ERG No. 3689.00.001.001

ELECTRICITY FRAMEWORK 5 YEAR REVIEW – CONTROL TECHNOLOGIES REVIEW

Final Report

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January 21, 2009

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LIST OF ACRONYMS

AA	advanced adiabatic
A/C	air-to-cloth
A/F	air-to-fuel
AFBC	atmospheric fluidized bed combustion
AHPC	Advanced Hybrid Particulate Collector
BACT	best available control technology
BACTEA	Best Available Control Technology Economically Achievable
BFB	bubbling fluidized bed
CAES	compressed air energy storage
CAMD	Clean Air Markets Division
CASA	Clean Air Strategic Alliance
CCS	carbon capture and storage
CFB	circulating fluidized bed
CI	carbon injection
Cl	chlorine
CO	carbon monoxide
CO_2	carbon dioxide
CTL	coal-to-liquids
DLE	dry low emissions
DLN	dry low NO_x
DOE	U.S. Department of Energy
EIA	Energy Information Administration
EPA	U.S. Environmental Protection Agency
EPRI	Electric Power Research Institute
ERG	Eastern Research Group
ESP	electrostatic precipitator
FF	fabric filter
FGD	flue gas desulfurization
HC	hydrocarbon
HCl	hydrochloric acid
Hg	mercury
HRSG	heat recovery steam generator
IFGR	induced flue gas recirculation
IGCC	integrated gasification combined cycle
ISCCS	Integrated Solar Combined Cycle System
kg	kilogram
kW	kilowatt
LAER	lowest achievable emission reductions
LNB	low-NO_x burner
LTMC	low temperature mercury control
MJ	megajoule
MMBtu	million British thermal units
MW	megawatt

LIST OF ACRONYMS (Continued)

NH_3	ammonia
NO	nitric oxide
NO_2	nitrogen dioxide
NO_x	nitrogen oxides
NSCR	non-selective catalytic reduction
NSR	New Source Review
OEM	Original Equipment Manufacturer
OAQPS	Office of Air Quality Planning and Standards
OFA	overfire air
OH	hydroxyl
PAC	Powdered Activated Carbon
PC	pulverized coal
PAFC	Phosphoric Acid Fuel Cell
PFBC	pressurized fluidized bed combustion
PM	particulate matter
ppmv	parts per million by volume
ppmvd	parts per million by volume dry
PRB	Powder River Basin
psi	pounds per square inch
PV	photovoltaic
RACT	reasonably available control technology
RBLC	RACT/BACT/LAER Clearinghouse
SCR	selective catalytic reduction
SD/FF	spray dryer absorber with fabric filter
SNCR	selective non-catalytic reduction
SO_2	sulfur dioxide
SRC	Solvent Refined Coal
TAC	total annual costs
TCI	total capital investment
TCR	total capital resources
ULNB	ultra low NO_x burners
WESP	wet electrostatic precipitator
WI	wet injection
ZEPPs	zero emissions power plants

1.0 INTRODUCTION

This document presents the results of a determination of the Best Available Technology Economically Achievable (BACTEA) for multiple greenfield build electricity generating technologies for use in the Province of Alberta, Canada, as requested by the Clean Air Strategic Alliance (CASA). Existing retrofit technologies are not assessed in this document.

The BACTEA analysis was conducted for control technologies used to reduce the emissions of four pollutants: nitrogen oxides (NO_x), sulfur dioxide (SO_2), particulate matter (PM), and mercury (Hg). Additionally, the energy requirements for any control technologies analyzed were identified, and the resulting greenhouse gas emissions were estimated. This document also discusses future technologies, control techniques, and the use of alternative fuels applicable to electric generating units.

The BACTEA determination was conducted for utility boilers and combustion turbines that are 25 megawatts (MW) or greater in size. The determination was also conducted for various fuels burned. Table 1-1 presents a matrix of pollutants analyzed, fuels burned, and combustion devices that are presented in this document.

Table 1-1. Pollutant BACTEA Analyses for Equipment/Fuel Combinations

Fuel Type	Boiler	Turbine
Coal	PM, NO_x, SO_2, Hg	Not applicable
Natural Gas	Not applicable	NO_x

This study does not include analyzing newly constructed boilers or turbines burning fuel oil. This decision was made because the cost of oil would make it an unlikely source for fueling new power plants, and because permits reviewed in the U.S. indicate that no new construction of boilers or turbines burning fuel oil are currently planned. Additionally, a study of electric generation capacity in Alberta does not project any new gas-fired boilers being built in the next 10 years.[1]

1.1 Methodology for Determining BACTEA

The BACTEA analyses followed procedures in the document “Guideline for Identification of Best Available Control Technology-Economically Achievable (BACTEA) for Ontario Regulation 194/05 “Industry Emissions-Nitrogen Oxides and Sulphur Dioxide”.[2] In summary, a BACTEA analysis consists of the following four steps:

- Identifying applicable control technologies
- Eliminating technically infeasible technologies
- Ranking control technologies
- Determining control costs and emission reductions

A BACTEA analysis is very similar to a Best Achievable Control Technology (BACT) analysis that is conducted for Prevention of Significant Determination (PSD) permits under U.S.

EPA regulations. Under BACT analyses control technologies that have been demonstrated for other industries or processes must be considered in addition to those technologies demonstrated for the specific industry in question. The control technologies considered for BACTEA analyses must be demonstrated for the specific industry and process in question. It is required that BACT analyses include other environmental impacts, for example increases in water pollution, waste disposal, or emissions. These impacts are not considered in a BACTEA analysis other than including any applicable costs in the costs of the controls, such as waste disposal costs.

The following discussion describes the considerations made for each step in the BACTEA process.

1.1.1 *Identification of Control Technologies*

All available control technologies potentially applicable to NO_x, SO_2, PM, or Hg control in boilers or turbines were identified from various data sources. These sources included the following:

- Recent permits issued for utilities using fuels similar to those in Alberta.
- U.S. Environmental Protection Agency's (EPA's) Reasonably Available Control Technology /Best Available Control Technology/Lowest Achievable Emission Reductions (RACT/BACT/LAER) or RBLC clearinghouse.[3]
- The report "Analysis of Combustion Controls for Reducing NO_x Emissions from Coal-fired EGU's in the WRAP Region." [4]
- Vendor information.
- Technical reviews and journal articles (including reports of actual operating experience), such as Power Magazine, Power Engineering, and the technical papers from the 2007 EPRI-EPA-DOE-AWMA Mega Symposium on Power Plant Air Pollution Control.

1.1.2 *Eliminating Technically Infeasible Technologies*

Following identification of the potential control technologies, the list was revised by removing those technologies that were either technically infeasible or not used at comparable facilities in North America. Technologies were removed on the basis of the following:

- Not being appropriate for the operational situations in Alberta (e.g., climate, load, fuel type available).
- Not being proven on a specific fuel.
- Not being commercially available.

1.1.3 *Rank Control Technologies*

Following the elimination of technically infeasible technologies, the remaining technologies were ranked. For the remaining technologies, the maximum and typical removal efficiencies/emission levels were determined, when available, for each of the applicable pollutants. The maximum removal efficiency represents the best performance the technology has been reported to achieve (e.g., from vendor information, permits, previous studies, journal articles, etc). If data are available, a range is provided to account for differences in fuel, pollutant loading in the fuel, and operation among the different sources of data reviewed. In some situations the maximum value might not be realistically achievable because emissions of the pollutant might be low, and any additional control would achieve an incrementally small reduction. The emission level achieved is the level that is typically required in state permits in the United States for the technology and pollutant. In many cases, this number is an average of the permit values from across the United States to account for regional differences in fuel and operation. The typical emission level is expressed as a 30-day rolling average (equivalent to a 720-hour rolling average) and is also expressed in English (pounds per million British thermal units [lb/MMBtu]), and metric (kilograms per megajoule [kg/MJ]) on an input basis, and metric units on a net output basis (kilogram per megawatt-hour [kg/MW-h]).

1.1.4 *Determine Control Costs and Emission Reductions*

After controls were ranked, costs and emission reductions for applying each of the controls were estimated by using model units. Model units were developed to represent the potential variations in operating parameters relevant to CASA (such as size, load, and hours of operation) of new units that will be constructed for the fuel/combustion device combinations presented in Table 1-1. The models were based on information provided in the draft report, "Alberta 10 Year Generation Outlook"[1], information on the U.S. EPA's Clean Air Markets Divisions (CAMD) Web site and the U.S. Department of Energy's (DOE) Energy Information Administration (EIA) Web site, journal articles, and ERG's experience with new boilers and turbines. The parameters representing significant variations that affect the control options, emission limits, and costs (e.g., type of fuel burned, size, load, and combustor type) were incorporated into the models.

Costs

Costs were then calculated for control options using cost algorithms or cost factors. The CUECost computer spreadsheet developed by EPA, other EPA studies (e.g, such as the U.S. EPA's Office of Air Quality Planning and Standards (OAQPS) Control Cost Manual), or regulatory background supporting documentation (e.g., the background documentation for the U.S. EPA's utility boiler standards), recent BACT analyses contained in air permit documentation, and technical journals were all reviewed to obtain cost information. Inputs for the cost algorithms were obtained from the studies previously mentioned, engineering judgment, and input by CASA.

Costs are presented in U.S. dollars ($) and standardized to a base year of 2007. Costs are presented as total capital investment (TCI)—also referred to as total capital resources (TCR)—and total annual costs (TAC). TCI is expressed in dollars, dollars per kilowatt (kW), and dollar per life of the control device ($\text{MW-hr}_{\text{lifetime}}$), and consists of the following:

- Purchased equipment costs – control device costs, auxiliary equipment costs, instrumentation, sales taxes, and freight.
- Direct installation costs – foundations and support, handling and erection, electrical, piping, insulation, and painting.
- Site preparation.
- Working capital.

- Indirect installation costs – engineering, construction and field expenses, contractor fees, start-up, and contingencies.

TAC is expressed in dollars per year and consists of the following:

- Operating costs – raw materials, utilities, waste treatment/disposal, labor, and maintenance.
- Indirect costs – overhead, property taxes, insurance, administrative charges, and capital recovery.

Capital recovery was calculated assuming the control equipment has a 20-year life at an interest rate of 7 percent. Construction labor costs were assumed to be $46/hour and operating labor costs were assumed to be $40/hour. Labor rates were based on input from CASA and recent publications specific to labor rates in Alberta, Canada and represent an average of rates for various types of skilled workers.[6] The elements of TCI and TAC were incorporated into the estimates included in this report when applicable to the control technology analyzed and also when available. A more detailed discussion of the cost sources used is provided in Sections 2.0 and 3.0, as well as Appendix B.

Emission Reductions

Emission reductions for each of the model facilities were calculated by applying the control effectiveness for each control technology being analyzed to the baseline emissions from the emission source (boiler or turbine). Baseline emissions of the pollutants of interest were calculated for each model by applying the emission level associated with the baseline control technology that is expected at newly constructed units (i.e., the controls that would be in place at a newly constructed unit). Except for NO_x control, the baseline level of control for all models was assumed to be uncontrolled. The energy consumed by control option auxiliary equipment (e.g., running motors, fans, pumps) and the resulting potential greenhouse gas emissions were also calculated.

1.2 Document Organization

Section 2.0 presents the BACTEA for boilers and Section 3.0 presents the BACTEA for combustion turbines. Section 2.0 is further divided into BACTEA determinations for NO_x, SO_2, PM, and Hg controls, while Section 3.0 addresses only NO_x controls. Each section discusses the relevant control technologies, ranks them on their effectiveness, and provides cost and emission reduction estimates from applying the technology. Section 4.0 presents an evaluation of future technologies, fuels, and control technologies.

References

1. Alberta 10 Year Outlook. Prepared by AMEC Americas Limited for Alberta Electric System Operator. August 18, 2006.

2. http://www.ene.gov.on.ca/envision/AIR/regulations/5169e.pdf.

3. http://cfpub1.epa.gov/rblc/htm/bl02.cfm.

4. Analysis of Combustion Controls for Reducing NO_x Emissions from Coal-fired EGU's in the WRAP Region. Final Report. August 4, 2005.

5. Wage Summary: Construction – Alberta 2007-2011. Prepared by Construction Labour Relations. August 25, 2008.

2.0 BACTEA FOR BOILERS

This section presents the BACTEA analyses conducted for boilers. Sections 2.1 through 2.4 show the analyses specific to NO_x, SO_2, PM, and Hg, respectively. Each section discusses the following topics: the control technologies identified for controlling the pollutant of interest, the control technologies eliminated from further consideration, the ranking of the remaining technologies, and the estimation of costs and emission reductions. Section 2.5 presents the impacts of using combinations of controls to reduce all the pollutants of interest. This section also discusses how the BACTEA analyses may be affected, given other fuels (such as petroleum coke, bitumen/asphaltene, syn gas, and refinery gas), and discusses the possible form that a regulation or permit could take, given the performance levels of the control technologies.

For the costs and emission reduction analyses conducted for boilers, model units were developed to represent the potential variations in operating parameters of new units that will be constructed for the fuel/combustion device combinations presented in Table 1-1. For boilers, the parameters that represent significant variations that affect the control options, emission limits, and costs are fuels burned, size ranges, and combustor types. Table 2-1 summarizes the model units developed for this analysis.

Table 2-1. Model Units for New Boilers

Model	Fuel	Class	Size Range (MW)	Combustor
1	Coal	Small	25-100	Tangential
2				Wall-fired
3		Medium	100-500	Tangential
4				Wall-fired
5		Large	500-1,000	Tangential
6				Wall-fired

Fuels Burned

The Alberta 10 year outlook generation report indicates that the majority of new electrical generating capacity in Alberta will be for coal-fired units. The Alberta 10 year outlook report also indicates that turbines firing natural gas, rather than gas-fired boilers, are projected to be built in Alberta.[1] The report does not project that any new oil-fired or gas-fired boilers will be built. Other information, such as the U.S. EPA's CAMD database and journal articles, indicate that new oil-fired boilers are unlikely to be built in North America. Consequently, the analysis in this report is limited to coal-fired units.

Size Ranges

The model boilers are subdivided into ranges to represent the sizes that could potentially be constructed. For coal- and gas-fired units, these model sizes are small (25 to 100 MW), medium (100 to 500 MW), and large (500 to 1,000 MW). These size ranges are based on information in the CAMD database and state permits and are on a net output basis. The cost and emission reductions analyses were conducted using the highest of each of the ranges: 100 MW, 500 MW, and 1,000 MW.

Combustor Types

A review of the CAMD database[2] indicates that approximately half the boilers constructed are tangential combustors and half are wall-fired combustors. Consequently, the model units are further subdivided into tangential and wall-fired. The primary effect of this distinction is on the baseline level of NO_x emissions between the different combustor types.

Efficiency and Load

The analyses in this section were conducted assuming new boilers constructed would be of the supercritical variety. A recent report for the EPA documents that supercritical boilers can achieve a typical net efficiency of 38 percent, while subcritical boilers can achieve a net efficiency of 36 percent.[3] Supercritical units have been in operation for a number of years and approximately 11 out of 18 units permitted in the U.S. since 2003 burn powder river basin (PRB) coal.[4] Another category of boilers, ultra-supercritical, has also been developed. These boilers have demonstrated efficiencies of greater than 40 percent, and some have reported efficiencies in the mid- to high- 40 percents.[3] However, the majority of ultra-supercritical boilers have been constructed overseas and only one has been permitted in the U.S. in the last five years burning PRB.[4] Therefore, they were not included in the analyses.

The U.S EPA's CUECost program defaults to a load factor of 65 percent (which accounts for hours of operation and capacity of the utilization). This typical load factor is an average for existing units and new units built. Therefore, it is not appropriate to assign this to a new unit. Based on ERG's experience, new units can typically run at 80 to 90 percent of load for the initial stage of their life. The load factor for the calculations in this chapter was set at 90 percent, assuming the supercritical units would be run as close to maximum load as possible. This assumption might need to be revised based on operating characteristics of new units in Alberta or to encapsulate the operation of the boiler over its lifetime instead of just in the initial years.

Other Factors

Initially, the cost estimates were going to incorporate an escalation factor for operation in cold weather climates. Discussions with vendors indicated that the additional costs would be for more insulation and some supplemental equipment to operate control devices better in colder climates. The vendors added, however, that the additional costs would be minimal compared to the cost of the control technology. Consequently, the analyses do not include an escalation factor for cold weather.[5]

The baseline levels of control for the model units assume that any new boiler constructed would incorporate the most current and best NO_x combustion controls. Consequently, the baseline emission levels incorporate the outlet NO_x levels from the combustion controls. The baseline level of control was assumed to be uncontrolled for SO_2, PM, and Hg. As a result, the costs and emission reduction analyses were conducted only for add-on control technologies.

The majority of permits reviewed were for units under construction. For units already in operation, facilities contacted indicated they were meeting the permit limits. Additionally, all permits reviewed were for facilities located in attainment areas. As a result, their permits reflect the best available control technology (BACT) rather than the lowest achievable emission rate (LAER). Appendix E summarizes operation information for boilers whose permit data were included in analyses for this document.

References

1. Alberta 10 Year Outlook. Prepared by AMEC Americas Limited for Alberta Electric System Operator. August 18, 2006.

2. http://camddataandmaps.epa.gov/gdm/.

3. Environmental Footprints and Costs of Coal Based Integrated Gasification Combined Cycle and PC Technologies. U.S. EPA, July 2006. EPA-430/R-06/006.

4. http://www.sierraclub.com/environmentallaw/coal/plantlist.asp.

5. Teleconferences between Bryan Lange, ERG and Sam Sarkar with Beltran Technologies, Inc., John Como with Siemens Environmental Systems & Services, and Flemming Hanson with Haldor Topsoe Inc., September 2008.

2.1 NO_x BACTEA

2.1.1 *Identification of NO_x Control Technologies*

The following technologies were identified from data sources reviewed as potentially applicable for reducing NO_x emissions from boilers:

- ECOTUBE
- ECT/CADM
- Induced Flue Gas Recirculation (IFGR)
- J-POWER ReACT
- Low NO_x Burners (LNB)
- Mobotec ROFA and ROTAMIX
- Non-Selective Catalytic Reduction (NSCR)
- NO_xStar and NO_xStar Plus
- Operational Modifications
- Optimization Software (Neural Networks)
- Oxygen Enhanced Combustion
- Overfire Air (OFA)
- Reburn
- NO_xOut Cascade (SNCR/SCR Hybrid)
- Selective Catalytic Reduction (SCR)
- Selective Non-Catalytic Reduction (SNCR)
- Ultra Low NO_x Burners (ULNB)

Brief descriptions of each technology as well as performance information and/or limitations of the technology are provided in Appendix A.

2.1.2 *Elimination of Infeasible Control Technologies*

The following NO_x control technologies were eliminated from consideration for further analysis:

ECOTUBE, ECT/CADM, J-Power ReACT, NSCR, NO_xStar and NO_xStar Plus, oxygen enhanced combustion

These technologies have not been commercially demonstrated on large boilers. Some have not been demonstrated in the U.S. ECT/CADM is used only on biomass and municipal solid waste units. Therefore, they have been excluded from the analysis.

IFGR

This control has a NO_x reduction efficiency of 20 percent. This effectiveness was considered to be low when compared to other technologies that achieved 40 and up to 90 percent reduction. Additionally, this technology has potential problems with flame instability. Therefore, it was excluded from the analysis.

Mobotec ROFA and ROTAMIX

These technologies are only in operation on a few units in North America and have not been demonstrated on Alberta-type coal. Additionally, they are more applicable to retrofits than new units.

Operational Modifications and Reburn

These technologies apply only to existing units, and were therefore excluded from the analysis.

LNB, ULNB, OFA, and optimization software (neural networks)

For new units, the baseline level of control would include the most up-to-date LNB technology, as well as optimization software. Therefore, these technologies were not considered as control options beyond the baseline level.

2.1.3 *Ranking of Control Technologies*

The remaining NO_x control technologies that were analyzed are SCR, SNCR, and NO_xOut Cascade. Both SCR and SNCR have been widely used, while NO_xOut Cascade is a relatively new technology that combines aspects of both SCR and SNCR. Table 2-2 summarizes the maximum NO_x control efficiency level that each technology can achieve, based on vendor data, permits, and previous EPA reports. Table 2-2 also provides the typical level of control that can be achieved based on requirements in state permits.

Table 2-2. Maximum Achievable and Typical Performance of Best Performing Commercially Demonstrated NO_x Control Technologies

Control Technology	Maximum Achievable Reduction Efficiency	Typical Emission Levels – Input Based (30-day rolling average)		Typical Emission Levels – Net Output Based (kg/MW-h)[f]
		lb/MMBtu	g/MJ	
SCR	86 to 92 percent[a]	0.078[d]	0.0336	0.32
SNCR	14 to 40 percent[b]	0.08, 0.10, 0.12[e]	0.0344, 0.043, 0.052[e]	0.33, 0.41, 0.49[e]
NO_xOut Cascade	65 to 75 percent[c]	0.078[d]	0.0336	0.32

[a] Srivastava, R. et al. Nitrogen Oxides Emission Control Options for Coal-Fired Electric Utility Boilers. Journal of the Air & Waste Management Association v55. September 2005.

[b] EPRI Report 1004727. SNCR Guidelines Update. November 2004. EPRI Project Manager R. Himes.

[c] According to Albanese, V.M., et al. "Hybridization of Urea-SNCR with SCR - A Fit for the Future" ICAC Clean Air Technologies & Strategies Conference Proceedings, 2005, hybrid SCR/SNCR technologies such as NO_xOut Cascade can achieve between 65 and 75 percent reductions in NO_x.

[d] Recent (since November 2004) permit limits have contained NO_x limits ranging from 0.07 to 0.1 lb/MMBtu (30-day rolling average) when using combustion controls in sequence with a SCR. The BACTEA limit reflects the average of six recent permit limits (since October 2004) using SCR and advanced combustion control technologies: Western Farmers Electric Coop Hugo Generating Station; Great Plain Energy Kansas City Power & Light Company IATAN Station Units 1 & 2; Xcel Energy Comanche Station; Omaha Public Power District Nebraska City Station; Wisconsin Public Service Weston Plant. To achieve the typical emission level, SCR would only need to reduce NO_x emissions by 44%. This BACTEA limit was also used for the NO_xOut Cascade control option. The analysis assumes that the NO_xOut Cascade control would perform no better than an SCR system.

[e] Since SNCR could not achieve a limit of 0.078 lb/MMBtu, the 0.08 lb/MMBtu limit was used to estimate emission reductions for SNCR. This is based on the maximum achievable reduction efficiency (40%) of SNCR from the baseline NO_x emission rate for a 100 MW unit. NO_X removal efficiencies from SNCR decrease as the unit's size increases. A removal efficiency of 14% was used for the 1,000 MW unit, and a removal efficiency of 27% was used for the 500 MW unit.

[f] Output levels were calculated assuming boilers have a net efficiency of 38%, based on a value for supercritical units quoted in a report for the EPA.[1]

SCR

The maximum reduction efficiency was obtained from a study of NO_x emission control performance.[2] The typical emission limit was determined from U.S. permits. Permit data from 2004 to 2007 indicates that states are requiring SCR to achieve emission levels ranging between 0.07 to 0.10 lb/MMBtu (0.03 to 0.043 g/MJ) with an average of 0.078 lb/MMBtu (0.033 g/MJ) on a 30-day rolling average basis.[3] At other averaging times, the stringency increases. For example, in one permit issued on July 31, 2008 the boiler was required to meet a NO_x level of 0.035 lb/MMBtu averaged over 365 days.

NO_XOut Cascade

A hybrid SNCR/SCR system was reported by the vendor to achieve an overall NO_x reduction of 65 to 75 percent.[4] Typical emission limits were assumed to be no better than what was achieved by SCR: 0.078 to 0.1 lb/MMBtu (0.03 to 0.043 g/MJ) on a 30-day rolling average basis.

SNCR

On small coal-fired units (i.e., less than 200 MW), SNCR has been demonstrated to achieve NO_x reductions ranging from 25 to 40 percent with acceptable levels of ammonia slip.[5] For larger boilers (i.e., greater than 300 MW), there are numerous challenges associated with applying SNCR. In particular, such boilers' large physical dimensions pose challenges for injecting and mixing the reagent with the flue gas. Pulverized coal-fired units have a much more limited furnace temperature window and poor lateral mixing, conditions that render SNCR less effective in these applications. The maximum efficiency values were determined from a 2004 Electric Power Research Institute (EPRI) report.[5] The typical emission level shown in this report, 0.08 lb/MMBtu (0.034 g/MJ), was based on the 40 percent reduction efficiency (i.e., the maximum level) for a 100 MW unit, because, in order to achieve similar emission levels as typical SCR installations, the maximum SNCR value would have to be used. Lower efficiencies, 27 percent and 14 percent were used for the 500 and 1,000 MW units. SNCR is not expected to be a technically feasible control option for achieving a BACTEA limit for larger-sized units.

2.1.4 *Estimation of Control Costs*

EPA's CUECost modeling tool was used to estimate the cost of installing an SCR or SNCR for NO_x control. CUECost inputs for coal type included parameters specific to Alberta coal as well as defaults for PRB coal. The base year for cost calculations in CUECost is 1998. These costs were scaled up to 2007 using a ratio of the 2007 to 1998 Chemical Engineering cost indices (525.4/388).[6] Costs for NO_xOut Cascade were based on vendor information. Further details of all cost calculations can be found in Appendix B. Costs for individual NO_x controls are summarized in Table 2-3.

Table 2-3. Summary of Costs for NO_x Controls for Tangential and Wall-fired Boilers

	100 MW				500 MW				1000 MW			
	TCI			TAC	TCI			TAC	TCI			TAC
Control	Million $	$/kW	$/MW-hr$_{lifetime}$[1]	Million $	Million $	$kW	$/MW-hr$_{lifetime}$[1]	Million $	Million $	$/kW	$/MW-hr$_{lifetime}$[1]	Million $
SCR	$21.8	$218	1.24	$2.9	$68.1	$136	0.78	$10.2	$114.9	$115	0.68	$18.3
SNCR	$1.6	$16	0.09	$0.4	$3.6	$7	0.04	$1.1	$5.6	$6	0.03	$1.8
NO_xOut Cascade	$5.8	$58	0.33	$1.0	$28.8	$57.5	0.33	$4.1	$57.5	$58	0.33	$8.0

[1] This cost is the total capital investment represented in $/MW over a 20 year lifetime of the control device. It **does not** represent capital recovery, which is the annualized cost of the capital expenditures over the equipment lifetime. Capital recovery incorporates an interest rate factor (7 percent for this analysis) to represent the cost of borrowing to pay for capital expenditures. Capital recovery is a component of total annual costs (TAC).

SCR

CUECost estimates SCR TCI costs based on the following components:

- Reactor housing
- Initial catalyst
- Ammonia (NH_3) or urea storage and injection system
- Additional flue gas handling equipment including ductwork and increased fan size;
- Enhanced air preheater

- Miscellaneous direct costs, including ash handling and water treatment

The TAC for SCR consists of operating and maintenance costs and capital recovery costs. Operating and maintenance costs include those for urea or NH_3 replacement, catalyst replacement and disposal, electricity, steam, labor, and maintenance. Capital recovery was calculated based on a 20-year life for the SCR equipment. However, the expected life may be lower, e.g., 15 years, resulting in a higher capital recovery than is calculated in this analysis.

Previous experience with the CUECost program has shown that the cost algorithms for SCR does not incorporate recent cost increases for labor and materials, and may significantly underestimate costs. Consequently, based on ERG's experience with SCR on other utility units and recent journal articles, [7-11] the SCR TCI estimates were escalated by an additional factor of two to account for the recent economic effects.

SNCR

CUECost was used to estimate the costs of an ammonia-based SNCR control system. The main equipment areas included in the TCI estimates are:

- Reagent receiving area
- Storage tanks and recirculation system
- Reagent injection system, including injectors, pumps, valves, piping, and distribution system; the control system; and air compressors

In addition, NH_3-based SNCR systems use electrically powered vaporizers to vaporize the NH_3 prior to injection. Total annual costs comprise operating labor, reagent, electricity, water, and steam requirements, as well as capital recovery costs.

NO_xOut Cascade

Because very little cost data are available concerning NO_xOut Cascade, these TCI costs are based on an average cost provided by one vendor on a $/kW basis.[12] For NO_xOut Cascade installation, the vendor provided an average cost of 57.5 US$/kW in 2007 dollars.[12] The TCI was calculated by multiplying the MW capacity of each model unit by the $/kW cost. The vendor suggested annual operating costs would be approximately two times that of SNCR. The TAC estimates include the operating costs and capital recovery.

2.1.5 *Estimation of Emission Reduction for Boilers*

The baseline NO_x emission for Alberta subbituminous coal-fired boilers was set to 0.14 lb/MMBtu (0.06 g/MJ) for tangentially fired units. This baseline reflects the 2003 average controlled NO_x emission rate from 23 tangential units using a modern LNB with both close-coupled and separated OFA as a primary control technology.[2] The baseline NO_x emission rate was also set to 0.14 lb/MMBtu (0.06 g/MJ) for wall-fired units. This baseline reflects the average controlled NO_x emission rate from four wall-fired units using a modern LNB with OFA as a primary control technology.[2] These baseline values assume that any new boiler will already be equipped with the most current LNB and combustion control system. Assuming a 90 percent operating loading (incorporating hours of operation and plant capacity factor) for a new unit, and a 38 percent efficiency in the boiler on a net basis, this results in baseline emissions of 450 metric tons per year for a 100 MW unit, 2,251 metric tons per year from a 500 MW unit, and 4,503 metric tons per year from a 1,000 MW unit.

The removal efficiency required to achieve a typical emission level of 0.078 lb/MMBtu (0.034 g/MJ) for NO_x was determined to be 44.3 percent. The decision to use 0.078 lb/MMBtu (0.034 g/MJ) as the limit was based on recent BACT limits for NO_x in U.S. permits for coal-fired utility boilers that ranged from 0.07 to 0.10 lb/MMBtu (0.03 to 0.043 g/MJ) when using state-of-the-art combustion controls in conjunction with SCR.[2] The required removal efficiency was used to determine the minimum emission reductions expected for each type of NO_x control, if that removal efficiency was technically feasible for a given control. Therefore, it was applicable for SCR and NO_xOut Cascade. For SNCR, which could not achieve the 44.3 percent reduction, the maximum reduction efficiency associated with SNCR (40 percent) was used for the 100 MW unit, the average of the reduction efficiency range (27 percent) was used for the 500 MW unit, and the minimum reduction efficiency associated with SNCR (14 percent) was used for the 1,000 MW unit. This assumption was made because literature sources indicated that SNCR becomes less effective for large units.

2.1.6 *Summary of Costs and Emission Reductions for NO_x Control*

The cost effectiveness of each NO_x control was determined by dividing its TAC by its annual emission reduction. The costs and emission reductions for NO_x controls are summarized in Table 2-4.

Table 2-4. Summary of Emission Reduction and Cost Effectiveness of NO_x Controls

Control	Boiler Size (MW)	Emission Reductions (metric tons/yr)	Cost Effectiveness ($/metric tons removed)
SCR	100	199	$14,600
	500	995	$10,300
	1,000	1,991	$9,200
SNCR	100	180	$2,130
	500	608	$1,730
	1,000	630	$2,920
NO_xOut Cascade	100	199	$5,040
	500	995	$4,150
	1,000	1,991	$4,030

It is important to note the limitations of this analysis, however. First, very little cost data were available for NO_xOut Cascade and only one piece of data was found relating to its emission reduction efficiency. Therefore, the reported cost effectiveness of this control option might not be as reliable as those for SCR or SNCR.

References

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2.2 SO_2 BACTEA

2.2.1 *Identification of SO_2 Control Technologies*

The following technologies were identified from data sources reviewed as potentially applicable for reducing SO_2 emissions from boilers:

- Activated Carbon Beds
- Circulating Dry Scrubber
- Coal Cleaning
- Furnace/Duct Reagent Injection
- J-Power ReACT
- Limestone Injection Gas Scrubbing
- Pahlman Process, EnviroScrub
- Spray Dryer Absorber with Fabric Filter (SD/FF)
- Wet Flue Gas Desulfurization (FGD)

Brief descriptions of each technology as well as performance information and/or limitations of the technology are provided in Appendix A.

2.2.2 *Elimination of Infeasible Control Technologies*

The following SO_2 control technologies were eliminated from consideration for further analysis:

Activated carbon beds, circulating dry scrubber, J-Power ReACT, limestone injection gas scrubbing, and Pahlman Process/EnviroScrub

These technologies have not been commercially demonstrated on large boilers. Some of the technologies have not been demonstrated on utility boilers in the U.S. Therefore, they have been excluded from the analysis.

Coal Cleaning, Furnace/Duct Reagent Injection

Physical coal cleaning is estimated to reduce the amount of SO_2 in the coal by 10 to 50 percent. Dry sorbent injection and furnace sorbent injection are once-through technologies designed as retrofit technologies to achieve 50 percent removal of SO_2. Other scrubber systems are capable of achieving a removal efficiency of much greater than 50 percent. Therefore, these were excluded from further analyses.

2.2.3 *Ranking of Control Technologies*

The remaining SO_2 control technologies that were analyzed in this report are SD/FF and FGD. Both have been widely used throughout North America. Table 2-5 summarizes the maximum SO_2 control efficiency/emission level that each technology can achieve, based on vendor data and previous EPA reports. Table 2-5 also provides the typical level of control that can be achieved based on requirements in state permits.

Table 2-5. Typical Performance of Best Performing Commercially Demonstrated SO_2 Control Technologies

Control Technology	Maximum Achievable Reduction Efficiency	Typical Emission Levels – Input Based (30-day rolling average)		Typical Emission Levels – Net Output Based (kg/MW-h)[c]
		lb/MMBtu	g/MJ	
Wet FGD	80 to 98 percent[a]	0.085[b]	0.037	0.35
Dry FGD (Spray Dryer/ Fabric Filter)	93 percent[b]	0.11[b]	0.047	0.45

[a] U.S. EPA AP-42 indicates that wet FGD systems can achieve 80 to 98 percent reductions depending on the inlet coal sulfur content. Recent units using low sulfur coals, such as PRB have been permitted with control efficiencies greater than 90 percent.

[b] Missouri Department of Natural Resources document listing public comments and State responses regarding issued Permit to Construct for Associated Electric *Cooperative,* Inc. Norborne Power Plant. February 22, 2008.

[c] Output levels were calculated assuming boilers have a net efficiency of 38 percent, based on a value for supercritical units quoted in a report for the EPA.[1]

Dry FGD (SD/FF)

SD/FFs have been shown to achieve 93 percent control efficiency over the long-term and have been demonstrated on large boilers burning low sulfur coals.[2] Recent (since October 2004) permits on low sulfur PRB coals have attained SO_2 limits ranging from 0.065 to 0.12 lb/MMBtu (0.028 to 0.052 g/MJ) on a 30-day rolling average when a SD/FF was selected for the control technology. The BACTEA limit reflects the average of six recent permit limits using dry FGD/spray dryer adsorbers: (Longleaf Energy Associates (Georgia); City Utilities of Springfield, Southwest Power Station (Missouri); Omaha Public Power District (Nebraska); Wisconsin Public Service Corp. Weston Unit 4 (Wisconsin); LS Power, Sandy Creek Energy Station (Texas); Black Hills Power and Light Wygen 2 (Wyoming); and Xcel Energy Comanche Station (Colorado)).[3]

Wet FGD

Wet FGD systems have been demonstrated on large supercritical pulverized coal (PC) boilers firing coal. U.S. EPA's AP-42 emission factor compilation indicates that wet FGD systems can achieve 80 to 98 percent reductions depending on the inlet coal sulfur content.[4] Recent (since October 2004) permits on low-sulfur PRB coals have contained SO_2 limits ranging from 0.065 to 0.1 lb/MMBtu (0.028 to 0.043 g/MJ) on a 30-day rolling average when a wet FGD technology was selected for the control technology. The BACTEA limit reflects the average of three recent permit limits using wet FGD technology (Hugo Unit 2 (Oklahoma); City Public Service of San Antonia Spruce 2 (Texas); and Intermountain Power Service Corporation (Utah)).[3]

2.2.4 *Estimation of Control Costs*

EPA's CUECost modeling tool was used to estimate the cost of installing a SD/FF or wet FGD system for SO_2 control. CUECost inputs for coal type included parameters specific to Alberta coal as well as defaults for PRB coal.

Three different unit capacities were considered: 100 MW, 500 MW, and 1,000 MW. The combustor design was determined to not have an effect on SO_2 emissions. Consequently, costs were estimated for three size capacities.

The base year for cost calculations in CUECost is 1998. These costs were scaled up to 2007, using a ratio of the 2007 to 1998 Chemical Engineering cost indices (525.4/388).[5] Further details for all cost calculations can be found in Appendix B.

Costs for individual SO_2 controls are summarized below in Table 2-6

Table 2-6. Summary of Costs (2007 $US) for SO_2 Control for Boilers

	100 MW				500 MW				1000 MW			
	TCI			TAC	TCI			TAC	TCI			TAC
Control	Million US$	$/kW	$/MW-hr$_{\text{lifetime}}$[1]	Million US$/yr	Million US$	$kW	$/MW-hr$_{\text{lifetime}}$[1]	Million US$/yr	Million US$	$/kW	$/MW-hr$_{\text{lifetime}}$[1]	Million US$/yr
SD/FF	$46.0	$460	2.63	$10.1	$82.1	$164	0.94	$26.1	$143.0	$143	0.82	$47.1
Wet FGD	$1,136	$1,136	6.49	$15.8	$203.0	$406	2.32	$33.0	$302.5	$303	1.73	$50.7

[1] This cost is the total capital investment represented in $/MW over a 20 year lifetime of the control device. It **does not** represent capital recovery, which is the annualized cost of the capital expenditures over the equipment lifetime. Capital recovery incorporates an interest rate factor (7 percent for this analysis) to represent the cost of borrowing to pay for capital expenditures. Capital recovery is a component of total annual costs (TAC).

SD/FF

The annual costs for a spray dryer system include capital recovery and annualized costs of power requirements, labor and maintenance. The capital costs involve five main equipment areas:

- Reagent Feed System - Receiving, Storage, Grinding
- SO_2 Removal - Spray Dryers, Tanks, Pumps
- Flue Gas Handling - Ductwork and I.D. Fan
- Waste / By-product Handling - Disposal, Storage
- Support Equipment - Electrical, Water, Air

The costs for a fabric filter are not included in the SO_2 analysis. In order to avoid double counting these costs, they are discussed in Section 2.3 under PM controls.

Wet FGD

The annual costs for a wet FGD system include capital recovery and annualized costs of power requirements, labor, and maintenance. The capital costs involve five main equipment areas:

- Reagent Feed - Receiving, Storage, Grinding
- SO_2 Removal - Absorbers, Tanks, Pumps
- Flue Gas Handling - Ductwork and I.D. Fan
- Waste / By-product Handling - Dewatering, Disposal/Storage, Washing
- Support Equipment - Electrical, Water, Air

Based on research of FGD systems installed in 2007, ERG found that CUECost significantly underestimates the cost of FGD. These cost shortfalls are a result of recent general inflation not incorporated in the 2007 Chemical Engineering index: supply shortages in several key raw material markets, including structural steel, nickel alloy, and industrial plastic resins markets; and an increase in demand from FGD suppliers. A 2007 Sargent and Lundy report indicates that recent contracts for FGD systems have been signed at prices 300 percent higher than contracts signed in 2002.[6] Further, this report notes that FGD costs increased by 25 percent between 2006 and 2007. Consequently, based on ERG's experience with FGD on other utility units, the FGD TCI estimates were escalated by an additional factor of two to account for the recent economic effects.

2.2.5 *Estimation of Emission Reduction for Boilers*

The baseline SO_2 emission for Alberta subbituminous coal-fired boilers was set as 0.82 lb/MMBtu (0.353 g/MJ) for both tangential units and wall-fired units. This baseline was calculated using the uncontrolled SO_2 emission factor equation from U.S. EPA's AP-42 document.[4] The factor is calculated by multiplying the sulfur content of the fuel by 38. CASA provided coal data from the Sheerness Station, Battle River Station (Unit 5), Genesee Station (Units 1 and 2), and Sundance station (Units 5 and 6).[7] The average sulfur content (0.43 percent weight) and heating value (23.28 MJ/kg or 10,008 Btu/lb) of Alberta subbituminous coal was used in this equation to calculate the baseline emission level for the model units. Assuming a 90 percent operating loading (incorporating hours of operation and plant capacity factor) for a new unit, and a 38 percent efficiency in the boiler on a net basis, this results in baseline emissions of 2,637 metric tons per year for a 100 MW unit, 13,190 metric tons per year from a 500 MW unit, and 26,370 metric tons per year from a 1,000 MW unit.

The removal efficiency required to achieve a typical emission level on a wet FGD of 0.085 lb/MMBtu (0.037 g/MJ) on a 30-day rolling average basis for SO_2 was determined to be 89.6 percent. The decision to use 0.085 lb/MMBtu (0.037 g/MJ) as the limit was based on recent (since October 2004) BACT limits for SO_2 in U.S. permits for coal-fired utility boilers that ranged from 0.065 to 0.12 lb/MMBtu (0.028 to 0.052 g/MJ) on a 30-day rolling average basis.[3] Recent BACT limits for SO_2, where a spray dryer/FF or dry FGD technology is proposed in U.S. permits, have a typical emission level of 0.11 lb/MMBtu (0.047 g/MJ) on a 30-day rolling average basis.[3] The removal efficiency required to achieve this emission level is 86.6 percent. The required removal efficiency was used to determine the minimum emission reductions expected for each type of SO_2 control. Table 2-7 summarizes the emission reduction calculated for each model unit and control option.

Table 2-7. Summary of Costs and Emission Reductions of SO_2 Control

Control	Boiler Size (MW)	Typical control performance	
		Emission Reductions (metric tons/yr)	Cost Effectiveness ($US/metric tons removed)
SD/FF	100	2,280	$4,400
	500	11,400	$2,280
	1,000	22,800	$2,060
Wet FGD	100	2,360	$6,700
	500	11,820	$2,800
	1,000	23,630	$2,140

2.2.6 *Summary of Costs and Emission Reductions for SO_2 Control*

The cost effectiveness of each SO_2 control was determined by dividing its TAC by its annual emission reduction. The costs and emission reductions for SO_2 controls are summarized in Table 2-7.

References

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2.3 PM BACTEA

PM composition and emission levels are a complex function of boiler firing configuration, boiler operation, pollution control equipment, and coal properties. Particulate matter comes in a range of sizes and designations. Particulate matter of 10 µm and smaller is called PM10 and particulate matter of 2.5 µm and under is known as PM2.5 or fine particle. PM can be further classified as either filterable PM or condensable PM. Filterable PM can be measured by traditional filter sampling methods (such as EPA Method 5) and is often classified as PM10 or PM2.5. Vapors and particles less than 0.3 microns pass through the filter. Condensable PM is formed when certain substances in stack gases cool, dilute, and condense (without chemical reaction) as the emitted plume mixes with ambient air. Condensable PM is always classified as PM2.5 or less. The controls discussed in this section are primarily effective in reducing filterable PM. Condensable PM control requires cooling of the stack gas prior to entering the PM control device or using a wet scrubber or FGD system.

2.3.1 *Identification of PM Control Technologies*

The following technologies were identified from data sources reviewed as potentially applicable for reducing filterable PM emissions from boilers:

- Electrostatic Precipitator (ESP)
- Wet Electrostatic Precipitator (WESP)
- Fabric Filter (FF)
- Centrifugal Collectors/Cyclones
- Wet Scrubber

Brief descriptions of each technology as well as performance information and/or limitations of the technology are provided in Appendix A.

2.3.2 *Elimination of Infeasible Control Technologies*

The following PM control technologies were eliminated from consideration for further analysis.

Centrifugal Collectors/Cyclones

Although they are commercially available, cyclones achieve significantly lower reductions than other PM control technologies. Generally, they cannot reduce emissions enough to comply with regulatory standards and are thus deemed technically infeasible.

Wet Scrubbers

Wet scrubbers achieve lower emission reductions than other PM control technologies and have not been used much on large boilers in the last 15 years. Additionally, wet scrubbers can present water pollution concerns.

2.3.3 *Ranking of Control Technologies*

ESPs and FFs are the most commonly used technology for filterable PM control. Additionally, a WESP can be used in combination with an ESP for the removal of smaller sizes of filterable PM and may also reduce condensable PM. These three configuration options for PM control are commercially available and technically feasible for coal-fired utility boilers. Members of CASA indicated that emission levels should be presented on a 30-day rolling average basis. The majority of permit levels are on a 3-hour basis, but three permits were identified on a 30-day basis. Table 2-8 summarizes the removal efficiencies achieved for PM controls based on a definition of PM that includes total filterable particulates and emission levels that correspond to total PM (filterable and condensable).

Table 2-8. Typical Performance of Best Performing Commercially Demonstrated PM Control Technologies

Control Technology	Maximum Achievable Reduction Efficiency	Typical Emission Levels – Input Based (30-day rolling average)		Typical Emission Levels – Net Output Based (kg/MW-h)[d]
		lb/MMBtu	g/MJ	
FF	99.0-99.99[a]	0.03[c]	0.013	0.12
ESP	99-99.5[a]	0.03[c]	0.013	0.12
ESP/WESP	99.0-99.86[b]	0.03[c]	0.013	0.12

[a] EPA, 2007. RACT/BACT/LAER Clearinghouse (RBLC) database. Values are for filterable PM.[2]

[b] BACT Analysis for Ely Energy Center, Nevada. Appendix B of Application for Operating Permit to Construct. October 2007.[5]

[c] Average of permit limits for Sandy Creek Energy Station; and Great Plains Energy Kansas City Power and Light Company IATAN Station Units 1&2. Permits for the Great Plains units indicated limits were for total PM filterable and condensable). However, the Sandy Creek permit did not indicate limits including condensable as well as filterable. For this analysis it is assumed the permits are on the same basis.

[d] Output levels were calculated assuming boilers have a net efficiency of 38 percent, based on a value for supercritical units quoted in a report for the EPA.[1]

Fabric Filter (FF)

Typical new FFs have filterable PM removal efficiencies ranging from 99 to 99.9 percent according to the RBLC database.[2] For example, Mid-American Energy Company[5] in Iowa, estimates a 99.7 percent efficiency for PRB coal. Omaha Public Power District[5] estimates a 99.9 percent PM removal efficiency for similar subbituminous coal. Both facilities use FFs for PM control. Fabric filters may remove some condensable PM if the stack gas entering the filter is cool enough that condensable PM is formed.

Electrostatic Precipitator (ESP)

According to the RBLC database[2], typical new ESPs have filterable removal efficiencies ranging from 99 to 99.5 percent. However, it is important to note that ESPs are generally less effective on low-sulfur coals because the ash has a high resistivity. Because Alberta coal has a similar sulfur content but much greater ash content than low-sulfur PRB coal[3,4], an ESP might not provide optimal PM removal for coal-fired utility boilers in Alberta. ESP are less effective than fabric filters for removing condensable PM.

Electrostatic Precipitator (ESP) and Wet Electrostatic Precipitator (WESP)

Little data are available on WESP removal efficiencies in utility boilers, because there have been only a few applications in the utility industry. The ESP/WESP combination is used on at least one utility boiler in the U.S. for removing condensable PM. According to the RBLC database, Thoroughbred Generating Company, LLC, in Kentucky, proposed a WESP as the BACT for PM with an estimated removal efficiency of 99.0 percent.[2] Ely Energy Center, in Nevada, estimates in their BACT analysis a filterable PM removal efficiency of 99.86 percent for WESPs.[5]

2.3.4 *Estimation of Control Costs*

EPA's CUECost modeling tool was used to estimate the cost of installing an ESP or FF for PM control. CUECost inputs for coal type included parameters specific to Alberta coal as well as defaults for PRB coal. For WESP installation, an average of costs reported in BACT permits was used to estimate cost per kilowatt-hour.

Three different unit capacities were considered: 100 MW, 500 MW, and 1000 MW. The combustor design was determined to not have an effect on the cost of PM controls.

The base year for cost calculations in CUECost is 1998. These costs were scaled up to 2007, using a ratio of the 2007 to 1998 Chemical Engineering cost indices (525.4/388).[6] Further details for all cost calculations can be found in Appendix B. Costs for individual PM controls are summarized in Table 2-9.

Table 2-9. Summary of Costs (2007 $US) for PM Controls for Boilers

	100 MW				500 MW				1000 MW			
	TCI			TAC	TCI			TAC	TCI			TAC
Control	**Million $US**	**$/kW**	**$/MW-hr$_{\text{lifetime}}$[1]**	**Million $US/yr**	**Million $US**	**$kW**	**$/MW-hr$_{\text{lifetime}}$[1]**	**Million $US/yr**	**Million $US**	**$/kW**	**$/MW-hr$_{\text{lifetime}}$[1]**	**Million $US/yr**
FF	$12.6	$126	0.72	$2.1	$40.5	$81	0.46	$7.5	$78.7	$79	0.45	$14.7
ESP	$12.0	$120	0.69	$1.7	$45.6	$91	0.52	$6.7	$84.5	$84	0.48	$12.5
WESP + ESP	$19.8	$198	1.13	$2.5	$84.4	$169	0.96	$10.4	$162.0	$162	0.92	$19.8

[1] This cost is the total capital investment represented in $/MW over a 20 year lifetime of the control device. It **does not** represent capital recovery, which is the annualized cost of the capital expenditures over the equipment lifetime. Capital recovery incorporates an interest rate factor (7 percent for this analysis) to represent the cost of borrowing to pay for capital expenditures. Capital recovery is a component of total annual costs (TAC).

FF (Pulse Jet)

CUECost estimates FF TCI costs based on the following components:

- FF
- Bags
- Ash handling system

- Fans
- Instruments and controls

The TAC for FF consists of operating and maintenance costs and capital recovery costs. Operating and maintenance costs include bag replacement, operating and maintenance labor, and electricity costs.

ESP

CUECost estimates ESP TCI costs based on the following components:

- ESP
- Ash handling system
- Fans
- Instruments and controls

The TAC for ESPs consists of operating and maintenance costs and capital recovery costs. Operating and maintenance costs include operating and maintenance labor, and electricity costs.

WESP/ESP

Because very little data is available concerning WESPs, the TAC for this control technology does not include any fixed or variable costs associated with equipment operation. Only the capital recovery factor is included. WESP and ESP costs are combined in the summary table because WESPs are typically used in conjunction with ESPs.

2.3.5 *Estimation of Emission Reduction for Boilers*

The baseline PM emission for Alberta subbituminous coal-fired boilers was set as 0.22 lb/MMBtu (0.095 g/MJ). This baseline was calculated using the uncontrolled PM emission factor equation from U.S. EPA's AP-42 document.[7] The AP-42 equation is based on EPA Method 5, which measures filterable PM. It does not include condensable PM. Data from stack surveys or emission factors from test reports were not used to calculate baseline emissions because information reviewed did not provide information on units burning Alberta coal.

The AP-42 factor is calculated by multiplying the ash content of the fuel by 10. CASA provided coal data from the Sheerness Station, Battle River Station (Unit 5), Genesee Station (Units 1 and 2), and Sundance Station (Units 5 and 6).[8] The average ash content (19.52 percent weight) and heating value (23.28 MJ/kg or 10,008 Btu/lb) of Alberta subbituminous coal was used in this equation to calculate the PM baseline for the model units. Assuming a 90 percent operating loading (incorporating hours of operation and plant capacity factor) for a new unit, and a 38 percent efficiency in the boiler on a net basis, this results in baseline total filterable PM emissions of 708 metric tons per year for a 100 MW boiler, 3,538 metric tons per year for a 500 MW boiler, and 7,076 metric tons per year for a 1000 MW boiler.

