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July 2, 1993

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Division of Air
Resources Management

Mr. Clair Fancy, P.E., Chief
Bureau of Air Regulation
Florida Department of Environmental Protection
2600 Blair Stone Road
Tallahassee, FL 32399-2400

Re: Okeelanta Power Limited Partnership
DEP File No. AC50-219413, PSD-FL-196

Dear Mr. Fancy:

On behalf of Okeelanta Power Limited Partnership (Okeelanta Power), we are submitting this letter and the following comments on the permit documents issued by the Department of Environmental Protection (DEP) on June 3, 1993, for the above-referenced facility. Our two major concerns are addressed below. A number of other issues are addressed in the addendum which is attached hereto and incorporated herein by reference. This letter does not fully address all of the issues we discussed during our meeting earlier today, but we will promptly provide you with the additional information that you requested in the next few days.

LIMITS ON FACILITY CAPACITY (74.9 MW)

In various documents, including the "Intent to Issue", "Technical Evaluation and Preliminary Determination", "Draft Permit", "BACT Determination" and "RACT Determination", the Department has reported the facility's capacity as 71.25 MW. It is unclear whether the 71.25 MW refers to a net or a gross generating capacity. As noted in the application and all supporting documents, the project will be designed with a gross generating capacity of 74.9 MW.

From discussions with staff at FDEP, we understand the 71.25 MW was selected to provide a significant margin with respect to the 75 MW limit on generating capacity. Roget's Dictionary of Electrical Terms defines "capacity" as a generating facility's "output in kilowatts under ordinary full load conditions." The Okeelanta facility will be designed such that its capacity, as defined above, will be 74.9 MW. Consequently, the electric generating equipment, namely the steam turbine generator, will be so sized. In addition, control systems will be installed to ensure that the capacity of the facility will not exceed 74.9 MW.

Any electric power plant, including particularly a cogeneration plant which supplies steam to a process, is subject to upset conditions (rapid, unexpected changes in steam flow or electrical load) which may result in an instantaneous increase in electrical output. The power plant control systems are designed to control most upset conditions; however, there may be some conditions which are outside the range of the control systems. As stated above, the Okeelanta facility will be designed to generate 74.9 MW under normal operating conditions and will have state-of-the-art controls to assure generation is limited to the design limit of 74.9 MW under most conditions, including many upset conditions. When an upset or abnormal condition occurs the facility's control system will maintain or, if necessary, promptly restore the electric generator's output to less than 75 MW by initiating one or more of the following:

- The steam turbine governor control valves will operate to maintain throttle steam flow and electrical output at the desired set point.
- If necessary, the steam output of the boilers can be reduced by operating the main steam line controls.
- If necessary, process steam can be bypassed around the steam turbine generator to the heat rejection system (condenser).
- If necessary, process steam can be vented to the atmosphere.
- As a last resort, the Control Room operator can initiate manual controls.

We believe that an upset condition which may result in an excursion of electrical generation above 75 MW will be infrequent, and of short duration, and normal operating conditions should be restored within minutes so that the facility's integrated hourly average will not exceed 75 MW. Consequently, we believe that it is appropriate to maintain the facility's capacity limit at 74.9 MW, subject to an additional requirement of meeting a short-term averaging condition (for example, one hour averaging time) with accompanying monitoring and reporting requirements. We will provide engineering plans, as well as a monitoring and reporting plan, to the Department within 30 days after they become available.

The New York Public Service Commission dealt with a similar question concerning an 80 MW threshold established by New York state law. See Case 91-E-0454, Proceeding on Motion of the Commission to Interpret and Enforce the Output Limitations Implementing the PSL §2(2-a) 80 MW Size Restriction (attached). In New York, there were 14 facilities with capacities at or near 79 MW that were at risk of exceeding the 80 MW threshold during upsets or compliance tests. In response to the arguments presented by those companies, the State of New York ultimately concluded that compliance with the 80 MW threshold could be demonstrated by measuring a facility's electrical output over a four-hour period. By analogy, the New York case helps to confirm our belief that a one-hour averaging period is reasonable, restrictive, and appropriate for demonstrating compliance with Florida's 75 MW threshold.

RESTRICTIONS ON WOOD WASTE

The draft permit states that any wood waste materials used as fuel shall be "free from painted and chemically treated wood, ... and special wastes...." (page 7, Specific Condition No. 12), which implies that any amount of these materials in the fuel is not permitted. This restriction is unnecessarily severe and restrictive.

"Special wastes" are defined by statue to include "yard trash" and "construction and demolition debris." See Section 403.703(33), Florida Statutes, as amended by Chapter 93-207, Laws of Florida (CS/HB 461). "Yard trash" is defined to mean:

> vegetative matter resulting from landscaping maintenance or land clearing operations and includes materials such as tree and shrub trimmings, grass clippings, palm fronds, trees, and tree stumps.

Section 17-701.200(87), Florida Administrative Code. "Construction and demolition debris" is defined to include "materials generally considered to be not water soluble and non-hazardous in nature, including but not limited to ... lumber, from the construction or destruction of a structure ...." Section 17-701.200(17), F.A.C. The term also includes "tree remains, trees, and other vegetative matter which normally results from land clearing or land development operations ...." Id.

Okeelanta Power intends to use clean wood wastes from yard trash and construction and demolition debris as fuel. There appears to be no legal or factual justification for the Department's proposed restriction on Okeelanta Power's ability to use these "special wastes." Therefore, we request that the term "special wastes" be deleted from Specific Condition No. 12.

As Okeelanta Power has explained to the Department in the past, Okeelanta Power will obtain its wood waste materials from a variety of suppliers, including several local governments. Okeelanta will use its best efforts to ensure that the wood waste materials are free from treated or painted wood. However, Okeelanta cannot guarantee that the fuel will be completely free from such materials at all times. Despite Okeelanta's best efforts, there is always the possibility that some painted or chemically treated wood will be present in the fuel supply. Moreover, Okeelanta has evaluated a "worst case" scenario in its application. Okeelanta considered the possibility that up to 3% of the wood in its fuel supply could consist of treated wood. The analysis of this scenario demonstrated that even the use of 3% treated wood would not result in any adverse air quality impacts. As shown in the application, all ambient air quality standards and the Department's Air Toxics Policy would be satisfied.

We believe the intent of Specific Condition No. 12 is to require that Okeelanta Power implement best efforts to prevent treated or painted wood from being combusted at this facility and ultimately to ensure that air quality standards are met. We are concerned that a literal interpretation of Specific Condition No. 12 would not allow any amount of treated or painted wood in the fuel supply. In order to assure the Department that best efforts are being implemented and air quality standards are being met, a two part compliance program will be performed. A protocol describing this program will be submitted to the Department 60 days prior to commencing operations.

The program will consist of the following elements:

1. A fuel quality inspection, testing, and management program (including daily visual inspections of the incoming wood material and regular inspections at the originating wood yard sites) will be implemented to control the amounts of treated and painted wood in the fuel and fuel concentration limits for arsenic, chromium, and copper will be established.

2. Stack testing for arsenic, chromium and copper emissions will be conducted every six months for the first two years of operations, as required for other pollutants under Specific Condition 24 of the draft permit. Based on the stack testing for arsenic, chromium, and copper, compliance with the No-Threat Levels for these compounds will be demonstrated.

We believe that as a result of implementing this plan, the requirement for ash monitoring, as stated in Specific Condition 12, is no longer necessary and should be deleted.

Attached for your consideration is a revised Specific Condition No. 12.

CONCLUSION

Thank you for your consideration of these comments. If you have any questions concerning these comments, please call me at 904-331-9000.

Sincerely,

David A. Buff

David A. Buff, M.E., P.E.
Principal Engineer

cc: Gus Cepero, Okeelanta Corp.
David Dee, Carlton-Fields
Jewell Harper, EPA
John Bunyak, NPS
Bevin Beaudet, PBCHU
Mark Carney, USGenCo
Richard Donelan
Clare Lardner
Stephanie Brooks--DEP District
File (2)

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STATE OF NEW YORK
PUBLIC SERVICE COMMISSION

At a Session of the Public Service
Commission held in the City of
Albany on April 1, 1992

COMMISSIONERS PRESENT:

Peter Bradford, Chairman
Harold A. Jerry, Jr.
Gail Garfield Schwartz
James T. McFarland
Henry G. Williams

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CASE 91-E-0454 - Proceeding on Motion of the Commission to Interpret and Enforce the Output Limitations Implementing the PSL §2(2-a) 80 MW Size Restriction.

ORDER INTERPRETING AND CLARIFYING THE 80 MW OUTPUT LIMITATIONS

(Issued and Effective April 22, 1992)

BY THE COMMISSION:

BACKGROUND

On August 27, 1991, a Notice Soliciting Comments was issued in the captioned proceeding, requesting comments on a Proposed Order interpreting and implementing the 80 MW limit set forth at PSL §2(2-a), which governs qualification as a cogeneration facility under state law. That standard has been enforced through imposition of the four output limitations, first promulgated in the Salt City Order[1] and later clarified in the Ramapo Order.[2] As set forth in the Ramapo Order, the output

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[1] Case 28689, Salt City Energy Venture, Order Approving Contract Subject to Conditions (Issued January 15, 1988).

[2] Case 28689, Consolidated Edison Company of New York, Inc. and Ramapo Cogeneration L.P., Order Granting Rehearing In Part (Issued February 1, 1989).

limitations mandate that a qualifying facility (QF) developer, in order to demonstrate compliance with the §2(2-a) limit, must:

A. Guarantee that the electric output will not exceed 80 MW for any quarter-hour period;

B. Cause to be installed and maintained in good operating condition, the metering and other equipment necessary for monitoring compliance with the above standard by the utility and for producing records for Staff review;

C. Agree that, in the event of two exceedences of the 80 MW limit within any five-year period, it shall lose status as a cogeneration facility under PSL §2(2-a), and as a result, forego entitlement (if relevant) to the statutory minimum rate set forth in PSL §66-c for the term of the contract; and

D. Agree that even if it loses status as a cogeneration facility under PSL §2(2-a) for a violation of the standards set forth [above], it will not seek to operate the facility at output levels above 80 MW.

These output limitations have been applied to contracts for facilities which, because sized near the 80 MW limit, could potentially exceed the limit. There are fourteen such contracts, including one between Niagara Mohawk Power Corporation (Niagara Mohawk) and Onondaga Cogeneration L.P. (Onondaga), approved before the limitations were first adopted.[1]

This proceeding was instituted, and comments were solicited, premised upon a petition from the Independent Power Producers of New York (IPPNY) requesting interpretation and clarification of the limitations. The Proposed Order resolved the questions IPPNY raised, but asked for comments before any

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[1] Twelve of these fourteen contracts, in effect as of April 1, 1992, are with Niagara Mohawk, one is with New York State Electric & Gas Corporation (NYSEG), and one is with Consolidated Edison Company of New York, Inc. (Con Edison). These facilities are listed in Appendix A.

decision was implemented. Because the comments raise a number of issues not addressed in the Proposed Order, it is replaced with this Order.

THE PROPOSED ORDER

IPPNY's Petition

In a petition filed April 29, 1991, IPPNY asks for an interpretation of the output limitations which it asserts would conform the limitations to reasonable capacity testing procedures. According to IPPNY, compliance with the limitations can be achieved only through installation of elaborate control systems that will continuously monitor plant performance and compensate for variations in generation output caused by either changes within the electricity grid or within the plant. IPPNY worries, however, that pre-commercial testing of control systems, such testing following a major repair, and annual Dependable Maximum Net Capability (DMNC) testing, could force developers to exceed the 80 MW limit, thereby violating the limitations. IPPNY submits that such forced exceedences should not trigger the penalties applicable under the limitations.

Properly tested control systems, IPPNY maintains, are essential to compliance with the limitations. As IPPNY describes it, these systems must be tested prior to a facility's commercial operation, and, during such tests, plant components must be run at their maximum output. IPPNY is concerned that the 80 MW limit might be exceeded as a consequence of that testing. IPPNY adds that the same considerations adhere following major repair or replacement of major plant components that might occur during the plant's operating lifetime.

IPPNY also relates that developers will conduct four-hour DMNC tests at least yearly to assess the capacity capability of their plants.[1] Interpreting the New York Power Pool (NYPP) DMNC testing procedures as measuring capability on a time-averaged basis over the four-hour period, IPPNY argues that it might be necessary for a facility to operate at more than 80 MW at some times during the four-hour period in order to compensate for other times when it is operating at less than 80 MW. Without an exception from the limitations, says IPPNY, production from a unit must be cycled down during a test, producing an artificially low result and thereby unfairly constraining facilities to a lower DMNC rating. For example, IPPNY insists, a unit might show a test result of 76 MW when it could actually achieve 79 MW. These circumstances, IPPNY asserts, are exacerbated during the summer DMNC tests, because adjustments may be made to the test results based on the average ambient temperature experienced during a utility's summer peak.

Giving these testing protocols, IPPNY contends that no purpose would be served by applying the output limitations during the tests. IPPNY believes developers could not evade the policy underlying the limitations through conducting frequent tests, because the number of such tests is limited under each developer's contract. To remove any incentive to evade the

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[1] According to IPPNY, the results of those tests will be used for a variety of purposes depending upon the provisions of the particular contract.

limitations through testing, however, IPPNY proposes that developers not be paid for electricity delivered in excess of the 80 MW limitation during a test.

IPPNY maintains that advising utilities to implement the limitations with flexibility during tests is the appropriate relief. As IPPNY sees it, this would defer technical issues to resolution between the utility and the developer, without further administrative intervention.

Resolution of Issues in The Proposed Order

The Proposed Order explains that we have continuing jurisdiction over the output limitations, which insure that the requirements of qualifying status under PSL §2(2-a) are met.[1] To effectuate that jurisdiction, the Proposed Order resolves some of the issues IPPNY raises. But IPPNY's petition also presents a number of questions that may intrude upon individual contract specifications, because contractual DMNC testing requirements vary significantly, depending upon the terms each utility and developer negotiated in each contract. Therefore, while generic policies were promulgated in the Proposed Order, it was also suggested that refinements to those policies could be made in order to reflect individual circumstances.

The Proposed Order addressed a number of issues related to capacity testing and the output limitations. As to pre-commercial testing, although the output limitations were to be

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[1] *See*, e.g., Case 90-E-1156, Niagara Mohawk Power Corporation and Seneca Power Partners L.P., Order Modifying Environmental Conditions (Issued April 29, 1991); Case 89-E-081, Order Denying Rehearing and Clarifying Prior Order (Issued December 12, 1989).

interpreted broadly because they enforce the statutory mandate incorporated in PSL §2(2-a), their application was limited to "operation" of a facility. Therefore, the Proposed Order states that tests conducted during the pre-commercial period should not be considered within "operation" of the plant, as defined under the limitations. The Proposed Order also advises the utilities to make arrangements with developers to avoid conflicts over exceedences during pre-operational and annual DMNC testing, and allows utilities to insist upon reasonable conditions circumscribing the time, number, extent and duration of such exceedences. Moreover, utilities were informed that in no event was a developer to be paid for any generation delivered in excess of 80 MW.

Similar reasoning was applied to the circumstances when a plant comes back on-line after a major repair. Under the proposed Order, however, definition of what constitutes a "major repair" was left to a later time. The Proposed Order again obligates utilities to craft reasonable conditions circumscribing exceedences during such testing, if they are to be excused, and advises that developers may not be paid for exceedences.

The Proposed Order also described potential conflicts between the output limitations and the terms of some individual contracts. These included the Onondaga - Niagara Mohawk contract, which was approved before the limitations were adopted, and the Niagara Mohawk - East Syracuse contract, which provided for a size increase that raised questions regarding the limitations. While the East Syracuse contract conflict has been resolved,[1] the Onondaga contract problem remains, and another developer, Selkirk Cogen Partners L.P. (Selkirk Cogen) reports an additional potential conflict.

POSITIONS OF THE PARTIES

CCE

Concerned Citizens for the Environment, Inc. (CCE) argues that overly-broad exemptions from the output limitations might "erode the ratepayers' federally protected right to receive fair and reasonable rates while benefitting from a safe, reliable, electric generating system."[2] Referring to the NYPP Report filed in Case 91-E-0237,[3] CCE argues that an excess of capacity supplies from state-qualifying cogenerators threatens to expose ratepayers to negative impacts, such as the burden of payments for unneeded capacity. According to CCE, tight controls must be placed on QF output to avoid exacerbating those negative impacts. Indeed, it believes that QFs should be fined for exceedences, and that the amount of the fines should be refunded to ratepayers through the fuel adjustment clause. CCE also urges that reporting requirements be imposed on QFs and that contracts be reopened for further review if a QF exceeds the output limitations on a recurring basis.

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[1] Case 91-E-0923, East Syracuse Generating Company, L.P., Declaratory Ruling (Issued October 2, 1991) (East Syracuse Ruling).

[2] CCE Comment, p.2.

[3] Case 91-E-0237, Report of the Member Corporations of the New York Power Pool Concerning 1991 Long-Run Avoided Cost Estimates (August 30, 1991).

CCE also interprets §2(2-a) as limiting our discretion. According to CCE, the authority to supervise the implementation of that statute cannot be delegated to the utilities. CCE urges strict interpretation and enforcement of the few legislative mandates applicable to QFs, who, it says, are not otherwise regulated, except by contract language.

Kamine

Kamine Syracuse Cogen Co., Inc. (Kamine), a developer planning to build a 79 MW facility, maintains that the definition of a "major repair" should be expanded. The developer believes that the definition propounded in the Proposed Order includes only instances where the entire plant undergoes substantial downtime. According to the developer, major repairs may also take place where performance from the plant is significantly reduced, but the plant does not cease production entirely. As Kamine describes it, its combined-cycle facility could undergo a major repair to the steam turbine component while the gas turbine component continues to operate. Under such circumstances, the developer says, bringing the steam turbine back into service would implicate testing procedures as rigorous as those that would be employed if the entire plant had been out of service.

Lavair

Lavair Cogeneration L.P., another 79 MW developer, supports IPPNY's position and asks that it be informed of the outcome of this proceeding.

Niagara Mohawk

Niagara Mohawk begins by noting that it is the purchasing utility for 12 of the 14 projects governed by the output limitations. Niagara Mohawk requests that it be accorded more flexibility in enforcing those limitations. It also questions whether DMNC testing upon return to service after a major repair implicates the same considerations as pre-operational testing.

According to the utility, it needs greater flexibility in implementing the limitations. It asks that it be authorized to measure a facility's output over a four-hour period instead of the fifteen minute interval contemplated under the first output limitation. The utility asserts the four-hour period is preferable, because it is the standard used to conduct DMNC tests and to measure output when determining the size of a facility.[1] The utility also argues that the four-hour output is easily determined from a facility's generation log book and is more reliable than measurement over the fifteen minute interval, which, it says, may be prone to telemetering errors.

Niagara Mohawk is also concerned that in order to comply with the limitations, developers will install devices which automatically trip a unit off-line if it appears that the 80 MW limit will be exceeded during a fifteen minute interval. Measuring output over the four-hour period instead, the utility believes, will allow the developer to average out such temporary

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[1] See Case 90-E-0238, American Ref-Fuel Company of Hempstead, Declaratory Ruling (Issued August 22, 1990).

exceedences, and avoid the potentially harmful consequences to system reliability of tripping a unit off-line with little or no notice. The utility also maintains that using the four-hour average would accommodate DMNC testing protocols, and that an exception to the limitations for exceedences that occur during such tests would not be needed if developers could average output (and exceedences) over the four-hour period.

Foreseeing emergency circumstances where operating a facility at its maximum output, even if it exceeds 80 MW, might benefit ratepayers, Niagara Mohawk also argues that the output limitations should be waived under extreme or unusual circumstances. For example, the utility says, extra output might be needed to sustain voltage in an area, to provide for system restoration, or to increase operating reserve during a severe generation shortage. The utility suggests that the limitations should be waived when it requests additional production from an 80 MW facility under such circumstances.

Niagara Mohawk also asks for guidance in implementing its contract with Selkirk Cogen. As the utility describes it, that developer intends to build two facilities at its Selkirk site, one selling to Niagara Mohawk and another selling to Con Edison. According to the utility, those two facilities might not be truly separate, and, if interconnected, would exceed the 80 MW limit, thereby violating the output limitations. Niagara Mohawk asks for clarification of application of the output limitations to this site.[1]

NYSEG

NYSEG is concerned that the Proposed Order's construction of the word "operation" might affect its 79 MW contract with South Corning. That contract, the utility relates, contains provisions requiring the developer to operate its facility in accordance with NYSEG's operating procedures. The utility claims these contract provisions should control electricity deliveries at all times, despite language in the Proposed Order indicating that the output limitations would not adhere to electricity deliveries made during a "pre-operational" period.

NYSEG also asks for clarification that the reasonable conditions it is to negotiate circumscribing exceedences during capacity testing also govern the length and size of each exceedence permitted during such a test. The utility believes that exceedences of undue length or amount in size over 80 MW could harm interconnection equipment, unless it were reconfigured to absorb the exceedence. NYSEG also requests permission to refuse to excuse an exceedence even during testing, if the exceedence is due to an equipment failure instead of caused by reasonable efforts to comply with testing protocols. Finally, NYSEG asserts that, although the output limitations have been

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[1] Niagara Mohawk's proposals to measure output over a four-hour period, and waive the limitations during unusual circumstances, were supported by IPPNY, in a letter dated November 8, 1991, and G.A.S., another 79 MW developer, in a letter from its agent, Entek Research, Inc., dated December 3, 1991.

imposed only on cogeneration facilities in the past, they should also be applied to any new energy-only contracts that might be entered into with small hydro facilities or alternate energy production facilities.

Selkirk Cogen

Selkirk Cogen Partners L.P. (Selkirk Cogen), the successor to J.M.C. Selkirk, Inc., criticizes some of the proposed policies expressed in the Proposed Order, and asks for relief specific to its project. The developer also supports Niagara Mohawk's proposed implementation of the output limitations.

Selkirk Cogen objects to two aspects of the proposed Order. First, it believes that it should be paid for electricity delivered in excess of 80 MW during testing periods. It claims the amount of such exceedences should be small, and that the utilities cannot be excused from paying for electricity delivered during exceedences, because PURPA, PSL §66-c and QF contracts require such payments.

Second, concerned that utilities might unfairly implement the conditions governing exceedences during testing, Selkirk Cogen contends that utilities should be required to publish their policies on these matters in advance, with an opportunity for administrative review. Without such guidelines, the developer believes that the numerous disputes between utilities and developers over these issues would have to be adjudicated.

Selkirk Cogen also complains that the Proposed Order erects an obstacle to development of its project. As the developer reads the Proposed Order, a state-qualifying facility sharing a site with a second facility must be "truly separate" from the second facility in order to comply with the output limitations. Selkirk Cogen reports it intends to build two facilities at its Selkirk site -- a 79 MW facility selling to Niagara Mohawk, qualifying under PSL §2(2-a),[1] and an approximately 277 MW (gross) facility selling to Con Edison that qualifies only under federal, and not state, law.[2] Although Selkirk admits that it plans to thermally interconnect the two facilities, it argues that this design configuration does not violate the output limitations or forfeit its qualification under PSL §2(2-a).

The developer describes its contract with Niagara Mohawk as requiring it to maintain state QF status under §2(2-a), or a rate penalty will be imposed. The developer relates, however, that the contract explicitly waives application of the penalty provision if sales exceeding 80 MW are made from a facility located at the site to Con Edison (or another third-party utility). The developer also claims its contract with Con Edison requires that utility to purchase all the electricity

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[1] The April 7, 1988 and February 14, 1990 Letter Orders approving this contract are cited in Appendix A.

[2] Case 28689, Consolidated Edison Co. of N.Y., Inc. and J.M.C. Selkirk, Inc., - Contract No. 346, Letter Order (Issued October 24, 1989) and Letter Order (Issued November 14, 1991).

produced at the site except that electricity sold to Niagara Mohawk.

The developer then cites the Con Edison Staff Memorandum accompanying approval of the Con Edison contract as noting that project plans called for the sharing of a steam turbine generator between the two facilities.[1] Selkirk also quotes the Niagara Mohawk Staff Memorandum accompanying approval of Supplement Nos. 1 and 2 to the Niagara Mohawk contract as stating that the rate penalty provision exception "is apparently in anticipation of the loss of state QF status that will occur when, if ever, the 252 MW Phase II [i.e., Con Edison] portion of this facility becomes operational."[2] According to Selkirk, the only condition placed upon its combined two-facility arrangement was that, upon entry of the Con Edison facility into service, the developer was required to provide firm security for both the existing and projected front-load under the contract.

Consequently, Selkirk concludes that a thermally and electrically interconnected facility was approved, and that two thermally and electrically isolated projects were not mandated. Selkirk also argues that, although the net generation standard of PSL §2(2-a) was imposed on its Niagara Mohawk contract, the four output limitations were not. According to Selkirk, approval of

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[1] Case 28689, Consolidated Edison Co. of N.Y., Inc. and J.M.C. Selkirk, Inc. - Contract No. 346, Staff Memorandum (Filed Session of October 18, 1989), p. 3.

[2] Case 28689, Niagara Mohawk Power Corporation and J.M.C. Selkirk, Inc. - Contract No. 524, Supp. Nos. 1 and 2, Staff Memorandum (Filed Session of February 14, 1990), p. 2.

the contract for that facility is therefore exempt from the fourth output limitation, which restricts future output to 80 MW.

