemical Mills. U. S. Environmental Protection Agency. Research Triangle Park, NC. Publication No. EPA-453/R-93-050a. October 1993.

49. Collection and Burning of Kraft Non-Condensible Gases--Current Practices, Operating Experience, and Important Aspects of Design and Operation. National Council of the Paper Industry for Air and Stream Improvement. New York. Technical Bulletin No. 469. August 29, 1985. p. 40.

50. An Investigation of Corrosion in Particulate Control Equipment. U. S. Environmental Protection Agency. Washington, DC. Publication No. EPA-340/1-81-002. February 1981. p. 38.

51. Reference 50, p. 51.

52. Hazardous Air Pollutant Emissions from Process Units in the Synthetic Organic Chemical Manufacturing Industry--Background Information for Proposed Standard. Volume 1B, Control Technologies. U. S. Environmental Protection Agency. Research Triangle Park, NC. Publication No. EPA-453/D-92-0166. November 1992. pp. 2-8 and 2-9.

53. Memorandum from Farmer, Jack R., EPA/CPB, to Ajax, B., et al., EPA/CPB. August 22, 1980. Thermal Incinerators and Flares.

54. Reference 52, p. 2-18.

55. Memorandum from Randall, D., Jones, R., and Nicholson, R., MRI, to Telander, J., EPA/MICG. April 25, 1995. Nationwide Baseline Emissions for Combustion Sources at Kraft and Soda Pulp Mills.

56. Final Environmental Impact Statement: Kraft Pulp Mills--Background Information for Promulgated Revisions to Standards. U. S. Environmental Protection Agency. Research Triangle Park, NC. Publication No. EPA-450/3-85-020. May 1986.

57. Memorandum from Eddinger, J., EPA/ISB, to Durkee, K., EPA/ISB. July 8, 1983. Meeting minutes from the April 14, 1983 meeting between kraft pulp mill representatives, the National Council of the Paper Industry for Air and Stream Improvement, Inc., and the U. S. Environmental Protection Agency. Discussion of problems mills have with TRS pickup in wet-bottom ESP's using unoxidized black liquor. Attachments.

58. Letter from Bringman, L., Environmental Elements Corp., to Nicholson, R., MRI. October 12, 1993. Budgetary pricing to convert from wet- to dry-bottom ESP.

59. RUST Environment & Infrastructure. MACT Cost Analysis Report for NCASI. Prepared for the National Council of the Paper Industry for Air and Stream Improvement, Inc. RUST Contract No. 35-4735. October 15, 1993. 47 p.

60. Telecon. Nicholson, R., MRI, with Bringman, L., Environmental Elements Corp. November 30, 1994. Information about the particulate return systems associated with kraft recovery furnace ESP's.

61. Memorandum from Soltis, V., MRI to Telander, J., EPA/ISB. July 14, 1992. Meeting minutes from the February 24, 1992 meeting between Gõtaverken Energy Systems, Inc., the U. S. Environmental Protection Agency, and Midwest Research Institute.

62. Memorandum from Ramsey M. and Nicholson, R., MRI, to Telander, J., EPA/ISB. November 3, 1992. Meeting minutes from the June 24, 1992 meeting between the Babcock & Wilcox Co., the U. S. Environmental Protection Agency, and Midwest Research Institute.

63. Barsin, J., The Babcock and Wilcox Co., Barberton, OH; R. Johnson, James River Corp., Camas, WA; and C. Rissler, James River Corp., St. Francisville, LA. The St. Francisville Recovery Low Odor Conversion and Capacity Upgrade. (Presented at the International Chemical Recovery Conference. Ottawa, Ontario, Canada. April 3-6, 1989.) The Babcock and Wilcox Co. Technical paper BR-1372. 11 p.

64. Green R. and G. Hough (eds.). Chemical Recovery in the Alkaline Pulping Process. 3rd Edition. Prepared by the Alkaline Pulping Committee of the Pulp Manufacture Division. Atlanta, GA, TAPPI Press. 1992. 196 p.

65. Garner, J., Jaako Pöyry. Conversion to Low Odor Improves Recovery Boiler Efficiency and Life. Pulp and Paper. 63(7):91-95. July 1989.

4.0 MODEL PROCESS UNITS, CONTROL OPTIONS, AND ENHANCED MONITORING OPTIONS

This chapter defines model process units, identifies HAP emission control options, and discusses enhanced monitoring options for combustion sources in the chemical recovery area at kraft pulp and paper mills. Model process units represent the types of units that currently exist in the industry and the types that may be constructed in the future. The control options represent demonstrated emission control techniques, and the enhanced monitoring options are methods of demonstrating continuous compliance for a particular control option. The use of model process units to characterize an industry allows the EPA to evaluate the environmental and energy impacts and costs of various control options for each combustion source. These impacts and costs are presented in Chapters 5 and 6, respectively.

4.1 MODEL PROCESS UNITS

This section presents the model process units that were developed for recovery furnaces, SDT's, BLO units, and lime kilns. The following parameters were evaluated to develop the process units: process-specific characteristics, level of PM emission control, stack gas characteristics, and HAP emission levels. The model process unit parameters are typical for equipment that currently exists or may be constructed in the kraft pulp industry.

4.1.1 Recovery Furnace Models

The nine model process units that were developed to characterize kraft recovery furnaces are presented in Table 4-1. Two PM emission levels were developed for each of the nine models

TABLE 4-1a (METRIC). RECOVERY FURNACE MODEL PROCESS UNITS AND PROCESS PARAMETERS[a]

Model No.	Evaporator type	Black liquor firing rate, kg BLS/d	Equivalent pulp production rate		PM emission control device[b]	Gas flow rate--ESP exit, m^3/sec actual	Exhaust gas temp., °C	Moisture content--ESP exit, %	HCl emissions, kg/d	Methanol emissions, kg/d	PM emissions, g/dscm @ 8% O_2
			ADMUP/d	ADMBP/d							
RF-1a	NDCE	0.7 MM	450	380	Dry ESP system	93.4	199	26	81.6	3.4	0.27
RF-1b	NDCE	0.7 MM	450	380	Dry ESP system	93.4	199	26	81.6	3.4	0.10
RF-2a	NDCE	1.2 MM	820	680	Dry ESP system	168	199	26	147	6.0	0.27
RF-2b	NDCE	1.2 MM	820	680	Dry ESP system	168	199	26	147	6.0	0.10
RF-3a	NDCE	1.8 MM	1,200	1,000	Dry ESP system	243	199	26	212	8.7	0.27
RF-3b	NDCE	1.8 MM	1,200	1,000	Dry ESP system	243	199	26	212	8.7	0.10
RF-4a	NDCE	0.7 MM	450	380	Wet ESP system	93.4	199	26	81.6	34.3	0.27
RF-4b	NDCE	0.7 MM	450	380	Wet ESP system	93.4	199	26	81.6	34.3	0.10
RF-5a	NDCE	1.2 MM	820	680	Wet ESP system	168	199	26	147	61.7	0.27
RF-5b	NDCE	1.2 MM	820	680	Wet ESP system	168	199	26	147	61.7	0.10
RF-6a	NDCE	1.8 MM	1,200	1,000	Wet ESP system	243	199	26	212	89.4	0.27
RF-6b	NDCE	1.8 MM	1,200	1,000	Wet ESP system	243	199	26	212	89.4	0.10
RF-7a	DCE	0.4 MM	270	230	Wet ESP system	56.2	160	32	49.0	69.4	0.18
RF-7b	DCE	0.4 MM	270	230	Wet ESP system	56.2	160	32	49.0	69.4	0.10
RF-8a	DCE	0.7 MM	450	380	Wet ESP system	93.4	160	32	81.6	116	0.18
RF-8b	DCE	0.7 MM	450	380	Wet ESP system	93.4	160	32	81.6	116	0.10
RF-9a	DCE	1.2 MM	820	680	Wet ESP system	168	160	32	147	208	0.18
RF-9b	DCE	1.2 MM	820	680	Wet ESP system	168	160	32	147	208	0.10

[a]Metric equivalents in this table were converted from the calculated English unit values given in Table 4-1b.
[b]"Wet ESP system" includes ESP's with black liquor or HAP-contaminated process water in the ESP bottom or PM return system.

TABLE 4-1b (ENGLISH). RECOVERY FURNACE MODEL PROCESS UNITS AND PROCESS PARAMETERS

Model No.	Evaporator type	Black liquor firing rate, lb BLS/d	Equivalent pulp production rate		PM emission control device[a]	Gas flow rate--ESP exit, acfm	Exhaust gas temp., °F	Moisture content, %	HCl emissions, lb/d	Methanol emissions, lb/d	PM emissions, gr/dscf @ 8% O_2
			ADTUP/d	ADTBP/d							
RF-1a	NDCE	1.5 MM	500	420	Dry ESP system	198,000	390	26	180	7.4	0.12
RF-1b	NDCE	1.5 MM	500	420	Dry ESP system	198,000	390	26	180	7.4	0.044
RF-2a	NDCE	2.7 MM	900	750	Dry ESP system	357,000	390	26	324	13.3	0.12
RF-2b	NDCE	2.7 MM	900	750	Dry ESP system	357,000	390	26	324	13.3	0.044
RF-3a	NDCE	3.9 MM	1,300	1,100	Dry ESP system	515,000	390	26	468	19.2	0.12
RF-3b	NDCE	3.9 MM	1,300	1,100	Dry ESP system	515,000	390	26	468	19.2	0.044
RF-4a	NDCE	1.5 MM	500	420	Wet ESP system	198,000	390	26	180	75.6	0.12
RF-4b	NDCE	1.5 MM	500	420	Wet ESP system	198,000	390	26	180	75.6	0.044
RF-5a	NDCE	2.7 MM	900	750	Wet ESP system	357,000	390	26	324	136	0.12
RF-5b	NDCE	2.7 MM	900	750	Wet ESP system	357,000	390	26	324	136	0.044
RF-6a	NDCE	3.9 MM	1,300	1,100	Wet ESP system	515,000	390	26	468	197	0.12
RF-6b	NDCE	3.9 MM	1,300	1,100	Wet ESP system	515,000	390	26	468	197	0.044
RF-7a	DCE	0.9 MM	300	250	Wet ESP system	119,000	320	32	108	153	0.08
RF-7b	DCE	0.9 MM	300	250	Wet ESP system	119,000	320	32	108	153	0.044
RF-8a	DCE	1.5 MM	500	420	Wet ESP system	198,000	320	32	180	255	0.08
RF-8b	DCE	1.5 MM	500	420	Wet ESP system	198,000	320	32	180	255	0.044
RF-9a	DCE	2.7 MM	900	750	Wet ESP system	357,000	320	32	324	459	0.08
RF-9b	DCE	2.7 MM	900	750	Wet ESP system	357,000	320	32	324	459	0.044

[a]"Wet ESP system" includes ESP's with black liquor or HAP-contaminated process water in the ESP bottom or PM return system.

(RF-1a and RF-1b, RF-2a and RF-2b, etc.). Model designations ending in "a" represent pre-new source performance standards (NSPS) furnaces with PM emissions above the NSPS PM limit promulgated in 1978; designations ending in "b" represent furnaces with PM emissions at or below the NSPS. Recovery furnaces have been characterized based on the following parameters: (1) type of final-stage black liquor evaporator (i.e., DCE or NDCE); (2) size (i.e., BLS firing rate and pulp production rate); (3) type of PM emission control device; (4) stack gas characteristics (i.e., flow rate, temperature, and moisture content); and (5) HAP emission levels (i.e., HCl, methanol as a surrogate for gaseous organic HAP's, and PM as a surrogate for PM HAP's). Each of these parameters is discussed below.

4.1.1.1 Evaporator Type. The recovery furnace models were characterized based on whether a DCE or an NDCE (i.e., concentrator) is used in the final stage of black liquor evaporation. Models RF-1 through RF-6 represent NDCE recovery furnaces, and models RF-7 through RF-9 represent DCE recovery furnaces. Separate models were developed for these two types of evaporators because the type of evaporator affects methanol emissions. As discussed in Chapter 2, higher methanol emissions are associated with DCE recovery furnace systems because methanol can be stripped from the black liquor in the DCE and in the BLO unit, which is only used with the DCE systems. Approximately 61 percent of existing recovery furnaces are NDCE recovery furnaces, and the remaining 39 percent are DCE recovery furnace systems.[1] Based on the current trend in the industry, all new recovery furnace installations are expected to be of the NDCE design.

4.1.1.2 Recovery Furnace Size. Model recovery furnaces were characterized by size, based on the BLS firing rate and equivalent unbleached and bleached pulp production rates. The BLS firing rates are expressed in terms of kg BLS/d (lb BLS/d). Equivalent unbleached pulp production rates are expressed in terms of air-dried megagrams of unbleached pulp per day (ADMUP/d)

(air-dried tons of unbleached pulp per day [ADTUP/d]). Similar units (air-dried megagrams of bleached pulp per day [ADMBP/d] and air-dried tons of bleached pulp per day [ADTBP/d]) are used to express the equivalent bleached pulp production rates. The equivalent pulp production rates are based on the BLS firing rates and conversion factors of 1,500 kg BLS/ADMUP (3,000 lb BLS/ADTUP) and 1,800 kg BLS/ADMBP (3,600 lb BLS/ADTBP).[2,3]

The models include three sizes of NDCE recovery furnaces and three sizes of DCE recovery furnaces. For the NDCE recovery furnace models, the BLS firing rates are 0.7 million (MM), 1.2 MM, and 1.8 MM kg BLS/d (1.5 MM, 2.7 MM, and 3.9 MM lb BLS/d). Equivalent unbleached pulp production rates are 450, 820, and 1,200 ADMUP/d (500, 900, and 1,300 ADTUP/d). Equivalent bleached pulp production rates are 380, 680, and 1,000 ADMBP/d (420, 750, and 1,100 ADTBP/d). For the DCE recovery furnace models, the BLS firing rates are 0.4 MM, 0.7 MM, and 1.2 MM kg BLS/d (0.9 MM, 1.5 MM, and 2.7 MM lb BLS/d). Equivalent unbleached pulp production rates are 270, 450, and 820 ADMUP/d (300, 500, and 900 ADTUP/d). Equivalent bleached pulp production rates are 230, 380, and 680 ADMBP/d (250, 420, and 750 ADTBP/d).

The distributions of BLS firing rates for DCE and NDCE recovery furnaces are shown in Figures 4-1a and 4-1b, respectively. The sample populations of DCE and NDCE recovery furnaces were divided into small, medium, and large model sizes. The median BLS firing rate within each size range was then selected as the production rate for the model. Figure 4-2a and 4-2b show the BLS firing rate ranges for DCE and NDCE recovery furnaces, respectively, for the small, medium, and large size categories and the corresponding medians for each size.

As shown in Figures 4-1 through 4-2, DCE recovery furnaces tend to be smaller than NDCE recovery furnaces. The difference in recovery furnace size reflects the fact that in recent years (i.e., since 1980), mills have installed larger recovery furnaces, and the majority of these mills selected the newer NDCE recovery furnace technology.[1,4] Consequently, the six model recovery furnace sizes reflect the size variation between the DCE

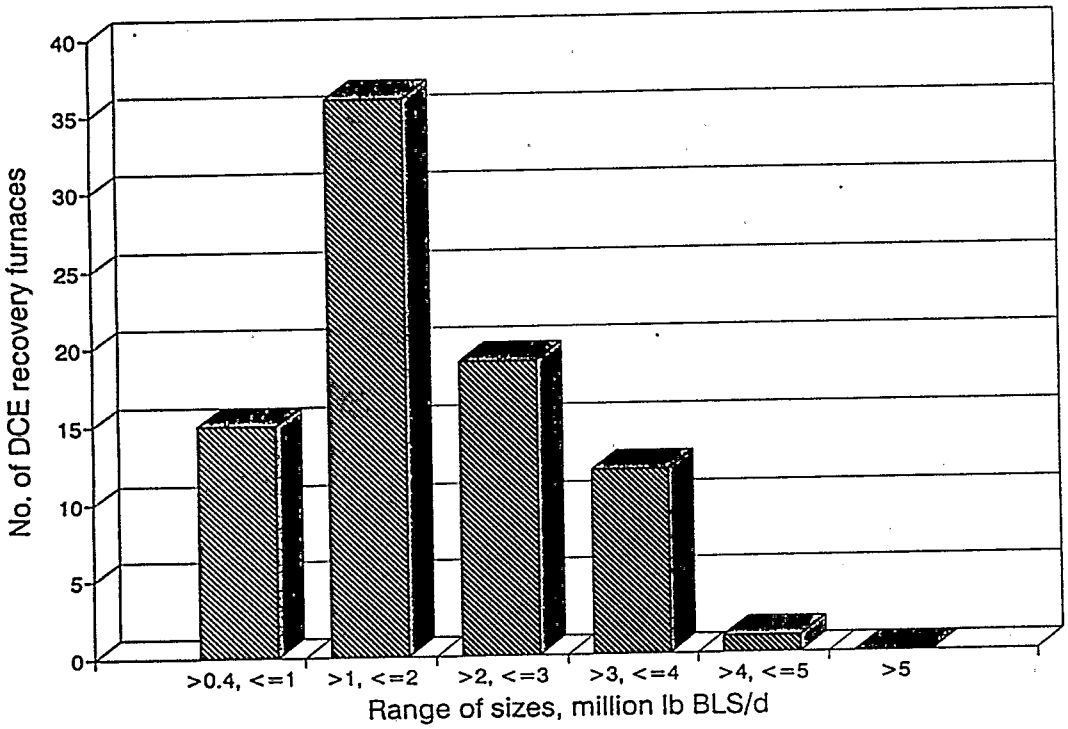

Figure 4-1a. Size distribution for DCE recovery furnaces.

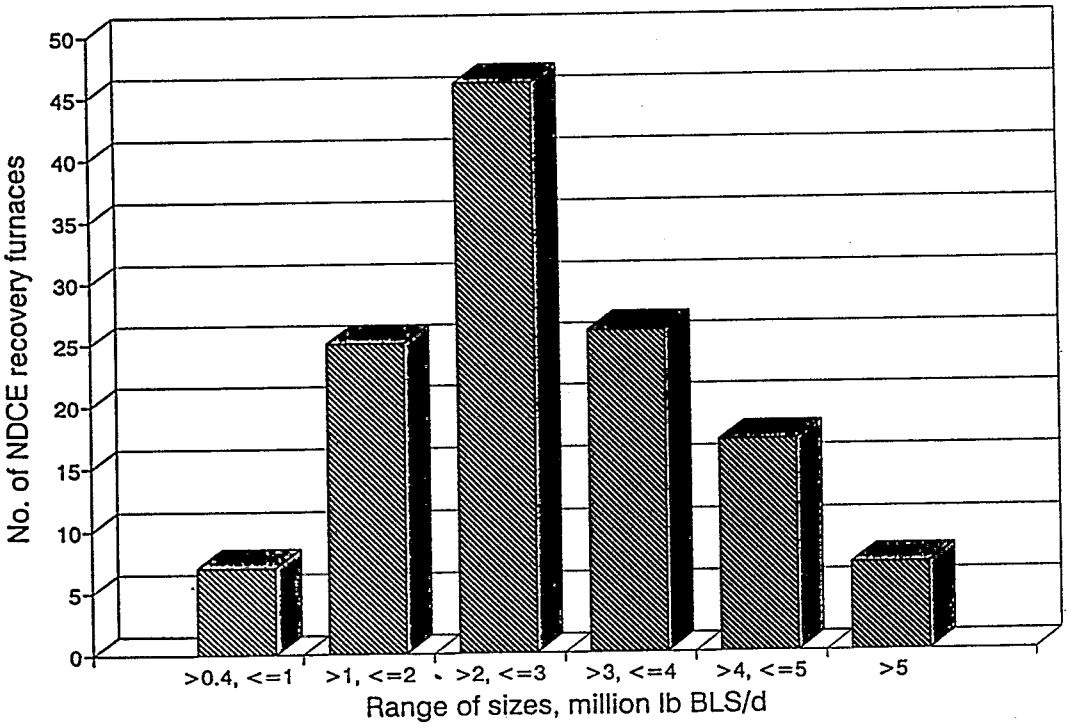

Figure 4-1b. Size distribution for NDCE recovery furnaces.

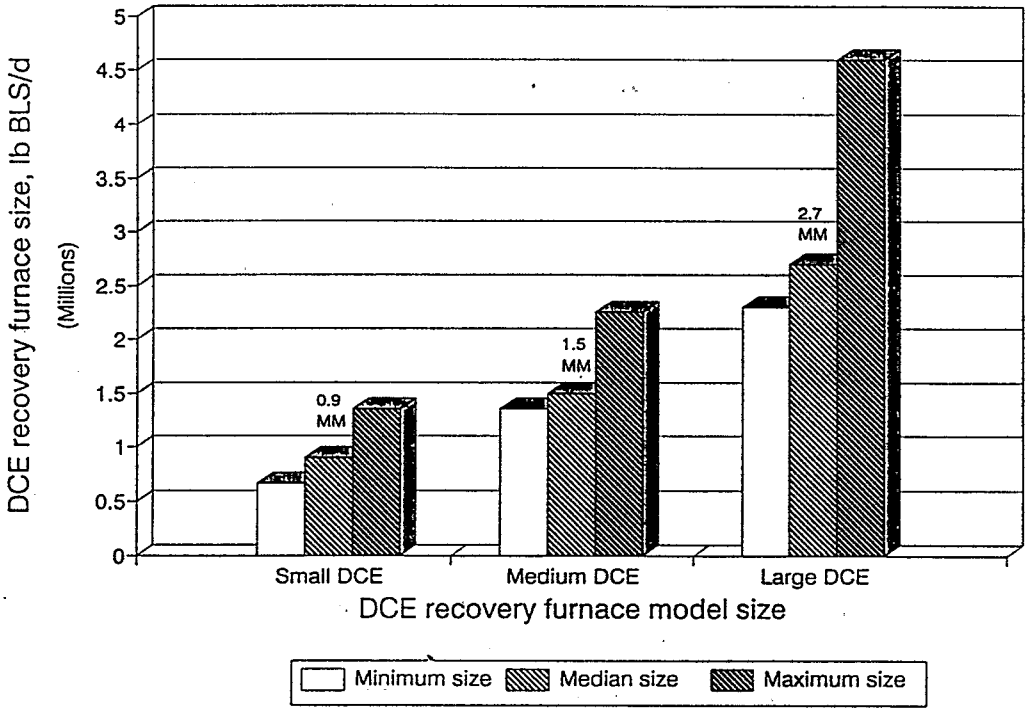

Figure 4-2a. DCE recovery furnace model size ranges.

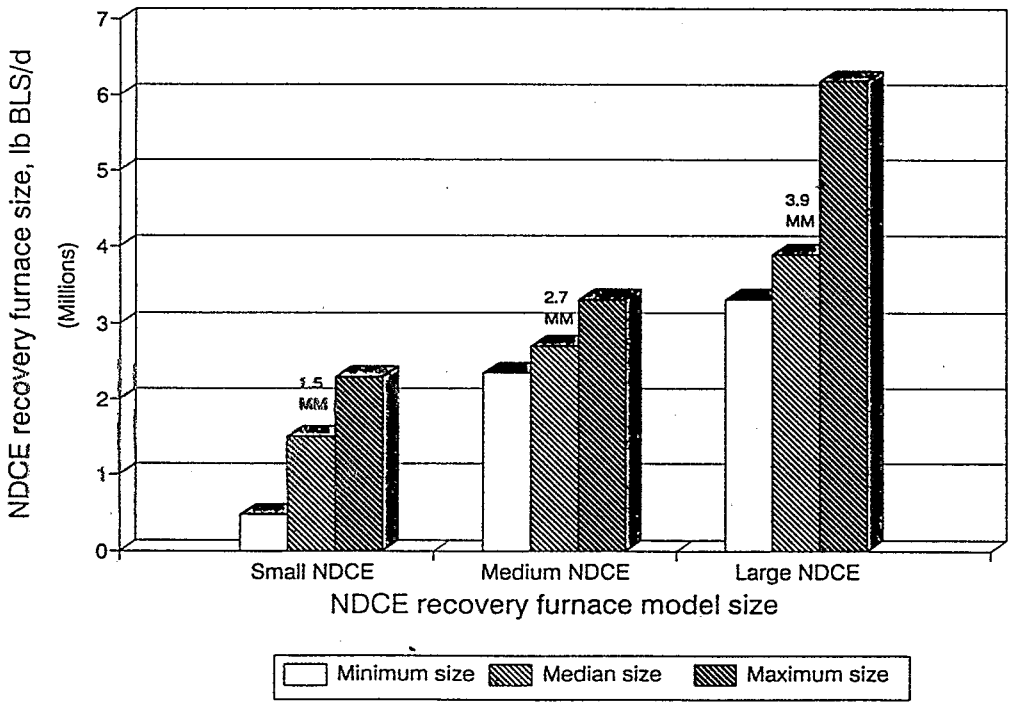

Figure 4-2b. NDCE recovery furnace model size ranges.

and NDCE recovery furnaces. The 0.7 and 1.2 MM kg BLS/d (1.5 and 2.7 MM lb BLS/d BLS firing rates for the medium (RF-8) and large (RF-9) model DCE recovery furnaces correspond to the small (RF-1 and RF-4) and medium (RF-2 and RF-5) model NDCE recovery furnaces, respectively. A separate small DCE recovery furnace model (RF-7) with a BLS firing rate of 0.4 MM kg BLS/d (0.9 MM lb BLS/d) and large NDCE recovery furnace models (RF-3 and RF-6) with a BLS firing rate of 1.8 MM kg BLS/d (3.9 MM lb BLS/d) were also selected to characterize the existing population of kraft recovery furnaces.

4.1.1.3 PM Emission Control Device. The PM emission control device for all of the model recovery furnaces is an ESP. Electrostatic precipitators are the most prevalent PM emission control device for recovery furnaces, installed on about 99 percent of recovery furnaces.[1] The model recovery furnaces are further characterized by the type of ESP system (i.e., with or without black liquor or HAP-contaminated process water in the ESP bottom or PM return system). The DCE recovery furnace systems are represented by models RF-7 through RF-9; NDCE recovery furnaces with dry ESP systems (i.e., dry-bottom ESP's with dry PM return systems) are represented by models RF-1 through RF-3; NDCE recovery furnaces with wet ESP systems (i.e., wet-bottom ESP's or dry-bottom ESP's with wet PM return systems) are represented by models RF-4 through RF-6. Nondirect contact evaporator recovery furnaces equipped with wet ESP systems emit higher quantities of gaseous organic HAP's than those equipped with dry ESP systems, due to stripping of HAP's from the black liquor or HAP-contaminated water in the ESP bottom or PM return system. However, new ESP's tend to have dry ESP systems.[5] For NDCE recovery furnaces, models were developed for both types of ESP systems. All of the DCE recovery furnace models have wet ESP systems because 94 percent of DCE recovery furnaces are equipped with wet-bottom ESP's, and information is not available to determine if the remaining 6 percent that are equipped with dry-bottom ESP's also have dry PM return systems.

4.1.1.4 Stack Gas Characteristics. Model recovery furnaces were characterized by stack gas characteristics (i.e., flow rate at the ESP exit, temperature, and moisture content). The gas flow rates were calculated from the BLS feed rates using conversion factors of 13.9 MJ/kg BLS (6,000 Btu/lb BLS) and 0.242 dscm at 0 percent O_2/MJ (9,000 dscf at 0 percent $O_2/10^6$ Btu).[3,6]

The exhaust gas temperatures used for the model recovery furnaces are based on ESP inlet gas temperatures from emissions tests.[2] Inlet temperature is an acceptable estimator for outlet temperature because the temperature does not vary significantly across an ESP. For model NDCE recovery furnaces, the stack gas temperature is 199°C (390°F). For model DCE recovery furnaces, the stack gas temperature is 160°C (320°F).

Stack gas moisture contents for the recovery furnace models are averages of measurements taken during emission tests of two NDCE recovery furnaces and two DCE recovery furnaces.[7] The average stack gas moisture content based on the two NDCE recovery furnace emission tests is 26 percent. The average stack gas moisture content based on the two DCE recovery furnace emission tests is 32 percent.

4.1.1.5 HAP Emissions. Model recovery furnaces were characterized by the levels of HCl, methanol, which is a surrogate for gaseous organic HAP's, and PM, which is a surrogate for PM HAP's. The HCl emission levels for the model recovery furnaces are based on the HCl emission factor of 1.20×10^{-4} kg of HCl/kg of BLS fired (1.20×10^{-4} lb of HCl/lb of BLS fired) presented in Chapter 2. As discussed in Chapter 2, HCl emissions from recovery furnace systems are highly variable, and insufficient data exist to determine the process operating parameters that significantly influence emissions. In addition, the available data on HCl emissions from NDCE and DCE recovery furnaces show a significant overlap. Therefore, the same emission factor is applied for both NDCE and DCE model recovery furnaces.

The methanol emission levels for model recovery furnaces are also based on emission factors presented in Chapter 2. Methanol emissions for the model DCE recovery furnaces are based on an emission factor of 1.70×10^{-4} kg methanol/kg BLS (1.70×10^{-4} lb methanol/lb BLS). For model NDCE recovery furnaces with wet ESP systems, methanol emission levels are based on an emission factor of 5.04×10^{-5} kg methanol/kg BLS (5.04×10^{-5} lb methanol/lb BLS). For model NDCE recovery furnaces with dry ESP systems, methanol emission levels are based on an emission factor of 4.93×10^{-6} kg methanol/kg BLS (4.93×10^{-6} lb methanol/lb BLS). Separate methanol emission levels were developed because DCE recovery furnaces have higher methanol emissions than NDCE recovery furnaces due to the stripping of methanol from the black liquor. Also, methanol emissions are higher from NDCE recovery furnaces with wet ESP systems than from NDCE recovery furnaces with dry ESP systems, due to the same stripping effects.

The NSPS for kraft pulp mills establishes a PM emission limit of 0.1 g/dscm (0.044 gr/dscf) at 8 percent O_2 that is applicable to recovery furnaces constructed, modified, or reconstructed after September 24, 1976.[8,9] Available data indicate only the year, not the month, that recovery furnaces were installed.[1,2] Consequently, this analysis assumes that those recovery furnaces installed during or after 1977 are subject to the NSPS, and those recovery furnaces installed before 1977 are not subject to the NSPS. Also, recovery furnaces subject to the NSPS are assumed to be in compliance with the NSPS PM standard.

The "a" model recovery furnaces shown in Table 4-1 characterize those recovery furnaces with PM emissions above the NSPS PM emission limit. Therefore, the "a" model PM emission levels were based on the PM emission concentrations for those recovery furnaces that are not subject to the NSPS and that emit more than the NSPS PM limit. Different PM levels were calculated for the DCE and NDCE recovery furnaces that are subject to these parameters. For NDCE recovery furnaces, the PM emission level is

0.27 g/dscm (0.12 gr/dscf). For DCE recovery furnaces, the PM emission level is 0.18 g/dscm (0.08 gr/dscf).[10]

The "b" model recovery furnaces represent those recovery furnaces with PM emissions at or below the NSPS PM limit of 0.10 g/dscm (0.044 gr/dscf). The "b" model recovery furnace PM emission level is equivalent to the NSPS limit and represents the maximum emission level for these recovery furnaces. Although only 30 percent of the recovery furnaces are subject to the NSPS, the majority (approximately 80 percent) of all recovery furnaces reportedly are meeting the NSPS PM emission level.[1,2] Therefore, the "b" models represent the majority of existing recovery furnaces.

4.1.2 Smelt Dissolving Tank Models

The seven model process units (SDT-1 through SDT-7) that were developed to characterize existing SDT's are presented in Table 4-2. Model process units SDT-1 through SDT-4 are also representative of those SDT's that are expected to be constructed in the future. The parameters selected to characterize typical SDT's are (1) size (i.e., BLS firing rate of the associated recovery furnace, equivalent pulp production rate, and smelt flow rate), (2) PM emission control device, (3) inlet and outlet gas stream characteristics (i.e., flow rate, temperature and moisture content), and (4) HAP emission levels (i.e., PM as a surrogate for PM HAP's).

4.1.2.1 SDT Size. The SDT size is indicated by the BLS firing rate and equivalent pulp production rate of the associated recovery furnace and the smelt flow rate. The smelt flow rate was calculated based on a conversion factor of 0.37 kg smelt/kg BLS (0.37 lb smelt/lb BLS).[2] An equivalent BLS firing rate was used as an indicator of SDT size because the SDT is an integral part of the kraft recovery furnace. The SDT models represent the majority of recovery furnace configurations, i.e., one SDT per recovery furnace. Approximately 8 percent of recovery furnaces have two SDT's.[1] The SDT models SDT-1 and SDT-5 correspond to the small DCE recovery furnace model (RF-7); SDT models SDT-2 and SDT-6 correspond to the medium DCE and small

TABLE 4-2a (METRIC). SMELT DISSOLVING TANK MODEL PROCESS UNITS AND PROCESS PARAMETERS[a]

Model No.	Equivalent black liquor firing rate, kg BLS/d	Equivalent pulp production rate		Smelt flow rate, kg/d	PM emission control device	PM emission control device inlet gas flow rate, m^3/sec actual @ 93°C	Stack gas flow rate, m^3/sec actual @ 77°C	Moisture content, %	PM emissions, kg/Mg BLS
		ADMUP/d	ADMBP/d						
SDT-1	0.4 MM	270	230	151,000	Wet scrubber	4.4	4.2	36	0.18
SDT-2	0.7 MM	450	380	252,000	Wet scrubber	7.4	7.1	36	0.18
SDT-3	1.2 MM	820	680	453,000	Wet scrubber	13.4	12.7	36	0.18
SDT-4	1.8 MM	1,200	1,000	655,000	Wet scrubber	19.3	18.4	36	0.18
SDT-5	0.4 MM	270	230	151,000	Mist eliminator	4.4	4.2	36	0.23
SDT-6	0.7 MM	450	380	252,000	Mist eliminator	7.4	7.1	36	0.23
SDT-7	1.2 MM	820	680	453,000	Mist eliminator	13.4	12.7	36	0.23

[a]Metric equivalents in this table were converted from the calculated English unit values given in Table 4-2b.

TABLE 4-2b (ENGLISH). SMELT DISSOLVING TANK MODEL PROCESS UNITS AND PROCESS PARAMETERS

Model No.	Equivalent black liquor firing rate, lb BLS/d	Equivalent pulp production rate		Smelt flow rate, lb/d	PM emission control device	PM emission control device inlet gas flow rate, acfm @ 200°F	Stack gas flow rate, acfm @ 170°F	Moisture content, %	PM emissions, lb/ton BLS
		ADTUP/d	ADTBP/d						
SDT-1	0.9 MM	300	250	333,000	Wet scrubber	9,400	9,000	36	0.37
SDT-2	1.5 MM	500	420	555,000	Wet scrubber	15,700	15,000	36	0.37
SDT-3	2.7 MM	900	750	999,000	Wet scrubber	28,300	27,000	36	0.37
SDT-4	3.9 MM	1,300	1,100	1,443,000	Wet scrubber	40,900	39,000	36	0.37
SDT-5	0.9 MM	300	250	333,000	Mist eliminator	9,400	9,000	36	0.46
SDT-6	1.5 MM	500	420	555,000	Mist eliminator	15,700	15,000	36	0.46
SDT-7	2.7 MM	900	750	999,000	Mist eliminator	28,300	27,000	36	0.46

NDCE recovery furnace models (RF-1, RF-4, and RF-8); SDT models SDT-3 and SDT-7 correspond to the large DCE and medium NDCE recovery furnace models (RF-2, RF-5, and RF-9); SDT model SDT-4 corresponds to the large NDCE recovery furnace models (RF-3 and RF-6).

4.1.2.2 PM Emission Control Device. Two types of emission control devices are predominantly used to control PM emissions from SDT's--wet scrubbers and mist eliminators. Wet scrubbers are used to control PM emissions from about 87 percent of the SDT's, and mist eliminators are used to control PM emissions from about 10 percent of the SDT's.[11] The type of PM emission control device will impact PM emission levels because wet scrubbers are generally more effective than mist eliminators at controlling PM emissions. The PM emission control device for SDT models SDT-1 through SDT-4 is a wet scrubber. These SDT models correspond to all four recovery furnace model sizes and include sizes for both DCE and NDCE recovery furnaces. The PM emission control device for SDT models SDT-5 through SDT-7 is a mist eliminator. These SDT models correspond to the three DCE recovery furnace model sizes and the small and medium NDCE recovery furnace model sizes. An SDT model with a mist eliminator that corresponds to the large NDCE recovery furnace model size was not included because less than 1 percent of recovery furnaces have this configuration.

4.1.2.3 Inlet and Outlet Gas Stream Characteristics. The model SDT stack gas flow rates represent typical gas flow rates within each size range. Available recovery furnace BLS firing rates and corresponding SDT gas flow rates were used to develop a factor of 1.04×10^{-5} m^3/sec/kg BLS/d (0.01 acfm/lb BLS/d).[2] The model gas flow rates calculated with this factor were then compared to, and found to be consistent with, the gas flow rates at actual mills with comparable BLS firing rates.

The model stack temperature of 77°C (170°F) is based on process information reported by mills and on measurements during emission tests.[1,12,13] The process information shows stack temperatures range from 49° to 100°C (120° to 212°F). The

emission test reports listed stack temperatures of 82°C (180°F) and 76°C (168°F).

The stack gas moisture content for model SDT's is 36 percent, which is the average moisture content from four emission tests.[7] The average moisture contents of the stack gases for the four SDT's that were tested are 35.5 percent, 46.7 percent, 25.4 percent, and 38.5 percent.[7]

The inlet gas flow rates were calculated from the stack gas flow rates and estimated inlet temperature and moisture content. The model inlet temperature of 93°C (200°F) was estimated based on available process information reported by individual mills.[2]

4.1.2.4 HAP Emissions. Particulate matter emissions, as a surrogate for PM HAP emissions, are characterized by the model SDT's. The models characterize SDT's with PM emissions above the NSPS PM emission limit for SDT's, i.e., 0.10 kg/Mg (0.20 lb/ton) BLS. Therefore, the model PM emission levels were based on the PM levels for those SDT's that are not subject to the NSPS and that emit more than the NSPS PM limit. The PM levels do not represent the PM emission performance of typical SDT's because approximately 75 percent of existing SDT's emit, on average, less than the NSPS.[2] The model SDT PM emission levels were selected so that the emission reduction potential and control costs for those SDT's with PM emission levels higher than the NSPS could be evaluated.

Scrubbers are generally more effective at removing PM than mist eliminators; therefore, different PM levels were calculated for SDT's with wet scrubbers and SDT's with mist eliminators. Average PM emissions are 0.18 kg/Mg (0.37 lb/ton) BLS if a wet scrubber is the PM emission control device, and 0.23 kg/Mg (0.46 lb/ton) BLS if a mist eliminator is the PM emission control device.[10]

4.1.3 Black Liquor Oxidation Unit Models

The three model process units that were developed to characterize existing BLO units are presented in Table 4-3. With the possible exception of the estimated one kraft pulp mill that operates DCE recovery furnaces but does not have a BLO unit, no

TABLE 4-3a (METRIC). BLACK LIQUOR OXIDATION UNIT MODEL PROCESS UNITS AND PROCESS PARAMETERS[a]

Model No.	Equipment type	Equivalent black liquor firing rate, kg BLS/d	Equivalent pulp production rate		Vent gas flow rate, m^3/sec actual @ 54°C	Moisture content, %	Air pollution control device	Methanol emissions, kg/d
			ADMUP/d	ADTBP/d				
BLO-1	2-stage, air-sparging	0.4 MM	270	230	4.2	35	Uncontrolled	70.8
BLO-2	2-stage, air-sparging	0.7 MM	450	380	8.5	35	Uncontrolled	118
BLO-3	2-stage, air sparging	1.2 MM	820	680	12.7	35	Uncontrolled	212

[a]Metric equivalents in this table were converted from the calculated English unit values given in Table 4-3b.

TABLE 4-3b (ENGLISH). BLACK LIQUOR OXIDATION UNIT MODEL PROCESS UNITS AND PROCESS PARAMETERS

Model No.	Equipment type	Equivalent black liquor firing rate, lb BLS/d	Equivalent pulp production rate		Vent gas flow rate, acfm @ 130°F	Moisture content, %	Air pollution control device	Methanol emissions, lb/d
			ADTUP/d	ADTBP/d				
BLO-1	2-stage, air-sparging	0.9 MM	300	250	8,900	35	Uncontrolled	156
BLO-2	2-stage, air-sparging	1.5 MM	500	420	18,000	35	Uncontrolled	260
BLO-3	2-stage, air-sparging	2.7 MM	900	750	26,900	35	Uncontrolled	467

new BLO units are expected to be built.[11] The parameters selected to characterize typical BLO units are (1) equipment type, (2) size (i.e., equivalent BLS firing rate and pulp production rate), (3) air pollution control device, (4) vent characteristics (i.e., flow rate and moisture content), and (5) HAP emission levels (i.e., methanol).

4.1.3.1 Equipment Type. Two types of BLO units are used at kraft pulp mills--air-sparging units and molecular O_2 systems. Models were developed only for air-sparging units, which account for an estimated 94 percent of the BLO units.[1,11] The BLO models were selected to characterize the majority of BLO systems; therefore, BLO models were not developed for molecular O_2 systems. Also, molecular O_2 BLO systems are essentially closed systems that do not emit HAP's.

4.1.3.2 BLO Size. Three model sizes for BLO systems were selected to correspond to the DCE recovery furnace model sizes. Therefore, the BLS firing rates and pulp production rates for the DCE recovery furnace models were selected as the parameters that reflect the size of the BLO unit. The equivalent BLS firing rates for models BLO-1, BLO-2, and BLO-3 are 0.4 MM, 0.7 MM, and 1.2 MM kg BLS/d (0.9 MM, 1.5 MM, and 2.7 MM lb BLS/d), respectively.

4.1.3.3 Air Pollution Control Device. Air emissions from approximately 95 percent of air-sparging BLO systems are uncontrolled.[11] Those few mills (about 2 mills) that control BLO emissions incinerate the BLO vent gases in a power boiler.[11] The BLO models were selected to characterize the majority of BLO systems; therefore, BLO model vent emissions are uncontrolled.

4.1.3.4 Vent Characteristics. The vent gas flow rates, temperature, and moisture content for the model BLO units are based on typical process data provided by mills for actual units within each model size range. The vent gas flow rates for the three model BLO units are 4.2, 8.5, and 12.7 m^3/sec (8,900, 18,000, and 26,900 acfm).[2] The exhaust gas temperature is 54°C (130°F).[11] The exhaust gas moisture content is 35 percent.[14,15]

4.1.3.5 HAP Emissions. As noted in Chapter 2, methanol emission levels associated with the BLO models were calculated using an emission factor of 1.73×10^{-4} kg/kg BLS (1.73×10^{-4} lb/lb BLS).

4.1.4 Lime Kiln Models

The six model process units (LK-1 through LK-6) that were developed to characterize rotary lime kilns are presented in Table 4-4. Current industry trends are (1) using ESP's rather than scrubbers to control PM emissions and (2) longer and larger diameter lime kilns.[11] Therefore, the majority of new lime kiln installations will likely be represented by the larger models LK-5 and LK-6. The parameters selected to characterize lime kilns are (1) size (i.e., equivalent pulp production rate, lime production rate, length, and diameter), (2) PM emission control device, (3) stack gas characteristics (i.e., flow rate, temperature and moisture content), and (4) HAP emission levels (i.e., PM as a surrogate for PM HAP's).

4.1.4.1 Lime Kiln Size. The lime kiln size is indicated by the equivalent pulp production rate, lime (i.e., CaO) production rate, length, and diameter.

The model sizes for lime kilns were determined using available lime production rates.[2] The size distribution, in Mg/d (ton/d) of lime, for lime kilns is shown in Figure 4-3. The sample population of lime kilns was divided into small, medium, and large model sizes. The median lime production rate within each size range was then selected as the lime production rate for the model. Based on the data presented in Figure 4-3, the three lime kiln model sizes are 90, 180, and 270 Mg/d (100, 200, and 300 ton/d) of CaO. The daily equivalent pulp production rate was estimated using a conversion factor of 275 kg CaO/ADMP (550 lb CaO/ADTP).[3]

The length and diameter were determined based on the lime production rates for the small, medium, and large models and available data on lime kiln dimensions at similar production rates.[2] Lime kiln lengths range from 37 to 137 m (120 to 450 ft) and average 82 m (270 ft); lime kiln diameters range from 2.1 to

TABLE 4-4a (METRIC). LIME KILN MODEL PROCESS UNITS AND PROCESS PARAMETERS[a]

Model No.	Equivalent pulp production rate, ADMP/d	CaO production rate, Mg/d	Length, m	Diameter, m	PM emission control device	PM emission control device inlet gas flow rate, m^3/sec actual @ 249°C	Stack gas flow rate & temp., m^3/sec actual @ °C	Moisture content, %	PM emissions, g/dscm @10% O_2
LK-1	320	90	61	2.7	Venturi scrubber	10.4	7.3 @ 71°C	30	0.27
LK-2	680	180	84	3.0	Venturi scrubber	20.1	14.2 @ 71°C	30	0.27
LK-3	1,000	270	107	3.7	Venturi scrubber	34.1	24.1 @ 71°C	30	0.27
LK-4	320	90	61	2.7	ESP	10.4	10.4 @ 249°C	25	0.15
LK-5	680	180	84	3.0	ESP	20.1	20.1 @ 249°C	25	0.15
LK-6	1,000	270	107	3.7	ESP	34.1	34.1 @ 249°C	25	0.15

[a]Metric equivalents in this table were converted from the calculated English unit values given in Table 4-4b.

TABLE 4-4b (ENGLISH). LIME KILN MODEL PROCESS UNITS AND PROCESS PARAMETERS

Model No.	Equivalent pulp production rate, ADTP/d	CaO production rate, ton/d	Length, ft	Diameter, ft	PM emission control device	PM emission control device inlet gas flow rate, acfm @ 480 °F	Stack gas flow rate & temp., acfm @ °F	Moisture content, %	PM emissions, gr/dscf @10% O_2
LK-1	350	100	200	9.0	Venturi scrubber	22,000	15,500 @ 160°F	30	0.12
LK-2	750	200	275	10	Venturi scrubber	42,500	30,000 @ 160°F	30	0.12
LK-3	1,100	300	350	12	Venturi scrubber	72,200	51,000 @ 160°F	30	0.12
LK-4	350	100	200	9.0	ESP	22,000	22,000 @ 480°F	25	0.067
LK-5	750	200	275	10	ESP	42,500	42,500 @ 480°F	25	0.067
LK-6	1,100	300	350	12	ESP	72,200	72,200 @ 480°F	25	0.067

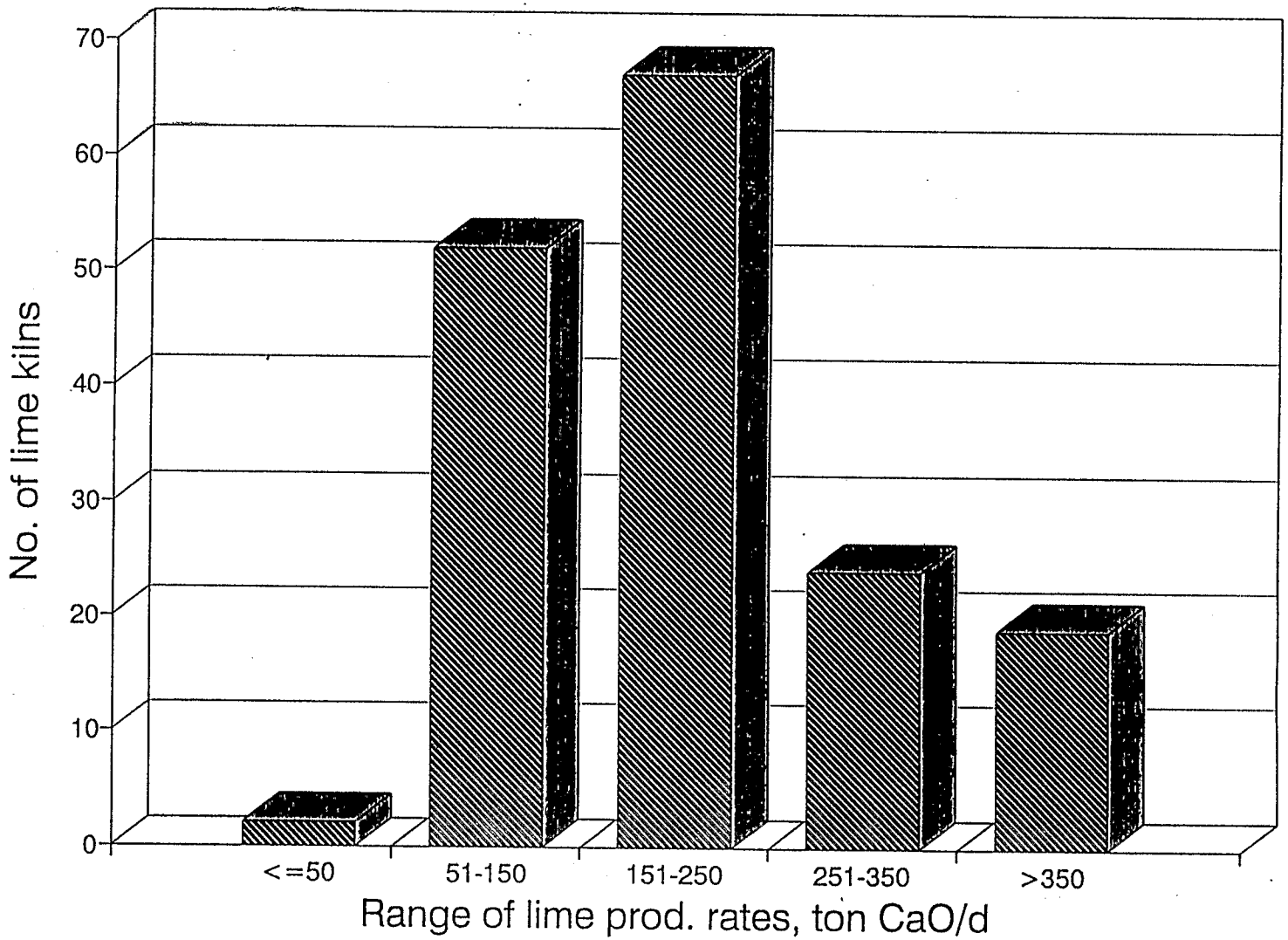

Figure 4-3. Size distribution for lime kilns.

4.6 m (7 to 15 ft) and average 3.2 m (10.5 ft).[11] The model lime kiln dimensions were selected to cover these ranges.

4.1.4.2 PM Emission Control Device. Both wet scrubbers and ESP's are used by the kraft pulp industry to control PM emissions from lime kilns.[11] Particulate matter emissions from the majority of lime kilns (90 percent) are controlled with wet scrubbers.[11] Venturi scrubbers are the most prevalent type of wet scrubber used (89 percent of all wet scrubbers used to control PM emissions are venturi scrubbers).[11] Particulate matter emissions from the remaining 10 percent of lime kilns are controlled by ESP's.[11] Lime kiln models were developed to characterize both types of PM emission control devices. The PM emission control device for models LK-1 through LK-3 is a venturi scrubber. For models LK-4 through LK-6, the PM emission control device is an ESP.

4.1.4.3 Stack Gas Characteristics. The stack gas flow rates for the lime kiln models with scrubbers represent typical gas flow rates within each size range based on available process data.[2] The model gas flow rates for the small, medium, and large models are 7.3, 14.2, and 24.1 m^3/sec (15,500, 30,000, and 51,000 acfm), respectively. The exhaust gas temperature of 71°C (160°F) is based on process data provided by mills.[2,11] The range of stack temperatures reported is 49° to 102°C (120° to 216°F).[11] The model exhaust gas moisture content of 30 percent is based on test data.[14]

Model inlet gas flow rates were calculated from the stack gas flow rates for the lime kiln models with scrubbers and estimated inlet temperature and moisture content. An inlet temperature of 250°C (480°F) was based on process data provided by mills.[11] The range of inlet temperatures is 150° to 315°C (300° to 600°F).[11] An inlet moisture content of 25 percent was assumed based on the moisture content from test results for a lime kiln with an ESP.[2] Because the gas stream temperature and moisture content do not vary significantly across the ESP, the inlet and outlet gas stream conditions are the same for the model lime kilns with ESP's.

4.1.4.4 HAP Emissions. The model lime kilns shown in Table 4-4 include PM emission levels as a surrogate for PM HAP emissions. The PM emission level for model lime kilns with wet scrubbers (LK-1 through LK-3) is 0.27 g/dscm (0.12 gr/dscf) at 10 percent O_2.[10] This emission level characterizes all lime kilns with PM emission levels above the NSPS PM emission limit for lime kilns firing natural gas, i.e., 0.15 g/dscm (0.067 gr/dscf); the emission level represents both natural gas-fired and oil-fired lime kilns. The PM level does not represent the PM emission performance of typical lime kilns because approximately 64 percent of existing lime kilns emit less than the NSPS limit for gas-fired lime kilns.[2] The model lime kiln PM emission level was selected so that the emission reduction potential and control costs for those lime kilns with PM emission levels higher than the NSPS could be evaluated.

Electrostatic precipitators are generally more effective at removing PM than venturi scrubbers; therefore, different PM levels were calculated for lime kilns with ESP's. Particulate matter emission levels for lime kiln models with ESP's (LK-4 through LK-6) are equivalent to the NSPS level, 0.15 g/dscm (0.067 gr/dscf). All existing lime kilns equipped with ESP's reportedly have PM emissions less than or equal to the NSPS PM limit for gas-fired lime kilns.[2]

4.2 CONTROL OPTIONS

Control options for recovery furnaces, SDT's, BLO units, and lime kilns were developed based on the emission control information presented in Chapter 3. This section identifies and briefly describes these control options. Controlled emission levels are presented for each applicable model process unit, along with model control device parameters, where applicable.

4.2.1 Recovery Furnace Control Options

Table 4-5 presents the four control options that have been evaluated for recovery furnaces. Table 4-5 also identifies the recovery furnace type that is affected by the control option and the pollutants that would be controlled. The four control options are (1) conversion of a DCE recovery furnace system to an

NDCE recovery furnace (i.e., low-odor conversion), (2) wet to dry ESP system conversion, (3) PM controls, and (4) addition of a packed-bed scrubber.

TABLE 4-5. RECOVERY FURNACE CONTROL OPTIONS

Control option	Affected recovery furnace type	Pollutants Controlled
Low-odor conversion[a]	DCE recovery furnaces	Gaseous organic HAP's and PM HAP's
Wet to dry ESP system conversion	NDCE recovery furnaces with wet ESP systems	Gaseous organic HAP's (e.g., methanol)
PM Controls:		
(a) ESP upgrade or replacement in order to meet NSPS PM emission level of 0.10 g/dscm (0.044 gr/dscf)	NDCE and DCE recovery furnaces currently emitting PM in quantities above NSPS PM emission level	PM HAP's
(b) ESP upgrade or replacement plus addition of packed-bed scrubber in order to meet PM level of 0.034 g/dscm (0.015 gr/dscf).	NDCE and DCE recovery furnaces currently emitting PM in quantities greater than 0.034 g/dscm (0.015 gr/dscf).	PM HAP's
Packed-bed scrubber	All recovery furnaces	HCl

[a]In addition to the elimination of the BLO unit and the conversion of the evaporator to the noncontact design, the low-odor conversion option also includes an ESP upgrade or replacement (including wet to dry ESP system conversion) to meet the applicable PM emission limit listed under "PM Controls."

4.2.1.1 Conversion of a DCE Recovery Furnace System to an NDCE Recovery Furnace. Converting a DCE recovery furnace system to an NDCE recovery furnace (or "low-odor conversion") was evaluated as a control option for reducing gaseous organic HAP emissions from DCE recovery furnace systems. Under this control option, the DCE is eliminated from the chemical recovery process and replaced with a concentrator, the BLO unit is eliminated, and the wet ESP system is converted to a dry ESP system. Based on the emission factors presented in Sections 4.1.1.5 and 4.1.3.5, this control option reduces methanol emissions by 99 percent. Because the DCE provides some PM control, as discussed in Chapter 3, conversion to an NDCE recovery furnace has the potential to increase PM emissions. Therefore, depending on the design characteristics of the ESP, an ESP upgrade or replacement may be required to achieve compliance with PM emission limits.

Two levels of PM control were included as part of the low-odor conversion option. A PM control level of 0.10 g/dscm (0.044 gr/dscf) at 8 percent O_2 (i.e., the NSPS limit) was evaluated for both "a" and "b" model recovery furnaces. Although "b" model recovery furnaces have a baseline of 0.10 g/dcsm (0.044 gr/dscf), conversion to an NDCE recovery furnace has the potential to increase PM emissions above the baseline. Therefore, PM controls are necessary to maintain a PM level of 0.10 g/dscm (0.044 gr/dscf) for the "b" model recovery furnaces. A control level of 0.034 g/dscm (0.015 gr/dscf) at 8 percent O_2 was also evaluated for both "a" and "b" model recovery furnaces. Tables 4-6 and 4-7 present the model recovery furnaces that were evaluated for this control option. Controlled emission levels also are included in these tables.

4.2.1.2 Wet to Dry ESP System Conversion. Converting wet ESP systems to dry ESP systems was evaluated as a control option for reducing methanol and other gaseous organic HAP emissions from NDCE recovery furnaces. With this control option, stripping of methanol from the black liquor or HAP-contaminated water in the ESP bottom or PM return system is eliminated. Using the methanol emission factors presented in Section 4.1.1.5 for wet and dry ESP systems, methanol emissions are reduced by 72 percent. The model recovery furnaces used to evaluate the impact of this option are presented in Table 4-8.

4.2.1.3 PM Emission Controls. One PM emission control option was evaluated that would reduce PM emissions from existing NDCE and DCE recovery furnace systems to the NSPS PM limit. The control option would involve either an ESP upgrade or an ESP replacement to meet the NSPS limit of 0.10 g/dscm (0.044 gr/dscf). The model recovery furnaces (NDCE and DCE "a" models) analyzed for this control option are presented in Table 4-9.

A second PM emission control option was evaluated that would reduce PM emissions from NDCE and DCE recovery furnaces to a more stringent level of 0.034 g/dscm (0.015 gr/dscf). The control option would involve (1) upgrading the existing ESP and

TABLE 4-6a (METRIC). RECOVERY FURNACE MODELS: LOW-ODOR CONVERSION CONTROL OPTION (INCLUDES WET TO DRY ESP SYSTEM CONVERSION AND PM CONTROL TO NSPS LEVEL)[a]

Model No.	Evaporator type	Equivalent pulp production rate		Black liquor firing rate, kg BLS/d	PM emission control device	Control option PM emission control device	Gas flow rate--ESP exit, m^3/sec actual	PM emissions, g/dscm	Controlled PM emissions, g/dscm	Methanol emissions, kg/d[b]	Controlled methanol emissions, kg/d[b]
		ADMUP/d	ADMBP/d								
RF-7a	DCE	270	230	0.4 MM	Wet ESP system	Dry ESP system	56.2	0.18	0.10	140	2.0
RF-8a	DCE	450	380	0.7 MM	Wet ESP system	Dry ESP system	93.4	0.18	0.10	234	3.4
RF-9a	DCE	820	680	1.2 MM	Wet ESP system	Dry ESP system	168	0.18	0.10	420	6.0

[a]Metric equivalents in this table were converted from the calculated English unit values given in Table 4-6b.
[b]Includes methanol emissions from recovery furnace stack and BLO vent.

TABLE 4-6b (ENGLISH). RECOVERY FURNACE MODELS: LOW-ODOR CONVERSION CONTROL OPTION (INCLUDES WET TO DRY ESP SYSTEM CONVERSION AND PM CONTROL TO NSPS LEVEL)

Model No.	Evaporator type	Equivalent pulp production rate		Black liquor firing rate, lb BLS/d	PM emission control device	Control option PM emission control device	Gas flow rate--ESP exit, acfm	PM emissions, gr/dscf	Controlled PM emissions, gr/dscf	Methanol emissions, lb/d[a]	Controlled methanol emissions, lb/d[a]
		ADTUP/d	ADTBP/d								
RF-7a	DCE	300	250	0.9 MM	Wet ESP system	Dry ESP system	119,000	0.08	0.044	309	4.4
RF-8a	DCE	500	420	1.5 MM	Wet ESP system	Dry ESP system	198,000	0.08	0.044	515	7.4
RF-9a	DCE	900	750	2.7 MM	Wet ESP system	Dry ESP system	357,000	0.08	0.044	926	13.3

[a]Includes methanol emissions from recovery furnace stack and BLO vent.

TABLE 4-7a (METRIC). RECOVERY FURNACE MODELS: LOW-ODOR CONVERSION CONTROL OPTION (INCLUDING WET TO DRY ESP SYSTEM CONVERSION AND PM CONTROL TO 0.034 G/DSCM)[a]

Model No.	Evaporator type	Equivalent pulp production rate		Black liquor firing rate, kg BLS/d	PM emission control device	Control option PM emission control device	Gas flow rate--ESP exit, m^3/sec actual	PM emissions, g/dscm	Controlled PM emissions, g/dscm	Methanol emissions, kg/d[b]	Controlled methanol emissions, kg/d[b]
		ADMUP/d	ADMBP/d								
RF-7a	DCE	270	230	0.4 MM	Wet ESP system	Dry ESP system	56.2	0.18	0.034	140	2.0
RF-7b	DCE	270	230	0.4 MM	Wet ESP system	Dry ESP system	56.2	0.10	0.034	140	2.0
RF-8a	DCE	450	380	0.7 MM	Wet ESP system	Dry ESP system	93.4	0.18	0.034	234	3.4
RF-8b	DCE	450	380	0.7 MM	Wet ESP system	Dry ESP system	93.4	0.10	0.034	234	3.4
RF-9a	DCE	820	680	1.2 MM	Wet ESP system	Dry ESP system	168	0.18	0.034	420	6.0
RF-9b	DCE	820	680	1.2 MM	Wet ESP system	Dry ESP system	168	0.10	0.034	420	6.0

[a]Metric equivalents in this table were converted from the calculated English unit values given in Table 4-7b.
[b]Includes methanol emissions from recovery furnace stack and BLO vent.

TABLE 4-7b (ENGLISH). RECOVERY FURNACE MODELS: LOW-ODOR CONVERSION CONTROL OPTION (INCLUDING WET TO DRY ESP SYSTEM CONVERSION AND PM CONTROL TO 0.015 GR/DSCF)[a]

Model No.	Evaporator type	Equivalent pulp production rate		Black liquor firing rate, lb BLS/d	PM emission control device	Control option PM emission control device	Gas flow rate--ESP exit, acfm	PM emissions, gr/dscf	Controlled PM emissions, gr/dscf	Methanol emissions, lb/d[a]	Controlled methanol emissions, lb/d[a]
		ADTUP/d	ADTBP/d								
RF-7a	DCE	300	250	0.9 MM	Wet ESP system	Dry ESP system	119,000	0.08	0.015	309	4.4
RF-7b	DCE	300	250	0.9 MM	Wet ESP system	Dry ESP system	119,000	0.044	0.015	309	4.4
RF-8a	DCE	500	420	1.5 MM	Wet ESP system	Dry ESP system	198,000	0.08	0.015	515	7.4
RF-8b	DCE	500	420	1.5 MM	Wet ESP system	Dry ESP system	198,000	0.044	0.015	515	7.4
RF-9a	DCE	900	750	2.7 MM	Wet ESP system	Dry ESP system	357,000	0.08	0.015	926	13.3
RF-9b	DCE	900	750	2.7 MM	Wet ESP system	Dry ESP system	357,000	0.044	0.015	926	13.3

[a]Includes methanol emissions from recovery furnace stack and BLO vent.

TABLE 4-8a (METRIC). RECOVERY FURNACE MODELS: WET TO DRY ESP SYSTEM CONVERSION CONTROL OPTION[a]

Model No.	Evaporator type	Black liquor firing rate, kg BLS/d	Equivalent pulp production rate		PM emission control device[b]	Control option PM emission control device	Gas flow rate--ESP exit, m^3/sec actual	Methanol emissions, kg/d	Controlled methanol emissions, kg/d
			ADMUP/d	ADMBP/d					
RF-4	NDCE	0.7 MM	450	380	Wet ESP system	Dry ESP system	93.4	34.3	3.4
RF-5	NDCE	1.2 MM	820	680	Wet ESP system	Dry ESP system	168	61.7	6.0
RF-6	NDCE	1.8 MM	1,200	1,000	Wet ESP system	Dry ESP system	243	89.4	8.7

[a]Metric equivalents in this table were converted from the calculated English unit values given in Table 4-8b.
[b]"Wet ESP system" includes ESP's with black liquor or HAP-contaminated process water in the ESP bottom or PM return system.

TABLE 4-8b (ENGLISH). RECOVERY FURNACE MODELS: WET TO DRY ESP SYSTEM CONVERSION CONTROL OPTION

Model No.	Evaporator type	Black liquor firing rate, lb BLS/d	Equivalent pulp production rate		PM emission control device[a]	Control option PM emission control device	Gas flow rate--ESP exit, acfm	Methanol emissions, lb/d	Controlled methanol emissions, lb/d
			ADTUP/d	ADTBP/d					
RF-4	NDCE	1.5 MM	500	420	Wet ESP system	Dry ESP system	198,000	75.6	7.4
RF-5	NDCE	2.7 MM	900	750	Wet ESP system	Dry ESP system	357,000	136	13.3
RF-6	NDCE	3.9 MM	1,300	1,100	Wet ESP system	Dry ESP system	515,000	197	19.2

[a]"Wet ESP system" includes ESP's with black liquor or HAP-contaminated process water in the ESP bottom or PM return system.

TABLE 4-9a (METRIC). RECOVERY FURNACE MODELS: PM CONTROL OPTIONS (0.10 G/DSCM)[a]

Model No.	Evaporator type	Black liquor firing rate, kg BLS/d	Equivalent pulp production rate		PM emission control device	Gas flow rate--ESP exit, m^3/sec actual	PM emissions, g/dscm @ 8% O_2	Controlled PM emissions, g/dscm @ 8% O_2
			ADMUP/d	ADMBP/d				
RF-1a	NDCE	0.7 MM	450	380	ESP	93.4	0.27	0.10
RF-2a	NDCE	1.2 MM	820	680	ESP	168	0.27	0.10
RF-3a	NDCE	1.8 MM	1,200	1,000	ESP	243	0.27	0.10
RF-7a	DCE	0.4 MM	270	230	ESP	56.2	0.18	0.10
RF-8a	DCE	0.7 MM	450	380	ESP	93.4	0.18	0.10
RF-9a	DCE	1.2 MM	820	680	ESP	168	0.18	0.10

[a]Metric equivalents in this table were converted from the calculated English unit values given in Table 4-9b.

TABLE 4-9b (ENGLISH). RECOVERY FURNACE MODELS: PM CONTROL OPTIONS (0.044 GR/DSCF)

Model No.	Evaporator type	Black liquor firing rate, lb BLS/d	Equivalent pulp production rate		PM emission control device	Gas flow rate--ESP exit, acfm	PM emissions, gr/dscf @ 8% O_2	Controlled PM emissions, gr/dscf @ 8% O_2
			ADTUP/d	ADTBP/d				
RF-1a	NDCE	1.5 MM	500	420	ESP	198,000	0.12	0.044
RF-2a	NDCE	2.7 MM	900	750	ESP	357,000	0.12	0.044
RF-3a	NDCE	3.9 MM	1,300	1,100	ESP	515,000	0.12	0.044
RF-7a	DCE	0.9 MM	300	250	ESP	119,000	0.08	0.044
RF-8a	DCE	1.5 MM	500	420	ESP	198,000	0.08	0.044
RF-9a	DCE	2.7 MM	900	750	ESP	357,000	0.08	0.044

installing a packed-bed scrubber or (2) installing a new ESP and a packed-bed scrubber. The control option would be applicable to both new and existing recovery furnaces. The model recovery furnaces analyzed for this control option are presented in Table 4-10.

Note that because the type of ESP bottom or PM return system does not affect PM emission levels, this parameter was not considered in evaluating either of the PM control options.

4.2.1.4 Packed-Bed Scrubber. The installation of packed-bed scrubbers was evaluated as an HCl control option for existing and new NDCE and DCE recovery furnaces, and for DCE recovery furnaces that have undergone a conversion to the NDCE furnace design. Packed-bed scrubbers are capable of controlling outlet HCl emissions by 99 percent or to levels less than or equal to 5 ppmv.[16,17] This emission level (i.e., 5 ppmv) corresponds to an emission factor of 6.20×10^{-5} kg HCl/kg BLS (6.20×10^{-5} lb HCl/lb BLS) for the model NDCE recovery furnaces and converted DCE's. The corresponding emission factor for the model DCE recovery furnaces is 6.53×10^{-5} kg HCl/kg BLS (6.53×10^{-5} lb HCl/lb BLS). Under this control option, HCl emissions from the model NDCE recovery furnaces (and converted DCE's) are reduced by 48 percent, and HCl emissions from the model DCE's are reduced by 46 percent. The model recovery furnaces that were analyzed for this control option are listed in Tables 4-11 and 4-12.

4.2.2 SDT Control Options

One PM emission control option was evaluated that would reduce PM emissions from existing SDT's to the NSPS PM limit. The control option would involve replacing the existing mist eliminator or existing scrubber with a new wet scrubber designed to meet the NSPS PM limit. The model SDT's analyzed for this control option are presented in Table 4-13.

A second PM emission control option was evaluated that would reduce PM emissions to a more stringent level of 0.06 kg/Mg BLS (0.12 lb/ton BLS). The control option would involve (1) replacing the existing mist eliminator or scrubber with a new

TABLE 4-10a (METRIC). RECOVERY FURNACE MODELS: PM CONTROL OPTIONS (0.034 G/DSCM)[a]

Model No.	Evaporator type	Black liquor firing rate, kg BLS/d	Equivalent pulp production rate		PM emission control device	Gas flow rate--ESP exit, m^3/sec actual	PM emissions, g/dscm @ 8% O_2	Controlled PM emissions, g/dscm @ 8% O_2
			ADMUP/d	ADMBP/d				
RF-1a	NDCE	0.7 MM	450	380	ESP	93.4	0.27	0.034
RF-1b	NDCE	0.7 MM	450	380	ESP	93.4	0.10	0.034
RF-2a	NDCE	1.2 MM	820	680	ESP	168	0.27	0.034
RF-2b	NDCE	1.2 MM	820	680	ESP	168	0.10	0.034
RF-3a	NDCE	1.8 MM	1,200	1,000	ESP	243	0.27	0.034
RF-3b	NDCE	1.8 MM	1,200	1,000	ESP	243	0.10	0.034
RF-7a	DCE	0.4 MM	270	230	ESP	56.2	0.18	0.034
RF-7b	DCE	0.4 MM	270	230	ESP	56.2	0.10	0.034
RF-8a	DCE	0.7 MM	450	380	ESP	93.4	0.18	0.034
RF-8b	DCE	0.7 MM	450	380	ESP	93.4	0.10	0.034
RF-9a	DCE	1.2 MM	820	680	ESP	168	0.18	0.034
RF-9b	DCE	1.2 MM	820	680	ESP	168	0.10	0.034

[a]Metric equivalents in this table were converted from the calculated English unit values given in Table 4-10b.

TABLE 4-10b (ENGLISH). RECOVERY FURNACE MODELS: PM CONTROL OPTIONS (0.015 GR/DSCF)

Model No.	Evaporator type	Black liquor firing rate, lb BLS/d	Equivalent pulp production rate		PM emission control device	Gas flow rate--ESP exit, acfm	PM emissions, gr/dscf @ 8% O_2	Controlled PM emissions, gr/dscf @ 8% O_2
			ADTUP/d	ADTBP/d				
RF-1a	NDCE	1.5 MM	500	420	ESP	198,000	0.12	0.015
RF-1b	NDCE	1.5 MM	500	420	ESP	198,000	0.044	0.015
RF-2a	NDCE	2.7 MM	900	750	ESP	357,000	0.12	0.015
RF-2b	NDCE	2.7 MM	900	750	ESP	357,000	0.044	0.015
RF-3a	NDCE	3.9 MM	1,300	1,100	ESP	515,000	0.12	0.015
RF-3b	NDCE	3.9 MM	1,300	1,100	ESP	515,000	0.044	0.015
RF-7a	DCE	0.9 MM	300	250	ESP	119,000	0.08	0.015
RF-7b	DCE	0.9 MM	300	250	ESP	119,000	0.044	0.015
RF-8a	DCE	1.5 MM	500	420	ESP	198,000	0.08	0.015
RF-8b	DCE	1.5 MM	500	420	ESP	198,000	0.044	0.015
RF-9a	DCE	2.7 MM	900	750	ESP	357,000	0.08	0.015
RF-9b	DCE	2.7 MM	900	750	ESP	357,000	0.044	0.015

TABLE 4-11a (METRIC). RECOVERY FURNACE MODELS: HCl CONTROL OPTION (PACKED-BED SCRUBBER)[a]

Model No.	Black liquor firing rate, kg BLS/d	Equivalent pulp production rate		PM emission control device	Gas flow rate--ESP exit, m^3/sec actual	Control option control device	HCl emissions, kg/d	Controlled HCl emissions, kg/d
		ADMUP/d	ADMBP/d					
RF-1	0.7 MM	450	380	ESP	93.4	ESP+packed-bed scrubber	81.6	42.2
RF-2	1.2 MM	820	680	ESP	168	ESP+packed-bed scrubber	147	75.7
RF-3	1.8 MM	1,200	1,000	ESP	243	ESP+packed-bed scrubber	212	110
RF-7	0.4 MM	270	230	ESP	56.2	ESP+packed-bed scrubber	49.0	26.7
RF-8	0.7 MM	450	380	ESP	93.4	ESP+packed-bed scrubber	81.6	44.5
RF-9	1.2 MM	820	680	ESP	168	ESP+packed-bed scrubber	147	79.8

[a]Metric equivalents in this table were converted from the calculated English unit values given in Table 4-11b.

TABLE 4-11b (ENGLISH). RECOVERY FURNACE MODELS: HCl CONTROL OPTION (PACKED-BED SCRUBBER)

Model No.	Black liquor firing rate, lb BLS/d	Equivalent pulp production rate		PM emission control device	Gas flow rate--ESP exit, acfm	Control option control device	HCl emissions, lb/d	Controlled HCl emissions, lb/d
		ADTUP/d	ADTBP/d					
RF-1	1.5 MM	500	420	ESP	198,000	ESP+packed-bed scrubber	180	93.0
RF-2	2.7 MM	900	750	ESP	357,000	ESP+packed-bed scrubber	324	167
RF-3	3.9 MM	1,300	1,100	ESP	515,000	ESP+packed-bed scrubber	468	242
RF-7	0.9 MM	300	250	ESP	119,000	ESP+packed-bed scrubber	108	58.8
RF-8	1.5 MM	500	420	ESP	198,000	ESP+packed-bed scrubber	180	98.0
RF-9	2.7 MM	900	750	ESP	357,000	ESP+packed-bed scrubber	324	176

TABLE 4-12a (METRIC). RECOVERY FURNACE MODELS: HCl CONTROL OPTION (PACKED-BED SCRUBBER AFTER LOW-ODOR CONVERSION)[a]

Model No.	Black liquor firing rate, kg BLS/d	Equivalent pulp production rate		Control device	Gas flow rate--ESP exit, m^3/sec actual	Control option control device	HCl emissions, kg/d	Controlled HCl emissions, kg/d
		ADMUP/d	ADMBP/d					
RF-7	0.4 MM	270	230	ESP	56.2	ESP+packed-bed scrubber	49.0	25.3
RF-8	0.7 MM	450	380	ESP	93.4	ESP+packed-bed scrubber	81.6	42.2
RF-9	1.2 MM	820	680	ESP	168	ESP+packed-bed scrubber	147	75.7

[a]Metric equivalents in this table were converted from the calculated English unit values given in Table 4-12b.

TABLE 4-12b (ENGLISH). RECOVERY FURNACE MODELS: HCl CONTROL OPTION (PACKED-BED SCRUBBER AFTER LOW-ODOR CONVERSION)

Model No.	Black liquor firing rate, lb BLS/d	Equivalent pulp production rate		Control device	Gas flow rate--ESP exit, acfm	Control option control device	HCl emissions, lb/d	Controlled HCl emissions, lb/d
		ADTUP/d	ADTBP/d					
RF-7	0.9 MM	300	250	ESP	119,000	ESP+packed-bed scrubber	108	55.8
RF-8	1.5 MM	500	420	ESP	198,000	ESP+packed-bed scrubber	180	93.0
RF-9	2.7 MM	900	750	ESP	357,000	ESP+packed-bed scrubber	324	167

TABLE 4-13a (METRIC). SMELT DISSOLVING TANK MODELS: PM CONTROL OPTIONS (0.10 KG/MG BLS)[a]

Model No.	Equivalent pulp production rate		Smelt flow rate, kg/d	PM emission control device	Control option PM emission control device	Inlet gas flow rate, m^3/sec actual	Moisture content, %	PM emissions, kg/Mg BLS	Controlled PM emissions, kg/Mg BLS
	ADMUP/d	ADMBP/d							
SDT-1	270	230	151,000	Wet scrubber	Wet scrubber	4.4	36	0.18	0.10
SDT-2	450	380	252,000	Wet scrubber	Wet scrubber	7.4	36	0.18	0.10
SDT-3	820	680	453,000	Wet scrubber	Wet scrubber	13.4	36	0.18	0.10
SDT-4	1,200	1,000	655,000	Wet scrubber	Wet scrubber	19.3	36	0.18	0.10
SDT-5	270	230	151,000	Mist eliminator	Wet scrubber	4.4	36	0.23	0.10
SDT-6	450	380	252,000	Mist eliminator	Wet scrubber	7.4	36	0.23	0.10
SDT-7	820	680	453,000	Mist eliminator	Wet scrubber	13.4	36	0.23	0.10

[a]Metric equivalents in this table were converted from the calculated English unit values given in Table 4-13b.

TABLE 4-13b (ENGLISH). SMELT DISSOLVING TANK MODELS: PM CONTROL OPTIONS (0.20 LB/TON BLS)

Model No.	Equivalent pulp production rate		Smelt flow rate, lb/d	PM emission control device	Control option PM emission control device	Inlet gas flow rate, acfm	Moisture content, %	PM emissions, lb/ton BLS	Controlled PM emissions, lb/ton BLS
	ADTUP/d	ADTBP/d							
SDT-1	300	250	333,000	Wet scrubber	Wet scrubber	9,400	36	0.37	0.20
SDT-2	500	420	555,000	Wet scrubber	Wet scrubber	15,700	36	0.37	0.20
SDT-3	900	750	999,000	Wet scrubber	Wet scrubber	28,300	36	0.37	0.20
SDT-4	1,300	1,100	1,443,000	Wet scrubber	Wet scrubber	40,900	36	0.37	0.20
SDT-5	300	250	333,000	Mist eliminator	Wet scrubber	9,400	36	0.46	0.20
SDT-6	500	420	555,000	Mist eliminator	Wet scrubber	15,700	36	0.46	0.20
SDT-7	900	750	999,000	Mist eliminator	Wet scrubber	28,300	36	0.46	0.20

wet scrubber or (2) installing a new wet scrubber. The control option would be applicable to both new and existing SDT's. The model SDT's analyzed for this control option are presented in Table 4-14.

4.2.3 BLO Unit Control Option

Two control options, (1) conversion of a DCE recovery furnace system to an NDCE recovery furnace and (2) incineration of BLO vent gases, were evaluated for controlling gaseous organic HAP emissions from air-sparging BLO units. Converting the DCE recovery furnace eliminates the BLO unit from the chemical recovery process. Thus, this control option results in a 100 percent reduction in emissions of HAP's, such as methanol, from BLO units. As discussed in Section 4.2.1.1, the overall methanol emission reduction for the recovery furnace system is approximately 99 percent. The model BLO units were not used to assess the impact of this control option; the economic and environmental impacts, such as cost for equipment removal and the associated emission reduction, were included in the impacts for the DCE recovery furnace models.

The second control option to reduce gaseous organic HAP emissions from BLO units is to incinerate the BLO emissions, most likely in a power boiler. This control option reduces methanol emissions from the BLO unit by 98 percent and from the DCE recovery furnace system by 49 percent. The model BLO units analyzed for this control option are presented in Table 4-15.

4.2.4 Lime Kiln Control Options

One PM emission control option was evaluated that would reduce PM emissions from existing lime kilns to the NSPS PM limit. The control option would involve replacing the existing scrubber with a new ESP. Because lime kilns with ESP's are already achieving the NSPS PM level, they are not included under this control option. The model lime kilns analyzed for this control option are presented in Table 4-16.

A second PM emission control option was evaluated that would reduce PM emissions from lime kilns to a more stringent level of 0.023 g/dscm (0.010 gr/dscf). The control option would involve

TABLE 4-14a (METRIC). SMELT DISSOLVING TANK MODELS: PM CONTROL OPTIONS (0.06 KG/MG BLS)[a]

Model No.	Equivalent pulp production rate		Smelt flow rate, kg/d	PM emission control device	Control option PM emission control device	Inlet gas flow rate, m^3/sec actual	Moisture content, %	PM emissions, kg/Mg BLS	Controlled PM emissions, kg/Mg BLS
	ADMUP/d	ADMBP/d							
SDT-1	270	230	151,000	Wet scrubber	Wet scrubber	4.4	36	0.18	0.06
SDT-2	450	380	252,000	Wet scrubber	Wet scrubber	7.4	36	0.18	0.06
SDT-3	820	680	453,000	Wet scrubber	Wet scrubber	13.4	36	0.18	0.06
SDT-4	1,200	1,000	655,000	Wet scrubber	Wet scrubber	19.3	36	0.18	0.06
SDT-5	270	230	151,000	Mist eliminator	Wet scrubber	4.4	36	0.23	0.06
SDT-6	450	380	252,000	Mist eliminator	Wet scrubber	7.4	36	0.23	0.06
SDT-7	820	680	453,000	Mist eliminator	Wet scrubber	13.4	36	0.23	0.06

[a]Metric equivalents in this table were converted from the calculated English unit values given in Table 4-14b.

TABLE 4-14b (ENGLISH). SMELT DISSOLVING TANK MODELS: PM CONTROL OPTIONS (0.12 LB/TON BLS)

Model No.	Equivalent pulp production rate		Smelt flow rate, lb/d	PM emission control device	Control option PM emission control device	Inlet gas flow rate, acfm	Moisture content, %	PM emissions, lb/ton BLS	Controlled PM emissions, lb/ton BLS
	ADTUP/d	ADTBP/d							
SDT-1	300	250	333,000	Wet scrubber	Wet scrubber	9,400	36	0.37	0.12
SDT-2	500	420	555,000	Wet scrubber	Wet scrubber	15,700	36	0.37	0.12
SDT-3	900	750	999,000	Wet scrubber	Wet scrubber	28,300	36	0.37	0.12
SDT-4	1,300	1,100	1,443,000	Wet scrubber	Wet scrubber	40,900	36	0.37	0.12
SDT-5	300	250	333,000	Mist eliminator	Wet scrubber	9,400	36	0.46	0.12
SDT-6	500	420	555,000	Mist eliminator	Wet scrubber	15,700	36	0.46	0.12
SDT-7	900	750	999,000	Mist eliminator	Wet scrubber	28,300	36	0.46	0.12

TABLE 4-15a (METRIC). BLACK LIQUOR OXIDATION UNIT MODELS: METHANOL CONTROL OPTION (INCINERATION)[a]

Model No.	Equivalent black liquor firing rate, kg BLS/d	Equivalent pulp production rate		Methanol control device	Control option control device	Methanol emissions, kg/d	Controlled methanol emissions, kg/d
		ADMUP/d	ADMBP/d				
BLO-1	0.4 MM	270	230	Uncontrolled	Incineration	70.8	1.4
BLO-2	0.7 MM	450	380	Uncontrolled	Incineration	118	2.4
BLO-3	1.2 MM	820	680	Uncontrolled	Incineration	212	4.2

[a]Metric equivalents in this table were converted from the calculated English unit values given in Table 4-15b.

TABLE 4-15b (ENGLISH). BLACK LIQUOR OXIDATION UNIT MODELS: METHANOL CONTROL OPTION (INCINERATION)

Model No.	Equivalent black liquor firing rate, lb BLS/d	Equivalent pulp production rate		Methanol control device	Control option methanol control device	Methanol emissions, lb/d	Controlled methanol emissions, lb/d
		ADTUP/d	ADTBP/d				
BLO-1	0.9 MM	300	250	Uncontrolled	Incineration	156	3.1
BLO-1	1.5 MM	500	420	Uncontrolled	Incineration	260	5.2
BLO-1	2.7 MM	900	750	Uncontrolled	Incineration	467	9.3

TABLE 4-16a (METRIC). LIME KILN MODELS: PM CONTROL OPTIONS (0.15 G/DSCM)[a]

Model No.	Equivalent pulp production rate, ADMP/d	CaO production rate, Mg/d	Length, m	Diameter, m	PM emission control device	Control option PM emission control device	PM emissions, g/dscm	Controlled PM emissions, g/dscm @ 10% O_2
LK-1	320	90	61	2.7	Venturi scrubber	ESP	0.27	0.15
LK-2	680	180	84	3.0	Venturi scrubber	ESP	0.27	0.15
LK-3	1,000	270	107	3.7	Venturi scrubber	ESP	0.27	0.15

[a]Metric equivalents in this table were converted from the calculated English unit values given in Table 4-16b.

