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Orlando Utilities Commission
Curtis H. Stanton Energy Center
Unit 2
Supplemental
Site Certification Application
Volume 1A

Introduction

The Supplemental Site Certification Application is being submitted by Orlando Utilities Commission (OUC), Florida Municipal Power Agency (FMPA), and Kissimmee Utility Authority (KUA) for the Unit 2 addition to the Curtis H. Stanton Energy Center (Stanton) in accordance with the Florida Electrical Power Plant Siting Act. The site was previously certified for an ultimate site development of approximately 2,000 MW of coal fired capacity at the same time that Unit 1 was certified (order entered December 15, 1982; DOAH Case No. 81-1431). The certified site included corridors for the railroad, transmission lines, makeup water supply pipelines, and the site access road. This supplemental application is filed pursuant to FS 403.517 regarding supplemental site certification applications. No DER instruction guide has been prepared for supplemental site certification applications, so this application has followed the outline of DER Form 17-1-211(1) for certification applications. Information in this supplemental application is supplied only for the purpose of assessing the need, construction, and operation of Stanton 2. This supplemental application is intended to serve the following purposes.

- Supplemental Site Certification for Unit 2 of the Stanton Energy Center including the Need for Power chapter.
- Revision to Best Available Control Technology (BACT) for the existing US Environmental Protection Agency approval of Stanton 2 under a two-unit phased construction Prevention of Significant Deterioration (PSD) permit and request for an amendment of the commence construction date for Unit 2 in the original PSD permit.
- Joint Application (US Army Corps of Engineers and Florida Department of Environmental Regulation) for dredge and fill permit for site development and construction of all associated facilities for Stanton 2 as may be required by the Clean Water Act.
- Permit applications for all other required state, regional, and local approvals.

A PSD permit (PSD-FL-084) was issued to OUC on June 10, 1982, for the phased construction of Units 1 and 2 at the Stanton Energy Center. Within this permit, the construction of Stanton 2 was scheduled to commence on July 1, 1990, with expected startup in January 1994. The authority to construct Stanton 2 under this permit will expire on January 1, 1992, unless construction has commenced by that date. This application includes the revised BACT and request for an extension in the commence construction deadline for a period of 18 months.

The Supplemental Site Certification Application comprises five volumes. The first four volumes contain Chapter 1 of the Supplemental Site Certification Application and are labeled 1A, 1B, 1C, and 1D. Chapter 1 contains the Public Service Commission Need for Power (NFP) Application portion of the Supplemental Site Certification Application. The Joint Need for Power Application is based on the needs of the joint participants in Stanton 2. The joint participants are Orlando Utilities Commission (OUC), Florida Municipal Power Agency (FMPA), and Kissimmee Utility Authority (KUA). Volumes 1A through 1D contain the following information.

- 1A--NFP Information Common to All Participants.
- 1B--NFP Information Specific to OUC.
- 1C--NFP Information Specific to FMPA.
- 1D--NFP Information Specific to KUA.

Appropriate appendices are included at the end of each volume. The last volume (Volume 2) contains Chapters 2.0 through 10.0 relating to all aspects of the Supplemental Site Certification Application other than the need for power.

APPLICANT INFORMATION

Applicants' Official Names and Mailing Addresses

Orlando Utilities Commission
P.O. Box 3193
Orlando, Florida 32802

Florida Municipal Power Agency
7201 Lake Ellenor Drive
Orlando, Florida 32809

Kissimmee Utility Authority
P.O. Box 423219
Kissimmee, Florida 34742-3219

Address of Official Headquarters

Orlando Utilities Commission
500 South Orange Avenue
Orlando, Florida 32801

Florida Municipal Power Agency
7201 Lake Ellenor Drive
Orlando, Florida 32809

Kissimmee Utility Authority
8 Broadway
Kissimmee, Florida 34741

Business Entity

Orlando Utilities Commission (OUC) is a statutory commission created by the legislature of the State of Florida as a separate part of the government of the City of Orlando. OUC has the full authority over the management and control of the electric light and water works parts of the City of Orlando. It has the power to undertake, among other things, the construction, operation, and maintenance of electric generation, transmission and distribution systems, and water production, transmission and distribution systems in order to meet the requirements of its customers.

Florida Municipal Power Agency (FMPA) is a joint agency formed pursuant to the Interlocal Cooperation Act and exercises powers under the Joint Power Act. FMPA has authority to undertake and finance electric projects and, among other things, to plan, finance, acquire, construct, reconstruct, own, lease, operate, maintain, repair, improve, extend, or otherwise participate jointly in those projects and to issue bonds or bond anticipation notes for the purpose of financing or refinancing the costs of such projects.

Kissimmee Utility Authority (KUA) is a public body, corporate and politic, duly organized, and legally existing as part of the government of the City of Kissimmee engaged in the generation, transmission, and distribution of electric power to persons within the service area.

Name and Titles of Chief Executive Officers

Orlando Utilities Commission

Jerry Chicone, Jr.	-	President
Royce B. Waldon	-	First Vice President
Susan T. McCaskill Little	-	Second Vice President
James H. Pugh, Jr.	-	Past President
Bill Frederick	-	Mayor
Theodore C. Pope	-	Executive Vice President and General Manager
Thomas B. Tart	-	General Counsel

Florida Municipal Power Agency

Dean G. Shaw	-	Chairman
Joseph M. Tardugno, Jr.	-	Vice Chairman
Harry M. Schindehette	-	Secretary - Treasurer
Vincente R. Ruano	-	Assistant Secretary - Treasurer

Kissimmee Utility Authority

Richard L. Hord	-	Chairman
Bob Bobroff	-	Director
Harry Lowenstein	-	Director
Arnold W. Jones	-	Director
George A. Gant	-	Director
John B. Pollet	-	Mayor (Executive Officio)

<u>Name, Address, and Telephone Number of Official Representative Responsible for Obtaining Certification</u>

Orlando Utilities Commission
Thomas B. Tart, General Counsel
500 South Orange Avenue
Orlando, Florida 32801
407-423-9123

<u>Site Location (County)</u>

Orange County

<u>Nearest Incorporated City</u>

Orlando, Florida

<u>Latitude and Longitude</u>

29° 29' North Latitude
81° 10' West Longitude

<u>UTMs (Center of Site)</u>

Northerly 1507528
Easterly 446825

<u>Section, Township, Range</u>

Sections 13, 14 (E 1/2), 23 (E 1/2), 24; Township T23S; Range R31E
Sections 18, 19; Township T23S; Range R32E

Location of Any Directly Associated Transmission Facilities (Counties)

Orange County

Nameplate Generating Capacity

Stanton 2 will have a nameplate gross generating capacity of 465 megawatts (MW) and is scheduled for commercial operation January 1, 1997.

Capacity of Proposed Additions and Ultimate Site Capacity

Stanton 3 and 4 have not been sized or scheduled at this time. The Stanton site was previously certified for an ultimate site capacity of approximately 2,000 MW.

1A.1.0 Overview and Summary

1A.1.1 Overview

~~Stanton Energy Center Unit 2 (Stanton 2) will be the second unit installed at the previously certified Stanton Energy Center site located approximately 14 miles southeast of Orlando, Florida.~~ The Stanton Energy Center site was certified for an ultimate capacity of approximately ~~2,000 MW based on four coal fired units. Stanton 2 is being planned as a replication of Stanton 1, with a net generating capacity of approximately 440 MW.~~

Both Stanton 1 and Stanton 2 will be jointly owned by Orlando Utilities Commission (OUC), the Florida Municipal Power Agency (FMPA), and Kissimmee Utility Authority (KUA). The existing ownership percentages of Stanton 1 and the proposed ownership percentages of Stanton 2 are shown below.

Owner	Stanton 1 percent	Stanton 2 percent
OUC	68.5542	75.0000
FMPA	26.6265	21.1686
KUA	4.8193	3.8314

Stanton 1 went into commercial operation July 1, 1987. Stanton 2 is scheduled for commercial operation on January 1, 1997.

1A.1.2 Summary

Stanton 2 is planned as a replication of Stanton 1, with an estimated 1997 installed capital cost of $522,000,000 excluding interest during construction, or $1,186/kW based on a 440 MW rating. The performance of Stanton 1 has been exceptional. The performance for Stanton 2 as a replicated unit is expected to be similar or better because of improvements incorporated. The replication process not only offers the opportunity to obtain a second unit with the same high level of performance, but also allows for significant capital cost savings. A number of common facilities were installed with Stanton 1 for Stanton 2. The ability to use these existing common facilities further reduces capital costs.

Stanton 1 has an average equivalent availability factor of over 84 percent for the first three years of operation, exceeding the design target of 83 percent. Operation and maintenance costs have been below average compared to similar units in the country, while the unit has achieved an actual full load net heat rate of 9,740 Btu/kWh.

Stanton 1 has also been a superb performer from an environmental perspective. Stanton 2 will incorporate further technological advancements such as low NO_x burners to further reduce emissions. The units will have complementary air quality control equipment which includes wet limestone scrubbers, electrostatic precipitators, and natural draft cooling towers. The units will utilize treated sewage effluent for cooling water. Stanton 1 consumes approximately 4.5 mgd of treated sewage water, helping to solve a critical environmental disposal problem for the area and perserve diminishing supplies of potable water. Stanton Energy Center has no offsite wastewater disposal. All ash and scrubber sludge are permanently disposed of onsite with fly ash being sold as an additive for cement.

The Stanton Energy Center site comprises 3,280 acres, providing an adequate buffer from the public while serving as a sanctuary for the endangered red-cockaded woodpecker. The construction of Stanton 2 will not do any damage to this sanctuary.

Since Stanton 2 is a replication of Stanton 1, the design is well defined and the projected capital costs are well known. Many of the major equipment contracts have been negotiated and signed pending approval of this application. The contracts cover more than 30 percent of the projected capital cost.

Stanton 2 will use the associated facilities constructed with Stanton 1. Stanton 2 will include a 14-mile 230 kV transmission line segment and a 1-mile alternate access road to the site both of which are to be located in the railroad corridor previously certified for Stanton Energy Center.

Subsection 1A.4.0 addresses the need for additional capacity in Peninsular Florida. The analyses indicate that Peninsular Florida will need over 1,700 MW of additional capacity in 1996 and 1997 above that which was shown to be needed through 1995 by the 1989 Planning Hearing. Furthermore, Stanton 2's low price coal fueled energy will be badly needed by 1997 to reduce consumption of higher priced natural gas and oil.

The following summarizes the Peninsular Florida winter capacity mix as of January 1, 1990, as reported in the Southeastern Electric Reliability Council Coordinated Bulk Power Supply Program IE-411, April 1, 1990.

	MW	Percent
Nuclear	3,920	12.6
Fossil Steam-Oil	7,822	25.1
Fossil Steam-Gas	5,270	16.9
Fossil Steam-Coal	8,467	27.2
Internal Combustion	287	0.9
Combustion Turbine	4,735	15.2
Combined Cycle	559	1.8
Hydro	47	0.2
	31,116	100.0

Nuclear and coal capacity comprise less than 40 percent of Peninisular Florida's capacity.

The remainder of the subsections in this volume provide details of the Stanton 2 project, economic projections including fuel price projections, discussions of the availability of fuel and transportation, alternatives to the construction for Stanton 2, and load flow analysis indicating the need for the additional 14-mile transmission line segment.

1A.2.0 Description of Project

1A.2.1 Description of Facilities

Stanton Energy Center Unit 2 will be a replicate and sister unit to Stanton Energy Center Unit 1, which went into commercial operation on July 1, 1987. Stanton 2 will be a net 440 MW electrical generating station owned by the Orlando Utilities Commission, the Florida Municipal Power Agency, and the Kissimmee Utility Authority. The new unit, like Stanton 1, will be operated by the Orlando Utilities Commission. The Stanton Energy Center is located in Orange County approximately 14 miles east-southeast of the City of Orlando, Florida, on a site which is conceptually designed as a four-unit facility and certified for an ultimate capacity of approximately 2,000 MW. Stanton 2 will be the second unit on this site and will be located north of Stanton 1. Stanton 2 is scheduled for commercial operation on January 1, 1997.

The Stanton Energy Center site encompasses approximately 3,280 acres of Orange County, Florida. Main site access is from the north off State Highway 50 and Alafaya Trail Road. Coal train access is from the south by a spur of the CSX Railroad. Power transmission lines leave the site to the north. The power is transmitted at 230 kV. More details of the site are discussed in Section 2 of the original and Supplemental Site Certification Applications.

Stanton Energy Center Unit 2 will comprise five areas: the powerblock, air quality control, water management, ash and sludge conditioning, and coal handling and storage. The water management area, ash and sludge conditioning area, and the coal handling and storage area are shared with Stanton 1.

1A.2.1.1 Powerblock

The powerblock includes the turbine area, the steam generator area, the Control Center Building, and the Administration and Plant Services Building. The turbine area is a 100-foot wide by 244-foot long building which encloses the turbine generator and related equipment. The turbine area roof is at Elevation 179 feet 6 inches, and the ground elevation at the powerblock is 82 feet. The building and enclosed steam turbine and generator will be positioned east of the steam generator with the turbine generator center line running north to south. In addition to the turbine generator, the turbine area equipment will include the surface condenser and the steam turbine-driven boiler feed pump and other auxiliary equipment items.

The steam generator area will be located west of and adjacent to the turbine area. The area will be approximately 160 feet wide by 224 feet long. The roof elevation will be 309 feet 0 inch. Major equipment located in the area will be the steam generator, deaerator, and storage tank, feedwater heaters, coal pulverizers, coal feeders and coal silos, air heater, and primary air and forced draft fans.

The Stanton 2 control center area will be an extension of the Stanton 1 Control Center Building located near the southeast corner of the steam generator area and beside the turbine area. The Stanton 2 extension area will be 28 feet wide by 100 feet long and will include four internal levels and a roof at Elevation 163 feet 6 inches. The Control Center Building will be completely enclosed.

The existing Administration and Plant Services Building is a common facility with Stanton 1 and is south of the Stanton 1 turbine area and adjacent to the southeast corner of the Stanton 1 steam generator. The Administration and Plant Services Building is approximately 156 feet wide by 208 feet long and has four internal floors. The ground floor includes maintenance shops, a machine shop, men's and women's locker rooms, and a storage room. The assembly floor provides meeting rooms and maintenance material storage. Office areas, a training room, classroom facilities, conference areas, and storage rooms are located on the top two floors of the building. A small addition to the plant services area is contemplated for the increased shop and maintenance personnel required for Stanton 2.

1A.2.1.2 Air Quality Control Area

The air quality control area will include the precipitator and Air Quality Control Building and equipment. The electrostatic precipitator will be located directly west of the steam generator area. The electrostatic precipitator will be 144 feet long by 176 feet wide and will be partially enclosed. A precipitator penthouse will be located above the precipitator.

The Air Quality Control Building will be located directly west of the electrostatic precipitator. The building will be 210 feet wide by 169 feet long and will be completely enclosed with a roof at Elevation 187 feet 0 inch. The Air Quality Control Building will enclose the induced draft fans and the scrubber equipment. The plant chimney will be located directly west of the Air Quality

Control Building. The chimney will be 550 feet tall; the same height as the chimney for Stanton 1.

1A.2.1.3 Water Management Area

The existing water management area is located south of the Stanton 1 steam generator, the Stanton 1 electrostatic precipitator, and the Stanton 2 Air Quality Control Building. The area includes the Water Management Building, the storage tanks, and neutralization basin.

The Water Management Building is approximately 72 feet wide by 126 feet long. It encloses the plant demineralizer equipment, a solids contact unit and equipment, six service water gravity filter cells and clearwell, and the chlorination equipment. Service water storage and demineralized water storage tanks are located west of the Water Management Building.

1A.2.1.4 Ash and Sludge Conditioning Area

The existing ash and sludge conditioning area is located west of the Stanton 1 air quality control area and east of the coal storage areas. The area includes a Sludge Conditioning Building, thickeners, sludge normal and emergency stockout facilities, ash settling tanks, and ash dewatering tanks. Additional equipment will be added to expand the system capabilities to process the Stanton 2 ash and sludge. The processing of scrubber effluent and fly ash takes place in this area. The Sludge Conditioning Building is approximately 104 feet wide by 186 feet long. The Sludge Conditioning Building includes the lime preparation and storage equipment, the sludge conditioning vacuum filters and equipment, and the ash conveying and mixing equipment.

1A.2.1.5 Coal Handling and Storage Area

The existing coal handling and storage area is designed for expansion for Stanton 2. The area is located along the west side of the plant site. The area includes an existing coal car unloading building, a coal transfer structure, an emergency stockout and reclaim hopper, a coal storage area with a track mounted stacker-reclaimer, a coal crusher building, and the interconnecting conveyors required to transport the coal all installed with Stanton 1.

The coal storage area is an open grade level storage system with storage on both sides of a stacker-reclaimer unit. The stacker-reclaimer operates from rails

Water is reserved onsite for fire protection purposes. The service water storage tanks are each provided with an internal standpipe to reserve the lower half of the tanks for fire protection water. A network of yard mains and hydrants provides fire water to the site. Three fire water pumps located in the Water Storage Pump Building provide the pressure and flow required.

The site is designed for zero wastewater discharge. All plant wastewater, including cooling tower blowdown, is reused or treated and no wastewater is discharged off the plant site; therefore, there is no impact to the Econlockhatchee River Basin.

1A.2.1.8 Coal and Oil Supply

The Stanton Energy Center is designed to burn various Appalachian and Illinois Basin coals delivered by unit train. The unloading of coal cars takes place while the train is moving. The system is sized to unload 3,500 tons per hour.

Fuel oil is delivered to the site by truck. No. 6 oil is used for steam generator ignitors during startup and for low load stabilization. No. 2 fuel oil is used for firing the auxiliary boiler and for mobile yard equipment. Fuel oil is stored in aboveground tanks located near the northwest corner of the recycle basin. A special berm containment is provided around the tanks. The containment area has an impermeable liner.

1A.2.1.9 Steam Cycle

The steam cycle will include a natural circulation steam generator which at maximum continuous rating supplies 3,305,000 lb per hour of superheated steam at 2,640 psig and 1,005 F. The steam from the steam generator will be supplied to a condensing type turbine consisting of a high-pressure section, intermediate-pressure section, and a low-pressure section. All three turbine sections will be in line and form a common shaft to drive the electrical generator. Exhaust from the high-pressure turbine will be returned to the reheat section of the steam generator before entering the intermediate-pressure turbine. The reheat steam flow at maximum continuous rating is 2,802,000 lb per hour at 645 psig and 1,005 F. The intermediate-pressure turbine will exhaust directly to the low-pressure turbine. The low-pressure turbine will be a double flow unit of six-stage design. The low-pressure turbine will exhaust to the condenser.

Extraction steam for regenerative feedwater heating will be provided from all three turbine sections. Feedwater Heater 8 will receive steam from an interstage extraction on the high-pressure turbine. Feedwater Heater 7 will receive steam from the high-pressure turbine exhaust (cold reheat). An interstage extraction from the intermediate-pressure turbine will provide steam to the deaerator (Feedwater Heater 5) and the boiler feed pump drive turbine. Extraction steam will be cascaded through the feedwater heater train to the deaerator and the condenser.

The condenser will reject cycle waste heat to the circulating water system, reduce the low-pressure turbine back pressure, and hold cycle water (condensation). The condenser will be of the single shell, two-pass divided water box construction.

Two full capacity condensate pumps will be provided and take suction from the condenser hot well and supply condensate quality water through the low-pressure heater train to the deaerator. A full capacity condensate polisher will be provided.

The deaerator will be the last heater in the low-pressure heater train. It will receive the heated condensate and remove air and other noncondensables. The deaerator will discharge to the deaerator storage tank. The deaerator storage tank will contain approximately 55,000 gallons of condensate available to the boiler feed pump and feedwater system. One full capacity, turbine driven boiler feed pump will take suction from the deaerator storage tank and discharge through the high-pressure heater train to the steam generator economizer inlet connection. The high-pressure feedwater train will include three levels of regenerative heating.

The turbine driven boiler feed pump will include a booster pump, also driven by the turbine, that assures adequate suction pressure to the main boiler feed pump during sudden load changes. The boiler feed pump turbine can use steam from either the main steam source or from the intermediate-pressure turbine exhaust. An electric motor driven startup boiler feed pump is used when steam is not available to the boiler feed pump turbine. The startup boiler feed pump will be sized to allow operation of Stanton 2 at 60 percent capacity if the main turbine driven boiler feed pump is unavailable.

1A.2.1.10 Steam Generator Equipment and Auxiliaries

The steam generator will be a Babcock & Wilcox, indoor drum type, balanced draft, parallel gas flow convection pass radiant boiler designed to burn pulverized bituminous coal from the Southern Appalachian and Illinois basins. It will also be designed to burn high fouling and high slagging coal to maintain fuel flexibility. The steam generator will have a guaranteed efficiency of 89.07 percent at 100 percent load firing the Appalachian design coal.

The maximum continuous rating (MCR) for this unit will be the guarantee load of 3,305,000 lb/h of main superheated steam at 2,640 psig and 1,005 F. This represents the 5 percent overpressure and 11 percent overflow condition for the steam turbine. The reheat steam conditions leaving this boiler at MCR will be 2,802,000 lb/h at 645 psig and 1,005 F.

The steam generator will be equipped with Babcock & Wilcox dual register Type XCL low NO_x burners with a guaranteed NO_x emission rate of 0.32 pound NO_x/MBtu, which represents over a 45 percent decrease below the current Prevention of Significant Deterioration (PSD) permit emission rates for the unit of 0.60 pound NO_x/MBtu.

1A.2.1.10.1 Combustion Air Fans. Two TLT Babcock radial flow fans will supply the primary air for the steam generator. Each fan will be capable of supplying the air requirements up to approximately 70 percent of MCR. Above MCR, both fans must be in service. These fans will take atmospheric suction through inlet silencers. Flow will be controlled by inlet vanes. Each fan will have a test block rating of 291,181 acfm when driven by a single-speed electric motor at 1,180 rpm.

Two forced draft fans of the double width, airfoil-bladed centrifugal type will provide secondary air to the furnace for combustion. The fans will take suction from atmosphere through inlet silencers. Each forced draft fan will be rated for 482,200 acfm at 17.7 inches wg test block. The forced draft fans will be driven by two-speed motors at either 711 rpm or 594 rpm.

Two induced draft fans of the double width, airfoil-bladed centrifugal type with double inlet will provide transport of the flue gas from the furnace through the air quality control equipment and to the chimney. The induced draft fans will be driven by two-speed motors at either 885 rpm or 705 rpm.

The forced draft and induced draft fans will be sized so that two fans operating at low speed are adequate to maintain boiler maximum continuous rated load point during operation with a clean air heater and the boiler unfouled by slag in the steam generator.

1A.2.1.10.2 Air Preheater. A fin tube type air preheater coil will be located in the ductwork between the forced draft fan discharge and the regenerative air heater inlet. The preheater will be designed to protect the air heater against corrosion by maintaining a minimum cold end average temperature. The coils will use condensing auxiliary steam for supply of heat. Condensate will be returned to the steam cycle at the flash tank.

1A.2.1.10.3 Air Heater. The air heater will be a Ljungstrom trisector regenerative type with vertical shaft. The air heater will normally be driven by either of two full capacity, 100-horsepower electric motors. An air driven adjustable speed motor is provided for maintenance purposes only. The air heater will be equipped with both soot blowing and water wash cleaning devices.

1A.2.1.10.4 Soot Blowing. The steam generator is provided with gas side soot blowers to remove ash and slag buildup, thereby improving heat transfer. The system will include 46 retractable soot blowers, 54 furnace wall soot blowers, and two air heater soot blowers. The soot blowers will use steam from the platen superheater outlet. Steam for the air heater will also be available from the plant auxiliary steam system.

1A.2.1.11 Turbine Equipment and Auxiliaries

The turbine will be a Westinghouse tandem compound (34-inch last-stage blades), two-flow exhaust turbine consisting of two primary sections: a combined single-flow high-pressure (HP) and single-flow intermediate-pressure (IP) section, and double-flow low-pressure (LP) section. The turbine will be designed to operate at 3,600 rpm and will be guaranteed to produce 427,420 kW when operating with a 2,400 psig throttle pressure at 1,000 F, reheating to 1,000 F, 3.0 inches Hg absolute condenser pressure and zero makeup. The 5 percent overpressure rating will be approximately 460 MW. The guaranteed net turbine heat rate at design condition will not exceed 7,834 Btu/kWh. The turbine will also be designed to provide extraction heating steam to four low-pressure feedwater heaters from the LP turbine; to a deaerator, boiler feed pump turbine, and air preheat coils from the IP turbine exhaust; to one high-pressure feedwater heater from the IP turbine; to one high-pressure feedwater heater from an intermediate stage of the HP turbine; and to one high-pressure feedwater heater from the HP turbine exhaust.

1A.2.1.11.1 Surface Condenser. The LP turbine will exhaust into a bottom mounted surface condenser. The surface condenser will act to maintain a reduced back pressure on the turbine and to remove waste heat from the turbine exhaust. The condenser will be a single shell with divided water box arranged for two passes on the tube side. The condenser tubes will be 1.0 inch in diameter, 22 BWG, ASTM 249, Type 316 stainless steel. The surface condenser will have a total effective surface area of 270,000 square feet. The condenser hot well will have a storage capacity of 25,000 gallons.

1A.2.1.11.2 Low-Pressure Feedwater Heaters. Four stages of LP feedwater heating will be employed. All heating steam will be provided by extraction from the LP turbine. The LP Heater No. 1 and 2 and all extractions from the turbine to the heater shell will be provided with power-assisted nonreturn check valves and motor-operated block valves to protect the turbine from accidental water induction. Heater drains will cascade from No. 4 to No. 3 to No. 2 and then to the flash tank. Steam from the flash tank enters Heater No. 1, and liquid drains will be discharged to the surface condenser through the drains cooler.

1A.2.1.11.3 Deaerator and Storage Tank. The deaerator will be a horizontal mounted cylindrical feedwater heater in which condensate makes direct contact with extraction steam from the intermediate-pressure turbine exhaust. The direct contact of condensate with steam will cause heating of the condensate and separation of the dissolved gasses. The deaerator will be open to atmosphere to permit venting of air and other noncondensables. The deaerator will have an integral vent condenser to reduce loss of steam. The deaerator will be designed for a maximum operating pressure of 270 psig and a maximum design temperature of 650 F. The deaerator will be approximately 8 feet in diameter and 35 feet long.

The deaerator condensate will drain into a bottom-mounted deaerator storage tank. The storage tank will provide condensate to the boiler feed pumps. The storage tank will be approximately 12 feet in diameter, 76 feet long, and have a normal operating storage volume of 55,000 gallons.

1A.2.1.11.4 High-Pressure Feedwater Heaters. Three stages of HP feedwater heating will be employed. Heating steam will be provided from the intermediate- and high-pressure turbine stages. An extraction from the IP turbine will provide steam to Feedwater Heater 6. The HP turbine exhaust (cold reheat) will provide steam to Feedwater Heater 7. An interstage extraction from the HP turbine will provide steam to Feedwater Heater 8. All extractions from the turbines to the high-pressure feedwater heaters will be provided with power-assisted nonreturn check valves and motor-operated block valves to protect the turbine from accidental water induction and turbine overspeed. The shell side drains will cascade from No. 8 to No. 7 to No. 6 and then to the deaerator. A feedwater heater drain pump will be provided to transfer drains to the deaerator below 63 percent load when adequate pressure does not exist. Upon high water level, all heaters will drain directly to the surface condenser.

1A.2.1.11.5 Boiler Feed Pump and Turbine. The boiler feed pump will be a Byron Jackson, four-stage horizontal, double suction, double case barrel, centrifugal, high-speed type designed for continuous operation in boiler feed service. The boiler feed pump will take suction from the booster boiler feed pump discharge and provide the required pressure at the steam generator economizer inlet. The boiler feed pump will be rated at 8,300 gpm and 7,750 foot, total head, at a speed of 5,800 rpm.

The boiler feed pump will be directly connected to a General Electric condensing, nonextraction steam driven boiler feed pump with bottom exhaust, seven impulse stages, tilting pad thrust bearing, elliptical main bearings, and labyrinth shaft seals. The boiler feed pump turbine will be arranged for connection of the exhaust end shaft extension to the boiler feed pump.

The boiler feed pump turbine will be designed for the steam conditions existing at the IP turbine to LP turbine crossover stage extraction of the turbine generator unit. The boiler feed pump turbine will also be designed for startup and low load operation when steam is supplied from the main steam feed to the main turbine.

The booster boiler feed pump will be a horizontal, double suction, single stage, centerline-mounted centrifugal pump with a radial split casing. The booster boiler feed pump will be driven by the boiler feed pump turbine through a speed reducing gear. The pump will be designed for continuous operation in boiler feed service.

The startup boiler feed pump will be a Byron Jackson, five-stage, diffuser type centrifugal pump. It is connected to a 7,000 horsepower, 1,800 rpm induction motor through a Voith adjustable variable-speed drive with integral speed increase gearing.

1A.2.1.12 Generator and Electrical Equipment

The generator will be a Westinghouse 24 kV, 13,000 ampere, 516,200 kVA, 2-pole, 3-phase, 60 hertz, hydrogen cooled, 3,600 rpm synchronous machine. The generator rating will be based on a minimum 0.9 power factor and a 0.58 short-circuit ratio. Maximum hydrogen cooling pressure will be 60 psig. Generator excitation will be provided by a main exciter alternator, permanent magnet pilot exciter, and fuse/diode wheel, all on a common shaft coupled to the generator field. Excitation current will be transmitted from the main exciter armature to a fuse/diode wheel and then to the generator field by means of conductors internal to the shaft. Voltage regulation will be accomplished by controlling the main exciter field current. The voltage regulator will employ static components.

Electrical power will be conducted from the generator terminals by a 3-phase self-cooled isolated phase bus duct to the main generator transformer. One generator phase will be conveyed on each bus duct. Each bus will consist of a 19-inch diameter aluminum conductor inside a 39-inch diameter enclosure. The enclosures will be spaced 54 inches center to center.

The generator transformer will be a three-phase, 60 hertz, two-winding, delta-wye connected unit. The generator transformer will provide step-up of the 24 kV generator voltage output to 230 kV for connection to the electrical transmission and distribution system. The transformer will be rated 537,600 kVA at maximum 65 C rise and 480,000 kVA at maximum 55 C rise.

1A.2.1.13 Air Quality Control Equipment

The Stanton 2 air quality control will be the subject of a Best Available Control Technology (BACT) demonstration during the amendment of the Stanton 2 PSD permit issued by the Environmental Protection Agency (EPA) in 1982. The proposed equipment will include a cold-side electrostatic precipitator, followed by a flue gas scrubber, and then dispersion by a chimney. The electrostatic precipitator will receive flue gas directly from the steam generator and remove particulate prior to forwarding the gas to the flue gas scrubbers. The scrubbers will function to remove sulfur dioxide by direct contact with an acid neutralizing slurry. The flue gas scrubbers additive system will prepare a slurry of limestone for the scrubbers.

1A.2.1.13.1 Electrostatic Precipitator. One full capacity cold-side electrostatic precipitator will be provided to remove fly ash from the combustion gas and reduce particulate emissions from the unit to comply with a proposed BACT limit of 0.020 lb/MBtu and a plume opacity of less than 10 percent. This emission rate is a 33 percent reduction form the current PSD permit emission rate. Particulate removal will be accomplished by a dc high voltage electrical field between high voltage discharge electrodes and ground collecting electrodes. Particles entering the electrical field will acquire an electrostatic charge which will cause them to migrate across the gas flow path to the collecting electrodes where the charge on the particle normally leaks to ground. The collected particles will be removed by periodic rapping of the electrodes. The particles will fall into hoppers below the combustion gas flow path and will be periodically removed by the fly ash system.

The electrostatic precipitator will be a Wheelabrator size 168/45/6 x 9/12/3 rigid frame type complete with all electrical and control equipment necessary for a complete operating unit. The precipitator will be provided with a weather enclosure completely covering the top of the precipitator and all externally top mounted equipment. The precipitator will be approximately 175 feet wide across the inlet, 116 feet deep, and 55 feet high.

The precipitator hoppers located on the bottom of the precipitator will collect the dust which falls from the collection system. There will be a total of 48 precipitator hoppers, each with a storage volume of 2,831 cubic feet. Each precipitator hopper will be provided with a hopper heating system.

The fly ash system will convey by vacuum transport fly ash collected in the precipitator hoppers, the economizer hoppers, and the air heater hoppers to the existing fly ash silo installed with Stanton 1.

Two parallel crosstied ash collecting systems will each serve 24 precipitator hoppers. Either system can service the entire fly ash unit by conveying continuously.

The mixture of ash and air will be conveyed to the fly ash separators where the heavier ash will be deposited into the silo. The conveying air will be passed through the fly ash exhauster filter to further clean it of the lighter particles. The conveying air will be passed through the exhausters to atmosphere.

The fly ash silo is equipped with fluidizing air blowers. These blowers will fluidize the fly ash and aid in ash removal.

The fly ash will normally be discharged to the scrubber solids system for disposal. The silo is also equipped with a rotary unloader to allow ash unloading into trucks.

The fly ash silo is 40 feet in diameter and 53 feet high. It has a net storage volume of 54,200 cubic feet. The silo has an approximate 72-hour collection capacity at the maximum fly ash collection rate.

1A.2.1.13.2 Flue Gas Scrubber. Sulfur dioxide (SO_2) will be removed from the steam generator flue gas by a direct contact spray type desulfurization system. The system will be designed and furnished by ABB Combustion Engineering Systems. The desulfurization system will comprise three ABB spray tower absorber modules and reaction tanks. The sizing of the modules will be based on two operating and one spare. The system will receive almost particulate-free flue gas from the electrostatic precipitator. As the gas enters the operating absorbers, it will be sprayed counter-currently with the slurry from the associated reaction tanks. Spray nozzles will disperse the slurry into fine droplets covering the entire cross-sectional area of the absorber, thus affording intimate contact between gas and liquid and effective absorption of SO_2 into the liquid phase.

After contacting the sprays, the cleaned gas temperature will be approximately 125 F, will be saturated with water, and will carry some moisture droplets. The removal of most of this entrained moisture will be accomplished in the demister section of the spray tower which will contain the bulk entrainment separator vanes and two series of chevron mist eliminator vanes. Upon exiting the mist eliminator, the treated flue gas will be directed to the chimney.

As the slurry falls to the bottom of the spray tower, it will absorb SO_2 from the flue gas. The slurry will drain from the tower into the reaction tank where the formation of insoluble calcium solids is promoted. Ground limestone additive will be added to the reaction tank under agitation by one top mounted Lightnin mixer.

1A.2.1.13.3 Reaction Tank. Each of the three reaction tanks will have a diameter of 57 feet, a height of 34 feet 6 inches, and an operating capacity of approximately 620,000 gallons, and will serve a single absorber module. The bottom of the absorber shell will hang in the reaction tank and will make a gastight seal with the slurry in the tank.

1A.2.1.15.3 Fly Ash and Lime Additive Subsystem. The fly ash and lime additive subsystem delivers fly ash and crushed quicklime to the stabilization mixers. Each must be present in the proper proportions and combined with the filter cake to harden the final product into a stabilized material in the landfill.

The system will be made up of three independent trains to handle both Stanton 1 and 2 scrubber solids. Each train consists of the following major components: a lime feeder, a lime crusher, a lime crusher exhaust fan, a fly ash feeder, a fly ash impact weighing element, screw conveyors, and slide gates needed to combine and transfer the fly ash and lime.

1A.2.1.15.4 Conveyor and Stabilization Mixer Subsystem. The conveyor and stabilization mixer subsystem will carry the dewatered filter cake from the vacuum filtration area to the mixer, where fly ash and lime will be mixed with the filter cake.

The subsystem will be made up of new, existing, and expanded equipment, including vacuum filter discharge conveyors, stabilization mixers and their associated dust collectors, stabilization mixer discharge conveyors, sludge transfer conveyors and their associated scales, a fixed stacker, and a radial stacker.

The processed material from the mixers will be discharged onto the mixer discharge conveyors. The mixer discharge conveyors will discharge to the sludge transfer conveyors. The material will then be fed to either the radial stacker or the fixed stacker for stockpiling of the end product. From the stockpiling area, the material will be loaded onto trucks for transport to the landfill area located onsite. The landfill area is approximately 312 acres. About 80 acres will be required for Stanton 2 combustion waste storage. The landfill area will accommodate four units for the total life of each unit.

1A.2.1.16 Associated Facilities

Offsite associated facilities for Stanton 2 include existing facilities constructed for Stanton 1, which will be used in the operation of both units, and a new 230 kV transmission line segment and a new alternate site access road. The existing facilities include the site access road, the cooling water makeup supply line, the railroad spur serving the site, and the transmission facilities which interconnect the site with the OUC transmission system. The new transmission line segment will be used to complete the interconnection of the site with the

1A.2.2 Fuel Supply

The coal for Stanton 1 is currently obtained from Blue Diamond Coal Company and delivered to the Stanton Energy Center by CSX Transportation. The contract with CSX Transportation has provisions to allow for the transportation of coal for Stanton 2. Subsections 1A.3.3 and 1A.3.4 discuss the availability of coal and transportation in detail.

OUC plans to seek bids for coal for Stanton 2 in the 1993 to 1994 time frame depending on market conditions. OUC intends to obtain two firm contracts covering the majority of the coal requirements while also allowing for some spot market purchases. OUC intends to obtain flexibility in the amount of coal to be delivered for each of the firm contracts to enable OUC to optimize purchases as market conditions change. OUC intends to simultaneously sign coal supply and transportation contracts after completion of the negotiations.

OUC desires to contract for a low-sulfur coal for Stanton 2 similar to the low-sulfur coal currently being burned in Stanton 1. However, OUC is designing Stanton 2 to burn the same wide range of coals as can be burned in Stanton 1 and is proposing BACT emission levels which would allow medium sulfur coals to be burned. This flexibility to obtain medium sulfur coals will help OUC to negotiate the purchase of low-sulfur coal at a reasonable price.

1A.2.3 Capital Costs

1A.2.3.1 Plant Cost

The capital cost for Stanton 2 for commercial operation on January 1, 1997, is estimated to be $522,000,000. Table 1A.2.3-1 presents the projected monthly cash flow requirements. This cost is based on replicating the Stanton 1 major equipment for Stanton 2. The effects on capital costs of replicating Stanton 1 major equipment for Stanton 2 are as follows.

- The number of spare parts required to be maintained in inventory can be reduced through duplicate equipment. While there will be a need to increase the number of some items of spare parts because of a two-unit demand, a complete set of duplicate spare parts is not required. This reduces capital as well as O&M costs.
- The initial training of operating and maintenance personnel will be reduced. While a limited number of new operators and maintenance personnel will require some training because of technical enhancements and upgrades, the existing personnel will already be familiar with these major items of equipment. This reduces the familiarity period and direct training expense in addition to early maintenance costs which are included in the capital cost of Stanton 2.
- Replicating major items of equipment will result in significant savings in both the design cost for the equipment, duplication in procurement of similar components, and the engineering cost of the buildings and associated facilities.

The decision to replicate Stanton 1 was based on a detailed planning study. The Curtis H. Stanton Energy Center Unit 2 Project Planning Study concluded that a total project savings of approximately $15 million could be realized by replicating the major items of equipment and materials as compared to bidding different designs. In order to confirm this savings, OUC has been receiving proposals and negotiating contracts for the replicated equipment. The contracts contain cancellation clauses resulting in minimum financial risk to OUC if regulatory approval is not obtained and cancellation of Stanton 2 is required. On the basis of proposals received on the equipment being replicated, the savings due to replicated equipment will actually be approximately $23 million.

There are numerous common facilities that Stanton 2 and Stanton 1 are sharing. Some of these facilities include the site access road; railroad; material handling system; makeup water supply pipeline and storage pond; water treatment; ash, scrubber sludge, and wastewater disposal; Site Security Building; Administration Building; Service Building; turbine room crane; emergency generator; fire protection; and coal cars. These common facilities costs were included in the capital cost of Stanton 1 and are not included in the capital cost of Stanton 2.

The estimated capital cost of Stanton 2 includes direct costs, indirect costs, escalation, and sales tax, and an appropriate amount for contingencies. Standard practices of estimating capital costs of coal fired power plants were used in developing the estimated capital cost of Stanton 2. Because Stanton 2 is a replicate of Stanton 1, the preliminary design has been established and decisions have been made regarding major plant and system parameters. Therefore, the estimate is more definitive than estimates normally developed at this stage of a project.

The estimate is developed on the basis of the firm lump sum contract basis that was used for Stanton 1. The detailed estimate for the 121 individual equipment and construction contracts is presented in Table 1A.2.3-2. Proposals have been received, and negotiations on the contract terms and cost have been initiated on the following equipment that is being replicated for Stanton 2. It is anticipated all pricing regarding these proposals will be established by April 1, 1991.

Specification	Description
61.1001	Concrete Chimney
61.4001	Structural Steel, Grating, and Coal Silos
62.0201	Particulate Removal Equipment
62.0202.1	Flue Gas Scrubber
62.0202.2	Sludge Conditioning Equipment
62.0401	Air Compressors
62.0601	Cooling Tower
62.1001	Turbine Generator
62.1801	Ash Handling System
62.2001	Boiler Feed Pump Turbine

Specification	Description
62.2602	Boiler Feed Pumps
62.2603	Circulating Water Pumps
62.3401	Steam Generator
65.0401	Condensate Polishing

The contracts for the above listed replicated equipment represent 30.7 percent of the Stanton 2 estimated cost.

In addition to the major equipment being replicated, preliminary proposals for the following Stanton 2 equipment have been received.

Specification	Description
61.2006	Duct Dampers
62.2615	General Service Pumps
62.3811	Forged Steel Valves
64.1401	Flue Gas Monitoring Equipment
65.0602	Water Quality Control System

The costs of the remaining contracts were estimated on the basis of the Stanton 1 final contract costs adjusted as required for the difference in quantity and scope. Adjustments have been included for costs due to escalation for the power generation industry including labor rates since Stanton 1 through the date of the Stanton 2 cost estimate. All current costs are based on a March 1, 1991, date. These prices are adjusted for escalation at 4.5 percent per year for the commercial operating date of January 1, 1997.

1A.2.3.2 Coal Pile

The space for the inactive coal pile for Stanton Energy Center is sized to store 400,000 tons, which is equal to approximately a 60-day supply for two units. In order to reduce carrying costs, OUC plans to minimize the use, and therefore the size, of the inactive storage pile except for times when there is potential for strikes or supply disruptions. In addition to reduced carrying charges, savings are also achieved through lowering coal handling costs. OUC plans to maintain a minimum supply of 15 to 20 days for Stanton 2 in active storage. A $3.5 million cost for the active coal pile is assumed on the basis of 1996 projected coal costs. For evaluation purposes, the carrying costs for this coal pile are added to the carrying costs for the plant costs.

1A.2.3.3 New 230 kV Transmission Line and Alternate Site Access Road

The capital cost for the new 230 kV transmission line from the Stanton Substation is $9 million in 1991 dollars, as shown in Table 1A.2.3-3. The new 230 kV transmission line is an associated facility of Stanton 2, and will be fully owned by OUC as part of OUC's transmission system. The cost of the new 230 kV transmission line is therefore not part of the Stanton 2 project cost in Subsection 1A.2.3.1.

The alternate site access road is also an associated facility of Stanton 2. The capital cost for the alternate site access road is included in the Stanton 2 capital cost in Subsection 1A.2.3.1.

1A.2.5 Heat Rate

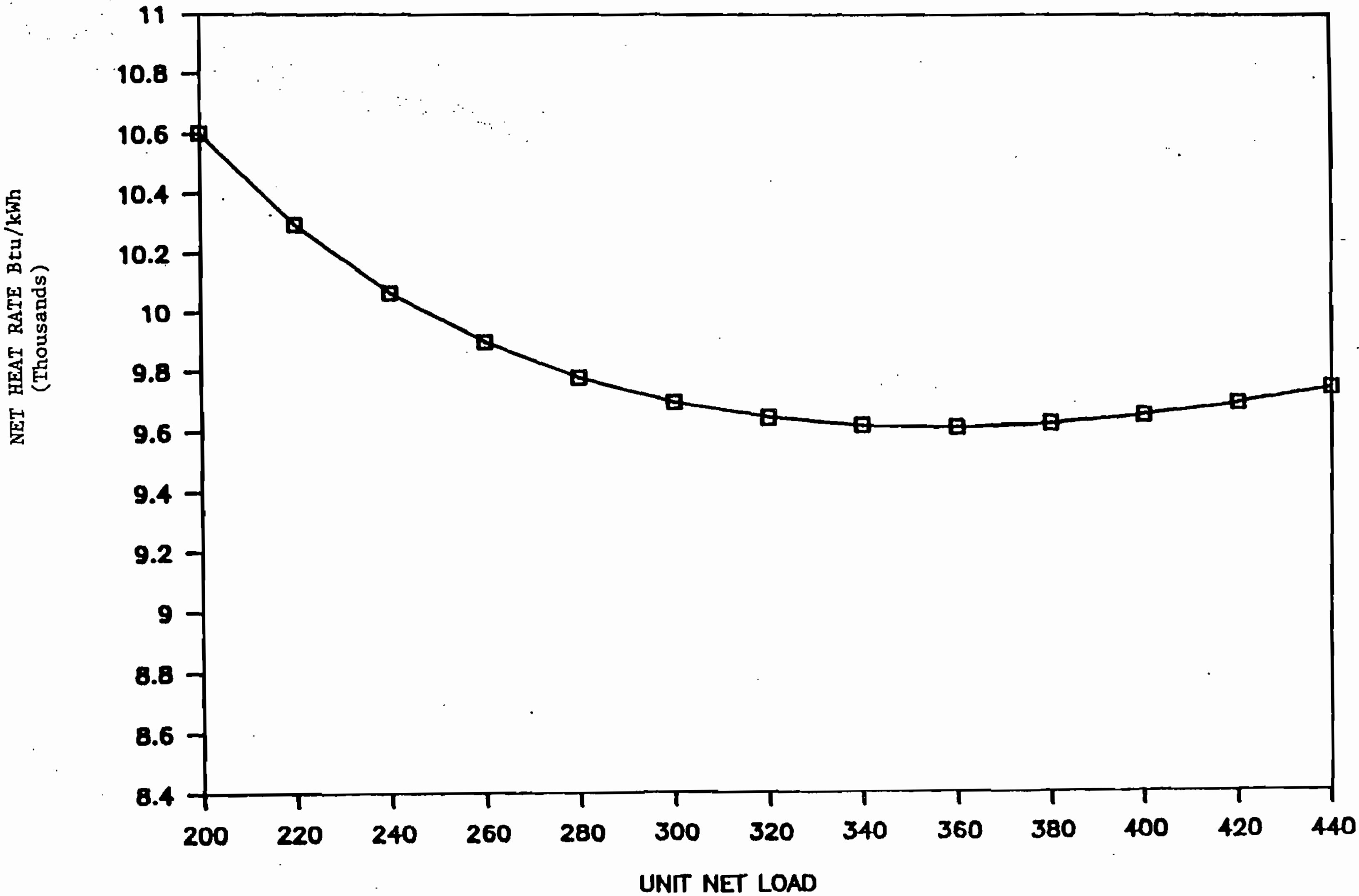
The estimated average net heat rate for Stanton Energy Center Unit 2 is as follows.

Net Load		Net Heat Rate
MW	Percent	Btu/kWh
440	100	9,740
330	75	9,640
220	50	10,310

The data presented for the respective loads were derived from actual data recorded during performance tests of Stanton Energy Center Unit 1. The heat rate curve provided in Figure 1A.2.5-1 was also developed from performance tests of Stanton 1. Since the major equipment from Stanton 1 is being replicated for Stanton 2, it is expected that Stanton 2 will meet, if not surpass, the performance of Stanton 1.