Selkirk asks for clarification that it may proceed with its development plans which, at a minimum, requires that the two facilities share a steam turbine, and so be thermally interconnected. The developer also argues that it should be permitted to electrically interconnect the two facilities, even though it admits it might be possible to supply Niagara Mohawk with electricity from only a single combustion turbine and to supply Con Edison with electricity exclusively from the remaining combustion turbines and the shared steam turbine. Finally, the developer claims that both projects are fully dispatchable, so that each utility's ratepayers are protected against the prospect of overpayments.

DISCUSSION AND CONCLUSION

IPPNY's proposals to clarify the output limitations, as modified in the Proposed Order, are generally acceptable. Niagara Mohawk's proposal to experiment with implementation of the output limitations also may be permitted in part. But if the utility's proposed method proves unworkable, both it and developers must be prepared to return to the method originally envisioned for implementation, upon reasonable notice to the affected QFs. The further clarification some utilities and QFs request concerning the principles governing testing exceedences is, however, not necessary. Finally, the output limitations are applicable to the Onondaga and Selkirk Cogen contracts, albeit that some flexibility in that application is appropriate.

Excusing Testing Exceedences

As IPPNY suggests, exceedences that occur during pre-operational testing, testing during return to service following a major repair, and DMNC testing may be excused.[1] Although, as CCE maintains, the output limitations are to be interpreted broadly because they enforce the 80 MW statutory mandate incorporated in PSL §2(2-a), they are primarily intended to control the amount of output for which a developer can receive payment. Because, as IPPNY concedes, developers may not be paid for exceedences which occur during the testing process, the output limitations need not adhere to such events.[2] This reasoning adheres whether the exceedence occurs during pre-operational tests, tests following major repairs, or other DMNC tests.

Utilities, however, may insist on conditions circumscribing such exceedences, if they are to be excused. Utilities may require notice before a test is conducted, and may reasonably limit their time, number, extent, and duration.[3] Moreover, in no event may a developer be paid for any generation delivered in excess of 80 MW, because that would violate both the

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[1] As NYSEG points out, the distinction drawn in the Proposed Order between "pre-operational" testing and other types of testing is confusing and unwarranted.

[2] For this reason, and because power purchase contracts already penalize developers for loss of §2(2-a) status, the further sanctions CCE proposes are redundant and so will not be adopted.

[3] In negotiating such conditions with QFs, utilities should remember that contractual provisions governing tests differ significantly, and the negotiated conditions should reflect analysis of individual contract language.

PSL §2(2-a) requirement that utility ratepayers' obligation to purchase from state-qualifying facilities be limited to 80 MW and the fourth output limitation, which restricts production to 80 MW. Selkirk Cogen's argument that it must be paid for electricity delivered during exceedences is rejected because it conflicts with that principle, and because neither federal law nor its contract require such payments.

Several parties petitioned for a more exact and detailed statement of rules which would govern testing exceedences. Although we are obligated to construe the Public Service Law and our policies, we need not regulate all of these details, which are a component of the ongoing utility - QF relationship. As responsible business entities, utilities and QFs should be able to resolve such technical issues without further intervention. Indeed, it has been explained that the utility - QF contractual relationship is no different from contractual arrangements that "involve fuel, labor or equipment...not subject to...regulation under normal conditions,"[1] and that "New York's statutory structure, sound regulatory practice, and ... limited staff resources"[2] preclude

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[1] Case 92-E-0032, Erie Energy Associates, Declaratory Ruling (Issued March 4, 1992), p. 2.

[2] Case 90-E-0775, Consolidated Edison Company of New York, Inc., et al., Order Accepting Contracts For Filing and Denying Petition (Issued December 10, 1990)(Hydro Quebec Order), p. 8.

ongoing supervision of such utility management issues.[1] Therefore, the requests for clarification are denied.[2]

Similarly, the description of "major repair" in the Proposed Order -- as a repair that causes substantial down-time at a facility -- adequately guides the parties. Since developers generally are not paid when their capacity is unavailable, there is no incentive to manufacture a "major repair" event in order to avoid the limitations. Of course, exceedences caused by minor equipment failures that do not result in substantial down-time may not be excused. As explained in the Salt City Clarification Order:

> The output limitations were established to force a developer to design its facility so that exceedences simply would not occur...[D]evelopers [may not] evade that responsibility, through claiming that an exceedence resulted from negligence or some other failure that was not intentional.[3]

Consequently, requests for further definition of the "major repair" terminology are denied as well.

<u>Niagara Mohawk's Proposal</u>

Niagara Mohawk asks that it be allowed to implement the output limitations through measuring output over a four-hour period instead of the 15-minute interval prescribed in the first limitation. IPPNY and several QFs support this proposal.

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[1] These management decisions, of course, are subject to later prudence review. Hydro Quebec Order, p. 6.

[2] NYSEG also asks for guidance in applying the limitations to future, hypothetical contracts. Such hypotheticals, however, are not ripe for review here, and NYSEG should negotiate future contracts in good faith.

[3] Case 28689, Salt City Energy Venture, <u>Order Granting In Part Petition For Clarification</u> (Issued November 1, 1988), p. 4.

Niagara Mohawk believes that its four-hour method would be easier to implement, and would avoid potential threats to system reliability. Maintaining that reliability, the utility asserts, could be difficult if an 80 MW unit were to trip off-line with little or no notice in order to forestall an exceedence measured over the 15-minute interval. If such an exceedence could be averaged out over the four-hour period, however, this threat could be avoided. Moreover, the utility claims that so averaging out temporary exceedences over the four-hour period would eliminate the need to excuse exceedences occurring during testing. These are sufficient reasons to allow the utility to experiment with implementation of its method.[1]

Both the utility and the developers are advised, however, that if the utility's method produces unforeseen consequences or requires the utility to absorb excessive amounts of generation, permission to use the method can be rescinded. In that event, developers must be prepared to reconfigure their facilities to abide by the 15-minute interval. While the experiment will not be terminated without reasonable notice, developers must take that possibility into account in designing their facilities.

Niagara Mohawk also suggests that the output limitations should be waived under unusual circumstances where use of additional generation might prove beneficial. This proposal conflicts with §2(2-a) and would undermine the integrity

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[1] Con Edison and NYSEG may join in this experiment at their option.

of the 80 MW limitation. Under the proposal, developers must be paid for production in excess of 80 MW, flatly contradicting the PSL §2(2-a) mandate circumscribing the benefits associated with state-qualifying status to facilities which restrict sales to 80 MW. Moreover, this proposal has been rejected before, when propounded by developers.[1] Consequently, Niagara Mohawk's request -- for permission to waive the limitations and purchase power in excess of 80 MW under unusual circumstances -- is denied.

The Onondaga Contract

The 79 MW Onondaga - Niagara Mohawk contract was approved before the limitations were adopted. Although the contract requires the developer to maintain §2(2-a) status, or it is voidable at the option of either party, it does not establish a method for demonstrating compliance with the 80 MW limit. As a result, given the language of §2(2-a), even one exceedence of the 80 MW limit by the developer would violate the statute, and at the utility's option, would invalidate the contract.

As discussed in the Proposed Order,[2] to ameliorate this harsh result, Niagara Mohawk is directed to amend that contract to provide for application of the output limitations, if the developer agrees. If the developer declines to accept the limitations, however, the utility is directed to treat any

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[1] Case 28689, Central Hudson Gas and Electric Corporation and Mid-Hudson Cogeneration L.P., Order Approving Contract Subject to Conditions (Issued August 30, 1988).

[2] Onondaga did not file comments in response to the Proposed Order.

exceedence of the 80 MW limit by this facility as a violation of §2(2-a).

The Selkirk Cogen Contract

Selkirk Cogen complains that the application of the output limitations described in the Proposed Order conflicts with the approvals of its contracts with Niagara Mohawk and Con Edison for two facilities at its Selkirk site. Those approvals, says Selkirk Cogen, exempt it from the limitations and allow it to thermally and electrically integrate its facilities. Although, contrary to this allegation, the output limitations were applied to the contract with Niagara Mohawk and remain in effect, the developer may be accorded some relief. As it points out, the approvals for its contracts were confusing, and should be clarified here.

The output limitations clearly were imposed on Selkirk Cogen's contract with Niagara Mohawk. The April 7, 1988 Letter Order approving that contract tied the approval to the January 15, 1988 Order in Case 28689 (i.e., the Salt City Order), where the output limitations were first promulgated.[1] Moreover, any confusion which might have existed over application of the output limitations was clarified in the Niagara Mohawk Staff Memorandum, which stated that:

> "The net generation standard adopted in Case 28689 for determining eligibility for state QF status would apply to this facility. Consequently, the output limitations, as modified in the Ramapo Order, apply to this contract."[2]

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[1] April 7, 1988 Letter Order, p. 2.

[2] Niagara Mohawk Staff Memorandum, p. 9.

Therefore, the output limitations[1] adhere to Selkirk Cogen's facility selling to Niagara Mohawk, and those limitations prohibit deliveries in excess of 80 MW from that facility.

Selkirk is correct, however, in contending that the approvals of its contract with Niagara Mohawk were confusing. In this respect, its circumstances are similar to those described in the East Syracuse Ruling. There, the developer asserted that it could, under its approvals, expand production of its facility beyond the 80 MW limit. It was decided, however, that the fourth output limitation, which prohibits production in excess of 80 MW, applied to that facility because:

> While compliance with that requirement is easily preserved when an entirely separate power plant component is subjected to that limitation, it makes less sense to assume that the limitation would be applied to a plant with integrated components, as East Syracuse surmises.[2]

Similar logic adheres to Selkirk Cogen. Like East Syracuse, however, Selkirk Cogen did not have adequate warning that its proposed arrangements were impermissible. Indeed, the approval language described as misleading in the East Syracuse Ruling resembles the approval language applied to Selkirk Cogen's Niagara Mohawk contract. Moreover, again like East Syracuse, Selkirk Cogen has moved forward diligently with project

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[1] The "net generation standard" and the "output limitations" terminologies both refer to the four conditions originally adopted in the Salt City Order. See, e.g., Kamine Syracuse April 20, 1988 Letter Order (cited in Appendix A); Case 28689, Niagara Mohawk Power Corporation and Dresser-Rand Company, Order Approving Contract Subject to Conditions and Denying Petition (Issued June 19, 1989), pp. 12-14.

[2] East Syracuse Ruling, p. 9.

development efforts based on its interpretation of the approvals. In fact, its Niagara Mohawk facility is expected to enter service before the end of the year. Therefore, Selkirk Cogen is entitled to relief similar to that accorded East Syracuse.

In the East Syracuse Ruling, that developer was allowed to proceed with its plans upon conditions that would effectuate the intent behind the output limitations and protect ratepayers. As CCE asserts, however, the output limitations enforce a statutory mandate that protects ratepayers from excessive payments for overproduction and so cannot be waived entirely. As a result, the relief accorded Selkirk Cogen must preserve the output limitations, including the ban against production in excess of 80 MW.

This can be accomplished by permitting Selkirk Cogen to build a facility that is thermally, but not electrically, interconnected. The developer proposed exactly that arrangement in its Environmental Information Report submitted in conjunction with the approval of its Con Edison contract.[1] As it stated there:

> A separate 79 MW turbine cogeneration project...at the [Selkirk] site...for the sale of 79 MW of electricity...to Niagara Mohawk...will be built and operated regardless of [the Con Edison project]....[The latter] project will be housed in buildings separate from the 79 MW project. However,...steam from the initial project...will be ducted to the new 110 MW steam turbine.[2]

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[1] Developers were required to file such reports as part of the contract approval process. Case 27834, Order Establishing Filing Requirements (Issued May 30, 1989).

[2] Case 28689, J.M.C. Selkirk, Inc., Environmental Information Report (Filed July 19, 1989), p. 2.

Construction of that facility, where one combustion turbine is dedicated to Niagara Mohawk while the others are dedicated to Con Edison, and waste heat from all those turbines drives a steam turbine dedicated to Con Edison, may be excused under these circumstances even though the facilities will not be entirely physically separate.[1] Selkirk Cogen could have used waste heat from the Niagara Mohawk turbine for purposes other than generating electricity for sale to that utility, even if the Con Edison facility were never built. Consequently, allowing the developer to generate electricity for sale to Con Edison using that heat may be permitted under these circumstances.

Electrically interconnecting the plants now, however, would violate the output limitations by forcing Niagara Mohawk to buy generation produced from plant components sized well in excess of 80 MW, and also deviates from the plant configuration the developer itself described. Moreover, as discussed in the East Syracuse Ruling, developers are bound to the size of facility described in their contract. An attempt by Selkirk Cogen to sell Con Edison production in excess of the 252 MW (net) limitation in that contract, by using electrical production from the combustion turbine dedicated to Niagara Mohawk, would constitute a size increase requiring entry into a new contract governing production from the "extra" turbine. Selkirk Cogen may preserve its existing contracts only if the plant components are

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[1] Other developers, however, who cannot point to implicit approval of their plans are now on notice that the 80 MW limit is violated unless facilities at the same site producing in excess of that amount are physically separate. East Syracuse Ruling, pp. 11-12.

dedicated in accordance with the developer's originally-proposed configuration.

In contrast to the East Syracuse contracts, however, further ratepayer protections need not be imposed here. Unlike the East Syracuse facilities, both of Selkirk Cogen's facilities are largely dispatchable, thereby sufficiently protecting ratepayers from the prospect of purchasing overpriced energy.

The Commission orders:

1. Consolidated Edison Company of New York, Inc., New York State Electric & Gas Corporation, and Niagara Mohawk Power Corporation are directed to enforce the 80 MW limit established by PSL §2(2-a), as applied to their power purchase contracts with state-qualifying facilities through the output limitations, in conformance with the discussion in the body of this Order.

2. This proceeding is continued.

By the Commission,

William J. Barnes

(SIGNED)

WILLIAM J. BARNES
Deputy Secretary

Appendix A

80 MW PSL §2(2-a) STATE-QUALIFYING CONTRACTS

1. *Onondaga*. Case 28689, Niagara Mohawk Power Corporation and Onondaga Cogeneration L.P. - Contract No. 507, *Letter Order* (Issued November 6, 1987).

2. *G.A.S.* Case 29292, Niagara Mohawk Power Corporation and G.A.S. Alternative Systems, Inc. - Contract No. 449, *Order Conditioning Contract* (Issued July 7, 1987) *and* *Order on Rehearing* (Issued January 26, 1988).

3. *Salt City*. Case 28689, Niagara Mohawk Power Corporation and Salt City Energy Venture - Contract No. 452, *Order Approving Contract Subject to Conditions* (Issued January 15, 1988) *and* *Order Denying Petitions For Rehearing and Motion For Clarification* (Issued May 27, 1988).

4. *J.M.C. Selkirk*. Case 28689, Niagara Mohawk Power Corporation and J.M.C. Selkirk, Inc. - Contract No. 524, *Letter Order* (Issued April 7, 1988) *and* *Letter Order* (Issued February 14, 1990) *and* *Letter Order* (Issued June 15, 1990).

5. *Kamine Syracuse*. Case 28689, Niagara Mohawk Power Corporation and Kamine Syracuse Cogen Co., Inc. - Contract No. 527, *Letter Order* (Issued April 20, 1988) *and* *Order Approving Contracts Subject to Conditions* (Issued March 12, 1990) *and* *Letter Order* (Issued November 21, 1990).

6. *Hadson*. Case 28689, Niagara Mohawk Power Corporation and Hadson Power Partners of Rensselaer (formerly Ultra Cogen Systems, Inc.) - Contract No. 540, *Order Approving Contract Subject to Conditions* (Issued September 12, 1988) *and* *Letter Order* (Issued October 31, 1990) *and* *Letter Order* (Issued March 5, 1991).

7. *Power City*. Case 28689, Niagara Mohawk Power Corporation and Power City Generating, Inc. - Contract No. 574, *Letter Order* (Issued February 6, 1989) *and* *Letter Order* (Issued September 12, 1989) *and* *Order Granting Rehearing in Part* (Issued March 12, 1990) *and* *Letter Order* (Issued January 15, 1991).

8. *Indeck Olean*. Case 28689, Niagara Mohawk Power Corporation and Indeck Energy Services of Olean, Inc. (formerly Dresser-Rand Co.) - Contract No. 576, *Order Approving Contract Subject to Conditions and Denying Petition* (Issued June 19, 1989) *and* *Letter Order* (Issued November 30, 1990) *and* *Letter Order* (Issued June 3, 1991).

9. Ag-Energy. Case 28689, Niagara Mohawk Power Corporation and Ag-Energy, Inc. - Contract No. 529, Letter Order (Issued September 27, 1989) and Letter Order (Issued August 10, 1990) and Letter Order (Issued December 17, 1990).

10. East Syracuse. Case 28689, Niagara Mohawk Power Corporation and East Syracuse Generating Co. L.P. (formerly Old River A-9) - Contract No. 588, Letter Order (Issued September 27, 1989) and Letter Order (Issued August 10, 1990) and Letter Order (Issued March 4, 1991).

11. Lavair. Case 28689, Consolidated Edison Co. of N.Y., Inc. and Lavair Cogeneration L.P. - Contract No. 343, Letter Order (Issued October 10, 1989) and Order Granting Rehearing in Part (Issued January 26, 1990) and Order Granting Petition (Issued July 2, 1990) and Letter Order (Issued September 20, 1990).

12. Kamine Beaver Falls. Case 28689, Niagara Mohawk Power Corporation and Kamine Beaver Falls Cogen, Inc. - Contract No. 613, Letter Order (Issued May 29, 1990) and Order Granting Rehearing (Issued September 21, 1990) and Letter Order (Issued April 22, 1991).

13. NCP. Case 28689, Niagara Mohawk Power Corporation and Northern Consolidated Power, Inc. - Contract No. 606, Letter Order (Issued July 23, 1990) and Order Clarifying Prior Order (Issued December 5, 1990)(Letter Order Pending).

14. South Corning. Case 28689, New York State Electric & Gas Corporation and South Corning Cogeneration, Inc. - Contract No. 481, Letter Order (Issued September 21, 1990) and Letter Order (Issued April 1, 1991).

DRAFT
7/2/93

OKEELANTA

Specific Condition No. 12.

The fuel used at the cogeneration facility shall not contain hazardous substances, hazardous wastes, biomedical wastes, or garbage.

The fuel used at the cogeneration facility shall not contain special wastes, except wood, lumber, trees, tree remains, bagasse, cane tops and leaves, and other clean vegetative and cellulose matter.

The permittee shall perform a daily visual inspection of any wood waste or similar vegetative matter that has been delivered to the facility for use as fuel. Any shipment observed to contain prohibited materials shall not be used as fuel, unless such materials can be readily segregated and removed from the wood waste and vegetative matter.

The permittee shall design and implement a testing program for the wood waste and other materials delivered to the facility for fuel. This program shall be submitted to the Department's Bureau of Air Regulation for review and approval at least 60 days before the commencement of operations of the cogeneration facility. At a minimum, the program shall provide for the routine testing of the fuel at the originating wood yard sites as well as at the cogeneration site, to ensure that the quantities of painted or chemically treated wood in the fuel are minimized. Fuel tests shall be conducted weekly for the first year of operations at the facility and monthly thereafter, if the Department determines on the basis of the prior test results that less frequent testing is appropriate. The permittee shall not use any delivered fuel that contains an amount of treated or painted wood which, if burned, would violate the Department's No Threat Levels (NTL). The fuel pile shall not contain more than 3% treated or painted wood at any time.

ADDENDUM FOR
OKEELANTA POWER LIMITED PARTNERSHIP

1. TECHNICAL EVALUATION AND PRELIMINARY DETERMINATION (TEPD)

Page 2, Section I.C. Emissions- The wording in the first paragraph should be revised to state that the facility will be restricted to less than 25 percent fossil fuel on a quarterly basis, as required by 40 CFR 60.42a.

Page 5, Section III. Technical Evaluation- Last paragraph, same concerns expressed above related to the 25 percent fossil fuel limitation.

3. DRAFT PERMIT

Page 1, 2nd paragraph, line 15- The phrase "(or equivalent)" should be placed after the phrase "an activated carbon injection system" in this sentence.

Page 8, Specific Condition 15- It is recommended that this condition be revised to read as follows:

> " The consumption of No. 2 fuel oil shall be less than 25 percent of the total heat input to each boiler unit in any calendar quarter. Not more than 73,714 tons of coal shall be burned at this facility during any 12-month period. The combined heat input for coal and oil shall be less than 25 percent of the heat input on a calendar quarter basis."

Page 8, Specific Condition 17- It is noted that during the initial startup, debugging and testing of the facility prior to commercial operation, the cogeneration boilers may operate when the existing Okeelanta boilers are also operating. We therefore request that this condition be modified to allow such operation during this period prior to commercial operation. This period shall not exceed 12 months following initial firing of fuel in the boilers.

Page 8, Specific Condition 18- The same concern for Boiler No. 16 as above for the existing boilers.

Page 9, Specific Condition 19.b- It is requested that the phrase "and covered" at the end of this condition be deleted. The coal pile for this facility will be small and covering the coal pile will not result in any significant reduction in fugitive PM emissions, while creating operational difficulties.

Page 9, Specific Condition 20- In the table of emission limits, the limit for nitrogen oxides for all three boilers should be "862.5" TPY, instead of "873.1" TPY.

It is also requested that the following emission limits be added to the table. These limits are felt to be necessary to comply with the NSPS, and to insure air quality standards are met.

- Sulfur dioxide, 3-hour average of 1.2 lb/MMBtu and 588.0 lb/hr for coal firing.
- Visible emission limit of 20% opacity, 6-minute average, except up to 27% opacity is allowed for up to 6 minutes in any 1-hour period.

Page 10, Specific Condition 21.a- It is requested that a provision be added to this condition to cover the situation where a fuel such as coal is not initially fired in the facility. The following wording is suggested to be added at the end of this condition: "If during the 180 day period after initial startup a specific fuel has not been fired in a unit, the permittee shall be required to conduct the performance tests within 60 days of achieving the maximum operating capacity on this fuel, but no later than 180 days after initially firing this fuel in the unit."

Page 10 and 11, Specific Condition 21.b- It is requested that the following changes be made in the listing of EPA stack sampling methods:

- Add Method 3A to Method 3
- Delete Method 7C and add Method 7E- Method 7C experiences interferences due to ammonia and SO_2.
- Delete Method 108- this method is not for mercury

A provision is also needed under "Compliance Requirements" to address excess emissions which may occur for more than 2 hours duration, under startup conditions. Due to the types of fuels used in this facility (e.g., biomass) and the limits on combustion related emissions such as CO, excess emissions could occur up to 6 hours during startup. The following wording is suggested to be added to the specific conditions: The permittee shall comply with the excess emissions rule contained in F.A.C. Rule 17-296.210. In addition, the permittee is allowed excess emissions during startup conditions, provided such excess emissions do not exceed a duration of four hours, and such emissions in excess of two hours do not exceed six (6) times per year.

4. BACT DETERMINATION

Page 1, Sulfur Dioxide- Note that the 25 percent of heat input limitation should be stated as <u>less than</u> 25 percent on a <u>calendar quarter</u> basis.

Page 3- BACT Determination By DEP- For sulfur dioxide emissions, same comment as above on fossil fuel heat input limitation.

RECEIVED

MAY 25 1993

Division of Air
Resources Management

May 25, 1993

Mr. Clair Fancy, P.E.
Bureau Chief
Bureau of Air Regulation
Florida Department of Environmental Regulation
2600 Blair Stone Road
Tallahassee, FL 32399-2400

Re: Okeelanta Power Limited Partnership and Osceola Power Limited Partnership

Dear Mr. Fancy:

This correspondence is in follow up to our meeting with you last week concerning the Okeelanta and Osceola cogeneration facilities. Based on our discussions, Okeelanta and Osceola are willing to further reduce total SO_2 emissions from the two facilities by reducing the annual emissions due to coal burning. This will result in a significant reduction in worst-case annual SO_2 emissions compared to the worst-case annual emissions presented in the original applications. In the original applications, worst-case annual SO_2 emissions due to both facilities combined totalled 2,772 TPY. This is the level allowed under federal new source performance standards, Subpart Da, for resource recovery facilities. To address the Department's concerns, Okeelanta and Osceola are willing to reduce the total annual SO_2 emissions to 1,507 TPY. In addition, their proposal eliminates the use of long-term averaging and bubbling of the two facilities.

Annual SO_2 Limits for Each Facility

Okeelanta and Osceola are willing to accept individual annual SO_2 emission limits for each facility. Combined, the total SO_2 emissions from the two facilities are 1,507 TPY. Okeelanta Power and Osceola Power are willing to accept the following permit limits:

1. Okeelanta Power--1,154.3 TPY of SO_2. This is equivalent to 0.21 lb/MMBtu at the maximum annual heat input of 11.06E+12 Btu/yr when burning 0.7 percent sulfur coal at a 16 percent annual capacity factor, with the remaining heat input due to biomass. It should be noted that burning 0.05 percent sulfur fuel oil at a 25 percent capacity factor results in only 68 TPY of SO_2 emissions and, therefore, Okeelanta wants to retain its ability to burn this fuel up to 25 percent capacity factor. Okeelanta Power is willing to accept a permit condition of 1,154 TPY SO_2 based on a 12-month rolling average.