Three recent BACT limits (since 2004) for PM in U.S. permits for coal-fired utility boilers that provided data on a 30-day average were reviewed. The permit levels ranged from

0.024 to 0.04 lb/MMBtu (0.009 to 0.017 g/MJ) for the following units: Sandy Creek Energy Associates Sandy Creek Energy Station; and Great Plains Energy Kansas City Power & Light Company IATAN Station Units 1&2.[2] The limit in 2 of the permits included condensable and filterable PM (Great Plains 1 and 2). The other permit did not specify whether the limit included condensables (Sandy Creek). For all three units a fabric filter is the prescribed PM control technology. The units also use controls (e.g., FGD) downstream of the FF to reduce acid gases. The FF will achieve high removal efficiencies of filterable PM, but not necessarily of condensable PM. Condensable PM removal is dependent on the temperature of the stack gas entering the FF, such that condensables will either remain in vapor form or be cooled sufficiently to be collected in the fabric filter. Additional condensable PM removal is achieved by the downstream controls. It is important to note that the EPA method referenced in the permits, Method 202, is undergoing review by EPA to address potential artifact formation in the sampling train which can lead to inaccurate results.

For this analysis, the PM limits for the 3 permits were averaged together using the assumption they were on the same basis (i.e., all 3 permit limits are for total PM). This results in an average value of 0.03 lb/MMBtu (0.013 g/MJ). It was also assumed that the majority of PM would be filterable PM. This assumption was based on a study done by EPA that reviewed permit limits for recently permitted coal-fired boilers across the U.S.[9] The study showed that the limits for both filterable and condensable PM are often the same or lower than the limits for filterable only, indicating the majority of the PM in the limits are most likely filterable PM. Additionally, the uncontrolled PM value was based upon total filterable PM; therefore, the analysis needed to be on the same basis. To achieve the limit of 0.03 lb/MMBtu (0.013 g/MJ), all three PM control options would need to achieve 86 percent reduction in filterable PM, based on using the calculated uncontrolled PM value of 0.22 lb/MMBtu (0.095 g/MJ), and the assumption that the permit levels represent mostly filterable PM. However, if the condensable portion were considerably more, the required removal efficiency for filterable PM would increase. For example, if the condensable portion were 50 percent of the total PM, then the total filterable PM level would be 0.015 lb/MMBtu (0.007 g/MJ), which would require 92-93 percent reduction in filterable PM. All three control technologies reviewed would be able to achieve this level.

2.3.6 *Summary of Costs and Emission Reductions for PM Control*

The cost effectiveness of each PM control was determined by dividing its TAC by its annual emission reduction. The costs and emission reductions for PM controls are summarized in Table 2-10.

When considering emissions control for PM alone, an ESP is clearly the most cost effective of the three options. It is important to note the limitations of this analysis, however. First of all, efficiency data collected is primarily from units that burn PRB coal, which reportedly has a lower ash content than Alberta subbituminous coal. The higher ash content of Alberta coal, and the resulting higher ash to sulfur ratio, can cause higher resistivity of particulates, making PM capture more difficult with an ESP. Another limitation to consider is that very little cost data were available for WESPs, and only two pieces of data were found relating to their emission reduction efficiency. Therefore, the reported cost effectiveness of this control option might not be as reliable as those for the ESP and FF. Finally, the assumptions made regarding filterable and condensable PM will effect the required performance of the controls analyzed. If the condensable portion is larger than what was assumed in the analysis a higher filterable PM reduction efficiency will be necessary. Additionally, the control of condensable PM has not been analyzed in this section. The impact of downstream controls, such as FGD, would need to be examined.

Table 2-10. Summary of PM Costs and Emission Reduction

Control	**Boiler Size (MW)**	**Typical control performance**	
		Emission Reductions (metric ton/yr)	**Cost Effectiveness ($/metric ton removed)**
FF	100	610	$3,500
	500	3,040	$2,470
	1,000	6,090	$2,420
ESP	100	610	$2,860
	500	3,040	$2,200
	1,000	6,090	$2,050
WESP + ESP	100	610	$4,060
	500	3,040	$3,400
	1,000	6,090	$3,250

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2.4 Hg BACTEA

2.4.1 *Identification of Hg Control Technologies*

Mercury controls are difficult to consider separately from controls for other pollutants, because some of the controls for NO_x, SO_2, and PM can have an effect on Hg emissions. For this reason, Hg controls are listed as a combination of controls for other pollutants, and as controls that may be used specifically for Hg. Few Hg-specific controls have been proven in the long term to be effective for large, coal-fired boilers. The following technologies were identified for the control of Hg, that are potentially applicable to utility boilers:

Multi-pollutant:

- ESP
- FF
- Wet FGD
- SDA
- J-POWER ReACT
- SCR catalysts

Hg-specific:

- Carbon Injection (CI)

Brief descriptions of each technology as well as performance information and/or limitations of the technology are provided in Appendix A.

2.4.2 *Elimination of Infeasible Control Technologies*

The following Hg control technologies were eliminated from consideration for further analysis:

ESP

Although generally effective for PM removal, ESPs show relatively low Hg removal efficiencies (10 percent or less) when compared with other potential Hg controls; therefore, ESPs are deemed infeasible for Hg control.

J-POWER ReACT

This multi-pollutant control technology has not been proven on PRB or Alberta coal. Additionally, there have been no commercial installations in North America. Due to its lack of commercial availability and proven effectiveness, J-POWER ReACT has been removed from the analysis.

SCR Catalysts

Gaseous compounds of Hg^{2+} are generally soluble in water, and they can be absorbed in the aqueous slurry of a wet FGD. Therefore, promoting the oxidation of elemental Hg (Hg^0) to Hg^{2+} upstream can effectively improve Hg removal. Although SCR catalysts can promote this oxidation for bituminous coal, it is thought that improving catalytic oxidation for low rank coals with SCR systems is not possible with the typical chemical composition of the flue gas or a lower catalyst temperature. Because the performance and applicability of SCR systems as a control for Hg for plants burning low rank coals is uncertain, SCR is considered infeasible as an effective Hg control in this analysis.[1]

2.4.3 *Ranking of Control Technologies*

The remaining technologies include FF, SD/FF, Wet FGD (in combination with ESPs and FFs), and CI. Technologies considered in this analysis have been demonstrated on utility boilers in North America, and, where data were available, specifically using PRB or Alberta coal. The exception to this is CI. Performance information from any full-scale operating utility boilers has not been identified. However, EPA and DOE databases and reports indicate that approximately 80 utility boilers in the U.S. are being equipped or are being permitted to use carbon injection (CI). Of these, approximately 60 burn subbituminous or PRB coal, and one burns Canadian subbituminous coal.[2] Consequently, we have included CI as a control technology for Hg. Typical removal efficiencies for the four control technologies considered are listed below.

FF

Mercury removal efficiency for FFs is highly dependent on fuel type.[3] This dependence on fuel type is mainly due to Hg speciation, which is the partitioning of Hg into elemental Hg vapor (Hg^0), ionic Hg (Hg^{2+}), and particulate Hg (Hg_p). PM control devices mostly capture Hg_p and Hg^{2+}, allowing much of the Hg^0 to pass through. Through complex reactions that occur as flue gas is cooled, the higher Cl content in bituminous coal allows more Hg^0 to be converted to Hg^{2+}, which is more easily captured by PM controls. For units that burn bituminous coal, Hg capture efficiency for FFs can be as high as 90 percent. However, for units burning subbituminous coal, which has a Cl content similar to that of most Alberta coals, FFs are typically no more than 70 percent efficient.

SDA/FF

Research has shown that for bituminous coal fired units, Hg removal is as high as 98 percent with the SDA/FF combination. However, with subbituminous coal fired units, Hg removal with this combination can be drastically less—as low as 20 percent[4]. This decrease is likely because in boilers burning subbituminous coals, which have a lower Cl content, most of the gaseous Hg produced is Hg^0 because less Cl is available for oxidation. Additionally, any hydrogen chloride (HCl) in the flue gas will be removed by the SDA, which also results in less Hg^0 oxidation and thus lower removal efficiency in the FF.

Wet FGD

Wet FGD systems are typically installed downstream of PM control devices. Because gaseous compounds of Hg^{2+} are generally soluble in water, they can be absorbed in the aqueous slurry of a wet FGD. Studies have shown that the combination of an ESP with a wet FGD can capture 29 percent or more of Hg from a unit burning subbituminous coal, and 49 to 79 percent of Hg from a unit burning bituminous coal (using a hot-side ESP or cold-side ESP, respectively). The studies also indicate that a combination of a FF and wet FGD can capture as much as 98 percent of the Hg from a unit burning bituminous coal.[5] Higher efficiencies can be attributed to increased oxidation of Hg^0 due to higher Cl content in the coal being burned.

Carbon Injection

Carbon injection is a control technology that is relatively new to utility boiler application. As such, limited data are available on its effectiveness in reducing Hg emissions. Complications arise when evaluating the removal efficiency of CI because it can be highly dependent on the type of PM control being used. Most efficiency data collected for Hg control analysis are based on field tests for units burning subbituminous or PRB coal. Most tests have focused on units with an ESP as the PM control because baseline Hg removal for an FF is typically higher, between 50 and 70 percent.[3] Table 2-11 summarizes the data available for Hg removal by CI for various pollutant control configurations.

Table 2-11. Summary of Hg Control Performance for Subbituminous Coal

Control Configuration[1]	Average Hg Percent Removal without CI (Baseline)	Hg Percent Removal with CI
FF	60[3]	No Data
SDA/FF	24[2]	54[4]
Wet FGD/FF	65[2,3]	No Data
Wet FGD/ESP	16[4]	65[2]

[1] NO_x control impacts on Hg collection efficiency for subbituminous coals have not been sufficiently proven, so are not considered in control configurations for Hg removal.

[2] U.S EPA, Control of Hg Emissions from Coal-fired Electric Utility Boilers. January 2004.

[3] Performance and Cost of Hg and Multipollutant Emission Control Technology Applications on Electric Utility Boilers. U.S. EPA. EPA-600/R-03-110. October 2003.

[4] EPRI Mercury Control Selection Guide: Final Report. September 2006.

Permit information from 6 units burning PRB coal showed Hg emission limits ranging from 15 to 20 lb/TWh gross (6.8 to 9.1 kg/TWh gross).[7] All the permits required the use of CI.

2.4.4 *Estimation of Control Costs*

Estimates for CI costs were obtained from the EPRI Mercury Control Selection Guide[4]. The total capital cost per kW, obtained from the guide, for a carbon additive injection system was $4 (US) with a base year of 2006. These costs were scaled up to 2007, using a ratio of the 2007 to 2006 Chemical Engineering cost indices (525.4/499.6).[5] Total annual costs comprise activated carbon replacement, operating and maintenance costs (including electricity costs), and capital recovery costs. The operating and maintenance and carbon replacement costs were provided for a 500 MW unit. The costs were then scaled to apply to 100 MW and 1,000 MW units based on the size ratio. Further details of all cost calculations can be found in Appendix B.

The costs for FF, SDA/FF, and wet FGD installation have been summarized in previous sections. Costs for CI, the Hg-specific control, are summarized below in Table 2-12 as they apply to each of the other control technologies installed.

Table 2-12. Summary of Costs (2007 $) for CI for Boilers with Other Controls

Control	100 MW		500 MW		1000 MW	
	TCI Million $	TAC Million $/yr	TCI Million $	TAC Million $/yr	TCI Million $	TAC Million $/yr
FF[1]	0.4	0.2	2.1	0.8	4.2	1.7
SDA/FF	0.4	0.2	2.2	0.8	4.4	1.7
ESP/FGD	0.4	0.2	2.2	1.1	4.4	2.1
FF/FGD[1]	0.4	0.2	2.1	0.8	4.2	1.7
ESP	0.44	0.4	2.2	2.0	4.4	4.0

[1] Operating costs were not provided in the EPRI report for CI applied to a FF/FGD system or FF. Operating costs were assumed to be at least equivalent to SD/FF costs. As a result, costs may be overstated for these systems.

2.4.5 *Estimation of Emission Reduction for Boilers*

The uncontrolled Hg emission factors from six units were provided by CASA (ranging from 12.9 to 29.7 kg/TWh). CASA provided coal data from the Sheerness Station, Battle River Station (Unit 5), Genesee Station (Units 1 and 2), and Sundance station (Units 5 and 6).[8] However, these uncontrolled values were developed on an output basis. In order to use this information, it was necessary to estimate the efficiency of each boiler, then determine the emission factors if the units were supercritical and achieved an efficiency of 38 percent. The efficiency of each unit was calculated by dividing the net generation by the thermal input of the coal utilized in the boiler. Both the net generation (GWh) and coal utilization (metric tons) were provided by CASA. The thermal input of the coal utilized was calculated from the higher heating value of the coal (provided by CASA) and converting units to be on the same basis as the net generation. The resulting efficiency was then used to develop a ratio of calculated efficiencies to assumed supercritical efficiency (38 percent), which was then applied to each output based emission factor to calculate an emission factor for supercritical units.

The resulting range of emission factors for a 38 percent efficient unit ranged from 8.6 to 21.7 kg/TWh. An average emission factor was then calculated from all 6 units to be 14.3 kg/TWh. Although the emission factors varied between the units, an average was used incorporating all the units because the ratios of Hg in the flyash to the Hg in the coals were similar, indicating that the only difference between the units was in the Hg content of the coal. Additionally, previous information provided by CASA indicated that the units controlled PM using only an ESP, which data shows is ineffective on its own in reducing vapor Hg emissions. However, the information provided by CASA shows that an average of 26 percent of the total mercury in the coal burned was captured in the fly ash by the ESP's and 74 percent was emitted. To accurately determine the amount of mercury captured by each control device reviewed in this section, it is first necessary to determine the uncollected mercury emissions (mercury in the fly ash plus mercury emissions). The estimation was made by the dividing the average calculated output emission factor by the average percent of Hg emitted (74%), resulting in 19.3 kg-TWh. Assuming a 90 percent operating load (incorporating hours of operation and plant capacity factor) for a new unit, results in an uncontrolled baseline emissions of Hg emissions of 34 lb/yr (15 kg/yr) for a 100 MW boiler, 168 lb/yr (76 kg/yr) for a 500 MW boiler, and 335 lb/yr (152 kg/yr) for a 1,000 MW boiler.

Permit data showed Hg limits ranging from 15 to 20 lb/TWh gross (6.8 to 9.1 kg/TWh gross). To meet these limits, Hg control would need to achieve 54 to 65 percent reduction. However, because Hg control technology is relatively new and data for units burning PRB coal or Alberta coal are not available, it cannot be determined that the permit limits are achievable by the units or are applicable to boilers in Alberta. Consequently, emission reductions were calculated by applying the average Hg reduction efficiency in Table 2-11 to the uncontrolled baseline emissions. Although the efficiencies in Table 2-11 were also not based on Alberta coal, they were based on real tests conducted on subbituminous coal and should provide a better basis for determining the potential emission reductions. Table 2-13 presents the emission reductions and additional costs resulting from applying CI, and the emission reductions from using the other controls only.

2.4.6 *Summary of Costs and Emission Reductions for Hg Control*

The cost effectiveness of CI control applied to units with existing controls was determined by dividing its TAC by its annual emission reduction. The costs and emission reductions for CI are summarized in Table 2-13.

Table 2-13. Summary of CI Costs and Emission Reduction Applied to Units with Other Controls

Other Controls	Boiler Size (MW)	Hg Emission Reduction of Other Controls (kg/yr)	Additional Carbon Injection Impacts	
			Hg Emission Reductions (kg/yr)	Cost Effectiveness ($/kg removed)
FF[1]	100	9	2	$72,800
	500	46	11	
	1,000	91	23	
SDA/FF	100	4	5	$36,800
	500	18	23	
	1,000	37	46	
ESP/Wet FGD	100	2	7	$28,200
	500	12	37	
	1,000	24	75	
FF/Wet FGD[1]	100	10	2	$109,000
	500	49	8	
	1,000	99	15	

[1] Hg reduction information was not available for CI applied to FF or FF/Wet FGD systems. For these controls, a value of 75 percent reduction was applied assuming that because a FF can achieve 50-70 percent reduction, CI applied would be able to achieve at least an additional 15-20 percent reduction. This additional reduction is less than achieved by applying CI to ESP/Wet FGD or SDA/FF.

It is important to note the limitations of this analysis. First, very limited data are available on the performance of the controls analyzed on Hg emissions. Because Hg control is being incorporated into boilers in North America, more information is expected in the near future that will refine the values used in the analysis. Additionally, cost information was obtained from one data source. While it is relatively recent, additional data points would be necessary to verify the values used in the analysis. Cost and efficiency information were not available for all the control device combinations of interest. The operating cost information for CI, in particular, was applied to 100 MW and 1,000 MW units by scaling information for a 500 MW unit. Because fabric filters achieve a high efficiency (when compared to other controls), tests were not conducted using CI on fabric filter control combinations. Consequently, a reduction efficiency of 75 percent was assumed for FF and FF in combination with a wet FGD. Insufficient data are available to determine if this was an appropriate methodology. Consequently, efficiencies and costs were transferred for like technologies. Finally, because CI is a relatively new technology, its effect on Alberta coal or PRB coal has not been proven. Until information is available for these coal types, assumptions for other coals must be used.

References

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2.5 Summary of Boilers BACTEA Results

This section presents the impacts of using combinations of controls to reduce all the pollutants of interest. This section also discusses how the BACTEA analyses may be affected, given other fuels (such as petroleum coke, bitumen/asphaltene, syngas, and refinery gas), and discusses the possible form that a regulation or permit could take, given the performance levels of the control technologies.

2.5.1 *Impacts of Combined Controls*

Costs and emission reductions for 18 different control combination options of NO_x, SO_2, PM, and Hg control were analyzed. The options include the following:

- SCR/Wet FGD/FF
- SCR/Wet FGD/ESP
- SCR/SD/FF
- SCR/Wet FGD/FF/CI
- SCR/Wet FGD/ESP/CI
- SCR/SD/FF/CI
- SNCR/Wet FGD/FF
- SNCR/Wet FGD/ESP
- SNCR/SD/FF
- SNCR/Wet FGD/FF/CI
- SNCR/Wet FGD/ESP/CI
- SNCR/SD/FF/CI
- NO_xOut/Wet FGD/FF
- NO_xOut/Wet FGD/ESP
- NO_xOut/SD/FF
- NO_xOut/Wet FGD/FF/CI
- NO_xOut/Wet FGD/ESP/CI
- NO_xOut/SD/FF/CI

Costs for the combined controls were assumed to be additive based on the information presented in Sections 2.1 through 2.4. Emission reductions for NO_x, SO_2, and PM were assumed to be independent from the controls for other pollutants. For example, based on information presented in Section 2.2, SO_2 emission reductions for the control devices analyzed were assumed to be unaffected by the PM and NO_x control devices. As discussed in Section 2.4, Hg controls are difficult to consider separately from controls for other pollutants, as some of the controls for NO_x, SO_2, and PM have an effect on Hg emissions. Section 2.4 indicates that for the controls analyzed, Hg reduction can occur from SD, FF, and wet FGD. In fact, the only controls currently being used to achieve Hg reduction are these controls with and without carbon injection to enhance the removal of Hg. Hg emission reductions were calculated including the combined effects of these controls with and without carbon injection using the information presented in Section 2.4.

Table 2-14 summarizes the control costs and emission reductions for a 100 MW unit. Table 2-15 summarizes the costs and emission reductions for a 500 MW unit. Table 2-16 summarizes the costs and emission reductions for a 1,000 MW unit.

2.5.2 *Greenhouse Gas Emissions*

Electricity is required to operate each identified control technology. The energy used by each control technology has been analyzed and the amount of greenhouse gas emissions associated with the additional energy usage is presented below.

Nearly every control system requires a fan to convey the exhaust gases through it and to overcome the pressure drop of the control system. In the spray dryer electricity is consumed by the water and solvent pumps. The ESP requires a high voltage static charge to attract particulate and electricity to power a mechanical system to clean the ESP plates.

The SNCR and SCR injection equipment, pumps, compressors, and control systems all require electricity. In addition to electricity requirements, both SNCR and SCR units will reduce the thermal efficiency of a boiler. Table 2-17 below shows the required electricity and the resulting greenhouse gas emissions.

Carbon Capture

Utilities worldwide are exploring ways to mitigate the production of greenhouse gases from the control and combustion device’s carbon capture. For details of the specific methods being explored, see the discussion in Section 3.4.5.

2.5.3 *Effects of Other Fuels Not Analyzed*

Utilities are cofiring alternative fuels in solid coal-fired boilers because such operations offer opportunities to lower fuel costs or to meet possible future mandates requiring renewable fuel usage. Several alternative fuels are discussed below.

Table 2-14. Summary of Control Costs and Emission Reductions for a 100 MW Unit

Option No.	**Control Configuration**	**TCI**		**TAC**	**Annual Emission Reductions**				
		$	$/kW	$/yr	*NO_x (mt)*	*SO_2 (mt)*	*PM (mt)*	*Hg (kg)*	*Total*
1	SCR/Wet FGD/FF	$147,990,000	$1,480	$20,870,000	200	2,360	610	10	3,170
2	SCR/Wet FGD/ESP	$147,420,000	$1,470	$18,740,000	200	2,360	610	2	3,170
3	SCR/SD/FF	$80,370,000	$800	$15,110,000	200	2,360	610	4	3,170
4	SCR/Wet FGD/FF/CI	$147,990,000	$1,480	$20,870,000	200	2,360	610	11	3,170
5	SCR/Wet FGD/ESP/CI	$147,420,000	$1,480	$18,740,000	200	2,360	610	9	3,170
6	SCR/SD/FF/CI	$80,370,000	$810	$15,110,000	200	2,280	610	9	3,090
7	SNCR/Wet FGD/FF	$127,830,000	$1,280	$18,340,000	180	2,360	610	10	3,150
8	SNCR/Wet FGD/ESP	$127,260,000	$1,270	$16,210,000	180	2,360	610	2	3,150
9	SNCR/SD/FF	$60,210,000	$600	$12,590,000	180	2,360	610	4	3,150
10	SNCR/Wet FGD/FF/CI	$127,830,000	$1,280	$18,340,000	180	2,360	610	11	3,150
11	SNCR/Wet FGD/ESP/CI	$127,260,000	$1,280	$16,210,000	180	2,360	610	9	3,150
12	SNCR/SD/FF/CI	$60,210,000	$610	$12,590,000	180	2,280	610	9	3,070
13	NO_xOut/Wet FGD/FF	$131,980,000	$1,320	$18,970,000	200	2,360	610	10	3,170
14	NO_xOut/Wet FGD/ESP	$131,410,000	$1,310	$16,840,000	200	2,360	610	2	3,170
15	NO_xOut/SD/FF	$64,350,000	$640	$13,210,000	200	2,360	610	4	3,170
16	NO_xOut/Wet FGD/FF/CI	$131,980,000	$1,320	$18,970,000	200	2,360	610	11	3,170
17	NO_xOut/Wet FGD/ESP/CI	$131,410,000	$1,320	$16,840,000	200	2,360	610	9	3,170
18	NO_xOut/SD/FF/CI	$64,350,000	$650	$13,210,000	200	2,280	610	9	3,090

mt – metric ton

Table 2-15. Summary of Control Costs and Emission Reductions for a 500 MW Unit

Option No.	**Control Configuration**	**TCI**		**TAC**	**Annual Emission Reductions**				
		$	$/kW	$/yr	*NO_x (mt)*	*SO_2 (mt)*	*PM (mt)*	*Hg (kg)*	*Total*
1	SCR/Wet FGD/FF	$311,610,000	$620	$50,810,000	1,000	11,820	3,060	49	15,880
2	SCR/Wet FGD/ESP	$316,750,000	$630	$43,290,000	1,000	11,820	3,060	12	15,880
3	SCR/SD/FF	$190,720,000	$380	$43,820,000	1,000	11,420	3,060	18	15,480
4	SCR/Wet FGD/FF/CI	$311,610,000	$630	$50,810,000	1,000	11,820	3,060	57	15,880
5	SCR/Wet FGD/ESP/CI	$316,750,000	$640	$43,290,000	1,000	11,820	3,060	49	15,880
6	SCR/SD/FF/CI	$190,720,000	$390	$43,820,000	1,000	11,420	3,060	41	15,480
7	SNCR/Wet FGD/FF	$247,050,000	$490	$41,620,000	610	11,820	3,060	49	15,490
8	SNCR/Wet FGD/ESP	$252,190,000	$500	$34,100,000	610	11,820	3,060	12	15,490
9	SNCR/SD/FF	$126,170,000	$250	$34,630,000	610	11,420	3,060	18	15,090
10	SNCR/Wet FGD/FF/CI	$247,050,000	$500	$41,620,000	610	11,820	3,060	57	15,490
11	SNCR/Wet FGD/ESP/CI	$252,190,000	$510	$34,100,000	610	11,820	3,060	49	15,490
12	SNCR/SD/FF/CI	$126,170,000	$260	$34,630,000	610	11,420	3,060	41	15,090
13	NO_xOut/Wet FGD/FF	$272,230,000	$540	$44,710,000	1,000	11,820	3,060	49	15,880
14	NO_xOut/Wet FGD/ESP	$277,370,000	$550	$37,190,000	1,000	11,820	3,060	12	15,880
15	NO_xOut/SD/FF	$151,340,000	$300	$37,710,000	1,000	11,420	3,060	18	15,480
16	NO_xOut/Wet FGD/FF/CI	$272,230,000	$550	$44,710,000	1,000	11,820	3,060	57	15,880
17	NO_xOut/Wet FGD/ESP/CI	$277,370,000	$560	$37,190,000	1,000	11,820	3,060	49	15,880
18	NO_xOut/SD/FF/CI	$151,340,000	$310	$37,710,000	1,000	11,420	3,060	41	15,480

mt – metric ton

Table 2-16. Summary of Control Costs and Emission Reductions for a 1,000 MW Unit

Option No.	**Control Configuration**	**TCI**		**TAC**	**Annual Emission Reductions**				
		$	$/kW	$/yr	*NO_x (mt)*	*SO_2 (mt)*	*PM (mt)*	*Hg (kg)*	*Total*
1	SCR/Wet FGD/FF	$496,150,000	$500	$83,650,000	1,990	23,640	6,110	99	31,740
2	SCR/Wet FGD/ESP	$501,910,000	$500	$68,930,000	1,990	23,640	6,110	24	31,740
3	SCR/SD/FF	$336,620,000	$340	$80,060,000	1,990	22,840	6,110	37	30,940
4	SCR/Wet FGD/FF/CI	$496,150,000	$500	$83,650,000	1,990	23,640	6,110	114	31,740
5	SCR/Wet FGD/ESP/CI	$501,910,000	$510	$68,930,000	1,990	23,640	6,110	99	31,740
6	SCR/SD/FF/CI	$336,620,000	$340	$80,060,000	1,990	22,840	6,110	114	30,940
7	SNCR/Wet FGD/FF	$386,840,000	$390	$67,210,000	630	23,640	6,110	99	30,380
8	SNCR/Wet FGD/ESP	$392,590,000	$390	$52,490,000	630	23,640	6,110	24	30,380
9	SNCR/SD/FF	$227,310,000	$230	$63,620,000	630	22,840	6,110	37	29,580
10	SNCR/Wet FGD/FF/CI	$386,840,000	$390	$67,210,000	630	23,640	6,110	114	30,380
11	SNCR/Wet FGD/ESP/CI	$392,590,000	$400	$52,490,000	630	23,640	6,110	99	30,380
12	SNCR/SD/FF/CI	$227,310,000	$230	$63,620,000	630	22,840	6,110	114	29,580
13	NO_xOut/Wet FGD/FF	$438,740,000	$440	$73,420,000	1,990	23,640	6,110	99	31,740
14	NO_xOut/Wet FGD/ESP	$444,500,000	$440	$58,700,000	1,990	23,640	6,110	24	31,740
15	NO_xOut/SD/FF	$279,210,000	$280	$69,830,000	1,990	22,840	6,110	37	30,940
16	NO_xOut/Wet FGD/FF/CI	$438,740,000	$440	$73,420,000	1,990	23,640	6,110	114	31,740
17	NO_xOut/Wet FGD/ESP/CI	$444,500,000	$450	$58,700,000	1,990	23,640	6,110	99	31,740
18	NO_xOut/SD/FF/CI	$279,210,000	$280	$69,830,000	1,990	22,840	6,110	114	30,940

mt – metric ton

Table 2-17. Summary of Energy Demand of Various Controls

Control	Units	Boiler Size (MW)		
		100	500	1,000
SCR	Usage (kWh/yr)	2,457,025	15,410,093	31,601,428
	$mtCO_2e^1$	2,096	13,145	26,956
SNCR	Usage (kWh/yr)	4,592,216	22,961,078	45,922,156
	$mtCO_2e^1$	3,917	19,586	39,172
Wet FGD	Usage (kWh/yr)	15,768	78,840	157,680
	$mtCO_2e^1$	13	67	135
Spray Dryer	Usage (kWh/yr)	5,519	27,594	55,188
	$mtCO_2e^1$	5	24	47
FF	Usage (kWh/yr)	4,178	25,119	49,879
	$mtCO_2e^1$	4	21	43
ESP	Usage (kWh/yr)	2,086	12,108	24,204
	$mtCO_2e^1$	2	10	21

[1] Calculation for tonnes of CO_2 are based on the 2006 Average Provincial CO_2 Electricity Emissions Factors (0.853 kg/kWh of CO_2e) taken from CAC Emission Summaries, 2006 Air Pollutant Emissions for Alberta, Version 1, April 2008. Emissions are based on 8,760 hours per year operations.

mt – metric ton

Petroleum Coke

The demand for light crude products such as transportation fuels exceeds that for heavy crude products. Petroleum coke, or "pet coke," can be used as an alternative to coal because of its high heat content, low ash content, ease of grinding, and low cost. However, pet coke is more difficult to burn than coal and has a high sulfur content that may offset its positive attributes. Its low volatility contributes to low reactivity, especially compared to coal, which leads to ignition instability.

Most pulverized coal-fired boilers are suited for pet coke blend fuel operations. Typically these boilers operate with a blend of 20 percent or less. Boiler efficiency is affected by the various characteristics of pet coke. Its high heating value and low moisture content improve boiler efficiency.

The higher peak flame temperatures associated with firing pet coke blends tend to increase production of thermal NO_x. In addition, the nitrogen content of pet coke is significantly higher than most coals, and an increase in fuel NO_x can be expected. Low NO_x burners that introduce air gradually along the path of combustion promote better combustion and higher flame stability of a low volatility fuel such as pet coke. The sulfur content of pet coke can be as high as 8 percent. The capacity of the FGD system often determines the maximum blend possible while meeting SO_2 requirements.

Although the ash produced through burning pet coke is low, it contains heavy metals. These heavy metal will likely deposit on the surface of SCR catalyst and react with the active sites. The deactivation of catalyst is rapid and usually irreversible. The presence of heavy metals such as vanadium at 50 to 60 percent in the ash, and nickel at concentrations of up to 12 percent in the ash, raise potential areas of concern over disposal.[1]

Pulverized coal boilers that blend coal and pet coke continue to be constructed and maintained. In 2004, Cross Generating Station in South Carolina was permitted to fire up to 30 percent pet coke by weight in its two 660 MW boilers. These boilers will be equipped with LNBs, SCRs, FGDs, and ESPs. A wall-fired boiler originally constructed in 1983 (Seminole Electric in Florida) was designed to fire a blend of coal and pet coke. In 2006, the burners and overfire air system were upgraded. Recent test data shows NO_x emissions less than 0.35 lb/MMBtu were achieved for full load operation while firing a blend of 70 percent Patiki coal and 30 percent petcoke. The NO_x reduction from the baseline level prior to the retrofit was approximately 35 percent.[2]

Wood and Wood Waste

No technical barriers prevent the use of wood in cofiring in power plants. Despite differences between coal and wood fuels, the equipment required for handling, storage, metering and feeding these materials is similar enough that cofiring has been proven feasible by a number of full-scale tests for both fuels.

Coal pulverizers can accept only a very small percentage of wood blended with coal. Experience suggests that the extreme upper limit is about 10 percent by weight or 5 percent by heat input. To get higher levels of heat input from wood in a cofired PC boiler, it is necessary to provide a separate sizing and firing system for the wood fuel, or to install a dump grate in the boiler.

The impact of cofiring wood and wood waste fuels with a dry primary fuel, such as coal, may also negatively impact boiler efficiency, net plant heat rate, and capacity. The slagging and fouling potential of wood and wood waste fuels are low compared to other waste product fuels.

Biomass co-firing may offer reductions in NO_x and SO_x. The SO_x improvement is easily calculated based on fuel sulfur content. There is some indication that the higher moisture and higher volatile content in biomass may reduce formation of NO_x by lowering flame temperature and increasing the staging of the combustion. The ash content of wood and wood wastes are low compared to coal. Wood ash, in most cases, is not toxic and does not contain pollutants, heavy metals, or mercury.[1]

2.5.4 *BACTEA Emission Limits*

Table 2-18 provides a summary of permit limits representing the BACT level of control in recent U.S. permits. The permits provided emission limits on an input basis. Output levels were calculated assuming boilers have a net efficiency of 38 percent, based on a value for supercritical unites quoted in a report for the EPA. For input limits, the minimum, maximum, and a typical performance level contained in permits for the various controls are shown. Only the typical performance level has been converted to an output limit and shown. Because mercury controls are relatively new, the permit limits are not based on as much experience as other types of controls. Additional data are likely needed to establish regulatory limitations for mercury.

Appendix E summarizes the operation status of the units with permits that were analyzed for this document. Seven out of the eleven units are under construction, and one unit has not begun construction. Three units are in operation and are in compliance with their 30-day rolling average limits. Additionally, all permits reviewed were for facilities located in attainment areas. As a result, their permits reflect the best available control technology (BACT) rather than the lowest achievable emission rate (LAER).

Table 2-18. Summary of BACT Levels for Pollutant/Control Combinations

Control Device	Input Basis (g/MJ)[d]			Typical Output (kg/MW-h)
	Minimum	Maximum	Typical Input	
NO_x Control				
SCR	0.03	0.43	0.034	0.32
NO_xOut Cascade	0.03	0.43	0.034	0.32
SNCR[a]	0.034, 0.043, 0.052			0.33, 0.41, 0.49
SO_2 Control				
Wet FGD	0.028	0.043	0.037	0.35
SD/FF	0.028	0.052	0.048	0.45
PM Control[c]				
FF	0.009	0.017	0.013	0.12
ESP	0.009	0.017	0.013	0.12
Hg Control				
Various w/CI[b]	7.3×10^{-7}	9.5×10^{-7}	9.1×10^{-7}	8.6×10^{-6}

[a] SNCR limit was based on the maximum achievable reduction efficiency (40 percent) of SNCR from the baseline NO_X emission rate for a 100 MW unit. NO_X removal efficiencies from SNCR decrease as the unit's size increases. A removal efficiency of 14 percent was used for the 1,000 MW unit, and a removal efficiency of 27 percent was used for the 500 MW unit.

[b] Limits are based on 6 recent permits issued in the U.S. for units burning PRB coal. Limits were originally on an output basis and were converted to input basis assuming the unit is 38 percent efficient on a net basis. Limits were not used in the cost and impacts calculations in Section 2.4 because Hg technology is relatively new and it cannot be determined if the permit limits are achievable by the units or are applicable to boilers in Alberta. Instead, control efficiencies, based on emission tests on Hg controls, were used to estimate costs and emission reductions.

[c] The PM limits are based on analytical tests that measure both filterable and condensible PM.

[d] Standards my be subject to exclusion periods during startup, shutdown, and certain malfunctions.

References

1. EPRI. Opportunity Fuels Guidebook, EPRI, Palo Alto, CA: 1998. Report TR-111487.

2. Conn, Richard. Low NO_x Combustion System Retrofit for Firing Coal/Petroleum Coke Blend Seminole Electric Units 1&2. Presented at PowerGen 2007 by Foster Wheeler North America Corporation.

3.0 BACTEA ANALYSIS FOR COMBUSTION TURBINES

The BACTEA analysis was conducted for control technologies used to reduce NO_x for both simple and combined-cycle combustion turbines. Small combustion turbines (less than 25 MW) have been excluded from the BACTEA analysis because they are not expected to be used by electrical utilities. In addition to size, ERG has included a peaking subcategory under the smaller sized turbines (25 to 150 MW). Peaking units are designed to generate energy on short notice and for relatively short periods of often operating on partial load. A base load unit provides the minimum constant level of electricity demand that a utility must meet.

This section presents ERG's BACTEA analysis using a step-by-step approach involving the following steps (the BACTEA analysis is described in detail in Section 1):

- Step 1—Identify all potential NO_x control technologies for utilities (i.e., the pertinent industry and processes).
- Steps 2 and 3—Eliminate technologies that are either not used at comparable facilities or are technologically infeasible.
- Step 4—Rank the remaining technologies based on effectiveness.
- Step 5—Evaluate control costs.
- Step 6—Select BACTEA.

This section, organized by the above steps, also includes a comparison of the control options by standardizing cost and removal results to a U.S. dollar per MW gross facility power output and dollar per tonne of NO_x reduction. Unless otherwise specified, all discussions of turbine power reflect gross output. For each BACTEA technology selected, ERG has specified emission limits that are reasonably achievable with proper engineering design and operation. Although the costs of control have been determined for natural gas-fired turbines, the effects of other fuels, such as syngas and refinery gas, on control technologies are included.

Information on the control technologies was gathered from U.S. permit applications, BACT analyses conducted by the permittee and submitted with the application, technical support documents for issued permits, EPA databases (such as the inventory of Combustion Turbine New Source Review (NSR) Construction Permits maintained by Region 4), and technical journals.

The turbine unit sizes in which the BACTEA analyses were conducted are based on a review of the DOE EIA database of operating turbines in the United States. ERG developed the size ranges to represent the sizes expected in the future and to provide ranges that are small enough that any member of a group would be well represented by the parameters assigned for that model to the group.

The turbine control cost estimates are based on EPA's Control Cost Manual 5th Edition (2000) adjusted to 2007 dollars and EPA background information documents supporting New Source Performance Standards (NSPS) for stationary combustion turbines, promulgated July 6, 2006. EPA cost conclusions were compared with EPRI-published material. EPRI, a nonprofit organization, supports multidiscipline research in emerging technologies. EPRI's members include Canadian utilities.

3.1 Selection of Control Technologies

NO_x formation in turbines occurs by three mechanisms. The principal mechanism with firing gas is thermal NO_x, which arises from the thermal dissociation and subsequent reaction of nitrogen (N_2) and oxygen (O_2) molecules in the combustion air. Most thermal NO_x is formed in high-temperature flame pockets downstream of the fuel injectors. The second mechanism, prompt NO_x, is formed from early reactions of nitrogen molecules in the combustion air and hydrocarbon radicals from the fuel. The third mechanism, fuel NO_x, stems from the reaction of fuel-bound nitrogen compounds with oxygen. Natural gas has negligible chemically bound fuel nitrogen (although some molecular nitrogen is present). Essentially all NO_x formed from natural gas combustion is thermal NO_x.

The simple cycle is the most basic operating cycle, using three main components: an air compressor, a combustor, and a power turbine. After the exhaust gases produce power in the power turbine, they are vented directly to the atmosphere without recovering their thermal energy. A combined-cycle turbine incorporates a heat recovery steam generator (HRSG) and steam turbine. The hot exhaust gas from the combustion turbine is directed to an HRSG to produce steam. The steam produced in the HRSG is used in a steam turbine, which also drives an electric generator. The HRSG may or may not have duct burners that supply additional heat to produce more steam.

Conventional turbine combustors are diffusion-controlled, where fuel and air are injected separately. Combustion occurs locally at stoichiometric interfaces, resulting in hot spots that produce high levels of NO_x. In contrast, lean premixed combustors (also known as dry low-NO_x or DLN combustors) mix the fuel and air prior to combustion.

In the DLN design, the air and fuel are premixed at a lean "air-to-fuel" (A/F) ratio prior to injection into the combustion zone. The fuel is combusted with excess air and the fuel is mixed well with the air prior to ignition; therefore, peak combustion temperatures and thermal NO_x are reduced. Currently, DLN combustion systems are standard equipment on many gas turbines.[1] Therefore, DLN is considered baseline control for the BACTEA analysis.

The majority of commercially available DLN combustors achieve NO_x reduction to 25 parts per million volume (ppmv). NO_x guarantees as low as 9 ppmv can be obtained.[2] Advanced designs and materials are continually being developed and often are first incorporated into the larger frame machines.

CASA workgroup members stated that aeroderivative turbines often are used as peaking units in Alberta. Aeroderivative turbines are not as likely to have low-NO_x combustion controls installed. Therefore, ERG’s baseline assumption, a DLN combustion system, does not apply as well to aeroderivative peaking units. Although we have not developed specific costing for these options, there are relatively low cost solutions to NO_x control for aeroderivative turbines that would achieve the assumed baseline level of control of 25 ppmv, such as water/steam injection or lean premix combustors. These are discussed throughout this section, with cost and emission reduction information provided where available.

Aeroderivative turbines are derived from flight services, including propulsion via turbojet, turbofan, or turboprop, and for helicopter power. Aeroderivative engines have higher efficiency and power density (kilowatt or horsepower per unit of air flow and machine volume) than their industrial, heavy-duty competitors. High-performance, high-efficiency aeroderivatives are fast-starting and tolerant to cycling, characteristics that make them suitable for peaking power and distributed generation applications. Lean premix combustors for aeroderivative turbines are referred to as dry low emissions (DLE). The principal of DLE and DLN is the same.

3.1.1 *Identification of Control Technologies (Step 1)*

The following section presents ERG's review of available NO_x control technologies for combustion turbines firing natural gas. Six control technologies were identified:

1. DLN with Catalytic Combustion (DLN2)

Catalytic technology features "flameless" combustion that occurs in a series of catalytic reactions to limit the temperature in the combustor. This allows complete mixing of the fuel and total air, with the combustion initiated by a catalytic surface and occurring at temperatures below those at which measurable amounts of NO_x form. Calytica Combustion Systems, Inc., performed a successful 27-month demonstration of its Xonon™ catalytic combustor in a 1.5 MW Kawasaki turbine. The system accumulated more than 20,000 hours of reliable operation at this site with NO_x emissions consistently below 3 ppmv.[3]

In August 2006, Kawasaki purchased the combustion technology and its related intellectual property. The catalytic combustor, the K-Lean or X-Combustor, is now available on a single model, a 1.4-MW turbine. Kawasaki has no plans to apply this technology to larger turbines. No other manufacturer makes a flameless combustor.

2. Non-selective Catalytic Reduction (NSCR)

NSCR technology is designed to simultaneously reduce NO_x and oxidize carbon monoxide (CO) and hydrocarbons (HCs) in the combustion gas to nitrogen, carbon dioxide (CO_2), and water. The catalyst, usually a noble metal, causes the reducing gases in the exhaust stream (hydrogen (H_2), methane (CH_4), and CO) to reduce both nitric oxide (NO) and nitrogen dioxide (NO_2) to nitrogen at a temperature between 800°F and 1,200°F (430°C and 650°C). To be effective, NSCR requires a low excess oxygen concentration in the exhaust gas stream because the oxygen must be depleted before the reduction chemistry can proceed.

3. SCONOx™

SCONOx™ is an oxidation catalyst-based technology that removes both NO_x and CO without the need for supplementary chemical reagents, such as NH_3. The SCONOx™ catalytic absorption system uses a potassium carbonate-coated catalyst to reduce NO_x emissions. The catalyst oxidizes CO to CO_2 and NO to NO_2 and potassium nitrates (KNO_3). The catalyst is regenerated by passing dilute hydrogen gas through the catalyst, which converts the KNO_2 and KNO_3 to K_2CO_3, water, and elemental nitrogen. The catalyst is renewed and available for further absorption, while the water and nitrogen are exhausted. The $SCONO_x$ system has demonstrated its ability to meet the same low emission rates as a conventional SCR/CO oxidation catalyst system without the use of NH_3. EMx™ (the second-generation of the SCONO$_x$™ NO_x Absorber technology) has been commercially demonstrated on several small (5 MW) gas turbines and a single 45 MW gas turbine in Redding, California, with NO_x emissions below 1.5 ppmv. [4]

4. Selective Catalytic Reduction (SCR)

SCR is a post-combustion gas treatment technique for reduction of NO and NO_2 in the turbine exhaust stream to molecular nitrogen, water, and oxygen. In the SCR process, aqueous or anhydrous NH_3 is used as the reducing agent and is injected into the flue gas upstream of the catalyst bed. The function of the catalyst is to lower the activation energy of the NO_x decomposition reaction. NO_x and NH_3 combine at the catalyst surface, forming an ammonium salt intermediate, which subsequently decomposes to produce elemental nitrogen and water. The NH_3/NO_x ratio can be varied to achieve the desired level of NO_x reduction. Increasing this ratio will not only further reduce NO_x emissions, but also will result in increased unreacted NH_3 that "slips" through the process into the atmosphere. Removal efficiencies are generally 80 to 95 percent. Technical literature shows the removal efficiency of a SCR system installed on a simple-cycle, gas-fired turbine can achieve an exhaust gas NO_x concentration of 3.4 ppmv with an NH_3 slip of 10 ppmv.[5] (Lower NO_x concentrations are achievable; however, NH_3 slip data were not available)

The catalyst's active surface is usually a noble metal, base metal (titanium or vanadium) oxide, or a zeolite-based material. Base metal catalysts have an operating temperature window for clean fuel applications of approximately 500°F to 800°F (260° to 426°C).

Turbines that operate in simple-cycle mode have exhaust gas temperatures ranging from approximately 850°F to 1,000°F (450°C to 540°C). For a base metal catalyst to be used on a simple-cycle turbine, the exhaust must be cooled first. Turbine heat recovery or dilution air systems can reduce exhaust gas temperatures to the proper operating range for the catalyst. An alternative is the use of other catalysts with higher temperature characteristics. The upper range of the temperature window can be increased to a maximum of 1,100°F (590°C) using a zeolite catalyst. The hot exhaust from the combustion turbine in a combined-cycle application has an opportunity to cool down in the HSRG and the base metal catalysts are more commonly used for combined-cycle turbines. SCR for combined-cycle turbines are often built into HSRG.

5. Selective Non-catalytic Reduction (SNCR)

SNCR is an add-on technology that involves the noncatalytic decomposition of NO_x in the flue gas to nitrogen and water using reducing agents, such as urea or NH_3. Since SNCR does not require a catalyst, the initial capital costs are lower than SCR. The reducing agent must be injected into the flue gas at a location in the unit that provides the optimum reaction temperature and residence time. The NH_3 process (e.g., trade name: Thermal DeNOX) requires a reaction temperature window of 1,600^{o}F to 2,200^{o}F (870°C to 1,200°C). In the urea process (e.g., trade name: NO$_X$OUT), the optimum temperature ranges from 1,600^{o}F to 2,100^{o}F (870°C to 1,150°C).

6. Water/Steam Injection

Water/steam injection is a mature technology, having been used since the 1970s to control NO_x emissions from gas turbines. Water/steam injection as a control technology involves the introduction of water or steam into the combustion zone. The injected fluid provides a heat sink, which absorbs some of the heat of reaction, causing a lower flame temperature. The lower flame temperature results in lower thermal NO_x formation. The water used for either approach needs to be demineralized thoroughly to avoid forming deposits and corrosion in the turbine expansion section.

The “water-to-fuel ratio” has a direct impact on the controlled NO_x emission rate and is generally controlled by the turbine inlet temperature and ambient temperature. The decision whether to use water versus steam injection depends on the availability and cost of steam, turbine performance, and maintenance impacts. Direct water impingement can result in rapid wear of the combustor liner. The impact of steam injection has been linked to a reduced life for the hot section parts due to the change in transport properties (added moisture increases heat transfer) and the increased compressor discharge pressure and temperature resulting from the added mass flow.[6]

Wet control technology, which was developed for combustors that had uncontrolled emissions of 100 ppmv or more, can reduce NO_x by 60 percent or more. Both water and steam increase the mass flow through the system and create a small amount of additional power. Wet control typically increases power output by 5 to 6 percent and decreases efficiency up to 4 percent.[7] Controlled NO_x emission levels are generally about 25 ppmv for gas turbines. There are some turbines in which manufacturers indicate an outlet concentration of 15 ppmvd of NO_x when using water/steam injection can be achieved. For example, the GE aeorderivative turbine LM2500 can achieve 15 ppmvd NO_x at 15% O_2 at a steam flow of about 10,000 kg/hr. Also, a CASA member indicated that their 3 aeroderivative GE LM6000 achieved actual emissions less than the 25 ppmvd guarantee.

The expected NO_x emissions from an aeroderivative turbine with no control are 175 ppmv. With steam or water, injection emission levels of 25 ppmv are achievable. DLE combustion systems are less proven on aeroderivative turbines; water and steam injection are less complex and have greater long-term durability.

3.1.2 *Elimination of Control Technologies (Step 2 & 3)*

Certain technologies were rejected from further analysis because the controls are technologically infeasible. For the BACTEA analysis, “infeasible” is defined as follows:

- Not appropriate for the operational situations in Alberta (e.g., climate, load, fuel type available).
- Not used at comparable facilities.
- Other technologies offer both improved turbine efficiency and lower removal efficiencies.

The following control options have been determined to be technically infeasible:

1. DLN2

XONON is considered technically infeasible because it is not commercially available for the turbine sizes typically found in Alberta. The combustor technology is available only on

Kawasaki model GPB15, a small 1.4-MW turbine. The flameless combustor has very different and proprietary controls (hardware and software) that are required to make the combustor function. The 18-MW model GPB180 is the largest gas turbine currently available by Kawasaki. Kawasaki has no plans to apply this technology to larger turbines. Also, the largest gas turbine that Kawasaki sells is 18 MW. Therefore, it is not expected that this technology will be available for gas turbines greater than 25 MW. No other manufacturer makes a flameless combustor.

2. NSCR

NSCR is considered technically infeasible because lean-burn DLN combustion included as standard equipment on natural gas-fired combustion turbines operates under fuel-lean conditions (relatively high excess oxygen). NSCR requires a low excess oxygen concentration in the exhaust gas stream to be effective. NSCR is technically infeasible on DLE aeroderivative turbines.

3. SCONO$_X$™

SCONO$_X$™ is considered technically infeasible because it is not commercially available. Although the technology has been installed and operated on one 45 MW turbine in California, it is not a mature technology. The majority of permitted and operating units are small 5-MW units. There are no known installations in low ambient temperature settings. The SCONO$_X$™ / EM$_X$™ technology can not be applied with predictable results.

The maximum catalyst operating temperature is 700°F (370°C). It should also be noted that the use of the SCONO$_X$™ catalyst for simple-cycle installations might be limited due to temperature. SCONO$_X$™ is also very sensitive to fuels other than natural gas; sulfur in other fuel types might coat or cover the catalyst active sites, reducing NO_x or NH_3 diffusion and necessitating frequent cleaning.

4. SNCR

SNCR is considered technically infeasible because of incompatibility with both the simple- and combined-cycle, turbine exhaust temperature range of 800°F to 1,000°F (425°C to 540°C). The optimum temperature range for SNCR is between 1,600°F and 2,100°F (870°C to 1,150°C). Additionally, the residence time required for the reaction is approximately 100 milliseconds, which is relatively slow for gas turbines. It might be feasible to initiate this reaction in the gas turbine (where operating temperatures fall within the reaction window) if suitable modifications and injection systems can be developed; however, this technology has not been applied to date. Aeroderivatives' turbine exhaust temperature, ranging from 700°F to 1,000°F (370°C to 540°C), is also outside the optimum temperature range for SNCR.

5. Water/Steam Injection

Water/steam injection is considered technically infeasible for this analysis because the baseline control has been chosen as a DLN burner. The majority of commercially available DLN combustors achieve NO_x reduction to 25 ppmv. NO_x guarantees as low at 9 ppmv can be obtained.[2] Water/steam injection can also obtain 25 ppmv levels of NO_x for some turbines, and even lower in some cases. However, wet injection is not expected to reduce NO_x lower than the DLN for most turbines; therefore, water/steam injection is considered technically infeasible for a turbine with a DLN burner installed.