STANTON

NET HEAT RATE vs. NET LOAD



NET HEAT RATE VS NET LOAD

Figure 1A.2.5-1

1A.2.6 Availability

The major equipment of Stanton 1 is being replicated for Stanton 2. One of the major incentives for replicating the Stanton 1 equipment is the reliability demonstrated by Stanton 1. The equivalent availability of Stanton 1 for the first three years of operation was 84.54 percent, and the equivalent forced outage rate was 4.76 percent. Stanton 2 is being designed on the basis that it will achieve an equivalent availability of 83 percent with an equivalent forced outage rate of 4 percent. Maintenance outages are assumed to be three weeks each in the spring and the fall, with every third spring maintenance outage increased to six weeks for a turbine inspection and overhaul.

1A.2.7 Schedule

The schedule for Stanton 2 is based on a 44-month construction period. Engineering would start 25 months before the start of construction to complete the design in advance of construction.With engineering well underway, there would be few unknowns during construction and there would be sufficient time to give consideration to a smaller number of construction contractors. To meet a January 1, 1997, commercial operation date, construction would start May 1, 1993, and engineering would start April 1, 1991.

The 44-month period for construction starts with eight months of foundation construction for the generator building complex. Foundation construction is followed by seven months of steel erection for the building, followed by boiler erection. Boiler erection is a major critical path item. Sixteen months after the start of boiler erection, the boiler should be ready to be hydro tested. This 16-month period is also used to erect the other major mechanical equipment, mechanical piping, and electrical equipment. Boiler hydro testing is followed by 10 months of other plant systems testing before initial operation. Commercial operation is expected to follow initial operation by three months.

The 25-month period for engineering before construction starts includes the time required to procure major equipment and to complete the detailed foundation design. Ten months of procurement is followed by 10 months of detailed design. Five months is allowed to bid, evaluate, and award the first construction package so that foundation construction can be initiated.

This project schedule is based on reasonable durations and a logical, efficient approach to the project. This approach will support completing the project on time with minimum total cost. The detailed schedule is presented in Figure 1A.2.7-1.

1A.3.3 Fuel Price Projections

Forecasts of nuclear and fossil fuels that are likely to be used for generating electricity were developed for the period 1990 to 2020. The forecasts are in real, or constant, 1989 dollars and in nominal dollars using the general inflation rates specified in Subsection 1A.3.2. The following discusses the basis of the price forecasts for coal, oil, natural gas, and nuclear fuels.

Numerous forecasts could be chosen for price projections. For evaluation purposes, the 1990 Annual Energy Outlook by the Energy Information Administration (EIA) has been chosen because it is a recent forecast and it compares favorably with other forecasts. EIA provides a comparison of their current forecast with other major forecasting firms such as DRI/McGraw-Hill, the WEFA Group, Gas Research Institute, and the American Gas Association. A comparison of fossil fuel forecasts from these other forecasters for the year 2010 and the EIA forecast is shown in Table 1A.3.3-1. The forecasts are similar; therefore, the EIA forecast should be representative of what the major forecasters predict for the future price of fossil fuels.

As in any forecast, uncertainties are present; therefore, the 1990 Annual Energy Outlook provides a base case and four other cases to account for these uncertainties. The four cases are low oil price, high oil price, low growth, and high growth, and are based on upper and lower bounds for world oil prices and the rate of economic growth, the key factors affecting energy use and price trends. To provide fuel price projections for most likely, high, and low growth scenarios, the base, low oil price, and high oil price scenarios developed by the EIA were used. High and low oil price scenarios are used instead of the low and high growth scenarios because they represent the highest and lowest price forecasts for oil. The 1990 Annual Energy Outlook forecasts fuel prices to the year 2010. To project fuel prices to the year 2020, the average annual escalation rate from 2005 to 2010 was assumed to continue until the year 2020.

The most likely fuel cost projections based on the 1990 Annual Energy Outlook base case are shown in Tables 1A.3.3-2 and 1A.3.3-3 in 1989 constant dollars and nominal dollars, respectively. The nominal prices shown in Table 1A.3.3-3 are based on applying the projected most likely GNP deflator escalation rates in Subsection 1A.3.2 to the 1989 constant dollar prices shown in Table 1A.3.3-2. Table 1A.3.3-4 indicates the low, average, and high 1990 actual costs for OUC along with the 1990 forecasted price from Table 1A.3.3-3. The extreme volatility of gas and oil prices within a single year points out the difficulty of determining an appropriate starting point for fuel price projections. Because of this large volatility and the fact that transportation costs represent a small portion of the delivered cost for gas and oil, the 1990 Annual Energy Outlook forecasts for the delivered cost of gas and oil are used directly without adjustment since they have been developed to be internally consistent. The following sections discuss the development of the forecast for each fuel.

1A.3.3.1 Coal

The coal being burned in Stanton 1 and McIntosh 3 is being purchased under long-term contracts. The contract for Stanton 1 is with the Blue Diamond Coal Company and is a requirements contract until July 1992. After July 1992, OUC can purchase a minimum of 600,000 tons per year and a maximum of 950,000 tons per year. The actual amount purchased in a given year will be based on the amount nominated by OUC. On an annual basis, OUC can increase or decrease the amount nominated by 75,000 tons over that nominated the previous year. The contract extends to 1997 with two additional five-year extensions at OUC's sole discretion. The contract for McIntosh 3 is a requirements contract. The contract expires in December 1991. OUC's coal contract is for a higher quality coal with significantly lower sulfur content than OUC is permitted to burn and as a result is more expensive than the least expensive coal that could be burned. OUC contracted for this coal to improve unit operation and reliability as well as to reduce SO_2 emissions. On an as-burned basis, the higher quality low-sulfur coal results in lower costs for scrubber additive and waste disposal, which offsets at least a portion of the lower cost of the higher sulfur coal. In addition, the higher quality low-sulfur coal results in higher unit availability, which further offsets the lower cost of the higher sulfur coal. The actual cost and quality of all the coal burned at Stanton 1 and McIntosh 3 during 1989 are shown in Table 1A.3.3-5.

The actual prices paid for Stanton 1 and McIntosh 3 coal are significantly above the average US price for 1989 of $1.46/MBtu from the 1990 Annual Energy Outlook. Part of the difference is accounted for by the higher quality of coal, but other factors contribute to the higher cost of coal for units in Florida. One of the factors is the higher cost of coal transportation to Florida. Another factor is the large amount of low cost western coal in the US average. While the 1990 Annual

Energy Outlook escalation rates are probably appropriate, the starting value must be adjusted to reflect the above factors.

For the low-sulfur coal burned in Stanton 1, the 1989 actual delivered price shown in Table 1A.3.3-5 was used as the starting point with the 1990 Annual Energy Outlook escalation rates applied. For the medium-sulfur coal, the starting point was the 1989 average actual contract delivered price for 1.5 to 2.0 percent sulfur coal to plants in the South Atlantic region as presented in the Cost and Quality of Fuels for Electric Utility Plants 1989. This adjustment essentially accounts for deleting much of the western coal that is included in the national average. A second adjustment is made to the starting point to account for the increase in transportation costs to Florida relative to the South Atlantic region. The adjustment is obtained by taking the ratio of the delivered 1989 cost of coal for Florida compared to the South Atlantic region. The adjustments result in a 1989 medium-sulfur coal cost of $1.67/MBtu to which the 1990 Annual Energy Outlook escalation rates are applied.

The passage of the 1990 Clean Air Act will place greater demands on the high quality low-sulfur coal that OUC is currently burning. As a result, an increase in the price of the coal is expected. Since the 1990 Annual Energy Outlook did not consider the effect of the 1990 Clean Air Act Amendments, a further adjustment for low-sulfur coal is necessary. While there are various opinions regarding the level of the price increase, several estimates have been used to select a consensus price of $1.37/MBtu f.o.b. mine in 2000 in 1989 dollars. These estimates are presented in Table 1A.3.3-6. The forecast of the low-sulfur coal was increased to this $1.37/MBtu f.o.b mine price in 2000, assuming that f.o.b mine costs represent 61 percent of the delivered coal cost based on the actual 1989 OUC f.o.b mine and transportation costs. Though it is likely that there will be a corresponding reduction in the price of medium- and high-sulfur coals due to the Clean Air Act, no reduction has been included in the forecasts.

The flexibility in the design of the Stanton units will allow them to burn the medium-sulfur coal at essentially the same emission levels as being actually achieved with the low-sulfur coal. OUC may, however, continue to burn the lower sulfur coal as long as the premium does not become excessive. With the commercial operation of Stanton 2 in 1997, significantly more coal will be required than is available under the Stanton 1 contract. For evaluation purposes and to be conservative, it will be assumed that OUC will continue to burn the same low-sulfur coal that it is currently burning. It is assumed that McIntosh 3 will burn the medium-sulfur coal projected in Table 1A.3.3-2. In reality, however, the specific escalation provisions in OUC's contract may allow it to be extended for the two 5-year extensions beginning in 1997 at prices below those for the low-sulfur coal in Table 1A.3.3-2.

1A.3.3.2 No. 6 Oil

The most likely forecast of No. 6 oil prices is based on the base end-use price for heavy oil to electric utilities from the 1990 Annual Energy Outlook.

1A.3.3.3 No. 2 Oil

The most likely forecast of No. 2 oil prices is based on the base price of distillate oil from the 1990 Annual Energy Outlook.

1A.3.3.4 Natural Gas

The most likely forecast of natural gas price is based on the base end-use price for electric utilities from the 1990 Annual Energy Outlook. No upward adjustment has been made to reflect the effect of the 1990 Clean Air Act Amendments.

1A.3.3.5 Nuclear

The 1990 Annual Energy Outlook does not forecast prices for nuclear fuel. The nuclear fuel forecast is taken from the 1989 Planning Hearing Generation Expansion Planning Studies by the Florida Electric Power Coordinating Group, Inc. (FLG) forecast adjusted to reflect the underlying general inflation rates presented in Subsection 1A.3.2.

1A.3.3.6 Low and High Forecasts

Low and high forecasts are developed by applying the low and high oil price scenarios (expressed in 1989 dollars) from the 1990 Annual Energy Outlook and the high growth and low growth inflation projections from Subsection 1A.3.2. The low and high forecasts in 1989 dollars and nominal dollars are shown in Tables 1A.3.3-7 and 1A.3.3-8 and in Tables 1A.3.3-9 and 1A.3.3-10, respectively.

Table 1A.3.3-1
Comparison of Fossil Fuel Price Forecasts for Year 2010
(1989 $)

	EIA Base Case	DRI	WEFA	GRI	AGA
World Oil Price (dollars per barrel)	36.90	32.09	33.59	33.68	46.59
Natural Gas Wellhead (dollars per thousand cubic feet)	5.63	4.48	5.12	5.10	4.47
Coal Minemouth (dollars per short ton)	28.55	29.56	36.12	NA	NA

Source: 1990 Annual Energy Outlook.

Acronyms: Energy Information Administration (EIA)
DRI/McGraw-Hill (DRI)
WEFA Group (WEFA)
Gas Research Institute (GRI)
American Gas Association (AGA)

Table 1A.3.3-2
Most Likely Delivered Fuel Price Projections Using
1990 Annual Energy Outlook Base Scenario
(1989 $/MBtu)

Year	Low-Sulfur Coal*	Medium-Sulfur Coal**	Natural Gas	No. 6 Oil	No. 2 Oil	Nuclear
1990	1.90	1.68	2.39	2.75	4.40	0.68
1991	1.93	1.71	2.48	2.88	4.56	0.68
1992	1.96	1.74	2.58	3.02	4.73	0.68
1993	1.99	1.76	2.68	3.17	4.90	0.68
1994	2.03	1.79	2.79	3.32	5.08	0.68
1995	2.06	1.82	2.90	3.48	5.27	0.68
1996	2.08	1.83	3.07	3.70	5.49	0.68
1997	2.09	1.85	3.24	3.92	5.72	0.68
1998	2.11	1.87	3.43	4.17	5.96	0.68
1999	2.13	1.88	3.62	4.43	6.21	0.68
2000	2.21	1.90	3.83	4.70	6.47	0.68
2001	2.23	1.92	4.03	4.86	6.66	0.68
2002	2.26	1.94	4.23	5.02	6.85	0.68
2003	2.29	1.97	4.45	5.19	7.05	0.68
2004	2.31	1.99	4.68	5.36	7.26	0.68
2005	2.34	2.01	4.92	5.54	7.47	0.68
2006	2.37	2.04	5.12	5.67	7.63	0.68
2007	2.39	2.06	5.33	5.80	7.78	0.68
2008	2.42	2.08	5.54	5.93	7.95	0.68
2009	2.45	2.10	5.77	6.07	8.11	0.68
2010	2.48	2.13	6.00	6.21	8.28	0.68
2011	2.50	2.15	6.24	6.35	8.45	0.68
2012	2.53	2.18	6.50	6.50	8.63	0.68
2013	2.56	2.20	6.76	6.65	8.81	0.68
2014	2.59	2.22	7.03	6.80	8.99	0.68
2015	2.62	2.25	7.32	6.96	9.18	0.68
2016	2.64	2.27	7.61	7.12	9.37	0.68
2017	2.67	2.30	7.92	7.29	9.56	0.68
2018	2.70	2.32	8.24	7.45	9.76	0.68
2019	2.73	2.35	8.58	7.63	9.97	0.68
2020	2.76	2.38	8.92	7.80	10.17	0.68

*0.7 to 1.0 percent sulfur.
**1.5 to 2.0 percent sulfur.

Table 1A.3.3-3
Most Likely Delivered Fuel Price Projections Using
1990 Annual Energy Outlook Base Scenario
($/MBtu)

Year	Low-Sulfur Coal*	Medium-Sulfur Coal**	Natural Gas	No. 6 Oil	No. 2 Oil	Nuclear
1990	1.99	1.76	2.50	2.87	4.59	0.71
1991	2.11	1.86	2.71	3.14	4.98	0.74
1992	2.23	1.97	2.93	3.43	5.37	0.77
1993	2.35	2.08	3.17	3.74	5.79	0.81
1994	2.49	2.20	3.42	4.07	6.24	0.85
1995	2.63	2.32	3.70	4.44	6.73	0.90
1996	2.76	2.43	4.07	4.91	7.29	0.95
1997	2.89	2.55	4.47	5.42	7.90	1.00
1998	3.03	2.68	4.92	5.98	8.56	1.05
1999	3.18	2.81	5.41	6.61	9.27	1.10
2000	3.43	2.95	5.95	7.30	10.05	1.16
2001	3.61	3.10	6.50	7.84	10.75	1.22
2002	3.80	3.26	7.11	8.43	11.51	1.28
2003	4.00	3.44	7.77	9.06	12.32	1.35
2004	4.20	3.61	8.50	9.74	13.19	1.42
2005	4.42	3.80	9.29	10.47	14.11	1.49
2006	4.65	4.00	10.06	11.14	14.98	1.56
2007	4.89	4.21	10.88	11.85	15.91	1.64
2008	5.14	4.42	11.78	12.61	16.89	1.73
2009	5.41	4.65	12.74	13.42	17.93	1.82
2010	5.69	4.89	13.79	14.27	19.03	1.91
2011	5.98	5.14	14.92	15.19	20.20	2.01
2012	6.29	5.41	16.15	16.16	21.45	2.11
2013	6.62	5.69	17.47	17.19	22.77	2.22
2014	6.96	5.98	18.91	18.30	24.18	2.33
2015	7.31	6.29	20.46	19.47	25.67	2.45
2016	7.69	6.61	22.14	20.71	27.25	2.57
2017	8.09	6.95	23.96	22.04	28.93	2.70
2018	8.51	7.31	25.93	23.45	30.71	2.84
2019	8.94	7.69	28.06	24.95	32.60	2.99
2020	9.41	8.08	30.36	26.55	34.61	3.14

*0.7 to 1.0 percent sulfur.
**1.5 to 2.0 percent sulfur.

Table 1A.3.3-4
1990 Range of OUC Oil and Gas Prices

	No. 2* ($/MBtu)	No. 6* ($/MBtu)	Gas ($/MBtu)
January	5.23	3.58	3.14
February	4.36	2.85	2.84
March	4.31	2.52	2.28
April	4.19	2.44	2.23
May	4.00	2.36	2.22
June	3.69	2.08	2.28
July	4.07	2.44	2.27
August	5.98	3.69	2.38
September	7.06	4.04	2.36
October	7.41	4.19	2.59
November	6.90	3.89	3.00
December	6.16	3.43	3.01
Minimum	3.69	2.08	2.22
Maximum	7.41	4.19	3.14
Average	5.28	3.12	2.55
AEO**	6.39	2.87	2.50

*Based on Platt's U.S. Gulf Coast prices. To develop a delivered cost to OUC, $0.20/MBtu was added to No. 6 oil and the No. 2 oil was increased by 12 percent. This is the typical price guideline used by OUC. The No. 2 oil is based on 5.825 MBtu/bbl and the No. 6 oil is based on 6.29 MBtu/bbl.

**Annual Energy Outlook.

Table 1A.3.3-5
Quality of Coal Burned at Stanton 1 and McIntosh 3 in 1989

Unit	Range	Tons (thousands)	Heating Value (Btu/lb)	Sulfur (percent)	SO_2 (lb/MBtu*)	Ash (percent)	Ash (lb/MBtu)	Delivered Cost ($/MBtu)
Stanton	Weighted Average	1,074.26	12,722	0.74	1.16	8.75	6.88	1.89
Stanton	Minimum		12,526	0.57	0.89	6.37	N/A	1.84
Stanton	Maximum		13,033	0.80	1.26	9.87	N/A	1.93
McIntosh	Weighted Average	724.00	12,298	1.29	2.09	11.01	8.95	2.12
McIntosh	Minimum		12,181	1.08	1.74	10.21	N/A	1.97
McIntosh	Maximum		12,395	1.67	2.74	11.39	N/A	2.29

*Before scrubbing.

Source: COALDAT Marketing Report: Utility Format Calendar Year 1989.

Table 1A.3.3-6
Compliance Coal Price Estimates for the Year 2000
(1989 $)

Source	Range ($/MBtu)	Average ($/MBtu)
1	1.35 - 1.46	1.40
2	1.44 - 1.60	1.52
3	--	1.20
Average		1.37

Sources:

(1) Rafael Villagran, Coal Outlook, June 4, 1990.

(2) Thomas A. Hewson, Jr., Utility Coal Market Dynamics and Compliance Decisions Under Acid Rain Legislation, June 1990.

(3) S. C. Suboleski, R. L. Frantz, R. V. Ramani, and R. Rao, Outlook for Coal Productivity, June 1990.

Table 1A.3.3-7
Low Delivered Fuel Price Projections Using
1990 Annual Energy Outlook Low Oil Price Scenario
(1989 $/MBtu)

Year	Low-Sulfur Coal*	Medium-Sulfur Coal**	Natural Gas	No. 6 Oil	No. 2 Oil	Nuclear
1990	1.90	1.68	2.17	2.40	3.99	0.68
1991	1.93	1.71	2.26	2.46	4.08	0.68
1992	1.96	1.74	2.35	2.52	4.17	0.68
1993	1.99	1.76	2.45	2.59	4.27	0.68
1994	2.03	1.79	2.55	2.65	4.36	0.68
1995	2.06	1.82	2.66	2.72	4.46	0.68
1996	2.08	1.84	2.81	2.87	4.64	0.68
1997	2.10	1.85	2.98	3.03	4.83	0.68
1998	2.12	1.87	3.15	3.20	5.02	0.68
1999	2.14	1.89	3.34	3.37	5.22	0.68
2000	2.21	1.91	3.53	3.56	5.43	0.68
2001	2.24	1.93	3.67	3.68	5.59	0.68
2002	2.26	1.95	3.82	3.80	5.76	0.68
2003	2.28	1.97	3.97	3.93	5.93	0.68
2004	2.31	1.99	4.13	4.06	6.10	0.68
2005	2.33	2.01	4.30	4.19	6.28	0.68
2006	2.35	2.03	4.44	4.27	6.38	0.68
2007	2.38	2.05	4.59	4.34	6.48	0.68
2008	2.40	2.07	4.74	4.42	6.58	0.68
2009	2.43	2.10	4.90	4.50	6.68	0.68
2010	2.45	2.12	5.06	4.58	6.79	0.68
2011	2.47	2.14	5.23	4.66	6.90	0.68
2012	2.50	2.16	5.40	7.01	7.01	0.68
2013	2.52	2.18	5.58	4.83	7.12	0.68
2014	2.55	2.20	5.76	4.92	7.23	0.68
2015	2.58	2.22	5.95	5.01	7.34	0.68
2016	2.60	2.25	6.15	5.10	7.46	0.68
2017	2.63	2.27	6.35	5.19	7.57	0.68
2018	2.65	2.29	6.57	5.28	7.69	0.68
2019	2.68	2.31	6.78	5.38	7.81	0.68
2020	2.71	2.34	7.01	5.47	7.94	0.68

*0.7 to 1.0 percent sulfur.
**1.5 to 2.0 percent sulfur.

Table 1A.3.3-8
High Delivered Fuel Price Projections Using
1990 Annual Energy Outlook High Oil Price Scenario
(1989 $/MBtu)

Year	Low-Sulfur Coal*	Medium-Sulfur Coal**	Natural Gas	No. 6 Oil	No. 2 Oil	Nuclear
1990	1.90	1.68	2.63	3.11	4.79	0.68
1991	1.93	1.71	2.68	3.32	5.00	0.68
1992	1.96	1.74	2.73	3.55	5.22	0.68
1993	1.99	1.76	2.78	3.79	5.46	0.68
1994	2.03	1.79	2.84	4.05	5.70	0.68
1995	2.06	1.82	2.89	4.32	5.95	0.68
1996	2.08	1.83	3.05	4.56	6.19	0.68
1997	2.09	1.85	3.21	4.81	6.44	0.68
1998	2.11	1.87	3.38	5.08	6.70	0.68
1999	2.13	1.88	3.57	5.36	6.98	0.68
2000	2.21	1.90	3.76	5.66	7.26	0.68
2001	2.23	1.92	3.94	5.91	7.51	0.68
2002	2.26	1.94	4.14	6.17	7.77	0.68
2003	2.28	1.96	4.34	6.44	8.04	0.68
2004	2.30	1.98	4.55	6.72	8.32	0.68
2005	2.33	2.00	4.77	7.02	8.61	0.68
2006	2.35	2.02	5.01	7.20	8.80	0.68
2007	2.38	2.05	5.25	7.39	9.00	0.68
2008	2.41	2.07	5.51	7.58	9.21	0.68
2009	2.43	2.09	5.78	7.78	9.42	0.68
2010	2.46	2.12	6.07	7.98	9.63	0.68
2011	2.49	2.14	6.37	8.19	9.85	0.68
2012	2.52	2.16	6.68	8.40	10.07	0.68
2013	2.55	2.19	7.01	8.62	10.30	0.68
2014	2.57	2.21	7.36	8.84	10.53	0.68
2015	2.60	2.24	7.72	9.07	10.77	0.68
2016	2.63	2.26	8.11	9.31	11.01	0.68
2017	2.66	2.29	8.51	9.55	11.26	0.68
2018	2.69	2.31	8.93	9.80	11.52	0.68
2019	2.72	2.34	9.37	10.05	11.78	0.68
2020	2.75	2.36	9.83	10.31	12.05	0.68

*0.7 to 1.0 percent sulfur.
**1.5 to 2.0 percent sulfur.

Table 1A.3.3-9
Low Delivered Fuel Price Projections Using
1990 Annual Energy Outlook Low Oil Price Scenario
($/MBtu)

Year	Low-Sulfur Coal*	Medium-Sulfur Coal**	Natural Gas	No. 6 Oil	No. 2 Oil	Nuclear
1990	1.99	1.76	2.27	2.51	4.17	0.71
1991	2.11	1.86	2.47	2.68	4.45	0.74
1992	2.20	1.94	2.63	2.82	4.67	0.76
1993	2.29	2.02	2.81	2.97	4.89	0.78
1994	2.38	2.10	3.00	3.12	5.12	0.80
1995	2.48	2.19	3.20	3.28	5.37	0.82
1996	2.57	2.27	3.47	3.54	5.73	0.84
1997	2.66	2.35	3.77	3.83	6.10	0.86
1998	2.75	2.43	4.09	4.15	6.51	0.88
1999	2.85	2.51	4.43	4.48	6.94	0.90
2000	3.01	2.60	4.81	4.85	7.40	0.92
2001	3.12	2.70	5.13	5.14	7.81	0.95
2002	3.23	2.79	5.47	5.44	8.24	0.97
2003	3.35	2.89	5.83	5.76	8.69	0.99
2004	3.47	3.00	6.22	6.10	9.17	1.02
2005	3.59	3.10	6.63	6.46	9.68	1.04
2006	3.72	3.21	7.02	6.74	10.08	1.07
2007	3.85	3.33	7.43	7.03	10.49	1.10
2008	3.99	3.44	7.87	7.34	10.93	1.12
2009	4.13	3.56	8.33	7.66	11.37	1.15
2010	4.27	3.69	8.83	7.99	11.84	1.18
2011	4.42	3.82	9.35	8.33	12.33	1.21
2012	4.58	3.96	9.90	8.70	12.84	1.24
2013	4.74	4.10	10.48	9.07	13.36	1.27
2014	4.91	4.24	11.10	9.47	13.91	1.30
2015	5.08	4.39	11.75	9.88	14.49	1.34
2016	5.26	4.54	12.44	10.31	15.08	1.37
2017	5.45	4.70	13.17	10.76	15.70	1.40
2018	5.64	4.87	13.95	11.22	16.35	1.44
2019	5.84	5.04	14.77	11.71	17.02	1.48
2020	6.04	5.22	15.64	12.22	17.72	1.51

*0.7 to 1.0 percent sulfur.
**1.5 to 2.0 percent sulfur.

Table 1A.3.3-10
High Delivered Fuel Price Projections Using
1990 Annual Energy Outlook High Oil Price Scenario
($/MBtu)

Year	Low-Sulfur Coal*	Medium-Sulfur Coal**	Natural Gas	No. 6 Oil	No. 2 Oil	Nuclear
1990	1.99	1.76	2.75	3.25	5.00	0.71
1991	2.11	1.86	2.92	3.62	5.46	0.74
1992	2.26	2.00	3.14	4.08	6.01	0.78
1993	2.42	2.14	3.38	4.60	6.62	0.82
1994	2.60	2.29	3.63	5.18	7.30	0.87
1995	2.78	2.46	3.91	5.84	8.04	0.92
1996	2.96	2.62	4.34	6.50	8.83	0.97
1997	3.15	2.78	4.83	7.24	9.69	1.02
1998	3.35	2.96	5.37	8.06	10.64	1.07
1999	3.57	3.15	5.97	8.98	11.68	1.13
2000	3.90	3.35	6.64	10.00	12.82	1.20
2001	4.16	3.58	7.35	11.01	14.00	1.26
2002	4.44	3.81	8.13	12.13	15.28	1.33
2003	4.73	4.07	9.00	13.36	16.68	1.40
2004	5.04	4.33	9.95	14.71	18.21	1.48
2005	5.38	4.62	11.01	16.21	19.88	1.56
2006	5.74	4.93	12.19	17.54	21.45	1.65
2007	6.12	5.26	13.50	18.99	23.14	1.74
2008	6.53	5.61	14.94	20.55	24.96	1.84
2009	6.96	5.99	16.54	22.24	26.93	1.94
2010	7.43	6.38	18.32	24.08	29.06	2.04
2011	7.92	6.81	20.28	26.06	31.35	2.16
2012	8.45	7.27	22.45	28.21	33.82	2.27
2013	9.02	7.75	24.85	30.53	36.49	2.40
2014	9.62	8.27	27.51	33.05	39.37	2.53
2015	10.26	8.82	30.46	35.77	42.47	2.67
2016	10.95	9.41	33.72	38.72	45.82	2.82
2017	11.68	10.04	37.33	41.91	49.44	2.97
2018	12.46	10.71	41.33	45.36	53.34	3.14
2019	13.29	11.43	45.76	49.10	57.55	3.31
2020	14.18	12.19	50.66	53.15	62.09	3.49

*0.7 to 1.0 percent sulfur.
**1.5 to 2.0 percent sulfur.

Table 3.3-11
1989 Planning Hearing Base Case Forecast
($/MBtu)

Year	Low-Sulfur Coal*	Medium-Sulfur Coal**	Natural Gas	No. 6 Oil***	No. 2 Oil	Nuclear
1990	2.14	2.03	2.38	3.10	4.65	0.71
1991	2.25	2.12	2.58	3.35	5.05	0.74
1992	2.38	2.25	2.92	3.68	5.58	0.77
1993	2.52	2.39	3.30	4.09	6.23	0.81
1994	2.68	2.54	3.72	4.54	6.99	0.85
1995	2.86	2.70	4.20	5.02	7.72	0.90
1996	3.05	2.88	4.68	5.54	8.56	0.95
1997	3.25	3.07	5.20	6.11	9.49	1.00
1998	3.46	3.27	5.75	6.72	10.47	1.05
1999	3.68	3.48	6.22	7.23	11.31	1.10
2000	3.91	3.70	6.87	7.77	12.20	1.16
2001	4.15	3.92	7.36	8.30	13.09	1.22
2002	4.40	4.16	7.88	8.87	14.04	1.28
2003	4.68	4.42	8.51	9.45	15.02	1.35
2004	4.96	4.69	9.09	10.06	16.05	1.42
2005	5.27	4.98	9.78	10.69	17.12	1.49
2006	5.60	5.29	10.41	11.34	18.23	1.56
2007	5.94	5.62	11.06	12.01	19.38	1.64
2008	6.31	5.97	11.85	12.70	20.57	1.73
2009	6.70	6.34	12.70	13.43	21.84	1.82
2010	7.12	6.74	13.61	14.21	23.18	1.91
2011	7.56	7.15	14.58	15.02	24.61	2.01
2012	8.03	7.60	15.62	15.89	26.13	2.11
2013	8.53	8.07	16.74	16.80	27.74	2.22
2014	9.06	8.57	17.93	17.77	29.44	2.33
2015	9.63	9.11	19.22	18.80	31.26	2.45
2016	10.23	9.67	20.59	19.88	33.18	2.57
2017	10.86	10.28	22.06	21.02	35.23	2.70
2018	11.54	10.92	23.64	22.23	37.40	2.84
2019	12.25	11.59	25.33	23.51	39.70	2.99
2020	13.02	12.32	27.14	24.87	42.15	3.14

*Less than 1.0 percent sulfur.
**1.0 to 2.0 percent sulfur.
***0.7 to 2.0 percent sulfur.

1A.3.4 Fuel Availability

This section reviews the availability of coal, natural gas, No. 2 oil, and No. 6 oil as potential fuels. For coal, the reserve base, fuel quality, and production capability will be discussed. Factors affecting the future price of these fuels have been discussed in Subsection 1A.3.3, Fuel Price Projections.

1A.3.4.1 Coal

1A.3.4.1.1 Quantity Required. Stanton 2, with a generating capability of 440 MW when operating at an average annual capacity factor of 70 percent, will require about 1.1 million tons of coal per year. This assumes that the coal has a heating value of 12,400 Btu/lb.

1A.3.4.1.2 Quality Required. Stanton 2 is designed to burn coals with a wide range of sulfur content ranging from 0.7 percent to 4.0 percent, which corresponds to uncontrolled sulfur dioxide emission potential rates of 1.1 lb SO_2/MBtu to 6.5 lb SO_2/MBtu, based on a 12,400 Btu/lb higher heating value. A discussion of the proposed BACT fuel for Stanton 2 can be found in Subsection 1A.3.4.3 of this Supplemental Site Certification Application.

1A.3.4.1.2.1 Quality of Potential Coal Sources. Figure 1A.3.4-1 compares western and eastern sources of coal and their uncontrolled sulfur dioxide emission potential. In the East, the primary source of low-sulfur coal is US Bureau of Mines District 8, which includes eastern Kentucky, western Virginia, and southwestern West Virginia. These coal producing areas are shown on Figure 1A.3.4-2, Coal Fields of the Conterminous United States. The average uncontrolled sulfur dioxide emission potential of this coal was 1.6 lb SO_2/MBtu. In the West, the primary source of low-sulfur coal is the Powder River Basin of Wyoming.

The low rank subbituminous coals from the Powder River Basin have a very low uncontrolled sulfur dioxide emission potential (less than 0.9 lb SO_2/MBtu). Because of their favorable geological conditions, the mining economics are also favorable, with current prices being less than $4.00 per ton. The primary negative characteristic of the Powder River Basin coal is its high moisture content (about 30 percent), which results in an as-received heating value of 8,000 to 9,000 Btu/lb. Considerable research and development are now being conducted on upgrading these coals by reducing their inherent moisture. These processes produce a bituminous rank coal with a heating value of nearly 11,000 Btu/lb. If this can be done in commercial quantities, the "market reach" of the Powder River Basin coals will be significant in the future.

Figures 1A.3.4-3 and 1A.3.4-4 show the uncontrolled sulfur dioxide emission potential of coals delivered to Florida and Georgia electric utilities in 1989. It is seen that coals from US Bureau of Mines District 8 in central Appalachia had the lowest uncontrolled sulfur dioxide emission potential (1.2 lb SO_2/MBtu - 1.5 lb SO_2/MBtu). It is also seen that foreign coals from Colombia and Venezuela have uncontrolled sulfur dioxide emission potentials of less than 1.2 lb SO_2/MBtu.

The primary alternative source to central Appalachian coals is the Illinois Basin (southern Illinois, southern Indiana, and west Kentucky), where the coals have uncontrolled sulfur dioxide emission potentials which ranged from 4.5 to 6 lb SO_2/MBtu in 1989.

1A.3.4.1.3 Clean Air Act Amendments and Changing Demand. The amendments to the Clean Air Act which were passed into law on November 15, 1990, call for a 10 million-ton reduction in the amount of sulfur dioxide emissions from 1980 levels by the year 2000 (Phase II). Those plants which are not in compliance will have to install flue gas desulfurization systems, switch to low-sulfur coal, or purchase allowances.

It is very difficult to estimate the additional demand for low-sulfur coal which will result from these regulations since "switch versus scrub" decisions have not yet been formalized. The actual amount of fuel switching will be dependent on the cost of retrofitting scrubbers compared to the anticipated premium for low-sulfur coal, the expected value of allowances that can be generated by over compliance, and the expected price of allowances that can be purchased to obviate the need for scrubbing or switching fuels.

The ability of low-sulfur coal reserves and production from central Appalachia to satisfy increased demand by 1995 (Phase I) and the year 2000 (Phase II) will depend upon the following factors.

- The amount of increased demand compared to current excess mine production capability and expansion capability within the central Appalachian coal region.
- The delivered cost of low-sulfur coal from the Powder River Basin of Wyoming.
- The delivered cost of foreign coals from Colombia and Venezuela to tidewater ports.

- Market prices for low-sulfur coals from central Appalachia compared to the cost of developing new mines to exploit undeveloped reserve blocks.
- The amount of increased demand for US origin low-sulfur coals from European utilities due to a reduction in domestic coal mining subsidies and the Europeans' desire to reduce sulfur dioxide emissions.

The interaction between these factors is quite complex. The current period of low prices and excess production capacity is likely to continue for the next few years. During this time, there will be opportunities to arrange long-term supply commitments at favorable prices and with favorable contract provisions.

1A.3.4.1.4 Potential Sources. The following subsections look at potential sources of coal supply from central Appalachia, the Illinois Basin, the Powder River Basin, and foreign sources. For each producing basin, a brief discussion is provided on reserves, coal quality, and production capability.

1A.3.4.1.4.1 Current Sources for Florida and Georgia Utilities. Figures 1A.3.4-5 and 1A.3.4-6 show the sources of coal delivered to Florida and Georgia electric utilities in 1989. The quantity and the predominant modes of transportation from each source are also shown. It is seen that the predominant source has been eastern Kentucky with delivery by rail. The coal from western Kentucky (US Bureau of Mines District 9) was the second largest source for Florida utilities and was predominantly delivered by river and gulf movements to west-coast Florida utilities (Florida Power Corporation and Tampa Electric Company).

There were also significant quantities of coal purchased from Colombia and Venezuela. Georgia utilities have received some coal from the distant Powder River Basin in Wyoming and Montana.

1A.3.4.1.4.2 Central Appalachia. The primary supply source of low-sulfur coal for consumers in the eastern US is the central Appalachian region, which comprises east Kentucky, southwestern West Virginia, and western Virginia. As previously discussed, east Kentucky is a major supply source to electric utilities in Florida and Georgia.

Reserve estimates for this region vary with the assumptions made regarding mineability and the market price level to cover the cost of producing and beneficiating (cleaning or washing) the coal. In 1982, the Kentucky Geological Survey estimated that eastern Kentucky contained 36 billion tons of remaining coal resources.[1] About 43 percent of these resources have an uncontrolled sulfur dioxide emission potential of less than 1.2 lb SO_2/MBtu.[2]

In reducing the resource base to economically recoverable resources, or reserve base, the level of mining technology employed and the market price available to cover extraction costs are important. This means that the amount of economically recoverable reserves can vary from time to time with market conditions.

At the present time, no agency of the federal government maintains coal reserve data. The last comprehensive estimate of coal reserves which emphasized low-sulfur reserves was made by the US Bureau of Mines in 1975.[2] In a recent Electric Power Research Institute sponsored study entitled "Reserves and Potential Supply of Low Sulfur Appalachian Coal," it was found that on the basis of current industry practice, the reserve estimates made by the US Bureau of Mines in a 1974 assessment had significant errors. For example, surface mineable reserves were found to be understated by a factor of 3 to 4. However, underground mineable reserves were overstated by a factor of 2.[3] Most of these errors were due to the way in which the feasibility of recovering underground mineable reserves was determined and the methodology used to calculate the cost of mining. However, the net result of these errors seemed to cancel and, therefore, the previously published economically recoverable reserve estimates should be fairly representative and, in fact, may understate the amount of reserves which are economically recoverable.

It has been estimated that at a 1984 price level of $35 per ton, there would be at least 20 billion tons of recoverable reserves with a sulfur content of less than 1.4 percent.[3] This compares with a value of slightly less than 20 billion tons estimated by the US Bureau of Mines.[3] For production levels of 264 million tons in 1989, the reserve base should be capable of supporting existing production levels for another 70 years.

The central Appalachian coals are desirable to electric utilities because of their low-sulfur content and relatively low ash content. These characteristics also make them desirable for producing metallurgical grade coking coals. Depending upon demand for metallurgical coking coal, some of these coals move in and out of the low-sulfur steam coal markets.

Of the 144 million tons of coal shipped to electric utilities in 1989, about 40 million tons (28 percent) had uncontrolled sulfur dioxide emission potentials of less than 1.2 lb SO_2/MBtu, about 90 million tons (63 percent) had uncontrolled sulfur dioxide emission potentials which ranged from 1.2 to 2.5 lb SO_2/MBtu, and only 10 million tons (7 percent) had uncontrolled sulfur dioxide emission potentials greater than 2.5 lb SO_2/MBtu.[4]

Table 1A.3.4-1 shows the quantity and quality of coal shipped from central Appalachia to electric utilities in 1989. As seen from this table, the major production was from east Kentucky. The ash averaged around 10 percent, the sulfur 1.6 lb SO_2/MBtu, and the heating value 12,450 Btu/lb.

A significant amount of coal production from the central Appalachian region is beneficiated. These beneficiation facilities remove the heavy mineral matter such as ash and pyrite (containing a major portion of the sulfur in coal) from the lighter, predominantly carbon particles using gravimetric separation processes. Many of the new mine developments in the mid and late 1990s will be equipped with coal beneficiation plants to produce the low-sulfur coals required by electric utilities who switch to low-sulfur coal to comply with the 1990 Clean Air Act Amendments.

With the increase in coal beneficiation plants, the high-sulfur middlings and refuse from these coal beneficiation plants still contain significant amounts of carbon and will be marketed at very attractive prices. With the proposed BACT emission rate, Stanton 2 will have the ability to burn these high-sulfur middlings and take advantage of this market.

1A.3.4.1.4.3 Illinois Basin. As seen from the map on Figure 1A.3.4-2, the Illinois Basin includes southern Illinois, southwestern Indiana, and western Kentucky. As seen from Figure 1A.3.4-1, the quality of this coal is not as good as that from the central Appalachian region. The uncontrolled sulfur dioxide emission potential ranges from 4.7 to 5.7 lb SO_2/MBtu compared to 1.6 lb SO_2/MBtu from central Appalachia. While the quality of coal from the Illinois Basin is not as desirable, there are very large reserves in this area. The US Bureau of Mines estimates remaining recoverable reserves amount to 110 billion tons which is significantly greater than the approximately 20 billion tons for the central Appalachian region.[5] Because of their greater thickness and the large area of the two predominant seams (Herrin No. 6 and Harrisburg-Springfield No. 5) making up these reserves, the mining economics are also more favorable than in central Appalachia. An indicator of the low production cost is current spot market prices for this high-sulfur coal, which range from \$14 to \$17 per ton in

Illinois and Indiana as compared to \$21 to \$32 per ton for low-sulfur central Appalachian coal.[6]

While the price of the Illinois Basin coals is attractive and the reserves are plentiful, the high average uncontrolled sulfur dioxide emission potential ranges from 4.7 to 5.7 lb SO_2/MBtu and will require high-sulfur dioxide removal using flue gas desulfurization.

Production from the Illinois Basin peaked at about 145 million annual tons in 1972.[7] With the advent of the Clean Air Act in 1970 and the New Source Performance Standards promulgated in 1971, and the subsequent preference for low-sulfur coals, many mining operations in the Illinois Basin have been shut down. With increased demand, some of these operations may be reactivated at minimal cost, compared to that required to develop new mines.

1A.3.4.1.4.4 Other Sources. The Powder River Basin, which covers portions of Wyoming and Montana, began to be exploited in the early 1970s because of its low-sulfur coal (average of 0.9 lb SO_2/MBtu) and favorable mining economics. Combined seam thicknesses often total 80 to 100 feet, with minimal interburden and overburden, often less than 50 feet thick. This results in very favorable mining economics with current f.o.b. mine prices less than \$4.00 per ton.

The western railroads have been quite progressive in reducing their cost structure through the use of unit coal trains, changes in work rules, and more favorable terrain compared to the mountainous Appalachian region. This has lead to very low transportation costs compared to eastern railroads. For example, the current cost estimate for moving coal from the Powder River Basin to barge loading terminals on the Mississippi River near St. Louis or on the Ohio River in southern Illinois ranges from \$14 to \$15 per ton. Many of the large power plants along the Ohio River which will have to reduce their sulfur dioxide emissions by the year 2000 may find it economically desirable to switch to low-sulfur Powder River Basin coals. Some of this coal will also be used in blends with higher rank eastern bituminous coals. The Southern Company has conducted successful test burns using Powder River Basin coal at its Scherer Plant in Georgia and Miller Plant in Alabama.

There are several researchers developing processes for reducing the high moisture content of this coal from its typical 30 percent to less than 10 percent. This would effectively increase the heating value from its low value of 8,000 to

8,500 Btu/lb up to about 11,000 Btu/lb. This would have a favorable impact on reducing transportation cost, reducing the derate associated with burning the low rank subbituminous coal in plants designed for bituminous coals, and improve boiler efficiencies with a lower moisture content coal. Several of these concepts have proved to be technologically and economically feasible at the pilot plant stage of development. If successful at the demonstration plant level, the market for Powder River Basin coals will be increased considerably and will have a material impact on the future demand for and price levels of central Appalachian low-sulfur coals.

For those power plants which have access to water delivery, coals from Colombia and Venezuela which have low production costs and low-sulfur dioxide emission potential (Figure 1A.3.4-1) will also constrain demand and price runups for low-sulfur central Appalachian coal. Although the OUC does not currently have water delivery capability for South America or Powder River Basin coals, the fact that other utilities in Florida and the southeast do will act as a constraining influence on future prices at the mine and for rail transportation from central Appalachia to Stanton 2.

1A.3.4.2 Natural Gas

1A.3.4.2.1 Reserves. When the availability of natural gas is evaluated as a fuel for baseload power generation, the quantity of reserves available for commitment and the replacement of production by new reserves must be considered. Historically, there has been a mixed record of the gas industry's ability to replace annual production with newly discovered reserves. According to the Energy Information Administration and the American Gas Association (AGA), the replacement level in the lower 48 states during the decade of 1970-1979 was 46 percent of production. By contrast, the replacement level was nearly 94 percent of production for the period 1980-1989.[8,9]

A major reason for the dramatic improvement in replacement is the Natural Gas Policy Act of 1978 (NGPA). The NGPA stimulated drilling activity with pricing incentives for new gas. Continued pricing motivation will be required to maintain drilling activity. According to the AGA, continued increases in exploratory drilling will be necessary to maintain an optimum level of reserve replacement.[8]

Florida is served primarily by the Florida Gas Transmission Company (FGT) pipeline system, with the vast majority of natural gas coming from the Texas, Louisiana, and offshore Gulf of Mexico region. According to the DOE, total new discoveries for the lower 48 states accounted for only 63 percent of all 1989 reserve additions.[8,10] New discoveries are solid additions to reserve figures, including extension of old reservoirs and new reservoir discoveries. By contrast, over one-third of 1989 reserve additions were softer and consisted of producer adjustments and net revisions to the existing reservoir base data. It is significant to note that nearly one-third of the lower 48 states' new discoveries were in the federal offshore Gulf of Mexico region. This reinforces its position as a dominant region for natural gas exploration. In addition, in 1989, gas production reached its highest level since 1984 with lower 48 state production of 16.58 trillion cubic feet. This is indicative of the strong demand for gas during 1989.[8]

The long-term availability of natural gas appears to be good if pricing remains high enough to encourage producers to drill exploratory wells replacing production. The demand for gas is increasing. If this continues, the pricing level will remain strong. In addition, the Clean Air Act Amendments of 1990 may further increase the demand for natural gas.

1A.3.4.2.2 Pipeline Capacity. The ability to deliver natural gas from the location of reserves is dependent on the availability of pipeline capacity. The only natural gas pipeline serving Peninsular Florida is the Florida Gas Transmission (FGT) system. The FGT system has been and continues to be fully allocated to existing demand. The current pipeline cannot accommodate increased firm service demands.

FGT is in the process of expanding its pipeline capability through a construction program known as the Phase II expansion. As part of the Phase II process, all customers were asked to project their increased natural gas requirements. Because consolidated requests for increased service were so large, FGT was unable to satisfy the requests with the Phase II expansion. FGT customer requests for service were decreased on a prorated basis by FGT. For example, OUC's initial request for approximately 7.2 billion cubic feet (bcf) per year was reduced 15 percent by FGT to an allocation of approximately 6.1 bcf per year. Likewise, Kissimmee Utility Authority's initial request of 3.1 bcf per year was reduced by nearly 42 percent to 1.8 bcf per year.

FGT is in the preliminary stage of an additional expansion to be known as Phase III. According to FGT officials, the allocation process for Phase III will be similar to Phase II. Customers will nominate to FGT the increase in firm service desired. If the requests from FGT customers are large enough, it is reasonable to assume FGT will again reduce on a pro rata basis the quantity of service available. This makes planning on increased natural gas availability very difficult and risky.

The cost of the new allocation of service is also unknown. FGT will be unable to make reasonable estimates of service costs until the Phase III expansion is defined in a regulatory application. Even then, the cost and design of service rates will be uncertain until approved by the appropriate regulatory agencies and accepted by FGT.

In summary, the availability of natural gas reserves to fuel baseload electric generation is subject to question. Though the current deliverability is good now, only strong gas prices can ensure exploration activity for reserve replacement. Demand for gas has increased and apparently it will continue to increase, thus reducing excess deliverability and potentially supporting high prices. Pipeline capacity does not exist today for large incremental loads. On the basis of past expansion allocations, sufficient future capacity is uncertain. In addition, should the capacity be installed, its cost is unknown until regulatory approval is obtained.

All of these factors must be considered when evaluating the reliability of natural gas as a fuel for generating unit capacity additions. Currently, when all of these factors are analyzed, natural gas cannot be considered a reliable fuel for alternatives to the addition of Stanton 2.