2. Osceola Power--353.2 TPY of SO_2. This is equivalent to 0.10 lb/MMBtu at the maximum annual heat input of 6.88E+12 Btu/yr when burning 0.7 percent sulfur coal at a 7 percent annual capacity factor, with the remaining heat input due to biomass. It should be noted that burning of 0.05 percent sulfur fuel oil at a 25 percent capacity factor results in only 41 TPY of SO_2 emissions and, therefore, Osceola wants to retain the ability to burn this fuel up to 25 percent

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capacity factor. Osceola Power is willing to accept a permit limit of 353 TPY SO_2 based on a 12-month rolling average.

The applicants are willing to accept these limitations as permit conditions if this would satisfy the Department's concerns and allow the permits to be issued promptly.

Overall SO_2 Limits

The revised conditions for the two facilities result in combined SO_2 emissions of 1,507 TPY. At a maximum heat input for the two facilities of 17.940×10^{12} Btu/yr (11.06×10^{12} Btu/yr for Okeelanta and 6.880×10^{12} Btu/yr for Osceola) when burning coal, the equivalent overall SO_2 emission rate is 0.168 lb/MMBtu. This emission rate is very close to what has been determined recently to be BACT for SO_2 for 100 percent coal-fired power plants (i.e., 0.17 lb/MMBtu for Bechtel Indiantown and OUC Stanton Unit 2).

Revised Tables

Based on the revised coal burning rates and SO_2 emissions, revised tables for the PSD applications have been prepared. These tables present the revised annual heat input rates and the revised emissions and emission limits for the two facilities.

Based on the revised SO_2 emission limits, the economic impact analysis included in the BACT evaluation has been revised. Since the previous BACT analysis has shown that burning coal with a sulfur content greater than 0.7 percent is not cost effective, the revised analysis was conducted for the 0.7 percent sulfur coal case only. The technologies previously analyzed were retained: dry sorbent injection, dry scrubbing and wet scrubbing. The revised BACT analysis is presented in the attached tables. The analysis demonstrates the most cost effective control technology is dry scrubbing, and results in a total cost effectiveness of $4,994/ton for Okeelanta Power and $10,487/ton for Osceola Power. These cost effectiveness values are well above the $2,000/ton value which has been considered to be reasonable in previous BACT determinations. Consequently, the use of dry scrubbing is not reasonable or appropriate as BACT in this case.

We appreciate meeting with you and the Bureau's staff to resolve the Department's concerns. Please call me at 904-331-9000 if you have any questions concerning this information.

Sincerely,

David A. Buff

David A. Buff, M.E., P.E.
Principal Engineer

cc: Gus Cepero, Okeelanta
David Dee, Carlton-Fields
Jewell Harper, EPA
John Bunyak, NPS
Bevin Beaudet, PBCHU
File (2)

OKEELANTA POWER

Table 2-2. Maximum Fuel Usage and Heat Input Rates, Okeelanta Power Limited Partnership Facility (Revised 05/18/93)

Fuel	Heat Input	Heat Transfer Efficiency (%)	Heat Output	Fuel Firing Rate
	Maximum Short-Term (per boiler)			
	(MMBtu/hr)		(MMBtu/hr)	
Biomass	715	68	486	168,236 lb/hr[a]
No. 2 Oil	490	85	417	3,551 gal/hr
Coal	490	85	417	40,833 lb/hr
	Annual Average (total all three boilers)			
	(Btu/yr)		(Btu/yr)	
NORMAL OPERATIONS				
Biomass	1.150E+13	68	7.820E+12	1,352,941 TPY[a]
No. 2 Oil	0	85	0	0 gal/yr
Coal	0	85	0	0 TPY
TOTAL	1.150E+13		7.820E+12	
25% OIL FIRING				
Biomass	8.118E+12	68	5.520E+12	955,059 TPY[a]
No. 2 Oil	2.706E+12	85	2.300E+12	19,608,696 gal/yr
Coal	0	85	0	0 TPY
TOTAL	1.082E+13		7.820E+12	
16% COAL FIRING				
Biomass	9.288E+12	68	6.316E+12	1,092,706 TPY[a]
No. 2 Oil	0	85	0	0 gal/yr
Coal	1.769E+12	85	1.504E+12	73,714 TPY
TOTAL	1.106E+13		7.820E+12	

Note: Total heat output required = 486 MMBtu/hr each boiler, and 7.820E+12 Btu/yr total all boilers. Fuels may be burned in combination, not to exceed indicated total heat outputs.

[a] Based on heating value for bagasse of 4,250 Btu/lb, wet basis.

Table 2-4. Proposed Emission Limits for the Okeelanta Power Facility (revised 05/18/93)

Pollutant	Emission Limit (lb/MMBtu)		
	Biomass	No.2 Oil	Bit. Coal
Particulate (TSP)	0.03	0.03	0.03
Particulate (PM10)	0.03	0.03	0.03
Sulfur Dioxide			
24-hour average	0.10	0.05	1.2
Annual average[a]	0.02	0.05	1.2
Nitrogen Oxides			
Annual average[a]	0.15	0.15	0.17
Carbon Monoxide			
8-hour average	0.35	0.2	0.2
Volatile Organic Compounds	0.06	0.03	0.03
Lead	2.5E-05	8.9E-07	6.4E-05
Mercury	[b]	2.4E-06	8.4E-06
Beryllium	--	3.5E-07	5.9E-06
Fluorides	--	6.3E-06	0.024
Sulfuric Acid Mist	0.003	0.0015	0.036

[a] Compliance based on 30-day rolling average, per 40 CFR 60, Subpart Da.

[b] Limits are $6.3\text{x}10^{-6}$ lb/MM Btu for bagasse and $0.29\text{x}10^{-6}$ lb/MM Btu for wood waste materials.

Table 2-5. Maximum Short-Term Emissions for the Okeelanta Power Cogeneration Facility (per boiler) (Revised 05/18/93)

	Biomass				No. 2 Fuel Oil				Coal				
Regulated Pollutant	Emission Factor (lb/MMBtu)	Ref.	Activity Factor (MMBtu/hr)	Maximum Emissions (lb/hr)	Emission Factor (lb/MMBtu)	Ref.	Activity Factor (MMBtu/hr)	Maximum Emissions (lb/hr)	Emission Factor (lb/MMBtu)	Ref.	Activity Factor (MMBtu/hr)	Maximum Emissions (lb/hr)	Maximum Emissions for any fuel (lb/hr)
Particulate (TSP)	0.03	1	715	21.5	0.03	1	490	14.7	0.03	1	490	14.7	21.5
Particulate (PM10)	0.03	1	715	21.5	0.03	1	490	14.7	0.03	1	490	14.7	21.5
Sulfur dioxide	0.10[a]	2	715	71.5[a]	0.05[a]	9	490	24.5[a]	1.2[a]	1	490	588.0[a]	588.0[a]
Nitrogen oxides	0.15[b]	3	715	107.3[b]	0.15[b]	3	490	73.5[b]	0.17[b]	3	490	83.3[b]	107.3[b]
Carbon monoxide	0.35[c]	4	715	250.3[c]	0.2[c]	4	490	98.0[c]	0.2[c]	4	490	98.0[c]	250.3[c]
VOC	0.06	4	715	42.9	0.03	4	490	14.7	0.03	4	490	14.7	42.9
Lead	2.5E-05	5	715	0.018	8.9E-07	10	490	0.0004	6.4E-05	12	490	0.031	0.031
Mercury	6.3E-06	6	715	0.0045	2.4E-06	11	490	0.00118	8.4E-06	13	490	0.0041	0.0045
Beryllium	--	7	715	--	3.5E-07	12	490	0.00017	5.9E-06	12	490	0.0029	0.0029
Fluorides	--	--	--	--	6.3E-06	14	490	0.003	0.024	14	490	11.8	11.8
Sulfuric acid mist	0.003	8	715	2.15	0.0015	8	490	0.74	0.036	8	490	17.6	17.6
Total reduced sulfur	--	--	--	--	--	--	--	--	--	--	--	--	--
Asbestos	--	--	--	--	--	--	--	--	--	--	--	--	--
Vinyl Chloride	--	--	--	--	--	--	--	--	--	--	--	--	--

[a] 24-hour average.

[b] 30-day rolling average.

[c] 8-hour average.

References:

1. Emission Factor based on NSPS 40CFR 60 Subpart Da.
2. Based upon maximum sulfur content of bagasse of 0.1%, dry basis (0.048%, wet basis).
3. Based on NO_x control system.
4. Based on boiler design.
5. No data available for bagasse; based on testing on wood fired boilers in California (Sassenrath, 1991).
6. Based on source testing at Okeelanta and Osceola, and 30% removal for mercury control system.
7. Emission Tests for Seminole Kraft (1990) and TAPPI Proceedings (1991).
8. Based on AP-42; 3% of SO_2 emissions.
9. Based on maximum sulfur content of No. 2 fuel oil.
10. Toxic Air Emission Factors, EPA, 1988 (EPA-450/2-88-006a).
11. Toxic Air Emission Factors, EPA, 1988 (EPA-450/2-88-006a), using 30% removal from mercury control system.
12. Estimating Air Toxic Emissions from Coal and Oil Combustion Sources (EPA -450/2-89-001) (1989).
13. Based on "Mercury Emissions to the Atmosphere in Florida" (KBN, 1992), and 30% removal from mercury and ESP control system.
14. Based on "Emissions Assessment of Conventional Stationary Combustion Sources: Volume V: Industrial Combustion Sources (EPA-600/7-81-003c).

Table 2-6. Maximum Annual Emissions for the Okeelanta Power Limited Partnership Facility (total all three boilers) (Revised 05/18/93)

	Biomass			No. 2 Fuel Oil			Coal			
Regulated Pollutant	Emission Factor (lb/MMBtu)	Activity Factor (E12 Btu/yr)	Annual Emissions (TPY)	Emission Factor (lb/MMBtu)	Activity Factor (E12 Btu/yr)	Annual Emissions (TPY)	Emission Factor (lb/MMBtu)	Activity Factor (E12 Btu/yr)	Annual Emissions (TPY)	Total Annual Emissions (TPY)
Normal Operations										
Particulate (TSP)	0.03	11.500	172.50	–	–	–	–	–	–	172.50 [a]
Particulate (PM10)	0.03	11.500	172.50	–	–	–	–	–	–	172.50 [a]
Sulfur dioxide	0.02	11.500	115.00	–	–	–	–	–	–	115.00
Nitrogen oxides	0.15	11.500	862.50	–	–	–	–	–	–	862.50 [a]
Carbon monoxide	0.35	11.500	2,012.50	–	–	–	–	–	–	2,012.50 [a]
VOC	0.06	11.500	345.00	–	–	–	–	–	–	345.00 [a]
Lead	2.5E-05	11.500	0.14	–	–	–	–	–	–	0.14
Mercury	b	11.500	b	–	–	–	–	–	–	0.0300
Beryllium	–	–	–	–	–	–	–	–	–	–
Fluorides	–	–	–	–	–	–	–	–	–	–
Sulfuric acid mist	0.00060	11.500	3.45	–	–	–	–	–	–	3.45
Total reduced sulfur	–	–	–	–	–	–	–	–	–	–
Asbestos	–	–	–	–	–	–	–	–	–	–
Vinyl Chloride	–	–	–	–	–	–	–	–	–	–
25% Oil Firing										
Particulate (TSP)	0.03	8.118	121.77	0.03	2.706	40.59	–	–	–	162.36
Particulate (PM10)	0.03	8.118	121.77	0.03	2.706	40.59	–	–	–	162.36
Sulfur dioxide	0.02	8.118	81.18	0.05	2.706	67.65	–	–	–	148.83
Nitrogen oxides	0.15	8.118	608.85	0.15	2.706	202.95	–	–	–	811.80
Carbon monoxide	0.35	8.118	1,420.65	0.2	2.706	270.60	–	–	–	1,691.25
VOC	0.06	8.118	243.54	0.03	2.706	40.59	–	–	–	284.13
Lead	2.5E-05	8.118	0.10	8.9E-07	2.706	0.001	–	–	–	0.10
Mercury	b	8.118	b	b	2.706	b	–	–	–	0.0300
Beryllium	–	–	–	3.5E-07	2.706	0.0005	–	–	–	0.00047
Fluorides	–	–	–	6.27E-06	2.706	0.0085	–	–	–	0.0085
Sulfuric acid mist	0.00060	8.118	2.44	0.0015	2.706	2.03	–	–	–	4.46
Total reduced sulfur	–	–	–	–	–	–	–	–	–	–
Asbestos	–	–	–	–	–	–	–	–	–	–
Vinyl Chloride	–	–	–	–	–	–	–	–	–	–
16% Coal Firing										
Particulate (TSP)	0.03	9.288	139.32	–	–	–	0.03	1.769	26.54	165.86
Particulate (PM10)	0.03	9.288	139.32	–	–	–	0.03	1.769	26.54	165.86
Sulfur dioxide	0.02	9.288	92.88	–	–	–	1.2	1.769	1,061.40	1,154.28 [a]
Nitrogen oxides	0.15	9.288	696.60	–	–	–	0.17	1.769	150.37	846.97
Carbon monoxide	0.35	9.288	1,625.40	–	–	–	0.2	1.769	176.90	1,802.30
VOC	0.06	9.288	278.64	–	–	–	0.03	1.769	26.54	305.18
Lead	2.5E-05	9.288	0.12	–	–	–	6.4E-05	1.769	0.06	0.17 [a]
Mercury	b	9.288	b	–	–	–	b	1.769	b	0.0300 [a]
Beryllium	–	–	–	–	–	–	5.9E-06	1.769	0.0052	0.0052 [a]
Fluorides	–	–	–	–	–	–	0.024	1.769	21.23	21.23 [a]
Sulfuric acid mist	0.00060	9.288	2.79	–	–	–	0.036	1.769	31.84	34.63 [a]
Total reduced sulfur	–	–	–	–	–	–	–	–	–	–
Asbestos	–	–	–	–	–	–	–	–	–	–
Vinyl Chloride	–	–	–	–	–	–	–	–	–	–

[a] Indicates maximum annual emission rate.
[b] Refer to text for explanation.

Table 2-8. Okeelanta Power Cogeneration Facility Maximum Annual PM Emission Rates for Fugitive Dust Sources (Revised 05/18/93)

Source	Uncontrolled Emission Factor (lb/ton)	Control	Control Efficiency (%)	Controlled Emission Factor (lb/ton)	Maximum Annual Thruput (tons/yr)	Maximum Annual PM(TSP) Emissions (tons/yr)	PM10 Size Mult.	Maximum Annual PM10 Emissions (tons/yr)
Coal Handling								
Railcar Unloading	0.00234	Enclosure	70	0.00070	73,714	0.026	0.35	0.009
Conveyor-to-Coal Pile	0.00234	None	0	0.00234	73,714	0.086	0.35	0.030
Reclaim Hopper	0.00234	Enclosure	90	0.00023	73,714	0.009	0.35	0.003
Conveyor-to-Crusher	0.00234	None	0	0.00234	73,714	0.086	0.35	0.030
Coal Crusher	0.02	Enclosure	70	0.00600	73,714	0.221	0.45	0.100
Crusher-to-Conveyor	0.00234	None	0	0.00234	73,714	0.086	0.35	0.030
Conveyor-to-Boiler Silo	0.00234	None	0	0.00234	73,714	0.086	0.35	0.030
Storage Pile	–	None	0	–	–	0.211 [a]	0.5	0.105 a
Coal Storage Pile Maintenance	0.90328	Watering	50	0.45164 b	14,600 [c]	3.297	0.35	1.154
Biomass Handling								
Truck Dump	0.00012	None	0	0.00012	1,352,941	0.083	0.35	0.029
Conveyor-to-Conveyor	0.00012	None	0	0.00012	1,352,941	0.083	0.35	0.029
Conveyor-to-Hog Tower	0.00012	None	0	0.00012	1,352,941	0.083	0.35	0.029
Hogger	0.02	Enclosed	95	0.00100	1,352,941	0.676	0.35	0.237
Hogger-to-Conveyor	0.00012	None	0	0.00012	1,352,941	0.083	0.35	0.029
Transfer Tower	0.00012	None	0	0.00012	1,352,941	0.083	0.35	0.029
Conveyor-to-Stacker	0.00012	None	0	0.00012	1,352,941	0.083	0.35	0.029
Stacking	0.00012	None	0	0.00012	1,352,941	0.083	0.35	0.029
Underpile Reclaim	0.00012	Enclosed	90	0.00001	1,352,941	0.008	0.35	0.003
Reclaimer-to-Conveyor	0.00012	None	0	0.00012	1,352,941	0.083	0.35	0.029
Transfer Tower	0.00012	None	0	0.00012	1,352,941	0.083	0.35	0.029
Conveyor-to-Boiler Feeders	0.00012	None	0	0.00012	1,352,941	0.083	0.35	0.029
Biomass Storage Pile	–	None	0	–	–	0.160	0.5	0.080
Biomass Storage Pile Maintenance	0.90328[b]	Watering	50	0.45164[b]	21,900[c]	4.945	0.35	1.731
Fly Ash Handling								
Fly Ash Transfer	0.00727	Enclosure or Watering	50	0.00364	43,294[d]	0.079	0.35	0.028
	TOTAL					10.804		3.859

[a] Refer to Appendix A and text for derivation.
[b] lb/VMT.
[c] Vehicle miles traveled per year.
[d] 1,352,941 TPY biomass at 3.20 percent ash; assumes all ash is flyash.

Table 2-10. Maximum Hourly Emissions of Non-Regulated Pollutants for the Okeelanta Power Cogeneration Facility (per boiler) (Revised 11/25/92)

	Biomass				No. 2 Fuel Oil				Coal				Maximum
Non Regulated Pollutant	Emission Factor (lb/MMBtu)	Ref	Activity Factor (MMBtu/hr)	Hourly Emissions (lb/hr)	Emission Factor (lb/MMBtu)	Ref	Activity Factor (MMBtu/hr)	Hourly Emissions (lb/hr)	Emission Factor (lb/MMBtu)	Ref	Activity Factor (MMBtu/hr)	Hourly Emissions (lb/hr)	Hourly Emission[a] (lb/hr)
Ammonia	0.0148	8	715	10.6	0.0148	8	490	7.3	0.048	8	490	23.6	23.6
Antimony	UD	3	715	--	2.32E-06	5	490	0.0011	3.49E-05	5	490	0.017	0.017
Arsenic	1.62E-04	10	715	0.116	5.00E-07	1	490	0.0002	2.64E-05	4	490	0.013	0.116
Barium	1.06E-04	3	715	0.076	6.69E-06	5	490	0.0033	7.44E-04	5	490	0.36	0.36
Bromine	1.47E-03	7	715	1.05	6.97E-06	5	490	0.00342	7.90E-04	5	490	0.387	1.05
Cadmium	5.43E-06	2	715	0.0039	1.58E-06	1	490	0.0008	1.36E-06	4	490	0.001	0.0039
Chromium	1.54E-04	10	715	0.110	1.39E-05	1	490	0.0068	1.66E-05	4	490	0.008	0.110
Chromium^{+6}	3.81E-05	9	715	0.027	2.78E-06	9	490	0.0014	3.32E-06	9	490	0.002	0.027
Cobalt	4.98E-04	7	715	0.356	1.17E-05	5	490	0.0058	7.20E-05	5	490	0.035	0.356
Copper	1.45E-04	10	715	0.104	4.20E-05	1	490	0.021	1.71E-04	4	490	0.084	0.104
Dioxin	6.93E-12	2	715	5.0E-09	--		490	--	--		490	--	5.5E-09
Furan	3.62E-10	2	715	2.6E-07	--		490	--	--		490	--	2.6E-07
Formaldehyde	6.56E-04	2	715	0.469	4.05E-04	1	490	0.20	2.20E-04	4	490	0.108	0.47
Hydrogen Chloride	3.70E-02	3	715	26.5	6.37E-04	6	490	0.312	7.90E-02	6	490	38.7	38.7
Indium	1.27E-04	7	715	0.091	--		490	--	--		490	--	0.091
Manganese	7.98E-04	2	715	0.57	3.08E-06	1	490	0.0015	3.10E-05	4	490	0.015	0.57
Molybdenum	2.54E-04	7	715	0.18	4.88E-06	5	490	0.0024	8.83E-05	5	490	0.043	0.18
Nickel	4.41E-05	2	715	0.032	4.76E-05	1	490	0.023	1.02E-03	4	490	0.50	0.50
Phosphorus	3.53E-04	3	715	0.25	5.81E-06	5	490	0.0028	8.60E-04	5	490	0.42	0.42
Selenium	UD	3	715	--	4.60E-06	1	490	0.0023	5.34E-05	5	490	0.026	0.026
Silver	2.94E-05	3	715	0.021	--		490	--	--		490	--	0.021
Thallium	UD	3	715	--	--		490	--	--		490	--	--
Tin	1.62E-04	7	715	0.12	3.30E-05	5	490	0.016	8.83E-05	5	490	0.043	0.12
Zinc	4.24E-04	2	715	0.30	6.69E-06	5	490	0.0033	3.49E-04	5	490	0.17	0.30
Zirconium	9.29E-05	7	715	0.066	--		490	--	--		490	--	0.066

Note: UD = undetectable levels in gas stream.

[a] Denotes maximum for any fuel.

References
1: Toxic Air Pollutant Emission Factors - A Compilation for Selected Air Toxic Compounds and Sources, Second Edition EPA-450/2-90-011 (1990).
2: Based on "Air Toxic Emissions from Wood Fired Boilers", C. Sassenrath, 1991 TAPPI Proceedings.
3: Based on stack test results of wood fired boilers and fuel analysis at Seminole Kraft Corporation (1990) equipped with wet scrubbers.
4: Estimating Emissions from Oil and Coal Combustion Sources EPA-450/2-89-001 (1989).
5: Emissions Assessment of Conventional Stationary Combustion Systems Volume V, 1981. Based on an uncontrolled spreader stoker design and then assuming 90% control from ESP.
6: Emissions Assessment of Conventional Stationary Combustion Systems Volume V, 1981. Based on an uncontrolled spreader stoker design.
7: EPA PM/VOC Speciation Database, updated October, 1989.
8: Based on maximum 20 ppm NH_3 in exhaust gases for biomass and No. 2 fuel oil; 65 ppm for coal.
9: Based upon stack test data at Dade County RRF, 1992, which indicated less than 20% of total chromium was chromium^{+6}.
10: Same as reference 2; includes 3% treated wood burning.

Source: KBN, 1992.