Because water/steam injection can often achieve NO_x outlet concentrations of 25 ppmvd or less, this technology is equivalent to the use of DLN burners in these applications. Section 3.4.2 addresses in more detail the baseline level of control assumption.

3.1.3 *Ranking of Control Technologies (Step 4)*

Currently, DLN combustion systems are standard equipment on many gas turbines.[1] Therefore, the DLN emission rate of 25 ppmv is considered baseline. The only control technology considered technically feasible for NO_x control in conjunction with the baseline level of control (i.e., DLN) is SCR. Technical literature shows that an SCR system installed on a simple-cycle, gas-fired turbine can achieve a NO_x concentration of 2.0 ppmv.[5] It is presumed that combined-cycle, gas-fired sources can achieve these same levels of emission reduction.

3.1.4 *Collateral Impacts*

Multiple environmental impacts occur from the use of SCR technology. As discussed previously, the more NH_3 that is used the more NO_x is removed; therefore, the NH_3/NO_x ratio can be varied to achieve the desired level of NO_x reduction. Generally, the more NH_3 used the more likely unreacted NH_3 will “slip” through the process into the atmosphere. Based on a sample of turbine permits, the NH_3 slip emission levels associated with the operation of the SCR are typically limited to less than 5 ppmvd.

In the SCR process, aqueous or anhydrous NH_3 is used as the reducing agent and is injected into the flue gas upstream of the catalyst bed. Additional products of combustion are generated due to the increased fuel combustion for generating power to pump and heat the reducing agent solution and to make up for the lost power caused by the catalyst backpressure. At the end of the catalyst life, it must be trucked offsite for disposal.

Ammonia can be supplied in any of three different forms: aqueous, anhydrous, or urea. EPA considers aqueous and anhydrous NH_3 to be hazardous material. Cold temperatures and concern for sensitive habitats complicate transportation, storage, and handling of NH_3. Larger SCR systems use anhydrous NH_3, requiring onsite storage of this chemical under pressure. It should be noted that many peaking turbines are located at unattended facilities. The need to store NH_3 at a site that might be unattended for substantial periods of time should be considered.

3.2 Estimation of Control Costs and Emission Reduction (Step 5)

3.2.1 *Control Cost Background*

The following section presents capital and annual costs for technologically feasible NO_x control technologies described above. ERG has developed costs for four size categories of turbines. The additional factors of operating cycle (peaking and base load) and supplemental heat (duct burners) have also been considered. The turbine units are based on a review of DOE data and were developed to evaluate the control technologies for a range of gas turbine sizes and loads. Each operating scenario and size combination that was considered is shown in Table 3-1.

Table 3-1. Model Units for New Gas Turbines

Class	Combustion Turbine Power Output Size Range (MW)	Combustion Turbine Power Output Representative Size (MW)	Operating Cycle	Total Facility Power Output (MW)[1]	Load
Small 1	25-75	50	Simple Cycle	50	Peaking
Small 2	75-150	113	Simple Cycle	113	Peaking
Small 1	25-75	50	Simple Cycle	50	Base
Small 2	75-150	113	Simple Cycle	113	Base
Medium	150-200	175	Simple Cycle	175	Base
Large 1	greater than 200	300	Simple Cycle	300	Base
Small 1	25-75	50	Combined Cycle	77	Base
Small 2	75-150	113	Combined Cycle	137	Base
Medium	150-200	175	Combined Cycle	269	Base
Large 1	greater than 200	300	Combined Cycle	462	Base
Small 1	25-75	50	Combined Cycle w/ Duct Burner	92	Base
Small 2	75-150	113	Combined Cycle w/ Duct Burner	208	Base
Medium	150-200	175	Combined Cycle w/ Duct Burner	323	Base
Large	greater than 200	300	Combined Cycle w/ Duct Burner	554	Base

[1] Total facility power output (MW) includes the power output from the combustion turbine generator and the steam turbine generator for combined cycle turbines. It was assumed that a steam turbine would increase the power output by 54% and a duct burner would add 20% additional power output. Therefore, the total facility power output of a combined cycle turbine with a duct burner = (Combustion Turbine Generator MW Power Output) x (1.54) x (1.20).

Size Ranges

ERG has subdivided combustion turbines into four size ranges. These sizes are small-1 (25-75 MW), small-2 (75-150 MW), medium (150-200 MW), and large (greater than 200 MW). Small combustion turbines (less than 25 MW) have been excluded because they are not expected to be used at electric utilities.

These ranges were established based on a review of the DOE EIA database of operating turbines in the United States. ERG developed the size ranges to represent the sizes expected in the future and to provide ranges that were small enough that any member of a group would be well represented by the parameters assigned to that group. The combined-cycle turbine sizes represent only the combustion turbine, not the total facility power.

Load

In addition to size, ERG included a peaking operating cycle subcategory under the two smaller sized turbines. Peaking units are designed to generate energy on short notice and for relatively short periods of time. Peaking units are used when all other units and energy sources are operating at maximum capability during peak hours or during unforeseen outages. Most utilities meet peak demand with a combination of resources (e.g., hydroelectric power, coal-fired boilers). ERG assumed 50 percent utilization for peaking units, or 4,200 hours per year. Although this utilization assumption might be higher than many individual utilities, it allows for a conservative estimation of cost effectiveness (dollar per tonne of NO_x reduction). If a combustion turbine and SCR are not operational, no removal occurs, and no return on capital costs investment is realized. ERG assumes a base load unit will operate 8,400 hours per year (24 hrs/day, 7 days/week, 50 weeks/yr).

Basic equipment for an operational combined-cycle package includes gas turbines, HRSG, steam turbine, and electric generators. The combined-cycle system incorporates two simple-cycle systems into one generation unit to maximize energy efficiency. Energy is produced in the first cycle using a gas turbine; then the heat that remains is used to create steam, which is run through a steam turbine. With respect to NO_x emissions, the only additional consideration for combined-cycle turbines is the use of a duct burner in the HRSG.

Duct Burners

When a gas turbine is equipped with an HRSG, it is common to utilize supplemental firing to increase steam generation or temperature. Equipment used to provide supplemental firing is often located in the duct between the combustion turbine exhaust diffuser and the HRSG inlet. These supplemental firing systems are often referred to as duct burners. Duct burners can burn several kinds of gaseous and liquid fuels and can be designed to support single- or dual-fuel capabilities. In this analysis, it was assumed that the duct burner would burn natural gas since natural gas is being used in the turbine. Approximately two-thirds of existing combined-cycle plants use duct burners to increase combined-cycle power output.[5] According to a review of recent construction permits, the size of a duct burner package is typically 18 to 25 percent of the combustion turbine input capacity. Model units assume duct burner packages are 20 percent of the combustion turbine input capacity.

Duct burner manufacturers have demonstrated that NO_x, CO, and unburned hydrocarbon production can be minimized for most applications with a duct burner design that produces certain gas flow dynamics. The level of NO_x produced by a duct burner burning natural gas is approximately 0.1 pound per million Btu of fuel burned. The ppmv level depends on the flowrate of gas turbine exhaust gases at which the burner is operating and thus, varies with the size of the turbine.[9]

The SCR catalyst is typically located inside the HSRG behind the high-pressure evaporator on the stack side, downstream of the duct burner. When a catalyst is used, HRSG performance should be evaluated at various modes of operation to ensure that the gas temperatures are within limits set by the catalyst supplier. For most combined-cycle applications, a medium-temperature catalyst type (e.g., vanadia/titanium catalyst on high-density honeycomb structure) with an operating temperature range between 500°F and 800°F (260° and 426°C) is used. [5]

3.2.2 *Control Cost Components*

The control cost estimation procedures are based on EPA's Control Cost Manual, 5th Edition (1996). SCR purchase costs are based on background information collected by EPA to support the NSPS for stationary gas turbines (40 CFR Part 60, Subpart GG) promulgated July 6, 2006. EPA cost conclusions were compared with EPRI published material. Where appropriate, costs were adjusted using Chemical Engineering plant cost index.

Purchased Equipment Cost

DOE background information contained basic equipment quotes from two SCR catalyst vendors for three turbine sizes (5, 25, and 150 MW).[10] To determine the costs of an SCR unit for other sizes of turbines, the costs were plotted as a function of combustion turbine size in MW to obtain a linear relationship. Although the DOE reference is almost nine years old, it continues to be cited by EPRI and EPA as an accurate characterization of control costs.[11, 12] In addition, equipment costs were adjusted to 2007 using the Chemical Engineering plant cost index.

Due to their high exhaust temperatures, it is assumed that simple-cycle turbines use a high temperature SCR catalyst. High temperature SCR is approximately 10 percent more costly than conventional SCR.

Basic equipment costs include an NH_3 injection skid, NH_3 storage equipment, and instrumentation. Capital costs include taxes, freight charges, and installation costs.

Direct and Indirect Installation Costs

Direct and indirect installation costs are estimated as a percentage of the purchased equipment cost as specified by OAQPS. The equations for each line-item cost are presented in Appendix B. Direct installation costs include costs for foundations and supports, erecting and handling the equipment, electrical work, piping, insulation, and painting. SCR will not require buildings, site preparation, offsite facilities, or land.

Indirect installation costs include costs such as construction and field expenses (i.e., costs for construction supervisory personnel, office personnel), startup and performance test costs (to get the control system running and to verify that it meets performance guarantees), and contingencies. Contingencies cover unforeseen costs that may arise, such as modification of equipment, escalation increases in equipment cost, or delays encountered in startup. Project contingency costs are assumed to equal 3 percent of purchased equipment.

Annual Costs

Direct annual costs include the purchase costs of SCR catalysts, reducing reagent (ammonia), electrical power, and labor necessary to maintain good operation. Indirect annual costs include overhead, property taxes, insurance, administrative changes and capital recovery. Total Annual Costs (TAC) are the sum of the direct and indirect annual costs.

The SCR reactor is a stationary device with no moving parts. Further, the SCR system incorporates only a few pieces of rotating equipment (e.g., pumps, motors). It is assumed that the existing plant staff spend 30 minutes per shift to maintain the SCR.[10] The facility operator and maintenance labor rates are obtained from Canadian Labour Relations literature.

The SCR catalyst reactor increases the back pressure on the turbine, which decreases the turbine power output by approximately 0.5 percent. The estimated costs account for the electrical demand of the NH_3 injection blower. The SCR catalyst cost and ammonia reagent usage are based on simple equations.

SCR catalyst vendors typically guarantee the catalyst life for a range of 10,000 to 30,000 hours. Where the catalyst life is 30,000 hours and the annual hours of operation are 8,400 catalyst life equals 3.6 years. ERG assumed the catalyst life for a base load unit as three years. ERG assumed 50 percent utilization for peaking units, or 4,200 hours per year. Given the same life expectancy for catalyst, catalyst should last 7-years at a peaking unit; however the additional life gained is uncertain given the frequent startup shutdown nature of peaking units and associated thermal stresses. All SCR systems include dampers, the complexity of which depends on whether the unit has been designed for base or peaking unit. Damper leakage can lead to degradation in catalyst activity. Dampers for a peaking unit are more complex than a damper for a base unit. Therefore, ERG has assumed a shorter life for catalyst used in peaking units of five years.

ERG did not find sufficient information to make a clear decision on a reduction of catalyst life for peaking units. Vendors reported that their catalyst could handle the stress, but other reports indicated that this could be a problem. In order to address the issue we used the shorter life estimate, but it is not completely clear if the life should be considered even shorter or if the catalyst life should be the same as for base units.

SCR catalysts can deactivate by any of three mechanisms: poisoning, masking, or plugging. Phosphorus compounds used in lubricating oils found in gas turbines can poison catalysts. Dual fuel operation can expose catalyst to fuel oil contaminants, such as sodium and potassium which can masks catalysts active sites. The deactivation of the catalyst is rapid and usually irreversible. The upstream ductwork must be fabricated so that no material, especially fibrous insulation, can get loose to plug the catalyst. The only solution to avoiding deactivation is to minimize concentrations of poisons, masking or plugging agent in flue gas. If properly managed, a peaking unit should have no additional catalyst deactivation than a comparable base unit.

The life of the SCR system depends on many factors, including operating environment, maintenance practices, and construction materials. To calculate the capital recovery costs, ERG assumed a 20-year expected useful life of the SCR system and a 7 percent discount rate. This assumes no salvage value can be taken for the system at the conclusion of its useful life. Even if it were reusable, the cost of disassembling the system into its components could be as high as the salvage value. The addition of add on control will have no affect on the useful life of a peaking or base load turbine.

3.2.3 *Control Cost*

Table 3-2 summarizes the estimated NO_x control costs. Individual component costs, including SCR purchased equipment cost, installation costs, and annual costs for each model unit, are included in Appendix A. Table 3-2 lists the TCI and TAC for each model.

The total gross facility power output of a combined-cycle plant includes the steam turbine generator and the combustion turbine generator. A combined-cycle plant produces 60 to 70 percent of the total plant power from the combustion turbine and 30 to 40 percent from the steam turbine.[13] The total capital investment costs for a combined cycle SCR are slightly less that of a simple cycle of the same size because of simple cycle turbine requires a high-temperature , higher cost catalyst. Table 3-2 also presents the total capital costs as a function of the total energy produced over the lifetime of the SCR ($/MW-hr). Using this statistical measure the combined cycle SCR is more affordable given that the steam turbine generator does not produce NO_x emissions.

Table 3-2. SCR Control Costs

ID	Model Unit Description	Total Facility Power Output (MW)[1]	Total Capital Investment (TCI)[2]			Total Annual Costs (TAC)[5]		
			Million $	Lifetime ($/MW-hr)[3]	Cost (Million $/MW capacity)[4]	Million $/yr	Per energy produced ($/MW-hr)[6]	Per power capacity (Million $/MW)[7]
1	50 MW - SC - Peak	50	2.62	0.62	0.052	0.68	3.26	0.014
2	113 MW - SC - Peak	113	4.91	0.52	0.044	1.34	2.84	0.012
3	50 MW - SC - Base	50	2.62	0.31	0.052	0.93	2.21	0.019
4	113 MW - SC - Base	113	4.91	0.26	0.044	1.82	1.93	0.016
5	175 MW - SC - Base	175	7.21	0.25	0.041	2.72	1.85	0.016
6	300 MW - SC - Base	300	11.8	0.23	0.039	4.51	1.79	0.015
7	50 MW - CC - Base	77	1.94	0.15	0.025	0.84	1.30	0.011
8	113 MW - CC - Base	173	3.50	0.12	0.020	1.64	1.12	0.009
9	175 MW - CC - Base	269	5.07	0.11	0.019	2.43	1.08	0.009
10	300 MW - CC - Base	462	8.20	0.11	0.018	4.02	1.04	0.009
11	50 MW - CC - Base w/ DB	92	2.19	0.14	0.024	0.94	1.21	0.010
12	113 MW - CC - Base w/ DB	208	4.07	0.12	0.020	1.85	1.06	0.009
13	175 MW - CC - Base w/ DB	323	5.94	0.11	0.018	2.77	1.02	0.009
14	300 MW - CC - Base w/ DB	554	9.70	0.10	0.018	4.61	0.99	0.008

SC = Simple Cycle; CC = Combined Cycle; DB = Duct Burner; all dollars are in US dollars

[1] Total facility power output (MW) includes the power output from the combustion turbine generator and the steam turbine generator, if applicable.
[2] TCI (Million $US) is the total capital investment cost associated with the SCR.
[3] TCI Lifetime ($/MW-hr) is capital cost of the SCR allocated to each MW-hr of energy produced over the life time of the SCR. It is equal to TCI divided by the total facility maximum energy output over the 20-year life of the equipment, where peaking units operate 4,200 hours per year, and base load units operate 8,400 hours per year.
[4] TCI Cost (Million $/MW capacity) is the TCI divided by the capacity of the total facility.
[5] TAC (Million $/yr) is the amortized TCI (capital recovery) plus the Direct Annual Costs associated with the SCR.
[6] TAC per energy produced ($/MW-hr annual) is the TAC divided by the annual total facility energy output.
[7] TAC per power capacity (Million $/MW) is the TAC divided by the capacity of the total facility.

3.3 Estimation of Emission Reductions

SCRs are generally capable of efficiencies of 80 to 95 percent. An SCR can be designed to achieve a targeted NO_x reduction by manipulating the reagent usage with respect to the stoichiometric ratio or increasing catalyst volume and the exhaust gas residence time.

DLN combustion systems are standard equipment on many gas turbines.[1] The majority of commercially available DLN combustors achieve NO_x reduction to 25 ppmv.[2] For turbines that do not come standard with a DLN, wet injection is available at a low cost that can also achieve the 25 ppmv. Therefore, ERG considered the DLN (wet injection) emission rate of 25 ppmv as baseline.

Technical literature shows that an SCR system installed on a gas-fired turbine can achieve NO_x concentration of 2.0 ppmv.[5] To achieve that level of control, the necessary SCR removal efficiency is 92 percent.

The duct burners contribute additional NO_x for combined-cycle turbines. Because the SCR is typically located downstream of the duct burner, the SCR will control emissions from the duct burner and combustion turbine. ERG assumed that a duct burner heat input capacity is 20 percent of the combustion turbine heat input capacity. To compensate for the additional NO_x, the catalyst volume has been increased by 20 percent. Table 3-3 shows the NO_x formation from the turbine and duct burner and the subsequent SCR removal.

Additional emissions associated with startup, shutdown, and malfunction were not estimated. At lower loads, turbines emit higher levels of NO_x. Also, there will be a time delay at startup for SCR equipment to reduce emissions at the expected level.

Table 3-3. NO_x Formation and Removal

ID	Model Unit Description	NO_x from DLN (tonne/yr)	NO_x from DB (tonne/ yr)	SCR NO_x Removed (tonne/ yr)	NO_x Emissions After Control (tonnes/yr)	NO_x Emissions After Control (kg/MW-hr)
1	50 MW - SC - Peak	91.9	-	84.6	7.35	0.035
2	113 MW - SC - Peak	207	-	190	16.5	0.035
3	50 MW - SC - Base	184	-	169	14.7	0.035
4	113 MW - SC - Base	414	-	381	33.1	0.035
5	175 MW - SC - Base	643	-	592	51.5	0.035
6	300 MW - SC - Base	1,103	-	1,015	88.3	0.035

Table 3-3. NO_x Formation and Removal (Continued)

ID	Model Unit Description	NO_x from DLN (tonne/yr)	NO_x from DB (tonne/ yr)	SCR NO_x Removed (tonne/ yr)	NO_x Emissions After Control (tonnes/yr)	NO_x Emissions After Control (kg/MW-hr)
7	50 MW - CC - Base	184	-	169	14.7	0.023
8	113 MW - CC - Base	414	-	381	33.1	0.023
9	175 MW - CC - Base	643	-	592	51.5	0.023
10	300 MW - CC - Base	1,103	-	1,015	88.3	0.023
11	50 MW - CC - Base w/ DB	184	13.0	181	15.7	0.020
12	113 MW - CC - Base w/ DB	414	29.3	407	35.4	0.020
13	175 MW - CC - Base w/ DB	643	45.5	634	55.1	0.020
14	300 MW - CC - Base w/ DB	1,103	78.0	1,087	94	0.020

SC = Simple Cycle; CC = Combined Cycle; DB = Duct Burner; DLN - dry low-NO_X combustor; SCR = Selective Catalytic Reduction; MW = Total facility power output including the combustion turbine, and the steam turbine.

The cost effectiveness in $/tonne of NO_x removed was developed for each unit. The cost effectiveness for a given model is calculated by dividing the total annual cost by the annual NO_x reduction in tonnes.

3.4 Summary of Costs and Emission Reductions

Table 3-4 summarizes the costs, emissions reduction, cost effectiveness ($/tonne), and the cost impact ($/kw) for installing SCR on turbines used for electricity generation at the various sizes and loads analyzed. The results show that it is more cost effective to install SCR on larger turbines than on smaller turbines. Also, it is more cost effective to install and operate an SCR on combined cycle systems with duct burners than combined cycle systems without duct burners. These results are not unexpected given that there are economies of scale for SCR units; as SCR increase in size their costs rise, but not in proportion to the amount of emissions reduced.

The cost effectiveness values for SCR units controlling peaking units are the highest; it is more costly to remove a tonne of NO_x for peaking units than for a base unit of comparable size. With peaking units operating only partially throughout a year, it is not as cost effective to install and operate an SCR. Although the SCR capital costs for peak and base units are similar, fewer tonnes are removed at a peak unit due to reduced operation.

The most cost effective operating scenario and size combination is an SCR installed at a 300 MW combined-cycle base unit with duct burners. Table 3-4 also shows that the relative costs for SCR are highest for the smallest turbines and that it decreases with increasing turbine size.

The NO_x removals shown in Table 3-4 above are based on a baseline emission rate of 25 ppmv (a DLN in most cases or possibly water/steam injection for some small turbines) and a 92 percent SCR removal efficiency which is necessary to achieve an after control emission rate of 2 ppmv. It should be noted that some manufacturers have guaranteed uncontrolled NO_x levels of 9 ppmv for certain large combustion turbines. Installation of an SCR on such a turbine will require a smaller SCR to achieve 2 ppmv and the tonnes of NO_x removed will be reduced by

64 percent the amount assumed in this study. The SCR will be less cost effective (higher $/tonne) than the turbines shown in the table above. As a rough estimate of the change in the cost effectiveness and using the 300 MW combined cycle case (#14), the emissions that would be reduced using an SCR to reduce the concentration of NO_x from 9 to 2 ppmv, would be about 391 tonnes. The SCR used on the 113 MW combined cycle case (#12) was modeled to reduce about that amount or 407 tonnes. Therefore, the annual costs for the SCR should be about the same or $1.85 million, with a cost effectiveness of about $4700/tonne ($1.85 million/391 tonnes). The assumption of inlet concentration seems to have less of an effect on the overall cost effectiveness values calculated than the operating hour assumption made for the peaking units.

Table 3-4. SCR Cost Effectiveness for Each Turbine Model

Model	Model Unit Description	NO_x Removed (tonne/yr)	Cost Effectiveness ($/tonne)	TAC (Million $/MW-hr annual)
1	50 MW - SC - Peak	84.6	8,099	3.26
2	113 MW - SC - Peak	190	7,054	2.84
3	50 MW - SC - Base	169	5,497	2.21
4	113 MW - SC - Base	381	4,795	1.93
5	175 MW - SC - Base	592	4,595	1.85
6	300 MW - SC - Base	1,015	4,442	1.79
7	50 MW - CC - Base	169	4,958	1.30
8	113 MW - CC - Base	381	4,297	1.12
9	175 MW - CC - Base	592	4,108	1.08
10	300 MW - CC - Base	1,015	3,964	1.04
11	50 MW - CC - Base w/ DB	181	5,166	1.21
12	113 MW - CC - Base w/ DB	407	4,549	1.06
13	175 MW - CC - Base w/ DB	634	4,372	1.02
14	300 MW - CC - Base w/ DB	1,087	4,238	0.99

TAC Cost (Million $/MW-hr annual) is the TAC divided by the annual total facility power output; TAC = Direct Annual Costs + amortized Total Capital Investment; ERG assumed a 20-year expected useful life of the SCR system and a 7 percent discount rate.

In addition to presenting these overall results of the turbine BACTEA analysis, this section highlights some key assumptions, additional considerations for rule or permit writing, and information specifically requested by CASA. The section is further divided into these subsections:

- 3.4.1 Peaking Units – The cost effectiveness results and the assumption on percent load for peaking units are discussed.

- 3.4.2 Baseline Level of Control – The assumption of DLN as baseline and the selection of 25 ppmvd as the performance level for DLN will be provided.

- 3.4.3 Recent BACT Decisions – This discussion includes a summary of recent BACT decisions, the performance level and averaging times required for simple cycle, combined cycle, aeroderivative, peaking units, as well as the required combustion or control equipment. The assumption in this analysis regarding the performance level of an SCR (2 ppmdv) is also discussed.

- 3.4.4 Alternative Fuels – Issues affecting the control of NO_x from turbines burning syngas or refinery gas are discussed.

- 3.4.5 Greenhouse Gases (GHG) – Methods for capturing carbon that might be used for turbines and their impact on NO_x emissions as well as the contribution of the SCR to GHG emissions are discussed.

3.4.1 *Peaking Units*

Installing SCR on peaking units could be much less cost effective (higher values) than presented in Table 3-5 because the costs and emission reductions for peaking units were based on the assumption that they operate half the year. This level of operation is very high for a peaking unit. For a turbine to be considered a peaking unit under the U.S. Acid Rain program it must average a load of 10 percent or less over the previous 3 years and not have any one year with the load more than 20 percent. However, based on a sample of turbine permits, hour limitations ranged from 1,300 to 6,300 hours per year for turbines. These permits did not specifically identify these turbines as peaking units in all cases; however, operating less than continuously in order to provide electricity when needed, seems to meet the intent of a peaking unit. In another U.S. EPA document, 200 hours per year was cited as typical for peaking units. Also, some turbines operate as load following or mid-load units that operate more frequently than peaking units, yet still much less than a full year. There are several possible operational load schemes in which turbines can operate. ERG selected 50 percent load (4,200 hours) as a worst case for peaking plants and to account for some of the other load operating practices. The operating costs of an SCR installed on a peaking unit operating fewer hours (less than 4,200 hours) will not be decrease proportional to the decrease in emissions reduced – notice the difference in costs in Table 3-3 between the peak and base units. The annual costs are reduced by 37 percent for the peaking unit, but the emissions are reduced by 50 percent, resulting in an increase cost effectiveness (cost per tonne) of 47 percent.

Given that 50 percent load could be five times higher (50 percent versus 10 percent) than the actual load of many peaking plants, the cost effectiveness shown in Table 3-5 for peaking units would be much higher if 10 percent load had been used as the estimate for peaking units. In addition to the emissions reductions being less for peaking units, the annual costs would also be less because of less catalyst used. Making these adjustments to the cost effectiveness value would result in a cost effectiveness of $34,300 and $30,500 for the 50 MW and 113 MW peaking units, respectively. When determining the level of control for non-base load turbines to be placed in a regulation or permit, it is important to consider this assumption of the analysis.

Also, because peaking units have many startups and shutdowns, and there is a lag in the time for an SCR to become fully effective when it starts up, the amount of emissions reduced are likely to be smaller than what was estimated in this study. SCR systems are often bypassed until an appropriate catalyst operating temperatures are reached. The information in the literature is mixed on how much an SCR will lag the startup of a turbine. It likely varies with the technology and setup of the system. Given the colder ambient temperature in Alberta the lag time is likely longer than in a warmer climate. Lag time cold be from 5 to 25 minutes. Also, emissions of NO_x from the turbine during startup and shutdown are typically higher than during normal conditions. The values shown for cost effectiveness ($/tonne) are very likely to be a low estimate of the cost per tonne given that in reality fewer emissions will be reduced during startup and because many peaking units operate less than 4,200 hours per year.

Startup and shutdown of turbines, especially peaking units is important to consider when determining peaking unit levels of control. Some permits or regulations exclude periods of startup and shutdown from the compliance determination of the outlet concentration limit; others allow for excess emissions during these periods, but limit the number of startups or the annual emissions (tones/yr).

3.4.2 *Baseline Level of Control*

One important aspect of the BACTEA analysis is the assumption that was made on the baseline level of control. In the analysis it was assumed that DLN burners would be standard on any turbine purchased for electric generation. For the large, heavy-duty turbines, this is clearly true. For example, the GE Model 7FA, Siemens model SG T5-8000 H, and the Mitsubishi model M501F3 are all manufactured with DLN systems as standard equipment. For many of the smaller combustion turbines that are less than 50 MW, DLN is not standard, but is provided as an optional configuration. However, there are units as small as 22 MW (e.g., Solar Titan 250) that come with a standard DLN. Because 25 MW is the smallest turbine being considered in this analysis, the smaller turbines that do not come with DLN as standard are a small portion of the population being studied in this BACTEA analysis.

In the United States, a New Source Performance Standard (NSPS), *Standards of Performance for Stationary Combustion Turbines*, 40 CFR Part 60, Subpart KKKK was promulgated in July 2006 requiring new turbines constructed after February 18, 2005 to meet certain restrictions. The turbines being analyzed in this BACTEA would be subject to a limit under Subpart KKKK of either 15 or 25 ppm at 15% O_2, depending on their size in units of heat input per hour. Using the efficiencies assumed by the U.S. EPA in the development of the NSPS, turbines less than 110 MW would be subject to the 25 ppm limit and those greater than 110 MW would be subject to a 15 ppm requirement, although the exact size of the turbine in MW would vary depending on the efficiency of the specific turbine. Regardless of the size of the turbine, it is clear that a facility in the United States would have to purchase a turbine with a DLN burner, wet injection, or install an SCR to meet the regulation. DLN burners are the most cost effective of these options as illustrated in the 1999 U.S. Department of Energy report discussed below.

During the review process, CASA reviewers questioned the baseline level of control assumption, citing aeroderivative peaking units as a class of turbines where the assumption did not hold. Aeroderivative turbines are often selected for peaking units and the concern was that this type of turbine does not often come with DLN. There are DLN-like burners for aeroderivative turbines from some manufacturers, which may be referred to as dry low emissions (DLE) that use similar lean, premix fuel systems as the DLN systems. These systems, either because of the size or design of the turbine, do not achieve NO_x output concentration levels as low as the DLN burners in the non-aeroderivative turbines (25 ppmv generally the lowest guaranteed by vendors for DLE in aeroderivative versus 9 ppmvd for DLN in some models of turbines), but they result in a relatively high percent reduction of NO_x. For example, on the GE models of aeroderivative turbines (LM1600, LM2000, LM2500, LM2500+, and LM6000) with DLE installed, GE guarantees an outlet concentration of 25 ppmvd at 15% O_2. This represents an 80 to 89 percent reduction of NO_x emissions depending on the model. Testing has also shown outlet concentrations as low as 15 ppmvd from the 50 MW GE LM6000, a 93 percent emissions reduction, but this is not guaranteed by GE. On these same models, GE guarantees the same outlet concentration (25 ppmvd at 15% O_2) using water or steam injection and an even lower 15 ppmvd using water or steam injection on their LM2500 model. Other manufacturers have similar guarantees. For example, the Rolls Royce line of aeroderivatives has a 25 ppmvd guarantee for their DLE and water injection controlled aeroderivative turbines. Pratt and Whitney does not appear to have DLE burners for their aeroderivative turbines, but can achieve 25 ppmvd outlet concentration with wet injection.

Although the baseline assumption does not perfectly address the smaller sized turbines, including aeroderivative, since combustion controls (i.e., DLN or DLE) are not always standard, controls for these turbines are readily accessible and relatively inexpensive, such that the baseline level of performance (25 ppmvd) can easily be achieved. In a 1999 report written for the U.S. Department of Energy, adjusted to 2007, costs for these control techniques are given for a 25 MW turbine as $282/tonne for DLN, $1,324/tonne for wet injection, and $4,763/tonne for SCR.[10] The values were also calculated for a 150 MW turbine resulting in $164/tonne for DLN, $640/tonne for wet injection, and $2,706/tonne for an SCR. The wet injection system used in this 1999 analysis was for a system that controlled to 42 ppm; a wet injection system controlling to a more modern 25 ppm outlet concentration would likely have the effect of reducing the $1,324/tonne and $640/tonne values (increasing its cost effectiveness), because of the additional emissions reduced would overwhelm the little additional costs. Adjusting the wet injection emission reductions to 25 ppm and assuming no additional costs, the cost effectiveness values for wet injection would be $1,176/tonne for a 25 MW turbine and $581 for a 150 MW turbine. Although it is the decision of the regulatory agency whether these values are considered reasonable for their particular situation, absent additional non-quantifiable factors, ERG believes that regulatory agencies would consider these cost effectiveness values reasonable. The results of recent BACT analyses for turbines at electric generating facilities support our belief; no permit was found with “no control” as the BACT level of control. See discussion on the results of our permit review.

Given the U.S. regulation, turbines meeting the regulatory levels (< 15 ppmv for turbines greater than about 110 MW and <25 ppmv for turbines less than about 110 MW and greater than 25 MW) are likely to become even more available and common in North America and globally in the future. Therefore, baseline level of control assumption of 25 ppmv is a reasonable assumption.

3.4.3 *Recent BACT Decisions*

Because the BACT and BACTEA analyses are very similar and the U.S. EPA maintains lists of BACT analyses, ERG identified several recent BACT analyses to use in the analysis of BACTEA for gas turbines. To identify recent BACT analyses, ERG searched the National Combustion Turbine database maintained by the U.S. EPA Region 4 New Source Review staff, searched in the U.S. EPA’s RACT/BACT/LAER Clearinghouse database, and searched the Internet. The results are grouped by peaking units, simple cycle turbines and combined cycle turbines.

Peaking Units

Table 3-5 lists the BACT analysis results from recent PSD permits for turbines that are most likely new peaking turbines. These were assumed to be being peaking units because the database or permit listed an hour or fuel limitation, or they were specifically identified as being a "peaking unit." One peaking unit is a combined cycle with an operating hour limit, but the rest are simple cycle turbines. The majority of the turbines on this table had a NO_x limit of 9 ppmvd at 15% O_2, including the three most recently permitted ones. DLN burners were considered BACT for five out of the eight permits; wet injection for one of the permits; and SCR for two of the permits.

In reviewing the BACT data for peaking units it is important to consider if there are any peaking turbines that would not be subject to BACT requirements; units that would have low enough emissions that BACT would be avoided. BACT is a preconstruction requirement for certain emission sources and are a part of a Prevention of Significant Determination (PSD) permit. PSD permits are required for new sources (completely new facilities that did not exist previously) that have an annual potential to emit [generally, the maximum hourly emissions every hour of the year (8,760 hours/yr)] of 100 tons per year of a criteria pollutant (NO_x, SO_2, CO, PM/PM_{10}, VOC, and lead). If a turbine is being added to an existing facility that already has PTE of a criteria pollutant greater than 100 tons per year, BACT would apply if the PTE of NO_x from the new turbine was greater than 40 tons per year.

The applicability requirements of PSD permitting is very complex and there are several other factors that can be considered. The information included here is meant to provide enough detail to aid the understanding of some of the limitations of the permit review in Table 3-5. The primary limitation that is important in the context of the BACTEA is that there are some turbines depending on their size, operating hours and outlet NO_x concentration that would not be subject to BACT. In general, these are small peaking units. Table 3-6 includes several scenarios in which BACT would not be required. Therefore, no additional control would be required unless they are subject to another rule. The scenarios in Table 3-6, are based on simple situations with the turbines being the only source of NO_x being added. For new facilities, the turbine shown would have estimated emissions of 100 tons per year; and for turbines added to existing facilities they would emit 40 tons per year.

Table 3-5. BACT Control Levels for Probable Peaking Units

Mode	Year	Turbine Model	Combustion Turbine Size (MW)	Limited Hours per Combustion Turbine	State	Number of Combustion Turbines	NO_x Limit (ppmvd @ 15% O_2)[1]	Averaging Time	BACT Control Method
SC	2007	GE 7EA	89	2,000[2]	OK	2	9	3-hr	DLN
SC	2007	GE 7EA	89	2,000	OK	4	9	3-hr	DLN
SC	2006	PG7241 FA	190	3,390	FL	1	9	24-hr	DLN
SC	2005	Pratt & Whitney FT8 (Twin Pack)[3]	60	2,000	FL	5	25[4]	24-hr	WI
SC	2004	GE LM 6000[3]	50	5,840	FL	2	5	24-hr	SCR
CC	2003		275[5]	6,250	IL	2	3.5 / 2.5	1-hr / 24-hr	DLN + SCR
SC	2003	GE 7FA	170	2,000	GA	3	9	3-hr	DLN
SC	2003	GE 7FA	170	3,000	SC	2	9 (12 w/PA)		DLN

CC – combined cycle; SC – simple cycle; DLN – dry low NO_x; WI-Wet Injection; SCR-selective catalytic reduction; BACT- best available control technology; hr- hour; ppmvd- parts per million by volume on a dry basis; and PA-power augmentation.

[1] Most permits had concentration limits specified in ppmvd @ 15% O_2 although some only specified ppm. For the purposes of this table, it has been assumed all concentration limits are in ppmvd @ 15% O_2.

[2] Limited by a fuel limit equivalent to about 2000 hours/year for each turbine.

[3] Aero-derivative turbine model.

[4] This limit appears to be unique in this group. Florida had initially drafted the permit with a lower limit (5 ppmvd) and requiring an SCR and the applicant filed for a Petition for Formal Administrative Hearing. The result of this filing is the limit shown. No information was found on the justification, however, the small size of the turbines and the hourly limit may have provided support for the change to a limit of 25 ppmvd.

[5] This capacity includes the additional power output of the steam turbine.

Table 3-6. Gas Turbines Likely Not to be Subject to BACT

Total Maximum Capacity Size (MW)	Maximum Total Annual Operating Hours	Construction	Outlet NO_x Concentration (ppmvd)
41	8400	New facility	15
25	8400	New facility	25
60	1000	New facility	87
30	2000	New facility	87
24	1000	Added to existing major facility	87
12	2000	Added to existing major facility	87
146	1000	New facility	35
73	2000	New facility	35
59	1000	Added to existing major facility	35
29	2000	Added to existing major facility	35
207	1000	New facility	25
104	2000	New facility	25
83	1000	Added to existing major facility	25
49	4200	New facility	25
41	2000	Added to existing major facility	25
20	4200	Added to existing major facility	25
345	1000	New facility	15
173	2000	New facility	15
138	1000	Added to existing major facility	15
82	4200	New facility	15
69	2000	Added to existing major facility	15
33	4200	Added to existing major facility	15

"New facility" – indicates that the turbine(s) is installed at a new facility where no other equipment is currently located; or the turbine(s) could be installed at an existing facility that has an existing PTE of any criteria pollutant of less than 100 tons per year (minor source).
"Added to existing major facility" – indicates that the turbine(s) is installed at an existing facility that has an existing PTE of one or more criteria pollutants of greater than 100 tons per year (major source).

To illustrate the use of this table, the installation of two turbines of 40 MW each will be used as an example. The "Total Maximum Capacity" column can reflect the capacity of one or more turbines. In this example, the Total Maximum Capacity is 80 MW. In order to install these two turbines in the U.S. without triggering BACT, the company would have to limit the number of hours the turbines operate. If the turbines have a DLN burner or wet injection they will likely emit NO_x at about 25 ppmvd. Based on Table 3-6, at 25 ppmvd outlet NO_x concentration, the turbines would have to be installed at a new facility and limited to about a 2,000 total operating hours. The capacity shown in the table for this scenario is 104 MW; therefore, the company could limit their hours a little higher than 2,000 hours and still not be subject to BACT (at 80 MW, less than 2,600 hours would keep the turbines out of a BACT review). If the company wanted to install the two turbines at an existing major facility (the existing facility already has a PTE of a criteria pollutant greater than 100 tons per year), they would have to limit the total hours of operation to about 1,000 hours in order not to be subject to a BACT requirement. If these limitations on the operating hours are not workable for the company, they may be able to find a better wet injection technology or DLN burner to achieve a lower outlet NO_x concentration. For example, if the company could lower the NO_x to 15 ppmvd, these two turbines could be installed at a new facility with a total operating hour limit of about 4,200; or they could be installed at an existing major facility with an operating hour limit somewhat higher than 1,000 hours (at 80 MW, less than 1,725 hours would avoid a BACT review).

Other Simple Cycle Turbines

Table 3-7 lists other simple cycle turbines that did not have hour limits listed in the data source. Some of these could be peaking units. Note that five of these turbines are aeroderivative turbines and all are required to install and operate an SCR. In fact, eight out of 12 permits listed in Table 3-7 require SCR to be installed. The most recent BACT for these turbines was set at 9 ppmvd at 15% O_2, although the most common performance level was 5 ppmvd at 15% O_2 the lowest was 3 ppmvd and the highest was 15 ppmvd.

Combined Cycle Turbines

Table 3-8 lists all of the BACT analyses found for combined cycle turbines. All of these permits require SCR as BACT (or LAER). The most common performance level in these permits and in 10 out of the most recent 13 is 2 ppmvd at 15% O_2. The highest level was 5 ppmvd at 15% O_2 that was included in three permits in 2003. Based on the data collected from 2002, only one of the 42 combined cycle BACT permits issued did not require SCR. In 2001, four out of 64 permits issued did not require SCR. It would be difficult to refute that SCR on a combined cycle turbine is BACT or BACTEA given no additional issues or situational factors exist.

Some combined cycle permits include annual hour limits to restrict the time when the turbine bypasses the HRSG and operates in simple cycle mode. This is significant because the SCR is built into the HRSG. Therefore, for these times, a second limit is often given for simple cycle operation that is not controlled by the SCR. Not all permits had an hour limit for simple cycle operation but did include a NO_x concentration limit for simple cycle operation. However, duel concentration limits have not been typical in the last 6 years. Only three permits out of the 33 identified by ERG issued in 2003 through present include an outlet concentration limit for both combined and simple cycle operation modes; and no permits in the last 4 years have had them. (One permit in 2005 included two concentration limits but this was to address the issue of transient load and was not for bypass of the SCR.)

Table 3-7. BACT Control Levels for Other Simple Cycle Turbines

Mode	Year	Turbine Model	Combustion Turbine Size (MW)	State	Number of Combustion Turbines	NO_x Limit (ppmvd @ 15% O_2)[1]	Averaging Time	BACT (LAER) Control Method
SC	2005	GE 7EA	80	KS	1	9		DLN
SC	2004	GE-LM 6000[2]	49	CA		3.5	3-hr	SCR (LAER)[3]
SC	2003	GE 7EA	82	CO	1	9 / 100	24-hr; SU / SD	DLN
SC	2003	GE 7EA	83.5	SC	3	9		DLN
SC	2003	GE 7FA	170	TX	1	3		SCR (LEAR)
SC	2003	GE F6B	35	TX	2	3.5		SCR (LAER)
SC	2003	GE-LM 6000[2]	48	TX	4	5		SCR
SC	2003	GE-LM 6000[2]	50	TX	1	5		SCR
SC	2003	GE-LM 6000[2]	50	TX	1	5		SCR
SC	2003	Pratt & Whitney FT8 (Twin Pack)[2]	55	WA	2	5	3-hr	SCR
SC	2003	Siemens 5	182.6	VA	4	15		DLN
SC	2003	SW 501F	180	TX	6	3.5		SCR

CC – combined cycle; SC – simple cycle; DLN – dry low NO_x; SCR-selective catalytic reduction; BACT- best available control technology; LAER-Lowest Achievable Emission Reduction; hr- hour; ppmvd- parts per million by volume on a dry basis; and SU/SD-this acronym was used in a permit database and its meaning is unknown, it was presented to accurately represent that the limits have an additional condition.

[1] Most permits had concentration limits specified in ppmvd @ 15% O_2, although some specified only ppm. For the purposes of this table, it has been assumed all concentration limits are in ppmvd @ 15% O_2.

[2] Aero-derivative turbine model.

[3] Many of the California Air Districts define BACT more consistent with the U.S. EPA federal definition of LAER. This permit indicates that this is "BACT"; however, since this is consistent with LAER in other locations of the United States we have listed it as "LAER".

Table 3-8. BACT Control Levels for Combined Cycle Turbines

Mode	Year	Turbine Model	Combustion Turbine Size (MW)	State	Number of Combustion Turbines	NO_x Limit (ppmvd @ 15%O_2)[1]	Average Time	BACT (LAER) Control Method
CC	2008	Siemens SGT6-5000F	312	CT	2	2	1-hr	DLN + SCR (LAER)
CC	2008	"F" Class	172	OR		2	3-hr	DLN + SCR
CC	2008	GE 7FA	150	FL		2	3-hr	DLN + SCR
CC	2008	M501G CTG		FL	3	2	24-hr	DLN + SCR
CC	2008	Siemens SGT6-5000F, GE 7FA, or GE207FA	312, 180, or 286	VA	2	2		DLN + SCR
CC	2008			LA	2	4	annual	DLN + SCR
CC	2007	SW 501G	245	NY		2	3-hr	DLN + SCR (LAER)
CC	2007	SW V84.3A	170	CA	2	2	3-hr	SCR (LAER)[3]
CC	2006	PG7241 FA	170	FL	1	2	24-hr	DLN + SCR
CC	2005	SW 501F	180	UT	2	2; 25 (transient load conditions)	3-hr	DLN + SCR (LAER)
CC	2005	GE 7FA	170	FL	4	2	24-hr	DLN + SCR
CC	2005	GE LM 6000[2]	50	NY	1	2.5	1-hr	SCR
CC	2005	GE 7FA	170	FL	2	2.5	24-hr	DLN + SCR
CC	2004	GE 7FA	170	UT	2	2.25 / 9; (25 transient load conditions)	3-hr CC / 18-hr SC	DLN + SCR
CC	2004	GE 7FA	170	NC	3	2.5 or 3.5 depends on history and PM levels	24-hr	SCR
CC	2004	SW 501FD	170	GA	2	3.5		SCR
CC	2003	Alstom GTX100	43	CA	1	2	1-hr	DLN + SCR (LAER)[3]
CC	2003	GE-PG7241FA	181	CA	1	2	3-hr	DLN + SCR (LAER)[3]
CC	2003	GE 7F or SW 501F	170	CA	2	2.5	1 hr	SCR (LAER)[3]

Table 3-8. BACT Control Levels for Combined Cycle Turbines (Continued)

Mode	Year	Turbine Model	Combustion Turbine Size (MW)	State	Number of Combustion Turbines	NO_x Limit (ppmvd @ 15%O_2)[1]	Average Time	BACT (LAER) Control Method
CC	2003	GE 7F or SW 501 F	170	CA	2	2	1 hr	SCR (LAER)[3]
CC	2003	GE PG 7241FA	153	CA	2	2.5	1 hr	SCR (LAER)[3]
CC	2003	GE 7FA	170	WA	1	2.5	1-hr	DLN + SCR
CC	2003	GE 7FA	170	WA	4	2.5	3-hr	DLN + SCR
CC	2003	"F" Class	180	OR	4	2.5	8-hr	DLN + SCR
CC	2003			OR	2	2.5	8-hr	DLN + SCR
CC	2003	SW 501FD	170	FL	2	2.5	24-hr	DLN + SCR
CC	2003	SW 501FD	170	FL	2	2.5	24-hr	SCR
CC	2003	GE 7FA	170	GA	4	2.5		DLN + SCR
CC	2003	GE 7FA	170	GA	2	2.5		SCR
CC	2003	GE 7FA	170	FL	4	2.5 / 9-15	24-hr; CC / SC	DLN + SCR; SC mode limited to 1000 hours
CC	2003	GE 7FA	170	FL	4	2.5 / 9-15	24-hr; CC / SC	DLN + SCR; SC mode limited to 1000 hours
CC	2003	GE 7EA	80	NE	2	3.5	3-hr	SCR
CC	2003	GE 7FA	170	TX	3	5		SCR
CC	2003	GE 7FA	170	TX	2	5		SCR
CC	2003	SW 501G	250	TX	2	5		SCR
CC	2003	SW 501F	180	WA	2	2	3-hr	DLN + SCR
CC	2003	SW 501F	180	WA	1	2	3-hr	DLN + SCR

CC – combined cycle; SC – simple cycle; DLN – dry low NO_x; SCR-selective catalytic reduction; BACT- best available control technology; LAER-Lowest Achievable Emission Reduction; hr- hour; PM-particulate matter; and ppmvd- parts per million by volume on a dry basis.

[1] Most permits had concentration limits specified in ppmvd @ 15% O_2, although some specified only ppm. For the purposes of this table, it has been assumed all concentration limits are in ppmvd @ 15% O_2.

[2] Aero-derivative turbine model.

[3] Many of the California Air Districts define BACT more consistent with the U.S. EPA federal definition of LAER. This permit indicates that this is "BACT"; however, since this is consistent with LAER in other locations of the United States we have listed it as "LAER".

Miscellaneous Considerations

Several items should be noted on Tables 3-5, 3-7, and 3-8. Information presented in these tables is from several sources, including summary databases. Data obtained from summary databases are more subject to human error because people have manually entered information from the permit. For PSD permits collected by ERG, a more thorough review was conducted. For items included in the summary databases that looked unusual, we attempted to acquire the permit and investigate the issue, but were not always successful. We completed data where possible. For instance, if we had the model number but no size for the gas turbine, we completed the size with a default. Size can vary for a turbine model, but not by a large amount. One item we did not change without verification is the BACT control method. Often "SCR" is listed as the control method (as opposed to "DLN + SCR") for turbines that are thought to have DLN as standard equipment, such as the GE model 7FA. We did not make this change because it was uncertain if the DLN was not listed because it was not considered part of BACT, if it was an oversight, if the BACT was conducted assuming the baseline of DLN, or if these turbine models were built in a previous year that did not have DLN. Most, if not all, of the permits checked for this indicated that both DLN and SCR used together were BACT. In those cases we revised the information in the table.

For most BACT limits found, ppmvd at 15% O_2 is listed as the emission limit used to state the performance level of BACT. Some permits use limits of pounds per hour, pounds per million Btu, or tonnes per year instead of ppmv. When there was no ppm limit, limits in other units are shown. Even for permits that had outlet concentration limits, many also had limits in other units. Usually this helped to regulate other factors; for instance, situations where different operating modes were used. Peak operation (100 percent load), power augmentation, oil firing, and periods of startup, shutdown, and malfunction were typical issues being addressed with the various units. It is important to realize that a ppmv limit does not limit the emissions of turbines to a specific emission rate. The amount of NO_x emitted from a 40 MW and 250 MW turbines are very different, even if they both have outlet concentrations of NO_x of 9 ppmv. Peaking and base load turbines will have very different annual emissions profiles even if they are the same size and subject to the same outlet concentration limit.

Startups, shutdowns, and malfunctions are usually addressed in regulations and permits. One method that is used is to exclude periods of startup, shutdown, and malfunction from the compliance determination of the outlet concentration limit, but include those events in the calculation of an annual emissions limit (tonnes/yr). The annual limit can roll on a monthly or quarterly basis so there are periodic checks instead of waiting until 12-months have gone by. Startup, shutdown, and malfunction provisions often specify what can be considered a startup, shutdown, or malfunction and what is considered a violation of the outlet concentration or other short term limit.

The averaging time for any limit is also important. Different averaging times can average out operational variability and are used with various units for addressing different operating models. Permits often use different averaging times for different limits. For example, a 3-hour average is used for the outlet concentration and a monthly average is used for the pound per hour limit. The most common averaging times appear to be 3-hour and 24-hour. Although the most stringent limit found is a 1-hour averaged limit for an outlet concentration limited to 2 ppmvd at

15% O_2. Longer averaging times do exist, including monthly and annually; but only one annual averaging time was found in the permits reviewed and no monthly limits.

Summary

To set the specific emission limit values to be used in permits or regulations, there are many things to consider and address. Based on our review of the costs and BACT limitations in the U.S., it appears that there could be four or more subcategories of turbines each with their own regulatory requirements, including:

1. Small Peaking units and/or units with few operating hours. The cost effectiveness for peaking units is much higher than the other categories of turbines. Smaller peaking units and ones that would have fewer operating hours would have even a higher cost effectiveness for an SCR then those shown in Table 3-4 for peaking units. Also, there is no evidence that this class of turbine has been required to install SCR or meet specific emission levels, since these peaking units emit low enough emissions to avoid BACT requirements. Therefore, Table 3-5 does not represent this category. It may be useful to define a “Small Peaking Unit” category that would be based on size, operating hours, emissions or a combination.

2. Peaking units that are large enough or operate enough to be represented by Table 3-5. These units have been required to meet more stringent requirements than category 1. However, the cost effectiveness for SCR is still relatively high, as seen in Table 3-4.

3. Other simple cycle turbines. The permit data seems to indicate that simple cycle turbines have not been historically limited to as low an outlet concentration as combined cycle, but recent (within the last 3 years) permits have not been identified for simple cycle.