1A.3.4.3 No. 2 Oil

The following subsections will discuss the current and future sources of No. 2 oil. This fuel could be used as startup fuel for a coal fired power plant, and as a backup or standby fuel for combustion turbines and combined cycle plants fueled with natural gas.

1A.3.4.3.1 Current Sources. During calendar year 1989, OUC purchased 11,000 barrels of No. 2 oil. This compares with purchases of 5,000 barrels in 1987 and 2,000 barrels in 1988.*

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*One barrel contains 42 gallons.

OUC currently purchases No. 2 fuel oil from marketers or oil companies on a cost, insurance, and freight (CIF) bid basis, with deliveries in trucks with a capacity of about 7,500 gallons. Most shipments originate at product terminals located at Taft, Port Canaveral, and sometimes Port Manatee. The sulfur content is less than 0.3 percent.

The refineries which produced this No. 2 fuel oil tend to be located along the Gulf Coast, such as Port Arthur/Houston and New Orleans/Baton Rouge. Some No. 2 oil has also come from offshore refineries in the Caribbean. The crude oil sources for these refineries may be domestic or foreign. During 1989, the US imported about 43 percent of its crude oil requirements.[11] During 1989, the Organization of Petroleum Exporting Countries (OPEC) controlled about 38 percent of world crude oil production, which averaged 60.3 million barrels per day.[12] OPEC countries controlled about 80 percent of total known world crude oil reserves which totaled 999 billion barrels in 1989.[12]

The Middle Eastern countries control about 65 percent of world crude oil reserves and they accounted for 27 percent of total world production in 1989.[12]

In terms of years of remaining reserves on a world-wide basis, the reserves-to-production ratio (known reserves/annual production) shows about 45 years of remaining reserves. However, this ratio can be misleading because the production history for the typical oil or gas well tends to be one of exponential depletion. That is, most of a well's reserves are produced in the first few years of its operation. Therefore, if production isn't replaced on an annual basis, the deliverability from producing wells can decline significantly and the potential for supply disruption is increased.

Unlike coal, the majority of oil transactions occur in the spot market. Market prices are primarily established on world wide supply and demand considerations over short periods of time. One of the primary uncertainties regarding short-term supply has been OPEC policies regarding production quotas.

1A.3.4.3.2 Future Sources. The future supply of No. 2 fuel oil is likely to come from the same refineries that currently satisfy OUC needs. Crude oil sources for these refineries are likely to change with a shift toward importing more from Latin American and Canadian producers.

1A.3.4.3.2.1 Impact of Clean Air Act Amendments. Clean Air Act Amendments passed by Congress in November 1990 called for a reduction in the sulfur content of highway diesel fuel to 0.05 percent sulfur by weight. This requirement is to be implemented by October of 1993. At the present time, most refiners produce a single grade of No. 2 fuel oil with a specification that satisfies the ASTM requirements for No. 2 heating oil and No. 2 diesel fuel. By producing a No. 2 fuel oil with properties that satisfy both specifications, they reduce distribution costs by avoiding double tankage and compartmentalization of delivery trucks. With the more stringent sulfur requirement for highway diesel fuel required by October 1993, it is not known whether refiners will produce a single No. 2 fuel oil with a 0.05 percent sulfur level or two fuels. One would be for off-highway diesel and heating oil use which could have a sulfur content up to the maximum ASTM allowed limit of 0.5 percent* and the other with 0.05 percent sulfur would be for highway diesel use.

Numerous considerations affect these decisions which will be made on a refinery specific basis. For example, some product pipelines have specifications on the maximum sulfur content allowed for fungible** fuels transported in their pipelines. The Colonial Pipe Line System currently limits the sulfur content of No. 2 fuel oil to 0.2 percent since it serves the mid-Atlantic markets which have such restrictions.

Should a single No. 2 fuel oil market develop with 0.05 percent maximum sulfur, the cost of No. 2 heating oil is likely to increase significantly. Should a tiered market develop, the higher sulfur fuels would sell at a substantial discount.

At this time, it is not known what air pollution control requirements would be required for power generation facilities using No. 2 fuel oil. It may be that future combustion turbines will have to be fired with a low-sulfur No. 2 highway diesel fuel, albeit at a higher price markup relative to the price of crude than has been suggested by past experience.

1A.3.4.4 No. 6 Oil

No. 6 or residual oil is often referred to as the "bottom of the barrel" after the lighter more profitable fractions have been removed from the crude. The lighter

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*As more states and local authorities lower the sulfur content of fuels for off-highway use, this upper limit may have less significance.

**Designates movable goods such as grains or precious metals where any unit or part of which can replace another unit.

fractions usually sell at a premium, and the heavy oil sells at a discount, However, if market conditions lead to a significant No. 6 oil discount relative to crude, refining processes can be modified to extract more light ends from the residual oil with the resulting "waste product" being a solid petroleum coke. The price of residual oil relative to crude oil tends to range from a low level governed by the capital and operating costs associated with coking or other refining processes to "destroy the resid" and the price of crude oil.

1A.3.4.4.1 Current Sources. In 1989, OUC purchased 474,000 barrels of No. 6 oil compared to 1,130,000 barrels in 1986, before commercial operation of Stanton 1.[13] OUC currently receives shipments of No. 6 oil for the Indian River Plant from refineries located along the Gulf Coast such as Port Arthur/Houston and New Orleans/Baton Rouge and other offshore refineries in the Caribbean. These shipments arrive at Port Canaveral in shipment lots which range from 100,000 to 200,000 barrels. The sulfur content of these shipments ranges from 1.0 to 1.8 percent. No. 6 oil for Stanton Energy Center, which is used during startup, is delivered by truck. The sulfur content is less than 0.7 percent. Since No. 6 oil is a product of the crude oil refining process, the crude sources would be the same as those for the other refined products discussed in Subsection 1A.3.4.3.1.

1A.3.4.4.2 Future Sources. The same refineries and distribution systems are likely to be used in the future for obtaining No. 6 oil. The crude oil sources are likely to shift in the manner discussed in Subsection 1A.3.4.3.2. One problem with shifting to some of the Latin American crudes, such as those from Venezuela, is that the specific gravity of the crude tends to be higher (API gravity is lower) than that of traditionally imported lighter crudes. This could increase the percentage of residual oil produced from a barrel of crude from the current 10 percent level, but it could also increase the economic incentive for installing coking units to recover more light ends which would destroy the excess residual oil.

1A.3.4.5 Coal Transportation

1A.3.4.5.1 Rail Transportation of Coal--Overview. The railroad servicing the Stanton Energy Center plant site is CSX Transportation, a unit of CSX Corporation. CSX Corporation is, in turn, a diversified enterprise operating within four primary lines of business (business units): transportation, energy, technology, and properties.

CSX Transportation (hereafter "CSXT") operates one of the largest coal-haul railroads in the United States and, through a wholly owned subsidiary (American Commercial Barge Lines), operates a major terminaling and large-haul system on the Ohio-Mississippi Rivers (and tributaries) of the eastern half of the nation. This rail and inland waterways transport system is linked to modern deep-water ports on the great lakes, Atlantic, and Gulf Coasts.

The CSXT rail network comprises approximately 18,800 route-miles of rail lines serving several hundred coal mines through the Appalachian coal region extending from north-central Alabama northeastward into southwestern Pennsylvania. CSXT also accesses major mines located in the Illinois Basin coal region of western Kentucky, southern Illinois, and southwestern Indiana. The locations of these coal regions are shown on Figure 1A.3.4-2.

In 1990, the total coal production in the United States was approximately 1.036 billion tons of which approximately 178.4 million tons (17.22 percent) were originated by CSXT railroads. Including traffic originated by other carriers raises the total coal tonnage handled by CSXT railroads to 186.6 million tons (18.01 percent of total US production). CSXT is the largest rail carrier of domestically produced coal in the United States. Additionally, approximately 24.1 million tons were handled by its subsidiary, American Commercial Barge Line Company on the inland waterways.[15]

CSXT is the dominant transporter of coal by rail into Florida. The movements use heavy-duty mainline trackage extending from both the Appalachian and Illinois Basin coal regions to access all rail-served utility power plants in Florida. These mainline corridors are shown on Figure 1A.3.4-7.

The CSXT rail system for coal movements to Florida utilities comprises two distinct routes--each having several partial alternative routings as discussed below.

1A.3.4.5.2 CSXT Appalachian Rail Corridor. The CSXT Appalachian Rail Corridor extends south from the major coal producing areas of eastern Kentucky with routing via Corbin, Kentucky; Knoxville, Tennessee; and Atlanta, Cordele, and Waycross, Georgia; to enter Florida at a point northwest of Callahan, Florida. This routing also picks up very limited tonnages of coals originating in Tennessee. Because of the large tonnages originating in eastern Kentucky, this routing is the primary corridor for unit train operations south from the Appalachian coal region to supply coal to Georgia and Florida destinations.

A partial alternative routing originates in the Big Sandy and Tug Fork Rivers area on the Kentucky-West Virginia border with routing south via Elkhorn City, Kentucky, and St. Paul and Speers Ferry, Virginia. This routing continues southward from Speers Ferry, Virginia, over former Clinchfield Railroad mainlines via Kingsport and Johnson City, Tennessee; Marion, North Carolina; to Laurens, South Carolina. At Laurens, South Carolina, the routing branches with one line running east to Savannah, Georgia, via Columbia, South Carolina, and the alternative line running south to Savannah via Greenwood, South Carolina, and Augusta, Georgia. After converging at Savannah, the routing continues southwest to rejoin the principal mainline at either Waycross or Folkston, Georgia. This corridor originates coals from mines located in extreme eastern Kentucky and southwestern Virginia.

The above principal corridors are crosstied in the coal regions by two lines extending between Beaver Junction and Blackey, Kentucky (the Rockhouse Creek-Deane Connection) and between Speers Ferry, Virginia, and Hagans, Kentucky (the Old Cumberland Valley Main-Norfolk Southern Connection). However, because of gradient, curvature, and other operational constraints, these crossties are not extensively used in movement to Florida destinations. A further partial alternative routing (which is also little used) extends from Greenwood, South Carolina, to rejoin the principal mainline corridor at Atlanta, Georgia.

A second partial alternative routing originates in the extreme western areas of West Virginia and runs north via Huntington, West Virginia, then west parallelling the south shore of the Ohio River before turning south at either Cattlettsburg or Newport, Kentucky to link up with the above corridors.

The Cattlettsburg to Newport line also serves as a fourth crosstie between the two north-south lines through the east Kentucky mine regions. Coal shipments occasionally move north to this Ohio River line before turning back to the south to relieve traffic blockages or congestion in the primary corridors. Because of its circuitous routing, this crosstie corridor carries minimal coal traffic destined for Florida. The CSXT Appalachian rail corridor originates major tonnages of coal for utilities located in the southeastern United States (North and South Carolina, Mississippi, Alabama, Georgia, and Florida utilities). These tonnages in 1989 totaled approximately 40.6 million tons. The CSXT Appalachian rail corridor also carries the major volume of rail coal destined for Florida (8.53 million tons in 1989), with the bulk of the coal originating in eastern Kentucky

(6.79 million tons) and Virginia (838 thousand tons). Originations from West Virginia mine areas are limited (608 thousand tons) because of higher transport costs associated with the longer haul distances to Florida utilities. These coal mine areas and 1989 origination tonnages for Florida use points are also shown relative to CSXT trackage on Figure 1A.3.4-8.[16]

The rail distance to Stanton Energy Center measured from Corbin, Kentucky, (the center for CSXT rail coal operations in the region) is 808 miles. Distances from Corbin to the various mine origins ranges up to 250 additional miles, depending upon location and routings, with the majority of the movements originating within 105 miles.

1A.3.4.5.3 CSXT Illinois Basin Rail Corridor. The Illinois Basin rail corridor extends from coal source areas in southern Illinois and southwestern Indiana southward--crossing the Ohio River at Evansville, Indiana--to service mines located on lateral branch lines extending west and east in western Kentucky from Madisonville, Kentucky. The corridor continues south via Nashville and Chattanooga, Tennessee, to intercept the Appalachian corridor routing near Cartersville, Georgia. This routing is little used for Florida coal traffic (less than 300 thousand tons in 1989) as the coals from these sources are (1) higher in sulfur content than the Appalachian Region coals and (2) generally move by water (barges on the Ohio, Mississippi, and Tennessee-Tombigbee Rivers) to gulf-port transload points for cross-gulf ocean-barge continuation to west-coast Florida ports.[17] The Illinois Basin coal origination areas are also shown relative to CSXT mainlines on Figure 1A.3.4-8.[17]

The rail distance to Stanton Energy Center measured from Madisonville, Kentucky, is about 923 miles. Distances from Madisonville, Kentucky, to the principal mine loadouts in western Kentucky range from 26 to 31 miles (in additional distance) and up to 90 miles for Illinois and Indiana sources.

1A.3.4.5.4 CSXT Coal Mainlines Within Florida. After entering Florida, the CSXT mainline divides at Callahan, Florida. Most of the traffic (approximately 5.35 million tons in 1989) moves southwestward to Starke, Florida, with the remainder (3.18 million tons) continuing south to Jacksonville, Florida. From Starke and Jacksonville, the coal movements branch to various utility power plants as depicted on Figure 1A.3.4-9.

Note that a significant tonnage of coal (3.03 million tons) also enters Florida by intracoastal barges at Port St. Joe, Florida. At Port St. Joe, the coal is transferred to unit trains and transported via the Apalachicola Northern Railroad (Port St. Joe to Chattahooche, Florida and CSXT (Chattahooche to Jacksonville, Florida) to intercept the primary corridor to Orlando, Florida at Jacksonville. This movement adds significant tonnage between Jacksonville and the Seminole Electric Cooperative, Inc., Palatka Plant (about 6 miles north of Palatka, Florida) to the rail corridor serving Stanton Energy Center.

A nominal tonnage of coal (340.6 thousand tons in 1989) is transported by CSXT from mines in western Kentucky to Gulf Power Company's Scholtz Plant west of Chattahoochee, Florida. This rail movement uses the CSXT Illinois Basin Rail Corridor, described in Subsection 1A.3.4.5.3, from the mines to Nashville, Tennessee. From Nashville, the routing continues south via Montgomery, Alabama, to Pensacola, Florida, and then east over the CSXT Panhandle mainline to the plant (Boykin, Florida). As this routing is totally detached from the CSXT coal routes to Florida utilities as outlined above, it has no impact on coal movements to Stanton Energy Center and is not depicted in Figure 1A.3.4-9.

1A.3.4.5.5 CSXT Track Standards and Current Traffic Levels. The CSXT mainlines identified above for the transport of coal to Florida are all heavy-haul corridors with facilities in place for the efficient and expeditious movement of unit coal trains. All rail lines are fully signaled with single- or double-line tracks. Installed trackwork is generally heavy-section rail (131 pounds per yard or higher) with high quality ties and ballast. Continuous welded rail (CWR) is in place in essentially all corridor segments carrying coal traffic to Stanton Energy Center. Exceptions include a short segment between Shelby and Elkhorn City, Kentucky, Big Stone Gap, Virginia, and Hagans, Kentucky, Kearney and Fairfax, South Carolina (via Augusta, Georgia) and south of Jacksonville, Florida (to approximately Orange Park, Florida). These segments are laid with intermediate rail sections (112 pounds per yard or higher weight in both jointed and CWR segments) and are likely to be upgraded to heavy-section CWR in the near future. Trackwork is maintained to high standards of line and grade to facilitate both general merchandise freight as well as unit train coal movements to and from the Georgia-Florida region.[18]

Traffic levels carried by the CSXT Appalachia to Florida mainlines are moderate to heavy in density. Levels in the Manchester, Kentucky, to Callahan, Florida, corridor ranged from 27.4 to 72.2 million gross tons in 1989, with the majority of the line segments carrying 45 to 65 million gross tons. Traffic direction is approximately 2 to 1 southbound versus northbound as could be expected for both coal and merchandise shipments bound to southeastern destinations.[19]

Traffic over the secondary CSXT Appalachian rail corridor to Florida (Beaver Jct., Kentucky, to Waycross, Georgia) is considerably lower, ranging between 14.4 and 37.7 million gross tons in 1989, with the bulk in the 25 to 35 million gross tons levels. As in the primary corridor, traffic is predominately southbound at ratios of roughly 2.5:1.

Traffic on the Evansville, Indiana, to Junta, Georgia, mainlines from Illinois Basin mines ranged from 23.8 to 38.0 million gross tons in 1989 and is also in a 2:1 southbound to northbound ratio.

The use of multiple north-south corridors between the coal source regions and southeastern US utility destinations allows the dispersal of movement flows to alleviate any "bottlenecks" in train operations. Present levels of coal traffic (when combined with non-coal traffic) are well within the capacity of the corridor line facilities as currently signaled and maintained. Movements generally encounter minimal delays and "fleeting" of trains to increase operational throughput are not used except in localized areas and for short durations.

1A.3.4.5.6 Projections of Future CSXT Rail Traffic. Projections of future trends in coal traffic on CSXT mainlines serving Florida utilities are dependent upon a number of major factors:

- Development of new coal-fueled generation facilities in the southeastern United States (primarily in Florida) having (or limited to) rail delivery for fuel.
- Extent and nature of fuel-switching (Illinois Basin to Appalachian coal region) employed by rail-dependent southeastern US utilities (principally the Southern Company operating subsidiaries) in response to Phase I compliance with 1990 Clean Air Act Amendments sulfur constraints (Phase I emission limits of 2.5 lb SO_2/MBtu by 1995). A related fuel-switching issue is the extent (tonnages) of market penetration of western United States (Powder River Basin) and foreign origin low-sulfur coals at specific utility plants in the southeastern United States.
- Type and nature of combustion and emissions control systems for coal-fueled facilities to be developed and installed to comply with Phase II limits (1.2 lb SO_2/MBtu by the year 2000) and the degree to which

- utilities return to "traditional" (pre-Clean Air Act Amendments of 1990) source areas for coal supply.

These issues are complex, interrelated, and subject to significant uncertainties in forecasting the future. Nevertheless, aspects of the above factors which will influence future CSXT coal traffic levels and operations are summarized below.

1A.3.4.5.6.1 Future Coal Capable Generation Projects in Florida. Coal capable power plants which have either received need for power certification or have had need for power certification hearings held before Florida Public Service Commission include the following.

	Net Capacity MW	Average Coal Use million tpy
FPL Martin Units 3 and 4 (Future GCC)	770	2.3
Indiantown Cogeneration, LP (PC)	330	1.0
AES Cedar Bay Cogeneration (CFB)	250	1.0

The coal sources for the above projects are forecast as being within the traditional coal supply areas (Appalachian and Illinois coal regions) presently used by Florida utilities. Transport of coal is projected as employing the CSXT system except for the FPL Martin Plant which will be served by both CSXT and FEC rail lines.

Stanton 2 is projected to consume an average of about 1.1 million tons of coal per year. Under the most probable scenario, this tonnage will be transported by unit trains along the Corbin-Jacksonville-Orlando CSXT rail corridor employed by Stanton 1 movements.

The impacts of additional coal traffic on the CSXT system generated by the above projects are speculative as they depend upon project in-service schedules, coal sourcing points, and procurement strategies which are not fully definable at this time. However, for purposes of this study, the following factors are assumed.

- All projects are commercial at forecast coal usages by year 2000.
- The FPC Martin and AES Cedar Bay plants are forecast as primarily sourcing higher-sulfur coals from the Illinois Basin coal region owning to advanced sulfur emissions reduction technologies inherent in the coal gasification and fluidized bed combustor technologies to be employed by these projects.

- The Indiantown and Stanton 2 projects are forecast as sourcing lower-sulfur coals from CSXT-served Appalachian coal region mines.

Assuming the development and operation of all of the above power generation projects in accord with the above projections, the resulting rail movements of coal over the Florida rail system by the year 2000 are shown on Figure 1A.3.4-10.

It should be noted that the CSXT Jacksonville-Orlando rail corridor is anticipated as accommodating only increased coal deliveries for Stanton 2. All other increases in Florida coal traffic are forecast as being routed via Callahan, Baldwin, Ocala, and Lakeland, Florida, to the various plants.

Additional coal-fueled generation projects to meet electrical needs in the late 1990s and later are in varying stages of the planning process for other locations within Florida. These projects have not progressed sufficiently to identify unit capacities, combustion technologies, and coal sourcing-transport linkages at this time.

1A.3.4.5.6.2 Future Coal Switching Strategies for Southeastern United States Utilities. The enactment of the Clean Air Act Amendments of 1990 has created a major degree of uncertainty with respect to future sourcing of coals for utilities located in the southeastern US. Virtually all utilities operating plants which lack modern flue gas SO_2 emissions control systems are evaluating the options of switching to lower sulfur sources versus the installation of flue gas desulfurization systems.

Scrub versus switch decisions are complex and dependent upon the interaction of site-specific variables and compliance with implementation regulations which are yet to be issued. The 1990 Clean Air Act Amendments identify 13 power plants located in the southeastern US (Alabama, Mississippi, Tennessee, and Florida) which will require facilities or operational modifications to comply with Phase I standards.[20] Four of the power plants are exclusively served by river barges or ocean vessels and are projected as exclusively continuing that delivery mode in future coal supply scenarios. The nine remaining power plants are served by rail either as captive destinations or as an alternative to barge deliveries. Five of the nine plants use Norfolk Southern (NS) as the delivering rail carrier. Four plants have CSXT service either as a captive destination (GPC Plant Bowen), joint service destination with NS (GPC McDonough Plant), or as an alternate to barge service as the primary coal transport mode (TVA Gallatin and

Johnsonville Plants). A forecast overview of the impacts on the coal traffic levels of the CSXT Appalachian to Florida rail corridors resulting from potential compliance strategies is summarized below.

The Alabama Power Company's E. C. Gaston Plant (jointly owned with Georgia Power Company) located near Wilsonville, Alabama, is projected to employ fuel switching for Phase I compliance. The plant is solely served by NS as the delivering carrier for coals originating primarily in Alabama and NS-origin areas of Tennessee and Kentucky. The switching in sourcing is anticipated as de-emphasizing the higher sulfur Alabama source regions in favor of lower-sulfur coals originating in NS-served areas of Tennessee, eastern Kentucky, and western Virginia. Impacts of these coal sourcing initiatives should be minimal with respect to CSXT-served coal areas and CSXT Kentucky to Florida coal traffic patterns.

The Tennessee Valley Authority (TVA) has two power plants (Gallatin and Johnsonville Plants) which are served by CSXT; however, these plants are also served by barge service from the Cumberland and Tennessee Rivers, respectively. The preponderance of coal received at these plants is delivered by barges originated on the Ohio River system in both the Illinois River and Central Appalachian coal regions.

TVA is forecast as employing a strategy to either switch to Powder River Basin coal or to retrofit flue gas desulfurization systems at their larger and more modern steam plants (Paradise and Cumberland) to free up sufficient sulfur dioxide allowances for use at other system plants under "substitution" provisions of the 1990 Clean Air Act Amendments.[21] This should allow continuation of coal procurements generally from historic source areas for all remaining TVA Phase 1 plant's with minimal switching to lower-sulfur coals within each area. In any case, any impacts of changes of coal procurements made by TVA should have only minor influence on market competition within the traditional east Kentucky, West Virginia, and southwestern Virginia coal regions and negligible impacts on rail coal traffic patterns for coals moving south to supply Georgia, North and South Carolina, and Florida utilities.

Georgia Power Company (GPC) has five power plants identified as "affected sources" under the 1990 Clean Air Act Amendments. Three of these plants (Hammond, Wansley, and Yates Plants) are served exclusively by the NS, one plant (Plant Bowen) is served solely by CSXT, and one plant (John McDonough Plant) is jointly served by both NS and CSXT. The strategy to be employed by

Georgia Power Company to meet Phase I requirements (while still under development) is forecast as probably incorporating the following elements.

- No installation of flue gas desulfurization at any existing power plant units.
- Integration of fuel switching strategies for sulfur emissions compliance with companion fuel switching strategies to favor system plants having the lowest fuel costs (i.e., "marginal fuel" method for dispatching coal-fueled generation implemented in January 1991).
- Switching of units having high fuel costs (principally Plant Scherer) to low-sulfur Powder River Basin source coals to both reduce costs and create sulfur dioxide allowances for transfer to other units in the GPC system.
- Higher levels of dispatch from other GPC power plants having both lower-sulfur level coal sourcing and delivered coal costs (i.e., Plants Bowen and Branch served by CSXT) at the expense of higher fuel cost plants (Hammond, Wansley, and Yates Plants served solely by NS).

Georgia Power Company initiated the above strategies in the second half of 1990 by switching procurements at certain plants (i.e., McDonough, Wansley, and Yates) from higher-sulfur Illinois Basin sources to lower-sulfur central Appalachian sources to the extent permitted by long-term contractual commitments. GPC also initiated in late 1989 a test firing program to evaluate Powder River Basin (Wyoming origin) low-sulfur coals to reduce high contractual costs of West Virginia origin coals at Plant Scherer and as blend coals with high-sulfur coals at McDonough, Wansley, and Yates Plants. These test firing programs are continuing and may lead to a significant tonnage of Powder River Basin coal (to 12 million tons per year) being procured for GPC plants for the near term.

The Powder River Basin coal is projected as backing down coal consumption from both Appalachian and Illinois Basin sources at GPU's plants having higher fuel costs (i.e., Branch, Hammond, Wansley, and Yates) all served by NS as the delivering rail carrier. These Powder River Basin rail movements will probably employ the NS as the delivering carrier from interchanges with western railroads at either Memphis, Tennessee (Union Pacific System) or Birmingham/Pride, Alabama (Burlington Northern Railroad). Plants served solely by CSXT (Bowen and Mitchell) are forecast to be only minimally impacted by any Powder River Basin coal procurements through 1995. Plants served jointly by CSXT and NS

(Branch and Atkinson-McDonough) are forecast as experiencing modest reductions in deliveries of Appalachian origin coals back-down of generation dispatch by other plants burning lower cost coals. Practically all losses in coal consumption are forecast as being incurred by NS-served mine areas and rail movements.

CSXT-serviced power plants located within North and South Carolina and Florida are either equipped with flue gas desulfurization systems or presently burn low-sulfur coals. These plants are in compliance with the 1990 Clean Air Act Amendments Phase I standards. These plants are forecast as continuing their historic coal sourcing practices, thereby imposing minimal changes in rail coal traffic patterns within the CSXT Appalachian rail corridors to Florida.

1A.3.4.5.6.3 Future Coal Procurement Strategies for Compliance with 1990 Clean Air Act Amendments Phase II Standards. Existing power plants will be required to comply with SO_2 emission limitations of 1.2 lb SO_2/MBtu by the year 2000 under Phase II standards of the Clean Air Act Amendments of 1990. These standards are forecast to result in the following compliance and fuels procurement strategies for southeastern US utilities in the late 1990s:

- Retrofit installation of flue gas desulfurization systems on existing generation units lacking SO_2 removal capability. In the absence of significant breakthroughs in reducing capital and operating costs of the desulfurization technologies, the systems will probably be installed at the larger capacity and newer/most efficient generating units.
- Low-sulfur (less than 1.2 lb SO_2/MBtu) coals will be switched from the above plants and used for older and smaller units operating in standby/ peaking service or, alternatively, the units will be retired.
- Powder River Basin coals will maintain market share established in the 1991 to 2000 period because of abundant reserves and mine capacity, favorable production and transport economics, and low-sulfur content (0.75 to 1.1 lb SO_2/MBtu) of this coal.

Most new plants constructed after the late 1970s and furnished with BACT emissions control technologies will be capable of using higher-sulfur content coals. This will help restore the use of some (although not all) of the historical higher-sulfur coal sources within the Illinois Basin and Appalachian coal regions. New coal cleaning and combustion technologies (most notably integrated coal gasification combined cycle) capable of meeting high emissions standards and environmental-resource usage constraints are forecast as appearing after year 2000. The development and maturation of these technologies will also tend to restore traditional coal sourcing practices.

1A.3.4.5.6.4 Summary of CSXT Rail Traffic Projections. The projections of future trends in coal traffic of CSXT mainlines between the anticipated source area (Central Appalachian coal region) and Stanton Energy Center are for modest increases through the year 2000 because of new rail-served projects (including Stanton 2) within Florida and slight growth in Savannah Electric and Georgia Power Company coal use for CSXT destination power plants.

Traffic projections after 2000 are more conjectural, depending upon the location and unit sizes of new coal-capable plants to be built in Georgia, North and South Carolina, and Florida and the market share lost by CSXT to water-borne foreign origin (or Tidewater US origin) coals and Powder River Basin origin coals delivered to Florida plant and port locations.

In view of the track capacities remaining above present-day traffic levels within the CSXT Appalachia to Florida rail corridors, no significant constraints on the efficient and timely movement of unit trains from either Illinois Basin or Central Appalachian coal regions to Stanton Energy Center are envisioned for the foreseeable future.

Table 1A.3.4-1
1989 Central Appalachia Coal Deliveries to US Utilities

State	Quantity (1,000 tons)	Heating Value (Btu/lb)	Sulfur (percent)	Sulfur ($lb\ SO_2/MBtu$)	Ash (percent)
East Kentucky	81,162	12,375	1.1	1.7	9.7
Northern Tennessee	4,019	12,587	1.4	2.3	9.6
West Virginia	17,965	12,894	1.1	1.8	9.4
Southwestern West Virginia	40,562	12,385	0.8	1.3	10.9
Total or Average	143,708	12,449	1.0	1.6	10.0

Source: Resource Data International.

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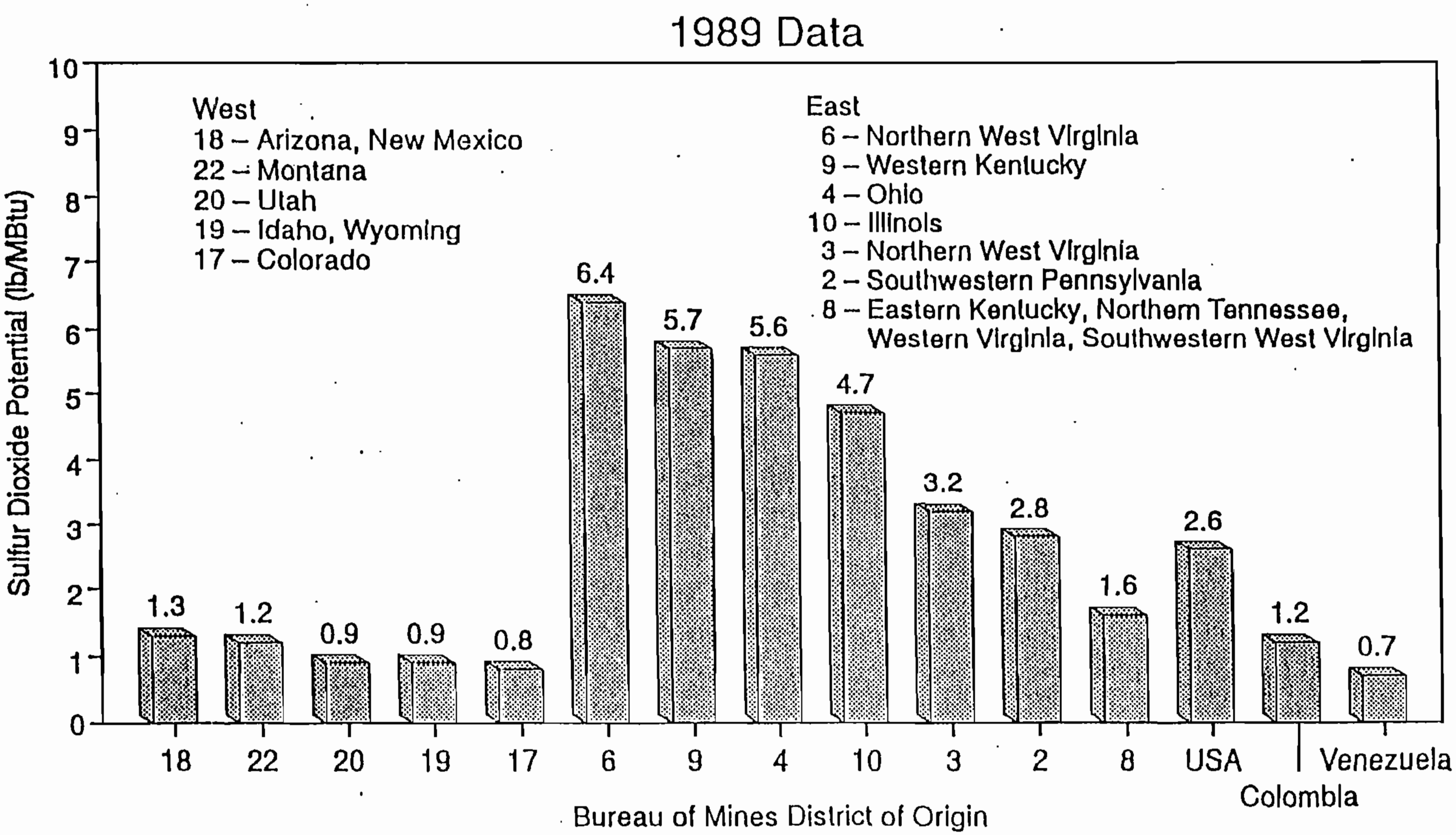
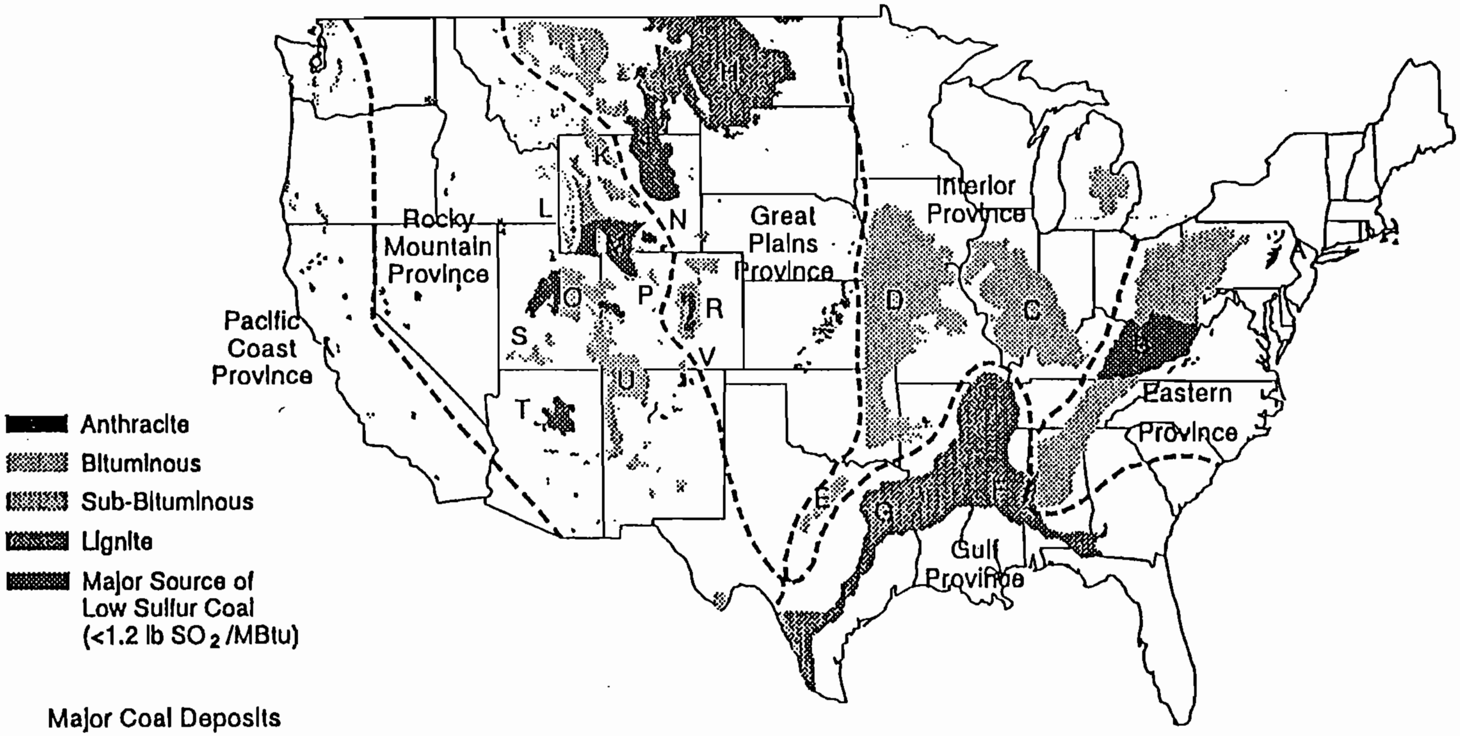


Source: Energy Information Administration, Cost and Quality of Fuels for Electric Utility Plants, 1989

WEST VS EAST – SO_2 EMISSION POTENTIAL

Figure 1A.3.4-1

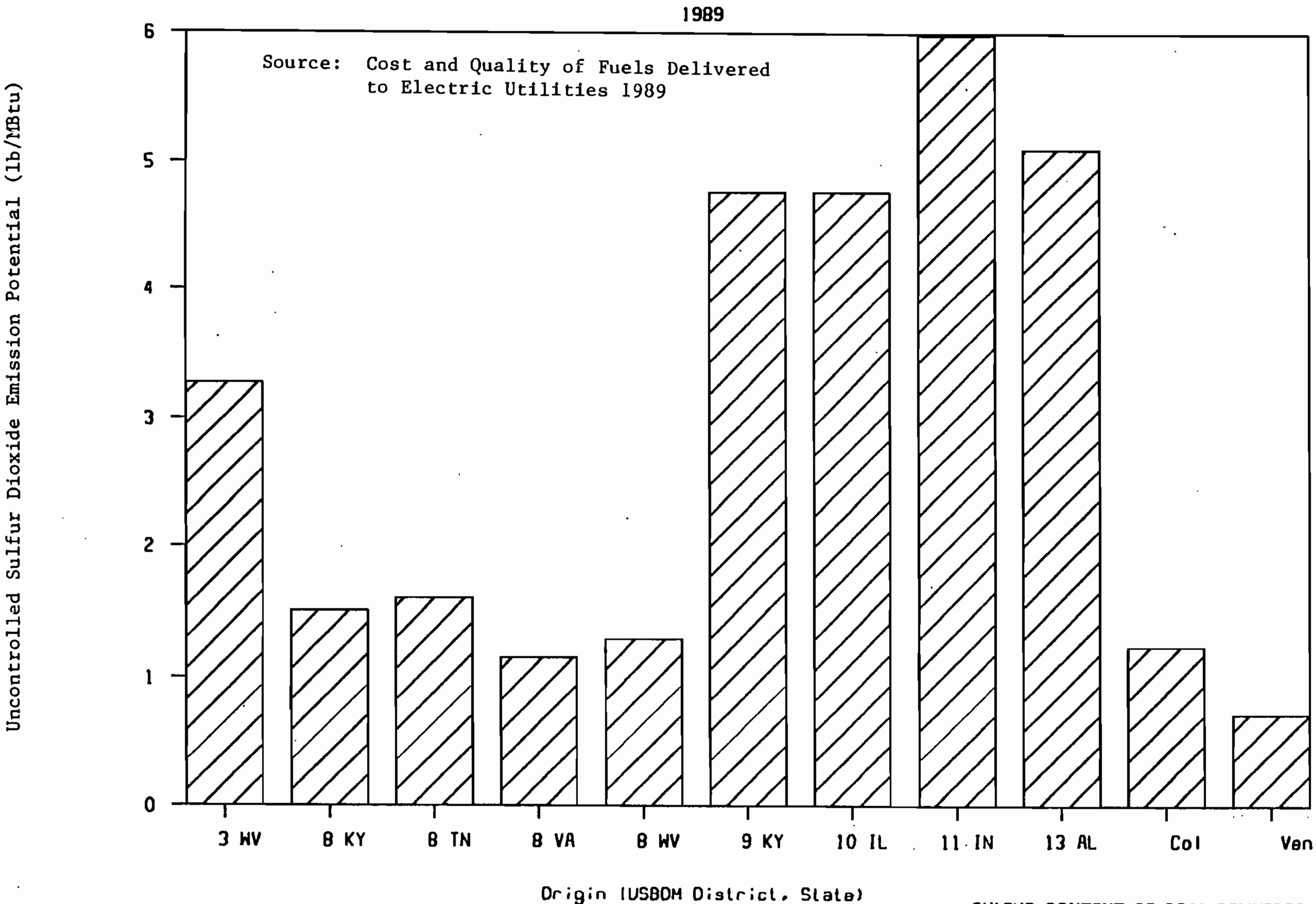


Major Coal Deposits

A - Pennsylvania Anthracite	F - Mississippi Region	L - Hams Fort Region	R - Denver Basin
B - Appalachia Region	G - Texas Region	M - Green River Region	S - Southwestern Utah Fields
C - Illinois Basin	H - Fort Union Region	N - Hanna Field	T - Black Mesa Region
D - Mid-Continent Region	J - Powder River Basin	O - Uinta Region	U - San Juan River Region
E - Southwestern Region	K - Bighorn Basin	P - North Park Region	V - Raton Mesa Region

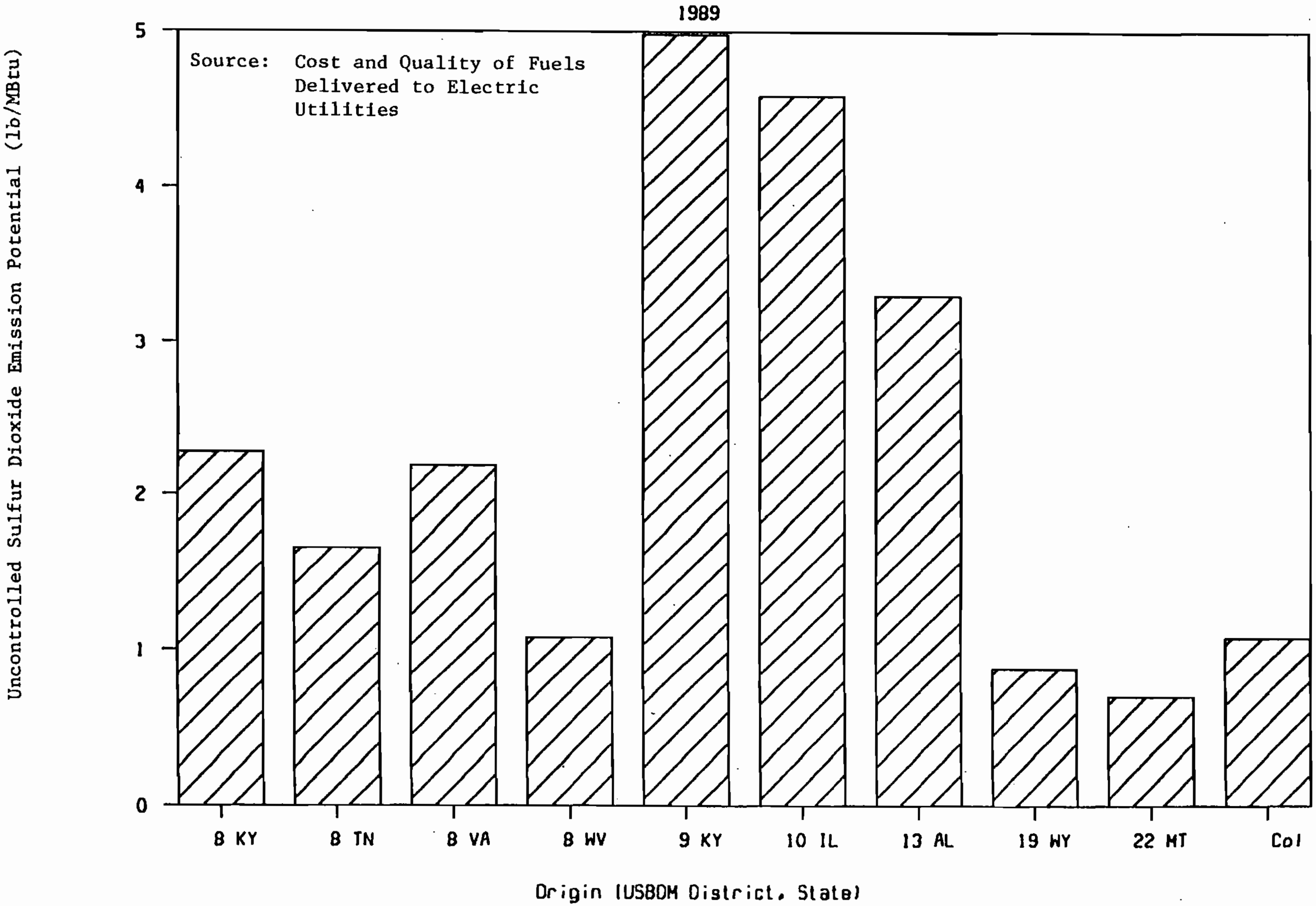
COAL FIELDS OF THE CONTERMINOUS UNITED STATE

Figure 1A.3.4-2



SULFUR CONTENT OF COAL DELIVERED
TO FLORIDA ELECTRIC UTILITIES

Figure 1A.3.4-3



SULFUR CONTENT OF COAL DELIVERED
TO GEORGIA ELECTRIC UTILITIES

Figure 1A.3.4-4

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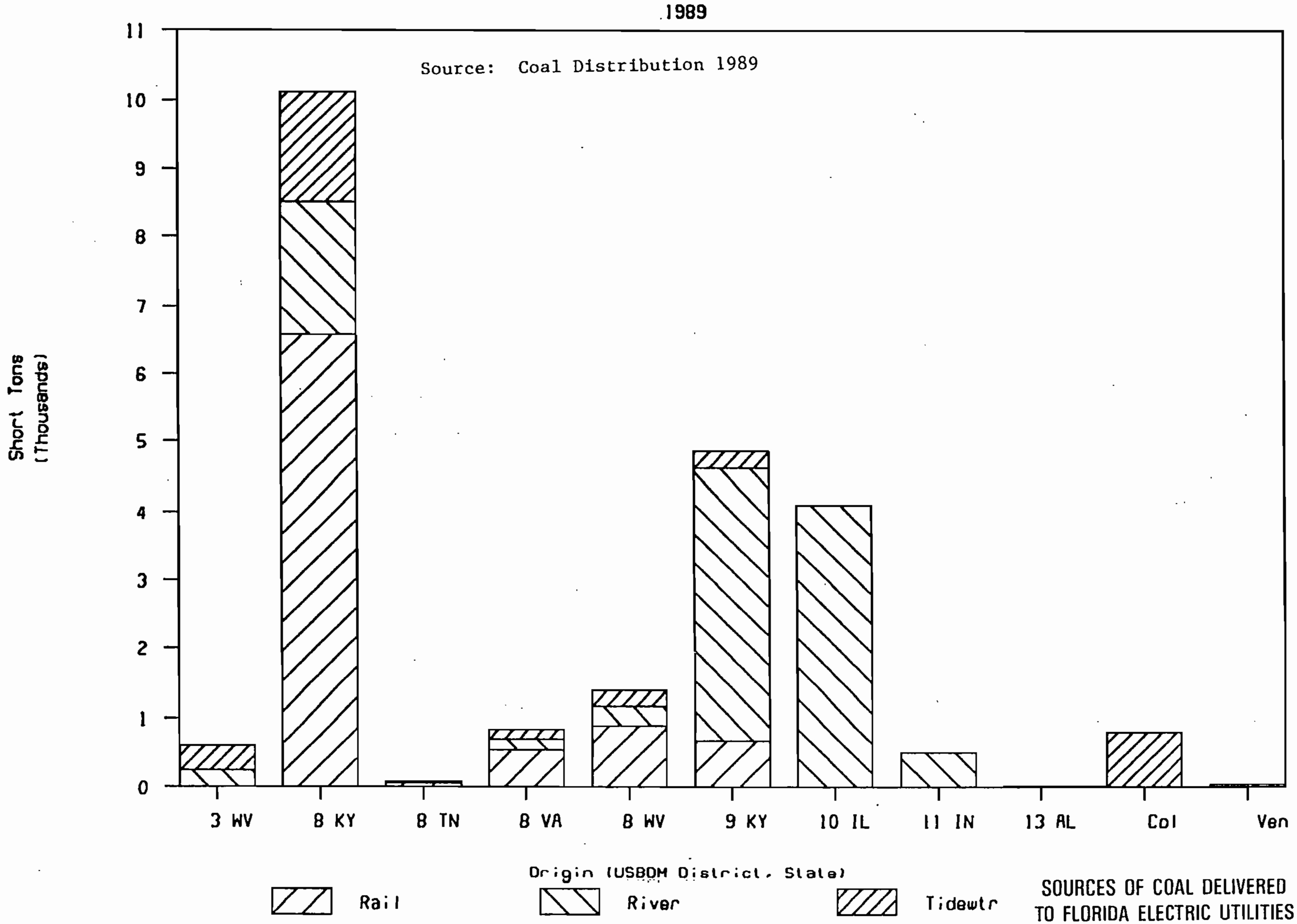
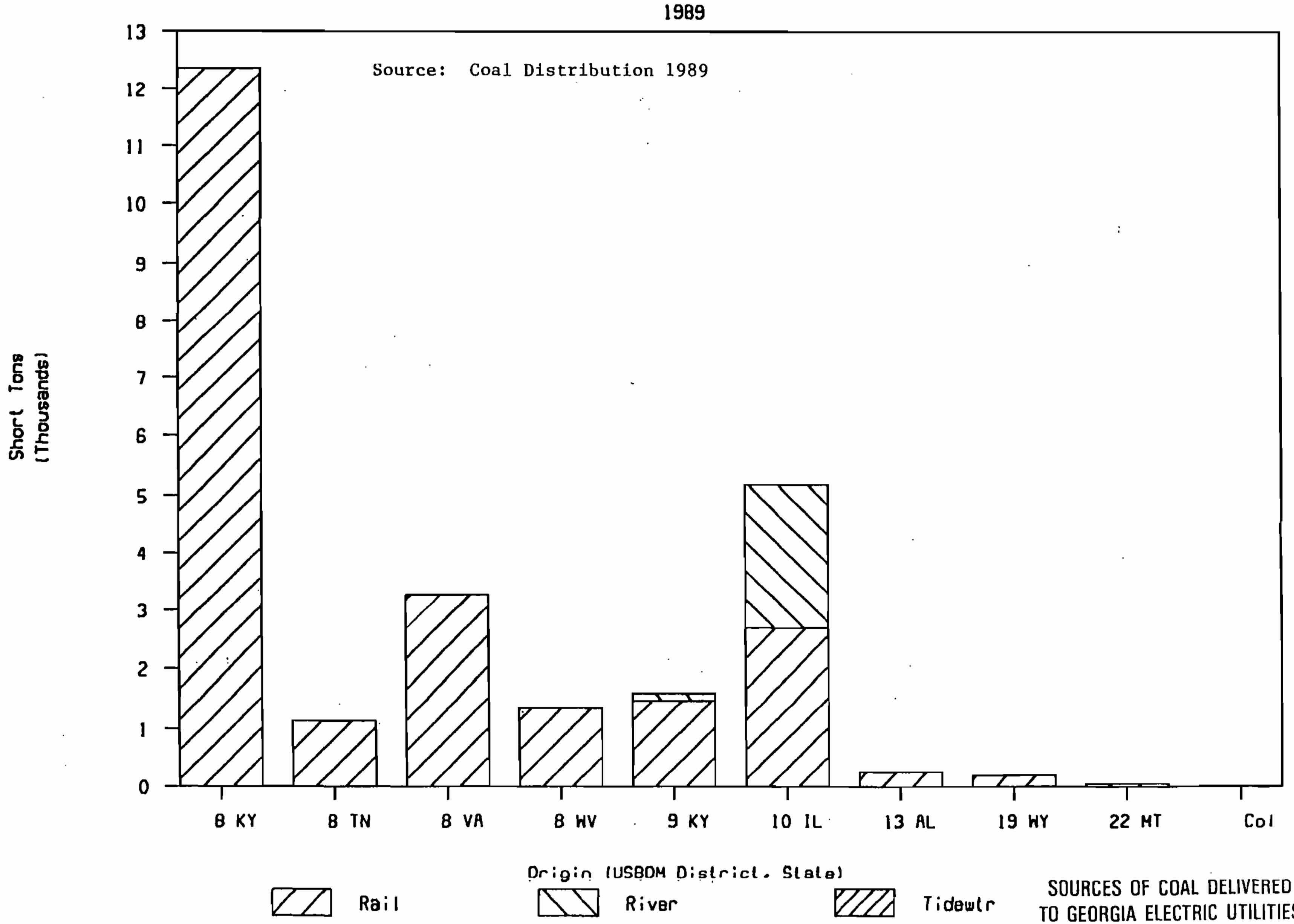
1A.3.4-29