TABLE 4-16b (ENGLISH). LIME KILN MODELS: PM CONTROL OPTIONS (0.067 GR/DSCF)

Model No.	Equivalent pulp production rate, ADTP/d	CaO production rate, ton/d	Length, ft	Diameter, ft	PM emission control device	Control option PM emission control device	PM emissions. gr/dscf	Controlled PM emissions, gr/dscf @ 10% O_2
LK-1	350	100	200	9.0	Venturi scrubber	ESP	0.12	0.067
LK-2	750	200	275	10	Venturi scrubber	ESP	0.12	0.067
LK-3	1,100	300	350	12	Venturi scrubber	ESP	0.12	0.067

(1) replacing the existing scrubber with an ESP, (2) upgrading the existing ESP, or (3) installing a new ESP. The more stringent PM control option is applicable to both new and existing lime kilns. The model lime kilns that were analyzed for this control option are presented in Table 4-17.

4.3 ENHANCED MONITORING OPTIONS

The practice of enhanced monitoring allows a facility to demonstrate continuous compliance with the emission limits established by the emission standard. The most direct means of monitoring compliance is the use of continuous emission monitors (CEM's) to measure the emissions of each pollutant on a continuous basis. In the event that CEM's for specific pollutants are not applicable because of cost and/or technology constraints, alternative approaches to ensure continuous compliance can be adopted. The best alternative approach is to use CEM's to monitor surrogate pollutants with emission profiles that closely match those of the pollutants of concern. Where CEM's are not applicable for surrogate pollutants, the next best option is to use process monitors to measure those process or add-on control device operating parameters that impact emissions of the pollutants of concern. In some cases, the very presence of specific processing equipment will ensure continuous compliance with the emission standard. Pollutant emissions can also be tested on a periodic basis (e.g., semiannually).

The following sections describe how the approach described above was used to develop the enhanced monitoring options for combustion sources in the pulp and paper industry. Table 4-18 summarizes the enhanced monitoring options for recovery furnaces, SDT's, BLO units, and lime kilns, based on the control options presented in Section 4.2. Enhanced monitoring options were not developed for control options other than those developed in Section 4.2; if a CEM is not applicable, facilities that choose to meet the emission limits through the application of other control options must develop an enhanced monitoring plan that demonstrates the ability of the selected parameter to gauge a change in emissions.

TABLE 4-17a (METRIC). LIME KILN MODELS: PM CONTROL OPTIONS (0.023 G/DSCM)[a]

Model No.	Equivalent pulp production rate, ADMP/d	CaO production rate, Mg/d	Length, m	Diameter, m	PM emission control device	Control option PM emission control device	PM emissions, g/dscm	Controlled PM emissions, g/dscm @ 10% O_2
LK-1	320	90	61	2.7	Venturi scrubber	ESP	0.27	0.023
LK-2	680	180	84	3.0	Venturi scrubber	ESP	0.27	0.023
LK-3	1,000	270	107	3.7	Venturi scrubber	ESP	0.27	0.023
LK-4	320	90	61	2.7	ESP	ESP	0.15	0.023
LK-5	680	180	84	3.0	ESP	ESP	0.15	0.023
LK-6	1,000	270	107	3.7	ESP	ESP	0.15	0.023

[a]Metric equivalents in this table were converted from the calculated English unit values given in Table 4-17b.

TABLE 4-17b (ENGLISH). LIME KILN MODELS: PM CONTROL OPTIONS (0.010 GR/DSCF)

Model No.	Equivalent pulp production rate, ADTP/d	CaO production rate, ton/d	Length, ft	Diameter, ft	PM emission control device	Control option control device	PM emissions, gr/dscf	Controlled PM emissions, gr/dscf @ 10% O_2
LK-1	350	100	200	9.0	Venturi scrubber	ESP	0.12	0.010
LK-2	750	200	275	10	Venturi scrubber	ESP	0.12	0.010
LK-3	1,100	300	350	12	Venturi scrubber	ESP	0.12	0.010
LK-4	350	100	200	9.0	ESP	ESP	0.067	0.010
LK-5	750	200	275	10	ESP	ESP	0.067	0.010
LK-6	1,100	300	350	12	ESP	ESP	0.067	0.010

TABLE 4-18. ENHANCED MONITORING OPTIONS

Emission source	Pollutants controlled	Control equipment	Enhanced monitoring options
Recovery furnace	PM/PM HAP's	ESP	Opacity monitor
			Monitoring of ESP operating parameters according to a monitoring plan prepared by the mill
		Wet scrubber[a]	Monitoring of scrubber operating parameters (i.e., pressure drop and scrubber liquid flow rate)
		Other[a]	Monitoring of selected parameter according to a monitoring plan prepared by the mill
		All	Periodic EPA Method 5, Method 29, or Method 17 tests
	Gaseous organic HAP's	NDCE recovery furnace with dry ESP system	Equipment in place
		Other[a]	Methanol CEM (e.g., FTIR)
	HCl	Packed-bed scrubber	HCl CEM
			Monitoring of scrubber operating parameters (i.e., scrubber liquid pH and flow rate)
		Other[a]	HCl CEM
			Monitoring of selected parameter according to a monitoring plan prepared by the mill
			Periodic EPA Method 26 tests
SDT	PM/PM HAP's	Wet scrubber	Monitoring of scrubber operating parameters (i.e., pressure drop and scrubber liquid flow rate)
		Other[a]	Monitoring of selected parameter according to a monitoring plan prepared by the mill
		All	Periodic EPA Method 5, Method 29, or Method 17 tests
BLO unit	Gaseous organic HAP's	NDCE recovery furnace with dry ESP system	Equipment in place
		Incineration	Equipment in place
			Incinerator temperature monitor
		Other[a]	Methanol CEM (e.g., FTIR)
Lime kiln	PM/PM HAP's	ESP	Opacity monitor
			Monitoring of ESP operating parameters according to a monitoring plan prepared by the mill
		Wet scrubber[a]	Monitoring of scrubber operating parameters (i.e., pressure drop and scrubber liquid flow rate)
		Other[a]	Monitoring of selected parameter according to a monitoring plan prepared by the mill
		All	Periodic EPA Method 5, Method 29, or Method 17 tests

[a]The emission source can comply with the applicable emission limit by another method or with existing equipment.

4.3.1 Recovery Furnace Enhanced Monitoring

Enhanced monitoring options that can be used to demonstrate compliance with recovery furnace emission limits for PM or PM HAP's, total gaseous organic HAP's, and HCl are presented in the following sections.

4.3.1.1 Enhanced Monitoring for PM or PM HAP's Controlled with an ESP. Because opacity is the surrogate measurement that best characterizes the level of recovery furnace PM emissions, installation of an opacity monitor after the ESP is one option being considered as a means of demonstrating compliance with a PM or PM HAP emission limit for recovery furnaces. For those recovery furnaces with a wet scrubber following the ESP, an opacity monitor must be located after the ESP but prior to the scrubber. Method 5, Method 29, or Method 17 compliance tests could be performed periodically as a substitute for an opacity monitor.

Another option being considered is for the facility to develop a monitoring plan that specifies ESP operating parameters to be monitored. Operating parameters for the ESP would be site-specific and would be based on the parameters measured during a three-run, EPA Method 5, Method 29, or Method 17 compliance test that showed the facility to be in compliance with the applicable PM or PM HAP emission limit. Under this option, operation outside the ranges of the ESP operating parameters would not represent noncompliance with the applicable emission limit but instead would require the facility to take corrective actions, if necessary, to return the ESP parameters to the levels established during the compliance test. The corrective action procedures would be documented in the facility's startup, shutdown, and malfunction plan.

4.3.1.2 Enhanced Monitoring for PM or PM HAP's Controlled with a Wet Scrubber. For those recovery furnaces that can comply with a PM or PM HAP emission limit with existing wet scrubbers, the use of an opacity monitor to demonstrate compliance with the PM or PM HAP emission limit may be inappropriate. The exhaust from the recovery furnace wet scrubber will have a high moisture

content and will interfere with the readings from an opacity monitor. Monitoring scrubber operating parameters (i.e., pressure drop and scrubber liquid flow rate) is an alternative enhanced monitoring option for showing compliance with a PM or PM HAP emission limit for recovery furnaces. The pressure drop and liquid flow rate are indirect measurements of the performance of the scrubber. Pressure drop and scrubber liquid flow rate levels would be site-specific and would be based on the operating parameters measured during a three-run, EPA Method 5, Method 29, or Method 17 compliance test that showed the facility to be in compliance with the applicable PM or PM HAP emission limit. Method 5, Method 29, or Method 17 compliance tests could also be performed periodically as a substitute for monitoring scrubber operating parameters.

4.3.1.3 Enhanced Monitoring for Gaseous Organic HAP's. Control of gaseous organic HAP emissions from recovery furnaces can be achieved by using NDCE recovery furnaces equipped with dry ESP systems. Therefore, enhanced monitoring for recovery furnace gaseous organic HAP emissions can be achieved simply by confirming that the recovery furnace is an NDCE recovery furnace with a dry ESP system. If the recovery furnace is a DCE recovery furnace or an NDCE recovery furnace equipped with a wet ESP system, the facility could measure methanol emissions with a methanol CEM (e.g., a fourier transform infrared [FTIR] spectroscopy monitoring system).

4.3.1.4 Enhanced Monitoring for HCl. Hydrochloric acid emissions can be measured directly using an HCl CEM. An HCl CEM could be installed after the packed-bed scrubber to demonstrate continuous compliance with an HCl emission standard. An HCl CEM could also be used after the ESP for those recovery furnaces that could comply with an HCl emission limit without a packed-bed scrubber. The feasibility of using HCl CEM's to demonstrate compliance with an HCl standard has not been determined. The low HCl concentrations and high moisture content associated with the recovery furnace flue gas may make the use of HCl CEM's more

difficult. However, additional information is needed before an HCl CEM can be definitively ruled out for recovery furnaces.

Because HCl emissions can be controlled with a packed-bed scrubber, monitoring scrubber operating parameters is another monitoring option being considered for those recovery furnaces that comply with an HCl emission limit using a packed-bed scrubber. The scrubber operating parameters to be monitored are the scrubber liquid pH and scrubber liquid flow rate. Scrubber liquid flow rate and pH levels would be site-specific and would be based on the operating parameters measured during a three-run, EPA Method 26 HCl compliance test that showed the facility to be in compliance with an HCl emission limit.

Alternative enhanced monitoring options are also available to demonstrate compliance for those recovery furnaces that could comply with an HCl emission limit without a packed-bed scrubber. One option would require the facility to develop a monitoring plan that specifies operating parameters to be monitored. The operating parameters would be site-specific and would be based on the parameters measured during a three-run, EPA Method 26 HCl compliance test that showed the facility to be in compliance with an HCl emission limit. Under this option, operating outside the ranges of the operating parameters would not represent noncompliance with the applicable emission limit but instead would require the facility to take corrective actions, if necessary, to return the parameters to the levels established during the compliance test. The corrective action procedures would be documented in the facility's startup, shutdown, and malfunction plan. A second option would require periodic Method 26 HCl compliance tests to demonstrate compliance.

4.3.2 Smelt Dissolving Tank Enhanced Monitoring

This section presents the enhanced monitoring options that can be used to demonstrate compliance with an SDT emission limit for PM or PM HAP's.

4.3.2.1 Enhanced Monitoring for PM or PM HAP's Controlled with a Wet Scrubber. Because the exhaust from the SDT wet scrubber will have a high moisture content and will interfere

with the readings from an opacity monitor, the use of an opacity monitor to demonstrate compliance with a PM or PM HAP emission limit for SDT's may be inappropriate. Monitoring scrubber operating parameters (i.e., pressure drop and scrubber liquid flow rate) is an alternative enhanced monitoring option for showing compliance with a PM or PM HAP emission limit for SDT's. The pressure drop and liquid flow rate are indirect measurements of the performance of the scrubber. Pressure drop and scrubber liquid flow rate levels would be site-specific and would be based on the operating parameters measured during a three-run, EPA Method 5, Method 29, or Method 17 compliance test that showed the facility to be in compliance with the applicable PM or PM HAP emission limit. Method 5, Method 29, or Method 17 compliance tests could also be performed periodically as a substitute for monitoring scrubber operating parameters.

4.3.3 Black Liquor Oxidation Unit Enhanced Monitoring

This section presents the enhanced monitoring options that can be used to demonstrate compliance with a total gaseous organic HAP emission limit for DCE recovery furnace systems (which include the BLO unit).

One control option presented for the BLO unit involves the removal of this piece of equipment from the chemical recovery process by converting a DCE recovery furnace to an NDCE recovery furnace equipped with a dry ESP system. Demonstrating that this conversion has been completed assures compliance with the applicable total gaseous organic HAP emission limit.

A second control option involves incineration of the BLO emissions. Enhanced monitoring for BLO incineration could be achieved simply by affirming that the BLO control equipment is in place. Another enhanced monitoring option would be for the facility to monitor the temperature of the power boiler or other incineration device.

4.3.4 Lime Kiln Enhanced Monitoring

Enhanced monitoring options that can be used to demonstrate compliance with a lime kiln emission limit for PM or PM HAP's are presented in the following sections.

4.3.4.1 Enhanced Monitoring for PM or PM HAP's Controlled with an ESP. Because opacity is the surrogate measurement that best characterizes the level of lime kiln PM emissions, installation of an opacity monitor after the ESP is one option being considered as a means of demonstrating compliance with a PM or PM HAP emission limit for lime kilns controlled with ESP's. For those lime kilns with a wet scrubber following the ESP, an opacity monitor must be located after the ESP but prior to the scrubber. Method 5, Method 29, or Method 17 compliance tests could be performed periodically as a substitute for an opacity monitor.

Another option being considered is for the facility to develop a monitoring plan that specifies ESP parameters to be monitored. Operating parameters for the ESP would be site-specific and would be based on the parameters measured during a three-run, EPA Method 5, Method 29, or Method 17 compliance test that showed the facility to be in compliance with the applicable PM or PM HAP emission limit. Under this option, operation outside the ranges of the ESP operating parameters would not represent noncompliance with the applicable emission limit but instead would require the facility to take corrective actions, if necessary, to return the ESP parameters to the levels established during the compliance test. The corrective action procedures would be documented in the facility's startup, shutdown, and malfunction plan.

4.3.4.2 Enhanced Monitoring for PM or PM HAP's Controlled with a Wet Scrubber. For those lime kilns that can comply with a lime kiln PM or PM HAP emission limit with existing wet scrubbers, the use of an opacity monitor to demonstrate compliance with a PM or PM HAP emission limit may be inappropriate. The exhaust from the lime kiln wet scrubber will have a high moisture content and will interfere with the readings from an opacity monitor. Monitoring scrubber operating parameters (i.e., pressure drop and scrubber liquid flow rate) is an alternative enhanced monitoring option for showing compliance with the applicable PM or PM HAP emission limit for lime kilns.

The pressure drop and liquid flow rate are indirect measurements of the performance of the scrubber. Pressure drop and scrubbing liquid flow rate levels would be site-specific and would be based on the operating parameters measured during a three-run, EPA Method 5, Method 29, or Method 17 compliance test that showed the facility to be in compliance with the applicable PM or PM HAP emission limit. Method 5, Method 29, or Method 17 compliance tests could also be performed periodically as a substitute for monitoring scrubber operating parameters.

4.4 REFERENCES FOR CHAPTER 4

1. Memorandum from Nicholson, R., MRI, to Telander, J., EPA/MICG. June 13, 1996. Addendum to Summary of Responses to the 1992 NCASI MACT Survey.

2. Memorandum from Soltis, V., MRI, to the project file. April 3, 1995. Kraft and Soda Pulp Mill Combustion Sources Data Base.

3. Someshwar, A. Compilation of "Air Toxic" Emission Data for Boilers, Pulp Mills, and Bleach Plants. National Council of the Paper Industry for Air and Stream Improvement, Inc., New York. Technical Bulletin No. 650. June 1993. 128 p.

4. Green, R. and G. Hough, (eds.). Chemical Recovery in the Alkaline Pulping Process. 3rd Edition. Prepared by the Alkaline Pulping Committee of the Pulp Manufacture Division. Atlanta, GA, TAPPI Press. 1992. 196 p.

5. Telecon. Nicholson, R., MRI, with Bringman, L, Environmental Elements Corp. November 30, 1994. Information about the particulate return systems associated with kraft recovery furnace ESP's.

6. Memorandum from Soltis, V., MRI, to the project file. October 20, 1994. Calculation of Recovery Furnace Stack Gas Flow Rate.

7. Proposed Standards of Performance for Kraft Pulp Mills. In: Standards Support and Environmental Impact Statement. Volume 1. U. S. Environmental Protection Agency. Research Triangle Park, NC. Publication No. EPA-450/2-76-014a. September 1976.

8. U. S. Environmental Protection Agency. New Source Performance Standards for Kraft Pulp Mills. 43 FR 7568. Washington, DC. U.S. Government Printing Office. February 23, 1978.

9. U. S. Environmental Protection Agency. New Source Performance Standards for Kraft Pulp Mills. 51 FR 18538. Washington, DC. U.S. Government Printing Office. May 20, 1986.

10. Memorandum from Holloway, T., MRI, to the project files. June 14, 1996. Summary of PM and HAP Metals Data.

11. Memorandum from Soltis, V., Nicholson, R., and Holloway, T., MRI, to Telander, J., EPA/MICG. July 29, 1994. Summary of Responses to the NCASI "MACT" Survey--Kraft and Soda Pulp Mills (Data Base Summary Memo).

12. Roy F. Weston, Inc. Texas Emissions Speciation Study Emission Test Results: Champion International Corp.--Sheldon, Texas. Prepared for Texas Paper Industry Environmental Committee. Report No. 06848-001-001. January 1993. Volume 3.

13. Roy F. Weston, Inc. Texas Emissions Speciation Study Emission Test Results: Simpson-Pasadena--Pasadena, Texas. Prepared for Texas Paper Industry Environmental Committee. Report No. 06848-001-001. January 1993. Volume 5.

14. Environmental Pollution Control, Pulp and Paper Industry, Part I, Air. U. S. Environmental Protection Agency. Cincinnati, OH. Publication No. EPA-625/7-76-001. October 1976.

15. Roy F. Weston, Inc. Emissions Testing of Combustion Processes in a Pulp and Paper Facility: Champion International Corp.--Roanoke Rapids, NC. Prepared for U. S. Environmental Protection Agency--Emission Measurement Branch. Research Triangle Park, NC. EMB Report No. 92-KPM-27. October 1992. Volume I.

16. Telecon. Soltis, V., MRI, with Bruno, J., AirPol, Inc. July 2, 1993. Information about scrubbers used to control HCl emissions.

17. Telecon. Soltis, V., MRI, with Sanders, D., Andersen 2000, Inc. August 11, 1993. Cost and efficiency information about HCl control for recovery furnaces.

5.0 MODEL PROCESS UNIT ENVIRONMENTAL AND ENERGY IMPACTS

This chapter discusses the environmental and energy impacts of controlling HAP emissions from new and existing combustion sources at kraft and soda pulp and paper mills. The environmental and energy impacts of each control option discussed in Chapter 4 are presented for the individual model combustion process units. The total nationwide environmental and energy impacts associated with each control option are presented in a separate memorandum.[1]

Section 5.1 of this chapter discusses the general approach used to determine the environmental and energy impacts associated with each control option. Sections 5.2 through 5.5 present the environmental and energy impacts of each control option for model process units representing recovery furnaces, BLO units, SDT's, and lime kilns. Section 5.6 contains the references cited in this chapter.

5.1 GENERAL APPROACH

This section introduces the types of environmental and energy impacts associated with the control options and discusses how these impacts were determined for each of the model process units. The types of impacts discussed include air pollution, energy consumption, water pollution, solid waste disposal, and other impacts (i.e., noise, visual, odor, and irreversible and irretrievable commitment of resources). Impacts were calculated for each model process unit on an annual basis. In calculating the annual impacts, all process units were assumed to operate 24 hr/d for 351 d/yr, which is equivalent to 8,424 operating hr/yr. This operating time accounts for 14 days of scheduled shutdown annually for maintenance and repair.

The procedure for estimating air pollution impacts is described in the following section. The general approaches used to develop energy impacts, water pollution impacts, solid waste disposal impacts, and other impacts are provided in Sections 5.1.2, 5.1.3, 5.1.4, and 5.1.5, respectively.

5.1.1 Air Pollution Impacts

This section presents the methodology used to determine both primary and secondary air impacts.

5.1.1.1 Primary Emissions. For this impact analysis, primary air impacts include the reduction of emissions directly attributable to the control option (i.e., the reduction of emissions due to the use of APCD's or process modifications). Primary emissions of PM, PM HAP's, gaseous organic HAP's, HCl, and SO_2 were estimated based on model concentrations and emission factors at baseline and at control levels. Gaseous organic HAP's include acetaldehyde, benzene, formaldehyde, methanol, methyl ethyl ketone, methyl isobutyl ketone, phenol, toluene, and xylenes.

As discussed later in this chapter, PM is used as a surrogate measure for PM HAP's, i.e, trace metals, such as antimony (Sb), arsenic (As), beryllium (Be), cadmium (Cd), chromium (Cr), cobalt (Co), lead (Pb), manganese (Mn), mercury (Hg), nickel (Ni), and selenium (Se). The primary emission estimates for PM were derived for each model recovery furnace and lime kiln by multiplying the model PM concentration by the model gas flow rate corrected to dry conditions at standard temperature and by 351 operating d/yr.

Primary emission estimates of gaseous organic HAP's, HCl, and SO_2 from model recovery furnaces and PM from model SDT's were determined by multiplying their respective model emission factors by the model BLS firing rates and by 351 operating d/yr.

Primary emission estimates of PM HAP's were determined as a percentage of PM for each model recovery furnace, SDT, and lime kiln; the percentages were derived based on a comparison of PM and PM HAP emissions for kraft and soda recovery furnaces, SDT's, and lime kilns. The percentages of PM were estimated to be

0.2 percent for recovery furnaces, 0.06 percent for SDT's, and 1.4 percent for lime kilns.[2]

The control level HCl emission factor was derived from an outlet HCl emission concentration of 5 ppmv guaranteed by packed-bed scrubber manufacturers for inlet HCl concentrations less than 500 ppmv.[3,4] However, the actual outlet HCl emission level achieved at a particular mill will be site-specific. The baseline HCl emission estimates (and percent HCl reduction) in this impact analysis may be underestimated because the baseline HCl emission factor includes emissions data from recovery furnaces with HCl emissions at or below 5 ppmv.

The SO_2 emission reduction obtained with the packed-bed scrubber was estimated for each model recovery furnace by applying a percent SO_2 reduction to average uncontrolled SO_2 emission estimates. Uncontrolled SO_2 emission factors for recovery furnaces have already been developed and are presented in the Air Pollution Engineering Manual.[5] As shown in the manual, the average uncontrolled SO_2 emission factor for NDCE recovery furnaces is 2.1 kilograms per air-dried megagram of pulp (kg/ADMP) (4.2 pounds per air-dried ton of pulp [lb/ADTP]).[5] Using a conversion factor of 1,700 kg BLS/ADMP (3,400 lb BLS/ADTP) (the average for both bleached and unbleached pulp mills), the average uncontrolled SO_2 emission factor for NDCE recovery furnaces is equivalent to 1.24×10^{-3} kg/kg BLS (1.24×10^{-3} lb/lb BLS).[6] The average uncontrolled SO_2 emission factor for DCE recovery furnaces is 1.8 kg/ADMP (3.5 lb/ADTP), which is equivalent to 1.03×10^{-3} kg/kg BLS (1.03×10^{-3} lb/lb BLS).[5] Based on information from individual mills, at least 50 percent SO_2 control was assumed to be achievable with packed-bed scrubbers.[7] The control level SO_2 emission estimates were determined using the uncontrolled emission factors and the 50 percent SO_2 control.

To estimate the incremental reduction in emissions for each primary pollutant (i.e., PM, PM HAP's, gaseous organic HAP's, HCl, and SO_2), the control level emission estimate for each

primary pollutant was compared to its corresponding baseline emission estimate.

5.1.1.2 Secondary Emissions. Secondary air impacts include the indirect or induced impacts resulting from implementing a control option. These indirect or induced impacts include the following:

(1) changes in power boiler emissions of criteria pollutants such as PM, SO_2, nitrogen oxides (NO_x), and carbon monoxide (CO) resulting from the generation of energy required to operate APCD's or other equipment included in the control options; and

(2) changes in power boiler emissions of SO_2 resulting from the incineration of TRS compounds present in the BLO vent gases routed to the power boiler to control gaseous organic HAP emissions.

The power boiler secondary emissions of PM, SO_2, NO_x, and CO resulting from the generation of electricity to operate APCD's or other equipment were estimated based on emission factors related to electricity usage. The emission factors used are 0.23 kilograms of PM per megawatt-hour (kg PM/MWh) (0.15 pounds of PM per million Btu [lb PM/MM Btu]), 1.1 kg SO_2/MWh (0.73 lb SO_2/MM Btu), 0.45 kg NO_x/MWh (0.29 lb NO_x/MM Btu), and 0.85 kg CO/MWh (0.55 lb CO/MM Btu).[8] The secondary emission estimates were calculated as a product of these emission factors and the model process unit electricity impacts discussed in the next section.

Secondary SO_2 emissions can be generated under the BLO vent gas control option when TRS compounds present in the vent gases are incinerated in a power boiler or other incineration device. All of the TRS from the BLO unit is assumed to be routed to a power boiler or other incineration device; none is "lost" as condensate along the way. In deriving the chemical equation for the conversion of TRS to SO_2, the TRS was assumed to be in the form of H_2S. During combustion, the H_2S would be combined with O_2 to produce SO_2 and H_2O. The chemical equation is as follows:

$$H_2S + \frac{3}{2}O_2 \rightarrow SO_2 + H_2O$$

As shown in this equation, one mole of SO_2 is formed for each mole of H_2S combusted. This molar ratio was used in the following equation to estimate the mass ratio of SO_2 to TRS:

$$(1\ kg\text{-}mole\ SO_2\ formed/1\ kg\text{-}mole\ H_2S\ combusted) \times$$
$$(1\ kg\text{-}mole\ H_2S/34.08\ kg\ H_2S) \times (64.06\ kg\ SO_2/kg\text{-}mole\ SO_2)$$
$$\approx 1.88\ kg\ SO_2/kg\ H_2S$$

The English unit equivalents for this equation are the same as the metric unit values presented in the equation. To determine the secondary SO_2 emission estimates, the ratio of SO_2 to TRS was multiplied by the estimate of TRS emissions incinerated by the BLO control system. These secondary SO_2 emission estimates represent worst-case estimates because some control of SO_2 emissions is likely to be achieved through the application of wet scrubbers on incineration devices such as power boilers and lime kilns that could be used to incinerate BLO vent gases.

To estimate the change in emissions for each secondary pollutant (i.e., PM, SO_2, NO_x, and CO), the control level emission estimate for each secondary pollutant was compared to its corresponding baseline emission estimate.

5.1.2 Energy Impacts

The energy impacts of the combustion source control options include changes in electricity and steam requirements. The increases in electricity requirements from the operation of APCD's were calculated using electricity cost equations from the EPA's Office of Air Quality Planning and Standards (OAQPS) Control Cost Manual.[9] These electricity requirements are divided into fan, pump, and operating electricity requirements and were calculated assuming 8,424 operating hr/yr. The electricity requirements to operate a BLO vent gas control system were calculated based on information from a kraft pulp mill.[10] The methods used to calculate the BLO control system energy impacts are described in Section 5.3.2.

The fan electricity requirement (applicable to ESP's and scrubbers) is equal to a numerical factor (0.00018) times the product of the gas flow rate, pressure drop, and operating hr/yr.[9] The gas flow rate varies with each model process unit. The pressure drop is based on information from individual mills.[6] Although the pressure drop is not the sole parameter that determines PM collection efficiency for lime kiln scrubbers, for the purposes of estimating impacts, the pressure drop was used as an indicator of PM collection efficiency for lime kiln scrubbers. (Note: A different scrubber design, rather than a higher pressure drop, was used to improve the PM collection efficiency for SDT scrubbers.)

The gas flow rate and pressure drop do not change for ESP's relative to current operation when the ESP's are upgraded or replaced to improve PM collection. Therefore, the fan electricity requirements for ESP's do not change relative to current operation. However, if a scrubber is added after an ESP, the gas flow rate would be reduced, thereby reducing the fan electricity requirements for the ESP.

The pump electricity requirement (applicable to packed-bed scrubbers) is equal to a numerical factor (0.000188) times the product of the liquid flow rate, amount of head pressure, and operating hr/yr divided by the pump-motor efficiency.[9] The liquid flow rate varies with each model process unit. A head pressure of 18 m (60 ft) and a pump-motor efficiency of 70 percent were assumed.

The operating electricity requirement (applicable to ESP's) is equal to a numerical factor (0.00194) times the product of the ESP plate area and operating hr/yr.[9] The ESP plate area is calculated as a product of the exhaust gas flow rate and the ESP SCA. The gas flow rate varies with each model process unit, and the SCA is based on information from individual mills.[6] Although the SCA is not the sole parameter that determines PM collection efficiency for ESP's, for the purposes of estimating impacts, the SCA was used as an indicator of PM collection efficiency.

Steam energy savings result from the increased steam flow that occurs with the gaseous organic HAP control option of converting the DCE recovery furnace system to an NDCE recovery furnace.[11] With NDCE recovery furnaces, more heat is available for steam production than with DCE recovery furnaces because combustion gases are not used to concentrate black liquor. The increased steam production resulting from converting the DCE recovery furnace system to the NDCE design reduces the power boiler steam production requirements. A fuel savings can be realized from this steam energy savings because the amount of fuel required for steam production in the power boiler decreases. The steam energy savings are partially offset by low-pressure steam requirements for the concentrator, which is a part of the NDCE recovery furnace system. The low-pressure steam would be used to concentrate the black liquor. These concentrator steam requirements were determined based on the low-pressure steam flow rate and the change in enthalpy between the entering feedwater and the steam leaving the superheater.[12]

Another energy savings from converting the DCE recovery furnace system to an NDCE recovery furnace is the electricity saved by removing the BLO unit. The BLO operating energy savings were determined by dividing the BLO electricity cost savings presented in Chapter 6 by the unit cost of electricity in the U. S. EPA Handbook: Control Technologies for Hazardous Air Pollutants.[13]

The exact methods used to determine the energy impacts of converting DCE recovery furnace systems to NDCE recovery furnaces are presented in Section 5.2.3.2 and corresponding tables.

5.1.3 Water Pollution Impacts

Some of the control options for recovery furnaces, SDT's, and lime kilns may have a significant impact on the amount of wastewater generated, treated, and disposed. If lime kilns reduce PM emissions by replacing the existing scrubber with a new ESP, the existing scrubber discharge would be completely eliminated. On the other hand, if recovery furnaces reduce HCl emissions by adding a packed-bed scrubber, a new scrubber

discharge would be generated. If SDT's reduce PM emissions by replacing the existing PM control device with a new scrubber, the wastewater discharge may be increased because of the increased amount of PM in the scrubber liquid. The additional discharge from the recovery furnace and SDT scrubbers may simply be recycled and used elsewhere in the plant. Alternatively, the scrubber discharge may require treatment and disposal. The wastewater disposal impacts and the exact methods used to estimate them are discussed for each of these control options in their respective sections.

5.1.4 Solid Waste Disposal Impacts

The recovery furnace and lime kiln PM control options include ESP's for PM control and, therefore, will generate a PM catch. The PM catch from the ESP may simply be recycled back into the process. Potential solid waste disposal impacts are discussed for each of these control options in their respective sections.

5.1.5 Other Impacts

Other impacts considered for the control options include noise impacts, visual impacts, odor impacts, and irreversible and irretrievable commitment of resources.

Noise impacts may occur as a result of changing the fan size to accommodate the new or upgraded APCD's. Visual impacts may occur as a result of reducing the amount of PM and moisture (i.e., steam) emitted from the stack. Odor impacts may occur as a result of reducing malodorous TRS emissions. Because the odor impacts were based on available TRS emissions data, these impacts were easily estimated quantitatively. However, little information was available to estimate noise, visual, and other impacts quantitatively; therefore, only a qualitative assessment of these impacts was done for each control option, where applicable.

Odor impacts were determined for the gaseous organic HAP control options by comparing TRS emissions at baseline and at control levels for those control options. The TRS emissions for DCE and NDCE recovery furnaces were derived from TRS emissions

data (in parts per million [ppm] as H_2S) provided by individual pulp and paper mills.[6] These TRS concentrations were converted to annual TRS emissions using the ideal gas law and the model BLS firing rates, assuming 351 operating d/yr. Total reduced sulfur emissions from BLO units are generally emitted in the range of 0.04 to 0.06 kg/ADMP (0.08 to 0.13 lb/ADTP).[14] A mid-range value of 0.05 kg/ADMP (0.10 lb/ADTP) was used to calculate the annual TRS emissions from model BLO units based on the model pulp production rates, assuming 351 operating d/yr. The mid-range value for BLO units was used because the only data available for BLO TRS emissions were the bounds of the range.

5.2 RECOVERY FURNACE CONTROL OPTIONS

The following sections discuss the model environmental and energy impacts of implementing control options designed to reduce emissions of PM (as a surrogate for PM HAP's), methanol (as a surrogate for gaseous organic HAP's), and HCl. These control options include PM controls (Section 5.2.1), wet to dry ESP system conversion (Section 5.2.2), conversion of a DCE recovery furnace system to an NDCE recovery furnace (Section 5.2.3), and addition of a packed-bed scrubber (Section 5.2.4). Table 5-1 presents the model recovery furnace sizes and operating parameters.

5.2.1 PM Controls

Two PM control options were evaluated for model NDCE recovery furnaces RF-1 through RF-6 and model DCE recovery furnaces RF-7 through RF-9. The control options apply to both new and existing recovery furnaces and are described below.

One PM control option that was evaluated would reduce PM emissions from existing recovery furnaces to the NSPS level of 0.10 g/dscm (0.044 gr/dscf). The PM control option evaluated would involve replacing or upgrading the recovery furnace ESP.

A second PM control option that was evaluated would reduce PM emissions from existing recovery furnaces to 0.034 g/dscm (0.015 gr/dscf). This more stringent PM control option would involve replacing or upgrading the recovery furnace ESP and adding a packed-bed scrubber. The second PM control option also

applies to new recovery furnaces; the option could be used to evaluate the impact on new sources subject to a more stringent standard (0.034 g/dscm [0.015 gr/dscf]) than the current NSPS.

The PM control impacts for model NDCE recovery furnaces with dry ESP systems (i.e., RF-1 through RF-3) are assumed to be identical to the PM control impacts for model NDCE recovery furnaces with wet ESP systems (i.e., RF-4 through RF-6) because PM emissions are not affected by whether or not black liquor is used in the ESP system. The environmental and energy impacts associated with the PM control options are presented for the model NDCE and DCE recovery furnaces in the following sections.

5.2.1.1 Air Pollution Impacts. This section presents the primary and secondary air impacts resulting from implementing PM controls.

5.2.1.1.1 Primary emissions. Although emission test data from recovery furnace ESP's on PM HAP performance are limited, data collected from other combustion sources on the relative performance of APCD's for PM and PM HAP's indicate that systems that achieve the greatest PM removal also provide the best performance for the HAP portion of the PM.[15] Therefore, PM performance can be used as a surrogate for PM HAP's. Because emission test data from recovery furnace ESP's indicate that PM emissions are reduced with PM controls, PM HAP emissions would also be reduced.[6] As stated in Section 5.1.1.1, PM HAP emissions from recovery furnaces were estimated to be 0.2 percent of PM based on a comparison of PM and PM HAP emission data for recovery furnaces.

Tables 5-1 and 5-2 present the operating parameters for model NDCE and DCE recovery furnaces, respectively. Tables 5-3 and 5-4 present PM concentrations and PM HAP emission factors for model NDCE and DCE recovery furnaces, respectively. Table 5-5 presents the annual PM and PM HAP emission estimates. Figures 5-1 and 5-2 illustrate the annual PM emission estimates for model NDCE and DCE recovery furnaces, respectively.

The baseline PM concentration for model NDCE recovery furnaces RF-1a through RF-6a is 0.27 g/dscm (0.12 gr/dscf).[2] By

controlling PM emissions from baseline to the NSPS level (0.10 g/dscm [0.044 gr/dscf]), PM emissions would be reduced by 63 percent. On an annual basis, PM emissions would be reduced by about 227 to 593 Mg/yr (250 to 654 ton/yr); PM HAP emissions would be reduced by about 0.5 to 1.2 Mg/yr (0.5 to 1.3 ton/yr). By controlling PM emissions from baseline to a more stringent PM control level of 0.034 g/dscm (0.015 gr/dscf), PM emissions would be reduced by 88 percent. On an annual basis, PM emissions would be reduced by about 313 to 818 Mg/yr (345 to 902 ton/yr); PM HAP emissions would be reduced by about 0.6 to 1.6 Mg/yr (0.7 to 1.8 ton/yr).

The baseline PM concentration for model NDCE recovery furnaces RF-1b through RF-6b is the NSPS PM level (0.10 g/dscm [0.044 gr/dscf]). By controlling PM emissions from the NSPS baseline to 0.034 g/dscm (0.015 gr/dscf), PM emissions would be reduced by 66 percent. On an annual basis, PM emissions would be reduced by about 86 to 225 Mg/yr (95 to 248 ton/yr); PM HAP emissions would be reduced by about 0.2 to 0.4 Mg/yr (0.2 to 0.5 ton/yr).

The baseline PM concentration for model DCE recovery furnaces RF-7a through RF-9a is 0.18 g/dscm (0.08 gr/dscf).[2] By controlling PM emissions from baseline to the NSPS level, PM emissions would be reduced by 45 percent. On an annual basis, PM emissions would be reduced by about 65 to 194 Mg/yr (71 to 214 ton/yr); PM HAP emissions would be reduced by about 0.1 to 0.4 Mg/yr (0.1 to 0.4 ton/yr). By controlling PM emissions from baseline to a more stringent control level of 0.034 g/dscm (0.015 gr/dscf), PM emissions would be reduced by 81 percent. On an annual basis, PM emissions would be reduced by about 117 to 350 Mg/yr (129 to 386 ton/yr); PM HAP emissions would be reduced by about 0.2 to 0.7 Mg/yr (0.3 to 0.8 ton/yr). The PM emission reductions at both PM control levels are lower for model DCE recovery furnaces than for model NDCE recovery furnaces because the baseline PM concentration for model DCE recovery furnaces is lower than that for model NDCE recovery furnaces.

The baseline PM concentration for model DCE recovery furnaces RF-7b through RF-9b is the NSPS PM level (0.10 g/dscm [0.044 gr/dscf]). By controlling PM emissions from the NSPS baseline to 0.034 g/dscm (0.015 gr/dscf), PM emissions would be reduced by 66 percent. On an annual basis, PM emissions would be reduced by about 52 to 156 Mg/yr (57 to 172 ton/yr); PM HAP emissions would be reduced by about 0.1 to 0.3 Mg/yr (0.1 to 0.3 ton/yr).

5.2.1.1.2 Secondary emissions. Secondary emissions of PM, SO_2, NO_x, and CO generated under the PM control options were estimated for model NDCE and DCE recovery furnaces. Table 5-6 presents the annual secondary emission estimates. As shown in the table, the secondary emissions generated under the PM control options are small. The increases in secondary PM emissions are especially insignificant compared to the reductions in primary PM emissions. The more stringent PM control option includes both an ESP upgrade and the addition of a packed-bed scrubber. The increases in secondary SO_2 emissions under the more stringent PM control option are insignificant compared with the large reductions in primary SO_2 emissions that result from adding the packed-bed scrubber. The reductions in primary SO_2 emissions using a packed-bed scrubber are presented in Table 5-12 and discussed in Section 5.2.4.1.1.

5.2.1.2 Energy Impacts. The energy requirements of the PM control options were estimated for model NDCE and DCE recovery furnaces. The energy requirements were estimated based on model SCA values for the ESP. Tables 5-1 and 5-2 present the baseline and control level SCA values for model NDCE and DCE recovery furnaces, respectively. Table 5-7 presents the annual energy requirements. As shown in the table, the annual energy requirements of the PM control options are small.

5.2.1.3 Water Pollution Impacts. Because the ESP operates on a dry basis, no water pollution impacts are associated with an ESP replacement or upgrade used to improve PM collection. Section 5.2.4.3 presents a discussion of the water pollution

impacts resulting from adding a packed-bed scrubber (which is included in the more stringent PM control option).

5.2.1.4 Solid Waste Disposal Impacts. As mentioned in Chapter 2, the PM catch from the recovery furnace ESP is primarily Na_2SO_4 (i.e., saltcake) and Na_2CO_3. These chemicals are subsequently added to the concentrated black liquor in a mix tank (i.e., recycled back into the process) in order to conserve chemicals. Approximately 95 percent of the Na_2SO_4 is recovered. The Na_2SO_4 recycled back to the recovery furnace is reduced to Na_2S, a cooking liquor chemical, in the reducing zone of the furnace. Reprocessing of the Na_2S and Na_2CO_3 into cooking liquor continues in the SDT. The recovery process was assumed to have sufficient capacity to absorb the additional PM resulting from the ESP replacement or upgrade. Thus, no solid waste disposal impacts are expected with an ESP upgrade or replacement. Also, no solid waste disposal impacts are expected with the addition of a packed-bed scrubber (which is included in the more stringent PM control option).

5.2.1.5 Other Impacts. Limited information was available to estimate quantitatively impacts such as noise, visual, odor, and irreversible and irretrievable commitment of resources. Beneficial visual impacts are expected to result from the reduced PM emissions coming out of the recovery furnace stack. However, there also may be negative visual impacts under the more stringent PM control option. The more stringent PM control option would involve adding a packed-bed scrubber, which would add to the moisture (i.e., steam) coming out of the stack. Adding a packed-bed scrubber also would require additional equipment (i.e., larger fans to overcome pressure drops and pumps) that would increase noise levels. However, these incremental noise increases are expected to be small compared to the typical background noise levels at pulp and paper mills. The other impacts, if any, are expected to be minimal as a result of implementing the recovery furnace PM control options.

5.2.2 Wet to Dry ESP System Conversion

Two control options were evaluated for reducing emissions of gaseous organic HAP's such as methanol from existing NDCE recovery furnaces. These control options are (1) converting an ESP system that uses unoxidized black liquor or HAP-contaminated process water in the ESP bottom or PM return system (referred to as a wet ESP system) to an ESP system that uses "clean" water (i.e., water uncontaminated with methanol and other gaseous organic HAP's) in the ESP bottom or PM return system; and (2) converting a wet ESP system to a dry-bottom ESP with a dry PM return system (referred to as a dry ESP system). With these two control options, the potential stripping of methanol and other gaseous organic HAP's from the black liquor or HAP-contaminated process water in the ESP system would be eliminated.

Only the impacts for the second control option, converting from a wet to a dry ESP system, were evaluated. This decision was based on (1) the lack of emissions data from ESP systems that use "clean" water in the ESP bottom or PM return system; and (2) the fact that very few mills use water in the ESP system.

The wet to dry ESP system conversion control option applies to model NDCE recovery furnaces RF-4 through RF-6, which represent existing NDCE recovery furnaces with wet ESP systems. These models represent existing NDCE recovery furnaces only, because no wet ESP systems are expected to be installed on new NDCE recovery furnaces. The environmental and energy impacts associated with this control option are presented in the following sections.

5.2.2.1 Air Pollution Impacts: Primary Emissions. This section presents the primary air impacts for model NDCE recovery furnaces RF-4 through RF-6 resulting from a wet to dry ESP system conversion. As discussed below in Section 5.2.2.2, no additional electricity requirements are expected in order to operate a dry ESP system. Therefore, no additional secondary emissions are expected to be generated from operation of a dry ESP system.

The impact of the wet to dry ESP system conversion on emissions of gaseous organic HAP's was evaluated. Table 5-3

presents the gaseous organic HAP emission factor for the model NDCE recovery furnaces. Table 5-8 and Figure 5-3 present the annual gaseous organic HAP emission estimates for model NDCE recovery furnaces.

Gaseous organic HAP emissions are reduced by about 72 percent with a wet to dry ESP system conversion. On an annual basis, gaseous organic HAP's are reduced by about 22 to 58 Mg/yr (25 to 64 ton/yr). Approximately 49 percent of these reductions result from the reduction of methanol emissions.

5.2.2.2 Energy Impacts. The tradeoff in horsepower requirements between the wet and dry ESP systems is expected to be approximately equal. For example, a wet-bottom ESP design requires four agitators and a recirculation pump, whereas a dry-bottom ESP design requires four drives for the drag system.[16] Therefore, additional energy requirements are not expected in order to operate dry ESP systems.

5.2.2.3 Water Pollution Impacts. The conversion to wet ESP systems that use "clean" water in the ESP bottom or PM return system was not considered as a control option for the reasons cited in Section 5.2.2. Therefore, no water impacts are expected for the wet to dry ESP system conversion control option.

5.2.2.4 Solid Waste Disposal Impacts. The wet to dry ESP system conversion control option is not expected to have an effect on PM control by the ESP. Furthermore, the PM catch from the ESP is recycled back into the process, and the recovery process is assumed to have sufficient capacity to absorb the additional PM. Therefore, no solid waste disposal impacts are expected for this control option.

5.2.2.5 Other Impacts. Limited information was available to estimate impacts such as noise, visual, and irreversible and irretrievable commitment of resources. The impacts, if any, are expected to be minimal under this control option.

Odor impacts were determined for the wet to dry ESP system conversion control option by comparing TRS emissions before and after the ESP conversion. Wet ESP systems are different from dry ESP systems in that they have the capability of increasing TRS

emissions in the furnace gas stream by stripping TRS from the black liquor.[17] Table 5-9 and Figure 5-4 present the annual TRS emission estimates.

The TRS emission estimates were derived from TRS emissions data (in ppm as H_2S) provided by individual pulp and paper mills.[6] Baseline TRS emissions were estimated based on the 10 percent trimmed mean TRS concentration (2.2 ppm) for NDCE recovery furnaces assumed to be equipped with wet ESP systems.[6] This trimmed mean is a compromise between the mean and the median. Using a trimmed mean with a moderate trimming proportion, such as 10 percent, yields a measure which is neither as sensitive to outlying values as the mean (since any small number of outlying values will be deleted before averaging) nor as insensitive as the median. The 10 percent trimmed mean was estimated by eliminating the smallest 10 percent and the largest 10 percent of the sample and then averaging what was left over.[18] Control level TRS emissions were estimated based on the available TRS concentration data (1 ppm) for an NDCE recovery furnace known to be equipped with a dry ESP system.[6] Using these baseline and control level numbers, it was estimated that TRS emissions would be reduced by approximately 55 percent under the wet to dry ESP system conversion control option. On an annual basis, the TRS emissions for models RF-4 through RF-6 would be reduced by about 3.0 to 7.8 Mg/yr (3.3 to 8.6 ton/yr).

5.2.3 Conversion of a DCE Recovery Furnace System to an NDCE Recovery Furnace

Converting a DCE recovery furnace system to an NDCE recovery furnace (or "low-odor conversion") was evaluated as a control option for reducing emissions of gaseous organic HAP's from DCE recovery furnace systems. The conversion of a DCE recovery furnace system to an NDCE design involves removing the DCE and BLO unit, adding a concentrator, and extending or replacing the boiler economizer. Additional upgrades are included in the low-odor conversion control option, i.e., an ESP upgrade to improve PM collection and a wet to dry ESP system conversion to reduce gaseous organic HAP emissions.

At stated in Chapter 3, the DCE provides some PM control. Therefore, with the removal of the DCE, the recovery furnace ESP often must be upgraded or replaced during a low-odor conversion in order to meet applicable PM emission limits. For the purposes of this impact analysis, an ESP upgrade PM control option that would maintain or reduce PM emissions to the NSPS level of 0.10 g/dscm (0.044 gr/dscf) has been evaluated for existing DCE recovery furnaces that have baseline PM emissions at or above the NSPS level. This PM control option applies to model DCE recovery furnaces RF-7a/7b through RF-9a/9b. These models represent existing sources only, because no new DCE recovery furnaces are expected to be built.

A PM control option that would reduce PM emissions to 0.034 g/dscm (0.015 gr/dscf) has also been evaluated for DCE recovery furnaces that have PM emissions at or below the NSPS level but greater than 0.034 g/dscm (0.015 gr/dscf). This PM control option includes an ESP upgrade coupled with the addition of a packed-bed scrubber and applies to model DCE recovery furnaces RF-7b through RF-9b.

The environmental and energy impacts associated with the low-odor conversion control option are presented for model DCE recovery furnaces in the following sections.

5.2.3.1 Air Pollution Impacts. This section presents the primary and secondary air impacts resulting from implementing the low-odor conversion control option.

5.2.3.1.1 Primary emissions. The impact on emissions of gaseous organic HAP's from converting a DCE recovery furnace system to an NDCE design and converting the wet ESP system to a dry ESP system was evaluated. The impact on PM and PM HAP emissions from upgrading the ESP was also studied. Table 5-2 presents the operating parameters, and Table 5-4 presents the model concentrations and emission factors for the model DCE recovery furnaces. Table 5-5 and Figure 5-2 present the annual emission estimates for PM and PM HAP's; Table 5-8 and Figure 5-3 present the annual emission estimates for gaseous organic HAP's.

The PM and PM HAP emission reductions achievable by upgrading the ESP for the model DCE recovery furnaces were discussed in Section 5.2.1.1.1.

With a low-odor conversion, gaseous organic HAP emissions from model DCE recovery furnace systems (including BLO units) would be reduced by about 93 percent; on an annual basis, emissions would be reduced by 69 to 206 Mg/yr (76 to 227 ton/yr). Approximately 70 percent of these reductions result from the reduction of methanol emissions.

5.2.3.1.2 Secondary emissions. Secondary emissions generated under the low-odor conversion control option were estimated for model DCE recovery furnaces. Table 5-10 presents the annual secondary emission estimates. The secondary emission reductions resulting from the removal of the BLO unit and the secondary emission increases resulting from the implementation of the low-odor conversion PM controls are discussed separately below.

As a result of the removal of the BLO unit, secondary emissions are reduced for model DCE recovery furnaces RF-7 through RF-9. The secondary PM emission estimates range from -567 to -1,100 kg/yr (-1,250 to -2,430 lb/yr); the secondary SO_2 emission estimates range from -2,770 to -5,350 kg/yr (-6,110 to -11,800 lb/yr); the secondary NO_x emission estimates range from -1,100 to -2,130 kg/yr (-2,430 to -4,700 lb/yr); and the secondary CO emission estimates range from -2,090 to -4,050 kg/yr (-4,600 to -8,920 lb/yr).

The procedures for estimating secondary emissions associated with the PM control options were presented previously in Section 5.2.1.1.2 and are the same for both converted and unconverted model DCE recovery furnaces; therefore, only the differences between the converted and unconverted models that impact secondary emissions are discussed below. The model SCA values for upgraded ESP's on DCE recovery furnaces converted to the NDCE design are higher than the model SCA values for ESP's on unconverted DCE recovery furnaces. Removal of the DCE, which removes a portion of the PM prior to the ESP, is the primary

reason for increasing the existing ESP's SCA after the low-odor conversion. The increased SCA results in an increase in the electricity requirements, and, subsequently, an increase in secondary emissions. Although the gas flow rates from converted DCE's are lower than the gas flow rates from unconverted DCE's (which reduces the fan electricity requirements), the associated decrease in electricity requirements is more than offset by the larger increase in electricity requirements due to the higher SCA. Therefore, the secondary emission estimates presented in Table 5-10 for converted DCE recovery furnaces are higher than those presented in Table 5-6 for unconverted DCE recovery furnaces. For example, the incremental PM secondary emissions for model RF-7a for the NSPS PM control option are equal to 90 kg/yr (199 lb/yr) for a converted DCE recovery furnace and 45 kg/yr (100 lb/yr) for an unconverted DCE recovery furnace. Although secondary PM emissions are higher when the PM control options are applied to the converted DCE recovery furnaces than to the unconverted DCE recovery furnaces, the secondary emissions are insignificant when compared to the primary PM emission reductions achieved with the PM control options.

5.2.3.2 Energy Impacts. Energy impacts for the low-odor conversion control option were estimated for model DCE recovery furnaces. The energy impacts include electricity impacts from both the elimination of the BLO unit and the PM controls and steam energy impacts from the furnace conversion. Table 5-2 presents the baseline and control level SCA values used to estimate the PM control electricity impacts for each model DCE recovery furnace. Table 5-11 presents the annual estimates of the energy impacts.

5.2.3.2.1 Electricity impacts. The incremental electricity impacts from the elimination of the BLO unit and the low-odor conversion PM controls are discussed separately below.

With the elimination of the BLO unit, the BLO operating electricity savings were estimated to range from -2,450 to -4,750 MWh/yr for models RF-7 through RF-9. The operating electricity savings were estimated by dividing the annual BLO

operating cost savings by the unit cost of electricity. The annual BLO operating cost savings are discussed in Chapter 6 and range from $147,000/yr to $285,000/yr for models RF-7 through RF-9. The unit cost of electricity is $0.06/kilowatt-hour (kWh).[13] Most of the BLO operating costs (about 60 percent) is for power to operate the blowers and pumps. The remaining 40 percent is for operating the reheater.[10]

The procedures for estimating the electricity requirements associated with the PM control options were presented previously in Section 5.2.1.1.3 and are the same for both converted and unconverted model DCE recovery furnaces; therefore, only the differences between the converted and unconverted models that impact electricity requirements are discussed below. As discussed in Section 5.2.3.1.2, the model SCA values for upgraded ESP's on converted DCE recovery furnaces are higher than the model SCA values for ESP's on unconverted DCE recovery furnaces. The increased SCA results in an increase in the electricity requirements. Although the gas flow rates from converted DCE recovery furnaces are lower than the gas flow rates from unconverted DCE recovery furnaces (which reduces the fan electricity requirements), the associated decrease in electricity requirements is more than offset by the larger increase in electricity requirements due to the higher SCA. Therefore, the electricity requirements for the PM control options are higher for the converted DCE recovery furnaces shown in Table 5-11 than for the unconverted DCE recovery furnaces shown in Table 5-7. For example, the electricity requirements for model RF-7a for the NSPS PM control option are equal to 391 MWh/yr for a converted DCE furnace and 191 MWh/yr for an unconverted DCE recovery furnace. Although electricity requirements are higher when the PM control options are applied to the converted DCE recovery furnaces than to the unconverted DCE recovery furnaces, these requirements are insignificant when compared to the energy savings obtained from removing the BLO unit or the energy requirements of pulp and paper mills.

5.2.3.2.2 Steam energy impacts. Implementing the low-odor conversion control option may increase steam energy production.[11] Table 5-11 presents the estimates for steam energy production for unconverted DCE recovery furnaces and the net increase in steam energy production after a low-odor conversion. The increased steam energy production for model DCE recovery furnaces RF-7 through RF-9 ranges from 1.97×10^8 megajoules per year (MJ/yr) (1.87×10^5 million Btu per year [MM Btu/yr]) to 5.91×10^8 MJ/yr (5.61×10^5 MM Btu/yr).

The net increase in steam energy production was estimated to be equal to the steam production from a low-odor conversion minus the steam requirements to operate the concentrator.

To determine the steam production from a low-odor conversion, the increase in the thermal efficiency that results from the low-odor conversion was estimated. The heat input to the system was calculated using a BLS heat content of 13,900 kJ/kg (6,000 Btu/lb) of BLS fired.[19] The increase in the thermal efficiency that results from a low-odor conversion was estimated to be 10 percentage points, based on average thermal efficiencies of 56 percent for DCE recovery furnaces and 66 percent for NDCE recovery furnaces.[20,21] Therefore, the increase in the amount of heat input that can be converted to steam with a low-odor conversion was estimated by multiplying the heat input by the 10 percentage point increase in thermal efficiency. The heat input value was divided by the thermal efficiency for a power boiler (about 85 percent) to determine the required energy input for a power boiler. The required energy input is equivalent to the increased steam production. The increased steam production was converted to annual steam production estimates for each model DCE recovery furnace, assuming 351 operating d/yr. The energy input for the power boiler can be converted to fuel savings by using the heat contents for specific power boiler fuels.

To determine the concentrator steam requirements, the steam flow for each model DCE recovery furnace was multiplied by the change in enthalpy between the entering feedwater (439.8 kJ/kg

[189.2 Btu/lb]) and the steam leaving the superheater (3,095.0 kJ/kg [1,331.5 Btu/lb]).[12] The model steam flows were based on a steam flow of approximately 4,500 kg/hr (10,000 lb/hr) of low-pressure steam for a 1.1 million kg BLS/d (2.4 million lb BLS/d) furnace.[22] The steam flow was extrapolated for each model DCE recovery furnace size, assuming that steam usage is proportional to the amount of black liquor concentrated.

5.2.3.3 Water Pollution Impacts. The conversion to ESP systems that use "clean" water in the ESP bottom or PM return system was not evaluated as a control option for the reasons cited in Section 5.2.2. Therefore, no water impacts are expected from the wet to dry ESP system conversion included under the low-odor conversion control option. The water pollution impacts resulting from adding a packed-bed scrubber (included under the more stringent PM control option) are discussed in Section 5.2.4.3.

5.2.3.4 Solid Waste Disposal Impacts. As noted in Section 5.2.1.4, the PM catch from the ESP is recycled back into the process. The recovery process was assumed to have sufficient capacity to absorb the additional PM resulting from upgrading the ESP. Therefore, no solid waste impacts are expected to be associated with the low-odor conversion PM control options.

5.2.3.5 Other Impacts. Odor impacts were determined for the low-odor conversion control option by comparing TRS emissions before and after the low-odor conversion. Table 5-9 presents the baseline TRS emission estimates for unconverted model DCE recovery furnace systems (which include BLO units) and the TRS emission reductions for converted model DCE recovery furnaces. Figure 5-4 illustrates the annual TRS emission estimates.

In order to determine the baseline TRS emissions from DCE recovery furnace systems, the baseline TRS emissions from DCE recovery furnaces and BLO units were combined. The DCE recovery furnace baseline TRS emission estimate was derived from TRS emissions data (in ppm as H_2S) provided by individual pulp and paper mills.[6] Based on the available TRS emissions data, the

10 percent trimmed mean TRS concentration for DCE recovery furnaces is approximately 12 ppm.[6] The BLO unit TRS emission estimate was derived from the pulp and paper NSPS supporting document.[14] Based on the bounds of TRS values given in the supporting document, a median value of 0.05 kg/ADMP (0.10 lb/ADTP) was used as the baseline TRS emission estimate for BLO units. The control level TRS emission estimate (1 ppm) for converted DCE recovery furnaces was derived from TRS emissions data provided for an NDCE recovery furnace known to be equipped with a dry ESP system.[6] Emissions of TRS compounds are expected to be reduced by approximately 94 percent under the low-odor conversion control option. On an annual basis, the TRS emission reductions would be 22 to 67 Mg/yr (25 to 74 ton/yr).

Limited information was available to estimate quantitatively impacts such as noise, visual, and irreversible and irretrievable commitment of resources. Beneficial visual impacts are expected under the low-odor conversion control option. The beneficial visual impacts include the reduction in PM emissions, the elimination of the BLO vent gas stacks, and the reduction in the amount of moisture (i.e., steam) coming out of the stack with the elimination of the DCE. This reduction in moisture content could be completely eliminated if the more stringent PM control option is implemented. The more stringent PM control option includes a packed-bed scrubber, which would increase the moisture content in the stack. Adding a packed-bed scrubber also requires additional equipment (i.e., larger fans to overcome pressure drops and pumps) that would increase noise levels. However, these incremental noise increases are expected to be small compared to the typical background noise levels at pulp and paper mills. The other impacts, if any, are expected to be minimal as a result of implementing the low-odor conversion control option.

5.2.4 Addition of Packed-Bed Scrubber

The addition of a packed-bed scrubber downstream of the ESP is included in two of the control options examined for recovery furnaces. These control options are (1) the use of an ESP plus a packed-bed scrubber to meet an outlet PM emission level of

0.034 g/dscm (0.015 gr/dscf); and (2) the use of a packed-bed scrubber to reduce HCl emissions from recovery furnaces.

The impacts of combining an ESP upgrade or ESP replacement with a packed-bed scrubber to control PM emissions to 0.034 g/dscm (0.015 gr/dscf) are presented in Section 5.2.1. The impacts presented in the sections below are based on the use of a packed-bed scrubber to control HCl emissions from recovery furnaces. The applicable model recovery furnaces for the packed-bed scrubber HCl control option are model NDCE and DCE recovery furnaces RF-1 through RF-9. The model DCE recovery furnaces include both DCE recovery furnaces converted to the NDCE design and unconverted DCE recovery furnaces. Tables 5-1 and 5-2 present the sizes and operating parameters for model NDCE and DCE recovery furnaces, respectively.

5.2.4.1 Air Pollution Impacts. This section presents the primary and secondary air impacts resulting from adding a packed-bed scrubber to control HCl emissions from model recovery furnaces.

5.2.4.1.1 Primary emissions. The impact of the packed-bed scrubber control option on emissions of HCl and another acid gas, SO_2, was evaluated for model recovery furnaces. Tables 5-3 and 5-4 present the HCl and SO_2 emission factors for model NDCE and DCE recovery furnaces, respectively. Table 5-12 presents the annual emission estimates for HCl and SO_2; Figures 5-5 and 5-6 illustrate the annual emission estimates for HCl and SO_2, respectively.

The incremental HCl emission reductions for model NDCE and converted model DCE recovery furnaces RF-1 through RF-9 range from 8.3 to 36 Mg/yr (9.1 to 40 ton/yr). The incremental HCl emission reductions for the unconverted model DCE recovery furnaces RF-7 through RF-9 are slightly lower because of different furnace characteristics (i.e., higher model gas flow rates). The HCl emission reductions for unconverted models RF-7 through RF-9 range from 7.8 to 24 Mg/yr (8.6 to 26 ton/yr).

The incremental SO_2 emission reductions for model NDCE and converted model DCE recovery furnaces RF-1 through RF-9 range

from 59 to 384 Mg/yr (65 to 423 ton/yr). The incremental SO_2 emission reductions for the unconverted model DCE recovery furnaces RF-7 through RF-9 are slightly higher because the control level SO_2 emission factor associated with DCE recovery furnaces is lower than the control level SO_2 emission factor associated with NDCE recovery furnaces. The SO_2 emission reductions for unconverted models RF-7 through RF-9 range from 74 to 221 Mg/yr (81 to 244 ton/yr). Overall, the SO_2 emission reduction achievable with a packed-bed scrubber is approximately 10 times greater than the HCl emission reduction.

Based on information from a scrubber manufacturer, emissions of gaseous HAP's such as methanol, formaldehyde, and acetaldehyde may be reduced with a packed-bed scrubber, but no emissions data were available to estimate potential reductions for these HAP's.[23] Therefore, HCl and SO_2 were assumed to be the only gaseous pollutants reduced with a packed-bed scrubber.

5.2.4.1.2 Secondary emissions. Secondary emissions generated under the packed-bed scrubber control option were estimated for model recovery furnaces. The baseline secondary emissions for the packed-bed scrubber control option are equal to zero (i.e., no packed-bed scrubber at baseline). Table 5-13 presents the incremental secondary emissions relative to baseline for each model recovery furnace.

The increases in secondary PM emissions from operating a recovery furnace packed-bed scrubber are insignificant compared to the reductions in primary PM emissions from implementing the recovery furnace PM control options. The secondary PM emission estimates for model NDCE and converted DCE recovery furnaces RF-1 through RF-9 range from 138 to 594 kg/yr (304 to 1,310 lb/yr). The secondary PM emission estimates for unconverted model DCE recovery furnaces RF-7 through RF-9 range from 145 to 435 kg/yr (320 to 960 lb/yr).

The increases in secondary SO_2 emissions from operating a recovery furnace packed-bed scrubber are insignificant compared to the reductions in primary SO_2 emissions from operating a packed-bed scrubber. For model NDCE/converted DCE recovery

furnaces RF-1 through RF-9, the secondary SO_2 emission estimates for the packed-bed scrubber control option range from 671 to 2,900 kg/yr (1,480 to 6,390 lb/yr). The secondary SO_2 emission estimates for unconverted model DCE recovery furnaces RF-7 through RF-9 range from 708 to 2,120 kg/yr (1,560 to 4,670 lb/yr).

For model NDCE and converted DCE recovery furnaces RF-1 through RF-9, the secondary NO_x emission estimates range from 266 to 1,150 kg/yr (587 to 2,540 lb/yr). For unconverted model DCE recovery furnaces RF-7 through RF-9, the secondary NO_x emission estimates range from 281 to 844 kg/yr (619 to 1,860 lb/yr).

For model NDCE and converted DCE recovery furnaces RF-1 through RF-9, the secondary CO emission estimates range from 503 to 2,190 kg/yr (1,110 to 4,820 lb/yr). For unconverted model DCE recovery furnaces RF-7 through RF-9, the secondary CO emission estimates range from 531 to 1,600 kg/yr (1,170 to 3,520 lb/yr).

5.2.4.2 Energy Impacts. The increases in energy impacts for the packed-bed scrubber control option were determined for model recovery furnaces. The baseline energy level for the packed-bed scrubber control option is equal to zero (i.e., no packed-bed scrubber at baseline). Table 5-14 presents the incremental energy impacts relative to baseline.

The energy impacts for the packed-bed scrubber control option range from 593 to 2,570 MWh/yr for NDCE/converted DCE models RF-1 through RF-9 and 625 to 1,870 MWh/yr for unconverted DCE models RF-7 through RF-9.

5.2.4.3 Water Pollution Impacts. Adding a packed-bed scrubber creates a new wastewater stream for recovery furnaces, thereby affecting the water balance at the mill. Whether there are any significant wastewater disposal impacts for this option depends on whether or not the scrubber discharge could be recycled and reused elsewhere in the mill. Wastewater impacts were estimated for model recovery furnaces assuming there was no prior recycle or reuse. Tables 5-1 and 5-2 present the model wastewater flow rates used to estimate wastewater impacts for model NDCE and DCE recovery furnaces, respectively. The baseline

wastewater impacts for the packed-bed scrubber control option are equal to zero (i.e., no packed-bed scrubber at baseline). Table 5-15 presents the incremental wastewater impacts relative to baseline.

Information was available from two sources to determine the amount of wastewater generated by the packed-bed scrubber control option. Wastewater flow rate values were estimated using procedures in the OAQPS Control Cost Manual and using wastewater flow rate values provided by a scrubber manufacturer.[9,23] The wastewater flow rate values calculated using the OAQPS Control Cost Manual are presented in Chapter 6 in Table 6-23. A scrubber manufacturer provided wastewater flow rate values of 21.24, 53.07, and 84.79 liters per minute (L/min) (5.61, 14.02, and 22.40 gallons per minute [gpm]) for three model recovery furnaces with scrubber inlet gas flow rates of 47.2, 118, and 189 m^3/sec (100,000, 250,000, and 400,000 acfm), respectively.[23] The wastewater flow rates from the scrubber manufacturer were corrected to correspond to the model sizes used in this impact analysis and then were averaged with the wastewater flow rates obtained using the OAQPS Control Cost Manual to determine the average wastewater impacts for the packed-bed scrubber control option. The increased wastewater impacts presented below for the recovery furnace packed-bed scrubber control option are insignificant compared to the reduced wastewater impacts presented in Section 5.5.1.3 for the lime kiln PM control option, which involves replacing existing wet scrubbers with ESP's.

The wastewater impacts for NDCE/converted DCE models RF-1 through RF-9 range from 6.1 to 26 million liters per year (L/yr) (1.6 to 7.0 million gallons per year [gal/yr]). The wastewater impacts for unconverted DCE models RF-7 through RF-9 range from 6.3 to 19 million L/yr (1.7 to 5.0 million gal/yr).

5.2.4.4 Solid Waste Disposal Impacts. No solid waste disposal impacts are expected from adding a packed-bed scrubber.

5.2.4.5 Other Impacts. Adding a packed-bed scrubber requires additional equipment (i.e., larger fans to overcome higher pressure drops and pumps) that would increase noise

levels. However, these incremental noise increases are expected to be small compared to the typical background noise levels at pulp and paper mills. Limited information is available to estimate quantitatively the visual, odor, and other impacts. However, because adding a packed-bed scrubber increases the moisture (i.e., steam) coming out of the stack, there may be negative visual impacts. Beneficial odor impacts may result from the reduction in acid gas emissions. The other impacts, if any, are expected to be minimal as a result of implementing this control option.

5.3 BLACK LIQUOR OXIDATION UNIT CONTROL OPTION

Two control options, (1) conversion of a DCE recovery furnace system to an NDCE recovery furnace and (2) incineration of BLO vent gases, were evaluated for controlling gaseous organic HAP emissions from air-sparging BLO units. The environmental and energy impacts of the first option--converting DCE recovery furnace systems to NDCE recovery furnaces--were presented in Section 5.2.3. The following sections present the environmental and energy impacts of the second BLO control option--incineration of BLO vent gases. This BLO control option applies to model BLO units BLO-1 through BLO-3, which represent existing BLO units associated with DCE recovery furnaces. These models represent only existing BLO units because no new DCE recovery furnace systems with BLO units are expected to be installed.

5.3.1 Air Pollution Impacts

This section presents the primary and secondary air impacts for model BLO units resulting from implementing the BLO vent gas control option.

5.3.1.1 Primary Emissions. The impact of the BLO vent gas control option on emissions of gaseous organic HAP's is discussed below for the model BLO units. Table 5-16 presents the operating parameters, as well as the model emission factors, for each model BLO unit. Table 5-17 and Figure 5-7 present the gaseous organic HAP annual emission estimates for model BLO units.

Gaseous organic HAP emissions are assumed to be reduced by about 98 percent from the model BLO units with the incineration

of BLO vent gases in a power boiler or other incineration device. For models BLO-1 through BLO-3, gaseous organic HAP emissions are estimated to be reduced by 29 to 86 Mg/yr (31 to 94 ton/yr). Approximately 85 percent of these reductions result from the reduction of methanol emissions.

5.3.1.2 Secondary Emissions. The secondary emissions generated under the BLO vent gas control option were estimated for the model BLO units. The baseline secondary emissions for the BLO vent gas control option are equal to zero (i.e., no BLO control at baseline). Table 5-18 presents the incremental secondary emissions relative to baseline.

For models BLO-1 through BLO-3, the secondary PM emission estimates range from 400 to 1,210 kg/yr (883 to 2,670 lb/yr); the secondary NO_x emission estimates range from 776 to 2,340 kg/yr (1,710 to 5,160 lb/yr); and the secondary CO emission estimates range from 1,470 to 4,440 kg/yr (3,240 to 9,780 lb/yr).

The secondary SO_2 emission estimates associated with controlling BLO vent gas emissions range from 9,710 to 29,200 kg/yr (21,400 to 64,300 lb/yr) for models BLO-1 through BLO-3. As discussed in Section 5.1.1.2, the secondary SO_2 emissions associated with the BLO vent gas control option include (1) SO_2 emissions resulting from the generation of energy required to collect and incinerate the BLO vent gases and (2) SO_2 emissions generated when the TRS in the BLO vent gases is combusted. All of the TRS collected is assumed to be combusted. Also, all of the SO_2 that is formed from the combustion of TRS compounds is assumed to be emitted to the atmosphere; this assumption is a worst-case assumption since many power boilers are equipped with scrubbers for SO_2 control.

5.3.2 Energy Impacts

The energy impacts for the BLO vent gas control option were estimated for the model BLO units. The baseline energy impacts for the BLO vent gas control option are equal to zero (i.e., no BLO control at baseline). Table 5-19 presents the incremental energy impacts relative to baseline.

The model energy impacts were estimated based on the information provided by an individual mill regarding the total horsepower requirements to operate the mill's BLO vent gas control equipment.[10] The energy requirements for the BLO control system at this mill were scaled for the model BLO units, assuming a direct relationship between BLO vent gas flow rate and energy requirements. The mill has a BLO vent gas flow rate of 7.7 m^3/sec (16,327 acfm).[6] Based on information supplied by the mill, 980 kilowatts (kW) (100 horsepower [hp]) are required to operate the mill water booster pump motor, 29 kW (3 hp) to operate the BLO condenser condensate pump motor, and 3,900 kW (400 hp) to operate the BLO off gas blower motor.[10] The annual energy requirements were estimated assuming the BLO system operates 8,424 hr/yr and range from 1,720 to 5,210 MWh/yr for models BLO-1 through BLO-3.

5.3.3 Water Pollution Impacts

Although some condensate is collected from the BLO vent gases, the quantity is negligible.[24] Therefore, the water impacts for the BLO control option are expected to be negligible.

5.3.4 Solid Waste Disposal Impacts

No solid waste impacts are expected to be associated with the BLO control option.

5.3.5 Other Impacts

Beneficial odor impacts are expected to result from implementing the BLO vent gas control option and are discussed below for the model BLO units. Table 5-20 and Figure 5-8 present the annual TRS emission estimates.

As discussed in Section 5.1.5, a mid-range value of 0.05 kg/ADMP (0.10 lb/ADTP) was used to calculate the baseline TRS emissions from BLO units.[14] The baseline TRS emissions are assumed to be reduced by about 98 percent with the incineration of BLO vent gases in a power boiler or other incineration device. On an annual basis, the TRS emissions were estimated to be reduced by about 4.1 to 12 Mg/yr (4.6 to 14 ton/yr) for models BLO-1 through BLO-3.

Beneficial visual impacts are expected to result from the elimination of BLO vent gas stacks. Limited information was available to estimate impacts such as noise and irreversible and irretrievable commitment of resources. The impacts, if any, are expected to be minimal as a result of implementing the BLO vent gas control option.

5.4 SMELT DISSOLVING TANK CONTROL OPTIONS

This section discusses the environmental and energy impacts resulting from implementing the SDT PM control options. Two PM control options that would reduce PM emissions from SDT's have been evaluated. The first option would reduce PM emissions from existing SDT's to the NSPS level of 0.10 kg/Mg (0.20 lb/ton) of BLS. The second option would reduce PM emissions from existing SDT's to a more stringent level of 0.06 kg/Mg (0.12 lb/ton) of BLS; scrubbers installed on new SDT's would also be required to meet a PM level of 0.06 kg/Mg (0.12 lb/ton) BLS.

For mills with existing SDT scrubbers, the environmental impacts of both PM control options were estimated based on replacing the existing scrubber with a new scrubber. These impacts were estimated for SDT models SDT-1 through SDT-4. The environmental impacts of installing scrubbers on new SDT's under the second, more stringent PM control option also apply to SDT-1 through SDT-4.

For mills with existing SDT mist eliminators, the environmental impacts of both PM control options were estimated based on replacing the existing mist eliminator with a new scrubber. These impacts were estimated for SDT models SDT-5 through SDT-7. The impacts of installing new mist eliminators were not examined because mist eliminators are not assumed to be installed on new SDT's.

5.4.1 PM Controls

Table 5-21 presents the model SDT sizes and operating parameters for the PM control options. The environmental and energy impacts associated with the control options are discussed in the following sections.

5.4.1.1 Air Pollution Impacts. This section presents the primary and secondary air impacts for model SDT's resulting from the implementation of the PM control options.

5.4.1.1.1 Primary emissions. Although emission test data from SDT's on PM HAP performance are limited, available information indicates that APCD's that achieve the greatest PM removal also provide the best performance for the HAP portion of the PM.[15] Therefore, PM performance can be used as a surrogate for PM HAP's. Because emission test data from SDT scrubbers indicate that PM emissions are reduced with PM controls, PM HAP emissions would also be reduced.[6] As stated in Section 5.1.1.1, PM HAP emissions from SDT's were estimated to be 0.06 percent of PM based on a comparison of PM and PM HAP emission data for SDT's.

Table 5-21 presents the model PM emission factors for models SDT-1 through SDT-7. The baseline PM emission factor is 0.18 kg/Mg (0.37 lb/ton) BLS for model SDT's with scrubbers (models SDT-1 through SDT-4).[2] The baseline PM emission factor is 0.23 kg/Mg (0.46 lb/ton) BLS for model SDT's with mist eliminators (models SDT-5 through SDT-7).[2] The control level PM emission factors are 0.10 kg/Mg (0.20 lb/ton) BLS and 0.06 kg/Mg (0.12 lb/ton) BLS.

Table 5-22 and Figure 5-9 present the annual PM and PM HAP emission estimates. By controlling PM emissions from a baseline of 0.18 kg/Mg (0.37 lb/ton) BLS to a control level of 0.10 kg/Mg (0.20 lb/ton) BLS, PM emissions would be reduced by 46 percent from models SDT-1 through SDT-4. On an annual basis, PM emissions would be reduced by about 12 to 53 Mg/yr (13 to 58 ton/yr); PM HAP emissions would be reduced by about 0.007 to 0.03 Mg/yr (0.008 to 0.03 ton/yr).

By controlling PM emissions from 0.18 kg/Mg (0.37 lb/ton) BLS to a more stringent control level of 0.06 kg/Mg (0.12 lb/ton) BLS, PM emissions would be reduced by 68 percent from models SDT-1 through SDT-4. On an annual basis, PM emissions would be reduced by about 18 to 78 Mg/yr (20 to 86 ton/yr); PM HAP

emissions would be reduced by about 0.01 to 0.05 Mg/yr (0.01 to 0.05 ton/yr).

By controlling PM emissions from a baseline of 0.23 kg/Mg (0.46 lb/ton) BLS to a control level of 0.10 kg/Mg (0.20 lb/ton) BLS, PM emissions would be reduced by 57 percent from models SDT-5 through SDT-7. On an annual basis, PM emissions would be reduced by about 19 to 56 Mg/yr (21 to 62 ton/yr); PM HAP emissions would be reduced by about 0.01 to 0.03 Mg/yr (0.01 to 0.04 ton/yr).

By controlling PM emissions from 0.23 kg/Mg (0.46 lb/ton) BLS to a more stringent control level of 0.06 kg/Mg (0.12 lb/ton) BLS, PM emissions would be reduced by 74 percent from models SDT-5 through SDT-7. On an annual basis, PM emissions would be reduced by about 24 to 73 Mg/yr (27 to 81 ton/yr); PM HAP emissions would be reduced by about 0.01 to 0.04 Mg/yr (0.02 to 0.05 ton/yr).

5.4.1.1.2 Secondary emissions. As discussed in the following section, there is only a small increase in energy requirements from implementing PM controls for SDT's. The increases in energy requirements from a scrubber replacement were considered negligible; because the secondary emissions were estimated based on the increase in energy requirements, they were also considered negligible and are not presented in this impact analysis for models SDT-1 through SDT-4. A different scrubber design, rather than a higher pressure drop, is used in this impact analysis to estimate the impacts for improved PM control for SDT scrubbers.

Although small, the difference in pressure drop between the existing mist eliminator and the new design scrubber replacing it was considered large enough to justify estimating the increase in APCD energy requirements for those model SDT's replacing existing mist eliminators with new wet scrubbers (models SDT-5 through SDT-7). The increases in secondary emissions were then estimated for those SDT models based on the increase in APCD energy requirements. Table 5-23 presents the annual secondary emission estimates for models SDT-5 through SDT-7. As shown in the table,

the increases in secondary emissions associated with replacing an existing mist eliminator with a new scrubber are small; the increases in secondary PM emissions are especially insignificant compared to the reductions in primary PM emissions associated with the SDT PM control options.

5.4.1.2 Energy Impacts. For those SDT's replacing existing scrubbers with new scrubbers, the resulting increases in energy requirements are not included in this impact analysis; the increases are considered negligible. The pressure drop for the existing scrubber design, i.e., 12 mm Hg (6.5 in. of H_2O), is only slightly lower than the pressure drop for the new scrubber design, i.e., 13 mm Hg (7 in. H_2O).[6,25] A different scrubber design, rather than a higher pressure drop, is used in this impact analysis to estimate the impacts for improved PM control for SDT scrubbers. Therefore, no energy impacts are presented for models SDT-1 through SDT-4.

As discussed in the previous section, the difference in pressure drop between the existing mist eliminator and the new design scrubber replacing it was considered large enough to justify estimating the incremental APCD energy requirements for models SDT-5 through SDT-7. Based on information from mills, the baseline pressure drop for existing mist eliminators is 1.3 mm Hg (0.7 in. H_2O); the pressure drop for the new scrubber design is estimated to be 13 mm Hg (7 in. H_2O).[6,25] The increase in pressure drop from 1.3 to 13 mm Hg (0.7 to 7 in. H_2O) was used to estimate the incremental energy requirements. Table 5-24 presents the annual energy requirements for models SDT-5 through SDT-7. As shown in the table, the incremental energy requirements associated with replacing an existing mist eliminator with a new scrubber are small.

5.4.1.3 Water Pollution Impacts. By increasing the amount of PM removed from the exhaust gases, the SDT PM control options would increase the amount of PM in the blowdown. Blowdown rates would have to increase because of the greater amount of PM, thereby increasing the amount of wastewater generated. Because wastewater from SDT scrubbers and mist eliminators would

typically be allowed to drain into the SDT to react with the smelt, the additional wastewater generated under this control option is expected to be reused. The SDT is assumed to have sufficient capacity to absorb the additional wastewater and PM resulting from implementing the PM control options; therefore, no wastewater impacts are expected.

5.4.1.4 Solid Waste Disposal Impacts. No solid waste disposal impacts are expected from implementation of the PM control options for SDT's.

5.4.1.5 Other Impacts. Beneficial visual impacts are expected from the reduced PM emissions coming out of the stack. Replacing an existing mist eliminator with an SDT scrubber that has a higher pressure drop would require a larger fan to overcome the higher pressure drop. The larger fan would increase noise levels. However, these incremental noise increases would be small compared to typical background noise levels at pulp and paper mills. Limited information is available to determine the other impacts; however, the impacts, if any, are expected to be minimal as a result of implementing the SDT PM control options.

5.5 LIME KILN CONTROL OPTIONS

This section discusses the environmental and energy impacts resulting from implementing control options designed to reduce PM emissions from lime kilns. Two PM control options were evaluated for existing and new lime kilns. These control options are described below.

One PM control option that has been evaluated for existing lime kilns would reduce PM emissions to the NSPS level of 0.15 g/dscm (0.067 gr/dscf). For existing lime kilns with wet scrubbers, the control option would involve replacing the existing scrubber with an ESP. The actual control device (i.e., ESP or high-efficiency scrubber) selected by a particular mill would actually be site-specific. The impacts for this PM control option were estimated for model lime kilns LK-1 through LK-3, which represent existing lime kilns controlled with wet scrubbers.

Based on PM emissions data supplied by mills, lime kilns controlled with ESP's already achieve a PM level of 0.15 g/dscm (0.067 gr/dscf).[6,25] Therefore, the impacts for the control option reducing PM emissions to 0.15 g/dscm (0.067 gr/dscf) were not estimated for lime kilns controlled with ESP's (represented by models LK-4 through LK-6).

A second PM control option that was evaluated for new and existing lime kilns would reduce PM emissions to 0.023 g/dscm (0.010 gr/dscf). For existing lime kilns with wet scrubbers, the control option would involve replacing the existing scrubber with an ESP; impacts would be estimated for model lime kilns LK-1 through LK-3. For existing lime kilns with ESP's, the control option would involve upgrading the existing ESP. For new lime kilns, the control option would involve installing a new ESP capable of achieving the 0.023 g/dscm (0.010 gr/dscf) PM level. The impacts for upgrading or installing an ESP were estimated for model lime kilns LK-4 through LK-6. The actual control device selected by a particular mill would actually be site-specific.

5.5.1 PM Controls

Table 5-25 presents the model lime kiln sizes and operating parameters used in estimating the impacts of the lime kiln PM control options. The environmental and energy impacts associated with the control options are presented in the following sections.

5.5.1.1 Air Pollution Impacts. This section presents the primary and secondary air impacts estimated for each model lime kiln.

5.5.1.1.1 Primary emissions. Emission test data from lime kiln ESP's on PM HAP performance are limited. As mentioned in Section 5.4.1.1.1, PM performance can be used as a surrogate for PM HAP's, and because emission test data from lime kiln ESP's indicate that PM emissions are reduced with PM controls, PM HAP emissions would also be reduced.[6] As stated in Section 5.1.1.1, PM HAP emissions from lime kilns were estimated to be 1.4 percent of PM based on a comparison of PM and PM HAP emission data for lime kilns.

Table 5-25 presents the PM emission concentrations for the PM control options. Table 5-26 and Figure 5-10 present the annual PM and PM HAP emission estimates.

The baseline PM concentration is 0.27 g/dscm (0.12 gr/dscf) for model lime kilns with wet scrubbers (models LK-1 through LK-3).[2] By controlling PM emissions from a baseline of 0.27 g/dscm (0.12 gr/dscf) to an NSPS control level of 0.15 g/dscm (0.067 gr/dscf), PM emissions would be reduced by 44 percent from models LK-1 through LK-3. On an annual basis, PM emissions would be reduced by about 16 to 53 Mg/yr (18 to 58 ton/yr); PM HAP emissions would be reduced by about 0.2 to 0.7 Mg/yr (0.2 to 0.8 ton/yr).

By controlling PM emissions from 0.27 g/dscm (0.12 gr/dscf) to a more stringent control level of 0.023 g/dscm (0.010 gr/dscf), PM emissions would be reduced by 92 percent from models LK-1 through LK-3. On an annual basis, PM emissions would be reduced by about 33 to 110 Mg/yr (37 to 121 ton/yr); PM HAP emissions would be reduced by about 0.5 to 1.5 Mg/yr (0.5 to 1.7 ton/yr).