Table 2-11. Maximum Annual Emissions of Non-Regulated Pollutants for the Okeelanta Power Cogeneration Facility (total all boilers) (Page 1 of 2) (Revised 05/18/93)

Non Regulated Pollutant	Biomass			No. 2 Fuel Oil			Coal			Total Annual Emission (TPY)
	Emission Factor (lb/MMBtu)	Activity Factor (E12 Btu/yr)	Annual Emissions (TPY)	Emission Factor (lb/MMBtu)	Activity Factor (E12 Btu/yr)	Annual Emissions (TPY)	Emission Factor (lb/MMBtu)	Activity Factor (E12 Btu/yr)	Annual Emissions (TPY)	
Normal Operations										
Ammonia	0.0148	11.500	85.1	–	–	–	–	–	–	85.1
Antimony	UD	11.500	–	–	–	–	–	–	–	–
Arsenic	5.58E-05	11.500	0.32	–	–	–	–	–	–	0.32 [a]
Barium	1.06E-04	11.500	0.61	–	–	–	–	–	–	0.61
Bromine	1.47E-03	11.500	8.45	–	–	–	–	–	–	8.5 [a]
Cadmium	5.43E-06	11.500	0.031	–	–	–	–	–	–	0.031 [a]
Chromium	5.54E-05	11.500	0.32	–	–	–	–	–	–	0.32 [a]
Chromium^{+6}	1.35E-05	11.500	0.078	–	–	–	–	–	–	0.078 [a]
Cobalt	4.98E-04	11.500	2.86	–	–	–	–	–	–	2.86 [a]
Copper	7.23E-05	11.500	0.42	–	–	–	–	–	–	0.42
Dioxin	6.93E-12	11.500	4.0E-08	–	–	–	–	–	–	4.0E-08 [a]
Furan	3.62E-10	11.500	2.1E-06	–	–	–	–	–	–	2.1E-06 [a]
Formaldehyde	6.56E-04	11.500	3.77	–	–	–	–	–	–	3.8 [a]
Hydrogen Chloride	3.70E-02	11.500	212.75	–	–	–	–	–	–	212.8
Indium	1.27E-04	11.500	0.73	–	–	–	–	–	–	0.73 [a]
Manganese	7.98E-04	11.500	4.59	–	–	–	–	–	–	4.6 [a]
Molybdenum	2.54E-04	11.500	1.46	–	–	–	–	–	–	1.5 [a]
Nickel	4.41E-05	11.500	0.25	–	–	–	–	–	–	0.25
Phosphorus	3.53E-04	11.500	2.03	–	–	–	–	–	–	2.03
Selenium	UD	11.500	–	–	–	–	–	–	–	–
Silver	2.94E-05	11.500	0.169	–	–	–	–	–	–	0.169 [a]
Thallium	UD	11.500	–	–	–	–	–	–	–	–
Tin	1.62E-04	11.500	0.93	–	–	–	–	–	–	0.93 [a]
Zinc	4.24E-04	11.500	2.44	–	–	–	–	–	–	2.44 [a]
Zirconium	9.29E-05	11.500	0.53	–	–	–	–	–	–	0.53 [a]
25% Oil Firing										
Ammonia	0.0148	8.118	60.1	0.0148	2.706	20.02	–	–	–	80.10
Antimony	UD	8.118	–	2.32E-06	2.706	0.0031	–	–	–	0.0031
Arsenic	5.58E-05	8.118	0.23	5.00E-07	2.706	0.0007	–	–	–	0.23
Barium	1.06E-04	8.118	0.43	6.69E-06	2.706	0.0091	–	–	–	0.44
Bromine	1.47E-03	8.118	5.967	6.97E-06	2.706	0.0094	–	–	–	5.976
Cadmium	5.43E-06	8.118	0.022	1.58E-06	2.706	0.0021	–	–	–	0.024
Chromium	5.54E-05	8.118	0.22	1.39E-05	2.706	0.0188	–	–	–	0.24
Chromium +6	1.35E-05	8.118	0.055	2.78E-06	2.706	0.0038	–	–	–	0.059
Cobalt	4.98E-04	8.118	2.02	1.17E-05	2.706	0.0159	–	–	–	2.04
Copper	7.23E-05	8.118	0.29	4.20E-05	2.706	0.0568	–	–	–	0.35
Dioxin	6.93E-12	8.118	2.8E-08	–	2.706	–	–	–	–	2.8E-08
Furan	3.62E-10	8.118	1.5E-06	–	2.706	–	–	–	–	1.5E-06
Formaldehyde	6.56E-04	8.118	2.7	4.05E-04	2.706	0.55	–	–	–	3.21
Hydrogen Chloride	3.70E-02	8.118	150.18	6.37E-04	2.706	0.8616	–	–	–	151.04

Table 2-11. Maximum Annual Emissions of Non-Regulated Pollutants for the Okeelanta Power Cogeneration Facility (total all boilers) (Page 2 of 2) (Revised 05/18/93)

	Biomass			No. 2 Fuel Oil			Coal			Total
Non Regulated Pollutant	Emission Factor (lb/MMBtu)	Activity Factor (E12 Btu/yr)	Annual Emissions (TPY)	Emission Factor (lb/MMBtu)	Activity Factor (E12 Btu/yr)	Annual Emissions (TPY)	Emission Factor (lb/MMBtu)	Activity Factor (E12 Btu/yr)	Annual Emissions (TPY)	Annual Emission (TPY)
Indium	1.27E-04	8.118	0.52	–	2.706	–	–	–	–	0.52
Manganese	7.98E-04	8.118	3.24	3.08E-06	2.706	0.0042	–	–	–	3.2
Molybdenum	2.54E-04	8.118	1.03	4.88E-06	2.706	0.0066	–	–	–	1.0
Nickel	4.41E-05	8.118	0.18	4.76E-05	2.706	0.0644	–	–	–	0.24
Phosphorus	3.53E-04	8.118	1.43	5.81E-06	2.706	0.0079	–	–	–	1.44
Selenium	UD	8.118	–	4.60E-06	2.706	0.0062	–	–	–	0.0062
Silver	2.94E-05	8.118	0.119	–	2.706	–	–	–	–	0.119
Thallium	UD	8.118	–	–	2.706	–	–	–	–	–
Tin	1.62E-04	8.118	0.66	3.30E-05	2.706	0.045	–	–	–	0.70
Zinc	4.24E-04	8.118	1.72	6.69E-06	2.706	0.0091	–	–	–	1.7
Zirconium	9.29E-05	8.118	0.38	–	2.706	–	–	–	–	0.38
<u>16% Coal Firing</u>										
Ammonia	1.48E-02	9.288	68.73	–	–	–	0.048	1.769	42.46	111.19 [a]
Antimony	UD	9.288	–	–	–	–	3.49E-05	1.769	0.031	0.031 [a]
Arsenic	5.58E-05	9.288	0.26	–	–	–	2.64E-05	1.769	0.023	0.28
Barium	1.06E-04	9.288	0.49	–	–	–	7.44E-04	1.769	0.66	1.15 [a]
Bromine	1.47E-03	9.288	6.83	–	–	–	7.90E-04	1.769	0.699	7.53
Cadmium	5.43E-06	9.288	0.025	–	–	–	1.36E-06	1.769	0.0012	0.026
Chromium	5.54E-05	9.288	0.257	–	–	–	1.66E-05	1.769	0.015	0.272
Chromium +6	1.35E-05	9.288	0.063	–	–	–	3.32E-06	1.769	0.003	0.066
Cobalt	4.98E-04	9.288	2.31	–	–	–	7.20E-05	1.769	0.064	2.4
Copper	7.23E-05	9.288	0.34	–	–	–	1.71E-04	1.769	0.15	0.49 [a]
Dioxin	6.93E-12	9.288	3.2E-08	–	–	–	–	1.769	–	3.2E-08
Furan	3.62E-10	9.288	1.7E-06	–	–	–	–	1.769	–	1.7E-06
Formaldehyde	6.56E-04	9.288	3.0	–	–	–	2.20E-04	1.769	0.19	3.24
Hydrogen Chloride	3.70E-02	9.288	171.828	–	–	–	7.90E-02	1.769	69.88	241.7 [a]
Indium	1.27E-04	9.288	0.59	–	–	–	–	1.769	–	0.59
Manganese	7.98E-04	9.288	3.71	–	–	–	3.10E-05	1.769	0.027	3.7
Molybdenum	2.54E-04	9.288	1.18	–	–	–	8.83E-05	1.769	0.078	1.3
Nickel	4.41E-05	9.288	0.21	–	–	–	1.02E-03	1.769	0.90	1.11 [a]
Phosphorus	3.53E-04	9.288	1.64	–	–	–	8.60E-04	1.769	0.76	2.40 [a]
Selenium	UD	9.288	–	–	–	–	5.34E-05	1.769	0.047	0.047 [a]
Silver	2.94E-05	9.288	0.137	–	–	–	–	1.769	–	0.137
Thallium	UD	9.288	–	–	–	–	–	1.769	–	–
Tin	1.62E-04	9.288	0.75	–	–	–	8.83E-05	1.769	0.078	0.83
Zinc	4.24E-04	9.288	1.97	–	–	–	3.49E-04	1.769	0.31	2.3
Zirconium	9.29E-05	9.288	0.43	–	–	–	–	1.769	–	0.43

Note: UD = undetectable levels in gas stream.

[a] Denotes maximum annual emissions for any fuel scenario.

Table 3-3. PSD Source Applicability Analysis for the Okeelanta Power Limited Partnership Facility (revised 05/18/93)

Regulated Pollutant	Baseline Emissions (TPY)	Cogeneration Facility Annual Emissions (TPY)	Net Change (TPY)	Significant Emission Rate (TPY)	PSD Applies ?
Particulate (TSP)	473.7	183.3[b]	-290.4	25	No
Particulate (PM10)	426.3	176.4[c]	-249.9	15	No
Sulfur Dioxide	748.3	1,154.3	406.0	40	Yes
Nitrogen Oxides	888.7	862.5	-26.2	40	No
Carbon Monoxide	10,388.0	2,012.5	-8,375.5	100	No
VOC	401.9	345.0	-56.9	40	No[a]
Lead	0.28	0.17	-0.11	0.6	No
Mercury	0.0292[d]	0.0300	0.0008	0.1	No
Beryllium	0.0004	0.0052	0.0048	0.0004	Yes
Fluorides	0.04	21.2	21.2	3	Yes
Sulfuric Acid Mist	22.4	34.6	6.4	7	No
Total Reduced Sulfur	--	--	0	10	No
Asbestos	--	--	0	0.007	No
Vinyl Chloride	--	--	0	0	No

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[a] Nonattainment review does not apply since there is no increase in VOC emissions.

[b] Includes 172.5 TPY from boilers and 10.8 TPY from fugitive dust sources.

[c] Includes 172.5 TPY from boilers and 3.9 TPY from fugitive dust sources.

[d] The estimated average annual emission rate for the 1990-1991 and 1991-1992 crop years is 0.0292 TPY. The highest annual emission rate for either of these years is 0.0300 TPY.

Table 6-18. Maximum Impacts of Toxic Pollutants for Okeelanta Power Cogeneration Facility (total all boilers) (Revised 05/18/93)

Pollutant	Maximum Hourly Emissions[a] (lb/hr)	Concentrations ($\mu g/m^3$)					
		8-Hour		24-Hour		Annual	
		Impact	NTL	Impact	NTL	Impact	NTL
Ammonia	70.8	3.9	180	3.0	43.2	–	–
Antimony	0.051	0.0028	5	0.002	1.2	0.0002	0.3
Arsenic	0.35	0.0163	2	0.01	0.48	0.000226 [b]	0.000230
Barium	1.08	0.0594	5	0.05	1.2	0.0033	50
Beryllium	0.0087	0.0005	0.02	0.0004	0.0048	0.00003	0.00042
Bromine	3.15	0.15	7	0.11	1.68	–	–
Cadmium	0.012	0.0005	0.5	0.0004	0.12	0.00003	0.00056
Chromium metals	0.33	0.0154	5	0.012	1.2	0.00087	1000
Chromium^{+6}	0.081	0.0041	0.5	0.003	0.12	0.000059 [c]	0.000083
Cobalt	1.07	0.05	0.5	0.04	0.12	–	–
Copper	0.31	0.01	10	0.01	2.4	–	–
Dioxins/Furans	8.0E-07	–	–	–	–	2.1E-09	2.2E-08
Fluoride	35.4	1.95	25	1.48	6	–	–
Formaldehyde	1.41	0.07	4.5	0.05	1.08	0.004	0.077
Hydrogen Chloride	116.1	6.39	70	4.84	16.8	0.360	7.0
Indium	0.27	0.01	1	0.01	0.24	–	–
Manganese	1.71	0.08	50	0.06	12	–	–
Mercury	0.0135	0.0007	0.5	0.0006	0.12	0.00004	0.3
Molybdenum	0.54	0.03	50	0.02	12	–	–
Nickel	1.50	0.08	0.5	0.06	0.12	0.0011 [d]	0.0042
Phosphorus	1.26	0.07	1	0.05	0.24	–	–
Selenium	0.08	0.004	2	0.003	0.48	–	–
Silver	0.06	0.003	0.1	0.002	0.024	0.0002	3
Sulfuric Acid Mist	52.8	2.9	10.0	2.2	2.4	–	–
Thallium	–	–	–	–	–	–	–
Tin	0.36	0.02	1	0.01	0.24	–	–
Zinc	0.90	0.04	10	0.03	2.4	–	–
Zirconium	0.20	0.009	50	0.01	12	–	–

Note: NTL = no-threat level.
Maximum concentrations determined with ISCST2 model and West Palm Beach meteorological data for 1982 to 1986.
Highest predicted concentration ($\mu g/m^3$) for a 10 g/s (79.365 lb/hr) emission rate:

8-hour = 4.369
24-hour = 3.310
Annual = 0.2459

[a] Total all three boilers.
[b] Based on maximum annual average emission rate of 0.32 TPY total all three boilers (avg. of 0.073 lb/hr).
[c] Based on maximum annual average emission rate of 0.078 TPY total all three boilers (avg. of 0.018 lb/hr).
[d] Based on maximum annual average emission rate of 1.56 TPY total all three boilers (avg. of 0.356 lb/hr).

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Table 2-2. Maximum Fuel Usage and Heat Input Rates, Osceola Power Limited Partnership Facility (Revised 05/03/93)

Fuel	Heat Input	Heat Transfer Efficiency (%)	Heat Output	Fuel Firing Rate	
	Maximum Short-Term (per boiler)				
	(MMBtu/hr)		(MMBtu/hr)		
Biomass	665	68	452	156,471	lb/hr*
No. 2 Oil	460	85	391	3,333	gal/hr
Coal	460	85	391	38,333	lb/hr
	Annual Average (total two boilers)				
	(Btu/yr)		(Btu/yr)		
NORMAL OPERATIONS					
Biomass	7.000E+12	68	4.760E+12	823,529	TPY*
No. 2 Oil	0	85	0	0	gal/yr
Coal	0	85	0	0	TPY
	----------		----------		
TOTAL	7.000E+12		4.760E+12		
25% OIL FIRING					
Biomass	4.941E+12	68	3.360E+12	581,294	TPY*
No. 2 Oil	1.647E+12	85	1.400E+12	11,934,783	gal/yr
Coal	0	85	0	0	TPY
	----------		----------		
TOTAL	6.588E+12		4.760E+12		
7% COAL FIRING					
Biomass	6.398E+12	68	4.351E+12	752,706	TPY*
No. 2 Oil	0	85	0	0	gal/yr
Coal	4.816E+11	85	4.093E+11	20,065	TPY
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TOTAL	6.880E+12		4.760E+12		

Notes: Total heat output required = 452 MMBtu/hr each boiler and 4.760E+12 Btu/yr total both boilers. Fuels may be burned in combination, not to exceed indicated total heat outputs.

* Based on heating value for bagasse of 4,250 Btu/lb.

Table 2-4. Proposed Emission Limits for the Osceola Power Facility (Revised 05/18/93)

Pollutant	Emission Limit (lb/MMBtu)		
	Biomass	No.2 Oil	Bit. Coal
Particulate (TSP)	0.03	0.03	0.03
Particulate (PM10)	0.03	0.03	0.03
Sulfur Dioxide			
24-hour average	0.10	0.05	1.2
Annual average[a]	0.02	0.05	1.2
Nitrogen Oxides			
Annual average[a]	0.12	0.12	0.17
Carbon Monoxide			
8-hour average	0.35	0.2	0.2
Volatile Organic Compounds	0.06	0.03	0.03
Lead	2.5E-05	8.9E-07	6.4E-05
Mercury	[b]	2.4E-06	8.4E-06
Beryllium	--	3.5E-07	5.9E-06
Fluorides	--	6.3E-06	0.024
Sulfuric Acid Mist	0.003	0.0015	0.036

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[a] Compliance based on 30-day rolling average, per 40 CFR 60, Subpart Da.

[b] Limits are $6.3\text{x}10^{-6}$ lb/MM Btu for bagasse and $0.29\text{x}10^{-6}$ lb/MM Btu for wood waste materials.

Table 2-5. Maximum Short-Term Emissions for Osceola Power Cogeneration Facility (per boiler) (Revised 05/18/93)

Regulated Pollutant	Biomass				#2 Oil				Coal				Maximum Emissions for any fuel (lb/hr)
	Emission Factor (lb/MMBtu)	Ref.	Activity Factor (MMBtu/hr)	Maximum Emissions (lb/hr)	Emission Factor (lb/MMBtu)	Ref.	Activity Factor (MMBtu/hr)	Maximum Emissions (lb/hr)	Emission Factor (lb/MMBtu)	Ref.	Activity Factor (MMBtu/hr)	Maximum Emissions (lb/hr)	
Particulate (TSP)	0.03	1	665	20.0	0.03	1	460	13.8	0.03	1	460	13.8	20.0
Particulate (PM10)	0.03	1	665	20.0	0.03	1	460	13.8	0.03	1	460	13.8	20.0
Sulfur dioxide	0.10 [a]	2	665	66.5 [a]	0.05	9	460	23.0 [a]	1.2	1	460	552.0 [a]	552.0 [a]
Nitrogen oxides	0.12 [b]	3	665	79.8 [b]	0.12 [b]	3	460	55.2 [b]	0.17 [b]	3	460	78.2 [b]	79.8 [b]
Carbon monoxide	0.35 [c]	4	665	232.8 [c]	0.2 [c]	4	460	92.0 [c]	0.2 [c]	4	460	92.0 [c]	232.8 [c]
VOC	0.06	4	665	39.9	0.03	4	460	13.8	0.03	4	460	13.8	39.9
Lead	2.5E-05	5	665	0.017	8.9E-07	10	460	0.0004	6.4E-05	12	460	0.029	0.029
Mercury	6.3E-06	6	665	0.0042	2.4E-06	11	460	0.0011	8.4E-06	13	460	0.0039	0.0042
Beryllium	–	7	–	–	3.5E-07	12	460	0.00016	5.9E-06	12	460	0.0027	0.0027
Fluorides	–	–	–	–	6.3E-06	14	460	0.003	0.024	14	460	11.0	11.0
Sulfuric acid mist	0.003	8	665	2.00	0.0015	8	460	0.7	0.036	8	460	16.6	16.6
Total reduced sulfur	–	–	–	–	–	–	–	–	–	–	–	–	–
Asbestos	–	–	–	–	–	–	–	–	–	–	–	–	–
Vinyl Chloride	–	–	–	–	–	–	–	–	–	–	–	–	–

[a] 24-hour average.

[b] 30-day rolling average.

[c] 8-hour average.

References:

1. Emission Factor based on NSPS 40CFR 60 Subpart Da.
2. Based upon maximum sulfur content of bagasse of 0.1 percent, dry basis (0.048 percent, wet basis).
3. Based on NO_x control system.
4. Based on boiler design.
5. No data available for bagasse; based on testing on wood fired boilers in California (Sassenrath, 1991).
6. Based on mercury content in sugar cane and mercury control system.
7. Emission Tests for Seminole Kraft (1990) and TAPPI Proceedings (1991).
8. Based on AP-42; 3% of SO_2 emissions.
9. Based on maximum sulfur content of No. 2 fuel oil.
10. Toxic Air Emission Factors EPA 1988 (EPA-450/2-88-006a).
11. Toxic Air Emission Factors EPA 1988 (EPA-450/2-88-006a), using 30% removal from mercury control system.
12. Estimating Air Toxic Emissions from Coal and Oil Combustion Sources (EPA -450/2-89-001) (1989).
13. Based on "Mercury Emissions to the Atmosphere in Florida" (KBN, 1992), and 30% removal from mercury and ESP control system.
14. Based on "Emissions Assessment of Conventional Stationary Combustion Sources: Volume V: Industrial Combustion Sources (EPA-600/7-81-003c).

Table 2-6. Maximum Annual Emissions for Osceola Power Limited Partnership Facility (total all boilers) (Revised 05/18/93)

Regulated Pollutant	Biomass			No. 2 Fuel			Coal			Total Annual Emissions (TPY)
	Emission Factor (lb/MMBtu)	Activity Factor (E12 Btu/yr)	Annual Emissions (TPY)	Emission Factor (lb/MMBtu)	Activity Factor (E12 Btu/yr)	Annual Emissions (TPY)	Emission Factor (lb/MMBtu)	Activity Factor (E12 Btu/yr)	Annual Emissions (TPY)	
Normal Operations										
Particulate (TSP)	0.03	7.000	105.00	–	–	–	–	–	–	105.00 [a]
Particulate (PM10)	0.03	7.000	105.00	–	–	–	–	–	–	105.00 [a]
Sulfur dioxide	0.02	7.000	70.00	–	–	–	–	–	–	70.00
Nitrogen oxides	0.12	7.000	420.00	–	–	–	–	–	–	420.00
Carbon monoxide	0.35	7.000	1,225.00	–	–	–	–	–	–	1,225.00 [a]
VOC	0.06	7.000	210.00	–	–	–	–	–	–	210.00 [a]
Lead	2.5E-05	7.000	0.09	–	–	–	–	–	–	0.09
Mercury	[b]	7.000	[b]	–	–	–	–	–	–	0.0161
Beryllium	–	–	–	–	–	–	–	–	–	0.00
Fluorides	–	–	–	–	–	–	–	–	–	0.00
Sulfuric acid mist	0.00060	7.000	2.10	–	–	–	–	–	–	2.10
Total reduced sulfur	–	–	–	–	–	–	–	–	–	–
Asbestos	–	–	–	–	–	–	–	–	–	–
Vinyl Chloride	–	–	–	–	–	–	–	–	–	–
25% Oil Firing										
Particulate (TSP)	0.03	4.941	74.12	0.03	1.647	24.71	–	–	–	98.82
Particulate (PM10)	0.03	4.941	74.12	0.03	1.647	24.71	–	–	–	98.82
Sulfur dioxide	0.02	4.941	49.41	0.05	1.647	41.18	–	–	–	90.59
Nitrogen oxides	0.12	4.941	296.46	0.12	1.647	98.82	–	–	–	395.28
Carbon monoxide	0.35	4.941	864.68	0.2	1.647	164.70	–	–	–	1,029.38
VOC	0.06	4.941	148.23	0.03	1.647	24.71	–	–	–	172.94
Lead	2.5E-05	4.941	0.06	8.9E-07	1.647	0.001	–	–	–	0.06
Mercury	[b]	4.941	[b]	[b]	1.647	[b]	–	–	–	0.0161
Beryllium	–	–	–	3.5E-07	1.647	0.0003	–	–	–	0.00029
Fluorides	–	–	–	6.27E-06	1.647	0.0052	–	–	–	0.005
Sulfuric acid mist	0.00060	4.941	1.48	0.0015	1.647	1.24	–	–	–	2.72
Total reduced sulfur	–	–	–	–	–	–	–	–	–	–
Asbestos	–	–	–	–	–	–	–	–	–	–
Vinyl Chloride	–	–	–	–	–	–	–	–	–	–
7% Coal Firing										
Particulate (TSP)	0.03	6.398	95.97	–	–	–	0.03	0.482	7.23	103.20
Particulate (PM10)	0.03	6.398	95.97	–	–	–	0.03	0.482	7.23	103.20
Sulfur dioxide	0.02	6.398	63.98	–	–	–	1.2	0.482	289.20	353.18 [a]
Nitrogen oxides	0.12	6.398	383.88	–	–	–	0.17	0.482	40.97	424.85 [a]
Carbon monoxide	0.35	6.398	1,119.65	–	–	–	0.2	0.482	48.20	1,167.85
VOC	0.06	6.398	191.94	–	–	–	0.03	0.482	7.23	199.17
Lead	2.5E-05	6.398	0.08	–	–	–	6.4E-05	0.482	0.02	0.10 [a]
Mercury	[b]	6.398	[b]	–	–	–	[b]	0.482	[b]	0.0161 [a]
Beryllium	–	–	–	–	–	–	5.9E-06	0.482	0.0014	0.0014 [a]
Fluorides	–	–	–	–	–	–	0.024	0.482	5.78	5.78 [a]
Sulfuric acid mist	0.00060	6.398	1.92	–	–	–	0.036	0.482	8.68	10.60 [a]
Total reduced sulfur	–	–	–	–	–	–	–	–	–	–
Asbestos	–	–	–	–	–	–	–	–	–	–
Vinyl Chloride	–	–	–	–	–	–	–	–	–	–

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[a] Denotes maximum annual emissions for any fuel scenario.

[b] Refer to text for explanation.

Table 2-8. Osceola Power Cogeneration Facility Maximum Annual PM Emission Rates for Fugitive Dust Sources (Revised 05/18/93)

Source	Uncontrolled Emission Factor (lb/ton)	Control	Control Efficiency (%)	Controlled Emission Factor (lb/ton)	Maximum Annual Thruput (tons/yr)	Maximum Annual PM(TSP) Emissions (tons/yr)	PM10 Size Mult.	Maximum Annual PM10 Emissions (tons/yr)
<u>Coal Handling</u>								
Railcar Unloading	0.00234	Enclosure	70	0.00070	20,065	0.007	0.35	0.002
Conveyor-to-Coal Pile	0.00234	None	0	0.00234	20,065	0.023	0.35	0.008
Reclaim Hopper	0.00234	None	90	0.00023	20,065	0.002	0.35	0.001
Conveyor-to-Crusher	0.00234	None	0	0.00234	20,065	0.023	0.35	0.008
Coal Crusher	0.02	Enclosure	70	0.00600	20,065	0.060	0.45	0.027
Crusher-to-Conveyor	0.00234	None	0	0.00234	20,065	0.023	0.35	0.008
Conveyor-to-Boiler Silo	0.00234	None	0	0.00234	20,065	0.023	0.35	0.008
Storage Pile	–	None	0	–	–	0.211 [a]	0.5	0.105 [a]
Coal Storage Pile Maintenance	0.90328	Watering	50	0.45164 [b]	14,600 [c]	3.297	0.35	1.154
<u>Biomass Handling</u>								
Truck Dump	0.00012	None	0	0.00012	823,529	0.050	0.35	0.018
Conveyor-to-Conveyor	0.00012	None	0	0.00012	823,529	0.050	0.35	0.018
Conveyor-to-Hog Tower	0.00012	None	0	0.00012	823,529	0.050	0.35	0.018
Hogger	0.02	Enclosed	95	0.00100	823,529	0.412	0.35	0.144
Hogger-to-Conveyor	0.00012	None	0	0.00012	823,529	0.050	0.35	0.018
Transfer Tower	0.00012	None	0	0.00012	823,529	0.050	0.35	0.018
Conveyor-to-Stacker	0.00012	None	0	0.00012	823,529	0.050	0.35	0.018
Stacking	0.00012	None	0	0.00012	823,529	0.050	0.35	0.018
Underpile Reclaim	0.00012	Enclosed	90	0.00001	823,529	0.005	0.35	0.002
Reclaimer-to-Conveyor	0.00012	None	0	0.00012	823,529	0.050	0.35	0.018
Transfer Tower	0.00012	None	0	0.00012	823,529	0.050	0.35	0.018
Conveyor-to-Boiler Feeders	0.00012	None	0	0.00012	823,529	0.050	0.35	0.018
Biomass Storage Pile	–	none	0	–	–	0.160	0.5	0.080
Biomass Storage Pile Maintenance	0.90328[b]	Watering	50	0.45164[b]	21,900 [c]	4.945	0.35	1.731
<u>Fly Ash Handling</u>								
Fly Ash Transfer	0.00727	Enclosure or Watering	50	0.00364	26,353 [d]	0.048	0.35	0.017
TOTAL						9.745		3.472

[a] Refer to Appendix A and text for derivation.
[b] lb/VMT
[c] Vehicle miles traveled per year.
[d] 823,529 TPY biomass at 3.20 percent ash; assumes all ash is fly ash.