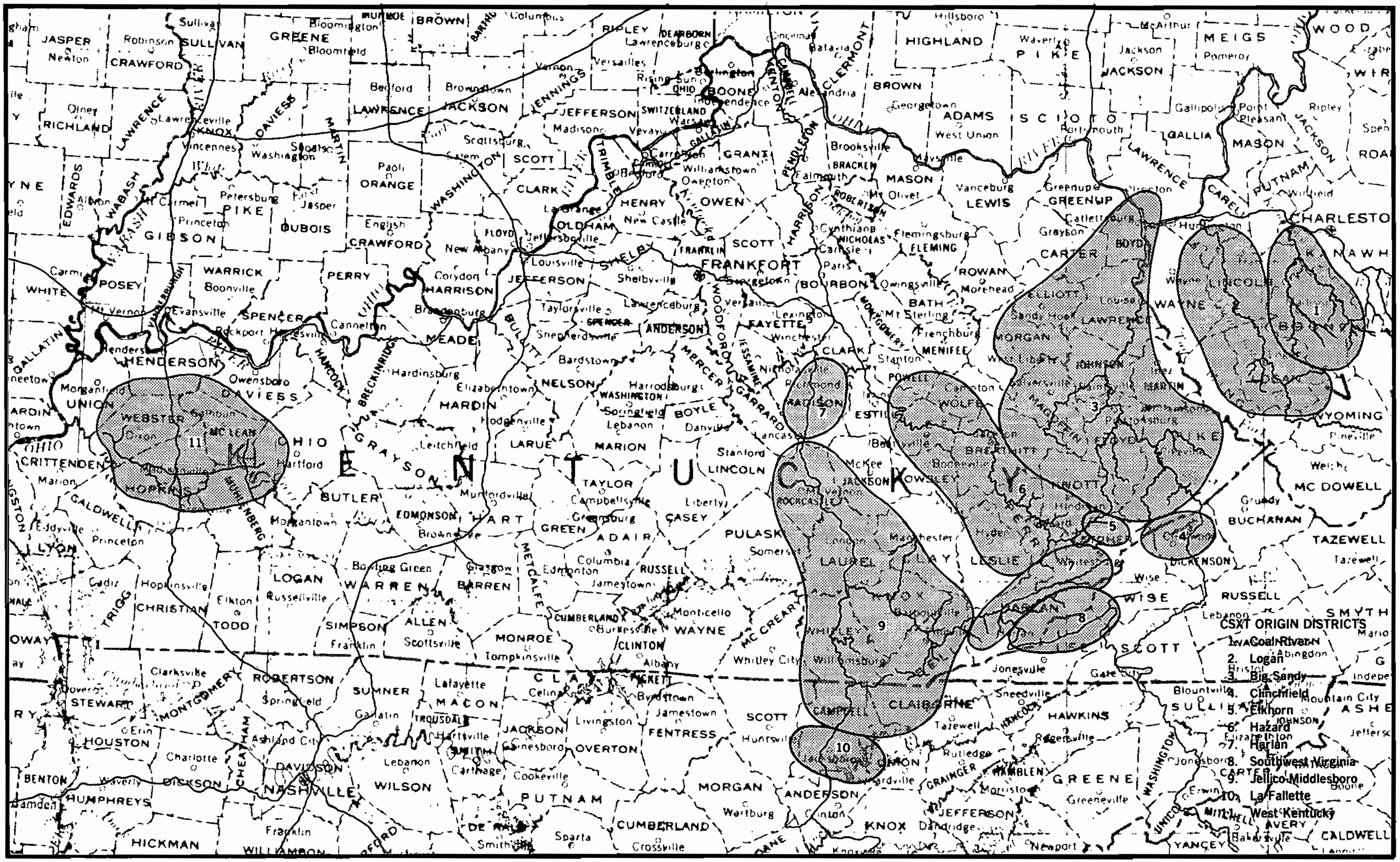
SOURCES OF COAL DELIVERED TO FLORIDA ELECTRIC UTILITIES

Figure 1A.3.4-5



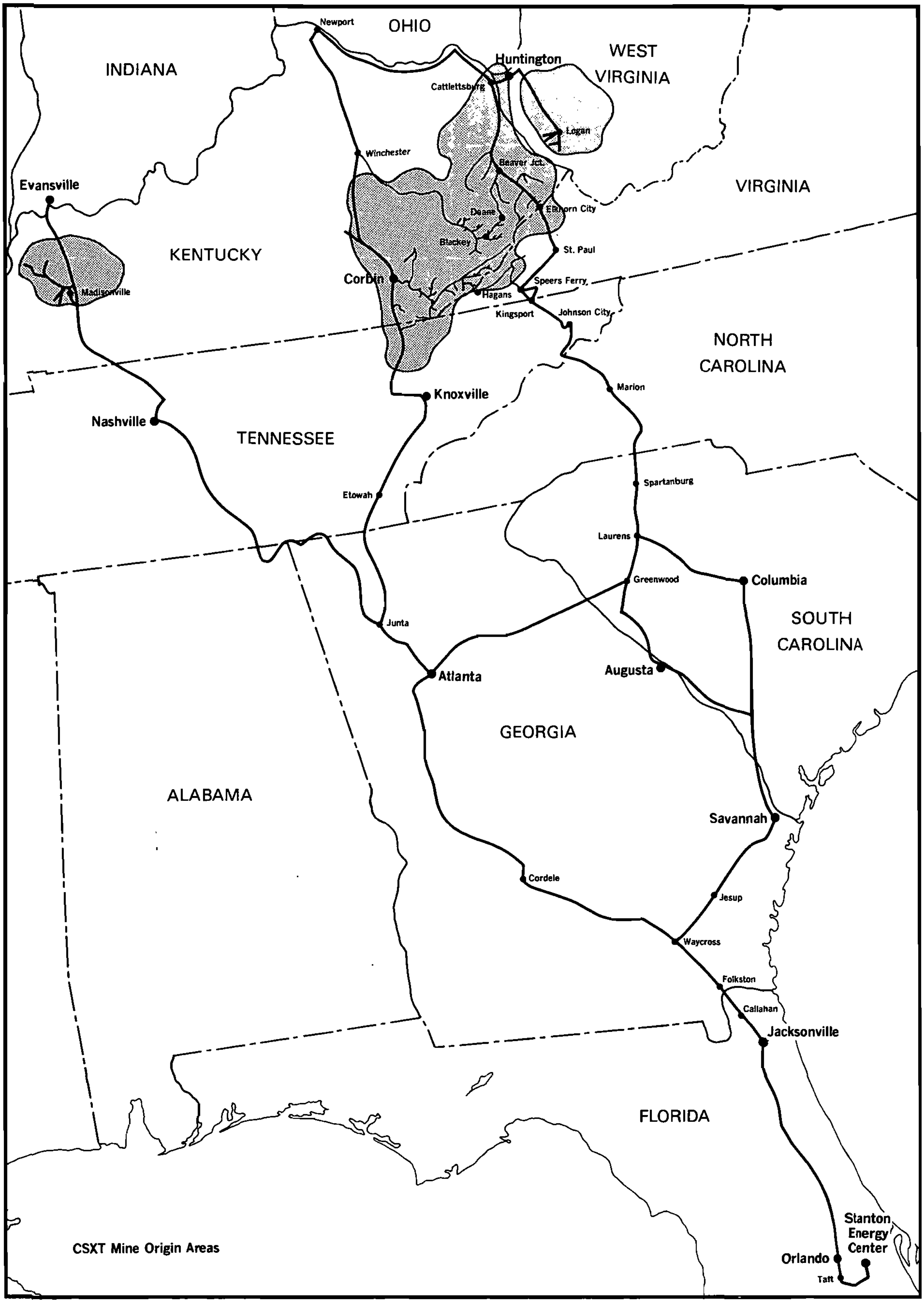
SOURCES OF COAL DELIVERED TO GEORGIA ELECTRIC UTILITIES

Figure 1A.3.4-6



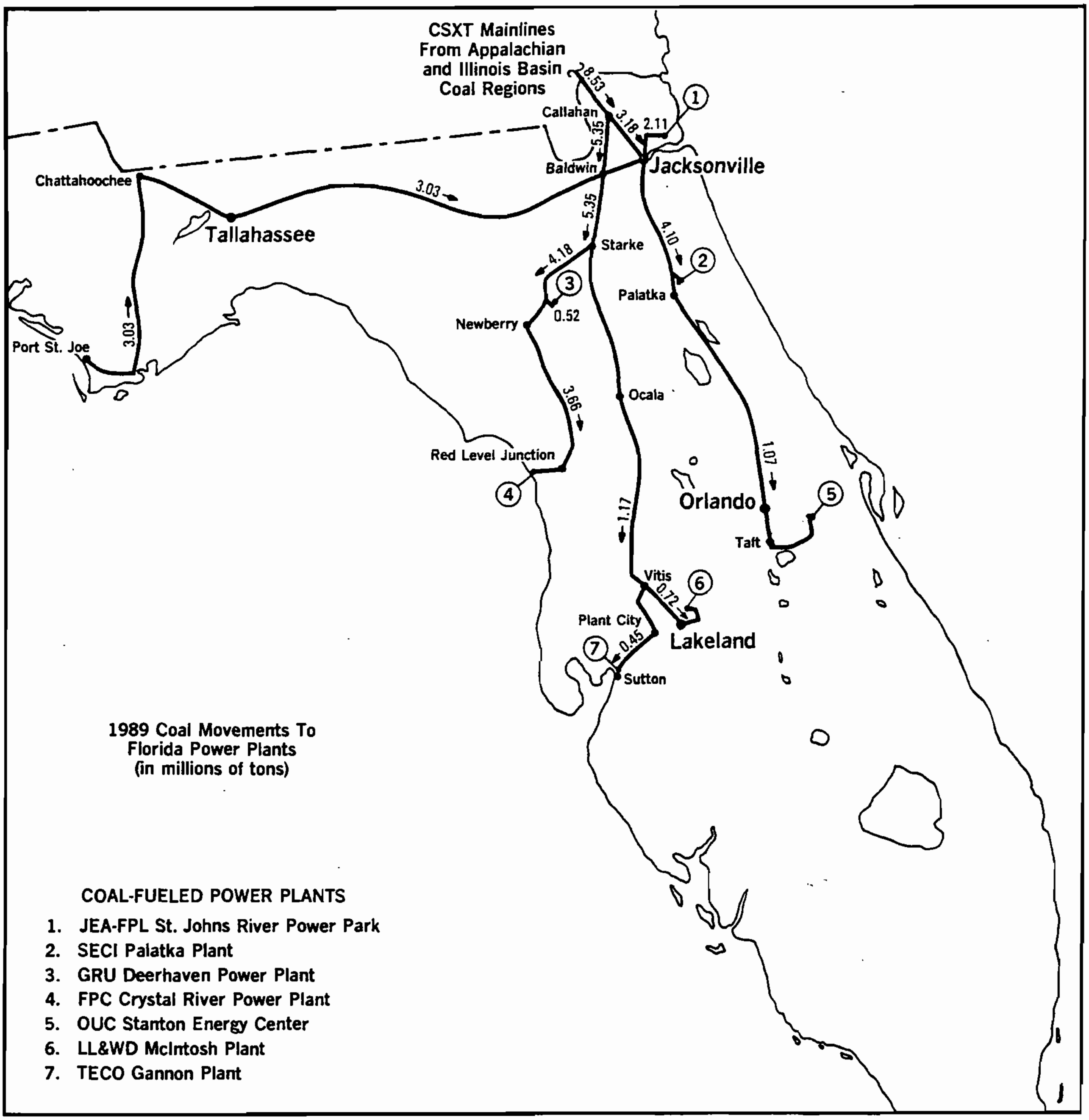
CSXT RAIL LINES AND MINE AREAS SUPPLYING COAL TO GEORGIA AND FLORIDA UTILITIES

FIGURE 1A.3.4-8



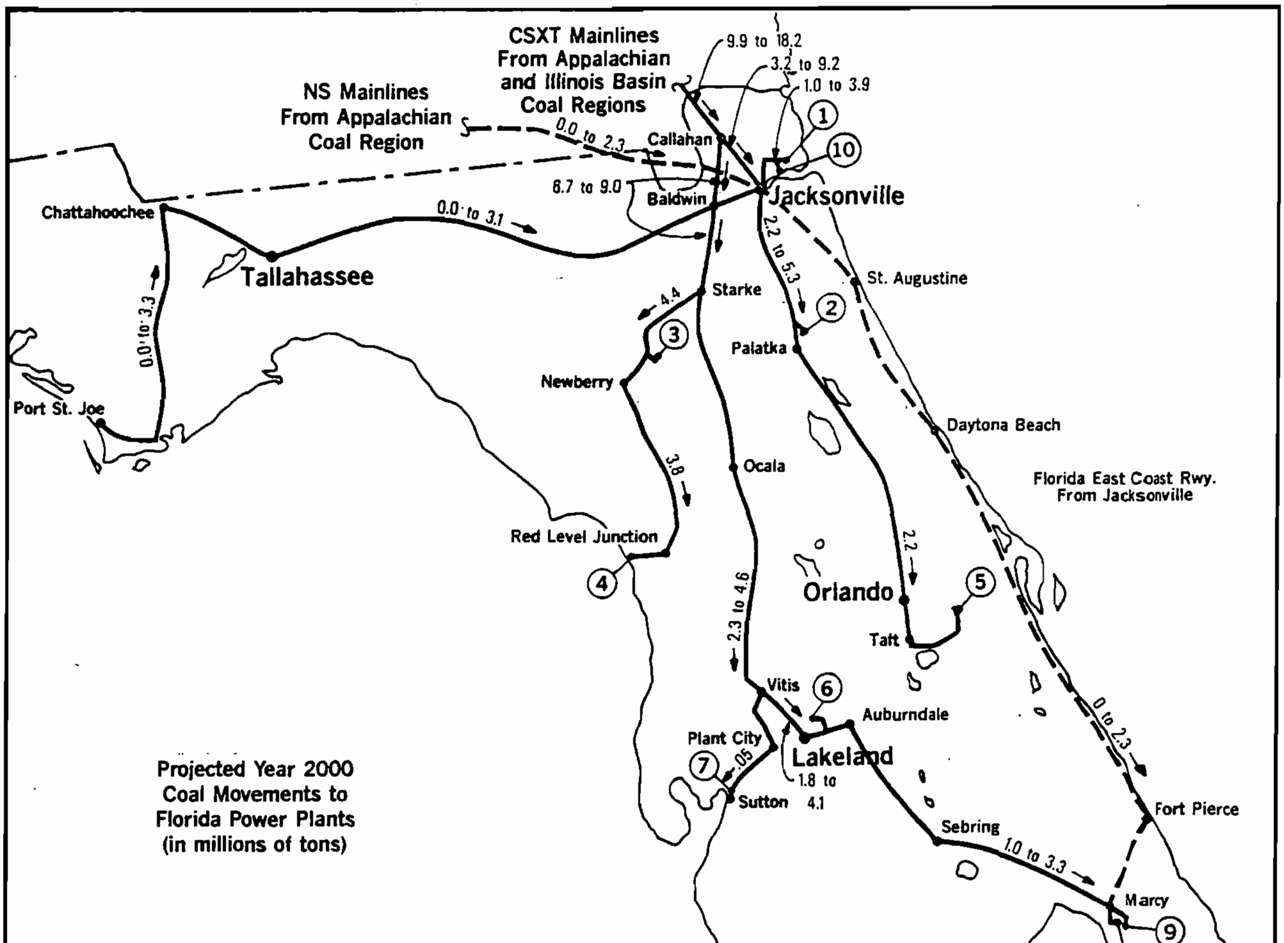
CSXT RAIL CORRIDORS BETWEEN APPALACHIAN AND ILLINOIS BASIN COAL REGIONS AND STANTON ENERGY CENTER

FIGURE 1A.3.4-7



CSXT RAIL CORRIDORS WITHIN FLORIDA

FIGURE 1A.3.4-9

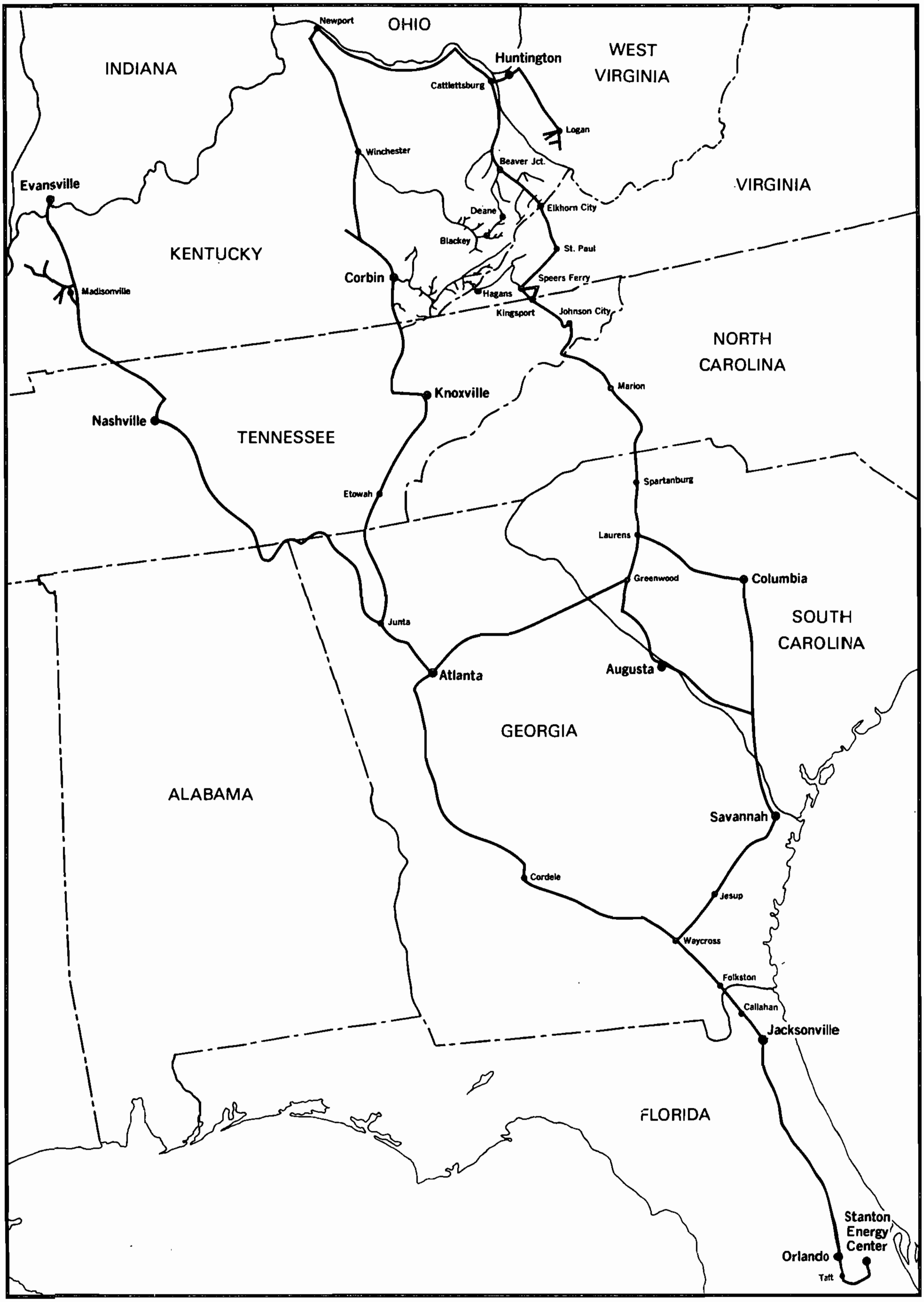


COAL-FUELED POWER PLANTS

1. JEA-FPL St. Johns River Power Park
2. SECI Palatka Plant
3. GRU Deerhaven Power Plant
4. FPC Crystal River Power Plant
5. OUC Stanton Energy Center
6. LL&WD McIntosh Plant
7. TECO Gannon Plant
8. FPL Martin Plant
9. Indiantown Cogeneration Plant
10. AES Cedar Bay Cogeneration Plant

CSXT AND FEC RAIL CORRIDORS WITH FLORIDA SERVING POWER PLANTS IN YEAR 2000

FIGURE 1A.3.4-10



FUTURE COAL TRAFFIC LEVELS IN
CSXT RAIL CORRIDORS BETWEEN APPALACHIAN AND
ILLINOIS BASIN COAL REGIONS AND STANTON ENERGY CENTER

FIGURE 1A.3.4-11

1A.3.5 References

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15. Energy Information Administration, "Preliminary 1990 Coal Production," US Department of Energy, January 1991 and personal communication with Marshall H. Bowen, Director, Utility Coal Marketing of CSXT, January 4, 1991.

16. Federal Energy Regulatory Commission (FERC) Form 423 monthly filings by Florida utilities by 1989 as compiled in the "Coaldat" data base published by Resources Data International, Inc.

17. FERC Form 423 filings (1989) for Florida utilities.

18. Personal communications with CSXT coal and facilities engineering personnel.

19. CSXT "Tonnage Chart for Year 1989," July 3, 1990.

20. "Table A - Affected Sources and Units in Phase I and Their Sulfur Dioxide Allowances," Section 404 of Title IV, Acid Deposition Control, Clean Air Act Amendments of 1990, P.L. 101-549 (November 14, 1990). The 13 power plants identified and their mode/rail carrier for coal deliveries are as follows:

Alabama	
TVA Colbert 1-5	Tennessee River MP 245.3 NS (after 1991)
APC E.C. Gaston 1-5	NS
Florida	
TECO Big Bend 1-3	Tampa Bay Ship Channel
GPC Crist 6 and 7	Escambia River MP 200.3
Georgia	
GPC Bowen 1-4	CSXT
GPC Hammond 1-4	NS
GPC J. McDonough 1 and 2	CSXT/NS
GPC Wansley 1 and 2	NS
GPC Yates 1-7	NS

Tennessee

TVA Allen 1-3	Mississippi River MP 720.3
TVA Cumberland 1 and 2	Cumberland River MP 104.0
TVA Gallatin 1-4	CSXT and Cumberland River MP 243.0
TVA Johnsonville 1-10	CSXT and Tennessee River MP 100.0

No plants located in North and South Carolina are included in the Table A listing.

21. Section 404(b) of Title IV, Acid Deposition Control, Clean Air Act Amendments of 1990, P.L. 101-549 (November 14, 1990).

1A.5.0 Supply-Side Alternatives

1A.5.1 Advanced Alternatives

This section presents summaries of advanced generation technologies and an evaluation of their applicability. Detailed descriptions of these technologies, along with references for the information sources, are presented in Appendix 1A.A. The assessment includes the following technologies.

- Coal Fueled Alternatives.
  - Gasification Combined Cycle.
  - Pressurized Fluidized Bed Combustion.
  - Advanced Pulverized Coal.
  - Gasification Fuel Cells.
  - Gasification Humid Air Turbine Cycle.
  - Coal Liquefaction.
  - Magnetohydrodynamics.
- Oil or Gas Fired Alternatives.
  - Steam Injected Combustion Turbine.
  - Humid Air Turbine.
  - Fuel Cells.
- Nuclear Alternatives.
  - Advanced Passive Light Water Reactors.
  - Modular Pressurized Heavy Water Reactor.
  - Module High-Temperature Gas-Cooled Reactor.
  - Liquid Metal Reactor.
  - Fusion.
- Renewable Energy Alternatives.
  - Wind Energy.
  - Solar Photovoltaic.
  - Solar Thermal.
  - Ocean Thermal.
  - Ocean Wave.
  - Ocean Tidal.
  - Geothermal.

- Energy Storage.
  - -- Battery.
  - -- Compressed Air.
  - -- Underground Pumped Hydro.

Cost estimates presented in this section were developed by Black & Veatch in January 1990 dollars unless otherwise designated. Capital costs are for overnight construction (not including interest during construction) unless noted.

Screening matrices for the five categories of technologies are presented in Tables 1A.5.1-1 through 1A.5.1-5. Three technologies were retained for screening: coal gasification combined cycle, solar thermal parabolic trough, and lead-acid battery storage.

1A.5.1.1 Coal Fueled Alternatives

The following coal fueled alternative technologies for power generation are discussed in this subsection.

- Gasification combined cycle.
- Pressurized fluidized bed combustion.
- Advanced pulverized coal.
- Gasification fuel cells.
- Gasification humid air turbine cycle.
- Coal liquefaction.
- Magnetohydrodynamics.

1A.5.1.1.1 Gasification Combined Cycle. A gasification combined cycle (GCC) system gasifies a solid fuel, producing a fuel gas for a combined cycle power generation system. Usable solid fuels include bituminous and subbituminous coals, or lignite. Fuel flexibility depends upon the gasifier used. Coal reacts in the gasifier with air or oxygen and water or steam to form raw syngas. A low Btu syngas (less than 200 Btu/scf, HHV basis) is produced by an air blown gasifier. A medium Btu gas (200 to 500 Btu/scf) is produced with an oxygen blown gasifier. The cleansed gas is used to fire a combined cycle power block. In an integrated gasification combined cycle (IGCC) system, steam generated in the heat recovery steam generator (HRSG) is augmented by steam produced in the gasification and gas cleanup system.

Large-scale demonstrations of IGCC plants include the 100 MW Cool Water plant in Daggett, California, which uses a Texaco gasifier, and the 160 MW GCC unit in Plaquemine, Louisiana, which uses a Dow gasifier. Shell is also developing a coal gasification process. Another large-scale gasifier application within the United States is the 13,000 ton per day Dakota Gasification Plant in North Dakota. The plant uses 16 Lurgi gasifiers with a lignite feed and produces synthetic natural gas. Lurgi gasifiers are also used in the SASOL I, II, and III plants in South Africa, where a total of 80,000 tons per day of coal are gasified to produce a number of products, including transportation fuels.

First commercial operation for GCC plants is projected to be in the mid-1990s. Commercial plants will range in capacity from 250 to 450 MW. Modular building blocks based on the General Electric Frame 7F combustion turbine are about 220 MW. Capital cost estimates for mature commercial scale (400 MW or larger) plants range from about \$1,120/kW to \$1,620/kW. Operation and maintenance (O&M) cost estimates are about \$40/kW (fixed) and \$0.80/MWh to \$4.7/MWh (variable). Heat rates depend upon the degree of integration and the economics driving equipment selection. Heat rates for planned commercial units range from about 8,240 to 9,500 Btu/kWh.

Emission levels of sulfur dioxide (SO_2) and nitrous oxide (NO_x) for a GCC have been reported to be only 10 to 15 percent of the New Source Performance Standards. Sulfur removal efficiency has exceeded 96 percent at prototype plants. These low levels of emission enhance the attractiveness of GCC systems in view of more stringent emission standards accompanying the 1990 Clean Air Act Amendment.

1A.5.1.1.2 Pressurized Fluidized Bed Combustion. Pressurized fluidized bed combustion (PFBC) is a variation of fluid bed technology in which combustion occurs in a pressure vessel at 10 to 15 atmospheres. The PFBC process involves burning crushed coal under high pressure in a sorbent bed, usually of limestone or dolomite. A compressor provides high-pressure combustion air at the bottom of the combustor to maintain the coal and sorbent in a highly turbulent suspended state. The turbulence promotes good particle mixing and the depth of bed allows long gas residence time, which leads to high combustion efficiency and SO_2 absorption.

Two basic cycle configurations are being considered: a turbocharged cycle and a combined cycle. In the turbocharged cycle, heat is extracted from the combustion gas into steam-cooled convective surfaces before entering the gas turbine, leaving sufficient energy to pressurize the system by driving a compressor, but not enough energy to generate electricity. Electric power is produced only in the steam cycle. In the combined cycle concept, the high temperature combustion gases are cleaned at the elevated temperature and then expanded to drive both the compressor and a turbine generator. The gas turbine exhaust gas is used to generate steam in a HRSG. The steam drives a steam turbine generator, which also generates electricity. PFBC combined cycle designs have more stringent gas cleanup and material requirements as a result of the elevated temperatures. The hot gas cleanup technology currently being used consists of cyclone separators; however, development of ceramic tubular and candle filters, hot electrostatic precipitators, and other forms of hot gas cleanup is progressing.

The pressurized fluidized bed technology is in the early demonstration stage. The American Electric Power Company recently began startup activities for a 70 MW PFBC combined cycle demonstration plant which repowers its Tidd Station in Ohio. Additional PFBC systems similar in size to the Tidd facility have recently been constructed in Sweden and in Spain. Additional systems to be built in the US are in the planning stage. It is projected that PFBC technology will be available commercially for repowering applications in the mid-to-late 1990s, and for larger-scale new construction applications in the late 1990s to early 2000s.

A capital cost estimate for a mature PFBC system is about $1,400/kW. O&M costs are about $30/kW-yr (fixed) and $6.40/MWh (variable). An estimated heat rate is about 8,700 Btu/kWh.

SO_2 emissions are controlled in the pressurized fluidized bed in a manner similar to that for atmospheric fluidized bed technology. Particulate emission is controlled by using hot gas cleanup consisting of cyclones and ceramic filters.

1A.5.1.1.3 Advanced Pulverized Coal. Advanced pulverized coal power plants make use of recent advances in major power plant components, including the steam generator, turbine generator, flue gas desulfurization (FGD) system, and flue gas reheat system. EPRI is developing a State-of-the-Art Power Plant (SOAPP) incorporating these advancements. SOAPP uses a spiral-wound supercritical steam generator which generates steam at 4,500 psig/1,100 F, with double reheat at 1,100 F. The steam turbine is a three-casing, tandem-compound unit which is designed for hybrid constant/sliding pressure operating with partial arc admission. There are two trains of 50 percent capacity feedwater heaters, with nine heaters in each train, including the deaerator. Turbine bypass facilitates unit startup. The FGD system design depends on the sulfur content of the design coal.

The high-sulfur coal FGD system uses a single 100 percent flow FGD absorber module based on commercial advanced limestone-gypsum FGD technology. A fabric filter is used for particulate removal. The low-sulfur FGD system uses two 50 percent flow lime spray dryer modules and a fabric filter.

Capital cost estimates for a mature technology SOAPP are about $1,370/kW for a 300 MW facility. The projected heat rate is about 8,800 Btu/kWh.

Various systems within the SOAPP are in different stages of development. A fully integrated SOAPP has not been fully designed or constructed at this time. Should development continue at a steady rate, commercialization might occur in the late 1990s.

1A.5.1.1.4 Gasification Fuel Cells. A gasification fuel cell plant combines the gasifier technologies discussed in Subsection 1A.5.1.1.1 with molten carbonate fuel cell technologies discussed in Subsection 1A.5.1.2.3. A clean medium Btu syngas fuel is supplied to a fuel cell by the gasifier and gas cleanup systems. The molten carbonate fuel cell generates electricity directly from the syngas (Subsection 1A.5.1.2.3). Heat recovered from the fuel cell is used to generate steam, which drives a steam turbine generator.

The gasification fuel cell technology is in the concept stage, with several EPRI studies having been conducted to estimate plant costs and performance. Molten carbonate fuel cell development is in the bench scale stage. Plans are underway for a 2 MW demonstration unit fired by natural gas. Gasification fuel cell systems are not likely to be commercially available until the late 1990s to early 2000s.

A cost estimate for a mature technology gasification fuel cell plant is $1,800/kW. The estimated heat rate for that plant is 7,400 Btu/kWh.

Emissions from a gasification fuel cell plant will be very low, similar to a GCC plant. NO_x emissions should be lower than for the GCC plant.

1A.5.1.1.5 Integrated Gasification Humid Air Turbine Cycle. An integrated gasification humid air turbine (IGHAT) cycle gasifies coal, producing a fuel gas for a humid air turbine (HAT) cycle. The HAT cycle concept is similar to that described for the gas fueled HAT in Subsection 1A.5.1.2.2. HAT cycles are not commercially available.

The coal gasification plant used in the IGHAT concept is similar to the ones used in IGCC plants described in Subsection 1A.5.1.1.1. In an IGHAT plant, the syngas is water quenched after it leaves the gasifier, and the resulting low level heat is recovered and used in the HAT cycle. This eliminates the need for expensive syngas coolers.

The net plant heat rate for an IGHAT plant is projected to be about 8,870 Btu/kWh for a 475 MW plant. A preliminary capital cost estimate for a 475 MW IGHAT facility is $1,160/kW.

Emission controls for sulfur removal described in the IGCC section of this report are applicable to IGHAT cycles. The HAT turbine used in the IGHAT cycle is expected to have NO_x emissions as low as 7 ppmvd (15 percent oxygen) or less without the use of selective catalytic reduction (SCR).

1A.5.1.1.6 Coal Liquefaction. Coal liquefaction processes have been under development for many years. The key objective of a coal liquefaction process is to upgrade the hydrogen content of the feed coal and to remove components of environmental concern. Early liquefaction processes were used in Germany during World War II to provide liquid fuels for military applications. Currently, liquefaction processes are being used in South Africa's SASOL facilities to provide transportation fuels for South Africa as a result of an embargo on oil shipments.

Developmental efforts within the US are aimed at reducing costs for coal derived liquid fuels. In addition to processes which produce synthetic crude oil, efforts are also being made to develop cost-effective coal derived methanol plants. The primary interest for utilities would be to supply fuels to oil fired plants should the price of oil escalate dramatically. In general, use of coal derived liquid fuels would be better suited to peaking applications, where a base load fuel production plant could provide storable fuels. Base load power generation plants would more likely use coal gasification processes rather than coal liquefaction processes to provide fuel.

It is projected that using mature liquefaction processes based on current pilot-scale technology, synthetic oil could be produced at a cost of about $40/barrel.

1A.5.1.1.7 Magnetohydrodynamics. Magnetohydrodynamics (MHD) is a fossil fuel fired technology in which the electrical generator is static (nonrotating) equipment. The MHD technology is in the early development stage. Coal or another fossil fuel is burned to produce hot gases at temperatures exceeding 4,500 F. Potassium carbonate is injected during the combustion process to increase the electrical conductivity of the combustion products. These hot gases are accelerated in a nozzle and pass through a rectangular channel, which is surrounded by a superconducting magnet. The electrically conductive gases moving in the magnetic field produce a direct current in the channel's electrodes. This direct current is subsequently converted to alternating current with a solid-state inverter. An MHD device would likely be used for the topping cycle of an MHD combined cycle system.

A heat rate projection for a mature MHD system is about 8,620 Btu/kWh. Capital and O&M costs for MHD systems are highly speculative because of the relatively early stage of development of the technology.

The MHD technology appears to be capable of adequately controlling NO_x and SO_2 emissions.

1A.5.1.2 Oil or Gas Fired Alternatives

Oil or gas fired alternatives considered in this subsection include the steam injected combustion turbine, humid air turbine (HAT) cycle, and fuel cells.

1A.5.1.2.1 Steam Injected Combustion Turbine (Power Augmentation). The steam injection combustion turbine uses steam injected into the combustor and turbine sections of a conventional combustion turbine to increase power output. This is called power augmentation and is also known as the Cheng cycle.

The injection steam has a cooling effect on the combustion turbine internals, allowing greater fuel burn rates. There is no steam turbine or associated equipment as in the combined cycle arrangement. The basic components for this cycle are the combustion turbine and the HRSG. Aero-derivative combustion turbines are typically used in power augmentation applications because of the higher compression ratio associated with the aero-derivative. The aero-derivative also has the ability to pass greater flows. Emissions control is similar to that for oil or gas fired simple cycle and combined cycle systems.

Several power augmentation units began operation in the early to mid-1980s. The cycle is considered a mature technology. However, the long-term impact of massive steam injection on combustion turbine service life is still being evaluated. Furthermore, large combustion turbines designed specifically for steam injection are not commercially available today.

Capital costs for the power augmentation cycle vary with unit size, amount of steam injected, and amount of steam to be produced for cogeneration. Capital cost estimates range from $800/kW for a 150 MW unit to $1,050/kW for a nominal 50 MW installation.

Steam injection can increase the power output of a combustion turbine by as much as 50 percent, while reducing the heat rate by as much as 17 percent.

1A.5.1.2.2 Natural Gas Fired Humid Air Turbine Cycle. The humid air turbine (HAT) cycle uses a combustion turbine that employs air intercooling and the substitution of water vapor for a portion of the excess air to the combustor to reduce the power requirement by the compressor. This reduction in air compression load significantly increases the net power output by the turbine. Intercooling and aftercooling provides some of the heat required for generating water vapor. Additional low temperature heat required for generating water vapor is also obtained from the turbine exhaust. A saturator generates water vapor in a multi-stage counter-current operation by directly contacting cooled compressed air with hot water. The humidified air is then further heated by the turbine exhaust prior to entering the combustor.

Natural gas fired HAT cycles are not commercially available. The HAT cycle is still in the early developmental stages, and a commercial prototype combustion turbine designed for the HAT cycle has not yet been built.

NO_x emissions are expected to be very low over the full range of capacity because of the high moisture levels in the air going to the combustor.

A preliminary capital cost estimate for a nominal 200 MW plant size is $630/kW. The net heat rate is projected to be 7,080 Btu/kWh.

1A.5.1.2.3 Fuel Cells. Fuel cells convert hydrogen rich fuels directly to electricity through electrochemical reactions. The reactants, fuel and oxidant (air or oxygen), are fed to separate anode and cathode electrodes. The electrodes are separated by an electrolyte that transports the ions generated by the anode reaction. This transport of ions creates an electric potential difference that promotes electron flow through an electric load. An inverter converts the direct current into alternating current.

The most common fuel cell concepts for power generation are the phosphoric acid fuel cell and the molten carbonate fuel cell. The solid oxide fuel cell is another advanced option, but it is not as well developed as the other two. Phosphoric acid and molten carbonate fuel cells are fueled with hydrogen, requiring a reforming step for the use of natural gas or methanol as the fuel. Some molten carbonate fuel cells have internal reforming capability, allowing the direct use of natural gas. Molten carbonate fuel cells are also able to operate with syngas from a coal gasification unit, allowing coal to be used as the fuel.

Phosphoric acid fuel cell technology is the most mature of the fuel cell technologies for power plant applications. However, operational problems such as long-term reliability remain to be solved. Phosphoric acid fuel cell power plants are expected to be commercially available and reliable in the 2000 to 2010 time period. Molten carbonate systems offer higher efficiencies than can be achieved with phosphoric acid systems, although the commercial development of these systems lags perhaps 5 to 10 years behind that of phosphoric acid systems. Small area molten carbonate cell stacks have been successfully tested, and scaleup to full area stacks is underway.

A capital cost projection for a mature, central station phosphoric acid fuel cell plant is about $1,000/kW. The net plant heat rate is projected to be 8,300 Btu/kWh. Combined cycle and molten carbonate fuel cell plants are projected to have significantly lower heat rates.

Because the fuel gas is necessarily very clean and combustion does not occur in a fuel cell, emissions of air pollutants, solid wastes, and contaminated wastewater are extremely low during operation.

1A.5.1.3 Nuclear Alternatives

This assessment consists of five advanced technologies that offer design modularity, reduced construction schedules, and potentially increased safety. The advanced technologies include the advanced passive light water reactor, modular pressurized heavy water reactor, the modular high-temperature gas-cooled reactor, the liquid metal reactor, and fusion.

Licensing procedures for new conventional and advanced generating plants are being revised. The Nuclear Regulatory Commission (NRC) is proposing a one-step licensing process for new nuclear power generation. With the one-step process, electric utilities would obtain a single permit approval for the nuclear power plant from the NRC before construction of the project. The approval would also include a license to operate, assuming the plant is built in accordance with the previously approved construction plans; however, this approval has been overturned in court and is subject to ongoing appeals. Some electric utilities prefer that a law be passed through Congress with similar one-step licensing provisions. The law would minimize potential challenges to the NRC's one-step procedure.

Table 1A.5.1-1
Coal Burning Technology Alternative Screening Matrix

Technology	Technical Maturity	Unit Resource	Unit Performance: Heat Rate (Btu/kWh)	Unit Performance: Capacity Factor (percent)	Environmental Considerations	Project Cost*: Capital ($/kW)	Project Cost*: O&M**	Time Frame of OUC Interest	Screening Comments
Coal Gasification Combined Cycle	Demonstration Scale with Nominal 100 MW and 160 MW in Operation	Abundant Resources of Coal Relatively Certain	8,240 to 9,500	Baseload	Air Emissions Much Lower Than Existing Requirements	1,120 to 1,620	$37/kW-y; $0.8/MWh to $4.70/MWh	Near-Term	Retain for Screening
Pressurized Fluidized Bed	Demonstration Plants at 70 MW, 79 MW, and 135 MW Under Construction	Abundant Resources of Coal and Limestone Relatively Certain	8,700	Baseload	Projected Compliance with Existing Requirements. Requires Hot Gas Cleanup of Particulates, a Developmental Issue.	1,300 to 1,400	$30/kW-y; $6.4/MWh	Near to Mid-Term	Drop from Screening; Inadequate Demonstration
Advanced Pulverized Coal	Not All Components Developed for SOAPP Conditions	Abundant Resources of Coal and Limestone Relatively Certain	8,800	Baseload	Projected Compliance with Existing Requirements	1,370	$30/kW-y; $3.2/MWh	Mid-Term	Drop from Screening; Inadequate Demonstration
Fuel Cells (Advanced with Coal Gasification)	Bench Scale	Abundant Resources of Coal Relatively Certain	7,500; Lower with Bottom Cycle	Intermediate or Baseload	Projected Compliance. Will Depend On Proven Gas Cleanup Similar to GCC or Hot Gas Cleanup Under Development	Not Available		Mid to Far-Term	Drop from Screening; Inadequate Demonstration
Humid Air Turbine Cycle (With Coal Gasification)	Not Developed	Abundant Resources of Coal Relatively Certain	8,800	Intermediate or Baseload	Projected Compliance. Will Depend On Proven Gas Cleanup Similar to GCC or Hot Gas Cleanup Under Development	1,160	Not Available	Mid-Term to Far-Term	Drop from Screening; Inadequate Demonstration
Coal Liquefaction	Demonstrated in South Africa	Abundant Resources of Coal Relatively Certain	Not Applicable	Not Applicable	Similar to GCC for Fuel Production Plant	Not Applicable		Mid-Term	Drop from Screening; High Cost Liquid Fuel Source Compared with Oil or Gas
Magnetohydrodynamics	Bench Scale	Abundant Resources of Coal Relatively Certain	8,620	Baseload	Projected for Compliance	1,810		Far-Term	Drop from Screening; Not Demonstrated to Date

*Costs in January 1990 dollars. Emerging technology costs typically are mature technology costs rather than present-day costs.