The baseline PM concentration for model lime kilns with ESP's (models LK-4 through LK-6) is the NSPS PM level (0.15 g/dscm [0.067 gr/dscf]). By controlling PM emissions from 0.15 g/dscm (0.067 gr/dscf) to a control level of 0.023 g/dscm (0.010 gr/dscf), PM emissions would be reduced by 85 percent from models LK-4 through LK-6. On an annual basis, PM emissions would be reduced by about 17 to 57 Mg/yr (19 to 63 ton/yr); PM HAP emissions would be reduced by about 0.2 to 0.8 Mg/yr (0.3 to 0.9 ton/yr).

5.5.1.1.2 <u>Secondary emissions</u>. The incremental secondary emissions were estimated based on the difference between the baseline and control level APCD energy requirements for lime kilns. Table 5-27 presents the annual secondary emission estimates associated with the PM control options.

Less energy is needed to operate ESP's than to operate scrubbers. Therefore, if the existing scrubber is replaced with an ESP to improve PM control, APCD energy requirements would be

reduced. Because the secondary emissions were estimated based on the APCD energy requirements, they would also be reduced. As shown in Table 5-27, the reductions in secondary emissions are small; the reductions in secondary PM emissions are especially small compared to the reductions in primary PM emissions under the lime kiln PM control options. The reductions are smaller under the more stringent PM control option since that option would require a larger ESP with more energy requirements.

If the existing ESP is upgraded to reduce PM emissions to a more stringent level, secondary emissions would be increased because the upgraded ESP would have higher energy requirements. As shown in Table 5-27, the increases in secondary emissions resulting from the ESP upgrade are small; the increases in secondary PM emissions are especially insignificant compared to the reductions in primary PM emissions associated with this more stringent PM control option.

5.5.1.2 Energy Impacts. Table 5-28 presents the annual APCD energy requirements associated with the lime kiln PM control options.

Because ESP's require less energy to operate than scrubbers, energy requirements are reduced for lime kilns that reduce PM emissions by replacing the existing scrubber with an ESP. The baseline energy requirements for the existing scrubber were determined based on a baseline pressure drop of 39 mm Hg (21 in. H_2O) for lime kiln scrubbers.[6] The energy requirements for ESP's controlling PM emissions to 0.15 g/dscm (0.067 gr/dscf) were determined based on an SCA of 90 $m^2/(m^3/sec)$ (460 ft^2/1,000 acfm) from an ESP with long-term PM emission data below 0.15 g/dscm (0.067 gr/dscf).[6,25] The energy requirements for ESP's controlling PM emissions to 0.023 g/dscm (0.010 gr/dscf) were determined based on an SCA of 220 $m^2/(m^3/sec)$ (1,120 ft^2/1,000 acfm) from an ESP with long-term PM emission data at or below 0.023 g/dscm (0.010 gr/dscf).[6,25]

As shown in Table 5-28, the reductions in energy requirements associated with replacing an existing scrubber with an ESP are small. The reductions are even smaller when baseline

PM emissions are controlled even further, to a level of 0.023 g/dscm (0.010 gr/dscf); additional energy is required for the ESP, thereby reducing the energy savings from replacing the existing scrubber with an ESP.

Energy requirements are increased if existing ESP's are upgraded to improve PM control. An increase in SCA from 90 $m^2/(m^3/sec)$ (460 ft^2/1,000 acfm) to 220 $m^2/(m^3/sec)$ (1,120 ft^2/1,000 acfm) was used to estimate the increase in energy requirements associated with an ESP upgrade. As shown in Table 5-28, the increases in energy requirements associated with an ESP upgrade are small.

5.5.1.3 Water Pollution Impacts. If a wet scrubber is replaced with an ESP to improve PM control, the wastewater stream from the scrubber would be eliminated, thereby affecting the water balance at the mill. Whether there would be any significant reduction in wastewater disposal for this option would depend on whether or not the scrubber discharge was previously recycled and reused. Wastewater impacts were estimated for model lime kilns assuming there was no prior recycle or reuse.

The wastewater impacts were estimated based on a factor of 2,250 kilograms of wastewater per oven-dried megagram of pulp (kg wastewater/ODMP) (4,500 pounds of wastewater per oven-dried ton of pulp [lb wastewater/ODTP]) for lime kiln scrubber blowdown.[21] This factor was converted to annual wastewater impacts using conversion factors of 1.0 kg H_2O/L (8.345 lb H_2O/gal) and 0.9 ODMP/ADMP (0.9 ODTP/ADTP) and multiplying by the product of the model ADMP/d (ADTP/d) and 351 operating d/yr. Table 5-29 presents the annual wastewater impacts for model lime kilns LK-1 through LK-3. With the replacement of existing wet scrubbers with ESP's, the wastewater discharge would be reduced; the reductions in wastewater impacts were estimated to range from -226 to -709 million L/yr (-60 to -187 million gal/yr).

Because ESP's operate on a dry basis, no water pollution impacts are associated with lime kiln PM control if the control involves an ESP upgrade.

5.5.1.4 Solid Waste Disposal Impacts. Dry PM catch would be generated if the existing scrubber is replaced with an ESP to improve PM control. Also, a larger PM catch would be generated if the existing ESP is upgraded to improve PM control. As stated in Chapter 2, existing lime kilns with ESP's return the PM catch directly to the lime kiln. The lime kiln is expected to have sufficient capacity to absorb the PM catch resulting from implementing the lime kiln PM control options.

5.5.1.5 Other Impacts. Beneficial visual impacts are expected as a result of the reduced PM emissions and reduced moisture (i.e., steam) coming out of the stack. The moisture content in the stack is lower with an ESP as the control device than with a scrubber. If the existing scrubber is replaced with an ESP to improve PM control, the noise from the larger fans used to overcome the higher pressure drop would be eliminated, thereby reducing noise impacts. However, the reduction in noise is not expected to be noticeable due to the high background noise levels typically associated with pulp and paper mills. Limited information is available to determine the other impacts, but the impacts, if any, are expected to be minimal as a result of implementing the lime kiln PM control options.

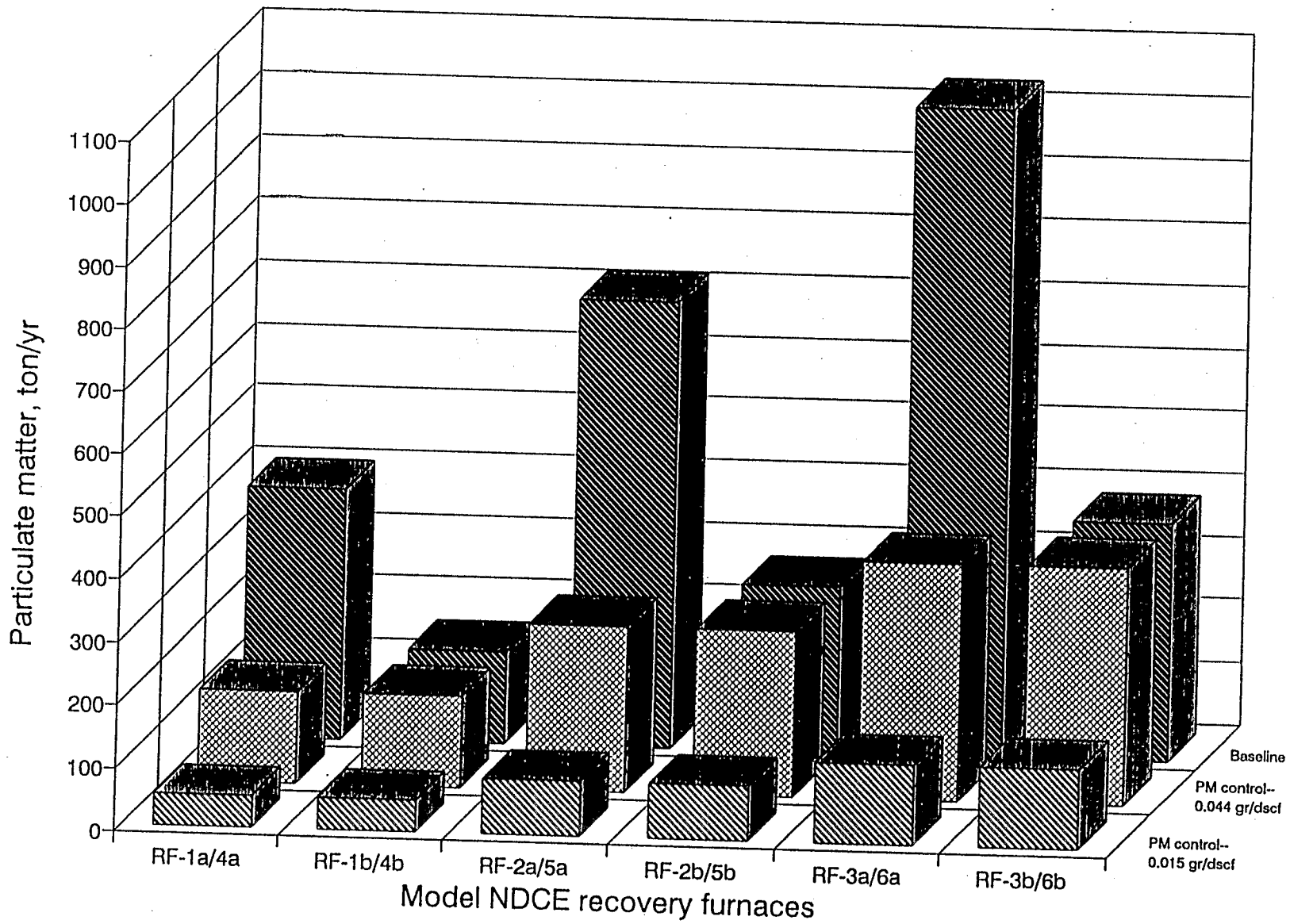

Figure 5-1. Particulate matter emissions from model NDCE recovery furnaces.

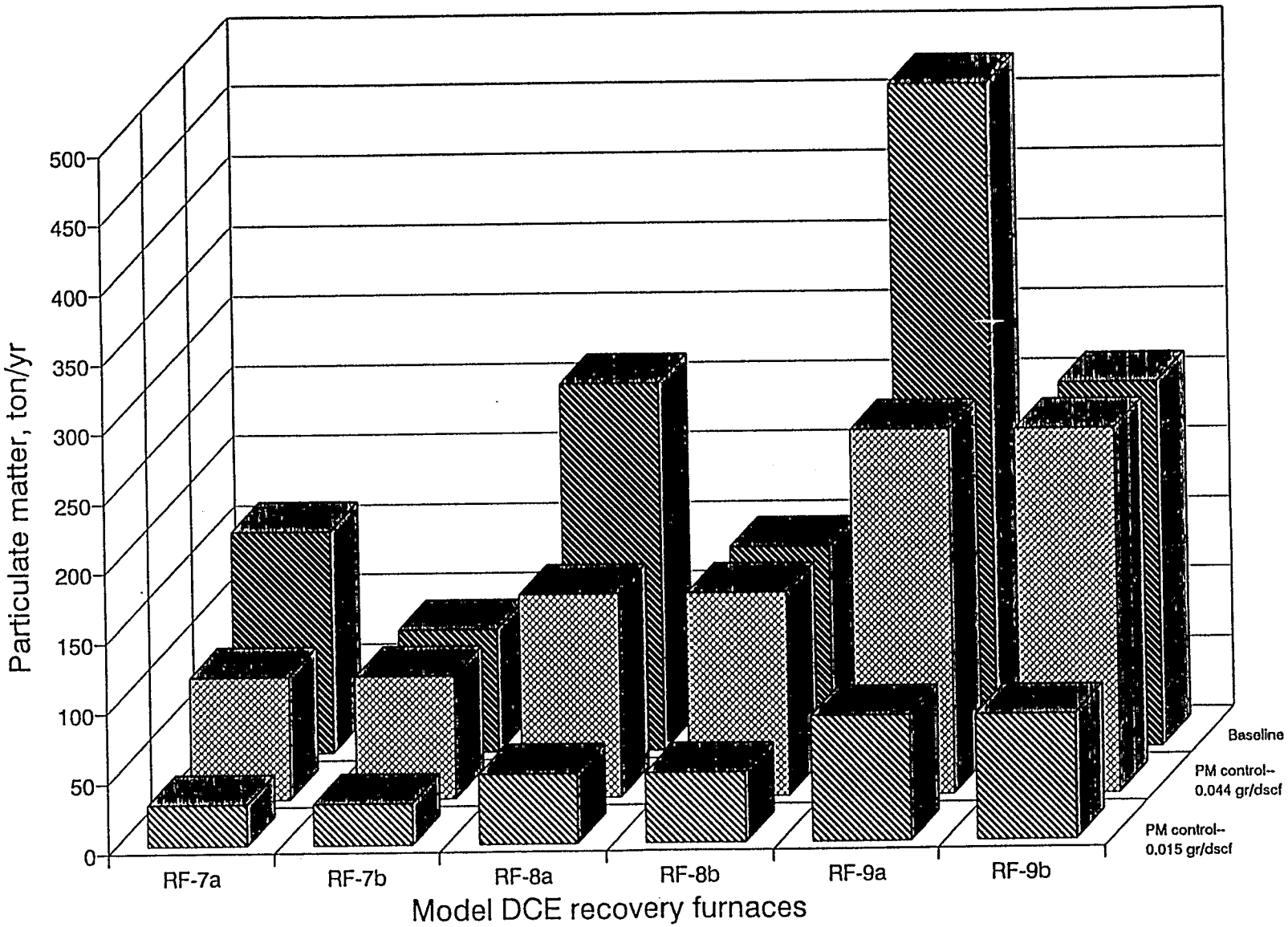

Figure 5-2. Particulate matter emissions from model DCE recovery furnaces.

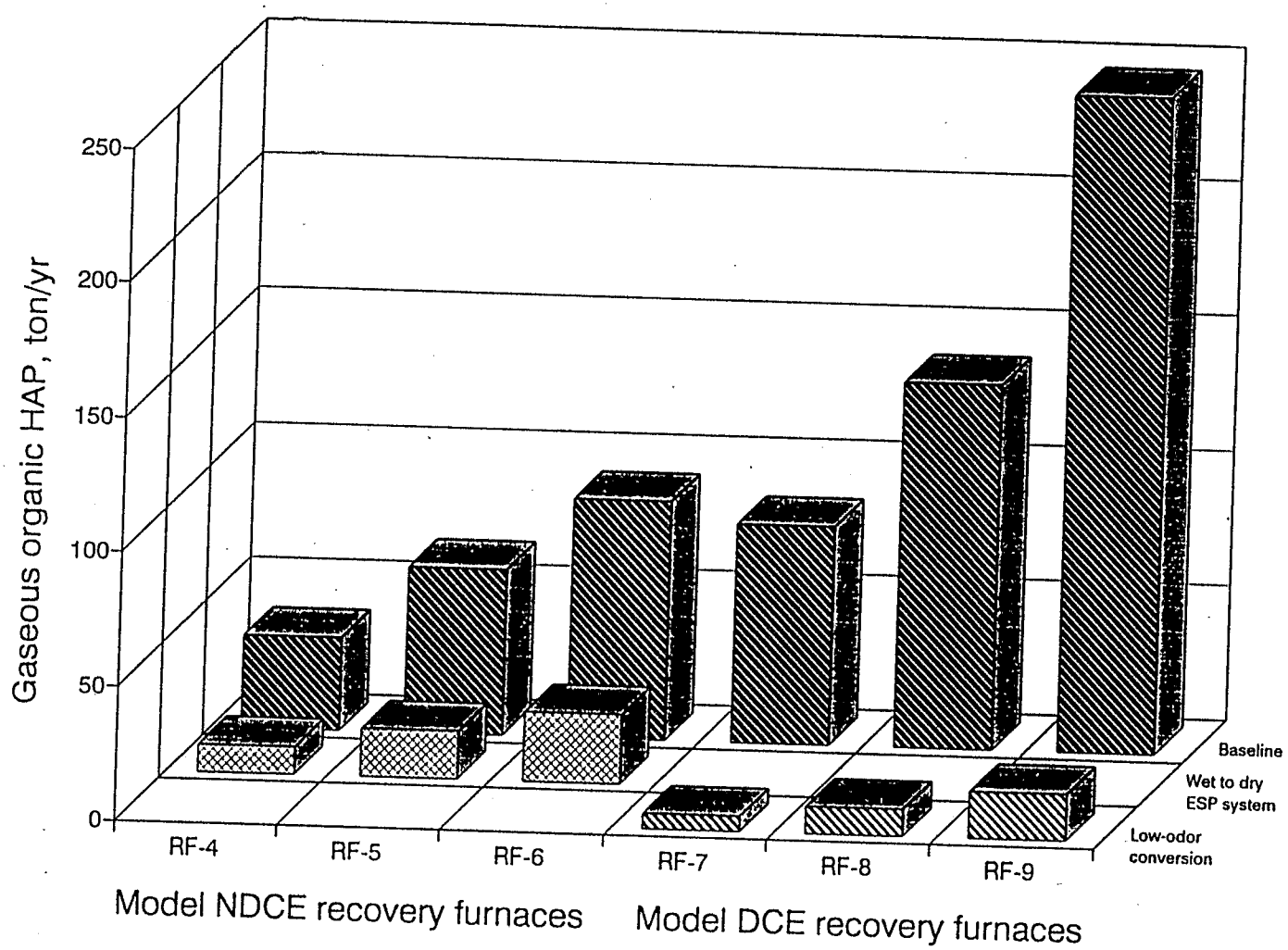

Figure 5-3. Gaseous organic HAP emissions from model recovery furnaces.

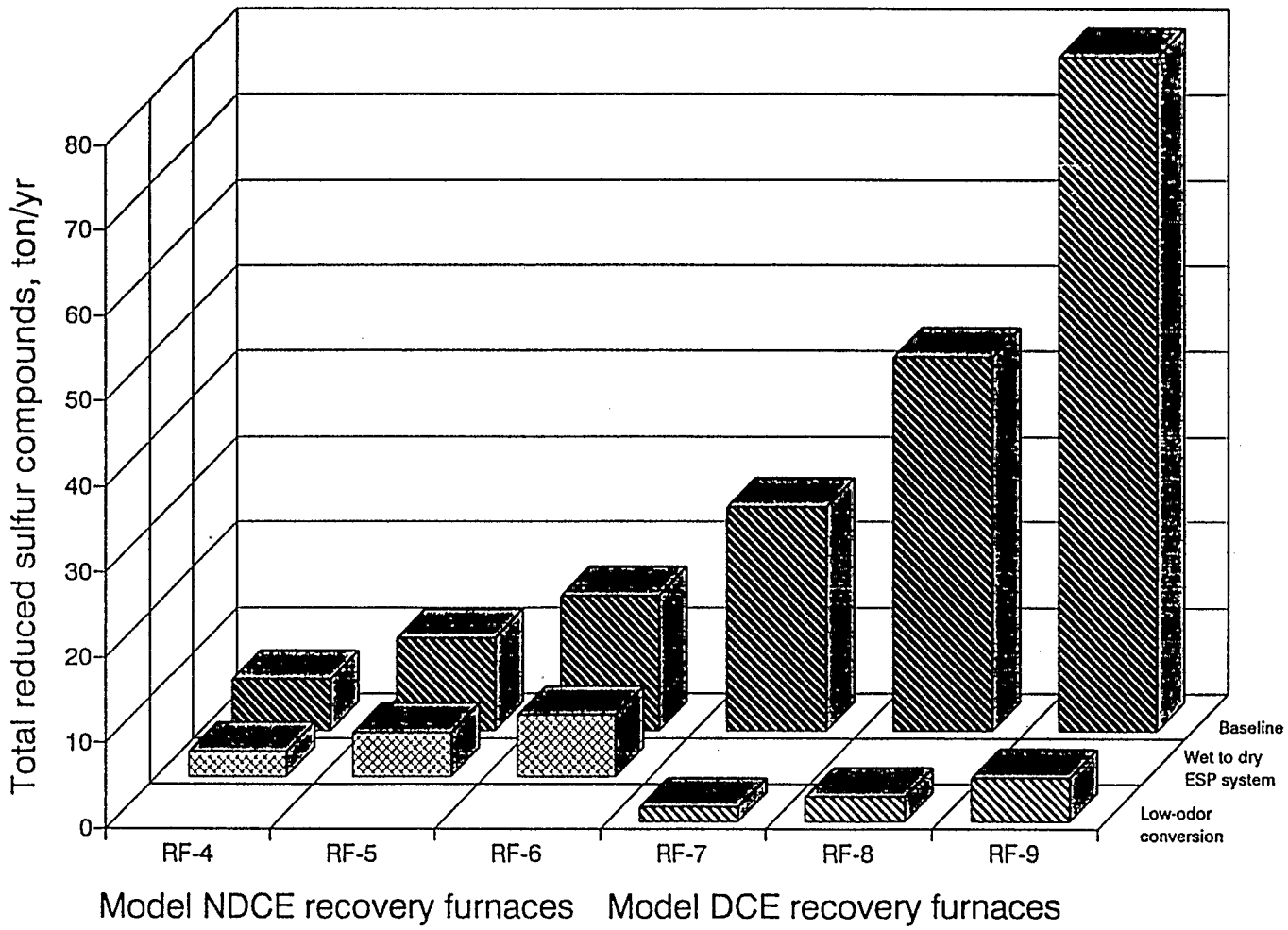

Figure 5-4. Total reduced sulfur emissions from model recovery furnace systems.

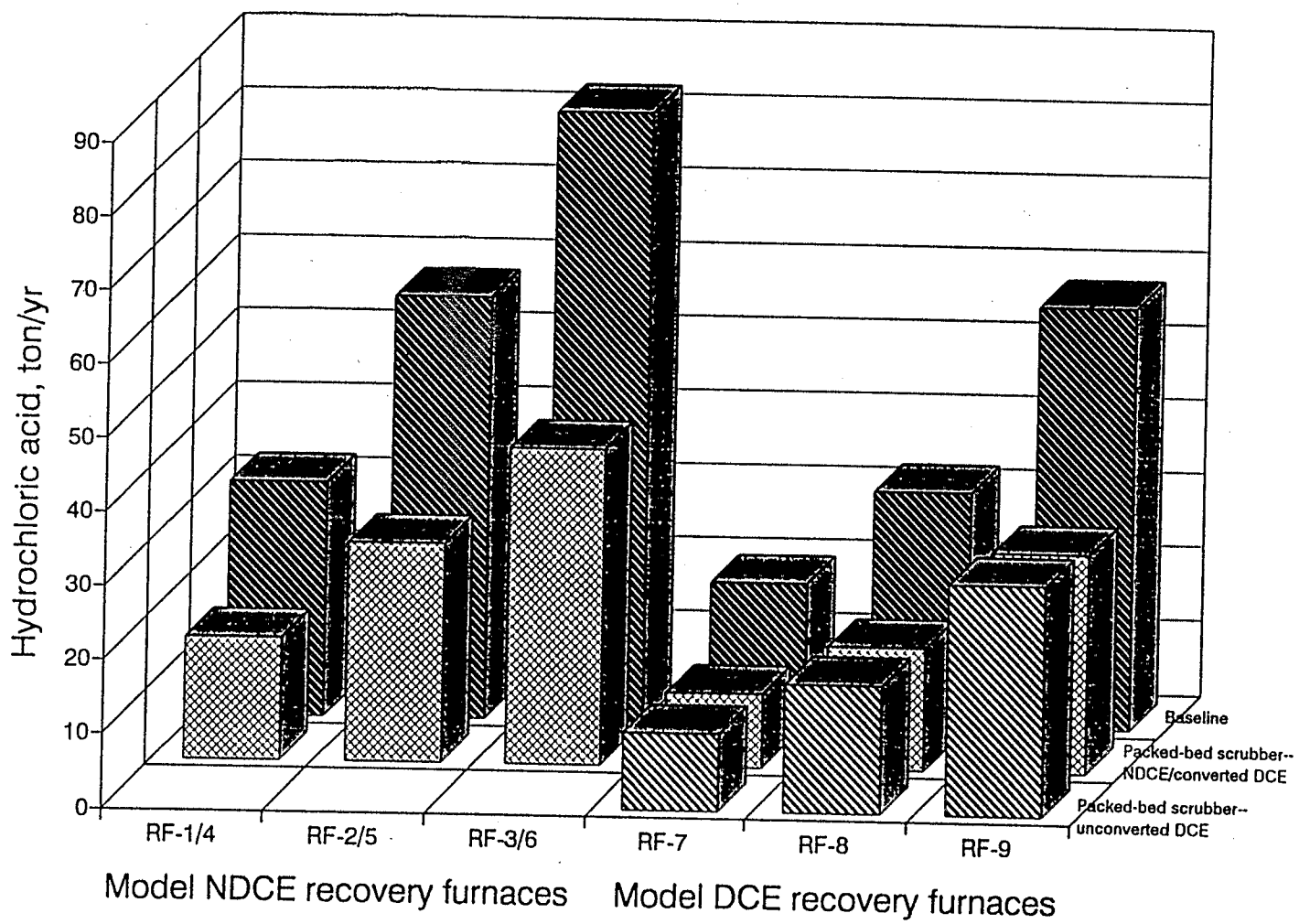

Figure 5-5. Hydrochloric acid emissions from model recovery furnaces.

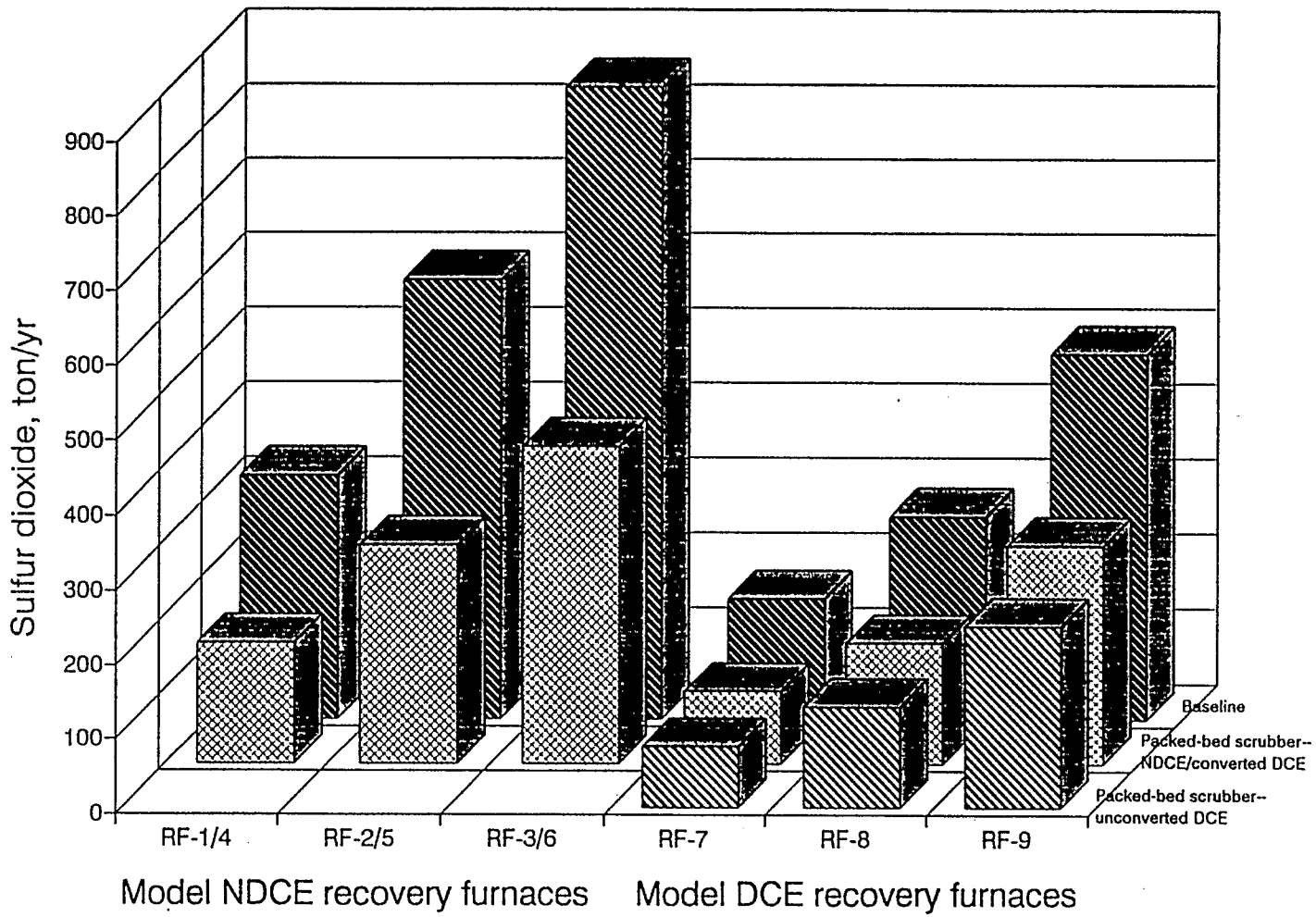

Figure 5-6. Sulfur dioxide emissions from model recovery furnaces.

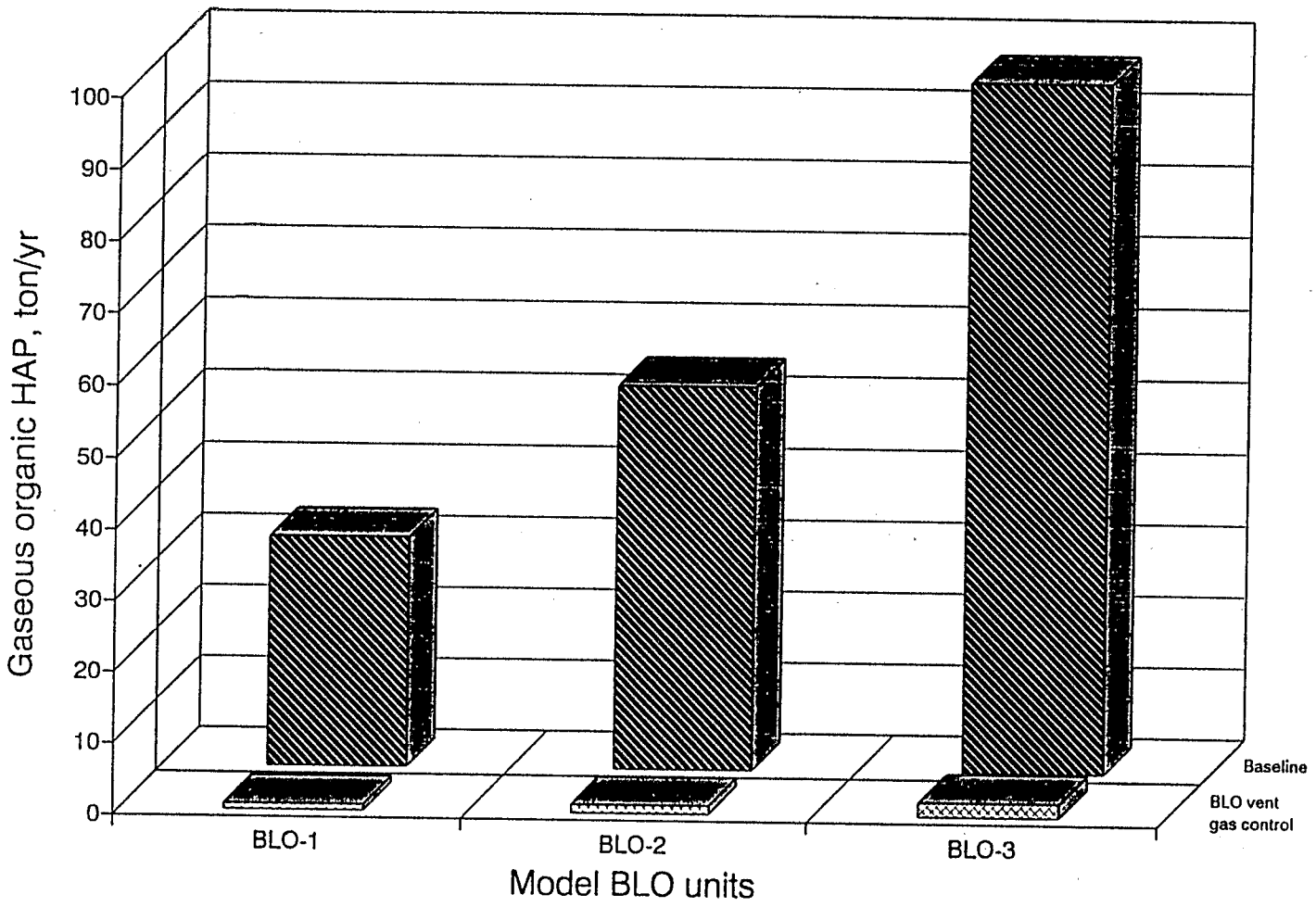

Figure 5-7. Gaseous organic HAP emissions from model BLO units.

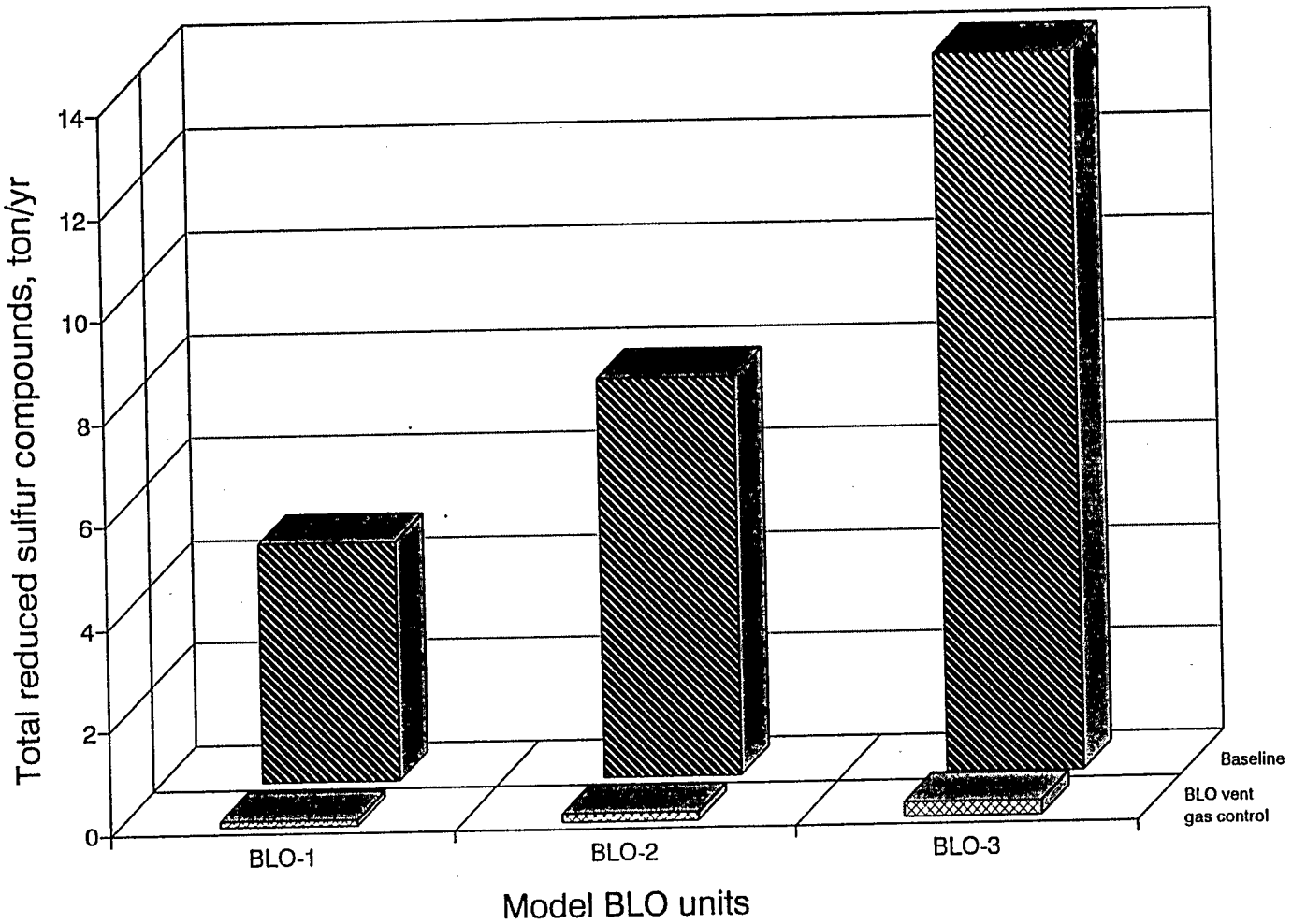

Figure 5-8. Total reduced sulfur emissions from model BLO units.

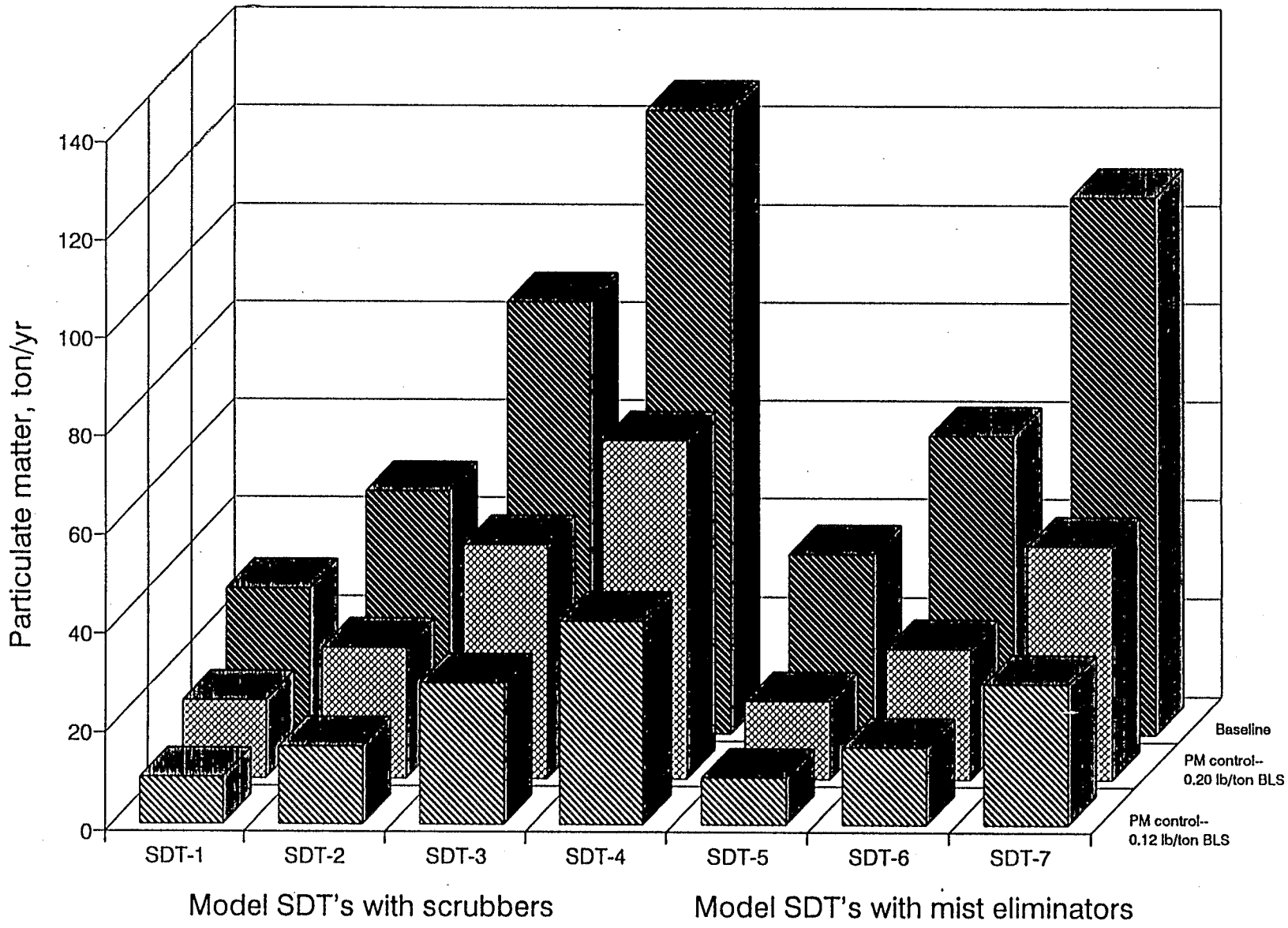

Figure 5-9. Particulate matter emissions from model smelt dissolving tanks.

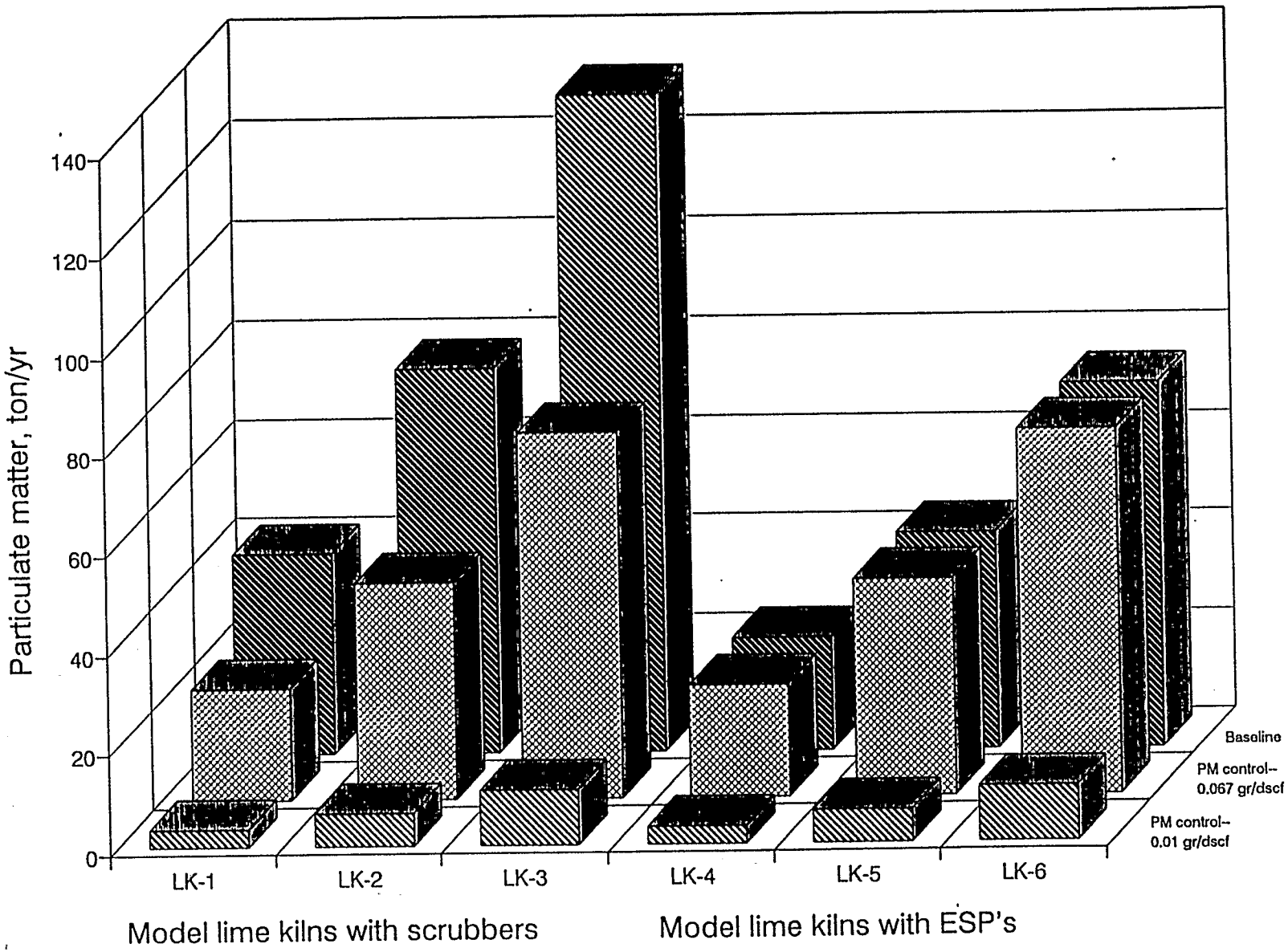

Figure 5-10. Particulate matter emissions from model lime kilns.

TABLE 5-1a (METRIC). MODEL NDCE RECOVERY FURNACE PARAMETERS[a]

Model recovery furnaces	Furnace type	Black liquor firing rate, kg BLS/d	Equivalent pulp production rate		Type of ESP system	Gas flow rate, m3/sec		Temperature, degrees C		Moisture content, %		ESP SCA, m2/(m3/sec)			Wastewater flow rate, L/min
			ADMUP/d	ADMBP/d		ESP exit	Quench outlet	ESP exit	Quench outlet	ESP exit	Quench outlet	Baseline	PM controls–0.10 g/dscm	PM controls–0.034 g/dscm	
RF-1a	NDCE	700,000	450	380	Dry	93.4	76	199	72	26	33	90	100	120	20.21
RF-1b	NDCE	700,000	450	380	Dry	93.4	76	199	72	26	33	100	100	120	20.21
RF-2a	NDCE	1,200,000	820	680	Dry	168	136	199	72	26	33	90	100	120	36.30
RF-2b	NDCE	1,200,000	820	680	Dry	168	136	199	72	26	33	100	100	120	36.30
RF-3a	NDCE	1,800,000	1,180	1,000	Dry	243	197	199	72	26	33	90	100	120	52.43
RF-3b	NDCE	1,800,000	1,180	1,000	Dry	243	197	199	72	26	33	100	100	120	52.43
RF-4a	NDCE	700,000	450	380	Wet	93.4	76	199	72	26	33	90	100	120	20.21
RF-4b	NDCE	700,000	450	380	Wet	93.4	76	199	72	26	33	100	100	120	20.21
RF-5a	NDCE	1,200,000	820	680	Wet	168	136	199	72	26	33	90	100	120	36.30
RF-5b	NDCE	1,200,000	820	680	Wet	168	136	199	72	26	33	100	100	120	36.30
RF-6a	NDCE	1,800,000	1,180	1,000	Wet	243	197	199	72	26	33	90	100	120	52.43
RF-6b	NDCE	1,800,000	1,180	1,000	Wet	243	197	199	72	26	33	100	100	120	52.43

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 5-1b.

TABLE 5-1b (ENGLISH). MODEL NDCE RECOVERY FURNACE PARAMETERS[a]

Model recovery furnaces	Furnace type	Black liquor firing rate, lb BLS/d	Equivalent pulp production rate		Type of ESP system	Gas flow rate, acfm		Temperature, degrees F		Moisture content, %		ESP SCA, ft2/1,000 acfm			Wastewater flow rate, gal/min
			ADTUP/d	ADTBP/d		ESP exit	Quench outlet	ESP exit	Quench outlet	ESP exit	Quench outlet	Baseline	PM controls–0.044 gr/dscf	PM controls–0.015 gr/dscf	
RF-1a	NDCE	1,500,000	500	420	dry	198,000	161,000	390	161	26	33	430	530	620	5.34
RF-1b	NDCE	1,500,000	500	420	dry	198,000	161,000	390	161	26	33	530	530	620	5.34
RF-2a	NDCE	2,700,000	900	750	dry	357,000	289,000	390	161	26	33	430	530	620	9.59
RF-2b	NDCE	2,700,000	900	750	dry	357,000	289,000	390	161	26	33	530	530	620	9.59
RF-3a	NDCE	3,900,000	1,300	1,100	dry	515,000	418,000	390	161	26	33	430	530	620	13.85
RF-3b	NDCE	3,900,000	1,300	1,100	dry	515,000	418,000	390	161	26	33	530	530	620	13.85
RF-4a	NDCE	1,500,000	500	420	wet	198,000	161,000	390	161	26	33	430	530	620	5.34
RF-4b	NDCE	1,500,000	500	420	wet	198,000	161,000	390	161	26	33	530	530	620	5.34
RF-5a	NDCE	2,700,000	900	750	wet	357,000	289,000	390	161	26	33	430	530	620	9.59
RF-5b	NDCE	2,700,000	900	750	wet	357,000	289,000	390	161	26	33	530	530	620	9.59
RF-6a	NDCE	3,900,000	1,300	1,100	wet	515,000	418,000	390	161	26	33	430	530	620	13.85
RF-6b	NDCE	3,900,000	1,300	1,100	wet	515,000	418,000	390	161	26	33	530	530	620	13.85

TABLE 5-2a (METRIC). MODEL DCE RECOVERY FURNACE PARAMETERS[a]

Model recovery furnaces	Furnace type	Black liquor firing rate, kg BLS/d	Equivalent pulp production rate		Type of ESP system	Gas flow rate, m3/sec		Temperature, degrees C		Moisture content, %		ESP SCA, m2/(m3/sec)			Wastewater flow rate, L/min
			ADMUP/d	ADMBP/d		ESP exit	Quench outlet	ESP exit	Quench outlet	ESP exit	Quench outlet	Baseline	PM controls--0.10 g/dscm	PM controls--0.034 g/dscm	
RF-7a	DCE	400,000	270	230	wet	56.2	48	160	74	32	37	70	90	100	12.45
	converted DCE				dry	56.2	45	199	72	26	33		100	120	12.11
RF-7b	DCE	400,000	270	230	wet	56.2	48	160	74	32	37	90	90	100	12.45
	converted DCE				dry	56.2	45	199	72	26	33		100	120	12.11
RF-8a	DCE	700,000	450	380	wet	93.4	80	160	74	32	37	70	90	100	20.74
	converted DCE				dry	93.4	76	199	72	26	33		100	120	20.21
RF-8b	DCE	700,000	450	380	wet	93.4	80	160	74	32	37	90	90	100	20.74
	converted DCE				dry	93.4	76	199	72	26	33		100	120	20.21
RF-9a	DCE	1,200,000	820	680	wet	168	144	160	74	32	37	70	90	100	37.29
	converted DCE				dry	168	136	199	72	26	33		100	120	36.30
RF-9b	DCE	1,200,000	820	680	wet	168	144	160	74	32	37	90	90	100	37.29
	converted DCE				dry	168	136	199	72	26	33		100	120	36.30

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 5-2b.

TABLE 5-2b (ENGLISH). MODEL DCE RECOVERY FURNACE PARAMETERS[a]

Model recovery furnaces	Furnace type	Black liquor firing rate, lb BLS/d	Equivalent pulp production rate		Type of ESP system	Gas flow rate, acfm		Temperature, degrees F		Moisture content, %		ESP SCA, ft2/1,000 acfm			Wastewater flow rate, gal/min
			ADTUP/d	ADTBP/d		ESP exit	Quench outlet	ESP exit	Quench outlet	ESP exit	Quench outlet	Baseline	PM controls–0.044 gr/dscf	PM controls–0.015 gr/dscf	
RF-7a	DCE	900,000	300	250	wet	119,000	102,000	320	165	32	37	330	430	520	3.29
	converted DCE				dry	119,000	96,500	390	161	26	33		530	620	3.20
RF-7b	DCE	900,000	300	250	wet	119,000	102,000	320	165	32	37	430	430	520	3.29
	converted DCE				dry	119,000	96,500	390	161	26	33		530	620	3.20
RF-8a	DCE	1,500,000	500	420	wet	198,000	170,000	320	165	32	37	330	430	520	5.48
	converted DCE				dry	198,000	161,000	390	161	26	33		530	620	5.34
RF-8b	DCE	1,500,000	500	420	wet	198,000	170,000	320	165	32	37	430	430	520	5.48
	converted DCE				dry	198,000	161,000	390	161	26	33		530	620	5.34
RF-9a	DCE	2,700,000	900	750	wet	357,000	306,000	320	165	32	37	330	430	520	9.85
	converted DCE				dry	357,000	289,000	390	161	26	33		530	620	9.59
RF-9b	DCE	2,700,000	900	750	wet	357,000	306,000	320	165	32	37	430	430	520	9.85
	converted DCE				dry	357,000	289,000	390	161	26	33		530	620	9.59

TABLE 5-3a (METRIC). MODEL NDCE RECOVERY FURNACE CONCENTRATIONS AND EMISSION FACTORS[a]

Model recovery furnace	Furnace type	Type of ESP system	Baseline PM, g/dscm	Control level PM, g/dscm		Baseline gaseous organic HAP's, kg/kg BLS	Control level gaseous organic HAP's, kg/kg BLS	Baseline acid gases, kg/kg BLS		Control level acid gases, kg/kg BLS	
				PM controls--0.10 g/dscm	PM controls--0.034 g/dscm			HCl	SO2	HCl	SO2
RF-1a	NDCE	dry	0.27	0.1	0.034	3.67E-05	3.67E-05	1.20E-04	1.24E-03	6.21E-05	6.18E-04
RF-1b	NDCE	dry	0.1	0.1	0.034	3.67E-05	3.67E-05	1.20E-04	1.24E-03	6.21E-05	6.18E-04
RF-2a	NDCE	dry	0.27	0.1	0.034	3.67E-05	3.67E-05	1.20E-04	1.24E-03	6.21E-05	6.18E-04
RF-2b	NDCE	dry	0.1	0.1	0.034	3.67E-05	3.67E-05	1.20E-04	1.24E-03	6.21E-05	6.18E-04
RF-3a	NDCE	dry	0.27	0.1	0.034	3.67E-05	3.67E-05	1.20E-04	1.24E-03	6.21E-05	6.18E-04
RF-3b	NDCE	dry	0.1	0.1	0.034	3.67E-05	3.67E-05	1.20E-04	1.24E-03	6.21E-05	6.18E-04
RF-4a	NDCE	wet	0.27	0.1	0.034	1.30E-04	3.67E-05	1.20E-04	1.24E-03	6.21E-05	6.18E-04
RF-4b	NDCE	wet	0.1	0.1	0.034	1.30E-04	3.67E-05	1.20E-04	1.24E-03	6.21E-05	6.18E-04
RF-5a	NDCE	wet	0.27	0.1	0.034	1.30E-04	3.67E-05	1.20E-04	1.24E-03	6.21E-05	6.18E-04
RF-5b	NDCE	wet	0.1	0.1	0.034	1.30E-04	3.67E-05	1.20E-04	1.24E-03	6.21E-05	6.18E-04
RF-6a	NDCE	wet	0.27	0.1	0.034	1.30E-04	3.67E-05	1.20E-04	1.24E-03	6.21E-05	6.18E-04
RF-6b	NDCE	wet	0.1	0.1	0.034	1.30E-04	3.67E-05	1.20E-04	1.24E-03	6.21E-05	6.18E-04

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 5-3b. Refer to Table 5-3b for footnotes.

TABLE 5-3b (ENGLISH). MODEL NDCE RECOVERY FURNACE CONCENTRATIONS AND EMISSION FACTORS

Model recovery furnace	Furnace type	Type of ESP system	Baseline PM, gr/dscf	Control level PM, gr/dscf		Baseline gaseous organic HAP's, lb/lb BLS (c)	Control level gaseous organic HAP's, lb/lb BLS (c)	Baseline acid gases, lb/lb BLS (d)		Control level acid gases, lb/lb BLS (e)	
				PM controls--0.044 gr/dscf	PM controls--0.015 gr/dscf			HCl	SO2	HCl	SO2
RF-1a	NDCE	dry	0.12	0.044	0.015	3.67E-05	3.67E-05	1.20E-04	1.24E-03	6.21E-05	6.18E-04
RF-1b	NDCE	dry	0.044	0.044	0.015	3.67E-05	3.67E-05	1.20E-04	1.24E-03	6.21E-05	6.18E-04
RF-2a	NDCE	dry	0.12	0.044	0.015	3.67E-05	3.67E-05	1.20E-04	1.24E-03	6.21E-05	6.18E-04
RF-2b	NDCE	dry	0.044	0.044	0.015	3.67E-05	3.67E-05	1.20E-04	1.24E-03	6.21E-05	6.18E-04
RF-3a	NDCE	dry	0.12	0.044	0.015	3.67E-05	3.67E-05	1.20E-04	1.24E-03	6.21E-05	6.18E-04
RF-3b	NDCE	dry	0.044	0.044	0.015	3.67E-05	3.67E-05	1.20E-04	1.24E-03	6.21E-05	6.18E-04
RF-4a	NDCE	wet	0.12	0.044	0.015	1.30E-04	3.67E-05	1.20E-04	1.24E-03	6.21E-05	6.18E-04
RF-4b	NDCE	wet	0.044	0.044	0.015	1.30E-04	3.67E-05	1.20E-04	1.24E-03	6.21E-05	6.18E-04
RF-5a	NDCE	wet	0.12	0.044	0.015	1.30E-04	3.67E-05	1.20E-04	1.24E-03	6.21E-05	6.18E-04
RF-5b	NDCE	wet	0.044	0.044	0.015	1.30E-04	3.67E-05	1.20E-04	1.24E-03	6.21E-05	6.18E-04
RF-6a	NDCE	wet	0.12	0.044	0.015	1.30E-04	3.67E-05	1.20E-04	1.24E-03	6.21E-05	6.18E-04
RF-6b	NDCE	wet	0.044	0.044	0.015	1.30E-04	3.67E-05	1.20E-04	1.24E-03	6.21E-05	6.18E-04

(a) For furnaces controlling PM emissions to the 0.044 gr/dscf, PM controls are an ESP upgrade or replacement.
(b) For furnaces controlling PM emissions to 0.015 gr/dscf, PM controls are an ESP upgrade or replacement plus addition of a packed-bed scrubber.
(c) The baseline emission factor for models RF-1 through RF-3 is the average emission factor for NDCE recovery furnaces with dry ESP systems.
The baseline emission factor for models RF-4 through RF-6 is the average emission factor for NDCE recovery furnaces with wet ESP systems.
The control level emission factor for models RF-1 through RF-6 is the average emission factor for NDCE recovery furnaces with dry ESP systems.
Gaseous organic HAP's include acetaldehyde, benzene, formaldehyde, methanol, methyl ethyl ketone, methyl isobutyl ketone, phenol, toluene, and xylenes. Methanol comprises 39 percent of the uncontrolled emissions and 13 percent of the controlled emissions.
(d) The baseline HCl emission factor is an average of HCl emission data from NDCE and DCE recovery furnaces.
The baseline SO2 emission factor is from an average SO2 emission factor of 4.2 lb/ADTP for NDCE recovery furnaces, assuming 3,400 lb BLS/ADTP.
(e) The control level HCl emission factor was derived from a guaranteed 5 ppm scrubber outlet concentration.
The control level SO2 emission factor was derived assuming 50 percent control of SO2 emissions with a packed-bed scrubber.

TABLE 5-4a (METRIC). MODEL DCE RECOVERY FURNACE CONCENTRATIONS AND EMISSION FACTORS[a]

Model recovery furnace	Furnace type	Type of ESP system	Baseline PM, g/dscm	Control level PM, g/dscm		Baseline gaseous organic HAP's, kg/kg BLS	Control level gaseous organic HAP's, kg/kg BLS	Baseline acid gases, kg/kg BLS		Control level acid gases, kg/kg BLS	
				PM controls--0.10 g/dscm	PM controls--0.034 g/dscm			HCl	SO2	HCl	SO2
RF-7a	DCE converted DCE	wet dry	0.18	0.1	0.034	5.15E-04	3.67E-05	1.20E-04	1.03E-03	6.53E-05 6.21E-05	5.15E-04 6.18E-04
RF-7b	DCE converted DCE	wet dry	0.1	0.1	0.034	5.15E-04	3.67E-05	1.20E-04	1.03E-03	6.53E-05 6.21E-05	5.15E-04 6.18E-04
RF-8a	DCE converted DCE	wet dry	0.18	0.1	0.034	5.15E-04	3.67E-05	1.20E-04	1.03E-03	6.53E-05 6.21E-05	5.15E-04 6.18E-04
RF-8b	DCE converted DCE	wet dry	0.1	0.1	0.034	5.15E-04	3.67E-05	1.20E-04	1.03E-03	6.53E-05 6.21E-05	5.15E-04 6.18E-04
RF-9a	DCE converted DCE	wet dry	0.18	0.1	0.034	5.15E-04	3.67E-05	1.20E-04	1.03E-03	6.53E-05 6.21E-05	5.15E-04 6.18E-04
RF-9b	DCE converted DCE	wet dry	0.1	0.1	0.034	5.15E-04	3.67E-05	1.20E-04	1.03E-03	6.53E-05 6.21E-05	5.15E-04 6.18E-04

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 5-4b. Refer to Table 5-4b for footnotes.

TABLE 5-4b (ENGLISH). MODEL DCE RECOVERY FURNACE CONCENTRATIONS AND EMISSION FACTORS

Model recovery furnace	Furnace type	Type of ESP system	Baseline PM, gr/dscf	Control level PM, gr/dscf		Baseline gaseous organic HAP's, lb/lb BLS (c)	Control level gaseous organic HAP's, lb/lb BLS (c)	Baseline acid gases, lb/lb BLS (d)		Control level acid gases, lb/lb BLS (e)	
				PM controls--0.044 gr/dscf	PM controls--0.015 gr/dscf			HCl	SO2	HCl	SO2
RF-7a	DCE converted DCE	wet dry	0.08	0.044	0.015	5.15E-04	3.67E-05	1.20E-04	1.03E-03	6.53E-05 6.21E-05	5.15E-04 6.18E-04
RF-7b	DCE converted DCE	wet dry	0.044	0.044	0.015	5.15E-04	3.67E-05	1.20E-04	1.03E-03	6.53E-05 6.21E-05	5.15E-04 6.18E-04
RF-8a	DCE converted DCE	wet dry	0.08	0.044	0.015	5.15E-04	3.67E-05	1.20E-04	1.03E-03	6.53E-05 6.21E-05	5.15E-04 6.18E-04
RF-8b	DCE converted DCE	wet dry	0.044	0.044	0.015	5.15E-04	3.67E-05	1.20E-04	1.03E-03	6.53E-05 6.21E-05	5.15E-04 6.18E-04
RF-9a	DCE converted DCE	wet dry	0.08	0.044	0.015	5.15E-04	3.67E-05	1.20E-04	1.03E-03	6.53E-05 6.21E-05	5.15E-04 6.18E-04
RF-9b	DCE converted DCE	wet dry	0.044	0.044	0.015	5.15E-04	3.67E-05	1.20E-04	1.03E-03	6.53E-05 6.21E-05	5.15E-04 6.18E-04

(a) For furnaces controlling PM emissions to 0.044 gr/dscf, PM controls are an ESP upgrade or replacement.
(b) For furnaces controlling PM emissions to 0.015 gr/dscf, PM controls are an ESP upgrade or replacement plus addition of a packed-bed scrubber.
(c) The baseline emission factor is the sum of the average emission factors for DCE recovery furnaces and BLO units.
The control level emission factor is the average emission factor for NDCE recovery furnaces with dry ESP systems.
Gaseous organic HAP's include acetaldehyde, benzene, formaldehyde, methanol, methyl ethyl ketone, methyl isobutyl ketone, phenol, toluene, and xylenes. Methanol comprises 67 percent of the uncontrolled emissions and 13 percent of the controlled emissions.
(d) The baseline HCl emission factor is an average of HCl emission data from NDCE and DCE recovery furnaces.
The baseline SO2 emission factor is from an average SO2 emission factor of 3.5 lb/ADTP for DCE recovery furnaces, assuming 3,400 lb BLS/ADTP.
(e) The control level HCl emission factor was derived from a guaranteed 5 ppm scrubber outlet concentration.
The control level SO2 emission factor was derived assuming 50 percent control of SO2 emissions with a packed-bed scrubber.

TABLE 5-5a (METRIC). PRIMARY PM AND PM HAP EMISSIONS FOR MODEL RECOVERY FURNACES[a]

Model recovery furnaces	Baseline emissions, Mg/yr		Control option	Control level emissions, Mg/yr		Emission reduction, Mg/yr		Emission reduction, %
	PM	PM HAP's		PM	PM HAP's	PM	PM HAP's	
NDCE								
RF-1a/4a	358	0.7	PM controls--0.10 g/dscm (b)	131	0.3	227	0.5	63
			PM controls--0.034 g/dscm (c)	45	0.09	313	0.6	88
RF-1b/4b	131	0.3	PM controls--0.034 g/dscm (c)	45	0.09	86	0.2	66
RF-2a/5a	645	1.3	PM controls--0.10 g/dscm (b)	236	0.5	408	0.8	63
			PM controls--0.034 g/dscm (c)	81	0.2	564	1.1	88
RF-2b/5b	236	0.5	PM controls--0.034 g/dscm (c)	81	0.2	156	0.3	66
RF-3a/6a	934	1.9	PM controls--0.10 g/dscm (b)	341	0.7	593	1.2	63
			PM controls--0.034 g/dscm (c)	116	0.2	818	1.6	88
RF-3b/6b	341	0.7	PM controls--0.034 g/dscm (c)	116	0.2	225	0.4	66
DCE								
RF-7a	144	0.3	PM controls--0.10 g/dscm (b)	79	0.2	65	0.1	45
			PM controls--0.034 g/dscm (c)	27	0.05	117	0.2	81
RF-7b	79	0.2	PM controls--0.034 g/dscm (c)	27	0.05	52	0.1	66
RF-8a	239	0.5	PM controls--0.10 g/dscm (b)	131	0.3	108	0.2	45
			PM controls--0.034 g/dscm (c)	45	0.09	194	0.4	81
RF-8b	131	0.3	PM controls--0.034 g/dscm (c)	45	0.09	87	0.2	66
RF-9a	431	0.9	PM controls--0.10 g/dscm (b)	236	0.5	194	0.4	45
			PM controls--0.034 g/dscm (c)	81	0.2	350	0.7	81
RF-9b	237	0.5	PM controls--0.034 g/dscm (c)	81	0.2	156	0.3	66

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 5-5b. Refer to Table 5-5b for footnotes, which include calculations.
(b) PM controls are an ESP upgrade or replacement.
(c) PM controls are an ESP upgrade or replacement plus a packed-bed scrubber.

TABLE 5-5b (ENGLISH). PRIMARY PM AND PM HAP EMISSIONS FOR MODEL RECOVERY FURNACES[a]

Model recovery furnaces	Baseline emissions, ton/yr		Control option	Control level emissions, ton/yr		Emission reduction, ton/yr		Emission reduction, %
	PM	PM HAP's		PM	PM HAP's	PM	PM HAP's	
NDCE								
RF-1a/4a	394	0.8	PM controls--0.044 gr/dscf (b)	145	0.3	250	0.5	63
			PM controls--0.015 gr/dscf (c)	49	0.1	345	0.7	88
RF-1b/4b	145	0.3	PM controls--0.015 gr/dscf (c)	49	0.1	95	0.2	66
RF-2a/5a	711	1.4	PM controls--0.044 gr/dscf (b)	261	0.5	450	0.9	63
			PM controls--0.015 gr/dscf (c)	89	0.2	622	1.2	88
RF-2b/5b	261	0.5	PM controls--0.015 gr/dscf (c)	89	0.2	172	0.3	66
RF-3a/6a	1,030	2.1	PM controls--0.044 gr/dscf (b)	376	0.8	654	1.3	63
			PM controls--0.015 gr/dscf (c)	128	0.3	902	1.8	88
RF-3b/6b	376	0.8	PM controls--0.015 gr/dscf (c)	128	0.3	248	0.5	66
DCE								
RF-7a	158	0.3	PM controls--0.044 gr/dscf (b)	87	0.2	71	0.1	45
			PM controls--0.015 gr/dscf (c)	30	0.06	129	0.3	81
RF-7b	87	0.2	PM controls--0.015 gr/dscf (c)	30	0.06	57	0.1	66
RF-8a	263	0.5	PM controls--0.044 gr/dscf (b)	145	0.3	119	0.2	45
			PM controls--0.015 gr/dscf (c)	49	0.1	214	0.4	81
RF-8b	145	0.3	PM controls--0.015 gr/dscf (c)	49	0.1	95	0.2	66
RF-9a	475	0.9	PM controls--0.044 gr/dscf (b)	261	0.5	214	0.4	45
			PM controls--0.015 gr/dscf (c)	89	0.2	386	0.8	81
RF-9b	261	0.5	PM controls--0.015 gr/dscf (c)	89	0.2	172	0.3	66

(a) PM emissions (ton/yr) = PM concentration (gr/dscf) x lb/7,000 gr x model gas flow rate (acfm) x (528R/[model temperature + 460F]) x (100% - model %H2O)/100% x 60 min/hr x 8,424 hr/yr x 1 ton/2,000 lb.
PM HAP emissions (ton/yr) = 0.2 percent of PM emissions.
(b) PM controls are an ESP upgrade or replacement.
(c) PM controls are an ESP upgrade or replacement plus a packed-bed scrubber.

TABLE 5-6a (METRIC). SECONDARY EMISSIONS FOR PM CONTROL FOR MODEL RECOVERY FURNACES[a]

Model recovery furnaces	Baseline emissions, kg/yr				Control option	Incremental emissions, kg/yr			
	PM	SO2	NOx	CO		PM	SO2	NOx	CO
NDCE									
RF-1a/4a	395	1,920	767	1,450	PM controls--0.10 g/dscm (b)	76	367	145	272
					PM controls--0.034 g/dscm (c)	267	1,300	517	980
RF-1b/4b	472	2,290	912	1,720	PM controls--0.034 g/dscm (c)	191	934	372	708
RF-2a/5a	712	3,470	1,380	2,610	PM controls--0.10 g/dscm (b)	136	662	263	499
					PM controls--0.034 g/dscm (c)	476	2,340	921	1,750
RF-2b/5b	848	4,130	1,640	3,110	PM controls--0.034 g/dscm (c)	340	1,670	658	1,250
RF-3a/6a	1,030	4,990	1,990	3,770	PM controls--0.10 g/dscm (b)	195	953	381	717
					PM controls--0.034 g/dscm (c)	694	3,400	1,340	2,530
RF-3b/6b	1,220	5,940	2,370	4,490	PM controls--0.034 g/dscm (c)	499	2,450	962	1,810
DCE									
RF-7a	193	940	372	708	PM controls--0.10 g/dscm (b)	45	218	86	163
					PM controls--0.034 g/dscm (c)	189	916	363	689
RF-7b	238	1,160	458	871	PM controls--0.034 g/dscm (c)	143	699	277	526
RF-8a	320	1,560	621	1,170	PM controls--0.10 g/dscm (b)	75	363	145	277
					PM controls--0.034 g/dscm (c)	315	1,530	608	1,160
RF-8b	395	1,920	767	1,450	PM controls--0.034 g/dscm (c)	240	1,170	463	880
RF-9a	576	2,810	1,120	2,120	PM controls--0.10 g/dscm (b)	136	658	263	494
					PM controls--0.034 g/dscm (c)	567	2,770	1,090	2,070
RF-9b	712	3,470	1,380	2,610	PM controls--0.034 g/dscm (c)	431	2,110	830	1,580

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 5-6b. See Table 5-6b for footnotes.
(b) Impacts were estimated for ESP upgrade/replacement.
(c) Impacts were estimated for ESP upgrade plus packed-bed scrubber.

TABLE 5-6b (ENGLISH). SECONDARY EMISSIONS FOR PM CONTROL FOR MODEL RECOVERY FURNACES[a]

Model recovery furnaces	Baseline emissions, lb/yr				Control option	Incremental emissions, lb/yr			
	PM	SO2	NOx	CO		PM	SO2	NOx	CO
NDCE									
RF-1a/4a	872	4,240	1,690	3,200	PM controls--0.044 gr/dscf (b)	168	810	320	600
					PM controls--0.015 gr/dscf (c)	588	2,870	1,140	2,160
RF-1b/4b	1,040	5,050	2,010	3,800	PM controls--0.015 gr/dscf (c)	420	2,060	820	1,560
RF-2a/5a	1,570	7,650	3,040	5,760	PM controls--0.044 gr/dscf (b)	300	1,460	580	1,100
					PM controls--0.015 gr/dscf (c)	1,050	5,150	2,030	3,860
RF-2b/5b	1,870	9,110	3,620	6,860	PM controls--0.015 gr/dscf (c)	750	3,690	1,450	2,760
RF-3a/6a	2,270	11,000	4,380	8,320	PM controls--0.044 gr/dscf (b)	430	2,100	840	1,580
					PM controls--0.015 gr/dscf (c)	1,530	7,500	2,960	5,580
RF-3b/6b	2,700	13,100	5,220	9,900	PM controls--0.015 gr/dscf (c)	1,100	5,400	2,120	4,000
DCE									
RF-7a	424	2,070	821	1,560	PM controls--0.044 gr/dscf (b)	100	480	189	360
					PM controls--0.015 gr/dscf (c)	416	2,020	799	1,520
RF-7b	524	2,550	1,010	1,920	PM controls--0.015 gr/dscf (c)	316	1,540	610	1,160
RF-8a	706	3,440	1,370	2,590	PM controls--0.044 gr/dscf (b)	166	800	320	610
					PM controls--0.015 gr/dscf (c)	694	3,380	1,340	2,550
RF-8b	872	4,240	1,690	3,200	PM controls--0.015 gr/dscf (c)	528	2,580	1,020	1,940
RF-9a	1,270	6,200	2,460	4,670	PM controls--0.044 gr/dscf (b)	300	1,450	580	1,090
					PM controls--0.015 gr/dscf (c)	1,250	6,100	2,410	4,570
RF-9b	1,570	7,650	3,040	5,760	PM controls--0.015 gr/dscf (c)	950	4,650	1,830	3,480

(a) Baseline and control level secondary emissions were estimated based on electricity requirements and emission factors. Calculations for energy impacts are presented in Table 5-7. The emission factors = 0.15 lb PM/MM Btu; 0.73 lb SO2/MM Btu; 0.29 lb NOx/MM Btu; and 0.55 lb CO/MM Btu. Numbers in parentheses represent negative values, indicating that secondary emissions are reduced by that amount.

(b) Impacts were estimated for ESP upgrade/replacement.

(c) Impacts were estimated for ESP upgrade plus packed-bed scrubber.

TABLE 5-7. ENERGY IMPACTS FOR PM CONTROL FOR MODEL RECOVERY FURNACES[a]

Model recovery furnaces	Baseline energy impacts, MWh/yr	Control option	Control level energy impacts, MWh/yr	Incremental energy impacts, MWh/yr (c)
NDCE				
RF-1a/4a	1,700	PM controls–0.10 g/dscm (0.044 gr/dscf) (b)	2,030	330
		PM controls–0.034 g/dscm (0.015 gr/dscf) (c)	2,850	1,150
RF-1b/4b	2,030	PM controls–0.034 g/dscm (0.015 gr/dscf) (c)	2,850	820
RF-2a/5a	3,070	PM controls–0.10 g/dscm (0.044 gr/dscf) (b)	3,650	580
		PM controls–0.034 g/dscm (0.015 gr/dscf) (c)	5,120	2,050
RF-2b/5b	3,650	PM controls–0.034 g/dscm (0.015 gr/dscf) (c)	5,120	1,470
RF-3a/6a	4,430	PM controls–0.10 g/dscm (0.044 gr/dscf) (b)	5,270	840
		PM controls–0.034 g/dscm (0.015 gr/dscf) (c)	7,410	2,980
RF-3b/6b	5,270	PM controls–0.034 g/dscm (0.015 gr/dscf) (c)	7,410	2,140
DCE				
RF-7a	829	PM controls–0.10 g/dscm (0.044 gr/dscf) (b)	1,020	191
		PM controls–0.034 g/dscm (0.015 gr/dscf) (c)	1,640	811
RF-7b	1,020	PM controls–0.034 g/dscm (0.015 gr/dscf) (c)	1,640	620
RF-8a	1,380	PM controls–0.10 g/dscm (0.044 gr/dscf) (b)	1,700	320
		PM controls–0.034 g/dscm (0.015 gr/dscf) (c)	2,730	1,350
RF-8b	1,700	PM controls–0.034 g/dscm (0.015 gr/dscf) (c)	2,730	1,030
RF-9a	2,490	PM controls–0.10 g/dscm (0.044 gr/dscf) (b)	3,070	580
		PM controls–0.034 g/dscm (0.015 gr/dscf) (c)	4,920	2,430
RF-9b	3,070	PM controls–0.034 g/dscm (0.015 gr/dscf) (c)	4,920	1,850

(a) Numbers in parentheses represent negative values, indicating that energy impacts are reduced by that amount. PM control energy impacts at baseline and at 0.044 gr/dscf = [0.00018 x model gas flowrate x 1 in. H2O pressure drop x 8,424 hr/yr x 1 MWh/1,000 kWh] + [0.00194 x model gas flow rate x SCA x 8,424 hr/yr x 1 MWh/1,000 kWh]. PM control energy impacts at 0.015 gr/dscf = [0.00018 x model gas flow rate x 4 in. H2O pressure drop x 8,424 hr/yr x 1 MWh/1,000 kWh] + [0.00194 x model gas flow rate x SCA x 8,424 hr/yr x 1 MWh/1,000 kWh] + [0.746 kW/hp x liquid flow rate x 60 ft head x spec. grav. H2O x 1/70% pump motor effici. x 8,424 hr/yr x 1 MWh/1,000 kWh].

(b) Impacts were estimated for ESP upgrade/replacement.

(c) Impacts were estimated for ESP upgrade/replacement plus packed-bed scrubber.

TABLE 5-8a (METRIC). PRIMARY GASEOUS ORGANIC HAP EMISSIONS FOR MODEL RECOVERY FURNACES[a]

Model recovery furnaces	Baseline emissions, Mg/yr	Control option	Control level emissions, Mg/yr	Emission reduction, Mg/yr	Emission reduction, %
NDCE					
RF-4	31	Wet to dry ESP system conversion	8.8	22	72
RF-5	56	Wet to dry ESP system conversion	16	40	72
RF-6	81	Wet to dry ESP system conversion	23	58	72
DCE					
RF-7	74	Low-odor conversion	5.3	69	93
RF-8	123	Low-odor conversion	8.8	114	93
RF-9	221	Low-odor conversion	16	206	93

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 5-8b. Refer to Table 5-8b for footnotes, which include calculations.

TABLE 5-8b (ENGLISH). PRIMARY GASEOUS ORGANIC HAP EMISSIONS FOR MODEL RECOVERY FURNACES[a]

Model recovery furnaces	Baseline emissions, ton/yr	Control option	Control level emissions, ton/yr	Emission reduction, ton/yr	Emission reduction, %
NDCE					
RF-4	34	Wet to dry ESP system conversion	9.7	25	72
RF-5	62	Wet to dry ESP system conversion	17	44	72
RF-6	89	Wet to dry ESP system conversion	25	64	72
DCE					
RF-7	81	Low-odor conversion	5.8	76	93
RF-8	136	Low-odor conversion	9.7	126	93
RF-9	244	Low-odor conversion	17	227	93

(a) Gaseous organic HAP emissions (ton/yr) = emission factor (lb/lb BLS) x model BLS firing rate (lb BLS/d) x 351 d/yr x 1 ton/2,000 lb. Gaseous organic HAP's include acetaldehyde, benzene, formaldehyde, methanol, methyl ethyl ketone, methyl isobutyl ketone, phenol, toluene, and xylenes. Methanol comprises 49 percent of the emission reduction for models RF-4 through RF-6 and 70 percent of the emission reduction for models RF-7 through RF-9.

TABLE 5-9a (METRIC). TOTAL REDUCED SULFUR COMPOUND EMISSIONS FOR MODEL RECOVERY FURNACES[a]

Model recovery furnaces	Baseline emissions, Mg/yr	Control option	Control level emissions, Mg/yr	Emission reduction, Mg/yr	Emission reduction, %
NDCE					
RF-4	5.5	Wet to dry ESP system conversion	2.5	3.0	55
RF-5	9.9	Wet to dry ESP system conversion	4.5	5.4	55
RF-6	14	Wet to dry ESP system conversion	6.5	7.8	55
DCE					
RF-7	24	Low-odor conversion	1.5	22	94
RF-8	40	Low-odor conversion	2.5	37	94
RF-9	71	Low-odor conversion	4.5	67	94

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 5-9b. Refer to Table 5-9b for footnotes.

TABLE 5-9b (ENGLISH). TOTAL REDUCED SULFUR COMPOUND EMISSIONS FOR MODEL RECOVERY FURNACES

Model recovery furnaces	Baseline emissions, ton/yr (a)	Control option	Control level emissions, ton/yr (b)	Emission reduction, ton/yr (b)	Emission reduction, %
NDCE					
RF-4	6.0	Wet to dry ESP system conversion	2.7	3.3	55
RF-5	11	Wet to dry ESP system conversion	5.0	5.9	55
RF-6	16	Wet to dry ESP system conversion	7.1	8.6	55
DCE					
RF-7	26	Low-odor conversion	1.6	25	94
RF-8	44	Low-odor conversion	2.7	41	94
RF-9	79	Low-odor conversion	4.9	74	94

(a) Baseline TRS for models RF-4 through RF-6 is based on a 10 percent trimmed TRS mean of 2.2 ppm taken from TRS emission data for NDCE recovery furnaces assumed to be equipped with wet ESP systems. The baseline TRS for models RF-7 through RF-9 is based on a 10 percent trimmed mean of 12 ppm taken from TRS emission data for DCE recovery furnaces plus an average TRS emission factor of 0.10 lb/ADTP for BLO units (assuming 3,400 lb BLS/ADT of bleached and unbleached pulp).

(b) TRS emission reduction = (baseline TRS) - (control level TRS). Control level TRS for models RF-4 through RF-9 is based on a TRS concentration of 1 ppm taken from TRS emission data for an NDCE recovery furnace known to be equipped with a dry ESP system.

TABLE 5-10a (METRIC). SECONDARY EMISSIONS FOR LOW-ODOR CONVERSION FOR MODEL DCE RECOVERY FURNACES[a]

Model recovery furnaces	Baseline emissions, kg/yr				Control option	Incremental emissions, kg/yr			
	PM	SO2	NOx	CO		PM	SO2	NOx	CO
RF-7	567	2,770	1,100	2,090	Low-odor conversion (b)	(567)	(2,770)	(1,100)	(2,090)
RF-8	776	3,770	1,500	2,840	Low-odor conversion (b)	(776)	(3,770)	(1,500)	(2,840)
RF-9	1,100	5,350	2,130	4,050	Low-odor conversion (b)	(1,100)	(5,350)	(2,130)	(4,050)
RF-7a	193	939	372	708	PM controls--0.10 g/dscm (c)	90	440	177	331
					PM controls--0.034 g/dscm (d)	205	998	394	748
RF-7b	238	1,160	458	871	PM controls--0.10 g/dscm (c)	45	222	91	168
					PM controls--0.034 g/dscm (d)	160	780	308	585
RF-8a	320	1,560	621	1,170	PM controls--0.10 g/dscm (c)	151	730	290	549
					PM controls--0.034 g/dscm (d)	342	1,660	662	1,260
RF-8b	395	1,920	767	1,450	PM controls--0.10 g/dscm (c)	76	367	145	272
					PM controls--0.034 g/dscm (d)	267	1,300	517	980
RF-9a	576	2,810	1,120	2,120	PM controls--0.10 g/dscm (c)	272	1,320	526	993
					PM controls--0.034 g/dscm (d)	612	2,980	1,180	2,250
RF-9b	712	3,470	1,380	2,610	PM controls--0.10 g/dscm (c)	136	662	263	499
					PM controls--0.034 g/dscm (d)	476	2,320	921	1,750

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 5-10b. See Table 5-10b for footnotes.
(b) Impacts were estimated based on removal of the BLO unit during the low-odor conversion.
(c) Impacts were estimated for low-odor conversion ESP upgrade.
(d) Impacts were estimated for low-odor conversion ESP upgrade plus packed-bed scrubber.

TABLE 5-10b (ENGLISH). SECONDARY EMISSIONS FOR LOW-ODOR CONVERSION FOR MODEL DCE RECOVERY FURNACES[a]

Model recovery furnaces	Baseline emissions, lb/yr				Control option	Incremental emissions, lb/yr			
	PM	SO2	NOx	CO		PM	SO2	NOx	CO
RF-7	1,250	6,110	2,430	4,600	Low-odor conversion (b)	(1,250)	(6,110)	(2,430)	(4,600)
RF-8	1,710	8,310	3,300	6,260	Low-odor conversion (b)	(1,710)	(8,310)	(3,300)	(6,260)
RF-9	2,430	11,800	4,700	8,920	Low-odor conversion (b)	(2,430)	(11,800)	(4,700)	(8,920)
RF-7a	424	2,070	821	1,560	PM controls--0.044 gr/dscf (c)	199	970	389	730
					PM controls--0.015 gr/dscf (d)	452	2,200	869	1,650
RF-7b	524	2,550	1,010	1,920	PM controls--0.044 gr/dscf (c)	100	490	200	370
					PM controls--0.015 gr/dscf (d)	353	1,720	680	1,290
RF-8a	706	3,440	1,370	2,590	PM controls--0.044 gr/dscf (c)	334	1,610	640	1,210
					PM controls--0.015 gr/dscf (d)	754	3,670	1,460	2,770
RF-8b	872	4,240	1,690	3,200	PM controls--0.044 gr/dscf (c)	168	810	320	600
					PM controls--0.015 gr/dscf (d)	588	2,870	1,140	2,160
RF-9a	1,270	6,200	2,460	4,670	PM controls--0.044 gr/dscf (c)	600	2,910	1,160	2,190
					PM controls--0.015 gr/dscf (d)	1,350	6,570	2,610	4,950
RF-9b	1,570	7,650	3,040	5,760	PM controls--0.044 gr/dscf (c)	300	1,460	580	1,100
					PM controls--0.015 gr/dscf (d)	1,050	5,120	2,030	3,860

(a) Baseline and control level secondary emissions were estimated based on electricity requirements and emission factors. Calculations for energy impacts are presented in Table 5-11b. The emission factors = 0.15 lb PM/MM Btu; 0.73 lb SO2/MM Btu; 0.29 lb NOx/MM Btu; and 0.55 lb CO/MM Btu. Numbers in parentheses represent negative values, indicating that secondary emissions are reduced by that amount.
(b) Impacts were estimated based on removal of the BLO unit during the low-odor conversion.
(c) Impacts were estimated for low-odor conversion ESP upgrade.
(d) Impacts were estimated for low-odor conversion ESP upgrade plus packed-bed scrubber.