Table 2-10. Maximum Hourly Emissions of Non-Regulated Pollutants for Osceola Power Cogeneration Facility (per boiler) (Revised 05/18/93)

	Biomass				No. 2 Fuel Oil				Coal				
Non Regulated Pollutant	Emission Factor (lb/MMBtu)	Ref	Activity Factor (MMBtu/hr)	Hourly Emissions (lb/hr)	Emission Factor (lb/MMBtu)	Ref	Activity Factor (MMBtu/hr)	Hourly Emissions (lb/hr)	Emission Factor (lb/MMBtu)	Ref	Activity Factor (MMBtu/hr)	Hourly Emissions (lb/hr)	Maximum Hourly Emission any fuel (lb/hr)
Ammonia	1.5E-02	8	665	10.0	4.8E-02	8	460	22.1	4.8E-02	8	460	22.1	22.1
Antimony	UD	3	665	–	2.32E-06	5	460	0.0011	3.49E-05	5	460	0.016	0.016
Arsenic	1.62E-04	10	665	0.11	5.00E-07	1	460	0.0002	2.64E-05	4	460	0.012	0.11
Barium	1.06E-04	3	665	0.07	6.69E-06	5	460	0.0031	7.44E-04	5	460	0.34	0.34
Bromine	1.47E-03	7	665	0.98	6.97E-06	5	460	0.00321	7.90E-04	5	460	0.363	0.98
Cadmium	5.43E-06	2	665	0.0036	1.58E-06	1	460	0.0007	1.36E-06	4	460	0.001	0.0036
Chromium	1.54E-04	10	665	0.10	1.39E-05	1	460	0.0064	1.66E-05	4	460	0.008	0.10
Chromium +6	3.81E-05	9	665	0.025	2.78E-06	9	460	0.0013	3.32E-06	9	460	0.002	0.025
Cobalt	4.98E-04	7	665	0.33	1.17E-05	5	460	0.0054	7.20E-05	5	460	0.033	0.33
Copper	1.45E-05	10	665	0.0096	4.20E-05	1	460	0.019	1.71E-04	4	460	0.079	0.079
Dioxin	7.18E-12	2	665	4.8E-09	–		460	–	–		460	–	4.8E-09
Furan	3.75E-10	2	665	2.5E-07	–		460	–	–		460	–	2.5E-07
Formaldehyde	6.77E-04	2	665	0.45	4.05E-04	1	460	0.19	2.20E-04	4	460	0.101	0.45
Hydrogen Chloride	3.70E-02	3	665	24.6	6.37E-04	6	460	0.293	7.90E-02	6	460	36.3	36.3
Indium	1.27E-04	7	665	0.084	–		460	–	–		460	–	0.084
Manganese	7.98E-04	2	665	0.53	3.08E-06	1	460	0.0014	3.10E-05	4	460	0.014	0.53
Molybdenum	2.54E-04	7	665	0.17	4.88E-06	5	460	0.0022	8.83E-05	5	460	0.041	0.17
Nickel	4.41E-05	2	665	0.029	4.76E-05	1	460	0.022	1.02E-03	4	460	0.47	0.47
Phosphorus	3.53E-04	3	665	0.23	5.81E-06	5	460	0.0027	8.60E-04	5	460	0.40	0.40
Selenium	UD	3	665	–	4.60E-06	1	460	0.0021	5.34E-05	5	460	0.025	0.025
Silver	2.94E-05	3	665	0.020	–		460	–	–	460	–	0.017	0.020
Thallium	UD	3	665	–	–		460	–	–	460	–	–	–
Tin	1.62E-04	7	665	0.11	3.30E-05	5	460	0.015	8.83E-05	5	460	0.041	0.11
Zinc	4.24E-04	2	665	0.28	6.69E-06	5	460	0.0031	3.49E-04	5	460	0.16	0.28
Zirconium	9.29E-05	7	665	0.062	–		460	–	–		460	–	0.062

Note: UD = undetectable levels in gas stream.

* Denotes maximum for any fuel scenario.

References
1: Toxic Air Pollutant Emission Factors - A Compilation for Selected Air Toxic Compunds and Sources, Second Edition EPA-450/2-90-011 (1990).
2: Based on "Air Toxic Emissions from Wood Fired Boilers", C. Sassenrath, 1991 TAPPI Proceedings.
3: Based on stack test results of wood fired boilers and fuel analysis at Seminole Kraft Corporation (1990) equipped with wet scrubbers.
4: Estimating Emissions from Oil and Coal Combustion Sources EPA-450/2-89-001 (1989).
5: Emissions Assessment of Conventional Stationary Combustion Systems Volume V, 1981. Based on an uncontrolled spreader stoker design and then assuming 90% control from ESP.
6: Emissions Assessment of Conventional Stationary Combustion Systems Volume V, 1981. Based on an uncontrolled spreader stoker design.
7: EPA PM/VOC Speciation Database, updated October, 1989.
8. Based on maximum 20 ppm NH_3 in exhaust gases, see text; 65 ppm for coal.
9: Based upon stack test data at Dade County RRF, 1992, which indicated less than 20% of total chromium was chromium^{+6}.
10: Same as reference 2; includes 3% treated wood burning.

Source: KBN, 1992.

Table 2-11. Maximum Annual Emissions of Non-Regulated Pollutants for Osceola Power Cogeneration Facility (total all boilers) (Page 1 of 2) (Revised 05/18/93)

Non Regulated Pollutant	Biomass			No. 2 Fuel Oil			Coal			Total Annual Emissions (TPY)
	Emission Factor (lb/MMBtu)	Activity Factor (E12 Btu/yr)	Annual Emissions (TPY)	Emission Factor (lb/MMBtu)	Activity Factor (E12 Btu/yr)	Annual Emissions (TPY)	Emission Factor (lb/MMBtu)	Activity Factor (E12 Btu/yr)	Annual Emissions (TPY)	
<u>Normal Operations</u>										
Ammonia	1.5E-02	7.000	52.5	–	–	–	–	–	–	52.5
Antimony	UD	7.000	–	–	–	–	–	–	–	–
Arsenic	5.58E-05	7.000	0.20	–	–	–	–	–	–	0.20 [a]
Barium	1.06E-04	7.000	0.37	–	–	–	–	–	–	0.37
Bromine	1.47E-03	7.000	5.15	–	–	–	–	–	–	5.1 [a]
Cadmium	5.43E-06	7.000	0.019	–	–	–	–	–	–	0.019 [a]
Chromium	5.54E-05	7.000	0.194	–	–	–	–	–	–	0.194 [a]
Chromium +6	1.35E-05	7.000	0.047	–	–	–	–	–	–	0.047 [a]
Cobalt	4.98E-04	7.000	1.74	–	–	–	–	–	–	1.74 [a]
Copper	7.23E-05	7.000	0.25	–	–	–	–	–	–	0.25
Dioxin	7.18E-12	7.000	2.5E-08	–	–	–	–	–	–	2.5E-08 [a]
Furan	3.75E-10	7.000	1.3E-06	–	–	–	–	–	–	1.3E-06 [a]
Formaldehyde	6.77E-04	7.000	2.37	–	–	–	–	–	–	2.37 [a]
Hydrogen Chloride	3.70E-02	7.000	129.50	–	–	–	–	–	–	129.5
Indium	1.27E-04	7.000	0.44	–	–	–	–	–	–	0.44 [a]
Manganese	7.98E-04	7.000	2.79	–	–	–	–	–	–	2.8 [a]
Molybdenum	2.54E-04	7.000	0.89	–	–	–	–	–	–	0.89 [a]
Nickel	4.41E-05	7.000	0.15	–	–	–	–	–	–	0.15
Phosphorus	3.53E-04	7.000	1.24	–	–	–	–	–	–	1.24
Selenium	UD	7.000	–	–	–	–	–	–	–	–
Silver	2.94E-05	7.000	0.103	–	–	–	–	–	–	0.103 [a]
Thallium	UD	7.000	–	–	–	–	–	–	–	–
Tin	1.62E-04	7.000	0.57	–	–	–	–	–	–	0.57 [a]
Zinc	4.24E-04	7.000	1.48	–	–	–	–	–	–	1.48 [a]
Zirconium	9.29E-05	7.000	0.33	–	–	–	–	–	–	0.33 [a]
<u>25% Oil Firing</u>										
Ammonia	1.5E-02	4.941	37.1	4.8E-02	1.647	39.5	–	–	–	76.6 [a]
Antimony	UD	4.941	–	2.32E-06	1.647	0.0019	–	–	–	0.0019
Arsenic	5.58E-05	4.941	0.138	5.00E-07	1.647	0.0004	–	–	–	0.138
Barium	1.06E-04	4.941	0.26	6.69E-06	1.647	0.0055	–	–	–	0.27
Bromine	1.47E-03	4.941	3.632	6.97E-06	1.647	0.0057	–	–	–	3.637
Cadmium	5.43E-06	4.941	0.013	1.58E-06	1.647	0.0013	–	–	–	0.015
Chromium	5.54E-05	4.941	0.137	1.39E-05	1.647	0.0115	–	–	–	0.149
Chromium+6	1.35E-05	4.941	0.033	2.78E-06	1.647	0.0023	–	–	–	0.035
Cobalt	4.98E-04	4.941	1.23	1.17E-05	1.647	0.0097	–	–	–	1.24
Copper	7.23E-05	4.941	0.18	4.20E-05	1.647	0.0346	–	–	–	0.21
Dioxin	7.18E-12	4.941	1.8E-08	–	1.647	–	–	–	–	1.8E-08
Furan	3.75E-10	4.941	9.3E-07	–	1.647	–	–	–	–	9.3E-07
Formaldehyde	6.77E-04	4.941	1.67	4.05E-04	1.647	0.33	–	–	–	2.00
Hydrogen Chloride	3.70E-02	4.941	91.41	6.37E-04	1.647	0.5244	–	–	–	91.93
Indium	1.27E-04	4.941	0.31	–	1.647	–	–	–	–	0.31
Manganese	7.98E-04	4.941	1.97	3.08E-06	1.647	0.0025	–	–	–	2.0
Molybdenum	2.54E-04	4.941	0.63	4.88E-06	1.647	0.0040	–	–	–	0.63
Nickel	4.41E-05	4.941	0.11	4.76E-05	1.647	0.0392	–	–	–	0.15
Phosphorus	3.53E-04	4.941	0.87	5.81E-06	1.647	0.0048	–	–	–	0.88

Table 2-11. Maximum Annual Emissions of Non-Regulated Pollutants for Osceola Power Cogeneration Facility (total all boilers) (Page 2 of 2) (Revised 05/18/93)

	Biomass			No. 2 Fuel Oil			Coal			
Non Regulated Pollutant	Emission Factor (lb/MMBtu)	Activity Factor (E12 Btu/yr)	Annual Emissions (TPY)	Emission Factor (lb/MMBtu)	Activity Factor (E12 Btu/yr)	Annual Emissions (TPY)	Emission Factor (lb/MMBtu)	Activity Factor (E12 Btu/yr)	Annual Emissions (TPY)	Total Annual Emissions (TPY)
Selenium	UD	4.941	--	4.60E-06	1.647	0.0038	--	--	--	0.0038
Silver	2.94E-05	4.941	0.073	--	1.647	--	--	--	--	0.073
Thallium	UD	4.941	--	--	1.647	--	--	--	--	--
Tin	1.62E-04	4.941	0.40	3.30E-05	1.647	0.027	--	--	--	0.43
Zinc	4.24E-04	4.941	1.05	6.69E-06	1.647	0.0055	--	--	--	1.1
Zirconium	9.29E-05	4.941	0.23	--	1.647	--	--	--	--	0.23
<u>7% Coal Firing</u>										
Ammonia	0.0148	6.398	47.35				4.80E-02	0.482	11.57	58.91
Antimony	UD	6.398	--	--	--	--	3.49E-05	0.482	0.008	0.008 [a]
Arsenic	5.58E-05	6.398	0.18	--	--	--	2.64E-05	0.482	0.006	0.18
Barium	1.06E-04	6.398	0.34	--	--	--	7.44E-04	0.482	0.18	0.52 [a]
Bromine	1.47E-03	6.398	4.70	--	--	--	7.90E-04	0.482	0.190	4.89
Cadmium	5.43E-06	6.398	0.017	--	--	--	1.36E-06	0.482	0.0003	0.018
Chromium	5.54E-05	6.398	0.177	--	--	--	1.66E-05	0.482	0.004	0.181
Chromium +6	1.35E-05	6.398	0.043	--	--	--	3.32E-06	0.482	0.001	0.044
Cobalt	4.98E-04	6.398	1.59	--	--	--	7.20E-05	0.482	0.017	1.6
Copper	7.23E-05	6.398	0.23	--	--	--	1.71E-04	0.482	0.04	0.27 [a]
Dioxin	7.18E-12	6.398	2.3E-08	--	--	--	--	0.482	--	2.3E-08
Furan	3.75E-10	6.398	1.2E-06	--	--	--	--	0.482	--	1.2E-06
Formaldehyde	6.77E-04	6.398	2.2	--	--	--	2.20E-04	0.482	0.05	2.22
Hydrogen Chloride	3.70E-02	6.398	118.363	--	--	--	7.90E-02	0.482	19.04	137.4 [a]
Indium	1.27E-04	6.398	0.41	--	--	--	--	0.482	--	0.41
Manganese	7.98E-04	6.398	2.55	--	--	--	3.10E-05	0.482	0.007	2.6
Molybdenum	2.54E-04	6.398	0.81	--	--	--	8.83E-05	0.482	0.021	0.83
Nickel	4.41E-05	6.398	0.14	--	--	--	1.02E-03	0.482	0.25	0.39 [a]
Phosphorus	3.53E-04	6.398	1.13	--	--	--	8.60E-04	0.482	0.21	1.34 [a]
Selenium	UD	6.398	--	--	--	--	5.34E-05	0.482	0.013	0.013 [a]
Silver	2.94E-05	6.398	0.094	--	--	--	--	0.482	--	0.094
Thallium	UD	6.398	--	--	--	--	--	0.482	--	--
Tin	1.62E-04	6.398	0.52	--	--	--	8.83E-05	0.482	0.021	0.54
Zinc	4.24E-04	6.398	1.36	--	--	--	3.49E-04	0.482	0.08	1.4
Zirconium	9.29E-05	6.398	0.30	--	--	--	--	0.482	--	0.30

Note: UD = undetectable levels in gas stream.

[a] Denotes maximum annual emissions for any fuel scenario.

Table 3-3. PSD Source Applicability Analysis for Osceola Power Limited Partnership Facility (Revised 05/18/93)

Regulated Pollutant	Baseline Emissions (TPY)	Cogeneration Facility Annual Emissions (TPY)	Net Change (TPY)	Significant Emission Rate (TPY)	PSD Applies ?
Particulate (TSP)	357.7	114.7[a]	-243.0	25	No
Particulate (PM10)	321.9	108.5[b]	-213.4	15	No
Sulfur dioxide	178.5	353.2	174.7	40	Yes
Nitrogen oxides	437.8	424.9	-12.9	40	No
Carbon monoxide	5,992.3	1,225.0	-4,767.3	100	No
Volatile org. compds.	208.6	210.0	1.4	40	No[c]
Lead	0.16	0.10	-0.06	0.6	No
Mercury	0.0158[d]	0.0161	0.0003	0.1	No
Beryllium	0.00002	0.00140	0.00138	0.0004	Yes
Fluorides	0.0079	5.78	5.8	3	Yes
Sulfuric acid mist	5.36	10.60	5.2	7	No
Total reduced sulfur	--	--	0	10	No
Asbestos	--	--	0	0.007	No
Vinyl Chloride	--	--	0	0	No

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[a] Includes 105.0 TPY from boilers and 9.7 TPY from fugitive dust emission sources.

[b] Includes 105.0 TPY from boilers and 3.5 TPY from fugitive dust emission sources.

[c] Nonattainment review does not apply since the increase in VOC emissions is less than 40 TPY.

[d] The estimated annual average emission rate for 1990 and 1991 is 0.0158 TPY. The highest annual emission rate for either of these years is 0.0161 TPY.

Table 6-18. Maximum Impacts of Toxic Pollutants for Osceola Power Cogeneration Facility (total both boilers) (revised 05/18/93)

Pollutant	Maximum Hourly Emissions[a] (lb/hr)	Concentrations ($\mu g/m^3$)					
		8-Hour		24-Hour		Annual	
		Impact	NTL	Impact	NTL	Impact	NTL
Ammonia	44.2	6.4	180	4.2	43.2	--	--
Antimony	0.032	0.005	5	0.003	1.2	0.0002	0.3
Arsenic	0.22	0.03	2	0.02	0.48	0.00029[b]	0.00023
Barium	0.68	0.10	5	0.07	1.2	0.004	50
Beryllium	0.0054	0.0008	0.02	0.0005	0.0048	0.00003	0.00042
Bromine	1.96	0.283	7	0.2	1.68	--	--
Cadmium	0.0072	0.001	0.5	0.0007	0.12	0.00005	0.00056
Chromium metals	0.20	0.03	5	0.02	1.2	0.001	1000
Chromium +6	0.050	0.01	0.5	0.005	0.12	0.000069[c]	0.000083
Cobalt	0.66	0.10	0.5	0.06	0.12	--	--
Copper	0.16	0.02	10	0.02	2.4	--	--
Dioxins/Furans	5.1E-07	--	--	--	--	3.2E-09	2.2E-08
Fluoride	22.0	3.2	25	2.1	6	--	--
Formaldehyde	0.90	0.1	4.5	0.09	1.08	0.006	0.077
Hydrogen Chloride	72.6	10.5	70	7.0	16.8	0.5	7.0
Indium	0.17	0.02	1	0.02	0.24	--	--
Manganese	1.06	0.15	50	0.1	12	--	--
Mercury	0.0084	0.001	0.5	0.0008	0.12	0.00005	0.3
Molybdenum	0.34	0.05	50	0.03	12	--	--
Nickel	0.94	0.14	0.5	0.09	0.12	0.0014[d]	0.0042
Phosphorus	0.80	0.12	1	0.08	0.24	--	--
Selenium	0.05	0.007	2	0.005	0.48	--	--
Silver	0.04	0.006	0.1	0.004	0.024	0.0003	3
Sulfuric Acid	33.2[e]	4.8	10.0	3.2	2.4	--	--
Thallium	--	--	--	--	--	--	--
Tin	0.22	0.03	1	0.02	0.24	--	--
Zinc	0.56	0.08	10	0.05	2.4	--	--
Zirconium	0.12	0.02	50	0.01	12	--	--

Note: NTL = no-threat level.
Maximum concentrations determined with ISCST2 model and West Palm Beach meteorological data for 1982 to 1986.
Highest predicted concentration ($\mu g/m^3$) for a 10 g/s (79.365 lb/hr) emission rate:
8-hour = 11.44, 24-hour = 7.62, and Annual = 0.50

[a] Total both boilers.
[b] Based on maximum annual average emission rate of 0.20 TPY total both boilers (avg. of 0.046 lb/hr).
[c] Based on maximum annual average emission rate of 0.047 TPY total both boilers (avg. of 0.011 lb/hr).
[d] Based on maximum annual average emission rate of 0.95 TPY total both boilers (avg. of 0.22 lb/hr).
[e] Will occur a maximum of 7 percent of the year, corresponding to the proposed coal-burning restrictions.

Revised BACT
Analysis
for
Okeelanta Power and
Osceola Power

Table A. BACT "Top-down" Hierarchy of SO2 Reduction Methods for Okeelanta Power and Osceola Power (Revised 05/21/93)

Top-down Ranking	Technology	Range of Control Effectiveness (%)	Control level for BACT Analysis (%)	SO2 Emission Level (lb/MMBtu)	Annual Emissions for Okeelanta Power* (TPY)	Annual Emissions for Osceola Power** (TPY)
1st	Wet Limestone Scrubber (0.7% S Coal)	80-95	95	0.06	53.1	14.5
2nd	Spray Dryer (0.7% S Coal)	80-92	92	0.10	84.9	23.1
3rd	Dry-Sodium Duct Injection (0.7% S Coal)	40-50	50	0.60	530.7	144.6
Baseline	Baseline (0.7 Wt% S Coal Uncontrolled)	--	--	1.20	1,061.4	289.2

* Total for the three boilers based on maximum heat input rate of 1.769 E+12 Btu/yr from coal firing only.

** Total for the two boilers based on maximum heat input rate of 0.482 E+12 Btu/yr from coal firing only.

Table B. Summary of Top-Down BACT Impact Analysis Results for SO2, Okeelanta Power (Revised 05/21/93).

				Environmental Impacts		Energy Impacts	Economic Impacts			
Control Alternative	Total SO2 Emissions (TPY)	Emissions Reduction Over Baseline (TPY)	Incremental Emission Reduction (TPY)	Air toxics impact? (Yes/No)	Adverse enviromental impacts? (Yes/No)	Additional Energy Requirements Electricity (MW-hr/yr)	Total Annualized Cost ($/yr)	Incremental Annualized Cost ($/yr)	Cost Effectiveness Over Baseline ($/ton)	Incremental Cost Effectiveness ($/ton)
Wet Limestone Scrubber (0.7% S Coal)	53.1	1,008.3	31.8	Yes	Yes	1,300	$8,996,906	$4,120,531	$8,923	$129,406
Spray Dryer (0.7% S Coal)	84.9	976.5	445.8	No	Yes	890	$4,876,375	$1,421,421	$4,994	$3,189
Dry-Sodium Duct Injection (0.7% S Coal)	530.7	530.7	530.7	No	Yes	273	$3,454,954	$3,454,954	$6,510	$6,510
Baseline (0.7 Wt% S Coal Uncontrolled)	1,061.4	--	--	--	--	--	--	--	--	--

Table C. Summary of Top-Down BACT Impact Analysis Results for SO2 and Acid Gases, Okeelanta Power (Revised 05/21/93).

			(a)	(b)	(a)+(b)		Economic Impacts			
Control Alternative	Total SO2 Emissions (TPY)	Other Acid Gas Emissions (TPY)	SO2 Emissions Reduction Over Baseline (TPY)	Other Acid Gas Emissions Reduction Over Baseline (TPY)	Combined Emissions Reduction Over Baseline (TPY)	Overall Incremental Emission Reduction (TPY)	Total Annualized Cost ($/yr)	Incremental Annualized Cost ($/yr)	Combined Cost Effectiveness Over Baseline ($/ton)	Incremental Cost Effectiveness ($/ton)
Wet Limestone Scrubber (0.7% S Coal)	53.1	17.8	1,008.3	105.1	1,113.4	29.7	$8,996,906	$4,120,531	$8,080	$138,636
Spray Dryer (0.7% S Coal)	84.9	15.7	976.5	107.2	1,083.7	445.8	$4,876,375	$1,421,421	$4,500	$3,189
Dry-Sodium Duct Injection (0.7% S Coal)	530.7	15.7	530.7	107.2	637.9	637.9	$3,454,954	$3,454,954	$5,416	$5,416
Baseline (0.7 Wt% S Coal Uncontrolled)	1,061.4	122.9	--	--	--	--	--	--	--	--

Table D. Summary of Top–Down BACT Impact Analysis Results for SO2, Osceola Power (Revised 05/21/93).

				Environmental Impacts		Energy Impacts	Economic Impacts			
Control Alternative	Total SO2 Emissions (TPY)	Emissions Reduction Over Baseline (TPY)	Incremental Emission Reduction (TPY)	Air toxics impact? (Yes/No)	Adverse enviromental impacts? (Yes/No)	Additional Energy Requirements Electricity (MW–hr/yr)	Total Annualized Cost ($/yr)	Incremental Annualized Cost ($/yr)	Cost Effectiveness Over Baseline ($/ton)	Incremental Cost Effectiveness ($/ton)
Wet Limestone Scrubber (0.7% S Coal)	14.5	274.7	8.7	Yes	Yes	210	$5,705,622	$2,915,416	$20,767	$336,032
Spray Dryer (0.7% S Coal)	23.1	266.1	121.5	No	Yes	140	$2,790,206	$1,042,039	$10,487	$8,579
Dry–Sodium Duct Injection (0.7% S Coal)	144.6	144.6	144.6	No	Yes	65	$1,748,167	$1,748,167	$12,090	$12,090
Baseline (0.7 Wt% S Coal Uncontrolled)	289.2	--	--	--	--	--	--	--	--	--

Table E. Summary of Top-Down BACT Impact Analysis Results for SO2 and Acid Gases, Osceola Power (Revised 05/21/93).

							Economic Impacts			
Control Alternative	Total SO2 Emissions (TPY)	Total Acid Gas Emissions (TPY)	(a) SO2 Emissions Reduction Over Baseline (TPY)	(b) Acid Gas Emissions Reduction Over Baseline (TPY)	(a)+(b) Combined Emissions Reduction Over Baseline (TPY)	Overall Incremental Emission Reduction (TPY)	Total Annualized Cost ($/yr)	Incremental Annualized Cost ($/yr)	Combined Cost Effectiveness Over Baseline ($/ton)	Incremental Cost Effectiveness ($/ton)
Wet Limestone Scrubber (0.7% S Coal)	14.5	4.9	274.7	28.6	303.4	8.1	$5,705,622	$2,915,416	$18,806	$360,017
Spray Dryer (0.7% S Coal)	23.1	4.3	266.1	29.2	295.3	121.5	$2,790,206	$1,042,039	$9,449	$8,579
Dry-Sodium Duct Injection (0.7% S Coal)	144.6	4.3	144.6	29.2	173.8	173.8	$1,748,167	$1,748,167	$10,057	$10,057
Baseline (0.7 Wt% S Coal Uncontrolled)	289.2	33.5	--	--	--	--	--	--	--	--

Table 1. Capital Cost Estimates for Alternative SO2 Control Systems for Okeelanta Power Using 0.7% S Coal (Revised 05/21/93).