**O&M costs do not include fuel.

Table 1A.5.1-2
Oil or Gas Burning Technology Alternative Screening Matrix

Technology	Technical Maturity	Unit Resource	Unit Performance: Heat Rate (Btu/kWh)	Unit Performance: Capacity Factor (percent)	Environmental Considerations	Project Cost*: Capital ($/kW)	Project Cost*: O&M**	Time Frame of OUC Interest	Screening Comments
Steam Injected Gas Turbine	Commercial for Smaller Sizes Using Aero-Derivative Turbines	Future of Economic Supply of Oil and/or Gas Not Certain. Also Requires Significant Water.	8,080 at 50 MW	Base or Intermediate Load	Same as Simple Cycle.	800 to 1,050	$5.4/kW-y; $0.5/MWh	Near-Term	Drop From Screening; Typically Used in Cogeneration Application Where Process Steam is Generated
Humid Air Cycle	Turbine Not Developed for HAT Cycle	Future of Economic Supply of Oil and/or Gas Not Certain. Also Requires Significant Water.	7,080	Base or Intermediate Load	Same as Simple Cycle.	630	Not Available	Mid-Term	Drop from Screening; Inadequate Demonstration
Fuel Cells	Demonstration	Future of Economic Supply of Natural Gas Not Certain	8,300	Multi Use	Very Low Emissions Because of Clean Fuel (Natural Gas) and Low Temperatures	Current 2,500. Projected 1,000	$8.4/kW-y; $5.2/MWh	Mid-Term	Drop from Screening; Inadequate Demonstration

*Costs in January 1990 dollars. Emerging technology costs typically are mature technology costs rather than present-day costs.

**O&M costs do not include fuel.

1A.5.2 Cost and Operating Characteristics for Conventional Alternatives

Cost and performance data for pulverized coal, atmospheric circulating fluidized bed, combustion turbine, and combined cycle units are provided in this section. Data were primarily obtained from Black & Veatch Generating Unit Characteristics--July 5, 1990.

1A.5.2.1 Conventional Pulverized Coal

Pulverized coal technology has been the standard of the coal fueled steam electric generating industry for several decades. In a conventional pulverized coal steam generating unit, the dry pulverized coal is entrained in a hot airstream and carried from the pulverizer through coal piping to the furnace, where it is ignited and burned in suspension. Radiant energy from the combustion process is absorbed by the furnace waterwalls. Downstream of the furnace, the flue gas flows through steam-and water-cooled convective heat transfer surfaces and then through a regenerative air heater. From the air heater, the flue gas flows through particulate removal and desulfurization equipment before entering the chimney and being exhausted to the atmosphere. Superheated steam that is generated in the steam-cooled convective heat transfer sections of the boiler is delivered to the steam turbine generator. Steam from the turbine exhaust is condensed, pumped to the steam generator operating pressure, heated by steam from turbine extractions, and fed back to the steam generating unit.

Sulfer dioxide (SO_2) emissions from pulverized coal units can be controlled through any number of flue gas desulfurization (FGD) technologies. The application of an FGD technology is dependent on the type of coal and its sulfur content, site-specific constraints, and other site environmental factors. High-sulfur coal fired units have been typically designed to achieve 90 percent SO_2 removal.

Control of nitrogen oxide (NO_x) emissions from pulverized coal fueled power plants in the US has been traditionally achieved through the use of combustion alternatives rather than installation of a specific nitrogen oxide removal process. However, installation of Thermal DeNO_x and Selective Catalytic Reduction (SCR) processes as well as use of low NO_x burners may become more commonplace in the future as proven technology is developed and reliability is demonstrated. Typical NO_x permit values for pulverized coal units have been

0.60 lb/MBtu. The latest low NO_x pulverized coal units have been permitted as low as 0.32 lb/MBtu.

Estimated performance and cost data for 300 MW, 425 MW, and 675 MW pulverized coal units are provided in Table 1A.5.2-1.

1A.5.2.2 Atmospheric Fluidized Bed

Atmospheric fluidized bed combustion (AFBC) systems are in an advanced state of development and are currently being applied in electric utility systems. The largest circulating AFBC in operation is a 96 MW unit in Trona, California. The largest unit scheduled for construction is a 165 MW unit at Point Aconi for Nova Scotia Power Corp.

In a circulating fluidized bed steam generating unit, fuel and sorbent are fed by gravity into a highly turbulent region of the combustion chamber where combustion air, feed fuel, sorbent, and recirculating solids are mixed. The reacting gas and solids flow upward and then enter a particulate separator, usually a cyclone, where the solids are separated and returned to the bottom of the combustion chamber. Flue gas leaves the cyclone and is cooled by a conventional convection bank and an air heater. Additional particulate removal is required before the gas is discharged to the stack.

Flue gas desulfurization is accomplished in fluidized bed combustion units by adding limestone to the bed and promoting high temperature reaction of calcium oxide (CaO) with SO_2. The level of SO_2 removal is increased by increasing the feed rate of limestone to the bed. On the basis of current operating experience with AFBC units, relatively high levels of excess limestone addition (Ca/S ratio of 2.0 to 3.0) are required to achieve 90 percent SO_2 removal. Higher SO_2 removal rates (95 percent) have not been demonstrated on a continuous basis, and are expected to require much higher limestone consumption rates.

Generation of NO_x in AFBC units is lower than that of pulverized coal units because of the lower combustion temperatures inherent to the process. With staged combustion, NO_x emissions in AFBCs can be controlled to less than 0.30 lb/MBtu.

A current drawback to the implementation of AFBC units over conventional pulverized coal units is the relatively higher limestone and waste disposal costs associated with AFBCs. Although each of the units generates about the same amount of fuel ash, the unreacted limestone in the combustion wastes is significantly higher for the AFBC. A conventional 150 MW pulverized coal plant will produce about 42,000 tons per year of desulfurization byproducts compared to 62,000 tons per year by a 150 MW AFBC unit. Since fly ash is not separately collected in the APBC, it cannot be marketed as is sometimes done at a pulverized coal unit.

As with most new or rapidly expanding technologies, when new equipment is constructed on the basis of successful design criteria used by significantly smaller equipment installations, operation of the larger equipment normally is accompanied by operating problems due to design shortcomings. The operation of fluidized bed boilers in the US has not been an exception to this rule. It is expected that new fluidized bed installations with capacities in excess of those in operation for several years will be the source of more forced outages than a steam generator of conventional design. After the first few years of operation, it is expected that maintenance requirements will be well defined, and preventive maintenance programs will enable the fluidized bed boilers to operate with availability comparable to that of conventional pulverized coal boilers. Fluidized bed systems lack the high maintenance pulverizers and scrubber equipment of conventional pulverized coal plants, but may experience higher boiler maintenance due to potentially greater heat transfer surface erosion.

Estimated performance and cost data for 300 MW and 425 MW circulating AFBC coal units are provided in Table 1A.5.2-2. Both units are based on two steam generators and one steam turbine.

1A.5.2.3 Combustion Turbines

Combustion turbines burning either natural gas and/or No. 2 oil are available in a wide variety of sizes and configurations. Combustion turbines are generally used for peaking and reserve purposes because of their relatively low capital cost, poorer full load heat rate, and the higher cost of fuel when compared to conventional baseload capacity. Combustion turbines have the added benefit of providing quick-start capability in certain configurations.

Combustion turbine design continues to be improved. Advanced designs such as the General Electric Frame 7FA (MS7001F) offer a lower heat rate than can be achieved with standard machines, mainly because of higher hot gas temperatures. The GE Frame 7FA compared to the Frame 7EA conventional design offers an increased firing temperature of 2,300 F, an increase in the number of combustion chambers, and a multiple combustor nozzle arrangement instead of a single nozzle, reducing combustion system wear. Estimated performance and cost data for the Frame 7EA and the Frame 7FA are provided in Table 1A.5.2-3.

1A.5.2.4 Combined Cycle

In a combustion turbine combined cycle facility, the hot exhaust gases from the combustion turbine pass through a heat recovery steam generator (HRSG). The steam generated by the HRSG is expanded through a steam turbine which, in turn, drives an additional generator. With a typical gas inlet temperature of 1,000 F, steam conditions of about 1,000 psi and 900 F can be produced by the HRSG. Typically, combined cycle installations can have more than one set of combustion turbines and HRSGs, with steam piped to a common steam turbine. Usually about two-thirds of the power is produced by the combustion turbine generator and about one-third is produced by the steam turbine generator.

Estimated performance and cost data for two Frame 7EA or 7FA combustion turbines, two unfired HRSGs, and one steam turbine are provided in Table 1A.5.2-4.

Table 1A.5.2-1
Pulverized Coal Technology Unit Cost and Performance Data

	Net Plant Output, MW*		
	300	425	675
Direct Capital Cost, 1990 $/kW**	1,210	1,070	895
Indirect Cost, percent	19.4	18.8	17.7
Fixed O&M Cost, 1990 $/kW-yr	19	17	15
Variable O&M, 1990 $/MWh	1.60	1.55	1.55
Net Plant Heat Rate, Btu/kWh			
Full load	9,870	9,820	9,730
75 percent load	10,080	10,010	9,940
50 percent load	10,760	10,690	10,560
25 percent load	12,800	12,730	12,630
Equivalent Availability, percent	83	83	83
Forced Outage Rate, percent	7	7	8
Planned Maintenance Outage, weeks/year	4	4	5
Construction Period, months	41	46	48

*Rated conditions are at valves wide open 5 percent over pressure (VWO-OP).

**Does not include escalation, indirect cost, or interest during construction. Based on first unit at a new site.

Table 1A.5.2-2
Circulating Atmospheric Fluidized Bed Technology Unit Cost and Performance Data

	Net Plant Output, MW*	
	300	425
Direct Capital Cost, 1990 $/kW**	1,140	1,000
Indirect Cost, percent	19.4	18.8
Fixed O&M Cost, 1990 $/kW-yr	17	15
Variable O&M, 1990 $/MWh	2.15	2.15
Net Plant Heat Rate, Btu/kWh		
Full load	10,260	10,210
75 percent load	10,480	10,410
50 percent load	11,190	11,120
25 percent load	13,310	13,240
Equivalent Availability, percent	83	83
Forced Outage Rate, percent	7	7
Planned Maintenance Outage, weeks/year	4	4
Construction Period, months	38	46

*Based on two boilers serving one steam turbine generator, (VWO-OP).

**Does not include escalation, indirect cost, or interest during construction. Based on first unit at a new site.

Table 1A.5.2-3
Combustion Turbine Technology Unit and Performance Data

	GE Frame 7EA	GE Frame 7FA
Net Plant Output, MW*	87.5	172.3
Direct Capital Cost, 1990 $/kW**	319	268
Indirect Cost, percent	13.0	12.0
Fixed O&M Cost, 1990 $/kW-yr	8.50	7.00
Variable O&M, 1990 $/MWh		
Peaking load, 5 percent capacity factor	3.7	3.7
Intermediate load, 25 percent capacity factor	1.3	1.3
Net Plant Heat Rate, Btu/kWh*		
Full load	12,271	11,221
75 percent load	12,614	11,590
50 percent load	13,905	12,887
25 percent load	18,991	18,008
Equivalent Availability, percent	88	88
Forced Outage Rate, percent	5	5
Planned Maintenance Outage, weeks/year	2.5	2.5
Construction Period, months	12	12

*59 F ambient temperature and 14.9 psi.

**Does not include escalation, indirect cost, or interest during construction.

Table 1A.5.2-4
Combined Cycle Technology Unit and Performance Data

	Combined Cycle with GE 7EA	Combined Cycle with GE 7FA
Net Plant Output, MW*	255	489
Direct Capital Cost, 1990 $/kW**	505	417
Indirect Cost, percent	13.1	13.1
Fixed O&M Cost, 1990 $/kW-yr	13.1	10.8
Variable O&M, 1990 $/MWh	1.80	1.80
Net Plant Heat Rate, Btu/kWh*		
Full load	8,409	7,912
75 percent load	8,514	8,097
50 percent load	9,336	8,896
25 percent load	12,959	12,097
Equivalent Availability, percent	86	86
Forced Outage Rate, percent	6	6
Planned Maintenance Outage, weeks/year	4	4
Construction Period, months	22	27

*Based on two combustion turbines, two unfired HRSGs, and one steam turbine at 59 F and 14.9 psi.

**Does not include escalation, indirect costs, or interest during construction.

Appendix 1A.A.0
Advanced Technology Descriptions

This appendix presents detailed infomation about the alternative generation technologies summarized in Subsection 1A.5.1. The assessment includes the following technologies.

- Coal Fired Alternatives.
  - Gasification Combined Cycle.
  - Pressurized Fluidized Bed Combustion.
  - Advanced Pulverized Coal.
  - Gasification Fuel Cells.
  - Gasification Humid Air Turbine Cycle.
  - Coal Liquefaction.
  - Magnetohydrodynamics.
- Oil or Gas Fired Alternatives.
  - Steam Injected Combustion Turbine.
  - Humid Air Turbine.
  - Fuel Cells.
- Nuclear Alternatives.
  - Advanced Passive Light Water Reactors.
  - Modular Pressurized Heavy Water Reactor.
  - Module High-Temperature Gas-Cooled Reactor.
  - Liquid Metal Reactor.
  - Fusion.
- Renewable Energy Alternatives.
  - Wind Energy.
  - Solar Photovoltaic.
  - Solar Thermal.
  - Ocean Thermal.
  - Ocean Wave.
  - Ocean Tidal.
  - Geothermal.

- Energy Storage
  - Battery
  - Compressed Air.
  - Underground Pumped Hydro.

The assessment of each technology includes a general description of the technology, its technological maturity, typical cost and performance characteristics, required resources, environmental impacts, and references used in producing the assessment. Capital costs are given in January 1990 dollars, and are for overnight construction, unless otherwise noted.

1A.A.1 Coal Fired Alternatives

The following coal fired alternative technologies for power generation are discussed in this report.

- Gasification combined cycle.
- Pressurized fluidized bed combustion.
- Advanced pulverized coal.
- Gasification fuel cells.
- Gasificaton humid air turbine cycle.
- Coal liquefaction.
- Magnetohydrodynamics.

1A.A.1.1 Gasification Combined Cycle

A gasification combined cycle (GCC) system gasifies a solid fuel, producing a fuel gas for a combined cycle power generation system. Usable solid fuels include bituminous, lignite, or subbituminous coals. Figure 1A.A.1-1 illustrates a GCC plant flow diagram. Coal may be delivered to a pressurized gasifier in a coal/water slurry or as a dry feed depending on the gasifier concept. The coal reacts with air or oxygen and water or steam to form raw syngas. A low Btu syngas (less than 200 Btu per scf, HHV basis) is produced by an air blown gasifier. A medium Btu gas (200 to 500 Btu per scf) is produced with an oxygen blown gasifier. The raw syngas is cooled to about 400 F to allow for particulate removal, and further cooled to 100 F to allow for acid gas removal. The cleansed gas is used in a combustion turbine. The turbine combustion gases are exhausted to a heat recovery steam generator (HRSG), where high-pressure steam is produced; the cooled combustion gases are then exhausted to the atmosphere. The high-pressure steam formed in the HRSG and in the syngas cooler is used to generate power in a bottoming steam cycle. In an integrated gasification combined cycle (IGCC) plant, steam produced in the gasifier portion of the plant is also transported to the HRSG.

1A.A.1.1.1 Technological Maturity. The first large-scale demonstration of an IGCC plant was the 100 MW Cool Water plant in Daggett, California.[1] That unit, which used a Texaco gasifier, operated from 1984 through 1989. Subsequently, Texaco obtained rights to purchase the plant from Southern California Edison. Recommissioning is planned for 1992, with plans for the unit to gasify a combined coal/municipal sewage sludge feed. It will produce electricity, CO_2, alcohol, and ammonia.[2] Dow currently operates a 160 MW GCC unit in Plaquemine, Louisiana.[3] Shell is also developing a coal gasification process.

The Cool Water unit began operation in mid-1984. In 1988, the maximum permitted feed rate was 1,200 tons per day at a rating of 122 MW (gross).[1] The design coal was Illinois No. 6, although a variety of coals were tested at Cool Water. Carbon conversions exceeded 97 percent for four different coals. The demonstrated heat rate was 10,950 Btu/kWh. The capacity factor for the first eight months of 1988 was 71 percent. During July and August 1988, the capacity factors were 98.0 and 90.6 percent, respectively. The availability of the gasifier was 79.3 percent in 1987.

The Dow Syngas Project uses a coal/water slurry with 52 to 54 percent concentration. The plant uses a gasifier with two stages to improve carbon use and cold gas efficiency. Plant availability for its first full year of operation was 42 percent. Plant availability from May to September 1988 was 52 percent. Production records through September 1988 showed a daily production of 92 percent capacity and a 30-consecutive-day capacity of 65 percent.[3,4]

The Shell Development Company is developing the Shell Coal Gasification Process (SCGP). Shell's current demonstration unit is SCGP-1, located in Deer Park, Texas. SCGP-1 can gasify 250 tons per day of bituminous coals and 400 tons per day of lignites. Commercial Shell gasifiers are expected to have capacities ranging from 1,000 to 3,000 tons of coal per day.[5]

Another large-scale gasifier application within the United States is the 13,000-ton per day Dakota Gasification Plant in North Dakota. The plant uses 14 Lurgi gasifiers with a lignite feed, and produces synthetic natural gas. Lurgi gasifiers are also used in the SASOL I, II, and III plants in South Africa, where a total of 80,000 tons per day of coal are gasified to produce a number of products, including transportation fuels.

First commercial operation for GCC plants is projected to be in about 1994.[6]

1A.A.1.1.2 Cost and Performance Characteristics. Capital cost estimates for GCC systems vary with system size, design coal, gasifier manufacturer, and degree of integration, as well as plant location. Lowest capital cost estimates are \$1,210/kW for Dow gasifier systems and \$1,120/kW for lignite and subbituminous burning plants with a Texas location.[7] No data are available in the public domain for a bituminous coal burning Dow gasifier facility in Florida. A capital cost estimate for a Shell gasifier facility burning bituminous coal at a Florida site is \$1,620/kW.[7] It is unknown to what extent these estimates reflect real cost differences.

O&M cost estimates vary in a manner similar to capital cost estimates. The low O&M cost estimate (for Dow) is \$37/kW-year (fixed) and \$0.8/MWh.[8]

Estimates of net plant heat rate vary with plant design (e.g., degree of integration and use of quench system versus heat recovery steam generation syngas cooling) and with the design coal. Projected heat rates range from around 8,200 to 9,630 Btu/kWh.

1A.A.1.1.3 Required Resources. The primary resources required for a GCC plant are coal and water. Coal requirements for the 400 GCC MW plant are about 3,800 tons per day at full load. The type of coal which can be burned depends on the gasifier. The Texaco gasifier currently operates on bituminous coal and has not been demonstrated to operate efficiently using subbituminous or lignite coals. The Dow gasifier was designed for, and has been operated with, subbituminous and lignite coals. Small-scale test burns have been made with a high-sulfur bituminous coal, with no operational or equipment problems encountered.[9] Test burns of several coals at SCGP-1 appear to validate predictions that the Shell gasifier will be flexible with regard to use of bituminous, subbituminous, or lignite coals.[10]

1A.A.1.1.4 Environmental Impacts. Emission levels of sulfur dioxide and nitrous oxide have been reported to be only 10 to 15 percent of the New Source Performance Standards.[3] Sulfur removal efficiency has exceeded 96 percent at prototype plants. These low levels of emission make GCC systems attractive in view of more stringent emission standards accompanying the 1990 Clean Air Act Amendments.

1A.A.1.1.5 References.

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2. W. H. Cummins, G. N. Richter, and J. S. Stevenson, "The Texaco Gasification Process: Status, New Applications and Proposed Developments," presented at the Ninth Annual EPRI Conference on Gasification Power Plants, Palo Alto, California, October 17-19, 1990.

3. R. D. Hudson and G. C. LeBlanc, "Environmental Monitoring at Dow's Coal Gasification Plant," presented at the Eighth EPRI Coal Gasification Contractors' Conference, Palo Alto, California, October 19-20, 1988.
4. R. M. Webb and K. W. Moser, "The Dow Syngas Project Recent Operating Experience," presented at the Eighth EPRI Coal Gasification Contractors' Conference, Palo Alto, California, October 19-20, 1988.
5. R. P. Jensen, U. Mahagaokar, and A. B. Krewinghaus, "SCGP--Progress in a Proven, Versatile, and Robust Technology," presented at the Ninth Annual EPRI Conference on Coal Gasification Power Plants, Palo Alto, California, October 17-19, 1990.
6. TAG: Technical Assessment Guide, Volume 1: Electricity Supply, EPRI P-587-L, November 1989.
7. Florida Power and Light Company's Study of Shell-Based GCC Power, Plants, EPRI GS-6176, January 1989.
8. Evaluation of a Dow-Based Gasification Combined Cycle Plant Using Low Rank Coals, EPRI GS-6318, April 1989.
9. M. W. Roll, R. J. Payonk, "Operation of the Dow Coal Gasification Process during 1990," presented at the Sixth Annual Pittsburgh Coal Conference, September 25-28, 1989.
10. A. B. Krewinghaus, P. C. Richards, "Coal Flexibility of the Shell Coal Gasification Process," presented at the Sixth Annual Pittsburgh Coal Conference, September 25-29, 1989.

1A.A.1.2 Pressurized Fluidized Bed Combustion

Pressurized fluidized bed combustion (PFBC) is a variation of fluid bed technology with combustion occurring in a pressure vessel at 10 to 15 atmospheres. The PFBC process shown on Figure 1A.A.1-2 involves burning crushed coal under high-pressure in a sorbent bed, usually of limestone or dolomite. A compressor provides high-pressure combustion air at the bottom of the combustor to maintain the coal and sorbent in a highly turbulent suspended state. The turbulence promotes good particle mixing, and the depth of bed allows long gas residence time, which leads to high combustion efficiency and sulfur dioxide absorption.[1]

Two basic cycle configurations are being considered: a turbocharged cycle and a combined cycle. In the turbocharged cycle, heat is extracted from the combustion gas into steam-cooled convective surfaces before entering the gas turbine, leaving sufficient energy to pressurize the system by driving a compressor, but not enough energy to generate electricity. Electric power is produced only in the steam cycle. In the combined cycle concept, the high temperature combustion gases are cleaned at the elevated temperature and then expanded to drive both the compressor and the turbine generator. The gas turbine exhaust gas is used to generate steam in a heat recovery steam generator (HRSG). PFBC combined cycle designs have more stringent gas cleanup and material requirements as a result of the elevated temperatures. The hot gas cleanup technology currently being used consists of cyclone separators; however, development of ceramic tubular and candle filters, hot electrostatic precipitators, and other forms of hot gas cleanup is progressing.

1A.A.1.2.1 Technological Maturity. The pressurized fluidized bed technology is entering the demonstration stage. Pilot facilities, which are presently not in operation, include the Grimethorpe facility in England and the 15 MWt ASEA Component Test Facility in Sweden. A 40 MWt turbocharged boiler was operated at the University of Auchen in West Germany. A 10 MWe facility is being operated by Ahlstrom (Pyropower) in Finland.

The American Electric Power Company recently completed a 70 MW PFBC combined cycle demonstration plant which repowers its Tidd Station in Ohio. That system began startup activity late in 1990. Additional PFBC systems similar in size to the Tidd facility have recently been constructed in Sweden and in Spain.[2] Additional systems to be built in the United States are in the planning stage.

A critical developmental issue for PFBC systems is the hot gas cleanup. The Tidd system will use cyclones, which are not adequate for small particle removal. Some gas turbine erosion is anticipated. Tidd will conduct a slip stream test of a ceramic filter. Various high temperature ceramic filters are under development in the United States and abroad.

It is projected that PFBC technology will be available commercially for repowering applications in the mid-to-late 1990s, and for larger-scale new construction applications in the late 1990s to early 2000s.

1A.A.1.2.2 Cost and Performance Characteristics. A cost estimate for a mature technology PFBC plant is about $1,400/kW.[7] An estimated heat rate is about 8,700 Btu/kWh. PFBC generating units are anticipated to be baseload units because their high efficiency (low heat rate) will cause them to be committed and dispatched with a high priority.

1A.A.1.2.3 Required Resources. Principal resource requirements for PFBC facilities are coal, water, and limestone or dolomite. Principal waste products are ash and desulfurization products. Land requirements for PFBC units are less than for conventional pulverized coal plants because the pressurized conditions provide combustion air mass flow requirements in smaller bed areas and volumes. Construction time and costs may be reduced by the modular nature of PFBC systems.[1]

1A.A.1.2.4 Environmental Impacts. SO_2 emissions are controlled in the pressurized fluidized bed in a manner similar to that for atmospheric fluidized bed technology. Particulate emission is controlled by using hot gas cleanup consisting of cyclones, ceramic filters, and conventional electrostatic precipitators.

1A.A.1.2.5 References.

1. J. D. McClung, ASEA Babcock PFBC, "Clean, Cost-Effective Kilowatts from Coal," presented to the 1988 Seminar on Fluidized Bed Combustion Technology for Utility Applications, Palo Alto, California, May 3-5, 1988.
2. "ASEA Babcock PFBC Update," A Quarterly Publication of ASEA Babcock, Vol. 1, No. 1, April 1988.
3. P. Almquist and B. Nordmark, "Update on Stockholm Energi's Vartan PFBC Combined Heat and Power Project," presented to the 1988 Seminar on Fluidized Bed Combustion Technology for Utility Applications, Palo Alto, California, May 3-5, 1988.
4. F. L. Kinsinger and D. K. McDonald of Babcock & Wilcox, "Combined Cycle using PFBC," presented to the American Power Conference, Chicago, Illinois, April 18-20, 1988.
5. D. K. McDonald and P. S. Weitzel of Babcock & Wilcox, "This PFBC Boiler for AEP's Tidd Plant Design for Performance and Operation," presented to the 1988 Seminar on Fluidized Bed Combustion Technology for Utility Applications, Palo Alto, California, May 3-5, 1988.
6. "ASEA Babcock PFBC Update," A Quarterly Publication of ASEA Babcock, Vol. 1, No. 2, Summer 1988.
7. TAG--Technical Assessment Guide, Volume 1: Electricity Supply-1986. EPRI, P-4463-SR, December 1986.

1A.A.1.3 Advanced Pulverized Coal

The advanced pulverized coal power plant concept makes use of recent advances in major power plant components, including the steam generator, turbine generator, flue gas desulfurization (FGD) system, and flue gas reheat system.[1,2] EPRI is developing a State-of-the-Art Power Plant, (SOAPP) incorporating these advancements. SOAPP uses a spiral-wound supercritical steam generator which generates steam at 4,500 psig/1,100 F, with double reheat at 1,100 F. The steam turbine is a three-casing, tandem-compound unit which is designed for hybrid constant/sliding pressure operating with partial arc admission. There are two trains of 50 percent capacity feedwater heaters, with nine heaters in each train, including the deaerator. Turbine bypass facilitates unit startup. The FGD system design depends on the sulfur content of the design coal. The high-sulfur coal FGD system uses a single 100 percent flow FGD absorber module based on commercial advanced limestone-gypsum FGD technology. A fabric filter is used for particulate removal. The low-sulfur FGD system uses two 50 percent flow lime spray dryer modules and a fabric filter.

1A.A.1.3.1 Technological Maturity.

Advanced pulverized coal plants at SOAPP steam conditions are in the conceptual stage. EPRI has undertaken a three-year developmental program which will survey and characterize SOAPP component technologies (1991), develop a plant conceptual design (1992), and perform sensitivity evaluations (1993). Detailed design and construction of an SOAPP facility could follow this program. Commercialization is not anticipated before 1997.[3]

1A.A.1.3.2 Cost and Performance Characteristics. An estimated cost for a 300 MW SOAPP facility is $1,370/kW.[1] O&M costs are estimated to be $30/kW-year (fixed) and $3.20/MWh (variable). The plant is estimated to have a full load net plant heat rate of 8,820 Btu/kWh.[1]

1A.A.1.3.3 Required Resources. The primary resources required for an advanced pulverized coal plant are similar to those for a conventional pulverized coal plant, except that the lower heat rate associated with the advanced design somewhat reduces requirements.

1A.A.1.3.4 Environmental Impacts. It is anticipated that emissions and wastes from SOAPP facilities will be similar to new pulverized coal plants using state-of-the-art emission control and waste management equipment.

1A.A.1.3.5 References.

1. TAG Technical Assessment Guide, Volume 1: Electric Supply--1989 (Revision 6), EPRI P-6587-L, November 1989.
2. G. L. Touchton, J. E. Cichanowicz, A. F. Armor, I. M. Torrens, M. W. McElroy, "State of the Art Pulverized Coal-Fired Power Plant," presented at the EPRI 2nd International Conference on Improved Coal Fired Power Plants, Palo Alto, California, November 2-4, 1988.
3. Charles McGowan, EPRI, in personal communication to L. E. Stoddard, Black & Veatch, January 31, 1991.

1A.A.1.4 Gasification Fuel Cells

A gasification fuel cell plant combines the gasifier technologies discussed in Subsection 1A.A.1.1 with molten carbonate fuel cell technologies discussed in Subsection 1A.A.2.3.[1,2] A clean medium Btu syngas fuel is supplied to a fuel cell by the gasifier and gas cleanup systems. The molten carbonate fuel cell generates electricity directly from the syngas (Subsection 1A.A.2.3). Heat recovered from the fuel cell is used to generate steam, which drives a steam turbine generator.

1A.A.1.4.1 Technological Maturity. The gasification fuel cell technology is in the concept stage, with several EPRI studies having been conducted to estimate plant costs and performance. Molten carbonate fuel cell development is in the bench-scale stage. Plans are underway for a 2 MW demonstration unit fired by natural gas. Gasification fuel cell systems are not likely to be commercially available until the late 1990s to early 2000s.

1A.A.1.4.2 Cost and Performance. A cost estimate for a mature technology gasification fuel cell plant is \$1,800/kW.[1] The estimated heat rate is 7,400 Btu/kWh.[1]

1A.A.1.4.3 Required Resources. Resource requirements for a gasification fuel cell system are likely to be similar to those for a GCC system.

1A.A.1.4.4 Environmental Impacts. Emissions from a gasification fuel cell plant will be very low, similar to a GCC plant. NO_x emissions should be lower than for the GCC plant.

1A.A.1.4.5 References.

1. TAG: Technical Assessment Guide, Volume 1, Electricity Supply, EPRI P-6587-L, November 1989.

2. Steven Myers, "Advanced Molten Carbonate Fuel Cell Systems Using BGL Gasification," presented at the Ninth Annual EPRI Conference on Gasification Power Plants, Palo Alto, California, October 17-19, 1990.

1A.A.1.5 Integrated Gasification Humid Air Turbine Cycle

An integrated gasification humid air turbine (IGHAT) cycle gasifies coal, producing a fuel gas for a humid air turbine (HAT) cycle. The HAT cycle used is similar to the one described in Subsection 1A.A.2.2. HAT cycles are not commercially available.

The coal gasification plant used is similar to the ones used in IGCC plants described in Subsection 1A.A.1.1. However, in an IGHAT plant the syngas is water quenched after it leaves the gasifier, and the resulting low level heat is recovered and used in the HAT cycle. This eliminates the need for expensive syngas coolers.[1] Figure 1A.A.1-3 shows a simplified flow diagram of a HAT cycle used in an IGHAT plant.

1A.A.1.5.1 Technology Maturity. IGHAT systems are not commercially available. A commercial prototype combustion turbine designed for the HAT cycle has not yet been built.

1A.A.1.5.2 Cost and Performance Characteristics. Because the IGHAT system has no steam cycle and expensive syngas coolers, the capital cost of an IGHAT plant is projected to be significantly less than a comparable sized IGCC plant. A preliminary capital cost estimate for a 475 MW IGHAT facility is \$1,160/kW.[2] The projected heat rate for the IGHAT plant is 8,700 Btu/kW.[2] Because the HAT cycle is thermodynamically superior to a combined cycle, the heat rate of an IGHAT plant is lower than that of a comparable sized IGCC plant.[1]

1A.A.1.5.3 Required Resources. Unit resource requirement for an IGHAT plant are similar to those for a similar sized IGCC plant. The major differences are that the land requirements are reduced because the bottoming steam cycle is eliminated. IGHAT consumes essentially the same amount of water as an IGCC plant.[2] However, larger water treatment facilities are required because the water vapor absorbed by the compressed air is discharged to the atmosphere and is not recycled.

1A.A.1.5.4 Environmental Impacts. Emission controls for sulfur removal described in Subsection 1A.A.1.1.4 of this appendix are applicable to IGHAT cycles. The HAT turbine used in the IGHAT cycle is expected to have NO_x emissions as low as 7 ppmvd (15 percent oxygen) or less without the use of SCR.[1]

1A.A.1.5.5 References.

1. "AFPS Developments," EPRI, Spring 1990, Issue 4.
2. A. D. Rao and J. R. Joiner, "A Technical and Economic Evaluation of the Humid Turbine Cycle," presented at the Ninth Annual EPRI Conference on Gasification Power Plants, Palo Alto, California, October 17-19, 1990.

1A.A.1.6 Coal Liquefaction

Shortages of oil in the 1970s and the prospect of future oil shortages has resulted in renewed efforts to synthesize liquid fuels from coal. Early liquefaction processes were developed and utilized in Germany during World War II when petroleum supply lines were severed. The technology for the principal process (Bergius) used in Germany has been lost; however, it is known that the thermal efficiency was quite low for that process. A second process developed during that time, the Fischer-Tropsch process, is currently being used in South Africa in the SASOL plants. The Fischer-Tropsch process is one of several indirect processes in which coal is first broken down into a synthesis gas, which is subsequently converted to liquid fuel or other chemicals. Indirect processes have an advantage of an extremely clean liquid product, but possess relatively low (45 to 60 percent) theoretical and demonstrated thermal efficiencies.

The relative low thermal efficiency of the indirect processes has caused the United States Department of Energy (DOE), along with private businesses, to focus attention on direct coal liquefaction processes. In direct processes, a coal slurry is heated under high pressure, and reacts with hydrogen (hydrogenated) to yield liquid hydrocarbons. Several types of direct processes are under development in the United States.

A third generic type of coal liquefaction process is pyrolysis. In this process, coal is heated to a very high temperature, which drives off the volatile matter and results in gaseous, liquid, and solid (char) products. A commercially available method for producing liquids by pyrolysis is the coke oven. A major disadvantage of this method is that it yields low liquid/high coke product ratio.

1A.A.1.6.1 Technology Maturity. Of the three generic methods for liquifying coal, only the indirect approach is currently used on a commercial scale. The SASOL I project near Sasolburg, South Africa, has been operational since 1955.

This state-owned plant, which uses the Fischer-Tropsch process (indirect), produces several thousand barrels of oil and petrochemical products per day. The indirect process used, however, results in a low thermal efficiency of 40 to 45 percent. The SASOL II and III projects are similar in concept to SASOL I. SASOL II and III primarily process coal into motor fuels, with each ton of coal yielding about 49 gallons of gasoline. Additional products include diesel fuel, ethylene, sulfur, coal tar derivatives, and ammonia. The three SASOL plants provide a significant portion of South Africa's liquid fuel requirements.

Because of the low thermal efficiency of the indirect method, and because of low liquid/high char yields of the pyrolysis method, significant development effort in the United States is being focussed on direct liquefaction technologies.

A primary emphasis within the United States using the indirect method has been methanol production. Methanol is generally produced from natural gas using commercially available processes; these processes can be used with the synthesis gas feedstock from a coal gasification process. However, developmental efforts are underway to produce methanol through processes which could be more cost-effective as coproduction applications. Development of these processes is still at the bench scale.

The largest full-time direct liquefaction research project in the United States is the Wilsonville Advanced Coal Liquefaction Research and Development Facility.[1,2] This facility, which is cosponsored by EPRI and DOE and managed by Southern Company Services, Incorporated, can process up to 6 tons of coal per day.

A 2.5 ton per day direct liquefaction facility is operated by British Coal Corporation at Point of Ayr in Great Britain.[3]

1A.A.1.6.2 Cost and Performance Characteristics. Costs of coal-derived liquids from existing commercial facilities (SASOL) are significantly higher than current costs of oil, although exact numbers are not available. Fuel costs based on a commercial-scale plant using the Wilsonville technology have been developed from a conceptual design. These fuels appear to be competitive with fuel costs when the price of crude oil is \$35 to \$40 per barrel.[1] The investment cost for the commercial-scale facility is about \$33,000 per barrel per day of capacity.[1]

Thermal efficiencies for the SASOL plants are low (45 to 60 percent). Thermal efficiencies for the Wilsonville technology on a commercial scale are estimated to be about 70 percent.[1]

1A.A.1.6.3 Required Resources. Coal liquefaction processes can use a variety of coals. The high reactivity and low cost of subbituminous coals make them attractive for the process, although problems have occurred with fouling using low rank coals in the Wilsonville facility.

1A.A.1.6.4 Environmental Impacts. Environmental impacts for coal-derived liquid fuels include the impacts of the facility producing the fuels as well as the combustion of the fuels at the power generation site. Emissions at the plants producing coal-derived liquid fuels are projected to be very low. Combustion tests of coal-derived liquid fuels show that emissions are similar to their petroleum-derived counterparts.

1A.A.1.6.5 References.

1. A. Basu, J. G. Masin, and N. G. Stewart, "Improvements in the Cost of Liquid Fuels from Direct Liquefaction," Proceedings of the Seventh Annual Pittsburgh Coal Conference, September 10-14, 1990.
2. Orlando Utilities Commission Long-Range Power Supply and Demand-Side Planning Study; Appendix B, Supply-Side Technologies, prepared by Southern Company Services, Inc., September 1988.
3. N. Robinson, S. A. Moore, G. M. Kimber, and M. D. Gray, "Transport Fuels from Coal: Modeling British Coal's Liquefaction Process," Proceedings of the Seventh Annual Pittsburgh Coal Conference, September 10-14, 1990.

1A.A.1.7 Magnetohydrodynamics

Magnetohydrodynamics (MHD) is a fossil fuel fired technology in which the electrical generator is static (nonrotating) equipment. Coal or another fossil fuel is burned to produce hot gases at temperatures exceeding 4,500 F.[1] Potassium carbonate is injected during the combustion process to increase the electrical conductivity of the combustion products.[2] These hot gases are accelerated in a nozzle and pass through a rectangular channel, which is surrounded by a superconducting magnet. The electrically conductive gases moving in the magnetic field produce a direct current in the channel's electrodes. This direct current is subsequently converted to alternating current with a solid-state inverter.[1] A general schematic of the MHD process is shown on Figure 1A.A.1-4.[3]

One of the more likely configurations for an MHD plant is to use the MHD generator as a topping cycle and a conventional steam power plant as the bottoming cycle. This configuration is attractive since the hot gases leave the MHD channel at a temperature of about 3,000 F. It is expected that an MHD steam, combined cycle plant will produce low emissions and have a thermal efficiency approaching 50 percent.[2]

The potassium present in the combustion products combines with the sulfur from the coal to form potassium sulfate. The potassium sulfate is precipitated and collected with ash in the steam generator and the electrostatic precipitator. It is then processed to regenerate the potassium carbonate seed.[3]

1A.A.1.7.1 Technological Maturity. The MHD technology as a whole is in the development stage. Power from an MHD test plant was first produced in 1959 with a 10-second long, 10 kW run. Numerous tests of MHD plants and components have been performed since then. The US Department of Energy currently operates two MHD test facilities: the Component Development and Integration Facility (CDIF) at Butte, Montana, and the Coal Fired Flow Facility (CFFF) at the University of Tennessee Space Institute in Tullahoma, Tennessee. Various private firms, including TRW and Westinghouse, are participating in MHD research, testing, and development.[3]

1A.A.1.7.2 Cost and Performance Characteristics. Capital and O&M costs for MHD systems are highly speculative because of the relatively early stage of development of the technology. The estimated capital cost for the direct fired MHD concept studied by Westinghouse Advanced Power Train is about $2,000/kW.[3] O&M costs for that plant are estimated to be about $9/MWh. The clean coal gas fired Westinghouse concept has an estimated capital cost of $2,900/kW. The O&M cost for that plant is about $12/MWh.

The unit performance of two commercial-scale MHD configurations has been addressed as part of the Westinghouse Advanced Power Train studies. The first configuration would burn coal directly and would include a 2,400 psig/ 1,000 F/997 F reheat steam bottoming cycle. Table 1A.A.1-1 presents data illustrating some of the major design performance parameters.[3]

The second configuration studied by Westinghouse would burn clean coal gas produced from a Texaco gasifier and would have a 1,450 psig/1,000 F/1,000 F reheat steam bottoming cycle. This concept results in a high percentage sulfur removal, but has relatively poor performance and high cost as currently configured. Some of the major design performance parameters are listed in the Table 1A.A.1-1.[3]

1A.A.1.7.3 Required Resources. The key resource for MHD is the fossil fuel, in this case coal. The direct burn concept discussed above has a coal feed rate of 4,600 tons/day (as received). The gasified coal concept has a coal feed of 5,700 tons/day of coal (as received) to produce 760,000 lb/h of clean coal gas at 525 psig and 570 F.[3]

Another resource is either potassium carbonate or potassium sulfate to serve as a seed to increase the electrical conductivity of the combustion products. The direct coal fired design uses 60,000 lb/h of potassium carbonate as the primary seed and 3,000 lb/h of potassium sulfate as makeup seed. In this design, potassium carbonate is regenerated from the potassium sulfate present in the ash that is produced during combustion of the coal. The clean (gasified) coal gas design uses a total of 34,600 lb/h of potassium sulfate, of which 33,000 lb/h is recovered from the ash and 1,700 lb/h is makeup seed.[3]

1A.A.1.7.4 Environmental Impacts. The MHD technology appears to be capable of adequately controlling NO_x and SO_2 emissions. NO_x is controlled by burning coal (or clean coal gas) in two stages. The first stage of combustion takes place in the primary combustor (before the MHD channel) where the coal is burned with approximately 85 to 90 percent of theoretical air. Although large amounts of NO_x are produced, the reducing conditions in the primary combustor cause the nitrogen oxides to decompose rapidly. The second stage of combustion occurs in the heat recovery steam generator (after the MHD channel) where the combustion products are further burned with approximately 105 percent of theoretical air. Because combustion takes place at lower temperatures with the cooler secondary air, less NO_x is produced.[4]

Sulfur dioxide removal is accomplished through the use of the potassium carbonate seed. The potassium combines with the sulfur in the combustion products, forming potassium sulfate. The sulfur is then removed from this compound during regeneration of the seed. In the direct coal fired design, 95 percent of the sulfur is removed from the combustion products and sulfur emissions are limited to 0.4 lb/MBtu.[3]

Solid waste disposal includes potassium sulfate, not a conventional waste for coal fueled stations. Specific disposal requirements are unknown at this time.

1A.A.1.7.5 References.

1. Combustion Fossil Power Systems, Combustion Engineering, 1981, pp. 24-34.
2. Energy Technology Status Report, California Energy Commission, October 1988, pp. 4-115 and 4-116.
3. Screening Evaluation of Advanced Power Cycles, EPRI AP-4826 Final Report, RP2477-1, November 1986, pp. 7-1 to 7-52.
4. J. N. Chapman, S. S. Strom, and Y. C. L. Wu, "MHD Steam Power-- Promise, Progress, and Problems," Mechanical Engineering, September 1981, pp. 30-37.