TABLE 5-11a (METRIC). ENERGY IMPACTS FOR LOW-ODOR CONVERSION FOR MODEL DCE RECOVERY FURNACES

Model recovery furnaces	Baseline energy impacts		Control option	Control level energy impacts		Incremental energy impacts	
	Electricity, MWh/yr	Steam energy, MJ/yr		Electricity, MWh/yr	Steam energy, MJ/yr	Electricity, MWh/yr	Steam energy, MJ/yr
RF-7	2,450	1.32E+09	Low-odor conversion (b)	0	1.51E+09	(2,450)	1.97E+08
RF-8	3,330	2.19E+09	Low-odor conversion (b)	0	2.52E+09	(3,330)	3.28E+08
RF-9	4,750	3.95E+09	Low-odor conversion (b)	0	4.54E+09	(4,750)	5.91E+08
RF-7a	829	--	PM controls--0.10 g/dscm (c)	1,220	--	391	--
			PM controls--0.034 g/dscm (d)	1,710	--	881	--
RF-7b	1,020	--	PM controls--0.10 g/dscm (c)	1,220	--	200	--
			PM controls--0.034 g/dscm (d)	1,710	--	690	--
RF-8a	1,380	--	PM controls--0.10 g/dscm (c)	2,030	--	650	--
			PM controls--0.034 g/dscm (d)	2,850	--	1,470	--
RF-8b	1,700	--	PM controls--0.10 g/dscm (c)	2,030	--	330	--
			PM controls--0.034 g/dscm (d)	2,850	--	1,150	--
RF-9a	2,490	--	PM controls--0.10 g/dscm (c)	3,650	--	1,160	--
			PM controls--0.034 g/dscm (d)	5,120	--	2,630	--
RF-9b	3,070	--	PM controls--0.10 g/dscm (c)	3,650	--	580	--
			PM controls--0.034 g/dscm (d)	5,120	--	2,050	--

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 5-11b. See Table 5-11b for footnotes, which include calculations.
(b) Electricity impacts based on BLO removal; steam energy impacts based on furnace conversion.
(c) Impacts were estimated for low-odor conversion ESP upgrade.
(d) Impacts were estimated for low-odor conversion ESP upgrade plus packed-bed scrubber.

TABLE 5-11b (ENGLISH). ENERGY IMPACTS FOR LOW-ODOR CONVERSION FOR MODEL DCE RECOVERY FURNACES[a]

Model recovery furnaces	Baseline energy impacts		Control option	Control level energy impacts		Incremental energy impacts	
	Electricity, MWh/yr	Steam energy, MM Btu/yr		Electricity, MWh/yr (c),(e)	Steam energy MM Btu/yr (f)	Electricity, MWh/yr	Steam energy, MM Btu/yr
RF-7	2,450	1.25E+06	Low-odor conversion (b)	0	1.44E+06	(2,450)	1.87E+05
RF-8	3,330	2.08E+06	Low-odor conversion (b)	0	2.39E+06	(3,330)	3.12E+05
RF-9	4,750	3.75E+06	Low-odor conversion (b)	0	4.31E+06	(4,750)	5.61E+05
RF-7a	829	–	PM controls–0.044 gr/dscf (c)	1,220	–	391	–
			PM controls–0.015 gr/dscf (d)	1,710	–	881	–
RF-7b	1,020	–	PM controls–0.044 gr/dscf (c)	1,220	–	200	–
			PM controls–0.015 gr/dscf (d)	1,710	–	690	–
RF-8a	1,380	–	PM controls–0.044 gr/dscf (c)	2,030	–	650	–
			PM controls–0.015 gr/dscf (d)	2,850	–	1,470	–
RF-8b	1,700	–	PM controls–0.044 gr/dscf (c)	2,030	–	330	–
			PM controls–0.015 gr/dscf (d)	2,850	–	1,150	–
RF-9a	2,490	–	PM controls–0.044 gr/dscf (c)	3,650	–	1,160	–
			PM controls–0.015 gr/dscf (d)	5,120	–	2,630	–
RF-9b	3,070	–	PM controls–0.044 gr/dscf (c)	3,650	–	580	–
			PM controls–0.015 gr/dscf (d)	5,120	–	2,050	–

(a) Numbers in parentheses represent negative values, indicating that energy impacts are reduced by that amount.
Baseline BLO energy impacts = annual BLO operating cost x kWh/$0.06 x 1 MWh/1,000 kWh. Control level impacts = 0.
PM control energy impacts at baseline and at 0.044 gr/dscf = [0.00018 x model gas flowrate x 1 in. H2O pressure drop x 8,424 hr/yr x 1 MWh/1,000 kWh] + [0.00194 x model gas flow rate x SCA x 8,424 hr/yr x 1 MWh/1,000 kWh].
PM control energy impacts at 0.015 gr/dscf = [0.00018 x model gas flow rate x 4 in. H2O pressure drop x 8,424 hr/yr x 1 MWh/1,000 kWh] + [0.00194 x model gas flow rate x SCA x 8,424 hr/yr x 1 MWh/1,000 kWh] + [0.746 kW/hp x liquid flow rate x 60 ft head x spec. grav. H2O x 1/70% pump motor effici. x 8,424 hr/yr x 1 MWh/1,000 kWh].
Unconverted DCE recovery furnace steam energy impacts = steam energy production for DCE.
Converted DCE recovery furnace steam energy = steam energy production for NDCE - concentrator steam requirements
Steam energy production = 6,000 Btu/lb BLS x model BLS firing rate (lb BLS/d) x 351 d/yr x furnace thermal effic. (56% DCE, 66% NDCE) x 1/(85% power boiler thermal effic.)
Concentrator steam requirements = steam flow (10,000 lb/hr) x [model BLS firing rate (lb BLS/d)/2.4 MM lb BLS/d] x change in enthalpy (1,331.5 Btu/lb-189.2 Btu/lb) x 8,424 hr/yr.
(b) Electricity impacts based on BLO removal; steam energy impacts based on furnace conversion.
(c) Impacts were estimated for low-odor conversion ESP upgrade.
(d) Impacts were estimated for low-odor conversion ESP upgrade plus packed-bed scrubber.

TABLE 5-12a (METRIC). PRIMARY ACID GAS EMISSIONS FOR MODEL RECOVERY FURNACES[a]

Model recovery furnaces	Baseline emissions, Mg/yr		Control option	Control level emissions, Mg/yr		Emission reduction, Mg/yr		Emission reduction, %	
	Hydrochloric acid	Sulfur dioxide		Hydrochloric acid	Sulfur dioxide	Hydrochloric acid	Sulfur dioxide	Hydrochloric acid	Sulfur dioxide
NDCE/converted DCE									
RF-1/4	29	295	Packed-bed scrubber	15	148	14	148	48	50
RF-2/5	52	531	Packed-bed scrubber	27	266	25	266	48	50
RF-3/6	75	767	Packed-bed scrubber	39	384	36	384	48	50
RF-7	17	148	Packed-bed scrubber	8.9	89	8.3	59	48	40
RF-8	29	246	Packed-bed scrubber	15	148	14	98	48	40
RF-9	52	443	Packed-bed scrubber	27	266	25	177	48	40
Unconverted DCE									
RF-7	17	148	Packed-bed scrubber	9.4	74	7.8	74	46	50
RF-8	29	246	Packed-bed scrubber	16	123	13	123	46	50
RF-9	52	443	Packed-bed scrubber	28	221	24	221	46	50

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 5-12b. Refer to Table 5-12b for footnotes, which include calculations.

TABLE 5-12b (ENGLISH). PRIMARY ACID GAS EMISSIONS FOR MODEL RECOVERY FURNACES[a]

Model recovery furnaces	Baseline emissions, ton/yr		Control option	Control level emissions, ton/yr		Emission reduction, ton/yr		Emission reduction, %	
	Hydrochloric acid	Sulfur dioxide		Hydrochloric acid	Sulfur dioxide	Hydrochloric acid	Sulfur dioxide	Hydrochloric acid	Sulfur dioxide
NDCE/converted DCE									
RF-1/4	32	325	Packed-bed scrubber	16	163	15	163	48	50
RF-2/5	57	585	Packed-bed scrubber	29	293	27	293	48	50
RF-3/6	82	845	Packed-bed scrubber	43	423	40	423	48	50
RF-7	19	163	Packed-bed scrubber	9.8	98	9.1	65	48	40
RF-8	32	271	Packed-bed scrubber	16	163	15	108	48	40
RF-9	57	488	Packed-bed scrubber	29	293	27	195	48	40
Unconverted DCE									
RF-7	19	163	Packed-bed scrubber	10	81	8.6	81	46	50
RF-8	32	271	Packed-bed scrubber	17	135	14	135	46	50
RF-9	57	488	Packed-bed scrubber	31	244	26	244	46	50

(a) HCl/SO2 emissions (ton/yr) = emission factor (lb/lb BLS) x model BLS firing rate (lb BLS/d) x 351 d/yr x 1 ton/2,000 lb.

TABLE 5-13a (METRIC). SECONDARY EMISSIONS FOR HCL CONTROL FOR MODEL RECOVERY FURNACES[a]

Model recovery furnaces	Control option	Incremental emissions, kg/yr			
		PM	SO2	NOx	CO
NDCE/converted DCE					
RF-1/4/8	Packed-bed scrubber	229	1,120	444	839
RF-2/5/9	Packed-bed scrubber	412	2,000	798	1,510
RF-3/6	Packed-bed scrubber	594	2,900	1,150	2,190
RF-7	Packed-bed scrubber	138	671	266	503
Unconverted DCE					
RF-7	Packed-bed scrubber	145	708	281	531
RF-8	Packed-bed scrubber	242	1,180	467	889
RF-9	Packed-bed scrubber	435	2,120	844	1,600

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 5-13b. See Table 5-13b for footnotes.

TABLE 5-13b (ENGLISH). SECONDARY EMISSIONS FOR HCL CONTROL FOR MODEL RECOVERY FURNACES[a]

Model recovery furnaces	Control option	Incremental emissions, lb/yr			
		PM	SO2	NOx	CO
NDCE/converted DCE					
RF-1/4/8	Packed-bed scrubber	506	2,460	978	1,850
RF-2/5/9	Packed-bed scrubber	908	4,420	1,760	3,330
RF-3/6	Packed-bed scrubber	1,310	6,390	2,540	4,820
RF-7	Packed-bed scrubber	304	1,480	587	1,110
Unconverted DCE					
RF-7	Packed-bed scrubber	320	1,560	619	1,170
RF-8	Packed-bed scrubber	533	2,600	1,030	1,960
RF-9	Packed-bed scrubber	960	4,670	1,860	3,520

(a) Secondary emissions are zero at baseline; therefore, control level and incremental secondary emission numbers are the same. The control level secondary emissions were estimated based on electricity requirements and emission factors. Calculations for energy impacts are presented in Table 5-14. The emission factors = 0.15 lb PM/MM Btu; 0.73 lb SO2/MM Btu; 0.29 lb NOx/MM Btu; and 0.55 lb CO/MM Btu.

TABLE 5-14. ENERGY IMPACTS FOR HCL CONTROL FOR MODEL RECOVERY FURNACES[a]

Model recovery furnaces	Control option	Incremental energy impacts, MWh/yr
NDCE/converted DCE		
RF-1/4/8	Packed-bed scrubber	988
RF-2/5/9	Packed-bed scrubber	1,770
RF-3/6	Packed-bed scrubber	2,570
RF-7	Packed-bed scrubber	593
Unconverted DCE		
RF-7	Packed-bed scrubber	625
RF-8	Packed-bed scrubber	1,040
RF-9	Packed-bed scrubber	1,870

(a) Baseline energy impacts are equal to zero. Packed-bed scrubber energy impacts = [0.00018 x model gas flow rate x 3 in. H2O pressure drop x 1 MWh/1,000 kWh x 8,424 hr/yr] + [0.746 kW/hp x liquid flow rate x 60 ft head x specific gravity H2O x hp/3,960 gal x 1/(70% pump motor effic.) x 1 MWh/1,000 kWh x 8,424 hr/yr].

TABLE 5-15a (METRIC). WASTEWATER IMPACTS FOR HCL CONTROL FOR MODEL RECOVERY FURNACES[a]

Model recovery furnaces	Control option	Incremental wastewater impacts, million L/yr
NDCE/converted DCE		
RF-1/4/8	Packed-bed scrubber	10
RF-2/5/9	Packed-bed scrubber	18
RF-3/6	Packed-bed scrubber	26
RF-7	Packed-bed scrubber	6.1
Unconverted DCE		
RF-7	Packed-bed scrubber	6.3
RF-8	Packed-bed scrubber	10
RF-9	Packed-bed scrubber	19

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 5-15b. See Table 5-15b for footnotes, which include calculations.

TABLE 5-15b (ENGLISH). WASTEWATER IMPACTS FOR HCL CONTROL FOR MODEL RECOVERY FURNACES[a]

Model recovery furnaces	Control option	Incremental wastewater impacts, million gal/yr
NDCE/converted DCE		
RF-1/4/8	Packed-bed scrubber	2.7
RF-2/5/9	Packed-bed scrubber	4.8
RF-3/6	Packed-bed scrubber	7.0
RF-7	Packed-bed scrubber	1.6
Unconverted DCE		
RF-7	Packed-bed scrubber	1.7
RF-8	Packed-bed scrubber	2.8
RF-9	Packed-bed scrubber	5.0

(a) Wastewater impacts are zero at baseline; therefore, control level and incremental wastewater impact numbers are the same. Control level wastewater impacts = wastewater flow rate (gpm) x 60 min/hr x 8,424 hr/yr.

TABLE 5-16a (METRIC). MODEL BLACK LIQUOR OXIDATION UNIT PARAMETERS AND EMISSION FACTORS[a]

Model BLO units	Equipment type	Black liquor firing rate, kg BLS/d	Equivalent pulp production rate		Vent gas flow rate, m3/sec	Vent gas temperature, degrees C	Moisture content, %	Baseline gaseous organic HAP's, kg/kg BLS	Control level gaseous organic HAP's, kg/kg BLS
			ADMUP/d	ADMBP/d					
BLO-1	2-stage, air sparging	400,000	270	230	4.2	54	35	2.03E-04	4.06E-06
BLO-2	2-stage, air sparging	700,000	450	380	8.5	54	35	2.03E-04	4.06E-06
BLO-3	2-stage, air sparging	1,200,000	820	680	12.7	54	35	2.03E-04	4.06E-06

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 5-16b.

TABLE 5-16b (ENGLISH). MODEL BLACK LIQUOR OXIDATION UNIT PARAMETERS AND EMISSION FACTORS

Model BLO units	Equipment type	Black liquor firing rate, lb BLS/d	Equivalent pulp production rate		Vent gas flow rate, acfm	Vent gas temperature, degrees F	Moisture content, %	Baseline gaseous organic HAP's, lb/lb BLS (a)	Control level gaseous organic HAP's, lb/lb BLS (a)
			ADTUP/d	ADTBP/d					
BLO-1	2-stage, air-sparging	900,000	300	250	8,900	130	35	2.03E-04	4.06E-06
BLO-2	2-stage, air-sparging	1,500,000	500	420	18,000	130	35	2.03E-04	4.06E-06
BLO-3	2-stage, air-sparging	2,700,000	900	750	26,900	130	35	2.03E-04	4.06E-06

(a) Gaseous organic HAP's include acetaldehyde, benzene, formaldehyde, methanol, methyl ethyl ketone, methyl isobutyl ketone, phenol, toluene, and xylenes. Methanol comprises 85 percent of the total. Assume 98 percent control with incineration of BLO vent gases.

TABLE 5-17a (METRIC). PRIMARY GASEOUS ORGANIC HAP EMISSIONS FOR MODEL BLACK LIQUOR OXIDATION UNITS[a]

Model BLO units	Baseline emissions, Mg/yr	Control option	Control level emissions, Mg/yr	Emission reductions, Mg/yr	Emission reductions, %
BLO-1	29	BLO vent gas control	0.6	29	98
BLO-2	48	BLO vent gas control	1.0	48	98
BLO-3	87	BLO vent gas control	1.7	86	98

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 5-17b. Refer to Table 5-17b for footnotes, which include calculations.

TABLE 5-17b (ENGLISH). PRIMARY GASEOUS ORGANIC HAP EMISSIONS FOR MODEL BLACK LIQUOR OXIDATION UNITS[a]

Model BLO units	Baseline emissions, ton/yr	Control option	Control level emissions, ton/yr	Emission reductions, ton/yr	Emission reductions, %
BLO-1	32	BLO vent gas control	0.6	31	98
BLO-2	53	BLO vent gas control	1.1	52	98
BLO-3	96	BLO vent gas control	1.9	94	98

(a) Gaseous organic HAP emissions (ton/yr) = emission factor (lb/lb BLS) x model BLS firing rate (lb BLS/d) x 351 d/yr x 1 ton/2,000 lb. Gaseous organic HAP's include acetaldehyde, benzene, formaldehyde, methyl ethyl ketone, methyl isobutyl ketone, phenol, toluene, and xylenes. Methanol comprises 85 percent of the total.

TABLE 5-18a (METRIC). SECONDARY EMISSIONS FOR MODEL BLACK LIQUOR OXIDATION UNITS[a]

Model BLO units	Control option	Incremental emissions, kg/yr			
		PM	SO2	NOx	CO
BLO-1	BLO vent gas control	400	9,710	776	1,470
BLO-2	BLO vent gas control	807	16,900	1,560	2,970
BLO-3	BLO vent gas control	1,210	29,200	2,340	4,440

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 5-18b. Refer to Table 5-18b for footnotes.

TABLE 5-18b (ENGLISH). SECONDARY EMISSIONS FOR MODEL BLACK LIQUOR OXIDATION UNITS[a]

Model BLO units	Control option	Incremental emissions, lb/yr			
		PM	SO2	NOx	CO
BLO-1	BLO vent gas control	883	21,400	1,710	3,240
BLO-2	BLO vent gas control	1,780	37,200	3,450	6,540
BLO-3	BLO vent gas control	2,670	64,300	5,160	9,780

(a) Secondary emissions are zero at baseline; therefore, control level and incremental secondary emission numbers are the same. Secondary emissions for PM, NOx, and CO were estimated based on energy and emission factors for PM, NOx, and CO. Secondary emissions for SO2 were estimated based on (1) energy requirements and the emission factor for SO2 and (2) 1.88 lb of SO2 generated for each lb of TRS combusted. Calculations for energy impacts are presented in Table 5-19. Emission factors = 0.15 lb PM/MM Btu; 0.73 lb SO2/MM Btu; 0.29 lb NOx/MM Btu; and 0.55 lb CO/MM Btu.

TABLE 5-19. ENERGY IMPACTS FOR MODEL BLACK LIQUOR OXIDATION UNITS

Model BLO units	Control option	Incremental energy impacts, MWh/yr
BLO-1	BLO vent gas control	1,720
BLO-2	BLO vent gas control	3,480
BLO-3	BLO vent gas control	5,210

(a) BLO control energy impacts = 503 hp x 0.746 kW/hp x 8,424 hr/yr x 1 MWh/1,000 kWh x model vent gas flow rate/16,327 acfm. The hp requirements and vent gas flow rate were provided by an individual mill that controls vent gas emissions.

TABLE 5-20a (METRIC). TOTAL REDUCED SULFUR COMPOUND EMISSIONS FOR MODEL BLACK LIQUOR OXIDATION UNITS[a]

Model BLO units	Baseline emissions, Mg/yr	Control option	Control level emissions, Mg/yr	Emission reduction, Mg/yr	Emission reduction, %
BLO-1	4.2	BLO vent gas control	0.08	4.1	98
BLO-2	7.0	BLO vent gas control	0.14	6.9	98
BLO-3	13	BLO vent gas control	0.25	12	98

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 5-20b. Refer to Table 5-20b for footnotes.

TABLE 5-20b (ENGLISH). TOTAL REDUCED SULFUR COMPOUND EMISSIONS FOR MODEL BLACK LIQUOR OXIDATION UNITS

Model BLO units	Baseline emissions, ton/yr (a)	Control option	Control level emissions, ton/yr (b)	Emission reduction, ton/yr	Emission reduction, % (b)
BLO-1	4.6	BLO vent gas control	0.09	4.6	98
BLO-2	7.7	BLO vent gas control	0.15	7.6	98
BLO-3	14	BLO vent gas control	0.28	14	98

(a) The baseline TRS emissions for BLO units are based on an average TRS emission factor of 0.10 lb/ADTP (assuming an average 3,400 lb BLS/ADT of bleached and unbleached pulp).
(b) The control level TRS emissions for BLO units are based on 98 percent TRS control.

TABLE 5-21a (METRIC). MODEL SMELT DISSOLVING TANK PARAMETERS AND PM EMISSION FACTORS[a]

Model SDT's	Baseline APCD	Control option APCD	Black liquor firing rate kg BLS/d	Equivalent pulp production rate, ADMUP/d	Equivalent pulp production rate, ADMBP/d	Stack gas flow rate, m3/sec	Scrubber inlet gas flow rate, m3/sec	Existing pressure drop, mm Hg	New pressure drop, mm Hg	Baseline PM, kg/Mg BLS	Control level PM, kg/Mg BLS: PM control-- 0.10 kg/Mg BLS	Control level PM, kg/Mg BLS: PM control-- 0.06 kg/Mg BLS
SDT-1	scrubber	scrubber	400,000	270	230	4.2	4.4	13	13	0.19	0.10	0.06
SDT-2	scrubber	scrubber	700,000	450	380	7.1	7.4	13	13	0.19	0.10	0.06
SDT-3	scrubber	scrubber	1,200,000	820	680	12.7	13.4	13	13	0.19	0.10	0.06
SDT-4	scrubber	scrubber	1,800,000	1,200	1,000	18.4	19.3	13	13	0.19	0.10	0.06
SDT-5	mist eliminator	scrubber	400,000	270	230	4.2	4.4	1.3	13	0.23	0.10	0.06
SDT-6	mist eliminator	scrubber	700,000	450	380	7.1	7.4	1.3	13	0.23	0.10	0.06
SDT-7	mist eliminator	scrubber	1,200,000	820	680	12.7	13.4	1.3	13	0.23	0.10	0.06

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 5-21b.

TABLE 5-21b (ENGLISH). MODEL SMELT DISSOLVING TANK PARAMETERS AND PM EMISSION FACTORS[a]

Model SDT's	Baseline APCD	Control option APCD	Black liquor firing rate lb BLS/d	Equivalent pulp production rate,		Stack gas flow rate, acfm	Scrubber inlet gas flow rate, acfm	Existing pressure drop, in. H2O	New pressure drop, in. H2O	Baseline PM, lb/ton BLS	Control level PM, lb/ton BLS	
				ADTUP/d	ADTBP/d						PM control-- 0.2 lb/ton BLS	PM control-- 0.12 lb/ton BLS
SDT-1	scrubber	scrubber	900,000	300	250	9,000	9,400	7	7	0.37	0.20	0.12
SDT-2	scrubber	scrubber	1,500,000	500	420	15,000	15,700	7	7	0.37	0.20	0.12
SDT-3	scrubber	scrubber	2,700,000	900	750	27,000	28,300	7	7	0.37	0.20	0.12
SDT-4	scrubber	scrubber	3,900,000	1,300	1,100	39,000	40,900	7	7	0.37	0.20	0.12
SDT-5	mist eliminator	scrubber	900,000	300	250	9,000	9,400	0.7	7	0.46	0.20	0.12
SDT-6	mist eliminator	scrubber	1,500,000	500	420	15,000	15,700	0.7	7	0.46	0.20	0.12
SDT-7	mist eliminator	scrubber	2,700,000	900	750	27,000	28,300	0.7	7	0.46	0.20	0.12

TABLE 5-22a (METRIC). PRIMARY PM AND PM HAP EMISSIONS FOR MODEL SMELT DISSOLVING TANKS[a]

Model SDT's	Baseline emissions, Mg/yr		Control option	Control level emissions, Mg/yr		Emission reduction, Mg/yr		Emission reduction, %
	PM	PM HAP's		PM	PM HAP's	PM	PM HAP's	
SDT-1	27	0.02	PM control--0.10 kg/Mg BLS (b)	14	0.009	12	0.007	46
			PM control--0.06 kg/Mg BLS (b)	8.6	0.005	18	0.01	68
SDT-2	44	0.03	PM control--0.10 kg/Mg BLS (b)	24	0.01	20	0.01	46
			PM control--0.06 kg/Mg BLS (b)	14	0.009	30	0.02	68
SDT-3	80	0.05	PM control--0.10 kg/Mg BLS (b)	43	0.03	37	0.02	46
			PM control--0.06 kg/Mg BLS (b)	26	0.02	54	0.03	68
SDT-4	115	0.07	PM control--0.10 kg/Mg BLS (b)	62	0.04	53	0.03	46
			PM control--0.06 kg/Mg BLS (b)	37	0.02	78	0.05	68
SDT-5	33	0.02	PM control--0.10 kg/Mg BLS (c)	14	0.009	19	0.01	57
			PM control--0.06 kg/Mg BLS (c)	8.6	0.005	24	0.01	74
SDT-6	55	0.03	PM control--0.10 kg/Mg BLS (c)	24	0.01	31	0.02	57
			PM control--0.06 kg/Mg BLS (c)	14	0.009	41	0.02	74
SDT-7	99	0.06	PM control--0.10 kg/Mg BLS (c)	43	0.03	56	0.03	57
			PM control--0.06 kg/Mg BLS (c)	26	0.02	73	0.04	74

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 5-22b. Refer to Table 5-22b for footnotes, which include calculations.

(b) PM control is replacement of existing scrubber with a new scrubber.

(c) PM control is replacement of existing mist eliminator with a new scrubber.

TABLE 5-22b (ENGLISH). PRIMARY PM AND PM HAP EMISSIONS FOR MODEL SMELT DISSOLVING TANKS[a]

Model SDT's	Baseline emissions, ton/yr		Control option	Control level emissions, ton/yr		Emission reduction, ton/yr		Emission reduction, %
	PM	PM HAP's		PM	PM HAP's	PM	PM HAP's	
SDT-1	29	0.02	PM control--0.20 lb/ton BLS (b)	16	0.009	13	0.008	46
			PM control--0.12 lb/ton BLS (b)	9.5	0.006	20	0.01	68
SDT-2	49	0.03	PM control--0.20 lb/ton BLS (b)	26	0.02	22	0.01	46
			PM control--0.12 lb/ton BLS (b)	16	0.009	33	0.02	68
SDT-3	88	0.05	PM control--0.20 lb/ton BLS (b)	47	0.03	40	0.02	46
			PM control--0.12 lb/ton BLS (b)	28	0.02	59	0.04	68
SDT-4	127	0.08	PM control--0.20 lb/ton BLS (b)	68	0.04	58	0.03	46
			PM control--0.12 lb/ton BLS (b)	41	0.02	86	0.05	68
SDT-5	36	0.02	PM control--0.20 lb/ton BLS (c)	16	0.009	21	0.01	57
			PM control--0.12 lb/ton BLS (c)	9.5	0.006	27	0.02	74
SDT-6	61	0.04	PM control--0.20 lb/ton BLS (c)	26	0.02	34	0.02	57
			PM control--0.12 lb/ton BLS (c)	16	0.009	45	0.03	74
SDT-7	109	0.07	PM control--0.20 lb/ton BLS (c)	47	0.03	62	0.04	57
			PM control--0.12 lb/ton BLS (c)	28	0.02	81	0.05	74

(a) PM emissions (ton/yr) = PM emission factor (lb PM/ton BLS) x BLS firing rate (lb BLS/d) x 1 ton BLS/2,000 lb BLS x 351 d/yr x 1 ton PM/2,000 lb PM. PM HAP emissions (ton/yr) = 0.06 percent of PM emissions.
(b) PM control is replacement of existing scrubber with a new scrubber.
(c) PM control is replacement of existing mist eliminator with a new scrubber.

TABLE 5-23a (METRIC). SECONDARY EMISSIONS FOR MODEL SMELT DISSOLVING TANKS[a]

Model SDT's	Baseline emissions, kg/yr				Control option	Incremental emissions, kg/yr			
	PM	SO2	NOx	CO		PM	SO2	NOx	CO
SDT-5	2.3	11	4.5	8.5	PM control (b)	21	102	40	76
SDT-6	3.9	19	7.5	14	PM control (b)	35	170	67	128
SDT-7	7.0	34	13	26	PM control (b)	63	306	121	230

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 5-23b. Refer to Table 5-23b for footnotes.

(b) Impacts were estimated based on replacement of existing mist eliminator with a new scrubber.

TABLE 5-23b (ENGLISH). SECONDARY EMISSIONS FOR MODEL SMELT DISSOLVING TANKS[a]

Model SDT's	Baseline emissions, lb/yr				Control option	Incremental emissions, lb/yr			
	PM	SO2	NOx	CO		PM	SO2	NOx	CO
SDT-5	5.1	25	9.9	19	PM control (b)	46	224	89	169
SDT-6	8.5	42	17	31	PM control (b)	77	374	149	282
SDT-7	15	75	30	56	PM control (b)	138	674	268	508

(a) Secondary emissions were estimated based on energy impacts and emission factors for PM, SO2, NOx, and CO. Calculations for energy impacts are presented in Table 5-24. Emission factors = 0.15 lb PM/MM Btu; 0.73 lb SO2/MM Btu; 0.29 lb NOx/MM Btu; and 0.55 lb CO/MM Btu.

(b) Impacts were estimated based on replacement of existing mist eliminator with a new scrubber.

TABLE 5-24. ENERGY REQUIREMENTS FOR MODEL SMELT DISSOLVING TANKS

Model SDT's	Baseline energy impacts, MWh/yr (a)	Control option	Control level energy impacts, MWh/yr (b)	Incremental energy impacts, MWh/yr (b)
SDT-5	10	PM control (c)	100	90
SDT-6	17	PM control (c)	167	150
SDT-7	30	PM control (c)	300	270

(a) Baseline mist eliminator energy impacts = 0.00018 x model inlet gas flow rate x 0.7 in. H2O pressure drop x 8,424 hr/yr x 1 MWh/1,000 kWh

(b) Incremental energy impacts = (control level scrubber energy impacts) - (baseline mist eliminator energy impacts). Control level scrubber energy impacts = 0.00018 x model inlet gas flow rate x 7 in. H2O pressure drop x 8,424 hr/yr x 1 MWh/1,000 kWh

(c) Impacts were estimated based on replacement of existing mist eliminator with a new scrubber.

TABLE 5-25a (METRIC). MODEL LIME KILN PARAMETERS AND PM CONCENTRATIONS[a]

Model lime kilns	Baseline APCD	Control level APCD	Lime production rate, Mg CaO/d	Equivalent pulp production rate ADMP/d	Gas flow rate, m3/sec		Temperature, degrees C		Moisture content, %	
					APCD inlet	APCD outlet	APCD inlet	APCD outlet	APCD inlet	APCD outlet
LK-1	scrubber	ESP	90	320	10	7.3	249	71	25	30
LK-2	scrubber	ESP	180	680	20	14	249	71	25	30
LK-3	scrubber	ESP	270	1,000	34	24	249	71	25	30
LK-4	ESP	ESP	90	320	10	10	249	249	25	25
LK-5	ESP	ESP	180	680	20	20	249	249	25	25
LK-6	ESP	ESP	270	1,000	34	34	249	249	25	25

Model lime kilns	Baseline pressure drop, mm Hg	ESP SCA, m2/(m3/sec)			Baseline PM, g/dscm	Control level PM, g/dscm	
		Baseline	PM control--0.15 g/dscm	PM control--0.023 g/dscm		PM control--0.15 g/dscm	PM control--0.023 g/dscm
LK-1	39	--	90	220	0.27	0.15	0.023
LK-2	39	--	90	220	0.27	0.15	0.023
LK-3	39	--	90	220	0.27	0.15	0.023
LK-4	--	90	90	220	0.15	0.15	0.023
LK-5	--	90	90	220	0.15	0.15	0.023
LK-6	--	90	90	220	0.15	0.15	0.023

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 5-25b.

TABLE 5-25b (ENGLISH). MODEL LIME KILN PARAMETERS AND PM CONCENTRATIONS[a]

Model lime kilns	Baseline APCD	Control level APCD	Lime production rate, ton CaO/d	Equivalent pulp production rate, ADTP/d	Gas flow rate, acfm		Temperature, degrees F		Moisture content, %	
					APCD inlet	APCD outlet	APCD inlet	APCD outlet	APCD inlet	APCD outlet
LK-1	scrubber	ESP	100	350	22,000	15,500	480	160	25	30
LK-2	scrubber	ESP	200	750	42,500	30,000	480	160	25	30
LK-3	scrubber	ESP	300	1,100	72,200	51,000	480	160	25	30
LK-4	ESP	ESP	100	350	22,000	22,000	480	480	25	25
LK-5	ESP	ESP	200	750	42,500	42,500	480	480	25	25
LK-6	ESP	ESP	300	1,100	72,200	72,200	480	480	25	25

Model lime kilns	Baseline pressure drop, in. H2O	ESP SCA, ft2/1,000 acfm			Baseline PM, gr/dscf	Control level PM, gr/dscf	
		Baseline	PM control--0.067 gr/dscf	PM control--0.01 gr/dscf		PM control--0.067 gr/dscf	PM control--0.01 gr/dscf
LK-1	21	--	460	1,120	0.12	0.067	0.01
LK-2	21	--	460	1,120	0.12	0.067	0.01
LK-3	21	--	460	1,120	0.12	0.067	0.01
LK-4	--	460	460	1,120	0.067	0.067	0.01
LK-5	--	460	460	1,120	0.067	0.067	0.01
LK-6	--	460	460	1,120	0.067	0.067	0.01

TABLE 5-26a (METRIC). PRIMARY PM AND PM HAP EMISSIONS FOR MODEL LIME KILNS[a]

Model lime kilns	Baseline emissions, Mg/yr		Control option	Control level emissions, Mg/yr		Emission reduction, Mg/yr		Emission reduction, %
	PM	PM HAP's		PM	PM HAP's	PM	PM HAP's	
LK-1	36	0.5	PM control--0.15 g/dscm (b)	20	0.3	16	0.2	44
			PM control--0.023 g/dscm (b)	3.0	0.04	33	0.5	92
LK-2	70	1.0	PM control--0.15 g/dscm (b)	39	0.6	31	0.4	44
			PM control--0.023 g/dscm (b)	5.9	0.08	65	0.9	92
LK-3	120	1.7	PM control--0.15 g/dscm (b)	67	0.9	53	0.7	44
			PM control--0.023 g/dscm (b)	10	0.1	110	1.5	92
LK-4	20	0.3	PM control--0.023 g/dscm (c)	3.0	0.04	17	0.2	85
LK-5	39	0.6	PM control--0.023 g/dscm (c)	5.9	0.08	33	0.5	85
LK-6	67	0.9	PM control--0.023 g/dscm (c)	10	0.1	57	0.8	85

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 5-26b. Refer to Table 5-26b for footnotes, which include calculations.

(b) PM control is replacement of existing scrubber with new ESP.

(c) PM control is upgrade of existing ESP.

TABLE 5-26b (ENGLISH). PRIMARY PM AND PM HAP EMISSIONS FOR MODEL LIME KILNS[a]

Model lime kilns	Baseline emissions, ton/yr		Control option	Control level emissions, ton/yr		Emission reduction, ton/yr		Emission reduction, %
	PM	PM HAP's		PM	PM HAP's	PM	PM HAP's	
LK-1	40	0.6	PM control--0.067 gr/dscf (b)	22	0.3	18	0.2	44
			PM control--0.01 gr/dscf (b)	3.3	0.05	37	0.5	92
LK-2	78	1.1	PM control--0.067 gr/dscf (b)	43	0.6	34	0.5	44
			PM control--0.01 gr/dscf (b)	6.5	0.09	71	1.0	92
LK-3	132	1.8	PM control--0.067 gr/dscf (b)	74	1.0	58	0.8	44
			PM control--0.01 gr/dscf (b)	11	0.2	121	1.7	92
LK-4	22	0.3	PM control--0.01 gr/dscf (c)	3.3	0.05	19	0.3	85
LK-5	43	0.6	PM control--0.01 gr/dscf (c)	6.5	0.09	37	0.5	85
LK-6	74	1.0	PM control--0.01 gr/dscf (c)	11	0.2	63	0.9	85

(a) PM emissions (ton/yr) = PM concentration (gr/dscf) x lb/7,000 gr x model inlet gas flow rate (acfm) x (528R/[model inlet temperature + 460F]) x (100% - model inlet %H2O)/100% x 60 min/hr x 8,424 hr/yr x 1 ton/2,000 lb. PM HAP emissions (ton/yr) = 1.4 percent of PM emissions.

(b) PM control is replacement of existing scrubber with new ESP.

(c) PM control is upgrade of existing ESP.

TABLE 5-27a (METRIC). SECONDARY EMISSIONS FOR MODEL LIME KILNS[a]

Model lime kilns	Baseline emissions, kg/yr				Control option	Incremental emissions, kg/yr			
	PM	SO2	NOx	CO		PM	SO2	NOx	CO
LK-1	163	794	315	599	PM control--0.15 g/dscm (b) PM control--0.023 g/dscm (b)	(116) (61)	(567) (299)	(225) (118)	(429) (226)
LK-2	314	1,530	608	1,150	PM control--0.15 g/dscm (b) PM control--0.023 g/dscm (b)	(225) (118)	(1,090) (576)	(435) (229)	(826) (431)
LK-3	535	2,600	1,030	1,960	PM control--0.15 g/dscm (b) PM control--0.023 g/dscm (b)	(383) (202)	(1,860) (980)	(739) (390)	(1,400) (739)
LK-4	46	225	90	170	PM control--0.023 g/dscm (c)	55	269	107	202
LK-5	89	435	173	328	PM control--0.023 g/dscm (c)	107	517	206	393
LK-6	152	739	294	558	PM control--0.023 g/dscm (c)	181	880	350	662

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 5-27b. Refer to Table 5-27b for footnotes.
(b) Impacts were estimated based on replacement of existing scrubber with a new ESP.
(c) Impacts were estimated based on upgrade of existing ESP.

TABLE 5-27b (ENGLISH). SECONDARY EMISSIONS FOR MODEL LIME KILNS[a]

Model lime kilns	Baseline emissions, lb/yr				Control option	Incremental emissions, lb/yr			
	PM	SO2	NOx	CO		PM	SO2	NOx	CO
LK-1	359	1,750	694	1,320	PM control--0.067 gr/dscf (b)	(257)	(1,250)	(496)	(946)
					PM control--0.01 gr/dscf (b)	(135)	(660)	(261)	(499)
LK-2	693	3,370	1,340	2,540	PM control--0.067 gr/dscf (b)	(496)	(2,410)	(959)	(1,820)
					PM control--0.01 gr/dscf (b)	(261)	(1,270)	(504)	(950)
LK-3	1,180	5,730	2,280	4,320	PM control--0.067 gr/dscf (b)	(845)	(4,100)	(1,630)	(3,090)
					PM control--0.01 gr/dscf (b)	(445)	(2,160)	(860)	(1,630)
LK-4	102	497	197	374	PM control--0.01 gr/dscf (c)	122	593	235	446
LK-5	197	960	381	723	PM control--0.01 gr/dscf (c)	235	1,140	455	867
LK-6	335	1,630	648	1,230	PM control--0.01 gr/dscf (c)	399	1,940	772	1,460

(a) Secondary emissions were estimated based on electricity requirements and emission factors for PM, SO2, NOx, and CO. Calculations for electricity requirements are presented in Table 5-28. Emission factors = 0.15 lb PM/MM Btu; 0.73 lb SO2/MM Btu; 0.29 lb NOx/MM Btu; and 0.55 lb CO/MM Btu.
Numbers in parentheses represent negative values, indicating that secondary emissions are reduced by that amount.
(b) Impacts were estimated based on replacement of existing scrubber with a new ESP.
(c) Impacts were estimated based on upgrade of existing ESP.

TABLE 5-28. ENERGY IMPACTS FOR MODEL LIME KILNS[a]

Model lime kilns	Baseline energy impacts, MWh/yr (b),(c)	Control option	Control level energy impacts, MWh/yr (c)	Incremental energy impacts, MWh/yr
LK-1	701	PM control--0.15 g/dscm (0.067 gr/dscf) (d)	199	(501)
		PM control--0.023 g/dscm (0.01 gr/dscf) (d)	437	(264)
LK-2	1,350	PM control--0.15 g/dscm (0.067 gr/dscf) (d)	385	(965)
		PM control--0.023 g/dscm (0.01 gr/dscf) (d)	844	(506)
LK-3	2,300	PM control--0.15 g/dscm (0.067 gr/dscf) (d)	654	(1,650)
		PM control--0.023 g/dscm (0.01 gr/dscf) (d)	1,430	(870)
LK-4	199	PM control--0.023 g/dscm (0.01 gr/dscf) (e)	437	238
LK-5	385	PM control--0.023 g/dscm (0.01 gr/dscf) (e)	844	459
LK-6	654	PM control--0.023 g/dscm (0.01 gr/dscf) (e)	1,430	776

(a) Numbers in parentheses represent negative values, indicating that energy impacts are reduced by that amount.

(b) Baseline scrubber energy impacts = 0.00018 x scrubber inlet gas flow rate x 21 in. H2O pressure drop x 8,424 hr/yr x 1 MWh/1,000 kWh

(c) Baseline and control level ESP energy impacts = [(0.00018 x ESP inlet gas flow rate x 1 in. H2O pressure drop) + (0.00194 x ESP inlet gas flow rate x ESP SCA)] x (8,424 hr/yr x 1 MWh/1,000 kWh)

(d) Impacts were estimated based on replacement of existing scrubber with a new ESP.

(e) Impacts were estimated based on upgrade of existing ESP.

TABLE 5-29a (METRIC). WASTEWATER IMPACTS FOR MODEL LIME KILNS[a]

Model lime kilns	Control option	Incremental wastewater impacts, million L/yr
LK-1	PM control (b)	(226)
LK-2	PM control (b)	(484)
LK-3	PM control (b)	(709)

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 5-29b. Refer to Table 5-29b for footnotes, which include calculations.
(b) Impacts were estimated based on replacement of the existing scrubber with a new ESP.

TABLE 5-29b (ENGLISH). WASTEWATER IMPACTS FOR MODEL LIME KILNS[a]

Model lime kilns	Control option	Incremental wastewater impacts, million gal/yr
LK-1	PM control (b)	(60)
LK-2	PM control (b)	(128)
LK-3	PM control (b)	(187)

(a) Control level wastewater impacts are zero. Therefore, incremental impacts are equal to baseline. Numbers in parentheses represent negative values, indicating that wastewater impacts are reduced by that amount. Wastewater impacts = 4,500 lb/ODTP x 0.9 ODTP/ADTP x ADTP/d x 351 d/yr x gal/8.345 lb
(b) Impacts were estimated based on replacement of the existing scrubber with a new ESP.

5.6 REFERENCES FOR CHAPTER 5

1. Memorandum from Holloway, T. and Nicholson, R., MRI, to Telander, J., EPA/MICG. October 9, 1996. Nationwide Costs, Environmental Impacts, and Cost-Effectiveness of Regulatory Alternatives for Kraft, Soda, Sulfite, and Semichemical Combustion Sources.

2. Memorandum from Holloway, T., MRI, to the project file. June 14, 1996. Summary of PM and HAP Metals Data.

3. Telecon. Soltis, V., MRI, with Sanders, D., Andersen 2000, Inc. August 11, 1993. Cost and efficiency information about HCl control for recovery furnaces.

4. Telecon. Soltis, V., MRI, with Bruno, J., AirPol, Inc. July 2, 1993. Information about scrubbers used to control HCl emissions.

5. Someshwar, A. and J. Pinkerton. Wood Processing Industry. In: Air Pollution Engineering Manual, Air & Waste Management Association. Buonicore, A. and W. Davis (eds.). New York, Van Norstrand Reinhold. 1992. pp. 835-849.

6. Memorandum from Soltis, V., MRI, to the project file. April 3, 1995. Kraft and Soda Pulp Mill Combustion Sources Data Base.

7. Memorandum from Holloway, T., MRI, to the project file. May 3, 1995. Sulfur Dioxide Control by Recovery Furnace Scrubbers.

8. Memorandum from Holloway, T., MRI, to the project file. May 21, 1996. Secondary Emission Calculations.

9. OAQPS Control Cost Manual. 4th Edition. U. S. Environmental Protection Agency. Research Triangle Park, NC Publication No. EPA-450/3-90-006. January 1990.

10. Letter and attachments from Black, C., Champion International Corp., Courtland, AL, to Crowder, J., EPA/ISB. June 10, 1993. Response to request for information on costs for the BLO vent gas control system at the Champion facility in Courtland, AL..

11. Kirby, M. Economic and Process Considerations in the Use of Oxygen for Black Liquor Oxidation. Union Carbide Canada Limited. Ontario, Canada. (Presented at the 21st Pulp and Paper ABCP Annual Meeting. Sao Paolo, Brazil. November 21-25, 1988.) 10 p.

12. Stultz, S. and J. Kitto (eds.). Steam: Its Generation and Use. 40th Edition. Babcock & Wilcox Co. New York. 1992. Chapter 26, pp. 3-7.

13. Handbook: Control Technologies for Hazardous Air Pollutants. U. S. Environmental Protection Agency. Cincinnati, OH. Publication No. EPA-625/6-91-014. June 1991. Chapter 4.

14. Proposed Standards of Performance for Kraft Pulp Mills. In: Standards Support and Environmental Impact Statement. Volume 1. U. S. Environmental Protection Agency. Research Triangle Park, NC. Publication No. EPA-450/2-76-014-a. September 1976. p. 2-9.

15. McIlvane, R. Removal of Heavy Metals and Other Utility Air Toxics. (Presented at the EPRI Hazardous Air Pollutant Conference. 1993.).

16. Letter from Burt, R., Radian Corp., to Holbrook, J. and Bringman, L., Environmental Elements Corp. July 23, 1985. Facsimile cover letter discussing utility and maintenance costs for wet- vs. dry-bottom ESP designs.

17. Memorandum from Eddinger, J., EPA/ISB, to Durkee, K., EPA/ISB. July 8, 1983. Meeting minutes from the April 14, 1993 meeting between kraft pulp mill representatives, the National Council of the Paper Industry for Air and Stream Improvement, Inc., and the U. S. Environmental Protection Agency. Discussion of problems mills have with TRS pickup in wet-bottom ESP's using unoxidized black liquor. Attachments.

18. Devore, J. Probability and Statistics for Engineering and the Sciences. Brooks/Cole Publishing Co. Monterey, California. 1982. pp. 15-17.

19. Someshwar, A. Compilation of "Air Toxic" Emission Data for Boilers, Pulp Mills, and Bleach Plants. National Council of the Paper Industry for Air and Stream Improvement, Inc. New York. Technical Bulletin No. 650. June 1993. 128 p.

20. Garner, J., Jaako Pöyry. Conversion to Low Odor Improves Recovery Boiler Efficiency and Life. Pulp and Paper. 63(7):91-95. July 1989.

21. Green, R. and G. Hough (eds.). Chemical Recovery in the Alkaline Pulping Process. 3rd Edition. Prepared by the Alkaline Pulping Committee of the Pulp Manufacture Division. Atlanta, GA, TAPPI Press, 1992. 196 p.

22. Telecon. Soltis, V., MRI, with Oscarsson, B., Götaverken Energy Systems, Inc. December 16, 1993. Concentrator steam costs.

23. Letter and attachments from Sanders, D., Andersen 2000, Inc., to Soltis, V., MRI. November 9, 1993. Preliminary proposal for three separate quench and packed-tower absorption systems to control hydrochloric acid gas emissions from recovery furnaces at kraft pulp and paper mills.

24. Memorandum from Nicholson, R. and Holloway, T., MRI, to Telander, J., EPA/MICG. September 6, 1995. Trip Report for S.D. Warren Co., Muskegon, MI.

25. Memorandum from Holloway, T., MRI, to the project file. July 16, 1996. State of Washington PM Data for Kraft Recovery Furnaces, Smelt Dissolving Tanks, and Lime Kilns.

6.0 MODEL PROCESS UNIT CONTROL AND ENHANCED MONITORING COSTS

This chapter discusses the costs of controlling HAP emissions from new and existing combustion sources at kraft and soda pulp and paper mills. The costs of each control option discussed in Chapter 4 are presented for the individual model combustion process units. Also presented are the costs of the enhanced monitoring options discussed in Chapter 4. The enhanced monitoring options are methods of demonstrating continuous compliance with the control options. The total nationwide costs associated with each control option are presented in a separate memorandum.[1]

Section 6.1 of this chapter presents the capital and annual costs for each control option for model process units representing recovery furnaces, BLO units, SDT's, and lime kilns. Section 6.2 discusses the costs of the enhanced monitoring options. Section 6.3 contains the references cited in this chapter.

6.1 CONTROL OPTION COSTS

This section discusses the general costing approach used to develop capital and annual costs for each control option and presents the estimated capital and annual costs of each control option as applied to the model process units. Section 6.1.1 describes the general costing approach. Section 6.1.2 provides the capital and annual costs associated with each recovery furnace control option and model recovery furnace. Sections 6.1.3, 6.1.4, and 6.1.5 provide the capital and annual costs applicable to the control options for model BLO units, SDT's and lime kilns, respectively.

6.1.1 General Costing Approach

A number of assumptions were made in deriving the costs for the control options. The year 1991 was used as the base year for all costs. Capital costs were adjusted to 1991 dollars using the Chemical Engineering Plant Cost Index or the Consumer Price Index, whichever was applicable.[2,3] Operating and maintenance personnel were assumed to work 3 shifts per day, 8 hours per shift, for 365 d/yr, which is equivalent to 8,760 operating hr/yr. All process units were assumed to operate 24 hr/d, for 351 d/yr, which is equivalent to 8,424 operating hr/yr. This operating time accounts for 14 days of scheduled shutdown annually for maintenance and repair. These 14 days could be combined into a single 2-week annual shutdown period to provide sufficient time for the modifications and upgrades both to existing equipment and APCD's needed to comply with the control options. If the scheduled 2-week shutdown did not provide sufficient time for the modifications and upgrades, then pulp production losses were calculated for the number of days beyond the 2-week shutdown that were needed to finish the work.

The procedure used for estimating pulp production losses is described in the following section. The general approaches used to develop capital and annual control costs are provided in Sections 6.1.1.2 and 6.1.1.3, respectively.

6.1.1.1 Estimation of Pulp Production Losses. Pulp production losses were calculated for each pulp type as a product of the pulp loss value and the total quantity of lost pulp production.

Assuming the kraft pulp mills achieve an earnings margin of 25 percent, the pulp loss value was estimated to be equal to approximately 25 percent of the total market value.[3] The 1989 market values of bleached and unbleached kraft pulp were estimated to be approximately $712 and $423, respectively, per ADMP ($646 and $384, respectively, per ADTP). The market values were adjusted to 1991 dollars using the Consumer Price Index.[3]

To estimate for each model recovery furnace the total quantity of lost pulp production associated with the extended

recovery furnace downtime, the number of days of shutdown beyond the scheduled 2-week shutdown were multiplied by the appropriate model pulp production rate. Model pulp production rates were determined for kraft bleached and unbleached pulp mills by multiplying model BLS firing rates by the appropriate correlation factor. A correlation factor of 1,800 kg BLS/ADMP (3,600 lb BLS/ADTP) was used for bleached pulp; a correlation factor of 1,500 kg BLS/ADMP (3,000 lb BLS/ADTP) was used for unbleached pulp.[4]

6.1.1.2 Development of Capital Costs. The following sources of cost information were used to develop capital costs (i.e., total capital investment [TCI]) for the control options being considered for kraft and soda pulp mill combustion sources:

1. Actual installed capital costs provided by individual kraft and soda pulp and paper mills;
2. Cost equations and quotes supplied by ESP and scrubber manufacturers;
3. Information supplied by recovery furnace manufacturers;
4. The U. S. EPA Handbook: Control Technologies for Hazardous Air Pollutants;[5] and
5. The U.S. EPA's OAQPS Control Cost Manual.[6]

Whenever possible, actual cost information from individual mills was used to develop the capital cost estimates. Because many combustion sources are subject to the NSPS for PM emissions from kraft and soda pulp mills, actual mill-specific PM control costs were already available for all of the control options involving control of PM HAP's. For those control options where actual costs were not available from individual mills or were only available from one mill, or where mill-specific costs varied widely, the EPA reference books (sources 4 and 5, above) were used.

In most cases, cost algorithms were developed, relating costs to the model process unit parameters (e.g., gas flow rate). In a few cases where a direct relationship between the capital cost and the model process unit was not apparent, the "six-tenths rule" was used to extrapolate costs from one model to another.

The "six-tenths rule" assumes a direct relationship between capital cost and capacity taken to the six-tenths power (i.e., $C_1/C_2 = (Q_1/Q_2)^{0.6}$, where C and Q are capital cost and capacity parameter, respectively). The capital costs were extrapolated using either gas flow rate or BLS firing rate as the capacity parameter.

6.1.1.3 Development of Annual Costs. Incremental total annual costs (ITAC) were derived using the annual cost model described in the OAQPS Control Cost Manual.[6] Incremental total annual costs refer to the incremental increase of total annual costs (TAC) over current operation. Total annual costs include both direct annual costs (DAC) and indirect annual costs (IAC). The cost components that comprise the DAC and IAC are discussed in the following sections.

6.1.1.3.1 Direct annual costs. The DAC include operating labor costs, maintenance labor and material costs, utility costs, and wastewater treatment costs.[6] Operating and maintenance labor costs were calculated as a product of the number of working hr/d to perform the required task, the number of operating d/yr, and the hourly wage. Operating and maintenance labor costs were calculated assuming 365 operating d/yr. The hourly wage and number of working hr/d vary with each control option and are based on information from the U. S. EPA's Handbook: Control Technologies for Hazardous Air Pollutants and the OAQPS Control Cost Manual.[5,6] The maintenance hourly wage is equal to approximately 1.5 times the operating hourly wage. The supervisory labor cost is approximately 15 percent of the operating labor cost. With the exception of ESP's, the maintenance materials cost is estimated as approximately 100 percent of the maintenance labor cost.[6] The maintenance materials cost for ESP's is estimated as 1 percent of the flange-to-flange purchased equipment cost (PEC) for ESP's, and the PEC is estimated as 0.6 times the TCI.[5]

Utility costs were broken down into electricity costs and water costs. Electricity costs were calculated as a product of the electricity unit cost and the electricity requirement. The

electricity unit cost was assumed to be $0.06/kWh.[5] Electricity requirements were divided into fan, pump, and operating electricity requirements and were calculated assuming 8,424 operating hr/yr.

The fan electricity requirement (applicable to ESP's and scrubbers) is equal to a numerical factor (0.00018) times the product of the gas flow rate, pressure drop, and operating hr/yr.[6] The gas flow rate varies with each model process unit. The pressure drop is based on information from mills.[4] Although the pressure drop is not the sole parameter that determines PM collection efficiency for lime kiln scrubbers, for the purposes of estimating costs, the pressure drop was used as an indicator of PM collection efficiency for lime kiln scrubbers. Note: A different scrubber design, rather than a higher pressure drop, was used to improve the PM collection efficiency for SDT scrubbers.

The gas flow rate and pressure drop do not change for ESP's relative to current operation when the ESP's are upgraded or replaced to improve PM collection. Therefore, the fan electricity requirements for ESP's do not change relative to current operation. However, if a scrubber is added after an ESP, the gas flow rate would be reduced, thereby reducing the fan electricity requirements for the ESP.

The pump electricity requirement (applicable to packed-bed scrubbers) is equal to a numerical factor (0.000188) times the product of the liquid flow rate, amount of head pressure, and operating hr/yr divided by the pump-motor efficiency.[6] The liquid flow rate varies with each model process unit. A head pressure of 18 m (60 ft) and a pump-motor efficiency of 70 percent were assumed.

The operating electricity requirement (applicable to ESP's) is equal to a numerical factor (0.00194) times the product of the ESP plate area and operating hr/yr.[6] The ESP plate area is calculated as a product of the exhaust gas flow rate and the ESP SCA. The gas flow rate varies with each process unit, and the SCA is based on information from mills.[4] Although the SCA is not

the sole parameter that determines PM collection efficiency for ESP's, for the purposes of estimating costs, the SCA was used as an indicator of PM collection efficiency. The cost analysis does not consider increasing the SCA to account for single-chamber ESP operation at reduced gas flow during maintenance situations. Instead, the regulation will allow for PM emission excursions during maintenance situations as part of the "Startup, Shutdown, Malfunction Plan" described in the General Provisions.[7]

Water costs were calculated as a product of the water unit cost and the water requirement. The water unit cost was assumed to be $\$0.05/m^3$ ($0.20/1,000 gal).[5] The water requirement is equal to a numerical factor (0.060) times the product of the gas flow rate and operating hr/yr.[6] The gas flow rate varies with each model process unit.

For some control options (e.g., PM controls that include upgrading or replacing an existing recovery furnace ESP or replacing an existing SDT scrubber with a new scrubber), the labor and maintenance costs are not expected to increase significantly, and, therefore, the ITAC is represented by the IAC plus the difference in electricity costs before and after implementation of the control options.

6.1.1.3.2 <u>Indirect annual costs</u>. The IAC include overhead costs, administrative charges, property taxes, insurance costs, and capital recovery costs. Overhead costs are approximately 60 percent of all labor and maintenance material costs.[6] Overhead costs are not applicable when labor and maintenance costs do not increase significantly. Administrative, insurance, and property tax costs are approximately 4 percent of the TCI.[6] Capital recovery costs are equal to a capital recovery factor (CRF) multiplied by the TCI. The CRF is estimated using the equation $CRF = [i(1 + i)^n]/[(1 + i)^n - 1]$, where i = interest rate (assumed to be 7 percent) and n = equipment life of the device being installed or modified.[6]

6.1.2 Recovery Furnace Control Options

The following sections discuss the model costs of four control options evaluated for recovery furnaces. These control options include PM controls (Section 6.1.2.1), wet to dry ESP system conversion (Section 6.1.2.2), conversion of a DCE recovery furnace system to an NDCE recovery furnace (Section 6.1.2.3), and addition of a packed-bed scrubber (Section 6.1.2.4).

6.1.2.1 PM Controls. Two PM control options were evaluated for model NDCE recovery furnaces RF-1 through RF-6 and model DCE recovery furnaces RF-7 through RF-9. The control options apply to new and existing recovery furnaces and are described below.

One PM control option that was evaluated would reduce PM emissions from existing recovery furnaces to the NSPS level of 0.10 g/dscm (0.044 gr/dscf). The PM control option evaluated would involve (1) replacing the recovery furnace ESP or (2) upgrading the recovery furnace ESP. The control equipment selected by a particular mill would be site-specific.

A second PM control option that was evaluated would reduce PM emissions from existing recovery furnaces to 0.034 g/dscm (0.015 gr/dscf). This more stringent PM control option would involve replacing or upgrading the recovery furnace ESP and adding a packed-bed scrubber. The second PM control option also applies to new recovery furnaces; the option could be used to evaluate the cost to new sources subject to a more stringent standard (0.034 g/dscm [0.015 gr/dscf]) than the current NSPS.

The PM control costs for existing recovery furnaces with baseline PM emissions above the NSPS level were estimated for model NDCE recovery furnaces RF-1a through RF-6a and model DCE recovery furnaces RF-7a through RF-9a. The PM control costs for new and existing recovery furnaces with baseline PM emissions at or below the NSPS level were estimated for model NDCE recovery furnaces RF-1b through RF-6b and model DCE recovery furnaces RF-7b through RF-9b.

The PM control costs for model NDCE recovery furnaces with dry ESP systems (i.e., RF-1 through RF-3) are assumed to be identical to the PM control costs for model NDCE recovery

furnaces with wet ESP systems (i.e., RF-4 through RF-6) because PM emissions are not affected by whether or not black liquor is used in the ESP bottom or PM return system. The capital and annual costs to replace or upgrade ESP's are presented for the model NDCE and DCE recovery furnaces in the following sections. The capital and annual costs to install a new packed-bed scrubber are presented in Section 6.1.2.4.

Based on the costs presented below, it is more expensive to replace an existing ESP than to upgrade it. However, a replacement may be necessary, depending on the age and condition of the existing ESP, in order to effectively control PM emissions. Site-specific conditions often dictate the cost-effectiveness of replacing or upgrading an ESP.

6.1.2.1.1 <u>ESP replacement: capital costs</u>. The ESP replacement costs for models RF-1a through RF-9a were calculated based on recent ESP replacement costs provided by individual pulp and paper mills (i.e., costs for ESP's replaced during or after 1989).[4] The cost to dispose of the existing ESP was assumed to be included in the new ESP costs provided by the individual mills. New ESP costs average $420/m^2$ ($39/ft^2$) of ESP plate area for NDCE and DCE recovery furnaces.[4] The ESP plate area for the new ESP was derived from the model exhaust gas flow rates for RF-1a through RF-9a and the SCA for the replacement ESP.

For NDCE recovery furnaces, an SCA of approximately 100 $m^2/(m^3/sec)$ (530 ft^2/1,000 acfm) was assumed based on ESP SCA information from NDCE recovery furnaces subject to the NSPS (i.e., furnaces installed or replaced during or after 1977 and required to have outlet PM emissions of 0.10 g/dscm [0.044 gr/dscf] or lower).[4] Because dry-bottom ESP's are used to control PM emissions from approximately 80 percent of NDCE recovery furnaces, it was assumed that all replacement ESP's for NDCE recovery furnaces would be of the dry-bottom design.[8] The ESP replacement costs for the three model NDCE recovery furnaces are presented in Table 6-1. The total capital costs range from $4.12 million to $10.7 million for RF-1a through RF-6a.

For DCE recovery furnaces, an SCA of approximately 90 $m^2/(m^3/sec)$ (430 ft^2/1,000 acfm) was assumed based on ESP SCA information from DCE recovery furnaces subject to the NSPS (i.e., furnaces installed or replaced during or after 1977 and required to have outlet PM emissions of 0.10 g/dscm [0.044 gr/dscf] or lower).[4] Because wet-bottom ESP's are used to control PM emissions from approximately 90 percent of DCE recovery furnaces, it was assumed that all new replacement ESP's for DCE recovery furnaces would be of the wet-bottom design.[8] The ESP replacement costs for the three model DCE recovery furnaces are presented in Table 6-1. The TCI costs range from $2.01 million to $6.03 million for RF-7a through RF-9a.

Installing a new fan and stack typically comprises approximately 20 percent of the total capital costs of installing a new ESP.[9] Because the fan and stack usually do not need to be replaced when an ESP is replaced, the ESP replacement costs stated above are estimated to be 80 percent of the cost of a completely new ESP. The costs to replace the fan and stack due to the addition of the packed-bed scrubber are presented in Section 6.1.2.4.2.

In some cases, installation of a replacement ESP can be achieved using a "roll-in" procedure, in which the ESP is erected adjacent to its final location and then rolled into position using roller assemblies or rubber-tired dollies. The roll-in technique has been used for a number of years on a wide range of equipment sizes.[9] However, the roll-in technique probably cannot be used for the majority of recovery furnaces due to the location of the ESP (e.g., an elevated position relative to the recovery furnace). Because the ESP replacement costs were based on actual ESP replacement costs provided by individual pulp and paper mills, it was assumed that they were based on recovery furnaces that cannot do ESP roll-ins. Therefore, no contingency factor is needed to adjust ESP replacement costs to account for the furnaces being unable to do ESP roll-ins. However, without the ESP roll-in, an ESP replacement probably could not be completed within the scheduled 2-week shutdown for maintenance. Because no

information is available on the time needed to complete an ESP replacement, the maximum time estimated to complete an ESP upgrade, as described in Section 6.1.2.1.3, was used. The estimated maximum time is 30 days. Therefore, pulp production losses for replacing the ESP were calculated for that period of time minus the 2-week (14-day) scheduled mill shutdown. Pulp production losses for the 16-day period are presented in Table 6-2.

For mills producing bleached pulp, the pulp production losses range from $1.19 million to $3.12 million for RF-1a through RF-6a and $710,000 to $2.13 million for RF-7a through RF-9a. For mills producing unbleached pulp, the pulp production losses range from $844,000 to $2.19 million for RF-1a through RF-6a and $506,000 to $1.52 million for RF-7a through RF-9a. The unbleached pulp production losses are lower for each model recovery furnace because unbleached pulp has a lower market value than bleached pulp.

6.1.2.1.2 <u>ESP replacement: incremental annual costs</u>. Labor and maintenance requirements and costs are assumed to be unchanged when the ESP is replaced. However, because the ESP replacement option includes an increase in ESP plate area, electricity costs are increased. The increase in ESP electricity costs resulting from the ESP replacement is based on an increase in the SCA.

For NDCE and DCE recovery furnaces, the baseline SCA values are approximately 90 $m^2/(m^3/sec)$ (430 ft^2/1,000 acfm) and 70 $m^2/(m^3/sec)$ (330 ft^2/1,000 acfm), respectively. These SCA values are the average values common to ESP's on NDCE and DCE recovery furnaces installed prior to 1977 (i.e., installed prior to the NSPS and not subject to the NSPS PM standard of 0.10 g/dscm [0.044 gr/dscf]) that also have PM emissions greater than 0.10 g/dscm (0.044 gr/dscf).[4] As a result of the ESP replacement, the baseline SCA would be increased to a value of approximately 100 $m^2/(m^3/sec)$ (530 ft^2/1,000 acfm) for NDCE recovery furnaces and 90 $m^2/(m^3/sec)$ (430 ft^2/1,000 acfm) for DCE recovery furnaces. These SCA values are the average values

common to ESP's on NDCE and DCE recovery furnaces installed during or after 1977 (i.e., installed after the NSPS and subject to the NSPS standard of 0.10 g/dscm [0.044 gr/dscf]) that also have PM emissions less than or equal to 0.10 g/dscm (0.044 gr/dscf) but greater than 0.034 g/dscm (0.015 gr/dscf) (the more stringent PM control level).[4]

The incremental annual costs for ESP replacements include the difference in electricity costs plus the indirect costs affected by the TCI, which include the administrative, property tax, and insurance costs plus the capital recovery cost. The lifetime of dry-bottom ESP's typically ranges from 12 to 15 years.[9] Therefore, an average 13.5-year life span was assumed when calculating the capital recovery cost for dry-bottom ESP's installed on NDCE recovery furnaces. The lifetime of wet-bottom ESP's is typically 10 years.[9] Therefore, an average 10-year life span was assumed when calculating the capital recovery cost for wet-bottom ESP's installed on DCE recovery furnaces.

The ITAC for the model NDCE and DCE recovery furnaces, excluding annualized pulp production losses (i.e., capital recovery costs for the pulp production losses), are presented in Table 6-1. The incremental annual costs range from $666,000/yr to $1.73 million/yr for RF-1a through RF-6a and $378,000/yr to $1.14 million/yr for RF-7a through RF-9a. The ITAC for the model NDCE and DCE recovery furnaces, including annualized bleached and unbleached pulp production losses, are presented in Table 6-2. The incremental annual costs, including annualized bleached pulp production losses, range from $805,000/yr to $2.09 million/yr for RF-1a through RF-6a and $764,000/yr through $1.99 million/yr for RF-7a through RF-9a. The incremental annual costs, including annualized unbleached pulp production losses, range from $764,000/yr to $1.99 million/yr for RF-1a through RF-9a and $451,000/yr to $1.36 million/yr for RF-7a through RF-9a.

6.1.2.1.3 <u>ESP upgrade (to 0.10 g/dscm [0.044 gr/dscf]): capital costs</u>. The ESP upgrade costs for existing NDCE and DCE recovery furnaces are based on May 1993 ESP upgrade costs supplied by an ESP manufacturer.[9] The ESP manufacturer supplied

costs for a model NDCE recovery furnace with an exhaust gas flow rate of 109 m^3/sec (230,000 acfm) and a model DCE recovery furnace with an exhaust gas flow rate of 160 m^3/sec (340,000 acfm).[9] According to the ESP manufacturer, the proposed upgrade would result in an ESP size adequate to meet an outlet PM level of 0.10 g/dscm (0.044 gr/dscf).[9]

The ESP manufacturer supplied ESP upgrade costs for two different ESP upgrade schedules (Schedules 1 and 2). Schedule 1 is a 21-day, 7-day/week ESP upgrade schedule, including about 14 days of recovery furnace outage and about 3 or 4 days of partial load on both sides of the outage.[9] Three days of partial load before and after the outage were assumed, for a total of 6 days of partial load. Assuming a 50 percent load for 6 days, there would be 3 days of downtime during the period of partial load. Added to the 14 days of recovery furnace outage, there is a total downtime of 17 days, which is 3 days of downtime beyond the annual 2-week shutdown for maintenance. Schedule 2 is a 30-day, 6-day/week upgrade schedule and has no periods of partial load. There would be 30 days of recovery furnace outage.[9] Schedule 2 would have 16 days of downtime beyond the annual 2-week shutdown. For the model NDCE recovery furnace, the cost for Schedule 1 is $1,292,000, and the cost for Schedule 2 is $1,259,000.[9] For the model DCE recovery furnace, the cost for Schedule 1 is $1,504,750, and the cost for Schedule 2 is $1,466,500.[9]

The nature of the ESP upgrades is identical for Schedules 1 and 2. A brief summary of the required modifications is presented in Table 6-3. The ESP upgrades for these schedules include replacing the weighted wire design with a rigid electrode design.[9] Rigid electrode ESP's generally operate at a higher voltage than weighted wire ESP's and, in some cases, may require replacement of the transformer and other controls. However, the ESP upgrade cost estimate does not include those costs that vary widely based on site-specific conditions, such as the need for transformer replacement.

The only difference between the two ESP upgrade schedules is that Schedule 2 allows a longer downtime because major modifications to the furnace would be performed at the same time. The costs of the recovery furnace modification are not included because they are not associated with the ESP upgrade. The ESP upgrade work for Schedule 2 would be extended over the 30-day shutdown period.

The ESP upgrade costs were adjusted to 1991 dollars and then scaled using the six-tenths power rule described in Section 6.1.1.2 to derive costs for the model NDCE and DCE recovery furnaces. The model ESP upgrade costs, without pulp production losses, for Schedules 1 and 2 are presented in Tables 6-4 and 6-5, respectively. The ESP upgrade capital costs for Schedule 1 range from $1.20 million to $2.12 million for RF-1a through RF-6a and $811,000 to $1.57 million for RF-7a through RF-9a. The ESP upgrade capital costs for Schedule 2 range from $1.16 million to $2.07 million for RF-1a through RF-6a and $791,000 to $1.53 million for RF-7a through RF-9a.

The pulp production losses associated with shutting down the mill to upgrade the ESP were calculated using the method described in Section 6.1.1.1. The ESP upgrade costs that include bleached and unbleached pulp production losses are presented in Tables 6-6 and 6-7, respectively. The pulp production losses for Schedules 1 and 2 are presented for the model recovery furnaces in Tables 6-6 and 6-7, respectively.

The bleached pulp production losses for Schedule 1 range from $224,000 to $585,000 for RF-1a through RF-6a and $133,000 to $399,000 for RF-7a through RF-9a. The unbleached pulp production losses for Schedule 1 range from $158,000 to $411,000 for RF-1a through RF-6a and $94,900 to $285,000 for RF-7a through RF-9a. The pulp production losses for mills producing unbleached pulp are lower than for mills producing bleached pulp because unbleached pulp has a lower market value than bleached pulp.

Because Schedule 2 is longer than Schedule 1, the pulp production losses are higher for Schedule 2 than for Schedule 1. The bleached pulp production losses for Schedule 2 range from

$1.19 million to $3.12 million for RF-1a through RF-6a and $710,000 to $2.13 million for RF-7a through RF-9a. The unbleached pulp production losses for Schedule 2 range from $844,000 to $2.19 million for RF-1a through RF-6a and $506,000 to $1.52 million for RF-7a through RF-9a.

6.1.2.1.4 ESP upgrade (to 0.10 g/dscm [0.044 gr/dscf]): incremental annual costs. Labor and maintenance costs are assumed to be unchanged when the ESP is upgraded to achieve a PM level of 0.10 g/dscm (0.044 gr/dscf). Therefore, the incremental annual costs for the ESP upgrade include only the electricity costs and the TCI-based indirect annual costs, which include administrative, property tax, insurance, and capital recovery costs. The capital recovery cost for NDCE recovery furnace ESP's is based on an average 13.5-year life span for dry-bottom ESP's.[9] The capital recovery cost for DCE recovery furnace ESP's is based on an average 10-year life span for wet-bottom ESP's.[9] Although the ESP plate area is not increased with an ESP upgrade, there will be additional electricity costs associated with the new ESP design. The actual increase in electricity costs with an ESP upgrade is not currently known. Therefore, it was assumed that the incremental increase in electricity costs for an upgraded ESP would be the same as for a replacement ESP. Because the electricity cost increases for the replacement ESP in Section 6.1.2.1.2 were based on an increase in SCA from baseline to control levels, the electricity cost increases for the upgraded ESP were calculated in the same manner.[4] The ITAC for the model recovery furnaces are presented in Tables 6-4 through 6-7.

Table 6-4 presents the ITAC for the Schedule 1 ESP upgrade for each model recovery furnace, without accounting for annualized pulp production losses. As shown in the table, the ITAC for the Schedule 1 ESP upgrade, without annualized pulp production losses, range from $207,000/yr to $383,000/yr for RF-1a through RF-6a and $159,000/yr to $322,000/yr for RF-7a through RF-9a. Table 6-5 presents the ITAC for the Schedule 2 ESP upgrade for each model recovery furnace, without accounting

for annualized pulp production losses. As shown in the table, the ITAC for the Schedule 2 ESP upgrade, without annualized pulp production losses, range from $202,000/yr to $375,000/yr for RF-1a through RF-6a and $156,000/yr to $314,000/yr for RF-7a through RF-9a.

Table 6-6 presents the ITAC for the Schedule 1 ESP upgrade for each model recovery furnace and includes estimated production losses associated with bleached and unbleached pulp. As shown in the table, the ITAC for the Schedule 1 ESP upgrade, including bleached pulp production losses, range from $233,000/yr to $452,000/yr for RF-1a through RF-6a and $178,000/yr to $379,000/yr for RF-7a through RF-9a. The ITAC for the Schedule 1 ESP upgrade, including unbleached pulp production losses, range from $226,000/yr to $431,000/yr for RF-1a through RF-6a and $173,000/yr to $363,000/yr for RF-7a through RF-9a.

Table 6-7 presents the ITAC for the Schedule 2 ESP upgrade for each model recovery furnace and includes estimated production losses associated with bleached and unbleached pulp. As shown in the table, the ITAC for the Schedule 2 ESP upgrade, including bleached pulp production losses, range from $341,000/yr to $740,000/yr for RF-1a through RF-6a and $257,000/yr to $617,000/yr for RF-7a through RF-9a. The ITAC for the Schedule 2 ESP upgrade, including unbleached pulp production losses, range from $300,000/yr to $631,000/yr for RF-1a through RF-6a and $228,000/yr to $530,000/yr for RF-7a through RF-9a.

6.1.2.1.5 <u>ESP upgrade (to 0.034 g/dscm [0.015 gr/dscf]): capital costs</u>. Control costs have been determined for those new and existing recovery furnaces controlling PM emissions from the NSPS level of 0.10 g/dscm (0.044 gr/dscf) to 0.034 g/dscm (0.015 gr/dscf). These costs would include an ESP upgrade cost and a packed-bed scrubber cost. Because no actual ESP upgrade costs were available for controlling PM emissions to 0.034 g/dscm (0.015 gr/dscf), the upgrade cost for the recovery furnace ESP was instead based on the incremental cost difference between an ESP capable of achieving a PM level less than or equal to 0.10 g/dscm (0.044 gr/dscf) but greater than 0.034 g/dscm

(0.015 gr/dscf) and an ESP capable of achieving a PM level less than or equal to 0.034 g/dscm (0.015 gr/dscf). This section presents the ESP upgrade costs; the packed-bed scrubber costs are presented in Section 6.1.2.4.

The ESP upgrade capital costs were based on recent ESP costs provided by individual pulp and paper mills (i.e., costs for ESP's installed or replaced during or after 1989).[4] The ESP costs average \$420/m^2 (\$39/ft^2) of ESP plate area for recovery furnaces.[4] To determine the ESP upgrade cost, this cost per ESP plate area was multiplied by the increase in ESP plate area assumed to reduce PM emissions from 0.10 g/dscm (0.044 gr/dscf) to 0.034 g/dscm (0.015 gr/dscf).

The ESP plate area for NDCE recovery furnace ESP's achieving a PM level less than or equal to 0.10 g/dscm (0.044 gr/dscf) but greater than 0.034 g/dscm (0.015 gr/dscf) is based on an average SCA of approximately 100 $m^2/(m^3/sec)$ (530 ft^2/1,000 acfm), as discussed in Section 6.1.2.1.2. The ESP plate area for DCE recovery furnace ESP's achieving a PM level less than or equal to 0.10 g/dscm (0.044 gr/dscf) but greater than 0.034 g/dscm (0.015 gr/dscf) is based on an average SCA of approximately 90 $m^2/(m^3/sec)$ (430 ft^2/1,000 acfm), as discussed in Section 6.1.2.1.2.[4] The ESP plate area for recovery furnace ESP's achieving a PM level less than or equal to 0.034 g/dscm (0.015 gr/dscf) is based on an SCA of approximately 120 $m^2/(m^3/sec)$ (620 ft^2/1,000 acfm). This is the SCA value for an ESP achieving a PM emission level of 0.034 g/dscm (0.015 gr/dscf) on a long-term basis.[4,10]

The capital cost attributable to the control option to install new recovery furnace ESP's capable of achieving a PM level of 0.034 g/dscm (0.015 gr/dscf) would not be the cost of a new ESP, but only that portion associated with controlling PM emissions from 0.10 g/dscm (0.044 gr/dscf) to 0.034 g/dscm (0.015 gr/dscf), which is the same as the cost to upgrade existing recovery furnace ESP's to achieve the same PM level of 0.034 g/dscm (0.015 gr/dscf).

The capital costs of the ESP upgrade used to achieve the 0.034 g/dscm (0.015 gr/dscf) PM level are lower than the capital costs of the ESP upgrade used to achieve the 0.10 g/dscm (0.044 gr/dscf) NSPS PM level discussed in Section 6.1.2.1.3. The capital costs for this more stringent ESP upgrade control option, excluding pulp production losses, are presented in Table 6-8 and range from $644,000 to $1.67 million for RF-1b through RF-6b and $387,000 to $1.16 million for RF-7b through RF-9b.

There are no pulp production losses for new recovery furnaces to install an upgraded ESP because the ESP is upgraded prior to installation. No information is currently available for existing recovery furnaces on the amount of time required to complete an ESP upgrade that would allow PM control to 0.034 g/dscm (0.015 gr/dscf). However, it was assumed that the pulp production losses could be as high as those for Schedule 1 of the ESP upgrade control option discussed in Section 6.1.2.1.3. The pulp production losses are presented in Table 6-9.

6.1.2.1.6 <u>ESP upgrade (to 0.034 g/dscm [0.015 gr/dscf]): incremental annual costs</u>. Labor and maintenance costs are assumed to be unchanged when the ESP is upgraded to achieve a PM level of 0.034 g/dscm (0.015 gr/dscf). Therefore, the incremental annual costs for the ESP upgrade include only the electricity costs and TCI-based indirect annual costs, which include administrative, property tax, insurance, and capital recovery costs. The capital recovery cost for NDCE recovery furnace ESP's is based on an average 13.5-year life span for dry-bottom ESP's. The capital recovery cost for DCE recovery furnace ESP's is based on an average 10-year life span for wet-bottom ESP's. The PM control electricity costs would include both ESP upgrade and packed-bed scrubber electricity costs. The ESP upgrade electricity costs are presented in this section; Section 6.1.2.4 presents the packed-bed scrubber electricity costs. The ESP upgrade electricity costs are based on an increase in SCA from approximately 100 $m^2/(m^3/sec)$ (530 ft^2/ 1,000 acfm) for NDCE recovery furnaces and 90 $m^2/(m^3/sec)$

(430 ft^2/1,000 acfm) for DCE recovery furnaces to approximately 120 $m^2/(m^3/sec)$ (620 ft^2/1,000 acfm) for all recovery furnaces.[4]

The electricity costs presented in this section are based on gas flow rates and pressure drops in the absence of a packed-bed scrubber. If a packed-bed scrubber was added after the ESP, the fan electricity costs for the ESP would change slightly from those presented in this section. However, because the total electricity costs are only a small fraction (approximately 10 percent) of the ITAC, a slight change in fan electricity costs would have a negligible effect on the ITAC.

The ITAC for the model recovery furnaces, excluding annualized pulp production losses, are presented in Table 6-8 and range from $117,000/yr to $304,000/yr for RF-1b through RF-6b and $80,300/yr to $241,000/yr for RF-7b through RF-9b. The ITAC for the model recovery furnaces, including annualized pulp production losses, are presented in Table 6-9. The incremental annual costs, including annualized bleached pulp production losses, range from $143,000/yr to $373,000/yr for RF-1b through RF-6b and $99,000/yr to $298,000/yr for RF-7b through RF-9b. The incremental annual costs, including annualized unbleached pulp production losses, range from $136,000/yr to $352,000/yr for RF-1b through RF-6b and $94,000/yr to $282,000/yr for RF-7b through RF-9b.

6.1.2.2 Wet to Dry ESP System Conversion. Two control options were evaluated for reducing emissions of gaseous organic HAP's such as methanol from existing NDCE recovery furnaces. These control options are (1) converting an ESP system that uses black liquor or HAP-contaminated process water in the ESP bottom or PM return system (referred to as a wet ESP system) to an ESP system that uses "clean" water (i.e., water uncontaminated with methanol and other gaseous organic HAP's) in the ESP bottom or PM return system; and (2) converting a wet ESP system to a dry-bottom ESP with a dry PM return system (referred to as a dry ESP system). With these two control options, the potential stripping of methanol and other gaseous organic HAP's from the black liquor

or contaminated process water in the ESP system would be eliminated.

Only the costs for the second control option, converting from a wet to a dry ESP system, were evaluated. This decision was based on (1) the uncertainty associated with the available cost estimates for converting to an ESP system that uses "clean" water in the ESP bottom or PM return system; and (2) the fact that very few mills use water in the ESP system.

A cost estimate is available for converting a wet-bottom ESP that uses black liquor in the ESP bottom to one that uses "clean" water in the ESP bottom. This cost estimate is significantly lower than the cost estimates available for converting to a dry-bottom ESP, but the accuracy of this cost estimate is questionable. According to a 1985 EPA estimate for a 900 ADMP/d (1,000 ADTP/d) mill, the capital cost to convert a wet-bottom ESP to one that uses water in the ESP bottom is $154,000. The annual cost is $67,000/yr.[11] These costs are lower than the wet- to dry-bottom conversion costs presented below. Several pulp and paper industry representatives commenting on the estimate indicated that the costs to evaporate the added water should be higher. They also noted that mills may experience a loss of production to evaporate the extra water if excess capacity was not available in the evaporators.[12]

The wet to dry ESP system conversion control option applies to model NDCE recovery furnaces RF-4 through RF-6, which represent existing NDCE recovery furnaces with wet ESP systems. These models represent existing NDCE recovery furnaces only, because no wet ESP systems are expected to be installed on new NDCE recovery furnaces.

The model costs for the wet to dry ESP system conversion control option are based on costs to convert NDCE recovery furnace wet-bottom ESP's to dry-bottom ESP's. For the purposes of this cost analysis, these wet- to dry-bottom ESP conversion costs are assumed to apply also to those NDCE recovery furnaces with dry-bottom ESP's and wet PM return systems. The ESP conversion costs may be lower than those presented if an ESP

upgrade to improve PM collection is also performed at the same time. However, no information is currently available on the extent of the cost reduction. The capital and annual costs for the wet to dry ESP system conversion control option are presented in the following sections for existing model NDCE recovery furnaces.

6.1.2.2.1 Capital costs. The wet to dry ESP system conversion capital costs are based on 1993 conversion costs from an ESP manufacturer.[13] These costs were adjusted to 1991 dollars and then scaled, using the six-tenths power rule, to derive costs for the three model NDCE recovery furnaces. The ESP model used by the manufacturer to develop the conversion costs was stated as being three fields in length and two chambers in width, each 6.1 m (20 ft) wide. The cost to remove the existing agitator paddles and liquor piping and install a perpendicular drag scraper system, shallow fallout hoppers, drag chain conveyors, and rotary valves was estimated to be $560,000 for the material and $285,000 for installation. The cost estimate does not include the cost associated with (1) any removal of asbestos, if applicable; (2) any piping beyond the rotary valves; or (3) any equipment beyond the rotary valves, such as an ash mixing tank with associated instrumentation. The ESP conversion costs from the ESP manufacturer are based on working two 10-hour shifts for about 10 days and converting both ESP chambers simultaneously.[13] Therefore, no downtime would be necessary beyond the annual 2-week shutdown, which means no pulp production losses would need to be included in the model ESP conversion costs presented below.