4. Combined cycle turbines (possibly two categories one with and one without duct burners). The permitting history also indicates that combined cycle turbines have been addressed as a separate category. Also, sizes of turbines may be necessary to consider in the regulation.

The Texas Commission on Environmental Quality (TCEQ) has published what they consider BACT for gas turbines. Table 3-9 shows how another group with regulatory responsibilities has addressed similar issues. This table does not address those peaking units that would not be subject to BACT.

Table 3-9. TCEQ Gas Turbine 2008
Best Available Control Technology (BACT) Requirements

Source Type	Minimum Acceptable Control for NO_x	Control Efficiency or Details
Gas-Fired Turbine Less than 2500 hours per year	9 - 15 ppmvd at 15% O_2	Dry Low NO_X burner
Gas-Fired Turbine Simple Cycle	5.0 ppmvd at 15% O_2	Dry Low NO_X burner, water, or steam, SCR
Gas-Fired Turbine Combined Cycle	2.0 ppmvd at 15% O_2, 24-hr average	Dry Low NO_X burner, water, or steam, SCR
Gas-Fired Turbine Combined Cycle with Duct Burner	2.0 ppmvd at 15% O_2, 24-hr average	Dry Low NO_X burner, water, or steam, SCR

3.4.4 *Alternate Fuels*

The sulfur content in pipeline quality natural gas is generally low (10 to 30 ppmv in the fuel), but distillate oil, refinery gas, as well as some low-Btu fuel gases such as syngas or landfill gas have sulfur contents that can complicate SCR catalyst choices. For sulfur-bearing fuels that produce greater than 1 ppmv SO_3 in the flue gas, the catalyst operating temperature range narrows to 315° to 400°C (600° to 800°F). Sulfur from fuel can also react with SCR ammonia to form ammonium bisulfate (NH_4HSO_4) and ammonium sulfate [$(NH_4)_2SO_4$]. The ammonium salts clog and foul the HRSG.[7]

The current designs of DLN burners are not appropriate for syngas and refinery gas fired combustion turbines. Due to low flame temperatures and fuel composition variability, DLN systems for syngas are particularly technically challenging. Several non-OEM suppliers are working on lean pre-mix combustion systems for syngas.[15] Current technology limits dictate that OEMs revert to conventional diffusion flame combustion systems and to the injection of diluent for NO_x control (i.e., water or steam injection).

Hydrogen, a typical constituent of syngas and refinery gas, can initiate flashback and combustion failure. The lower caloric content of synthesis gas requires the combustor be able to process more than five times the fuel flow relative to a natural gas combustor. The quantity of diluent needed to achieve acceptable NO_x concentrations is equal to or greater than the quantity of syngas.[16] Full fuel burnout and linear temperatures can be achieved by co-firing natural gas during startup and system upsets. Refinery gas has a higher calorie content than syngas so this is less of an issue for this fuel, but it does contain a significant amount of hydrogen.

In an IGCC facility, the gasification and combined cycle technologies are integrated so that the syngas generated by gasification becomes the fuel burned in the combustion turbine. Steam produced in the gasification area is piped to the HRSG, adding to the steam produced in the combined cycle power block and augmenting steam power generation (similar to a duct burner). Nitrogen from the air separation unit that is not used in an oxygen-blown gasifier is usually injected into the combustion turbine for NO_x reduction (the nitrogen is a diluent that cools the flame) and for power augmentation (the additional mass flow increases power generation). Emission limits for IGCC facilities permitted in the U.S. range from 15 - 25 ppmv.[17]

IGCC combustion turbines are capable of firing low heat value fuels with varying hydrogen content. Currently, four coal-based IGCC units are in operation around the world, with 30+ unit-years of experience. The essential characteristics of these plants are shown in Table 3-10. SCR has not been applied to any of these IGCC plants.

Table 3-10. Coal-Based IGCC Units

Project Name / Location	Combustion Turbine	Net Output MW	Startup Date
Wabash River/ Vigo County, IN	GE 7 FA	262	October 1995
Tampa Electric/ Polk County, FL	GE 7 FA	250	September 1996
Nuon Power/ The Netherlands	Siemens V 94.2	253	January 1994
ELCOGAS/ Puertollano, Spain	Siemens V 94.3	300	December 1997

The sulfur content of syngas and refinery gas raises SCR feasibility issues such as catalyst poisoning, and plugging and corrosion problems with the HRSG. SCR has been proposed in permit applications. One operating unit (Nuon Power in the Netherlands) and two sources not yet operating (Cash Creek Generation LLC in Henderson County, KY and Duke Energy in Edwardsport, IN) are considering using an SCR.[17] Impurities contained in the feedstock or formed during the gasification process can be removed from the syngas more effectively and efficiently before it is burned in the combustion turbine.

Several IGCC units are in service at oil refineries, using petroleum residuals as feedstocks. Three heavy oil IGCC plants in Italy entered full commercial service in 2001. Each Italian IGCC plant is equipped with acid gas and sulfur recovery technologies to clean the gas before it is burned in the turbine and then exhaust sent to the SCR. Other heavy oil IGCC plants are in operation in the United States (Texas), Singapore, and Japan. Table 3-11 shows two Italian units and one Japanese unit equipped with an SCR.[18]

Table 3-11. Heavy Oil-Based IGCC Units

Project Name / Location	Acid gas removal tech./ Total sulfur (H_2S + COS)	Combustion Turbine	Net Output MW	Start–up Date
ISAB Spa/ Sicily, Italy	MDEA <15 ppmv	Siemens V 94.2K	510	July 1999
Api Energia/ Falconara, Italy	Selexol unit <30 ppmv	ABB 13 E2	260	February 2000
Negishi refinery/ Japan	scrubbing+ADIP Lurgi OxyClaus 99.8% recovery	MHI 701 F	342	June 2003

Gasification is also being used on Canadian tar sands. Conventional crude oil is normally extracted from the ground by drilling oil wells. Because heavy oil and bitumen flow very slowly, if at all, toward producing wells under normal reservoir conditions, the sands must be extracted by mining or in-situ methods such as steam assisted gravity drainage (SAGD).

Recovering and refining tar sands are very energy intensive processes. Historically large amounts of natural gas have been used to provide the steam and power. Gasification of the less valuable and heavier tar sands processing residues such as petroleum coke or asphaltenes can provide the required steam, power, and hydrogen.

The most popular gas turbine utilized at oil sands facilities over the past decade has been the General Electric Frame 7EA, a nominal 85 MW unit. Two of these units are installed at the Syncrude Aurora Mine, two are installed at the Albian Sands Muskeg River Mine and one is planned for the CNRL Horizon Project. These units have also been utilized at heavy oil and In-situ (includes CSS and SAGD) projects such as Primrose and will be used at the OPTI/Nexen Long Lake Project. Some oil sands developments have also used larger gas turbines such as the facility at MacKay River and the Suncor Plant. [19]

DLN may not be appropriate for distillate oil combustion turbines. Distillate oil burns at a flame temperature that is approximately 150°F higher than that of natural gas and produces higher NO_x emissions. Many duel-fuel combustion gas turbines are equipped with steam or water injection systems.

3.4.5 *Greenhouse Gas (GHG)*

Electricity is required to operate the SCR. Pumps and fans inject the reducing agent into the flue gas upstream of the catalyst bed. In addition to electricity requirements, the SCR catalyst reactor increases the back-pressure on the turbine, which decreases the turbine power output by approximately 0.5 percent.

It is assumed that this reduction in the turbines electricity generation capacity and electricity requirements are compensated for by combusting more fuel in a comparable natural gas fired combustion turbine. To calculate additional fuel usage, the efficiency of each model turbine was incorporated. Table 3-12 shows the required electricity and the resulting greenhouse gas emissions.

Table 3-12. SCR Required Energy and Global Warming Impact

Model	Model Unit Description	Required Electricity (MWh-yr)	Required Additional Fuel (MMBtu-yr)	CO_2 (tonne/yr)	CH_4 (tonne/yr)	N_2O (tonne/yr)	CO_2e [1] (tonne/yr)
1	50 MW - Peak	1,092	9,930	2,408	0.19	0.07	2,432
2	113 MW - Peak	2,426	20,207	4,900	0.38	0.13	4,950
3	50 MW - Base	2,184	19,860	4,816	0.38	0.13	4,865
4	113 MW - Base	4,851	40,414	9,801	0.77	0.27	9,900
5	175 MW - Base	7,518	69,348	16,818	1.31	0.46	16,987
6	300 MW - Base	12,810	115,054	27,902	2.18	0.76	28,183

[1] 100-year Global Warming Potentials (GWP) from U.S. EPA - 2008 Inventory of Greenhouse Gas Emissions and Sinks

Carbon Capture

In order to mitigate the production of GHGs from the control and combustion devices, carbon capture is currently being explored by utilities worldwide. The methods being explored include the following: absorption, adsorption, membrane separation, frosting, and biological fixation.

Chemical absorption is the most mature technology used to capture CO_2. In chemical absorption, CO_2 in flue gases is captured by special solvents and separated. When the CO_2 loaded solvents are heated to 212° - 280°F (100° - 140°C), CO_2 is released, compressed, and transported to a storage site while the CO_2 lean solvent is reused. The heat to regenerate the solvent, electricity to drive pumps, and compressors consume a significant amount of energy. Absorption solvents are traditionally aqueous solutions, but they can also be used in a dry form. Chemical solvents, such as alkanolamines, have been utilized for many years by the chemical and petrochemical industries.

The chemical solvents that are currently used to capture CO_2 also interact chemically with NO_2 and SO_x to form stable chemical compounds that cannot be decomposed during the normal solvent regeneration process. As a result, NO_2 is also removed from the flue gases together with the CO_2. Chemical solvents can remove up to 85 percent of the NO_2 in the flue gas. Developments of new solvents optimization CO_2 selectivity and cost are expected.

The thermal efficiency of a combined-cycle plant with CO_2 capture is lower than that of an equivalent conventional combined-cycle plant. The reduction in efficiency due to CO_2 capture is estimated in the order of 7 percent.[20] Efficiency penalties have also been imposed to power plants in the past, as a result of the implementation of acid gas removal technologies, such as FGD and SCR. This reduction in net power generation capacity can be compensated either by expanding the existing plant and combusting more fuel.

Activated carbon is the second carbon capture technology being explored. Activated carbon is the most common adsorbent with many uses in industry. Depending on how the carbon is activated it can be tuned to select target chemicals of a particular size. The use of adsorbents to capture CO_2 has been investigated for both pre-combustion and post-combustion systems. Ideal adsorbents should exhibit high transfer rates, high regenerability, and high capacity for CO_2 in a flue gas environment 140° - 180°F (60° - 80°C). For adsorption to be economical compared to other capture methods, improved adsorbents are needed with properties that allow for thermal swing adsorption (TSA) or low pressure swing adsorption (PSA).

Membrane separation is a third capture technology that is used in a variety of industrial applications including gas separation from natural gas streams. The use of membranes to capture CO_2 from flue gas is being actively pursued using two types of membranes separation: molecular sieve membrane and solution-diffusion membrane. Research is focused on increasing the permeability and selectivity of the membranes.

Low temperature separation is also under exploration for carbon capture. Low temperature separation is used by industry to isolate and purify chemicals. This separation method is based on the fact that different chemicals and gases freeze at different temperatures. While technically possible to remove CO_2 from combustion gases, the energy required to cool large volumes of flue gas would be prohibitively expensive. The conceptual design was presented in papers by Ecole de Mines de Paris (Paris School of Mines) researchers at the 6th and 7th Greenhouse Gas Control Technology Meetings. Ecole de Mines de Paris reports that the total energy required to capture 90 percent of CO_2 would require a 10.8 percent reduction in net plant electrical output.

Finally, biological carbon capture describes the process of using plants to capture, react and fix CO_2 through the photosynthesis reaction. Plants, trees and aquatic flora account for tonnes of carbon dioxide uptake each year. Researchers at EniTecnologie in Italy conducted a field experiment of CO_2 uptake by algae in an engineered lagoon. Another way to grow algae is with a photobioreactor. The capitol and operational costs for photobioreactors are higher but they allow for a faster growth rate, and better temperature control than lagoons. GreenFuel is a Massachusetts company that designed a capture system using algae in photobioreactors.[21]

Regardless of which capture technology is used, CO_2 separated from the exhaust gas must eventually be stored or otherwise prevented from entering the atmosphere. Permanent storage solutions include underground injection into geologic formations such as depleted oil or gas fields for enhanced oil recovery, deep aquifers, or deep coal seams. In these cases, CO_2 must be prepared for pipeline transportation. Moisture and oxygen must be removed to prevent oxidation of the pipe, and other flue gas species (NO_x and SO_x) must be reduced to prevent corrosion and phase separation. Therefore, there may be a need for higher removal.

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4.0 EVALUATION OF FUTURE TECHNOLOGIES, FUELS, AND CONTROL TECHNOLOGIES

This chapter discusses future electricity generating technologies, alternative fuels, and future control technologies. In order to evaluate these three topics, ERG conducted an Internet literature review. Details of this Internet literature review are provided below.

4.1 Introduction of Technology Review

Prior to commencing the Internet literature review, several guidelines were established for conducting the review:

- Technologies/fuels must be commercially available in 2020. In order to meet this timeline, pilot and demonstration projects for these technologies/fuels must be functional at the current time or within the next few years. Technologies/fuels that are still at the research level or a theoretical concept were not included; for example, commercial nuclear fusion was not included.

- Technologies/fuels must be fundamentally new. Traditional technologies/fuels that are already utilized at the commercial level and will only see incremental improvements over time were not reviewed.

- Technologies/fuels that are infeasible in Alberta were excluded. For instance, ocean-based forms of renewable energy (i.e., power generated from tides, waves, marine currents, ocean thermal energy, and salinity gradients) were not included because they are physically unavailable in Alberta.

- Although the results of the Internet literature review are broad and far-reaching, it is possible that the findings are not absolutely comprehensive. However, attempts were made to address and research all possible technologies/fuels.

The results of the Internet literature review are presented in Section 4.2 (Future Generating Technologies and Alternative Fuels) and Section 4.3 (Future Control Technologies).

4.2 Future Generating Technologies and Alternative Fuels

The future generating technologies and alternative fuels that were investigated are presented in this section. Future generating technologies and alternative fuels are discussed together in this section due to some overlap between the two topics. A total of 15 different future generating technologies and alternative fuels were identified during the Internet literature review. These future generating technologies and alternative fuels included the following:

- Integrated gasification combined cycle (IGCC)
- Supercritical and ultra-supercritical technology
- Pressurized fluidized bed combustion (PFBC)
- Advanced circulating fluidized bed (CFB)

- Carbon capture and storage (CCS)
- Lignite fuel enhancement
- Coal liquefaction
- Coal gasification
- Gasification and thermal recovery
- Advanced biomass gasification
- High-pressure oxy-fuel power generation
- Photovoltaic (PV) concentration
- Integrated solar combined cycle
- Phosphoric acid fuel cell (PAFC) systems
- Compressed air energy storage (CAES)

Each of these future generating technologies and alternative fuels are briefly described below. Some representative projects are included in Table 4-1. The table includes the following information for each project: name, location, fuel used, generation size, operation start date, and current status. Detailed page-length summaries for each of the projects presented in Table 4-1 are included in Appendix C. The summaries of each project include the following information: project description, controls, benefits, plant efficiency, emissions, problems, costs, and references. In some instances, this information could not be located.

4.2.1 *Integrated Gasification Combined Cycle (IGCC)*

IGCC is a technology that turns fuel into synthetic gas (i.e., syngas) and then removes impurities from the syngas prior to combustion in a turbine. This results in lower emissions of SO_2, particulates, and mercury. It also results in improved combustion efficiency compared to conventional pulverized coal. The gasification process can produce syngas from high-sulfur coal, heavy petroleum residues, and biomass. The plant is termed "integrated" because its syngas is produced in a gasification unit in the plant and has been optimized for the plant's combined cycle turbine. The gasification process produces heat, which is reclaimed by steam "waste heat boilers" and then used by steam turbines.

Currently, the main obstacle for IGCC is its extremely high capital cost, which can be higher than $3,593/kW. In general, IGCC is viewed as too risky for private investors and typically requires government subsidies. Some studies have indicated that the capitol costs for IGCC plants to be 20 to 47 percent higher than traditional coal plants. IGCC plants also have potential water contamination problems due to the gas cleaning process. For example, coal gasification wastewater has an average pH of 9.8, similar to the pH of hand soap (pure water has a pH of 7.0). There have also been some chronic reliability issues with plant gasifiers.

There are currently only two commercial-size coal-based IGCC power plants in the United States, with a few more overseas. Additional facilities have been proposed in the United States and Europe but have not begun construction. Currently in the United States, five IGCC projects have been cancelled and four IGCC projects are on hold.[11] There are currently no commercial "carbon capture-ready" IGCC plants. These plants require a different design configuration to efficiently separate carbon from the syngas prior to combustion. IGCC technology will likely be available at the commercial-scale in 2020. There are no commercial "carbon capture-ready" technologies for IGCC plants.

Table 4-1. Summary of Investigated Future Electric Generating Technologies and Alternative Fuels

Technology	Name	Location	Fuel	Size	Operation Start	Status
Integrated Gasification Combined Cycle (IGCC)	Buggenum	The Netherlands	Coal	253 MW	1994	Operating
	Wabash River Power Station	West Terre Haute, Indiana, USA	Coal	262 MW	1995	Operating
	Polk Power Station	Tampa, Florida, USA	Coal	250 MW	1996	Operating
	Puertollano	Spain	Coal	300 MW	1997	Operating
	Pinon Pine	Reno, Nevada, USA	Coal	100 MW	1998	Failed
	Priolo Gargallo	Italy	Coal	521 MW	1999	Operating
Advanced Steam/Water Cycle, Ultra-Supercritical (USC) Boiler	Advanced 700 °C PF Power Plant	Europe	Coal	400 MW	Scheduled - 2013	Planning/Under construction
	Huaneng Yuhuan Power Plant	Xiaqingtang, Zhejiang, China	Coal	4x1,000 MW units	Scheduled – 2 units in 2007, 2 units in 2008	Under construction/ operating
Pressurized Fluidized Bed Combustion (PFBC)	Vartan	Sweden	Coal	135 MW	1991	Operating
	Tidd	Brilliant, Ohio, USA	Coal	70 MW	1991	Operating
	Escatron	Spain	Coal	80 MW	1992	Operating

Table 4-1. Summary of Investigated Future Electric Generating Technologies and Alternative Fuels (Continued)

Technology	Name	Location	Fuel	Size	Operation Start	Status
Pressurized Fluidized Bed Combustion (PFBC) (Cont.)	Wakamatsu	Japan	Coal	71 MW	1994	Operating
	Cottbus	Germany	Coal	80 MW	1999	Operating
	Osaki	Japan	Coal	250 MW	2000	Operating
	Karita	Japan	Coal	350 MW	2001	Operating
	Wangjia Power Plant	China	Coal	15 MW	2006	Operating (trial basis)
Advanced Circulating Fluidized Bed (CFB)	Western Greenbrier Co-Production Demonstration Project	Rainelle, West Virginia, USA	Coal mining waste materials	100 MW	Scheduled – December 2009	Cancelled (June 14, 2008)
Carbon Capture and Storage	Elsam Power Plant	Esbjerg, Denmark	Coal	420 MW	March 2006	Active
	FutureGen	Mattoon, Illinois, USA	Coal	275 MW	Scheduled – 2012	Uncertain – U.S. Department of Energy funding withdrawn (January 29, 2008)
Lignite Fuel Enhancement	Great River Energy (GRE) Spiritwood Station	Spiritwood, North Dakota, USA	Lignite	62 MW base (37 MW peak)	Scheduled - 2010	Under construction
Coal Liquefaction Technology	Gilberton Coal-to-Clean Fuels and Power Co-Production Project	Gilberton, Pennsylvania, USA	Coal waste	41 MW	Unknown	Project delayed for unspecified reasons

Table 4-1. Summary of Investigated Future Electric Generating Technologies and Alternative Fuels (Continued)

Technology	Name	Location	Fuel	Size	Operation Start	Status
Coal Gasification	Power Systems Development Facility (PSDF)	Wilsonville, Alabama, USA	Coal	Unknown	1996	Operating
	Advanced Hydrogen Turbine Development Project	Orlando, Florida, USA	Coal	Unknown	Phase 1 (2006-2007); Phase 2 (2008-2011)	Under development
Gasification and Thermal Recovery	Philadelphia Biosolids Recycling Center	Philadelphia, Pennsylvania, USA	Various wastes	Unknown	Unknown	Operating
Advanced Biomass Gasification	Emery Recycling	Salt Lake City, Utah, USA	Municipal solid waste, animal waste, and agricultural residues	Unknown	2001	Research and development project; candidate for future commercialization
	Sebesta Blomberg	Roseville, Minnesota, USA	Barley residues and corn stover	Unknown	2001	Research and development project; candidate for future commercialization
	Alliant Energy	Lansing, Iowa, USA	Corn stover	Unknown	2001	Research and development project; candidate for future commercialization
	United Technologies Research Center	East Hartford, Connecticut, USA	Clean wood residues and natural gas	Unknown	2001	Research and development project; candidate for future commercialization
	Carolina Power and Light	Raleigh, North Carolina, USA	Clean wood residues	Unknown	2001	Research and development project; candidate for future commercialization
High-Pressure Oxy-Fuel Power Generation	Clean Energy Systems (CES) Kimberlina Power Plant	Bakersfield, California, USA	Natural gas and various alternate fuels	20 MW (to expand to 80 MW in 2008)	2005	Active
Photovoltaic (PV) Concentration	EUCLIDES-THERMIE Plant	Tenerife, Canary Islands, Spain	Solar	480 kWp (kW- peak)	1998	Active

Table 4-1. Summary of Investigated Future Electric Generating Technologies and Alternative Fuels (Continued)

Technology	Name	Location	Fuel	Size	Operation Start	Status
Integrated Solar Combined Cycle	Hassi R'Mel	Hassi R'Mel, Algeria	Solar and natural gas	150 MW turbine and 25 MW array	2010	Under construction
Phosphoric Acid Fuel Cell (PAFC) Systems	HydroGen/ASHTA Chemicals	Ashtabula, Ohio, USA	Hydrogen	400 kW	2008	Active
Compressed Air Energy Storage (CAES)	Alabama Electric Corporation (AEC) CAES	McIntosh, Alabama, USA	Natural gas	110 MW	1991	Active
	Norton Energy Storage (NES)	Norton, Ohio, USA	Natural gas	2,700 MW	Announced 2001; construction not started	Planned
	Iowa Stored Energy Park (ISEP)	Fort Dodge, Iowa, USA	Wind and fossil fuels	200 MW (CAES); 100 MW (wind farm)	2011	Planned

4.2.2 *Supercritical and Ultra-Supercritical Technology*

New pulverized coal combustion systems utilizing supercritical and ultra-supercritical technology operate at increasingly higher temperatures and pressures and therefore, achieve higher efficiencies than conventional units and realize significant CO_2 reductions. Supercritical steam-cycle technology has been used for decades and is becoming the system of choice for new commercial coal-fired plants in many countries. Recent plants built in Europe and Asia use supercritical boiler-turbine technology, and China has made this standard on all new plants larger than 600 MW. Research and development is under way for ultra-supercritical units operating at even higher efficiencies (i.e., potentially up to 50 percent). The introduction of ultra-supercritical technology has been driven over recent years in countries such as Denmark, Germany, and Japan, to achieve improved plant efficiencies and reduce fuel costs. Research is focusing on the development of new steels for boiler tubes and on high alloy steels that minimize corrosion.[49] Ultra-supercritical technology will likely be commercially available in 2020.

Supercritical phase is the state of a substance when there is no clear distinction between the liquid and the gaseous phase (i.e., they are a homogenous fluid). Water reaches this state at a pressure above 22.1 megapascals (MPa). The cycle medium becomes a single-phase fluid with homogeneous properties, thereby eliminating the need to separate steam from water in the boiler. Current technology permits efficiencies that exceed 45 percent, depending on cooling conditions. Options to increase the efficiency above 50 percent in ultra-supercritical power plants rely on elevated steam conditions, as well as on improved process and component quality. It should be noted that the information reviewed did not indicate whether the efficiency was on a net or gross basis. Additionally, the documents reviewed reflect overseas installations, which are likely using lower heating value to assess efficiency, where as higher heating value is used in the U.S. If put on the same basis (higher heating value) the reporting efficiencies would be reduced.

Advanced steel types must be used for components such as the boiler and the live steam and hot reheat steam piping that are in direct contact with steam under elevated conditions. Steam conditions up to 30 MPa/600 °C/620 °C can be achieved using steels with 12 percent chromium content, while conditions up to 31.5 MPa/620 °C/620 °C can be achieved using Austenite, which is a proven, but expensive, material. Nickel-based alloys (e.g., Inconel) could potentially permit conditions up to 35 MPa/700 °C/720 °C, thereby yielding efficiencies up to 48 percent.[17]

Details on how the efficiency percents were calculated were not provided in the references reviewed. These efficiencies will vary depending on gross versus net heat capacity of fuel and the lower heating value versus higher heating value to the fuel.

4.2.3 *Pressurized Fluidized Bed Combustion (PFBC)*

PFBC technology uses a combustion process similar to that of atmospheric fluidized bed combustion (AFBC), but the boiler operates at higher than atmospheric pressure (0.5 to 2 MPa), the gas is cleaned downstream from the PFBC boiler, and the gas is expanded in a gas turbine. Coal and sorbent are introduced at the bottom of the PFBC boiler, where the coal is burned and the sorbent reacts with the SO_2 to form $CaSO_4$. A series of cyclones or other type of hot gas cleanup device removes the particles from the flue gas, which expands in a gas turbine and is then directed to the stack through an HRSG and a conventional particulate removal device (i.e., ESP or baghouse).

PFBC has multiple advantages over the conventional pulverized coal or AFBC technology, including the following:

- High SO_2 and NO_x removal
- Ability to use low-quality fuels
- Compact design suitable for shop fabrication and modular construction
- Easier to construct at an existing power plant than an AFBC because of the small space requirements
- Potential for achieving higher plant efficiency (up to 45 percent) than conventional pulverized coal or AFBC (36.5 percent)
- Lower capital costs than IGCC or pulverized coal with wet scrubbers.

Field demonstration of the PFBC technology has proven to achieve more than 90 percent removal of SO_2. PFBC technology is in various phases of demonstration in four different power plants in the United States. Currently, research and development is being conducted related to the following PFBC areas: hot gas cleanup technology; coal and sorbent preparation and feed systems; and effects of the PFBC boiler-gas contaminants on gas turbine performance, reliability, and life expectancy. Projections of capital costs for PFBC range from $1,150 to 1,250/kW.[50] Recently, suppliers have not been actively pushing technology, but instead, favoring IGCC and supercritical PCC.

4.2.4 *Advanced Circulating Fluidized Bed (CFB)*

The advanced CFB technology is used to convert coal mining waste materials into electricity. In addition, steam for industrial use and/or district heating can be generated using this process. Some of the ash generated as a by-product from this process can also be sold, and the remaining quantity of ash can be used for remediation of acid water formation. The advanced CFB incorporates an inverted cyclone separator and mid-support structure designs to reduce assembly time (6 to 8 weeks), lower material costs (60 percent less than structural steel tonnage), and provide a smaller footprint (30 to 40 percent) than conventional designs.

Waste coal and limestone are simultaneously fed to the CFB, which raises steam by passing water through water walls lining the CFB. The limestone removes the bulk of the sulfur in the coal feedstock, while solids are entrained and recirculated via the cyclone separators to enhance limestone and carbon utilization. An economizer located downstream of the cyclones recovers additional heat from the flue gas. SNCR, flash dryer absorber, and a baghouse provide additional control of NO_x, SO_2, PM, and Hg. Steam from the CFB boiler drives a steam turbine.

A demonstration plant for this technology type was started in mid-2007 in West Virginia with DOE funding assistance. It was assumed that this technology would be made commercially available by the 2012-2015 timeframe. Due to funding issues, however, this demonstration project was discontinued in early 2008; therefore, the future outlook is unclear.

4.2.5 *Carbon Capture and Storage (CCS)*

CCS is an approach for reducing carbon emissions by capturing CO_2 from large point sources, such as fossil fuel power plants, and storing it instead of releasing it into the atmosphere. Although CO_2 has been previously injected into geological formations for various purposes, the long-term storage of CO_2 is a relatively untried concept. Storage of the CO_2 is envisioned either in deep geological formations, in deep ocean masses, or in the form of mineral carbonates. In the case of deep ocean storage, there is a risk of greatly increasing ocean acidification, which already exists due to the excess CO_2 already in the atmosphere and oceans. Geological formations are currently considered the most promising sequestration sites.

CCS applied to a modern conventional power plant could potentially reduce CO_2 emissions by approximately 80 to 90 percent compared to a plant without CCS.[22] Capturing and compressing CO_2 requires much energy, however, and would increase the fuel needs of a coal-fired plant with CCS by about 25 percent. The cost of this extra fuel, as well as storage and other system costs, are estimated to increase the costs of energy from a power plant with CCS by 21 to 91 percent, depending on the specific circumstances. These cost estimates apply to purpose-built plants near a storage location; applying the CCS technology to pre-existing plants or plants far from a storage location would be even more expensive.

This technology will likely not be commercially available before 2020 because considerable fundamental research still needs to be conducted.

4.2.6 *Lignite Fuel Enhancement*

An estimated 45 percent of world's coal reserves consist of lignite. Although lignite is comparatively inexpensive and low in sulfur, it is a low-grade coal with a typical moisture content in the range of 25 to 40 percent and a low relative energy output compared to other types of coal. Even though current lignite-fired power plants are designed to burn high moisture coal, a reduction of up to 15 percent moisture content in lignite would result in significant improvements.

Lignite fuel enhancement focuses on this moisture-content reduction. Fuel moisture has many effects on unit operation, performance, and emissions. As fuel moisture decreases, the fuel's heating value increases so that less coal needs to be fired to produce the same electric power, thereby reducing the burden on the coal handling system. Drier coal is also easier to convey, which reduces maintenance costs and increases availability of the coal to the handling system. When the crushed coal is gravity-fed into bunkers, the drier coal flows more readily than the wet coal, causing fewer feed hopper bridging and plugging problems. In addition, drier coal is easier to pulverize, so that less mill power is needed to achieve the same coal fineness. Finally, more complete drying of coal can be achieved in the mill, which results in an increased mill exit temperature, better conveyance of coal in the coal pipes, and fewer coal pipe-plugging problems.[38]

The lignite fuel enhancement process uses waste heat from the power plant condenser to drive a bubbling fluidized-bed coal dryer.[13] The dryer removes approximately 25 percent of the coal's moisture before the coal is fed into the power plant boiler. This low-moisture coal not only improves the power plant efficiency, but it also results in reduction of emissions of Hg, NO_x, SO_2, PM, and CO_2. This technology will be commercially available by 2012.

4.2.7 *Coal Liquefaction*

The coal liquefaction process is also known as coal-to-liquids (CTL). The coal liquefaction process converts coal into synthetic liquid fuels and requires coal to be in contact with a hydrogen environment at high temperature and pressure. There are various liquefaction processes that are in use around the world, including: Bergius process, Fischer-Tropsch process, Solvent Refined Coal (SRC) process, and Sasol process.

All of these liquid fuel production methods release CO_2 during the conversion process—far more than is released in the extraction and refinement of liquid fuel production from petroleum. If these methods were adopted to replace declining petroleum supplies, CO_2 emissions would be greatly increased on a global scale. To avoid excess emissions, carbon capture methods would need to be used in conjunction with liquefaction methods. Different coal liquefaction technologies are currently commercially available around the world; additional new liquefaction technologies are also being tested.

4.2.8 *Coal Gasification*

Coal gasification is the process of breaking down coal into smaller molecular weight molecules by subjecting it to high temperature and pressure using steam and measured amounts of oxygen. This leads to the production of syngas, a mixture primarily consisting of CO and hydrogen (H_2), but it can also contain other gaseous constituents depending on the type of feedstock. This syngas can be cleaned up relatively easily, thus leading to a cleaner burning fuel than traditional coal. Syngas produced in this way can be burned in a gas turbine rather than in a boiler used to drive a steam turbine.

The main component of a gasification-based system is the gasifier. A gasifier converts hydrocarbon feedstock into gaseous components by applying heat under pressure in the presence of steam. The amount of air or oxygen available inside the gasifier is carefully controlled so that only a relatively small portion of the fuel burns completely. This “partial oxidation” process provides the heat. Instead of burning, most of the carbon-containing feedstock is chemically broken apart by the gasifier’s heat and pressure, setting into motion chemical reactions that produce syngas. See Section 4.2.1 above for additional gasification details.

4.2.9 *Gasification and Thermal Recovery*

Gasification is the thermochemical conversion of an organic solid into a combustible gas. Applying gasification technology to supply the auxiliary fuel for thermal oxidation eliminates the need for fossil fuel and stabilizes energy costs. Organic solids such as wood waste, plastic, sludge, and agricultural and municipal wastes can be used with this technology.

The gasifier converts waste wood or other biomass materials into a hot, combustible syngas, with calorific constituents consisting primarily of CO and H_2. This gasification process is usually carried out in an oxygen-starved environment. Hot syngas leaving the gasifier is directed to a thermal oxidizer/recovery unit where volatile gaseous constituents along with toxic and noxious wastes are destroyed. The resulting combustion products are primarily CO_2 and water vapor. After exiting the thermal oxidizer, the hot combustion products are then split and sent to an HRSG.

4.2.10 *Advanced Biomass Gasification*

Gasification is a thermochemical process that converts biomass materials into gaseous components. The resulting product of the gasification process is produced gas, which contains CO, H_2, CH_4, and other inert gases. Produced gas can be used as a source of alternate fuel in a variety of applications, including power generation. The main component of this process is called the gasifier. Biomass gasifiers are reactors that heat biomass in a low-oxygen environment to produce a fuel gas that contains from 20 to 50 percent of the heat content of natural gas. The produced gas or BioMethane is a renewable fuel with properties similar to natural gas. It is usually fed into existing gas boilers and/or gas turbine augments to generate electricity. There are various methods for advanced biomass gasification. The two most common methods are Bubbling Fluidized Bed (BFB) and CFB.

The BFB and CFB systems both utilize a “bed” (typically sand), which is then “fluidized” by high-pressure combustion air. The primary difference between the BFB and the CFB systems is that the BFB system normally operates in a reduced oxygen environment. In addition, BFB technology does not have as great an ability to absorb SO_2. BFB systems are normally selected to burn lower quality fuels with high volatile matter. CFB systems are a relatively new and evolving technology, which has become a very efficient method of generating low-cost electricity while generating electricity with very low emissions and environmental impacts. In a CFB combustion process, fuel is mixed with limestone and fired in a process resembling a boiling fluid. The limestone removes the sulfur and converts it into an environmentally benign powder that is removed with the ash.

4.2.11 *High-Pressure Oxy-Fuel Power Generation*

High-pressure, oxy-fuel combustion technology is based on mature, proven rocket technology. Zero emissions power plants (ZEPPs) can be built based upon the integration of this oxy-fuel combustion technology into conventional power generation systems. The advantages of ZEPPs include the following: compact and lower cost equipment; potential greater cycle efficiencies with advanced turbines; complete carbon capture and sequestration of the CO_2 effluent; and zero emissions (or ultra low emissions when the exhaust is vented to the atmosphere, as in a peaking power plant). The core of this technology is an oxygen-fuel combustor that burns a clean gaseous and/or liquid fuel with gaseous oxygen at near-stoichiometric conditions in the presence of recycled coolant (water, steam, or CO_2). The combustion products are primarily a mixture of steam and CO_2 at high temperature and pressure. Acceptable fuels include natural gas, syngas from coal, refinery residues, landfill gas, bio-digester gases, and renewable fuels such as glycerin from biodiesel production facilities.

The combustion products will drive conventional or advanced steam turbines or modified aero-derivative gas turbines operating at high temperatures and intermediate pressures. Every component in this process, except for the oxy-fuel combustor/gas generator and reheater, is commercially proven and is standard in power generation or other commercial application. Oxy-fuel combustion is also a very promising technology to be used in conjunction with CCS. The timeframe for this technology to be available on a commercial scale is 2017-2020.

4.2.12 *Photovoltaic (PV) Concentration*

PV cells convert sunlight into electricity and are being used mainly in Europe. Many of these plants are integrated with agriculture, and some use innovative tracking systems that follow the sun's daily path across the sky to generate more electricity than conventional fixed-mounted systems. There are no fuel costs or emissions during operation of the power stations.

PV cells are typically combined into modules that hold about 40 cells; about 10 of these modules are mounted in PV arrays that can measure up to several meters on a side. These flat-plate PV arrays can be mounted at a fixed angle facing south, or they can be mounted on a tracking device that follows the sun, allowing them to capture the most sunlight over the course of a day. About 10 to 20 PV arrays can provide enough power for a household; for large electric utility or industrial applications, hundreds of arrays can be interconnected to form a single, large PV system.

Some solar cells are designed to operate with concentrated sunlight. These cells are built into concentrating collectors that use a lens to focus the sunlight onto the cells. This approach has both advantages and disadvantages compared with flat-plate PV arrays. The main idea is to use very little of the expensive semiconducting PV material while collecting as much sunlight as possible. However, the use of concentrating collectors is limited to only very sunny areas, because the lenses must be pointed at the sun. Some concentrating collectors are designed to be mounted on simple tracking devices, but most require sophisticated tracking devices, which further limit their use to electric utilities, industries, and large buildings.

This technology is commercially available in some parts of the world, primarily in Europe. Further research is being conducted to increase the efficiency of PV cells and arrays.

4.2.13 *Integrated Solar Combined Cycle System (ISCCS)*

ISCCS plants combine solar energy with traditional combined-cycle plants. Generally, a parabolic trough concentrating solar power array is used in conjunction with the combined-cycle gas turbine plant, which cuts carbon emissions compared to a traditional power plant. The output from the solar array will be used to run the steam turbine to produce electricity. ISCCS plants use solar heat for steam generation and gas turbine waste heat for preheating/superheating the steam. The ISCCS plants can approximately double steam turbine capacity. However, when solar energy is not available, then the steam turbine must run at part load, which reduces overall efficiency.

Some studies have indicated the cost for an ISCCS plant to be about 3.1¢ (US)/kWh as compared to 2.4¢ (US)/kWh for a traditional fossil fuel-fired combined cycle plant.[18] It is expected that this technology will be commercially available by 2015.

4.2.14 *Phosphoric Acid Fuel Cell (PAFC) System*

A PAFC is a type of fuel cell that uses liquid phosphoric acid as an electrolyte, housed by a ceramic matrix such as silicon carbide. The electrodes are made of carbon paper coated with a finely dispersed platinum catalyst.

PAFC technology has the most extensive track record for operational experience of any of the fuel cell technologies, with more than 300 systems (mostly 100 to 200 kW) installed worldwide. These fuel cells are not affected by CO impurities in the hydrogen stream. These cells need to be operated continuously since phosphoric acid solidifies at 40 ºC. Currently, 100 kW cells are being used in commercial and institutional settings for air and water heating. Some PAFC manufacturers have announced plans to upgrade the cells in the 1,000 kW range for distributed generation and cogeneration applications. This technology should be commercially available in the immediate future (i.e., 2008-2009).[5]

4.2.15 *Compressed Air Energy Storage (CAES)*

In a CAES process, air is compressed using low-cost, off-peak electricity, and stored in a deep underground geological formation (e.g., depleted gas wells, salt caverns, abandoned mines) for later use in generating electricity. When energy is needed, the stored air will be released, heated, and used to drive generating turbines. The electricity it produces can be used as needed, especially during high-demand peak hours. This process uses less fuel than a conventional combustion-turbine facility. Currently there are two operating CAES facilities in the world located in Germany and Alabama; an additional facility in Iowa is in the design phase.

Using off-peak electricity, air is compressed to around 1,000 pounds per square inch (psi) (or 70 times atmospheric pressure), which raises its temperature to more than 600 ºC (1,100 ºF). This is far too hot to pump underground, so the air is cooled to about 50 ºC (120 ºF). When the air needs to be released to generate electricity, it is reheated to turn the turbine. Unfortunately, this results in only 50 percent efficiency for the CAES process. However, current research indicates that an advanced adiabatic CAES (AA-CAES) process can be used, in which the heat that is removed from the air is also stored and later used to reheat the air as it is discharged to the turbine. The efficiency of the AA-CAES process could potentially increase to 70 percent. An AA-CAES demonstrator project is expected around 2012.[42] Commercial availability of this technology could be possible by 2020.

4.3 Future Control Technologies

The future control technologies that were investigated are presented in this section. A total of 10 different control technologies were identified during the Internet literature review. Most of these technologies focused on mercury control (either individually or with other pollutants). These future control technologies included the following:

- TOXECON™
- Membrane-based up-flow wet electrostatic precipitation
- Non-thermal plasma-based multi-pollutant control (electro-catalytic oxidation – ECO)
- Amended Silicates™
- Activated CI

- MerCAP^{TM}
- Mercury oxidation catalysts
- Low temperature mercury control (LTMC)
- Wet scrubbing mercury removal technology
- Advanced hybrid particulate collector (Advanced Hybrid^{TM})

Each of the future control technologies is briefly described below. Some representative projects are included in Table 4-2. Table 4-2 includes the following information for each control technology: target pollutant, location, emission reduction, project timeline, and current status. Detailed page-length summaries for each of the technologies presented in Table 4-2 are included in Appendix D. The summaries of each technology include the following information: technology description, benefits, problems, costs, and references. In some instances, this information could not be located. In general, the control technologies that are presented in this section are either commercially available at the present time or will be in the near future (i.e., in the 2010-2012 timeframe). Some of these control technologies are undergoing additional research and development in order to develop future improvements.

4.3.1 *TOXECON™*

The TOXECON™ technology was developed by EPRI. This control technology injects reagents and/or sorbents (including powdered activated carbon for mercury control, and others for NO_x and SO_x control) into the inlet duct of a secondary particulate control device that is downstream of the existing primary particulate control device. This configuration thus segregates the ash collected in the primary particulate control device from the ash/reagent/sorbent mixture collected in the secondary (downstream) particulate control device, preserving the marketability of the fly ash from the primary particulate control device. In the TOXECON™ process, the secondary particulate control device is generally a pulse jet fabric filter dust collector. EPRI is also currently in the process of demonstrating their new TOXECON II technology. Demonstration is currently underway with funding from DOE. The new technology is expected to reduce mercury emissions by up to 90 percent. Results from demonstration plants have also showed decrease in up to 30 percent of SO_2 emissions and up to 70 percent of NO_x.

4.3.2 *Membrane-Based Up-Flow Wet Electrostatic Precipitation*

ESPs use electrostatic forces to separate dust particles from exhaust fumes. A number of high-voltage, direct-current discharge electrodes are placed between grounded collecting electrodes. The exhaust flows through the passage created between the two sets of electrodes. The airborne particles receive a negative charge as they pass through the ionized files and get attracted to the grounded electrode and are removed from the exhaust stream. The ESP plates are then cleaned by vibrating or rapping either continuously or at predetermined intervals to remove the collected PM.

In most wet precipitators (both tubular and flat-plate), the collection surface normally has the form of a sheet of metal or plastic. Therefore, flushing liquid (typically water) passing over the surface tends to “bead” due to both surface tension effects, as well as the initial geometric surface imperfections. Because the flushing liquid cannot be uniformly distributed over the surface, this beading can lead to channeling and formation of “dry spots” of collected particles.

The resulting build-up of collected material causes the precipitator electrical performance to degrade. As a result, current flow is inhibited, which results in increased emissions from that section of the ESP.

Table 4-2. Summary of Investigated Future Control Technologies

Technology	Target Pollutant	Location	Emission Reduction	Project Timeline	Status
TOXECON^{TM}	Hg, SO_2, NO_x, PM	Presque Isle Power Plant (We Energies), Marquette, Michigan, USA	Hg – 90%; SO_2 – 30%; and NO_x – 70%	Start – April 2004; Completion – 2009	Under development
Membrane-Based Up-Flow Wet Electrostatic Precipitation	Hg, SO_3, $PM_{2.5}$	Bruce Mansfield Plant (Penn Power), Shippingport, Pennsylvania, USA	Hg (elemental) – 33%; Hg (oxidized) – 82%; Hg (particulate) – 100%; H_2SO_4 – 93%; $PM_{2.5}$ – 96%	Start – October 2002; Completion – March 2004	Testing completed
Non-Thermal Plasma-Based Multi-Pollutant Control (Electro-Catalytic Oxidation – ECO)	Hg, NO_x, SO_2, $PM_{2.5}$	R.E. Burger Generation Station (Ohio Edison), Akron, Ohio, USA	Hg – 80-90%; NO_x – 90%; SO_2 – 98%; $PM_{2.5}$ – 95%	Start – September 2001; Completion – September 2004	Testing completed
Amended Silicates^{TM}	Hg	Miami Fort Station (Duke Energy), North Bend, Ohio, USA	Hg – 40%	Test – 1st Quarter 2006	Testing completed
Activated Carbon Injection	Hg	Stanton Station (Great River Energy), Stanton, North Dakota, USA	Hg – >90% (chemically treated sorbents); 75% (non-treated sorbents)	Unknown	Testing completed
MerCAP^{TM}	Hg	Stanton Station (Great River Energy), Stanton, North Dakota, USA	Hg – 30-35%	Start – September 2003; Completion – September 2006	Testing completed
Mercury Oxidation Catalysts	Hg	Coal Creek Station (Great River Energy), Underwood, North Dakota, USA and J.K. Spruce Power Plant (City Public Service of San Antonio), San Antonio, Texas, USA	Hg – 12-98% oxidation (dependent upon catalyst type); 82% total Hg capture	Start – October 2002; Completion – April 2005	Testing completed
Low Temperature Mercury Control (LTMC)	Hg	R. Paul Smith Station (Allegheny Energy), Williamsport, Maryland, USA	Hg – 90%	Completion – December 2008	Testing currently being conducted

Table 4-2. Summary of Investigated Future Control Technologies (Continued)

Technology	Target Pollutant	Location	Emission Reduction	Project Timeline	Status
Wet Scrubbing Mercury Removal Technology	Hg	Endicott Station (Michigan South Central Power Agency), Litchfield, Michigan, USA and Zimmer Station (Duke Energy), Moscow, Ohio, USA	Hg – 51-79%	Start – October 2000; Completion – June 2002	Testing completed
Advanced Hybrid Particulate Collector (Advanced Hybrid^{TM})	$PM_{2.5}$	Big Stone Power Plant (Otter Tail Power company), Milbank, South Dakota, USA	PM – 99.9% (from 0.01 to 50 µm)	Start – October 2002; Completion – December 2005	Testing completed

Membrane-based up-flow wet ESP is a new type of wet precipitator that solves these problems. Instead of the traditional metal-collecting electrodes, fabric membranes are used. Tests indicate that membranes made from materials that transport liquid by capillary action are effective collection electrodes. Capillary flow promotes well-distributed water flow both vertically and horizontally, which is necessary for particle collection, removal, and transport. This solves a major historical problem with wet ESPs.

The cost of installation for a membrane-based wet ESP is significantly lower than the traditional ESP by as much as 60 percent. Also, the emission reduction efficiency of the membrane ESP is almost 20 percent higher than that of a traditional ESP.

4.3.3 *Non-Thermal Plasma-Based Multi-Pollutant Control (Electro-Catalytic Oxidation – ECO)*

The ECO system is a four-stage pollution control process that integrates several technologies to remove high levels of the primary air pollutants generated by coal-fired power plants. Following are the four stages of ECO technology:

- Stage 1—A dielectric barrier discharge reactor that oxidizes NO and elemental Hg
- Stage 2—An absorber that removes SO_2 and NO_2
- Stage 3—A WESP used to collect aerosols and fine particles
- Stage 4—A co-product treatment system for removal of Hg and ash from the liquid co-product stream prior to produce solid commercial grade fertilizer.

The ECO process is designed for installation downstream of an existing particulate collection device. The process utilizes a dielectric barrier discharge reactor to oxidize pollutants in the flue gas stream, followed by an NH_3 wet scrubber and wet ESP for removal of oxidized pollutants and by-products. The technology produces a marketable ammonium sulfate/ammonium nitrate fertilizer by-product.

Flue gas flows from the existing ESP or FF into an ECO reactor, where hydroxyl (OH) radicals and atomic oxygen (O) are formed by the reaction of high-energy electrons with water and oxygen molecules. These radicals then react with pollutants in the flue gas, oxidizing Hg to mercuric oxide (HgO), SO_2 to sulfuric acid (H_2SO_4), and NO_x to nitric acid (HNO_3). Approximately 90 percent of NO in the flue gas is oxidized to NO_2 and HNO_3 at this stage. Less than 10 percent of SO_2 in the gas is oxidized to form SO_3, which ultimately forms sulfuric acid. The oxidized flue gas is then sent to a double loop NH_3 scrubber where the final products are removed from the flue gas stream. In the upper loop of the scrubber, NH_3 is added to form ammonium sulfate and nitrate by-products by reaction with sulfuric and nitric acid. The lower loop of the scrubber cools and saturates the flue gas and concentrates the by-products for removal. Oxidation air is added to the lower loop to further oxidize sulfites and nitrites. The flue gas then enters a mist eliminator that removes entrained droplets from the gas prior to entering a WESP located in the top portion of the scrubber. The WESP collects acid aerosols, PM, and HgO, while collecting and recycling excess NH_3 as an aqueous solution. The soluble ammonium sulfate and nitrate by-products exit the scrubber in a bleed stream that is treated by filtration to remove ash and insoluble metal compounds. The stream is passed through an activated carbon absorption bed where mercury compounds are absorbed onto the bed. The purified by-product stream is then sent to a crystallizer where well-defined sulfate and nitrate crystals are formed by evaporation. The crystals are then dried and granulated to form a fertilizer by-product that could be marketable depending on quality and location. Pilot testing results indicate the ability of the ECO process to remove 90 percent of NO_x, 98 percent of SO_2, 80 to 90 percent of Hg, and 99.9 percent of PM_{10}.[14]

4.3.4 *Sorbent Injection*

Sorbent injection is a cheap alternative to remove mercury efficiently without installing expensive equipment. Sorbents are injected into the flue gas of coal-fired power plants to remove mercury. The gas-phase mercury in the flue gas contacts the injected sorbent and attaches to the surface. The sorbent with the mercury attached is then collected by the existing particle control device, such as an ESP or fabric filter. The type of particulate control equipment installed at the plant is a key parameter defining both the amount of sorbent that is required and the ultimate limitation of the amount of mercury that can be removed. Sorbent injection rate also depends on many other different variables, including the concentrations of Hg and acid gases (e.g., SO_2, NO_2, and HCl) in the combustion gas, Hg speciation (as Hg^0 or Hg^{2+}), temperature, fly ash properties, and the particle size, composition, reactivity, and capacity of the sorbent.[21] Two specific sorbents are described below.