Table 1A.A.1-1

Predicted Design Parameters for MHD Facilities

Parameter	Direct Burn MHD	Coal Gasification MHD
Net Plant Output, MW	500	470
MHD Power Output, MW	191	136
Steam Turbine Power Output, MW	401	465
Coal Gas Expander Output, MW	--	30
Internal Power Consumption, MW	92	161
Net Heat Rate, Btu/kWh	8,616	11,289
Overall System Efficiency, percent	39.6	30.2
Sulfur Removal, percent	95	99.3

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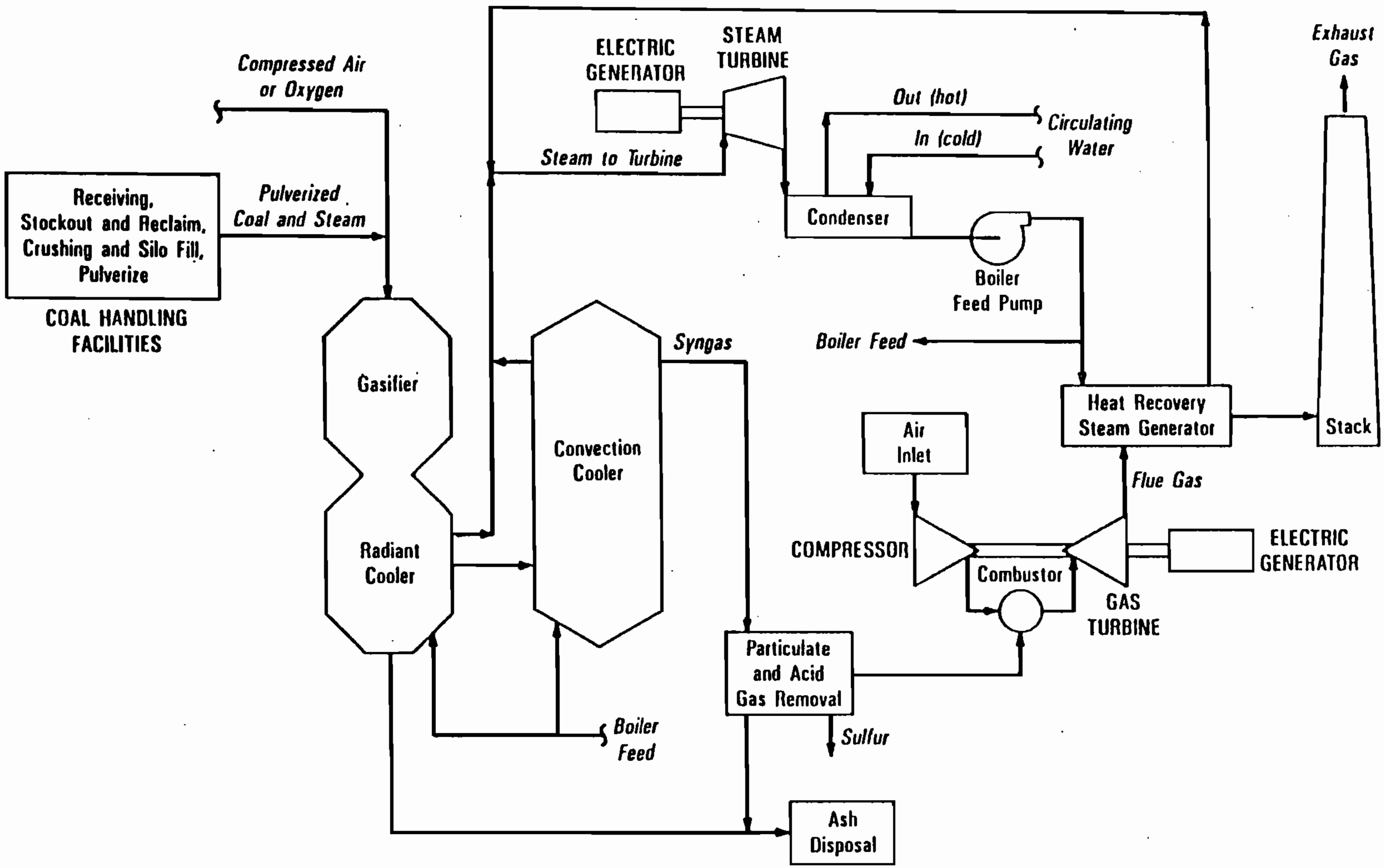
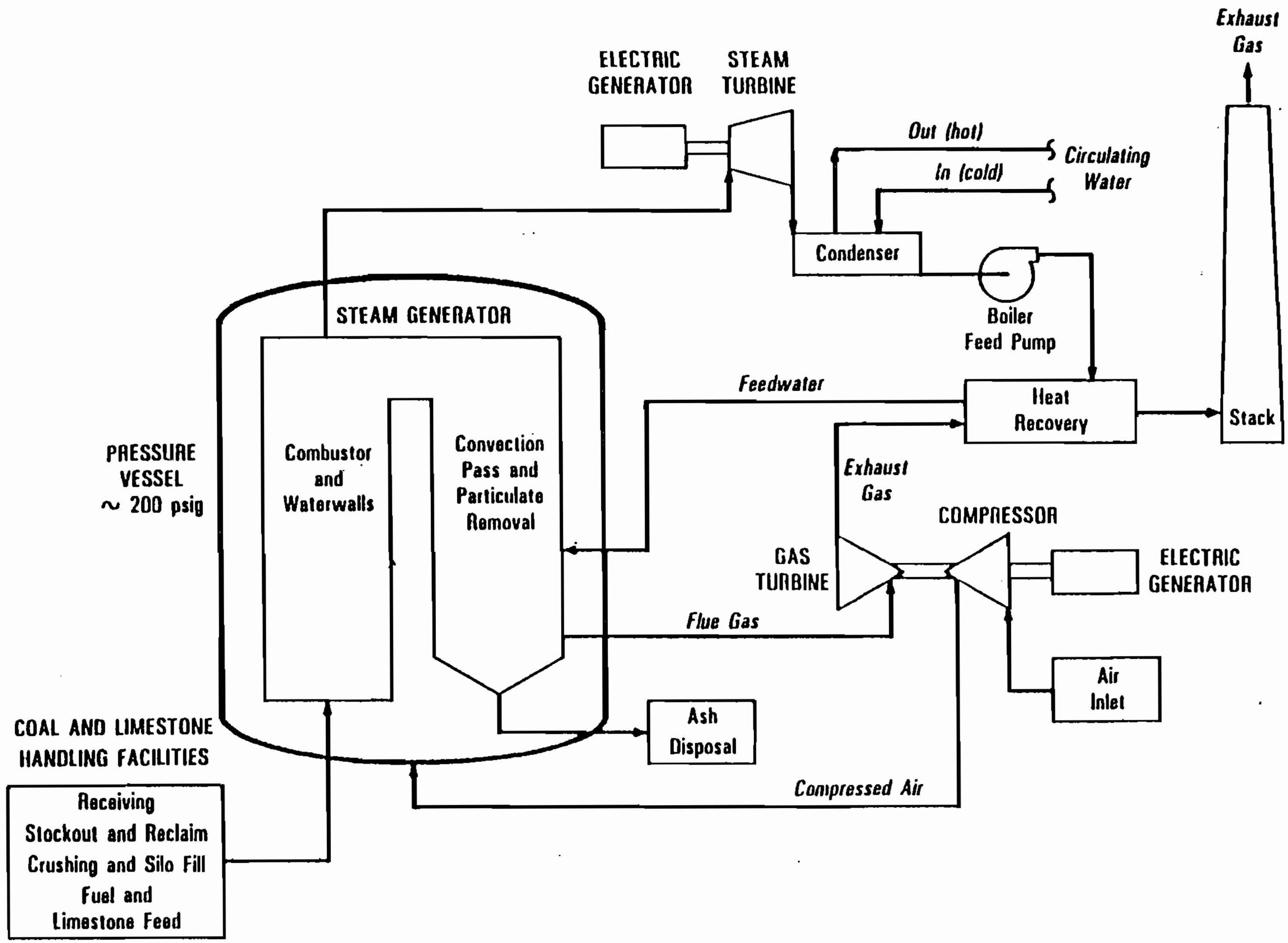


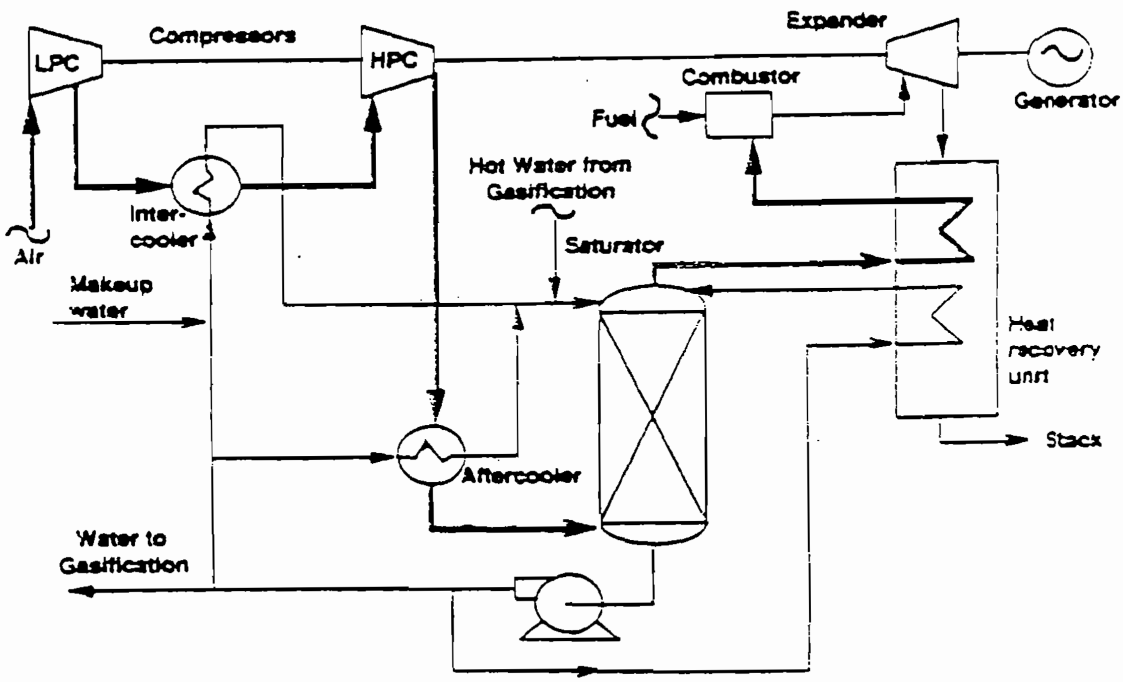
GASIFICATION COMBINED CYCLE PLANT
FLOW DIAGRAM

Figure 1A.A.1-1



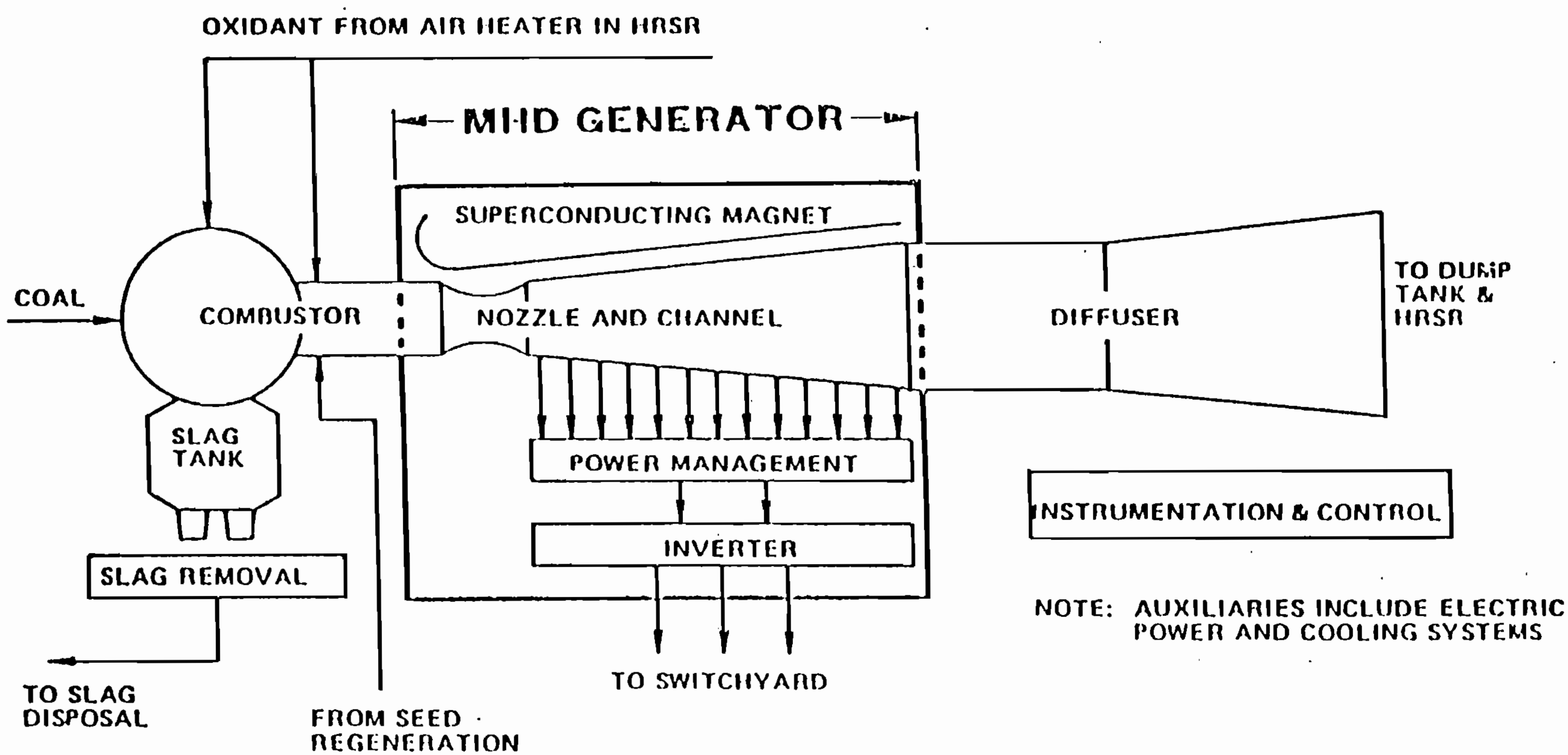
PRESSURIZED FLUIDIZED BED PLANT
COMBUSTION SYSTEM PLANT FLOW DIAGRAM

Figure 1A.A.1-2



INTEGRATED COAL GASIFICATION
HUMID AIR TURBINE CYCLE FLOW DIAGRAM

Figure 1A.A.1-3



POWER TRAIN SCHEMATIC SHOWING MHD GENERATOR

Figure 1A.A.1-4

Source: Reference 3

1A.A.2 Oil or Gas Fired Alternatives

Oil or gas fired alternatives considered in this section include the steam injected combustion turbine, humid air turbine (HAT) cycle, and fuel cells.

1A.A.2.1 Steam Injected Combustion Turbine (Power Augmentation)

The steam injection combustion turbine uses steam injected into the combustor and turbine sections of a conventional combustion turbine to increase power output. This is called power augmentation and is also known as the Cheng cycle.

The injection steam has a cooling effect on the combustion turbine internals, allowing greater fuel burn rates. There is no steam turbine or associated equipment as in the combined cycle arrangement. The basic components for this cycle, shown on Figure 1A.A.2-1, are the combustion turbine and the heat recovery steam generator (HRSG).

Aero-derivative combustion turbines are typically used in power augmentation applications because of the higher compression ratio associated with the aero-derivative. The aero-derivative also has the ability to pass greater flows.

The HRSG is required to produce steam for injection into the combustion turbine and for process use. The typical application of the power augmentation cycle is a cogeneration unit where excess power can be produced during low cogeneration demand and/or peak electrical demand periods.

1A.A.2.1.1 Technological Maturity. Several power augmentation units began operation in the early to mid-1980s. The cycle is considered a mature technology. However, the long-term impact of massive steam injection on combustion service life is still being evaluated. Furthermore, combustion turbines designed specifically for steam injection are not commercially available today.

1A.A.2.1.2 Cost and Performance Characteristics. Capital costs for the power augmentation cycle vary with unit size, amount of steam injected, and amount of steam to be produced for cogeneration. Preliminary capital cost estimates range from $800/kW for a 150 MW unit to $1,050/kW for a nominal 50 MW installation.

Based on manufacturers' estimates, the steam injected combustion turbine maintenance costs would increase about 15 to 20 percent over simple cycle combustion turbine maintenance costs because the greater mass flow through the turbine results in increased frequency of maintenance.

The construction schedule for this cycle would be similar to that of the combined cycle unit.

The power augmentation cycle is very flexible with respect to the amount of steam injected relative to output desired. Steam injection can increase combustion turbine power output over 50 percent and decrease heat rate up to 20 percent, depending on the size and design of the unit and amount of steam injection. A General Electric LM5000 system with 133,000 pounds per hour of steam injection is shown in Table 1A.A.2-1 as an example of power augmentation performance.[1] This includes both high- and low-pressure steam injection based on typical unfired HRSG performance.[2]

1A.A.2.1.3 Required Resources. Unit resource requirements for a steam injected combustion turbine are very similar to those for a combined cycle unit. The major differences are that a significant continuous water source and a larger water treatment facility are required because the steam injected into the combustion turbine is discharged to the atmosphere and is not recycled. Land requirements are reduced because the steam turbine and cooling tower are not required.

1A.A.2.1.4 Environmental Impacts. Emissions control described in the simple cycle and combined cycle sections of this report are directly applicable to combustion turbines using steam injection for power augmentation.

1A.A.2.1.5 References.

1. "Aero-derivative Gas Turbine Performance, Emissions, and STIG," GE Gas Turbine Reference Library, GER-3572, 1988.
2. General Electric Model PG7111(EA) Gas Turbine Estimated Performance, 499HA733, July 8, 1988.

1A.A.2.2 Natural Gas Fired Humid Air Turbine Cycle

The humid air turbine (HAT) cycle uses a combustion turbine that employs air intercooling and the substitution of water vapor for a portion of the excess air to the combustor to reduce the power requirement by the compressor. This reduction in air compression load significantly increases the net power output by the turbine.[1]

Figure 1A.A.2-2 shows the basic HAT cycle. Intercooling and aftercooling provides some of the heat required for generating water vapor. Additional low temperature heat required for generating water vapor is also obtained from the turbine exhaust. A saturator generates water vapor in a multi-stage counter-current operation by directly contacting cooled compressed air with hot water.[1] The humidifed air is then further heated by the turbine exhaust prior to entering the combustor.

1A.A.2.2.1 Technological Maturity. Natural gas fired HAT cycles are not commercially available. The HAT cycle is still in the early developmental stages and a commercial prototype combustion turbine designed for the HAT cycle has not yet been built.

1A.A.2.2.2 Cost and Performance Characteristics. A preliminary capital cost estimate for a nominal 200 MW plant size is $630/kW.[1] Table 1A.A.2-2 shows the expected performance of a natural gas fired HAT cycle.[1] The performance of a combined cycle using a GE 7F combustion turbine is also shown for comparison. The HAT cycle heat rate is lower than that of a combined cycle.

1A.A.2.2.3 Required Resources. Unit resource requirements for a HAT cycle are comparable to those of a similar capacity combined cycle. Land requirements are reduced because the steam turbine and cooling tower are not required. The water requirements are comparable to that of a combined cycle plant with cooling towers.[3] Larger water treatment facilities are required by the HAT cycle because the water vapor absorbed by the compressed air is discharged to the atmosphere and is not recycled.

1A.A.2.2.4 Environmental Impacts. NO_x emissions are expected to be very low over the full range of capacity due to the high moisture levels in the air going to the combustor.

1A.A.2.2.5 References.

1. A. D. Rao and J. R. Joiner, "A Technical and Economic Evaluation of Turbine Cycle," presented at the Ninth Annual EPRI Conference on Gasification Power Plants, Palo Alto, CA, October 17-19, 1990.
2. "AFPS Developments," Electric Power Research Institute, Spring 1990, Issue 4.
3. Ashok D. Rao and Thomas R. Morton, "Perspective for Advance High Efficiency Cycles using Gas Turbines," 1989 EPRI Conference on Technologies for Producing Electricity in the 21st Century, San Francisco, CA, October 30-November 2, 1989.

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Unit 2
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1D.8.2 Advanced Alternatives

A detailed discussion of advanced alternatives is presented in Subsection 1A.5.0. A total of 27 potentially viable advanced alternative technologies are discussed to determine their applicability as an alternative to Stanton 2. Most were eliminated because of the lack of adequate demonstration of the technology, their comparatively high cost versus conventional alternatives or because Florida is not well suited to the technology's requirements. Of the 27 advanced alternatives reviewed, only three were carried forward to the screening curve analysis by OUC. These three were a coal gasification combined cycle unit, a solar thermal parabolic trough, and lead-acid batteries.

The results of OUC's screening curve analysis indicated that none of these advanced alternatives were cost-effective for OUC. Furthermore, a utility as small as KUA cannot afford the risk associated with advanced alternatives. Since KUA has similar economics to OUC and since KUA cannot afford the risk associated with advanced alternatives, the advanced alternatives will not be considered further as an alternative to Stanton 2.

1D.8.3 Cost and Operating Characteristics for Conventional Alternatives

A system planning study must evaluate the viability of other generation alternatives before making conclusions as to the cost-effectiveness of a proposed capacity addition. Among the more important alternatives are those generation alternatives fitting a utility's requirements that are commonly used in the industry and that have a proven record regarding cost, reliability, and fuel availability. These are termed conventional alternatives.

Given KUA's capacity requirements and fuel price and availability considerations, the conventional alternatives evaluated for the KUA system are a coal fired atmospheric fluidized bed unit, a combustion turbine, and a combined cycle unit. Cost and performance data for each of these units are provided in this section. The cost estimates provided in this section include the offsite and onsite facilities required for operation of these units at the Cane Island site.

1D.8.3.1 Atmospheric Fluidized Bed Combustion Unit

Atmospheric fluidized bed combustion (AFBC) systems are in an advanced state of development and are currently being applied in electric utility systems. The largest single circulating AFBC unit in operation is a 96 MW unit in Trona, California, which was designed by Black & Veatch. The largest single unit scheduled for construction is a 165 MW unit at Point Aconi for Nova Scotia Power Corporation.

In a circulating fluidized bed steam generating unit, fuel and sorbent are fed by gravity into a highly turbulent region of the combustion chamber where combustion air, feed fuel, sorbent, and recirculating solids are mixed. The reacting gas and solids flow upward and then enter a particulate separator, usually a cyclone, where the solids are separated and returned to the bottom of the combustion chamber. Flue gas leaves the cyclone and is cooled by a conventional convection bank and an air heater. Additional particulate removal is required before the gas is discharged to the stack.

Flue gas desulfurization is accomplished in fluidized bed combustion units by adding limestone to the bed and promoting high temperature reaction of calcium oxide with sulfur dioxide (SO_2). Removal of SO_2 is increased by increasing the feed rate of limestone to the bed. Current operating experience with AFBC units indicates that relatively high levels of excess limestone addition (Ca/S ratio of 2.0

to 3.0) are required to achieve 90 percent SO_2 removal. Higher SO_2 removal rates (95 percent) have not been demonstrated on a continuous basis, and are expected to require much higher limestone consumption rates.

A current drawback to implementation of AFBC units over conventional pulverized coal units is the relatively high waste disposal costs associated with AFBCs. Although each of the units generates about the same amount of fly ash, the desulfurization byproducts generated are about 50 percent higher for AFBC units as compared with those of the pulverized coal unit. Since fly ash is not separately collected in the AFBC, it cannot be marketed as is sometimes done at a pulverized coal unit.

Generation of nitrogen oxides (NO_x) in AFBC units is lower than that of pulverized coal units because of the lower combustion temperatures inherent to the process.

AFBC units have accumulated several years of operating experience in the 50 MW size range appropriate for consideration by KUA. It is expected that maintenance requirements will be well defined and preventive maintenance programs will enable fluidized bed boilers to operate with availability comparable to that of conventional pulverized coal units. However, AFBC units may experience higher boiler maintenance due to the potential for greater heat transfer surface erosion.

Estimated performance and cost data for a 50 MW circulating AFBC coal unit are provided in Table 1D.8.3-1. The cost estimate includes estimates for offsite rail facilities and onsite water treatment facilities. The water treatment facilities are sized for treatment of sewage treatment effluent for makeup to the demineralizer (cycle water), plant service, and circulating water systems. The AFBC unit would utilize the common facilities at the Cane Island site as presented in the <u>KUA Combustion Turbine Site Feasibility Study</u>. The costs associated with these common facilities, such as the demineralizer system and water storage tanks, are considered a capital cost burden on the first combustion turbine unit installed at the site and are not included in the AFBC capital cost estimate.

1D.8.3.2 Combustion Turbine Unit

Combustion turbines typically burn natural gas and/or No. 2 oil and are available in a wide variety of sizes and configurations. This section provides cost and performance estimates for the addition of a single combustion turbine unit at the Cane Island site. The estimates are based on a General Electric PG6541(B) combustion turbine (Frame 6) with an International Standards Organization (ISO) rating of 38.3 MW on natural gas at 59 F ambient temperature and sea level. The rating would be 43.3 MW at normal winter peak conditions of 25 F. Performance estimates are based on burning natural gas fuel in a standard combustor using water injection for NO_x control to 42 ppm.

Combustion turbines are generally used for peaking and reserve purposes because of their relatively low capital cost, poorer full load heat rate, and the higher cost of fuel when compared to conventional baseload capacity. Combustion turbines have the added benefit of providing quick-start capability in certain configurations.

Estimated performance and cost data for the Frame 6 combustion turbine system are provided in Table 1D.8.3-2. The cost estimates are based on the installation of a second combustion turbine at the Cane Island site as described in the KUA Combustion Turbine Site Feasibility Study. Only those costs associated with the installation of the second combustion turbine power block are included in the direct capital cost estimate. The cost of the common facilities at the Cane Island site as shown in the above-mentioned study, although sized to support two Frame 6 combustion turbines, are considered a capital cost burden for the first combustion turbine which will already be installed at the site.

1D.8.3.3 Combined Cycle Unit

In a combustion turbine combined cycle facility, the hot exhaust gases from the combustion turbine pass through a heat recovery steam generator (HRSG). The steam generated by the HRSG is expanded through a steam turbine which, in turn, drives an additional generator. With a typical gas inlet temperature of 1,000 F into the HRSG, steam conditions of about 1,000 psi and 900 F can be produced by the HRSG.

Combustion turbine combined cycle systems typically burn natural gas and/or No. 2 oil and are available in a wide variety of sizes and configurations. This section provides cost and performance estimates for the addition of a combustion turbine combined cycle unit at the Cane Island site. The estimates are based on a Frame 6 combustion turbine in combined cycle operation with a single HRSG and steam turbine. This unit configuration can produce a net plant output of approximately 60 MW. Performance estimates are based on burning natural gas fuel in a standard combustor using water injection for NO_x control to 42 ppm, with no supplemental firing in the HRSG.

Estimated performance and cost data for the 60 MW combined cycle system are provided in Table 1D.8.3-3. The cost estimates are based on the installation of a second Frame 6 combustion turbine with a single HRSG and steam turbine at the Cane Island site. Only those costs associated with the installation of the combined cycle power block and the water treatment and storage facilities of sewage treatment effluent are included in the direct capital cost estimate. The water treatment facilities are sized for treatment of 1,008,000 gpd of sewage treatment effluent as makeup water for the demineralizer (injection and cycle water), plant service, and circulating water. The common facilities at the Cane Island site, although sized to support two Frame 6 combustion turbines, are considered a capital cost burden for the first combustion turbine installed at the site.

Table 1D.8.3-1
Circulating Atmospheric Fluidized Bed Technology
Unit Cost and Performance Data

Net Plant Output, MW*	50
Direct Capital Cost, 1990$/kW**	2,410
Indirect Cost, percent	18.0
Fixed O&M Cost, 1990$/kW-yr	47
Variable O&M, 1990$/MWh	3.00
Net Plant Heat Rate, Btu/kWh	
Full load	10,360
75 percent load	10,570
50 percent load	11,240
25 percent load	13,330
Equivalent Availability, percent	85
Forced Outage Rate, percent	6
Planned Maintenance Outage, weeks/year	4
Construction Period, months	24

*Based on one boiler serving one steam turbine generator, VWO-OP.

**Does not include escalation, indirect cost, or interest used during construction.

Table 1D.8.3-2
Combustion Turbine Unit and Performance Data

	GE Frame 6
Net Plant Output, MW*	38.3
Direct Capital Cost, 1990$/kW**	280
Indirect Cost, percent	12.0
Fixed O&M Cost, 1990$/kW-yr	9.50
Variable O&M, 1990$/MWh	
Peaking load, 5 percent capacity factor	3.7
Intermediate load, 25 percent capacity factor	1.3
Net Plant Heat Rate, Btu/kWh HHV*	
Full load	12,760
75 percent load	13,530
50 percent load	15,250
25 percent load	21,060
Equivalent Availability, percent	89
Forced Outage Rate, percent	5
Planned Maintenance Outage, weeks/year	2
Construction Period, months	10

*59 F ambient temperature and 14.9 psi.

**Does not include escalation, indirect cost, or interest during construction.

Table 1D.8.3-3
Combined Cycle Unit and Performance Data

	CC With GE 6
Net Plant Output, MW*	60
Direct Capital Cost, 1990$/kW**	670
Indirect Cost, percent	13.0
Fixed O&M Cost, 1990$/kW-yr	12.0
Variable O&M, 1990$/MWh	1.80
Net Plant Heat Rate, Btu/kWh HHV*	
Full load	8,620
75 percent load	8,870
50 percent load	9,860
25 percent load	10,300
Equivalent Availability, percent	86
Forced Outage Rate, percent	6
Planned Maintenance Outage, weeks/year	4
Construction Period, months	16

*Based on one combustion turbine, one unfired HRSG, and one steam turbine at 59 F and 14.9 psi.

**Does not include escalation, indirect costs, or interest during construction.

1D.8.4 Combined Cycle Conversion

Conversion of the proposed Cane Island combustion turbine to combined cycle operation could provide additional baseload capacity for the KUA system. However, the conversion of the Cane Island combustion turbine to combined cycle would effectively eliminate the peaking capacity of the combustion turbine requiring it to be replaced with other peaking capacity. For that reason, it is more reasonable to evaluate the addition of a new combined cycle unit as described in Subsection 1D.8.3 rather than convert the proposed Cane Island combustion turbine to combined cycle.

ORLANDO UTILITIES COMMISSION

CURTIS H. STANTON ENERGY CENTER
UNIT 2

SUPPLEMENTAL SITE CERTIFICATION APPLICATION

VOLUME 2

Introduction

The Supplemental Site Certification Application is being submitted by Orlando Utilities Commission (OUC), Florida Municipal Power Agency (FMPA), and Kissimmee Utility Authority (KUA) for the Unit 2 addition to the Curtis H. Stanton Energy Center (Stanton) in accordance with the Florida Electrical Power Plant Siting Act. The site was previously certified for an ultimate site development of approximately 2,000 MW of coal fired capacity at the same time that Unit 1 was certified (order entered December 15, 1982; DOAH Case No. 81-1431). The certified site included corridors for the railroad, transmission lines, makeup water supply pipelines, and the site access road. This supplemental application is filed pursuant to FS 403.517 regarding supplemental site certification applications. No DER instruction guide has been prepared for supplemental site certification applications, so this application has followed the outline of DER Form 17-1-211(1) for certification applications. Information in this supplemental application is supplied only for the purpose of assessing the need, construction, and operation of Stanton 2. This supplemental application is intended to serve the following purposes.

- Supplemental Site Certification for Unit 2 of the Stanton Energy Center including the Need for Power chapter.
- Revision to Best Available Control Technology (BACT) for the existing US Environmental Protection Agency approval of Stanton 2 under a two-unit phased construction Prevention of Significant Deterioration (PSD) permit and request for an amendment of the commence construction date for Unit 2 in the original PSD permit.
- Joint Application (US Army Corps of Engineers and Florida Department of Environmental Regulation) for dredge and fill permit for site development and construction of all associated facilities for Stanton 2 as may be required by the Clean Water Act.
- Permit applications for all other required state, regional, and local approvals.

A PSD permit (PSD-FL-084) was issued to OUC on June 10, 1982, for the phased construction of Units 1 and 2 at the Stanton Energy Center. Within this *permit, the construction of Stanton 2 was scheduled to commence on* July 1, 1990, with expected startup in January 1994. The authority to construct Stanton 2 under this permit will expire on January 1, 1992, unless construction has commenced by that date. This application includes the revised BACT and request for an extension in the commence construction deadline for a period of 18 months.

The Supplemental Site Certification Application comprises five volumes. The first four volumes contain Chapter 1 of the Supplemental Site Certification Application and are labeled 1A, 1B, 1C, and 1D. Chapter 1 contains the Public Service Commission Need for Power (NFP) Application portion of the Supplemental Site Certification Application. The Joint Need for Power Application is based on the needs of the joint participants in Stanton 2. The joint participants are Orlando Utilities Commission (OUC), Florida Municipal Power Agency (FMPA), and Kissimmee Utility Authority (KUA). Volumes 1A through 1D contain the following information.

- 1A--NFP Information Common to All Participants.
- 1B--NFP Information Specific to OUC.
- 1C--NFP Information Specific to FMPA.
- 1D--NFP Information Specific to KUA.

Appropriate appendices are included at the end of each volume. The last volume (Volume 2) contains Chapters 2.0 through 10.0 relating to all aspects of the Supplemental Site Certification Application other than the need for power.

APPLICANT INFORMATION

Applicants' Official Names and Mailing Addresses

Orlando Utilities Commission
P.O. Box 3193
Orlando, Florida 32802

Florida Municipal Power Agency
7201 Lake Ellenor Drive
Orlando, Florida 32809

Kissimmee Utility Authority
P.O. Box 423219
Kissimmee, Florida 34742-3219

Address of Official Headquarters

Orlando Utilities Commission
500 South Orange Avenue
Orlando, Florida 32801

Florida Municipal Power Agency
7201 Lake Ellenor Drive
Orlando, Florida 32809

Kissimmee Utility Authority
8 Broadway
Kissimmee, Florida 34741

Business Entity

Orlando Utilities Commission (OUC) is a statutory commission created by the legislature of the State of Florida as a separate part of the government of the City of Orlando. OUC has the full authority over the management and control of the electric light and water works parts of the City of Orlando. It has the power to undertake, among other things, the construction, operation, and maintenance of electric generation, transmission and distribution systems, and water production, transmission and distribution systems in order to meet the requirements of its customers.

Florida Municipal Power Agency (FMPA) is a joint agency formed pursuant to the Interlocal Cooperation Act and exercises powers under the Joint Power Act. FMPA has authority to undertake and finance electric projects and, among other things, to plan, finance, acquire, construct, reconstruct, own, lease, operate, maintain, repair, improve, extend, or otherwise participate jointly in those projects and to issue bonds or bond anticipation notes for the purpose of financing or refinancing the costs of such projects.

Kissimmee Utility Authority (KUA) is a public body, corporate and politic, duly organized, and legally existing as part of the government of the City of Kissimmee engaged in the generation, transmission, and distribution of electric power to persons within the service area.

Name and Titles of Chief Executive Officers

Orlando Utilities Commission

Jerry Chicone, Jr.	-	President
Royce B. Waldon	-	First Vice President
Susan T. McCaskill Little	-	Second Vice President
James H. Pugh, Jr.	-	Past President
Bill Frederick	-	Mayor
Theodore C. Pope	-	Executive Vice President and General Manager
Thomas B. Tart	-	General Counsel

Florida Municipal Power Agency

Dean G. Shaw	-	Chairman
Joseph M. Tardugno, Jr.	-	Vice Chairman
Harry M. Schindehette	-	Secretary - Treasurer
Vincente R. Ruano	-	Assistant Secretary - Treasurer

Kissimmee Utility Authority

Richard L. Hord	-	Chairman
Bob Bobroff	-	Director
Harry Lowenstein	-	Director
Arnold W. Jones	-	Director
George A. Gant	-	Director
John B. Pollet	-	Mayor (Executive Officio)

Name, Address, and Telephone Number of Official Representative Responsible for Obtaining Certification

Orlando Utilities Commission
Thomas B. Tart, General Counsel
500 South Orange Avenue
Orlando, Florida 32801
407-423-9123

Site Location (County)

Orange County

Nearest Incorporated City

Orlando, Florida

Latitude and Longitude

29° 29' North Latitude
81° 10' West Longitude

UTMs (Center of Site)

Northerly 1507528
Easterly 446825

Section, Township, Range

Sections 13, 14 (E 1/2), 23 (E 1/2), 24; Township T23S; Range R31E
Sections 18, 19; Township T23S; Range R32E

Location of Any Directly Associated Transmission Facilities (Counties)

Orange County

Nameplate Generating Capacity

Stanton 2 will have a nameplate gross generating capacity of 465 megawatts (MW) and is scheduled for commercial operation January 1, 1997.

Capacity of Proposed Additions and Ultimate Site Capacity

Stanton 3 and 4 have not been sized or scheduled at this time. The Stanton site was previously certified for an ultimate site capacity of approximately 2,000 MW.

2.3.3 *Site Water Budget and Area Uses*

Detailed information regarding the site and region water budgets was presented in Sections 2.5 and 2.6 of the original SCA. A discussion of area water uses was included in Subsection 2.2.3 of the original SCA. Water supply wells installed in the region since submittal of the original SCA are shown on Figure 2.3-1. Three water supply wells have been installed within one mile of the site. The well owners and specific well information are summarized in Table 2.3-1.

2.3.4 *Surficial Hydrology*

Descriptions of regional and local surface waters are included in Subsection 2.5.1 of the original SCA.

2.3.5 *Vegetation/Land Use*

A discussion of the project area's plant communities is included in Subsection 2.7.2 of the original SCA. The existing land use features are described in Subsection 2.2.2 of the original SCA.

2.3.6 *Ecology*

Detailed descriptions of the ecological communities in the site area are included in Section 2.7 of the original SCA.

2.3.7 *Meteorology and Ambient Air Quality*

Discussions of regional climate, site meteorology, and site air quality are included in Sections 2.6 and 2.8 of the original SCA. These discussions include information obtained from the Stanton 1 preconstruction onsite monitoring program, which began operation in May 1980. OUC has continued to operate onsite monitoring with some changes in program configuration. Stanton 1 was put into commercial service in July 1987. Even with the operation of this new source, measurements taken at the site during the entire period of approximately 10 years have been low, well below applicable ambient air quality standards.

Because the initial Prevention of Significant Deterioration (PSD) permit was issued for both Stanton 1 and 2, the Florida Department of Environmental Regulation (FDER) has indicated that preconstruction air monitoring will not be a requirement for the approval of the requested construction schedule extension for Stanton 2.

2.3.8 Noise

A detailed discussion of ambient noise levels at the Stanton Energy Center site and in the immediate vicinity was included in Subsection 2.9.1 of the original SCA.

2.3.9 Other Environmental Features

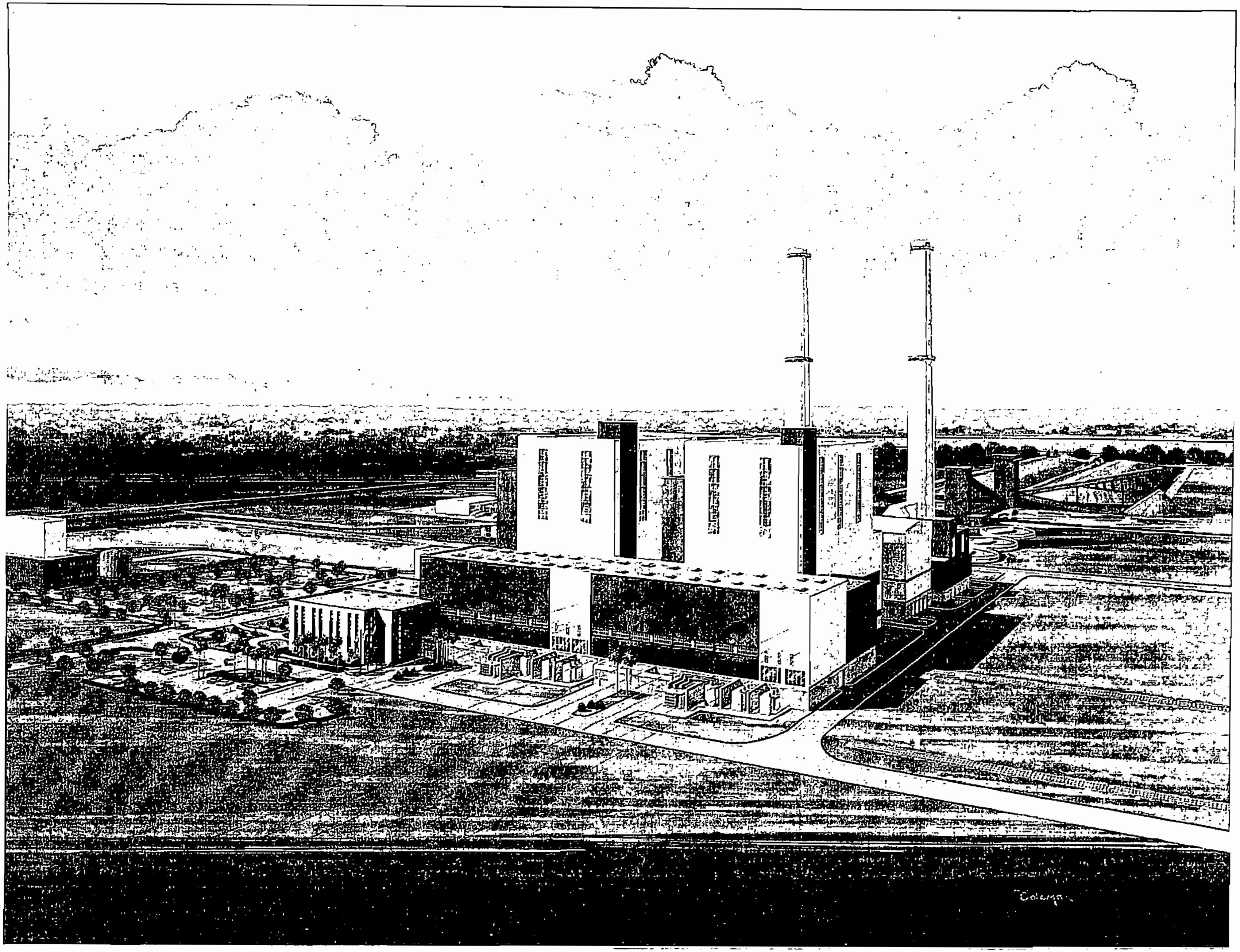
No other features pertinent to the environmental evaluation of the proposed Stanton 2 addition have been identified.

3.0 The Plant and Directly Associated Facilities

The information presented in this chapter is provided only for the purposes of construction and operation of Stanton 2 as discussed and qualified in the Introduction.

3.1 Background

An artist's rendering of the Stanton Energy Center site with the proposed Stanton 2 added is shown on Figure 3.1-1. Stanton 2 will be a 465 MW gross, 440 MW net, pulverized coal fueled steam/electric power plant. This size is totally consistent with the original ultimate site certification. New Stanton 2 facilities will include sulfur dioxide removal equipment, electrostatic precipitator, chimney, cooling tower, and an expansion of the cooling tower blowdown treatment system. Other facilities previously constructed for Stanton 1 will also be used for Stanton 2. These include the onsite ponds and basins; materials handling and storage systems for coal, oil, limestone, lime, and combustion wastes; administration building; warehousing; and other common support facilities.



ARTIST'S RENDITION OF
CURTIS H. STANTON ENERGY CENTER — 2 UNITS

Figure 3.1-1

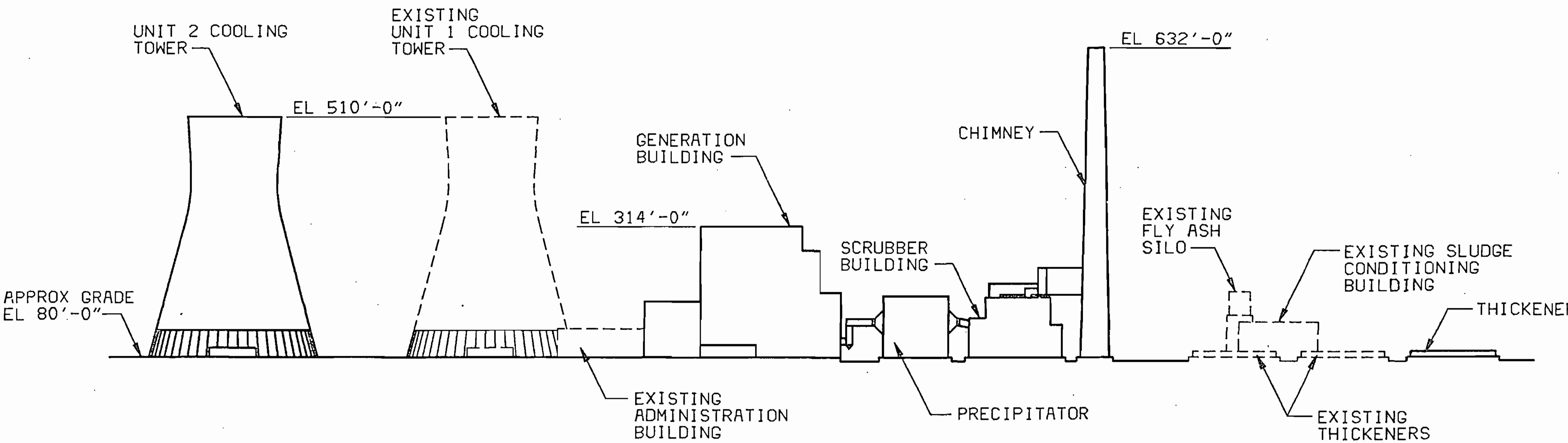
3.2 Site Layout

Figure 3.2-1 shows the proposed layout for the Stanton Energy Center with the addition of Stanton 2 and associated facilities. The new onsite facilities will occupy approximately 9 acres of the 3,280-acre site.

A general plant profile of the total two-unit power plant is shown on Figure 3.2-2. This profile is based on the elevations of the various facilities as viewed from the north looking south. The elevations and dimensions shown on Figure 3.2-2 are based on Stanton 1 sizes and considered preliminary. Some changes may occur as detailed engineering design proceeds.

The Stanton Energy Center is designed so that no wastewater discharges to surface waters will be necessary. A detailed description of the water supply, uses, and management is presented in Section 3.5.

Release points for fugitive dust and combustion gas wastes were identified and discussed in the original SCA. The only new source of gaseous wastes will be the 550-foot chimney for Stanton 2. Further, this new air emissions source was previously evaluated with regard to its effect on air quality and is included as an approved source in the Prevention of Significant Deterioration (PSD) permit issued for the project during the initial permitting process. A discussion of the proposed air quality control systems for Stanton 2 is presented in Section 3.4.



PLANT VIEW LOOKING SOUTH

GENERATION FACILITIES PROFILE

Figure 3.2-2

3.3 Fuel

3.3.1 *Fuel Types and Qualities*

The primary fuel for the Stanton Energy Center will continue to be bituminous coal. Although coal supply contracts for Stanton 2 have not been finalized, a design basis coal has been developed for use on the project. Table 3.3-1 presents the typical and ranges of selected properties of the design basis coal. All other properties are expected to fall within the ranges provided for coal quality in Table 3.2-1 of the original SCA. These coal properties provide a "worst case" design basis that will provide OUC with system operating flexibility to burn any coal with properties in the ranges given.

The Stanton 2 steam generator will be started with No. 6 fuel oil. During periods of low load operation, No. 6 fuel oil will also be used for flame stabilization.

7,276 mmBtu/hr coal 4,005 mmBtu/hr

3.3.2 *Fuel Quantities*

With a heating value range of 11,000 to 13,350 Btu/lb, the maximum coal consumption rate will be 150 to 193 tons/h. At a 70 percent capacity factor, the annual coal consumption for Stanton 2 will be 975,000 to 1,190,000 tons per year. Estimates of No. 6 fuel oil usage for cold and hot startups with Stanton 2 are the same as those discussed for Stanton 1 in Subsection 3.2.2 of the original SCA.

3.3.3 *Fuel Transportation*

Coal will continue to be transported to the plant site by rail via CSX Transportation, Inc., as described in Subsection 3.2.3 of the original SCA. An additional two to three trains per week will be required to supply coal for Stanton 2. No. 6 fuel oil will continue to be received by trucks as has been the case for Stanton 1.

3.3.4 *Coal Handling and Storage*

The coal handling system described in Subsection 3.2.4 of the original SCA will serve both Stanton 1 and 2. The existing system as constructed consists of unloading, stocking, reclaiming, and storage facilities. The system will be unchanged except for the addition of new silo fill conveyors and plant silos to serve Stanton 2. Figure 3.2-3 of the original SCA shows a perspective of these facilities.

3.3.5 *Fuel Oil Storage and Handling*

The No. 6 fuel oil required by Stanton 2 will be stored in the previously installed onsite tanks, described in Subsection 3.2.5 of the original SCA.

3.3.6 *Alternate Fuel Types*

As in the case of Unit 1, no special design features have been included in the design of Stanton 2 to allow burning of alternate fuels.

Table 3.3-1
Design Basis Coal Properties

Ultimate Analysis	Typical
Carbon	67.0 percent
Hydrogen	4.50 percent
~~Sulfur~~	~~2.5 percent~~
Moisture	7.5 percent
Nitrogen	1.29 percent
Chlorine	0.11 percent
Oxygen	5.1 percent
Ash	12.00 percent
Higher Heating Value	12,400 Btu/lb

3.4 Air Emissions and Controls

It is OUC's philosophy for the construction, operation, and maintenance of facilities, to focus on safety, reliability, and redundancy, all accomplished while maintaining an environmentally responsible posture. These goals are achieved by following a course whose bounds are well within the conservative constraints of prudent utility practice.

Following this philosophy, OUC avoids using unproven technologies or technologies applied in an unproven manner. Conservative prudent utility practice requires a diversity of fuels and maximum fuel flexibility within OUC's generation system. Further, OUC will operate its units well below permitted emission levels where this is consistent with energy, environmental, and economic considerations.

Stanton 2 fits into this OUC policy by its duplication of the highly successful and reliable Stanton 1. However, even with this duplication, Stanton 2 will maintain an environmentally responsible posture by the application of advanced but proven control technologies to yield emission rates well below those in the Stanton 1 permit.

Stanton 1 and 2 are both designed as baseload units with load following capabilities. Together they make up 45.1 percent of OUC's generation capacity. Stanton 1 has both design capabilities and permit limitations which allow fuel flexibility. Stanton 2 is being designed with the same design capabilities. Therefore, fuel flexibility in the Stanton 2 permit limitations is important to OUC's successful philosophy.

Another factor demonstrating the unique and special nature of Stanton 2 is the location of the Stanton Energy Center away from all air quality sensitive areas (PSD Class I, nonattainment, and other major increment consumers).

3.4.1 Air Emission Types and Sources

The types and sources of air emissions are the same as previously noted in the Orlando Utilities Commission's Stanton Energy Center Unit 1 Site Certification Application, Subsection 3.7.1.

3.4.2 Air Emission Controls

3.4.2.1 Fugitive Dust. All fugitive dust controls are the same as previously noted in the Orlando Utilities Commission's Stanton Energy Center Unit 1 Site Certification Application, Subsection 3.7.2.1.

3.4.2.2 Nitrogen Oxides. In the combustion process, nitrogen oxides (NO_x) are formed in the high temperature regions of the boiler in and around the flame zone by oxidation of both atmospheric nitrogen and nitrogen in the fuel. Formation of NO_x can be reduced by lowering peak combustion temperatures and by limiting the amount of excess air available to the fuel.

Nitrogen oxides emissions will be controlled by using low NO_x burners and other features designed to limit NO_x formation during combustion. These design features will include the following.

- Compartmented wind box (improved combustion control).
- Large furnace and widely spaced burners (reduced temperatures).
- Overfire air distribution at the burners.
- Staged combustion.
- Modified coal pulverizers for a finer grind.

The large furnace and widely spaced burners increase the burner firing zone absorption area and decrease peak combustion temperatures, thus minimzing NO_x formation.

The steam generator will be designed (and guaranteed by the steam generator manufacturer) to maintain nitrogen oxides emissions to 0.32 lb NO_x per million Btu of heat input (lb/MBtu). This emission compares to a Stanton 1 emission limit of 0.60 lb/MBtu.

3.4.2.3 Particulate. Particulate emissions will be limited through the use of an electrostatic precipitator. The electrostatic precipitator will be located directly downstream of the steam generator air heater. The design of the precipitator is based on meeting a particulate emission limit of 0.02 lb/MBtu when burning the bituminous coal as listed in Table 3.3-1. This emission compares to a Stanton 1 emission limit of 0.03 lb/MBtu. The precipitator design will also include margins to help assure that the emission standards will be met under off-design operating conditions.