The recovery furnace size used by the ESP manufacturer in calculating the wet-bottom ESP conversion costs was stated to be about 600 to 900 ADMP/d (700 to 1,000 ADTP/d)[13]. A size of 800 ADMP/d (900 ADTP/d) (the size for the mid-size model NDCE recovery furnace RF-5) was used to scale the costs. The capital costs for the three model NDCE recovery furnaces are presented in Table 6-10. The wet to dry ESP system conversion costs range from $596,000 to $1.06 million for RF-4 through RF-6.

6.1.2.2.2 Incremental annual costs. Direct annual costs are not assumed to increase as a result of the wet to dry ESP system conversion. Although there may be some extra maintenance costs, they are expected to be small compared to the increases in capital recovery and other indirect costs. Past dry-bottom ESP designs were associated with higher maintenance costs. Changes in designs have eliminated many of those problems.[14] The costs from the ESP manufacturer used to develop the model costs are based on the modern design. Furthermore, because wet-bottom ESP designs are associated with greater corrosion, switching to dry-bottom ESP designs results in a longer life span for the ESP. Utility costs (i.e., electricity) also do not change significantly because of an equal trade-off in horsepower requirements between the wet and dry ESP system designs.[14]

Based on these assumptions, the ITAC for wet to dry ESP system conversions should only include those indirect annual costs affected by the TCI (i.e., administrative, property tax, insurance, and capital recovery costs). Similar to the ESP replacement costs, the capital recovery cost is also based on an average 13.5-year life for dry-bottom ESP's operating on NDCE recovery furnaces. The ITAC for the model NDCE recovery furnaces are presented in Table 6-10 and range from $93,500/yr to $166,000/yr for RF-4 through RF-6.

6.1.2.3 Conversion of a DCE Recovery Furnace System to an NDCE Recovery Furnace. Converting a DCE recovery furnace system to an NDCE recovery furnace (or "low-odor conversion") was evaluated as a control option for reducing gaseous organic HAP emissions from DCE recovery furnace systems. The conversion of a DCE recovery furnace system to an NDCE design involves removing the DCE and BLO unit, adding a concentrator, and extending or replacing the boiler economizer. Capital and annual costs have been evaluated for these three tasks. Additional upgrades are included in the low-odor conversion control option, i.e., an ESP upgrade to improve PM collection and a wet to dry ESP system conversion to reduce gaseous organic HAP emissions. Separate capital and annual costs were developed for the ESP upgrade and

the wet to dry ESP system conversion. The ESP conversion costs may be lower if an ESP upgrade is also performed at the same time as the ESP conversion. However, no information is currently available on the extent of the cost reduction. Therefore, the costs were developed in the same way as those developed for recovery furnaces in Sections 6.1.2.1 and 6.1.2.2.

Often other upgrades are performed at the same time as a low-odor conversion. These upgrades usually provide additional cost savings because of increased efficiency, increased process capacity, and improved performance and safety. Possible upgrades include combustion air system improvements, composite tubing, and emergency drain and flame safety systems.[15] The capital costs and annual cost savings associated with these additional upgrades have not been evaluated as part of the low-odor conversion control option. The pulp production credits associated with increased process capacity were not included in the low-odor conversion total annual cost estimate because the additional capacity increases may require significant modifications (e.g., expanding the recovery furnace bed or modifying the air system), which would require additional capital expenses.

The low-odor conversion total annual cost estimates also do not include (1) DCE maintenance cost savings, (2) ESP maintenance cost savings, and (3) higher solids firing cost benefits for the reasons described below.

Maintenance requirements associated with the DCE are eliminated with the removal of this piece of equipment during a low-odor conversion. The lower maintenance requirements associated with an NDCE recovery furnace increase furnace availability, which allows for higher utilization of recovery furnace capacity without additional costs. The DCE maintenance cost savings were not included in the low-odor conversion cost estimates because sufficient data are not available to quantify the cost savings.

Lower corrosion rates are associated with NDCE recovery furnace ESP's than with DCE recovery furnace ESP's. The lower corrosion rates for NDCE recovery furnace ESP's are the result of

a higher-temperature, lower-moisture content gas stream from NDCE recovery furnaces compared to DCE recovery furnaces, as well as from the predominant use of dry-bottom ESP's. Therefore, converting to an NDCE design would eliminate the need for more frequent ESP replacement resulting from ESP corrosion. The ESP maintenance cost savings were not included in the low-odor conversion cost estimates because sufficient data are not available to quantify the cost savings.

Because concentrators can achieve higher BLS concentrations than DCE's (i.e., 75 to 80 percent vs. 65 percent), converting to an NDCE recovery furnace provides the mill with an opportunity to increase the solids content of the black liquor fired in the furnace.[4,16] However, increasing the solids content of the black liquor to the upper limits requires additional capital expenses, such as modifications to the fuel delivery system to handle a more viscous liquid. Neither the potential cost credits nor the additional capital expenses and any associated maintenance costs associated with higher solids firing are included in the low-odor conversion cost estimates.

Particulate matter control costs are included in the low-odor conversion cost estimates. With the removal of the DCE, which provides some PM control, as stated in Chapter 3, the ESP often must be upgraded or replaced during a low-odor conversion in order to meet applicable PM emission limits. For the purposes of this cost analysis, an ESP upgrade PM control option that would maintain or reduce PM emissions to the NSPS level of 0.10 g/dscm (0.044 gr/dscf) has been evaluated for those existing DCE recovery furnaces that have PM emissions at or above the NSPS level. This PM control option applies to model DCE recovery furnaces RF-7a/7b through RF-9a/9b. These models represent existing sources only, because no new DCE recovery furnaces are expected to be built.

A PM control option that would reduce PM emissions to 0.034 g/dscm (0.015 gr/dscf) has also been evaluated for DCE recovery furnaces that have PM emissions at or below the NSPS level. This PM control option includes an ESP upgrade coupled

with the addition of a packed-bed scrubber and applies to model DCE recovery furnaces RF-7b through RF-9b, which, as cited above, represent only existing sources.

The capital and annual costs for low-odor conversions are presented in the following sections for model DCE recovery furnaces.

6.1.2.3.1 Capital costs: introduction. The total capital cost of the low-odor conversion control option includes purchase and installation costs of the extended economizer with associated soot blowers and ash handling equipment; demolition costs for the DCE and BLO unit; purchase and installation costs of the black liquor concentrator; ESP upgrade capital costs; and wet to dry ESP system conversion capital costs.

6.1.2.3.2 Capital costs: economizer expansion, demolition, and concentrator. The cost of economizer expansion and demolition has been estimated at $6.5 million for a mid-size recovery furnace. The cost estimate is based on cost data from three sources--two recovery furnace manufacturers and one kraft pulp mill, at which three low-odor conversions were completed over a 3-year period. A 20 percent contingency factor was added to the supplier costs to account for site-specific tie-in work. Where applicable, the available cost data were adjusted to 1991 dollars and scaled for a 0.7 million kg BLS/d (1.5 million lb BLS/d) furnace using the six-tenths power rule. The $6.5 million cost estimate is the average of the adjusted costs from the three sources--$6 million, $4.8 million, and $8.6 million.[4,17,18]

The concentrator costs are based on cost information from a concentrator manufacturer. The equipment costs for a falling film concentrator range from $1.5 to $3 million. Total capital costs, including installation, are approximately two times the equipment cost.[19] Based on this information, the average total capital cost estimate for a concentrator is approximately $4.5 million. The concentrator capital cost estimate does not include liquor storage and piping costs or the cost for the addition of a cooling tower cell.

The total economizer expansion, demolition, and concentrator cost for the mid-size model DCE recovery furnace (i.e., RF-8) is approximately $11 million, equal to the sum of the $6.5 million for the economizer expansion and demolition and the $4.5 million for the concentrator. The six-tenths power rule was used to calculate the capital costs for the small and large model DCE recovery furnaces (i.e., RF-7 and RF-9). The low-odor conversion capital costs (excluding the pulp production losses) are presented in Table 6-11. The capital costs, excluding the ESP upgrade and wet to dry ESP system conversion costs, range from $8.09 million to $15.7 million for model DCE recovery furnaces RF-7 through RF-9.

6.1.2.3.3 Capital costs: ESP upgrade. Electrostatic precipitator upgrade costs to achieve total outlet PM emissions of 0.10 g/dscm (0.044 gr/dscf) have been determined for the applicable DCE recovery furnace models. The ESP upgrade capital costs to control PM to NSPS levels for model DCE recovery furnaces RF-7a through RF-9a are derived from the Schedule 1 ESP manufacturer costs discussed in Section 6.1.2.1.3 and presented in Table 6-4.[9] The Schedule 1 costs were chosen because the ESP upgrade could be completed within the scheduled time for the low-odor conversion. The six-tenths power rule was used to calculate the capital costs for the model DCE recovery furnaces. The ESP upgrade costs to control PM emissions to NSPS levels for the model DCE recovery furnaces are presented in Table 6-11, excluding pulp production losses. The ESP upgrade costs range from $881,000 to $1.70 million for RF-7a through RF-9a. The bleached and unbleached pulp production losses are presented in Table 6-12. Although NSPS PM emission levels are associated with model DCE recovery furnaces RF-7b through RF-9b at baseline, once they are converted to NDCE recovery furnaces, the ESP's must be upgraded in order to maintain PM emissions at NSPS levels. The ESP upgrade costs presented above for model furnaces RF-7a through RF-9a were applied, as a worst-case cost estimate, for model DCE recovery furnaces RF-7b through RF-9b.

Control costs to achieve total outlet PM emissions of 0.034 g/dscm (0.015 gr/dscf) have been determined for the applicable DCE recovery furnace models. These PM control costs would include both ESP upgrade costs and packed-bed scrubber costs. This section presents the ESP upgrade costs; the packed-bed scrubber costs are presented in Section 6.1.2.4. The capital costs for an ESP upgrade to control PM emissions to 0.034 g/dscm (0.015 gr/dscf) were estimated by summing the Schedule 1 ESP upgrade costs presented above and the ESP upgrade costs presented in Table 6-8 and discussed in Section 6.1.2.1.5. These costs apply to model DCE recovery furnaces RF-7a through RF-9a. The ESP upgrade costs for model DCE recovery furnaces are presented in Table 6-13, excluding pulp production losses, and range from $9.80 million to $19.4 million for RF-7a through RF-9a. The bleached and unbleached pulp production losses are presented in Table 6-14. Using the same reasoning stated in the previous paragraph, the ESP upgrade costs presented above for model recovery furnaces RF-7a through RF-9a were applied, as a worst-case cost estimate, for model DCE recovery furnaces RF-7b through RF-9b.

6.1.2.3.4 Capital costs: wet to dry ESP system conversion. The wet to dry ESP system conversion costs for model DCE recovery furnaces RF-7 through RF-9 are based on ESP manufacturer costs presented in Table 6-10 and discussed in Section 6.1.2.2.1 for converting NDCE recovery furnace wet ESP systems to the dry ESP system design.[13] The ESP manufacturer costs were converted to 1991 dollars and then scaled for the model DCE recovery furnaces using the six-tenths power rule. The conversion costs include the costs to remove the existing agitator paddles and liquor piping and install a perpendicular drag scraper system, shallow fallout hoppers, drag chain conveyors, and rotary valves but do not include the costs for asbestos removal or equipment or piping beyond the rotary valves (e.g., an ash mixing tank and associated equipment).[13] The ESP system conversion costs for the model DCE recovery furnaces are presented in Table 6-11 and range from $439,000 to $849,000 for RF-7 through RF-9.

6.1.2.3.5 Capital costs: total costs. The total capital costs for the low-odor conversion option are equal to the sum of the economizer expansion, demolition, and concentrator costs plus the ESP upgrade and wet to dry ESP system conversion costs. These costs, excluding pulp production losses, are presented in Table 6-11 and range from $9.41 million to $18.2 million for RF-7a/7b through RF-9a/9b.

6.1.2.3.6 Capital costs: pulp production losses. The production losses attributed to a low-odor conversion are site-specific and depend on factors such as liquor storage capacity, liquor trade or sell options, and coordination with scheduled mill shutdowns.[15] Pulp production losses were calculated assuming an average additional shutdown period of 11 days beyond the scheduled 2-week shutdown period (i.e., a total of 25 days of downtime). The pulp production losses were calculated using the market values of bleached and unbleached pulp discussed in Section 6.1.1.1 and an earnings margin of 25 percent.[3]

The average 25-day shutdown was estimated based on the following information:

1. A time frame for completion of 21 to 30 days with proper pre-shutdown planning and prefabrication;[15]

2. A case study where one mill completed three low-odor conversions over a 4-week outage (i.e., 31 days);[20] and

3. A case study that involved two shutdowns; the first shutdown was for several days to relocate ductwork, and the second was for approximately 10 days to tie in the new systems.[21]

The total capital costs that include pulp production losses associated with the low-odor conversion are presented in Tables 6-12 and 6-14. For mills producing bleached pulp, the pulp production losses associated with a low-odor conversion range from $488,000 to $1.46 million for RF-7a/7b through RF-9a/9b. For mills producing unbleached pulp, the pulp production losses associated with a low-odor conversion range from $348,000 to $1.04 million for RF-7a/7b through RF-9a/9b.

6.1.2.3.7 Incremental annual costs: introduction. The ITAC estimate includes the following six primary components:

1. Capital recovery;
2. Administrative costs, taxes, and insurance;
3. Steam production credits;
4. Operating cost savings for the BLO;
5. Concentrator steam costs; and
6. Operating electricity costs for the ESP.

The annual costs of these six primary components were estimated for the three model DCE recovery furnaces and summed to determine the ITAC. As discussed in Section 6.1.2.3, certain other potential cost savings were not quantified. These potential cost savings include (1) maintenance cost savings resulting from eliminating the DCE; (2) ESP replacement cost savings resulting from a less corrosive exit gas stream associated with NDCE recovery furnaces; and (3) pulp production credits for those mills that choose to provide for additional capacity during the low-odor conversion.

6.1.2.3.8 Incremental annual costs: capital recovery. For a low-odor conversion, the capital recovery costs for the model DCE recovery furnaces are based on the following:

1. An equipment life of 20 years;
2. The model capital costs presented in Tables 6-11 through 6-14;
3. The model bleached and unbleached pulp production losses incurred during construction, which are presented in Tables 6-12 and 6-14; and
4. An interest rate of 7 percent.

Total capital recovery costs were calculated for each PM control level for the following three scenarios:

1. Without annualized pulp production losses (Scenario 1);
2. With annualized bleached pulp production losses (Scenario 2); and
3. With annualized unbleached pulp production losses (Scenario 3).

Design parameters for each model furnace are presented in Table 6-15.

The low-odor conversion capital recovery costs for Scenarios 1, 2, and 3 for model DCE recovery furnaces (including an ESP upgrade to control PM emissions to NSPS levels) are presented in Tables 6-16, 6-17, and 6-18, respectively. For Scenario 1, capital recovery costs range from $918,000/yr to $1.78 million/yr for RF-7a/7b through RF-9a/9b. For Scenario 2, capital recovery costs range from $964,000/yr to $1.92 million/yr for RF-7a/7b through RF-9a/9b. For Scenario 3, capital recovery costs range from $951,000/yr to $1.88 million/yr for RF-7a/7b through RF-9a/9b.

The low-odor conversion capital recovery costs for Scenarios 1, 2, and 3 for model DCE recovery furnaces (including an ESP upgrade to control PM emissions to 0.034 g/dscm [0.015 gr/dscf]) are presented in Tables 6-19, 6-20, and 6-21, respectively. For Scenario 1, capital recovery costs range from $963,000/yr to $1.91 million/yr for RF-7a/7b through RF-9a/9b. For Scenario 2, capital recovery costs range from $1.01 million/yr to $2.05 million/yr for RF-7a/7b through RF-9a/9b. For Scenario 3, capital recovery costs range from $996,000/yr to $2.01 million/yr for RF-7a/7b through RF-9a/9b.

6.1.2.3.9 Incremental annual costs: administrative, taxes, and insurance costs. Administrative, tax, and insurance costs were estimated as 4 percent of the TCI and are presented in each of the low-odor conversion annual cost tables, starting with Table 6-16.[6]

6.1.2.3.10 Incremental annual costs: steam production credits. Steam production credits result from the improved steam flow that occurs with a low-odor conversion.[22] The steam production credit is assumed to be equal to the cost of the power boiler fuel that has been displaced by black liquor in steam generation. It was assumed that mills would first reduce the use of higher-cost power boiler fuels, i.e., natural gas or fuel oil. Therefore, only the reduction in the use of natural gas and fuel oil was considered in determining the displaced fuel cost and not

reductions in the use of lower-cost hogged wood or coal. To determine the displaced fuel cost, the increase in the thermal efficiency that results from a low-odor conversion was estimated.

The increase in the thermal efficiency that results from a low-odor conversion is estimated to be 10 percentage points. The supporting data for this estimate are as follows:

1. Direct contact evaporator recovery furnaces operate at thermal efficiencies of 53 to 58 percent, whereas NDCE recovery furnaces operate at thermal efficiencies of 63 to 68 percent;[15] and

2. The thermal efficiency of a recovery furnace increases approximately 1 percentage point for every 22°C (40°F) drop in exit temperature.[16] The difference in the exit flue gas temperatures before and after a low-odor conversion is about 200°C (400°F), which corresponds to a 10 percentage point increase in thermal efficiency. For DCE recovery furnaces, the exit flue gas temperature is 371°C to 427°C (700°F to 800°F).[15] This high temperature range is needed to operate the DCE. For NDCE recovery furnaces, the design exit flue gas temperature range is 177°C to 357°C (350°F to 375°F).[16] The exit flue gas temperature for NDCE recovery furnaces is limited by the optimum operable range for the ESP (163°C to 204°C [325°F to 400°F]) and recovery furnace operating and design parameters (163°C [325°F] minimum).[16]

The steam production credit estimates are presented in each of the low-odor conversion annual cost tables, starting with Table 6-16. If natural gas is the displaced steam generation fuel, the average steam production credit is estimated to range from $758,000/yr to $2.27 million/yr for RF-7a/7b through RF-9a/9b. If oil is the displaced fuel, the average steam production credit is estimated to range from $1.19 million/yr to $3.58 million/yr for RF-7a/7b through RF-9a/9b. These estimates are based on the following information:

1. A thermal efficiency increase of 10 percentage points (from 56 to 66 percent);

2. Model BLS firing rates of 0.7, 1.2, and 1.8 million kg BLS/d (0.9, 1.5, and 2.7 million lb BLS/d), with a BLS heat content of 13,900 kJ/kg (6,000 Btu/lb);[8]

3. A natural gas heat content of 38,100 kJ/m^3 (1,024 Btu/ft^3) and cost of \$0.12/m^3 (\$3.48/1,000 ft^3);[23,24]

4. A fuel oil heat content of 40,400 kJ/L (145,000 Btu/gal) and cost of \$0.20/L (\$0.77/gal);[23,25] and

5. A power boiler thermal efficiency of 85 percent.[21]

6.1.2.3.11 Incremental annual costs: BLO operating cost savings. The high operating costs associated with air-sparging BLO units are eliminated with the removal of the BLO unit during a low-odor conversion. Most of the BLO operating costs (about 60 percent) is for power to operate the blowers and pumps. The remaining 40 percent is for operating the reheater.[26] These cost savings are included as a credit in the total annual cost estimate. The annual operating costs of a BLO system that oxidizes black liquor for a DCE recovery furnace range from \$103,000 to \$309,000 for RF-7 through RF-9. The cost savings from removal of the BLO unit are presented in each of the low-odor conversion annual cost tables, beginning with Table 6-16. The cost savings are based on total annual costs of \$251,900/yr for a BLO unit that oxidizes black liquor fired in two DCE recovery furnaces with a total black liquor firing rate of 1.0 million kg BLS/d (2.2 million lb BLS/d)[4].

6.1.2.3.12 Incremental annual costs: concentrator steam costs. The concentrator that replaces the DCE in a low-odor conversion uses low-pressure steam to evaporate moisture from the black liquor. The vapor from the concentrator can be used for additional evaporation at lower black liquor solids levels, or can be used to heat water. Concentrator steam costs were estimated for each model DCE recovery furnace, assuming that steam usage is proportional to the amount of black liquor concentrated. The annual concentrator steam costs for the model recovery furnaces range from \$57,800 to \$173,000 for RF-7 through RF-9. The concentrator steam costs are presented in each of the low-odor conversion annual cost tables, beginning with

Table 6-16. Concentrator steam costs are based on the following information:

1. A steam requirement of approximately 4,500 kg/hr (10,000 lb/hr) of low-pressure steam for a 1.1 million kg BLS/d (2.4 million lb BLS/d) furnace;[27] and

2. A cost of $4.02/Mg ($3.65/ton) for low-pressure steam.[28]

6.1.2.3.13 Incremental annual costs: electricity costs. The increase in electricity costs for upgrading the ESP to maintain or reduce PM emissions to the NSPS level of 0.10 g/dscm (0.044 gr/dscf) was estimated for model DCE recovery furnaces RF-7a/7b through RF-9a/9b. The actual increase in electricity costs for the ESP upgrade is not known. It was assumed that the incremental increase in electricity costs for an upgraded ESP would be the same as for a replacement ESP. Because the electricity cost increases for the replacement ESP in Section 6.1.2.1.2 were based on an increase in SCA, the electricity cost increases for the upgraded ESP were calculated in the same manner.

For DCE recovery furnaces RF-7a through RF-9a, with baseline PM emissions above 0.10 g/dscm (0.044 gr/dscf), the ESP electricity costs were estimated based on an increase in SCA values from approximately 70 $m^2/(m^3/sec)$ (330 ft^2/1,000 acfm) to approximately 100 $m^2/(m^3/sec)$ (530 ft^2/1,000 acfm).[4] For DCE recovery furnaces RF-7b through RF-9b, with baseline PM emissions less than or equal to 0.10 g/dscm (0.044 gr/dscf) but greater than 0.034 g/dscm (0.015 gr/dscf), the ESP electricity costs were estimated based on an increase in SCA values from approximately 90 $m^2/(m^3/sec)$ (430 ft^2/1,000 acfm) to approximately 100 $m^2/(m^3/sec)$ (530 ft^2/1,000 acfm).[4] The increase in ESP electricity costs resulting from the maintenance or control of PM emissions to the NSPS level are presented in Tables 6-16 through 6-18 for each of the model DCE recovery furnaces and range from $23,300/yr to $70,000/yr for RF-7a through RF-9a and $11,700/yr to $35,000/yr for RF-7b through RF-9b.

The increase in electricity costs resulting from the implementation of PM controls to reduce PM emissions to

0.034 g/dscm (0.015 gr/dscf) was estimated for model DCE recovery furnaces RF-7a/7b through RF-9a/9b. The PM control electricity costs would include both ESP upgrade and packed-bed scrubber electricity costs. The ESP upgrade electricity costs are presented in this section; Section 6.1.2.4 presents the packed-bed scrubber electricity costs.

For DCE recovery furnaces RF-7a through RF-9a, with baseline PM emissions above 0.10 g/dscm (0.044 gr/dscf), the increase in ESP electricity costs is based on an increase in the SCA from a baseline value of approximately 70 $m^2/(m^3/sec)$ (330 ft^2/1,000 acfm) to approximately 120 $m^2/(m^3/sec)$ (620 ft^2/1,000 acfm).[4] For DCE recovery furnaces RF-7b through RF-9b, with baseline PM emissions less than or equal to 0.10 g/dscm (0.044 gr/dscf) but greater than 0.034 g/dscm (0.015 gr/dscf), the ESP electricity costs were estimated based on an increase in SCA values from approximately 90 $m^2/(m^3/sec)$ (430 ft^2/1,000 acfm) to approximately 120 $m^2/(m^3/sec)$ (620 ft^2/1,000 acfm).[4] The increase in ESP electricity costs resulting from the control of PM emissions to 0.034 g/dscm (0.015 gr/dscf) are presented in Tables 6-19 through 6-21 for each of the model DCE recovery furnaces and range from $33,100/yr to $99,200/yr for RF-7a through RF-9a and $21,400/yr to $64,200/yr for RF-7b through RF-9b.

As stated in Section 6.1.2.2.2, electricity costs do not change significantly when a wet ESP system is converted to the dry ESP system design because of an equal trade-off in horsepower requirements between the wet and dry ESP system designs. Therefore, no electricity costs for the wet to dry ESP system conversion are presented in this cost analysis.

6.1.2.3.14 <u>Incremental annual costs: total costs</u>. Incremental total annual costs were estimated for converting model DCE recovery furnaces to an NDCE design for each of the PM control levels, each of the scenarios, and each of the displaced fuels. The scenarios were discussed in Section 6.1.2.3.8. Scenario 1 excludes annualized pulp production losses from the ITAC; Scenario 2 includes annualized bleached pulp production

losses in the ITAC; and Scenario 3 includes annualized unbleached pulp production losses in the ITAC.

The ITAC estimates for Scenario 1 for DCE recovery furnaces with controlled PM emissions at NSPS levels are presented in Table 6-16. For Scenario 1, with natural gas as the displaced steam generation fuel, the ITAC range from $520,000/yr to $170,000/yr for RF-7a through RF-9a and $508,000/yr to $140,000/yr for RF-7b through RF-9b. If fuel oil is the displaced steam generation fuel, the ITAC for Scenario 1 range from a cost of $90,000/yr for RF-7a to a cost savings of $1.14 million/yr for RF-9a; for RF-7b through RF-9b, the ITAC for Scenario 1 range from a cost of $80,000/yr to a cost savings of $1.17 million/yr.

The ITAC estimates for Scenario 2 for DCE recovery furnaces with controlled PM emissions at NSPS levels are presented in Table 6-17. For Scenario 2, with natural gas as the displaced steam generation fuel, the ITAC estimates range from $560,000/yr to $310,000/yr for RF-7a through RF-9a and $548,000/yr to $280,000/yr for RF-7b through RF-9b. If fuel oil is the displaced steam generation fuel, the ITAC for Scenario 2 range from $130,000/yr for RF-7a to $120,000/yr for RF-7a to a cost savings of $1.00 million/yr for RF-9a; for RF-7b through RF-9b, the ITAC for Scenario 2 range from a cost of $120,000/yr to a cost savings of $1.03 million/yr.

The ITAC estimates for Scenario 3 for DCE recovery furnaces with controlled PM emissions at NSPS levels are presented in Table 6-18. For Scenario 3, with natural gas as the displaced steam generation fuel, the ITAC range from $550,000/yr to $270,000/yr for RF-7a through RF-9a and $538,000/yr to $240,000/yr for RF-7b through RF-9b. If fuel oil is the displaced steam generation fuel, the ITAC for Scenario 3 range from a cost of $120,000/yr for RF-7a to a cost savings of $1.04 million/yr for RF-9a; for RF-7b through RF-9b, the ITAC for Scenario 3 range from a cost of $110,000/yr to a cost savings of $1.07 million/yr.

The ITAC estimates for Scenario 1 for DCE recovery furnaces with controlled PM emissions of 0.034 g/dscm (0.015 gr/dscf) are presented in Table 6-19. For Scenario 1, with natural gas as the displaced steam generation fuel, the ITAC range from $590,000/yr to $380,000/yr for RF-7a through RF-9a and $578,000/yr to $350,000/yr for RF-7b through RF-9b. If fuel oil is the displaced steam generation fuel, the ITAC for Scenario 1 range from a cost of $160,000/yr to a cost savings of $930,000/yr for RF-9a; for RF-7b through RF-9b, the ITAC for Scenario 1 range from a cost of $150,000/yr to a cost savings of $960,000/yr.

The ITAC estimates for Scenario 2 for DCE recovery furnaces with controlled PM emissins of 0.034 g/dscm (0.015 gr/dscf) are presented in Table 6-20. For Scenario 2, with natural gas as the displaced steam generation fuel, the ITAC range from $630,000/yr to $520,000/yr for RF-7a through RF-9a and $618,000/yr to $490,000/yr for RF-7b through RF-9b. If fuel oil is the displaced steam generation fuel, the ITAC for Scenario 2 range from a cost of $200,000/yr for RF-7a to a cost savings of $790,000/yr for RF-9a; for RF-7b through RF-9b, the ITAC for Scenario 2 range from a cost of $190,000/yr to a cost savings of $820,000/yr.

The ITAC estimates for Scenario 3 for DCE recovery furnaces with controlled PM emissions of 0.034 g/dscm (0.015 gr/dscf) are presented in Table 6-21. For Scenario 3, with natural gas as the displaced steam generation fuel, the ITAC range from 620,000/yr to $480,000/yr for RF-7a through RF-9a and $608,000/yr to $450,000/yr for RF-7b for RF-9b. If fuel oil is the displaced steam generation fuel, the ITAC for Scenario 3 range from a cost of $190,000/yr for RF-7a to a cost savings of $830,000/yr for RF-9a; for RF-7b through RF-9b, the ITAC for Scenario 3 range from a cost of $180,000/yr to a cost savings of $860,000/yr.

6.1.2.4 <u>Addition of Packed-Bed Scrubber</u>. The addition of a packed-bed scrubber downstream of the ESP is included in two of the control options examined for recovery furnaces. These control options are (1) the use of an ESP plus a packed-bed scrubber to meet an outlet PM emission level of 0.034 g/dscm

(0.015 gr/dscf); and (2) the use of a packed-bed scrubber to reduce HCl emissions from recovery furnaces.

The costs of replacing or upgrading ESP's to control PM emissions are presented in Sections 6.1.2.1 and 6.1.2.3 for NDCE and DCE recovery furnaces. This section discusses the design and cost of packed-bed scrubbers for nine model recovery furnaces. The design and cost of packed-bed scrubbers are presented for the three model DCE recovery furnaces both with and without a low-odor conversion. The applicable model recovery furnaces for the packed-bed scrubber control option are RF-1 through RF-9. Exhaust gas stream parameters for each model and associated absorber are shown in Table 6-22.

6.1.2.4.1 Packed-bed scrubber design. Because only limited information was available from a scrubber manufacturer regarding the design parameters associated with a scrubber used to control HCl emissions from kraft recovery furnaces, the model packed-bed scrubbers were designed based on the procedures presented in Chapter 9 of the OAQPS Control Cost Manual for counterflow towers.[6,29] Two assumptions were made to simplify the packed-bed scrubber design analysis. First, it was assumed that the gas stream exiting the ESP is cooled to saturation by a water spray before it enters the packed-bed scrubber. As a result, the gas stream flow rates into and out of the packed-bed scrubber are the same. Insufficient design information was available from the OAQPS Control Cost Manual and a scrubber manufacturer to include in the design analysis a quench chamber for cooling the gas stream. A second simplifying assumption was that the diffusivity of HCl in the gas stream is approximated by the diffusivity of HCl in the air.

Because the model gas flow rates are large and the inlet HCl concentrations are low (only 9.7 ppmv for model NDCE and converted model DCE recovery furnaces and 9.2 ppmv for unconverted model DCE recovery furnaces), the diameters of the model towers are 11 to 23 times the height of the packing. The packing height was about 0.60 m (1.54 ft) for each model NDCE and converted model DCE recovery furnace and 0.43 m (1.41 ft) for

each unconverted model DCE recovery furnace. The tower diameters ranged from 5.2 to 11 m (17 to 36 ft) for model NDCE and converted model DCE recovery furnaces and 5.4 to 9.3 m (18 to 30 ft) for unconverted model DCE recovery furnaces. Design and operating parameters for each model recovery furnace packed-bed scrubber are shown in Table 6-23. The algorithm showing the procedures and equations used to determine the packed-bed scrubber design parameters is presented in a separate memorandum.[30]

6.1.2.4.2 Capital costs. Capital costs were calculated based on procedures presented in the OAQPS Control Cost Manual.[6] The unit costs used in calculating the capital and annual costs were derived from the OAQPS Control Cost Manual and from background information for the Medical Waste Incinerator standard.[6,31] The unit costs are shown in Table 6-24. Capital costs in 1991 dollars are presented in Table 6-25 for each model recovery furnace. These costs were compared to those obtained from a scrubber manufacturer.[29]

Capital costs consist of purchased equipment and installation costs. Purchased equipment costs consist of equipment, instrumentation, sales tax, and freight costs. Equipment includes a fiberglass reinforced polyester stack, fiberglass absorber tower, 5-cm (2-in.) randomly packed ceramic Raschig rings, a liquid recirculating pump, an induced draft fan, and a fan motor.

The addition of a packed-bed scrubber may result in additional dissolved solids loading to the wastewater treatment system. For those mills with restrictive total dissolved solids (TDS) effluent limitations, the additional solids loading from the scrubber may require internal process measures to reduce dissolved loading from other areas of the mill. These internal process measures will have associated engineering, equipment, and construction costs. However, because the additional costs are so site-specific, they cannot be estimated on a model basis and may even be offset by any heat recovery benefits realized as a result of adding an HCl scrubber.

All equipment costs were estimated as functions of various design parameters. Specifically, the stack cost was based on the length and diameter of the stack; the packed tower cost was based on the surface area of the tower; the packing cost was based on the volume of packing; the fan cost was based on the impeller diameter; the fan motor cost was based on the horsepower rating; and the pump cost was based on the design liquid flow rate.[5,6] In most cases, the referenced costing equations were developed for much smaller equipment. Thus, the costs for the models in this analysis were developed by extrapolating well beyond the largest parameter value for which the equations were developed.[30]

Instrumentation, sales tax, and freight were estimated to be equal to 18 percent of the equipment costs. Installation was estimated to be equal to 120 percent of the purchased equipment costs.[6] Total capital costs were estimated for each of the model recovery furnaces. The model furnaces include both new and existing furnaces. For model NDCE recovery furnaces RF-1 through RF-6, the total capital costs range from $1.10 million to $2.58 million. For unconverted model DCE recovery furnaces RF-7 through RF-9, the total capital costs range from $736,000 to $1.93 million. The total capital costs for the model DCE recovery furnaces converted to the NDCE design are slightly lower because of the change in furnace characteristics that occurs after a low-odor conversion. The total capital costs for the converted model DCE recovery furnaces RF-7 through RF-9 are identical to those for comparably sized model NDCE recovery furnaces and range from $707,000 to $1.85 million.

Capital costs for a packed-bed scrubber were also obtained from a scrubber manufacturer. The manufacturer provided packed-bed scrubber capital costs of $895,000, $1.69 million, and $2.30 million for three model furnaces with scrubber inlet gas flow rates of 47.2, 118, and 189 m^3/sec (100,000, 250,000, and 400,000 acfm), respectively. These costs include the cost for a quench, packed tower, mist eliminator, recirculation system (including two recirculation pumps and recirculation piping), instrumentation, engineering costs, and exhaust stack.[29]

Corrected for differences in gas flow rates from the model costs presented in the previous paragraph, the costs from the scrubber manufacturer differ by approximately 5 to 25 percent from the OAQPS costs presented in the previous paragraph for comparably sized model furnaces. This cost difference is within the ±30 percent range of accuracy that OAQPS costs should have, according to the OAQPS Control Cost Manual.[6] However, it should be noted that the OAQPS costs estimated here do not include quench costs.

6.1.2.4.3 Annual costs. Annual costs for the packed-bed scrubber were also calculated based on procedures presented in the OAQPS Control Cost Manual. Unit costs used in the annual cost calculations are shown in Table 6-24. Total annual costs in 1991 dollars are presented in Table 6-26 for the model recovery furnaces.

Annual costs were developed for labor, maintenance materials, water, caustic, wastewater disposal, electricity, overhead, property taxes, insurance, administrative charges, and capital recovery. Operator labor and maintenance labor were both assumed to be 0.5 hr per 8-hr shift, with three shifts per day. Supervisory labor costs were estimated to be equal to 15 percent of the operator labor costs. Maintenance materials costs were estimated to be equal to 100 percent of the maintenance labor costs.[6] The wastewater flow rate was estimated based on the assumption that the NaCl concentration in the recirculating water would be limited to 10 percent by weight. As a result, blowdown is approximately 0.08 to 0.09 percent of the recirculating liquid flow rate. Makeup water is needed for evaporative cooling before the packed-bed scrubber and to replace the blowdown losses. A stoichiometric amount of caustic is needed to react with all of the HCl in the exhaust gas stream; an additional amount of caustic must be added to react with the SO_2 also present in the exhaust gas stream (assuming 50 percent SO_2 control). Electricity usage by the fan was based on the gas flow rate out of the quench shown in Table 6-22 and the pressure drop and fan-motor efficiency shown in Table 6-23. Electricity usage by

the pump was based on the gas flow rate out of the quench shown in Table 6-22, the pump-motor efficiency shown in Table 6-23, and an assumed pressure head of 18 m (60 ft). Overhead costs were estimated to be equal to 60 percent of all labor and maintenance materials costs.[6] Collectively, property taxes, insurance, and administrative charges were estimated to be equal to 4 percent of the TCI.[6] Capital recovery was estimated to be equal to a CRF times the TCI.[6] The CRF is 0.1098, based on an equipment life of 15 years and an interest rate of 7 percent.

For model NDCE recovery furnaces RF-1 through RF-6, the TAC (as shown in Table 6-26) range from $348,000/yr to $790,000/yr. For model unconverted DCE recovery furnaces RF-7 through RF-9, the TAC range from $229,000/yr to $554,000/yr. The TAC for the model DCE recovery furnaces converted to the NDCE design are slightly higher because of higher costs for caustic, water, and wastewater disposal. The higher costs are a result of the higher SO_2 emission factor included in the equations for those costs. Based on the limited information available, SO_2 emissions are slightly higher, on average, from NDCE recovery furnaces than from DCE recovery furnaces.[32] The TAC for the converted model DCE recovery furnaces RF-7 through RF-9 are identical to those for comparably sized model NDCE recovery furnaces and range from $234,000/yr to 571,000/yr.

6.1.3 Black Liquor Oxidation Unit Control Options

Two control options, conversion of a DCE recovery furnace system to an NDCE recovery furnace and incineration of BLO vent gases, were evaluated for controlling gaseous organic HAP emissions from air-sparging BLO units. The cost of the first option--converting DCE recovery furnace systems to NDCE recovery furnaces--was presented in Section 6.1.2.3. The following section presents the capital and annual costs of the second BLO control option--incineration of BLO vent gases. This BLO control option applies to model BLO units BLO-1 through BLO-3, which represent existing BLO units associated with DCE recovery furnaces. These models represent only existing BLO units because

no new DCE recovery furnace systems with BLO units are expected to be installed.

6.1.3.1 Capital Costs. The total capital costs to collect BLO vent gases and incinerate them in a power boiler or other incineration device are based on a 1990 BLO control cost estimate of $4.8 million supplied by industry for a 730 ADMP/d (800 ADTP/d) kraft pulp mill.[33] The 1990 BLO control cost estimate includes piping, fans, condensers, and safety-related equipment, such as flame arrestors, rupture disks, etc. No major power boiler modifications (such as scrubber modifications) are included. The BLO control cost estimate was adjusted to 1991 dollars and then scaled using the six-tenths power rule described in Section 6.1.1.2 to derive BLO control costs for the model BLO units. A conversion factor of 1,700 kg BLS/ADMP (3,400 lb BLS/ADTP) (the average for bleached and unbleached pulp mills together) was assumed in scaling the cost. Design parameters for each model BLO unit are presented in Table 6-27. The model BLO control capital costs are presented in Table 6-28. The BLO control capital costs range from $2.5 million to $4.83 million for BLO-1 through BLO-3. Because the BLO collection and incineration system for one mill was installed within a 1-week maintenance shutdown, it was assumed that no downtime would be necessary beyond the annual 2-week shutdown used in determining costs.[34] As a result, no pulp production losses are included in these costs.

6.1.3.2 Annual Costs. Annual costs were estimated for operating and supervisory labor, maintenance labor and materials, electricity, steam, and indirect costs (e.g., overhead, administrative, taxes, insurance, and capital recovery). The BLO control annual costs were estimated for each model BLO unit based on the following sources:

1. Annual operating requirements provided by one mill (Mill A) for its BLO vent gas control system;[26]

2. Operating labor costs from the U. S. EPA Handbook: Control Technologies for Hazardous Air Pollutants;[5] and

3. Supervisory labor and maintenance costs and indirect costs from the OAQPS Control Cost Manual.[6]

The operating labor costs were estimated assuming 0.5 hours per shift per condenser for two condensers for three 8-hour shifts per day.[5] The operating labor hourly rate was assumed to be $17/hr.[5] The annual operating labor hours were assumed to be 365 d/yr. The supervisory labor costs were assumed to be 15 percent of the operating labor costs.[6] The maintenance labor costs were estimated at 1.5 times the operating labor costs.[6] The maintenance materials costs were estimated at 100 percent of the maintenance labor costs.[6]

Electricity costs were estimated based on the total kW (hp) requirements to operate the BLO vent gas control equipment at Mill A and scaled for the model BLO units assuming a direct relationship between BLO vent gas flow rate and electricity costs. For Mill A, with a BLO vent gas flow rate of 7.7 m^3/sec (16,327 acfm), 980 kW (100 hp) are required to operate the mill water booster pump motor, 29 kW (3 hp) to operate the BLO condenser condensate pump motor, and 3,900 kW (400 hp) to operate the BLO off gas blower motor.[26] The model electricity costs were estimated assuming 8,424 operating hr/yr and $0.06/kWh.

Steam costs were estimated based on the steam requirements and unit steam cost for the BLO off-gas reheater at Mill A--730 kg steam/hr (1,600 lb steam/hr) and $7/Mg of steam ($3/1,000 lb of steam), respectively--and scaled for the model BLO units assuming a direct relationship between BLO vent gas flow rate and steam costs.[26] The model steam costs were estimated assuming 8,424 operating hr/yr.

Indirect costs were estimated using assumptions in the OAQPS Control Cost Manual. Overhead costs were estimated as 60 percent of labor and maintenance costs. Administrative, taxes, and insurance costs were estimated as 4 percent of the TCI. Capital recovery costs were calculated as the product of the CRF and the TCI.[6] The CRF is 0.1424, based on a 10-yr equipment life for the ductwork and condenser and an interest rate of 7 percent.[5,24]

Total annual costs are presented in Table 6-28 and range from $681,000/yr to $1.32 million/yr for BLO-1 through BLO-3.

6.1.4 Smelt Dissolving Tank Control Options

Two PM control options that would reduce PM emissions from SDT's have been evaluated. The first option would reduce PM emissions from existing SDT's to the NSPS level of 0.10 kg/Mg (0.20 lb/ton) BLS. The second option would reduce PM emissions from existing SDT's to a more stringent level of 0.06 kg/Mg (0.12 lb/ton) BLS; the second option also applies to new SPTs; the option could be used to evaluate the cost to new sources subject to a more stringent standard (0.06 kg/Mg [0.12 lb/ton] BLS) than the current NSPS.

For mills with existing SDT scrubbers, the costs of both PM control options were estimated based on replacing the existing scrubber with a new scrubber. These costs were estimated for SDT models SDT-1 through SDT-4. For mills with new SDT scrubbers, the costs of installing scrubbers under the second, more stringent PM control option also apply to SDT-1 through SDT-4. For the purposes of this cost analysis, the capital cost to install a new SDT scrubber capable of meeting 0.06 kg/Mg (0.12 lb/ton) BLS was assumed to be the same as the cost to replace an existing scrubber with a new scrubber capable of meeting 0.06 kg/Mg (0.12 lb/ton) BLS. However, that may be an overestimate because the capital cost that would be attributable to the 0.06 kg/Mg (0.12 lb/ton) BLS control option would only be that cost associated with controlling PM emissions from the current NSPS level of 0.10 kg/Mg (0.2 lb/ton) BLS to 0.06 kg/Mg (0.12 lb/ton) BLS. Such a cost is more similar to a scrubber modification cost than a scrubber replacement cost.

For mills with existing SDT mist eliminators, the costs of both PM control options were estimated based on replacing the existing mist eliminator with a new scrubber. These costs were estimated for SDT models SDT-5 through SDT-7. The costs of installing new mist eliminators were not examined because mist eliminators are not assumed to be installed on new SDT's.

Note: For at least one scrubber type (wetted-wheel scrubbers), the scrubber replacement capital and annual costs would be approximately equal to the cost of replacing a mist eliminator with a scrubber. Similar to mist eliminators, the total capital costs would be based on replacement of the entire scrubber system. Also similar to mist eliminators, the incremental annual costs would include electricity costs based on replacing an existing scrubber that has a low pressure drop with another that has a significantly higher pressure drop, as well as indirect costs based on a capital cost of replacing the entire scrubber system. The capital and annual costs to replace wetted-wheel scrubbers will not be presented in this cost analysis because there is insufficient information to estimate the costs and because only 13 percent of SDT's have wetted-wheel scrubbers.[4]

The following sections present the capital and annual costs to replace existing wet scrubbers and mist eliminators with new wet scrubbers under the SDT PM control options.

6.1.4.1 Replacement of Existing Scrubber with New Scrubber: Capital Costs. The conditions under which replacing a scrubber is more cost-effective than modifying an existing scrubber are very site-specific. To be conservative, only the cost to replace a scrubber was evaluated. The cost to dispose of the existing scrubber was assumed to be included in the scrubber replacement costs. The capital costs to replace the scrubber are based on recent costs provided by a pulp and paper mill for an SDT scrubber.[4] The available long-term PM emissions data for the scrubber show that it is capable of consistently meeting both of the SDT PM emission limits--0.10 kg/Mg (0.20 lb/ton) BLS and 0.06 kg/Mg (0.12 lb/ton) BLS.[10] The SDT scrubber is a packed-tower scrubber with a gas flow rate of 3.8 m^3/sec (8,071 acfm) at 60 to 70 percent of recovery furnace capacity. The costs provided by the pulp and paper mill are packed-tower scrubber costs and scrubber modification costs. The packed-tower scrubber cost is $280,000 (1988 dollars); the scrubber modification cost is $50,000 (1991 dollars).[4] The 1988 scrubber cost was adjusted

to 1991 dollars using the Chemical Engineering Plant Cost Index and then summed with the scrubber modification costs to obtain the total capital cost in 1991 dollars. The six-tenths rule was used to extrapolate the total capital cost for each of the model SDT's. Based on information from a scrubber manufacturer, the on-site work time to replace a scrubber is about 2 days.[35] Therefore, no downtime beyond the annual 2-week shutdown is necessary for the scrubber replacement, which means that no pulp production losses are expected for this control option.

The model scrubber replacement costs are presented in Table 6-29. Total capital investment costs for the SDT models SDT-1 through SDT-4 range from $292,000 to $706,000.

6.1.4.2 Replacement of Existing Scrubber with New Scrubber: Incremental Annual Costs. The ITAC for each SDT model include the indirect costs associated with the TCI of the new scrubber (i.e., administrative, property tax, and insurance costs plus the capital recovery costs). The administrative, taxes, and insurance costs were estimated to be equal to 4 percent of the TCI.[6] A properly designed and maintained venturi scrubber can operate for 20 years.[35] Other types of scrubbers may have different life spans. To be conservative, a 15-year life span for replacement scrubbers was assumed. The capital recovery cost was estimated based on a 15-year scrubber life and a 7 percent interest rate. The increase in electricity costs from the scrubber replacement is not included in the ITAC estimate because it is not significant relative to the total annual cost. The average pressure drop for existing SDT scrubbers with PM emissions greater than 0.10 kg/Mg (0.20 lb/ton) BLS (i.e., baseline pressure drop) is approximately the same as the average pressure drop for those SDT scrubbers capable of meeting both SDT PM emission limits (0.10 kg/Mg [0.20 lb/ton] BLS) and 0.06 kg/Mg [0.12 lb/ton] BLS) on a long-term basis.[4,10] The baseline and control level pressure drops are 12 mm Hg (6.5 in. H_2O) and 13 mm Hg (7 in. H_2O), respectively.[4] All other direct costs (i.e., costs for operating labor, maintenance, water, and wastewater treatment) are assumed to be the same as those incurred by the

existing scrubber and, therefore, are not included in the ITAC estimate.

The ITAC for the four model SDT's are presented in Table 6-29. The ITAC estimates range from $43,800/yr to $106,000/yr for SDT-1 through SDT-4.

6.1.4.3 Replacement of Existing Mist Eliminator with New Scrubber: Capital Costs. The capital costs to replace an existing mist eliminator with a new scrubber include the costs for the new scrubber and all auxiliary equipment (i.e., fans, ductwork, etc.) that would be required at a new source. Based on information from a scrubber manufacturer, a completely new scrubber system would cost about twice as much as replacing only the scrubber, as described in Section 6.1.4.1.[35] The cost to dispose of the existing mist eliminator was assumed to be included in these replacement costs. Therefore, the capital costs for the three model SDT's are equal to twice the scrubber replacement costs for the three corresponding size model SDT's presented in Section 6.1.4.1. It is assumed that no downtime beyond the annual 2-week shutdown is necessary for replacing a mist eliminator with a scrubber.

The design parameters for the three SDT models are presented in Table 6-30. Capital costs for replacing a mist eliminator with a new scrubber are presented in Table 6-31. Total capital investment costs for the SDT models SDT-5 through SDT-7 range from $584,000 to $1.13 million.

6.1.4.4 Replacement of Existing Mist Eliminator with New Scrubber: Incremental Annual Costs. Incremental total annual costs were estimated for each of the three model SDT's. The ITAC for each SDT model include both direct and indirect annual costs, with the exception of operating labor. No changes in operator and supervisor personnel were assumed to be required. The increase in maintenance labor costs was estimated, assuming 3 hr/d at a wage rate of $25/hr.[6] Maintenance materials were estimated at 100 percent of maintenance labor costs.[6] As discussed in Section 6.1.1.3.1, for the purposes of calculating costs, pressure drop is used as an indicator of PM collection

efficiency. Electricity requirements and costs increase as a result of increasing the pressure drop to reduce PM emissions to 0.10 kg/Mg (0.20 lb/ton) BLS and to 0.06 kg/Mg (0.12 lb/ton) BLS. The average pressure drop is 1.3 mm Hg (0.7 in. H_2O) for those SDT mist eliminators with PM emissions above 0.06 kg/Mg (0.12 lb/ton) BLS (i.e., baseline pressure drop); the average pressure drop is 13 mm Hg (7 in. H_2O) for those SDT scrubbers capable of meeting both PM emission limits (0.10 kg/Mg [0.20 lb/ton] and 0.06 kg/Mg [0.12 lb/ton] BLS) on a long-term basis.[4,10] Therefore, an increase in pressure drop from 1.3 to 13 mm Hg (0.7 to 7 in. H_2O) was used to estimate the increase in electricity requirements and resulting increase in electricity costs. The overhead cost was estimated to be equal to 60 percent of the total maintenance cost.[6] The administrative, taxes, and insurance costs were estimated to be equal to 4 percent of the TCI.[6] The capital recovery cost was estimated to be equal to the product of a CRF and the TCI.[6] The CRF is 0.1098, based on a 15-year scrubber life and a 7 percent interest rate.

The ITAC for the SDT models SDT-5 through SDT-7 are presented in Table 6-31 and range from $190,000/yr to $301,000/yr.

6.1.5 Lime Kiln Control Options

Two PM control options have been evaluated for existing and new lime kilns. One PM control option that has been evaluated for existing lime kilns would reduce PM emissions to the NSPS level for gas-fired lime kilns--0.15 g/dscm (0.067 gr/dscf). For existing lime kilns with wet scrubbers, the control option would involve replacing the existing scrubber with an ESP. However, the actual control device (e.g., ESP or high-efficiency scrubber) selected by a particular mill would be site-specific. The costs for this PM control option were estimated for lime kiln models LK-1 through LK-3, which represent existing lime kilns controlled with wet scrubbers.

Based on PM emissions data supplied by mills, lime kilns controlled with ESP's already achieve a PM level of 0.15 g/dscm (0.067 gr/dscf).[4,10] Therefore, costs were not estimated for the

control option reducing PM emissions to 0.15 g/dscm (0.067 gr/dscf) for lime kilns controlled with ESP's (represented by models LK-4 through LK-6).

A second PM control option that was evaluated for new and existing lime kilns would reduce PM emissions to 0.023 g/dscm (0.010 gr/dscf). For existing lime kilns with wet scrubbers, the control option would involve replacing the existing scrubber with an ESP; costs would be estimated for models LK-1 through LK-3. For existing lime kilns with ESP's, the control option would involve upgrading the existing ESP. For new lime kilns, the control option would involve installing a new ESP capable of achieving the 0.023 g/dscm (0.010 gr/dscf) PM level. The costs for upgrading or installing a new ESP were estimated for models LK-4 through LK-6. The actual control device selected by a particular mill would actually be site-specific. The capital and annual costs for each of these options are presented in the following sections.

6.1.5.1 Replacement of Existing Scrubber with ESP: Capital Costs. The costs of replacing a scrubber with an ESP were calculated based on recent ESP costs provided by individual pulp and paper mills.[4] The cost to dispose of the existing scrubber was assumed to be included in the new ESP costs provided by the individual mills. New lime kiln ESP costs average $\$484/m^2$ ($\$45/ft^2$) of ESP plate area.[4] The ESP plate area for each model lime kiln was derived from the model gas flow rate at the ESP inlet and the SCA for the new ESP. The SCA for model lime kiln ESP's meeting a PM level of 0.15 g/dscm (0.067 gr/dscf) is estimated to be 90 $m^2/(m^3/sec)$ (460 ft^2/1,000 acfm).[4] The SCA value is based on the SCA of an actual lime kiln ESP for which long-term PM emissions data are available to demonstrate that its PM emissions are consistently at or below 0.15 g/dscm (0.067 gr/dscf).[10] The SCA for model lime kiln ESP's meeting a PM level of 0.023 g/dscm (0.010 gr/dscf) is estimated to be 220 $m^2/(m^3/sec)$ (1,120 ft^2/1,000 acfm).[4] The SCA value is based on the SCA of an actual lime kiln ESP for which long-term PM

emissions data are available to demonstrate that its PM emissions are consistently at or below 0.023 g/dscm (0.010 gr/dscf).[10]

Installing a new fan and stack typically comprises approximately 20 percent of the total capital costs of installing a new ESP. Because the fan and stack usually do not need to be replaced when a scrubber is replaced with an ESP, the cost for replacing a scrubber with an ESP is only about 80 percent of the cost of a completely new ESP.[9] Based on information from an ESP manufacturer, the lifetime of the replacement ESP is about 15 years.[36]

Table 6-32 presents the model replacement costs to control PM emissions to 0.15 g/dscm (0.067 gr/dscf). The costs range from $457,000 to $1.50 million for LK-1 through LK-3. Table 6-33 presents the model replacement costs to control PM emissions to 0.023 g/dscm (0.010 gr/dscf). The costs range from $1.11 million to $3.65 million for LK-1 through LK-3.

It was assumed that installation of the ESP could be completed within the 2-week scheduled shutdown, and, therefore, pulp production losses were not included in these cost estimates. Further information is needed to determine the validity of this assumption (e.g., the lack of available space complicating ESP installation and thereby increasing costs).

6.1.5.2 Replacement of Existing Scrubber with ESP: Incremental Annual Costs. To determine the incremental annual costs of replacing the existing scrubber with a new ESP for each model lime kiln, the annual costs for operating the existing scrubber were subtracted from the annual costs for operating a new ESP. To be conservative, the TCI-related indirect costs for the existing scrubber (i.e., the administrative, property tax, insurance, and capital recovery costs) were not included in the cost comparison.

The incremental annual costs of replacing the existing scrubber with a new ESP are presented in Tables 6-32 and 6-33. Design parameters and costs for the existing scrubber at baseline PM levels are presented in Tables 6-34 and 6-35, respectively. Because over 80 percent of lime kiln scrubbers are venturi

scrubbers, the scrubber design parameters presented in Table 6-34 are based on a venturi scrubber as the baseline control device.[8] Design parameters for the new ESP are presented in Table 6-36, and the ESP costs at the 0.15 g/dscm (0.067 gr/dscf) and 0.023 g/dscm (0.010 gr/dscf) PM levels are presented in Tables 6-37 and 6-38, respectively.

Direct annual costs (i.e., operating labor costs, maintenance costs, and utility costs) are reduced significantly when the existing scrubber is replaced with an ESP. Overhead costs are also reduced because they are a function of the labor and maintenance costs. For model lime kilns LK-1 through LK-3 controlling PM emissions to 0.15 g/dscm (0.067 gr/dscf), the reduction in direct annual costs and overhead costs obtained by switching from a scrubber to an ESP was greater than the TCI-related indirect annual costs for the new ESP. As a result, the ITAC are actually cost savings of $104,000/yr to $53,000/yr for LK-1 through LK-3. The ITAC savings for the model lime kilns are presented in Table 6-32.

For model lime kilns LK-1 through LK-3 controlling PM emissions to 0.023 g/dscm (0.010 gr/dscf), the reduction in direct annual costs and overhead costs was less than the TCI-related indirect costs for the ESP. As a result, there are annual costs associated with replacing the existing scrubber with a new ESP, ranging from $17,400/yr to $342,000/yr for LK-1 through LK-3. The TCI-related indirect annual costs for the ESP were strongly influenced by the high SCA value used to estimate the TCI. The SCA for new ESP's meeting a PM level of 0.023 g/dscm (0.010 gr/dscf) is approximately 220 $m^2/(m^3/sec)$ (1,120 ft^2/1,000 acfm).[4] The ITAC for the model lime kilns are presented in Table 6-33.

Approximately 15 percent of lime kiln scrubbers for which data are available have pressure drops below 19 mm Hg (10 in. H_2O), or are low-energy, low-pressure, or ejector type scrubbers.[4] These scrubbers operate at a lower pressure drop than the venturi scrubber used as the baseline control device. The annual costs to replace these low-pressure drop scrubbers

with ESP's would be higher than the costs and cost saving presented above. The additional annual costs for replacing low-pressure drop scrubbers were estimated based on the difference in pressure drop between the low-pressure drop scrubbers and the venturi scrubber used as the baseline control device. The average pressure drop is 7.5 mm Hg (4 in. H_2O) for those low-pressure drop lime kiln scrubbers with PM emissions greater than 0.15 g/dscm (0.067 gr/dscf). The average pressure drop is 39 mm Hg (21 in. H_2O) for the baseline venturi scrubber. To estimate the incremental annual costs of replacing low-pressure drop lime kiln scrubbers with ESP's, annual costs of $34,500/yr for LK-1, $65,700 for LK-2, and $112,000/yr for LK-3 should be added to the annual costs and cost savings presented above.

6.1.5.3 Upgrade of Existing ESP: Capital Costs. Because no actual ESP upgrade costs were available for controlling PM emissions from a level of 0.15 g/dscm (0.067 gr/dscf) to a level of 0.023 g/dscm (0.010 gr/dscf), the upgrade cost for the lime kiln ESP was instead based on the incremental cost difference between an ESP capable of achieving a PM level of 0.15 g/dscm (0.067 gr/dscf) and one capable of achieving a PM level of 0.023 g/dscm (0.010 gr/dscf).

The ESP upgrade capital costs were based on the recent ESP costs provided by individual pulp and paper mills (i.e., costs for ESP's installed or replaced during or after 1989).[4] The new lime kiln ESP costs average $484/m^2$ ($45/ft^2$) of ESP plate area.[4] To determine the ESP upgrade cost, this cost per ESP plate area was multiplied by the increase in ESP plate area assumed to reduce PM emissions from 0.15 g/dscm (0.067 gr/dscf) to 0.023 g/dscm (0.010 gr/dscf). The ESP plate area for each model lime kiln was derived from the model gas flow rate at the ESP inlet and the ESP SCA. The model SCA for an ESP meeting a long-term PM level of 0.15 g/dscm (0.067 gr/dscf) is approximately 90 $m^2/(m^3/sec)$ (460 ft^2/1,000 acfm).[4,10] The model SCA for an ESP meeting a long-term PM level of 0.023 g/dscm (0.010 gr/dscf) is approximately 220 $m^2/(m^3/sec)$ (1,120 ft^2/1,000 acfm).[4,10]

For new lime kilns installing ESP's to control PM emissiosn to 0.023 g/dscm (0.010 gr/dscf), the capital cost attributable to the control option would not be the cost of a new ESP but only that portion associated with controlling PM emissions from the current NSPS level of 0.15 g/dscm (0.067 gr/dscf) to 0.023 g/dscm (0.010 gr/dscf). Such a cost would be the same as the cost to upgrade existing lime kiln ESP's to achieve the same PM level of 0.023 g/dscm (0.010 gr/dscf).

The capital costs, excluding pulp production losses, for the ESP upgrade control option for new and existing model lime kilns are presented in Table 6-39 and range from $654,000 to $2.15 million for LK-4 through LK-6. No information is currently available for existing lime kilns on the amount of time required to complete an ESP upgrade that would achieve a PM control level of 0.023 g/dscm (0.010 gr/dscf). As a result, no pulp production losses were estimated for this ESP upgrade control option.

6.1.5.4 Upgrade of Existing ESP: Incremental Annual Costs. Labor and maintenance costs are assumed to be unchanged when the ESP is upgraded to achieve a PM level of 0.023 g/dscm (0.010 gr/dscf). Therefore, the incremental annual costs for the ESP upgrade include only the increase in electricity costs and the TCI-based indirect annual costs, which include administrative, property tax, insurance, and capital recovery costs. The capital recovery cost is based on an average 15-year life span for lime kiln ESP's.[36] The increased electricity costs are based on an increase in SCA from 90 $m^2/(m^3/sec)$ (460 ft^2/1,000 acfm) to 220 $m^2/(m^3/sec)$ (1,120 ft^2/1,000 acfm).[4] The ITAC for the new and existing model lime kilns, without taking pulp production losses into account, are presented in Table 6-39 and range from $112,000/yr to $369,000/yr for LK-4 through LK-6.

6.2 ENHANCED MONITORING COSTS

The following sections present the estimated costs for the enhanced monitoring options discussed in Chapter 4. Table 6-40 presents a summary of the enhanced monitoring costs, and Table 6-41 presents the itemized capital and annual costs for

opacity monitors and HCl CEM's. Enhanced monitoring costs were not estimated for control options other than those presented in Chapter 4; if facilities choose to meet the emission limits through the application of other control options, and a CEM is not applicable because of cost and/or technology constraints, the facilities must develop an enhanced monitoring plan that demonstrates the ability of the selected parameter to gauge a change in emissions.

6.2.1 Recovery Furnace Enhanced Monitoring

The following sections present the costs of enhanced monitoring options that can be used to demonstrate compliance with recovery furnace emission limits for PM or PM HAP's, total gaseous organic HAP's, and HCl.

6.2.1.1 Enhanced Monitoring for PM or PM HAP's Controlled with an ESP. Because opacity is the surrogate measurement that best characterizes the level of recovery furnace PM emissions, installation of an opacity monitor after the ESP is one option being considered as a means of demonstrating compliance with a PM or PM HAP emission limit for recovery furnaces. For those recovery furnaces with a wet scrubber following the ESP, an opacity monitor must be located after the ESP but prior to the scrubber. A computer program distributed by the Emission Measurement Technical Information Center (EMTIC) of EPA was used to estimate capital and annual costs for an opacity monitor.

The capital cost from EMTIC to purchase and install an in-situ opacity monitor (i.e., an opacity monitor that measures emissions in the stack or duct) is approximately $34,800. The capital costs include planning, selecting the type of equipment, providing support facilities, PEC, installing and checking CEM's, performance specification tests (certification), and preparing the quality assurance/quality control (QA/QC) plan required by appendix F (40 CFR 60). The PEC includes the cost to purchase a data acquisition system (DAS) which includes data reduction and reporting hardware/software.[37]

The annual costs from EMTIC equal $16,500/yr and include costs for operating and maintenance, reporting and recordkeeping,

and annual review and update.[37] Administrative, insurance, property tax, and capital recovery costs were estimated separately and added to the annual costs from the EMTIC program. The administrative, insurance, and property tax costs were calculated as 4 percent of the TCI, based on guidance in the OAQPS Control Cost Manual.[6] The capital recovery cost was calculated as a product of a CRF and the TCI.[6] The CRF was calculated assuming a 20-year equipment life and 7 percent interest. The total annual cost for the opacity monitor is approximately $21,200/yr.

Method 5, Method 29, or Method 17 compliance tests could be performed periodically (e.g., semiannually) as a substitute for an opacity monitor. The estimated cost for one three-run, EPA Method 5 compliance test is $8,500.[37] The estimated cost for one three-run, EPA Method 29 compliance test is $12,000.[38] No costs are available for a three-run, EPA Method 17 compliance test. If performed semiannually, the Method 5 tests would cost $17,000/yr; the Method 29 tests would cost $24,000/yr.

Another option being considered is for the facility to develop a monitoring plan that specifies ESP operating parameters to be monitored. Operating parameters for the ESP would be site-specific and would be based on the parameters measured during a three-run, EPA Method 5, Method 29, or Method 17 compliance test that showed the facility to be in compliance with the applicable PM or PM HAP emission limit. The cost to implement a monitoring plan has not been estimated. The estimated costs for three-run, EPA Method 5 or Method 29 compliance tests were discussed in the previous paragraph. No costs are available for a three-run, EPA Method 17 compliance test.

6.2.1.2 <u>Enhanced Monitoring for PM or PM HAP's Controlled with a Wet Scrubber</u>. For those recovery furnaces that can comply with a PM or PM HAP emission limit with existing wet scrubbers, the use of an opacity monitor to demonstrate compliance with the PM emission limit may be inappropriate. The exhaust from the recovery furnace wet scrubber will have a high moisture content and will interfere with the readings from an opacity monitor.

Monitoring scrubber operating parameters (i.e., pressure drop and scrubber liquid flow rate) is an alternative enhanced monitoring option for showing compliance with a PM or PM HAP emission limit for recovery furnaces. The pressure drop and liquid flow rate are indirect measurements of the performance of the scrubber.

The pressure drop across a wet scrubber can be determined by using a basic magnehelic gauge coupled with the S-type pitot tube and would require manual reading. The pressure drop can also be determined using a similar arrangement with an electronic magnehelic gauge that produces a digital signal. The cost for a manual read-out system would be less than $300. The additional cost for the electronic magnehelic gauge is not currently known.

There are various techniques for measuring liquid flow rate in a wet scrubber. These techniques include ultra-sonic detection mounted externally on the in-flowing water pipe and turbine devices that are mounted within the pipe, both of which generate an electrical signal that can be logically displayed in a control room. The cost for these flow-measuring techniques can vary from $2,000 to $25,000, depending on the device sensitivity and distance from the control room.[37]

Pressure drop and scrubber liquid flow rate levels would be site-specific and would be based on the operating parameters measured during a three-run, EPA Method 5, Method 29, or Method 17 compliance test that showed the facility to be in compliance with the applicable PM or PM HAP emission limit. Method 5, Method 29, or Method 17 compliance tests could also be performed periodically (e.g., semiannually) as a substitute for monitoring scrubber operating parameters. The estimated costs for three-run, EPA Method 5 and Method 29 compliance tests were discussed in Section 6.2.1.1. No costs are available for a three-run, EPA Method 17 compliance test.

6.2.1.3 <u>Enhanced Monitoring for Gaseous Organic HAP's</u>. Control of gaseous organic HAP emissions from recovery furnaces can be achieved by using NDCE recovery furnaces equipped with dry ESP systems. Therefore, enhanced monitoring for recovery furnace gaseous organic HAP emissions can be achieved simply by

confirming that the furnace is an NDCE recovery furnace with a dry ESP system. No costs are associated with this enhanced monitoring option.

If the recovery furnace is a DCE recovery furnace or an NDCE recovery furnace equipped with a wet ESP system, the facility could measure methanol emissions with a methanol CEM (e.g, a fourier transform infrared, or FTIR, spectroscopy monitoring system). The capital and annual costs for an FTIR monitor have been estimated to be approximately 160,000 and 71,500/yr, respectively.[39]

6.2.1.4 Enhanced Monitoring for HCl. Hydrochloric acid emissions can be measured directly using an HCl CEM. An HCl CEM could be installed after the packed-bed scrubber to demonstrate continuous compliance with an HCl emission standard. An HCl CEM could also be used after the ESP for those recovery furnaces that could comply with an HCl emission limit without a packed-bed scrubber. The capital and annual costs for an HCl CEM were determined using the EMTIC CEM cost program.

The capital cost from EMTIC to purchase and install an extractive HCl monitor (i.e., an HCl CEM that extracts a sample and transports the sample through a conditioning system and into a gas analyzer) is approximately $126,900.[37] The capital costs include planning, selecting the type of equipment, providing support facilities, PEC, installing and checking CEM's, performance specification tests (certification), and preparing the QA/QC plan required by Appendix F (40 CFR 60). The PEC costs include the cost for a DAS for data reduction and reporting. A performance specification for HCl monitors is currently under development for the CFR. However, one has already been developed by the Northeast States for Coordinated Air Use Management, or NESCAUM, and is being evaluated by EPA.[37]

The direct annual costs from EMTIC equal $60,300/yr and include costs for operating and maintenance, reporting and recordkeeping, and annual review and update.[37] Administrative, insurance, property tax, and capital recovery costs were estimated separately and added to the program costs. The

administrative, insurance, and property tax costs were calculated as 4 percent of the TCI, based on guidance in the OAQPS Control Cost Manual.[6] The capital recovery cost was calculated as a product of a CRF and the TCI.[6] The CRF was calculated assuming a 20-year equipment life and 7 percent interest.[6] The total annual cost for the HCl CEM is approximately $77,400/yr.

The feasibility of using HCl CEM's to demonstrate compliance with an HCl standard has not been determined. The low HCl concentrations and high moisture content associated with the recovery furnace flue gas may make the use of HCl CEM's more difficult. However, additional information is needed before an HCl CEM can be definitively ruled out for recovery furnaces.

Because HCl emissions can be controlled with a packed-bed scrubber, monitoring scrubber operating parameters is another monitoring option being considered for those recovery furnaces that comply with an HCl emission limit using a packed-bed scrubber. The scrubber operating parameters to be monitored are scrubber liquid pH and scrubber liquid flow rate. The cost for a pH monitoring system for scrubber water is approximately $5,000.[37] The cost to measure scrubber liquid flow rate was discussed in Section 6.2.1.2 and ranges from $2,000 to $25,000, depending on the devices' sensitivity and distance from the control room.[37] Scrubber liquid flow rate and pH levels would be site-specific and would be based on the operating parameters measured during a three-run, EPA Method 26A HCl compliance test that showed the facility to be in compliance with an HCl emission limit. Method 26A must be used when the measurement point is downstream of a wet scrubber. The estimated cost for one three-run, EPA Method 26 HCl compliance test is $9,100 if performed alone, or $600 if performed in conjunction with EPA Method 5, Method 29, or Method 17 testing.[40] The $600 would be in addition to the cost of the PM or PM HAP testing.

Alternative enhanced monitoring options are also available to demonstrate compliance for those recovery furnaces that could comply with an HCl emission limit without a packed-bed scrubber. One option would require the facility to develop a monitoring

plan that specifies operating parameters to be monitored. Operating parameters would be site-specific and would be based on the parameters measured during a three-run, EPA Method 26 or 26A HCl compliance test that showed the facility to be in compliance with the applicable HCl emission limit. Either Method 26 or 26A could be used if the measurement point does not follow a wet scrubber. The cost to implement a monitoring plan has not been estimated. The estimated cost for a three-run EPA Method 26 HCl compliance test would be the same as the estimated cost for a three-run, EPA Method 26A HCl compliance test, which was discussed in the previous paragraph. A second option would require periodic (e.g., annual) Method 26 or 26A HCl compliance tests to demonstrate compliance. The annual cost of EPA Method 26 or 26A HCl compliance testing is $9,100/yr if the test is performed annually and without EPA Method 5, Method 29, or Method 17 testing.[40] The annual cost is $600/yr if performed annually and in conjunction with EPA Method 5, Method 29, or Method 17 testing.[40] The $600 would be in addition to the annual cost of the PM or PM HAP testing.

6.2.2 Black Liquor Oxidation Unit Enhanced Monitoring

This section presents the costs of the enhanced monitoring options that can be used to demonstrate compliance with a total gaseous organic HAP emission limit for DCE recovery furnace systems (which include the BLO unit).

One control option presented for the BLO unit involves the removal of this piece of equipment from the chemical recovery process by converting a DCE recovery furnace system equipped with a wet ESP system to an NDCE recovery furnace equipped with a dry ESP system. Demonstrating that this conversion has been completed assures compliance with the applicable total gaseous organic HAP emission limit. No costs are associated with this enhanced monitoring option.

A second control option involves incineration of the BLO emissions. Enhanced monitoring for BLO incineration could be achieved simply by affirming that the BLO control equipment is in place. No cost is associated with this enhanced monitoring

option. Another enhanced monitoring option would be for the facility to monitor the temperature of the power boiler or other incineration device. The cost of a temperature monitor is assumed to be zero because temperature monitoring is already conducted by mills.

6.2.3 Smelt Dissolving Tank Enhanced Monitoring

This section presents the costs of the enhanced monitoring options that can be used to demonstrate compliance with an SDT emission limit for PM or PM HAP's.

6.2.3.1 Enhanced Monitoring for PM or PM HAP's Controlled with a Wet Scrubber. Because the exhaust from the SDT wet scrubber will have a high moisture content and will interfere with the readings from an opacity monitor, the use of an opacity monitor to demonstrate compliance with a PM or PM HAP emission limit for SDT's may be inappropriate. Monitoring scrubber operating parameters (i.e., pressure drop and scrubber liquid flow rate) is an alternate enhanced monitoring option for showing compliance with a PM or PM HAP emission limit for SDT's. The pressure drop and liquid flow rate are indirect measurements of the performance of the scrubber. The costs to monitor pressure drop and scrubber liquid flow rate were discussed in Section 6.2.1.2.

Pressure drop and scrubber liquid flow rate levels would be site-specific and would be based on the operating parameters measured during a three-run, EPA Method 5, Method 29, or Method 17 compliance test that showed the facility to be in compliance with the applicable PM or PM HAP emission limit. Method 5, Method 29, or Method 17 compliance tests could also be performed periodically (e.g., semiannually) as a substitute for monitoring scrubber operating parameters. The estimated costs for three-run, EPA Method 5 and Method 29 compliance tests were discussed in Section 6.2.1.1. No costs are available for a three-run, EPA Method 17 compliance test.

6.2.4 Lime Kiln Enhanced Monitoring

The following sections present the costs of the enhanced monitoring options that can be used to demonstrate compliance with a lime kiln emission limit for PM or PM HAP's.

6.2.4.1 Enhanced Monitoring for PM or PM HAP's Controlled with an ESP. Because opacity is the surrogate measurement that best characterizes the level of lime kiln PM emissions, installation of an opacity monitor is one option being considered as a means of demonstrating compliance with a PM or PM HAP emission limit for lime kilns controlled with ESP's. For those lime kilns with a wet scrubber following the ESP, an opacity monitor must be located after the ESP but prior to the scrubber. The capital and annual costs for the opacity monitor were discussed in Section 6.2.1.1. Method 5, Method 29, or Method 17 compliance tests could be performed periodically (e.g., semiannually) as a substitute for an opacity monitor. The estimated costs for three-run, EPA Method 5 and Method 29 compliance tests were discussed in Section 6.2.1.1. No costs are available for a three-run, EPA Method 17 compliance test.

Another option being considered is for the facility to develop a monitoring plan that specifies ESP parameters to be monitored. Operating parameters for the ESP would be site-specific and would be based on the parameters measured during a three-run, EPA Method 5, Method 29, or Method 17 compliance test that showed the facility to be in compliance with the applicable PM or PM HAP emission limit. The cost to implement a monitoring plan has not been estimated. The estimated costs for three-run, EPA Method 5 and Method 29 compliance tests were discussed in Section 6.2.1.1. No costs are available for a three-run, EPA Method 17 compliance test.

6.2.4.2 Enhanced Monitoring for PM or PM HAP's Controlled with a Wet Scrubber. For those lime kilns that can comply with a lime kiln PM or PM HAP emission limit with existing wet scrubbers, the use of an opacity monitor to demonstrate compliance with a PM or PM HAP emission limit may be inappropriate. The exhaust from the lime kiln wet scrubber will

have a high moisture content and will interfere with the readings from an opacity monitor. Monitoring scrubber operating parameters (i.e., pressure drop and scrubber liquid flow rate) is an alternative enhanced monitoring option for showing compliance with the applicable PM or PM HAP emission limit for lime kilns. The pressure drop and liquid flow rate are indirect measurements of the performance of the scrubber. The costs to monitor pressure drop and scrubber liquid flow rate were discussed in Section 6.2.1.2.

Pressure drop and scrubber liquid flow rate levels would be site-specific and would be based on the operating parameters measured during a three-run, EPA Method 5, Method 29, or Method 17 compliance test that showed the facility to be in compliance with the applicable PM or PM HAP emission limit. Method 5, Method 29, or Method 17 compliance tests could also be performed periodically (e.g., semiannually) as a substitute for monitoring scrubber operating parameters. The estimated costs for three-run, EPA Method 5 and Method 29 compliance tests were discussed in Section 6.2.1.1. No costs are available for a three-run, EPA Method 17 compliance test.

TABLE 6-1a (METRIC). COSTS OF ESP REPLACEMENT TO CONTROL PM TO 0.10 G/DSCM FOR MODEL RECOVERY FURNACES (EXCLUDING PULP PRODUCTION LOSSES)[a]

Model recovery furnaces	Pulp type	Furnace type	Black liquor firing rate, kg BLS/d	Equivalent ADMP/d	Gas flowrate--ESP exit, m3/sec @8% O2	ESP plate area, m2	New SCA, m2/(m3/sec)	Downtime, d	TCI, $	Electricity, $/yr	A,I & T, $/yr	Capital recovery, $/yr	ITAC, $/yr
RF-1a/4a	Bl	NDCE	700,000	380	93.4	9,806	105	30	$4,120,000	$19,400	$165,000	$482,000	$666,000
RF-2a/5a	Bl	NDCE	1,200,000	680	168	17,638	105	30	$7,430,000	$35,000	$297,000	$869,000	$1,200,000
RF-3a/6a	Bl	NDCE	1,800,000	1,000	243	25,512	105	30	$10,700,000	$50,500	$428,000	$1,250,000	$1,730,000
RF-7a	Bl	DCE	400,000	230	56.2	4,794	85	30	$2,010,000	$11,700	$80,400	$286,000	$378,000
RF-8a	Bl	DCE	700,000	380	93.4	7,967	85	30	$3,350,000	$19,400	$134,000	$477,000	$630,000
RF-9a	Bl	DCE	1,200,000	680	168	14,331	85	30	$6,030,000	$35,000	$241,000	$859,000	$1,140,000
RF-1a/4a	Unbl	NDCE	700,000	450	93.4	9,806	105	30	$4,120,000	$19,400	$165,000	$482,000	$666,000
RF-2a/5a	Unbl	NDCE	1,200,000	820	168	17,638	105	30	$7,430,000	$35,000	$297,000	$869,000	$1,200,000
RF-3a/6a	Unbl	NDCE	1,800,000	1,200	243	25,512	105	30	$10,700,000	$50,500	$428,000	$1,250,000	$1,730,000
RF-7a	Unbl	DCE	400,000	270	56.2	4,794	85	30	$2,010,000	$11,700	$80,400	$286,000	$378,000
RF-8a	Unbl	DCE	700,000	450	93.4	7,967	85	30	$3,350,000	$19,400	$134,000	$477,000	$630,000
RF-9a	Unbl	DCE	1,200,000	820	168	14,331	85	30	$6,030,000	$35,000	$241,000	$859,000	$1,140,000

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 6-1b. Refer to Table 6-1b for footnotes, which include calculations.

TABLE 6-1b (ENGLISH). COSTS OF ESP REPLACEMENT TO CONTROL PM TO 0.044 GR/DSCF FOR MODEL RECOVERY FURNACES (EXCLUDING PULP PRODUCTION LOSSES)[a]

Model recovery furnaces	Pulp type	Furnace type	Black liquor firing rate, lb BLS/d	Equivalent ADTP/d	Gas flowrate--ESP exit, acfm @8% O2	ESP plate area, ft2 (b)	New SCA, ft2/1,000 acfm	Downtime, d	TCI, $ (c)	Electricity, $/yr (d)	A,I & T, $/yr (e)	Capital recovery, $/yr (f)	ITAC, $/yr (g)
RF-1a/4a	BI	NDCE	1,500,000	420	198,000	105,600	533	30	$4,120,000	$19,400	$165,000	$482,000	$666,000
RF-2a/5a	BI	NDCE	2,700,000	750	357,000	190,400	533	30	$7,430,000	$35,000	$297,000	$869,000	$1,200,000
RF-3a/6a	BI	NDCE	3,900,000	1,100	515,000	274,667	533	30	$10,700,000	$50,500	$428,000	$1,250,000	$1,730,000
RF-7a	BI	DCE	900,000	250	119,000	51,567	433	30	$2,010,000	$11,700	$80,400	$286,000	$378,000
RF-8a	BI	DCE	1,500,000	420	198,000	85,800	433	30	$3,350,000	$19,400	$134,000	$477,000	$630,000
RF-9a	BI	DCE	2,700,000	750	357,000	154,700	433	30	$6,030,000	$35,000	$241,000	$859,000	$1,140,000
RF-1a/4a	Unbl	NDCE	1,500,000	500	198,000	105,600	533	30	$4,120,000	$19,400	$165,000	$482,000	$666,000
RF-2a/5a	Unbl	NDCE	2,700,000	900	357,000	190,400	533	30	$7,430,000	$35,000	$297,000	$869,000	$1,200,000
RF-3a/6a	Unbl	NDCE	3,900,000	1,300	515,000	274,667	533	30	$10,700,000	$50,500	$428,000	$1,250,000	$1,730,000
RF-7a	Unbl	DCE	900,000	300	119,000	51,567	433	30	$2,010,000	$11,700	$80,400	$286,000	$378,000
RF-8a	Unbl	DCE	1,500,000	500	198,000	85,800	433	30	$3,350,000	$19,400	$134,000	$477,000	$630,000
RF-9a	Unbl	DCE	2,700,000	900	357,000	154,700	433	30	$6,030,000	$35,000	$241,000	$859,000	$1,140,000

(a) All costs in $1991
(b) ESP plate area = model gas flowrate x new SCA
(c) TCI = ESP plate area x ($39/ft2 plate area)
(d) For NDCE furnaces, electricity = 0.00194 x model gas flowrate x (533 - 433 ft2)/1,000 acfm x 8,424 hr/yr x $0.06/kWh
For DCE furnaces, electricity = 0.00194 x model gas flowrate x (433 - 333 ft2)/1,000 acfm x 8,424 hr/yr x $0.06/kWh
(e) Administrative, insurance, and taxes (A,I & T) = 0.04 x TCI
(f) For NDCE furnaces, capital recovery = 0.1169 CRF x TCI (based on 13.5-yr ESP life and 7% interest)
For DCE furnaces, capital recovery = 0.1424 CRF x TCI (based on 10-yr ESP life and 7% interest)
(g) Incremental total annual cost (ITAC) = Electricity + A,I & T + capital recovery

TABLE 6-2a (METRIC). COSTS OF ESP REPLACEMENT TO CONTROL PM TO 0.10 G/DSCM FOR MODEL RECOVERY FURNACES (INCLUDING PULP PRODUCTION LOSSES)[a]

Model recovery furnaces	Pulp type	Furnace type	Black liquor firing rate, kg BLS/d	Equivalent ADMP/d	Gas flowrate--ESP exit, m3/sec @8% O2	ESP plate area, m2	New SCA, m2/(m3/sec)	Downtime, d	TCI, $	Pulp production losses, $	TCI+ production losses, $	Electricity, $/yr	A,I & T, $/yr	Capital recovery, $/yr	ITAC, $/yr
RF-1a/4a	BI	NDCE	700,000	380	93.4	9,808	105	30	$4,120,000	$1,190,000	$5,310,000	$19,400	$165,000	$621,000	$805,000
RF-2a/5a	BI	NDCE	1,200,000	660	168	17,638	105	30	$7,430,000	$2,130,000	$9,560,000	$35,000	$297,000	$1,120,000	$1,450,000
RF-3a/6a	BI	NDCE	1,800,000	1,000	243	25,512	105	30	$10,700,000	$3,120,000	$13,800,000	$50,500	$428,000	$1,610,000	$2,090,000
RF-7a	BI	DCE	400,000	230	56.2	4,794	85	30	$2,010,000	$844,000	$4,960,000	$19,400	$165,000	$580,000	$764,000
RF-8a	BI	DCE	700,000	380	93.4	7,967	85	30	$3,350,000	$1,520,000	$8,950,000	$35,000	$297,000	$1,050,000	$1,380,000
RF-9a	BI	DCE	1,200,000	660	168	14,331	85	30	$6,030,000	$2,190,000	$12,900,000	$50,500	$428,000	$1,510,000	$1,990,000
RF-1a/4a	Unbl	NDCE	700,000	450	93.4	9,808	105	30	$4,120,000	$844,000	$4,960,000	$19,400	$165,000	$580,000	$764,000
RF-2a/5a	Unbl	NDCE	1,200,000	820	168	17,638	105	30	$7,430,000	$1,520,000	$8,950,000	$35,000	$297,000	$1,050,000	$1,380,000
RF-3a/6a	Unbl	NDCE	1,800,000	1,200	243	25,512	105	30	$10,700,000	$2,190,000	$12,900,000	$50,500	$428,000	$1,510,000	$1,990,000
RF-7a	Unbl	DCE	400,000	270	56.2	4,794	85	30	$2,010,000	$506,000	$2,520,000	$11,700	$80,400	$359,000	$451,000
RF-8a	Unbl	DCE	700,000	450	93.4	7,967	85	30	$3,350,000	$844,000	$4,190,000	$19,400	$134,000	$597,000	$750,000
RF-9a	Unbl	DCE	1,200,000	820	168	14,331	85	30	$6,030,000	$1,520,000	$7,550,000	$35,000	$241,000	$1,080,000	$1,360,000

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 6-2b. Refer to Table 6-2b for footnotes, which include calculations.