Amended Silicates™

Amended Silicates™ is a powdered, noncarbon, mercury-control sorbent that is developed and manufactured by Amended Silicates, LLC. This new sorbent has the advantage that it removes mercury at least as efficiently as does activated carbon, but it does not contaminate the collected fly ash such that it can no longer be sold as a by-product. This feature is seen as a major cost advantage over activated carbon. Pilot tests at coal-burning power plants with mercury using Amended Silicates™ have shown mercury removal of up to 96 percent.[1]

Activated Carbon

In this injection method, Powdered Activated Carbon (PAC) sorbent is injected into the flue gas at a location preceding the particulate control device. The PAC sorbent binds with the mercury in the flue gas in the duct and in the particulate control device. Subsequently, the mercury-containing PAC is captured in the particulate control device. Activated carbons are nonhazardous, processed, carbonaceous products, having a porous structure and a large internal surface area. The main features of activated carbon sorbent are its low capital cost, low sorbent usage, low carbon usage, low energy consumption, ability to be safely disposed, and ease to retrofit to any system. Activated carbon as a sorbent is capable of removing up to 90 percent of the mercury in flue gas.[21]

4.3.5 *MerCAP™*

Mercury Control Adsorption Process (MerCAP™), developed by EPRI, uses noble metal gases to adsorb Hg. The general concept for MerCAP is to place fixed sorbent structures into a flue gas stream to adsorb mercury and then, as the sorbent surfaces becomes saturated, thermally regenerate the sorbent and recover the mercury. The fixed structures can be made of a sorbent or coated with a sorbent material such as activated carbon or metals that can amalgamate with mercury such as gold and silver. One example includes parallel gold-coated plates. Mercury forms an amalgam with the gold and is removed from the flue gas flowing past the plates. Some pilot test programs utilized an electroplated layer of gold on a stainless steel screen (substrate). The captured mercury can be subsequently sequestered using a carbon canister or cryogenic trap during thermal regeneration of the substrates. Recent work has shown that chemical desorption of the mercury from the gold-coated substrates is also a feasible regeneration technology. Field testing results have indicated over 90 percent mercury capture using MERCAP.[10]

4.3.6 *Low Temperature Mercury Control (LTMC)*

The LTMC process controls mercury by cooling the flue gas temperature to about 220°F and absorbing the mercury on the carbon inherent in the fly ash. However, the main concern with this process is the corrosion caused by acid condensation on equipment surfaces at reduced temperatures. Recent field studies have indicated a new technology that utilizes this concept for mercury removal, while eliminating the threat of related corrosion. The technology involves the low-temperature removal of mercury combined with removal of acid gases by reaction with magnesium hydroxide ($Mg(OH)_2$). It has been shown that $Mg(OH)_2$ is a highly effective reagent for the removal of SO_3 from flue gas. Additionally, research indicates that $Mg(OH)_2$ produced as a by-product in a Thiosorbic lime scrubber is a more effective reagent than commercial grade $Mg(OH)_2$, making it possible to utilize by-product from other scrubbers as reagent in the process. The $Mg(OH)_2$ is injected upstream of the air preheater to prevent corrosion of downstream equipment by SO_3 condensation. The flue gas temperature is cooled in the preheater to 220-250°F, allowing mercury to adsorb onto fly ash for collection in an existing particulate collection device.

Field testing results have concluded that over 90 percent Hg is removed by LTMC. Depending on the amount of $Mg(OH)_2$ injected into the system, 78 to 85 percent removal of SO_3 was achieved. Leaching tests were performed on the fly ash, and the captured mercury showed stability in pH ranges of 3 to7 and around 140° F. Also, LTMC costs an order of magnitude less than current systems that inject activated carbon.[14]

4.3.7 *Wet Scrubbing Mercury Removal*

Wet FGD systems are currently installed on about 25 percent of the coal-fired utility generating capacity in the United States, representing about 15 percent of the total number of coal-fired units. In a wet FGD, the flue gas enters a large vessel (spray tower or absorber), where it is sprayed with water slurry (approximately 10 percent lime or limestone). The calcium in the slurry reacts with the SO_2 to form calcium sulfite or calcium sulfate. A portion of the slurry from the reaction tank is pumped into the thickener, where the solids settle before going to a filter for final dewatering to about 50 percent solids. The calcium sulfite waste product is usually mixed with fly ash (at an approximate 1:1 ratio) and fixative lime (approximately 5 percent) and disposed of in landfills. Alternatively, gypsum can be produced from FGD waste, which is a useful by-product.

The Wet Scrubbing Mercury Removal technology relies on the addition of very small amounts of a liquid reagent to existing wet FGD units to achieve increased mercury removal along with traditional SO_2 removal. The benefits of this technology include cost effectiveness, multipollutant control, easy retrofit to existing units, and no impact on by-product disposal or usage (McDonald et al., 2003).

4.3.8 *Advanced Hybrid Particulate Collector*

The Advanced Hybrid Particulate Collector (AHPC) combines the best features of ESPs and baghouses in a manner that has not been done before. The AHPC concept consists of a combination of FF and ESP in the same control device, providing significant synergism between the two collection methods, both in the particulate collection step and in the transfer of the dust to the hopper. In particular, specific anticipated benefits of this approach are the following:

- Solves the problem of excessive fine-particle emissions found with traditional ESPs
- Provides ultra-high collection efficiencies (99.99 percent), even at high air-to-cloth (A/C) ratios
- Solves the problem of re-entrainment and re-collection of dust in conventional pulse-jet baghouses caused by close bag spacing and the effect of cleaning one row of bags at a time
- Solves the problem of chemical attack on bags, making it suitable to be used for all U.S. coals
- Requires significantly less total collection area compared to traditional ESPs or baghouses
- Is suitable for new installations or as a retrofit to existing installations.

A conventional ESP has two collecting or earth-grounded plates with a row of discharge electrodes between them. In contrast, the AHPC incorporates a row of pulse-jet filter bags centered in parallel between two rows of discharge electrodes. The inlet flue gas is directed to the area between the discharge electrode and the collecting plate on both sides of the filter bag row. The ESP ionization zone collects most of the PM on the plate. Particles escaping the ESP zone are collected on the surface on the filter bags. The flue gas then passes through the filter bags' cross-section, exiting to the AHPC clean air plenum.

In tests conducted to date, the AHPC has demonstrated collection efficiencies greater than 99.99 percent for PM between 0.01 and 50 μm and can operate at filter bag A/C ratios up to 12 cfm/ft^2, while requiring 60 to 75 percent fewer filter bags than a conventional baghouse. Because the FF provides ultra-high collection efficiency of fine PM, large ESP collecting plate areas are not required. This results in a smaller system footprint and lower capital costs than traditional particulate removal equipment.

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Appendix A

Summary Descriptions of Control Technologies

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Summary Descriptions of Control Technologies

NO_x Controls

ECOTUBE

The ECOTUBE™ system is a hybrid NO_x reduction system that combines separated overfire air (SOFA) and SNCR technologies. Retractable lance tubes that penetrate the boiler above the primary burner zone inject high-velocity air, which creates turbulent airflow and increases the residence time for the air/fuel mixture. The water-cooled ECOTUBEs are retracted from the boiler on a regular basis and cleaned to remove layers of soot and other depositions.[3]

Electric Charge Transfer (ECT) / Compartment Air Distribution Monitoring System (CADM)

The electric charge transfer (ECT) is a coal flow distribution and velocity measurement device. The compartment air distribution monitoring system (CADM) matches air flow and coal flow. The ECT and CADM systems are estimated by the vendor to achieve an 8 to 20 percent additional NO_x reduction.[4] The ECT can also monitor coal fineness on a mill basis for unburned carbon (UBC) or mill maintenance purposes. ECT can works in conjunction with other combustion controls.

Induced Flue Gas Recirculation (IFGR)

The Induced Flue Gas Recirculation (IFGR) technology flue gas recirculated and mixed with combustion air in the furnace. The resulting dilution in the flame decreases the temperature and availability of oxygen, thereby reducing thermal NO_x formation. The technology is also know as wind box FGR (WFGR). Flue gas recirculation can achieve a NO_x reduction efficiency of less 20%. This is because the ratio of thermal-NO_x to total NO_x emissions is relatively low in coal-fired plants. Excessive flue gas recirculation can also result in flame instability problems and increased steam temperatures.[3]

J-POWER ReACT

The ReACT process utilizes a moving bed of activated coke (AC) with ammonia injection to simultaneously remove SO_2, NO_x, and mercury from the flue gas. Spent activated coke from the adsorption process is regenerated and recycled back to the adsorber, while SO_2 rich gas is sent to a by-product recovery unit and processed into a saleable sulfuric acid or gypsum by-product. The ReACT system is designed to be placed downstream of an existing particulate control device. J-POWER EnTech reports ammonia injection into the regenerator is necessary if more than 50% NO_x removal is required. The ReACT technology has been installed on 14 commercial units to date, including 4 coal-fired utility boilers and other industrial flue gas applications. All commercial installations of the technology are located abroad, with the majority located in Japan.[6]

Low NO_x Burners

Low NO_x Burners (LNB) are designed to control the mixing of fuel and air to achieve what amounts to staged combustion. This staged combustion reduces both flame temperature and oxygen concentration during some phases of combustion that lowers both thermal NO_x and fuel NO_x production. Average NO_x reduction ranges from 35 to 45 percent for tangential- and wall-fired boilers equipped with LNB. Sub-bituminous-fired plants tend to operate at a higher NO_x removal efficiency than bituminous-fired plants.[7]

Mobotec ROFA and ROTAMIX

Mobotec provides a NO_x reduction system that combines OFA and SNCR technologies into an integrated system. The system uses a modified OFA system with improved mixing characteristics achieved through adding a rotation to the OFA. This system is called ROFA™ - Rotating Opposed Firing Air. ROTAMIX™, consisting of adding urea or ammonia injection into the ROFA air nozzles, can be added to the system. The extra mixing produced by combining the ROFA nozzles with the reagent injection results in improved mixing and a more homogeneous temperature profile in the boiler.[18] The ROFA/Rotamix system was installed on Carolina Power & Light's Cape Fear Unit 6, a 174 MW t-fired twin boiler burning medium sulfur, eastern bituminous coal. At full load NO_x reductions were 71 percent.[8]

Non-Selective Catalytic Reduction (NSCR)

Non-selective Catalytic Reduction (NSCR) technology is designed to simultaneously reduce NO_x and oxidize CO and hydrocarbons in the combustion gas to nitrogen, carbon dioxide, and water. The catalyst, usually a noble metal, causes the reducing gases in the exhaust stream to reduce NO_x at a temperature between 800 and 1,200°F. NSCR requires a low excess oxygen (1%) concentration in the exhaust gas stream to be effective because the oxygen must be depleted before the reduction chemistry can proceed.[9] Coal units cannot operate less than 1% excess air.

NO_xStar and NO_xStar Plus

The NO_xStar process continuously injects controlled quantities of an ammonia based reagent with relatively small amounts of hydrocarbon (typically natural gas) into the convective pass of an operating boiler. The injection grid is comprised of a permanent array of feed lances attached to the pendant heat transfer assemblies in the upper furnace gas pass. This arrangement optimizes reagent distribution throughout the flue gas. At the elevated temperatures (1600 to 1800°F) the hydrocarbon auto-ignites to form a plasma of free radicals that auto-catalyzes the reaction of ammonia and NO_x to produce nitrogen and water vapor. In 2004, NO_xStar was installed at a Tennessee Valley Authority (TVA) 200 MW t-fired boiler; no results have been published8 (8). To achieve the vendor advertized 50 percent NO_x reduction an instillation will likely require a single layer of in-duct catalyst (NO_x Star Plus).[10]

Operational Modifications

Changing certain boiler operational parameters can create conditions in the furnace that will lower NO_x production. Examples include: Burners out of service (BOOS), Low Excess Air (LEA), and biased burner firing (BF).

BOOS involves withholding fuel flow to all or part of the top row of burners so that only air is allowed to pass through. BOOS simulates air staging, or OFA conditions, and limits NO_x formation by lowering the oxygen level in the burner area. LEA involves operating at the lowest possible excess air level without interfering with good combustion, and BF involves injecting more fuel to some burners (typically the lower burners) while reducing fuel to other burners (typically the upper burners) to create staged combustion conditions in the furnace. Depending on boiler specific operating conditions, operational modifications could lead to a NO_x reduction efficiencies of 10 to 20 percent.[11] These controls are used as retrofits on existing units.

Optimization Software (Neural Networks)

Optimization generally relies on the use of a computer program to determine the optimum set points for a potentially large number of components in the fuel and air feed systems (e.g., mills, dampers, fans) to reduce NO_x emissions and improve heat rate. Industry experience shows that the use of optimization software can reduce NO_x emissions 5 to 30 percent.

Known suppliers of combustion optimization software include: Lehigh University (PEPCP), Radian & SCS (EPRI & PowerGen), Pegasus Technologies (Neural Networks), Praxis Engineers (PECOS), Pavillion Technologies (Process Insights), NeuCo (ProcessLink), DHR Technologies (TOPAZ), and ULTRAMAX.[12]

Oxygen Enhanced Combustion

Oxygen enhanced combustion (OEC) is a technique developed by Praxair to reduce NO_x emissions from PC-fired boilers, while improving combustion characteristics such as loss-on-ignition (LOI). The technique works by replacing a small portion (5 to 10 percent) of the combustion air in the primary combustion zone of specified burners with oxygen. In order to support this concept, Praxair is also developing an oxygen transport membrane (OTM) process that uses pressurized ceramic membranes for separation of oxygen from air. Two utility boilers, City Utilities' James River Unit 3 and Northeast Utilities' Mt. Tom Generating Station demonstrated a NO_x reduction of 40 percent.[13]

Overfire Air (OFA)

OFA is air that is injected into the furnace above the normal combustion zone. Generally when OFA is employed, the burners are operated at a lower than normal air-to-fuel ratio that reduces NO_x formation. OFA, which is frequently used in conjunction with LNBs, completes the combustion process at a lower temperature. Since the early to mid-1990's, boosted over fire air systems, referred to as advance over fire air (AOFA), began to be operated. The AOFA system adds air ports to several walls of the furnace, in addition to just the burner walls.

Reburn

In the reburning process, part of the boiler fuel input (typically 10-25%) is added in a separate reburn zone. Heat input to the main combustion zone is reduced while reburn fuel (natural gas, oil, or pulverized coal) is injected with either air or flue gas to create a fuel-rich zone. Existing boiler shapes or designs can reduce effectiveness of modifications. This technology is used as a retrofit technology.

NO_xOUT CASCADE (SNCR/SCR Hybrid)

The NO_xOUT CASCADE® process is a multistage system, employing both urea-based Selective Non-Catalytic Reduction (SNCR) and a compact Selective Catalytic Reduction (SCR) component. SNCR is typically applied in the furnace, where high temperatures initiate the breakdown of urea which leads to effective NO_x reduction. The SNCR technology is limited to temperatures high enough to insure low ammonia breakthrough. At high furnace temperatures performance can be lessened by competing reactions. The hybridized SNCR injects chemical in cooler regions where NO_x reductions improve. The ammonia slip feeds a compact or in-duct SCR reactor which further reduces NO_x while limiting the cost associated with a larger catalyst. A hybrid SNCR/SCR system can achieve an overall NO_x reduction of 65 to 75 percent.[16]

Selective Catalytic Reduction (SCR)

Selective catalytic reduction is a process that involves post-combustion removal of NO_x from flue gas utilizing a catalytic reactor. In the SCR process, ammonia injected into the flue gas reacts with nitrogen oxides and oxygen to form nitrogen and water vapor. The function of the catalyst is to effectively lower the activation energy of the NO_x decomposition reaction to about 375° to 750°F, depending on the specific catalyst and other contaminants in the flue gas.[17] A typical SCR removal efficiency on a PRB unit today is 60% with 2 ppm of ammonia slip. Some units can achieve 90% or more reduction depending on boiler load and hour of operation.

Selective Non-Catalytic Reduction (SNCR)

Selective non-catalytic reduction is based on a gas-phase homogeneous reaction that involves the injection of an amine-based compound into the flue gas within an appropriate temperature range for reduction of NO_x. In an SNCR system, NO_x reduction does not take place in the presence of a catalyst, but rather is driven by the thermal decomposition of ammonia or urea and the subsequent reduction of NO_x. Consequently, the SNCR process operates at higher temperatures than the SCR process. For the ammonia injection process, the necessary temperature range is 1,700 to 1,900°F.[17] On small coal-fired units (i.e., less than 200 MW), SNCR has been demonstrated to achieve NO_x reductions ranging from 25 to 50% with acceptable levels of ammonia slip. For larger boilers (i.e., greater than 300 MW), there are numerous challenges associated with applying SNCR. In particular, such boilers' large physical dimensions pose challenges for injecting and mixing the reagent with the flue gas.[17] PC-fired units have a much more limited furnace temperature window and poor lateral mixing, conditions that render SNCR less effective in these applications.[11]

Ultra Low NO_x Burners

ALSTOM Power under a U.S. DOE contract has developed the Ultra-Low NO_x (ULB) burner, a refinement of its TDS 2000 burner, for use on t-fired boilers. This can achieve NO_x emissions of 0.15 lb/MMBtu with bituminous coal and 0.10 lb/MMBtu with sub-bituminous coal. To date, burners incorporating aspects of the ULN design are installed in around 20-power plants firing sub-bituminous coal.[20]

Babcock and Wilcox and Fuel Tech under a U.S. DOE contract have been investigating improvements to low-NO_x burners for used in wall-fired boilers. The approach combined B&W's

DRB-4Z burner with Fuel Tech's NO_xOUT urea-based SNCR technology. To date, DRB-4Z burners are installed in approximately 12 power plants, but no units include the Fuel Tech SNCR system.[21]

NO_x Controls for Gas Turbines

Dry Low NOx

Turbine DLN combustors utilize multistage premix combustors where the air and fuel is mixed at a lean (high oxygen) fuel-to-air ratio. The excess air in the lean mixture acts as a heat sink, which lowers peak combustion temperatures, ensures a more homogeneous mixture, and reduces NO_x formation rates. The word "dry" signifies that NO_x control is achieved without the need for water or steam injection. Turbine combustors generally can not be operated in lean premix DLN mode when fuel oil is burned. The primary NO_x control method for fuel oil combustion remains wet injection.[1] For older gas turbines, it may not be possible to retrofit DLN technology as a means to control NO_x emissions.[2]

Inlet Air Heating

Inlet Air Heating is a control method that heats the air entering the combustion chamber above ambient temperatures. Inlet air heating can be achieved by any of the following methods: exhaust gas mixing with the inlet air, heat exchangers in the exhaust and inlet ducts, and electric heating coils in the inlet duct. Combustion control methods used for NO_x emission control in Solar gas turbines (SoLoNO_x) have proven experience at maintaining NO_x emission levels at 25 ppm at ambient air temperatures above 0 ºF.[5] May derate the turbine.

SCONO_x

SCONO_x is an oxidation catalyst based technology that removes both NO_x and CO without the need of supplementary chemical reagents (e.g., ammonia). The SCONO_x system has demonstrated its ability to meet the same low emission rates as a conventional SCR/CO Catalyst system without the use of ammonia. The use of the SCONO_x catalyst is limited for simple cycle installations due to the maximum operating temperature; 700°F. Due to its precious metal catalyst and complex mechanical system, SCONO_x is much more expensive when compared to an SCR/CO catalyst. There is concern regarding scale-up problems that could occur from 25 MW gas turbine to a 250 MW gas turbine.[15]

Selective Catalytic Reduction (SCR)

Aqueous or anhydrous ammonia is injected into the turbine exhaust gas, which then passes through the catalyst reactor NO_x is converted to nitrogen, water, and oxygen. Typical ammonia slip permit limits are 10 ppmvd, although certain newer permits have limits of 5 ppmvd.[19] The majority of the gas turbine NO_x abatement systems operating around the world use SCR.[9]

Selective Non-Catalytic Reduction (SNCR)

Simple-cycle turbines normally operate in the range from 800 to 1,000°F, achieving the required reaction temperature represents the main technical difficulty for application of SNCR to turbines. High temperature catalysts have been developed.[20]

Water/Steam Injection

Water/Steam Injection as a control technology involves the introduction of water or steam into the combustion zone. The injected fluid provides a heat sink which absorbs some of the heat of reaction, causing a lower flame temperature. The lower flame temperature results in lower thermal NO_x formation. The process requires approximately 0.8 to 1.0 pound of water or steam per pound of fuel burned.[5]

XONON

The Xonon technology is installed as an integral part of the combustor. Rather than combust the fuel in a flame, the Xonon system combusts the fuel at lower temperatures using a catalyst. Fuel and air are thoroughly mixed prior to entering a catalyst region that acts to combust the fuel, releasing its energy. The Xonon catalyst module consists of a channel structure whereby the fuel-air mixture readily passes through the channels coated with the catalyst. As fuel and oxygen molecules contact the channel walls, the molecules and catalyst interact and are rearranged at temperatures well below those of flame combustion. Nitrogen molecules are not involved in the Xonon chemistry and pass through the channels unchanged, thereby preventing the formation of NO_x. There are currently three commercial installations of the Xonon technology (1.4 MW, 1.5 MW and 1.4 MW gas turbines). The technology is guaranteed to reduce NO_x emissions to below 3 ppm.[23] XONON is “No NO_x” spelled backwards.

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SO_2 Controls

Coal Cleaning

Physical coal cleaning is estimated to reduce the amount of SO_2 in the coal by 10 to 50 percent, and costs vary from $1 to $10 per ton, depending on the coal quality. Additionally, there are advanced coal cleaning technologies including advanced physical cleaning, aqueous-phase pretreatment, selective agglomeration, and organic-phase pretreatment. Coal cleaning may lower the uncontrolled SO_2 emission rate but not to the extent of a FGD system.

Flue Gas Desulfurization (FGD)

FGD systems utilize chemical reactions to reduce the concentrations of SO_2 in the flue gas. FGD systems can be characterized as either wet, dry, or semi-dry.

Wet FGD

A Wet FGD sprays a reagent and water slurry into the flue gas using an absorber tower. The flue gas becomes saturated with water vapor and the SO_2 is removed by reacting with the slurry. In the United States, quicklime, and limestone are the commonly used reagents. Ammonia and magnesium enhanced lime reagents are also used on a more limited basis. SO_2 reacts with the quicklime to form calcium sulfite and then compressed air is injected into retention tank to oxidize the calcium sulfite to form calcium sulfate. The slurry is then recirculated through the system and de-watered. If limestone is used, the SO_2 reacts with the some of the lime to form calcium hydroxide. Limestone is less expensive than lime, and so limestone is more commonly used, especially for medium to higher sulfur coals.

Wet FGD systems have been demonstrated on large supercritical PC-boilers firing coal. U.S. EPAs AP-42 document indicates that wet FGD systems can achieve 80 to 98% reductions depending on the inlet coal sulfur content. Recent units using low sulfur coals, such as PRB have been permitted with control efficiencies greater than 90%.[1]

Semi-Dry and FGD

Dry FGD systems may use either a dry powder or semi-dry slurry to remove SO_2. A semi-dry system uses hydrated lime mist slurry to form calcium sulfite solids and excess reagent. The solids are collected using a fabric filter, as was discussed in the section on particulate matter control devices. There are three types of dry FGD systems as discussed below. As of February 22, 2008 eleven out of 16 recent permit applications for power plants burning low sulfur coal were using dry FGD system.[2]

Spray Dryer Absorber

A spray dryer absorber (SDA) sprays a slurry of lime and water into the tower. The water evaporates and a dry by-product results. These by-products are collected in a fabric filter, as discussed in the section under PM control. SDAs are expected to achieve 93% control efficiency over the long-term.[1] Median design efficiencies for wet and dry FGD systems are identical based on EPA findings.[2] SDAs have been demonstrated on large boilers burning low sulfur coals.

Dry Sorbent Injection

Dry sorbent injection sprays a powdered absorbent into the furnace or flue gas stream. The control efficiencies of these systems range from 40 to 85% depending on various factors including sorbent type, stoichiometry, amount of recycled reagent, temperature and plant configuration. Dry sorbent injection has been demonstrated on boilers burning low sulfur coals, however, the control efficiencies achieved are less than the other SO_2 controls.

Circulating Dry Scrubber

A circulating dry scrubber uses hydrated lime to remove SO_2. The flue gas is circulated around the lime for adsorption of the SO_2. The by-product solids and excess reagent are collected using either a fabric filter or ESP as mentioned in the particulate controls section. Over 90 percent of the collected solids in the ESP or FF contain unreacted lime. This type of system has only been used on small pulverized coal-fired boilers (50-300 MWe.); there is no operating experience for large units. The AES Greenridge Plant in Torrey, New York is installing a system which plans to get a 95% reduction in SO_2 emissions from a coal with greater than 2% sulfur.[3]

Emerging Technologies

Activated Carbon Beds use activated carbon as a dry sorbent to remove SO_2 from the flue gas. As the activated carbon becomes saturated with SO_2, it is regenerated, and the SO_2 is released as a stream of gaseous SO_2. Commercial sales of activated carbon bed technology have not been identified.

Limestone Injection Gas Scrubbing (LIDS) combines sorbent injection and dry scrubber technologies. Limestone reagent is injected into the furnace, and the excess reagent continues through the system until it is collected in the particulate collection system (e.g., a fabric filter). A portion of the collected ash is slurried with water and fed to the dry scrubber, where most of the SO_2 is removed. The LIDS technology is still under development to improve performance and has not been demonstrated for controlling SO_2 emissions on a large-scale, coal-fired boiler.

Pahlman Process, EnviroScrub is a multi-pollutant control technology that simultaneously controls NO_x and SO_2. The process uses a spray dryer absorber where a proprietary Pahlmanite™ scrubber material contacts the exhaust stream. The exhaust stream then passes through a "baghouse reaction chamber" where the Pahlmanite™ material is removed prior to the final exhaust stack. This technology is currently in the pilot-scale stage of development.

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PM Controls

Centrifugal Collectors / Cyclones

Cyclones use centrifugal force to separate PM from the gas stream. Often, multiple cyclones are operated together in order to achieve higher efficiencies with larger throughput.[4] These multiple-cyclone separators, also known as multiclones, are longer and have smaller diameters, which create longer residence times and greater centrifugal force. However, smaller particles do not have enough mass the generate the force required for separation, so these devices are only effective in removing particles greater than 10 microns in size.[5] Therefore, they are primarily used before the principal control mechanism in order to recover larger particles from the exhaust stream.[3]

Cyclone collector efficiencies range only from 80 to 90 percent, and although they are commercially available, they cannot reduce emissions enough to comply with standards.

Electrostatic Precipitator (ESP)

In an ESP, an electric field is produced that imparts a negative electric charge to solid particles in the flue gas stream. These negatively charged particles then migrate to positively charged collection electrodes, or plates. Rapping mechanisms operated intermittently dislodge the collected particles from the plates and allow them to fall and settle into a hopper located below the ESP. An ESP's performance is highly dependent on the resistivity of the particles, which is influenced by particle composition and size distribution, flue gas characteristics, and particle loading.[2] Typically, an ESP can have a PM collection efficiency of 99 percent or more.[4]

Fabric Filter (FF) / Baghouse

A FF system, or baghouse, collects particulate as the flue gas passes through a fabric filter media such as woven cloths or felts. The filters are typically configured in cylindrical bags, hundreds or thousands of which are contained in a housing structure that has inlets and outlets for the flue gas. PM collection occurs through several mechanisms, including gravitational settling, direct impaction, inertial impaction, diffusion, and electrostatic attraction, as the flue gas enters the cylinders from the bottom and flows upward, through the fabric to the housing outlet. The groups of bags are cleaned periodically by shaker cleaning, reverse-air cleaning, or pulse-jet cleaning. Pulse-jet cleaning is currently considered the best practice for cleaning.[2]

For fuels with higher sulfur content, difficulties have been experienced with FFs, but FFs are usually ranked as the most efficient PM control technology for low-sulfur fuel applications, having a collection efficiency of 99.5 percent or greater. They are commercially available and have been shown to reduce PM emissions from large pulverized coal boilers using sub-bituminous or PRB coal.[3]

Wet Electrostatic Precipitator (WESP)

A WESP operates in the same way as an ESP, but a continuous or intermittent spray is used to wash the collectors instead of mechanical rapping. WESPs are commonly used in applications where the flue gas has a high moisture content or is below the dew point.[3] They will typically be installed downstream of a WFGD system, where the flue gas is already saturated. An added benefit to WESPs is a lack of re-entrainment after contact with the wet walls, which makes a WESP well suited fine particulate or acid mist applications. The humidity, temperature, and particulate characteristics in WESPs allow for excellent fuel flexibility, but it is necessary to investigate water chemistry, scaling, and corrosion potential when considering WESP as a control option.[5]

The collection efficiency of a WESP is generally 99 percent or more. WESPs are commercially available, and although there have been only a few applications in the utility industry, they are a technically feasible control option for coal fired units.[4]

Wet Scrubber

In a wet scrubber system, PM is removed from the flue gas stream principally by inertial impaction onto water droplets that are sprayed into the stream. This impaction may occur through impingement, diffusion, or condensation. A reagent may also slurried with water and sprayed into the exhaust stream. A venturi scrubber constricts the gas in a narrow throat section, which increasing the velocity of the gas, helping to atomize the scrubbing liquid and improve gas-liquid contact. A cyclone is usually used to collect the wetted PM and excess liquid droplets.[3]

A wet venturi scrubber can be up to 95 percent efficient at reducing PM emissions but cannot reduce emissions enough to comply with standards. [3,4] High energy wet scrubbers have had no applications to large coal fired boilers in the last 15 years.[1,6] For these reasons, wet scrubbers are deemed technically infeasible.

1. Duke Power: PSD Permit Application for the Cliffside Steam Station Units 6 & 7 -Volume I, North Carolina. December 2005.

2. Air Construction Permit Application and Prevention of Significant Deterioration Analysis for FPL Glades Power Park, Glades County, Florida. December 2006.

3. Air Construction/Prevention of Significant Deterioration Permit Application for Taylor Energy Center, Florida. May 2007.

4. BACT Analysis for Ely Energy Center, Nevada. Appendix B of Application for Operating Permit to Construct – Volume I. October 2007.

5. Best Available Retrofit Technology (BART) Analysis for Portland General Electric, Boardman Plant, Oregon. November 2007.

6. Application for Prevention of Significant Deterioration Permit for the Desert Rock Energy Facility, Navaho Nation. May 2004.

Hg Controls

Multi-pollutant Controls

Varying levels of mercury emission reduction can be achieved through controls installed for other pollutants, such as FGD systems for SO_2 control, SCR systems for NO_x control, and PM control systems. Below are brief descriptions of the effectiveness of these devices for reducing Hg emissions.

FFs and ESPs

Hg removal efficiencies using FFs have been shown to be greater than efficiencies for ESPs, although efficiency is highly dependent on fuel type.[1] This dependence on fuel type is mainly due to mercury speciation, which is the partitioning of Hg into elemental mercury vapor (Hg^0), ionic mercury (Hg^{2+}), and particulate mercury (Hg_p). PM control devices mostly capture Hg_p and Hg^{2+}, allowing much of the Hg^0 to pass through. Through complex reactions that occur as flue gas is cooled, the higher Cl content in bituminous coal allows more Hg^0 to be converted to Hg^{2+}, which is more easily captured by PM controls.[3,4]

For units that burn bituminous coal, Hg capture efficiency for FFs can be as high as 90 percent, but for units burning subbituminous coal, FFs are typically no more than 70 percent efficient. With ESPs, Hg removal efficiencies are only 3 to 6 percent for subbituminous coal and vary from 9 to 36 percent for bituminous coal. ESPs are less effective in Hg capture because there is less contact between gaseous Hg and fly ash in these devices.[3,4]

FGD Systems

Dry FGD systems, which are typically spray dryer absorbers (SDAs), are often installed upstream of FFs,. Research has shown that for bituminous coal fired units, Hg removal is as high as 98 percent with the SDA/FF combination. However, with subbituminous coal fired units, Hg removal with this combination can be drastically less, as low as 24 percent. This decrease is likely because for boilers burning subbituminous coals, which have a lower Cl content, most of the gaseous mercury produced is Hg^0 because less Cl is available for oxidation. Additionally, any HCl in the flue gas will be removed by the SDA, which also results in less Hg^0 oxidation and thus lower removal efficiency in the FF.[4]

Wet FGD systems are typically installed downstream of PM control devices. Because gaseous compounds of Hg^{2+} are generally soluble in water, they can be absorbed in the aqueous slurry of a wet FDG. It has been shown that the combination of a PM control with a wet FGD can capture 29 percent or more of Hg from a unit burning subbituminous coal, and can capture as much as 98 percent of the Hg from a unit burning bituminous coal. Higher efficiencies can be attributed to increased oxidation of Hg^0 due to higher Cl content in the coal being burned.[5]

SCR Catalysts

Increasing the amount of Hg^{2+} relative to total Hg in the flue gas should improve overall mercury capture by the FGD system, because Hg^{2+} is easier for the system to capture. Therefore, promoting the oxidation of Hg^0 to Hg^{2+} upstream can effectively improve mercury removal by the wet FGD. Studies show that under certain conditions SCR catalysts will promote this oxidation, in particular for bituminous coal. In most cases, around 90 percent of mercury was in the oxidized form when burning bituminous coal. For subbituminous coals, however, it is thought that improving catalytic oxidation with SCR systems is not possible with the typical chemical composition of the flue gas or a lower catalyst temperature.[4] The performance and applicability of SCR systems as a control for mercury for plants burning low rank coals is uncertain, but SCR impact on mercury control is still being studied in on-going, full-scale tests.[1]

Mercury-Specific Controls

Carbon Injection

Currently, activated carbon injection (CI) is the most extensively investigated mercury-specific control methods for coal-fired power plants. With this technology, powdered activated carbon (PAC) is injected into the gas stream before the gas enters a PM control device. The PAC sorbent binds with mercury in the flue gas and is captured, along with the mercury, in the PM control device. The efficiency of mercury capture is typically greater for FFs as opposed to ESPs because of the increased particle-gas contact in the filter cakes on the bag surfaces in a FF. The choice of carbon for a specific application will depend on the total Hg concentration, flue gas composition, and the method of capture.[1,5]

Various short term full-scale tests of CI technologies have been conducted, and some long-term studies are in progress. However, the completed trials may not encompass the range of control/fuel combinations that would be necessary to demonstrate widely achievable Hg emission reduction.

References

1. U.S. EPA: http://www.epa.gov/mercury/control_emissions/tech_exist.htm

2. U.S. EPA: http://www.epa.gov/mercury/control_emissions/tech_merc_specific.htm

3. Performance and Cost of Mercury and Multipollutant Emission Control Technology Applications on Electric Utility Boilers. EPA- 600/R-03-110. October 2003. http://www.epa.gov/appcdwww/aptb/EPA600R03110.pdf)

4. Control of Mercury Emissions from Coal-Fired Electric Utility Boilers. EPA Office of Research and Development. January 2004. http://www.epa.gov/ttn/atw/utility/hgwhitepaperfinal.pdf

5. Control of Mercury Emissions from Coal Fired Electric Utility Boilers: An Update. U. S. EPA Office of Research and Development. February 2005.

Appendix B

Cost Inputs and Outputs

INPUTS FOR CONTROL COSTS IN CUECOST MODEL

			Cue-Cost Default Value	**Alberta Model Value**	**Default Used in Alberta Model (Y/N)**
General Plant Technical Inputs					
Location - State	Abbrev.	All States	PA	Alberta	
MW Equivalent of Flue Gas to Control System	MW	100-2000	500	100, 500, and 1,000	
Net Plant Heat Rate (w/o APC)	Btu/kWhr		10,500	10,500	Yes
Plant Capacity Factor	%	40-90%	65%	90%	
Percent Excess Air in Boiler	%		120%	120%	Yes
Air Heater Inleakage	%		12%	12%	Yes
Air Heater Outlet Gas Temperature	°F		300	300	Yes
Inlet Air Temperature	°F		80	80	Yes
Ambient Absolute Pressure	In. of Hg		29.4	29.4	Yes
Pressure After Air Heater	In. of H2O		-12	-12	Yes
Moisture in Air	lb/lb dry air		0.013	0.013	Yes
Ash Split:					Yes
Fly Ash	%		80%	61%	
Bottom Ash	%		20%	39%	
Seismic Zone	Integer	1-5	1	1	Yes
Retrofit Factor	Integer	1.0-3.0	1.3	1	
(1.0 = new, 1.3 = medium, 1.6 = difficult)					
Select Coal	Integer	1-8	1	8 - User Specified Alberta Coal - see Right for assumptions	
Is Selected Coal a Powder River Basin Coal?	Yes / No	See Column K	Yes	Yes	Yes
Economic Inputs					
Cost Basis -Year Dollars	Year		1998	2007	
Sevice Life (levelization period)	Years		30	20	
Inflation Rate	%		3.00%	3.00%	Yes
After Tax Discount Rate (current $'s)	%		9.20%	9.20%	Yes
AFDC Rate (current $'s)	%		10.80%	10.80%	Yes
First-year Carrying Charge (current $'s)	%		22.30%	22.30%	Yes
Levelized Carrying Charge (current $'s)	%		16.90%	16.90%	Yes
First-year Carrying Charge (constant $'s)	%		15.70%	15.70%	Yes
Levelized Carrying Charge (constant $'s)	%		11.70%	11.70%	Yes
Sales Tax	%		6%	6%	Yes
Escalation Rates:					
Consumables (O&M)	%		3%	3%	Yes
Capital Costs:					
Is Chem. Eng. Cost Index available?	Yes / No		Yes	Yes	Yes
If "Yes" input cost basis CE Plant Index.	Integer		388	525.4	
If "No" input escalation rate.	%		3%	3%	Yes
Construction Labor Rate	$/hr		$35	$46	
Prime Contractor's Markup	%		3%	3%	Yes
Operating Labor Rate	$/hr		$30	$40	
Power Cost	Mills/kWh		25	25	Yes
Steam Cost	$/1000 lbs		3.5	3.5	Yes

			Cue-Cost Default Value	Alberta Model Value	Default Used in Alberta Model (Y/N)
Limestone Forced Oxidation (LSFO) Inputs					
SO2 Removal Required	%	90-98%	95%	90%	
L/G Ratio	gal / 1000 acf	95-160	125	125	Yes
Design Scrubber with Dibasic Acid Addition?	Integer	1 or 2	2	2	Yes
(1 = yes, 2 = no)					Yes
Adiabatic Saturation Temperature	°F	100-170	127	127	Yes
Reagent Feed Ratio	Factor	1.0-2.0	1.05	1.1	
(Mole CaCO3 / Mole SO2 removed)					Yes
Scrubber Slurry Solids Concentration	Wt. %		15%	15%	Yes
Stacking, Landfill, Wallboard	Integer	1,2,3	1	2	
(1 = stacking, 2 = landfill, 3 = wallboard)					
Number of Absorbers	Integer	1-6	1	1 or 2, depending on unit capacity	
(Max. Capacity = 700 MW per absorber)					
Absorber Material	Integer	1 or 2	1	1	Yes
(1 = alloy, 2 = RLCS)					
Absorber Pressure Drop	in. H2O		6	6	Yes
Reheat Required ?	Integer	1 or 2	1	1	Yes
(1 = yes, 2 = no)					
Amount of Reheat	°F	0-50	25	25	Yes
Reagent Bulk Storage	Days		60	60	Yes
Reagent Cost (delivered)	$/ton		$15	$15	Yes
Landfill Disposal Cost	$/ton		$30	$30	Yes
Stacking Disposal Cost	$/ton		$6	$6	Yes
Credit for Gypsum Byproduct	$/ton		$2	$2	Yes
Maintenance Factors by Area (% of Installed Cost)					
Reagent Feed	%		5%	5%	Yes
SO2 Removal	%		5%	5%	Yes
Flue Gas Handling	%		5%	5%	Yes
Waste / Byproduct	%		5%	5%	Yes
Support Equipment	%		5%	5%	Yes
Contingency by Area (% of Installed Cost)					
Reagent Feed	%		20%	20%	Yes
SO2 Removal	%		20%	20%	Yes
Flue Gas Handling	%		20%	20%	Yes
Waste / Byproduct	%		20%	20%	Yes
Support Equipment	%		20%	20%	Yes
General Facilities by Area (% of Installed Cost)					
Reagent Feed	%		10%	10%	Yes
SO2 Removal	%		10%	10%	Yes
Flue Gas Handling	%		10%	10%	Yes
Waste / Byproduct	%		10%	10%	Yes
Support Equipment	%		10%	10%	Yes
Engineering Fees by Area (% of Installed Cost)					
Reagent Feed	%		10%	10%	Yes
SO2 Removal	%		10%	10%	Yes
Flue Gas Handling	%		10%	10%	Yes
Waste / Byproduct	%		10%	10%	Yes
Support Equipment	%		10%	10%	Yes

			Cue-Cost Default Value	Alberta Model Value	Default Used in Alberta Model (Y/N)
Lime Spray Dryer (LSD) Inputs					
SO2 Removal Required	%	90-95%	90%	89%	
Adiabatic Saturation Temperature	°F	100-170	127	127	Yes
Flue Gas Approach to Saturation	°F	10.-50	20	20	Yes
Spray Dryer Outlet Temperature	°F	110-220	0.25	147	
Reagent Feed Ratio	Factor	Calc. Based on %S	0.90	0.90	
(Mole CaO / Mole Inlet SO2)					
Recycle Rate	Factor	Calculated	30	30	Yes
(lb recycle / lb lime feed)					
Recycle Slurry Solids Concentration	Wt. %	10-50	35%	35%	Yes
Number of Absorbers	Integer	1-7	2	1 or 2, depending on unit capacity	
(Max. Capacity = 300 MW per spray dryer)					
Absorber Material	Integer	1 or 2	1	1	Yes
(1 = alloy, 2 = RLCS)					
Spray Dryer Pressure Drop	in. H2O		5	5	Yes
Reagent Bulk Storage	Days		60	60	Yes
Reagent Cost (delivered)	$/ton		$75	$75	Yes
Dry Waste Disposal Cost	$/ton		$30	$30	Yes
Maintenance Factors by Area (% of Installed Cost)					
Reagent Feed	%		5%	5%	Yes
SO2 Removal	%		5%	5%	Yes
Flue Gas Handling	%		5%	5%	Yes
Waste / Byproduct	%		5%	5%	Yes
Support Equipment	%		5%	5%	Yes
Contingency by Area (% of Installed Cost)					
Reagent Feed	%		20%	20%	Yes
SO2 Removal	%		20%	20%	Yes
Flue Gas Handling	%		20%	20%	Yes
Waste / Byproduct	%		20%	20%	Yes
Support Equipment	%		20%	20%	Yes
General Facilities by Area (% of Installed Cost)					
Reagent Feed	%		10%	10%	Yes
SO2 Removal	%		10%	10%	Yes
Flue Gas Handling	%		10%	10%	Yes
Waste / Byproduct	%		10%	10%	Yes
Support Equipment	%		10%	10%	Yes
Engineering Fees by Area (% of Installed Cost)					
Reagent Feed	%		10%	10%	Yes
SO2 Removal	%		10%	10%	Yes
Flue Gas Handling	%		10%	10%	Yes
Waste / Byproduct	%		10%	10%	Yes
Support Equipment	%		10%	10%	Yes

			Cue-Cost Default Value	**Alberta Model Value**	**Default Used in Alberta Model (Y/N)**
Particulate Control Inputs					
Outlet Particulate Emission Limit	lbs/MMBtu		0.03	0.03	Yes
Fabric Filter:					
Pressure Drop	in. H2O		6	6	Yes
Type (1 = Reverse Gas, 2 = Pulse Jet)	Integer		2	2	Yes
Gas-to-Cloth Ratio	ACFM/ft^2		3.5	3.5	Yes
Bag Material (RGFF fiberglass only)	Integer		2	2	Yes
(1 = Fiberglass, 2 = Nomex, 3 = Ryton)					
Bag Diameter	inches	5 - 14	6	6	Yes
Bag Length	feet	15 - 35	20	20	Yes
Bag Reach			3	3	Yes
Compartments out of Service	%		10%	10%	Yes
Bag Life	Years	1 - 10	5	5	Yes
Maintenance (% of installed cost)	%		5%	5%	Yes
Contingency (% of installed cost)	%		20%	20%	Yes
General Facilities (% of installed cost)	%		10%	10%	Yes
Engineering Fees (% of installed cost)	%		10%	10%	Yes
ESP:					
Strength of the electric field in the ESP = *E*	kV/cm		10.0	10.0	Yes
Plate Spacing	in.		12	12	Yes
Plate Height	ft.		36	36	Yes
Pressure Drop	in. H2O		3	3	Yes
Maintenance (% of installed cost)	%		5%	5%	Yes
Contingency (% of installed cost)	%		20%	20%	Yes
General Facilities (% of installed cost)	%		10%	10%	Yes
Engineering Fees (% of installed cost)	%		10%	10%	Yes
NOx Control Inputs					
Selective Catalytic Reduction (SCR) Inputs					
NH3/NOX Stoichiometric Ratio	NH3/NOX	0.7-1.0	0.9	D	Yes
NOX Reduction Efficiency	Fraction	0.60-0.90	0.70	0.078	
Inlet NOx	lbs/MMBtu		0.9	0.14	
Space Velocity (Calculated if zero)	1/hr		0	D	Yes
Overall Catalyst Life	years	2-5	3	D	Yes
Ammonia Cost	$/ton		206	352	
Catalyst Cost	$/ft3		356	D	Yes
Solid Waste Disposal Cost	$/ton		11.48	30	
Maintenance (% of installed cost)	%		1.5%	D	Yes
Contingency (% of installed cost)	%		20%	D	Yes
General Facilities (% of installed cost)	%		5%	D	Yes
Engineering Fees (% of installed cost)	%		10%	D	Yes
Number of Reactors	integer		2	D	Yes
Number of Air Preheaters	integer		1	D	Yes

			Cue-Cost Default Value	**Alberta Model Value**	**Default Used in Alberta Model (Y/N)**
Selective NonCatalytic Reduction (SNCR) Inputs					
Reagent	integer	1:Urea 2:Ammonia	1	D	Yes
Number of Injector Levels	integer		3	D	Yes
Number of Injectors	integer		18	D	Yes
Number of Lance Levels	integer		0	D	Yes
Number of Lances	integer		0	D	Yes
Steam or Air Injection for Ammonia	integer	1: Steam, 2: Air	1	D	Yes
NOX Reduction Efficiency	fraction	0.30-0.70	0.50	0.4 for 100 MW; 0.27 for 500 MW; 0.14 for 1000 MW	
Inlet NOx	lbs/MMBtu		0.9	0.14	
NH3/NOX Stoichiometric Ratio	NH3/NOX	0.8-2.0	1.2	D	Yes
Urea/NOX Stoichiometric Ratio	Urea/NOX	0.8-2.0	1.2	D	Yes
Urea Cost	$/ton		254	D	Yes
Ammonia Cost	$/ton		206	352	
Water Cost	$/1,000 gal		0.5	D	Yes
Maintenance (% of installed cost)	%		1.5%	D	Yes
Contingency (% of installed cost)	%		20%	D	Yes
General Facilities (% of installed cost)	%		5%	D	Yes
Engineering Fees (% of installed cost)	%		10%	D	Yes

*All $ here represent US Dollars.