The design conditions are essentially the same as previously noted in the Stanton 1 Site Certification Application, Subsection 3.7.2.3.

3.4.2.4 Sulfur Dioxide. The flue gas desulfurization (FGD) system will consist of a multi-module wet limestone spray tower scrubber located downstream of the induced draft fans. The system will have three 50 percent capacity modules with a bypass system. The FGD system will be designed to limit sulfur dioxide emissions to 0.32 lb/MBtu on a 30-day rolling average basis. The proposed 3-hour and 24-hour emission limits are 0.85 lb/MBtu and 0.67 lb/MBtu, respectively. These emissions compare to a Stanton 1 2-hour emission limit of 1.2 lb/MBtu and a 3-hour emission limit of 1.14 lb/MBtu. The scrubber design will also include margins to assure that the emission standards will be met under off-design operating conditions.

The design conditions for the scrubber are essentially the same as previously noted in the Stanton 1 Site Certification Application, Subsection 3.7.2.4.

3.4.3 Best Available Control Technology Analysis

The 1977 Clean Air Act Amendments establish revised conditions for the approval of preconstruction permit applications under the Prevention of Significant Deterioration (PSD) program. One of these requirements is that the best available control technology (BACT) be installed for all pollutants regulated under the Act. Under the revised Act, BACT determinations must be made on a case-by-case basis considering technical, economic, energy, and environmental impacts for various BACT alternatives (rather than automatically applying a specific Federal New Source Performance Standard). To bring consistency to the BACT process, the EPA has authorized development of guidance documents on the use of a "top-down" approach to BACT determinations.* This BACT analysis is based on draft guidance documents issued by the EPA in March 1990.

The first step in a top-down BACT analysis is to determine, for the pollutant in question, the most stringent control alternative available for a similar source or source category (lowest achievable emission rate [LAER] technology). If it can be shown that this level of control is infeasible on the basis of technical, economic, energy, and environmental impacts for the source in question, then the next most stringent level of control is identified and similarly evaluated. This

*US EPA memorandum from J. C. Potter (Assistant Administrator for Air and Radiation) to Regional Administrators, December 1, 1987.

process continues until the BACT level under consideration cannot be eliminated by any technical, economic, energy, or environmental consideration.

This analysis supports the selection of BACT for the OUC Stanton 2 project (440 MW net) regarding the control of particulate, sulfur dioxide, nitrogen oxides, carbon monoxide, volatile organic compounds (VOC), lead, and applicable noncriteria pollutant emissions.

3.4.3.1 Basis of Analysis. The following is a summary of the requirements and assumptions on which this BACT evaluation is based.

- Federal and state ambient air quality standards, emission limitations, significant deterioration increments, solid waste standards, and the requirements of other applicable regulations will be met.
- Federal New Source Performance Standards (NSPS) establish limiting criteria for pollutant emissions from the Stanton 2 project.
- The Stanton Energy Center project is intended to be a baseloaded facility with load following capabilities. With consideration of the large relative portion of Orlando Utilities Commission's generating capacity represented by the Stanton Energy Center, the operating reliability of the air quality control system (AQCS) cannot limit overall unit reliability. Therefore, this reliability consideration may preclude the use of innovative or developmental control technologies.
- The Stanton Energy Center is located in a Class II area which is designated as attainment for all applicable PSD pollutants. In addition, the Stanton Energy Center is not located adjacent to (within its zone of influence) any nonattainment areas.
- The BACT analysis is based on the economic criteria and the coal quality data listed in Tables 3.4-1 and 3.4-2, respectively.
- Costs for OUC Stanton 2 electrostatic precipitator (particulate emissions control) and wet limestone scrubber (flue gas desulfurization) systems reflect virtual duplication of Stanton 1 systems. As such, the costs for these systems presented in the BACT analysis are lower than for a "green field" installation at a new station because of reduced engineering costs. Since air quality control alternatives to these technologies do not have this advantage, their costs will be estimated assuming a "green field" facility.

3.4.3.2 Particulate Emissions Control. The objective of this analysis is to determine BACT for particulate removal alternatives for the Stanton 2 project. This analysis evaluates BACT for both total particulate and fine particulate (PM_{10} emissions).

Additional Requirements and Assumptions.

- Federal New Source Performance Standards limit particulate emissions to 0.03 lb/MBtu, and opacity to a maximum of 20 percent.
- The particulate removal system is designed to meet the 24-hour PM_{10} ambient standard of 150 micrograms per cubic meter, not to be exceeded more than once per year, and the PM_{10} annual primary ambient standard of 50 micrograms per cubic meter.
- A review of information contained in the BACT/LAER Clearinghouse (1985 and 1990 editions) indicates that the most stringent particulate emission limit issued to date is a requirement of 0.012 lb/MBtu for a proposed California coal fired project using a fabric filter.

Particulate Removal Methods. Two particulate removal systems have demonstrated removal efficiencies on pulverized coal fired boilers: electrostatic precipitators and fabric filters.

Operating experience obtained with fabric filters during the last decade has indicated that these devices are extremely effective particulate removal devices. Fabric filters have been the technology of choice for a number of recent BACT and lowest achievable emission rate (LAER) determinations. Fabric filters use fabric bags as filters to collect particulate. The particulate laden flue gas enters a fabric filter compartment and passes through collected particulate and filter bags. The collected particulate forms a cake on the bag which greatly enhances the bag's filtering efficiency. Filter bags can be cleaned by any one of three methods: reverse gas, shake-deflate (reduced reverse gas flow with gentle mechanical shaking of the bag), or pulse jet. Dislodged particulate collects in hoppers beneath the bags for subsequent removal by the ash handling system.

In general, pulse jet fabric filters offer cost savings, compared to reverse gas and shake-deflate fabric filters, on units sized to treat less than 300,000 to 500,000 acfm of flue gas. In addition, it is not expected that pulse jet fabric filters will be any more effective than reverse gas and shake-deflate fabric filters since these devices generally operate at higher cloth velocities (air-to-cloth ratio). Therefore, on the basis of relative economics for a facility the size of Stanton 2 (1.6 million acfm) pulse jet fabric filters will not be considered for use.

With proper design, either reverse gas or shake-deflate fabric filters are capable of meeting a particulate emission requirement of 0.02 lb/MBtu. With a number of design considerations (described subsequently) either of these fabric filter alternatives is capable of meeting a LAER emission requirement of 0.012 lb/MBtu. Since costs are very similar for these two alternatives, a reverse gas fabric filter will be evaluated as the base case fabric filter.

Electrostatic precipitators are the most widely used particulate removal devices for coal fired power plants. Electrostatic precipitators remove particulate matter from the flue gas stream by charging fly ash particulates with very high dc voltage and subsequently attracting these particles to oppositely charged collecting plates. A layer of collected particulate forms on the collecting plates (electrodes) and is removed periodically by rapping the electrodes. The collected particulate drops into hoppers below the precipitator and is periodically removed by the fly ash handling system.

Although more difficult to properly design, precipitators can be equally effective as fabric filters at limiting particulate emissions. However, at lower particulate emission limits (i.e., 0.012 lb/MBtu), design considerations become more difficult and the relative economics for a precipitator become prohibitive. For the types of coal under consideration, it is expected that a precipitator could compete effectively with fabric filters down to an emission limit of 0.02 lb/MBtu.

The following are the alternative particulate control technologies evaluated consistent with a top-down approach.

- LAER Alternative--Reverse gas fabric filter designed to achieve an emission rate of 0.012 lb/MBtu.
- BACT Alternative 1--Electrostatic precipitator designed to achieve an emission rate of 0.02 lb/MBtu.
- BACT Alternative 2--Reverse gas fabric filter designed to achieve an emission rate of 0.02 lb/MBtu.

Economic Evaluation of Particulate Removal Alternatives.

Technical Design Criteria. Fabric filter design criteria are presented in Table 3.4-3 and electrostatic precipitator design criteria are presented in Table 3.4-4. Design criteria for the purpose of this analysis are developed for two emission requirements: 0.012, and 0.02 lb/MBtu. These design criteria are presented for the purpose of establishing the capital and operating costs for the economic comparison of the particulate removal alternatives.

The physical size of an electrostatic precipitator is determined by the particulate and flue gas properties, gas flow, and the required collection efficiency. The most significant particulate property affecting precipitator design is fly ash resistivity, which varies with the moisture content, the chemical composition, and the temperature of the fly ash and flue gas.

As emission limits are lowered, the specific collecting area, total collecting area, and the total number of transformer/rectifiers will increase in electrostatic precipitator designs. The maintenance will also increase as emission limits are lowered. The precipitator electrode alignment, efficient rapping of electrodes, and the electrical stability of the transformer/rectifiers must also be maintained on a more regular basis to meet lower emission requirements. Considering the wide range of coal characteristics anticipated for the plant, a precipitator sized to meet an outlet emission limit of 0.02 lb/MBtu would require a relatively high specific collection area of 743 square feet per 1,000 acfm of flue gas.

Fabric filters are sized primarily on the basis of flue gas flow rate and the design cloth velocity (acfm of flue gas per square foot of cloth area or ft/min). A net cloth velocity of 2.3 ft/min (two compartments out of service, one for cleaning, one for maintenance) is typical for reverse gas fabric filters used to meet an emission requirement of 0.02 lb/MBtu. The selection of a filter medium (cloth) is also important in meeting a specified emission requirement. Fabric filters designed to meet emission requirements of 0.02 lb/MBtu typically use filter bags made of woven fiberglass with an acid-resistant finish.

Although fabric filters cannot be specifically designed to meet a particulate emission requirement (as compared to electrostatic precipitators), it is possible to minimize emissions if certain design changes and quality control measures are taken. Therefore, as fabric filter outlet emission requirements are lowered, certain real capital cost additions can be identified.

A significant amount of the particulate that escapes from a fabric filter results from construction deficiencies. Faulty welds attaching the tubesheet to the walls of the compartment, or thimbles to the tubesheet, allow leakage. Flue gas leaks increase emissions significantly. In addition, inproper attachment of the bags to the tubesheet can allow flue gas to slip from beneath the cuff of the bag. Therefore, as emission requirements are tightened, quality control efforts must be increased to ensure gastight construction and tight tolerances between thimble and bag cuffs.

In addition to fabric filter construction quality control, operation and maintenance procedures must be rigorous in order to meet stringent emissions requirements. Fabric filter bag life can become a significant parameter that directly affects a facility's ability to comply with these emission requirements. Typically, for a unit that operates to meet a 0.02 lb/MBtu emission requirement, bag life ranges from three to five years. As a bag ages, fabric fibers may become abraded and brittle. Therefore, as a bag goes through numerous cleaning cycles, the clearance between woven fabric fibers tends to increase, causing increased particulate penetration through the bags. Accordingly, as particulate emission requirements are reduced, bag changes are required more frequently.

As previously mentioned, reverse gas fabric filters typically use filter bags constructed of woven fiberglass with an acid resistant finish. On applications with low emission requirements (less than 0.02 lb/MBtu), the penetration of particulate from a typical fiberglass bag may become significant. Woven fiberglass bags laminated with a Gore-tex membrane have, in a limited number of applications, minimized particulate bleed-through (penetration) relative to conventional woven fiberglass bags. Therefore, as emission requirements become more strict, the contingency for changeout to Gore-tex filter bags increases.

To ensure compliance with an emission requirement of 0.012 lb/MBtu throughout the life of the plant, it is recommended that design cloth velocities be reduced (increasing the amount of cloth area in the fabric filter). Cloth velocity is a measurement of volumetric gas flow (acfm) per square foot of cloth area. A lower cloth velocity lowers the drag coefficient through the cake built up on the filter bags. A lower coefficient of drag minimizes particulate penetration through filter bags. A net cloth velocity of 2.1 ft/min (two compartments out of service, one for cleaning, one for maintenance) is recommended to comply with an emission requirement of 0.012 lb/MBtu.

A rigorous quality control program must be adhered to during construction to meet an outlet particulate emission requirement of 0.012 lb/MBtu. More frequent inspection visits to the fabrication shop and the construction site will be required to identify potential welding and material defects that may enable flue gas to slip by filter bags untreated. In addition to more frequent inspection of materials and welding, die penetrant or hydro testing of all tubesheet welds will be required.

In addition to increased quality control, the manufacturer is likely to add cost to his contract to account for the increased risk of failing guarantee requirements. This risk money would be held in reserve for the possibility of being required to rebag with Gore-tex bags. Therefore, increased risk money is included in the cost of the 0.012 lb/MBtu alternative to cover rebagging of the fabric filters with Gore-tex filter bags.

Capital and Annual Costs. Comparative costs for a fabric filter particulate removal system designed for a 0.012 lb/MBtu and a 0.02 lb/MBtu particulate emission, and an electrostatic precipitator particulate removal system designed for a 0.02 lb/MBtu particulate emission are presented in Table 3.4-5. The costs presented in Table 3.4-5 are total costs for a complete particulate removal system installed downstream of a pulverized coal fired boiler at Stanton 2.

Capital costs are separated into several categories including electrostatic precipitator, fabric filter, waste handling, ductwork, and differential induced draft (ID) fans. Electrostatic precipitators and fabric filter costs include inlet and outlet plenums, poppet dampers (fabric filter only), electrical and control, and foundations and enclosures. Differential ID fan costs account for the additional fan capacity required to overcome draft losses through the particulate removal systems. Waste handling costs include the solids storage silo, solids blowers, piping, and valves. The capital cost includes contingency, escalation, indirects, and allowance for funds used during construction (see Table 3.4-1 for economic evaluation criteria). Capital costs range from $48 million for a precipitator designed for an outlet emission rate of 0.02 lb/MBtu to $58 million for a fabric filter designed to meet a 0.012 lb/MBtu emission limit.

Levelized annual operating costs include maintenance, operating personnel, and energy. Total levelized annual costs are calculated as the sum of the levelized annual operating costs and the levelized annual fixed charges on capital investment. Levelized annual costs range from $8.7 million to $12 million for a precipitator (0.02 lb/MBtu) and fabric filter (0.012 lb/MBtu), respectively.

Other Considerations. Electrostatic precipitators are more effective than fabric filters at limiting the emission of particulate sized less than 10 microns (PM_{10}). Approximately 92 percent of a total particulate emission rate from a fabric filter is of fine particulate, less than 10 microns in size. Alternatively, precipitator PM_{10}

emissions constitute only 67 percent of the total emission rate. This fraction is based on information presented in the EPA's "Compilation of Air Pollutant Emission Factors," AP-42, September 1985. However, to estimate maximum ambient impacts, dispersion modeling of PM_{10} emissions from Stanton 2 is performed assuming that 100 percent of the 0.02 lb/MBtu emission rate consists of particulate less than 10 microns in size.

An additional advantage for electrostatic precipitators is that they do not require time to condition their removal efficiency. A precipitator sized to limit outlet emissions to 0.02 lb/MBtu should be capable of meeting that limit immediately. However, as discussed previously, fabric filters rely on both the filter bag and a residual dust cake to attain optimum filtering efficiency. The ultimate filter medium for the fabric filter is this residual dust cake. Until an adequate residual dust cake is established, it is likely that fabric filter emissions will exceed 0.02 lb/MBtu. The development of this residual dust cake can take anywhere from two to six months, depending on dust cake characteristics. This period of noncompliance is likely to reoccur every three to five years whenever a rebagging occurs.

A disadvantage of an electrostatic precipitator is its energy consumption. As indicated in Tables 3.4-3 and 3.4-4, the precipitator consumes 85 percent more energy than a fabric filter sized to meet the same emission requirement. However, this additional energy requirement represents only 0.2 percent of the total unit power output.

Conclusions. A fabric filter designed to meet a particulate emission limit of 0.012 lb/MBtu has the highest evaluated cost. Total levelized annual costs for this LAER alternative are \$2.9 million and \$2.7 million higher than for an electrostatic precipitator and a fabric filter, respectively, designed to meet a 0.02 lb/MBtu emission limit. These additional costs result in an incremental removal cost in excess of \$19,000 per ton of particulate removed (as compared to the electrostatic precipitator case).

In addition, a precipitator will result in lower PM_{10} emissions and more consistent emissions performance than a fabric filter. However, a precipitator would consume more energy than a fabric filter. This increased energy requirement is equivalent to only 0.2 percent of plant power output.

Therefore, based on economics and environmental considerations, an electrostatic precipitator designed to meet an emission requirement of 0.02 lb/MBtu represents BACT for Stanton 2. This level of control is 33 percent less than the Stanton 1 emission limit of 0.03 lb/MBtu.

3.4.3.3 Sulfur Dioxide Emissions Control. The objective of this section is to determine BACT for sulfur dioxide (SO_2) emission control alternatives for the Stanton 2 project.

Additional Requirements and Assumptions.

- Federal New Source Performance Standards (NSPS), applicable to Stanton 2 when firing the design coal presented in Table 3.4-2 requires the facility to meet a 1.2 lb/MBtu SO_2 emission rate. Compliance with this requirement is determined on a 30-day rolling average basis.
- FGD for pulverized coal (PC) fired boilers will be accomplished by either a wet lime or limestone scrubbing system, or a lime spray dryer system.
- A review of information contained in the BACT/LAER Clearinghouse (1985 and 1990 editions) indicates that the most restrictive SO_2 removal permit requirement issued to date is 96.2 percent for a proposed circulating fluidized bed (CFB) boiler project in California. Stanton 2 is proposed to be a pulverized coal (PC) fired project.
- Fluidized bed boilers are not available in the size necessary for Stanton 2, and therefore will not be considered further.
- A review of information contained in the BACT/LAER Clearinghouse indicates that the most restrictive SO_2 removal permit requirement for a pulverized coal installation is 95 percent for a proposed installation in Nevada. Flue gas desulfurization at this facility will be provided by a wet lime scrubber. Therefore, the LAER alternative for a pulverized coal fired source such as Stanton 2 would be a wet lime scrubber.
- Compliance with an SO_2 removal requirement based on a 30-day rolling average requires that the SO_2 removal system routinely maintain a removal efficiency in excess of the permitted removal requirement. If the FGD system were designed to operate exactly at the required 30-day removal efficiency, any upset in system operation that reduced SO_2 removal would cause 30 days of noncompliance. This requires that typical FGD systems located downstream of a PC boiler be operated at a removal rate at least 3 percent higher than the overall removal requirement to account for periods of system upset.

Flue Gas Desulfurization Methods. A number of post-combustion FGD processes have demonstrated SO_2 removal capabilities for use downstream of a pulverized coal fired boiler. However, wet scrubber and spray dryer systems are the most widely used FGD systems. In addition, these FGD systems are favored because of their simplicity of operation and equivalent removal capabilities compared to relatively complex byproduct recovery FGD systems. In addition, byproduct recovery systems require a market for their end product of sulfur or sulfuric acid. These markets do not exist in Orlando. Therefore, byproduct recovery systems are not a suitable alternative for Stanton 2.

Wet lime or limestone scrubbing and lime spray drying FGD systems have the advantage of using widely available calcium based additives compared to remotely located sodium based additives (almost all active sodium mines are located in Wyoming). Therefore, the cost of sodium delivered to the Stanton Energy Center site (approximately \$200 to \$250 per ton) would be prohibitive compared to the cost of lime or limestone (\$80 and \$8 per ton, respectively). In addition, the use of sodium based additives increases the complexity and cost of waste disposal due to the high solubility of sodium wastes (increased potential for groundwater contamination due to leachate problems). Considering the location of Stanton 2 in Florida, sodium based FGD alternatives are not a feasible additive for use at Stanton 2.

Currently, 118 utility units with a combined capacity of 53,800 MW are in operation with wet scrubbers using either lime or limestone. In addition, 17 utility units with a combined capacity of 10,500 MW are under construction or under contract to use these wet scrubbing technologies. Lime and limestone wet scrubbers represent about 80 percent (MW basis) of the FGD system capacity in operation, under construction, or under contract in the United States.

During the last decade, the lime spray dryer process has been used on a number of new PC boiler installations. This FGD process absorbs SO_2 through the use of a spray absorber dryer module followed by a fabric filter. A benefit of the spray dryer process compared to wet scrubber FGD systems is the dry waste product, resulting in less complicated and less expensive waste disposal.

Both wet scrubbers and spray dryers are capable of very high SO_2 removal efficiencies. Because of the highly alkaline nature of lime, wet lime scrubbers are capable of up to 97 percent SO_2 removal. Considering an adequate control margin of 3 percent (to ensure reliability during process control upsets), a wet lime scrubber should be capable of meeting an outlet emission requirement of 0.24 lb/MBtu (94 percent removal).

The lime spray drying technology is capable of up to 95 percent removal because of the less efficient nature of its SO_2 removal reaction. Accordingly, lime spray dryers should be capable of maintaining compliance with an outlet emission requirement of 0.32 lb/MBtu (92 percent removal). Wet limestone scrubbers are also capable of up to 95 percent SO_2 removal. Considering an adequate 3 percent control margin, wet limestone scrubbers should also be capable of meeting an SO_2 emission requirement of 0.32 lb/MBtu (92 percent removal).

The following are the alternative FGD technologies evaluated consistent with a top-down approach.

- LAER Alternative--Wet lime scrubber designed to achieve an SO_2 emission rate of 0.24 lb/MBtu on a 30-day rolling average basis.
- BACT Alternative 1--Wet limestone scrubber system designed to achieve an SO_2 emission rate of 0.32 lb/MBtu on a 30-day rolling average basis.
- BACT Alternative 2--Lime spray dryer system designed to achieve an SO_2 emission rate of 0.32 lb/MBtu on a 30-day rolling average basis.

Table 3.4-6 lists estimated sulfur dioxide emissions for the various SO_2 removal alternatives when burning the typical coal. Should shorter averaging periods be desired, emission rates should be increased to account for decreased potential for compliance.

Economic Evaluation of FGD Alternatives. To determine relative economics, each FGD alternative is evaluated on a total air quality control system (AQCS) basis. The AQCS includes the following subsystems.

- Additive storage and preparation.
- Flue gas desulfurization.
- Particulate removal.
- Flue gas supply and exhaust.
- Waste storage and conditioning.

Capital costs are based on FGD systems designed to meet SO_2 removal requirements when burning the worst case coal (high sulfur and low heating value). Operating costs are based on FGD systems operated to meet SO_2 removal requirements when burning the typical coal.

Technical Design Criteria.

Wet Lime Scrubber AQCS. Figure 3.4-1 shows the equipment included in a PC boiler/wet lime scrubber AQCS that would be designed for Stanton 2. With this system, flue gas exiting the air heater passes through electrostatic precipitators and is directed by induced draft (ID) fans to the absorber modules (spray towers). The ID fans are located between the electrostatic precipitators and the absorber modules to minimize particulate erosion and water vapor condensation on fan internals.

Wet lime absorber modules serve as the contact zone where the alkaline additive absorbs the SO_2 from the flue gas. Recycle pumps spray the lime slurry counter-current to the direction of the flue gas flow. The resultant reaction products flow downward through the spray tower into the reaction tank while the flue gas flows out of the absorber module and into the stack. Table 3.4-7 lists selected design parameters for wet lime scrubber AQCS.

The scrubber module diameter listed in Table 3.4-7 is based on a flue gas velocity limit through the module of 10 feet per second (fps). At velocities above 10 fps, mist eliminator equipment performance degrades quickly. For consideration of overall plant reliability, system design is based on the use of three 50 percent capacity modules.

The preparation of lime slurry is accomplished by the additive storage and preparation system. With this system, pebble lime is stored in silos to protect it from moisture. Lime from storage silos is hydrated in a slaker/classifier system for feed to the slurry storage tanks (24-hour capacity). Additive from the slurry storage tank is transported to absorber module reaction tanks by additive feed pumps.

To convert the liquid waste to a solid waste product for disposal, blowdown from the absorber module reaction tanks is pumped to a thickener for primary dewatering. The decanted water from the thickener is reused in the reaction tanks and to slurry additional lime, while the underflow from the thickener is pumped to vacuum filters for additional dewatering. Thickened sludge from the vacuum filters is mixed with fly ash to form a product suitable for transport to disposal. Wastes are transported by trucks to an onsite landfill disposal location.

Wet Limestone Scrubber AQCS. Figure 3.4-2 shows the equipment included in a wet limestone scrubber AQCS that would be designed for Stanton 2. With the exception of additive preparation and adipic acid addition, a wet limestone system process flow sheet is very similar to a wet lime scrubber AQCS. Additive preparation differences are due to the low solubility of limestone allowing on-ground bulk storage and requiring ball mills for preparing additive slurry. Adipic acid is required to enhance removal efficiency when higher sulfur coals are burned. The adipic acid tends to buffer slurry pH enhancing liquid phase alkalinity. Table 3.4-8 lists selected design parameters for the wet limestone scrubber AQCS.

Lime Spray Dryer AQCS. Figure 3.4-3 shows the scope of equipment included in a lime spray dryer AQCS. Table 3.4-9 lists the design parameters used to evaluate the lime spray dryer AQCS.

The lime spray dryer AQCS is a two-stage process that removes both sulfur dioxide and particulate from the flue gas through the use of a spray dryer/absorber followed by a fabric filter. The absorber modules serve as the initial contact zone where alkaline additive and SO_2 in the flue gas react to form dry reaction products. The majority of reaction products formed in the spray dryer flow out of the absorber modules and into the fabric filter for removal with the fly ash. The ID fans are located between the fabric filters and the stack to minimize particulate erosion on fan internals.

The absorber modules are sized on the basis of gas flow rate and residence time. Residence times of approximately 10 seconds have proved sufficient to ensure adequate reaction product drying. The atomizers, which disperse the additive slurry, are sized on the basis of additive and tempering water feed necessary to achieve the required SO_2 removal level and outlet gas temperature.

Flue gas temperatures at the fabric filter inlet must be sufficiently high to avoid corrosion in the fabric filter and in other downstream equipment. Low flue gas temperatures can also cause condensation of cementatious fly ash materials on the filter bags, severely affecting bag life and fabric filter operation. Adjustment of the spray dryer module approach temperature (number of degrees that the spray dryer operates above the saturation temperature) determines the spray dryer module outlet gas temperature. The amount of water added to the slurry is adjusted to control the spray dryer module outlet gas temperature. For the same SO_2 removal efficiency, a higher approach temperature results in greater lime consumption. Lime consumption increases as a result of a reduction in the SO_2 removal reaction efficiency at the higher approach temperature. An approach temperature of 40 F results in a fabric filter inlet gas temperature of approximately 165 F. An inlet gas temperature of 165 F is sufficiently high to protect the fabric filter and other downstream equipment.

The preparation of lime for use as additive in a spray dryer AQCS is accomplished by the additive storage and preparation system. With this system, pebble lime is stored in silos to protect it from moisture. Lime from storage silos is hydrated in a slaker/classifier system for feed to the slurry storage tanks (24-hour capacity). Additive from the slurry storage tank is pumped to the additive feed tank.

Since a significant portion of the lime feed does not initially react with the SO_2 in the flue gas stream, a portion of the solids collected in the fabric filter is returned and mixed with fresh lime slurry so that unreacted lime or alkalinity contained in the fly ash can be utilized. The lime and recycled solids are blended in a recycle slurry mix tank and pumped to the additive feed tanks.

The solids collected in the fabric filter, which are not recycled, are collected in the solids storage silo and subsequently transported by trucks to an onsite landfill.

Capital Costs. Table 3.4-10 lists the estimated capital costs for the alternative AQCS when the coal listed in Table 3.4-2 is burned. The table shows the capital costs for a complete SO_2 and particulate removal system. Sulfur dioxide and particulate removal costs are based on the design parameters listed in Tables 3.4-7, 3.4-8, and 3.4-9. Economic criteria used to develop capital costs are listed in Table 3.4-1.

Capital costs for air quality control system alternatives range from \$111 million, for the lime spray dryer AQCS (0.32 lb/MBtu emission), to \$130 million, for the wet limestone AQCS designed for an emission rate of 0.32 lb/MBtu SO_2. The costs in Table 3.4-10 are separated into five categories. The following paragraphs describe the costs included in each of these categories.

Additive storage and preparation. Additive storage and preparation capital costs include all equipment necessary to store and prepare the additive for use in the SO_2 removal process.

A wet ball mill/classifier system is used to obtain slurry of constant properties for use in the wet limestone scrubber modules. The wet lime and lime spray dryer systems use slakers for additive slurry preparation. Slurry is stored in a storage tank designed to hold 24 hours of additive at peak consumption. For the wet limestone scrubber, wet lime scrubber and lime spray dryer AQCS, fresh additive is pumped from the slurry storage tank to the absorber reaction tanks (wet lime and limestone) or the additive feed tank (lime spray dryer).

A recycle system is included for the lime spray dryer to utilize unreacted additive and residual fly ash alkalinity. Solids from the fabric filters are stored in either of two recycle feed bins, each sized for six hours of average recycle feed requirements. Additive from the slurry storage tanks and the recycle mix tanks are combined in the additive feed tanks. The combined lime/recycle material slurry is then pumped to absorber head tanks.

Costs in this category for the alternative AQCS include a reclaim hopper with vibratory grizzly and mechanical conveyors (wet limestone AQCS), portable pneumatic conveyors (wet lime scrubber and lime spray dryer AQCS), additive storage silos, ball mills (limestone slurry), weigh belt feeders, slakers (lime slurry), slurry storage tanks, additive feed system, recycle system (lime spray dryer AQCS), piping, valves, electrical and control equipment, and foundations and enclosures.

Flue gas desulfurization. Flue gas desulfurization capital costs include all equipment necessary for desulfurization of the flue gas with prepared alkaline additive.

Wet limestone and lime scrubber module costs are estimated assuming rubber-lined carbon steel vessels. Reaction tanks are sized for 10 minutes of slurry retention. FGD capital costs for the wet limestone and lime scrubber AQCS include scrubber modules, reaction tanks, recirculation pumps, miscellaneous tanks and pumps, piping, valves, electrical and control equipment, and foundations and enclosures.

The wet limestone scrubber shows a lower than expected capital cost because it is a virtual duplicate of the Stanton 1 wet limestone scrubber, requiring only replicate engineering and equipment drawings.

Absorber module costs for the lime spray dryer AQCS are estimated assuming carbon steel vessels. Costs for the lime spray dryer AQCS include absorber modules, atomizers, foundations and enclosures, piping, valves, and electrical and control equipment.

Particulate removal. Consistent with the results presented in Subsection 3.4.3.2, particulate removal costs for the wet limestone and the wet lime AQCS include costs for an electrostatic precipitator. The electrostatic precipitator is designed for a SCA of 743 ft^2 per 1,000 acfm of gas flow with a flue gas velocity of 3.5 feet per second.

Particulate removal costs for the lime spray dryer AQCS include costs for two 12-compartment, reverse gas cleaned fabric filters. The fabric filters are designed for a maximum net cloth velocity of 2.3 ft/min (one compartment out of service for cleaning and one out of service for maintenance).

In addition to precipitator and fabric filter costs, particulate removal costs include inlet and outlet ductwork, ash handling equipment, foundations and enclosures, and electrical and control equipment.

Flue gas supply and exhaust. The flue gas supply and exhaust capital cost category includes ductwork needed to route the flue gas to the ID fans, absorber modules (if applicable), the particulate removal system, and the stack. This cost also includes incremental ID fan capacity capable of overcoming additional flow resistance created by the flue gas desulfurization and particulate removal systems.

Waste storage and conditioning. Waste storage and conditioning capital costs include all equipment necessary for transportation, separation, storage, and conditioning of wastes in preparation for transportation to the offsite disposal location.

For the wet limestone and wet lime scrubber AQCS, blowdown from the absorber module reaction tanks is pumped to the thickener for primary dewatering. Thickener overflow is reused in the reaction tanks and to slurry additional limestone or lime, depending on the process. The underflow from the thickener is pumped to a surge tank in preparation for secondary dewatering. Secondary dewatering is accomplished by vacuum filters. Thickened sludge from the vacuum filters is mixed with fly ash to form a product suitable for transport to disposal. Costs in this category include a primary and secondary dewatering system, a sludge/fly ash mixing system, conveyors for transport of conditioned waste products and stockout of the waste mixture, piping, valves, electrical and control equipment, and foundations and enclosures.

For the lime spray dryer AQCS, waste solids from absorber modules and fabric filter hoppers are transported and stored in elevated solids storage silos. It is assumed for this analysis that wastes will be conditioned with water. Conditioning with water fixates the waste as water reacts with unused quantities of lime contained in the waste products, thereby controlling fugitive dust. Waste conditioning capital costs for the lime spray dryer AQCS include solids handling equipment, solids storage silos, waste conditioners, piping, valves, electrical and control equipment, and foundations and enclosures.

Levelized Annual Operating Costs. Table 3.4-11 lists the levelized annual operating costs for the air quality control system alternatives. Levelized annual costs reflect the effects of escalation and present worth discounting on future operating cost expenditures. First year operating costs are multiplied by the levelization factor listed in Table 3.4-1 to obtain a levelized annual operating cost.

Levelized annual operating costs listed in Table 3.4-11 range from $26 million for the 0.32 lb/MBtu SO_2 emission rate wet limestone scrubber AQCS, to $44 million for the lime spray dryer AQCS also designed for an emission rate of 0.32 lb/MBtu SO_2.

Operating personnel costs include personnel required for additive preparation, flue gas desulfurization, particulate removal, and waste conditioning operations. Personnel costs, including salary and benefits, are based on a 1991 labor cost of $43,333 per employee year and a 4.75 percent escalation rate. Maintenance personnel costs are included in the maintenance cost described below.

Maintenance costs are estimates of material and labor required to operate alternative AQCS. Maintenance costs are a major contributor to operating costs and vary proportionally with the amount of equipment installed. It is not likely that the maintenance expense shown in Table 3.4-11 would occur during the first few years of unit operation, but is representative of average annual maintenance costs over the life of the plant.

Additive requirements for the AQCS alternatives are determined on the basis of the SO_2 removal requirements and on actual reaction stoichiometrics obtained from operational and experimental data. Additive costs are based on a 1991 limestone cost of $8 per ton, and on a 1991 pebble lime cost of $80 per ton.

Energy costs are also included to account for alternative AQCS auxiliary power requirements. Energy costs are calculated based on operation of AQCS equipment, and the costs associated with operating ID fans to overcome the differential pressure drop caused by the operation of the AQCS.

Annual waste disposal costs are based on the use of a subcontractor to transport and dispose of wastes. Waste solids will be transported by trucks to an onsite landfill.

Total Levelized Annual Costs. In addition to levelized annual operating costs, Table 3.4-11 presents a levelized annual cost summary. The total levelized annual cost allows comparison of alternative AQCS. The total levelized annual cost is calculated as the sum of fixed charges on capital investments and operating costs. Total levelized annual costs range from $36 million for the wet limestone scrubber AQCS designed for an emission rate of 0.32 lb/MBtu SO_2, to $52 million for the lime spray dryer AQCS also designed for an emission rate of 0.32 lb/MBtu SO_2.

Other Considerations. As indicated in the assumptions for this section, it is expected that a 3 percent SO_2 removal control margin between expected and required performance is necessary to ensure compliance during periods of process upset or equipment outages. For contemporary FGD systems, the fundamental element for noncompliance is one of process control. At a target SO_2 removal of 94 to 95 percent for a wet limestone scrubber AQCS (approaching the practical limits of this technology), the distribution of daily efficiencies becomes skewed. Although it would not be unusual for a scrubber targeting 94 percent removal to drift to a 91 percent daily removal rate, it is much less likely that a 97 percent daily removal would occur.

To maintain consistent compliance, the margin between "target" and "30-day average" (compliance) must be large enough to allow for this potential performance shift. Statistical analysis of operating FGD systems correlating performance and reliability have indicated that the appropriate minimum margin is 3 percent to maintain compliance with a 30-day rolling average.

This concern for the Stanton Energy Center is further confirmed by the fact that the plant is designed for zero discharge of plant wastewater. Accordingly, there is a high degree of makeup water quality variability complicating FGD process chemistry (especially with respect to chlorides control). The ability of OUC to achieve or exceed 30-day rolling average removal limitations would be severely compromised by requiring an unduly high compliance level.

Energy Evaluation of Alternatives. The lime spray dryer AQCS has the lowest energy demand of FGD alternatives. At peak demand, this difference represents 1.1 percent and 1.8 percent of total plant power output as compared to the wet lime scrubber AQCS and the wet limestone scrubber AQCS, respectively.

Conclusions. A wet lime scrubber AQCS designed for an emission rate of 0.24 lb/MBtu SO_2 has a total levelized annual cost of $47 million. Levelized annual costs are $10 million higher than a wet limestone scrubber AQCS designed for an emission rate of 0.32 lb/MBtu SO_2. The additional costs for a wet lime AQCS result in an incremental removal cost of $6,900 per ton of SO_2 removed, to go from an emission rate of 0.32 lb/MBtu SO_2 to 0.24 lb/MBtu SO_2. The lime spray dryer AQCS has the highest levelized annual cost of $52 million.

On the basis of economics and environmental considerations, a wet limestone scrubber AQCS designed for an emission rate of 0.32 lb/MBtu SO_2 on a 30-day rolling average is considered to represent BACT for use at Stanton 2. In addition, to accommodate process control and equipment reliability problems as well as provide for some fuel quality flexibility, it is proposed that 3-hour and 24-hour emission requirements of 0.85 lb SO_2/MBtu and 0.67 lb SO_2/MBtu, respectively, be allowed.

3.4.3.4 Nitrogen Oxides, Carbon Monoxide, and VOC Emissions Control. The objective of this analysis is to determine BACT for nitrogen oxides (NO_x), carbon monoxide (CO), and volatile organic compound (VOC) emissions. Because of the mutually dependent formation characteristics of NO_x, CO, and VOC (expressed as total nonmethane hydrocarbons) emissions, it is necessary to consider BACT concurrently for these emissions.

Additional Requirements and Assumptions.

- Nitrogen oxide emissions are limited by New Source Performance Standards to 0.60 lb/MBtu of heat input to the boiler for bituminous coal. The coal listed in Table 3.4-2 is a bituminous coal.
- There are no coal fired boiler NSPS limiting the emission of CO or VOC.
- A review of information contained in the BACT/LAER Clearinghouse (1985 and 1990 editions) indicates that the most stringent NO_x emission limit issued to date is 0.043 lb/MBtu for a proposed project located in California. The installation will use a circulating fluidized bed boiler with a selective noncatalytic reduction system.
- Fluidized bed boilers are not available in the size necessary for Stanton 2, and, therefore, will not be considered further.
- A review of information contained in the BACT/LAER Clearinghouse indicates that the most restrictive NO_x emission requirement for a pulverized coal installation is 0.44 lb/MBtu for a plant in Arizona. NO_x emissions from this facility are limited through the use of combustion controls.
- The most stringent CO emission limit issued to date is a requirement of 0.014 lb/MBtu for a project operating in Florida. This unit limits CO emissions through the use of combustion controls consistent with meeting a NSPS NO_x emission limit of 0.60 lb/MBtu.

- The most stringent VOC emission limit is a requirement of 0.003 lb/MBtu for a project operating in Virginia. This unit limits VOC emissions through the use of combustion controls consistent with meeting a NSPS NO_x emission of 0.60 lb/MBtu.
- Since NO_x emissions are the dominant pollutant with regard to total impact, this analysis will be based on optimizing combustion controlled emissions to minimize NO_x emissions.

Emission Control Alternatives. Nitrogen oxides and CO/VOC emission controls are divided into two categories: in-furnace formation control and post-combustion emission reduction. In-furnace combustion control processes reduce the quantity of NO_x and CO/VOC formed during the combustion process. Post-combustion NO_x controls reduce a portion of the NO_x exiting the boiler to nitrogen and water. Post-combustion CO/VOC emission controls oxidize a portion of these pollutants to carbon dioxide and water.

In-Furnace Combustion Control. Nitrogen oxides are formed by the oxidation of nitrogen contained in the fuel (fuel NO_x) and in the combustion air (thermal NO_x). Nitrogen oxide emissions are limited by lowering combustion temperatures, minimizing excess combustion air, and staging combustion. Carbon monoxide and volatile organic compounds are formed by incomplete combustion of coal. Increasing combustion temperatures, increasing excess air, and better fuel/air mixing during combustion minimize CO and VOC emissions while increasing NO_x emissions.

The commercial installation of low NO_x burners over the last several years represents an advance in the control of NO_x emissions from pulverized coal fired boilers. Low NO_x burners reduce NO_x formation in the boiler by maintaining a reducing atmosphere at the coal nozzle and diverting additional combustion air (to complete combustion) to secondary air registers. This staged combustion primarily inhibits the formation of fuel NO_x.

The NO_x emission rate of 0.32 lb/MBtu, based on current pulverized coal combustion controls utilizing advanced design burners and associated peripherals, represents over a 45 percent decrease below Stanton 1 emission requirements of 0.60 lb/MBtu. Consistent with the use of these combustion controls minimizing NO_x emissions, carbon monoxide and volatile organic compound emissions are expected to be 0.15 lb/MBtu and 0.015 lb/MBtu, respectively. Further decreases in CO and VOC emissions will result in NO_x emission increases.

Post-Combustion Emissions Reduction Systems. Nitrogen oxide emissions from a coal fired boiler can be reduced by use of either a selective catalytic reduction (SCR) or a selective noncatalytic reduction system (SNCR). These systems are the only potentially viable post-combustion NO_x emission reduction technologies that can be considered for installation on pulverized coal boilers.

SCR Systems. In an SCR system, ammonia is injected into the flue gas stream just upstream of a catalytic reactor. The ammonia molecules in the presence of the catalyst dissociate reducing a significant portion of the NO_x into nitrogen and water. SCR systems may potentially reduce NO_x emissions by as much as 70 to 90 percent.

The ammonia (either aqueous or anhydrous) is received and stored as a liquid. The ammonia is vaporized and subsequently injected into the flue gas by either compressed air or steam carrier. Injection of the ammonia must occur at temperatures between 600 and 800 F. Therefore, the system is logically located between the economizer outlet and the air heater inlet. The SCR catalyst is housed in a reactor vessel which is separate from the boiler. An economizer bypass may be required to maintain the reactor temperature during low load operation. This will reduce boiler efficiency at lower loads.

Ammonia is a hazardous material. Therefore, ammonia must be handled and stored with extreme care. Working on and around ammonia equipment will cause operational personnel to be less productive and functional than under normal working conditions.

SCR systems have been used predominately on Japanese and West German gas, oil, and coal fired boilers. Coal fired boilers that have utilized SCR have all burned low sulfur (less than 1.3 percent) coals with relatively low ash contents. There are no coal fired boilers using SCR systems in the United States.

In addition to fuel quality and safety concerns, SCR systems will experience problems with unreacted ammonia slippage. SCR systems generally have ammonia slip rates of between 5 and 10 ppm. Unreacted ammonia and sulfur trioxide can react to form ammonia bisulfate and ammonia sulfate salts. These sticky substances can severely affect downstream equipment. Air heaters could suffer pluggage problems and fabric filters could experience bag blinding if these substances were present in the flue gas. In addition, fly ash tends to erode the catalyst, leading to premature failures, and a number of trace metals have detrimental effects on catalyst reactivity. In general, United States coals contain higher levels of sulfur, ash, and trace metals than the coals used in Japan and West Germany. The sulfur, ash, and trace metal contents of United States fuels could significantly affect the performance and operating reliability of an SCR system. It has been estimated that SCR systems burning United States coal could experience a catalyst life of one year or shorter. Catalyst costs account for over 60 percent of the initial capital cost of an SCR system.

In summary, based on the eastern United States coals being considered for use at Stanton 2, and based on the complete lack of SCR experience with these coals, this analysis will not consider the use of SCR.

SNCR Systems. Selective noncatalytic NO_x reduction systems rely on the appropriate reagent injection temperature and good reagent/gas mixing rather than a catalyst to achieve NO_x reductions. SNCR systems can use either ammonia (Thermal $DeNO_x$) or urea (NO_xOUT) as reagents.

Ammonia for a Thermal $DeNO_x$ system is stored as a liquid. Subsequently, the ammonia is vaporized and injected into the flue gas using either compressed air or steam as a carrier. The ammonia then reacts with the NO_x to form nitrogen and water. Reagents for SNCR systems are injected in the backpass (convective portion) of the boiler.

Urea for a NO_xOUT system is stored as a 50 percent solution in water. This solution is atomized at the injection point to optimize mixing. In the flue gas, the urea molecule dissociates to form two molecules of NH_3 (ammonia). The NH_3 reacts with NO_x to form nitrogen and water. Urea would be injected at a similar location to an ammonia based SNCR system.

The optimum temperature range for injection of ammonia or urea is 1,550 to 1,900 F. The NO_x reduction efficiency of the SNCR system decreases rapidly at temperatures outside this range. Operation below this temperature window results in excessive ammonia emissions. Operation above this temperature window results in increased NO_x emissions. A pulverized coal boiler operates at a temperature of between 2,500 and 3,000 F. Therefore, the optimum temperature window in a pulverized coal fired boiler occurs somewhere in the backpass of the boiler. To further complicate matters, this temperature location will change as a function of unit load. In addition, residence times in this temperature window are limited, further detracting from optimum performance.

SNCR systems are a less efficient NO_x reduction system than SCR systems. In general, SNCR systems on pulverized coal fired boilers will only be capable of between 40 and 50 percent NO_x reduction. The major site specific considerations that limit the NO_x emission reduction potential of SNCR systems include boiler temperature profile, the coal's sulfur and chlorine contents, and the geometry of the boiler (affecting effective additive distribution).

Both SNCR processes require more than twice the theoretical amount of reagent to achieve these NO_x reductions. Accordingly, SNCR systems produce significant quantities of unreacted ammonia. A portion of this ammonia decomposes into nitrogen and water. However, any ammonia that does not decompose exits the system as ammonia slip. SNCR systems installed on pulverized coal boilers would have ammonia slips of between 10 and 50 ppm.

Ammonia slip will either exit the system through the stack or condense onto the fly ash collected in the electrostatic precipitator. Unless stack emissions are in excess of 50 ppm, it is not likely that a noticeable odor will occur. However, fly ash will absorb some of the ammonia from the flue gas stream and will tend to be odorous. Accordingly, if an SNCR system is used, commercial sale of fly ash will not be possible because of the ammonia contamination. Stanton 1 has historically been capable of selling all ash production for use in the concrete industry. It was expected that Stanton 2 would be similarily capable. However, should an SNCR system be required, the potential for fly ash sales from Stanton 2 would be eliminated due to ammonia contamination. As a result, this contaminated fly ash must be disposed of in an onsite landfill, incurring additional cost.

Close control of SNCR system ammonia or urea injection in a pulverized coal fired boiler is difficult. Tube spacings, temperature profiles, and the physical size of a pulverized coal fired boiler such as Stanton 2 greatly complicate additive injection. These problems are likely to result in additional ammonia slip emissions or diminished performance. In addition, reliable continuous ammonia emission monitors have proved to be highly unreliable. Without ammonia monitors, it is not possible to optimize reagent injection through feedback control by ammonia slip measurements. This also results in higher ammonia slip emissions.

Similar to SCR systems, unreacted ammonia and sulfur trioxide can react to form ammonia bisulfate and ammonia sulfate salts. Based on the SNCR injection location and higher levels of ammonia slip, there is a higher potential to foul equipment more severely in an SNCR system than in an SCR system. In addition, the formation of ammonia salts will increase the fine particulate (less than

10 microns) loading to the fabric filter. Therefore, if an SNCR system is used, it is likely that PM_{10} emissions will increase.

An additional technical concern with the use of an SNCR system is the creation of an ammonia chloride plume (typically brown in color). It has been documented that for fuels with significant chloride content (greater than approximately 0.05 percent), ammonia slips of 5 ppm and higher will result in a continuous ammonia chloride plume. The ammonia chlorides do not increase opacities measured by the continuous emissions monitor, but would nonetheless be visible to the human eye. This would be a significant negative aesthetic impact for use of an SNCR system. It is likely that ammonia slips will exceed 5 ppm unless NO_x reduction efficiencies are maintained at 30 percent or less for Stanton 2.