TABLE 6-2b (ENGLISH). COSTS OF ESP REPLACEMENT TO CONTROL PM TO 0.044 GR/DSCF FOR MODEL RECOVERY FURNACES (INCLUDING PULP PRODUCTION LOSSES)[a]

Model recovery furnaces	Pulp type	Furnace type	Black liquor firing rate, lb BLS/d	Equivalent ADTP/d	Gas flowrate--ESP exit, acfm @8% O2	ESP plate area, ft2 (b)	New SCA, ft2/1,000 acfm	Downtime, d	TCI, $ (c)	Pulp production losses, $ (d)	TCI+ production losses, $	Electricity, $/yr (e)	A,I & T, $/yr (f)	Capital recovery, $/yr (g)	ITAC, $/yr (h)
RF-1a/4a	BI	NDCE	1,500,000	420	198,000	105,600	533	30	$4,120,000	$1,190,000	$5,310,000	$19,400	$165,000	$621,000	$805,000
RF-2a/5a	BI	NDCE	2,700,000	750	357,000	190,400	533	30	$7,430,000	$2,130,000	$9,560,000	$35,000	$297,000	$1,120,000	$1,450,000
RF-3a/6a	BI	NDCE	3,900,000	1,100	515,000	274,667	533	30	$10,700,000	$3,120,000	$13,800,000	$50,500	$428,000	$1,610,000	$2,090,000
RF-7a	BI	DCE	900,000	250	119,000	51,567	433	30	$2,010,000	$710,000	$2,720,000	$11,700	$80,400	$387,000	$479,000
RF-8a	BI	DCE	1,500,000	420	198,000	85,800	433	30	$3,350,000	$1,190,000	$4,540,000	$19,400	$134,000	$646,000	$799,000
RF-9a	BI	DCE	2,700,000	750	357,000	154,700	433	30	$6,030,000	$2,130,000	$8,160,000	$35,000	$241,000	$1,160,000	$1,440,000
RF-1a/4a	Unbl	NDCE	1,500,000	500	198,000	105,600	533	30	$4,120,000	$844,000	$4,960,000	$19,400	$165,000	$580,000	$764,000
RF-2a/5a	Unbl	NDCE	2,700,000	900	357,000	190,400	533	30	$7,430,000	$1,520,000	$8,950,000	$35,000	$297,000	$1,050,000	$1,380,000
RF-3a/6a	Unbl	NDCE	3,900,000	1,300	515,000	274,667	533	30	$10,700,000	$2,190,000	$12,900,000	$50,500	$428,000	$1,510,000	$1,990,000
RF-7a	Unbl	DCE	900,000	300	119,000	51,567	433	30	$2,010,000	$506,000	$2,520,000	$11,700	$80,400	$359,000	$451,000
RF-8a	Unbl	DCE	1,500,000	500	198,000	85,800	433	30	$3,350,000	$844,000	$4,190,000	$19,400	$134,000	$597,000	$750,000
RF-9a	Unbl	DCE	2,700,000	900	357,000	154,700	433	30	$6,030,000	$1,520,000	$7,550,000	$35,000	$241,000	$1,080,000	$1,360,000

(a) All costs in $1991
(b) ESP plate area = model gas flowrate x new SCA
(c) TCI = ESP plate area x ($39/ft2 plate area)
(d) Pulp production losses = (25% gross profit margin) x ($646/ton bleached pulp or $384/ton unbleached pulp) x (136.2 $1991/124 $1989) x (30 d downtime-14 d scheduled downtime) x ADTP/d
(e) For NDCE furnaces, electricity = 0.00194 x model gas flowrate x (533 - 433 ft2)/1,000 acfm x 8,424 hr/yr x $0.06/kWh
For DCE furnaces, electricity = 0.00194 x model gas flowrate x (433 - 333 ft2)/1,000 acfm x 8,424 hr/yr x $0.06/kWh
(f) Administrative, insurance, and taxes (A,I & T) = 0.04 x TCI
(g) For NDCE furnaces, capital recovery = 0.1169 CRF x (TCI + production losses) (based on 13.5-yr ESP life and 7% interest)
For DCE furnaces, capital recovery = 0.1424 CRF x (TCI + production losses) (based on 10-yr ESP life and 7% interest)
(h) Incremental total annual cost (ITAC) = Electricity + A,I & T + capital recovery

TABLE 6-3. SUMMARY OF ASSUMPTIONS USED IN ESP UPGRADE COSTS[9]

NDCE Recovery Furnace ESP Characteristics:	
Existing ESP: • The existing ESP is assumed to contain 46 lanes on 10-inch centers (weighted wire design). • The existing ESP is 3 fields long, with each field 10 feet long and 30 feet tall. • The existing ESP has a parallel drag scraper arrangement (dry-bottom).	New ESP: • The ESP is rebuilt with 42 lanes on 11-inch centers (rigid electrode design). • The size will be adequate for 99.4 percent PM removal, yielding an outlet PM residual of 0.044 gr/dscf at 8 percent oxygen. • The ESP will be 3 fields long, with each field 10 feet long and 30 feet high.
DCE Recovery Furnace ESP Characteristics:	
Existing ESP: • The existing ESP is assumed to contain 50 lanes on 10-inch centers (weighted wire design). • The existing ESP is 3 fields long, with each field 10 feet long and 30 feet tall. • The existing ESP has a steel shell design (wet-bottom).	New ESP: • The ESP is rebuilt with 46 lanes on 11-inch centers (rigid electrode design). • The size will be adequate for 98.8 percent PM removal, yielding an outlet PM residual of 0.044 gr/dscf at 8 percent oxygen. • The ESP will be 3 fields long, with each field 10 feet long and 32 feet high.

Modifications:

- There is single-chamber operation while the other chamber is washed down, the roof removed, and the internals gutted.
- The recovery furnace is taken off line; work continues on rebuild; one chamber is finished.
- The recovery furnace is at partial load through rebuilt chamber; the other chamber is finished.

Other Assumptions:

- The existing casing is in acceptable condition; only minor replating is needed.
- Wet- and dry-bottoms are reused "as is," except for minor repairs. Transformers, controls, and inlet and outlet ductwork are also reused.
- The roof insulation is replaced, but all other insulation is acceptable.
- The site location is in the southeast United States. As such, all field labor is based on merit shop, 1993 wage rates and work rules.
- No removal of asbestos or polychlorinated biphenyls (PCB's) is required.
- The scrap is taken to a site within the plant gates.
- There are no perforated plates.
- The ductwork and stack do not obstruct crane mobility.
- The ESP is at grade.
- There is no low voltage.

TABLE 6-4a (METRIC). COSTS OF SCHEDULE 1 ESP UPGRADE TO CONTROL PM TO 0.10 G/DSCM FOR MODEL RECOVERY FURNACES (EXCLUDING PULP PRODUCTION LOSSES)[a]

Model recovery furnaces	Pulp type	Furnace type	Black liquor firing rate, kg BLS/d	Equivalent ADMP/d	Gas flowrate--ESP exit, m3/sec @ 8% O2	Downtime, d	TCI, $	Electricity, $/yr	A,I & T, $/yr	Capital recovery, $/yr	ITAC, $/yr
RF-1a/4a	BI	NDCE	700,000	380	93.4	17	$1,200,000	$19,400	$48,000	$140,000	$207,000
RF-2a/5a	BI	NDCE	1,200,000	680	168	17	$1,700,000	$35,000	$68,000	$199,000	$302,000
RF-3a/6a	BI	NDCE	1,800,000	1,000	243	17	$2,120,000	$50,500	$84,800	$248,000	$383,000
RF-7a	BI	DCE	400,000	230	56.2	17	$811,000	$11,700	$32,400	$115,000	$159,000
RF-8a	BI	DCE	700,000	380	93.4	17	$1,100,000	$19,400	$44,000	$157,000	$220,000
RF-9a	BI	DCE	1,200,000	680	168	17	$1,570,000	$35,000	$62,800	$224,000	$322,000
RF-1a/4a	Unbl	NDCE	700,000	450	93.4	17	$1,200,000	$19,400	$48,000	$140,000	$207,000
RF-2a/5a	Unbl	NDCE	1,200,000	820	168	17	$1,700,000	$35,000	$68,000	$199,000	$302,000
RF-3a/6a	Unbl	NDCE	1,800,000	1,200	243	17	$2,120,000	$50,500	$84,800	$248,000	$383,000
RF-7a	Unbl	DCE	400,000	270	56.2	17	$811,000	$11,700	$32,400	$115,000	$159,000
RF-8a	Unbl	DCE	700,000	450	93.4	17	$1,100,000	$19,400	$44,000	$157,000	$220,000
RF-9a	Unbl	DCE	1,200,000	820	168	17	$1,570,000	$35,000	$62,800	$224,000	$322,000

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 6-4b. Refer to Table 6-4b for footnotes, which include calculations.

TABLE 6-4b (ENGLISH). COSTS OF SCHEDULE 1 ESP UPGRADE TO CONTROL PM TO 0.044 GR/DSCF FOR MODEL RECOVERY FURNACES (EXCLUDING PULP PRODUCTION LOSSES)[a]

Model recovery furnaces	Pulp type	Furnace type	Black liquor firing rate, lb BLS/d	Equivalent ADTP/d	Gas flowrate-- ESP exit, acfm @ 8% O2	Downtime, d	TCI, $ (b)	Electricity, $/yr (c)	A,I & T, $/yr (d)	Capital recovery, $/yr (e)	ITAC, $/yr (f)
RF-1a/4a	BI	NDCE	1,500,000	420	198,000	17	$1,200,000	$19,400	$48,000	$140,000	$207,000
RF-2a/5a	BI	NDCE	2,700,000	750	357,000	17	$1,700,000	$35,000	$68,000	$199,000	$302,000
RF-3a/6a	BI	NDCE	3,900,000	1,100	515,000	17	$2,120,000	$50,500	$84,800	$248,000	$383,000
RF-7a	BI	DCE	900,000	250	119,000	17	$811,000	$11,700	$32,400	$115,000	$159,000
RF-8a	BI	DCE	1,500,000	420	198,000	17	$1,100,000	$19,400	$44,000	$157,000	$220,000
RF-9a	BI	DCE	2,700,000	750	357,000	17	$1,570,000	$35,000	$62,800	$224,000	$322,000
RF-1a/4a	Unbl	NDCE	1,500,000	500	198,000	17	$1,200,000	$19,400	$48,000	$140,000	$207,000
RF-2a/5a	Unbl	NDCE	2,700,000	900	357,000	17	$1,700,000	$35,000	$68,000	$199,000	$302,000
RF-3a/6a	Unbl	NDCE	3,900,000	1,300	515,000	17	$2,120,000	$50,500	$84,800	$248,000	$383,000
RF-7a	Unbl	DCE	900,000	300	119,000	17	$811,000	$11,700	$32,400	$115,000	$159,000
RF-8a	Unbl	DCE	1,500,000	500	198,000	17	$1,100,000	$19,400	$44,000	$157,000	$220,000
RF-9a	Unbl	DCE	2,700,000	900	357,000	17	$1,570,000	$35,000	$62,800	$224,000	$322,000

(a) All costs in $1991
(b) For NDCE furnaces, TCI = ($1,292,000)/[(230,000 acfm/model gas flowrate) ^ 0.6] x (361.3 $1991/357 $May 1993)
For DCE furnaces, TCI = ($1,504,750)/[(340,000 acfm/model gas flowrate) ^ 0.6] x (361.3 $1991/357 $May 1993)
(c) Electricity costs for ESP upgrade assumed to equal electricity costs for ESP replacement.
For NDCE furnaces, electricity = 0.00194 x model gas flowrate x (533 - 433 ft2)/1,000 acfm x 8,424 hr/yr x $0.06/kWh
For DCE furnaces, electricity = 0.00194 x model gas flowrate x (433 - 333 ft2)/1,000 acfm x 8,424 hr/yr x $0.06/kWh
(d) Administrative, insurance, and taxes (A,I & T) = 0.04 x TCI
(e) For NDCE furnaces, capital recovery = 0.1169 CRF x TCI (based on 13.5-yr ESP life and 7% interest)
For DCE furnaces, capital recovery = 0.1424 CRF x TCI (based on 10-yr ESP life and 7% interest)
(f) Incremental total annual cost (ITAC) = Electricity + A,I & T + capital recovery

TABLE 6-5a (METRIC). COSTS OF SCHEDULE 2 ESP UPGRADE TO CONTROL PM TO 0.10 G/DSCM FOR MODEL RECOVERY FURNACES (EXCLUDING PULP PRODUCTION LOSSES)[a]

Model recovery furnaces	Pulp type	Furnace type	Black liquor firing rate, kg BLS/d	Equivalent ADMP/d	Gas flowrate--ESP exit, m3/sec @ 8% O2	Downtime, d	TCI, $ (b)	Electricity, $/yr (c)	A,I & T, $/yr (d)	Capital recovery, $/yr (e)	ITAC, $/yr (f)
RF-1a/4a	BI	NDCE	700,000	380	93.4	30	$1,160,000	$19,400	$46,400	$136,000	$202,000
RF-2a/5a	BI	NDCE	1,200,000	680	168	30	$1,660,000	$35,000	$66,400	$194,000	$295,000
RF-3a/6a	BI	NDCE	1,800,000	1,000	243	30	$2,070,000	$50,500	$82,800	$242,000	$375,000
RF-7a	BI	DCE	400,000	230	56.2	30	$791,000	$11,700	$31,600	$113,000	$156,000
RF-8a	BI	DCE	700,000	380	93.4	30	$1,070,000	$19,400	$42,800	$152,000	$214,000
RF-9a	BI	DCE	1,200,000	680	168	30	$1,530,000	$35,000	$61,200	$218,000	$314,000
RF-1a/4a	Unbl	ESP	700,000	450	93.4	30	$1,160,000	$19,400	$46,400	$136,000	$202,000
RF-2a/5a	Unbl	ESP	1,200,000	820	168	30	$1,660,000	$35,000	$66,400	$194,000	$295,000
RF-3a/6a	Unbl	ESP	1,800,000	1,200	243	30	$2,070,000	$50,500	$82,800	$242,000	$375,000
RF-7a	Unbl	DCE	400,000	270	56.2	30	$791,000	$11,700	$31,600	$113,000	$156,000
RF-8a	Unbl	DCE	700,000	450	93.4	30	$1,070,000	$19,400	$42,800	$152,000	$214,000
RF-9a	Unbl	DCE	1,200,000	820	168	30	$1,530,000	$35,000	$61,200	$218,000	$314,000

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 6-5b. Refer to Table 6-5b for footnotes, which include calculations.

TABLE 6-5b (ENGLISH). COSTS OF SCHEDULE 2 ESP UPGRADE TO CONTROL PM TO 0.044 GR/DSCF FOR MODEL RECOVERY FURNACES (EXCLUDING PULP PRODUCTION LOSSES)[a]

Model recovery furnaces	Pulp type	Furnace type	Black liquor firing rate, lb BLS/d	Equivalent ADTP/d	Gas flowrate-- ESP exit, acfm @ 8% O2	Downtime, d	TCI, $ (b)	Electricity, $/yr (c)	A,I & T, $/yr (d)	Capital recovery, $/yr (e)	ITAC, $/yr (f)
RF-1a/4a	BI	NDCE	1,500,000	420	198,000	30	$1,160,000	$19,400	$46,400	$136,000	$202,000
RF-2a/5a	BI	NDCE	2,700,000	750	357,000	30	$1,660,000	$35,000	$66,400	$194,000	$295,000
RF-3a/6a	BI	NDCE	3,900,000	1,100	515,000	30	$2,070,000	$50,500	$82,800	$242,000	$375,000
RF-7a	BI	DCE	900,000	250	119,000	30	$791,000	$11,700	$31,600	$113,000	$156,000
RF-8a	BI	DCE	1,500,000	420	198,000	30	$1,070,000	$19,400	$42,800	$152,000	$214,000
RF-9a	BI	DCE	2,700,000	750	357,000	30	$1,530,000	$35,000	$61,200	$218,000	$314,000
RF-1a/4a	Unbl	NDCE	1,500,000	500	198,000	30	$1,160,000	$19,400	$46,400	$136,000	$202,000
RF-2a/5a	Unbl	NDCE	2,700,000	900	357,000	30	$1,660,000	$35,000	$66,400	$194,000	$295,000
RF-3a/6a	Unbl	NDCE	3,900,000	1,300	515,000	30	$2,070,000	$50,500	$82,800	$242,000	$375,000
RF-7a	Unbl	DCE	900,000	300	119,000	30	$791,000	$11,700	$31,600	$113,000	$156,000
RF-8a	Unbl	DCE	1,500,000	500	198,000	30	$1,070,000	$19,400	$42,800	$152,000	$214,000
RF-9a	Unbl	DCE	2,700,000	900	357,000	30	$1,530,000	$35,000	$61,200	$218,000	$314,000

(a) All costs in $1991

(b) For NDCE furnaces, TCI = ($1,259,000)/[(230,000 acfm/model gas flowrate) ^ 0.6] x (361.3 $1991/357 $May 1993)
For DCE furnaces, TCI = ($1,466,500)/[(340,000 acfm/model gas flowrate) ^ 0.6] x (361.3 $1991/357 $May 1993)

(c) Electricity costs for ESP upgrade assumed to equal electricity costs for ESP replacement.
For NDCE furnaces, electricity = 0.00194 x model gas flowrate x (533 - 433 ft2)/1,000 acfm x 8,424 hr/yr x $0.06/kWh
For DCE furnaces, electricity = 0.00194 x model gas flowrate x (433 - 333 ft2)/1,000 acfm x 8,424 hr/yr x $0.06/kWh

(d) Administrative, insurance, and taxes (A,I & T) = 0.04 x TCI

(e) For NDCE furnaces, capital recovery = 0.1169 CRF x TCI (based on 13.5-yr ESP life and 7% interest)
For DCE furnaces, capital recovery = 0.1424 CRF x TCI (based on 10-yr ESP life and 7% interest)

(f) Incremental total annual cost (ITAC) = Electricity + A,I & T + capital recovery

TABLE 6-6a (METRIC). COSTS OF SCHEDULE 1 ESP UPGRADE TO CONTROL PM TO 0.10 G/DSCM FOR MODEL RECOVERY FURNACES (INCLUDING PULP PRODUCTION LOSSES)[a]

Model recovery furnaces	Pulp type	Furnace type	Black liquor firing rate, kg BLS/d	Equivalent ADMP/d	Gas flowrate--ESP exit, m3/sec @8% O2	Downtime, d	TCI, $	Pulp production losses, $	TCI+ production losses, $	Electricity, $/yr	A,I & T, $/yr	Capital recovery, $/yr	ITAC, $/yr
RF-1a/4a	BI	NDCE	700,000	380	93.4	17	$1,200,000	$224,000	$1,420,000	$19,400	$48,000	$166,000	$233,000
RF-2a/5a	BI	NDCE	1,200,000	680	168	17	$1,700,000	$399,000	$2,100,000	$35,000	$68,000	$245,000	$348,000
RF-3a/6a	BI	NDCE	1,800,000	1,000	243	17	$2,120,000	$585,000	$2,710,000	$50,500	$84,800	$317,000	$452,000
RF-7a	BI	DCE	400,000	230	56.2	17	$811,000	$133,000	$944,000	$11,700	$32,400	$134,000	$178,000
RF-8a	BI	DCE	700,000	380	93.4	17	$1,100,000	$224,000	$1,320,000	$19,400	$44,000	$188,000	$251,000
RF-9a	BI	DCE	1,200,000	680	168	17	$1,570,000	$399,000	$1,970,000	$35,000	$62,800	$281,000	$379,000
RF-1a/4a	Unbl	NDCE	700,000	450	93.4	17	$1,200,000	$158,000	$1,360,000	$19,400	$48,000	$159,000	$226,000
RF-2a/5a	Unbl	NDCE	1,200,000	820	168	17	$1,700,000	$285,000	$1,990,000	$35,000	$68,000	$233,000	$336,000
RF-3a/6a	Unbl	NDCE	1,800,000	1,200	243	17	$2,120,000	$411,000	$2,530,000	$50,500	$84,800	$296,000	$431,000
RF-7a	Unbl	DCE	400,000	270	56.2	17	$811,000	$94,900	$906,000	$11,700	$32,400	$129,000	$173,000
RF-8a	Unbl	DCE	700,000	450	93.4	17	$1,100,000	$158,000	$1,260,000	$19,400	$44,000	$179,000	$242,000
RF-9a	Unbl	DCE	1,200,000	820	168	17	$1,570,000	$285,000	$1,860,000	$35,000	$62,800	$265,000	$363,000

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 6-6b. Refer to Table 6-6b for footnotes, which include calculations.

TABLE 6-6b (ENGLISH). COSTS OF SCHEDULE 1 ESP UPGRADE TO CONTROL PM TO 0.044 GR/DSCF FOR MODEL RECOVERY FURNACES (INCLUDING PULP PRODUCTION LOSSES)[a]

Model recovery furnaces	Pulp type	Furnace type	Black liquor firing rate, lb BLS/d	Equivalent ADTP/d	Gas flowrate--ESP exit, acfm @8% O2	Downtime, d	TCI, $ (b)	Pulp production losses, $ (c)	TCI+ production losses, $	Electricity, $/yr (d)	A,I & T, $/yr (e)	Capital recovery, $/yr (f)	ITAC, $/yr (g)
RF-1a/4a	BI	NDCE	1,500,000	420	198,000	17	$1,200,000	$224,000	$1,420,000	$19,400	$48,000	$166,000	$233,000
RF-2a/5a	BI	NDCE	2,700,000	750	357,000	17	$1,700,000	$399,000	$2,100,000	$35,000	$68,000	$245,000	$348,000
RF-3a/6a	BI	NDCE	3,900,000	1,100	515,000	17	$2,120,000	$585,000	$2,710,000	$50,500	$84,800	$317,000	$452,000
RF-7a	BI	DCE	900,000	250	119,000	17	$811,000	$133,000	$944,000	$11,700	$32,400	$134,000	$178,000
RF-8a	BI	DCE	1,500,000	420	198,000	17	$1,100,000	$224,000	$1,320,000	$19,400	$44,000	$188,000	$251,000
RF-9a	BI	DCE	2,700,000	750	357,000	17	$1,570,000	$399,000	$1,970,000	$35,000	$62,800	$281,000	$379,000
RF-1a/4a	Unbl	NDCE	1,500,000	500	198,000	17	$1,200,000	$158,000	$1,360,000	$19,400	$48,000	$159,000	$226,000
RF-2a/5a	Unbl	NDCE	2,700,000	900	357,000	17	$1,700,000	$285,000	$1,990,000	$35,000	$68,000	$233,000	$336,000
RF-3a/6a	Unbl	NDCE	3,900,000	1,300	515,000	17	$2,120,000	$411,000	$2,530,000	$50,500	$84,800	$296,000	$431,000
RF-7a	Unbl	DCE	900,000	300	119,000	17	$811,000	$94,900	$906,000	$11,700	$32,400	$129,000	$173,000
RF-8a	Unbl	DCE	1,500,000	500	198,000	17	$1,100,000	$158,000	$1,260,000	$19,400	$44,000	$179,000	$242,000
RF-9a	Unbl	DCE	2,700,000	900	357,000	17	$1,570,000	$285,000	$1,860,000	$35,000	$62,800	$265,000	$363,000

(a) All costs in $1991
(b) For NDCE furnaces, TCI = ($1,292,000)/[(230,000 acfm/model gas flowrate) ^ 0.6] x (361.3 $1991/357 $May 1993)
For DCE furnaces, TCI = ($1,504,750)/[(340,000 acfm/model gas flowrate) ^ 0.6] x (361.3 $1991/357 $May 1993)
(c) Pulp production losses = (25% gross profit margin) x ($646/ton bleached pulp or $384/ton unbleached pulp) x (136.2 $1991/124 $1989) x (17 d downtime-14 d scheduled downtime) x ADTP/d
(d) Electricity costs for ESP upgrade assumed to equal electricity costs for ESP replacement.
For NDCE furnaces, electricity = 0.00194 x model gas flowrate x (533 - 433 ft2)/1,000 acfm x 8,424 hr/yr x $0.06/kWh
For DCE furnaces, electricity = 0.00194 x model gas flowrate x (433 - 333 ft2)/1,000 acfm x 8,424 hr/yr x $0.06/kWh
(e) Administrative, insurance, and taxes (A,I & T) = 0.04 x TCI
(f) For NDCE furnaces, capital recovery = 0.1169 CRF x (TCI + production losses) (based on 13.5-yr ESP life and 7% interest)
For DCE furnaces, capital recovery = 0.1424 CRF x (TCI + production losses) (based on 10-yr ESP life and 7% interest)
(g) Incremental total annual cost (ITAC) = Electricity + A,I & T + capital recovery

TABLE 6-7a (METRIC). COSTS OF SCHEDULE 2 ESP UPGRADE TO CONTROL PM TO 0.10 G/DSCM FOR MODEL RECOVERY FURNACES (INCLUDING PULP PRODUCTION LOSSES)[a]

Model recovery furnaces	Pulp type	Furnace type	Black liquor firing rate, kg BLS/d	Equivalent ADMP/d	Gas flowrate-- ESP exit, m3/sec @8% O2	Downtime, d	TCI, $	Pulp production losses, $	TCI+ production losses, $	Electricity, $/yr	A,I & T, $/yr	Capital recovery, $/yr	ITAC, $/yr
RF-1a/4a	BI	NDCE	700,000	380	93.4	30	$1,160,000	$1,190,000	$2,350,000	$19,400	$46,400	$275,000	$341,000
RF-2a/5a	BI	NDCE	1,200,000	680	168	30	$1,660,000	$2,130,000	$3,790,000	$35,000	$66,400	$443,000	$544,000
RF-3a/6a	BI	NDCE	1,800,000	1,000	243	30	$2,070,000	$3,120,000	$5,190,000	$50,500	$82,800	$607,000	$740,000
RF-7a	BI	DCE	400,000	230	56.2	30	$791,000	$710,000	$1,500,000	$11,700	$31,600	$214,000	$257,000
RF-8a	BI	DCE	700,000	380	93.4	30	$1,070,000	$1,190,000	$2,260,000	$19,400	$42,800	$322,000	$384,000
RF-9a	BI	DCE	1,200,000	680	168	30	$1,530,000	$2,130,000	$3,660,000	$35,000	$61,200	$521,000	$617,000
RF-1a/4a	Unbl	NDCE	700,000	450	93.4	30	$1,160,000	$844,000	$2,000,000	$19,400	$46,400	$234,000	$300,000
RF-2a/5a	Unbl	NDCE	1,200,000	820	168	30	$1,660,000	$1,520,000	$3,180,000	$35,000	$66,400	$372,000	$473,000
RF-3a/6a	Unbl	NDCE	1,800,000	1,200	243	30	$2,070,000	$2,190,000	$4,260,000	$50,500	$82,800	$498,000	$631,000
RF-7a	Unbl	DCE	400,000	270	56.2	30	$791,000	$506,000	$1,300,000	$11,700	$31,600	$185,000	$228,000
RF-8a	Unbl	DCE	700,000	450	93.4	30	$1,070,000	$844,000	$1,910,000	$19,400	$42,800	$272,000	$334,000
RF-9a	Unbl	DCE	1,200,000	820	168	30	$1,530,000	$1,520,000	$3,050,000	$35,000	$61,200	$434,000	$530,000

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 6-7b. Refer to Table 6-7b for footnotes, which include calculations.

TABLE 6-7b (ENGLISH). COSTS OF SCHEDULE 2 ESP UPGRADE TO CONTROL PM TO 0.044 GR/DSCF FOR MODEL RECOVERY FURNACES (INCLUDING PULP PRODUCTION LOSSES)[a]

Model recovery furnaces	Pulp type	Furnace type	Black liquor firing rate, lb BLS/d	Equivalent ADTP/d	Gas flowrate--ESP exit, acfm @8% O2	Downtime, d	TCI, $ (b)	Pulp production losses, $ (c)	TCI+ production losses, $	Electricity, $/yr (d)	A,I & T, $/yr (e)	Capital recovery, $/yr (f)	ITAC, $/yr (g)
RF-1a/4a	BI	NDCE	1,500,000	420	198,000	30	$1,160,000	$1,190,000	$2,350,000	$19,400	$46,400	$275,000	$341,000
RF-2a/5a	BI	NDCE	2,700,000	750	357,000	30	$1,660,000	$2,130,000	$3,790,000	$35,000	$66,400	$443,000	$544,000
RF-3a/6a	BI	NDCE	3,900,000	1,100	515,000	30	$2,070,000	$3,120,000	$5,190,000	$50,500	$82,800	$607,000	$740,000
RF-7a	BI	DCE	900,000	250	119,000	30	$791,000	$710,000	$1,500,000	$11,700	$31,600	$214,000	$257,000
RF-8a	BI	DCE	1,500,000	420	198,000	30	$1,070,000	$1,190,000	$2,260,000	$19,400	$42,800	$322,000	$384,000
RF-9a	BI	DCE	2,700,000	750	357,000	30	$1,530,000	$2,130,000	$3,660,000	$35,000	$61,200	$521,000	$617,000
RF-1a/4a	Unbl	NDCE	1,500,000	500	198,000	30	$1,160,000	$844,000	$2,000,000	$19,400	$46,400	$234,000	$300,000
RF-2a/5a	Unbl	NDCE	2,700,000	900	357,000	30	$1,660,000	$1,520,000	$3,180,000	$35,000	$66,400	$372,000	$473,000
RF-3a/6a	Unbl	NDCE	3,900,000	1,300	515,000	30	$2,070,000	$2,190,000	$4,260,000	$50,500	$82,800	$498,000	$631,000
RF-7a	Unbl	DCE	900,000	300	119,000	30	$791,000	$506,000	$1,300,000	$11,700	$31,600	$185,000	$228,000
RF-8a	Unbl	DCE	1,500,000	500	198,000	30	$1,070,000	$844,000	$1,910,000	$19,400	$42,800	$272,000	$334,000
RF-9a	Unbl	DCE	2,700,000	900	357,000	30	$1,530,000	$1,520,000	$3,050,000	$35,000	$61,200	$434,000	$530,000

(a) All costs in $1991
(b) For NDCE furnaces, TCI = ($1,259,000)/[(230,000 acfm/model gas flowrate) ^ 0.6] x (361.3 $1991/357 $May 1993)
For DCE furnaces, TCI = ($1,466,500)/[(340,000 acfm/model gas flowrate) ^ 0.6] x (361.3 $1991/357 $May 1993)
(c) Pulp production losses = (25% gross profit margin) x ($646/ton bleached pulp or $384/ton unbleached pulp) x (136.2 $1991/124 $1989) x (30 d downtime-14 d scheduled downtime) x ADTP/d
(d) Electricity costs for ESP upgrade assumed to equal electricity costs for ESP replacement.
For NDCE furnaces, electricity = 0.00194 x model gas flowrate x (533 - 433 ft2)/1,000 acfm x 8,424 hr/yr x $0.06/kWh
For DCE furnaces, electricity = 0.00194 x model gas flowrate x (433 - 333 ft2)/1,000 acfm x 8,424 hr/yr x $0.06/kWh
(e) Administrative, insurance, and taxes (A,I & T) = 0.04 x TCI
(f) For NDCE furnaces, capital recovery = 0.1169 CRF x (TCI + production losses) (based on 13.5-yr ESP life and 7% interest)
For DCE furnaces, capital recovery = 0.1424 CRF x (TCI + production losses) (based on 10-yr ESP life and 7% interest)
(g) Incremental total annual cost (ITAC) = Electricity + A,I & T + capital recovery

TABLE 6-8a (METRIC). COSTS OF ESP UPGRADE TO CONTROL PM FROM 0.10 G/DSCM TO 0.034 G/DSCM FOR MODEL RECOVERY FURNACES (EXCLUDING PULP PRODUCTION LOSSES)[a]

Model recovery furnaces	Pulp type	Furnace type	Black liquor firing rate, kg BLS/d	Equivalent ADMP/d	Gas flowrate--ESP exit, m3/sec @8% O2	Increase in ESP plate area, m2	Increase in SCA, m2/(m3/sec)	Downtime, d	TCI, $	Electricity, $/yr	A,I & T, $/yr	Capital recovery, $/yr	ITAC, $/yr
RF-1b/4b	BI	NDCE	700,000	380	93.4	1,532	16	17	$644,000	$16,200	$25,800	$75,300	$117,000
RF-2b/5b	BI	NDCE	1,200,000	680	168	2,756	16	17	$1,160,000	$29,200	$46,400	$136,000	$212,000
RF-3b/6b	BI	NDCE	1,800,000	1,000	243	3,986	16	17	$1,670,000	$42,100	$66,800	$195,000	$304,000
RF-7b	BI	DCE	400,000	230	56.2	922	16	17	$387,000	$9,720	$15,500	$55,100	$80,300
RF-8b	BI	DCE	700,000	380	93.4	1,532	16	17	$644,000	$16,200	$25,800	$91,700	$134,000
RF-9b	BI	DCE	1,200,000	700	168	2,756	16	17	$1,160,000	$29,200	$46,400	$165,000	$241,000
RF-1b/4b	Unbl	NDCE	700,000	450	93.4	1,532	16	17	$644,000	$16,200	$25,800	$75,300	$117,000
RF-2b/5b	Unbl	NDCE	1,200,000	820	168	2,756	16	17	$1,160,000	$29,200	$46,400	$136,000	$212,000
RF-3b/6b	Unbl	NDCE	1,800,000	1,200	243	3,986	16	17	$1,670,000	$42,100	$66,800	$195,000	$304,000
RF-7b	Unbl	DCE	400,000	270	56.2	922	16	17	$387,000	$9,720	$15,500	$55,100	$80,300
RF-8b	Unbl	DCE	700,000	450	93.4	1,532	16	17	$644,000	$16,200	$25,800	$91,700	$134,000
RF-9b	Unbl	DCE	1,200,000	820	168	2,756	16	17	$1,160,000	$29,200	$46,400	$165,000	$241,000

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 6-8b. Refer to Table 6-8b for footnotes, which include calculations.

TABLE 6-8b (ENGLISH). COSTS OF ESP UPGRADE TO CONTROL PM FROM 0.044 GR/DSCF TO 0.015 GR/DSCF FOR MODEL RECOVERY FURNACES (EXCLUDING PULP PRODUCTION LOSSES)[a]

Model recovery furnaces	Pulp type	Furnace type	Black liquor firing rate, lb BLS/d	Equivalent ADTP/d	Gas flowrate--ESP exit, acfm @8% O2	Increase in ESP plate area, ft2 (b)	Increase in SCA, ft2/1,000 acfm	Downtime, d	TCI, $ (c)	Electricity, $/yr (d)	A,I & T, $/yr (e)	Capital recovery, $/yr (f)	ITAC, $/yr (g)
RF-1b/4b	BI	NDCE	1,500,000	420	198,000	16,500	83	17	$644,000	$16,200	$25,800	$75,300	$117,000
RF-2b/5b	BI	NDCE	2,700,000	750	357,000	29,750	83	17	$1,160,000	$29,200	$46,400	$136,000	$212,000
RF-3b/6b	BI	NDCE	3,900,000	1,100	515,000	42,917	83	17	$1,670,000	$42,100	$66,800	$195,000	$304,000
RF-7b	BI	DCE	900,000	250	119,000	9,917	83	17	$387,000	$9,720	$15,500	$55,100	$80,300
RF-8b	BI	DCE	1,500,000	420	198,000	16,500	83	17	$644,000	$16,200	$25,800	$91,700	$134,000
RF-9b	BI	DCE	2,700,000	750	357,000	29,750	83	17	$1,160,000	$29,200	$46,400	$165,000	$241,000
RF-1b/4b	Unbl	NDCE	1,500,000	500	198,000	16,500	83	17	$644,000	$16,200	$25,800	$75,300	$117,000
RF-2b/5b	Unbl	NDCE	2,700,000	900	357,000	29,750	83	17	$1,160,000	$29,200	$46,400	$136,000	$212,000
RF-3b/6b	Unbl	NDCE	3,900,000	1,300	515,000	42,917	83	17	$1,670,000	$42,100	$66,800	$195,000	$304,000
RF-7b	Unbl	DCE	900,000	300	119,000	9,917	83	17	$387,000	$9,720	$15,500	$55,100	$80,300
RF-8b	Unbl	DCE	1,500,000	500	198,000	16,500	83	17	$644,000	$16,200	$25,800	$91,700	$134,000
RF-9b	Unbl	DCE	2,700,000	900	357,000	29,750	83	17	$1,160,000	$29,200	$46,400	$165,000	$241,000

(a) All costs in $1991
(b) Increase in ESP plate area = model gas flowrate x increase in SCA
(c) TCI = Increase in ESP plate area x ($39/ft2 plate area)
(d) For NDCE furnaces, electricity = 0.00194 x model gas flowrate x (617 - 533 ft2)/1,000 acfm x 8,424 hr/yr x $0.06/kWh
For DCE furnaces, electricity = 0.00194 x model gas flowrate x (517 - 433 ft2)/1,000 acfm x 8,424 hr/yr x $0.06/kWh
(e) Administrative, insurance, and taxes (A,I & T) = 0.04 x TCI
(f) For NDCE furnaces, capital recovery = 0.1169 CRF x TCI (based on 13.5-yr ESP life and 7% interest)
For DCE furnaces, capital recovery = 0.1424 CRF x TCI (based on 10-yr ESP life and 7% interest)
(g) Incremental total annual cost (ITAC) = Electricity + A,I & T + capital recovery

TABLE 6-9a (METRIC). COSTS OF ESP UPGRADE TO CONTROL PM FROM 0.10 G/DSCM TO 0.034 G/DSCM FOR MODEL RECOVERY FURNACES (INCLUDING PULP PRODUCTION LOSSES)[a]

Model recovery furnaces	Pulp type	Furnace type	Black liquor firing rate, kg BLS/d	Equivalent ADMP/d	Gas flowrate--ESP exit, m3/sec @8% O2	Increase in ESP plate area, m2	Increase in SCA, m2/(m3/sec)	Downtime, d	TCI, $	Pulp production losses, $	TCI+ production losses, $	Electricity, $/yr	A,I & T, $/yr	Capital recovery, $/yr	ITAC, $/yr
RF-1b/4b	BI	NDCE	700,000	380	93.4	1,532	16	17	$644,000	$224,000	$868,000	$16,200	$25,800	$101,000	$143,000
RF-2b/5b	BI	NDCE	1,200,000	680	168	2,756	16	17	$1,160,000	$399,000	$1,560,000	$29,200	$46,400	$182,000	$258,000
RF-3b/6b	BI	NDCE	1,800,000	1,000	243	3,986	16	17	$1,670,000	$585,000	$2,260,000	$42,100	$66,800	$264,000	$373,000
RF-7b	BI	DCE	400,000	230	56.2	922	16	17	$387,000	$133,000	$520,000	$9,700	$15,500	$74,000	$99,000
RF-8b	BI	DCE	700,000	380	93.4	1,532	16	17	$644,000	$224,000	$868,000	$16,200	$25,800	$124,000	$166,000
RF-9b	BI	DCE	1,200,000	680	168	2,756	16	17	$1,160,000	$399,000	$1,560,000	$29,200	$46,400	$222,000	$298,000
RF-1b/4b	Unbl	NDCE	700,000	230	93.4	1,532	16	17	$644,000	$158,000	$802,000	$16,200	$25,800	$93,800	$136,000
RF-2b/5b	Unbl	NDCE	1,200,000	820	168	2,756	16	17	$1,160,000	$285,000	$1,450,000	$29,200	$46,400	$170,000	$246,000
RF-3b/6b	Unbl	NDCE	1,800,000	1,200	243	3,986	16	17	$1,670,000	$411,000	$2,080,000	$42,100	$66,800	$243,000	$352,000
RF-7b	Unbl	DCE	400,000	270	56.2	922	16	17	$387,000	$94,900	$482,000	$9,700	$15,500	$68,600	$94,000
RF-8b	Unbl	DCE	700,000	450	93.4	1,532	16	17	$644,000	$158,000	$802,000	$16,200	$25,800	$114,000	$156,000
RF-9b	Unbl	DCE	1,200,000	820	168	2,756	16	17	$1,160,000	$285,000	$1,450,000	$29,200	$46,400	$206,000	$282,000

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 6-9b. Refer to Table 6-9b for footnotes, which include calculations.

TABLE 6-9b (ENGLISH). COSTS OF ESP UPGRADE TO CONTROL PM FROM 0.044 GR/DSCF TO 0.015 GR/DSCF FOR MODEL RECOVERY FURNACES (INCLUDING PULP PRODUCTION LOSSES)[a]

Model recovery furnaces	Pulp type	Furnace type	Black liquor firing rate, lb BLS/d	Equivalent ADTP/d	Gas flowrate-- ESP exit, acfm @8% O2	Increase in ESP plate area, ft2 (b)	Increase in SCA, ft2/1,000 acfm	Downtime, d	TCI, $ (c)	Pulp production losses, $ (d)	TCI+ production losses, $	Electricity, $/yr (e)	A,I & T, $/yr (f)	Capital recovery, $/yr (g)	ITAC, $/yr (h)
RF-1b/4b	BI	NDCE	1,500,000	420	198,000	16,500	83	17	$644,000	$224,000	$868,000	$16,200	$25,800	$101,000	$143,000
RF-2b/5b	BI	NDCE	2,700,000	750	357,000	29,750	83	17	$1,160,000	$399,000	$1,560,000	$29,200	$46,400	$182,000	$258,000
RF-3b/6b	BI	NDCE	3,900,000	1,100	515,000	42,917	83	17	$1,670,000	$585,000	$2,260,000	$42,100	$66,800	$264,000	$373,000
RF-7b	BI	DCE	900,000	250	119,000	9,917	83	17	$387,000	$133,000	$520,000	$9,700	$15,500	$74,000	$99,000
RF-8b	BI	DCE	1,500,000	420	198,000	16,500	83	17	$644,000	$224,000	$868,000	$16,200	$25,800	$124,000	$166,000
RF-9b	BI	DCE	2,700,000	750	357,000	29,750	83	17	$1,160,000	$399,000	$1,560,000	$29,200	$46,400	$222,000	$298,000
RF-1b/4b	Unbl	NDCE	1,500,000	500	198,000	16,500	83	17	$644,000	$158,000	$802,000	$16,200	$25,800	$93,800	$136,000
RF-2b/5b	Unbl	NDCE	2,700,000	900	357,000	29,750	83	17	$1,160,000	$285,000	$1,450,000	$29,200	$46,400	$170,000	$246,000
RF-3b/6b	Unbl	NDCE	3,900,000	1,300	515,000	42,917	83	17	$1,670,000	$411,000	$2,080,000	$42,100	$66,800	$243,000	$352,000
RF-7b	Unbl	DCE	900,000	300	119,000	9,917	83	17	$387,000	$94,900	$482,000	$9,700	$15,500	$68,600	$94,000
RF-8b	Unbl	DCE	1,500,000	500	198,000	16,500	83	17	$644,000	$158,000	$802,000	$16,200	$25,800	$114,000	$156,000
RF-9b	Unbl	DCE	2,700,000	900	357,000	29,750	83	17	$1,160,000	$285,000	$1,450,000	$29,200	$46,400	$208,000	$282,000

(a) All costs in $1991
(b) Increase in ESP plate area = model gas flowrate x increase in SCA
(c) TCI = Increase in ESP plate area x ($39/ft2 plate area)
(d) Pulp production losses = (25% gross profit margin) x ($646/ton bleached pulp or $384/ton unbleached pulp) x (136.2 $1991/124 $1989)x (17 d downtime-14 d scheduled downtime) x ADTP/d
(d) For NDCE furnaces, electricity = 0.00194 x model gas flowrate x (617 - 533 ft2)/1,000 acfm x 8,424 hr/yr x $0.06/kWh
For DCE furnaces, electricity = 0.00194 x model gas flowrate x (517 - 433 ft2)/1,000 acfm x 8,424 hr/yr x $0.06/kWh
(f) Administrative, insurance, and taxes (A,I & T) = 0.04 x TCI
(g) For NDCE furnaces, capital recovery = 0.1169 CRF x (TCI + production losses) (based on 13.5-yr ESP life and 7% interest)
For DCE furnaces, capital recovery = 0.1424 CRF x (TCI + production losses) (based on 10-yr ESP life and 7% interest)
(h) Incremental total annual cost (ITAC) = Electricity + A,I & T + capital recovery

TABLE 6-10a (METRIC). WET- TO DRY-BOTTOM ESP CONVERSION COSTS FOR MODEL NDCE RECOVERY FURNACES[a]

Model recovery furnaces	Pulp type	Black liquor firing rate, kg BLS/d	Equivalent ADMP/d	Baseline ESP type	Control option ESP type	Gas flowrate--ESP exit, m3/sec @8% O2	TCI, $	A,I & T, $/yr	Capital recovery, $/yr	ITAC, $/yr
RF-4a/4b	BI	700,000	380	wet-bottom	dry-bottom	93.4	$596,000	$23,800	$69,700	$93,500
RF-5a/5b	BI	1,200,000	680	wet-bottom	dry-bottom	168	$849,000	$34,000	$99,200	$133,000
RF-6a/6b	BI	1,800,000	1,000	wet-bottom	dry-bottom	243	$1,060,000	$42,400	$124,000	$166,000
RF-4a/4b	Unbl	700,000	450	wet-bottom	dry-bottom	93.4	$596,000	$23,800	$69,700	$93,500
RF-5a/5b	Unbl	1,200,000	820	wet-bottom	dry-bottom	168	$849,000	$34,000	$99,200	$133,000
RF-6a/6b	Unbl	1,800,000	1,200	wet-bottom	dry-bottom	243	$1,060,000	$42,400	$124,000	$166,000

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 6-10b. Refer to Table 6-10 for footnotes, which include calculations.

TABLE 6-10b (ENGLISH). WET- TO DRY-BOTTOM ESP CONVERSION COSTS FOR MODEL NDCE RECOVERY FURNACES[a]

Model recovery furnaces	Pulp type	Black liquor firing rate, lb BLS/d	Equivalent ADTP/d	Baseline ESP type	Control option ESP type	Gas flowrate--ESP exit, acfm @8% O2	TCI, $ (b)	A,I & T, $/yr (c)	Capital recovery, $/yr (d)	ITAC, $/yr (e)
RF-4a/4b	BI	1,500,000	420	wet-bottom	dry-bottom	198,000	$596,000	$23,800	$69,700	$93,500
RF-5a/5b	BI	2,700,000	750	wet-bottom	dry-bottom	357,000	$849,000	$34,000	$99,200	$133,000
RF-6a/6b	BI	3,900,000	1,100	wet-bottom	dry-bottom	515,000	$1,060,000	$42,400	$124,000	$166,000
RF-4a/4b	Unbl	1,500,000	500	wet-bottom	dry-bottom	198,000	$596,000	$23,800	$69,700	$93,500
RF-5a/5b	Unbl	2,700,000	900	wet-bottom	dry-bottom	357,000	$849,000	$34,000	$99,200	$133,000
RF-6a/6b	Unbl	3,900,000	1,300	wet-bottom	dry-bottom	515,000	$1,060,000	$42,400	$124,000	$166,000

(a) All costs in $1991
(b) TCI = $845,000/(357,000 acfm/model gas flowrate) ^ 0.6 x (361.3 $1991/359.4 $July 1993)
(c) Administrative, insurance, and taxes (A,I & T) = 0.04 x TCI
(d) Capital recovery = 0.1169 CRF x TCI (based on 13.5-yr ESP life and 7% interest)
(e) Incremental total annual cost (ITAC) = A,I & T + capital recovery

TABLE 6-11a (METRIC). CAPITAL COSTS OF LOW-ODOR CONVERSION (INCLUDING ESP UPGRADE TO CONTROL PM TO 0.10 G/DSCM) FOR MODEL DCE RECOVERY FURNACES (EXCLUDING PULP PRODUCTION LOSSES)[a]

Model recovery furnaces	Pulp type	Black liquor firing rate, kg BLS/d	Equivalent ADMP/d	Gas flowrate--ESP exit, m3/sec @ 8% O2	Downtime, d	Economizer expansion, demolition, $	Concentrator, $	Low-odor conversion, $	ESP upgrade, $	Wet- to dry-bottom ESP conversion, $	TCI, $
RF-7a/7b	BI	400,000	230	56.2	25	$4,780,000	$3,310,000	$8,090,000	$881,000	$439,000	$9,410,000
RF-8a/8b	BI	700,000	380	93.4	25	$6,500,000	$4,500,000	$11,000,000	$1,200,000	$596,000	$12,800,000
RF-9a/9b	BI	1,200,000	680	168	25	$9,250,000	$6,400,000	$15,700,000	$1,700,000	$849,000	$18,200,000
RF-7a/7b	Unbl	400,000	270	56.2	25	$4,780,000	$3,310,000	$8,090,000	$881,000	$439,000	$9,410,000
RF-8a/8b	Unbl	700,000	450	93.4	25	$6,500,000	$4,500,000	$11,000,000	$1,200,000	$596,000	$12,800,000
RF-9a/9b	Unbl	1,200,000	820	168	25	$9,250,000	$6,400,000	$15,700,000	$1,700,000	$849,000	$18,200,000

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 6-11b. Refer to Table 6-11b for footnotes, which include calculations.

TABLE 6-11b (ENGLISH). CAPITAL COSTS OF LOW-ODOR CONVERSION (INCLUDING ESP UPGRADE TO CONTROL PM TO 0.044 GR/DSCF) FOR MODEL DCE RECOVERY FURNACES (EXCLUDING PULP PRODUCTION LOSSES)[a]

Model recovery furnaces	Pulp type	Black liquor firing rate, lb BLS/d	Equivalent ADTP/d	Gas flowrate--ESP exit, acfm @ 8% O2	Downtime, d	Economizer expansion, demolition, $	Concentrator, $	Low-odor conversion, $ (b)	ESP upgrade, $ (c)	Wet- to dry-bottom ESP conversion, $ (d)	TCI, $ (e)
RF-7a/7b	BI	900,000	250	119,000	25	$4,780,000	$3,310,000	$8,090,000	$881,000	$439,000	$9,410,000
RF-8a/8b	BI	1,500,000	420	198,000	25	$6,500,000	$4,500,000	$11,000,000	$1,200,000	$596,000	$12,800,000
RF-9a/9b	BI	2,700,000	750	357,000	25	$9,250,000	$6,400,000	$15,700,000	$1,700,000	$849,000	$18,200,000
RF-7a/7b	Unbl	900,000	300	119,000	25	$4,780,000	$3,310,000	$8,090,000	$881,000	$439,000	$9,410,000
RF-8a/8b	Unbl	1,500,000	500	198,000	25	$6,500,000	$4,500,000	$11,000,000	$1,200,000	$596,000	$12,800,000
RF-9a/9b	Unbl	2,700,000	900	357,000	25	$9,250,000	$6,400,000	$15,700,000	$1,700,000	$849,000	$18,200,000

(a) All costs in $1991

(b) Low-odor conversion cost = economizer expansion and DCE demolition cost + concentrator cost

Economizer expansion and DCE demolition cost = $6,500,000/[(1,500,000 lb BLS/d/model BLS firing rate) ^ 0.6]

Concentrator cost = $4,500,000/[(1,500,000 lb BLS/d/model BLS firing rate) ^ 0.6]

(c) ESP upgrade cost (to 0.044 gr/dscf) = $1,292,000/[(230,000 acfm/model gas flowrate) ^ 0.6] x (361.3 $1991/357 $May 1993)

(d) Wet- to dry-bottom ESP conversion cost = $845,000/[(357,000 acfm/model gas flowrate) ^ 0.6] x (361.3 $1991/359.4 $July 1993)

(e) TCI = low-odor conversion cost + ESP upgrade cost + wet- to dry-bottom ESP conversion cost

TABLE 6-12a (METRIC). CAPITAL COSTS OF LOW-ODOR CONVERSION (INCLUDING ESP UPGRADE TO CONTROL PM TO 0.10 G/DSCM) FOR MODEL DCE RECOVERY FURNACES (INCLUDING PULP PRODUCTION LOSSES)[a]

Model recovery furnaces	Pulp type	Black liquor firing rate, kg BLS/d	Equivalent ADMP/d	Gas flowrate--ESP exit, m3/sec @ 8% O2	Downtime, d	Economizer expansion, demolition, $	Concentrator, $	Low-odor conversion, $	ESP upgrade, $	Wet- to dry-bottom ESP conversion, $	TCI, $	Pulp production losses, $	TCI+ production losses, $
RF-7a/7b	Bl	400,000	230	56.2	25	$4,780,000	$3,310,000	$8,090,000	$881,000	$439,000	$9,410,000	$488,000	$9,900,000
RF-8a/8b	Bl	700,000	380	93.4	25	$6,500,000	$4,500,000	$11,000,000	$1,200,000	$596,000	$12,800,000	$820,000	$13,600,000
RF-9a/9b	Bl	1,200,000	680	168	25	$9,250,000	$6,400,000	$15,700,000	$1,700,000	$849,000	$18,200,000	$1,460,000	$19,700,000
RF-7a/7b	Unbl	400,000	270	56.2	25	$4,780,000	$3,310,000	$8,090,000	$881,000	$439,000	$9,410,000	$348,000	$9,760,000
RF-8a/8b	Unbl	700,000	450	93.4	25	$6,500,000	$4,500,000	$11,000,000	$1,200,000	$596,000	$12,800,000	$580,000	$13,400,000
RF-9a/9b	Unbl	1,200,000	820	168	25	$9,250,000	$6,400,000	$15,700,000	$1,700,000	$849,000	$18,200,000	$1,040,000	$19,200,000

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 6-12b. Refer to Table 6-12b for footnotes, which include calculations.

TABLE 6-12b (ENGLISH). CAPITAL COSTS OF LOW-ODOR CONVERSION (INCLUDING ESP UPGRADE TO CONTROL PM TO 0.044 GR/DSCF) FOR MODEL DCE RECOVERY FURNACES (INCLUDING PULP PRODUCTION LOSSES)[a]

Model recovery furnaces	Pulp type	Black liquor firing rate, lb BLS/d	Equivalent ADTP/d	Gas flowrate--ESP exit, acfm @ 8% O2	Downtime, d	Economizer expansion, demolition, $	Concentrator, $	Low-odor conversion, $ (b)	ESP upgrade, $ (c)	Wet- to dry-bottom ESP conversion, $ (d)	TCI, $ (e)	Pulp production losses, $ (f)	TCI+ production losses, $
RF-7a/7b	BI	900,000	250	119,000	25	$4,780,000	$3,310,000	$8,090,000	$881,000	$439,000	$9,410,000	$488,000	$9,900,000
RF-8a/8b	BI	1,500,000	420	198,000	25	$6,500,000	$4,500,000	$11,000,000	$1,200,000	$596,000	$12,800,000	$820,000	$13,600,000
RF-9a/9b	BI	2,700,000	750	357,000	25	$9,250,000	$6,400,000	$15,700,000	$1,700,000	$849,000	$18,200,000	$1,460,000	$19,700,000
RF-7a/7b	Unbl	900,000	300	119,000	25	$4,780,000	$3,310,000	$8,090,000	$881,000	$439,000	$9,410,000	$348,000	$9,760,000
RF-8a/8b	Unbl	1,500,000	500	198,000	25	$6,500,000	$4,500,000	$11,000,000	$1,200,000	$596,000	$12,800,000	$580,000	$13,400,000
RF-9a/9b	Unbl	2,700,000	900	357,000	25	$9,250,000	$6,400,000	$15,700,000	$1,700,000	$849,000	$18,200,000	$1,040,000	$19,200,000

(a) All costs in $1991
(b) Low-odor conversion cost = economizer expansion and DCE demolition cost + concentrator cost
Economizer expansion and DCE demolition cost = $6,500,000/[(1,500,000 lb BLS/d/model BLS firing rate) ^ 0.6]
Concentrator cost = $4,500,000/[(1,500,000 lb BLS/d/model BLS firing rate) ^ 0.6]
(c) ESP upgrade cost (to 0.044 gr/dscf) = $1,292,000/[(230,000 acfm/model gas flowrate) ^ 0.6] x (361.3 $1991/357 $May 1993)
(d) Wet- to dry-bottom ESP conversion cost = $845,000/[(357,000 acfm/model gas flowrate) ^ 0.6] x (361.3 $1991/359.4 $July 1993)
(e) TCI = low-odor conversion cost + ESP upgrade cost + wet- to dry-bottom ESP conversion cost
(f) Production losses = (25% gross profit margin) x ($646/ton bleached pulp or $384/ton unbleached pulp) x (136.2 $1991/124 $1989) x (25 d downtime-14 d scheduled downtime) x ADTP/d

TABLE 6-13a (METRIC). CAPITAL COSTS OF LOW-ODOR CONVERSION (INCLUDING ESP UPGRADE TO CONTROL PM TO 0.034 G/DSCM) FOR MODEL DCE RECOVERY FURNACES (EXCLUDING PULP PRODUCTION LOSSES)[a]

Model recovery furnaces	Pulp type	Black liquor firing rate, kg BLS/d	Equivalent ADMP/d	Gas flowrate--ESP exit, m3/sec @ 8% O2	Downtime, d	Economizer expansion, demolition, $	Concentrator, $	Low-odor conversion, $	ESP upgrade, $	Wet- to dry-bottom ESP conversion, $	TCI, $
RF-7a/7b	BI	400,000	230	56.2	25	$4,780,000	$3,310,000	$8,090,000	$1,270,000	$439,000	$9,800,000
RF-8a/8b	BI	700,000	380	93.4	25	$6,500,000	$4,500,000	$11,000,000	$1,840,000	$596,000	$13,400,000
RF-9a/9b	BI	1,200,000	680	168	25	$9,250,000	$6,400,000	$15,700,000	$2,860,000	$849,000	$19,400,000
RF-7a/7b	Unbl	400,000	270	56.2	25	$4,780,000	$3,310,000	$8,090,000	$1,270,000	$439,000	$9,800,000
RF-8a/8b	Unbl	700,000	450	93.4	25	$6,500,000	$4,500,000	$11,000,000	$1,840,000	$596,000	$13,400,000
RF-9a/9b	Unbl	1,200,000	820	168	25	$9,250,000	$6,400,000	$15,700,000	$2,860,000	$849,000	$19,400,000

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 6-13b. Refer to Table 6-13b for footnotes, which include calculations.

TABLE 6-13b (ENGLISH). CAPITAL COSTS OF LOW-ODOR CONVERSION (INCLUDING ESP UPGRADE TO CONTROL PM TO 0.015 GR/DSCF) FOR MODEL DCE RECOVERY FURNACES (EXCLUDING PULP PRODUCTION LOSSES)[a]

Model recovery furnaces	Pulp type	Black liquor firing rate, lb BLS/d	Equivalent ADTP/d	Gas flowrate--ESP exit, acfm @ 8% O2	Downtime, d	Economizer expansion, demolition, $	Concentrator, $	Low-odor conversion, $ (b)	ESP upgrade, $ (c)	Wet- to dry-bottom ESP conversion, $ (d)	TCI, $ (e)
RF-7a/7b	BI	900,000	250	119,000	25	$4,780,000	$3,310,000	$8,090,000	$1,270,000	$439,000	$9,800,000
RF-8a/8bb	BI	1,500,000	420	198,000	25	$6,500,000	$4,500,000	$11,000,000	$1,840,000	$596,000	$13,400,000
RF-9a/9b	BI	2,700,000	750	357,000	25	$9,250,000	$6,400,000	$15,700,000	$2,860,000	$849,000	$19,400,000
RF-7a/7b	Unbl	900,000	300	119,000	25	$4,780,000	$3,310,000	$8,090,000	$1,270,000	$439,000	$9,800,000
RF-8a/8b	Unbl	1,500,000	500	198,000	25	$6,500,000	$4,500,000	$11,000,000	$1,840,000	$596,000	$13,400,000
RF-9a/9b	Unbl	2,700,000	900	357,000	25	$9,250,000	$6,400,000	$15,700,000	$2,860,000	$849,000	$19,400,000

(a) All costs in $1991

(b) Low-odor conversion cost = economizer expansion and DCE demolition cost + concentrator cost
Economizer expansion and DCE demolition cost = $6,500,000/[(1,500,000 lb BLS/d/model BLS firing rate) ^ 0.6]
Concentrator cost = $4,500,000/[(1,500,000 lb BLS/d/model BLS firing rate) ^ 0.6]

(c) ESP upgrade cost (to 0.015 gr/dscf) = ESP upgrade cost (to 0.044 gr/dscf) + [(increase in ESP plate area) x ($39/ft2 plate area)]

(d) Wet- to dry-bottom ESP conversion cost = $845,000/[(357,000 acfm/model gas flowrate) ^ 0.6] x (361.3 $1991/359.4 $July 1993)

(e) TCI = low-odor conversion cost + ESP upgrade cost + wet- to dry-bottom ESP conversion cost

TABLE 6-14a (METRIC). CAPITAL COSTS OF LOW-ODOR CONVERSION (INCLUDING ESP UPGRADE TO CONTROL PM TO 0.034 G/DSCM) FOR MODEL DCE RECOVERY FURNACES (INCLUDING PULP PRODUCTION LOSSES)[a]

Model recovery furnaces	Pulp type	Black liquor firing rate, kg BLS/d	Equivalent ADMP/d	Gas flowrate--ESP exit, m3/sec @ 8% O2	Downtime, d	Economizer expansion, demolition, $	Concentrator, $	Low-odor conversion, $	ESP upgrade, $	Wet- to dry-bottom ESP conversion, $	TCI, $	Pulp production losses, $	TCI+ production losses, $
RF-7a/7b	BI	400,000	230	56.2	25	$4,780,000	$3,310,000	$8,090,000	$1,270,000	$439,000	$9,800,000	$488,000	$10,300,000
RF-8a/8b	BI	700,000	380	93.4	25	$6,500,000	$4,500,000	$11,000,000	$1,840,000	$596,000	$13,400,000	$820,000	$14,200,000
RF-9a/9b	BI	1,200,000	680	168	25	$9,250,000	$6,400,000	$15,700,000	$2,860,000	$849,000	$19,400,000	$1,460,000	$20,900,000
RF-7a/7b	Unbl	400,000	270	56.2	25	$4,780,000	$3,310,000	$8,090,000	$1,270,000	$439,000	$9,800,000	$348,000	$10,100,000
RF-8a/8b	Unbl	700,000	450	93.4	25	$6,500,000	$4,500,000	$11,000,000	$1,840,000	$596,000	$13,400,000	$580,000	$14,000,000
RF-9a/9b	Unbl	1,200,000	820	168	25	$9,250,000	$6,400,000	$15,700,000	$2,860,000	$849,000	$19,400,000	$1,040,000	$20,400,000

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 6-14b. Refer to Table 6-14b for footnotes, which include calculations.

TABLE 6-14b (ENGLISH). CAPITAL COSTS OF LOW-ODOR CONVERSION (INCLUDING ESP UPGRADE TO CONTROL PM TO 0.015 GR/DSCF) FOR MODEL DCE RECOVERY FURNACES (INCLUDING PULP PRODUCTION LOSSES)[a]

Model recovery furnaces	Pulp type	Black liquor firing rate, lb BLS/d	Equivalent ADTP/d	Gas flowrate--ESP exit, acfm @ 8% O2	Downtime, d	Economizer expansion, demolition, $	Concentrator, $	Low-odor conversion, $ (b)	ESP upgrade, $ (c)	Wet- to dry-bottom ESP conversion, $ (d)	TCI, $ (e)	Pulp production losses, $ (f)	TCI+ production losses, $
RF-7a/7b	BI	900,000	250	119,000	25	$4,780,000	$3,310,000	$8,090,000	$1,270,000	$439,000	$9,800,000	$488,000	$10,300,000
RF-8a/8b	BI	1,500,000	420	198,000	25	$6,500,000	$4,500,000	$11,000,000	$1,840,000	$596,000	$13,400,000	$820,000	$14,200,000
RF-9a/9b	BI	2,700,000	750	357,000	25	$9,250,000	$6,400,000	$15,700,000	$2,860,000	$849,000	$19,400,000	$1,460,000	$20,900,000
RF-7a/7b	Unbl	900,000	300	119,000	25	$4,780,000	$3,310,000	$8,090,000	$1,270,000	$439,000	$9,800,000	$348,000	$10,100,000
RF-8a/8b	Unbl	1,500,000	500	198,000	25	$6,500,000	$4,500,000	$11,000,000	$1,840,000	$596,000	$13,400,000	$580,000	$14,000,000
RF-9a/9b	Unbl	2,700,000	900	357,000	25	$9,250,000	$6,400,000	$15,700,000	$2,860,000	$849,000	$19,400,000	$1,040,000	$20,400,000

(a) All costs in $1991

(b) Low-odor conversion cost = economizer expansion and DCE demolition cost + concentrator cost

Economizer expansion and DCE demolition cost = $6,500,000/[(1,500,000 lb BLS/d/model BLS firing rate) ^ 0.6]

Concentrator cost = $4,500,000/[(1,500,000 lb BLS/d/model BLS firing rate) ^ 0.6]

(c) ESP upgrade cost (to 0.015 gr/dscf) = ESP upgrade cost (to 0.044 gr/dscf) + [(increase in ESP plate area) x ($39/ft2 plate area)]

(d) Wet- to dry-bottom ESP conversion cost = $845,000/[(357,000 acfm/model gas flowrate) ^ 0.6] x (361.3 $1991/359.4 $July 1993)

(e) TCI = low-odor conversion cost + ESP upgrade cost + wet- to dry-bottom ESP conversion cost

(f) Production losses = (25% gross profit margin) x ($646/ton bleached pulp or $384/ton unbleached pulp) x (136.2 $1991/124 $1989) x (25 d downtime-14 d scheduled downtime) x ADTP/d

TABLE 6-15a (METRIC). MODEL DCE RECOVERY FURNACE/ESP DESIGN PARAMETERS

Model furnace/ESP design parameters	Model Recovery Furnaces					
	RF-7a	RF-7b	RF-8a	RF-8b	RF-9a	RF-9b
Black liquor firing rate, kg BLS/d	400,000	400,000	700,000	700,000	1,200,000	1,200,000
Equivalent ADMP/d						
Bleached pulp	230	230	380	380	680	680
Unbleached pulp	270	270	450	450	820	820
Gas flow rate--ESP exit, m3/sec	56.2	56.2	93.4	93.4	168	168
ESP SCA, m2/(m3/sec)						
Baseline	66	85	66	85	66	85
PM controls--0.10 g/dscm	105	105	105	105	105	105
PM controls--0.034 g/dscm	121	121	121	121	121	121
ESP plate area, m2						
Baseline	3,688	4,794	6,129	7,967	11,024	14,331
PM controls--0.10 g/dscm	5,900	5,900	9,806	9,806	17,638	17,638
PM controls--0.034 g/dscm	6,822	6,822	11,338	11,338	20,394	20,394
Operating days per year	351	351	351	351	351	351
Operating hours per year	8,424	8,424	8,424	8,424	8,424	8,424

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 6-15b.

TABLE 6-15b (ENGLISH). MODEL DCE RECOVERY FURNACE/ESP DESIGN PARAMETERS

Model furnace/ESP design parameters	Model Recovery Furnaces					
	RF-7a	RF-7b	RF-8a	RF-8b	RF-9a	RF-9b
Black liquor firing rate, lb BLS/d	900,000	900,000	1,500,000	1,500,000	2,700,000	2,700,000
Equivalent ADTP/d						
Bleached pulp	250	250	420	420	750	750
Unbleached pulp	300	300	500	500	900	900
Gas flow rate--ESP exit, acfm	119,000	119,000	198,000	198,000	357,000	357,000
Exhaust gas temperature, degrees F	320	320	320	320	320	320
Moisture content--ESP exit, %	32	32	32	32	32	32
ESP SCA, ft2/1,000 acfm						
Baseline	333	433	333	433	333	433
PM controls--0.044 gr/dscf	533	533	533	533	533	533
PM controls--0.015 gr/dscf	617	617	617	617	617	617
ESP plate area, ft2						
Baseline	39,667	51,567	66,000	85,800	119,000	154,700
PM controls--0.044 gr/dscf	63,467	63,467	105,600	105,600	190,400	190,400
PM controls--0.015 gr/dscf	73,383	73,383	122,100	122,100	220,150	220,150
Operating days per year	351	351	351	351	351	351
Operating hours per year	8,424	8,424	8,424	8,424	8,424	8,424

TABLE 6-16. SCENARIO 1: ANNUAL COSTS OF LOW-ODOR CONVERSION (INCLUDING ESP UPGRADE TO CONTROL PM TO 0.10 G/DSCM [0.044 GR/DSCF]) FOR MODEL DCE RECOVERY FURNACES (EXCLUDING ANNUALIZED PULP PRODUCTION LOSSES)[a]

Costs	Model Recovery Furnaces					
	RF-7a	RF-7b	RF-8a	RF-8b	RF-9a	RF-9b
Total Capital Investment (TCI)	$9,410,000	$9,410,000	$12,800,000	$12,800,000	$18,200,000	$18,200,000
Direct Annual Costs (DAC)						
Steam production credits (b)						
Natural gas	($758,000)	($758,000)	($1,260,000)	($1,260,000)	($2,270,000)	($2,270,000)
Fuel oil	($1,190,000)	($1,190,000)	($1,990,000)	($1,990,000)	($3,580,000)	($3,580,000)
BLO operating cost savings (c)	($103,000)	($103,000)	($172,000)	($172,000)	($309,000)	($309,000)
Concentrator steam costs (d)	$57,800	$57,800	$96,300	$96,300	$173,000	$173,000
ESP operating electricity costs (e)	$23,300	$11,700	$38,800	$19,400	$70,000	$35,000
Total DAC						
Natural gas	($780,000)	($792,000)	($1,300,000)	($1,320,000)	($2,340,000)	($2,370,000)
Fuel oil	($1,210,000)	($1,220,000)	($2,030,000)	($2,050,000)	($3,650,000)	($3,680,000)
Indirect Annual Costs (IAC)						
Administrative, taxes, and insurance (f)						
Low-odor conversion	$324,000	$324,000	$440,000	$440,000	$628,000	$628,000
ESP upgrade	$35,200	$35,200	$48,000	$48,000	$68,000	$68,000
Wet- to dry-bottom ESP conversion	$17,600	$17,600	$23,800	$23,800	$34,000	$34,000
Total	$377,000	$377,000	$512,000	$512,000	$730,000	$730,000
Capital recovery (g)						
Low-odor conversion	$764,000	$764,000	$1,040,000	$1,040,000	$1,480,000	$1,480,000
ESP upgrade	$103,000	$103,000	$140,000	$140,000	$199,000	$199,000
Wet- to dry-bottom ESP conversion	$51,300	$51,300	$69,700	$69,700	$99,200	$99,200
Total	$918,000	$918,000	$1,250,000	$1,250,000	$1,780,000	$1,780,000
Total IAC	$1,300,000	$1,300,000	$1,760,000	$1,760,000	$2,510,000	$2,510,000
Total Annual Costs						
Natural gas	$520,000	$508,000	$460,000	$440,000	$170,000	$140,000
Fuel oil	$90,000	$80,000	($270,000)	($290,000)	($1,140,000)	($1,170,000)

(a) All costs in $1991. Numbers in parentheses represent cost savings.
(b) Steam production credit calculations (either natural gas or fuel oil):
Natural gas credit = (model BLS firing rate) x (6,000 Btu/lb) x (1 d/24 hr) x (66%-56% thermal efficiciency)/(85% power boiler efficiciency) x (1 ft3/1,024 Btu) x (8,424 hr/yr) x ($3.48/1,000 ft3)
Fuel oil credit = (model BLS firing rate) x (6,000 Btu/lb) x (1 d/24 hr) x (66%-56% thermal efficiency)/(85% power boiler efficiency) x (1 gal/144,000 Btu) x (8,424 hr/yr) x ($0.77/gal)
(c) BLO operating cost savings = $251,900/yr x (model BLS firing rate/2,200,000 lb BLS/d)
(d) Concentrator steam costs = $54,000/yr x (model BLS firing rate/2,400,000 lb BLS/d)
(e) ESP operating electricity cost = 0.00184 x increase in ESP plate area x 8,424 hr/yr x $0.06/kWh
(f) Administrative, taxes, and insurance = 0.04 x TCI
(g) For low-odor conversion, capital recovery cost = 0.0944 x TCI (based on 20-yr remaining service life and 7% interest)
For ESP upgrade and wet- to dry-bottom ESP conversion, capital recovery cost = 0.1169 x TCI (based on 13.5-yr ESP life and 7% interest)

TABLE 6-17. SCENARIO 2: ANNUAL COSTS OF LOW-ODOR CONVERSION (INCLUDING ESP UPGRADE TO CONTROL PM TO 0.10 G/DSCM [0.044 GR/DSCF]) FOR MODEL DCE RECOVERY FURNACES (INCLUDING BLEACHED PULP PRODUCTION LOSSES)[a]

Costs	Model Recovery Furnaces					
	RF-7a	RF-7b	RF-8a	RF-8b	RF-9a	RF-9b
Total Capital Investment (TCI)	$9,900,000	$9,900,000	$13,600,000	$13,600,000	$19,700,000	$19,700,000
Direct Annual Costs (DAC)						
Steam production credits (b)						
Natural gas	($758,000)	($758,000)	($1,260,000)	($1,260,000)	($2,270,000)	($2,270,000)
Fuel oil	($1,190,000)	($1,190,000)	($1,990,000)	($1,990,000)	($3,580,000)	($3,580,000)
BLO operating cost savings (c)	($103,000)	($103,000)	($172,000)	($172,000)	($309,000)	($309,000)
Concentrator steam costs (d)	$57,800	$57,800	$96,300	$96,300	$173,000	$173,000
ESP operating electricity costs (d)	$23,300	$11,700	$38,800	$19,400	$70,000	$35,000
Total DAC						
Natural gas	($780,000)	($792,000)	($1,300,000)	($1,320,000)	($2,340,000)	($2,370,000)
Fuel oil	($1,210,000)	($1,220,000)	($2,030,000)	($2,050,000)	($3,650,000)	($3,680,000)
Indirect Annual Costs (IAC)						
Administrative, taxes, and insurance (e)						
Low-odor conversion	324,000	324,000	440,000	440,000	628,000	628,000
ESP upgrade	35,200	35,200	48,000	48,000	68,000	68,000
Wet- to dry-bottom ESP conversion	17,600	17,600	23,800	23,800	34,000	34,000
Total	$377,000	$377,000	$512,000	$512,000	$730,000	$730,000
Capital recovery (f)						
Low-odor conversion	$764,000	$764,000	$1,040,000	$1,040,000	$1,480,000	$1,480,000
ESP upgrade	$103,000	$103,000	$140,000	$140,000	$199,000	$199,000
Wet- to dry-bottom ESP conversion	$51,300	$51,300	$69,700	$69,700	$99,200	$99,200
Production losses	$46,100	$46,100	$77,400	$77,400	$138,000	$138,000
Total	$964,000	$964,000	$1,330,000	$1,330,000	$1,920,000	$1,920,000
Total IAC	$1,340,000	$1,340,000	$1,840,000	$1,840,000	$2,650,000	$2,650,000
Total Annual Costs						
Natural gas	$560,000	$548,000	$540,000	$520,000	$310,000	$280,000
Fuel oil	$130,000	$120,000	($190,000)	($210,000)	($1,000,000)	($1,030,000)

(a) All costs in $1991. Numbers in parentheses represent cost savings.
(b) Steam production credit calculations (either natural gas or fuel oil):
Natural gas credit = (model BLS firing rate) x (6,000 Btu/lb) x (1 d/24 hr) x (66%-56% thermal efficiciency)/(85% power boiler efficiciency) x (1 ft3/1,024 Btu) x (8,424 hr/yr) x ($3.48/1,000 ft3)
Fuel oil credit = (model BLS firing rate) x (6,000 Btu/lb) x (1 d/24 hr) x (66%-56% thermal efficiency)/(85% power boiler efficiency) x (1 gal/144,000 Btu) x (8,424 hr/yr) x ($0.77/gal)
(c) BLO operating cost savings = $251,900/yr x (model BLS firing rate/2,200,000 lb BLS/d)
(d) Concentrator steam costs = $54,000/yr x (model BLS firing rate/2,400,000 lb BLS/d)
(e) ESP operating electricity cost = 0.00194 x increase in ESP plate area x 8,424 hr/yr x $0.06/kWh
(f) Administrative, taxes, and insurance = 0.04 x TCI
(g) For low-odor conversion, capital recovery cost = 0.0944 x TCI (based on 20-yr remaining service life and 7% interest)
For ESP upgrade and wet- to dry-bottom ESP conversion, capital recovery cost = 0.1169 x TCI (based on 13.5-yr ESP life and 7% interest)

TABLE 6-18. SCENARIO 3: ANNUAL COSTS OF LOW-ODOR CONVERSION (INCLUDING ESP UPGRADE TO CONTROL PM TO 0.10 G/DSCM [0.044 GR/DSCF]) FOR MODEL DCE RECOVERY FURNACES (INCLUDING UNBLEACHED PULP PRODUCTION LOSSES)[a]

Costs	Model Recovery Furnaces					
	RF-7a	RF-7b	RF-8a	RF-8b	RF-9a	RF-9b
Total Capital Investment (TCI)	$9,760,000	$9,760,000	$13,400,000	$13,400,000	$19,200,000	$19,200,000
Direct Annual Costs (DAC)						
Steam production credits (b)						
Natural gas	($758,000)	($758,000)	($1,260,000)	($1,260,000)	($2,270,000)	($2,270,000)
Fuel oil	($1,190,000)	($1,190,000)	($1,990,000)	($1,990,000)	($3,580,000)	($3,580,000)
BLO operating cost savings (c)	($103,000)	($103,000)	($172,000)	($172,000)	($309,000)	($309,000)
Concentrator steam costs (d)	$57,800	$57,800	$96,300	$96,300	$173,000	$173,000
ESP operating electricity costs (e)	$23,300	$11,700	$38,800	$19,400	$70,000	$35,000
Total DAC						
Natural gas	($780,000)	($792,000)	($1,300,000)	($1,320,000)	($2,340,000)	($2,370,000)
Fuel oil	($1,210,000)	($1,220,000)	($2,030,000)	($2,050,000)	($3,650,000)	($3,680,000)
Indirect Annual Costs (IAC)						
Administrative, taxes, and insurance (f)						
Low-odor conversion	324,000	324,000	440,000	440,000	628,000	628,000
ESP upgrade	35,200	35,200	48,000	48,000	68,000	68,000
Wet- to dry-bottom ESP conversion	17,600	17,600	23,800	23,800	34,000	34,000
Total	377,000	377,000	512,000	512,000	730,000	730,000
Capital recovery (g)						
Low-odor conversion	$764,000	$764,000	$1,040,000	$1,040,000	$1,480,000	$1,480,000
ESP upgrade	$103,000	$103,000	$140,000	$140,000	$199,000	$199,000
Wet- to dry-bottom ESP conversion	$51,300	$51,300	$69,700	$69,700	$99,200	$99,200
Production losses	$32,900	$32,900	$54,800	$54,800	$98,200	$98,200
Total	951,000	951,000	1,300,000	1,300,000	1,880,000	1,880,000
Total IAC	$1,330,000	$1,330,000	$1,820,000	$1,820,000	$2,610,000	$2,610,000
Total Annual Costs						
Natural gas	$550,000	$538,000	$520,000	$500,000	$270,000	$240,000
Fuel oil	$120,000	$110,000	($210,000)	($230,000)	($1,040,000)	($1,070,000)

(a) All costs in $1991. Numbers in parentheses represent cost savings.
(b) Steam production credit calculations (either natural gas or fuel oil):
Natural gas credit = (model BLS firing rate) x (6,000 Btu/lb) x (1 d/24 hr) x (66%-56% thermal efficiciency)/(85% power boiler efficiciency) x (1 ft3/1,024 Btu) x (8,424 hr/yr) x ($3.48/1,000 ft3)
Fuel oil credit = (model BLS firing rate) x (6,000 Btu/lb) x (1 d/24 hr) x (66%-56% thermal efficiency)/(85% power boiler efficiency) x (1 gal/144,000 Btu) x (8,424 hr/yr) x ($0.77/gal)
(c) BLO operating cost savings = $251,900/yr x (model BLS firing rate/2,200,000 lb BLS/d)
(d) Concentrator steam costs = $54,000/yr x (model BLS firing rate/2,400,000 lb BLS/d)
(e) ESP operating electricity cost = 0.00194 x increase in ESP plate area x 8,424 hr/yr x $0.06/kWh
(f) Administrative, taxes, and insurance = 0.04 x TCI
(g) For low-odor conversion, capital recovery cost = 0.0944 x TCI (based on 20-yr remaining service life and 7% interest)
For ESP upgrade and wet- to dry-bottom ESP conversion, capital recovery cost = 0.1169 x TCI (based on 13.5-yr ESP life and 7% interest)

TABLE 6-19. SCENARIO 1: ANNUAL COSTS OF LOW-ODOR CONVERSION (INCLUDING ESP UPGRADE TO CONTROL PM TO 0.034 G/DSCM [0.015 GR/DSCF]) FOR MODEL DCE RECOVERY FURNACES (EXCLUDING ANNUALIZED PULP PRODUCTION LOSSES)[a]

Costs	Model Recovery Furnaces					
	RF-7a	RF-7b	RF-8a	RF-8b	RF-9a	RF-9b
Total Capital Investment (TCI)	$9,600,000	$9,600,000	$13,400,000	$13,400,000	$19,400,000	$19,400,000
Direct Annual Costs (DAC)						
Steam production credits (b)						
Natural gas	($758,000)	($758,000)	($1,263,000)	($1,260,000)	($2,270,000)	($2,270,000)
Fuel oil	($1,190,000)	($1,190,000)	($1,990,000)	($1,990,000)	($3,580,000)	($3,580,000)
BLO operating cost savings (c)	($103,000)	($103,000)	($172,000)	($172,000)	($309,000)	($309,000)
Concentrator steam costs (d)	$57,800	$57,800	$96,300	$96,300	$173,000	$173,000
ESP operating electricity costs (e)	$33,100	$21,400	$55,000	$35,600	$99,200	$64,200
Total DAC						
Natural gas	($770,000)	($782,000)	($1,280,000)	($1,300,000)	($2,310,000)	($2,340,000)
Fuel oil	($1,200,000)	($1,210,000)	($2,010,000)	($2,030,000)	($3,620,000)	($3,650,000)
Indirect Annual Costs (IAC)						
Administrative, taxes, and insurance (f)						
Low-odor conversion	324,000	324,000	440,000	440,000	628,000	628,000
ESP upgrade	50,800	50,800	73,600	73,600	114,000	114,000
Wet- to dry-bottom ESP conversion	17,600	17,600	23,800	23,800	34,000	34,000
Total	392,000	392,000	537,000	537,000	776,000	776,000
Capital recovery (g)						
Low-odor conversion	$764,000	$764,000	$1,040,000	$1,040,000	$1,480,000	$1,480,000
ESP upgrade	$148,000	$148,000	$215,000	$215,000	$334,000	$334,000
Wet- to dry-bottom ESP conversion	$51,300	$51,300	$69,700	$69,700	$99,200	$99,200
Total	963,000	963,000	1,320,000	1,320,000	1,910,000	1,910,000
Total IAC	$1,360,000	$1,360,000	$1,860,000	$1,860,000	$2,690,000	$2,690,000
Total Annual Costs						
Natural gas	$590,000	$578,000	$580,000	$560,000	$380,000	$350,000
Fuel oil	$160,000	$150,000	($150,000)	($170,000)	($930,000)	($960,000)

(a) All costs in $1991. Numbers in parentheses represent cost savings.
(b) Steam production credit calculations (either natural gas or fuel oil):
Natural gas credit = (model BLS firing rate) x (6,000 Btu/lb) x (1 d/24 hr) x (66%-56% thermal efficiciency)/(85% power boiler efficiciency) x (1 ft3/1,024 Btu) x (8,424 hr/yr) x ($3.48/1,000 ft3)
Fuel oil credit = (model BLS firing rate) x (6,000 Btu/lb) x (1 d/24 hr) x (66%-56% thermal efficiency)/(85% power boiler efficiency) x (1 gal/144,000 Btu) x (8,424 hr/yr) x ($0.77/gal)
(c) BLO operating cost savings = $251,900/yr x (model BLS firing rate/2,200,000 lb BLS/d)
(d) Concentrator steam costs = $54,000/yr x (model BLS firing rate/2,400,000 lb BLS/d)
(e) ESP operating electricity cost = 0.00194 x increase in ESP plate area x 8,424 hr/yr x $0.06/kWh
(f) Administrative, taxes, and insurance = 0.04 x TCI
(g) For low-odor conversion, capital recovery cost = 0.0944 x TCI (based on 20-yr remaining service life and 7% interest)
For ESP upgrade and wet- to dry-bottom ESP conversion, capital recovery cost = 0.1169 x TCI (based on 13.5-yr ESP life and 7% interest)

TABLE 6-20. SCENARIO 2: ANNUAL COSTS OF LOW-ODOR CONVERSION (INCLUDING ESP UPGRADE TO CONTROL PM TO 0.034 G/DSCM [0.015 GR/DSCF]) FOR MODEL DCE RECOVERY FURNACES (INCLUDING BLEACHED PULP PRODUCTION LOSSES)[a]

Costs	Model Recovery Furnaces					
	RF-7a	RF-7b	RF-8a	RF-8b	RF-9a	RF-9b
Total Capital Investment (TCI)	$10,300,000	$10,300,000	$14,200,000	$14,200,000	$20,900,000	$20,900,000
Direct Annual Costs (DAC)						
Steam production credits (b)						
Natural gas	($758,000)	($758,000)	($1,260,000)	($1,260,000)	($2,270,000)	($2,270,000)
Fuel oil	($1,190,000)	($1,190,000)	($1,990,000)	($1,990,000)	($3,580,000)	($3,580,000)
BLO operating cost savings (c)	($103,000)	($103,000)	($172,000)	($172,000)	($309,000)	($309,000)
Concentrator steam costs (d)	$57,800	$57,800	$96,300	$96,300	$173,000	$173,000
ESP operating electricity costs (d)	$33,100	$21,400	$55,000	$35,600	$99,200	$64,200
Total DAC						
Natural gas	($770,000)	($782,000)	($1,280,000)	($1,300,000)	($2,310,000)	($2,340,000)
Fuel oil	($1,200,000)	($1,210,000)	($2,010,000)	($2,030,000)	($3,620,000)	($3,650,000)
Indirect Annual Costs (IAC)						
Administrative, taxes, and insurance (e)						
Low-odor conversion	324,000	324,000	440,000	440,000	628,000	628,000
ESP upgrade	50,800	50,800	73,600	73,600	114,000	114,000
Wet- to dry-bottom ESP conversion	17,600	17,600	23,800	23,800	34,000	34,000
Total	392,000	392,000	537,000	537,000	776,000	776,000
Capital recovery (f)						
Low-odor conversion	$764,000	$764,000	$1,040,000	$1,040,000	$1,480,000	$1,480,000
ESP upgrade	$148,000	$148,000	$215,000	$215,000	$334,000	$334,000
Wet- to dry-bottom ESP conversion	$51,300	$51,300	$69,700	$69,700	$99,200	$99,200
Production losses	$46,100	$46,100	$77,400	$77,400	$138,000	$138,000
Total	1,010,000	1,010,000	1,400,000	1,400,000	2,050,000	2,050,000
Total IAC	$1,400,000	$1,400,000	$1,940,000	$1,940,000	$2,830,000	$2,830,000
Total Annual Costs						
Natural gas	$630,000	$618,000	$660,000	$640,000	$520,000	$490,000
Fuel oil	$200,000	$190,000	($70,000)	($90,000)	($790,000)	($820,000)

(a) All costs in $1991. Numbers in parentheses represent cost savings.
(b) Steam production credit calculations (either natural gas or fuel oil):
Natural gas credit = (model BLS firing rate) x (6,000 Btu/lb) x (1 d/24 hr) x (66%-56% thermal efficiciency)/(85% power boiler efficiciency) x (1 ft3/1,024 Btu) x (8,424 hr/yr) x ($3.48/1,000 ft3)
Fuel oil credit = (model BLS firing rate) x (6,000 Btu/lb) x (1 d/24 hr) x (66%-56% thermal efficiency)/(85% power boiler efficiency) x (1 gal/144,000 Btu) x (8,424 hr/yr) x ($0.77/gal)
(c) BLO operating cost savings = $251,900/yr x (model BLS firing rate/2,200,000 lb BLS/d)
(d) Concentrator steam costs = $54,000/yr x (model BLS firing rate/2,400,000 lb BLS/d)
(e) ESP operating electricity cost = 0.00194 x increase in ESP plate area x 8,424 hr/yr x $0.06/kWh
(f) Administrative, taxes, and insurance = 0.04 x TCI
(g) For low-odor conversion and production losses, capital recovery cost = 0.0944 x TCI (based on 20-yr remaining service life and 7% interest)
For ESP upgrade and wet- to dry-bottom ESP conversion, capital recovery cost = 0.1169 x TCI (based on 13.5-yr ESP life and 7% interest)

TABLE 6-21. SCENARIO 3: ANNUAL COSTS OF LOW-ODOR CONVERSION (INCLUDING ESP UPGRADE TO CONTROL PM TO 0.034 G/DSCM [0.015 GR/DSCF]) FOR MODEL DCE RECOVERY FURNACES (INCLUDING UNBLEACHED PULP PRODUCTION LOSSES)[a]

Costs	Model Recovery Furnaces					
	RF-7a	RF-7b	RF-8a	RF-8b	RF-9a	RF-9b
Total Capital Investment (TCI)	$10,100,000	$10,100,000	$14,000,000	$14,000,000	$20,400,000	$20,400,000
Direct Annual Costs (DAC)						
Steam production credits (b)						
Natural gas	($758,000)	($758,000)	($1,260,000)	($1,260,000)	($2,270,000)	($2,270,000)
Fuel oil	($1,190,000)	($1,190,000)	($1,990,000)	($1,990,000)	($3,580,000)	($3,580,000)
BLO operating cost savings (c)	($103,000)	($103,000)	($172,000)	($172,000)	($309,000)	($309,000)
Concentrator steam costs (d)	$57,800	$57,800	$96,300	$96,300	$173,000	$173,000
ESP operating electricity costs (e)	$33,100	$21,400	$55,000	$35,600	$99,200	$64,200
Total DAC						
Natural gas	($770,000)	($782,000)	($1,280,000)	($1,300,000)	($2,310,000)	($2,340,000)
Fuel oil	($1,200,000)	($1,210,000)	($2,010,000)	($2,030,000)	($3,620,000)	($3,650 000)
Indirect Annual Costs (IAC)						
Administrative, taxes, and insurance (f)						
Low-odor conversion	324,000	324,000	440,000	440,000	628,000	628,000
ESP upgrade	50,800	50,800	73,600	73,600	114,000	114,000
Wet- to dry-bottom ESP conversion	17,600	17,600	23,800	23,800	34,000	34,000
Total	392,000	392,000	537,000	537,000	776,000	776,000
Capital recovery (g)						
Low-odor conversion	$764,000	$764,000	$1,040,000	$1,040,000	$1,480,000	$1,480,000
ESP upgrade	$148,000	$148,000	$215,000	$215,000	$334,000	$334,000
Wet- to dry-bottom ESP conversion	$51,300	$51,300	$69,700	$69,700	$99,200	$99,200
Production losses	$32,900	$32,900	$54,800	$54,800	$98,200	$98,200
Total	996,000	996,000	1,380,000	1,380,000	2,010,000	2,010,000
Total IAC	$1,390,000	$1,390,000	$1,920,000	$1,920,000	$2,790,000	$2,790,000
Total Annual Costs						
Natural gas	$620,000	$608,000	$640,000	$620,000	$480,000	$450,000
Fuel oil	$190,000	$180,000	($90,000)	($110,000)	($830,000)	($860,000)

(a) All costs in $1991. Numbers in parentheses represent cost savings.
(b) Steam production credit calculations (either natural gas or fuel oil):
Natural gas credit = (model BLS firing rate) x (6,000 Btu/lb) x (1 d/24 hr) x (66%-56% thermal efficiciency)/(85% power boiler efficiciency) x (1 ft3/1,024 Btu) x (8,424 hr/yr) x ($3.48/1,000 ft3)
Fuel oil credit = (model BLS firing rate) x (6,000 Btu/lb) x (1 d/24 hr) x (66%-56% thermal efficiency)/(85% power boiler efficiency) x (1 gal/144,000 Btu) x (8,424 hr/yr) x ($0.77/gal)
(c) BLO operating cost savings = $251,900/yr x (model BLS firing rate/2,200,000 lb BLS/d)
(d) Concentrator steam costs = $54,000/yr x (model BLS firing rate/2,400,000 lb BLS/d)
(e) ESP operating electricity cost = 0.00194 x increase in ESP plate area x 8,424 hr/yr x $0.06/kWh
(f) Administrative, taxes, and insurance = 0.04 x TCI
(g) For low-odor conversion and production losses, capital recovery cost = 0.0944 x TCI (based on 20-yr remaining service life and 7% interest)
For ESP upgrade and wet- to dry-bottom ESP conversion, capital recovery cost = 0.1169 x TCI (based on 13.5-yr ESP life and 7% interest)

TABLE 6-22a (METRIC). GAS AND LIQUID STREAM PARAMETERS FOR RECOVERY FURNACE MODEL PROCESS UNITS[a]

Parameters	Model NDCE/Converted DCE Recovery Furnaces				Model DCE Recovery Furnaces		
	RF-1/4/8	RF-2/5/9	RF-3/6	RF-7	RF-7	RF-8	RF-9
Gas Stream Properties:							
ESP Outlet:							
Flow rate, m^3/sec	93	168	243	56	56	93	168
Temperature, °C	199	199	199	199	160	160	160
Moisture content, %	26	26	26	26	32	32	32
Out of Quench:							
Flow rate, m^3/sec	76	136	197	45	48	80	144
Temperature, °C	72	72	72	72	74	74	74
Moisture content, %	33	33	33	33	37	37	37
Density, kg/m^3	0.8946	0.8946	0.8946	0.8946	0.8762	0.8762	0.8762
Molecular weight, g/gmole	25.3	25.3	25.3	25.3	25.0	25.0	25.0
Viscosity, kg/m-hr	0.0617	0.0617	0.0617	0.0617	0.0607	0.0607	0.0607
HCl concentration, ppmv	9.7	9.7	9.7	9.7	9.2	9.2	9.2
HCl emissions, kg/d	82	147	212	49	49	82	147
Out of Absorber:							
HCl concentration, ppmv	5	5	5	5	5	5	5
Liquid Stream Properties:							
HCl concentration in entering liquid, gmole HCl per gmole pollutant-free liquid	0	0	0	0	0	0	0
Density, kg/m^3	999	999	999	999	999	999	999
Molecular weight, g/gmole	18	18	18	18	18	18	18
Pollutant Properties:							
Diffusivity of HCl in air, m^2/hr	0.0673	0.0673	0.0673	0.0673	0.0673	0.0673	0.0673
Diffusivity of HCl in water, m^2/hr	9.4E-6	9.4-6	9.4E-6	9.4E-6	9.4E-6	9.4E-6	9.4E-6

[a]Metric equivalents in this table were converted from the calculated English unit values given in Table 6-22b.