COAL PROPERTIES USED IN CUECOST MODEL

Properties of Alberta Subbituminous Coal -

Parameter	**Units**	**Value**	**Note**
Coal Cost	$/MMBtu	1.50	Default from CueCost
PROXIMATE ANALYSIS (ASTM, as rec'd)			
Moisture - Enter below in Ultimate Analysis			
Volatile Matter	wt%	**31.39**	PRB value*
Fixed Carbon	wt%	**33.05**	PRB value*
Ash - Enter below in Ultimate Analysis			
		105.70	
COAL ULTIMATE ANALYSIS (ASTM, as rec'd)			
Moisture	wt%	**21.74**	
Carbon	wt%	**48.18**	PRB value*
Hydrogen	wt%	**3.31**	PRB value*
Nitrogen	wt%	**0.70**	PRB value*
Chlorine	wt%	**0.01**	
Sulfur	wt%	**0.43**	
Ash	wt%	**19.52**	
Oxygen	wt%	**6.11**	Difference of 100 and sum of all other analytes
TOTAL	wt%	100.00	
Modified Mott Spooner HHV (Btu/lb) - ***calc***	Btu/lb	8,589	
COAL ASH ANALYSIS (ASTM, as rec'd)			
SiO2	wt%	**35.51**	PRB value*
Al2O3	wt%	**17.11**	PRB value*
TiO2	wt%	**1.26**	PRB value*
Fe2O3	wt%	**6.07**	PRB value*
CaO	wt%	**26.67**	PRB value*
MgO	wt%	**5.30**	PRB value*
Na2O	wt%	**1.68**	PRB value*
K2O	wt%	**2.87**	PRB value*
P2O5	wt%	**0.97**	PRB value*
SO3	wt%	**1.56**	PRB value*
Other Unaccounted for	wt%	**1.00**	PRB value*
TOTAL	wt%	100.00	

*PRB value from CueCost was used since no value was provided for Alberta

SCR (high-dust) - Preliminary

		100 MW	500 MW	1000 MW
Ammonia Injection Rate	lb/hr	49	245	490
Space Velocity	1/hr	3,171	3,171	3,171
Gross Catalyst Volume	ft^3	4,292	21,460	42,921

SCR Capital Costs

		100 MW	500 MW	1000 MW
Cost Basis (Year)		***2007***	***2007***	***2007***
Reactor Housing and Installation	$	2,335,271	5,130,193	7,200,081
Ammonia Handling and Injection	$	487,415	1,058,772	1,478,764
Flue Gas Handling:Ductwork and Fans	$	2,220,924	6,786,080	10,978,266
Air Preheater Modifications	$	464,259	1,682,428	2,929,276
Misc. Other Direct Capital Costs	$	305,668	563,670	778,529
Equipment Capital Cost Subtotal	$	$5,813,538	$15,221,142	$23,364,917
Instruments & Controls	$	$116,271	$304,423	$467,298
Taxes	$	$348,812	$913,269	$1,401,895
Freight	$	$290,677	$761,057	$1,168,246
Total Direct Cost		$6,569,298	$17,199,890	$26,402,356
Total Direct Cost with Retrofit Factor	$	$6,569,298	$17,199,890	$26,402,356
General Facilities	$	$328,465	$859,995	$1,320,118
Engineering Fees	$	$656,930	$1,719,989	$2,640,236
Contingency	$	$1,313,860	$3,439,978	$5,280,471
Total Plant Cost (TPC)	$	$8,868,553	$23,219,852	$35,643,181
Total Plant Cost (TPC) w/ Prime Contractor's Markup	$	$9,134,609	$23,916,448	$36,712,476
Total Cash Expended (TCE)	$	$9,134,609	$23,226,615	$35,139,451
Allow. for Funds During Constr. (AFDC)	$	$0	$2,546,699	$5,957,800
Total Plant Investment (TPI)	$	$9,134,609	$25,773,314	$41,097,251
Preproduction Costs	$	$206,328	$589,363	$951,772
Inventory Capital				
Initial Ammonia(60 days)	$	$11,172	$55,858	$111,716
Initial Catalyst	$	$1,529,444	$7,647,220	$15,294,440
Total Capital Requirement (TCR)	$	**$10,881,553**	**$34,065,755**	**$57,455,179**
	$/kW	$109	$68	$57

SCR O&M Costs

		100 MW	500 MW	1000 MW
Cost Basis (Year)		***2007***	***2007***	***2007***
Ammonia	$/yr	67,960	339,802	679,603
Catalyst Replacement	$/yr	509,815	2,549,073	5,098,147
Catalyst Disposal	$/yr	1,030	5,151	10,301
Electricity	$/yr	61,426	385,252	790,036
High-dust SCR Steam	$/yr	5,084	25,628	51,307

Operating Labor	$/yr		75,092	160,900	268,160
Maintenance	$/yr		133,028	348,298	534,648
Total O&M Costs	$/yr		**853,435**	**3,814,103**	**7,432,202**

SNCR - Preliminary

			100 MW	500 MW	1000 MW
Number of Wall Injectors	integer		18	18	18
Number of Lances	integer		0	0	0
Urea Injection Rate	lb/hr		115	573	1147
Ammonia Injection Rate	lb/hr		65	327	653

SNCR Capital Costs

			100 MW	500 MW	1000 MW
Cost Basis (Year)			***2007***	***2007***	***2007***
Urea Based SNCR Costs					
Urea Storage & Handling	$		$150,288	$294,032	$392,576
Urea Injection	$		$411,903	$411,903	$411,903
Controls/Miscellaneous	$		$165,149	$222,149	$293,400
Air Heater Modifications	$		$377,592	$1,308,059	$2,233,688
Ammonia Based SNCR Costs					
Ammonia Storage, Handling, Injection, Controls	$		$0	$0	$0
Air Heater Modifications	$		$0	$0	$0
Total Direct Cost	$		$1,104,931	$2,236,143	$3,331,567
Total Direct Cost with Retrofit Factor	$		$1,104,931	$2,236,143	$3,331,567
General Facilities	$		$55,247	$111,807	$166,578
Engineering Fees	$		$110,493	$223,614	$333,157
Contingency	$		$220,986	$447,229	$666,313
Total Plant Cost (TPC)	$		$1,491,657	$3,018,793	$4,497,616
Total Plant Cost (TPC) w/ Prime Contractor's Markup	$		$1,536,407	$3,109,357	$4,632,544
Total Cash Expended (TCE)	$		$1,536,407	$3,019,673	$4,434,053
Allow. for Funds During Constr. (AFDC)	$		$0	$331,094	$751,782
Total Plant Investment (TPI)	$		$1,536,407	$3,350,767	$5,185,835
Preproduction Costs	$		$50,933	$131,649	$223,350
Inventory Capital	$		$18,872	$94,361	$188,721
Total Capital Requirement (TCR)	$		**$1,606,212**	**$3,576,777**	**$5,597,906**
	$/kW		$16.1	$7.2	$5.6

SNCR O&M Costs

			100 MW	500 MW	1000 MW
Cost Basis (Year)			***2007***	***2007***	***2007***
Operating and Supervisory Labor	$/yr		87,600	87,600	87,600
Maintenance Labor and Materials	$/yr		22,375	45,282	67,464
Reagent	$/yr		114,805	574,027	1,148,054
Electricity	$/yr		2,206	2,206	2,206

Water	$/yr		4,428	4,428	4,428
Steam (for steam atomization)	$/yr		-	-	-
Total O&M Costs	$/yr		**231,413**	**713,542**	**1,309,751**

Low NOX Burner Technology Capital Costs

			100 MW	500 MW	1000 MW
Cost Basis (Year)			***2007***	***2007***	***2007***
Total Capital Requirement with Retrofit (TCR)	**$**		**$1,720,479**	**$8,602,397**	**$17,204,793**
	$/kW		$17.2	$17.2	$17.2

Low NOX Burner Technology O&M Costs

			100 MW	500 MW	1000 MW
Cost Basis (Year)			***2007***	***2007***	***2007***
Maintenance Labor	$/yr		13,764	68,819	137,638
Maintenance Materials	$/yr		20,646	103,229	206,458
Control, Administration, Overhead	$/yr		4,129	20,646	41,292
Total O&M Costs	$/yr		**38,539**	**192,694**	**385,387**

Natural Gas Reburning - Preliminary

			100 MW	500 MW	1000 MW
Fraction of heat input as reburn fuel	fraction		0.15	0.15	0.15
Bottom Ash Rate	tons/yr		1844	9221	18442

Natural Gas Reburning Capital Costs

			100 MW	500 MW	1000 MW
Cost Basis (Year)			***2007***	***2007***	***2007***
Gas Pipeline from Fenceline to Boiler	$		$688,807	$1,818,946	$3,231,620
Fuel Injectors, Overfire Air Ports and Associated Piping, Valves, Windbox and Control Dampers	$		$1,903,680	$4,589,174	$8,081,741
Total Direct Cost	$		$2,592,487	$6,408,121	$11,313,361
Total Direct Cost with Retrofit Factor	$		$2,592,487	$6,408,121	$11,313,361
General Facilities	$		$51,850	$128,162	$226,267
Engineering Fees	$		$259,249	$640,812	$1,131,336
Contingency	$		$518,497	$1,281,624	$2,262,672
Total Plant Cost (TPC)	$		$3,422,083	$8,458,719	$14,933,637
Total Plant Cost (TPC) w/ Prime Contractor's Markup	$		$3,524,745	$8,712,481	$15,381,646
Total Cash Expended (TCE)	$		$3,524,745	$8,461,183	$14,722,586
Allow. for Funds During Constr. (AFDC)	$		$0	$927,733	$2,496,175
Total Plant Investment (TPI)	$		$3,524,745	$9,388,915	$17,218,761

Preproduction Costs	$		$74,472	$196,864	$360,075
Inventory Capital	$		$0	$0	$0
Total Capital Requirement (TCR)	$		**$3,599,217**	**$9,585,780**	**$17,578,837**
	$/kW		$36.0	$19.2	$17.6
Natural Gas Reburning O&M Costs			100 MW	500 MW	1000 MW
Cost Basis (Year) →			***2007***	***2007***	***2007***
Electrical Consumption Savings	$/yr		(39,237)	(196,187)	(392,375)
Maintenance	$/yr		51,252	126,673	223,689
Waste Disposal Savings	$/yr		(3,176)	(15,879)	(31,757)
Natural Gas Consumption	$/yr		1,005,801	5,029,007	10,058,013
Total O&M Costs			**1,014,640**	**4,943,613**	**9,857,570**

LSFO Material Balance - Preliminary

			100 MW	500 MW	1000 MW
Flue Gas, Downstream of ID Fans					
Temperature	°F		295	295	295
Pressure	in. H2O		10	10	10
Flow Rate	SCFM		254,007	1,270,033	2,540,066
Flow Rate	ACFM		366,173	1,830,865	3,661,730
CO2	lb/hr		214,735	1,073,676	2,147,353
N2	lb/hr		811,940	4,059,702	8,119,403
SO2	lb/hr		1,050	5,251	10,503
O2	lb/hr		64,641	323,206	646,412
HCl	lb/hr		7	37	74
Other Gases	lb/hr		308	1,542	3,083
H2O	lb/hr		76,485	382,427	764,853
Fly Ash	lb/hr		32	158	315
Total (gas only)	lb/hr		1,169,168	5,845,840	11,691,681
Flue Gas, to Absorber					
Temperature	°F		295	295	295
Pressure	in. H2O		10	10	10
Flow Rate	SCFM		127,003	1,270,033	1,270,033
Flow Rate	ACFM		183,087	1,830,865	1,830,865
CO2	lb/hr		107,368	1,073,676	1,073,676
N2	lb/hr		405,970	4,059,702	4,059,702
SO2	lb/hr		525	5,251	5,251
O2	lb/hr		32,321	323,206	323,206
HCl	lb/hr		4	37	37
Other Gases	lb/hr		154	1,542	1,542
H2O	lb/hr		38,243	382,427	382,427
Fly Ash	lb/hr		16	158	158
Total (gas only)	lb/hr		584,584	5,845,840	5,845,840
Flue Gas, from Absorbers (total)					
Temperature	°F		127	127	127
Pressure	in. H2O		4	4	4
Flow Rate	SCFM		132,770	1,323,336	1,327,697
Flow Rate	ACFM		151,018	1,505,223	1,510,183
CO2	lb/hr		108,038	1,077,041	1,080,384
N2	lb/hr		409,194	4,075,820	4,091,939
SO2	lb/hr		-448	385	-4,480
O2	lb/hr		33,050	326,852	330,497
HCl	lb/hr		0	0	0
Other Gases	lb/hr		154	1,542	1,542
H2O	lb/hr		52,147	521,470	521,470
Fly Ash	lb/hr		16	158	158
Total (gas only)	lb/hr		602,135	6,003,110	6,021,351
Heat Capacities					
O2	Btu/lbmol°F		7.213	7.213	7.213
CO2	Btu/lbmol°F		9.354	9.354	9.354
N2	Btu/lbmol°F		6.999	6.999	6.999
H2O	Btu/lbmol°F		8.069	8.069	8.069
NO	Btu/lbmol°F		7.164	7.164	7.164
SO2	Btu/lbmol°F		9.830	9.830	9.830
HCl	Btu/lbmol°F		6.951	6.951	6.951
O2	Btu/lb°F		0.225	0.225	0.225
CO2	Btu/lb°F		0.213	0.213	0.213
N2	Btu/lb°F		0.250	0.250	0.250
H2O	Btu/lb°F		0.448	0.448	0.448
NO	Btu/lb°F		0.239	0.239	0.239
SO2	Btu/lb°F		0.153	0.153	0.153
HCl	Btu/lb°F		0.191	0.191	0.191
Reheated Gas Temperature:	°C		66.7	66.7	66.7
	K		339.7	339.7	339.7
FGD Outlet Temperature:	°C		52.8	52.8	52.8
	K		325.8	325.8	325.8
Total Btu/hr	Btu/hr		3,898,723	38,866,938	38,987,234

Hot Reheat Air					
Temperature	°F		440	440	440
Pressure	in. H2O		1	1	1
Flow Rate	SCFM		11,970	119,328	119,697
Flow Rate	ACFM		21,031	209,658	210,307
N2	lb/hr		41,068	409,413	410,680
O2	lb/hr		12,385	123,469	123,851
H2O	lb/hr		704	7,019	7,040
Total	lb/hr		54,157	539,900	541,571
Heat Capacities of Hot Reheat Air					
O2	Btu/lbmol°F		7.332	7.332	7.332
N2	Btu/lbmol°F		7.113	7.113	7.113
H2O	Btu/lbmol°F		8.338	8.338	8.338
O2	Btu/lb°F		0.229	0.229	0.229
N2	Btu/lb°F		0.254	0.254	0.254
H2O	Btu/lb°F		0.463	0.463	0.463
Heated Temperature:	°C		226.7	226.7	226.7
	K		499.7	499.7	499.7
FGD Outlet Temperature:	°C		66.7	66.7	66.7
	K		339.7	339.7	339.7
Heat Capacities of Inlet Reheat Air					
O2	Btu/lbmol°F		7.304	7.304	7.304
N2	Btu/lbmol°F		7.087	7.087	7.087
H2O	Btu/lbmol°F		8.276	8.276	8.276
O2	Btu/lb°F		0.228	0.228	0.228
N2	Btu/lb°F		0.253	0.253	0.253
H2O	Btu/lb°F		0.459	0.459	0.459
Heated Temperature:	°C		226.7	226.7	226.7
	K		499.7	499.7	499.7
Inlet Air Temperature:	°C		26.7	26.7	26.7
	K		299.7	299.7	299.7
Required Heat	Btu/hr		4,873,404	48,583,673	48,734,043
Oxidation Air (total)					
Temperature	°F		60	60	60
Pressure	in. H2O		0	0	0
Flow Rate	SCFM		939	4,697	9,393
Flow Rate	ACFM		956	4,780	9,560
N2	lb/hr		3,224	16,119	32,237
O2	lb/hr		972	4,861	9,722
H2O	lb/hr		55	273	545
Total (gas only)	lb/hr		4,250	21,252	42,505
Limestone to Ball Mill					
Temperature	°F		80	80	80
Wt.% Solids	wt. %		100%	100%	100%
Inerts	lb/hr		82	412	825
CaCO3	lb/hr		1,673	8,363	16,725
Total	lb/hr		1,755	8,775	17,550
Limestone Slurry to Limestone Slurry Tank					
Temperature	°F		90	90	90
Flow Rate	GPM		6.3	31.3	62.6
Wt.% Solids	wt. %		40%	40%	40%
Inerts	lb/hr		82	412	825
CaCO3	lb/hr		1,673	8,363	16,725
H2O	lb/hr		2,633	13,163	26,325
Total	lb/hr		4,388	21,938	43,875
Limestone Slurry to Reaction Mix Tank (total)					
Temperature	°F		68	68	68
Flow Rate	GPM		14.6	73.1	146.1
Wt.% Solids	wt. %		20%	20%	20%
Inerts	lb/hr		82	412	825
CaCO3	lb/hr		1,673	8,363	16,725
H2O	lb/hr		7,020	35,100	70,200
Total	lb/hr		8,775	43,875	87,750

Slurry to Absorber

Temperature	°F		126	126	126
Flow Rate	GPM		18,647	185,854	186,467
Wt.% Solids	wt. %		15%	15%	15%
CaSO3*1/2H2O	lb/hr		0	0	0
CaSO4*2H2O	lb/hr		1,476,536	14,716,868	14,765,362
Inerts	lb/hr		46,566	464,128	465,657
CaCl2	lb/hr		6,359	63,381	63,590
CaCO3	lb/hr		80,101	798,383	801,014
H2O	lb/hr		9,120,853	90,908,975	91,208,534
Total	lb/hr		10,730,416	106,951,735	107,304,158

Slurry from Rxn Tank to Thickener

Temperature	°F		126	126	126
Flow Rate	GPM		33.0	165.2	330.3
Wt.% Solids	wt. %		15%	15%	15%
CaSO3*1/2H2O	lb/hr		0	0	0
CaSO4*2H2O	lb/hr		2,615	13,077	26,155
Inerts	lb/hr		82	412	825
CaCl2	lb/hr		11	56	113
CaCO3	lb/hr		142	709	1,419
H2O	lb/hr		16,156	80,782	161,564
Total	lb/hr		19,008	95,038	190,075

LSFO Equipment Capital Costs

			100 MW	500 MW	1000 MW
Cost Basis (Year)			**2007**	**2007**	**2007**
		Sizing Criteria			
Reagent Feed System	$	kpph Reag.	$9,799,235	$10,399,592	$11,067,980
Ball Mill & Hydroclone System	$	TPH Reag.	$2,538,432	$2,645,779	$2,781,507
DBA Acid Tank (pump, heater, agitator)	$	gpm DBA	$0	$0	$0
SO2 Removal System	$	kpph SO2	$2,629,240	$2,820,685	$3,048,323
Absorber Tower	$	kACFM	$11,755,001	$21,527,454	$43,054,908
Spray Pumps	$	slurry gpm	$430,349	$3,244,695	$3,251,057
Flue Gas Handling System	$	*	$4,110,920	$9,081,988	$9,901,999
ID Fans	$	ACFM	$1,023,198	$3,077,477	$6,154,955
Waste / Byproduct Handling System	$	kpph SO2	$1,088,188	$1,247,695	$1,436,100
Thickener System	$	TPH solids	$172,541	$242,179	$329,227
Support Equipment	$	MW	$1,850,663	$2,634,311	$3,256,361
Chimney	$	ACFM	$2,919,099	$6,290,306	$6,297,219
TOTAL	$		$38,316,866	$63,212,161	$90,579,637

* *Based on flue gas flow and reheat temperature.*

Capital Costs with Retrofit Factors

Reagent Feed System	$		$9,799,235	$10,399,592	$11,067,980
Ball Mill & Hydroclone System	$		$2,538,432	$2,645,779	$2,781,507
DBA Acid Tank (pump, heater, agitator)	$		$0	$0	$0
SO2 Removal System	$		$2,629,240	$2,820,685	$3,048,323
Absorber Tower	$		$11,755,001	$21,527,454	$43,054,908
Spray Pumps	$		$430,349	$3,244,695	$3,251,057
Flue Gas Handling System	$		$4,110,920	$9,081,988	$9,901,999
ID Fans	$		$1,023,198	$3,077,477	$6,154,955
Waste / Byproduct Handling System	$		$1,088,188	$1,247,695	$1,436,100
Thickener System	$		$172,541	$242,179	$329,227
Support Equipment	$		$1,850,663	$2,634,311	$3,256,361
Chimney	$		$2,919,099	$6,290,306	$6,297,219
TOTAL	$		$38,316,866	$63,212,161	$90,579,637

General Facilities

Reagent Feed System	$		$979,923	$1,039,959	$1,106,798
Ball Mill & Hydroclone System	$		$253,843	$264,578	$278,151
DBA Acid Tank (pump, heater, agitator)	$		$0	$0	$0
SO2 Removal System	$		$262,924	$282,068	$304,832
Absorber Tower	$		$1,175,500	$2,152,745	$4,305,491
Spray Pumps	$		$43,035	$324,470	$325,106
Flue Gas Handling System	$		$411,092	$908,199	$990,200
ID Fans	$		$102,320	$307,748	$615,495
Waste / Byproduct Handling System	$		$108,819	$124,769	$143,610
Thickener System	$		$17,254	$24,218	$32,923
Support Equipment	$		$185,066	$263,431	$325,636
Chimney	$		$291,910	$629,031	$629,722
TOTAL	$		$3,831,687	$6,321,216	$9,057,964
Engineering Fees					
Reagent Feed System	$		$979,923	$1,039,959	$1,106,798
Ball Mill & Hydroclone System	$		$253,843	$264,578	$278,151
DBA Acid Tank (pump, heater, agitator)	$		$0	$0	$0
SO2 Removal System	$		$262,924	$282,068	$304,832
Absorber Tower	$		$1,175,500	$2,152,745	$4,305,491
Spray Pumps	$		$43,035	$324,470	$325,106
Flue Gas Handling System	$		$411,092	$908,199	$990,200
ID Fans	$		$102,320	$307,748	$615,495
Waste / Byproduct Handling System	$		$108,819	$124,769	$143,610
Thickener System	$		$17,254	$24,218	$32,923
Support Equipment	$		$185,066	$263,431	$325,636
Chimney	$		$291,910	$629,031	$629,722
TOTAL	$		$3,831,687	$6,321,216	$9,057,964
Contingency					
Reagent Feed System	$		$1,959,847	$2,079,918	$2,213,596
Ball Mill & Hydroclone System	$		$507,686	$529,156	$556,301
DBA Acid Tank (pump, heater, agitator)	$		$0	$0	$0
SO2 Removal System	$		$525,848	$564,137	$609,665
Absorber Tower	$		$2,351,000	$4,305,491	$8,610,982
Spray Pumps	$		$86,070	$648,939	$650,211
Flue Gas Handling System	$		$822,184	$1,816,398	$1,980,400
ID Fans	$		$204,640	$615,495	$1,230,991
Waste / Byproduct Handling System	$		$217,638	$249,539	$287,220
Thickener System	$		$34,508	$48,436	$65,845
Support Equipment	$		$370,133	$526,862	$651,272
Chimney	$		$583,820	$1,258,061	$1,259,444
TOTAL	$		$7,663,373	$12,642,432	$18,115,927
Total Plant Cost (TPC)	$		$53,643,612	$88,497,026	$126,811,492
Total Plant Cost (TPC) w/ Prime Contractor's Markup	$		$55,252,921	$91,151,936	$130,615,836
Total Cash Expended (TCE)	$		$55,252,921	$88,522,801	$125,019,318
Allow. for Funds During Constr. (AFDC)	$		$0	$9,706,148	$21,196,693
Total Plant Investment (TPI)	$		$55,252,921	$98,228,949	$146,216,011
Preproduction Costs	$		$1,539,611	$3,174,676	$4,857,418
Inventory Capital	$		$18,954	$94,770	$189,540
Total Capital Requirement (TCR)	$		**$56,811,486**	**$101,498,395**	**$151,262,969**
	$/kW		$568	$203	$151

Maintenance Cost by Area

			100 MW	500 MW	1000 MW
TPC w/o Retrofit Factor					
Reagent Feed System	$		$14,110,898	$14,975,413	$15,937,892
Ball Mill & Hydroclone System	$		$3,655,342	$3,809,922	$4,005,370
DBA Acid Tank (pump, heater, agitator)	$		$0	$0	$0
SO2 Removal System	$		$3,786,106	$4,061,786	$4,389,585
Absorber Tower	$		$16,927,201	$30,999,534	$61,999,067
Spray Pumps	$		$619,703	$4,672,361	$4,681,522
Flue Gas Handling System	$		$5,919,724	$13,078,062	$14,258,879
ID Fans	$		$1,473,405	$4,431,568	$8,863,135
Waste / Byproduct Handling System	$		$1,566,991	$1,796,680	$2,067,985
Thickener System	$		$248,459	$348,738	$474,087
Support Equipment	$		$2,664,955	$3,793,408	$4,689,161
Chimney	$		$4,203,502	$9,058,040	$9,067,996
TOTAL	$		$55,176,287	$91,025,512	$130,434,677
First Year Maintenance Costs					
Reagent Feed System	$		$705,545	$748,771	$796,895
Ball Mill & Hydroclone System	$		$182,767	$190,496	$200,268
DBA Acid Tank (pump, heater, agitator)	$		$0	$0	$0
SO2 Removal System	$		$189,305	$203,089	$219,479
Absorber Tower	$		$846,360	$1,549,977	$3,099,953
Spray Pumps	$		$30,985	$233,618	$234,076
Flue Gas Handling System	$		$295,986	$653,903	$712,944
ID Fans	$		$73,670	$221,578	$443,157
Waste / Byproduct Handling System	$		$78,350	$89,834	$103,399
Thickener System	$		$12,423	$17,437	$23,704
Support Equipment	$		$133,248	$189,670	$234,458
Chimney	$		$210,175	$452,902	$453,400
TOTAL	$		$2,758,814	$4,551,276	$6,521,734

LSFO O&M Data and Costs

			100 MW	500 MW	1000 MW
Cost Basis (Year)		→	***2007***	***2007***	***2007***
Parameters					
Reagent Required	lbs/hr		1,755	8,775	17,550
	lbs/MMBtu		1.671	1.671	1.671
DBA Required	lbs/hr		0.0	0.0	0.0
Percent SO2 Removal	%		93%	93%	93%
FGD Sludge to Disposal	lbs/hr, dry		2,851	14,256	28,511
Steam to FGD System	lbs/hr		5,699	56,814	56,990
Total FGD Power Consumption	kW		2,000	10,000	20,000
FGD Byproduct	lbs/hr		0	0	0
Fixed O&M Costs					
Number of Operators			9	28	45
(40 hrs/week)					
Operating Labor Cost **	$/yr		$786,223	$2,340,090	$3,743,161
Maint. Labor & Matls. Cost	$/yr		$2,758,814	$4,551,276	$6,521,734
Admin. & Support Labor	$/yr		$566,924	$1,248,180	$1,905,556
TOTAL	$/yr		$4,111,961	$8,139,546	$12,170,451
Variable Operating Costs **					
Reagent Costs	$/yr		$103,773	$518,866	$1,037,733
DBA Costs	$/yr		$0	$0	$0
Disposal Costs	$/yr		$337,174	$1,685,872	$3,371,744

Credit for Byproduct	$/yr		$0	$0	$0
Steam Costs	$/yr		$157,257	$1,567,717	$1,572,570
Power Costs	$/yr		$394,200	$1,971,000	$3,942,000
TOTAL	$/yr		$992,405	$5,743,456	$9,924,046

*** These costs assume inputs are in current dollars (no escalation included).*

Intermediate Material Balance Calcs.

			100 MW	500 MW	1000 MW
Sulfite Reaction					
SO2	lbmole/hr		15.19	75.96	151.91
CaCO3	lbmole/hr		15.19	75.96	151.91
H2O	lbmole/hr		7.60	37.98	75.96
CaSO3*1/2H2O	lbmole/hr		15.19	75.96	151.91
CO2	lbmole/hr		15.19	75.96	151.91
SO2	lb/hr		973	4,866	9,732
CaCO3	lb/hr		1,520	7,602	15,205
H2O	lb/hr		137	684	1,368
CaSO3*1/2H2O	lb/hr		1,962	9,810	19,619
CO2	lb/hr		669	3,343	6,686
Sulfate Reaction					
CaSO3*1/2H2O	lbmole/hr		15.19	75.96	151.91
O2	lbmole/hr		7.60	37.98	75.96
H2O	lbmole/hr		22.79	113.93	227.87
CaSO4*2H2O	lbmole/hr		15.19	75.96	151.91
CaSO3*1/2H2O	lb/hr		1,962	9,810	19,619
O2	lb/hr		243	1,215	2,430
H2O	lb/hr		411	2,053	4,105
CaSO4*2H2O	lb/hr		2,615	13,077	26,155
Water in Absorber					
Mole Fraction H2O in Absorber	%		0.1389	0.1389	0.1389
Moles H2O in Absorber	lbmole		2,894.54	28,945.35	28,945.35
DBA Feed Calculations					
SO2 Removed	lbs/hr		973	4,866	9,732
DBA Added	lbs/hr		0.00	0.00	0.00
DBA Added	GPM		0.00	0.00	0.00

LSD Material Balance - Preliminary

		100 MW	500 MW	1000 MW
Flue Gas, Downstream of Air Heater				
Temperature	°F	300	300	300
Pressure	in. H2O	-12	-12	-12
Flow Rate	SCFM	254,007	1,270,033	2,540,066
Flow Rate	ACFM	389,483	1,947,414	3,894,828
CO2	lb/hr	214,735	1,073,676	2,147,353
N2	lb/hr	811,940	4,059,702	8,119,403
SO2	lb/hr	1,050	5,251	10,503
O2	lb/hr	64,641	323,206	646,412
HCl	lb/hr	7	37	74
Other Gases	lb/hr	308	1,542	3,083
H2O	lb/hr	76,485	382,427	764,853
Fly Ash	lb/hr	14,685	73,427	146,854
Total (gas only)	lb/hr	1,169,168	5,845,840	11,691,681
Flue Gas, to Spray Dryer				
Temperature	°F	300	300	300
Pressure	in. H2O	-12	-12	-12
Flow Rate	SCFM	63,502	635,016	635,016
Flow Rate	ACFM	97,371	973,707	973,707
CO2	lb/hr	53,684	536,838	536,838
N2	lb/hr	202,985	2,029,851	2,029,851
SO2	lb/hr	263	2,626	2,626
O2	lb/hr	16,160	161,603	161,603
HCl	lb/hr	2	19	19
Other Gases	lb/hr	77	771	771
H2O	lb/hr	19,121	191,213	191,213
Fly Ash	lb/hr	3,671	36,714	36,714
Total (gas only)	lb/hr	292,292	2,922,920	2,922,920
Flue Gas, from Spray Dryers (total)				
Temperature	°F	147	147	147
Pressure	in. H2O	-17	-17	-17
Flow Rate	SCFM	292,982	1,373,113	2,745,597
Flow Rate	ACFM	363,486	1,703,544	3,406,310
CO2	lb/hr	214,735	1,073,676	2,147,353
N2	lb/hr	811,940	4,059,702	8,119,403
SO2	lb/hr	306	1,528	3,057
O2	lb/hr	64,595	322,973	645,947
HCl	lb/hr	0	0	0
Other Gases	lb/hr	308	1,542	3,083
H2O	lb/hr	187,634	676,800	1,351,812
Fly Ash	lb/hr	39,423	197,117	394,234
Total (gas only)	lb/hr	1,279,518	6,136,221	12,270,655
Flue Gas Downstream of Particulate Control Device				
Temperature	°F	147	147	147
Pressure	in. H2O	-23	-23	-23
Flow Rate	SCFM	292,679	1,371,602	2,742,576
Flow Rate	ACFM	368,886	1,728,732	3,456,674
CO2	lb/hr	214,735	1,073,676	2,147,353
N2	lb/hr	811,940	4,059,702	8,119,403
SO2	lb/hr	120	598	1,195
O2	lb/hr	64,583	322,915	645,831

HCl	lb/hr		0	0	0
Other Gases	lb/hr		308	1,542	3,083
H2O	lb/hr		186,832	672,794	1,343,800
Fly Ash	lb/hr		32	158	315
Total (gas only)	lb/hr		1,278,519	6,131,226	12,260,664
Flue Gas Downstream of ID Fans					
Temperature	°F		152	152	152
Pressure	in. H2O		1	1	1
Flow Rate	SCFM		292,679	1,371,602	2,742,576
Flow Rate	ACFM		349,680	1,638,728	3,276,706
CO2	lb/hr		214,735	1,073,676	2,147,353
N2	lb/hr		811,940	4,059,702	8,119,403
SO2	lb/hr		120	598	1,195
O2	lb/hr		64,583	322,915	645,831
HCl	lb/hr		0	0	0
Other Gases	lb/hr		308	1,542	3,083
H2O	lb/hr		186,832	672,794	1,343,800
Fly Ash	lb/hr		32	158	315
Total (gas only)	lb/hr		1,278,519	6,131,226	12,260,664
Lime to Ball Mill					
Temperature	°F		60	60	60
Wt.% Solids	wt. %		100%	100%	100%
Inerts	lb/hr		92	462	924
CaO	lb/hr		831	4,157	8,314
Total	lb/hr		924	4,619	9,238
Water to Ball Mill					
Temperature	°F		60	60	60
Flow Rate	GPM		6	30	61
Wt.% Solids	wt. %		0%	0%	0%
H2O	lb/hr		3,046	15,229	30,457
Total	lb/hr		3,046	15,229	30,457
Lime Slurry to Head Tanks (Total)					
Temperature	°F		110	110	110
Flow Rate	GPM		6	31	61
Wt.% Solids	wt. %		30%	30%	30%
Ca(OH)2	lb/hr		1,098	5,492	10,985
Inerts	lb/hr		92	462	924
H2O	lb/hr		2,779	13,893	27,787
Total	lb/hr		3,970	19,848	39,695
Lime Slurry from Head Tank					
Temperature	°F		110	110	110
Flow Rate	GPM		30	304	304
Wt.% Solids	wt. %		35%	35%	35%
CaSO3*1/2H2O	lb/hr		560	5,602	5,602
CaSO4*2H2O	lb/hr		249	2,489	2,489
Flyash / Inerts	lb/hr		5,989	59,891	59,891
Ca(OH)2	lb/hr		286	2,864	2,864
CaCl2	lb/hr		214	2,143	2,143
H2O	lb/hr		13,299	132,989	132,989
Total	lb/hr		20,598	205,978	205,978
Lime Slurry to Atomizer					

Temperature	°F		104	104	104
Flow Rate	GPM		58	331	331
Wt.% Solids	wt. %		21%	33%	33%
CaSO3*1/2H2O	lb/hr		560	5,602	5,602
CaSO4*2H2O	lb/hr		249	2,489	2,489
Flyash / Inerts	lb/hr		5,989	59,891	59,891
Ca(OH)2	lb/hr		286	2,864	2,864
CaCl2	lb/hr		214	2,143	2,143
H2O	lb/hr		27,869	147,559	147,559
Total	lb/hr		35,168	220,549	220,549
Solids from Spray Dryers (Total)					
Temperature	°F		150	150	150
Wt.% Solids	wt. %		98%	98%	98%
CaSO3*1/2H2O	lb/hr		337	1,683	3,367
CaSO4*2H2O	lb/hr		150	748	1,496
Flyash / Inerts	lb/hr		3,855	19,276	38,552
Ca(OH)2	lb/hr		28	138	277
CaCl2	lb/hr		11	56	113
H2O	lb/hr		89	447	894
Total	lb/hr		4,470	22,349	44,698
Baghouse/ESP Solids to Recycle					
Particulate Removal Efficiency	%		99.92%	99.92%	99.92%
Temperature	°F		150	150	150
Wt.% Solids	wt. %		98%	98%	98%
CaSO3*1/2H2O	lb/hr		1,904	9,520	19,040
CaSO4*2H2O	lb/hr		846	4,231	8,461
Flyash / Inerts	lb/hr		20,009	100,044	200,089
Ca(OH)2	lb/hr		19	97	195
CaCl2	lb/hr		846	4,231	8,461
H2O	lb/hr		465	2,324	4,649
Total	lb/hr		23,243	116,217	232,433
Recycle Solids to Slurry Tank					
Temperature	°F		150	150	150
Wt.% Solids	wt. %		98%	98%	98%
CaSO3*1/2H2O	lb/hr		2,241	11,203	22,407
CaSO4*2H2O	lb/hr		996	4,979	9,957
Flyash / Inerts	lb/hr		23,864	119,320	238,641
Ca(OH)2	lb/hr		47	236	471
CaCl2	lb/hr		857	4,287	8,574
H2O	lb/hr		554	2,771	5,543
Total	lb/hr		28,559	142,796	285,592
Blowdown Water to Recycle Solids Tank					
Temperature	°F		60	60	60
Flow Rate	GPM		100	498	996
Wt.% Solids	wt. %		0%	0%	0%
H2O	lb/hr		49,863	249,313	498,627
Total	lb/hr		49,863	249,313	498,627
Recycle Slurry to Head Tanks (Total)					
Temperature	°F		110	110	110
Flow Rate	GPM		116	580	1,161
Wt.% Solids	wt. %		35%	35%	35%
CaSO3*1/2H2O	lb/hr		2,241	11,203	22,407

CaSO4*2H2O	lb/hr		996	4,979	9,957
Flyash / Inerts	lb/hr		23,864	119,320	238,641
Ca(OH)2	lb/hr		47	236	471
CaCl2	lb/hr		857	4,287	8,574
H2O	lb/hr		50,417	252,085	504,169
Total	lb/hr		78,422	392,109	784,219
Dry Solids					
Temperature	°F		100	100	100
Wt.% Solids	wt. %		98%	98%	98%
CaSO3*1/2H2O	lb/hr		1,405	7,023	14,047
CaSO4*2H2O	lb/hr		624	3,121	6,242
Flyash / Inerts	lb/hr		14,750	73,750	147,500
Ca(OH)2	lb/hr		14	72	144
CaCl2	lb/hr		23	113	225
H2O	lb/hr		343	1,715	3,429
Total	lb/hr		17,159	85,794	171,587
Solids to Landfill					
Temperature	°F		100	100	100
Wt.% Solids	wt. %		80%	80%	80%
CaSO3*1/2H2O	lb/hr		1,405	7,023	14,047
CaSO4*2H2O	lb/hr		624	3,121	6,242
Flyash / Inerts	lb/hr		14,750	73,750	147,500
Ca(OH)2	lb/hr		14	72	144
CaCl2	lb/hr		23	113	225
H2O	lb/hr		4,204	21,020	42,039
Total	lb/hr		21,020	105,099	210,197
LSD Equipment Capital Costs			100 MW	500 MW	1000 MW
Cost Basis (Year)			***2007***	***2007***	***2007***
		Sizing Criteria			
Reagent Feed System	$	*	$5,485,007	$6,453,105	$7,588,887
SO2 Removal System	$	Wt. % S	$2,421,462	$1,987,146	$2,421,462
Spray Dryers	$	kACFM	$12,218,749	$23,537,426	$47,074,851
Flue Gas Handling System	$	kACFM	$2,500,723	$5,189,006	$9,599,878
ID Fans	$	ACFM	$795,630	$2,289,270	$4,577,822
Waste / Byproduct Handling System	$	kpph SO2	$1,432,950	$1,432,950	$1,432,950
Support Equipment	$	MW	$2,184,237	$3,255,403	$3,856,431
Chimney	$	ACFM	$3,863,693	$6,471,346	$8,155,965
TOTAL	$		$30,902,451	$50,615,652	$84,708,247
** Based on lbs/hr of lime feed and GPM of lime slurry.*					
Capital Costs with Retrofit Factors					
Reagent Feed System	$		$5,485,007	$6,453,105	$7,588,887
SO2 Removal System	$		$2,421,462	$1,987,146	$2,421,462
Spray Dryers	$		$12,218,749	$23,537,426	$47,074,851
Flue Gas Handling System	$		$2,500,723	$5,189,006	$9,599,878
ID Fans	$		$795,630	$2,289,270	$4,577,822
Waste / Byproduct Handling System	$		$1,432,950	$1,432,950	$1,432,950
Support Equipment	$		$2,184,237	$3,255,403	$3,856,431
Chimney	$		$3,863,693	$6,471,346	$8,155,965

TOTAL	$		$30,902,451	$50,615,652	$84,708,247
General Facilities					
Reagent Feed System	$		$548,501	$645,311	$758,889
SO2 Removal System	$		$242,146	$198,715	$242,146
Spray Dryers	$		$1,221,875	$2,353,743	$4,707,485
Flue Gas Handling System	$		$250,072	$518,901	$959,988
ID Fans	$		$79,563	$228,927	$457,782
Waste / Byproduct Handling System	$		$143,295	$143,295	$143,295
Support Equipment	$		$218,424	$325,540	$385,643
Chimney	$		$386,369	$647,135	$815,597
TOTAL	$		$3,090,245	$5,061,565	$8,470,825
Engineering Fees					
Reagent Feed System	$		$548,501	$645,311	$758,889
SO2 Removal System	$		$242,146	$198,715	$242,146
Spray Dryers	$		$1,221,875	$2,353,743	$4,707,485
Flue Gas Handling System	$		$250,072	$518,901	$959,988
ID Fans	$		$79,563	$228,927	$457,782
Waste / Byproduct Handling System	$		$143,295	$143,295	$143,295
Support Equipment	$		$218,424	$325,540	$385,643
Chimney	$		$386,369	$647,135	$815,597
TOTAL	$		$3,090,245	$5,061,565	$8,470,825
Contingency					
Reagent Feed System	$		$1,097,001	$1,290,621	$1,517,777
SO2 Removal System	$		$484,292	$397,429	$484,292
Spray Dryers	$		$2,443,750	$4,707,485	$9,414,970
Flue Gas Handling System	$		$500,145	$1,037,801	$1,919,976
ID Fans	$		$159,126	$457,854	$915,564
Waste / Byproduct Handling System	$		$286,590	$286,590	$286,590
Support Equipment	$		$436,847	$651,081	$771,286
Chimney	$		$772,739	$1,294,269	$1,631,193
TOTAL	$		$6,180,490	$10,123,130	$16,941,649
Total Plant Cost (TPC)	$		$43,263,431	$70,861,913	$118,591,545
Total Plant Cost (TPC) w/ Prime Contractor's M	$		$44,561,334	$72,987,770	$122,149,292
Total Cash Expended (TCE)	$		$44,561,334	$70,882,552	$116,915,541
Allow. for Funds During Constr. (AFD	$		$0	$7,771,969	$19,822,719
Total Plant Investment (TPI)	$		$44,561,334	$78,654,522	$136,738,260
Preproduction Costs	$		$1,391,185	$3,209,471	$5,754,172
Inventory Capital	$		$49,884	$249,418	$498,836
Total Capital Requirement (TCR)	$		**$46,002,403**	**$82,113,411**	**$142,991,268**
	$/kW		$460	$164	$143

Maintenance Cost by Area

			100 MW	500 MW	1000 MW
TPC w/o Retrofit Factor					
Reagent Feed System	$		$7,898,410	$9,292,471	$10,927,997
SO2 Removal System	$		$3,486,906	$2,861,491	$3,486,906
Spray Dryers	$		$17,594,999	$33,893,893	$67,787,786
Flue Gas Handling System	$		$3,601,041	$7,472,168	$13,823,825
ID Fans	$		$1,145,707	$3,296,548	$6,592,064
Waste / Byproduct Handling System	$		$2,063,448	$2,063,448	$2,063,448
Support Equipment	$		$3,145,301	$4,687,780	$5,553,260
Chimney	$		$5,563,717	$9,318,739	$11,744,590
TOTAL	$		$44,499,529	$72,886,539	$121,979,875
First Year Maintenance Costs					
Reagent Feed System	$		$394,920	$464,624	$546,400
SO2 Removal System	$		$174,345	$143,075	$174,345
Spray Dryers	$		$879,750	$1,694,695	$3,389,389
Flue Gas Handling System	$		$180,052	$373,608	$691,191
ID Fans	$		$57,285	$164,827	$329,603
Waste / Byproduct Handling System	$		$103,172	$103,172	$103,172
Support Equipment	$		$157,265	$234,389	$277,663
Chimney	$		$278,186	$465,937	$587,230
TOTAL	$		$2,224,976	$3,644,327	$6,098,994

LSD O&M Data and Costs

			100 MW	500 MW	1000 MW
Cost Basis (Year)			***2007***	***2007***	***2007***
Parameters					
Reagent Required	lbs/hr		924	4,619	9,238
	lbs/MMBtu		0.880	0.880	0.880
Percent SO2 Removal	%		89%	89%	89%
FGD Solids - dry	lbs/hr		16,816	84,079	168,158
- wetted	lbs/hr		21,020	105,099	210,197
Fresh Water to FGD	gpm		6	30	61
Blowdown Water to FGD	gpm		136	566	1,103
Total FGD Power Consumption	kW		700	3,500	7,000
Fixed O&M Costs					
Number of Operators			8	20	25
(40 hrs/week)					
Operating Labor Cost **	$/yr		$645,584	$1,702,739	$2,091,759
Maint. Labor & Matls. Cost	$/yr		$2,224,976	$3,644,327	$6,098,994
Admin. & Support Labor	$/yr		$460,672.35	$948,141	$1,359,407
TOTAL	$/yr		$3,331,233	$6,295,207	$9,550,160
Variable Operating Costs **					
Reagent Costs	$/yr		$273,113	$1,365,563	$2,731,127
Disposal Costs	$/yr		$1,988,635	$9,943,174	$19,886,347
Credit for Byproduct	$/yr		$0	$0	$0

Steam Costs	$/yr		$0	$0	$0
Fresh Water Costs	$/yr		$1,728	$8,638	$17,275
Power Costs	$/yr		$137,970	$689,850	$1,379,700
TOTAL	$/yr		$2,401,445	$12,007,225	$24,014,449
** *These costs assume inputs are in current dollars (no escalation included).*					

Fabric Filter - Preliminary

		100 MW	500 MW	1000 MW
Flue Gas, Upstream of Fabric Filter				
Temperature	°F	147	147	147
Pressure	in. H2O	-17	-17	-17
Flow Rate	SCFM	292,982	1,373,113	2,745,597
Flow Rate	ACFM	363,486	1,703,544	3,406,310
CO2	lb/hr	214,735	1,073,676	2,147,353
N2	lb/hr	811,940	4,059,702	8,119,403
SO2	lb/hr	306	1,528	3,057
O2	lb/hr	64,595	322,973	645,947
HCl	lb/hr	0	0	0
Other Gases	lb/hr	308	1,542	3,083
H2O	lb/hr	187,634	676,800	1,351,812
Fly Ash	lb/hr	39,423	197,117	394,234
Total (gas only)	lb/hr	1,279,518	6,136,221	12,270,655
Total Fabric Required	Ft^2	103,853	486,727	973,231
Surface Area per Bag	Ft^2	31.4	31.4	31.4
Required No. of Bags (no spare compartments)		3,306	15,493	30,979
Final No. of Bags		3,636	17,042	34,077
No. of Casings		1	1	2
Fabric Filter Dimensions (per Casing)	Ft^2	3,570	16,731	16,727
Length	Ft	84	183	183
Width	Ft	42	91	91

Capital Cost

		100 MW	500 MW	1000 MW
Cost Basis (Year)	→	***2007***	***2007***	***2007***
Fabric Filter	$	$2,790,010	$8,339,028	$16,675,353
Bags	$	$259,395	$1,215,701	$2,430,847
Ash Handling System	$	$929,442	$3,466,560	$6,637,958
ID Fan(s)	$	$395,948	$1,030,158	$1,581,903
Equipment Cost Subtotal	$	$4,374,795	$14,051,448	$27,326,061
Instruments & Controls	$	$87,496	$281,029	$546,521
Taxes	$	$262,488	$843,087	$1,639,564
Freight	$	$218,740	$702,572	$1,366,303
Purchased Equipment Cost Subtotal	$	$4,943,518	$15,878,136	$30,878,449
Installation	$	$3,312,157	$10,638,351	$20,688,561
Total Direct Cost	$	$8,255,675	$26,516,488	$51,567,011
Total Direct Cost with Retrofit Factor	$	$8,255,675	$26,516,488	$51,567,011
General Facilities	$	$825,568	$2,651,649	$5,156,701
Engineering Fees	$	$825,568	$2,651,649	$5,156,701
Contingency	$	$1,651,135	$5,303,298	$10,313,402
Total Plant Cost (TPC)	$	$11,557,945	$37,123,083	$72,193,815
Total Plant Cost (TPC) w/ Prime Contractor's M	$	$11,904,683	$38,236,775	$74,359,629
Total Cash Expended (TCE)	$	$11,731,314	$37,679,929	$73,276,722

Allow. for Funds During Constr. (AFDC)	$		$624,129	$2,004,646	$3,898,466
Total Plant Investment (TPI)	$		$12,355,443	$39,684,575	$77,175,188
Preproduction Costs	$		$247,109	$793,692	$1,543,504
Inventory Capital	$		$0	$0	$0
Total Capital Requirement (TCR)	$		**$12,602,552**	**$40,478,267**	**$78,718,692**
	$/kW		$126.0	$81.0	$78.7

O&M Data and Costs

			100 MW	500 MW	1000 MW
Cost Basis (Year)			***2007***	***2007***	***2007***
Power Required Excluding ID Fan(s)	kW		163	721	1,409
ID Fan Power for FF Delta P	kW		367	1,720	3,440
Total Power	kW		530	2,442	4,849
Power Cost **	$/yr		$104,460	$627,964	$1,246,981
Maintenance Costs	$/yr		$577,897	$1,856,154	$3,609,691
Periodic Replacement Items	$/yr		$259,395	$1,215,701	$2,430,847
First Year Cost. Bags Replaced Every			*5 Years*	*5 Years*	*5 Years*

** *These costs assume inputs are in current dollars (no escalation included).*

ESP

			100 MW	500 MW	1000 MW
Flue Gas, Upstream of ESP					
Temperature	°F		147	147	147
Pressure	in. H2O		-17	-17	-17
Flow Rate	SCFM		292,982	1,373,113	2,745,597
Flow Rate	ACFM		363,486	1,703,544	3,406,310
CO2	lb/hr		214,735	1,073,676	2,147,353
N2	lb/hr		811,940	4,059,702	8,119,403
SO2	lb/hr		306	1,528	3,057
O2	lb/hr		64,595	322,973	645,947
HCl	lb/hr		0	0	0
Other Gases	lb/hr		308	1,542	3,083
H2O	lb/hr		187,634	676,800	1,351,812
Fly Ash	lb/hr		39,423	197,117	394,234
Total (gas only)	lb/hr		1,279,518	6,136,221	12,270,655
Inlet Particulate Loading	lb/hr		39,423	197,117	394,234
	gr/ft3		12.65	13.50	13.50
Overall PM Collection Efficiency	%	η	99.9%	99.9%	99.9%
ESP Requirements					
k	dimensionless		0.451	0.451	0.451
E	kV/cm		10.0	10.0	10.0
Ash Composition					
Na2O	wt% in Ash		1.68	1.68	1.68
Fe	wt% in Ash		6.07	6.07	6.07
MgO	wt% in Ash		31.97	31.97	31.97
CaO	wt% in Ash		0.00	0.00	0.00

Flue Gas Composition					
H2O	Vol%		23.70	17.83	17.81
SO2	ppm		109	113	113
SO3	ppm		0.8	0.8	0.8
Flue Gas Temperature					
TF	°F		147	147	147
TC	°C		64	64	64
TK	Kelvin		337	337	337
Resistivity Calculations					
Volume Resistivity					
rv1	Log10(ohm-cm)		9.13	9.13	9.13
rv2	Log10(ohm-cm)		8.89	8.89	8.89
iv	Log10(ohm-cm)		1.95	1.95	1.95
rv	Log10(ohm-cm)		14.81	14.81	14.81
rv(TK=1000/2.4)	Log10(ohm-cm)		12.35	12.35	12.35
Volume Resistivity	ohm-cm		6.4904E+14	6.4904E+14	6.4904E+14
Surface Resistivity					
rs1	Log10(ohm-cm)		10.27	10.27	10.27
rs2	Log10(ohm-cm)		8.39	9.14	9.14
rs3	Log10(ohm-cm)		7.16	7.91	7.91
rs4	Log10(ohm-cm)		6.92	7.67	7.67
rs4(with E=12)	Log10(ohm-cm)		6.86	7.61	7.61
rs0	Log10(ohm-cm)		9.87	9.88	9.88
rs	Log10(ohm-cm)		2.81	4.56	4.57
Surface Resistivity	ohm-cm		6.4305E+02	3.6653E+04	3.7186E+04
rvs1	ohm-cm		6.4305E+02	3.6653E+04	3.7186E+04
rs2(with H2O=10)	Log10(ohm-cm)		1.01E+01	1.01E+01	1.01E+01
rs3(with H2O=10)	Log10(ohm-cm)		8.91	8.91	8.91
rs4(with H2O = 10)	Log10(ohm-cm)		8.67	8.67	8.67
rs4(with H2O = 10, E = 12)	Log10(ohm-cm)		8.61	8.61	8.61
rs0(with H2O = 10)	Log10(ohm-cm)		9.88	9.88	9.88
rs(with H2O = 10, TK = 1000/2.4)	Log10(ohm-cm)		9.08	9.08	9.08
rvs1(with H2O=10, TK = 1000/2.4)	ohm-cm		1.1934E+09	1.1934E+09	1.1934E+09
rvs2	Log10(ohm-cm)		9.32	9.32	9.32
Acid Resistivity					
ia1	Log10(ohm-cm)		0.29	0.29	0.29
ra1	Log10(ohm-cm)		7.44	7.44	7.44
sa			-2.05	-2.05	-2.05
ia2	Log10(ohm-cm)		8.67	8.67	8.67
ra2	Log10(ohm-cm)		8.87	8.83	8.83
acid_v	index value		1	1	1
atom_v	index value		1	1	1
sa1			-4.74	-4.74	-4.74
ia3	Log10(ohm-cm)		20.24	20.20	20.20
ra3	Log10(ohm-cm)		6.18	6.14	6.14
ra	Log10(ohm-cm)		6.65	6.62	6.62
Acid Resistivity	ohm-cm		4.4180E+06	4.1400E+06	4.1390E+06
Total Resistivity					
rvsa	ohm-cm		6.4295E+02	3.6332E+04	3.6855E+04

Migration Velocity					
w_k	1000ft/minute		0.102	0.088	0.088
Specific Collection Area					
SCA	ft^2/1000acfm		**766**	**882**	**883**
Total Collector Plate Area	ft^2		278,511	1,502,866	3,006,657
ESP Footprint Area	ft^2		7,736	41,746	83,518
Capital Cost			100 MW	500 MW	1000 MW
Cost Basis (Year)	→		***2007***	***2007***	***2007***
ESP	$		$2,529,108	$10,114,289	$18,834,775
Ash Handling System	$		$929,442	$3,466,560	$6,637,958
ID Fan(s)	$		$395,948	$1,030,158	$1,581,903
Equipment Cost Subtotal	$		$3,854,498	$14,611,007	$27,054,637
Instruments & Controls	$		$77,090	$292,220	$541,093
Taxes	$		$231,270	$876,660	$1,623,278
Freight	$		$192,725	$730,550	$1,352,732
Purchased Equipment Cost Subtotal	$		$4,355,583	$16,510,438	$30,571,740
Installation	$		$3,528,022	$13,373,455	$24,763,109
Total Direct Cost	$		$7,883,606	$29,883,893	$55,334,849
Total Direct Cost with Retrofit Factor	$		$7,883,606	$29,883,893	$55,334,849
General Facilities	$		$788,361	$2,988,389	$5,533,485
Engineering Fees	$		$788,361	$2,988,389	$5,533,485
Contingency	$		$1,576,721	$5,976,779	$11,066,970
Total Plant Cost (TPC)	$		$11,037,048	$41,837,450	$77,468,788
Total Plant Cost (TPC) w/ Prime Contractor's M	$		$11,368,159	$43,092,574	$79,792,852
Total Cash Expended (TCE)	$		$11,202,604	$42,465,012	$78,630,820
Allow. for Funds During Constr. (AFDC)	$		$596,001	$2,259,222	$4,183,315
Total Plant Investment (TPI)	$		$11,798,604	$44,724,234	$82,814,134
Preproduction Costs	$		$235,972	$894,485	$1,656,283
Inventory Capital	$		$0	$0	$0
Total Capital Requirement (TCR)	$		**$12,034,576**	**$45,618,719**	**$84,470,417**
	$/kW		$120.3	$91.2	$84.5
O&M Data and Costs			100 MW	500 MW	1000 MW
Cost Basis (Year)	→		***2007***	***2007***	***2007***
Power Required Excluding ID Fan(s)	kW		81	317	633
ID Fan Power for FF Delta P	kW		184	860	1,720
Total Power	kW		265	1,177	2,353
Power Cost **	$/yr		$52,148	$302,698	$605,092
Maintenance Costs	$/yr		$551,852	$2,091,873	$3,873,439

** *These costs assume inputs are in current dollars (no escalation included).*

Table B-3a. Turbine Cost Inputs

Inputs	Cost		Units	Reference	Capital Recovery @ 7%
Basic SCR Equipment	High temperature catalyst	1.5 M - 6.7 M	US dollars	U.S. DOE	
	Conventional catalyst	1.2 M - 5.6 M	US dollars	U.S. DOE	
Operating Period	Peaking	4,200	hrs/yr	Engineering Judgment	
	Base	8,400	hrs/yr	Engineering Judgment	
	Duct Burner	8,400	hrs/yr	Engineering Judgment	
Operator and Maintenance Labor	-	40	US $/hr	Canadian Labour Relations	
Electricity Cost	-	0.06	US $/kw-h	Engineering Judgment	
Catalyst Replace Cost	-	400	US $/ft3	Engineering Judgment	
Catalyst Life	Peaking	5	years	Engineering Judgment	0.24389
	Base	3	years	Engineering Judgment	0.38105
Catalyst Disposal	-	15	US $/ft3	U.S. DOE	
Ammonia	-	397	US $/tonne	U.S. DOE	
Capital Recovery	-	7	interest rate	U.S. EPA	
	-	20	years of equipment life	Engineering Judgment	0.09439
NOx Concentration, Exhaust from DLN	-	25	ppmv	Engineering Judgment	
Control Effectiveness	SCR	92.0	percent	Engineering Judgment	
NOx Concentration, Exhaust from SCR	-	2.0	ppmv	Engineering Judgment	

Table B-3b. Cost Inflation Factors

Chemical Engineering Plant Cost Index (CEPCI)	Cost Basis
390.6	1999
525.4	2007

Table B-3c. SCR Costs vs. Combustion Turbine Output

Conventional SCR		High Temperature SCR	
Turbine Output Capacity (MW)	Basic Equipment ($)	Turbine Output Capacity (MW)	Basic Equipment ($)
5	240,000	5	380,000
23	660,000	23	730,000
170	2,100,000	170	3,000,000
slope	10,703	slope	15,710
intersept	293,599	intersept	333,116
correlation	0.99312	correlation	0.99972

Class	Size Range	Avg. Size	NOx Control Applied	Basic Equipment ($)
Small 1	25-75	50	High Temp SCR	1,504,686
Small 2	75-150	112.5	High Temp SCR	2,825,447
Medium	150-200	175	High Temp SCR	4,146,207
Large 1	greater than 200	300	High Temp SCR	6,787,728
Small 1	25-75	50	Conventional SCR	1,114,762
Small 2	75-150	112.5	Conventional SCR	2,014,560
Medium	150-200	175	Conventional SCR	2,914,358
Large 1	greater than 200	300	Conventional SCR	4,713,953
Small 1	25-75	50	Conventional SCR w/ DB*	1,258,729
Small 2	75-150	113	Conventional SCR w/ DB*	2,338,487
Medium	150-200	175	Conventional SCR w/ DB*	3,418,244
Large 1	greater than 200	300	Conventional SCR w/ DB*	5,577,759

e duct burner packages are 20 percent of the combustion turbine input capacity.