Cost Items	Cost Factors	Spray Dryer	Wet Scrubber
DIRECT CAPITAL COSTS (DCC):			
(1) Purchased Equipment			
(a) Basic Equipment*	Vendor Quote	$3,990,000	$11,400,000
(b) Auxiliary Equipment	included	included	included
(c) Structure Support	0.10 x (1a)	$399,000	$1,140,000
(d) Instrumentation & Controls	0.12 x (1a)	$478,800	$1,368,000
(e) Freight[1]	0.05 x (1a .. 1d)	$243,390	$695,400
(f) Sales Tax (Florida)	0.06 x (1a .. 1d)	$292,068	$834,480
(g) Subtotal	(1a .. 1f)	$5,403,258	$15,437,880
(2) Direct Installation[1]	0.30 x (1a .. 1f)	$1,620,977	$4,631,364
(3) Upsizing ESP for Particulate Control			
(a) ESP and Auxiliary Equip. Upsizing	See Note 1	$210,000	--
(b) Instrumentation & Controls	0.1 x (3a)	$21,000	--
(c) Freight[1]	0.05 x (3a .. 3b)	$11,550	--
(d) Sales Tax (Florida)	0.06 x (3a .. 3b)	$13,860	--
(e) Direct Installation Costs[2]	0.67 x (3a .. 3d)	$171,795	--
(4) Upsizing Ash Handling System			
(a) Cost of Upsizing for 3 Boilers	See Note 2	$40,000	--
Total DCC:	(1) + (2) + (3) + (4)	$7,492,440	$20,069,244
INDIRECT CAPITAL COSTS (ICC):			
(5) Indirect Installation			
(a) Engineering & Supervision[1]	0.10 x (DCC)	$749,244	$2,006,924
(b) Construction & Field Expenses[1]	0.10 x (DCC)	$749,244	$2,006,924
(c) Contruction Contractor Fee[1]	0.05 x (DCC)	$374,622	$1,003,462
(d) Contigencies[1]	0.20 x (DCC)	$1,498,488	$4,013,849
(6) Other Indirect Costs			
(a) Startup & Testing[1]	0.03 x (DCC)	$224,773	$602,077
(b) Working Capital	30-day DOC**	$123,580	$180,088
Total ICC:	(5) + (6)	$3,719,951	$9,813,325
TOTAL CAPITAL INVESTMENTS (TCI):	DCC + ICC	$11,212,391	$29,882,569

* For the spray dryers, the basic equipment cost for three units are the average of budgetary quotations from ABB-Flakt, Joy Environmental Systems, and United McGill. The basic equipment cost for the limestone wet scrubbers was based on budgetary pricing from ABB Environmental Systems.

** 30 days of direct operating costs, calculated from the annualized cost Table 2 (i.e., total DOC/12 months).

[1] Based on catalytic incinerators, from OAQPS Control Cost Manual, Fourth Edition (1990).

[2] Based on Capital Cost Factors for ESP, from OAQPS Control Cost Manual, Fourth Edition (1990).

Note 1: 10 percent increase in size for the ESP at $7,000 for 1 percent increase, from information supplied by Research-Cottrell.

Note 2: Capital cost estimates of the ash handling system prior to the upsizing in order to handle the additional ash generated from the spray dryer process are approximately $400,000. The upsizing of the ash handling system will add an additional 10% cost or approximately $40,000 to the spray dryer option. The wet FDG option does not involve any upsizing of the ESP ash handling system.

Table 2. Annualized Cost Estimates for Alternative SO2 Control Systems for Okeelanta Power Using 0.7% S Coal (Revised 05/21/93).

Cost Items	Basis	Spray Dryer	Wet Scrubber
DIRECT OPERATING COSTS (DOC):			
(1) Labor			
Operator[2]	\$22/hr; approx. 3,630 hr/yr for 3 boilers	\$79,860	--
	\$22/hr; approx. 5,470 hr/yr for 3 boilers	--	\$120,340
Supervisor[1]	15% of operator cost	\$11,979	\$18,051
(2) Maintenance[2]	5% of direct capital cost	\$374,622	\$1,003,462
(3) Replacement Parts	3% of direct capital cost	\$224,773	\$602,077
(4) Utilities			
(a) Electricity	\$85 per MW-hr; 890 and 1,300 MW-hr	\$75,650	\$110,500
(b) Water	\$0.27 / 1,000 gal; 32.3 and 38.2 mil. gal.	\$8,721	\$10,314
(5) Raw Chemicals			
(a) Limestone (97% purity)	\$32 / ton delivered for 2,400 TPY	--	\$76,800
(b) Hydrated Lime (74% purity)	\$140 / ton delivered for 3,500 TPY	\$490,000	--
(6) Differential Fuel Cost or Credit			
(a) 0.7% S Coal	No Cost Adjustment	\$0	\$0
(7) Solid Disposal	\$27 / ton for approx. 8,050 TPY	\$217,350	--
(8) Sludge Disposal	\$27 / ton for approx. 8,130 TPY	--	\$219,510
Total DOC		\$1,482,955	\$2,161,055
INDIRECT OPERATING COSTS (IOC):			
(7) Overhead[1]	60% of operating labor & maintenance	\$1,120,668	\$778,655
(8) Property Taxes[1]	1% of total capital investment	\$112,124	\$298,826
(9) Insurance[1]	1% of total capital investment	\$112,124	\$298,826
(10) Administration[1]	2% of total capital investment	\$224,248	\$597,651
Total IOC		\$1,569,164	\$1,973,958
CAPITAL RECOVERY COSTS (CRC)	CRF of 0.1627 times TCI	\$1,824,256	\$4,861,894
ANNUALIZED COSTS (AC):	DOC + IOC + CRC	\$4,876,375	\$8,996,906

[1] Based on catalytic incinerators, from OAQPS Control Cost Manual, Fourth Edition.

[2] Based on maximum of 25% coal firing for the Flo-Energy cogeneration facility.

Table 3. Cost Estimates for Sorbent Injection Systems for Okeelanta Power Using 0.7% S Coal (Revised 05/21/93).

Cost Items	Basis	Furnace Injection Lime Sorbent	Duct Injection Lime Sorbent	Duct Injection Sodium Sorbent
DIRECT CAPITAL COSTS (DCC):				
(1) Dry Sorbent Injection System	See Note 1	$5,880,000	$4,704,000	$4,410,000
(a) Flue Gas Humidification System		Included	Included	--
(2) Upsizing ESP for Particulate Control				
(a) Percent Increase in Size	Based on ESP's Vendor Estimate	55%	60%	25%
(b) Cost of Upsizing for 3 Boilers	$7,000 per 1% Increase per boiler	$1,155,000	$1,260,000	$525,000
(3) Upsizing Ash Handling System				
(a) Cost of Upsizing for 3 Boilers	10% of current system cost. See Note 2.	$40,000	$40,000	$40,000
Subtotal of DCC	(1) + (2b) + (3a)	$7,075,000	$6,004,000	$4,975,000
INDIRECT CAPITAL COSTS (ICC):				
(4) Indirect Installation				
(a) Engineering & Supervision [1]	0.20 x (DCC)	$1,415,000	$1,200,800	$995,000
(b) Construction & Field Expenses [1]	0.20 x (DCC)	$1,415,000	$1,200,800	$995,000
(c) Contruction Contractor Fee [1]	0.10 x (DCC)	$707,500	$600,400	$497,500
(d) Contingencies	0.25 x (DCC)	$1,768,750	$1,501,000	$1,243,750
(5) Other Indirect Costs				
(a) Start-up, Perf. Test & Model Study [1]	0.04 x (DCC)	$283,000	$240,160	$199,000
(b) Working Capital	30-day DOC	$118,274	$118,274	$118,274
Total ICC:	(4) + (5)	$5,707,524	$4,861,434	$4,048,524
TOTAL CAPITAL INVESTMENTS (TCI):	DCC + ICC	$12,782,524	$10,865,434	$9,023,524
DIRECT OPERATING COSTS (DOC):				
(6) Labor				
Operator [2]	$22/hr; 5,500 hr/yr total for 3 boilers	$121,000	$121,000	$121,000
Supervisor [1]	15% of operator cost	$18,150	$18,150	$18,150
(7) Maintenance [2]	0.05 x (DCC)	$353,750	$300,200	$248,750
(8) Replacement Parts	0.03 x (DCC)	$212,250	$180,120	$149,250
(9) Utilities				
(a) Electricity for ESP	$85 per MW-hr	$17,111	$18,666	$7,778
for Auxiliary Equipment	$85 per MW-hr	$33,934	$37,019	$15,425
(b) Humidification Water	$0.27 / 1,000 gal	$9,356	$9,356	--
(c) Steam Lost or Reheat	\$6.19/10^3 lb	$41,431	--	--
(10) Raw Chemicals				
(a) Hydrated Lime (74% purity)	$140 / ton delivered for approx. 4,600 TPY	$495,880	$495,880	--
(b) Sodium Bicarbonate (98% purity)	$200 / ton delivered for approx. 4,650 TPY	--	--	$716,100
(11) Differential Fuel Cost or Credit				
(a) 0.7% S Coal	None	$0	$0	$0
(12) Solid Disposal	$27 / ton for Lime; $50 /ton for NaHCO3-type	$116,424	$122,661	$116,694
Total DOC		$1,419,285	$1,303,052	$1,393,146
INDIRECT OPERATING COSTS (IOC):				
(13) Overhead [1]	60% of operating labor & maintenance	$295,740	$263,610	$232,740
(14) Property Taxes [1]	1% of total capital investment	$127,825	$108,654	$90,235
(15) Insurance [1]	1% of total capital investment	$127,825	$108,654	$90,235
(16) Administration [1]	2% of total capital investment	$255,650	$217,309	$180,470
Total IOC		$807,041	$698,227	$593,681
CAPITAL RECOVERY COSTS (CRC)	CRF of 0.1627 times TCI	$2,079,717	$1,767,806	$1,468,127
ANNUALIZED COSTS (AC):	DOC + IOC + CRC	$4,306,043	$3,769,085	$3,454,954

1 Based on Capital Cost Factors for ESP, from OAQPS Control Cost Manual, Fourth Edition (1990).

[2] Based on a maximum 86,944 TPY of coal firing for Okeelanta Power.

Note 1: Capital cost factors are $4,000/MMBtu of heat input for Furnace Injection and $3,500/MMBtu of heat input for Duct Injection (both using lime) projecting from Babcock and Wilcox's cost analysis summary for Ohio's Edison Power Plant located at Edgewater, Ohio. The capital cost factor for the sodium sorbent injection process is $3,000/MMBtu of heat input based on estimations from Colorado Springs Utilities Company and Publlic Service Company of Colorado. Total Capital Investment for the dry sorbent injection process was calculated using these cost factors for a total of three 490-MMBtu boilers for the proposed Okeelanta Power.

Note 2: Capital cost estimates of the ash handling system prior to the upsizing in order to handle the additional ash generated from the DSI process are approximately $400,000.

Table 4. Capital Cost Estimates for Alternative SO2 Control Systems for Osceola Power Using 0.7% S Coal (Revised 05/21/93).

Cost Items	Cost Factors	Spray Dryer	Wet Scrubber
DIRECT CAPITAL COSTS (DCC):			
(1) Purchased Equipment			
(a) Basic Equipment*	Vendor Quote	$2,660,000	$7,600,000
(b) Auxiliary Equipment	included	included	included
(c) Structure Support	0.10 x (1a)	$266,000	$760,000
(d) Instrumentation & Controls	0.12 x (1a)	$319,200	$912,000
(e) Freight[1]	0.05 x (1a .. 1d)	$162,260	$463,600
(f) Sales Tax (Florida)	0.06 x (1a .. 1d)	$194,712	$556,320
(g) Subtotal	(1a .. 1f)	$3,602,172	$10,291,920
(2) Direct Installation[1]	0.30 x (1a .. 1f)	$1,080,652	$3,087,576
(3) Upsizing ESP for Particulate Control			
(a) ESP and Auxiliary Equip. Upsizing	See Note 1	$140,000	--
(b) Instrumentation & Controls	0.1 x (3a)	$14,000	--
(c) Freight[1]	0.05 x (3a .. 3b)	$7,700	--
(d) Sales Tax (Florida)	0.06 x (3a .. 3b)	$9,240	--
(e) Direct Installation Costs[2]	0.67 x (3a .. 3d)	$114,530	--
(4) Upsizing Ash Handling System			
(a) Cost of Upsizing for 2 Boilers	See Note 2	$26,000	--
Total DCC:	(1) + (2) + (3) + (4)	$4,994,293	$13,379,496
INDIRECT CAPITAL COSTS (ICC):			
(5) Indirect Installation			
(a) Engineering & Supervision[1]	0.10 x (DCC)	$499,429	$1,337,950
(b) Construction & Field Expenses[1]	0.10 x (DCC)	$499,429	$1,337,950
(c) Contruction Contractor Fee[1]	0.05 x (DCC)	$249,715	$668,975
(d) Contigencies[1]	0.20 x (DCC)	$998,859	$2,675,899
(6) Other Indirect Costs			
(a) Startup & Testing[1]	0.03 x (DCC)	$149,829	$401,385
(b) Working Capital	30-day DOC**	$50,016	$99,335
Total ICC:	(5) + (6)	$2,447,277	$6,521,493
TOTAL CAPITAL INVESTMENTS (TCI):	DCC + ICC	$7,441,570	$19,900,989

* For the spray dryers, the basic equipment cost for three units are the average of budgetary quotations from ABB-Flakt, Joy Environmental Systems, and United McGill. The basic equipment cost for the limestone wet scrubbers was based on budgetary pricing from ABB Environmental Systems.

** 30 days of direct operating costs, calculated from the annualized cost Table 2 (i.e., total DOC/12 months).

[1] Based on catalytic incinerators, from OAQPS Control Cost Manual, Fourth Edition (1990).

[2] Based on Capital Cost Factors for ESP, from OAQPS Control Cost Manual, Fourth Edition (1990).

Note 1: 10 percent increase in size for the ESP at $7,000 for 1 percent increase, from information supplied by Research-Cottrell.

Note 2: Capital cost estimates of the ash handling system prior to the upsizing in order to handle the additional ash generated from the spray dryer process are approximately $260,000. The upsizing of the ash handling system will add an additional 10% cost or approximately $26,000 to the spray dryer option. The wet FDG option does not involve any upsizing of the ESP ash handling system.

Table 5. Annualized Cost Estimates for Alternative SO2 Control Systems for Osceola Power Using 0.7% S Coal (Revised 05/21/93).

Cost Items	Basis	Spray Dryer	Wet Scrubber
DIRECT OPERATING COSTS (DOC):			
(1) Labor			
Operator[2]	$22/hr; approx. 870 hr/yr for 2 boilers	$19,140	--
	$22/hr; approx. 1,300 hr/yr for 2 boilers	--	$28,600
Supervisor[1]	15% of operator cost	$2,871	$4,290
(2) Maintenance[2]	5% of direct capital cost	$249,715	$668,975
(3) Replacement Parts	3% of direct capital cost	$149,829	$401,385
(4) Utilities			
(a) Electricity	$85 per MW-hr; 140 and 210 MW-hr	$11,900	$17,850
(b) Water	$0.27 / 1,000 gal; 18.4 and 21.5 mil. gal.	$1,450	$1,696
(5) Raw Chemicals			
(a) Limestone (97% purity)	$32 / ton delivered for 560 TPY	--	$17,920
(b) Hydrated Lime (74% purity)	$140 / ton delivered for 820 TPY	$114,800	--
(6) Differential Fuel Cost or Credit			
(a) 0.7% S Coal	No Cost Adjustment	$0	$0
(7) Solid Disposal	$27 / ton for approx. 1,870 TPY	$50,490	--
(8) Sludge Disposal	$27 / ton for approx. 1,900 TPY	--	$51,300
Total DOC		$600,194	$1,192,015
INDIRECT OPERATING COSTS (IOC):			
(7) Overhead[1]	60% of operating labor & maintenance	$681,605	$479,676
(8) Property Taxes[1]	1% of total capital investment	$74,416	$199,010
(9) Insurance[1]	1% of total capital investment	$74,416	$199,010
(10) Administration[1]	2% of total capital investment	$148,831	$398,020
Total IOC		$979,268	$1,275,716
CAPITAL RECOVERY COSTS (CRC)	CRF of 0.1627 times TCI	$1,210,744	$3,237,891
ANNUALIZED COSTS (AC):	DOC + IOC + CRC	$2,790,206	$5,705,622

[1] Based on catalytic incinerators, from OAQPS Control Cost Manual, Fourth Edition.

[2] Based on a maximum 20,069 TPY of coal firing for Osceola Power.

MAY 17, 93

Clair Fancy	DER/BAQM	904 488 1344
GUS R. CEPERO	OKEELANTA CORP.	407-996-9072 x-204
JORGE Dominicis	" "	655-6303
David Buff	KBN Eng.	904-331-9000
Willard Hanks	DER/BAR	904/488-1344
David Dee	Carlton Fields	904 - 224 - 1585
CLAIRE LARDNER	OGC/DER	904 488·9730
PAT COMER	OGC/DER	904 488 9730
PRESTON LEWIS	FDER/DARM/BAR	904 488-1344
Howard L. Rhodes	DER/	904-488-0114

Patty

INTEROFFICE MEMORANDUM

Date: 12-May-1993 10:34am
From: William Congdon TA
CONGDON_W
Dept: Office General Couns
Tel No: 904/488-9730
SUNCOM: sc 278-9730

TO: Patricia Comer TAL (COMER_P)
TO: Claire Lardner TAL (LARDNER_C)

CC: Clair Fancy TAL (FANCY_C)

Subject: Meeting on air permit

David Dee has scheduled a meeting with Claire Fancy for Monday the 17th @ 4 pm to discuss pending permit applications for two co-generation plants. The plants are sugar industry related (I think).

Claire L., please attend as we may end up in litigation over the applications and they are in your district. Pat, I would like a program atty. to attend also, if Claire Fancy thinks we need more air legal expertise than Claire L. now has. I'll leave the attendance of a 2nd atty. to you and C.F.

Steis ok

KBN

FACSIMILE COVER SHEET

DATE: May 13, 1993

TO: Clair Fancy

ORGANIZATION: FDER

FAX NUMBER: 904 922-6979 TELEPHONE NUMBER: 904 488-1344

FROM: Dave Buff

TOTAL NUMBER OF PAGES: 2 (including cover page)

MESSAGE/INSTRUCTIONS:

PROJECT NUMBER: 12118-0100 FAX OPERATOR:

(✓) This is the ONLY form of delivery of the transmitted document.

( ) The original of the transmitted document will be sent by:

- ( ) US Mail
- ( ) Overnight delivery
- ( ) Other:

Return original to DAB

cc: Project File _____ yes _____ no

KBN Engineering and Applied Sciences, Inc.
1034 NW 57th Street, Gainesville, Florida, 32605, U.S.A.
Main Office: Phone (904) 331-9000 FAX (904) 332-4189
Canopy Park: Phone (904) 331-9000 FAX (904) 331-3368

AF1/FAX (07/92)

12118A1/WAI-1
5/12/93

WAIVER OF 90 DAY TIME LIMIT
UNDER SECTIONS 120.60(2) AND 403.0876, FLORIDA STATUTES

License (Permit, Certification) Application No. AC50-219795 and AC50-219413

Applicant's Name: Osceola Power Limited Partnership and Okeelanta Power Limited Partnership

With regard to the above referenced application, the applicant hereby with full knowledge and understanding of applicant's rights under Sections 120.60(2) and 403.0876, Florida Statutes, waives the right to have the application approved or denied by the State of Florida Department of Environmental Regulation within the 90 day time period prescribed by law. Said waiver is made freely and voluntarily by the applicant, with full knowledge, and without any pressure or coercion by anyone employed by the State of Florida Department of Environmental Regulation.

This waiver shall expire on the 25 day of May, 1993.

The undersigned is authorized to make this waiver on behalf of the applicant.

David A. Buff
Signature 5/13/93

David A. Buff, P.E.
Name (Please Type or Print)

Revised April, 1990

5-13-93

WAIVER OF 90 DAY TIME LIMIT
UNDER SECTIONS 120.60(2) and 403.0876, FLORIDA STATUTES

License (Permit, Certification) Application No. AC50-219795 & AC50-219413

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This waiver shall expire on the 4th day of June 1993.

The undersigned is authorized to make this waiver on behalf of the applicant.

G R Cepero
Signature

G. R. CEPERO
Name (Please Type or Print)

Revised April, 1990

May 6, 1993

RECEIVED
MAY 07 1993
Division of Air
Resources Management

Mr. Clair Fancy, P.E., Chief
Bureau of Air Regulation
Florida Department of Environmental Regulation
2600 Blair Stone Road
Tallahassee, FL 32399-2400

Re: Okeelanta Power Limited Partnership and Osceola Power Limited Partnership

Dear Mr. Fancy:

As requested by Mr. Gus Cepero of Okeelanta, I am forwarding to you revised tables from the Okeelanta and Osceola cogeneration permit applications. These tables present the revised annual heat input rates and the revised annual emissions for the two facilities, and the revised PSD source applicability analysis. As presented in Mr. Cepero's letter to the Department, the maximum amount of coal burning has been reduced at each of the facilities. At Okeelanta Power, annual coal burning would be limited to 2.086E+12 Btu/yr, equivalent to a 19 percent capacity factor. At Osceola Power, annual coal burning would be limited to 0.482E+12 Btu/yr, equivalent to a 7 percent capacity factor. In addition, the fuel oil sulfur content has been reduced to 0.05 percent. Please note that with these revisions, the two facilities are no longer subject to PSD for sulfuric acid mist emissions.

If you have any questions concerning these revise tables, please call me at 904-331-9000.

Sincerely,

David A. Buff

David A. Buff, M.E., P.E.
Principal Engineer

DAB/tyf

cc: Preston Lewis, DER
Howard Rhodes, DER
Gus Cepero, Okeelanta Corp.
David Dee, Carlton-Fields
Jewell Harper, EPA
John Bunyak, NPS
Bevin Beaudet, PBCHU
File (2)
J. Harper
D. Knowles
B. [illegible]

12118A1/10

KBN ENGINEERING AND APPLIED SCIENCES, INC.
1034 Northwest 57th Street Gainesville, Florida 32605 904/331-9000 FAX: 904/332-4189

Table 2-2. Maximum Fuel Usage and Heat Input Rates, Okeelanta Power Limited Partnership Facility (Revised 05/03/93)

Fuel	Heat Input	Heat Transfer Efficiency (%)	Heat Output	Fuel Firing Rate
	Maximum Short-Term (per boiler)			
	(MMBtu/hr)		(MMBtu/hr)	
Biomass	715	68	486	168,236 lb/hr[a]
No. 2 Oil	490	85	417	3,551 gal/hr
Coal	490	85	417	40,833 lb/hr
	Annual Average (total all three boilers)			
	(Btu/yr)		(Btu/yr)	
NORMAL OPERATIONS				
Biomass	1.150E+13	68	7.820E+12	1,352,941 TPY[a]
No. 2 Oil	0	85	0	0 gal/yr
Coal	0	85	0	0 TPY
TOTAL	1.150E+13		7.820E+12	
25% OIL FIRING				
Biomass	8.118E+12	68	5.520E+12	955,059 TPY[a]
No. 2 Oil	2.706E+12	85	2.300E+12	19,608,696 gal/yr
Coal	0	85	0	0 TPY
TOTAL	1.082E+13		7.820E+12	
19% COAL FIRING				
Biomass	8.893E+12	68	6.047E+12	1,046,235 TPY[a]
No. 2 Oil	0	85	0	0 gal/yr
Coal	2.086E+12	85	1.773E+12	86,917 TPY
TOTAL	1.098E+13		7.820E+12	

Note: Total heat output required = 486 MMBtu/hr each boiler, and 7.820E+12 Btu/yr total all boilers. Fuels may be burned in combination, not to exceed indicated total heat outputs.

[a] Based on heating value for bagasse of 4,250 Btu/lb, wet basis.