TABLE 6-22b (ENGLISH). GAS AND LIQUID STREAM PARAMETERS FOR RECOVERY FURNACE MODEL PROCESS UNITS

Parameters	Model NDCE/Converted DCE Recovery Furnaces				Model DCE Recovery Furnaces		
	RF-1/4/8	RF-2/5/9	RF-3/6	RF-7	RF-7	RF-8	RF-9
Gas Stream Properties:							
ESP Outlet:							
Flow rate, acfm	198,000	357,000	515,000	119,000	119,000	198,000	357,000
Temperature, °F	390	390	390	390	320	320	320
Moisture content, %	26	26	26	26	32	32	32
Out of Quench[a]:							
Flow rate, acfm	161,000	289,000	417,000	96,500	102,000	170,000	307,000
Temperature, °F	161	161	161	161	165	165	165
Moisture content, %	33	33	33	33	37	37	37
Density, lb/ft^3	0.05587	0.05587	0.05587	0.05587	0.05470	0.05470	0.05470
Molecular weight, lb/lbmole	25.3	25.3	25.3	25.3	25.0	25.0	25.0
Viscosity, lb/ft-hr	0.0415	0.0415	0.0415	0.0415	0.0408	0.0408	0.0408
HCl concentration, ppmv	9.7	9.7	9.7	9.7	9.2	9.2	9.2
HCl emissions, lb/d	180	324	468	108	108	180	324
Out of Absorber:							
HCl concentration, ppmv	5	5	5	5	5	5	5
Liquid Stream Properties:							
HCl concentration in entering liquid, lbmole HCl per lbmole pollutant-free liquid	0	0	0	0	0	0	0
Density, lb/ft^3	62.4	62.4	62.4	62.4	62.4	62.4	62.4
Molecular weight, lb/lbmole	18	18	18	18	18	18	18
Pollutant Properties:							
Diffusivity of HCl in air, ft^2/hr	0.725	0.725	0.725	0.725	0.725	0.725	0.725
Diffusivity of HCl in water, ft^2/hr	1.02E-4	1.02E-4	1.02E-4	1.02E-4	1.02E-4	1.02E-4	1.02E-4

[a]To simplify the packed-bed scrubber design analysis, it was assumed that the gas stream exiting the ESP was cooled to saturation with water sprays.

TABLE 6-23a (METRIC). PACKED-BED SCRUBBER DESIGN AND OPERATING PARAMETERS[a]

Packed-Bed Scrubber Parameters	Model NDCE/Converted DCE Recovery Furnaces				Model DCE Recovery Furnaces		
	RF-1/4/8	RF-2/5/9	RF-3/6	RF-7	RF-7	RF-8	RF-9
Packed-Bed Scrubber Design Parameters:							
Cross-sectional area, m^2	36	64	93	21	22	37	67
Vessel diameter, m	6.7	9.0	11	5.2	5.4	6.9	9.3
Packing height, m	0.47	0.47	0.47	0.47	0.43	0.43	0.43
Tower height, m	8.4	11	13	6.8	6.9	8.5	11
Surface area, m^2	249	433	615	155	161	259	452
Pressure drop, mm Hg	2.4	2.4	2.4	2.4	2.2	2.2	2.2
Auxiliary Equipment:							
Stack height, m	30.5	30.5	30.5	30.5	30.5	30.5	30.5
Stack diameter, m	3.1	4.1	5.0	2.4	2.5	3.2	4.3
Fan impeller diameter, m	2.7	3.6	4.1	2.2	2.3	2.8	3.7
Fan motor efficiency, fraction	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Pump motor efficiency, fraction	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Power for fan, kW	34.4	62.3	89.5	20.6	20.2	33.6	60.7
Power for pump, kW	28.1	50.6	73.1	16.9	17.7	29.5	53.1
Wastewater flow rate, L/min	6.25	11.2	16.2	3.75	3.21	5.36	9.65
Caustic addition (dry NaOH), kg/hr	25.6	46.1	66.6	15.4	13.2	22.0	39.6
Column Operation:							
Minimum wetting rate, m^2/hr	0.12	0.12	0.12	0.12	0.12	0.12	0.12
Fraction of flooding gas velocity	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Operating hours per year							
Equipment	8,424	8,424	8,424	8,424	8,424	8,424	8,424
Labor	8,760	8,760	8,760	8,760	8,760	8,760	8,760
Operating labor requirement, hr/d	1.5	1.5	1.5	1.5	1.5	1.5	1.5
Maintenance labor requirement, hr/d	1.5	1.5	1.5	1.5	1.5	1.5	1.5

[a]Metric equivalents in this table were converted from the calculated English unit values given in Table 6-23b.

TABLE 6-23b (ENGLISH). PACKED-BED SCRUBBER DESIGN AND OPERATING PARAMETERS

Packed-Bed Scrubber Parameters	Model NDCE/Converted DCE Recovery Furnaces				Model DCE Recovery Furnaces		
	RF-1/4/8	RF-2/5/9	RF-3/6	RF-7	RF-7	RF-8	RF-9
Packed-Bed Scrubber Design Parameters:							
Cross-sectional area, ft^2	384	690	998	231	242	403	725
Vessel diameter, ft	22.1	29.6	35.6	17.1	17.6	22.7	30.4
Packing height, ft	1.54	1.54	1.54	1.53	1.41	1.41	1.41
Tower height, ft	27.5	35.2	41.3	22.4	22.7	27.9	35.8
Surface area, ft^2	2,679	4,657	6,624	1,672	1,735	2,791	4,864
Pressure drop, in H_2O	1.3	1.3	1.3	1.3	1.2	1.2	1.2
Auxiliary Equipment:							
Stack height, ft	100	100	100	100	100	100	100
Stack diameter, ft	10.1	13.6	16.3	7.8	8.1	10.4	14.0
Fan impeller diameter, in	105	140	160	85	90	110	145
Fan motor efficiency, fraction	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Pump motor efficiency, fraction	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Power for fan, kW	34.4	62.3	89.5	20.6	20.2	33.6	60.7
Power for pump, kW	28.1	50.6	73.1	16.9	17.7	29.5	53.1
Wastewater flow rate, gpm	1.651	2.971	4.291	0.990	0.849	1.415	2.548
Caustic addition (dry NaOH), lb/hr	56.5	102	147	33.9	29.1	48.4	87.2
Column Operation:							
Minimum wetting rate, ft^2/hr	1.3	1.3	1.3	1.3	1.3	1.3	1.3
Fraction of flooding gas velocity	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Operating hours per year Equipment Labor	 8,424 8,760	 8,424 8,760	 8,424 8,760	 8,424 8,760	 8,424 8,760	 8,424 8,760	 8,424 8,760
Operating labor requirement, hr/d	1.5	1.5	1.5	1.5	1.5	1.5	1.5
Maintenance labor requirement, hr/d	1.5	1.5	1.5	1.5	1.5	1.5	1.5

TABLE 6-24a (METRIC). UNIT COSTS FOR PACKED-BED SCRUBBER[a]

Parameters	Value
Operator wage rate, $/hr	15.64
Water cost, $/m^3	0.05
Electricity, $/kWh	0.06
Wastewater disposal, $/m^3	1.0
Caustic (dry NaOH), $/Mg	441
Packing material, $/m^3	706

[a]Metric equivalents in this table were converted from the calculated English unit values given in Table 6-24b.

TABLE 6-24b (ENGLISH). UNIT COSTS FOR PACKED-BED SCRUBBER[a,b]

Parameters	Value
Operator wage rate, $/hr	15.64
Water cost, $/1,000 gal	0.2
Electricity, $/kWh	0.06
Wastewater disposal, $/1,000 gal	3.8
Caustic (dry NaOH), $/ton	400
Packing material, $/ft^3	20

[a]OAQPS Control Cost Manual
[b]Medical Waste Incinerators--Background Information for Proposed Standards and Guidelines: Model Plant Description and Cost Report for New and Existing Facilities. U. S. Environmental Protection Agency. Research Triangle Park, NC. Publication No. EPA-453/R-94-044a. July 1994. p. 62.

TABLE 6-25. PACKED-BED SCRUBBER CAPITAL COSTS FOR MODEL RECOVERY FURNACES[a]

Costs	Model NDCE/Converted DCE Recovery Furnaces				Model DCE Recovery Furnaces		
	RF-1/4/8	RF-2/5/9	RF-3/6	RF-7	RF-7	RF-8	RF-9
Equipment Costs[b]:							
Tower	$308,139	$535,566	$761,709	$192,226	$199,517	$320,931	$559,395
Packing	$11,810	$21,196	$30,685	$7,084	$6,809	$11,366	$20,409
Pump	$27,888	$50,112	$72,480	$16,776	$17,575	$29,268	$52,653
Fan	$35,639	$53,007	$63,733	$26,624	$28,809	$38,001	$55,637
Fan motor	$2,414	$3,935	$5,297	$1,583	$1,557	$2,367	$3,848
Stack	$36,394	$48,868	$58,695	$28,214	$29,075	$37,505	$50,360
Total Equipment Cost (TEC):	$422,283	$712,685	$992,598	$272,507	$283,343	$439,438	$742,303
Purchased Equipment Cost (PEC)[c]:	$498,294	$840,968	$1,171,266	$321,559	$334,344	$518,536	$875,918
Total Capital Investment (TCI)[d]:	$1,100,000	$1,850,000	$2,580,000	$707,000	$736,000	$1,140,000	$1,930,000

[a]All costs in $1991

[b]Equipment cost equations:

FRP tower = 115 x surface area of column

Packing = volume of packing x ($20/ft^3 of packing)

Pump = liquid flow rate x ($16/gpm)

Fan = 57.9 x (fan impeller diameter)$^{1.38}$

Fan motor = 104 x $(E_{fan}/0.746)^{0.821}$

Stack = 36 x stack diameter x stack height

[c]Purchased equipment cost = 1.18 x (TEC)

[d]Total capital investment = 2.2 x (PEC)

TABLE 6-26. PACKED-BED SCRUBBER ANNUAL COSTS FOR MODEL RECOVERY FURNACES[a]

Costs	Model NDCE/Converted DCE Recovery Furnaces				Model DCE Recovery Furnaces		
	RF-1/4/8	RF-2/5/9	RF-3/6	RF-7	RF-7	RF-8	RF-9
Total Capital Investment (TCI):	$1,100,000	$1,850,000	$2,580,000	$707,000	$736,000	$1,140,000	$1,930,000
Direct Annual Costs (DAC):							
Operating labor[b]:							
Operator	$8,563	$8,563	$8,563	$8,563	$8,563	$8,563	$8,563
Supervisor	$1,284	$1,284	$1,284	$1,284	$1,284	$1,284	$1,284
Maintenance[c]:							
Labor	$9,419	$9,419	$9,419	$9,419	$9,419	$9,419	$9,419
Material	$9,419	$9,419	$9,419	$9,419	$9,419	$9,419	$9,419
Utilities[d]:							
Electricity (fan and pump)	$31,598	$57,062	$82,216	$18,949	$19,150	$31,888	$57,515
Water	$7,803	$14,069	$20,297	$4,690	$3,481	$5,793	$10,444
Caustic[e]:	$95,145	$171,262	$247,378	$57,087	$48,957	$81,596	$146,872
Wastewater disposal[f]:	$3,170	$5,706	$8,242	$1,902	$1,631	$2,719	$4,894
Total DAC:	$166,401	$276,784	$386,818	$111,313	$101,904	$150,681	$248,410
Indirect Annual Costs (IAC):							
Overhead[g]:	$17,211	$17,211	$17,211	$17,211	$17,211	$17,211	$17,211
Admin., taxes and insurance[h]:	$43,850	$74,005	$103,071	$28,297	$29,422	$45,631	$77,081
Capital recovery[i]:	$120,368	$203,144	$282,931	$77,676	$80,764	$125,258	$211,587
Total IAC:	$181,429	$294,360	$403,213	$123,184	$127,397	$188,100	$305,879
Total Annual Costs (TAC):	$348,000	$571,000	$790,000	$234,000	$229,000	$339,000	$554,000

[a]All costs in $1991
[b]Operator labor = 1.5 hr/d x 365 d/yr x $15.64/hr; supervisor labor = 15% of operator labor cost
[c]Maintenance labor = 1.5 hr/d x 365 d/yr x (1.1 x $15.64/hr); maintenance materials = 100% of maintenance labor cost
[d]Electricity (fan and pump) = (E_{fan} + E_{pump}) x 8,424 hr/yr x $0.06/kWh
Water = [wastewater flow rate + (water vapor out of quench - water vapor out of ESP)] (gpm) x 60 min/hr x 8,424 hr/yr x $0.20/gal H_2O
[e]Caustic = caustic addition (lb/hr) x 8,424 hr/yr x 1 ton/2,000 lb x $400/ton caustic
[f]Wastewater disposal = wastewater flow rate (gpm) x 60 min/hr x 8,424 hr/yr x $3.80/1,000 gal wastewater
[g]Overhead = 0.6 x (maintenance cost + operating labor cost)
[h]Administrative, taxes, and insurance = 0.04 x TCI
[i]Capital recovery = 0.1098 x TCI (based on 15-yr scrubber life and 7% interest)

TABLE 6-27a (METRIC). MODEL BLACK LIQUOR OXIDATION UNIT DESIGN PARAMETERS[a]

Design parameters	Model BLO units		
	BLO-1	BLO-2	BLO-3
Black liquor firing rate, kg BLS/d	400,000	700,000	1,200,000
Equivalent ADMP/d			
Bleached pulp	230	380	680
Unbleached pulp	270	450	820
Vent gas flowrate, m3/sec @ 54 degrees C	4.2	8.5	12.7
Operating days per year			
Equipment	351	351	351
Labor	365	365	365
Operating hours per year			
Equipment	8,424	8,424	8,424
Labor	8,760	8,760	8,760

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 6-27b.

TABLE 6-27b (ENGLISH). MODEL BLACK LIQUOR OXIDATION UNIT DESIGN PARAMETERS[a]

Design parameters	Model BLO units		
	BLO-1	BLO-2	BLO-3
Black liquor firing rate, lb BLS/d	900,000	1,500,000	2,700,000
Equivalent ADTP/d			
Bleached pulp	250	420	750
Unbleached pulp	300	500	900
Vent gas flowrate, acfm @ 130 degrees F	8,900	18,000	26,900
Operating days per year			
Equipment	351	351	351
Labor	365	365	365
Operating hours per year			
Equipment	8,424	8,424	8,424
Labor	8,760	8,760	8,760

TABLE 6-28. CAPITAL AND ANNUAL COSTS OF COLLECTION AND INCINERATION OF BLO VENT GASES FOR MODEL BLACK LIQUOR OXIDATION UNITS[a]

Costs	Model BLO units		
	BLO-1	BLO-2	BLO-3
Total Capital Investment (TCI) (b)	2,500,000	3,390,000	4,830,000
Direct Annual Costs (DAC)			
Operating labor (c)			
Operator	$18,600	$18,600	$18,600
Supervisor	$2,790	$2,790	$2,790
Maintenance (d)			
Labor	$27,400	$27,400	$27,400
Material	$27,400	$27,400	$27,400
Electricity (e)	$103,000	$209,000	$312,000
Steam (f)	$22,000	$44,600	$66,600
Total DAC	$179,000	$285,000	$388,000
Indirect Annual Costs (IAC)			
Overhead (g)	$45,700	$45,700	$45,700
Administrative, taxes and insurance (h)	$100,000	$136,000	$193,000
Capital recovery (i)	$356,000	$483,000	$688,000
Total IAC	$502,000	$665,000	$927,000
Total Annual Costs	$681,000	$950,000	$1,320,000

(a) All costs in $1991.
(b) TCI = $4,800,000 x [(model BLS rate)/(800 ADTP/d x 3,400 lb BLS/ADTP)] ^ 0.6 x (361.3 $1991/357.6 $1990)
(c) Operating labor = 0.5 hr/shift/condenser x 3 shifts/d x 2 condensers x 365 d/yr x $17/hr
Supervisor labor = 15% of operator labor
(d) Maintenance labor = 0.5 hr/shift/condenser x 3 shifts/d x 2 condensers x 365 d/yr x $25/hr
Maintenance materials = 100 percent of maintenance labor
(e) Electricity costs = Mill A electricity costs x (model vent gas flowrate/Mill A vent gas flowrate)
Mill A electricity costs = (100 hp + 3 hp + 400 hp) x 0.746 kW/hp x 8,424 hr/yr x $0.06/kWh
The Mill A electricity costs are based on 100 hp to operate the mill water booster pump motor, 3 hp to operate the BLO condenser condensate pump motor, and 400 hp to operate the BLO off gas blower motor.
The Mill A vent gas flowrate = 16,327 acfm
(f) Steam cost = Mill A steam cost x (model vent gas flowrate/Mill A vent gas flowrate)
Mill A steam cost = 1,600 lb steam/hr x 8,424 hr/yr x $3/1,000 lb steam
The Mill A steam cost is based on steam requirements of 1,600 lb steam/hr for the BLO off gas reheater.
The Mill A vent gas flowrate = 16,327 acfm
(g) Overhead = 0.6 x (labor + maintenance)
(h) Administrative, taxes, and insurance = 0.04 x TCI
(i) Capital recovery = 0.1424 x TCI, based on 7% interest and 10-yr equip. life for ductwork and condenser.

TABLE 6-29a (METRIC). SCRUBBER REPLACEMENT COSTS FOR MODEL SDT'S[a]

Model SDT	Pulp type	Black liquor firing rate, kg BLS/d	Equivalent ADMP/d	Baseline APCD	Control Option APCD	Gas flowrate--APCD inlet, m3/sec	TCI, $	A,I & T, $/yr	Capital recovery, $/yr	ITAC, $/yr
SDT-1	BI	400,000	230	scrubber	scrubber	4.4	$292,000	$11,700	$32,100	$43,800
SDT-2	BI	700,000	380	scrubber	scrubber	7.4	$398,000	$15,900	$43,700	$59,600
SDT-3	BI	1,200,000	680	scrubber	scrubber	13.4	$566,000	$22,600	$62,100	$84,700
SDT-4	BI	1,800,000	1,000	scrubber	scrubber	19.3	$706,000	$28,200	$77,500	$106,000
SDT-1	Unbl	400,000	270	scrubber	scrubber	4.4	$292,000	$11,700	$32,100	$43,800
SDT-2	Unbl	700,000	450	scrubber	scrubber	7.4	$398,000	$15,900	$43,700	$59,600
SDT-3	Unbl	1,200,000	820	scrubber	scrubber	13.4	$566,000	$22,600	$62,100	$84,700
SDT-4	Unbl	1,800,000	1,200	scrubber	scrubber	19.3	$706,000	$28,200	$77,500	$106,000

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 6-29b. Refer to Table 6-29b for footnotes, which include calculations.

TABLE 6-29b (ENGLISH). SCRUBBER REPLACEMENT COSTS FOR MODEL SDT'S[a]

Model SDT	Pulp type	Black liquor firing rate, lb BLS/d	Equivalent ADTP/d	Baseline APCD	Control Option APCD	Gas flowrate--APCD inlet, acfm	TCI, $ (b)	A,I & T, $/yr (c)	Capital recovery, $/yr (d)	ITAC, $/yr (e)
SDT-1	BI	900,000	250	scrubber	scrubber	9,400	$292,000	$11,700	$32,100	$43,800
SDT-2	BI	1,500,000	420	scrubber	scrubber	15,700	$398,000	$15,900	$43,700	$59,600
SDT-3	BI	2,700,000	750	scrubber	scrubber	28,300	$566,000	$22,600	$62,100	$84,700
SDT-4	BI	3,900,000	1,100	scrubber	scrubber	40,900	$706,000	$28,200	$77,500	$106,000
SDT-1	Unbl	900,000	300	scrubber	scrubber	9,400	$292,000	$11,700	$32,100	$43,800
SDT-2	Unbl	1,500,000	500	scrubber	scrubber	15,700	$398,000	$15,900	$43,700	$59,600
SDT-3	Unbl	2,700,000	900	scrubber	scrubber	28,300	$566,000	$22,600	$62,100	$84,700
SDT-4	Unbl	3,900,000	1,300	scrubber	scrubber	40,900	$706,000	$28,200	$77,500	$106,000

(a) All costs in $1991
(b) TCI = ($280,000 x [361.3 $1991/342.5 $1988] + $50,000) x (model inlet gas flowrate/[8,071 acfm/0.65]) ^ 0.6
(c) Administrative, insurance, and taxes (A,I & T) = 0.04 x TCI
(d) Capital recovery = 0.1098 x TCI (based on 15-yr scrubber life and 7% interest)
(e) Incremental total annual cost (ITAC) = A,I & T + capital recovery

TABLE 6-30a (METRIC). MODEL SDT/SCRUBBER DESIGN PARAMETERS[a]

Design parameters	Model SDT's		
	SDT-5	SDT-6	SDT-7
Baseline APCD:	mist eliminator	mist eliminator	mist eliminator
Control option APCD:	scrubber	scrubber	scrubber
Black liquor firing rate, kg BLS/d:	400,000	700,000	1,200,000
Equivalent ADMP/d:			
Bleached pulp	230	380	680
Unbleached pulp	270	450	820
Gas flowrate--APCD inlet (Q), m3/sec:	4.4	7.4	13.4
Existing pressure drop, mm Hg:	1.3	1.3	1.3
New pressure drop, mm Hg:	13	13	13
Operating days per year:			
Equipment	351	351	351
Labor	365	365	365
Operating hours per year:			
Equipment	8,424	8,424	8,424
Labor	8,760	8,760	8,760

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 6-30b.

TABLE 6-30b (ENGLISH). MODEL SDT/SCRUBBER DESIGN PARAMETERS

Design parameters	Model SDT's		
	SDT-5	SDT-6	SDT-7
Baseline APCD:	mist eliminator	mist eliminator	mist eliminator
Control option APCD:	scrubber	scrubber	scrubber
Black liquor firing rate, lb BLS/d:	900,000	1,500,000	2,700,000
Equivalent ADTP/d:			
Bleached pulp	250	420	750
Unbleached pulp	300	500	900
Gas flow rate--APCD inlet, acfm:	9,400	15,700	28,300
Existing pressure drop, in. H2O:	0.7	0.7	0.7
New pressure drop, in. H20:	7	7	7
Operating days per year:			
Equipment	351	351	351
Labor	365	365	365
Operating hours per year:			
Equipment	8,424	8,424	8,424
Labor	8,760	8,760	8,760

TABLE 6-31. CAPITAL AND ANNUAL COSTS TO REPLACE MIST ELIMINATORS WITH SCRUBBERS FOR MODEL SDT'S[a]

Costs	Model SDT's		
	SDT-5	SDT-6	SDT-7
Total Capital Investment (TCI) (b):	$584,000	$796,000	$1,130,000
Direct Annual Costs (DAC):			
Operating labor (c):			
Operator	$37,200	$37,200	$37,200
Supervisor	$5,580	$5,580	$5,580
Maintenance (d):			
Labor	$27,400	$27,400	$27,400
Material	$27,400	$27,400	$27,400
Utilities (e):			
Electricity	$5,420	$9,050	$16,300
Water	$9,500	$15,900	$28,600
Wastewater treatment (f):	$0	$0	$0
Total DAC:	$113,000	$123,000	$142,000
Indirect Annual Costs (IAC):			
Overhead (g):	$32,900	$32,900	$32,900
Admin., taxes and insurance (h):	$23,400	$31,800	$45,200
Capital recovery (i):	$64,100	$87,400	$124,000
Total IAC:	$120,000	$152,000	$202,000
Total Annual Costs (TAC):	$233,000	$275,000	$344,000
Incremental Total Annual Costs (ITAC) (j):	$190,000	$232,000	$301,000

(a) All costs in $1991
(b) New scrubber TCI (models 5-7) = [scrubber replacement TCI (models 1-3)] x 2
(c) Operator labor = 6 hr/d x 365 d/yr x $17/hr
Supervisor labor = 15% of operator labor
(d) Maintenance labor = 3 hr/d x 365 d/yr x $25/hr
Maintenance materials = 100% of maintenance labor
(e) Electricity = 0.000181 x inlet gas flowrate x (7 - 0.7 in. H2O) x 8,424 hr/yr x $0.06/kWh
Water = 0.6 x inlet gas flowrate x 8,424 hr/yr x $0.20/1,000 gal
(f) Recycled to SDT
(g) Overhead = 0.6 x maintenance
(h) Administrative, taxes, and insurance = 0.04 x TCI
(i) Capital recovery = 0.1098 x TCI (based on 15-yr scrubber life and 7% interest)
(j) ITAC = TAC - operating labor costs.

TABLE 6-32a (METRIC). COSTS TO REPLACE SCRUBBERS WITH ESP'S TO CONTROL PM TO 0.15 G/DSCM FOR MODEL LIME KILNS[a]

Model lime kilns	Lime rate, Mg/d	Equivalent ADMP/d	Baseline APCD	Control option APCD	Gas flowrate--ESP exit, m3/sec @8% O2	ESP plate area, m2	SCA, m2/(m3/sec)	TCI, $	A,I & T, $/yr	Capital recovery, $/yr	DAC & overhead cost savings, $/yr	ITAC, $/yr
LK-1	90	320	scrubber	ESP	10	909	91	$457,000	$18,300	$50,200	($172,000)	($104,000)
LK-2	180	680	scrubber	ESP	20	1,818	91	$883,000	$35,300	$97,000	($215,000)	($82,700)
LK-3	270	1,000	scrubber	ESP	34	3,090	91	$1,500,000	$60,000	$165,000	($278,000)	($53,000)

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 6-32b. Refer to Table 6-32b for footnotes, which include calculations. Numbers in parentheses represent cost savings.

TABLE 6-32b (ENGLISH). COSTS TO REPLACE SCRUBBERS WITH ESP'S TO CONTROL PM TO 0.067 GR/DSCF FOR MODEL LIME KILNS[a]

Model lime kilns	Lime rate, ton/d	Equivalent ADTP/d	Baseline APCD	Control option APCD	Gas flowrate--ESP exit, acfm @8% O2	ESP plate area, ft2 (b)	SCA, ft2/1,000 acfm	TCI, $ (c)	A,I & T, $/yr (d)	Capital recovery, $/yr (e)	DAC & overhead cost savings, $/yr (f)	ITAC, $/yr (g)
LK-1	100	350	scrubber	ESP	22,000	10,157	462	$457,000	$18,300	$50,200	($172,000)	($104,000)
LK-2	200	750	scrubber	ESP	42,500	19,621	462	$883,000	$35,300	$97,000	($215,000)	($82,700)
LK-3	300	1,100	scrubber	ESP	72,200	33,332	462	$1,500,000	$60,000	$165,000	($278,000)	($53,000)

(a) All costs in $1991. Numbers in parentheses represent cost savings.
(b) ESP plate area = model gas flowrate x SCA/1,000 acfm
(c) TCI = ESP plate area x $45/ft2 plate area
(d) Administrative, insurance, and taxes (A,I & T) = 0.04 x TCI
(e) Capital recovery = 0.1098 CRF x TCI (based on 15-yr ESP life and 7% interest)
(f) Direct annual cost (DAC) and overhead cost savings = (ESP DAC + ESP overhead costs) - (scrubber DAC + scrubber overhead costs)
(g) Incremental total annual cost (ITAC) = DAC and overhead cost savings + A,I & T + capital recovery

TABLE 6-33a (METRIC). COSTS TO REPLACE SCRUBBERS WITH ESP'S TO CONTROL PM TO 0.023 G/DSCM FOR MODEL LIME KILNS[a]

Model lime kilns	Lime rate, Mg/d	Equivalent ADMP/d	Baseline APCD	Control option APCD	Gas flowrate--ESP exit, m3/sec @8% O2	ESP plate area, m2	SCA, m2/(m3/sec)	TCI, $	A,I & T, $/yr	Capital recovery, $/yr	DAC & overhead cost savings, $/yr	ITAC, $/yr
LK-1	90	320	scrubber	ESP	10	2,210	221	$1,110,000	$44,400	$122,000	($149,000)	$17,400
LK-2	180	680	scrubber	ESP	20	4,420	221	$2,150,000	$86,000	$236,000	($172,000)	$150,000
LK-3	270	1,000	scrubber	ESP	34	7,514	221	$3,650,000	$146,000	$401,000	($205,000)	$342,000

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 6-33b.

TABLE 6-33b (ENGLISH). COSTS TO REPLACE SCRUBBERS WITH ESP'S TO CONTROL PM TO 0.010 GR/DSCF FOR MODEL LIME KILNS[a]

Model lime kilns	Lime rate, ton/d	Equivalent ADTP/d	Baseline APCD	Control option APCD	Gas flowrate--ESP exit, acfm @8% O2	ESP plate area, ft2 (b)	SCA, ft2/1,000 acfm	TCI, $ (c)	A,I & T, $/yr (d)	Capital recovery, $/yr (e)	DAC & overhead cost savings, $/yr (f)	ITAC, $/yr (g)
LK-1	100	350	scrubber	ESP	22,000	24,699	1,123	$1,110,000	$44,400	$122,000	($149,000)	$17,400
LK-2	200	750	scrubber	ESP	42,500	47,714	1,123	$2,150,000	$86,000	$236,000	($172,000)	$150,000
LK-3	300	1,100	scrubber	ESP	72,200	81,058	1,123	$3,650,000	$146,000	$401,000	($205,000)	$342,000

(a) All costs in $1991
(b) ESP plate area = model gas flowrate x SCA/1,000 acfm
(c) TCI = ESP plate area x ($45/ft2 plate area)
(d) Administrative, insurance, and taxes (A,I & T) = 0.04 x TCI
(e) Capital recovery = 0.1098 CRF x TCI (based on 15-yr ESP life and 7% interest)
(f) Direct annual cost (DAC) and overhead cost savings = (ESP DAC + ESP overhead costs) - (scrubber DAC + scrubber overhead costs)
(g) Incremental total annual cost (ITAC) = DAC and overhead cost savings + A,I & T + capital recovery

TABLE 6-34a (METRIC). MODEL LIME KILN/SCRUBBER DESIGN PARAMETERS[a]

Design parameters	Model Lime Kilns		
	LK-1	LK-2	LK-3
Baseline APCD	scrubber	scrubber	scrubber
Lime production rate, Mg CaO/d	90	180	270
Equivalent ADMP/d	320	680	1,000
Gas flowrate--APCD inlet, m3/sec	10	20	34
Pressure drop, mm Hg	39	39	39
Operating days per year			
Equipment	351	351	351
Labor	365	365	365
Operating hours per year			
Equipment	8,424	8,424	8,424
Labor	8,760	8,760	8,760

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 6-34b.

TABLE 6-34b (ENGLISH). MODEL LIME KILN/SCRUBBER DESIGN PARAMETERS

Design parameters	Model Lime Kilns		
	LK-1	LK-2	LK-3
Baseline APCD	scrubber	scrubber	scrubber
Lime production rate, ton CaO/d	100	200	300
Equivalent ADTP/d	350	750	1,100
Gas flowrate--APCD inlet, acfm	22,000	42,500	72,200
Pressure drop, in. H20	21	21	21
Operating days per year			
Equipment	351	351	351
Labor	365	365	365
Operating hours per year			
Equipment	8,424	8,424	8,424
Labor	8,760	8,760	8,760

TABLE 6-35. ANNUAL COSTS FOR EXISTING LIME KILN SCRUBBERS[a]

Costs	Model Lime Kilns		
	LK-1	LK-2	LK-3
Direct Annual Costs (DAC)			
Operating labor (b)			
Operator	$37,200	$37,200	$37,200
Supervisor	$5,580	$5,580	$5,580
Maintenance (c)			
Labor	$27,400	$27,400	$27,400
Material	$27,400	$27,400	$27,400
Utilities (d)			
Electricity	$42,000	$81,200	$138,000
Water	$22,200	$43,000	$73,000
Wastewater treatment (e)	$0	$0	$0
Total DAC	$162,000	$222,000	$309,000
Indirect Annual Costs (IAC)			
Overhead (f)	$58,500	$58,500	$58,500
Total Annual Costs	$221,000	$281,000	$368,000

(a) All costs in $1991
(b) Operator labor = 6 hr/d x 365 d/yr x $17/hr
Supervisor labor = 15% of operator labor
(c) Maintenance labor = 3 hr/d x 365 d/yr x $25/hr
Maintenance materials = 100% of maintenance labor
(d) Electricity = 0.00018 x model inlet gas flowrate x pressure drop x 8,424 hr/yr x $0.06/kWh
Water = 0.06 x model inlet gas flowrate x 8,424 hr/yr x $0.20/1,000 gal
(e) Recycled to mud washer
(f) Overhead = 0.6 x (labor + maintenance)

TABLE 6-36a (METRIC). MODEL LIME KILN/ESP DESIGN PARAMETERS[a]

Design parameters	Model Lime Kilns		
	LK-1	LK-2	LK-3
Control option APCD	ESP	ESP	ESP
Lime production rate, Mg CaO/d	90	180	270
Equivalent ADMP/d	320	680	1,000
Gas flowrate--ESP exit, m3/sec	10	20	34
ESP plate area, m2			
PM controls--0.15 g/dscm	909	1,818	3,090
PM controls--0.023 g/dscm	2,210	4,420	7,514
SCA, m2/(m3/sec)			
PM controls--0.15 g/dscm	91	91	91
PM controls--0.023 g/dscm	221	221	221
Pressure drop, mm Hg	2	2	2
Operating days per year			
Equipment	351	351	351
Labor	365	365	365
Operating hours per year			
Equipment	8,424	8,424	8,424
Labor	8,760	8,760	8,760

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 6-36b.

TABLE 6-36b (ENGLISH). MODEL LIME KILN/ESP DESIGN PARAMETERS

Design parameters	Model Lime Kilns		
	LK-1	LK-2	LK-3
Control option APCD	ESP	ESP	ESP
Lime production rate, ton CaO/d	100	200	300
Equivalent ADTP/d	350	750	1,100
Gas flowrate--ESP exit, acfm	22,000	42,500	72,200
ESP plate area (A), ft2			
PM controls--0.067 gr/dscf	10,157	19,621	33,332
PM controls--0.01 gr/dscf	24,699	47,714	81,058
SCA, ft2/1,000 acfm			
PM controls--0.067 gr/dscf	462	462	462
PM controls--0.01 gr/dscf	1,123	1,123	1,123
Pressure drop, in. H20	1	1	1
Operating days per year			
Equipment	351	351	351
Labor	365	365	365
Operating hours per year			
Equipment	8,424	8,424	8,424
Labor	8,760	8,760	8,760

TABLE 6-37. ANNUAL COSTS FOR NEW LIME KILN ESP'S CONTROLLING PM TO 0.15 G/DSCM (0.067 GR/DSCF)[a]

Costs	Model Lime Kilns		
	LK-1	LK-2	LK-3
Total Capital Investment (TCI) (b):	$457,000	$883,000	$1,500,000
Direct Annual Costs (DAC):			
Operating labor (c):			
Operator	$13,100	$13,100	$13,100
Supervisor	$1,970	$1,970	$1,970
Coordinator	$4,320	$4,320	$4,320
Maintenance (d):			
Labor	$838	$1,620	$2,750
Material	$2,740	$5,300	$9,000
Electricity (e):			
Electricity-fan	$2,000	$3,870	$6,570
Electricity-operating	$10,000	$19,200	$32,700
Waste disposal:	$0	$0	$0
Total DAC:	$35,000	$49,400	$70,400
Indirect Annual Costs (IAC):			
Overhead (f):	$13,800	$15,800	$18,700
Admin., taxes and insurance (g):	$18,300	$35,300	$60,000
Capital recovery (h):	$50,200	$97,000	$165,000
Total IAC:	$82,300	$148,000	$244,000
Total Annual Costs:	$117,000	$197,000	$314,000

(a) All costs in $1991
(b) TCI = ESP plate area x $45/ft2 plate area
(c) Operator labor = 3 hr/day x 365 d/yr x $12/hr
Supervisor labor = 15% of operator labor
Coordinator labor = 33% of operator labor
(d) Maintenance labor = 0.0825 x ESP plate area
Maintenance material = 0.01 x (0.6 x TCI)
(e) Electricity-fan = 0.00018 x model gas flowrate x pressure drop x 8,424 hr/yr x $0.06/kWh
Electricity-operating = 0.00194 x ESP plate area x 8,424 hr/yr x $0.06/kWh
(f) Overhead = 0.6 x (labor + maintenance)
(g) Administrative, taxes, and insurance = 0.04 x TCI
(h) Capital recovery = 0.1098 x TCI (based on 15-yr ESP life and 7% interest)

TABLE 6-38. ANNUAL COSTS FOR NEW LIME KILN ESP'S CONTROLLING PM TO 0.023 G/DSCM (0.010 GR/DSCF)[a]

Costs	Model Lime Kilns		
	LK-1	LK-2	LK-3
Total Capital Investment (TCI) (b):	$1,110,000	$2,150,000	$3,650,000
Direct Annual Costs (DAC):			
Operating labor (c):			
Operator	$13,100	$13,100	$13,100
Supervisor	$1,970	$1,970	$1,970
Coordinator	$4,320	$4,320	$4,320
Maintenance (d):			
Labor	$2,040	$3,940	$6,690
Material	$6,660	$12,900	$21,900
Electricity (e):			
Electricity-fan	$2,000	$3,870	$6,570
Electricity-operating	$24,200	$46,800	$79,500
Waste disposal:	$0	$0	$0
Total DAC:	$54,300	$86,900	$134,000
Indirect Annual Costs (IAC):			
Overhead (f):	$16,900	$21,700	$28,800
Admin., taxes and insurance (g):	$44,400	$86,000	$146,000
Capital recovery (h):	$122,000	$236,000	$401,000
Total IAC:	$183,000	$344,000	$576,000
Total Annual Costs:	$237,000	$431,000	$710,000

(a) All costs in $1991
(b) TCI = ESP plate area x $45/ft2 plate area
(c) Operator labor = 3 hr/day x 365 d/yr x $12/hr
Supervisor labor = 15% of operator labor
Coordinator labor = 33% of operator labor
(d) Maintenance labor = 0.0825 x ESP plate area
Maintenance material = 0.01 x (0.6 x TCI)
(e) Electricity-fan = 0.00018 x model gas flowrate x pressure drop x 8,424 hr/yr x $0.06/kWh
Electricity-operating = 0.00194 x ESP plate area x 8,424 hr/yr x $0.06/kWh
(f) Overhead = 0.6 x (labor + maintenance)
(g) Administrative, taxes, and insurance = 0.04 x TCI
(h) Capital recovery = 0.1098 x TCI (based on 15-yr ESP life and 7% interest)

TABLE 6-39a (METRIC). COSTS TO UPGRADE ESP'S TO CONTROL PM TO 0.023 G/DSCM FOR MODEL LIME KILNS[a]

Model lime kilns	Lime prod., Mg/d	Equivalent ADMP/d	Gas flowrate-- ESP exit, m3/sec @8% O2	Increase in ESP plate area, m2	Increase in SCA, m2/(m3/sec)	TCI, $	Electricity, $/yr	A,I & T, $/yr	Capital recovery, $/yr	ITAC, $/yr
LK-4	90	320	10	1,301	130	$654,000	$14,300	$26,200	$71,800	$112,000
LK-5	180	680	20	2,602	130	$1,260,000	$27,500	$50,400	$138,000	$216,000
LK-6	270	1,000	34	4,424	130	$2,150,000	$46,800	$86,000	$236,000	$369,000

(a) Metric equivalents in this table were converted from the calculated English unit values given in Table 6-39b.

TABLE 6-39b (ENGLISH). COSTS TO UPGRADE ESP'S TO CONTROL PM TO 0.010 GR/DSCF FOR MODEL LIME KILNS[a]

Model lime kilns	Lime prod., ton/d	Equivalent ADTP/d	Gas flowrate--ESP exit, acfm @8% O2	Increase in ESP plate area, ft2 (b)	Increase in SCA, ft2/1000 acfm	TCI, $ (c)	Electricity, $/yr (d)	A,I & T, $/yr (e)	Capital recovery, $/yr (f)	ITAC, $/yr (g)
LK-4	100	350	22,000	14,543	661	$654,000	$14,300	$26,200	$71,800	$112,000
LK-5	200	750	42,500	28,094	661	$1,260,000	$27,500	$50,400	$138,000	$216,000
LK-6	300	1,100	72,200	47,726	661	$2,150,000	$46,800	$86,000	$236,000	$369,000

(a) All costs in $1991
(b) Increase in ESP plate area = flow rate x (1,123 - 462)ft2/1,000 acfm
(c) TCI = Increase in ESP plate area x ($45/ft2 plate area)
(d) Electricity = (0.00194 x flow rate x [1,123 - 462]ft2/1,000 acfm SCA) x 8,424 hr/yr x $0.06/kWh
(e) Administrative, insurance, and taxes (A,I & T) = 0.04 x TCI
(f) Capital recovery = 0.1098 CRF x TCI (based on 15-yr ESP life and 7% interest)
(g) Incremental total annual cost (ITAC) = Electricity + A,I & T + capital recovery

TABLE 6-40. SUMMARY OF ENHANCED MONITORING COSTS

Enhanced monitoring system	Cost
Opacity monitor	Capital cost: $34,800 Annual cost: $21,200/yr
Methanol CEM (FTIR)	Capital cost: $160,000 Annual cost: $71,500/yr
HCl CEM	Capital cost: $126,900 Annual cost: $77,400/yr
EPA Method 5 PM compliance test (three runs)	Initial compliance test cost: $8,500 Semiannual testing annual cost: $17,000/yr
EPA Method 29 PM compliance test (three runs)	Initial compliance test cost: $12,000 Seminnual testing annual cost: $24,000/yr
EPA Method 26 or 26A HCl compliance test (three runs)[a]	Initial compliance test cost: $9,100[a] $600[b] Annual testing annual cost: $9,100/yr[a] $600[b]
Monitoring plan	Cost not available
Confirmation of NDCE recovery furnace with dry ESP system	$0
Confirmation of BLO control equipment	$0
Temperature monitor for power boiler or other incineration device[c]	$0
pH monitoring system[d]	$5,000
Flow rate measurement system[d,e]	$2,000 - $25,000
Magnehelic gauge, S-type pitot tube (pressure drop measurement)[d]	Manual read-out system: $300 Digital read-out system: Cost not available

[a]Cost assumes EPA Method 26 or 26A HCl testing is performed alone.
[b]Cost assumes EPA Method 26 or 26A HCl testing is performed in conjunction with EPA Method 5, Method 29, or Method 17 testing.
[c]Cost of monitoring power boiler or incinerator temperature is assumed to be zero because temperature monitoring is already conducted by mills.
[d]Annual costs were not available for monitoring pH and flow rate of scrubber liquid or for monitoring scrubber pressure drop.
[e]Cost depends on the device sensitivity and distance from the control room.

TABLE 6-41. OPACITY AND HCL CONTINUOUS EMISSION MONITOR COSTS[a]

Parameters	CEM costs	
	Opacity CEM	HCl CEM
Capital costs		
Planning	$1,200	$4,600
Select type of equipment	$4,000	$10,500
Provide support facilities	$100	$13,100
Purchased equipment cost	$21,500	$61,500
Install and check CEM's	$300	$11,100
Performance specification tests (certification)	$1,400	$14,300
Prepare QA/QC plan	$6,300	$11,800
Total capital investment	$34,800	$126,900
Annual costs		
Operating and maintenance	$7,000	$8,500
Annual RATA[b]	$0	$10,000
Supplemental RATA	$0	$9,500
Quarterly CGA's[c]	$0	$3,500
Reporting and recordkeeping	$5,900	$11,600
Annual review and update	$3,600	$17,200
Administrative, insurance, and property taxes[d]	$3,300	$12,000
Capital recovery[e]	$1,400	$5,100
Total annual cost	$21,200	$77,400

[a]All costs are based on EMTIC's CEM program, except for administrative, insurance, property tax, and capital recovery costs, which are based on procedures from the OAQPS Control Cost Manual.
[b]Relative accuracy test audit
[c]Cylinder gas audits
[d]Administrative, taxes, and insurance = 0.04 x TCI
[e]Capital recovery = 0.0944 x TCI (based on 20-yr equipment life and 7% interest)

6.3 REFERENCES FOR CHAPTER 6

1. Memorandum from Holloway, T. and Nicholson, R., MRI, to Telander, J., EPA/MICG. October 9, 1996. Nationwide Costs, Environmental Impacts, and Cost-Effectivness of Regulatory Alternatives for Kraft, Soda, Sulfite, and Semichemical Combustion Sources.

2. Economic Indicators--CE Plant Cost Index. Chemical Engineering (New York). 1960-1993.

3. Memorandum from Mathias, S., EPA/CEIS, to Telander, J., EPA/ISB. February 2, 1994. Value of Lost Production days for Kraft Pulp Mills.

4. Memorandum from Soltis, V., MRI, to the project file. April 3, 1995. Kraft and Soda Pulp Mill Combustion Sources Data Base.

5. Handbook: Control Technologies for Hazardous Air Pollutants. U. S. Environmental Protection Agency. Cincinnati, OH. Publication No. EPA-625/6-91-014. June 1991. Chapter 4.

6. OAQPS Control Cost Manual. 4th Edition. U. S. Environmental Protection Agency. Research Triangle Park, NC. Publication No. EPA-450/3-90-006. January 1990.

7. U. S. Environmental Protection Agency. National Emission Standards for Hazardous Air Pollutants: General Provisions. 59 FR 12408 et seq. Washington, DC. U.S. Government Printing Office. March 16, 1994.

8. Memorandum from Soltis, V., Nicholson, R., Holloway, T., MRI, to Telander, J., EPA/MICG. July 29, 1994. Summary of Responses to the NCASI "MACT" Survey--Kraft and Soda Pulp Mills (Data Base Summary Memo).

9. Memorandum from Nicholson, R., MRI, to Telander, J., EPA/ISB. May 28, 1993. Meeting minutes from the March 30, 1993 meeting between Environmental Elements Corp. and Midwest Research Institute.

10. Memorandum from Holloway, T., MRI, to the project file. July 16, 1996. State of Washington PM Data for Kraft Recovery Furnaces, Smelt Dissolving Tanks, and Lime Kilns.

11. Memorandum from Radian Corp. to EPA/ISB. September 18, 1985. Meeting minutes from the June 10, 1985 meeting between kraft pulp mill representatives, the National Council of the Paper Industry for Air and Stream Improvement, Inc., the U. S. Environmental Protection Agency, and Radian Corp.

12. Letter and attachments from Eddinger, J., EPA/ISB, to Blosser, R., National Council of the Paper Industry for Air and Stream Improvement. May 22, 1985. Preliminary cost estimates for evaluating the three control options to bring wet-bottom ESP's into compliance with the NSPS.

13. Letter from Bringman, L., Environmental Elements Corp., to Nicholson, R., MRI. October 12, 1993. Budgetary pricing to convert from wet- to dry-bottom ESP.

14. Letter from Burt, R., Radian Corp., to Holbrook, J. and Bringman, L., Environmental Elements Corp. July 23, 1985. Facsimile cover letter discussing utility and maintenance costs for wet- vs. dry-bottom ESP designs.

15. Garner, J., Jaako Pöyry. Conversion to Low Odor Improves Recovery Boiler Efficiency and Life. Pulp and Paper. 63(7):91-95. July 1989.

16. Green, R. and G. Hough (eds.). Chemical Recovery in the Alkaline Pulping Process. 3rd Edition. Prepared by the Alkaline Pulping Committee of the Pulp Manufacture Division. Atlanta, GA, TAPPI Press, 1992. 196 p.

17. Memorandum from Ramsey, M. and Nicholson, R., MRI, to Telander, J., EPA/ISB. November 3, 1992. Meeting minutes from the June 24, 1992 meeting between The Babcock and Wilcox Co., the U. S. Environmental Protection Agency, and Midwest Research Institute.

18. Telecon. Soltis, V., MRI, with Morgan, P., ABB Combustion Engineering Services, Inc. August 6, 1993. Information about cascade evaporators and low-odor conversion costs.

19. Telecon. Soltis, V., MRI, with Blake, J., Goslin-Birmingham. August 27, 1993. Concentrator costs.

20. Williamson, F., Chief Power Engineer, Harmac Pulp Division. Low Odor Boiler Conversion--By Threes. In: TAPPI Kraft Recovery Operations Short Course. TAPPI Press. January 3-8, 1993. pp. 229-247.

21. Hein, A. Kraft Recovery Boiler: Conventional to Low-Odor Conversion. Pulp & Paper Canada. 88(12):223-226. December 1987.

22. Kirby, M. Economic and Process Considerations in the Use of Oxygen for Black Liquor Oxidation. Union Carbide Canada Limited. Ontario, Canada. (Presented at the 21st Pulp and Paper ABCP Annual Meeting. Sao Paolo, Brazil. November 21-25, 1988.) 10 p.

23. Perry, R., D. Green, and J. Maloney (eds.). Perry's Chemical Engineer's Handbook. 6th edition. 1984. pp. 9-16, 9-18.

24. Pulp, Paper, and Paperboard Industry--Background Information for Proposed Air Emission Standards, Manufacturing Processes at Kraft, Sulfite, Soda, and Semi-Chemical Mills. U. S. Environmental Protection Agency. Research Triangle Park, NC. Publication No. EPA-453/R-93-050a. October 1993.

25. National Petroleum News 1991 Fact Book. Volume 83, No. 7. June 1991. p. 114.

26. Letter and attachments from Black, C., Champion International Corp., Courtland, AL, to Crowder, J., EPA/ISB. June 10, 1993. Response to request for information on costs for the BLO vent gas control system at the Champion facility in Courtland, AL.

27. Telecon. Soltis, V., MRI, with Oscarsson, B., Gõtaverken Energy Systems, Inc. December 16, 1993. Concentrator steam costs.

28. Memorandum from Rovansek, W., Radian Corp., Herndon, VA, to the record. September 1, 1993. Pulp and Paper Operating Costs.

29. Letter and attachments from Sanders, D., Andersen 2000, Inc., to Soltis, V., MRI. November 9, 1993. Preliminary proposal for three separate quench and packed-tower absorption systems to control HCl gas emissions from recovery furnaces at kraft pulp and paper mills.

30. Memorandum from Randall, D. and Holloway, T., MRI, to the project file. October 4, 1994. Absorber Design and Cost Algorithm.

31. Medical Waste Incinerators--Background Information for Proposed Standards and Guidelines: Model Plant Description and Cost Report for New and Existing Facilities. U. S. Environmental Protection Agency. Research Triangle Park, NC. Publication No. EPA-453/R-94-044a. July 1994. p. 62.

32. Someshwar, A. and J. Pinkerton. Wood Processing Industry. In: Air Pollution Engineering Manual, Air and Waste Management Association. Buonicore, A. and W. Davis (eds.). New York, Van Norstrand Reinhold. 1992. pp. 835-849.

33. RUST Environment & Infrastructure. MACT Cost Analysis Report for NCASI. Prepared for the National Council of the Paper Industry for Air and Stream Improvement, Inc. RUST Contract No. 35-4735. October 15, 1993. 47 p.

34. Memorandum from Nicholson, R. and Holloway, T., MRI, to Telander, J., EPA/MICG. September 6, 1995. Trip Report for S.D. Warren Co., Muskegon, MI.

35. Telecon. Soltis, V., MRI, with Bruno, J., Airpol, Inc. April 22 and 27, 1993. Information on scrubbers used to control particulate matter emissions from lime kilns and smelt dissolving tanks at pulp and paper mills.

36. Memorandum from Nicholson, R., MRI, to Telander, J., EPA/MICG. June 15, 1995. Meeting minutes from the April 12, 1993 meeting between Research-Cottrell and Midwest Research Institute.

37. Memorandum from Toney, M., EPA/EMB, to Telander, J., EPA/ISB. November 24, 1993. Enhanced Monitoring for Pulp and Paper Combustion Sources.

38. Memorandum from Segall, R. EPA/EMB, to Copland, R., EPA/SDB. October 14, 1992. Final Report on Emission Measurement and Continuous Monitoring for Medical Waste Incinerator Study.

39. Memorandum from Toney, M., EPA/EMC, to Telander, J., EPA/MICG. January 13, 1995. Costs of Enhanced Monitoring for Pulp and Paper Combustion Sources.

40. Telecon. Harris, V., MRI-NCO, with Hosenfeld, J., MRI-KCO. May 26, 1995. Costs of EPA Method 26 stack tests.

APPENDIX A.
EVOLUTION OF THE TECHNICAL SUPPORT DOCUMENT

The purpose of this study was to develop a basis for supporting proposed NESHAP for combustion sources at kraft and soda pulp and mills. This study was initiated in June of 1991 and focused on the following combustion equipment: chemical recovery furnaces (including DCE's), SDT's, BLO systems, and lime kilns. To accomplish the objectives of this study, the following technical data were acquired for each type of combustion equipment: (1) equipment design and operating parameters, (2) types and quantities of HAP's emitted, (3) types and costs of emission control techniques (including both process changes/controls and add-on air pollution control devices), and (4) the effectiveness of these control techniques in reducing HAP emissions. The bulk of the information was gathered from the following sources:

1. Technical literature;
2. Responses to a maximum achievable control technology (MACT) survey distributed by NCASI to U.S. pulp and paper mills;
3. Plant visits;
4. Emission test reports supplied by individual mills, State agencies, and NCASI;
5. Recovery furnace manufacturers; and
6. Manufacturers of ESP's and wet scrubbers.

Significant events relating to the evolution of the TSD are listed in Table A-1.

TABLE A-1. EVOLUTION OF THE TECHNICAL SUPPORT DOCUMENT

Date	Event
7/1/91-7/31/91	Contacted State and local agencies to identify well-controlled combustion sources
8/1/91	Initial meeting with NCASI and AFPA regarding combustion sources NESHAP and "MACT" survey
9/17/91	Site visit to Champion International Corp., Roanoke Rapids, NC
9/23/91	Site visit to Procter and Gamble, Perry, FL
9/24/91	Meeting with Florida Department of Environmental Regulation to discuss kraft pulp and paper mills in Florida and obtain copies of permits
9/25/91	Site visit to Packaging Corp., Clyattville, GA
10/28/91	Meeting with NCASI and AFPA to discuss industry profile
1/10/92	Site visit to Longview Fibre Co., Longview, WA
1/14/92	Site visit to James River Corp., Camas, WA
2/7/92	NCASI MACT survey mailed to industry by NCASI
2/24/92	Meeting with Gõtaverken to discuss design and operation of kraft recovery furnaces
4/9/92	Public meeting held in Washington, DC to discuss status of NESHAP and effluent guidelines for pulp and paper mills
4/30/92	Initial responses to NCASI MACT survey received by EPA, and kraft combustion sources data base created to tabulate information in these responses
5/30/92	Additional responses to NCASI MACT survey received by EPA
6/24/92	Meeting with Babcock & Wilcox to discuss design and operation of kraft recovery furnaces
7/2/92	Draft summary of responses to NCASI MACT survey prepared
9/22/92	Site visit to MEAD Coated Board, Phenix City, AL
9/23/92	Site visit to Champion International Corp., Courtland, AL
9/30/92	Public meeting in Washington, DC, to discuss status of NESHAP and effluent guidelines for pulp and paper mills
10/8/92-10/31/92	Source test at Champion International Corp., Roanoke Rapids, NC, to identify HAP's emitted from a recovery furnace, smelt dissolving tank, black liquor oxidation system, and lime kiln
11/20/92	Meeting with NCASI and industry representatives to discuss NCASI's source test plan for combustion sources
3/30/93	Meeting with Environmental Elements Corp., Baltimore, MD, to discuss ESP designs, modifications, and upgrades, and associated costs
4/12/93	Meeting with Research-Cottrell, Somerville, NJ, to discuss ESP designs, modifications, and upgrades, and associated costs
5/24/93	First EPA Work Group meeting to discuss status and direction of combustion sources NESHAP
6/14/93	Monthly particulate matter emission data for kraft recovery furnaces, lime kilns, and SDT's received from the State of Washington (1989-1992)
6/15/93	Emission test reports for Texas kraft mills received from NCASI
8/16/93	Letters sent to Andersen 2000, Inc., and AirPol regarding the effectiveness of packed-bed scrubbers on HCl removal from kraft recovery furnace flue gas
9/13/93	Completed review and summary of State regulations regarding PM and gaseous HAP emissions from kraft combustion sources
10/12/93	Meeting with NCASI and industry to discuss methanol emissions and effectiveness of steam strippers

TABLE A-1. (continued)

Date	Event
10/31/93	Review of first draft emission test report from NCASI-sponsored testing at kraft pulp completed
12/2/93	Mill-specific control costs for MACT floor options submitted to EPA for impact analysis
12/16/93	Meeting with NCASI and industry to discuss status of NCASI source test program, subcategories, MACT floors, and proposed format of the standard
1/1/94	Preliminary drafts of TSD Chapter 1 through 6 completed
2/3/94	Meeting with NCASI and industry to discuss status and findings of NCASI's test program
2/17/94	Preliminary draft of TSD Chapter 6 (Control Costs) sent to NCASI for review
3/11/94	Mill-specific control costs for beyond-the-floor options submitted to EPA for impact analysis
3/21/94	Industry comments on preliminary draft of TSD Chapter 6 received
4/11/94	Response to industry comments on preliminary draft of TSD Chapter 6 completed
4/27/94	Preliminary draft of Appendix B completed
5/12/94	An updated version of monthly particulate matter emission data for kraft recovery furnaces, lime kilns, and SDT's received from the State of Washington (1988-1994)
7/13/94	Received and reviewed report from industry summarizing results from an industry survey of recovery furnace manufacturers regarding time needed for recovery furnace conversions that included a list of all kraft and soda recovery furnaces valid through December 1994
7/29/94	All HAP emissions data from NCASI's test program received to date compiled into emissions data base for the project
7/29/94	Preliminary regulatory alternatives developed based on the subcategorization of recovery furnaces for the source category
8/24/94	Combustion sources data base revised to account for mill-wide and recovery furnace shutdowns and furnaces installed since the original MACT survey was completed based on an industry inventory of kraft and soda recovery furnaces
8/31/94	Emissions data received from International Paper Co. compiled and added to emissions data base for the project
9/1/94	Meeting with industry representatives to discuss the MACT standards under consideration and the vendor capacity and resource requirements to rebuild/replace recovery furnaces.
9/14/94	Site visit to Federal Paper Board, Augusta, GA
9/21/94	Site visit to S.D. Warren Co., Muskegon, MI
9/30/94	Received and reviewed HCl study (Technical Bulletin No. 674) conducted by NCASI
10/4/94	Additional mill-specific control costs for MACT floor and beyond-the-floor options submitted to EPA for impact analysis
11/30/94	Second Work Group meeting to provide technical update
12/20/94	Third Work Group meeting to discuss the preliminary regulatory alternatives, the impacts and economic analyses of those alternatives, and the EPA recommended regulatory alternative; examined use of bubble approach to reduce the control costs for particulate control under the EPA recommended regulatory alternative
1/18/95	Teleconference meeting with OECA on enhanced monitoring
1/31/95	Teleconference meeting with States on issues associated with the bubble approach
3/7/95	Teleconference meeting with OECA on enhanced monitoring protocols
4/20/95	Teleconference meeting with OECA on enhanced monitoring

TABLE A-1. (continued)

Date	Event
5/22/95	Meeting with industry representatives regarding control options for the Pulp and Paper NESHAP and the definition of a "clean mill"
8/2/95	Meeting with representatives of the Institute of Clean Air Companies to discuss the status of the NESHAP
2/29/96	Submitted revised mill-specific control costs for new regulatory alternatives to EPA for impact analysis
2/29/96	Submitted inputs for HCl risk analysis to EPA to determine applicability of Section 112(d)(4) of the CAA to HCl controls for pulp and paper combustion sources
4/12/96	Nationwide environmental impacts inputs submitted to EPA for benefits analysis
6/17/96	Meeting with industry representatives to discuss applicability threshold and PM HAP emisison limits for new sources
7/8/96	Teleconference meeting with OECA on enhanced monitoring
7/26/96	Draft TSD submitted to EPA and industry representatives for review
7/31/96	Teleconference meeting with industry representatives regarding enhanced monitoring approaches
9/11/96	Fourth Work Group meeting to provide update on status of the NESHAP
9/12/96	NCASI comments on the TSD received

APPENDIX B.

EMISSION MEASUREMENT AND CONTINUOUS MONITORING FOR PULP AND PAPER COMBUSTION SOURCES

B.1 INTRODUCTION

Between July 1991 and November 1994, EPA gathered emissions data from a variety of information sources in support of a NESHAP for combustion sources in the pulp and paper industry. Sources of information included:

1. An emissions test program sponsored by NCASI;
2. An emissions test program for the Texas Paper Industry Environmental Committee (TPIEC);
3. An emissions test program sponsored by the International Paper Company; and
4. Selected emissions test reports in state agency compliance files.

The industry-sponsored test programs were specifically designed to provide data to: (1) determine the mass emission rates of HAP's from recovery furnaces, BLO units, SDT's, and lime kilns; and (2) determine control efficiencies of APCD's for HAP's.

Emissions testing was performed primarily using manual methods. Locations sampled included the inlet and outlet of the control device(s) where pollution controls were applied, and the effluent stack where they were not applied.

This appendix defines the methods used in the test programs to collect emissions data, discusses alternatives for monitoring emissions to indicate continuous compliance with any proposed

standards, and recommends procedures to demonstrate compliance with proposed standards.

B.2 MEASUREMENT METHODS USED IN DATA COLLECTION

This section presents a summary of measurement methods used to generate emissions data in approximately 29 source test reports. The data in these reports were used to support the development of emissions standards for combustion sources in the pulp and paper industry.

B.2.1 Scope of Test Programs

The types of samples and data collected at each test site, the measurement locations, and test methods used are summarized in Tables B-1 through B-8.[1-29] Tables B-2 through B-8 summarize information and data collected as part of International Paper Company's nine-mill emissions test program. Table B-1 summarizes information contained in the remaining 20 emission test reports. Sample types and sampling locations are discussed in Section 2.1. The specific test methods used for collecting and analyzing the samples are discussed in Section 2.2.

B.2.1.1 Types of Samples and Data Collected. The samples collected during the test programs were primarily flue gas samples before or after emission control devices. Because a variety of source emissions test reports were used, a disparity exists in the number and type of pollutants measured at each source.

The source tests contained data for the following pollutants:

1. Metals, including antimony (Sb), arsenic (As), beryllium (Be), cadmium (Cd), cobalt (Co), chromium (Cr) (total), lead (Pb), manganese (Mn), mercury (Hg), nickel (Ni), selenium (Se), thallium (Tl);
2. SO_2;
3. H_2S, methyl mercaptan, dimethyl sulfide, carbon disulfide, sulfuric acid;
4. Terpenes, a-terpinol, a,b-pyrene, 3-curene, p-cyrene;
5. HCl;
6. CO;

7. NO_x;

8. Polychlorinated dibenzo-p-dioxins and dibenzofurans (PCDD/PCDF);

9. Selected organic HAP's, such as acrolein, acetaldehyde, acetophenone, acetone, methanol, carbon tetrachloride, chloroform, chlorobenzene, chloromethane, bromomethanol, isobutyl ketone, methyl ethyl ketone, methylene chloride, 1,2-dichloromethane, trichlorofluoromethane, bromodichloromethane, formaldehyde, 1,1,1-dichloroethylene, 1,1,2-trichloroethane, 1,2,4-trichloroethane, tetrachloroethylene, styrene, toluene, xylenes, benzene; and

10. Semivolatile organics, such as acenaphthene, pyrene, benzo-a-anthracene, chrysene, benzo-b-fluoroanthene, benzo-a-pyrene, dibenzo-a,h-anthracene, benzo-g,h,i-pyrene, indeno-1,2,3-pyrene.

B.2.1.2 Emission and Process Sampling Locations. Flue gas stream samples consisted of uncontrolled gas streams from BLO tanks, SDT's, and recovery furnaces, and controlled streams from ESP outlets and scrubbers following lime kilns and recovery furnaces. Additionally, one test report contained analysis of a quench water process stream. This process stream was analyzed for PCDD/PCDF and metals.

B.2.2 Sampling and Analytical Methods Used

Sampling and analysis methods used at each mill are given in Tables B-1 and B-2. The discussion in this section groups the sampling and analytical methods into three categories: (1) organic HAP methods, (2) metal HAP methods, and (3) HCl methods.

B.2.2.1 Organic HAP Methods. The methods used to gather organic HAP emissions data included EPA Methods 18 and 23 of 40 CFR 60 Appendix A; Methods 0010, 0011, and 0030 of SW-846; and two methods developed by NCASI. One of the methods developed by NCASI and validated by EPA Method 301 made use of a heated summa canister source sampling train (HSCSST) for collection of the gas sample for later laboratory analysis by gas chromatography (GC) and flame ionization detector (FID) or mass selective detector

(MSD).[1,2,30,31] This method is capable of measuring a variety of organic HAP's. The HSCSST used a Summa® canister, heated to 130°C, for collecting gaseous samples. The system employed for the analysis of the Summa® canisters was consistent with the requirements of EPA Method TO14, except that the system bypasses the Nafion® permeation dryer.

The other method developed by NCASI and validated by EPA Method 301 made use of a water-filled impinger followed by silica gel traps for capturing methanol and other water-soluble organics such as acetaldehyde, acetone, and methyl ethyl ketone.[1,32] The collected samples are later analyzed by GC/FID. The method is very similar to EPA's Method 308 for methanol, which is described in Section 4.4.

Differences among the methods include the ability to provide information on multiple pollutants and detection limits. Methods 18, 0010, 0030, and the HSCSST provided information on a greater number of pollutants. Methods 23 (dioxins), 0011 (aldehydes), and 308 (methanol) provided information on a very limited number of pollutants. Detection limits for most of the methods were generally acceptable. However, some of the data gathered using EPA Method 18 did not provide adequate detection limits because the GC set-up variables were not optimized.

B.2.2.2 Metal HAP Methods. At most plants, metal HAP compounds were collected and analyzed using EPA draft Method 29 (also known as draft Method 0012 of SW-846, the EPA combined metals train, or the EPA multimetals train). Differences were noted in some of the older test reports with respect to blank correction procedures and in the treatment of nondetect data. These differences were reconciled before the data were used further.[33,34] The CARB Method 425 was used at six plants for determining hexavalent Cr emissions.[16,17,29] At one plant, EPA Method 12 and a modified Method 5 train were used to determine Ni and Mn, and hexavalent Cr emissions, respectively.[10] At this plant, a stainless steel probe was used instead of a glass probe. For this reason, high levels of Cr and Ni were present in the

blanks. The Cr and Ni emissions data at this mill were removed from the NESHAP data pool.

B.2.2.3 Hydrogen Chloride Methods. Hydrogen chloride emissions were determined using EPA Methods 26 and 26A in 40 CFR Part 60 Appendix A. The two methods are very similar, except that in Method 26A the gas sample is collected isokinetically, whereas in Method 26 a nonisokinetic procedure is used. Method 26A is preferred at locations downstream of scrubbers where acid-gas mists are present.

B.3 MONITORING SYSTEMS AND DEVICES

This section presents the various types of continuous monitoring systems and monitoring devices, including data acquisition and data processing systems, readily available on an "off-the-shelf" basis that could be used to monitor emissions. The applicability of these systems to the affected facilities or similar facilities is discussed. This discussion covers such factors as the accuracy, precision, repeatability, reliability, complexity, maintenance, the difficulty of installation and operation of these systems, and costs (capital and annualized). Cost estimates include the costs of setting up and operating the monitoring system, but not the cost of reporting any data collected. Performance specifications are provided, also, if available.

B.3.1 Opacity Monitoring

Opacity monitoring equipment has been routinely used to monitor the opacity of emissions from combustion sources for many years. These monitors are comparatively simple and are easy to install. They are usually installed in the breeching duct or in the discharge stack downstream of any particulate control device. Choice of an appropriate measurement location depends upon source-specific factors, such as effluent ductwork configuration, presence of vibration, ease of access to the location, etc. A detailed discussion of the accuracy, precision, repeatability, reliability, and maintenance requirements for these instruments will not be provided because this monitoring technique is well established.

Estimated initial costs (purchase, installation, performance testing, quality assurance program plan preparation) and annual costs (operation & maintenance, recordkeeping and reporting, program update, quality assurance) for an opacity monitor installed at a stack location are $39,600 and $18,800, respectively (all dollar amounts are in 1994 dollars).[35]

Performance specifications for opacity monitors are contained in Performance Specification 1, Appendix B, 40 CFR Part 60. These requirements were first promulgated on October 6, 1975; substantial revisions were issued on March 30, 1983.

B.3.2 Hydrogen Chloride Monitoring

Emissions monitoring equipment for measuring gaseous HCl is commercially available; monitoring techniques include gas filter correlation (GFC) infrared analyzers, FTIR, ion mobility spectroscopy (IMS), ultraviolet (UV) spectroscopy, on-line specific ion electrodes (SIE), and mass spectrometric techniques. The mass spectrometric technique is not included in the discussion that follows because of its high cost. Detection limits for these instruments range from the low part per billion (ppb) range (e.g., IMS) to the low ppm range (e.g., FTIR). The FTIR and SIE systems are fairly complex systems, whereas the GFC and IMS are not. All of these systems are comparatively new, and, therefore, source-specific installation instructions are not well developed. All systems are prone to difficulties if condensed water is present in the sample delivery lines. Sample lines and other sample delivery components must be relatively inert to moist HCl and should be heated to prevent condensation. For HCl concentrations in the low ppm range, some HCl absorption in the sample delivery system is likely to occur upon monitoring system start up. However, after an initial conditioning period, the HCl instrument readings should become stable. Permapure dryers are the preferred method of moisture removal where moisture contents are high. Gas dilution to reduce the moisture content below the dew point of the gas mixture may also be used. However, the latter technique is not always successful if cold spots are present in the sample delivery system. The gas

dilution technique will dilute the sample below the HCl quantitation limit for some of the instrumental methods.

Limited information can be provided at this time about the accuracy, precision, repeatability, and reliability of HCl monitoring systems as applied to gas streams from pulp and paper mills. It should be noted that none of the aforementioned methods were developed for, or have yet been successfully applied to, measuring HCl emissions from recovery furnaces at kraft pulp mills. However, performance characteristics of GFC monitoring systems at hazardous waste incineration facilities have been studied by EPA and have been found to be acceptable.[36]

The NCASI completed an evaluation of a TECO 15 GFC monitoring system using the dilution approach for moisture removal.[37] Method 26 data were gathered concurrently with the monitoring data. The investigators were unable to establish a good correlation between the GFC monitoring data and the EPA Method 26 data. Several factors may have been responsible for the poor results. For example, the HCl content of the gas stream may have been reduced to levels below the detection limit of the instrument by the dilution gases. It is possible to compensate for the lower gas concentrations by increasing the optical path length. However, sufficient details were not provided in the study to confirm this. Loss of HCl in the sampling lines is also a possibility due to the high moisture contents. Improved HCl recoveries may be achieved by using other moisture management schemes, such as a Nafion® drier at or near the stack. The Nafion® drier would remove moisture while the gas stream is still hot and significantly reduce the potential for water condensation in the sampling lines. Furthermore, the use of the Nafion® drier as an alternative to dilution gases lessens the potential for reducing the HCl concentrations below the GFC monitor detection limits.

Based on information generated in an EPA study of GFC monitoring systems, the initial costs (purchase, installation, performance testing, quality assurance program plan preparation) and annual costs (operation & maintenance, recordkeeping and

reporting, program update, quality assurance) for a GFC monitor installed at a stack location are $144,500 and $68,700, respectively.[35,36] The estimated initial costs and annual costs associated with an FTIR monitoring system are $162,000 and $50,000, respectively, and those for an IMS system are approximately $149,500 and $68,700, respectively.[38,39] The SIE systems have not been costed because their performance characteristics are not acceptable.

Performance specifications for HCl monitoring systems in stationary sources are contained in Performance Specification 13, which was proposed on April 19, 1996 along with regulations for hazardous waste combustion systems.[40]

B.3.3 Methanol Monitoring

Emissions monitoring equipment for measuring vaporous methanol is currently commercially available; HAP specific monitors include on-line GC/FID and FTIR. The methanol quantitation limit for the GC/FID system is 2 to 3 ppm.[1] If lower detection limits are needed, an Entech concentrator can be used.

A GC-based monitoring system, called the hazardous organic mass emission rate (HOMER) monitor, is another type of monitoring system that combines EPA Methods 25A and 18 using a direct interface.[41] A total hydrocarbon (THC) analyzer provides a continuous measurement of the THC concentration, as propane, in an extracted gas sample. The analyzer output is recorded by a personal computer-based data acquisition system (PC/DAS). A gas chromatograph equipped with a FID is programmed to semi-continuously analyze the same gas sample stream every 6 minutes, with the GC/FID results being recorded by the PC-DAS. The HOMER software program converts the continuous THC signal to provide a continuous read out of the speciated VOC concentrations (in this case, methanol) in the gas stream using the GC/FID results. This is done by determining the relative response factors (RRF's) for each target VOC on the THC analyzer from gaseous standards of the compound at known concentrations. The GC/FID response determines the proportion of each target VOC in the gas stream. The EPA

recently sponsored a demonstration study of this monitoring system and showed that the HOMER monitor is a viable way to provide continuous information on speciated organic compounds.[41]

For emission sources where the correlation between methanol emissions and THC emissions can be established, continuous on-line FID monitors are readily available as surrogate monitors.

Current technical data indicate that the above monitoring systems are applicable for monitoring methanol emissions from SDT's and recovery furnace vents, as well as other pulp and paper vent streams.[1,2] Although no long-term continuous monitoring data are available, sufficient short-term integrated data have been gathered to establish that both the GC/FID system and the FTIR system can successfully measure methanol emissions from the above sources.[1,2,42] Data were recently submitted to EPA to demonstrate that the techniques can meet EPA Method 301 criteria.[30,31,42] It should be noted that none of the monitoring systems suggested for continuous monitoring of methanol emissions have been used for long-term monitoring of combustion sources at kraft or soda pulp mills. Therefore, long-term studies are required to determine the accuracy, precision, repeatability, and costs of setting up and operating the monitoring systems.

The initial costs (purchase, installation, performance testing, quality assurance program plan preparation) and annual costs (operation & maintenance, recordkeeping and reporting, program update, quality assurance) for a HOMER monitoring system installed at a stack location are $135,300 and $64,500, respectively.[39] The estimated costs associated with an THC monitoring system would be approximately $115,300 and $59,800, respectively.[39] The estimated costs associated with a FTIR monitoring system would be approximately $162,000 and $50,000, respectively.[39]

Performance specification 8A for THC systems can be found in the background information document which supports proposed regulations for hazardous waste combustion systems.[43] Additional performance specifications for THC systems and GC systems were proposed on October 22, 1993 in 58 FR 54648.[44] The FTIR sampling

and analysis protocols for measuring gaseous organic compounds, such as methanol, can be found in EPA Method 318.[45,46]

B.3.4 Parametric Monitoring Systems

Scrubber performance is easily monitored with off-the-shelf monitoring equipment for pH and liquid flow rate. There are various techniques for measuring liquid flow rate in a wet scrubber including ultra sonic detection mounted externally on the inflowing water pipe and turbine devices that are mounted within the pipe, both of which generate an electrical signal that can be logically displayed in a control room.[38] For liquid streams containing less than 1 percent solids, a paddle wheel sensor can be used to measure the flow rate. If the solids content of the liquid flow stream is greater than 1 percent, an insertion type of sensor must be used to determine flow rate.