Table B-3d. SCR Catalyst Volume vs. Percent Reduction

80% NOx reduction	Catalyst Volume		94% NOx Reduction	Catalyst Volume	
Turbine Output Capacity (MW)	(cubic feet)	MW to Catalyst Volume Ratio	Turbine Output Capacity (MW)	(cubic feet)	MW to Catalyst Volume Ratio
42	730	17	42	1,320	31
27	600	22	27	1,040	39
27	980	36	27	1,746	64
30	750	25	30	1,134	38

MW to Catalyst Volume Ratio	25.1	42.9
Target Reduction	80	94

slope	1.27
intersept	-76.6

MW to Catalyst Volume Ratio	40.4
Target Reduction	92
MW to Catalyst Volume Ratio for Combined Cycle with Duct Burner	48.5
Target Reduction	92

Table B-3e. Detailed SCR for Simple Cycle Peaking Turbine with High Temperature Catalyst.

Turbine Model			Small 1	Small 2
Turbine Gross Output (MWh)			50	113
Total Power Gross Output (WMh)	(Combustion turbine generator)		50	113
Efficiency (%)			37.5%	41.0%
Turbine Input (MWh)	(Combustion turbine generator) / (Efficiency)		133	275
Direct Capital Costs (DC):				
Purchased Equip. Cost (PE):		Source		
Basic Equipment (A):		EPA	1,504,686	2,825,447
Ammonia injection skid and storage	0.00 x A	EPA	included	included
Instrumentation	0.00 x A	EPA	included	included
Taxes and freight:	0.08 x A	OAQPS	120,375	226,036
PE Total:		OAQPS	1,625,061	3,051,482
Direct Installation Costs (DI):				
Foundation & supports:	0.08 x PE	OAQPS	130,005	244,119
Handling and erection:	0.14 x PE	OAQPS	227,509	427,208
Electrical:	0.04 x PE	OAQPS	65,002	122,059
Piping:	0.02 x PE	OAQPS	32,501	61,030
Insulation:	0.01 x PE	OAQPS	16,251	30,515
Painting:	0.01 x PE	OAQPS	16,251	30,515
DI Total:			487,518	915,445
Total Direct Cost (PE + DI):			2,112,579	3,966,927
Indirect Costs (IC):				
Engineering:	0.10 x PE	OAQPS	162,506	305,148
Construction and field expenses:	0.05 x PE	OAQPS	81,253	152,574
Contractor fees:	0.10 x PE	OAQPS	162,506	305,148
Start-up:	0.02 x PE	OAQPS	32,501	61,030
Performance testing:	0.01 x PE	OAQPS	16,251	30,515
Contingencies:	0.03 x PE	OAQPS	48,752	91,544
IC Total:			503,769	945,960
Total Capital Investment (TCI = DC + IC):			2,616,348	4,912,887
Direct Annual Costs (DAC):				
Operating Costs (O):	24 hrs/day, 7 days/week, 25 weeks/yr			
Operator:	0.5 hr/shift; 40 $/hr for operator pay	OAQPS	10,500	10,500
Supervisor:	15% of operator	OAQPS	1,575	1,575
Maintenance Costs (M):				
Labor:	0.5 hr/shift; 40 $/hr for labor pay	OAQPS	10,500	10,500
Material:	100% of labor cost:	OAQPS	10,500	10,500
Utility Costs:				
Perf. loss:	SCR increases back-pressure, decreasing output 0.5%			
Electricity cost	0.06 ($/kwh) performance loss cost penalty	variable	63,000	141,750
Ammonia inject skid:	blower (5 kw) + NH3/H2O pump (5 or 10 kw)	-	2,520	3,780
Equipment:				
Catalyst replacement:	catalyst per MW, $400/ft3, 5 yr. life	EPRI	197,042	443,344
Catalyst disposal:	$15/ft3*30 ft3/MW*MW*0.24389 (5 yr amortized)	OAQPS	5,488	12,347
Ammonia:	($/tonne) [tons NH3 = tons NOx * (17/46)]	variable	12,403	27,907
Total DAC:			313,528	662,203
Indirect Annual Costs (IAC):				
Overhead:	60% of O&M	OAQPS	19,845	19,845
Administrative:	0.02 x TCI	OAQPS	52,327	98,258
Insurance:	0.01 x TCI	OAQPS	26,163	49,129
Property tax:	0.01 x TCI	OAQPS	26,163	49,129
Capital recovery:	7% interest rate; 20 yrs - period			
	0.09439 x TCI	OAQPS	246,957	463,727
Total IAC:			371,456	680,088
Total Annual Cost (DAC + IAC):			684,984	1,342,291
Size of Duct Burner (MMBtu/hr)			-	-
NOx Attributed to Duct Burner (lb/MMBtu heat input)			-	-
NOx Attributed to Duct Burner (tonne/yr)			-	-
NOx Attributed to Duct Burner (kg/yr)			-	-
NOx Concentration, Exhaust from DLN (ppm)			25	25
NOx Concentration, Exhaust from DLN (lb/MWh)			0.97	0.97
NOx Concentration, Exhaust from DLN (tonne/yr)			91.9	207
NOx Concentration, Exhaust from DLN (kg/yr)			91,928	206,837
NOx Concentration, Exhaust after Duct Burner (tonne/yr)			-	-
NOx Concentration, Exhaust after Duct Burner (kg/yr)			-	-
NOx Removed by SCR (tonne/yr)			84.6	190
NOx Removed by SCR (kg/yr)			84,573	190,290
Cost Effectiveness ($/tonne)			8,099	7,054
Cost Effectiveness ($/kg)			8.10	7.05
Cost Impact ($/kW)			13.70	11.93
NOx Emissions (tonne/yr)			7.35	16.5
NOx Emissions (kg/yr)			7,354	16,547
NOx Emissions (lbs/hr)			10.3	21.2
NOx Emissions (kg/hr)			4.67	9.62
Total Power Gross Output (WMh)			50.0	113
NOx Concentration, Exhaust from DLN (kg/MWh)			0.44	0.44
NOx Emissions (kg/MWh)			0.04	0.04

Table B-3f. Detailed SCR for Simple Cycle Base Load Turbine with High Temperature Catalyst.

Turbine Model			Small 1	Small 2	Medium	Large 1
Turbine Gross Output (MWh)			50	113	175	300
Total Power Gross Output (WMh)	(Combustion turbine generator)		50	113	175	300
Efficiency (%)			37.5%	41.0%	37.0%	38.0%
Turbine Input (MWh)	(Combustion turbine generator) / (Efficiency)		133	275	473	789
Direct Capital Costs (DC):						
Purchased Equip. Cost (PE):		Source				
Basic Equipment (A):		EPA	1,504,686	2,825,447	4,146,207	6,787,728
Ammonia injection skid and storage	0.00 x A	EPA	included	included	included	included
Instrumentation	0.00 x A	EPA	included	included	included	included
Taxes and freight:	0.08 x A	OAQPS	120,375	226,036	331,697	543,018
PE Total:		OAQPS	1,625,061	3,051,482	4,477,904	7,330,746
Direct Installation Costs (DI):						
Foundation & supports:	0.08 x PE	OAQPS	130,005	244,119	358,232	586,460
Handling and erection:	0.14 x PE	OAQPS	227,509	427,208	626,907	1,026,304
Electrical:	0.04 x PE	OAQPS	65,002	122,059	179,116	293,230
Piping:	0.02 x PE	OAQPS	32,501	61,030	89,558	146,615
Insulation:	0.01 x PE	OAQPS	16,251	30,515	44,779	73,307
Painting:	0.01 x PE	OAQPS	16,251	30,515	44,779	73,307
DI Total:			487,518	915,445	1,343,371	2,199,224
Total Direct Cost (PE + DI):			2,112,579	3,966,927	5,821,275	9,529,970
Indirect Costs (IC):						
Engineering:	0.10 x PE	OAQPS	162,506	305,148	447,790	733,075
Construction and field expenses:	0.05 x PE	OAQPS	81,253	152,574	223,895	366,537
Contractor fees:	0.10 x PE	OAQPS	162,506	305,148	447,790	733,075
Start-up:	0.02 x PE	OAQPS	32,501	61,030	89,558	146,615
Performance testing:	0.01 x PE	OAQPS	16,251	30,515	44,779	73,307
Contingencies:	0.03 x PE	OAQPS	48,752	91,544	134,337	219,922
IC Total:			503,769	945,960	1,388,150	2,272,531
Total Capital Investment (TCI = DC + IC):			2,616,348	4,912,887	7,209,425	11,802,501
Direct Annual Costs (DAC):						
Operating Costs (O):	24 hrs/day, 7 days/week, 50 weeks/yr					
Operator:	0.5 hr/shift; 40 $/hr for operator pay	OAQPS	21,000	21,000	21,000	21,000
Supervisor:	15% of operator	OAQPS	3,150	3,150	3,150	3,150
Maintenance Costs (M):						
Labor:	0.5 hr/shift; 40 $/hr for labor pay	OAQPS	21,000	21,000	21,000	21,000
Material:	100% of labor cost:	OAQPS	21,000	21,000	21,000	21,000
Utility Costs:						
Perf. loss:	SCR increases back-pressure, decreasing output 0.5%					
Electricity cost	0.06 ($/kwh) performance loss cost penalty	variable	126,000	283,500	441,000	756,000
Ammonia inject skid:	blower (5 kw) + NH3/H2O pump (5, 10, 15 or 20 kw)	-	5,040	7,560	10,080	12,600
Equipment:						
Catalyst replacement:	catalyst per MW, $400/ft3, 3 yr. life	EPRI	307,855	692,674	1,077,494	1,847,132
Catalyst disposal:	$15/ft3*30 ft3/MW*MW*0.38105 (3 yr amortized)	OAQPS	8,574	19,291	30,008	51,442
Ammonia:	($/tonne) [tons NH3 = tons NOx * (17/46)]	variable	24,806	55,814	86,822	148,838
Total DAC:			538,425	1,124,989	1,711,553	2,882,161
Indirect Annual Costs (IAC):						
Overhead:	60% of O&M	OAQPS	39,690	39,690	39,690	39,690
Administrative:	0.02 x TCI	OAQPS	52,327	98,258	144,188	236,050
Insurance:	0.01 x TCI	OAQPS	26,163	49,129	72,094	118,025
Property tax:	0.01 x TCI	OAQPS	26,163	49,129	72,094	118,025
Capital recovery:	7% interest rate; 20 yrs - period					
	0.09439 x TCI	OAQPS	246,957	463,727	680,498	1,114,038
Total IAC:			391,301	699,933	1,008,565	1,625,828
Total Annual Cost (DAC + IAC):			929,726	1,824,922	2,720,118	4,507,989
Size of Duct Burner (MMBtu/hr)			-	-	-	-
NOx Attributed to Duct Burner (lb/MMBtu heat input)			-	-	-	-
NOx Attributed to Duct Burner (tonne/yr)			-	-	-	-
NOx Attributed to Duct Burner (kg/yr)			-	-	-	-
NOx Concentration, Exhaust from DLN (ppm)			25	25	25	25
NOx Concentration, Exhaust from DLN (lb/MWh)			0.97	0.97	0.97	0.97
NOx Concentration, Exhaust from DLN (tonne/yr)			184	414	643	1,103
NOx Concentration, Exhaust from DLN (kg/yr)			183,855	413,674	643,493	1,103,131
NOx Concentration, Exhaust after Duct Burner (tonne/yr)			-	-	-	-
NOx Concentration, Exhaust after Duct Burner (kg/yr)			-	-	-	-
NOx Removed by SCR (tonne/yr)			169	381	592	1,015
NOx Removed by SCR (kg/yr)			169,147	380,580	592,014	1,014,881
Cost Effectiveness ($/tonne)			5,497	4,795	4,595	4,442
Cost Effectiveness ($/kg)			5.50	4.80	4.59	4.44
Cost Impact ($/kW)			18.59	16.22	15.54	15.03
NOx Emissions (tonne/yr)			14.7	33.1	51.5	88.3
NOx Emissions (kg/yr)			14,708	33,094	51,479	88,251
NOx Emissions (lbs/hr)			10.3	21.2	36.5	61.0
NOx Emissions (kg/hr)			4.67	9.62	16.56	27.65
Total Power Gross Output (WMh)			50	113	175	300
NOx Concentration, Exhaust from DLN (kg/MWh)			0.44	0.44	0.44	0.44
NOx Emissions (kg/MWh)			0.04	0.04	0.04	0.04

Table B-3g. Detailed SCR for Combined Cycle Base Load Turbine with Conventional Catalyst.

Turbine Model			Small 1	Small 2	Medium	Large 1
Turbine Gross Output (MWh)			50	113	175	300
Total Power Gross Output (WMh)	(Combustion turbine generator) + (Steam turbine generator)		77	173	269	462
Efficiency (%)			37.5%	41.0%	37.0%	38.0%
Turbine Input (MWh)	(Combustion turbine generator) / (Efficiency)		133	275	473	789
Direct Capital Costs (DC):						
Purchased Equip. Cost (PE):		Source				
Basic Equipment (A):		EPA	1,114,762	2,014,560	2,914,358	4,713,953
Ammonia injection skid and storage	0.00 x A	EPA	included	included	included	included
Instrumentation	0.00 x A	EPA	included	included	included	included
Taxes and freight:	0.08 x A	OAQPS	89,181	161,165	233,149	377,116
PE Total:		OAQPS	1,203,943	2,175,724	3,147,506	5,091,070
Direct Installation Costs (DI):						
Foundation & supports:	0.08 x PE	OAQPS	96,315	174,058	251,800	407,286
Handling and erection:	0.14 x PE	OAQPS	168,552	304,601	440,651	712,750
Electrical:	0.04 x PE	OAQPS	48,158	87,029	125,900	203,643
Piping:	0.02 x PE	OAQPS	24,079	43,514	62,950	101,821
Insulation:	0.01 x PE	OAQPS	12,039	21,757	31,475	50,911
Painting:	0.01 x PE	OAQPS	12,039	21,757	31,475	50,911
DI Total:			361,183	652,717	944,252	1,527,321
Total Direct Cost (PE + DI):			1,565,126	2,828,442	4,091,758	6,618,390
Indirect Costs (IC):						
Engineering:	0.10 x PE	OAQPS	120,394	217,572	314,751	509,107
Construction and field expenses:	0.05 x PE	OAQPS	60,197	108,786	157,375	254,553
Contractor fees:	0.10 x PE	OAQPS	120,394	217,572	314,751	509,107
Start-up:	0.02 x PE	OAQPS	24,079	43,514	62,950	101,821
Performance testing:	0.01 x PE	OAQPS	12,039	21,757	31,475	50,911
Contingencies:	0.03 x PE	OAQPS	36,118	65,272	94,425	152,732
IC Total:			373,222	674,475	975,727	1,578,232
Total Capital Investment (TCI = DC + IC):			1,938,348	3,502,916	5,067,485	8,196,622
Direct Annual Costs (DAC):						
Operating Costs (O):	24 hrs/day, 7 days/week, 50 weeks/yr					
Operator:	0.5 hr/shift; 40 $/hr for operator pay	OAQPS	21,000	21,000	21,000	21,000
Supervisor:	15% of operator	OAQPS	3,150	3,150	3,150	3,150
Maintenance Costs (M):						
Labor:	0.5 hr/shift; 40 $/hr for labor pay	OAQPS	21,000	21,000	21,000	21,000
Material:	100% of labor cost:	OAQPS	21,000	21,000	21,000	21,000
Utility Costs:						
Perf. loss:	SCR increases back-pressure, decreasing output 0.5%					
Electricity cost	0.06 ($/kwh) performance loss cost penalty	variable	126,000	283,500	441,000	756,000
Ammonia inject skid:	blower (5 kw) + NH3/H2O pump (5, 10, 15 or 20 kw)	EPA	5,040	7,560	10,080	12,600
Equipment:						
Catalyst replacement:	catalyst per MW, $400/ft3, 3 yr. life	EPRI	307,855	692,674	1,077,494	1,847,132
Catalyst disposal:	$15/ft3*30 ft3/MW*MW*0.38105 (3 yr amortized)	OAQPS	8,574	19,291	30,008	51,442
Ammonia:	($/tonne) [tons NH3 = tons NOx * (17/46)]	variable	24,806	55,814	86,822	148,838
Total DAC:			538,425	1,124,989	1,711,553	2,882,161
Indirect Annual Costs (IAC):						
Overhead:	60% of O&M	OAQPS	39,690	39,690	39,690	39,690
Administrative:	0.02 x TCI	OAQPS	38,767	70,058	101,350	163,932
Insurance:	0.01 x TCI	OAQPS	19,383	35,029	50,675	81,966
Property tax:	0.01 x TCI	OAQPS	19,383	35,029	50,675	81,966
Capital recovery:	7% interest rate; 20 yrs - period					
	0.09439 x TCI	OAQPS	182,961	330,640	478,320	773,679
Total IAC:			300,185	510,447	720,709	1,141,234
Total Annual Cost (DAC + IAC):			838,610	1,635,436	2,432,262	4,023,395
Size of Duct Burner (MMBtu/hr)			-	-	-	-
NOx Attributed to Duct Burner (lb/MMBtu heat input)			-	-	-	-
NOx Attributed to Duct Burner (tonne/yr)			-	-	-	-
NOx Attributed to Duct Burner (kg/yr)			-	-	-	-
NOx Concentration, Exhaust from DLN (ppm)			25	25	25	25
NOx Concentration, Exhaust from DLN (lb/MWh)			0.97	0.97	0.97	0.97
NOx Concentration, Exhaust from DLN (tonne/yr)			184	414	643	1,103
NOx Concentration, Exhaust from DLN (kg/yr)			183,855	413,674	643,493	1,103,131
NOx Concentration, Exhaust after Duct Burner (tonne/yr)			-	-	-	-
NOx Concentration, Exhaust after Duct Burner (kg/yr)			-	-	-	-
NOx Removed by SCR (tonne/yr)			169	381	592	1,015
NOx Removed by SCR (kg/yr)			169,147	380,580	592,014	1,014,881
Cost Effectiveness ($/tonne)			4,958	4,297	4,108	3,964
Cost Effectiveness ($/kg)			4.96	4.30	4.11	3.96
Cost Impact ($/kW)			10.90	9.45	9.03	8.72
NOx Emissions (tonne/yr)			14.7	33.1	51.5	88.3
NOx Emissions (kg/yr)			14,708	33,094	51,479	88,251
NOx Emissions (lbs/hr)			10.3	21.2	36.5	61.0
NOx Emissions (kg/hr)			4.67	9.62	16.56	27.65
Total Power Gross Output (WMh)			77	173	269	462
NOx Concentration, Exhaust from DLN (kg/MWh)			0.28	0.28	0.28	0.28
NOx Emissions (kg/MWh)			0.023	0.023	0.023	0.023

Table B-3h. Detailed SCR for Combined Cycle Base Load Turbine with Conventional Catalyst and Duct Burners.

Turbine Model			Small 1	Small 2	Medium	Large 1
Turbine Gross Output (MWh)			50	113	175	300
Total Power Gross Output (WMh)	(Combustion turbine generator) + (Steam turbine generator) + (Duct burne		92	208	323	554
Efficiency (%)			37.5%	41.0%	37.0%	38.0%
Turbine Input (MWh)	(Combustion turbine generator) / (Efficiency)		133	275	473	789
Direct Capital Costs (DC):						
Purchased Equip. Cost (PE):		Source				
Basic Equipment (A):		EPA	1,258,729	2,338,487	3,418,244	5,577,759
Ammonia injection skid and storage	0.00 x A	EPA	included	included	included	included
Instrumentation	0.00 x A	EPA	included	included	included	included
Taxes and freight:	0.08 x A	OAQPS	100,698	187,079	273,460	446,221
PE Total:		OAQPS	1,359,428	2,525,566	3,691,704	6,023,980
Direct Installation Costs (DI):						
Foundation & supports:	0.08 x PE	OAQPS	108,754	202,045	295,336	481,918
Handling and erection:	0.14 x PE	OAQPS	190,320	353,579	516,839	843,357
Electrical:	0.04 x PE	OAQPS	54,377	101,023	147,668	240,959
Piping:	0.02 x PE	OAQPS	27,189	50,511	73,834	120,480
Insulation:	0.01 x PE	OAQPS	13,594	25,256	36,917	60,240
Painting:	0.01 x PE	OAQPS	13,594	25,256	36,917	60,240
DI Total:			407,828	757,670	1,107,511	1,807,194
Total Direct Cost (PE + DI):			1,767,256	3,283,236	4,799,215	7,831,174
Indirect Costs (IC):						
Engineering:	0.10 x PE	OAQPS	135,943	252,557	369,170	602,398
Construction and field expenses:	0.05 x PE	OAQPS	67,971	126,278	184,585	301,199
Contractor fees:	0.10 x PE	OAQPS	135,943	252,557	369,170	602,398
Start-up:	0.02 x PE	OAQPS	27,189	50,511	73,834	120,480
Performance testing:	0.01 x PE	OAQPS	13,594	25,256	36,917	60,240
Contingencies:	0.03 x PE	OAQPS	40,783	75,767	110,751	180,719
IC Total:			421,423	782,925	1,144,428	1,867,434
Total Capital Investment (TCI = DC + IC):			2,188,679	4,066,161	5,943,643	9,698,608
Direct Annual Costs (DAC):						
Operating Costs (O):	24 hrs/day, 7 days/week, 50 weeks/yr					
Operator:	0.5 hr/shift; 40 $/hr for operator pay	OAQPS	21,000	21,000	21,000	21,000
Supervisor:	15% of operator	OAQPS	3,150	3,150	3,150	3,150
Maintenance Costs (M):						
Labor:	0.5 hr/shift; 40 $/hr for labor pay	OAQPS	21,000	21,000	21,000	21,000
Material:	100% of labor cost:	OAQPS	21,000	21,000	21,000	21,000
Utility Costs:						
Perf. loss:	SCR increases back-pressure, decreasing output 0.5%					
Electricity cost	0.06 ($/kwh) performance loss cost penalty	variable	126,000	283,500	441,000	756,000
Ammonia inject skid:	blower (5 kw) + NH3/H2O pump (5, 10, 15 or 20 kw)	EPA	5,040	7,560	10,080	12,600
Equipment:						
Catalyst replacement:	catalyst per MW; 3 yr. life	EPRI	369,426	831,209	1,292,992	2,216,558
Catalyst disposal:	$15/ft3*30 ft3/MW*MW*0.38105 (3 yr amortized)	OAQPS	8,574	19,291	30,008	51,442
Ammonia:	($/tonne) [tonne NH3 = tonne NOx * (17/46)]	variable	26,561	59,762	92,963	159,365
Total DAC:			601,751	1,267,472	1,933,193	3,262,115
Indirect Annual Costs (IAC):						
Overhead:	60% of O&M	OAQPS	39,690	39,690	39,690	39,690
Administrative:	0.02 x TCI	OAQPS	43,774	81,323	118,873	193,972
Insurance:	0.01 x TCI	OAQPS	21,887	40,662	59,436	96,986
Property tax:	0.01 x TCI	OAQPS	21,887	40,662	59,436	96,986
Capital recovery:	7% interest rate; 20 yrs - period					
	0.09439 x TCI	OAQPS	206,589	383,805	561,020	915,452
Total IAC:			333,827	586,141	838,456	1,343,086
Total Annual Cost (DAC + IAC):			935,577	1,853,613	2,771,649	4,605,201
Size of Duct Burner (MMBtu/hr)	(Combustion turbine generator) x 1.20		34.1	77	119	205
NOx Attributed to Duct Burner (lb/MMBtu heat input)			0.10	0.10	0.10	0.10
NOx Attributed to Duct Burner (tonne/yr)			13.0	29.3	45.5	78.0
NOx Attributed to Duct Burner (kg/yr)			13,004	29,260	45,515	78,026
NOx Concentration, Exhaust from DLN (ppm)			25	25	25	25
NOx Concentration, Exhaust from DLN (lb/MWh)			0.97	0.97	0.97	0.97
NOx Concentration, Exhaust from DLN (tonne/yr)			184	414	643	1,103
NOx Concentration, Exhaust from DLN (kg/yr)			183,855	413,674	643,493	1,103,131
NOx Concentration, Exhaust after Duct Burner (tonne/yr)			197	443	689	1,181
NOx Concentration, Exhaust after Duct Burner (kg/yr)			196,860	442,934	689,008	1,181,157
NOx Removed by SCR (tonne/yr)			181	407	634	1,087
NOx Removed by SCR (kg/yr)			181,111	407,499	633,888	1,086,665
Cost Effectiveness ($/tonne)			5,166	4,549	4,372	4,238
Cost Effectiveness ($/kg)			5.17	4.55	4.37	4.24
Cost Impact ($/kW)			10.14	8.92	8.58	8.31
NOx Emissions (tonne/yr)			15.7	35.4	55.1	94.5
NOx Emissions (kg/yr)			15,749	35,435	55,121	94,493
NOx Emissions (lbs/hr)			10.6	21.8	37.5	62.6
NOx Emissions (kg/hr)			4.79	9.9	17.0	28.4
Total Power Gross Output (WMh)			92.3	208	323	554
NOx Concentration, Exhaust from DLN (kg/MWh)			0.24	0.24	0.24	0.24
NOx Emissions (kg/MWh)			0.021	0.021	0.021	0.021

Table B-3i. Turbine Cost Summary.

Model	Class	Size Range	Unit Size (MWh)	Power Gross Output (WMh)	Operating Cycle	NOx Control Applied	Load	Life of Equipment (hrs)	Total Capital Investment ($)	Total Annual Cost ($)	Direct Annual Costs ($)	NOx from DLN (tonne/yr)	NOx from DB (tonne/yr)	SCR NOx Removed (tonne/yr)	Actual NOx Emissions (tonne/yr)	NOx Emissions (kg/hr)	Cost Effectiveness ($/tonne)	Cost Impact ($/kW)
1	Small 1	25-75	50	50	SC	High Temp SCR	Peaking	84,000	2,616,348	684,984	313,528	91.9	-	84.6	7.35	4.67	8,099	13.7
2	Small 2	75-150	113	113	SC	High Temp SCR	Peaking	84,000	4,912,887	1,342,291	662,203	207	-	190	16.5	9.62	7,054	11.9
3	Small 1	25-75	50	50	SC	High Temp SCR	Base	168,000	2,616,348	929,726	538,425	184	-	169	14.7	4.67	5,497	18.6
4	Small 2	75-150	113	113	SC	High Temp SCR	Base	168,000	4,912,887	1,824,922	1,124,989	414	-	381	33.1	9.62	4,795	16.2
5	Medium	150-200	175	175	SC	High Temp SCR	Base	168,000	7,209,425	2,720,118	1,711,553	643	-	592	51.5	16.6	4,595	15.5
6	Large 1	greater than 200	300	300	SC	High Temp SCR	Base	168,000	11,802,501	4,507,989	2,882,161	1,103	-	1,015	88.3	27.6	4,442	15.0
7	Small 1	25-75	50	77	CC	Conventional SCR	Base	168,000	1,938,348	838,610	538,425	184	-	169	14.7	4.67	4,958	10.9
8	Small 2	75-150	113	173	CC	Conventional SCR	Base	168,000	3,502,916	1,635,436	1,124,989	414	-	381	33.1	9.62	4,297	9.4
9	Medium	150-200	175	269	CC	Conventional SCR	Base	168,000	5,067,485	2,432,262	1,711,553	643	-	592	51.5	16.6	4,108	9.0
10	Large 1	greater than 200	300	462	CC	Conventional SCR	Base	168,000	8,196,622	4,023,395	2,882,161	1,103	-	1,015	88.3	27.6	3,964	8.7
11	Small 1	25-75	50	92	CC w/ DB	Conventional SCR	Base	168,000	2,188,679	935,577	601,751	184	13.0	181	15.7	4.79	5,166	10.1
12	Small 2	75-150	113	208	CC w/ DB	Conventional SCR	Base	168,000	4,066,161	1,853,613	1,267,472	414	29.3	407	35.4	9.9	4,549	8.9
13	Medium	150-200	175	323	CC w/ DB	Conventional SCR	Base	168,000	5,943,643	2,771,649	1,933,193	643	45.5	634	55.1	17.0	4,372	8.6
14	Large 1	greater than 200	300	554	CC w/ DB	Conventional SCR	Base	168,000	9,698,608	4,605,201	3,262,115	1,103	78.0	1,087	94	28.4	4,238	8.3

Table 3-2. SCR Control Costs

Model	Total Power Gross	Total Capital Investment			Total Annual Cost		
	Output (WM-hr)	(Million $US)	Cost (Million $/MW-hr lifetime)	Cost (Million $/MW-hr capacity)	(Million $US/yr)	Cost (Million $/MW-hr annual)	Cost (Million $/MW-hr capacity)
1	50	2.62	0.62	0.052	0.68	3.26	0.014
2	113	4.91	0.52	0.044	1.34	2.84	0.012
3	50	2.62	0.31	0.052	0.93	2.21	0.019
4	113	4.91	0.26	0.044	1.82	1.93	0.016
5	175	7.21	0.25	0.041	2.72	1.85	0.016
6	300	11.8	0.23	0.039	4.51	1.79	0.015
7	77	1.94	0.15	0.025	0.84	1.30	0.011
8	173	3.50	0.12	0.020	1.64	1.12	0.009
9	269	5.07	0.11	0.019	2.43	1.08	0.009
10	462	8.20	0.11	0.018	4.02	1.04	0.009
11	92	2.19	0.14	0.024	0.94	1.21	0.010
12	208	4.07	0.12	0.020	1.85	1.06	0.009
13	323	5.94	0.11	0.018	2.77	1.02	0.009
14	554	9.70	0.10	0.018	4.61	0.99	0.008

NOx from DLN (kg/MWh)	NOx Emissions (kg/MWh)
0.44	0.035
0.44	0.035
0.44	0.035
0.44	0.035
0.44	0.035
0.44	0.035
0.28	0.023
0.28	0.023
0.28	0.023
0.28	0.023
0.24	0.021
0.24	0.021
0.24	0.021
0.24	0.021

Table B-3j. SCR Additional Energy Usage and Resulting Greenhouse Gas Emissions.

Model	Class	Size Range	Unit Size (MWh)	Load	Efficiency (%)	Required Electricity (a) (MWh-yr)	Required Additional Fuel (MMBtu-yr)	CO2 Emissions (tonne/yr)	CH4 Emissions (tonne/yr)	N2O Emissions (tonne/yr)	CO2e (b) Emissions (tonne/yr)
1	Small 1	25-75	50	Peaking	0.38	1,092	9,930	2,408	0.19	0.07	2,432
2	Small 2	75-150	113	Peaking	0.41	2,426	20,207	4,900	0.38	0.13	4,950
3	Small 1	25-75	50	Base	0.38	2,184	19,860	4,816	0.38	0.13	4,865
4	Small 2	75-150	113	Base	0.41	4,851	40,414	9,801	0.77	0.27	9,900
5	Medium	150-200	175	Base	0.37	7,518	69,348	16,818	1.31	0.46	16,987
6	Large 1	greater than 200	300	Base	0.38	12,810	115,054	27,902	2.18	0.76	28,183

(a) - SCR electricity consumption (i.e., ammonia inject skid pumps and blowers) and performance loss.
(b) - 100-year Global Warming Potentials (GWP) from U.S. EPA - 2008 Inventory of Greenhouse Gas Emissions and Sinks

Table B-3k. DLN Turbine Greenhouse Gas Emission Factors

	Emission Factors (c)	
CO2	110	lb/MMBtu - Fuel Input
N2O	0.003	lb/MMBtu - Fuel Input
CH4	8.60E-03	lb/MMBtu - Fuel Input

(c) AP-42 emission factors from Chapter 3.1 - Stationary Gas Turbines (April 2000)

Table B-3l. Greenhouse Gas Global Warming Potentials

	100-year Global Warming Potentials
CO2	1
N2O	21
CH4	310

(d) Global Warming Potentials (GWP) are calculated over 100 year time horizon from IPCC Climate Change 1995.

SCR costs were developed for 14-models, each representing different turbine sizes, operating cycles (simple and combined cycle), supplemental heat (duct burners), and load (peaking and base). Operating cycle and supplemental heat had no direct impact on energy usages; therefore only six unique energy usages rates were calculated. There are 4 size categories of turbines and 2-peaking units. Peaking units are designed to generate energy on short notice and for relatively short periods of time. ERG has assumed 50% utilization for peaking units or 4,200 hours per year. Ammonia pumps and motors operate only when the peaking unit is generating power.

Turbine BACTEA Cost Detail; APPENDIX B-3

1) PURPOSE:

Estimate capital and annual costs for installations of technologically feasible NO_x control technologies on model turbine units.

Model units have been designed to provide ranges that were small enough that any member of a group would be well represented by the parameters assigned for that model to the group. ERG has developed a matrix of costs parameters: turbine size, operating cycle (simple and combined cycle), supplemental heat (duct burners), and load (peaking and base). Each operating scenario, size combination that was considered is shown in Table B.3-1.

Table B.3-1. Model Units for New Gas Turbines

Class	Size Range (MW)	Representative Size (MW)	Operating Cycle	Load
Small 1	25-75	50	Simple Cycle	Peaking
Small 2	75-150	113	Simple Cycle	Peaking
Small 1	25-75	50	Simple Cycle	Base
Small 2	75-150	113	Simple Cycle	Base
Medium	150-200	175	Simple Cycle	Base
Large 1	greater than 200	300	Simple Cycle	Base
Small 1	25-75	50	Combined Cycle	Base
Small 2	75-150	113	Combined Cycle	Base
Medium	150-200	175	Combined Cycle	Base
Large 1	greater than 200	300	Combined Cycle	Base
Small 1	25-75	50	Combined Cycle w/ Duct Burner	Base
Small 2	75-150	113	Combined Cycle w/ Duct Burner	Base
Medium	150-200	175	Combined Cycle w/ Duct Burner	Base
Large 1	greater than 200	300	Combined Cycle w/ Duct Burner	Base

2) INPUT

Costs for basic Selective Catalytic Reduction (SCR) equipment were obtained from the report "Cost Analysis of NO_x Control Alternatives for Stationary Gas Turbines," prepared for the U.S. Department of Energy (DOE), by ONSITE SYCOM Energy Corporation.[1] The report presented capital and annual costs of high temperature SCR for three different sizes of turbines, 5, 25, and 150 MW. In order to determine the costs of SCR for other sizes of turbines, the costs were plotted as a function of turbine size in MW to obtain a linear relationship. The resulting functions, in 1999 U.S. dollars, are shown below. The correlation coefficient for each relationship is greater than 0.99.

High Temp. Catalyst SCR Equipment = $15,710 x (MW) + $333,116
Conventional Catalyst SCR Equipment = $10,703 x (MW) + $293,599
Conventional Catalyst SCR Equipment with Duct Burner
= $10,703 x (MW) x 1.20 + $293,599

Although the DOE reference is almost 9-years old it continues to be cited by the utility industry trade group, Electric Power Research Institute (EPRI), and the U.S. EPA as a characterization of control costs. See EPRI publication "Technical and Economic Assessment of Combined Heat and Power Technologies for Commercial Customer Applications," published in March of 2003 and "Cost Impact of Proposed NSPS for Stationary Combustion Turbines" prepared for the U.S. EPA Office of Air Quality Planning and Standards, by Alpha-Gamma Technologies, Inc. in February of 2005.

Table B.3-2 below presents the inputs needed to determine capital and annual costs for each model turbine unit. Costs inputs have been adjusted to 2007 using the chemical engineering plant cost index.

Table B.3-2. Capital and Annual Cost Inputs

Input	Description	Cost	Reference
Basic SCR Equipment	High temp. catalyst	1.7 M - 8.1 M	U.S. DOE[1]
	Conventional catalyst	1.2 M - 5.6 M	U.S. DOE[1]
Operating Period	Peaking	4,200 hrs/yr	Engineering Judgment
	Base	8,400 hrs/yr	Engineering Judgment
	Duct Burner	8,400 hrs/yr	Engineering Judgment
Operator and Maintenance Labor	-	$40 per hr	Canadian Labour Relations [2]
Electricity Cost	-	$0.06 per kW-h	Engineering Judgment
Catalyst Replace Cost	-	$400 per ft3	Engineering Judgment
Catalyst Life	Peaking	5 years	Engineering Judgment
	Base	3 years	Engineering Judgment
Catalyst Disposal	-	$15 per ft3	U.S. DOE[1]
Ammonia	-	$397 per tonne	U.S. DOE[1]
Capital Recovery	-	7% interest rate	U.S. EPA[3]
	-	20 years of equipment life	Engineering Judgment
NO_x Concentration, Exhaust from DLN	-	25 ppmv	Engineering Judgment
Control Effectiveness	SCR	92%	Engineering Judgment
NO_x Concentration, Exhaust from SCR	-	2.0 ppmv	Engineering Judgment

3) ASSUMPTIONS

The required catalyst volume for the SCR was obtained from an EPRI article "Simple-Cycle SCR Costs," published in the CT Experience and Intelligence Report, October 2002.[4] The article presented SCR statistics for four (4) simple-cycle turbines. Reported statistics included catalyst volumes associated with 80% and 94% removal efficiencies. To achieve the desired NO_x exhaust gas concentration of 2.0 ppmv with a baseline of 25 ppmv, a 92% reduction is required. To determine the required volume, a linear relationship was developed from the available data. The resulting function, in 2002 U.S. dollars, is shown below.

Catalyst Volume (cubic foot per MW of CT output) = 1.27 x (target reduction) - 76.6

Therefore, the SCR catalyst volume to achieve a removal efficiency of 92% for a combustion turbine (CT) is 40.4 cubic feet for each MW of CT output.

The following assumptions are necessary to calculate control cost:

a) SCR control systems are to be installed on new turbines. Retrofit cost considerations such as additional ductwork or flue gas heaters necessary to make the SCR system work properly have been excluded. In a combined cycle combustion turbine (CCCT) configuration the SCR is inserted into existing heat recovery steam generator (HRSG).

b) Due to their high exhaust temperatures, it is assumed that simple cycle turbines use high temperature SCR catalyst.

c) SCR maintenance and operation will require a total of 1.0 hour per shift of labor. [1]

d) Required ammonia is equal to tonne of NO_x removed multiplied by 17/46.[1]

e) The SCR catalyst reactor will increase the back-pressure on the turbine, which decreases the turbine power output by approximately 0.5 percent. Annual costs include the cost penalty of this performance loss by assuming $0.06 per kW. [1]

f) According to construction permits from several states, issued from 2001 to 2008, the size of a duct burner package in a CCCT configuration is typically between 18% - 25% of the CT input capacity. Model units assume duct burner packages are 20% of the combustion turbine input capacity.

g) The total gross power output of a combined cycle plant includes the steam turbine and the combustion turbine. A combined cycle plant produces 60-70% of the total plant power from the combustion turbine and 30-40% from the steam turbine.[5]

h) The required SCR catalyst volume to achieve a removal efficiency of 92% from a CCCT with a duct burner is 48.5 ft3 of catalyst for each MW of CT output. Catalyst volume and SCR equipment costs have been increased by 20% to account for duct burner NO_x emissions.

i) The NO_x emission rate from a duct burner is 0.1 lb/MMBtu.[6]

4) TECHNICAL APPROACH

Direct and indirect installation costs are estimated as a percentage of the purchased equipment cost. The equations for each cost line-item are presented Table B.3-3 below.

Table B.3-3. Cost Factors

Item	Equation	Description	Reference
Direct Capital Costs (DC):	-	-	-
Purchased Equip. Cost (PE):	-	-	-
Basic Equipment:	A	Input (see above)	U.S. DOE[1]
Ammonia injection skid and storage	0.00 x A	Included in basic equipment	U.S. DOE[1]
Instrumentation	0.00 x A	Included in basic equipment	U.S. DOE[1]
Taxes and freight:	0.08 x A	Sales tax and delivery	U.S. EPA[3]
Direct Installation Costs (DI):	-	-	-
Foundation & supports:	0.08 x PE	for equipment, ductwork, piping	U.S. EPA[3]
Handling and erection:	0.14 x PE		
Electrical:	0.04 x PE	wiring and conduit to connect to plant supply systems	U.S. EPA[3]
Piping:	0.02 x PE		U.S. EPA[3]
Insulation:	0.01 x PE		U.S. EPA[3]
Painting:	0.01 x PE		U.S. EPA[3]
Indirect Costs (IC):	-	-	
Engineering:	0.10 x PE	-	U.S. EPA[3]
Construction and field expenses:	0.05 x PE	costs for construction, office and supervisory personnel, rental of temporary offices, etc.	U.S. EPA[3]
Contractor fees:	0.10 x PE	for construction and engineering firms involved	U.S. EPA[3]
Start-up:	0.02 x PE	-	U.S. EPA[3]
Performance testing:	0.01 x PE	verify SCR meets performance guarantees	U.S. EPA[3]
Contingencies:	0.03 x PE	redesign of equipment, delays in startup, increased labor cost	U.S. EPA[3]
Total Capital Investment (TCI):	DC + IC	-	-
Direct Annual Costs (DAC):	-	-	-
Operating Costs (O):	-	Input (see above)	-
Operator:	0.5 hr/shift	Assumption (see above)	U.S. DOE[1]
Supervisor:	Operator x 0.15	-	U.S. DOE[1]
Maintenance Costs (M):	-	-	-
Labor:	0.5 hr/shift	Assumption (see above)	U.S. DOE[1]

Table B.3-3. Cost Factors (Continued)

Item	Equation	Description	Reference
Material:	Labor x 1.00	replacement parts, lubricants, gaskets, seals	U.S. DOE[1]
Utility Costs:	-	-	-
Performance Loss:	decreased output 0.5%	Assumption (see above)	U.S. DOE[1]
Ammonia inject skid:	(5 kW blower + pump) x Operating Hrs		U.S. DOE[1]
Equipment:	-	-	-
Catalyst replacement:	3 or 5 yr. life	Input (see above)	U.S. DOE[1]
Catalyst disposal:	$15 per ft3	Input (see above)	U.S. DOE[1]
Ammonia Injection:	NO_x tonne x 17/46; $397 per tonne	Assumption and Input (see above)	U.S. DOE[1]
Indirect Annual Costs (IAC):	-	-	-
Overhead:	60% of O&M	Workmen's comp, vacations, employee amenities, plant lighting, parking areas, etc.	U.S. EPA[3]
Administrative:	0.02 x TCI	Accounting, R&D, etc.	U.S. EPA[3]
Insurance:	0.01 x TCI	-	U.S. EPA[3]
Property tax:	0.01 x TCI	-	U.S. EPA[3]
Capital recovery:	7%; 20-yr life	Input (see above)	-
Total Annual Cost:	(DAC + IAC)	-	-

5) NO_x REMOVED

NO_x emissions for turbines are typically presented as parts per million (ppm) reported at 15% O_2 in the exhaust stack. The concentration in the exhaust exiting the dry low-NO_x (DLN) combustors was converted from ppmv to lb/MWh using following equation:[7]

$$NO_x\text{, Exhaust from DLN (lb/MWh)} = \frac{(\text{(ppm @15\% } O_2\text{) x (10,500 Btu HHV/kWh heat rate))}}{\text{(272 ppm/(lb/MMBtu) x 1000)}}$$

To convert the concentration from lb/MWh exiting the DLN combustors to tons/yr the following equation was used:

$$NO_x\text{, Exhaust from DLN (tons/yr)} = \frac{\text{(lb/MWh) x (Turbine Output MW) x (Operating hours)}}{\text{2000 lbs}}$$

To calculate the tonne of NO_x removed annually by SCR the following equation was used:

$$NO_x\text{ removed (tonne/yr)} = \text{(tonne/yr of } NO_x\text{) x (SCR removal efficiency)}$$

6) EFFECTIVENESS

One measure of cost effectiveness was calculated by dividing the total annual cost by the annual NO_x reduction, in tonne. Total annual costs were calculated as the sum of all annual operating costs and annualized capital costs (see discussion and Table B.3-3 above). The following equation was used:

$$\text{Cost Effectiveness (\$/tonne)}: \frac{NO_x \text{ removed (tonne/yr)}}{\text{Total Annual Cost (\$)}}$$

A second measure of cost effectiveness was calculated by dividing the total annual cost by the turbine power output, in kilowatts. The following equation was used:

$$\text{Cost Effectiveness (\$/kW)}: \frac{\text{Turbine Output MW x 1000}}{\text{Total Annual Cost (\$)}}$$

7) DETAILED COST

Presented below are a series of tables. The tables represent the detailed costs and outputs.

References

1. ONSITE SYCOM Energy Corporation. Cost Analysis of NO_x Control Alternatives for Stationary Gas Turbines. Prepared for the U.S. Department of Energy, Contract No. DE-FC02-97CHIO877. November 5, 1999.

2. Wage Summary: Construction - Alberta 2007-2011. Prepared by Construction Labour Relations. August 25, 2008.

3. U.S. EPA. Control Cost Manual 5th Edition. Section 4.2 NO_x Post-Combustion. EPA/452/B-02-001. October 2000.

4. EPRI. Combustion Turbine Experience and Intelligence Report: 2002, Combustion Turbine/Combined Cycle Technology Developments, Reliability Issues, and Related Market Conditions, EPRI, Palo Alto, CA: 2002. 1004640.

5. Reale, Michael. GE Energy - New High Efficiency Simple Cycle Gas Turbine – GE's LMS100™. GER-4222A. June 2004. Published online <http://www.gepower.com/prod_serv/products/tech_docs/en/downloads/ger4222a.pdf>

6. U.S. EPA, Office of Air Quality Planning and Standards, Alternative Control Techniques (ACT) Document - NO_x Emissions from Stationary Gas Turbines, EPA-453/R-93-007, January 1993.

7. EPRI. Assessment of Emerging Low-Emissions Technologies for Combustion-Based Distributed Generators, EPRI, Palo Alto, CA: 2005. 1011341.

Appendix C

Future Technologies and Alternative Fuels – Sample Projects

Technology: Integrated Gasification Combined Cycle (IGCC)
Project Name: Various
Location: The Netherlands (Buggenum), USA (West Terre Haute, Indiana; Tampa, Florida; Reno, Nevada), Spain (Puertollano), and Italy (Priolo Gargallo)
Fuel: Coal
Size: 100-521 MW
Start: Various
Status: Operating (except for Reno, Nevada)

Description: Gasification converts carbon-containing material into a synthetic gas (primarily composed of carbon monoxide and hydrogen). The "syngas" is then either used as a fuel to generate electricity or steam, or used as a feedstock chemical in the petrochemical/refining industry. Sulfur is removed from the syngas either as sulfuric acid or in elemental form – both products are marketable. In addition, the high temperature of the gasification process converts ash and other inert materials into granular solids that can utilized in the building and construction industries.

Controls: No data found

Benefits: Can also be used for hydrogen generation. Identified as a suitable technology for carbon sequestration. Generated solid wastes are only 10-20% of those generated by other conventional power generation technologies, such as circulating fluidized bed combustion (CFBC) and pulverized coal boilers with flue gas desulfurization (FGD).

Plant Efficiency: 40-45%

Emissions: Reductions – SO_2 (90%+), NO_x (90%+), CO_2 (10-20%)

Problems: Current costs are relatively high ($1,600/kW) versus $1,250/kW for a standard pulverized coal facility and $550/kW for a conventional combined cycle gas turbine. Gasifiers tend to be unreliable and subject to stoppages and breakdowns. Less efficient with lower grade fuels.

Costs: $1,600/kW

Technology: Advanced Steam/Water Cycle, Ultra Super-Critical (USC) Boiler
Project Name: Advanced 700 °C PF Power Plant
Location: Europe
Fuel: Coal
Size: 400 MW
Start: 2013 (scheduled)
Status: Planning/Under construction

Description: The aim of the project is to break the "steel barrier" and introduce new nickel-based super alloys for the highest temperatures in the steam cycle, which will be in the range of 700 °C, thus boosting the coal-to-wire efficiency into the range of 52-55% depending on site and fuel conditions, compared to an efficiency of 35% for a typical coal-fired power plant. The project is technologically very advanced but even if it may succeed technically, the economy of the project is challenged by other fuels and technologies and to ensure the success of the whole project the complete plant structure will be reviewed to find other more cost effective ways of arranging the major components (i.e. boiler and turbine). This kind of advanced overall plant architecture has been called compact design. The original idea of this project was generated during ELSAM's R&D program of the mid 1990s and further promoted through a joint European program named COST. Now, 40 partners of which 26 are industrial companies, including leading boiler and turbine manufacturers, material and steel manufacturers, utilities and material test laboratories. The organization of the project has called for the formation of three main groups: a turbine group, a boiler group and a process group. The turbine group is responsible for the materials and design of the turbine; the boiler group is subdivided into three subgroups responsible for boiler design, boiler materials and combustion; and the process group is responsible for the cycle optimization, balance-of-plant, overall economy, etc. Following a preliminary design and feasibility study in parallel with initial materials testing, design and testing of large critical components will have to be performed as well as qualification of all optimized materials. The overall project is divided into 6 phases over a period of 15 years. Phase 1 started in 1998 and concluded at the end of 2003 and focused on a feasibility study and a program for the development of new materials. Phase 2 started at the beginning of 2002 and concluded in 2006 and concentrated on basic component design, as well as continued materials research. Construction is scheduled from 2010 to 2012 with operation beginning in 2013.