As previously described for SCR systems, ammonia is a hazardous material. Accordingly, this material for a Thermal $DeNO_x$ type SNCR must be handled and stored with extreme care. Working on and around ammonia equipment will cause operational personnel to be less productive and functional than under normal working conditions.

An additional disadvantage of a NO_xOUT type SNCR system is higher carbon monoxide emissions. Carbon molecules released from the urea molecule during decomposition to ammonia can react to form carbon monoxide. Equipment supplier estimates indicate that CO emissions could increase by as much as 10 to 20 percent.

Despite the potential problems, a review of information contained in the BACT/LAER Clearinghouse (1985 and 1990 editions) provided a number of California projects that were required to use SNCR systems. However, all of these projects are smaller fluidized bed boilers. Fluidized bed boilers provide a more optimum reaction environment for NO_x reduction operations. In addition, because of nonattainment status and California's unique air quality problems, these limitations are more representative of LAER determinations. All of the facilities operating with SNCR burn coal with very low sulfur and chloride contents (approximately 0.5 and 0.03 percent or less, respectively, in fluidized bed boilers). Fluidized bed boilers provide an optimum environment for the use of SNCR systems because of prolonged residence times at the appropriate reaction temperature.

With the relatively high sulfur and chlorine content of the coal available for use at Stanton 2, it is recommended that SNCR systems designed for 40 percent (outlet emission of 0.19 lb/MBtu) and 30 percent (outlet emission of 0.22 lb/MBtu) NO_x reduction be evaluated for use. An SNCR system designed for higher NO_x reductions would have higher ammonia slip emissions and a higher probability of an ammonia chloride plume, and would run a significantly higher risk of lower unit reliability as a result of the possibility of equipment fouling from ammonia salts.

CO and VOC Emissions Reduction Systems. Lower CO and VOC emissions are possible if boiler temperatures are increased. However, NO_x formation would increase. Therefore, consistent with the approach of evaluating BACT for CO and VOC emissions based on BACT for NO_x, increasing combustion temperatures to limit CO and VOC emissions is not an option.

A catalytic CO and VOC emissions reduction method is available for use on the exhaust from combustion turbines and petroleum refining operations. The process oxidizes CO, resulting in the emission of carbon dioxide and water. The process is a straight catalytic oxidation/reduction reaction requiring no additives. However, the platinum coated catalyst is extremely expensive.

This process has never been applied to a coal fired power plant. The catalytic reaction is effective at a temperature of approximately 700 F. In pulverized coal boilers, a temperature of 700 F is available just upstream of the air heater. However, because of the potential for erosion and pluggage of the platinum catalyst by abrasive combustion products, and poisoning of the catalyst by trace metals in the fly ash, this process is unsuited to coal fired applications, and is, therefore, considered not technically feasible for Stanton 2.

Economic Evaluation of Alternatives. Table 3.4-12 lists the estimated emission of NO_x, CO, VOC, and ammonia for the NO_x emission control alternatives. Table 3.4-13 lists the estimated total capital and annual cost for installing an SNCR NO_x emission reduction system on Stanton 2. The table shows all costs for a complete ammonia based SNCR system. It is expected that costs for a urea based system would be approximately equivalent to those for an ammonia based system. The costs listed are incremental costs assuming a base case of combustion controls for NO_x emission control. Economic criteria used to develop these costs are listed in Table 3.4-1.

The capital costs include ammonia receiving, storage, and injection equipment; technology licensing fees; and balance-of-plant costs. Balance-of-plant costs include foundations, dikes, structural steel, piping, wash water system for air heater, and electrical and control equipment. In addition, because of safety considerations regarding the use and storage of ammonia, fire protection and other safety equipment costs were included.

Incremental levelized annual operating costs for an SNCR system are also presented in Table 3.4-13. Operating costs include operating personnel, maintenance, ammonia additive, electric energy, and demand costs, as well as loss of fly ash sales and fly ash landfill costs.

Installing an SNCR system would add approximately $14 million and $11 million to the capital cost of Stanton 2 for 40 percent and 30 percent NO_x reduction systems, respectively. The total levelized annual cost for an SNCR system would be approximately $6.5 million and $5.5 million for 40 percent and 30 percent NO_x reduction systems, respectively. These costs result in an incremental NO_x reduction cost of $2,700 per ton (40 percent reduction--2,403 tons reduced per year) and $3,100 per ton (30 percent reduction--1,802 tons reduced per year) as compared to use of combustion controls to achieve an NO_x emission of 0.32 lb/MBtu.

Energy Evaluation of Alternatives. An SNCR system consumes both electrical and steam energy. An ammonia based SNCR system would require approximately 2,200 kW of electrical energy. This represents approximately 0.5 percent of total plant power output.

Environmental Evaluation of Alternatives. Areas surrounding Stanton 2 are classified as attainment areas for NO_x, CO, and VOC. Modeling analyses based on NO_x and VOC emission rates of 0.32 lb/MBtu and 0.012 lb/MBtu, respectively, indicate that ambient impacts of emissions from Stanton 2 were below impacts predicted in the original Stanton 1 Site Certification Application.

Operation of a selective noncatalytic reduction system to meet an NO_x emission limit of 0.19 lb/MBtu (40 percent reduction) will likely result in excessive ammonia slip emissions of between 20 and 50 ppm. Accordingly, this ammonia slip in conjunction with chloride emissions will result in the formation of a visible ammonia chloride plume. An SNCR system operated to limit NO_x emissions to 0.22 lb/MBtu (30 percent reduction) will likely have ammonia slip emissions below 5 to 10 ppm. Operation of an SNCR system to meet this NO_x emission is less likely to result in any visible ammonia chloride emissions from the plant.

Conclusions. Advances in the control of NO_x from pulverized coal boilers enable the project to lower anticipated NO_x emissions from the Stanton 1 emission limit of 0.6 lb/MBtu to 0.32 lb/MBtu. This level is more than 45 percent lower than the Stanton 1 emission limit of 0.60 lb/MBtu and 27 percent lower than the lowest NO_x emission limit on record (BACT/LAER Clearinghouse) for a pulverized coal boiler. Consistent with this NO_x emission, carbon monoxide and VOC emissions are expected to be 0.15 and 0.015 lb/MBtu, respectively.

Selective catalytic reduction systems are insufficiently developed for use on pulverized coal fired boilers burning United States coal. Selective noncatalytic reduction systems could possibly be used on Stanton 2. However, SNCR systems are not demonstrated on pulverized coal boilers burning coals with sulfur contents greater than 0.5 percent. A higher coal sulfur content results in larger amounts of ammonia bisulfate and ammonia sulfate being produced when an SNCR system is used. It is likely that these relatively sticky compounds will deposit on downstream equipment detrimentally affecting unit reliability. Ammonia salts that do exit the stack will largely consist of particles less than 10 microns.

Reagent injection control for SNCR systems is not precise. Therefore, ammonia slip emissions of between 10 ppm (27 lb/h) and 50 ppm (135 lb/h) can be expected. Fly ash will absorb some of the ammonia from the flue gas stream and will tend to be odorous. Like Stanton 1, it was anticipated that fly ash from Stanton 2 would be sold. Use of an SNCR system on Stanton 2 would eliminate the environmentally sound practic of selling fly ash for reuse in the concrete industry.

In addition, use of an ammonia based system will result in handling and storage of a hazardous material on the Stanton 2 site. Alternatively, use of a urea based system will result in increased CO emissions.

Use of an SNCR system (designed to achieve 40 percent NO_x reduction) at Stanton Unit 2 is estimated to cost \$6.5 million annually. This results in an incremental NO_x reduction cost of \$2,700 per ton. Ammonia slip emissions from this system of 20 ppm are likely to result in a visible ammonia chloride plume. This is a significant concern considering the location of the Stanton Energy Center in Orlando. NO_x reduction must be lowered to eliminate the potential for an ammonia chloride plume. NO_x reduction must be decreased to 30 percent.

This results in an annual cost of approximately \$5.5 million (incremental reduction cost of \$3,100 per ton).

In addition to the costs identified in Table 3.4-13, a requirement for an SNCR system on Stanton 2 would limit the operating reliability of the unit. Use of this system would increase the mechanical complexity of the plant as well as impacting downstream equipment operability and reliability. This decreased plant reliability could result in significant additional cost impacts. These cost impacts are not reflected in this analysis.

The preceding discussion strongly supports that on the basis of technical, economic, energy, and environmental considerations, combustion controls designed to meet an NO_x emission requirement of 0.32 lb/MBtu represents BACT for Stanton 2 and SNCR should not be applied to this installation.

3.4.3.5 Lead and Noncriteria Pollutant Emissions Control. An additional requirement of BACT analyses is the evaluation of control technologies for lead, Prevention of Significant Deterioration (PSD) noncriteria pollutants, and other hazardous air pollutants that may occur. Coal contains a number of trace elements which may be volatilized during combustion. In addition, a number of other organic emissions can also occur as a byproduct of combustion. The EPA has identified a list of potential hazardous air pollutants from coal fired combustion ("Control Technologies for Hazardous Air Pollutants," EPA/625/6-86/014, September 1986). This section discusses the control of these emissions from Stanton 2.

Coal does not contain asbestos or vinyl chloride, and none is formed during combustion. Therefore, asbestos and vinyl chloride emissions do not require further consideration since annual emissions will be less than PSD significance levels.

Hydrogen sulfide and reduced sulfur compounds form in a reducing atmosphere. Combustion in a pulverized coal fired boiler occurs in an oxidizing atmosphere. Therefore, emissions of these compounds will be less than PSD significance levels.

An additional benefit of particulate removal and flue gas desulfurization air quality control efforts is the removal of a number of the hazardous air pollutants from the flue gas stream. Removal occurs as a result of either condensation of trace emissions from the flue gas onto fly ash particles, or absorption by the scrubbing liquor. Control of organic emissions occur as a result of complete combustion in the boiler (consistent with the control of carbon monoxide and volatile organic compound emissions). Table 3.4-14 lists estimated emissions for lead, PSD noncriteria pollutants, and other hazardous air pollutants identified by the EPA. Emission estimates listed in Table 3.4-14 are based on coal trace element concentrations, expected removal efficiencies, and other emission factors from available literature.

Coal trace element concentrations vary significantly between coal suppliers. Since a coal supplier has not been selected for Stanton 2, it is necessary to estimate these trace concentrations independently. Where possible, concentrations were estimated on the basis of information contained in the EPA publication "Estimating Air Toxics Emissions from Coal and Oil Combustion Sources" (EPA-450/2-89-001). In the absence of information from that source, concentrations were estimated from values contained in "Trace Elements in Coal" (Vlado Valkovic, CRC Press, 1983).

Expected removal efficiencies were derived from emission test results from similar facilities. The removal efficiencies listed in the table should be representative. However, it should be noted that there is not an abundant amount of information available to predict removal performance.

Formaldehyde, radionuclide, and polycyclic organic matter (POM) emissions are based on emission factors from the EPA publication "Estimating Air Toxics Emissions from Coal and Oil Combustion Sources." Estimates of phenol and pyridine emissions were based on information contained in the EPA publication "Emissions of Reactive Volatile Organic Compounds from Utility Boilers" (EPA-600/7-80-111).

BACT regarding these trace emissions will occur as part of control technologies (BACT) for particulate, sulfur dioxide, carbon monoxide, and volatile organic compound emissions.

3.4.3.6 Summary. The following is a summary of BACT for Stanton 2 and the associated emission rates.

- Sulfur dioxide--A wet limestone scrubber AQCS designed to meet an SO_2 emission limit of 0.32 lb/MBtu.
- Nitrogen oxides, CO, and VOC--Combustion controls designed to meet an NO_x emission requirement of 0.32 lb/MBtu for NO_x, 0.15 lb/MBtu for CO, and 0.015 lb/MBtu for VOCs.

- Particulate--An electrostatic precipitator designed to meet a 0.02 lb/MBtu (0.01 gr/dscf) emission limit.

3.4.4 Design Data for Control Equipment

Control equipment design data are included as part of the detailed BACT analyses contained in Subsection 3.4.3.

3.4.5 Design Philosophy

In general, air quality control system designs are determined based on conservative design parameters. Parameters are developed to ensure adequate performance to equal or better emission requirements. Where necessary, adequate spares (i.e., 50 percent spare capacity in the FGD system) are provided to ensure the operating reliability of the plant. Specific details of the design philosphy can be found in the detailed BACT analyses contained in Subsection 3.4.3.

Table 3.4-1
Economic Evaluation Criteria

Item	Value
Fuel Burn Rate	4,286 MBtu/h
Initial Operation	January 1997
Economic Recovery Period	35 years
Contingency Cost Factor	10 percent
Capital Escalation Rate	4.5 percent
O&M Escalation Rate	4.75 percent
Additive Escalation Rate	4.75 percent
Levelized Fixed Charge Rate[a]	7.90 percent
Present Worth Discount Rate	7.03 percent
Levelization Factor[b]	1.687
Indirects Cost Factor	16 percent
Allowance for Funds Used During Construction	7.10 percent
Capacity Factor	100 percent
1991 Pebble Lime Cost	80 $/ton
1991 Limestone Cost	8 $/ton
1991 Labor Cost	43,333 $/man-year
1991 Energy Cost	47.59 mills/kWh
1991 Waste Disposal Cost	10 $/ton

[a]Calculations are based on the economic recovery period, cost of money, and margins for insurance and taxes.

[b]Calculations are based on the economic recovery period, escalation rate, and present worth discount rate.

Table 3.4-2
Coal Quality Analysis

Ultimate Analysis	Typical
Carbon	67.0 percent
Hydrogen	4.50 percent
Sulfur	2.5 percent
Moisture	7.5 percent
Nitrogen	1.29 percent
Chlorine	0.11 percent
Oxygen	5.1 percent
Ash	12.00 percent
Higher Heating Value	12,400 Btu/lb

Table 3.4-3
Fabric Filter Design Parameters[a]

Parameter	0.12 lb/MBtu Particulate Emission	0.02 lb/MBtu Particulate Emission
Inlet Gas Flow, acfm	1,636,900	1,636,900
Gas Temperature, F	290	290
Gas Pressure Drop, in. wg	8.0	8.0
Fabric Filter Units	2	2
Compartments Per Unit	12	12
Bags Per Compartment	450	406
Total Number of Bags	10,800	9,744
Filter Area		
Per bag, ft^2	96	96
Per compartment, ft^2	43,200	38,980
Total, ft^2	1,036,800	935,400
Cloth Velocity		
All compartments on-line, ft/min	1.58	1.75
Two compartments out-of-service, ft/min	2.10	2.30
Peak Demand,[b] kW	2,770	2,680

[a]Design parameters are based on one (440 MW net) unit.

[b]Also includes differential ID fan power to overcome fabric filter draft losses.

Table 3.4-4
Electrostatic Precipitator Design Parameters[a]

Parameter	0.02 lb/MBtu Particulate Emission
Inlet Gas Flow, acfm	1,636,900
Gas Temperature, F	290
Gas Velocity, fps	3.5
Aspect Ratio	1.8
Specific Collecting Area, ft^2/1,000 acfm	743
Total Collecting Area, ft^2	1,326,000
Number of Transformer Rectifiers	48
Peak Demand,[b] kW	3,470

[a]Design parameters are based on one (440 MW net) unit.

[b]Also includes differential ID fan power to overcome electrostatic precipitator draft losses.

Table 3.4-5
Capital and Annual Costs of Particulate Removal Systems[a]

	Fabric Filter 0.012 lb/MBtu Particulate Emission ($1,000)	Electrostatic Precipitator 0.02 lb/MBtu Particulate Emission ($1,000)	Fabric Filter 0.02 lb/MBtu Particulate Emission ($1,000)
Capital Costs			
Fabric filter	29,100	NA	25,700
Electrostatic precipitator	NA	22,530	NA
Ductwork and differential ID fans	3,190	4,040	3,190
Waste handling	1,010	1,330	1,000
1991 capital cost	33,300	27,900	29,890
Contingency	3,330	2,790	2,990
1991 direct capital cost	36,630	30,690	32,880
Escalation	7,460	6,250	6,690
Direct capital cost	44,090	36,940	39,570
Indirects	7,050	5,910	6,330
AFUDC	6,680	5,600	6,000
1997 total capital cost	57,820	48,450	51,900
Levelized Annual Costs			
Operating personnel	470	470	470
Maintenance	3,430	790	1,270
Energy	2,860	3,350	2,760
Demand	170	210	170
1997 levelized annual operating cost	6,930	4,820	4,670
Fixed charges on capital	4,570	3,830	4,100
1997 total levelized annual cost	11,500	8,650	8,770
Incremental Removal Cost, $/ton	19,180	Base	NA

[a]Costs are for particulate removal systems installed downstream of a 440 MW net unit.

Table 3.4-6
Sulfur Dioxide Emissions

Alternative	Uncontrolled Emission[a] lb/MBtu	Controlled Emission Rate lb/MBtu	Annual Emission[b] tpy
PC Boiler/Wet Lime AQCS	4.03	0.24	4,506
PC Boiler/Wet Limestone AQCS	4.03	0.32	6,008
PC Boiler/Lime Spray Dryer AQCS	4.03	0.32	6,008

[a]Uncontrolled emissions are based on a typical case fuel sulfur content of 2.5 percent and a higher heating value of 12,400 Btu/lb.

[b]Annual emissions are based on a 100 percent capacity factor.

Table 3.4-7
Selected Wet Lime Scrubber AQCS Design Parameters[a]

Parameter	
Outlet SO_2 Emission, lb/MBtu	0.24
System Inlet Gas Flow, acfm	1,556,000
Inlet Flue Gas Temperature, F	290
Number of Electrostatic Precipitator Units	1
Gas Velocity, ft/sec	3.5
Specific Collection Area, ft^2/1,000 acfm	743
Total Collecting Area, ft^2	1,326,000
Number of Collecting Fields	6
Module Diameter, feet	37
Operating/Spare Modules	2/1
Water Usage, gpm	458
Liquid/Gas Ratio, gal/1,000 acfm	75
Module Outlet Temperature, F	125
Additive Molar Ratio,[b] mol Ca/mol S	1.10
Lime Consumption, tph	8.7
AQCS Peak Demand[c]	13,220

[a]All values are for an AQCS located downstream of one full size (440 MW net) pulverized coal boiler.

[b]Moles of calcium per mole of sulfur in the coal.

[c]Includes all equipment associated with SO_2 and particulate removal system operation including differential ID fan power to overcome AQCS draft losses.

Table 3.4-8
Selected Wet Limestone Scrubber AQCS Design Parameters[a]

Parameter	
Outlet SO_2 Emission, lb/MBtu	0.32
System Inlet Gas Flow, acfm	1,556,000
Inlet Flue Gas Temperature, F	290
Number of Electrostatic Precipitator Units	1
Gas Velocity, ft/sec	3.5
Specific Collection Area, ft^2/1,000 acfm	743
Total Collecting Area, ft^2	1,326,000
Number of Collecting Fields	6
Module Diameter, feet	37
Operating/Spare Modules	2/1
Liquid/Gas Ratio, gal/1,000 acfm	100
Water Usage, gpm	462
System Outlet Temperature, F	125
Additive Molar Ratio[b]	1.12
Limestone Consumption, tph	16.4
AQCS Peak Demand, kW[c]	16,150

[a]All values are for an AQCS located downstream of one full size (440 MW net) pulverized coal boiler.

[b]Moles of calcium per mole of sulfur in the coal.

[c]Includes all equipment associated with SO_2 and particulate removal system operation including differential ID fan power to overcome AQCS draft losses.

Table 3.4-9
Selected Lime Spray Dryer AQCS Design Parameters[a]

Parameter	
Outlet SO_2 Emissions, lb/MBtu	0.32
System Inlet Gas Flow, acfm	1,636,900
Flue Gas Temperature, F	290
Module Diameter, ft	48
Operating/Spare Modules	3/1
Water Usage, gpm	421
Module Outlet Temperature, F	165
Additive Molar Ratio[b], mol Ca/mol S	1.6
Lime Consumption, tph	12.4
Fabric Filter Inlet Gas Flow, acfm	1,425,100
Fabric Filter Compartments	12
Number of Bags per Compartment	338
Total Number of Bags	8,110
Filter Area per Bag, ft^2	96
Cloth Velocity, ft/min	
All Compartments On-Line	1.8
Two Compartments Out-of-Service	2.4
AQCS Peak Demand[c]	8,600

[a]All values are for an AQCS located downstream of one full size (440 MW Net) pulverized coal boiler.

[b]Moles of calcium per mole of sulfur in the coal.

[c]Includes all equipment associated with SO_2 and particulate removal system operation including differential ID fan power to overcome AQCS draft losses.

Table 3.4-10
Capital Costs of AQCS Alternatives[a]

	Wet Lime AQCS 0.24 lb/MBtu SO_2 Emission Rate $1,000	Wet Limestone AQCS 0.32 lb/MBtu SO_2 Emission Rate $1,000	Lime Spray Dryer AQCS 0.32 lb/MBtu SO_2 Emission Rate $1,000
Additive Storage and Preparation	9,180	10,560	10,210
Flue Gas Desulfurization	27,300	31,680	22,730
Particulate Removal	22,530	22,530	17,950
Flue Gas Supply and Exhaust	7,500	7,500	9,870
Waste Storage and Conditioning	2,630	2,630	2,900
1991 Capital Cost	69,140	74,900	63,660
Contingency	6,910	7,490	6,370
Direct Capital Cost	76,050	82,390	70,030
Escalation	15,480	16,770	14,260
Direct Capital Cost	91,530	99,160	84,290
Indirects	14,640	15,870	13,490
Interest During Construction	13,870	15,030	12,770
1997 Total Capital Cost	120,040	130,060	110,550

[a]Costs are total for one (440 MW net) unit.

Table 3.4-11
Levelized Annual Costs of AQCS Alternatives[a]

	Wet Lime AQCS 0.24 lb/MBtu SO_2 Emission Rate $1,000	PC Boiler/Wet Limestone AQCS 0.32 lb/MBtu SO_2 Emission Rate $1,000	PC Boiler/ Lime Spray Dryer AQCS 0.32 lb/MBtu SO_2 Emission Rate $1,000
Operating Personnel	1,730	1,730	1,570
Maintenance	4,210	4,150	5,680
Additive	15,700	2,960	22,470
Energy	5,530	6,450	3,530
Demand	820	1,000	530
Waste Disposal	9,080	9,710	9,930
1997 Total Levelized Annual Operating Cost	37,070	26,000	43,710
Fixed Charges on Capital'	9,480	10,270	8,730
1997 Total Levelized Annual Cost	46,550	36,270	52,440
Incremental Removal Cost, $/ton	6,870	Base	NA

[a]Costs are total for one (440 MW net) unit.

Table 3.4-12
Nitrogen Oxides, Carbon Monoxide, VOC, and Ammonia Emissions

Alternative	Uncontrolled Emission lb/MBtu	Reduction Rate percent	Emission Rate lb/MBtu	Annual Emission tpy
Post-Combustion NO_x Controls				
NO_x emissions	0.32	40	0.192	3,604
Ammonia emissions (20 ppm)	0.0128	NA	0.0128	240
NO_x emissions[a]	0.32	30	0.224	4,205
Ammonia emissions (10 ppm)	0.0064	NA	0.0064	120
Combustion Controls Only				
NO_x emissions	0.32	NA	0.32	5,943
CO emissions	0.15	NA	0.15	2,816
VOC emissions	0.015	NA	0.015	282
Ammonia emissions	0	NA	0	0

[a]SNCR NO_x reduction limited to 30 percent to minimize ammonia slip and to avoid the potential of an ammonia chloride plume.

Table 3.4-13
SNCR System Capital and Annual Costs

	40 Percent Reduction SNCR System $1,000	30 Percent Reduction SNCR System $1,000
Capital Costs		
SNCR system	5,130	4,320
Balance-of-plant	2,730	2,300
1991 capital cost	7,860	6,620
Contingency	790	660
Direct capital cost	8,650	7,280
Escalation	1,720	1,450
Direct capital cost	10,370	8,730
Indirects	1,660	1,400
Interest during construction	1,570	1,320
1997 total capital cost	13,600	11,450
Levelized Annual Costs		
Operating personnel	260	260
Maintenance	560	470
Additive	2,310	1,730
Loss in fly ash sales	980	980
Landfill costs of fly ash	290	290
Energy	950	800
Demand	110	100
1997 total annual operating cost	5,460	4,630
Fixed charges on capital	1,070	900
1997 total levelized annual cost	6,530	5,530

Table 3.4-14
Estimated Lead and Noncriteria Pollutant Emissions

Pollutant	Uncontrolled Emissions: Average Emission (lb/h)	Uncontrolled Emissions: Worst Case Emission (lb/h)	Removal Rate (percent)	Controlled Emissions: Average Emission (lb/h)	Controlled Emissions: Worst Case Emission (lb/h)
Lead	2.9	13	95	0.16	0.64
Beryllium	0.78	2.2	99	0.0088	0.022
Fluorine	32	181	99	0.36	1.8
Mercury	0.083	0.46	90	0.0094	0.046
Sulfuric Acid Mist	0.86	1.2	50	149	179
Antimony	0.59	1.9	99	0.0066	0.019
Arsenic	7.7	44	95	0.43	2.2
Barium	69	164	99	0.78	1.6
Cadmium	1.9	17	90	0.21	1.7
Chromium	9.4	53	90	1.1	5.3
Cobalt	3.4	19	95	0.19	0.95
Copper	6.3	21	90	0.71	2.1
Hydrogen Chloride	0.038	0.078	80	88	160
Manganese	35	273	95	1.9	14
Nickel	9.2	36	90	1.0	3.6
Phosphorus	52	292	90	5.8	29
Zinc	8.6	49	90	0.97	4.9
Formaldehyde	NA	NA		0.56	0.56
Phenol	NA	NA		3.2	3.2
Polycyclic Organic Matter	NA	NA		0.017	0.017
Pyridine	NA	NA		3.2	14
Radionuclides					0.47 μCi/h

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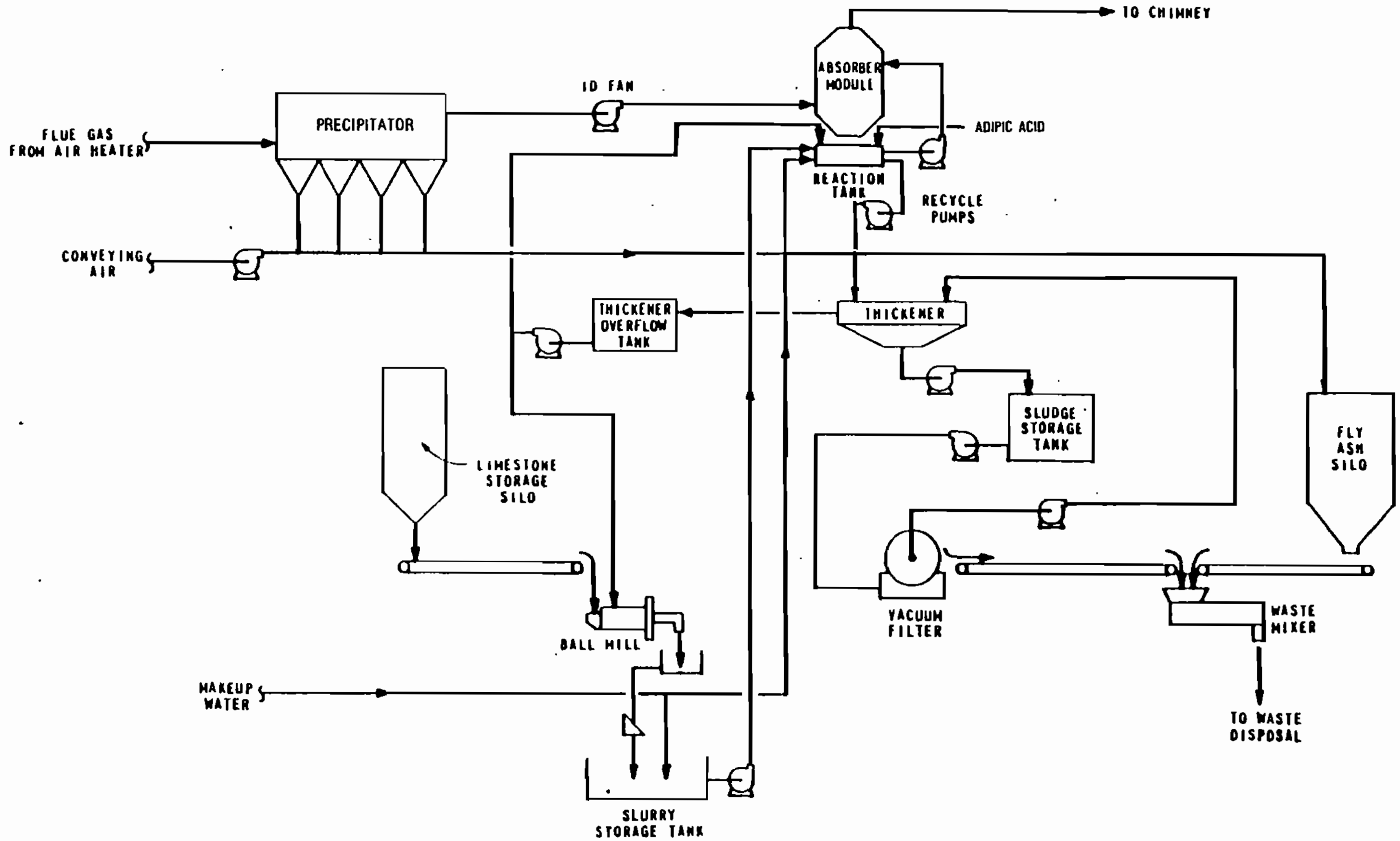
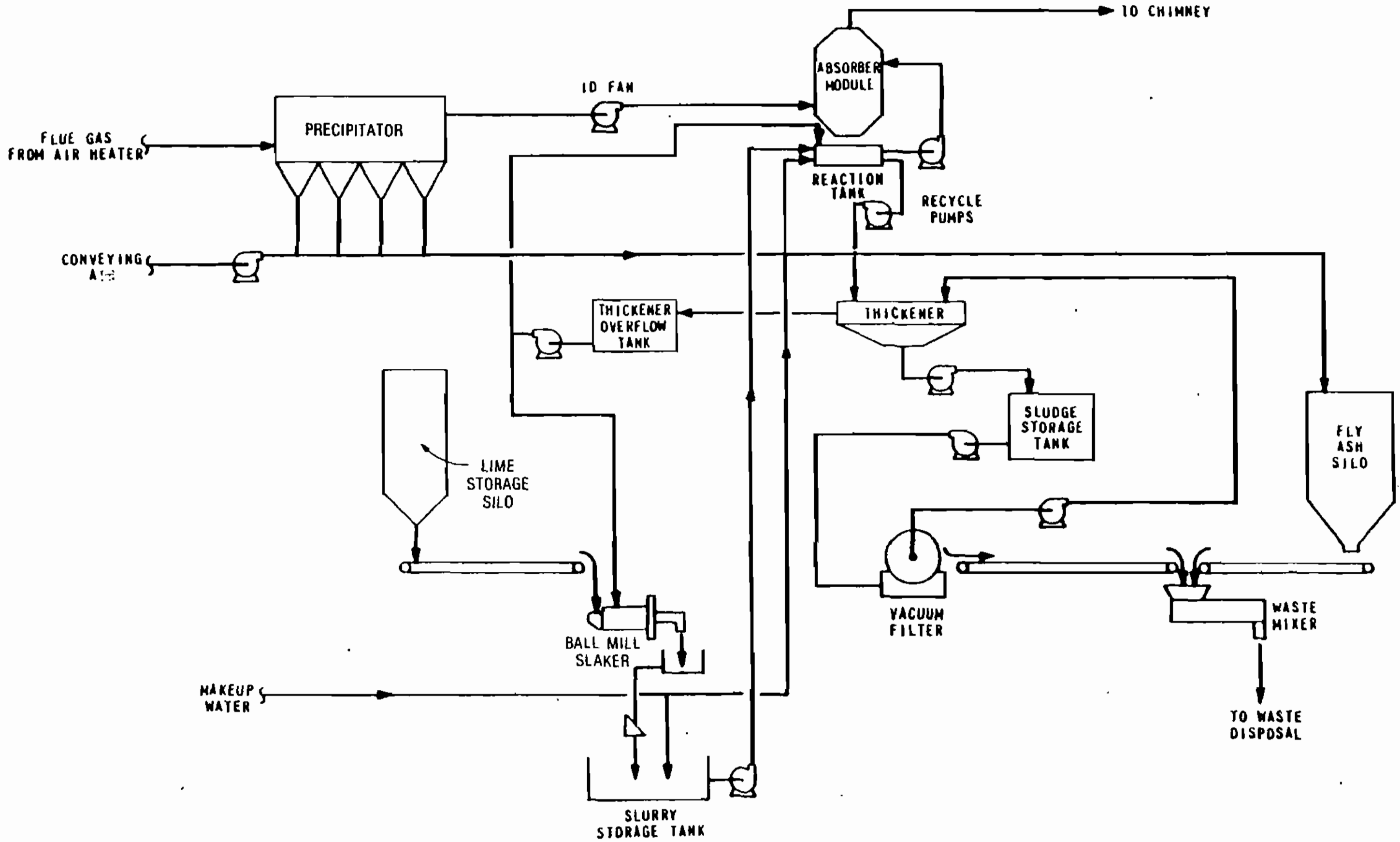


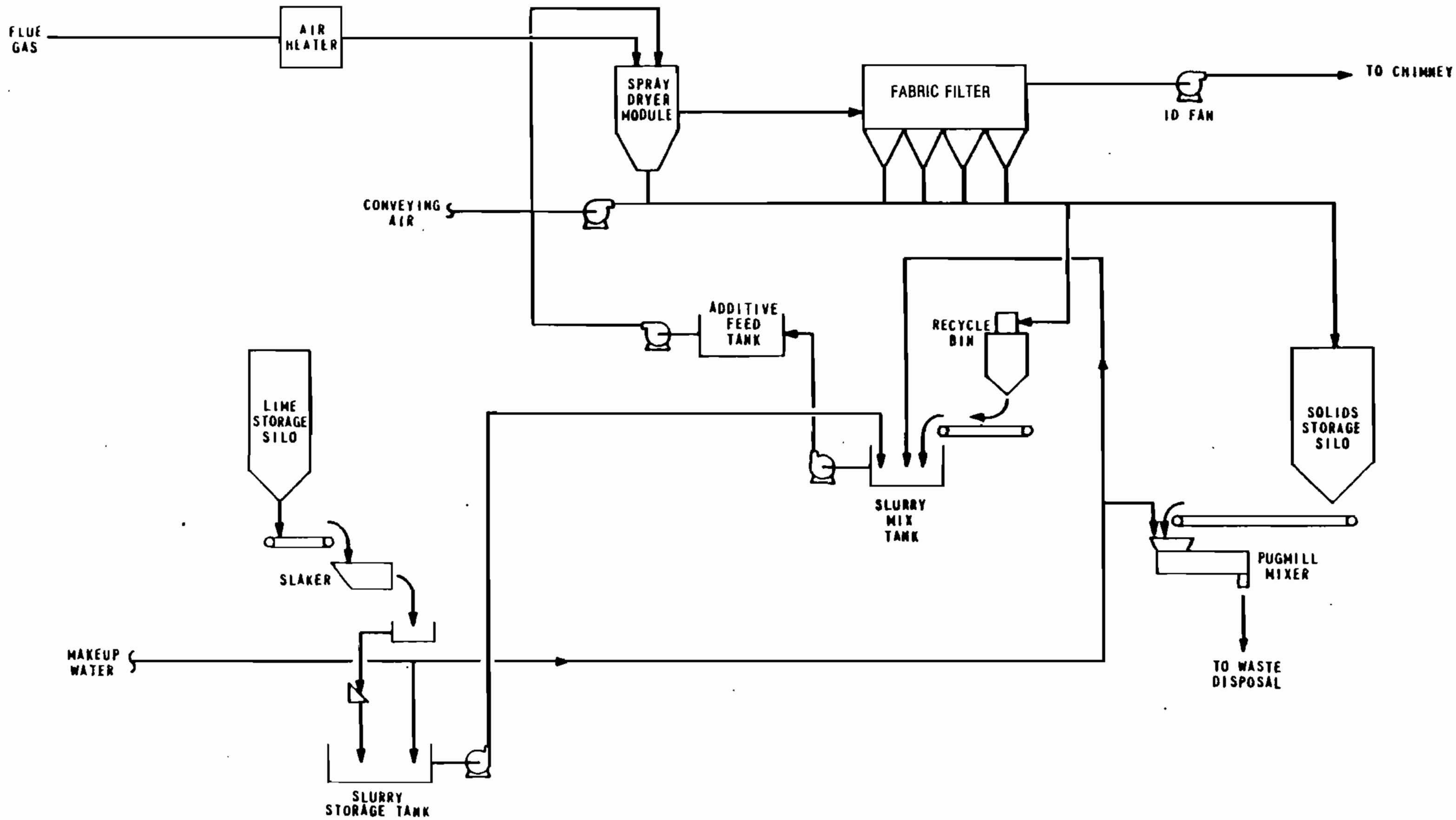
WET LIMESTONE SCRUBBER
AIR QUALITY CONTROL SYSTEM

Figure 3.4-1



WET LIME SCRUBBER
AIR QUALITY CONTROL SYSTEM

Figure 3.4-2



LIME SPRAY DRYER
AIR QUALITY CONTROL SYSTEM

Figure 3.4-3

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4.5 Air Impacts

There will be temporary and minor air quality impacts during the construction phase for Stanton 2. These impacts will include the generation of fugitive dust and equipment exhaust emissions. The local air quality impact will be minimized by the application of appropriate dust suppression control methods, such as water spraying. The impacts will end when the construction activities are completed and the facilities are ready for operation.

Fugitive dust emissions will be associated with the construction of buildings and roads onsite. These construction activities will include land clearing, ground excavation, cut and fill operations, and actual construction of facilities. There can be considerable variation in daily fugitive dust emissions depending upon the level of activity, the nature of the operations, and prevailing weather conditions. Vehicular traffic at the construction site will produce a large portion of the emissions.

Stanton 2 will be constructed on the existing Stanton Energy Center site which is fairly isolated and has a reasonable buffer area due to the large size of the site. Fugitive dust impacts should be localized and of short duration.

The operation of construction equipment will cause a minor, temporary impact on the local air quality. The use of construction equipment will be short-term, and air quality impacts will cease at the completion of construction.

Air quality impacts related to the construction of the transmission line and alternate access road are discussed in Chapter 6.

5.6 Air Quality Impacts

Air quality impacts for the operation of Stanton 1 and 2 were included in Section 5.5 of the original SCA. However, because combustion parameters and emissions rates for Stanton 2 have been revised, additional dispersion modeling was performed to verify that air quality impacts remain below all applicable PSD increments and ambient air quality standards.

5.6.1 Description of Pollutant Emissions

5.6.1.1 Steam Generator Emissions. Pollutant emissions for Stanton 1 and 2 were summarized in Table 5.5-1 of the original SCA. The sulfur dioxide (SO_2) emission rate was 1.14 lb SO_2/MBtu heat input. Particulate matter and nitrogen oxides (NO_x) emissions were equal to the New Source Performance Standards (NSPS) levels of 0.03 and 0.60 lb/MBtu heat input, respectively. Carbon monoxide (CO) emissions were based on an emission factor recommended by EPA of 1.0 lb CO/ton of coal fired.

The revised emission rates for Stanton 2 reflect new pollutant control technology as determined by the revised Best Available Control Technology (BACT) assessment, included in Section 3.4 of this application. The Stanton 2 SO_2 emission rate is 0.32 lb SO_2/MBtu heat input on a 30-day rolling average basis. This rate will also be appropriate on an annual basis. The maximum 3-hour and 24-hour SO_2 emission rates are proposed to be 0.85 and 0.67 lb SO_2/MBtu heat input, respectively. The Stanton 2 particulate matter and NO_x emission rates also reflect improved pollutant removal rates resulting from advances in technology. The particulate matter emission rate is 0.02 lb particulate matter/MBtu heat input. The NO_x emission rate is 0.32 lb NO_2/MBtu heat input. Carbon monoxide emissions for Stanton 2 will reflect more current methods of estimating emission rates than previous EPA recommended estimates. The Stanton 2 CO emission rate is 0.15 lb CO/MBtu heat input.

Table 5.6-1 presents emission rate data for Stanton 1 and the proposed Stanton 2 steam generators. These rates were used in the revised dispersion modeling analysis.

5.6.1.2 Emissions from Other Station Sources. Emissions from other station sources were included in Subsection 5.5.1.2 of the original SCA. These emissions and associated impacts are not expected to change significantly.

5.6.2 Revised Impacts of Stack Emissions

The revised impacts of stack emissions on ambient air quality concentrations of SO_2, NO_2, TSP, and CO were evaluated using currently accepted EPA computer dispersion modeling methods. The primary objective of this analysis was to verify that the reduced emissions from Stanton 2, coupled with minor changes in exhaust gas parameters, would result in equal or lower ambient air quality impact levels.

5.6.2.1 Dispersion Modeling Methodology. The dispersion modeling methodology for the revised impact analysis differs somewhat from the original analysis. This was necessary because of changes in EPA's modeling guidelines since the original SCA was issued.

5.6.2.1.1 Screening Modeling. Previous screening-level modeling, using the EPA approved PTMAX model, had shown that the boiler operating at maximum capacity would result in the highest ground level concentrations. This result was assumed to continue. Therefore, no additional screening modeling was performed.

5.6.2.1.2 Refined Modeling. Refined dispersion modeling was originally performed using EPA's CRSTER model. The revised dispersion modeling uses an updated version of the CRSTER model known as the Industrial Source Complex Short Term (ISCST) model. Like the CRSTER model, the ISCST model uses site-specific meteorological data to predict ground level concentrations at various user defined receptor locations.

5.6.2.1.3 Meteorological Data. The original dispersion modeling used surface meteorological data from Orlando, Florida, and upper air mixing height data from nearby Tampa, Florida, for the years 1974-1978. The revised dispersion modeling uses more current data for the years 1981-1985 from the same surface and upper air stations.

5.6.2.1.4 Stack Parameters. Exhaust gas parameters for Stanton 2 changed slightly as a result of revised coal characteristics and pollution control technologies. Table 5.6-2 presents the exhaust gas parameters for both units.

5.6.2.1.5 Receptor Locations. The original dispersion modeling used a polar receptor array with one receptor on each of 36 radials (one every 10 degrees by azimuth) at downwind distances that were required to give a representative spacial coverage. The revised modeling analysis uses the same approach. The downwind spacing of the polar rings starts at 450 meters (the nearest property fence line) and locates additional rings at 100-meter intervals out to 2,000 meters. Between 2,000 and 4,000 meters, the ring spacing was increased to 200 meters.

5.6.2.2 Modeling Results. Tables 5.6-3 through 5.6-5 present the results of the revised dispersion modeling analysis. Table 5.6-3 shows the maximum predicted ground level pollutant concentrations for the existing Stanton 1 and revised Stanton 2 emissions. The table also lists the location, year, and period when the concentrations were expected to occur.

Table 5.6-4 presents a similar comparison of predicted ground level concentrations and PSD Class II increments. Again, each pollutant's impact is below the applicable level.

Table 5.6-5 presents a comparison of the predicted ground level concentrations determined in the original and revised dispersion modeling studies. The modeling results show a net air quality improvement for every pollutant except carbon monoxide.

The apparent increase in CO impacts is due solely to the lower emission factor recommended by EPA at the time of the original SCA submittal. The previously used emission factor of 1.0 lb CO/ton of coal fired results in an emission rate lower than rates currently used based on manufacturer's experience.

Table 5.6-1
Steam Generator Emission Rates for Units 1 and 2

Pollutant	Unit 1	Unit 2
Sulfur Dioxide, lb/MBtu		
Long-term emission rate	1.14	0.32
24-hour emission rate	1.14	0.67
3-hour emission rate	1.14	0.85
Nitrogen Oxides, lb/MBtu	0.60	0.32
Particulate Matter, lb/MBtu		
TSP	0.03	0.02
PM_{10}	--	0.02
Carbon Monoxide		
lb/MBtu	--	0.15
lb/ton coal[a]	1.00	--

[a]Emission estimate was based on recommended emission factor from EPA's document AP-42, applicable at the time of the original SCA submittal.

Table 5.6-2
Exhaust Gas Parameters for Units 1 and 2

Parameter	Unit 1	Unit 2
Stack Height, ft	550	550
Stack Exit Temperature, F	126.5	123.9
Stack Exit Flow, acfm	1,202,867	1,310,120
Stack Exit Diameter, ft	19	19
Stack Exit Velocity, fpm	4,242.5	4,620.8
Heat Input, MBtu/h	4,183	4,286

Table 5.6-3
Maximum Predicted Ground Level Pollutant Concentrations from the Two-Unit Operation

Pollutant	Averaging Period	Impact ($\mu g/m^3$)	Distance (meters)	Direction (degrees)	Year	Period (day/h)
SO_2	Annual	3	2,800	130	1985	--
	24-hour[a]	84	900	20	1984	243/1
	3-hour[a]	508	900	260	1981	176/5
NO_2	Annual	2	2,800	130	1985	--
PM	Annual	2	(See Note Below)			
	24-hour[a]	27	(See Note Below)			
CO	8-hour[a]	26	900	20	1984	183/2
	1-hour[a]	94	900	240	1983	232/13

[a]Impacts represent the highest, second-highest pollutant concentrations for the five-year period 1981-1985.

Note: Maximum annual and highest, second-highest 24-hour ground level particulate impacts result from fugitive dust and material handling particulate matter emissions. Because these emissions are not expected to change, impacts from the original SCA have been used.

Table 5.6-4
Comparison of PSD Class II Increments with Predicted Ground Level Pollutant Concentrations from the Two-Unit Operation

Pollutant	PSD Class II Increment ($\mu g/m^3$)	Predicted Impacts ($\mu g/m^3$)	Percent of Increment Consumed
SO_2			
Annual	20	3	15
24-hour	91	84	92
3-hour	512	508	99
PM			
Annual	19	2	11
24-hour	37	27	73

Table 5.6-5
Comparison of Ground Level Pollutant Impacts from the Original SCA and the Current SCA

Pollutant	Original SCA Pollutant Concentrations[a] ($\mu g/m^3$)	Revised Pollutant Concentrations[b] ($\mu g/m^3$)
SO_2		
Annual	6	3
24-hour	85	84
3-hour	509	508
NO_2		
Annual	3	2
PM		
Annual	2	2
24-hour	27	27
CO		
8-hour	12[c]	26
1-hour	44[c]	94

[a]Modeled with 1974-1979 meteorological data.

[b]Modeled with 1981-1985 meteorological data.

[c]CO impacts from the original SCA were listed as only $<<1$ mg/m^3. The values listed are given in $\mu g/m^3$ and are two times the impact from Stanton 1 based on the revised modeling.

5.7 Noise

The noise impacts of the ultimate (4-unit) development of the Stanton Energy Center were included in Subsection 5.7.2.1 of the original SCA, and they will not be discussed further.

9.0 Coordination

Orlando Utilities Commission Government Contacts

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Gregg M. Worley	Chemical/Environmental Engineer
Lewis Nagler	Regional Meteorologist

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Perry Oldenburg	Deceased

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Dwight Jenkins	Hydrologist III

10.1.5 PSD Applications/Permits

A PSD permit for phased construction of both Stanton 1 and 2 was issued by EPA Region IV during the initial permitting process. An updated BACT analysis is required for Stanton 2 and is included as part of Section 3.4 of this application. In addition, air quality dispersion modeling has been conducted for Stanton 1 and 2. Stanton 2 was modeled at both the emission rate provided in the original application and at the rate proposed as BACT in this application in order to demonstrate a reduced level of air quality impact and compliance with ambient air quality standards and PSD increments. A description of the modeling analysis and results is presented in Section 5.6.

10.1.6 Coastal Zone Management Certification

Not applicable to this project because proposed facilities will not be located in a coastal county.