Table 2-6. Maximum Annual Emissions for the Okeelanta Power Limited Partnership Facility (total all three boilers) (Revised 05/03/93)

Regulated Pollutant	Biomass			No. 2 Fuel Oil			Coal			Total Annual Emissions (TPY)
	Emission Factor (lb/MMBtu)	Activity Factor (E12 Btu/yr)	Annual Emissions (TPY)	Emission Factor (lb/MMBtu)	Activity Factor (E12 Btu/yr)	Annual Emissions (TPY)	Emission Factor (lb/MMBtu)	Activity Factor (E12 Btu/yr)	Annual Emissions (TPY)	
Normal Operations										
Particulate (TSP)	0.03	11.500	172.50	–	–	–	–	–	–	172.50 [a]
Particulate (PM10)	0.03	11.500	172.50	–	–	–	–	–	–	172.50 [a]
Sulfur dioxide	0.02	11.500	115.00	–	–	–	–	–	–	115.00
Nitrogen oxides	0.15	11.500	862.50	–	–	–	–	–	–	862.50 [a]
Carbon monoxide	0.35	11.500	2,012.50	–	–	–	–	–	–	2,012.50 [a]
VOC	0.06	11.500	345.00	–	–	–	–	–	–	345.00 [a]
Lead	2.5E-05	11.500	0.14	–	–	–	–	–	–	0.14
Mercury	b	11.500	b	–	–	–	–	–	–	0.0262
Beryllium	–	–	–	–	–	–	–	–	–	–
Fluorides	–	–	–	–	–	–	–	–	–	–
Sulfuric acid mist	0.00060	11.500	3.45	–	–	–	–	–	–	3.45
Total reduced sulfur	–	–	–	–	–	–	–	–	–	–
Asbestos	–	–	–	–	–	–	–	–	–	–
Vinyl Chloride	–	–	–	–	–	–	–	–	–	–
25% Oil Firing										
Particulate (TSP)	0.03	8.118	121.77	0.03	2.706	40.59	–	–	–	162.36
Particulate (PM10)	0.03	8.118	121.77	0.03	2.706	40.59	–	–	–	162.36
Sulfur dioxide	0.02	8.118	81.18	0.05	2.706	67.65	–	–	–	148.83
Nitrogen oxides	0.15	8.118	608.85	0.15	2.706	202.95	–	–	–	811.80
Carbon monoxide	0.35	8.118	1,420.65	0.2	2.706	270.60	–	–	–	1,691.25
VOC	0.06	8.118	243.54	0.03	2.706	40.59	–	–	–	284.13
Lead	2.5E-05	8.118	0.10	8.9E-07	2.706	0.001	–	–	–	0.10
Mercury	b	8.118	b	b	2.706	b	–	–	–	0.0262
Beryllium	–	–	–	3.5E-07	2.706	0.0005	–	–	–	0.00047
Fluorides	–	–	–	6.27E-06	2.706	0.0085	–	–	–	0.0085
Sulfuric acid mist	0.00060	8.118	2.44	0.0015	2.706	2.03	–	–	–	4.46
Total reduced sulfur	–	–	–	–	–	–	–	–	–	–
Asbestos	–	–	–	–	–	–	–	–	–	–
Vinyl Chloride	–	–	–	–	–	–	–	–	–	–
19% Coal Firing										
Particulate (TSP)	0.03	8.893	133.40	–	–	–	0.03	2.086	31.29	164.69
Particulate (PM10)	0.03	8.893	133.40	–	–	–	0.03	2.086	31.29	164.69
Sulfur dioxide	0.02	8.893	88.93	–	–	–	1.2	2.086	1,251.60	1,340.53 [a]
Nitrogen oxides	0.15	8.893	666.98	–	–	–	0.17	2.086	177.31	844.29
Carbon monoxide	0.35	8.893	1,556.28	–	–	–	0.2	2.086	208.60	1,764.88
VOC	0.06	8.893	266.79	–	–	–	0.03	2.086	31.29	298.08
Lead	2.5E-05	8.893	0.11	–	–	–	6.4E-05	2.086	0.07	0.18 [a]
Mercury	b	8.893	b	–	–	–	b	2.086	b	0.0262 [a]
Beryllium	–	–	–	–	–	–	5.9E-06	2.086	0.0062	0.0062 [a]
Fluorides	–	–	–	–	–	–	0.024	2.086	25.03	25.03 [a]
Sulfuric acid mist	0.00060	8.893	2.67	–	–	–	0.036	2.086	37.55	40.22 [a]
Total reduced sulfur	–	–	–	–	–	–	–	–	–	–
Asbestos	–	–	–	–	–	–	–	–	–	–
Vinyl Chloride	–	–	–	–	–	–	–	–	–	–

---

[a] Indicates maximum annual emission rate.
[b] Refer to text for explanation.

Table 3-3. PSD Source Applicability Analysis for the Okeelanta Power Limited Partnership Facility (revised 05/03/93)

Regulated Pollutant	Baseline Emissions (TPY)	Cogeneration Facility Annual Emissions (TPY)	Net Change (TPY)	Significant Emission Rate (TPY)	PSD Applies ?
Particulate (TSP)	473.7	183.6[b]	-290.1	25	No
Particulate (PM10)	426.3	176.4[c]	-249.9	15	No
Sulfur Dioxide	748.3	1,340.5	592.2	40	Yes
Nitrogen Oxides	888.7	862.5	-26.2	40	No
Carbon Monoxide	10,388.0	2,012.5	-8,375.5	100	No
VOC	401.9	345.0	-56.9	40	No[a]
Lead	0.28	0.18	-0.10	0.6	No
Mercury	0.0256[d]	0.0262	0.0006	0.1	No
Beryllium	0.0004	0.0062	0.0058	0.0004	Yes
Fluorides	0.04	25.0	25.0	3	Yes
Sulfuric Acid Mist	22.4	40.2	6.4	7	No
Total Reduced Sulfur	--	--	0	10	No
Asbestos	--	--	0	0.007	No
Vinyl Chloride	--	--	0	0	No

---

[a] Nonattainment review does not apply since there is no increase in VOC emissions.

[b] Includes 172.5 TPY from boilers and 11.1 TPY from fugitive dust sources.

[c] Includes 172.5 TPY from boilers and 3.9 TPY from fugitive dust sources.

[d] The estimated average annual emission rate for the most recent 2 years is 0.0256 TPY. The highest annual emission rate for either of the last 2 years is 0.0262 TPY.

Table 2-2. Maximum Fuel Usage and Heat Input Rates, Osceola Power Limited Partnership Facility (Revised 05/03/93)

Fuel	Heat Input	Heat Transfer Efficiency (%)	Heat Output	Fuel Firing Rate
	Maximum Short-Term (per boiler)			
	(MMBtu/hr)		(MMBtu/hr)	
Biomass	665	68	452	156,471 lb/hr[a]
No. 2 Oil	460	85	391	3,333 gal/hr
Coal	460	85	391	38,333 lb/hr
	Annual Average (total two boilers)			
	(Btu/yr)		(Btu/yr)	
NORMAL OPERATIONS				
Biomass	7.000E+12	68	4.760E+12	823,529 TPY[a]
No. 2 Oil	0	85	0	0 gal/yr
Coal	0	85	0	0 TPY
	----------		----------	
TOTAL	7.000E+12		4.760E+12	
25% OIL FIRING				
Biomass	4.941E+12	68	3.360E+12	581,294 TPY[a]
No. 2 Oil	1.647E+12	85	1.400E+12	11,934,783 gal/yr
Coal	0	85	0	0 TPY
	----------		----------	
TOTAL	6.588E+12		4.760E+12	
7% COAL FIRING				
Biomass	6.398E+12	68	4.351E+12	752,706 TPY[a]
No. 2 Oil	0	85	0	0 gal/yr
Coal	4.816E+11	85	4.093E+11	20,065 TPY
	----------		----------	
TOTAL	6.880E+12		4.760E+12	

Notes: Total heat output required = 452 MMBtu/hr each boiler and 4.760E+12 Btu/yr total both boilers. Fuels may be burned in combination, not to exceed indicated total heat outputs.

[a] Based on heating value for bagasse of 4,250 Btu/lb.

Table 2-6. Maximum Annual Emissions for Osceola Power Limited Partnership Facility (total all boilers) (Revised 05/03/93)

	Biomass			No. 2 Fuel			Coal			Total
Regulated Pollutant	Emission Factor (lb/MMBtu)	Activity Factor (E12 Btu/yr)	Annual Emissions (TPY)	Emission Factor (lb/MMBtu)	Activity Factor (E12 Btu/yr)	Annual Emissions (TPY)	Emission Factor (lb/MMBtu)	Activity Factor (E12 Btu/yr)	Annual Emissions (TPY)	Annual Emissions (TPY)
Normal Operations										
Particulate (TSP)	0.03	7.000	105.00	–	–	–	–	–	–	105.00 [a]
Particulate (PM10)	0.03	7.000	105.00	–	–	–	–	–	–	105.00 [a]
Sulfur dioxide	0.02	7.000	70.00	–	–	–	–	–	–	70.00
Nitrogen oxides	0.12	7.000	420.00	–	–	–	–	–	–	420.00
Carbon monoxide	0.35	7.000	1,225.00	–	–	–	–	–	–	1,225.00 [a]
VOC	0.06	7.000	210.00	–	–	–	–	–	–	210.00 [a]
Lead	2.5E-05	7.000	0.09	–	–	–	–	–	–	0.09
Mercury	b	7.000	b	–	–	–	–	–	–	0.0139
Beryllium	–	–	–	–	–	–	–	–	–	0.00
Fluorides	–	–	–	–	–	–	–	–	–	0.00
Sulfuric acid mist	0.00060	7.000	2.10	–	–	–	–	–	–	2.10
Total reduced sulfur	–	–	–	–	–	–	–	–	–	–
Asbestos	–	–	–	–	–	–	–	–	–	–
Vinyl Chloride	–	–	–	–	–	–	–	–	–	–
25% Oil Firing										
Particulate (TSP)	0.03	4.941	74.12	0.03	1.647	24.71	–	–	–	98.82
Particulate (PM10)	0.03	4.941	74.12	0.03	1.647	24.71	–	–	–	98.82
Sulfur dioxide	0.02	4.941	49.41	0.05	1.647	41.18	–	–	–	90.59
Nitrogen oxides	0.12	4.941	296.46	0.12	1.647	98.82	–	–	–	395.28
Carbon monoxide	0.35	4.941	864.68	0.2	1.647	164.70	–	–	–	1,029.38
VOC	0.06	4.941	148.23	0.03	1.647	24.71	–	–	–	172.94
Lead	2.5E-05	4.941	0.06	8.9E-07	1.647	0.001	–	–	–	0.06
Mercury	b	4.941	b	b	1.647	b	–	–	–	0.0139
Beryllium	–	–	–	3.5E-07	1.647	0.0003	–	–	–	0.00029
Fluorides	–	–	–	6.27E-06	1.647	0.0052	–	–	–	0.005
Sulfuric acid mist	0.00060	4.941	1.48	0.0015	1.647	1.24	–	–	–	2.72
Total reduced sulfur	–	–	–	–	–	–	–	–	–	–
Asbestos	–	–	–	–	–	–	–	–	–	–
Vinyl Chloride	–	–	–	–	–	–	–	–	–	–
7% Coal Firing										
Particulate (TSP)	0.03	6.398	95.97	–	–	–	0.03	0.482	7.23	103.20
Particulate (PM10)	0.03	6.398	95.97	–	–	–	0.03	0.482	7.23	103.20
Sulfur dioxide	0.02	6.398	63.98	–	–	–	1.2	0.482	289.20	353.18 [a]
Nitrogen oxides	0.12	6.398	383.88	–	–	–	0.17	0.482	40.97	424.85 [a]
Carbon monoxide	0.35	6.398	1,119.65	–	–	–	0.2	0.482	48.20	1,167.85
VOC	0.06	6.398	191.94	–	–	–	0.03	0.482	7.23	199.17
Lead	2.5E-05	6.398	0.08	–	–	–	6.4E-05	0.482	0.02	0.10 [a]
Mercury	b	6.398	b	–	–	–	b	0.482	b	0.0139 [a]
Beryllium	–	–	–	–	–	–	5.9E-06	0.482	0.0014	0.0014 [a]
Fluorides	–	–	–	–	–	–	0.024	0.482	5.78	5.78 [a]
Sulfuric acid mist	0.00060	6.398	1.92	–	–	–	0.036	0.482	8.68	10.60 [a]
Total reduced sulfur	–	–	–	–	–	–	–	–	–	–
Asbestos	–	–	–	–	–	–	–	–	–	–
Vinyl Chloride	–	–	–	–	–	–	–	–	–	–

---

[a] Denotes maximum annual emissions for any fuel scenario.

[b] Refer to text for explanation.

Table 3-3. PSD Source Applicability Analysis for Osceola Power Limited Partnership Facility (Revised 05/03/93)

Regulated Pollutant	Baseline Emissions (TPY)	Cogeneration Facility Annual Emissions (TPY)	Net Change (TPY)	Significant Emission Rate (TPY)	PSD Applies ?
Particulate (TSP)	357.7	115.1[a]	-242.6	25	No
Particulate (PM10)	321.9	108.6[b]	-213.3	15	No
Sulfur dioxide	178.5	353.2	174.7	40	Yes
Nitrogen oxides	437.8	424.9	-12.9	40	No
Carbon monoxide	5,992.3	1,225.0	-4,767.3	100	No
Volatile org. compds.	208.6	210.0	1.4	40	No[c]
Lead	0.16	0.10	-0.06	0.6	No
Mercury	0.0137[d]	0.0139	0.0002	0.1	No
Beryllium	0.00002	0.00140	0.00138	0.0004	Yes
Fluorides	0.0079	5.78	5.8	3	Yes
Sulfuric acid mist	5.36	10.60	5.2	7	No
Total reduced sulfur	--	--	0	10	No
Asbestos	--	--	0	0.007	No
Vinyl Chloride	--	--	0	0	No

[a] Includes 105.0 TPY from boilers and 10.1 TPY from fugitive dust emission sources.

[b] Includes 105.0 TPY from boilers and 3.6 TPY from fugitive dust emission sources.

[c] Nonattainment review does not apply since the increase in VOC emissions is less than 40 TPY.

[d] The estimated annual average emission rate for the most recent 2 years is 0.0137 TPY. The highest annual emission rate for either of the last 2 years is 0.0139 TPY.

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316 ROYAL POINCIANA PLAZA
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FAX: (407) 659-3206

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Resources Ma[illegible]

May 4, 1993

Mr. Clair Fancy
Bureau Chief
Bureau of Air Regulation
Florida Department of Environmental Regulation
2600 Blair Stone Road
Tallahassee, FL 32399-2400

RE: Okeelanta Power Limited Partnership and Osceola Power Limited Partnership

Dear Mr. Fancy:

During a telephone discussion between David Buff of KBN and your staff last week concerning the pending draft air construction permits for the Okeelanta and Osceola cogeneration facilities, several issues were raised. The main issues appeared to be the overall magnitude of SO_2 emissions from the facilities, enforcement of a long-term average emission limit for SO_2, and the tying together of the two facilities. Presented below are suggested permit conditions which we believe would resolve the Department's concerns in these areas. These conditions reflect a significant reduction in worst-case year SO_2 emissions as compared to the worst-case year in our applications. In the applications, worst-year SO_2 emissions due to both facilities totalled 2,772 TPY. This is the level allowed under Sub-Part DA of the Federal NSPS for resource recovery facilities without scrubbers. Our proposal would reduce these emissions to less than 1,700 TPY. Additionally, our proposal will eliminate long-term averaging and the bubble concept.

Overall SO_2 Limits

The proposed conditions would limit maximum annual SO_2 emissions from all sources for the two facilities combined to 1,693.7 TPY. For a maximum heat input for the two facilities combined of 17.860×10^{12} Btu/yr (10.98×10^{12} Btu/yr for Okeelanta and 6.880×10^{12} Btu/yr for Osceola) when burning coal, the equivalent overall SO_2 emission rate is 0.19 lb/MMBtu. This emission rate is very close to what has been determined recently to be BACT for SO_2 for 100 percent coal-fired power plants (i.e., Bechtel Indiantown and OUC Stanton Unit 2). Additionally, the BACT cost effectiveness analysis included in the application demonstrates that the cost of control technology ranges from $3,100 to $25,000 per ton per year.

Individual, Annual Limits

In view of the Department's concerns related to tying the two facilities together in any manner, Okeelanta and Osceola are willing to accept limits on the individual facilities which would total 1,693.7 TPY. However, in order to provide more flexibility in burning supplementary fuels at Okeelanta, it is desired to permit the facilities for the following:

1. Okeelanta Power-- 1,340.5 TPY of SO_2 from all sources. This is equivalent to 0.24 lb/MMBtu at the maximum annual heat input of 10.98×10^{12} Btu/yr when burning 0.7% sulfur coal at an annual 19% capacity factor, with the remaining heat input due to biomass. Annual coal burning would be limited to 2.086×10^{12} Btu/yr, equivalent to a 19% capacity factor.

2. Osceola Power-- 353.2 TPY of SO_2 from all sources. This is equivalent to 0.10 lb/MMBtu at the maximum annual heat input of 6.88×10^{12} Btu/yr when burning 0.7% sulfur coal at an annual 7% capacity factor, with the remaining heat input due to biomass. Annual coal burning would be limited to 0.482×10^{12} Btu/yr, equivalent to a 7% capacity factor.

We are willing to accept these limitations as permit conditions if this would satisfy the Department's concerns and allow the permits to be issued.

It is noted that the 0.24 lb/MMBtu limit at Okeelanta, although somewhat higher than the Indiantown limit, is lower than the recently determined BACT by EPA Region IV for the OUC Stanton Unit #2 PSD permit.

I have asked David Buff of KBN to forward directly to you the tables which illustrate the calculation of maximum annual emissions and emission rates for each facility.

Fuel Oil

You also indicated the Department's intent to permit, as BACT, fuel oil with a sulfur content of no more than 0.05%. Although Okeelanta has concerns regarding the availability and price of this not yet available fuel, Okeelanta is willing to accept a permit condition which would require the use of 0.05% sulfur oil as soon as such fuel becomes generally available in the commercial stream as a boiler fuel. Since the facilities will not be constructed for another two years, and EPA is requiring this fuel to be used in all on-road diesel engines starting in October 1993, this new fuel will likely be available at a reasonable price. Any SO_2 emissions from the use of fuel oil would be included in the overall limits proposed above.

If this proposal does not meet with the Department's concurrence, we would like to schedule a meeting with yourself and your staff at your earliest convenience. We can then further discuss these issues as well as the time frame for issuing the draft permit.

Please call me at 407-996-9072, Ext. 204 or David Buff at 904-331-9000 if you have any questions concerning this information.

Sincerely,

Gus Cepero
Vice President

xc: Preston Lewis, FDER
Howard Rhodes, FDER
David Buff, KBN
David Dee, Carlton-Fields
Jewell Harper, EPA
John Bunyak, NPS
Bevin Beaudet, PBCHU
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W. Knowles, SFD
B. Brooks, SED
C. Holladay

OKEELANTA CORPORATION

Administration
P.O. Box 86
South Bay, FL 33493

Fax Transmittal Sheet

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Howard Rhodes

From: Gus Cepero

Date: 5/5/93

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POST OFFICE BOX 1059
PALM BEACH, FLORIDA 33480

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May 4, 1993

Mr. Clair Fancy
Bureau Chief
Bureau of Air Regulation
Florida Department of Environmental Regulation
2600 Blair Stone Road
Tallahassee, FL 32399-2400

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2. Osceola Power-- 353.2 TPY of SO_2 from all sources. This is equivalent to 0.10 lb/MMBtu at the maximum annual heat input of $6.88x10^{12}$ Btu/yr when burning 0.7% sulfur coal at an annual 7% capacity factor, with the remaining heat input due to biomass. Annual coal burning would be limited to $0.482x10^{12}$ Btu/yr, equivalent to a 7% capacity factor.

We are willing to accept these limitations as permit conditions if this would satisfy the Department's concerns and allow the permits to be issued.

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Sincerely,

Gus Cepero
Vice President

xc: Preston Lewis, FDER
Howard Rhodes, FDER
David Buff, KBN
David Dee, Carlton-Fields
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February 17, 1993

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Division of Air
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Mr. Willard Hanks
Bureau of Air Regulation
Florida Department of Environmental Regulation
2600 Blair Stone Road
Tallahassee, FL 32399-2400

Re: Flo-Energy and Sol-Energy Cogeneration Facility
Palm Beach County
AC50-219413, PSD-FL-196; AC50-219795, PSD-FL-197
[illegible] [illegible]

Dear Mr. Hanks:

This letter contains our response to the two letters dated December 24, 1992 and January 12, 1993, from the Palm Beach County Health Unit of HRS. This letter also contains our answer to a question raised by Cleve Holliday, FDER, during a telephone conversation about the ambient air quality impacts of sulfuric acid emissions from the proposed facilities and questions raised by Willard Hanks on applicability of nonattainment new source review and RACT requirements. Some additional information concerning BACT for SO_2 also is being provided.

Please call if you have any questions concerning this information. I am forwarding copies of this letter to EPA Region IV, Palm Beach County, and the National Park Service.

Sincerely,

David A. Buff

David A. Buff, M.E., P.E.
Principal Engineer
Florida Registration No. 19011

Seal

cc: Gus Cepero, Okeelanta
Don Schaberg, Okeelanta
David Dee, Carlton Fields
Jewell Harper, EPA Region IV
John Bunyak, NPS
Frank Gargiulo, PBCHU
David Knowles, FDER Ft. Myers
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1 ☒ Bill Sender
2 ☐ Bill Recipient's FedEx Acct No
3 ☐ Bill 3rd Party FedEx Acct No
4 ☐ Bill Credit Card
5 ☐ Cash/Check
[redacted]

4 **SERVICES** (Check only one box)

Priority Overnight (Delivery by next business morning†)	Standard Overnight (Delivery by next business afternoon†)
11 ☐ YOUR PACKAGING	51 ☐ YOUR PACKAGING
16 ☐ FEDEX LETTER *	56 ☐ FEDEX LETTER *
12 ☒ FEDEX PAK *	52 ☐ FEDEX PAK *
13 ☐ FEDEX BOX	53 ☐ FEDEX BOX
14 ☐ FEDEX TUBE	54 ☐ FEDEX TUBE

Economy Two-Day (Second business day†)
30 ☐

Government Overnight (Restricted to authorized users only)
46 ☐ GOV'T LETTER
☐ GOV'T PACKAGE

70 ☐ OVERNIGHT FREIGHT (Confirmed reservation) (Delivery commitment may be later in some areas) (150 lbs+)
TWO-DAY FREIGHT ** (Declared value limit $100) (Call for delivery schedule)

5 **DELIVERY AND SPECIAL HANDLING** (Check services required)
1 ☐ HOLD FOR PICK-UP (Fill in Box H)
2 ☐ DELIVER WEEKDAY
3 DELIVER SATURDAY (Extra charge) (Not available to all locations) ☐
4 ☐ DANGEROUS GOODS (Extra charge)
5 ☐
6 ☐ DRY ICE lbs
7 ☐ OTHER SPECIAL SERVICE
8 ☐
9 ☐ SATURDAY PICK-UP (Extra charge)
10 ☐
11 ☐
12 ☐ HOLIDAY DELIVERY (If offered) (Extra charge)

6 PACKAGES | WEIGHT In Pounds Only
Total | Total
[redacted]

DIM SHIPMENT (Chargeable Weight)
☐ lbs
___ × ___ × ___ = ___

Received At
☐ Shipper's Door ☐ Drop Box
☐ B.S.C.
☐ On-Call Stop ☐ Station

Emp No | Date
☐ Cash Received
☐ Return Shipment
☐ Third Party ☐ Chg To Del ☐ Chg To Hold
Street Address
City | State | Zip
Received By
X
Date/Time Received | FedEx Employee Number

7 Release Signature

FedEx Emp No | Date/Time

Federal Express Use
Base Charges
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Other 1
Other 2
Total Charges

REVISION DATE 4/91
PART #137204 FXEM 6/91
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RESPONSES TO LETTER DATED DECEMBER 24, 1992
FROM THE PALM BEACH COUNTY HRS HEALTH UNIT

1. BACT DETERMINATION FOR SO_2

Comment: We are confused as to the type of coal that was assumed for [the BACT] analysis of the three add-on controls:

- Furnace Injection W/Lime
- Duct Injection W/Lime
- Duct Injection W/Sodium

Was the 0.7% sulfur by weight coal used for the determination of the additional costs for these controls, or was a higher sulfur coal used? A higher sulfur coal would seemingly be less expensive and, with the assumed control efficiencies, could still provide less SO_2 emissions.

Response: Low sulfur coal (0.7% sulfur) was used throughout the BACT analysis to provide the greatest reduction in SO_2 emissions possible. A higher sulfur coal would be more expensive, but it would create higher SO_2 emissions. Coal cost is dependent on a number of factors, including length of contract, shipping distance, and type of coal. A review of the Department of Energy fuel cost report shows that 1.5%, 2.5% and 3.5% sulfur coal delivered to Florida are about \$3/ton, \$5/ton and \$7/ton, respectively, less than the cost of 0.7% sulfur coal. Flo-Energy would use a maximum of 112,750 tons per year (TPY) of coal, thus the maximum cost savings would be between \$338,000 and \$790,000 per year. Sol-Energy would have a maximum coal usage of 68,625 TPY, and the maximum annual savings would be between \$206,000 and \$480,000. These savings would be offset by increased capital and operational costs due to higher reagent injection rates, higher waste disposal amounts, increased energy and water requirements, and additional upsizing of the ESP and ash handling system.

To specifically analyze the economic impact of using higher sulfur coal, a detailed economic analysis was performed and is included in Attachment A. The energy and environmental impacts were also analyzed. It is concluded from the analysis that the economic, energy, and environmental impacts resulting from higher sulfur coal usage are adverse, and that the economic impacts are unreasonable. Therefore, the use of compliance coal not to exceed 25% heat input on an annual basis with SO_2 emissions not to exceed 1.2 lb/MMBtu, is BACT for the proposed facilities.

2. EMERGENCY AMMONIA RELEASE

Comment: In 1992, an ice block plant in Pahokee, Florida had just such a release during filling of the ammonia tank. (The facility uses anhydrous ammonia as a refrigerant.) The pressure relief valve released a large cloud of ammonia which slowly dissipated. Several people were exposed to relatively high levels of ammonia, with two persons being admitted to the local hospital.