Controls: No data found

Benefits: Increasing plant efficiency from 35% up to 55% will yield a reduction of greenhouse gases of roughly 30%.

Plant Efficiency: Efficiency estimated at about 55%

Emissions: No data found

Problems: The new nickel-based materials to be used in headers, pipelines, and turbine components must have creep rapture strength values of at least 100 N/mm^2 over 100,000 hours of operation. Significant materials research is required to ensure sufficiently high creep rupture strength potential. In addition, there are questions regarding whether the new materials can be manufactured into pipes and whether they can be welded.

Costs: No data found

Technology: Advanced Steam/Water Cycle, Ultra Super-Critical (USC) Boiler
Project Name: Huaneng Yuhuan Power Plant
Location: Xiaqingtang, Zhejiang, China
Fuel: Coal
Size: 4 1,000 MW units
Start: 2007 (2 units); 2008 (2 units) (scheduled)
Status: Under construction/operating

Description: First ultra-supercritical plant to be constructed in China. First phase of plant (two units) was scheduled for completion in 2007, but unable to confirm this. A second ultra-supercritical plant (Waigaoqiao III) was awarded by the Shanghai Municipal Electric Power Company to Siemens. Completion expected in 2009.

Controls: No data found

Benefits: Increased plant efficiency – estimated at 45%

Plant Efficiency: No data found

Emissions: No data found

Problems: No data found

Costs: No data found

Technology: Pressurized Fluidized Bed Combustion (PFBC)
Project Name: Various
Location: Sweden (Vartan), Spain (Escatron), Germany (Cottbus), USA (Tidd), China (Wangjia), and Japan (Wakamatsu, Osaki, and Karita)
Fuel: Coal
Size: 70-350 MW (15 MW pilot-scale)
Start: Various
Status: Operating

Description: Fluidized bed combustion burns coal on a rapidly-flowing bed of moving air which allows complete coal combustion at relatively low temperatures and allows virtually any combustible fuels to be used. In the PFBC plant, steam is generated in a pressurized boiler; the pressurized coal combustion system heats steam for used in a steam turbine, and produces a hot gas supplied to a gas turbine.

Controls: No data found

Benefits: Can be used with low grade coal or mixed with other combustible fuels (e.g., biomass, coal waste, etc.). SO_2 removal more effective than post-combustion method, such as flue gas desulfurization. NO_x emissions lower due to lower operating temperatures.

Plant Efficiency: 40-45%

Emissions: Reductions – SO_2 (90%+), NO_x (90%+), CO_2 (10-20%)

Problems: Relatively small plant sizes. Reliability is variable because plants built largely as demonstration projects. Suppliers have not been actively pushing technology, but instead, favoring IGCC and supercritical PCC.

Costs: $1,300/kW

Technology: Advanced Circulating Fluidized-Bed (CFB)
Project Name: Western Greenbrier Co-Production Demonstration Project
Location: Rainelle, West Virginia, USA
Fuel: Bituminous coal mining waste material
Size: 100 MW (net)
Start: Unknown
Status: Project delayed for unknown reasons

Description: Advanced CFB used to convert approximately 4,000 tons/day of coal mining waste materials (“gob”) into 100 MW (net) of electricity. Additional by-products include: up to 20,000 pounds/hour of steam/hot water for industrial use and district heating, alkaline ash for remediation of acid water formation, and alkaline ash for co-production of salable products (e.g., structural bricks, cements, and aggregates). Waste coal and limestone are simultaneously fed to the CFB, which raises steam by passing water through water walls lining the CFB. The limestone removes the bulk of the sulfur in the coal feedstock, and the solids are entrained and re-circulated via the cyclone separators to enhance limestone and carbon utilization. An economizer located downstream of the cyclones recovers additional heat from the flue gas. Steam from the CFB boiler drives a nominal 100-MW (net) steam turbine. Bottom ash and a small portion of the fly ash are collected and returned to the source of the feedstock. The mildly alkaline nature of the ash assists in neutralizing the acid runoff from the waste pile, alleviating a significant environmental problem.

Controls: Limestone (SO_2); cyclones (PM); Selective non-catalytic reduction (SNCR), fly ash dryer absorber, and a baghouse (NO_x, SO_2, PM, and Hg)

Benefits: The advanced CFB incorporates an inverted cyclone separator and mid-support structure designs to reduce assembly time (6-8 weeks), lower material costs (60 percent less structural steel tonnage), and provide a smaller footprint (30-40% less) than conventional designs.

Plant Efficiency: No data found

Emissions: No data found

Problems: No data found

Costs: $215 million (50% U.S. DOE and 50% Western Greenbrier Co-Generation [WGC], LLC)

Technology: Carbon Capture and Storage
Project Name: Elsam Power Plant
Location: Esbjerg, Denmark
Fuel: Coal
Size: 420 MW
Start: March 2006
Status: Active

Description: The pilot project at the Elsam Power Station will demonstrate new technology for capturing carbon dioxide (CO_2) emissions as they are produced by power stations and then storing them underground. The pilot unit captures most of the CO_2 in the flue gases emitted by the coal-fired power station. The flue gases to be treated are directed to an absorber, where they are then mixed with a solvent. Having more affinity with the CO_2 molecules than with the other components of the flue gases, the solvent captures nearly 90 percent of the CO_2 in the flue gases. The CO_2-rich solvent is then fed to a regenerator. The device is heated to 120 °C (248 °F) in order to break the bonds between the CO_2 and the solvent. The CO_2 is then isolated and transported to its storage place and the solvent is restored to its initial form as CO_2-poor solvent and reintroduced into the absorber with more flue gases to be treated. This system is intended to capture one metric ton of CO_2 per hour. Four underground storage locations will be tested as part of the Elsam Power Station pilot project – three depleted oil and gas reservoirs located in the North Sea, in the Mediterranean Sea, and in Austria, and one saline aquifer located in the North Sea.

Controls: CO_2 capture and storage

Benefits: The cost of conventional processes for CO_2 capture in the flue gases of large industrial facilities, already operational in Japan, is estimated at 50-60 euros per metric ton of CO_2. The Elsam industrial pilot is expected to halve the cost per ton of CO_2 avoided, to between 20-30 euros per metric ton of CO_2.

Plant Efficiency: No data found

Emissions: No data found

Problems: A major problem that has been encountered is the frequent plugging of the solvent filter by gypsum and fly ash.
Costs: The total pilot project cost of 16 million euros is about half funded by the European Commission, with the remainder being funded by private partners. Work on capture technology will absorb 70 percent of the project's budget, and the remaining 30 percent will be spent on the storage of the captured CO_2 gas.

Technology: Carbon Capture and Storage
Project Name: FutureGen
Location: Mattoon, Illinois, USA
Fuel: Coal
Size: 275 MW
Start: 2012
Status: Uncertain – U.S. Department of Energy (DOE) funding withdrawn (January 29, 2008)

Description: FutureGen was a public-private partnership to build a near zero-emissions coal-fired power plant to produce electricity and hydrogen, while capturing and permanently storing carbon dioxide underground. FutureGen was intended to combine and test several new technologies in a single location, including coal gasification, emissions controls, hydrogen production, electricity generation, and CO_2 capture and storage (CCS). Integrated gasification combined cycle (IGCC) was the core technology behind FutureGen. IGCC power plants use two turbines (i.e., a gas and a steam turbine) to produce electric power more efficiently than pulverized coal plants. IGCC plants also make it easier to capture CO_2 for carbon sequestration. FutureGen was to capture CO_2 produced during the gasification process and pump it into deep rock formations thousands of feet under ground.

Controls: CO_2 capture and storage

Benefits: FutureGen specifically targeted rock formations containing saline water, as these are one of the most abundant types of geologic formations that can be used to store CO_2 worldwide. A study by the Global Energy Technology Strategy Program estimates the storage capacity of these saline rock formations in the U.S. to be 2,970 gigatons of CO_2, compared to a capacity of 77 gigatons of carbon dioxide for all other types of reservoirs, such as depleted gas fields. Focusing on rock formations with saline water was intended to help ensure that the lessons learned from the project are broadly transferable throughout the U.S. and around the world.

Plant Efficiency: No data found

Emissions: No data found

Problems: Funding from the DOE was withdrawn from the project on January 29, 2008. The principal reason given was rising costs. The FutureGen Alliance is continuing to push forward, in spite of the cancelled DOE funding. However, the future of the project is uncertain.
Costs: Gross project costs were estimated to be $1.5-1.8 billion. The original cooperative agreement between the DOE and the FutureGen Alliance established a distribution of 74% DOE funding and 26% private industry funding (U.S. and foreign utilities). The foreign governments of China, Japan, South Korea, India, and Australia also expressed interest in participating and sharing project costs.

Technology: Lignite Fuel Enhancement
Project Name: Great River Energy (GRE) Spiritwood Station
Location: Spiritwood, North Dakota, USA
Fuel: Lignite
Size: 62 MW (base); 37 MW (peak)
Start: 2010 (scheduled)
Status: Under construction

Description: The project will use a waste heat dryer to reduce the moisture of lignite coal from nearly 40 percent down to 25 percent. The coal drying technology was initially developed at GRE's Coal Creek Station. The technology utilizes water drawn from the cooling tower, which captures heat from the steam condenser in the boiler circuit, raising the temperature to about 120 °F. The heated water is routed to an air heater before returning to the plant cooling water circuit. Ambient air is heated in the air heater to about 105 °F and subsequently used as the fluidizing media in the fluidized-bed dryer to provide heat along with hot water. In practice, a two-stage dryer is used to enhance heat transfer. The Spiritwood Station will supply steam to a nearby malt plant and ethanol plant. The water source for the facility will come from multiple sources, but wastewater from the malt plant will be the primary source.

Controls: U.S. Best Available Control Technologies (BACT) will be used.

Benefits: This technology uses heat (that would otherwise be lost out the stack) to upgrade the low-rank coal feedstock, thereby enhancing plant efficiency and performance. Moreover, high-moisture content coals can contribute to corrosion of ductwork, and place an energy penalty on fans that move the vaporized water and pulverizers that process the moisture in the coal. GRE's upgrading process improves plant economics and reduces plant heat loss (decreases heat rate), increases efficiency, and thereby reduces emissions of carbon dioxide (CO_2), mercury, nitrogen oxides (NO_x), sulfur dioxide (SO_2), and particulate matter (PM) per unit of energy produced.

Plant Efficiency: Efficiency estimated at about 66 percent (combined heat and power)

Emissions: No data found

Problems: No data found

Costs: $276 million

Technology: Shell Oxygen-Blown, Entrained-Bed Gasifier; SASOL FT Coal Liquefaction Technology
Project Name: Gilberton Coal-to-Clean Fuels and Power Co-Production Project
Location: Gilberton, Pennsylvania, USA
Fuel: Anthracite coal mining waste material
Size: 41 MW
Start: Unknown
Status: Project delayed for unspecified reasons

Description: This project will use the gasification of 4,700 tons/day of coal waste to produce 41 MW of power and 5,000 barrels/day of clean liquid transportation fuel, including high-cetane diesel fuel and naphtha that contain no sulfur or aromatics. Coal waste is from abandoned anthracite culm piles, which will be processed at a rate of 1 million tons/year. In the conversion process, coal waste is fed to an oxygen-blown, entrained-bed gasifier that applies heat and pressure, transforms the ash constituent of the coal waste into an inert vitreous slag, and converts the hydrocarbon and sulfur constituents primarily into carbon monoxide (CO), hydrogen (H_2), carbonyl sulfide (COS), and hydrogen sulfide (H_2S). This raw synthesis gas is cleaned in a patented Rectisol process, which removes nearly all of the COS and H_2S. Clean synthesis gas (CO and H_2) is either shifted by the addition of steam to carbon dioxide (CO_2) and H_2 for separation, or used directly for power generation and liquid fuel production. Power is generated in a gas turbine, which in turn provides process heat and steam for a slurry-phase reactor. The reactor produces high-cetane diesel fuel and naphtha that contain no sulfur or aromatics. Naphtha can either be upgraded to a high-octane, clean burning reformulated gasoline or used as sulfur-free on-board reforming feed for fuel cell-powered vehicles.

Controls: No data found

Benefits: The process scheme is very flexible, allowing use of a broad range of feedstock (coal, coal waste, petroleum coke, biomass, and blends thereof), and facilitating carbon separation/capture for sequestration by keeping CO_2 streams concentrated. The project addresses the environmental issues associated with abandoned coal waste pile, while providing an alternative source of transportation fuels. If successful, this project is of sufficient scale to reduce technical, business, and financial risks, clearing the way for subsequent applications.

Plant Efficiency: Approximately 45 percent (based on total energy input and energy value of plant's products)

Emissions: No data found

Problems: Project delays appear to be due to funding and/or environmental issues.

Costs: $612,480,000 (16% U.S. DOE and 84% WMPI PTY, LLC)

Technology: Coal Gasification
Project Name: Power Systems Development Facility (PSDF)
Location: Wilsonville, Alabama, USA
Fuel: Coal
Size: Unknown
Start: 1996
Status: Operating

Description: The PSDF includes an engineering scale demonstration of key components of an Integrated Gasification Combined Cycle (IGCC) power plant, including a KBR Transport Gasifier, a Siemens hot gas filter using candle-type filter elements, syngas cooling, and high pressure solids handling systems. These components are designed at sufficient size to provide data for commercial scale-up. The KBR Transport Gasifier is a circulating fluidized bed reactor operating at higher circulation rates and riser densities than conventional circulating bed units, resulting in higher throughput, better mixing, and higher mass and heat transfer rates. Since the gasifier uses a dry feed system, it is well-suited for high moisture fuels such as sub-bituminous and lignite coals, but can also process some higher-rank coals. Virtually all the particulate from the syngas exiting the gasifier is removed by the downstream filter. System commissioning and initial test campaigns were performed in combustion mode from 1996 to 1999. Gasification operation began in late 1999, with four gasification commissioning tests completed by early 2001, and there have been 17 test campaigns since. As of May 2007, the PSDF gasification process had been operated for more than 10,000 hours. Filter element materials must be compatible with gasification operating conditions and be able to withstand system upsets. Many different types of elements have been tested at the PSDF, including monolithic ceramic, ceramic composite, sintered metal powder, and sintered metal fiber. The ceramic elements were primarily used in early combustion-mode tests at temperatures around 1400 ºF. The filter operating temperature was reduced to around 750 ºF for gasification, making it possible to use the more durable metal elements. In addition to semi-commercial scale testing, the PSDF has slip-stream testing capability for cost effective technology screening. Future PSDF work will include (1) scale-up and continued development of several CO_2 capture technologies being developed either at DOE's NETL facility, at private R&D laboratories or at PSDF as well as; (2) support for FutureGen; (3) efforts to enhance the coal feeding systems to enable wider ranges of coal as well as biomass to be economically and reliably introduced into many different versions of IGCC technology under consideration commercially today; and (4) synergistic use of DOE & PSDF capabilities in technology modeling screening, and engineering/economic assessments.

Coal Gasification converts carbon-containing material into a synthetic gas (primarily composed of carbon monoxide and hydrogen).

Controls: No data found

Benefits: No data found

Plant Efficiency: No data found

Emissions: No data found

Problems: No data found.

Costs: No data found

Technology: Coal Gasification
Project Name: Advanced Hydrogen Turbine Development Project
Location: Orlando, Florida, USA
Fuel: Coal
Size: Unknown
Start: 2006-2007 (Phase 1); 2008-2011 (Phase 2)
Status: Under development

Description: The objective of this project is to design and develop a fuel flexible (i.e., coal-derived hydrogen or syngas) gas turbine for IGCC applications that meets U.S. Department of Energy (DOE) turbine performance goals. The overall DOE Advanced Power System goal is to conduct, by 2010, the research and development (R&D) necessary to produce coal-based IGCC power systems with high efficiency (45-50%), near-zero emissions (less than 2 ppmv NO_x at 15% O_2) and competitive capital cost (less than $1,000/kW). To meet this 2010 goal the DOE Fossil Energy Turbine Program has as a primary objective through this project to demonstrate 2 to 3 percentage points improvement in combined cycle (CC) efficiency and a final goal of 3 to 5 percentage points improvement in CC efficiency (by 2015) above current state of the art CC turbines in IGCC applications with less than 2 ppmv NO_x (at 15% O_2). This project is expected to be implemented by 2012. The development timeline includes the following: Phase 1 (Studies and Conceptual Design) – 2006-2007; Phase 2 (Development and Detailed Design) – 2008-2011; Phase 3a (Manufacturing and Plant Construction) – 2012-2014; and Phase 3b (Initial Full Scale Plant Operations) – 2015-2018.

Controls: No data found.

Benefits: Development of coal-based IGCC power systems ready for CO_2 sequestration with high efficiency, near zero emissions and competitive plant capital cost.

Plant Efficiency: Goal of 45-50%

Emissions: Goal of 2 ppmv NO_x at 15% O_2

Problems: No data found.

Costs: No data found

Technology: Gasification and Thermal Recovery
Project Name: Philadelphia Biosolids Recycling Center
Location: Philadelphia, Pennsylvania, USA
Fuel: Various
Size: Unknown
Start: Unknown
Status: Operating

Description: The gasification and thermal recovery technology is a waste-to-energy conversion technology that uses intense direct heat in the absence of oxygen to convert organic material into sterile residues and recoverable syngas and thermal energy. Other potential applications include the processing of landfill gas, poultry litter, tires, paper, plastic, sludge, and a variety of other agricultural and municipal solid wastes. Wastes are continually loaded and conveyed through a retort chamber at rates that vary based on the composition of the wastes. Using an annular chamber that is adjacent to the retort chamber, the waste is indirectly heated to temperatures from 1000-1850 ºF (538-1010 ºC) until gasified (not incinerated). Wastes remain in the retort chamber until all volatile material is gasified and drawn onto a secondary chamber, called the thermal oxidizer, where volatile gases are destroyed. The unit discharges any remaining solids from the retort through air locks to ensure that oxygen is not drawn into the retort. These discharged materials are usually sterile and nonleachable. When the hot gases reach the thermal oxidizer, they are combusted with oxygen, and toxic and noxious wastes are eliminated. The resulting emissions are mostly in the form of carbon dioxide and water vapor due to the high temperatures of 1600-2250 ºF (871-1232 ºC) of the thermal oxidizer. Trace amounts of nitrous oxide, sulfur dioxide, carbon monoxide, and oxygen may remain, but they rarely exceed stringent air emissions criteria. This technology can generate sufficient amounts of steam to power a turbine for cogeneration.

Controls: No data found.

Benefits: The process can potentially reduce up to 95% of the initial waste material, leaving only a minimal amount of the waste stream to be discharged as solid material. Observations have shown that the technology is viable and sufficiently developed for the purpose of converting a broad spectrum of potential waste materials into recoverable syngas and thermal energy.

Plant Efficiency: No data found.

Emissions: No data found.

Problems: The main drawbacks of this technology include potential exposure of workers to pressurized flammable gas and steam, high voltage power, heavy waste materials, and high temperatures from the reactor vessel.

Costs: No data found

Technology: Advanced biomass gasification research and development
Project Name: Various
Location: USA – Salt Lake City, Utah; Roseville, Minnesota; Lansing, Iowa; East Hartford, Connecticut; Raleigh, North Carolina
Fuel: Various
Size: Unknown
Start: Various
Status: Research and development projects; candidates for future commercialization

Description: Emery Recycling (Salt Lake City, Utah) – using segregated municipal solid waste, animal waste, and agricultural residues to test new IGCC and integrated gasification and fuel cell (IGFC) concepts with a new gasifier. Sebesta Blomberg (Roseville, Minnesota) – using barley residues and corn stover from a malting facility to develop an atmospheric gasifier with a gas turbine. Alliant Energy (Lansing, Iowa) – using corn stover to develop a new combined-cycle concept that utilizes a fluidized bed pyrolyzer. United Technologies Research Center (East Hartford, Connecticut) – using clean wood residues and natural gas to test a biomass gasifier couple with an aero-derivative turbine with fuel cell and steam turbine options. Carolina Power and Light (Raleigh, North Carolina) – using clean wood residues to develop a biomass gasification process that will produce a reburning fuel stream for utility boilers.

Controls: No data found

Benefits: Reduces reliance on fossil fuels.

Plant Efficiency: No data found

Emissions: No data found

Problems: Commercialization potential of research and development project not known.

Costs: No data found

Technology: High-Pressure Oxy-Fuel Power Generation
Project Name: Clean Energy Systems (CES) Kimberlina Power Plant
Location: Bakersfield, California, USA
Fuel: Natural gas and various alternate fuels
Size: 20 MW
Start: 2005
Status: Active

Description: Initially this plant was used as a demonstration facility for CES' 20 MW gas generator operating on natural gas and pure oxygen. Subsequently, the facility has been used to demonstrate the gas generator on various alternative fuels including gas mixtures that simulate syngas derived from coal and biomass gasification, glycerin (a by-product of bio-diesel), and MSAR (Multiphase Superfine Atomized Residue – a stabilized emulsion of heavy petroleum refining residuals in water). An existing 5.5 MW Elliott steam turbine is driven by the CES gas generator to generate electrical power. The Kimberlina Power Plant continues to be used to perform demonstrations of CES' high-pressure oxy-fuel power generation technology using natural gas for interested parties. However, increasing emphasis is being focused on test demonstrations with alternative fossil- and bio-derived fuels. Tests have been performed in which glycerin has been co-fired with natural gas. Additional tests indicate that the technology can also be applied to other gas mixtures such as biomass-derived syngas, landfill gas, bio-digester gases, refinery off-gases, various low heating value gaseous fuels, and hydrogen-rich fuels. CES has designed and is fabricating a gas generator system rated at 170 MW. A J79 gas turbine is also being modified to operate directly on the drive gases produced by the newly designed gas generator. These components will be installed and tested at the Kimberlina facility starting early in 2008 at power levels up to 80 MW.

Controls: No data found.

Benefits: Exploratory tests indicate that the technology can be used to effectively dispose of low-value bio-fuel by-product with complete CO_2 capture. If the captured CO_2 were sequestered, the generated power would represent a "negative carbon balance" power cycle (i.e., a net removal of CO_2 from the atmosphere).

Plant Efficiency: No data found.

Emissions: No data found.

Problems: No data found.

Costs: No data found.

Technology: Photovoltaic (PV) Concentration
Project Name: EUCLIDES-THERMIE Power Plant
Location: Tenerife, Canary Islands, USA
Fuel: Solar
Size: 480 kW (peak)
Start: November 1998
Status: Active

Description: The PV concentration grid-connected power plant consists of 14 collector arrays, each 84 meters long, parallel to each other. The collector arrays are parabolic troughs with one axis tracking, oriented north/south and parallel to the ground. Each parabolic trough consists of 140 mirrors and 138 specially-designed PV receiver modules. The receiver modules are cooled with a passive heat sink. Every two contiguous troughs are connected to a 60 kVA inverter. All the seven inverters are then connected in parallel to the primary transformer, which connects the plant to the grid; avoiding any intermediate transformers has resulted in saving around 4% of the overall energy of the plant. The mirrors used in the troughs are made of reflective sheets glued onto aluminum panels. Three different materials were tested to be used for the reflective surfaces, and the acrylic silver film, ECP305, was chosen to be employed.

Controls: Not applicable

Benefits: No data found.

Plant Efficiency: 8.4% (10.95% at Madrid prototype)

Emissions: Not applicable

Problems: Windy and salty conditions have reduced the optical efficiency by about 20%, due to deposits on the mirrors and receivers. Misalignment of the mirrors and excessive gap between the mirrors has also proven to reduce the efficiency of the plant. Excessive wind conditions (90 kilometers/hour and above) have resulted in a small number of mirrors coming unglued from the panel. High level of grounding failure in receivers (i.e., 58 failures out of 1,932 total receivers, or 3%).

Costs: 2.13 million euro total cost; 4.45 euro/W_{peak} (future projection of 3.84 euro/W_{peak})

Technology: Integrated Solar Combined Cycle Power Plant
Project Name: New Energy Algeria (NEAL) Hassi R'Mel
Location: Northern Algeria
Fuel: Solar and natural gas
Size: 150 MW combined cycle gas turbine with 25 MW solar array
Start: 2010
Status: Under construction (Construction started July 2007)

Description: The Hassi R'mel integrated solar combined cycle power station is a hybrid power station that is the first of its kind. The plant will combine a 25 MW parabolic trough concentrating solar power array (covering an area of over 180,000 m^2) with a 150 MW combined cycle gas turbine plant. The solar array will provided complementary thermal energy to the combined cycle gas turbine. The solar field consists of 216 solar collectors in 54 loops. The inlet heat transfer fluid temperature will be 560 °F, while the outlet temperature will be 740 °F.

Controls: No data found

Benefits: Carbon emissions will be lower compared to a traditional power station. The solar array investment will be minimized due to shared components with the combined cycle portion of the power station.

Plant Efficiency: No data found

Emissions: No data found

Problems: No data found

Costs: $177 million

Technology: Phosphoric Acid Fuel Cell (PAFC) Systems
Project Name: HydroGen/ASHTA Chemicals
Location: Ashtabula, Ohio, USA
Fuel: Hydrogen
Size: 400 kW
Start: 2008
Status: Active

Description: HydroGen successfully fully started up a full scale 400 kilowatt (kW) commercial demonstration PAFC plant at a chlor-alkali facility. The plant serves as a field prototype for the multi-megawatt fuel cell plants currently being developed by HydroGen for near term commercial deployment, and is designed to use by-product hydrogen produced by the chemical facility to produce and feed electrical power back to the facility for its internal use. The start-up follows the successful design, engineering and construction of the fuel cell plant at the chemical facility, and the manufacture, delivery, and installation of the 400 kW PAFC module. The plant will now be taken through a program of testing and optimization that will include tests of all start-up, operational, and shut-down modes at progressively higher power levels, followed by a reliability run to demonstrate system robustness under commercial operating conditions.

Controls: No data found.

Benefits: Potential sites can be found among “hydrogen-available” market segments.

Plant Efficiency: No data found.

Emissions: No data found.

Problems: No data found.

Costs: The Ohio Department of Development provided a grant of $1,250,000 to support design and construction of the plant. Unsure of total plant cost.

Technology: Compressed Air Energy Storage
Project Name: Iowa Stored Energy Park (ISEP)
Location: Fort Dodge, Iowa, USA
Fuel: Wind and fossil fuels
Size: Unknown
Start: 2011
Status: Planned

Description: ISEP will use the energy from a large wind farm located in Iowa where there are good wind resources. This wind power will be used to store compressed air in an underground geologic structure. During peak power demands, the stored air will be released, mixed with a fuel and used to power combustion turbines that produce environmentally friendly and economical electricity. ISEP will utilize some of the latest innovations in the generation of electricity. Wind turbines, deep underground air storage, and efficient combustion turbines will be used to take the variability of wind and turn it into clean energy on demand. Testing and analysis of the ability to store the air underground is being conducted at potential locations. When the results from the studies are completed, the project will move into the design phase, with construction to follow. There are only two existing CAES facilities – a 290 MW plant in Huntorf, Germany (built in 1978) and a 110 plant in McIntosh, Alabama, USA. A larger CAES facility (i.e., 2,700 MW) was proposed for Norton, Ohio, USA in 2001; however, construction has not begun. Unlike these three facilities, ISEP will utilize an underground aquifer (instead of an underground cavern) and will supplement fossil fuels with wind power.

Controls: No data found.

Benefits: During off-peak periods, cheaper excess power can be used for air compression. Air storage will use an underground aquifer rather than an underground cavern; aquifer requires less pressure and less energy.

Plant Efficiency: No data found.

Emissions: No data found.

Problems: No data found.

Costs: $215 million. ISEP is a joint project of municipal utilities in Iowa, Minnesota, North Dakota, and South Dakota. Additional funding for this project will also come from U.S. DOE.

Appendix D

Future Control Technologies – Sample Projects

Technology: TOXECON^{TM}
Target Pollutants: Hg, SO_2, NO_x, PM
Location: Presque Isle Power Plant (We Energies), Marquette, Michigan, USA
Emission Reduction: Hg – 90%; SO_2 – 30%; and NO_x – 70%
Project Timeline: Start – April 2004; completion - 2009
Status: Under development

Description: We Energies of Milwaukee, Wisconsin, will design, install, operate and evaluate the TOXECON process as an integrated system to control emissions of mercury, particulate matter (PM), sulfur dioxide (SO_2) and nitrogen oxides (NO_x) during the operations of its Presque Isle Power Plant. In the patented TOXECON process, sorbents are injected into a power plant's exhaust stream to soak up the pollutants so they can be captured and prevented from being released to the atmosphere. The project's key objectives include: achieving very high levels of mercury removal, increasing the collection efficiency of particulate matter, and determining the viability of sorbent injection for SO_2 and NO_x control, all while maximizing the use of coal combustion by-products. An additional feature allows the system configuration to keep the activated carbon sorbent separate from the captured fly ash, which can be sold for reuse in concrete applications.

Benefits: Likely to become primary mercury control choice for users of Western coals; only choice for units burning any coal type with hot-side electrostatic precipitators.

Emissions: When completed in 2009, the project is expected to reduce mercury emissions at the plant by 90 percent; it will capture about 80 pounds per year of mercury. It also will eliminate 1,145 tons per year of SO_2 and 428 tons per year of NO_x – reductions of 30 percent and 70 percent, respectively, above and beyond what Presque Isle currently removes.

Problems: No data found

Costs: The \$52.9 million TOXECON project will be managed by DOE’s National Energy Technology Laboratory. DOE's share of the cost of the five-year project is \$24.8 million, and We Energies' share is \$28.1 million.

Technology: Membrane-Based Up-Flow Wet Electrostatic Precipitation
Target Pollutants: Hg, SO_3, $PM_{2.5}$
Location: Bruce Mansfield Plant (Penn Power), Shippingport, Pennsylvania, USA
Emission Reduction: Hg (elemental) – 33%; Hg (oxidized) – 82%; Hg (particulate) – 100%; H_2SO_4 – 93%; $PM_{2.5}$ – 96%
Project Timeline: Start – October 2002; completion – March 2004
Status: Testing completed

Description: The project involved conducting detailed tests of metallic and new membrane collection material within a wet electrostatic precipitator (WESP). The project's overall objectives were to compare the performance of metallic collecting surfaces to the performance of membrane (fabric) collecting surfaces in a WESP, in terms of their efficiency in removing fine particles, acid aerosols, and mercury from an actual power plant flue gas stream; and to determine the relative durability and overall cost-effectiveness of the membrane collectors versus metallic collectors. This project utilized the existing 2-field WESP infrastructure installed at the plant as a baseline to compare the membrane technology. A new 2-field membrane WESP similar in design to the metallic WESP was installed alongside the existing metallic WESP. Test results demonstrated the membrane material had no adverse impact and achieved similar removal levels as that for the metallic unit. Differences between removal efficiencies is attributable to either averaging differences due to the limited number of test runs and/or improved collection properties of the membrane collection material.

Benefits: The membrane materials are less expensive than the metallic corrosion resistant alloys used in the metallic WESP.

Emissions: Results indicated a higher removal efficiency for the membrane WESP compared to the metallic WESP.

Problems: The disadvantages with the membrane materials are, long-term deterioration of the material, potential plugging of the membrane material from salt build-up, potential to blow holes in the membrane from arcing if not saturated, and potential fire hazard if there is a loss of water within the WESP.

Costs: No data found

Technology: Non-Thermal Plasma-Based Multi-Pollutant Control (Electro-Catalytic Oxidation – ECO)
Target Pollutants: Hg, NO_x, SO_2, $PM_{2.5}$
Location: R.E. Burger Generation Station (Ohio Edison), Akron, Ohio, USA
Emission Reduction: Hg – 80-90%; NO_x – 90%; SO_2 – 98%; $PM_{2.5}$ – 95%
Project Timeline: Start – September 2001; completion – September 2004
Status: Testing completed

Description: Electro-Catalytic Oxidation (ECO) is an integrated air pollution control technology that achieves major reductions in emissions of multiple pollutants (i.e., Hg, NO_x, SO_2, fine particulate matter) from the flue gas emitted by coal-fired power plants. The patented technology also reduces emissions of air toxic compounds such as arsenic and lead as well as acid gases such as hydrochloric acid (HCl). An ECO pilot unit was installed and has been in operation since February 2002. The pilot treats 1500 to 3000 scfm of flue gas drawn from the boilers. In commercial operation, the ECO process is to be installed downstream of a power plant's existing electrostatic precipitator or fabric filter. It treats flue gas in four steps to achieve multi-pollutant removal. In the first process step, a barrier discharge reactor oxidizes gaseous pollutants to higher oxides. For example, nitric oxide is oxidized to nitrogen dioxide and nitric acid, a small portion of the sulfur dioxide is converted to sulfuric acid, and mercury is oxidized to mercuric oxide. Following the barrier discharge reactor is an ammonia-based scrubber, which removes unconverted sulfur dioxide and the nitrogen dioxide produced in the barrier discharge, creating an ammonium sulfate-ammonium nitrate solution. A wet electrostatic precipitator (WESP) follows the scrubber. It, along with the scrubber, captures acid aerosols produced by the discharge reactor, fine particulate matter and oxidized mercury. The WESP also captures aerosols generated in the scrubber. Finally, liquid effluent produced by the scrubber contains dissolved ammonium sulfate and nitrate salts, along with Hg and captured particulate matter. It is sent to a co-product recovery system, which includes filtration to remove ash and activated carbon adsorption for Hg removal. The treated co-product stream, free of Hg and ash, can be processed to form ammonium sulfate nitrate (ASN) fertilizer in crystal, granular, or liquid form. A booster fan moves flue gas through the barrier discharge reactor then the absorber/WESP tower. Aqueous ammonia and water is added to the upper loop of the absorber to maintain scrubbing pH and density. Liquid is transferred from the upper loop to the lower loop to maintain upper loop scrubbing conditions. Particulate matter and aerosols captured in the WESP are also drained to the lower loop through condensation in the WESP and periodic washing. Evaporation of water in the lower loop concentrates the liquid to a near saturated ammonium sulfate-nitrate (ASN) solution. The co-product stream is then drawn from the lower loop and processed to produce commercial grade fertilizer. Water removed from the co-product is then returned to the ECO process.

Benefits: The ECO equipment has a much smaller footprint than conventional control equipment and can be easily installed at space-constrained sites.

Emissions: No data found

Problems: No data found.

Costs: The capital costs for the ECO system are estimated to be approximately $250 per kW including balance of plant modifications. Levelized O&M costs are estimated to be 2.0 to 2.5 mils/kW.

Technology: Amended Silicates^{TM}
Target Pollutants: Hg
Location: Miami Fort Station (Duke Energy), North Bend, Ohio, USA
Emission Reduction: Hg – 40%
Project Timeline: Test – 1st Quarter 2006
Status: Testing completed

Description: Amended Silicates^{TM}, a powdered, noncarbon mercury-control sorbent, was tested at Duke Energy's Miami Fort Station, Unit 6 during the first quarter of 2006. Unit 6 is a 175 MW boiler with a cold-side electrostatic precipitator (ESP). The plant burns run-of-the-river eastern bituminous coal with typical ash contents ranging from 8-15% and sulfur contents from 1.6-2.6% on an as-received basis. The performance of the Amended Silicates sorbent was compared with that for powdered activated carbon (PAC). The trial began with a period of baseline monitoring during which no sorbent was injected. Sampling during this and subsequent periods indicated mercury capture by the native fly ash was less than 10%. After the baseline period, Amended Silicates sorbent was injected at several different ratios, followed by a 30-day trial at a fixed injection ratio of 5-6 lb/MMACF. After this period, PAC was injected to provide a comparison. Approximately 40% mercury control was achieved for both the Amended Silicates sorbent and PAC at injection ratios of 5-6 lbs/MMACF. Higher injection ratios did not achieve significantly increased removal. Similar removal efficiencies have been reported for PAC injection trials at other plants with cold-side ESPs, most notably for plants using medium to high sulfur coal.

Benefits: Sorbent injection did not detrimentally impact plant operations and testing confirmed that the use of Amended Silicates sorbent does not degrade fly ash quality (unlike PAC).
Emissions: 40% Hg emissions control.

Problems: No data found.

Costs: The cost for mercury control using either PAC or Amended Silicates sorbent was estimated to be equivalent if fly ash sales are not a consideration. However, if the plant did sell fly ash, the effective cost for mercury control could more than double if those sales were no longer possible, due to lost by-product sales and additional cost for waste disposal. Accordingly, the use of Amended Silicates sorbent could reduce the overall cost of mercury control by 50% or more versus PAC for locations where fly ash is sold as a by-product.

Technology: Activated Carbon Injection
Target Pollutants: Hg
Location: Stanton Station (Great River Energy), Stanton, North Dakota, USA
Emission Reduction: Hg – >90% (chemically treated sorbents); 75% (non-treated sorbents)
Project Timeline: Unknown
Status: Testing completed

Description: Carbon injection technology is one of the most studied mercury control technologies available for coal-fired power plants. The technology has shown the capability to achieve fairly high mercury removals in plants burning specific coals and employing specific pollution control devices. For plants firing lignite fuels, injection of untreated activated carbons has shown lower mercury removal performance than for other coals, such as bituminous coals. The low chlorine and high calcium content of lignite fuel results in low reactivity between the mercury and sorbent, thereby resulting in poorer performance. The reactivity between mercury and sorbent can be improved by using chemically-treated carbons or by using furnace additives to increase the chloride content of the flue gas. Six sorbents were tested on Stanton Station Unit 10 for their ability to remove flue gas mercury. Two of the chemically treated sorbents (ST BAC and Norit E-3) were able to achieve mercury removals greater than 90% across the spray dryer/baghouse combination at sorbent injection rates as low as 1.5 lb/MMacf. In contrast, the non-treated sorbents were limited to 75% mercury removal, even at injection rates as high as 6 lb/MMacf. The Norit E-3 sorbent was selected for 24 days of continuous sorbent injection testing. During the long-term tests, an injection rate of 1 lb/MMacf was maintained in order to achieve 65-75% mercury removal across the spray dryer/baghouse combination. The resulting outlet mercury concentration was in the general range of 2.5-3.5 $\mu g/Nm^3$.

Benefits: No data found.

Problems: The most notable change in plant operation during sorbent injection was a doubling of the baghouse cleaning frequency. The increased particulate load to the baghouse was estimated to be less than 0.2%. Therefore, it is unlikely the increase in cleaning frequency is attributable to sorbent injection.

Costs: No data found.

Technology: MerCAP^{TM}
Target Pollutants: Hg
Location: Stanton Station (Great River Energy), Stanton, North Dakota, USA
Emission Reduction: Hg – 30-35%
Project Timeline: Start – September 2003; Completion – September 2006
Status: Testing completed

Description: The general concept for MerCAP^{TM} is to place fixed structure sorbents into a flue gas stream to adsorb mercury and then, as the sorbent surfaces becomes saturated, regenerate the sorbent and recover the mercury. Results from modeling studies and field testing of a single-plate, gold-coated MerCAP probe have indicated that high mercury removals can be achieved over relatively short plate lengths at very high flue gas velocities. Since the gold sorbent can efficiently capture elemental mercury, whereas existing wet or dry flue gas desulfurization (FGD) units are better suited to capture oxidized mercury, a promising initial retrofit application of the MerCAP technology is for "polishing" of elemental mercury downstream of FGD devices. Small-scale tests have also indicated that gold-coated plates can be thermally or chemically regenerated without degradation of the adsorption capacity. The system is very flexible via alteration of plate length, spacing, and flue gas flow rate. Disposal of mercury-contaminated sorbents is minimized. MerCAP technology has been in continuous operation for over 5,300 hours at Stanton Station. The first 1,700 hours of service were with ND lignite coal and the remaining service hours with PRB coal. Mercury removal efficiencies with acid pre-treated plates and 1-inch spacing have averaged 30-35% during this time period. Acid pre-treated plates removed mercury more efficiently than untreated plates; regeneration via acid-washing also improved the mercury removal performance of the plates. Tighter plate spacing (½-inch vs. 1-inch) also improved mercury removal performance. MerCAP substrates were subjected to three thermal regeneration cycles with no measurable negative impact on the mercury capture performance; removal efficiencies actually increased slightly after each regeneration cycle.

Benefits: No data found.

Problems: Lower fuel sulfur levels associated with PRB fuel have required less reagent slurry injection into the spray dryer FGD than for lignite. However, the higher flue gas temperatures associated with low rates of slurry injection seem to adversely affect the mercury capture efficiency of the gold sorbent. The relationship of plant slurry feed rates and spray dryer outlet temperatures to MerCAP performance is being further investigated. During both thermal and chemical regeneration, only a fraction of the mercury theoretically collected on the plates was recaptured. Reasons for this discrepancy are also being investigated.

Costs: $1,725,716 ($1,113,216 – DOE; $612,500 – non-DOE)

Technology: Mercury Oxidation Catalyst
Target Pollutants: Hg
Location: Coal Creek Station (Great River Energy), Underwood, North Dakota, USA and J.K. Spruce Power Plant (City Public Service of San Antonio), San Antonio, Texas, USA
Emission Reduction: Hg – 12-98% oxidation (dependent upon catalyst type); 82% total Hg capture
Project Timeline: Start – October 2002; Completion – April 2005
Status: Testing completed

Description: The mercury control process under development uses catalyst materials in honeycomb form to promote the oxidation of elemental mercury in the flue gas from coal-fired power plants that have wet lime or limestone flue gas desulfurization (FGD) systems. Oxidized mercury is removed in the wet FGD absorbers and leaves with the by-products from the FGD system. The pilot-scale tests were conducted for at least 14 months at each of two sites to provide longer-term catalyst life data. An important finding during these tests was that, even though the mercury oxidation catalyst pilot unit was installed downstream of a high-efficiency electrostatic precipitator (ESP) at Coal Creek Station (CCS), fly ash builds up over time and began to plug flue gas flow through the horizontal catalyst cells. Sonic horns were installed in each catalyst compartment and appeared to limit fly ash buildup, at least for the three catalysts of most interest for process commercialization. Over nearly 21 months of operation, a palladium-based catalyst slowly declined in activity, with initial elemental mercury oxidation percentages of 95% across the catalyst, but declining to 67% at the end of the period. A carbon-based catalyst began with almost 98% elemental mercury oxidation across the catalyst, but declined in activity to achieve about 79% oxidation after nearly 13 months in service. The other two catalysts were less active. An SCR-type catalyst (titanium/vanadium) declined from an initial value of 67% to only 26% elemental mercury oxidation after nearly 21 months in flue gas service, while an experimental fly-ash-based catalyst declined to only 12% oxidation after nearly 17 months. The fly-ash-based catalyst was not cleaned effectively by the sonic horn as described above, so the loss of activity observed may be exacerbated by the buildup of CCS fly ash deposits in the horizontal catalyst cells. The palladium-based and SCR-type catalysts were effectively regenerated at the end of the long-term test by flowing 600 ºF air at a slow rate through the catalyst overnight, with much of their original activity for elemental mercury oxidation being restored. The carbon-based catalyst was not observed to regenerate at this temperature. No attempt was made to regenerate the fly-ash-based catalyst due to the observed fly ash buildup within the catalyst cells. From pilot testing at both plants, it was determined that a plant configuration with a baghouse rather than an ESP for particulate control was particularly advantageous for the activated carbon injection process and disadvantageous for mercury oxidation catalysts.

Benefits: No data found.

Problems: As described above, the pilot-scale results showed the catalysts could not sustain 90% or greater oxidation of elemental mercury in the flue gas for a period of two years.

Costs: No data found.

Technology: Low Temperature Mercury Control (LTMC)
Target Pollutants: Hg
Location: R. Paul Smith Station (Allegheny Energy), Williamsport, Maryland, USA
Emission Reduction: Hg – 90%
Project Timeline: Completion – December 2008
Status: Testing currently being conducted

Description: Low temperature mercury control (LTMC) has the ability to reduce mercury emissions by over 90%, as was recently demonstrated by CONSOL R&D on a slip-stream pilot plant at the Allegheny Energy Mitchell Station under DOE contract. The next step is to demonstrate the performance, operability, and economics on a full-scale utility boiler. In addition, this project will demonstrate that magnesium hydroxide ($Mg(OH)_2$) slurry injection into the flue gas reduces sulfur trioxide (SO_3) concentration sufficiently to avoid corrosion at the low-temperature conditions, and will demonstrate that water spray humidification can maintain the electrostatic precipitator (ESP) performance under low SO_3 conditions. The LTMC process controls mercury by cooling the flue gas temperature to approximately 220 °F and absorbing the mercury on the carbon inherent in the fly ash. The host site will be the AE R. Paul Smith Unit 4, which is a nominal 88 MW bituminous coal-fired unit. R. Paul Smith fly ash, with an LOI between 15 to 20%, is ideally suited for mercury capture. The flue gas exiting the boiler from this unit is divided into two ducts, each equipped with its own air heater and ESP. The LTMC process will be installed on both ducts. The flue gas temperature will be reduced from a nominal 280 °F to 220 °F using water sprays inserted into the existing ductwork. To prevent corrosion, magnesium hydroxide slurry will be sprayed into the flue gas with the water sprays. A two-month test will collect operating data on mercury removal and balance of plant impacts from the process. The project will include an economic analysis, including estimates of capital costs, and fixed and variable operating and maintenance (O&M) costs.

Benefits: This technology has the potential to remove over 90% of the flue gas mercury at a cost at least an order of magnitude lower (on a $/lb mercury removed basis) than activated carbon injection. The technology is suitable for retrofitting to existing and new plants, and, although it is best suited to bituminous coal-fired plants, it may have some applicability to the full range of coal types.

Problems: No data found.

Costs: No data found.

Technology: Wet Scrubbing Mercury Removal Technology
Target Pollutants: Hg
Location: Endicott Station (Michigan South Central Power Agency), Litchfield, Michigan, USA and Zimmer Station (Duke Energy), Moscow, Ohio, USA
Emission Reduction: Hg – 51-79%
Project Timeline: Start – October 2000; Completion – June 2002
Status: Testing completed

Description: The goal of this project is to commercialize methods for the control of mercury in coal-fired electric utility systems equipped with wet flue gas desulfurization (wet FGD). The wet scrubbing mercury removal technology relies on the addition of very small amounts of a liquid reagent to existing wet FGD units to achieve increased mercury removal

Benefits: The wet scrubbing mercury removal technology is easily retrofitable to existing units. The equipment necessary to add a liquid reagent to existing wet FGD units is very minimal. This allows for a practical system with a very small system footprint and ease of use. There is the potential for multiple pollutant control beyond mercury. Effective mercury sequestration in the wet scrubbing solid by-product is expected with this technology. Very small amounts of reagent are used in the process, therefore virtually no impact on the by-product is seen based on the reagent alone. The additional mercury sequestered because of the increased mercury capture should have very little to no impact on the disposal or use of the by-product based on the results of pilot scale tests.

Problems: No data found.

Costs: Based on preliminary economic analyses completed to date, the B&W/MTI technology has the potential to be significantly more cost-effective than activated carbon for the same level of removal.

Technology: Advanced Hybrid Particulate Collector (Advanced Hybrid^{TM})
Target Pollutants: $PM_{2.5}$
Location: Big Stone Power Plant (Otter Tail Power Company), Milbank, South Dakota, USA
Emission Reduction: PM – 99.9% (from 0.01 to 50 µm)
Project Timeline: Start – October 2002; Completion – December 2005
Status: Testing completed

Description: The technology consists of fabric filter bags interspersed with perforated electrostatic precipitator (ESP) plates and electrodes in the same housing. The filter bags can achieve greater collection of very fine particles than can the ESP plates, while the ESP plates can capture dust that is re-entrained due to back-pulsing of the fabric filter bags.

Benefits: The combination of these two technologies (ESP and filtration) in the patented Advanced Hybrid^{TM} technology uses the ESP portion to capture the bulk of the particles (as much as 90%) and allows the filter bags to be made out of highly efficient membrane materials because of a reduction in filtration surface required as compared to conventional pulse-jet type fabric filters. As such, it is anticipated that the particulate control device can operate at 2.5 - 4 times the throughput of conventional fabric filters.

Problems: No data found.

Costs: This technology offers the potential to increase fine particle ($PM_{2.5}$) collection efficiency by one or two orders of magnitude (i.e., 99.99% to 99.999% removal) at a cost that is roughly comparable to conventional particulate control technology.

Appendix E

Operation Status of Units Reviewed for Boiler BACTEA

Appendix E. Operation Status of Units Reviewed for Boiler BACTEA

Company or Facility Name	Location	County in Attainment for NO_x, SO_2, and PM?	Contact	Unit Status
Western Farmers Electric Coop - Hugo Generating Station	Choctaw County, OK	Yes	Charles Coller (580) 873-2201	Unable to Contact
Great Plains Energy - Kansas City Power & Light Company - IATAN Station	Platte County, MO	Yes	Paul Ling (816) 556-2200	Under construction; unit 1 expected to come online in 2009 and unit 2 projected to come online in 2010.
Great Plains Energy - Kansas City Power & Light Company - IATAN Station	Platte County, MO	Yes	Paul Ling (816) 556-2200	Under construction; unit 1 expected to come online in 2009 and unit 2 projected to come online in 2010.
Xcel Energy Comanche Station, Colorado	Pueblo County, CO	Yes	Gary Magno (720) 497-2112	Under construction; projected start up summer of 2009
Omaha Public Power District (OPPD) - Nebraska City Station	Otoe County, NE	Yes	http://www.oppd.com/AboutUs/22_002695	Latter stages of construction and is expected to begin commercial production of electricity by May of 2009
Wisconsin Public Service (WPS) - Weston Plant	Marathon County, WI	Yes	Connie Lawniczak (920) 433-1140	**Operational since mid-2008**; compliance testing completed (has a problem meeting SO_2 limit during startups (over the 3 hr-avg), but company has resolved this problem by finding a way to bring the dry scrubber on earlier and not destroy it from high temperatures). Although in compliance, the company is also working on a Hg optimization program.
City Public Service of San Antonio, Spruce 2	Bexar County, TX	Yes	http://www.cpsenergy.com/About_CPS_Energy/Who_We_Are/History/History_of_CPS_Energy.asp http://www.sourcewatch.org/index.php?title=Spruce_Unit_2 http://www.burnsmcd.com/portal/page/portal/Internet/Service/Power_Generation1/Power%20Gen%20PD%20Repository/JK%20Spruce%20Unit%202	Under construction; projected start up August 2009

Appendix E. Operation Status of Units Reviewed for Boiler BACTEA (Continued)

Company or Facility Name	Location	County in Attainment for NO_x, SO_2, and PM?	Contact	Unit Status
Longleaf Energy Associates	Early County, GA	Yes	http://www.sourcewatch.org/index.php?title=Longleaf http://www.earlycountynews.com/news/2008/1217/front_page/001.html	Not begun construction due to litigation
Intermountain Power Service Corporation	Sevier County, UT	Yes	Blaine Ipson (435) 864-6406 http://www.intermountainpower.com/About_Us.html	Unit 2 upgrade completed in spring 2004 and **operational**; no issues meeting permitted limits
City Utilities of Springfield, Southwest Power Station	Greene County, MO	Yes	http://www.cityutilities.net/sw2/timeline.htm http://www.power-technology.com/projects/springfield/	Unit 2 is expected to come on line in 2010
LS Power, Sandy Creek Energy Station	McLennan County, TX	Yes	http://www.sourcewatch.org/index.php?title=Sandy_Creek_Plant	Construction has begun, but heavy litigation; projected to go into service 2012
Black Hills Power and Light (Wygen 2)	Campbell County, WY	Yes	http://www.redorbit.com/news/business/1211556/black_hills_corporation_announces_commencement_of_wygen_ii_power_plant/index.html Tim Rogers (605) 721-2286	**Entered into commercial service on January 1, 2008**; still completing testing, so far unit shows compliance with the permitted limits.