The cost of the liquid flow rate monitoring equipment can vary from $2,000 to $25,000 depending on the device sensitivity and device location from the control room.[35] A complete paddle wheel flowrate monitoring system will cost approximately $1,200.[38] The initial cost of a flow rate monitoring system with an insertion type sensor is approximately $3,750.[38] The initial cost for a pH monitoring system is estimated to be $5,000.[38]

B.4 PERFORMANCE TEST METHODS

The following subsections discuss the test methods recommended for measuring emissions of particulates, metal HAP's, opacity, methanol, and HCl. This discussion covers available sampling procedures and analytical techniques. Costs to conduct performance testing using each of the methods are also presented. These costs include preparation of the final test report. In cases where alternate test methods or procedures have been evaluated, comments are provided on the relationships between the data obtained by the alternate methods and the data obtainable by the proposed reference methods. As appropriate, factors that could prevent the use of the reference methods to determine compliance are identified.

B.4.1 Particulate Matter and Metal HAP Emissions

Emissions of HAP metals and PM can be measured simultaneously using EPA Method 29.[47,48] Because of potential losses of Hg during sample desiccation, it is recommended that PM and Hg not be measured simultaneously. If these two species must both be measured, one of the two following options should be used:

1. Method 5[49] or Method 17[49] for PM and Method 29[48] for Hg and other trace metals of interest, or

2. Method 101A[48] for Hg and Method 29[48] for PM and other trace metals of interest.

Method 3, using the integrated sampling techniques, or Method 3A are recommended for measuring O_2 and CO_2.[49] The concentrations of these gases are necessary for correcting the particulate matter emission rate to a standard basis.

The sampling and analytical cost for three Method 5 runs with associated Method 3 testing is estimated to be $8,500 (all dollar amounts are expressed in 1994 dollars).[47] Method 17 may be used as an alternative to Method 5 if a constant value of 0.009 g/dscm (0.004 gr/dscf) is added to the results of Method 17 and the stack temperature is no greater than 205°C (400°F).[49] No sampling and analytical cost is available for Method 17 test runs. If a three-run Method 29 test for all trace metals and Hg is run in conjuction with a Method 5 test for PM, the additional cost of the Method 29 test would be approximately $8,750.[47] The estimated cost for three Method 29 runs for PM and trace metals with associated Method 3 testing is $12,000.[47] The additional cost for three Method 101A runs in conjunction with PM/metals testing is estimated at $5,500.[47]

B.4.2 Opacity

Method 9 should be used to determine the opacity of emissions.[49] The estimated cost for conducting three Method 9 observations in conjunction with PM emissions testing is $1,060.[47]

B.4.3 Hydrogen Chloride Emissions

Methods 26 or 26A should be used to determine the HCl emissions.[49] If the measurement point is downstream of a wet scrubber, Method 26A must be used. The estimated cost to conduct a three-run performance test using Method 26 or 26A is $9,100.[50] However, if the Method 26A test is run in conjunction with a Method 5 test, the additional cost for the Method 26A test would be approximately $600 (above the cost of a Method 5 test) for analyzing the HCl impingers after the PM filters.[50]

B.4.4 Methanol Emissions

Method 308 or the Methanol Sampling Train (MST) should be used to determine methanol emissions.[32,51] The Method 308 sampling train consists of an unheated Teflon® line, two 30-milliliter (mL) water-filled midget impingers in an ice bath, and one glass adsorbent tube packed in two sections containing 520 milligrams (mg) and 260 mg of silica gel, respectively. The gas sample is nonisokinetically extracted from the sampling point in the stack, and methanol is collected in deionized distilled water and adsorbed on silica gel. The sample is returned to the laboratory, where the methanol in the water fraction is separated from other organic compounds with a GC and is then measured by an FID. The fraction adsorbed on the silica gel is extracted with an aqueous solution of n-propanol and is then separated and measured by GC/FID. The analytical limit of quantitation (LOQ) in the water impinger samples is 0.50 nanograms (ng) (equivalent to about 0.5 ppm methanol in a 20-liter (L) sample of source gas). The overall LOQ for the method is 1.6 ppm of methanol for a 20-L sample of source gas.[51]

The MST consists of a glass-lined heated probe, two condensate knockout traps in an ice bath, and three sorbent cartridges packed with Anasorb® 747. A 1:1 mixture of carbon disulfide (CS_2) and N,N-dimethylformamide (DMF) is used to desorb methanol from the Anasorb® samples. The condensate traps also collect a significant amount of methanol, if water is present. Condensate and desorption samples are analyzed by GC/FID. The analytical LOQ in the CS_2/DMF samples is 0.69 ng (equivalent to

about 0.5 ppm methanol in a 20-L sample of source gas). The overall LOQ for the MST is 3 ppm of methanol for a 20-L sample of source gas.[51]

The estimated cost to conduct a three-run performance test using Method 308 in conjunction with other performance testing is $9,500.[35] A three-run performance test using the MST in conjunction with other performance testing should cost about $9,500 and offers the additional advantage that other chemical compounds can be measured simultaneously using the same sampling train, if desired.[52] Compounds in the emission matrix can be identified by mass spectral library matching if the desorbed samples are analyzed on a GC equipped with an MSD and appropriate computer software.

TABLE B-1. SUMMARY OF PULP AND PAPER COMBUSTION SOURCES TEST PROGRAMS

Test site/dates	Process unit and control system description	Sampling location	Sample type/method	Pollutants/parameters measured
NCASI Test Program: Mill C 3/19/93-3/24/93	Smelt dissolving tank w/AirPol venturi scrubber on vent; scrubber is single-pass (once-through) and uses weak wash as the scrubbing fluid; spent scrubbing fluid is routed to the smelt dissolving tank.	Smelt dissolving tank scrubber outlet	Flue gas/heated Summa® canister source sampling train	Acetone, methanol, carbon tetrachloride, chlorobenzene, chloroform, dichloromethane, 1,2-dichloroethane, 1,2-dichloroethylene, 1,1,1-trichloroethane, 1,1,2-trichloroethane, 1,2,4-trichlorobenzene, tetrachloroethylene, styrene, toluene, xylene, benzene, alpha- and beta-pinene, methyl isobutyl ketone, methyl ethyl ketone
			Flue gas/National Council midget impinger train	Acetaldehyde, methanol, acetone, methyl ethyl ketone
	Babcock & Wilcox NDCE Recovery Furnace (firing rate: 1,200 tons BLS/day) equipped w/ dry-bottom Koppers electrostatic precipitator (ESP)	Recovery furnace ESP outlet	Flue gas/heated Summa® canister source sampling train	Acetone, methanol, carbon tetrachloride, chlorobenzene, chloroform, dichloromethane, 1,2-dichloroethane, 1,2-dichloroethylene, 1,1,1-trichloroethane, 1,1,2-trichloroethane, 1,2,4-trichlorobenzene, tetrachloroethylene, styrene, toluene, xylene, benzene, alpha- and beta-pinene, methyl isobutyl ketone, methyl ethyl ketone
NCASI Test Program: Mill H 6/9/93-6/11/93	Three sources tested for identical pollutants: Combustion Engineering cascade DCE recovery furnace (firing rate: 2.4 million lb BLS/day at 64-69 percent dry solids, corresponding to a wood pulping rate of 800 ton/day), with 2 combustion air zones; Wheelabrator-Frye wet- bottom ESP with 50 percent solids oxidized liquor. Natural gas used for auxiliary fuel. 2-stage Black Liquor Oxidation (BLO) system; Retention times: 1st stage: 2 hours 2nd stage: 3 hours Liquor temperature: 1st stage: 190°F 2nd stage: 175°F Combustion Engineering NDCE recovery furnace (firing rate: 2.4 million lb BLS/day at 64-68 percent dry solids) with 2 combustion air zones; Wheelabrator-Frye wet bottom ESP w/68 percent solids unoxidized black liquor; combustion air and auxiliary fuel configuration identical to No. 3 recovery furnace.	BLO tank vents Nos. 1 and 2; No. 3 recovery furnace ESP outlet No. 4 recovery furnace ESP outlet	Flue gas/heated Summa® canister source sampling train	Acetaldehyde, acetone, methanol, methyl mercaptan, carbon tetrachloride, chlorobenzene, chloroform, methylene chloride, 1,2-dichloroethane, 1,2-dichloroethylene, dimethyl sulfide, dimethyl disulfide, 1,1,1-trichloroethane, 1,1,2-trichloroethane, 1,2,4-trichlorobenzene, trichloroethylene, tetrachloroethylene, styrene, toluene, m,p-xylene, o-xylene, benzene, terpenes, methyl isobutyl ketone, methyl ethyl ketone, acrolein

TABLE B-1. (continued)

Test site/dates	Process unit and control system description	Sampling location	Sample type/method	Pollutants/parameters measured
Champion-Lufkin, TX 4/29/92 - 9/16/92	Fuller gas-fired rotary lime kiln, diameter = 10.5', length = 275', chain section length = 37.5'; produces 125-140 tons/d CaO. Controlled with venturi scrubber (max pressure drop: 19" H_2O) which uses fresh water for makeup and throughout the causticizing system. Precoat drum washer (length = 14', diameter = 10') used for mud washing. NCG's are burned in the kiln. (NOTE: all listed emission test methods were run both with and without NCG burning in the kiln).	Kiln stack	Flue gas/M16	Hydrogen sulfide, methyl mercaptan, dimethyl sulfide, carbon disulfide dimethyl disulfide
			Flue gas/M18	Methanol, ethanol, acetone, 2-propanol, 2-butanone, chloroform, benzene, bromodichloromethane, toluene, ethyl benzene, m-, p- xylene, o-xylene, cumene, alpha-pinene, beta-pinene, 3-carene, terpenes (unspecified), p-cymene
			Flue gas/SW-846 M0030	Chloromethane, bromomethane, methylene chloride, acetone, carbon disulfide, chloroform, trichlorofluoromethane, dimethyl disulfide, dimethyl sulfide, n-hexane, 2-butanone, bromodichloromethane, trichloroethene, benzene, toluene, ethylbenzene, a-pinene, b-pinene, p-cymene
			Flue gas/SW-846 M0011	Formaldehyde, acetaldehyde, acetone, acetophenone, methyl isobutyl ketone, acrolein, benzaldehyde
			Flue gas/M29	Antimony, arsenic, barium, beryllium, cadmium, chromium, copper, lead, manganese, mercury, nickel, silver, selenium, thallium
			Flue gas/M26	Hydrogen chloride

TABLE B-1. (continued)

Test site/dates	Process unit and control system description	Sampling location	Sample type/method	Pollutants/parameters measured
Champion-Lufkin, TX (continued)	Gotaverken NDCE Recovery furnace (black liquor firing rate: 1000 lb BLS/hr or 177 gal/min, saltcake addition rate: 720 lb/hr, black liquor fired at 68% solids); dry-bottom Flakt ESP.	Recovery Furnace No. 14 ESP outlet	Flue gas/M29	Antimony, arsenic, barium, beryllium, cadmium, chromium, copper, lead, manganese, mercury, nickel, phosphorous, silver, selenium, thallium
			Flue gas/M26	Hydrogen chloride
			Flue gas/M26A	Sulfuric acid mist
	Smelt dissolving tank vent on the No. 14 recovery furnace; emissions controlled with low pressure venturi scrubber (pressure drop = -1 in w.c.); scrubber liquor flowrate = 200 gal/min).	Smelt dissolving tank vent scrubber outlet	Flue gas/M29	Antimony, arsenic, barium, beryllium, cadmium, chromium, copper, lead, manganese, mercury, nickel, phosphorous, silver, selenium, thallium
Champion-Sheldon, TX 6/9/92 - 6/24/92	Two-stage black liquor oxidation system Retention times: 1st stage: 460 min 2nd stage: 140 min. Air provided by compressor to each stage. Compressor flow rates: primary stage: 6,760 CFM, secondary stage: 9,000 CFM. Vent gases pass through four Impco foam breakers on the primary stage, and cyclone separator on the secondary stage.	BLO vent	Flue gas/M16	Hydrogen sulfide, methyl mercaptan, dimethyl sulfide, carbon disulfide dimethyl disulfide
			Flue gas/M18	Methanol, ethanol, acetone, 2-propanol, 2-butanone, chloroform, benzene, bromodichloromethane, toluene, ethyl benzene, m-, p- xylene, o-xylene, cumene, alpha-pinene, beta-pinene, 3-carene, terpenes (unspecified), p-cymene
			Flue gas/SW-846 M0030	Chloromethane, bromomethane, methylene chloride, acetone, carbon disulfide, chloroform, trichlorofluoromethane, dimethyl disulfide, dimethyl sulfide, n-hexane, 2-butanone, bromodichloromethane, trichloroethene, benzene, dibromomethane, toluene, ethylbenzene, o-xylene, m-/p-xylene, a-pinene, b-pinene, p-cymene

TABLE B-1. (continued)

Test site/dates	Process unit and control system description	Sampling location	Sample type/method	Pollutants/parameters measured
Champion-Sheldon, TX (continued)	Combustion Engineering DCE recovery furnace [firing rate: 2,100,000 lb black liquor solids (BLS) per day, corresponding to a wood pulping rate of 575 tons/d; primary and secondary combustion air zones; BL fired at 65 percent solids by weight; 0.003 gal/lb BLS sodium added in the form of sodium acetate; 4 natural gas-fired starter burners located just above primary air ports; brine from tall oil plant added to liquor; BL Btu content = 5800 Btu/lb BLS; recovery furnace temperature range: 1850-1950 °F.] Equipped with ESP; PM concentration in flue gases typically around 0.026 gr/SDCF. <u>Representative process conditions</u>: liquor firing rate: 42 tons BLS/hour sulfidity (green liquor): 65-70% Excess air or O_2 in furnace: 1-2% Makeup chemicals and addition rates: Sodium acetate @ 4 gal/min* CO, NO_x, SO_2, TRS in the furnace @ 8% O_2: CO: 175-275 ppm NO_x: 40 ppm SO_2: 75-100 ppm TRS: 10-15 ppm *(NOTE: test methods M16, M18, M26 and M0011 were also performed under the conditions of no sodium acetate being added to the furnace.)	Recovery furnace ESP outlet	Flue gas/M16 M0030	Hydrogen sulfide, methyl mercaptan, dimethyl sulfide, carbon disulfide, dimethyl disulfide
			Flue gas/M18	Methanol, ethanol, acetone, 2-propanol, 2-butanone, chloroform, benzene, bromodichloromethane, toluene, ethyl benzene, m-, p- xylene, o-xylene, cumene, alpha-pinene, beta-pinene, 3-carene, terpenes (unspecified), p-cymene

TABLE B-1. (continued)

Test site/dates	Process unit and control system description	Sampling location	Sample type/method	Pollutants/parameters measured
Champion-Sheldon, TX (continued)	DCE recovery furnace (continued)	Recovery furnace ESP outlet (continued)	Flue gas/SW-846 M0030	Chloromethane, bromomethane, methylene chloride, acetone, carbon disulfide, chloroform, trichlorofluoromethane, dimethyl disulfide, dimethyl sulfide, n-hexane, 2-butanone, bromodichloromethane, trichloroethene, benzene, dibromomethane, toluene, ethylbenzene, o-xylene, m-/p-xylene, a-pinene, b-pinene, p-cymene
			Flue gas/M26	Hydrogen chloride
			Flue gas/M26A	Sulfuric acid
			Flue gas/SW-846 M0011	Formaldehyde, acetaldehyde, acetone, acetophenone, 2-butanone, methyl isobutyl ketone, acrolein, benzaldehyde
			Flue gas/M29	Antimony, arsenic, barium, beryllium, cadmium, chromium, copper, lead, manganese, mercury, nickel, phosphorous, silver, selenium, thallium
	Smelt dissolving tank, equipped with Munters mist eliminators (scrubbing liquid: fresh water); sprays weak wash to flue gas; weak wash contains unstripped condensate, used to dissolve smelt; fumes from tall oil plant enter the tank vent at a point 10' below the bottom showers. Representative process conditions: BLS firing rate: 42 tons BLS/hr Scrubbing liquid flow rate: 30 gal/min Pressure drop in particulate control device: 8 in. H_2O Particulate control device: Chevron type demister	Smelt dissolving tank demister outlet	Flue gas/M16	Hydrogen sulfide, methyl mercaptan, dimethyl sulfide, carbon disulfide dimethyl disulfide

TABLE B-1. (continued)

Test site/dates	Process unit and control system description	Sampling location	Sample type/method	Pollutants/parameters measured
Champion-Sheldon, TX (continued)	Smelt dissolving tank, continued	Smelt dissolving tank demister outlet, continued	Flue gas/M18	Methanol, ethanol, acetone, 2-propanol, 2-butanone, chloroform, benzene, bromodichloromethane, toluene, ethyl benzene, m-, p- xylene, o-xylene, cumene, alpha-pinene, beta-pinene, 3-carene, terpenes (unspecified), p-cymene
			Flue gas/M29	Antimony, arsenic, barium, beryllium, cadmium, chromium, copper, lead, manganese, mercury, nickel, phosphorous, silver, selenium, thallium
Inland-Orange Orange, TX 7/13/92 - 7/21/92	Traylor rotary natural gas-fired lime kiln, diameter = 12', length = 260'. Noncondensable gases are normally burned in the kiln. Kiln produces 283 tons/d CaO (equivalent to 1,400 tons/d of unbleached pulp). PM emission controlled w/ variable throat venturi scrubber using fresh mill water. Scrubber pressure drop =14 in w.c. Makeup rate= 68 gal/min Representative process conditions: hot-end temp: 2,284-2,425°F cold-end temp: 721-768°F oxygen in kiln: 4.9%-6.3% lime mud solids: 69.4%-74.4% Na content of mud: 8.3-12.8%	Lime kiln scrubber outlet	Flue gas/M16	Hydrogen sulfide, methyl mercaptan, dimethyl sulfide, carbon disulfide dimethyl disulfide
			Flue gas/M18	Methanol, ethanol, acetone, 2-propanol, 2-butanone, chloroform, benzene, bromodichloromethane, toluene, ethyl benzene, m-, p- xylene, o-xylene, cumene, alpha-pinene, beta-pinene, 3-carene, terpenes (unspecified), p-cymene
			Flue gas/SW-846 M0030	Bromomethane, methylene chloride, acetone, carbon disulfide, acrolein, dibromomethane, 2-butanone, benzene, toluene, styrene

TABLE B-1. (continued)

Test site/dates	Process unit and control system description	Sampling location	Sample type/method	Pollutants/parameters measured
Simpson-Pasadena-Pasadena, TX 5/18/92-6/25.92	Babcock & Wilcox cyclone DCE recovery furnace with two chamber, eight-field/chamber wet-bottom ESP. Particulate level entering ESP: 15 - 20 gr/CF. Black liquor is oxidized in 2 stages and fired at 62 to 65% solids. Furnace temperature: >2000 °F. Pulp production capacity: 550 tons/d BLS density: 12 lb/gal Liquor firing rate: 60.9 - 66.7 lb BLS/hr Sulfidity (green liquor): 29.9% BL Btu content: 5830 Btu/lb Makeup chemicals: sodium hydrosulfide plus sodium hydroxide Emission data: O_2: 7.8% - 8.3% SO_2: 58 ppm - 111 ppm TRS: 0.7 ppm - 1.1 ppm	Recovery furnace ESP stack	Flue gas/M16	Hydrogen sulfide, methyl mercaptan, dimethyl sulfide, carbon disulfide dimethyl disulfide
			Flue gas/M18	Methanol, ethanol, acetone, 2-propanol, 2-butanone, chloroform, benzene, bromodichloromethane, toluene, ethyl benzene, m-, p- xylene, o-xylene, cumene, alpha-pinene, beta-pinene, 3-carene, terpenes (unspecified), p-cymene
			Flue gas/SW-846 M0030	Chloromethane, bromomethane, methylene chloride, acetone, carbon disulfide, chloroform, trichlorofluoromethane, dimethyl disulfide, dimethyl sulfide, iodomethane, isooctane, n-hexane, 2-butanone, benzene, toluene, ethylbenzene, styrene, m-p- xylene, acrolein, a-pinene, b-pinene, p-cymene
			Flue gas/SW-846 M0011	Formaldehyde, acetaldehyde, acetone, acetophenone, 2-butanone, methyl isobutyl ketone, acrolein, benzaldehyde

TABLE B-1. (continued)

Test site/dates	Process unit and control system description	Sampling location	Sample type/method	Pollutants/parameters measured
Simpson-Pasadena Pasadena, TX continued	DCE recovery furnace, continued	Recovery furnace ESP stack, continued	Flue gas/SW-846 M0010	A-pinene, b-pinene, 1,2,4-trichlorobenzene, a-terpineol, naphthalene, di-n-butylphthalate, phenol, 2-methylphenol, 1,2,4-trichlorobenzene, naphthalene, di-n-butylphthalate, bis(2-ethylhexyl)phthalate
	Smelt dissolving tank with scrubber; uses fresh water as scrubbing medium.	Smelt dissolving tank scrubber outlet	Flue gas/M16	Hydrogen sulfide, methyl mercaptan, dimethyl sulfide, carbon disulfide, dimethyl disulfide
			Flue gas/M18	Methanol, ethanol, acetone, 2-propanol, 2-butanone, chloroform, benzene, bromodichloromethane, toluene, ethyl benzene, m-, p- xylene, o-xylene, cumene, alpha-pinene, beta-pinene, 3-carene, terpenes (unspecified), p-cymene
			Flue gas/SW-846 M0030	Chloromethane, bromomethane, methylene chloride, acetone, carbon disulfide, chloroform, dimethyl disulfide, dimethyl sulfide, iodomethane, isooctane, n-hexane, 2-butanone, 2-chloro-1,3-butadiene, bromodichloromethane, trichloroethane, tert-butyl methyl ether, tetrachloroethene, benzene, toluene, ethylbenzene, styrene, o-xylene, m-p- xylene, a-pinene, b-pinene, cumene, p-cymene
			Flue gas/SW-846 M0010	A-pinene, b-pinene, 1,2,4-trichlorobenzene, a-terpineol, naphthalene, di-n-butylphthalate, butylbenzylphthalate, 2-methylnaphthalene, benzyl alcohol, 1,2,4-trichlorobenzene, naphthalene, di-n-butylphthalate, bis(2-ethylhexyl)phthalate

TABLE B-1. (continued)

Test site/dates	Process unit and control system description	Sampling location	Sample type/method	Pollutants/parameters measured
Simpson-Pasadena, TX (continued)	Natural gas-fired rotary lime kiln, diameter = 11.5', length = 275, chain section length = 75'. No NCG's burned in kiln. CaO production rate: 195 tons/d. PM emissions controlled with 2-chamber dry bottom ESP with 6 fields per chamber. Eimco belt filter (length = 16', diameter = 10') used for mud washing and dewatering. Representative process conditions: CaO production: 8.1 tons CaO/hr Fuel: natural gas at 56,000 cf/hr Btu/ton product: 7.1 MMBtu/ton of lime Cold end temp: 360°F Hot end temp: 2000°F Solids content of mud: 70% Na_2S and soda content of mud: 0.29% Na_2O	Lime kiln ESP outlet	Flue gas/M16	Hydrogen sulfide, methyl mercaptan, dimethyl sulfide, carbon disulfide, dimethyl disulfide
			Flue gas/M18	Methanol, ethanol, acetone, 2-propanol, 2-butanone, chloroform, benzene, bromodichloromethane, toluene, ethyl benzene, m-, p- xylene, o-xylene, cumene, alpha-pinene, beta-pinene, 3-carene, terpenes (unspecified), p-cymene
			Flue gas/SW-846 M0011	Formaldehyde, acetaldehyde, acetone, acetophenone, 2-butanone, methyl isobutyl ketone, acrolein, benzaldehyde

TABLE B-1. (continued)

Test site/dates	Process unit and control system description	Sampling location	Sample type/method	Pollutants/parameters measured
Temple-Inland Silsbee, TX 7/9/92 - 7/21/92	Combustion Engineering recovery furnace, modified to noncontact configuration. Average firing rate: 1.5 million lb BLS/day, corresponding to a wood pulping rate of 500 tons/d. Primary, secondary, and tertiary air. Black liquor fired at 68% - 70% solids. No makeup sulfur necessary. Recovery boiler temperatures usually >1800°F. Wet bottom ESP; exhaust particulate concentration: 0.047 gr/SDCF. Representative process conditions: liquor firing rate: 120 gal/min Sulfidity (green liquor): 4.30 Primary, secondary: 66% primary, 34% secondary Excess air or O_2 in furnace: 7% Black liquor Btu content: 5600 Btu/lb NO_x: 0.10 lb/MMBtu @ 39 ppm TRS: 10 ppm average SO_2: 64 ppm	Recovery furnace ESP outlet	Flue gas/M16	Hydrogen sulfide, methyl mercaptan, dimethyl sulfide, carbon disulfide dimethyl disulfide
			Flue gas/M18	Methanol, ethanol, acetone, 2-propanol, 2-butanone, chloroform, benzene, bromodichloromethane, toluene, ethyl benzene, m-, p- xylene, o-xylene, cumene, alpha-pinene, beta-pinene, 3-carene, terpenes (unspecified), p-cymene
			Flue gas/SW-846 M0030	Methylene chloride, acetone, n-hexane, 2-butanone, benzene, dibromomethane, toluene, styrene, o-xylene, m-p- xylene, a-pinene, b-pinene, p-cymene
			Flue gas/SW-846 M0011	Formaldehyde, acetaldehyde, acetone, acetophenone, 2-butanone, methyl isobutyl ketone, acrolein, benzaldehyde

TABLE B-1. (continued)

Test site/dates	Process unit and control system description	Sampling location	Sample type/method	Pollutants/parameters measured
NCASI Test Program: Mill G 1/27/94-1/28/94	Babcock & Wilcox NDCE recovery furnace; 4 oscillating liquor guns. Black liquor fired at 65% solids, Btu value = 5400 Btu/lb. 2050 lb BLS produced per ton oven dried pulp. BL inorganic content = 47%. BL from softwood kraft, hardwood kraft, semi-chem pulp lines are mixed before evaporation. Brine used for sulfur make-up. Brine added to BL between evaporators and concentrators. A wet-bottom precipitator, with BL in the bottom, is used for particulate control. Test run-average process information: BLS firing rate: 1343 ton BLS/day BL temperature: 235 °F Air distribution: Primary: 39% Secondary: 45% Tertiary: 16% Boiler CO conc.: 46 ppm Boiler O_2 conc.: 0.97% Gas Exit Temp: 393 °F	Recovery furnace ESP outlet	Flue gas/heated Summa® canister source sampling train	Acetaldehyde, acetone, acrolein, methanol, mercaptan, carbon tetrachloride, chlorobenzene, chloroform, dichloromethane, 1,2-dichloroethane, 1,2-dichloroethylene, dimethyl disulfide, dimethyl sulfide, 1,1,1-trichloroethane, 1,1,2-trichloroethane, 1,2,4-trichlorobenzene, tetrachlorobenzene, styrene, toluene, m-,p- xylene, o-xylene, benzene, terpenes, methyl isobutyl ketone, methyl ethyl ketone, n-hexane
			Flue gas/National Council Midget Impinger Train	Acetaldehyde, methanol, acetone, methyl ethyl ketone, formaldehyde

TABLE B-1. (continued)

Test site/dates	Process unit and control system description	Sampling location	Sample type/method	Pollutants/parameters measured
NCASI Test Program: Mill D 5/8/93-5/11/93	Babcock & Wilcox short-body DCE recovery furnace (No. 1 furnace), design capacity: 2.1 million lb/day BLS. BL exits BLOx system at 50% BLS, flows through wet-bottom ESP. Liquor then passes through cyclone-type DCE, where solids content is raised to 65% prior to feed to recovery furnace. Test run-average process conditions: firing rate: 2.7 million lb BLS/day BL temperature: 216 °F BLS concentration: 66.6% Air distribution: primary: 34% secondary: 45% tertiary: 21% Boiler CO conc.: 7 ppm Boiler O_2 conc.: 3.6%	Recovery furnace ESP outlet	Flue gas/heated Summa® canister source sampling train	Acetone, methanol, methyl mercaptan, carbon tetrachloride, chlorobenzene, chloroform, dichloromethane, 1,2,-dichloroethane, 1,2-dichloroethylene, dimethyl disulfide, dimethyl sulfide, 1,1,1-trichloroethane, 1,2,4-trichlorobenzene, tetrachloroethylene, trichloroethylene, styrene, toluene, m-,p- xylene, o-xylene, benzene, terpenes, methyl isobutyl ketone, methyl ethyl ketone
			Flue gas/National Council Midget Impinger Train	Acetaldehyde, methanol, acetone, methyl ethyl ketone, formaldehyde
4/27/93 - 4/30/93	Two-stage black liquor oxidation (BLO) system for No. 1 DCE recovery furnace; air sparged through heavy black liquor in 2 stages connected in series. BL in second stage fed to foam tank, then to No. 1 furnace. During testing, considerably more air was fed to the first oxidation stage than to the second stage (oxidizing BL equally in each stage caused firing problems in the recovery furnace). Design air flow rate = 550 ACFM @ 72°F; oxidation efficiency = 85%; retention time at operating temperature = 9.4 hours Test run-average process conditions: BL flow rate: 236 gal/min BL temperature: 215 °F BL solids conc.: 48%	BLO tank vents Nos. 1 and 2; BLO foam tank vents A and B	Flue gas/heated Summa® canister source sampling train	Acetone, methanol, methyl mercaptan, carbon tetrachloride, chlorobenzene, chloroform, dichloromethane, 1,2-dichloroethane, 1,2-dichloroethyene, dimethyl disulfide, dimethyl sulfide, 1,1,1-trichloroethane, 1,1,2-trichloroethane, 1,2,4-trichlorobenzene, tetrachloroethylene, trichloroethylene, styrene, toluene, m-,p-, xylenes, o-xylenes, benzene, terpenes, methyl isobutyl ketone, methyl ethyl ketone

TABLE B-1. (continued)

Test site/dates	Process unit and control system description	Sampling location	Sample type/method	Pollutants/parameters measured
NCASI test program: Mill D (continued)	BLO system, continued	BLO system vents, continued	Flue gas/National Council Midget Train	Acetaldehyde, methanol, acetone, methyl ethyl ketone, formaldehyde
4/29/93-5/4/93	No. 1 and No. 2 smelt dissolving tanks. SDT vent gases are routed through orifice scrubber, with weak wash as scrubbing medium. Scrubber effluent flows into the smelt dissolving tanks. Test run-average operating conditions: No. 1 tank: Recovery firing rate: 1.62 million lb BLS/day Scrubber makeup flow rate: N/A No. 2 tank: Recovery firing rate: 2.89 million lb BLS/day Scrubber makeup flow rate: 377 gal/min	No. 1 and No. 2 smelt dissolving tank scrubber outlets	Flue gas/heated Summa® canister source sampling train	Acetone, methanol, methyl mercaptan, carbon tetrachloride, chlorobenzene, chloroform, dichloromethane 1,2-dichloroethylene, dimethyl disulfide, dimethyl sulfide, 1,1,1-trichloroethane, 1,1,2-trichlorobenzene, tetrachlorobenzene, tetrachloroethylene, trichloroethylene, styrene, toluene, m-,p-xylene, o-xylene, benzene, terpenes, methyl isobutyl ketone, methyl ethyl ketone
			Flue gas/National Council Midget Impinger Train	Acetaldehyde, methanol, acetone, methyl ethyl ketone, formaldehyde
Potlatch Corp.; Cloquet, MN 4/26/89-4/27/89	No. 8 B&W NDCE Recovery Boiler - steaming capacity: 250,000 lb/hr. PM emissions controlled with Flakt electrostatic precipitator (maximum outlet loading: 0.023 gr/dscf; inlet loading: 10 gr/dscf; maximum opacity: 10% @ 182,000 ACFM; specific collection area = 28 ft^2 CA/KACFM; gas velocity = 2.49 ft/sec; treatment time = 11.82 seconds). Cleaned flue gas exhausted to atmosphere through radial steel stack, height = 249;', internal diameter = 7' - 10'.	Recovery furnace ESP outlet	Flue gas/M12	Nickel, manganese
			Flue gas/Modified M5	Chromium species

TABLE B-1. (continued)

Test site/dates	Process unit and control system description	Sampling location	Sample type/method	Pollutants/parameters measured
Potlatch Corp., (continued)	NDCE recovery furnace, continued	Recovery furnace ESP outlet, continued	Flue gas/Modified M5	Hydrogen chloride
			Flue gas/modified NIOSH Method 3500	Formaldehyde
			Flue gas/SW-846 M0030	Chloroform
			Flue gas/M8270	Acenaphthene, pyrene, benzo-a-anthracene, chrysene, benzo-b-fluoranthene, benzo-a-pyrene, dibenzo-a,h, i-pyrene, indeno-1,2,3-pyrene
			Flue gas/modified M5; SW846 M0010; M8290	polychlorinated dibenzo-p-dioxins (PCDD's) and polychlorinated dibenzofurans (PCDF's)
NCASI test program: Mill J 8/4/93-8/8/93	Three sources were tested for same pollutants: No. 3 DCE recovery recovery furnace w/ wet-bottom ESP and venturi scrubber. Black liquor fired at 55% solids. No. 4 NDCE recovery furnace w/ dry-bottom ESP. Black liquor fired at 63% solids. No. 1 lime kiln, natural gas or oil-fired, produces 250 tons CaO per day; venturi scrubber on outlet.	No. 3 recovery furnace ESP outlet No. 4 recovery furnace ESP outlet No. 1 lime kiln scrubber outlet	Flue gas/ heated Summa® canister source sampling train	Acetaldehyde, acetone, acrolein, methanol, methyl mercaptan, carbon tetrachloride, chlorobenzene, chloroform, dichloromethane, 1,2-dichloroethane, 1,2-dichloroethylene, dimethyl disulfide, dimethyl sulfide, 1,1,1-trichloroethane, 1,1,2-trichloroethane, 1,2,4-trichlorobenzene, tetrachloroethylene, trichloroethylene, styrene, toluene, m-,p-xylene, o-xylene, benzene, terpenes, methyl isobutyl ketone, methyl ethyl ketone
			Flue gas/National Council Midget Impinger Train	Acetaldehyde, methanol, acetone, methyl ethyl ketone, formaldehyde
NCASI test program: Mill K 10/1/93-10/5/93	B&W NDCE recovery furnace; 4 oscillating liquor guns; black liquor fired at 67% solids; black liquor BTU value = 5800 BTU/lb. PM emissions controlled with dry-bottom ESP. Test run average operating conditions: BLS firing rate: 1560 tons/d BL temp: 239°F BL solids content: 68.7% CO emissions: 17 ppm O_2 content in furnace: 3.7% Flue gas exit temp: 378°	Recovery furnace ESP outlet	Flue gas/heated Summa® canister sources sampling train	Acetaldehyde, acetone, acrolein, methanol, methyl mercaptan, carbon tetrachloride, chlorobenzene, chloroform, dichloromethane, 1,2-dichloroethane, 1,2-dichloroethylene, dimethyl disulfide, dimethyl sulfide, 1,1,1-trichloroethane, 1,1,2-trichloroethane, 1,2,4-trichlorobenzene, tetrachloroethylene, trichloroethylene, styrene, toluene, m-,p-xylene, o-xylene, benzene, terpenes, methyl isobutyl ketone, methyl ethyl ketone, n-hexane
			Flue gas/National Council Midget Impinger Train	Acetaldehyde, methanol, acetone, methyl ethyl ketone, formaldehyde

TABLE B-1. (continued)

Test site/dates	Process unit and control system description	Sampling location	Sample type/method	Pollutants/parameters measured
NCASI test program: Mill L 11/11/93-11/18/93	Babcock & Wilcox DCE recovery furnace (No. 1) w/direct contact cyclone evaporators and wet bottom precipitator. Precipitator uses concentrated BL (solids content: 52 percent). Representative operating conditions: Firing rate: 1005 tons BLS/day BLS content: 62% - 63% BL Btu content: 5550 Btu/lb BL temperature: 240°F Furnace oxygen content: 2.5%-3.0% Stack oxygen content: 6.0%-6.5% Test run-specific operating parameters (average): BLS firing rate: 1013 tons BLS/day BL solids contents: 63% BL temperature: 215°F	Recovery furnace ESP outlet	Flue gas/heated Summa® canister source sampling train	Acetaldehyde, acetone, acrolein, methanol, methyl mercaptan, carbon tetrachloride, chlorobenzene, chloroform, dichloromethane, 1,2-dichloroethane, 1,2-dichloroethylene, dimethyl disulfide, dimethyl sulfide, 1,1,1-trichloroethane, 1,1,2-trichloroethane, 1,2,4-trichlorobenzene, tetrachloroethylene, trichloroethylene, styrene, toluene, m-,p-xylene, o-xylene, benzene, terpenes, methyl isobutyl ketone, methyl ethyl ketone, n-hexane
			Flue gas/National Council Midget Impinger Train	Acetaldehyde, methanol, acetone, methyl ethyl ketone, formaldehyde
NCASI test program: Mill L (continued)	No. 3 Lime Kiln, 300'Lx11.5'D, slope = 7/16" per foot. Retention time = 2.5 hrs, process rate = 275 tons lime/day; 1 rpm. Mechanical dust collector + venturi scrubber + cyclonic separator on gas outlet stream. Mill water (200 gal/min) used as scrubbing medium. Test run-specific operating parameters: Production: 277 tons CaO/day Lime mud solids content: 75%-80%.	Lime kiln cyclonic separator outlet	Flue gas/National Council Midget Impinger Train	Acetaldehyde, methanol, acetone, methyl ethyl ketone, formaldehyde

TABLE B-1. (continued)

Test site/dates	Process unit and control system description	Sampling location	Sample type/method	Pollutants/parameters measured
NCASI test program: Mill M 12/7/93 - 12/16/93	Two sources tested for same pollutants: 1) Babcock & Wilcox NDCE Recovery Furnace w/dry bottom ESP, firing rate = 5.4 million pounds of BLS/day. BL at 73% solids is fed to product flash tank, then to furnace. Representative operating parameters: BLS Production: 3600 lb BLS/ODTP Liquor sulfidity: 25%-30% Soda and sulfur makeup: Saltcake from R8 ClO_2 generator, added to precipitator mix tank No. of liquor guns: 4 Auxiliary fuel: Natural gas or No. 1 fuel oil Combustion air: 4 levels of combustion air BLS firing rate: 237,000 lb BLS/hr BL temp: 252°F ESP O_2: 1.5 % Stack TRS: 1ppm	Recovery furnace ESP outlet Inlet and outlet of lime kiln scrubber	Flue gas/Summa® canister source sampling train	Acetaldehyde, acetone, acrolein, methanol, methyl mercaptan, carbon tetrachloride, chlorobenzene, chloroform, dichloromethane, 1,2-dichloroethane, 1,2-dichloroethylene, dimethyl disulfide, dimethyl sulfide 1,1,1-trichloroethane, 1,1,2-trichloroethane, 1,2,4-trichlorobenzene, tetrachloroethylene, trichloroethylene, styrene, toluene, m-,p-xylene, o-xylene, benzene, terpenes, methyl isobutyl ketone, methyl ethyl ketone, n-hexane
NCASI test program: Mill M (continued)	(continued) 2) Lime kiln; can use either natural gas or fuel oil; receives NCG's; PM emissions controlled with single chamber, 3-field ESP followed by wet scrubber with 8% caustic for SO_2 removal. Test run-specific operating parameters (average): Kiln throughput: 553 tons CaO/day Cold end temp: 1210°F Hot end temp: 2535°F Exit TRS concentration: 3.9 ppm Exit VOC concentration: 8.7 ppm Exit SO_2 concentration: 56 ppm	(continued) Recovery furnace ESP outlet Lime kiln cyclonic separator outlet	Flue gas/National Council Midget Impinger Train	Acetaldehyde, methanol, acetone, methyl ethyl ketone, formaldehyde

TABLE B-1. (continued)

Test site/dates	Process unit and control system description	Sampling location	Sample type/method	Pollutants/parameters measured
Federal Paper Board August, GA 5/30/91	Two sources tested for HCl: 1) No. 2 recovery furnace, DCE (cyclone) with wet-bottom ESP; BLS firing rate: ≈1.6 million lb BLS/d; BLS content = 65%. Test run-specific operating parameters (average): Stack temperature: 351°F Stack flow rate: 99,255 DCFM O_2 concentration: 8.4% 2) No. 3 recovery furnace, NDCE with dry-bottom ESP; BLS firing rate ≈4.65 million lb BLS/d; BLS content = 65%. Test run-specific operating parameters (average): Stack temperature: 356°F Stack flow rate: 175,414 DCFM O_2 concentration: 9.2%	No. 2 recovery furnace ESP outlet No. 3 recovery furnace ESP outlet	Flue gas/M26	Hydrogen chloride
Consolidated Papers 11/29/89- 12/1/89	Recovery furnace No. 2; DCE (cascade) w/ wet-bottom ESP; design capacity: 1.5 million lb BLS/d Test run-specific operating parameters (average): Pulp production: 858 tons/d BL firing rate: 106 gal/min BL solids concentration: 70% Stack temperature: 262°F O_2 concentration: 6% Stack gas flow: 51,000 dscfm Average total PM emissions: 0.0069 gr/dscf	Recovery furnace ESP outlet	Flue gas/CARB M425	Total PM, Chromium species
			Flue gas/EPA proposed multi-metals method	Barium, beryllium, cadmium, chromium, nickel

TABLE B-1. (continued)

Test site/dates	Process unit and control system description	Sampling location	Sample type/method	Pollutants/parameters measured
Boise-Cascade, International Falls, MN 2/4/91 - 2/9/91	B&W NDCE Recovery furnace w/ dry-bottom ESP; can be fired with BLS, natural gas, or distillate oil; firing rate: 2,300,000 lb BLS/hr. Environmental Elements ESP with 8 fields, design particulate removal efficiency: 99.9%. Test run-specific operating parameters (average): Steam generation: 322,433 lb/hr BL solids firing rate: 43.4 tons/hr O_2 concentration: 7.2% CO concentration: 0.0% Flue gas temperature: 417 °F	Recovery furnace ESP outlet	Flue gas/EPA document "Methodology for the Det. of Metals Emissions in Exhaust Gases from Stationary Source Combustion Processes"	Nickel, manganese, total chromium
			Flue gas/EPA document "Suggested Operating Procedure for Det. of Formaldehyde Emissions from Stationary Sources"	Formaldehyde
Boise Cascade (continued)	NDCE recovery furnace, (continued)	Recovery furnace ESP outlet, (continued)	Flue gas/NCASI "Method for the Sampling and Analysis of Chloroform Emitted Directly to the Atmosphere from Bleach Plant Point Sources"	Chloroform
			Flue gas/M26	Hydrochloric acid
			Flue gas/modified M5, w/ M8290	dioxins, furans and PAH's
			Flue gas/CARB M425	hexavalent chromium

TABLE B-1. (continued)

Test site/dates	Process unit and control system description	Sampling location	Sample type/method	Pollutants/parameters measured
NCASI test program: Mill N 11/0/293	Lime Kiln; lime mud from precoat filter fed to kiln, clean mill water used as shower water on precoat filter; contaminated evaporator condensate used in the lime mud clarifier; foul condensate from evaporators is steam stripped, and gases from steam stripper burned in the kiln; all other NCG's burned in NCG thermal oxidizer. Kiln design firing rate: 400 tons CaO/day. PM emissions from kiln are controlled with an ESP Operating conditions during sampling (average): Production rate: 301 tons CaO/day Lime mud solids content: 74% Natural gas firing rate: 75,700 scfm Hot end temp: 2,141°F Cold end temp: 406°F Exit TRS: 2.6 ppm Exit O_2: 11%	Lime kiln ESP outlet	Flue gas/heated Summa® canister	Acetaldehyde, acrolein, acetone, methanol, methyl mercaptan, carbon tetrachloride, chlorobenzene, chloroform, dichloromethane, 1,2-dichloroethane, 1,2-dichloroethylene, dimethyl disulfide, dimethyl sulfide, 1,1,1-trichloroethane, n-hexane, 1,1,2-trichloroethane, 1,2,4-trichlorobenzene, tetrachloroethylene, styrene, toluene, m,p-xylene, o-xylene, benzene, terpenes, methyl isobutyl ketone, methyl ethyl ketone
			Flue gas/National Council Midget Impinger Train	Acetaldehyde, methanol, formaldehyde, acetone, methyl ethyl ketone

TABLE B-1. (continued)

Test site/dates	Process unit and control system description	Sampling location	Sample type/method	Pollutants/parameters measured
Westvaco-Covington, VA 2/5/90 - 4/02/93	Five sources were tested for methanol and acetone on a yearly basis over a 3-year period. These sources include: 1) No. 1 C-E DCE Recovery Furnace (cascade DCE); black liquor solids firing rate = 4.6 million lb/day at 72% solids; PM emissions controlled with wet-bottom ESP, 2 chambers w/ 5 fields per chamber; each ESP chamber has separate stack 2) Smelt dissolving tanks associated with the No. 1 DCE recovery furnace; furnace has two SDT's; emissions from SDT vents controlled with identical venturi scrubbers. Prior to 11/91, "unstripped" evaporator condensates were used as part of the SDT scrubbing liquor. After 11/91, condensate used in the scrubbers was routed through a stripper (for methanol removal) before being sent to the scrubbers. (3) No. 2 Gotaverken NDCE recovery furnace; black liquor solids firing rate = 6 million lb/day at 75% solids; PM emissions controlled with a dry-bottom ESP; ESP has two chambers, 3 fields per chamber. (4) Smelt dissolving tank associated with the No. 2 NDCE recovery furnace; emissions from SDT vent controlled with venturi scrubber. (5) BLOx system: single-stage air-sparging system, uncontrolled vent.	Outlet of DCE recovery furnace ESP-2 separate stacks Outlets of both of the No. 1 DCE recovery furnace SDT scrubbers No. 2 NDCE recovery furnace ESP outlet Outlet of the No. 2 NDCE recovery furnace SDT scrubber BLOx system vent	Flue gas/modified NIOSH Method 2000	Methanol, acetone
Louisiana-Pacific, Samoa, CA 12/13/90	Smelt dissolving tank	SDT exhaust stack	Flue gas/EPA combined metals train	Arsenic, total chromium, copper, cadmium, mercury, selenium, beryllium, nickel, lead, manganese
			Flue gas/CARB 425	Hexavalent chromium
Simpson Paper Co., Eureka, CA 12/10/90	Babcock and Wilcox DCE Recovery Furnace (Firing rate: 60 tons BLS/day) equipped with an ESP and wet scrubber.		Flue gas/EPA combined metals train	Arsenic, total chromium, copper, cadmium, mercury, selenium, beryllium, nickel, lead, mangansese
			Flue gas/CARB 425	Hexavalent chromium

TABLE B-1. (continued)

Test site/dates	Process unit and control system description	Sampling location	Sample type/method	Pollutants/parameters measured
Simpson Paper Co., Anderson, CA 11/27/90	Lime kiln		Flue gas/EPA combined metals train	Arsenic, total chromium, copper, cadmium, mercury, selenium, beryllium, nickel, lead, manganese
			Flue gas/CARB 425	Hexavalent chromium
Gaylord Container Corp., Antioch, CA 11/14/90	NDCE Recovery Furnace equipped with a dry bottom ESP. Unknown PM return system.		Flue gas/EPA combined metals train	Arsenic, total chromium, copper, cadmium, mercury, selenium, beryllium, nickel, lead, manganese
			Flue gas/CARB 425	Hexavalent chromium

TABLE B-2. SUMMARY OF COMBUSTION SOURCES TEST PROGRAM AT INTERNATIONAL PAPER MILLS

Process unit and control system description[a]	Sampling location	Sample type	Method Used to Measure Pollutants												
			Metals[b]	Total hydro-carbons	Ammonia	Hydro-chloric acid	Speciated organics[c]	Sulfuric acid mist	Total reduced sulfur	Nitrogen oxides	Carbon monoxide	Sulfur dioxide	Aldehydes and ketones[d]	Dioxins and furans[e]	Oxygen
Moss Point Mill, Moss Point, MS Test dates: 04/28/93-05/15/93															
Nos. 1 & 2 recovery furnaces	Combined stack outlet	Flue gas	M29			M26	M18				M10	M6			
No. 3 East recovery furnace	Outlet	Flue gas	M29			M26	M18					M6			
No. 3 West recovery furnace	Outlet	Flue gas	M29			M26	M18				M10	M6			
Nos. 1 and 2 smelt dissolving tanks	Scrubber outlet	Flue gas			Mod M6		M18								
	Scrubber solution	Liquid					M25D								
Black liquor oxidation tank	Vent	Flue gas					M18						RTI DNPH		
Lime kiln	Scrubber outlet	Flue gas	M29				M18								
	Scrubber outlet	Liquid					M25D								
Androscoggin Mill, Jay, ME Test dates: 07/09/93-07/23/93															
Recovery furnace	Combined stack	Flue gas				M26	M18								
No. 1 smelt dissolving tank	Scrubber outlet	Flue gas					M18					M6			
	Weak wash scrubber solution	Liquid					M25D								
A-side lime kiln	Scrubber outlet	Flue gas					M18								

TABLE B-2. (continued)

Process unit and control system description[a]	Sampling location	Sample type	Method Used to Measure Pollutants												
			Metals[b]	Total hydro-carbons	Ammonia	Hydro-chloric acid	Speciated organics[c]	Sulfuric acid mist	Total reduced sulfur	Nitrogen oxides	Carbon monoxide	Sulfur dioxide	Aldehydes and ketones[d]	Dioxins and furans[e]	Oxygen
Erie Mill, Erie PA Test dates: 08/26/93-08/31/93															
Smelt dissolving tank	Scrubber outlet, East	Flue gas		M25A	Mod M6		M18								
Recovery furnace		Flue gas		M25A		M26	M18								
Lime Kiln	Venturi scrubber	Flue gas		M25A			M18				M10				
Pine Bluff Mill, Pine Bluff, AR Test dates: 03/19/93-04/06/93															
No. 3 recovery furnace	Outlet	Flue gas	M29			M26	M18	NCASI 106			M10	M6			
No. 4 recovery furnace	Outlet	Flue gas	M29			M26	M18	NCASI 106			M10	M6			
No. 3 smelt dissolving tank	Outlet	Flue gas			Mod. M6		M18								
No. 4 smelt dissolving tank	Outlet	Flue gas					M18								
	Scrubber liquor	liquid					M25D								
Black liquor oxidation tank	Vent	Flue gas					M18		M16				RTI DNH		
Lime kiln	Scrubber outlet	Flue gas	M29				M18				M10				
Georgetown Mill, Georgetown, SC Test dates: 12/02/92-12/15/92 & 01/08/93-01/13/93															
No. 2 Recovery furnace	Scrubber inlet	Flue gas	M29											M23	
	Scrubber outlet	Flue gas	M29											M23	
	Scrubber liquid	Liquid					M25D								

TABLE B-2. (continued)

Process unit and control system description[a]	Sampling location	Sample type	Method Used to Measure Pollutants												
			Metals[b]	Total hydro-carbons	Ammonia	Hydro-chloric acid	Speciated organics[c]	Sulfuric acid mist	Total reduced sulfur	Nitrogen oxides	Carbon monoxide	Sulfur dioxide	Aldehydes and ketones[d]	Dioxins and furans[e]	Oxygen
No. 2 Smelt dissolving tank	Scrubber outlet	Flue gas		M25A	Mod. M6		M18		M16						
	Scrubber solution	Liquid					M25D & M305								
Black liquor oxidation system	First stage vent	Flue gas		M25A			M18		M16				RTI DNPH		
	Second stage vent	Flue gas		M25A			M18		M16						
No. 1 lime kiln	Scrubber outlet	Flue gas	M29	M25A			M18							M23	
	Scrubber solution	Liquid					M25D								
No. 2 lime kiln	Scrubber outlet	Flue gas		M25A			M18								
	Scrubber solution	Liquid					M25D & M305								
Ticonderoga Mill, Ticonderoga, NY Test dates: 10/21/93-11/03/93															
Recovery furnace		Flue gas	M29	M25A		M26	M18		M16	M7E	M10	M6	RTI DNPH	M23	M3A
Smelt dissolving tank	Vent	Flue gas		M25A	Mod. M6		M18		M16						
		Liquid					M25D & M305								
Lime kiln		Flue gas	M29	M25A			M18		M16	M7E	M10			M23	M3A
		Liquid					M25D								
Riverdale Mill, Selma, AL Test dates: 09/09/93-09/24/93															
No. 3 lime kiln	Scrubber outlet	Flue gas					M18								
		Process liquid					M25D								
No. 1 recovery furnace		Flue gas				M26	M18								
No. 2 recovery furnace		Flue gas				M26	M18								

TABLE B-2. (continued)

Process unit and control system description[a]	Sampling location	Sample type	Method Used to Measure Pollutants												
			Metals[b]	Total hydro-carbons	Ammonia	Hydro-chloric acid	Speciated organics[c]	Sulfuric acid mist	Total reduced sulfur	Nitrogen oxides	Carbon monoxide	Sulfur dioxide	Aldehydes and ketones[d]	Dioxins and furans[e]	Oxygen
No. 1 smelt dissolving tank	Scrubber outlet	Flue gas					M18		M16						
No. 2 smelt dissolving tank	Scrubber outlet	Flue gas					M18		M16						
Natchez Mill, Natchez, MS Test dates: 03/04/93-04/01/93															
No. 7S recovery furnace	ESP outlet	Flue gas	M29												
No. 7N recovery furnace	ESP outlet	Flue gas	M29												
No. 4 smelt dissolving tank	Scrubber outlet	Flue gas			Mod. M6		M18								
		Scrubber liquid					M25D								
No. 5 smelt dissolving tank	Scrubber outlet	Flue gas			Mod. M6		M18								
		Scrubber liquid					M25D								
No. 6 smelt dissolving tank	Scrubber outlet	Flue gas			Mod. M6		M18								
		Scrubber liquid					M25D								
No. 7S smelt dissolving tank	Scrubber outlet	Flue gas			Mod. M6		M18								
		Process liquid					M25D								
Black liquor oxidation	Vent	Flue gas					M18						RTI DNPH		
		Process liquid					M25D								
No. 3 lime kiln	Scrubber outlet	Flue gas	M29												

TABLE B-2. (continued)

Process unit and control system description[a]	Sampling location	Sample type	Method Used to Measure Pollutants												
			Metals[b]	Total hydro-carbons	Ammonia	Hydro-chloric acid	Speciated organics[c]	Sulfuric acid mist	Total reduced sulfur	Nitrogen oxides	Carbon monoxide	Sulfur dioxide	Aldehydes and ketones[d]	Dioxins and furans[e]	Oxygen
Mobile Mill, Mobile, AL Test date: 04/28/93-05/17/93															
No. 1 recovery furnace	Stack A outlet	Flue gas	M29			M26	M18				M10	M6			
No. 1 recovery furnace	Stack B outlet	Flue gas	M29			M26	M18				M10	M6			
No. 4 recovery furnace	Outlet	Flue gas	M29			M26	M18				M10	M6			
No. 1 smelt dissolving tank	Scrubber outlet	Flue gas			Mod. M6		M18								
No. 4 smelt dissolving tank	Scrubber outlet	Flue gas			Mod. M6		M18								
Black liquor oxidation	Vent	Flue gas					M18						RTI DNPH		
No. 1 lime kiln	Scrubber outlet	Flue gas	M29				M18								
No. 3 lime kiln	Scrubber outlet	Flue gas	M29				M18								

[a]See Table 3 for average process conditions during the tests of each process unit.
[b]See Table 4 for the specific metals analyzed by Method 29 at each mill.
[c]See Table 5 for the volatile organic compounds speciated with Method 18 testing at each mill. See Table 6 for the volatile organic compounds speciated by liquid injection using either M25D or M305.
[d]See Table 7 for the aldehydes and ketones analyzed by the RTI DNPH procedure at each mill.
[e]See Table 8 for the dioxin and furan congeners analyzed using Method 23.

TABLE B-3. AVERAGE PROCESS DATA FROM EMISSIONS TESTING AT INTERNATIONAL PAPER MILLS

Mill/process unit	Pulp production rate, ADTUBPH	Gas exit temperature, °F	Gas exit flow rate, dscfh	Gas moisture content, % v/v	Gas oxygen content, % v/v	Gas carbon dioxide content, % v/v
Moss Point Mill						
Nos. 1 & 2 recovery furnaces stack outlet	17.44	329	7,085,664	32	12.0	9.2
No. 3 East recovery furnace outlet	16.72	242	1,744,008	34	13.2	8.0
No. 3 West recovery furnace outlet	16.72	255	1,758,572	35	13.2	7.8
Nos. 1 & 2 smelt dissolving tank scrubber outlet	17.44	133	1,542,931	16	20.8	0.0
Black liquor oxidation vent	32.29	179	1,750,248	50	20.9	0.0
Lime kiln scrubber outlet	34.32	162	1,664,038	33	8.2	18.0
Androscoggin Mill						
Recovery furnace	40.35	191	19,143,100	18	9.0	12.5
No. 1 smelt dissolving tank	16.24	175	384,656	40	20.9[a]	0.0
A-side lime kiln	21.86	162	972,759	31	12.0	17.7
Erie Mill						
Smelt dissolving tank scrubber outlet, east	31.25	148	677,982	23	20.9	0.0
Recovery furnace	31.25	358	6,165,520	19	7.5	13.5
Lime kiln venturi scrubber outlet	31.25	161	1,393,670	31	10.4	12.1
Pine Bluff MIll						
No. 3 recovery furnace outlet	15.17	335	4,350,588	28	9.6	9.5
No. 4 recovery furnace outlet	48.65	317	4,250,988	42	7.1	11.2
No. 3 smelt dissolving tank scrubber outlet	16.71	152	498,800	25.37	20.83	0.0
No. 4 smelt dissolving tank scrubber outlet	51.92	141	1,834,000	20.93	20.80	0.0
Black liquor oxidation tank vent	69.17	182	1,613,403	53.40	20.80	0.0
Lime kiln scrubber outlet	32.25	165	1,210,800	32	8.3	11.5

TABLE B-3. (continued)

Mill/process unit	Pulp production rate, ADTUBPH	Gas exit temperature, °F	Gas exit flow rate, dscfh	Gas moisture content, % v/v	Gas oxygen content, % v/v	Gas carbon dioxide content, % v/v
Georgetown Mill						
No. 2 recovery furnace scrubber inlet	46.13	348	9,600,000[b]	36.16	9.80	10.08
No. 2 recovery furnace scrubber outlet	36.88	160	9,600,000[b]	30.55	6.68	14.08
No. 2 smelt dissolving tank scrubber outlet	61.15	175	502,200[c]	51.0	20.8	0.0
Black liquor oxidation system first stage vent	70.98	180	505,000	51.0	20.8	0.0
Black liquor oxidation system second stage vent	70.98	150	547,000	25.0	20.8	0.0
No. 1 lime kiln scrubber outlet	58.2	153	1,010,750	0.0	7.08	17.13
No. 2 lime kiln scrubber outlet	60.05	159	1,665,000	31.1	12.5	15.3
Ticonderoga Mill						
Recovery furnace	17.42	408	3,990,000	24.5	5.5	12.0
Smelt dissolving tank	28.90	127	466,500	25.1	20.8	0
Lime kiln	29.80	160	694,500	25.3	6.25	20.25
Riverdale Mill						
No. 3 lime kiln scrubber outlet	58.50	169	771,000	38	5.5	19.0
No. 1 recovery furnace	23.30	380	6,228,416	21	5.2	10.7
No. 2 recovery furnace	28.50	358	9,015,060	25	8.2	11.2
No. 1 smelt dissolving tank scrubber outlet	23.30	161	634,572	32	20.8[a]	0.0
No. 2 smelt dissolving tank scrubber outlet	28.50	174	531,800	41	20.8[a]	0.0
Natchez Mill						
No. 7S recovery furnace ESP outlet	23.59	365	3,876,000	25	7.6	12.2
No. 7N recovery furnace ESP outlet	23.59	368	4,356,000	25	5.9	13.9
No. 4 smelt dissolving tank scrubber outlet	7.69	144	982,000	21	20.8	0.0
No. 5 smelt dissolving tank scrubber outlet	7.69	145	780,000	21	20.8	0.0
No. 6 smelt dissolving tank scrubber outlet	18.47	152	766,000	27	20.8	0.0
No. 7S smelt dissolving tank scrubber outlet	11.80	151	1,137,000	22	20.8	0.0
Black liquor oxidation vent	33.80	176	978,000	47	20.8	0.0
No. 3 lime kiln scrubber outlet	19.13	166	1,632,000	30	9.7	9.0

TABLE B-3. (continued)

Mill/process unit	Pulp production rate, ADTUBPH	Gas exit temperature, °F	Gas exit flow rate, dscfh	Gas moisture content, % v/v	Gas oxygen content, % v/v	Gas carbon dioxide content, % v/v
Mobile Mill						
No. 1 recovery furnace Stack A outlet	33.20	337	3,712,152	32	3.9	14.8
No. 1 recovery furnace Stack B outlet	32.67	337	3,828,036	34	7.2	13.0
No. 4 recovery furnace outlet	13.41	304	3,533,208	31	10.2	9.7
No. 1 smelt dissolving tank scrubber outlet	31.92	158	946,140	31	20.8	0.0
No. 4 smelt dissolving tank scrubber outlet	14.83	161	270,960	33	20.8	0.0
Black liquor oxidation vent	59.95	174	717,729	44	20.8	0.0
No. 1 lime kiln scrubber outlet	17.42	165	686,832	35	7.5	16.3
No. 3 lime kiln scrubber outlet	18.58	170	497,652	40	6.1	16.9

[a]This oxygen content was assumed in the test report.
[b]The gas flow rate was estimated based on historical data; it was not measured during the emission test.
[c]The gas flow rate was measured a few days after the emissions sampling.

TABLE B-4. METALS ANALYZED BY EPA METHOD 29 IN TESTING AT INTERNATIONAL PAPER MILLS[a]

Metals	Mills[b]			
	Moss Paint and Mobile	Pine Bluff	Georgetown	Natchez
Antimony	✓	✓	✓	✓
Arsenic	✓	✓	✓	✓
Beryllium	✓	✓	✓	✓
Cadmium	✓	✓	✓	✓
Chromium	✓	✓	✓	✓
Cobalt	✓	✓	✓	✓
Copper		✓		✓
Lead	✓	✓	✓	✓
Manganese	✓		✓	
Mercury	✓	✓	✓	✓
Nickel	✓	✓	✓	✓
Selenium	✓	✓	✓	✓
Silver		✓		

[a]The same metals were analyzed for in all samples at a particular mill, regardless of the process unit.
[b]No metals testing was conducted at the Androscoggin, Erie, and Riversdale Mills. At Ticonderoga, the metals compounds were not identified.

TABLE B-5. VOLATILE ORGANIC COMPOUNDS SPECIATED BY METHOD 18 IN TESTING AT INTERNATIONAL PAPER MILLS[a]

Volatile organic compounds	Mills							
	Moss Point	Androscoggin	Erie	Pine Bluff and Natchez	Georgetown	Ticonderoga	Mobile	Riverdale
Acetaldehyde	✓	✓	✓					✓
Acetone	✓	✓	✓			✓	✓	✓
Acetone/Acrolein[b]				✓	✓			
Acetophenone	✓	✓	✓	✓	✓	✓	✓	✓
Acrolein		✓	✓			✓	✓	✓
Benzene	✓	✓	✓	✓	✓	✓	✓	✓
Carbon tetrachloride	✓	✓	✓	✓	✓	✓	✓	✓
Catechol	✓							
Chlorobenzene	✓	✓	✓	✓	✓	✓		✓
Chloroform	✓	✓	✓	✓	✓	✓	✓	✓
m-Cresol						✓		
o-Cresol	✓	✓	✓	✓	✓	✓	✓	✓
Cumene	✓	✓	✓		✓		✓	✓
1,2-Dichlorethane	✓					✓		
1,2-Dichloroethylene	✓					✓		
Dimethyl disulfide	✓	✓	✓	✓	✓	✓	✓	✓
Dimethyl sulfide	✓	✓	✓	✓	✓	✓	✓	✓
Ethanol	✓	✓	✓	✓	✓		✓	✓
Ethyl benzene	✓	✓	✓			✓	✓	✓
Ethyl benzene/ m,p-cresol[b]				✓	✓			
Hexachlorocyclopentadiene	✓			✓	✓	✓		
Hexachloroethane				✓	✓	✓		
Hexachloroethane/ m,p-cresol[b]	✓	✓	✓				✓	
n-Hexane	✓	✓	✓	✓	✓	✓	✓	✓
Isopropyl alcohol	✓	✓	✓			✓	✓	✓
Isopropyl alcohol/acrolein				✓	✓			
Methanol	✓	✓	✓			✓	✓	✓
Methanol/acetaldehyde				✓	✓			
Methly chloroform	✓	✓	✓	✓	✓	✓	✓	✓
Methylene chloride	✓	✓	✓	✓	✓	✓	✓	✓
Methyl ethyl ketone	✓	✓	✓	✓	✓	✓	✓	✓
Methyl isobutyl ketone	✓	✓	✓	✓	✓	✓	✓	✓
Methyl mercaptan	✓	✓	✓	✓	✓	✓	✓	✓
Napthalene/a-terpineol[b]	✓	✓	✓	✓	✓		✓	
Phenol	✓	✓	✓		✓	✓	✓	✓
a-Pinene					✓	✓		
b-Pinene					✓	✓		
a-Terpineol						✓		

TABLE B-5. (continued)

Volatile organic compounds	Mills							
	Moss Point	Androscoggin	Erie	Pine Bluff and Natchez	Georgetown	Ticonderoga	Mobile	Riverdale
Terpenes	✓	✓	✓	✓			✓	✓
Toluene	✓	✓	✓	✓	✓	✓	✓	✓
Trichlorobenzene	✓					✓		
1,1,2-trichloroethane	✓	✓	✓	✓	✓	✓	✓	✓
Trichloroethylene	✓	✓	✓	✓	✓		✓	✓
m,p-Xylene	✓	✓	✓	✓	✓	✓	✓	✓
o-Xylene	✓	✓	✓	✓	✓	✓	✓	✓

[a]All samples analyzed by Method 18 at a particular mill were analyzed for the same compounds, regardless of the process unit from which the samples were collected.
[b]Compounds listed together were coelutants.

TABLE B-6. VOLATILE ORGANIC COMPOUNDS SPECIATED BY LIQUID INJECTION TESTING AT INTERNATIONAL PAPER MILLS

Volatile organic compounds	Mills[a]							
				Georgetown		Ticonderoga		
	Moss Point	Androscoggin	Pine Bluff and Natchez	Recovery furnace & lime kiln No. 1	Smelt dissolving tank & lime kiln No. 2	Smelt dissolving tank	Lime kiln	Riverdale
Total VOC		✓		✓		✓	✓	
Total chlorine				✓		✓	✓	
Total hydrocarbons				✓		✓	✓	
Acetaldehyde	✓		✓					✓
Acetone	✓		✓					✓
Acetone/Acrolein[b]								
Acetophenone	✓		✓					✓
Acrolein			✓					✓
Benzene	✓		✓					✓
Carbon tetrachloride	✓		✓					✓
Carbon tetrachloride/ benzene[b]					✓	✓		
Catechol	✓		✓					
Catechol/Naphthalene[b]					✓	✓		
Chlorobenzene	✓		✓		✓	✓		✓
Chloroform	✓		✓					✓
Chloroform/n-Hexane[b]					✓	✓		
m,p-Cresol			✓		✓	✓		
o-Cresol	✓		✓		✓	✓		✓
Cumene	✓							
Cumene/a-terpineol[b]								✓
1,2-Dichloroethane	✓		✓		✓	✓		
1,2-Dichloroethylene	✓		✓		✓	✓		
Dimethyl disulfide	✓		✓					✓
Dimethyl sulfide	✓		✓					✓
Ethanol	✓		✓					✓
Ethyl benzene	✓		✓		✓	✓		✓
Ethyl benzene/ m,p-Cresol[b]								
Hexachlorocyclo-pentadiene	✓		✓		✓	✓		✓
Hexachloroethane			✓					
Hexachloroethane/ m,p-Cresol[b]	✓							✓
n-Hexane	✓		✓					✓
Isopropyl alcohol	✓		✓					✓

TABLE B-6. (continued)

Volatile organic compounds	Mills[a]							
				Georgetown		Ticonderoga		
	Moss Point	Androscoggin	Pine Bluff and Natchez	Recovery furnace & lime kiln No. 1	Smelt dissolving tank & lime kiln No. 2	Smelt dissolving tank	Lime kiln	Riverdale
Isopropyl alcohol/ Acrolein[b]								
Methanol	✓		✓					✓
Methanol/ Acetaldehyde[b]								
Methyl chloroform	✓		✓					✓
Methylene chloride	✓		✓		✓	✓		✓
Methyl ethyl ketone	✓		✓		✓	✓		✓
Methyl isobutyl ketone	✓		✓		✓	✓		✓
Methyl mercaptan	✓		✓					✓
Naphthalene			✓					
Naphthalene/ a-Terpineol[b]	✓							✓
Phenol	✓				✓	✓		✓
a-Pinene					✓	✓		
b-Pinene					✓	✓		
a-Terpineol			✓		✓	✓		
Terpenes	✓		✓					✓
Toluene	✓		✓					✓
Trichlorobenzene	✓		✓		✓	✓		
1,1,2-Trichloroethane	✓		✓		✓	✓		✓
Trichloroethylene	✓		✓					✓
m,p-Xylene	✓		✓		✓	✓		✓
o-Xylene	✓		✓		✓	✓		✓

[a]No liquid samples were collected from the recovery furnaces, lime kilns, smelt disolving tanks, or black liquor oxidation tanks at the Mobile and Erie Mills.
[b]Compounds listed together were coelutants.

TABLE B-7. ALDEHYDE AND KETONE COMPOUNDS ANALYZED BY RTI DNPH PROCEDURE IN TESTING AT INTERNATIONAL PAPER MILLS

Aldehydes and ketones	Mills[a]	
	Moss Point, Pine Bluff, Natchez, and Mobile	Georgetown
Acetaldehyde	✓	✓
Acetone	✓	✓
Acetophenone	✓	✓
Acrolein	✓	✓
Benzaldehyde	✓	✓
Crotonaldehyde	✓	
Cyclohexanone	✓	
Formaldehyde	✓	✓
Methyl ethyl ketone	✓	✓
Methyl isobutyl ketone	✓	✓
Pentanal	✓	
Propionaldehyde	✓	✓

[a]No aldehyde and ketone testing was conducted at the Androscoggin, Erie, and Riverdale Mills. The aldehyde and ketone compounds analyzed by this method at the Ticonderoga Mill were not identified.

TABLE B-8. DIOXIN AND FURAN COMPOUNDS ANALYZED BY METHOD 23 AT INTERNATIONAL PAPER MILLS[a]

2378	- Tetrachlorodibenzo-p-dioxin (TCDD)
12378	- Pentachlorodibenzo-p-dioxin (PeCDD)
123478	- Hexachlorodibenzo-p-dioxin (HxCDD)
123678	- Hexachlorodibenzo-p-dioxin (HxCDD)
123789	- Hexachlorodibenzo-p-dioxin (HxCDD)
1234678	- Heptachlorodibenzo-p-dioxin (HpCDD)
	Octachlorodibenzo-p-dioxin (OCDD)
2378	- Tetrachlorodibenzofuran (TCDF)
12378	- Pentachlorodibenzofuran (PeCDF)
23478	- Pentachlorodibenzofuran (PeCDF)
123478	- Hexachlorodibenzofuran (HxCDF)
123678	- Hexachlorodibenzofuran (HxCDF)
234678	- Hexachlorodibenzofuran (HxCDF)
123789	- Hexachlorodibenzofuran (HxCDF)
1234678	- Heptachlorodibenzofuran (HpCDF)
1234789	- Heptachlorodibenzofuran (HpCDF)
	Octachlorodibenzofuran (OCDF)
Total	Tetrachlorodibenzo-p-dioxins (TCDD)
Total	Pentachlorodibenzo-p-dioxins (PeCDD)
Total	Hexachlorodibenzo-p-dioxins (HxCDD)
Total	Heptachlorodibenzo-p-dioxins (HpCDD)
Total	Tetrachlorodibenzofurans (TCDF)
Total	Pentachlorodibenzofurans (PeCDF)
Total	Hexachlorodibenzofurans (HxCDF)
Total	Heptachlorodibenzofurans (HpCDF)

[a]This is a list of congeners from the analysis at the Georgetown Mill. At Ticonderoga, the congeners were not identified.

B.5 REFERENCES FOR APPENDIX B

1. Roy F. Weston, Inc. Hazardous Air Pollutant Emission Inventory for Mill C. Prepared for National Council of the Paper Industry for Air and Stream Improvement, Inc. Revised Draft Final. October 11, 1993. Volume 1, Section 4.6.

2. Roy F. Weston, Inc. Hazardous Air Pollutant Emission Inventory for Mill H. Prepared for National Council of the Paper Industry for Air and Stream Improvement, Inc. Draft Final. December 8, 1993. Volume 1, Sections 4.4-4.6.

3. Roy F. Weston, Inc. Texas Emissions Speciation Study Emission Test Results: Champion International Corp., Lufkin, Texas. Prepared for Texas Paper Industry Environmental Committee. Report No. 06848-001-001. January 1993. Volume 2.

4. Roy F. Weston, Inc. Texas Emissions Speciation Study Emission Test Results: Champion International Corp., Sheldon, Texas. Prepared for Texas Paper Industry Environmental Committee. Report No. 06848-001-001. January 1993. Volume 3.

5. Roy F. Weston, Inc. Texas Emissions Speciation Study Emission Test Results: Inland-Orange, Orange, Texas. Prepared for Texas Paper Industry Environmental Committee. Report No. 06848-001-001. January 1993. Volume 4.

6. Roy F. Weston, Inc. Texas Emissions Speciation Study Emission Test Results: Simpson-Pasadena, Pasadena, Texas. Prepared for Texas Paper Industry Environmental Committee. Report No. 06848-001-001. January 1993. Volume 5.

7. Roy F. Weston, Inc. Texas Emissions Speciation Study Emission Test Results: Temple-Inland, Inland, Texas. Prepared for Texas Paper Industry Environmental Committee. Report No. 06848-001-001. January 1993. Volume 6.

8. Roy F. Weston, Inc. Hazardous Air Pollutant Emission Inventory for Mill G. Prepared for National Council of the Paper Industry for Air and Stream Improvement, Inc. Draft Final. May 20, 1994. Volume 1, Section 4.7.

9. Roy F. Weston, Inc. Hazardous Air Pollutant Emission Inventory for Mill D. Prepared for National Council of the Paper Industry for Air and Stream Improvement, Inc. Draft Final. December 13, 1993. Volume 1, Sections 4.4 and 4.6.

10. Letter and attachments from Hershey, R., Potlatch Corp., to Thorvig, L., Minnesota Pollution Control Agency. July 28, 1989. Transmittal of a copy of Interpoll Laboratories Report No. 9-2757 containing the results of the April 1989 emissions test on the No. 8 recovery furnace at Potlatch Corp.'s Cloquet, Minnesota mill.

11. Roy F. Weston, Inc. Hazardous Air Pollutant Emission Inventory for Mill J. Prepared for National Council of the Paper Industry for Air and Stream Improvement, Inc. Draft Final. December 30, 1993. Volume 1, Section 4.2.

12. Roy F. Weston, Inc. Hazardous Air Pollutant Emission Inventory for Mill K. Prepared for National Council of the Paper Industry for Air and Stream Improvement, Inc. Draft Final. April 19, 1994. Volume 1, Section 4.3.

13. Roy F. Weston, Inc. Hazardous Air Pollutant Emission Inventory for Mill L. Prepared for National Council of the Paper Industry for Air and Stream Improvement, Inc. Draft Final. March 17, 1994. Volume 1, Section 4.8.

14. Roy F. Weston, Inc. Hazardous Air Pollutant Emission Inventory for Mill M. Prepared for National Council of the Paper Industry for Air and Stream Improvement, Inc. Draft Final. May 25, 1994. Volume 1, Section 4.2.

15. Letter and attachments from McGaughey, J., Radian Corporation, to Pearson, J., Federal Paper Board Company, Inc. June 11, 1991. Transmittal of HCl emissions test data for the May 1991 emissions test of the Nos. 2 and 3 recovery furnaces at Federal Paper Board's Augusta, Georgia mill.

16. Letter and attachments from Reitter, A., Consolidated Papers, Inc., to Telander, J., EPA/ISB. June 22, 1993. Transmittal of Interpoll Laboratories Report No. 1-3472E containing the results of the December 1991 emissions tests on the Nos. P2 and P3 power and R1 recovery boiler stacks at the Consolidated Paper Kraft Plant in Wisconsin Rapids, Wisconsin.

17. Letter and attachment from Reimer, D., Boise Cascade Paper, to Telander, J., EPA/ISB. June 13, 1993. Transmittal of Source Test Report: Boise Cascade Paper, International Falls, Minnesota, Recovery Boiler Noncriteria Pollutant Tests, February 5-9, 1991, Volume II.

18. Roy F. Weston, Inc. Hazardous Air Pollutant Emission Inventory for Mill N. Prepared for National Council of the Paper Industry for Air and Stream Improvement, Inc. Draft Final. April 26, 1994. Volume 1, Section 4.2.

19. Letter and attachments from Morse, T., Westvaco Bleached Board Division to Telander, J., EPA/ISB. September 22, 1993. Transmittal of the complete emissions test report from the February 1990 methanol emissions test and the emission test summaries from the April 1990 to April 1993 methanol emissions tests for the Nos. 1 and 2 recovery furnaces, the Nos. 1 and 2 SDT's, the BLO unit, and the lime kiln at Westvaco's Covington, Virginia mill.

20. Roy F. Weston, Inc. Hazardous Air Pollutant Emissions Study. Prepared for International Paper. Moss Point, Mississippi. September 1993.

21. Roy F. Weston, Inc. Hazardous Air Pollutant Emissions Study. Prepared for International Paper. Androscoggin Mill. Jay, Maine. November 1993.

22. Roy F. Weston, Inc. Hazardous Air Pollutant Emissions Study. Prepared for International Paper. Erie, Pennsylvania. December 1993.

23. Roy F. Weston, Inc. Hazardous Air Pollutant Emissions Study. Prepared for International Paper. Pine Bluff, Arkansas. July 1993.

24. Roy F. Weston, Inc. Hazardous Air Pollutant Emissions Study. Prepared for International Paper. Georgetown, South Carolina. July 1993.

25. Roy F. Weston, Inc. Hazardous Air Pollutant Emissions Study. Prepared for International Paper. Ticonderoga, New York. April 1993.

26. Roy F. Weston, Inc. Hazardous Air Pollutant Emissions Study. Prepared for International Paper. Riverdale Mill. Selma, Alabama. December 1993.

27. Roy F. Weston, Inc. Hazardous Air Pollutant Emissions Study. Prepared for International Paper. Natchez, Mississippi. September 1993.

28. Roy F. Weston, Inc. Hazardous Air Pollutant Emissions Study. Prepared for International Paper. Mobile, Alabama. September 1993.

29. Air Toxics Emission Inventory Report, California Kraft Pulp Mills. Prepared for California Air Resources Board. Sacramento, California. Undated. Appendix B: Pooled Air Toxics Source Test Program for Kraft Pulp Mills. The appendix contains the November 14, 1990 emissions test report for Gaylord Container Corp., Antioch, California; the November 27, 1990 emissions test report for Simpson Paper Company, Anderson, California; the December 10, 1990 emissions test report for Simpson Paper Company, Fairhaven, California; and the December 13, 1990 emissions test report for Louisiana Pacific Corp., Samoa, California.

30. Roy F. Weston, Inc. Field Evaluation of the Heated Summa® Canister Source Sampling Train. Recovery Furnace, Mill D. Prepared for the National Council of the Paper Industry for Air and Stream Improvement, Inc. July 1993.

31. Roy F. Weston, Inc. Field Evaluation of the Heated Summa® Canister Source Sampling Train. Smelt Dissolving Tank Vent, Mill C. Prepared for the National Council of the Paper Industry for Air and Stream Improvement, Inc. July 1993.

32. Environmental Protection Agency. Test Methods, Proposed Method 308. Code of Federal Regulations. Title 40, Chapter I, Subchapter C, Part 63, Appendix A (draft). December 17, 1993.

33. Memorandum from Harris, V., MRI, to the project file. July 15, 1996. Revisions to Champion Sheldon and Champion Lufkin Metals Emissions Data.

34. Memorandum from Harris, V., MRI, to the project file. May 16, 1996. Review Findings of Metals Blank Data.

35. Memorandum from Toney, M., EPA/EMB, to Telander, J., EPA/ISB. November 24, 1993. Enhanced Monitoring for Pulp and Paper Combustion Sources.

36. Shanklin, S., L. Cone., and S. Steinsberger, Entropy Environmentalists, Inc. Report: Evaluation of HCl Measurement Techniques at a Hazardous Waste Incinerator. Prepared for the U. S. Environmental Protection Agency, Research Triangle Park. EPA Contract No. 68-02-4442. EPA/ORD/AREAL. Undated.

37. Someshwar, A., A Study of Kraft Recovery Furnace Hydrochloric Acid Emissions. National Council of the Paper Industry for Air and Stream Improvement, Inc. New York. Technical Bulletin No. 674. August 1994. pp. 11-13.

38. Memorandum from Toney, M., EPA/EMC, to Telander, J. EPA/MICG. January 13, 1995. Costs of Enhanced Monitoring for Pulp and Paper Combustion Sources.

39. HAP CEMS Data Base, Emission Measurement Technical Information Center Bulletin Board, Office of Air Quality Planning and Standards Technology Transfer Network, U. S. Environmental Protection Agency, Research Triangle Park, North Carolina.

40. U. S. Environmental Protection Agency. Proposed Revisions to Federal Regulations for Hazardous Waste Combustion Devices. Code of Federal Regulations. Title 40, Chapter I, Subchapter C, Part 264/265, Subpart O and Part 266, Subpart H. April 19, 1996.

41. Steinsberger, S., W. Buynitzky, W. DeWees, J. Knoll, M. Midgett, and M. Hartman, U. S. EPA Evaluation of Hazardous Organic Mass Emission Rate (HOMER) System for Measurement of Hazardous Organic NESHAP Emissions. Submitted for presentation at the 86th Annual meeting of the Air and Waste Management Association. Unpresented. June 1993.

42. Leavy, J., Analytical Sciences. Hazardous Air Pollutant Testing at Natchez Mill by Merlab. International Paper Technical Service Report No. 93.558T. Program/Project No. 3271-14. June 16, 1993.

43. Draft Technical Support Document for HWC MACT Standards. Volume IV: Compliance with the Proposed MACT Standards. Appendix G. Performance Specification 8A--Performance Specification for Total Hydrocarbon CEMS. U. S. Environmental Protection Agency. Office of Solid Waste and Emergency Response. February 1996.

44. U.S. Environmental Protection Agency. Test Methods, Proposed Performance Specifications 101 and 102. Code of Federal Regulations. Title 40, Chapter C, Proposed Part 64, Appendix A. October 22, 1993.

45. U.S. Environmental Protection Agency. Test Methods, Proposed Method 318. Measurement of Gaseous Formaldehyde, Phenol, and Methanol Emissions by FTIR Spectroscopy. Code of Federal Regulations. Title 40, Chapter C, Part 63, Appendix A (draft). February 3, 1995.

46. U.S. Environmental Protection Agency. Test Methods, EPA Protocol for the Use of Extractive Fourier Transform Infrared (FTIR) Spectrometry in Analyses of Gaseous Emissions from Stationary Industrial Sources. Code of Federal Regulations. Title 40, Chapter C, Part 63, Appendix A (draft). February 3, 1995.

47. Memorandum from Segall, R., EPA/EMB, to Copland, R., EPA/SDB. October 14, 1992. Final Report on Emission Measurement and Continuous Monitoring for Medical Waste Incinerator Study.

48. U.S. Environmental Protection Agency. Test Methods, Method 29. Code of Federal Regulations. Title 40, Chapter I, Subchapter C, Part 60, Appendix A. April 25, 1996.

49. U. S. Environmental Protection Agency. Test Methods. Code of Federal Regulations. Title 40, Chapter I, Subchapter C, Part 60, Appendix A. July 1, 1993.

50. Telecon. Harris, V., MRI-NCO, with Hosenfeld, J., MRI-KCO. May 26, 1995. Costs of EPA Method 26 stack tests.

51. Peterson, M., B. Pate, E. Rickman, R. Jayanty, F. Wilshire, and J. Knoll. Validation of a Test Method for the Measurement of Methanol Emissions from Stationary Sources. Journal of the Air and Waste Management Association (Pittsburgh). 45:3-11. January 1995.

52. Telecon. Harris, V., MRI, with Peterson, M., Research Triangle Institute. May 24, 1995. Costs of Methanol Sampling Train (MST).

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16. ABSTRACT

National emission standards for hazardous air pollutants (NESHAP) are being proposed for the pulp and paper industry under authority of Section 112(d) of the Clean Air Act as amended in 1990. This technical support document provides technical data and information such as industry and equipment descriptions, analyses of effectiveness and costs of emission control systems, and estimates of environmental impacts of emission control options, that were used in the development of the proposed NESHAP for chemical recovery combustion sources at kraft and soda pulp mills. A NESHAP for noncombustion sources in the pulp and paper industry is being developed concurrently, and information on these sources is contained in separate documents.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS		b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
black liquor black liquor oxidation chemical recovery combustion source hazardous air pollutants kraft pulp mill	lime kiln particulate matter recovery furnace smelt dissolving tank soda pulp mill	NESHAP air pollution control pulp and paper mills	

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