Response: This comment concerns in-plant safety, which is regulated by the Occupational Safety and Health Administration (OSHA). As discussed in our letter dated December 4, 1992, an ammonia storage system, if used, will meet the American National Standards Institute (ANSI) K-61.1-1989 standard safety requirements for the storage and handling of anhydrous ammonia. The ANSI construction standards are designed to minimize the potential for an emergency release, but ANSI also established requirements and procedures to be used in the event of such a release. The procedures are designed to minimize the potential effects on persons exposed to ammonia fumes. Several pertinent parts of the ANSI standards are provided in Attachment B.

We are not familiar with the ammonia handling system and safety procedures used at the Pahokee ice plant and, therefore, we cannot comment on the problems encountered there.

3. ACTIVATED CARBON INJECTION (ACI)

Comment: Most of the literature on ACI indicates significantly increased control efficiencies when the operating temperature is kept below 300°F. Would it be possible for the applicant to reduce the operating temperature to this level without affecting performance or other control mechanisms?

Response: This question regarding mercury control has been addressed in David Dee's letter to the Department dated December 23, 1992.

RESPONSE TO DER FAX DATED JANUARY 5, 1993

1. Comment: Revised Table 6-18 for the Sol-Energy project shows arsenic impacts exceeding the no-threat level. Please discuss any measures that can be taken to mitigate this impact.

   Response: The maximum annual arsenic emissions and impacts are a result of using very conservative assumptions to predict worst-case conditions. Among other things, we assumed that 3 percent of the wood waste to be received at the facility would contain treated wood. Treated wood that contains chromium copper arsenate has an arsenic emission factor that is higher than clean wood waste. Assuming 3 percent treated wood is very conservative because Flo-Energy or Sol-Energy will not knowingly accept treated wood at the facility and, under the conditions imposed by Palm Beach County, they cannot knowingly accept such wood. Pursuant to the County's conditions, all contracts with wood waste suppliers must contain a provision that the wood waste must be substantially free from treated wood. Flo-Energy and Sol-Energy will also have the ability to inspect the suppliers' storage yards to insure they are taking measures to remove treated wood from the wood waste material. These measures will minimize the potential for treated wood to be present in the wood waste and, therefore, will reduce the potential for arsenic to be present.

   Even using these worst-case assumptions, the predicted maximum annual impact of arsenic (0.00029 $\mu g/m^3$) for the Sol-Energy facility exceeds FDER's proposed no threat level (NTL) (0.00023 $\mu g/m^3$) by only a very small margin. However, this maximum impact is approximately 10 times less than the EPA promulgated impact level of 0.0023 $\mu g/m^3$, which is based on a 1 in 100,000 risk for a permanent resident living 70 years at the point of maximum impact. The potential impacts drop to levels below the NTL at a distance of 4 km from the proposed facility. Since there are no residences or other dwellings located within 4 km of the sugar mill, FDER's NTL for long-term residential exposure should not be applied in this case.

2. Comment: Provide the impacts of sulfuric acid for comparison to the no-threat levels.

   Response: Based on the information already provided in the application, the maximum predicted impacts of sulfuric acid are as follows:

	Flo-Energy	Sol-Energy
Maximum sulfuric acid emissions (lb/hr)	52.8	33.2
8-hour impact ($\mu g/m^3$)	2.9	4.8
8-hour NTL ($\mu g/m^3$)	10.0	10.0
24-hour impact ($\mu g/m^3$)	2.2	3.2
24-hour NTL ($\mu g/m^3$)	2.4	2.4
Annual impact ($\mu g/m^3$)	0.16	0.2
Annual NTL ($\mu g/m^3$)	--	--

As shown, the maximum predicted impacts of the potential sulfuric acid emissions for the Flo-Energy or Sol-Energy projects are below the applicable FDER NTLs, except for sulfuric acid for the 24-hour averaging time for Sol-Energy. Modeling of the Sol-Energy facility alone demonstrates that the NTL for sulfuric acid will not be exceeded beyond 2.4 km from the facility. As discussed above for arsenic, there are no residential or other dwellings located within this area, only sugar cane fields. This impact will be mitigated also by the fact that the modeled sulfuric acid emissions, which are based upon 100 percent coal firing for the entire year, will occur less than 25 percent of the time during the year, since coal burning is limited to 25 percent firing during an annual period.

RESPONSES TO LETTER DATED JANUARY 12, 1993 FROM PALM BEACH COUNTY HRS

Comment: See attached Palm Beach County letter dated January 12, 1993.

Response: The permit applications and BACT analysis in this case have evaluated three scenarios that cover the potential range of operating conditions. The BACT analysis has addressed (a) the very worst potential case for a single year, which is 1,700 TPY SO_2 emitted at Flo-Energy only, with no SO_2 emitted at Sol-Energy; (b) a similar worst-case for a single year, with 1,072 TPY emitted at Sol-Energy and no SO_2 emitted at Flo-Energy; and (c) the worst case over the life of the projects, which is 1,000 TPY SO_2 for both facilities combined. We believe the analysis for the third case with 1,000 TPY emitted from both facilities combined, is more representative of actual operating conditions than the other extreme cases for the following reasons:

1. Coal is much more expensive than bagasse or other biomass fuels and, therefore, its use will be minimized.
2. A major portion of the biomass needs are already being procured, even though the facilities will not begin operating until 1996. The total biomass needs of the two facilities from outside sources, when considering the bagasse supplied from the sugar mills, will be between 500,000 and 600,000 TPY, assuming 100 percent biomass fuel firing. The applicants are in the process of negotiating contracts with Palm Beach County and Dade County to supply biomass fuel. Dade County would supply approximately 150,000 TPY of biomass. Palm Beach County would supply at least 85,000 TPY of biomass. The applicant's feasibility study conducted in 1991 concluded that large/ample quantities of biomass are available within reasonable distances of the proposed facilities. We have found that independent suppliers are eager to enter into long-term biomass supply agreements with Flo-Energy and Sol-Energy. For example, Flo-Energy is presently receiving over 60,000 TPY from these suppliers. Flo-Energy and Sol-Energy have not negotiated additional supplies because the fuel will not be needed until 1996, but the applicants are confident that all of their biomass needs can be met during normal conditions.
3. Based on historical trends and planned changes in cane harvesting methods, there is a strong likelihood that the supply of bagasse and cane trash will

increase from presently assumed levels over the next several years and certainly over the life of the project.

4. Given these facts, we believe the amount of coal burned at the facilities will be much less than 25 percent of the annual heat input, and the average SO_2 emissions will be much less than 1,500 TPY.
5. The BACT economic analysis should reflect long-term costs. Since the projects will limit SO_2 emissions to an average of 1,000 TPY on a long-term basis, this scenario should be given maximum consideration.

It should be noted that, even if 1,500 TPY of SO_2 emissions for both facilities is used for the BACT analysis, the cost effectiveness values for the SO_2 controls are very high and are economically unreasonable.

RESPONSE TO FDER FAX DATED JANUARY 13, 1993

Comment: What will be the maximum electrical production capacity of Sol-Energy?

Response: The normal, expected net electrical output delivered to Florida Power & Light Company (FPL) under the terms of the Power Purchase Agreement will be between 42 and 45 megawatts (MW). However, as stated in the response package for Sol-Energy dated December 4, 1993, Sol-Energy desires the option to produce, at times, up to 60 MW of electrical energy. This will be possible in the off-season when steam is not being sent to the sugar mill. This can be accomplished without increasing the size of the boilers or increasing air emissions. A relatively small increase in the size and cost of the steam turbine would allow Sol-Energy to produce 60 MW. All of the conditions in the county's zoning approval would be met with the higher rate of electrical production. Consequently, Sol-Energy wishes to obtain FDER approval for 60 MW of electrical production. If granted, Sol-Energy will discuss this increase with the county to determine whether the county will authorize the higher rate of energy production, subject to all of the county's other permit conditions.

ADDITIONAL INFORMATION FOR FLO-ENERGY AND SOL-ENERGY PROJECTS

During a meeting with EPA Region IV on November 4, 1992, EPA asked for additional information about two experimental SO_2 control techniques, which were identified by EPA Region IV personnel as "NOXSO" and "ADVACATE". This submittal contains the information requested about these two systems.

NOXSO Process--The NOXSO process is currently under development by NOXSO Corporation, MK-Ferguson Company, W.R. Grace & Company, and Ohio Edison Power. The process uses a post-combustion or add-on control device installed downstream from the particulate control unit, (e.g., an ESP for a coal-fired boiler). The NOXSO process employs a dry and regenerative sorbent to simultaneously remove NO_x and SO_2 from the boiler flue gas.

The sorbent is a high surface γ-alumina substrate impregnated with sodium, which combines with moisture to form sodium hydroxide (NaOH) at the reactive sites on the substrate surface. The flue gas must first be quenched to 120°C by spraying water into the duct work. As the flue gas contacts the sorbent, both NO_x and SO_2 react with these NaOH sites to form $NaNO_3$ and $NaSO_3$, respectively, and are absorbed onto the surface of the sorbent. The flue gas then leaves the NOXSO device, passes through a cyclone collector to remove spent sorbent, and continues on to the stack.

The $NaNO_3$ and $NaSO_3$-filled collected sorbent can then be regenerated via a two-step process for removing the NO_x and SO_2. The NO_x is released from the sorbent through a hot gas (660°F) heating process, and is then recycled back to the furnace. After the NO_x removal step, the sorbent is delivered to a fluidized regenerative bed where natural gas is used to reduce the sulfur compounds in the sorbent into SO_2 and H_2S. This off-gas stream is then processed by a Claus-type reactor where SO_2 and H_2S are converted to elemental sulfur. Solid sulfur powder is the byproduct of the NOXSO process.

The NOXSO process has been tested only at the pilot-scale level. The effects of the NO_x recycling on the boiler combustion process is still being investigated separately by Babcock & Wilcox at the Pittsburgh Combustion Technology Center. It has been theorized that the NO_x recycling with the NOXSO system will reach an equilibrium when the NO_x formation in the boiler is at a steady state. Currently, the Department of Energy Clean Coal Technology Program, the

Ohio Coal Development Office, the Electric Power Research Institute, the Gas Research Institute, and the East Ohio Gas Company are funding a project to conduct a full-scale demonstration of the NOXSO Process. This first full-scale demonstration will be conducted on a 115-MW boiler at Ohio Edison's Niles Power Plant in northeastern Ohio. Construction is scheduled to begin in early 1993, with plant startup scheduled in May 1994.

The NOXSO flue gas treatment system is not applicable to the proposed Flo-Energy and the Sol-Energy cogeneration projects for the following reasons:

1. Currently, the NOXSO technology is not commercially available.
2. The NOXSO process has been targeted for large coal-fired power plants using pulverized coal boiler and cyclone boiler technology. There has been no data developed for spreader stoker boilers such as those proposed for the Flo-Energy and Sol-Energy projects. Commercial implementation of the NOXSO process on a spreader stoker boiler is not planned and will not be achieved in the near future.
3. The sorbent regeneration step for the NOXSO process requires a high temperature (660°F) air stream and natural gas. Additionally, a Claus plant would have to be installed to convert the captured SO_2 and H_2S gasses into sulfur. Thus, more capital will be required for the additional fuel and equipment. Natural gas is not available at the Flo-Energy and Sol-Energy sites.

ADVACATE Process–The ADVACATE process is a technology being developed by the Air and Energy Engineering Research Laboratory (AEERL) of EPA and Acurex Corporation. This flue gas desulfurization process uses an **ADVAnced siliCATE** (ADVACATE) calcium silicate sorbent material. Calcium silicate is the main constituent in flyash. The ADVACATE sorbent is produced by mixing ground flyash with a hydrated lime $Ca(OH)_2$ slurry at a weight ratio of 3 to 1. The sorbent is injected into the flue gas duct upstream from the particulate control device.

Between 1987 and 1989, pilot test and optimization tests of the ADVACATE process were performed in batches. These tests were conducted on boilers equipped with fabric filters and they resulted in up to 80 percent SO_2 removal. However, these pilot testing programs did not include continuous recycle and none were performed on boilers equipped with an ESP for particulate control (EPA/600/D-90/147, 1990). Overall, the SO_2 removal rate was approximately 50 percent at normal flue gas temperature (194°F) and approximately 60 percent at saturation temperature (20°F) (EPA/600/J-92/047, 1992). Currently, the ADVACATE process is being applied in a field test on a 10-MW pilot plant at TVA's Shawnee facility. A full-scale demonstration will follow during the next 3 to 4 years prior to commercialization.

The ADVACATE process is not applicable for the proposed Flo-Energy and the Sol-Energy Cogeneration projects for the following reasons:

1. Currently, the ADVACATE technology is not commercially available.
2. The ADVACATE process has been developed for boilers equipped with fabric filters. The system's performance is unknown for boilers equipped with an ESP.
3. It is projected that the cost for the ADVACATE Process will be slightly higher than the cost of a lime sorbent injection system. It was estimated that the cost effectiveness for the lime sorbent injection system would be approximately $3,400 per ton of SO_2 removed for the proposed Flo-Energy and Sol-Energy projects. Therefore, the ADVACATE process will be even more economically infeasible for these proposed projects.

Additional literature on the NOXSO and ADVACATE processes is contained in Attachment C.

RESPONSE TO FDER QUESTIONS RELATED TO NONATTAINMENT REVIEW AND APPLICABILITY OF RACT REQUIREMENTS

Comment: On February 4, 1993, Mr. Willard Hanks of FDER requested clarification on the applicability of nonattainment new source review for VOC emissions for the proposed projects. The two projects are located in a designated nonattainment area for ozone. Mr. Hanks also inquired about the applicability of the recently promulgated reasonably available control technology (RACT) requirements for this moderate ozone nonattainment area.

Response: The applicability of PSD and nonattainment review regulations were discussed in Sections 3.4.1 and 3.4.2 of the air permit applications filed for Flo-Energy and Sol-Energy. As noted in the applications, nonattainment review for ozone is not required for the proposed cogeneration projects. The proposed projects are modifications to existing major sources of VOC emissions. FDER 17-212.500(2)(d)4.a., F.A.C., provides that a proposed modification to a major facility is subject to the preconstruction review requirements in Section 17-212.500(4) if the facility to be modified would be subject to preconstruction review pursuant to Rule 17-212.500(2)(d)2. "if it were itself a proposed new facility and modification would result in a significant net emissions increase [as set forth in Rule 17-212.500(2)(e)2., F.A.C.] of the affected pollutant." In this case, the modification will not result in a significant net emissions increase for the affected pollutant (VOC) and, therefore, the requirements for new source review for nonattainment areas are not applicable to the proposed facilities. Specifically, the Flo-Energy facility will result in a net decrease in VOC emissions; the Sol-Energy facility will increase VOC emissions by only 1.4 tons per year, which is far less than the significant emission rate for VOC of 40 tons per year. See Table 17-212.400-2, F.A.C. These FDER rules have been discussed with Mr. Larry George of FDER, who confirmed that nonattainment new source review should not be required for the proposed cogeneration projects.

According to the recently adopted RACT rules (Rule 17-296.570, F.A.C.), which apply to Dade, Broward and Palm Beach counties, the RACT requirements will apply to the proposed projects. The new rule requires that a RACT determination be made for all major NO_x and VOC emitting facilities which have not undergone nonattainment or prevention of significant deterioration (PSD) new source review for these two pollutants. Since Flo-Energy and Sol-Energy will not be subject to nonattainment or PSD new source review for NO_x or VOC, the RACT requirements will apply.

However, the rules clearly indicate that the RACT requirements and determination procedure are operating permit requirements, and not construction permit requirements. Therefore, it does not appear necessary to address the RACT requirements for these two projects at this time. Nevertheless, the applicants want to address the RACT requirements at this time so that the RACT technology and emission limits can be established prior to construction. The proposed RACT for the two facilities is presented below.

In regard to NO_x emissions, the two facilities will employ selective non-catalytic reduction (SNCR) to reduce NO_x emissions. SNCR is a technology which is being applied to wood-waste and municipal solid waste combustors as BACT (best available control technology under PSD new source review). Since by definition RACT is less stringent than BACT, SNCR technology is also acceptable as RACT (RACT being less stringent than BACT). The proposed RACT emission limits for NO_x are based on SNCR and are contained in the permit applications:

Flo-Energy-	0.15 lb/MM Btu for biomass and No. 2 fuel oil
	0.17 lb/MM Btu for coal
Sol-Energy-	0.12 lb/MM Btu for biomass and No. 2 fuel oil
	0.17 lb/MM Btu for coal

In regard to VOC emissions, the only feasible method of controlling VOC emissions from biomass and coal combustion sources is good combustion practices to minimize VOC emissions (while also minimizing NO_x emissions). Flo-Energy and Sol-Energy will achieve this through modern boiler design, which optimizes combustion efficiency. Therefore, the proposed RACT technology for VOC emissions is proper boiler operation to minimize VOC emissions (while meeting the NO_x emission limits). The proposed RACT emission limits for VOC are those contained in the permit applications:

Flo-Energy-	0.06 lb/MM Btu for biomass
	0.03 lb/MM Btu for coal and No. 2 fuel oil
Sol-Energy-	0.06 lb/MM Btu for biomass
	0.03 lb/MM Btu for coal and No. 2 fuel oil

Florida Department of Environmental Regulation
Twin Towers Office Bldg. ● 2600 Blair Stone Road ● Tallahassee, Florida 32399-2400
Lawton Chiles, Governor
Carol M. Browner, Secretary

FAX TRANSMITTAL SHEET

NAME(S): DAVID BUFF

DEPARTMENT/COMPANY: KBN

DATE: 1-13-93

PHONE: 904/332-4189

TOTAL NUMBER OF PAGES, INCLUDING COVER PAGE: 5

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FROM: WILLARD HANKS

DIVISION OF AIR RESOURCES MANAGEMENT

BUREAU: AIR REGULATIONS

OFFICE PHONE: 904/488-1344 FAX PHONE: (904)922-6979

SENDER: WILLARD HANKS

COMMENTS: PLEASE COMMENT ON THE ISSUES RAISED BY THE COUNTY:
① BASIS BACT FOR SO_2
② AGREEMENT THAT SOL-ENERGY WOULD BE A MAX OF 50 MW

HAVE A NICE DAY!

HRS
STATE OF FLORIDA
DEPARTMENT OF HEALTH AND REHABILITATIVE SERVICES

ESE-WPB

January 12, 1993

Mr. Willard Hanks
Florida Department of Environmental Regulation
Division of Air Resources
Bureau of Air Regulation
2600 Blair Stone Road
Twin Towers Office Building
Tallahassee, Florida 32399-2400

Re: Comments on the Letter from Carlton, Fields, Ward, Emanuel, Smith & Cutler, P.A., the Law Firm Representing the Two Cogeneration Projects of the "Flo-Sun" Corporation

Dear Mr. Hanks:

Our office received a copy of the above referenced letter on January 4, 1993 and offers the following comments:

1. The applicant suggests that the BACT analysis should be based on the Palm Beach County imposed long term average SO_2 emission limitation of 1000 TPY on the combined projects. This seems unreasonable since this limitation is based on the average SO_2 emissions over the life of the facility (estimated at 30 years) and is largely not enforceable. The county zoning conditions also provide SO_2 limits based on the biomass fuel that is supplied by the county. If the county makes at least 200,000 tons of biomass per year available, the applicant may emitt a maximum of 1500 TPY of SO_2, not to exceed an average of 1300 TPY for a five year period. If the county cannot make this quantity of biomass available, the applicant may emitt up to 1700 TPY of SO_2, not to exceed 1500 TPY for a ten year period. These limitations appear to be more enforceable, and therefore, we believe that at least 1500 TPY of SO_2 emissions should be used to determine BACT for these projects.

2. It is mentioned in the above referenced letter that BACT regulations indicate DER and EPA <u>may</u> consider actual operating conditions when determining BACT. From the information provided in the applications, it is difficult to determine what might constitute "actual" operating conditions for these facilities in the future. The applicant provides scenarios of only biomass fuel firing, 25% oil firing and 25% coal firing. The primary factor in predicting coal and oil usage is the availability of the biomass fuel which is entirely unknown, as recognized by the county and the applicant. If the biomass fuel is not available, the facilities will surely burn quantities of coal and/or oil necessary to profitably run these units. Therefore, we could realistically expect the following cases:

- Biomass fuel is readily available and SO_2 emissions are actually reduced
- Only a portion of the required biomass fuel is available and coal is used at only one facility

DISTRICT IX
PALM BEACH COUNTY HEALTH UNIT • P.O. BOX 29 • WEST PALM BEACH, FLORIDA 33402
LAWTON CHILES GOVERNOR

- Only a portion of the required biomass fuel is available and coal is used at both facilities
- Biomass availability is poor; bagasse is burned during sugar cane season and 25% coal is fired in the off-season.

It is impossible to predict which case will eventually define "actual" operating conditions in the future.

3. Consider the following information found in the applications, which assumes that both sites will operate at 25% Coal Firing:

	HEAT INPUT (BTU x 10^{12})		TPY OF FUEL	
	Biomass	Coal	Biomass	Coal
Sol-Energy	4.941	1.647	581,294	68,625
Flo-Energy	8.118	2.706	955,059	112,750
Total	13.059	4.353	1,536,353	181,375

1,536,353 TPY of Biomass
-1,130,000 TPY of Biomass from Sugar Mills
- 200,000 TPY of Biomas from PBC County Proposal
*206,353 TPY of Additional Biomass to Operate at capacity

*(This is about 1,754,000 MMBTU/Year of heat input, or about 10% of the total required heat input, assuming 4250 BTU/lb fuel, similar to bagasse).

In other words, if these projects secured twice the amount of biomass that Palm Beach County is proposing to supply, they would still have to fire 25% coal to come close to the design capacity. This is another reason for assuming that the maximum firing of coal should be used for determining BACT.

Thank you for the opportunity to comment on these applications.

Sincerely,

For the Division Director
Environmental Science and Engineering

Jeffery J. Koerner
Jeffery F. Koerner, Engineer IV
Air Pollution Contorl Section

FJG/JFK/lh

cc: AP-Source File

Petitioner: Osceola Farms Company
Agent: Daniel D. Ross, Esq.
Telephone: 655-6303

LOCATION: The subject property is located approximately one (1) mile north of the intersection of US 98 and Hatton Highway, eight (8) miles east of the City of Pahokee, in the Agricultural Production (AP) Zoning District.

REQUEST: The Osceola Farms Company has applied for a special exception to allow a public and private utility service (electrical power facility) in the AP Zoning District.

Uses proposed on site include general offices and industrial uses required to construct and operate the proposed facility, a reverse osmosis facility (for potable water), a waste management system, fuel storage, water detention, internal circulation and parking.

PETITION SUMMARY:

Proposed is a 42 mega watt co-generation electrical power facility on a 50 acre site located adjacent to an existing sugar mill. The existing sugar refinery includes a power facility which burns bagasse, a residue of the sugar grinding operation, and supplemental fuels to generate the electricity required by the sugar mill grinding and refining process. The existing power facility generates only enough electricity to operate the mill and is operational only during portions of the year.

The petitioner proposes to replace and upgrade the existing electric power facility and continue to generate steam to run the sugar mill. Excess electricity will be sold to the Florida Power and Light Company (FPL). The proposed facility will be in operation throughout the year and will burn bagasse and biomass waste (i.e. vegetative debris) brought to the site.

The petitioner has submitted a request to amend the application to request approval of a 42 net mega watt power facility (Exhibit A.1). To ensure that the net amount is attained the petitioner has requested that the application be amended to reflect a maximum 50 mega watt facility. In addition, the letter submitted by the applicant provides further details of the request, including accessory uses, disposal of waste water, potable water, sewage effluent and a definition of biomass waste.

ISSUES SUMMARY:

1. Comprehensive Plan: The Planning Division has reviewed the petition and determined that the request is consistent with the Comprehensive Plan.

2. Concurrency: The site has been issued a Concurrency Reservation Certificate for 217,800 square feet (7,500 s.f. general office, 210,300 s.f. general industrial).

3. Zoning: The subject property is zoned AP (Agricultural Production). The proposed use is allowed as a special exception in the AP Zoning District, subject to section 500.42 of the Zoning Code (Exhibit A.2).

4. Development of Regional Impact (DRI) Status: This petition has been reviewed to determine if it constitutes a DRI. Based on the standards contained within Chapter 380.06, F.S., the State of Florida Law governing such developments, this petition does not exceed the defined thresholds (Exhibit B).

H. **LIGHTING**

1. All outdoor lighting used to illuminate the premises and identification signs shall be of low intensity, shielded and directed downward. (BUILDING - CODE ENF)

I. **PARKING**

1. Vehicle parking shall be limited to the parking areas designated on the approved site plan. No parking of vehicles shall be permitted in landscaped areas, right-of-way or interior drives. (CODE ENF)

J. **TRANSMISSION LINES**

1. Transmission line easements required by this facility to be utilized for the construction of above-ground transmission lines shall not be located within 1,000 feet of any property designated as residential by the Palm Beach County Comprehensive Plan. Transmission line easements required by this facility to be utilized for the construction or installation of underground transmission lines shall not be located within 200 feet of any property designated as residential by the Palm Beach County Comprehensive Plan. Plans for final transmission line corridors shall be submitted to the Zoning Director prior to final adoption. (ZONING)

2. All transmission lines required by this facility are to be constructed in accordance with the National Electric Safety Code. (BUILDING)

3. All transmission lines leaving the site and required by this facility shall not exceed 138 KV. (BUILDING)

* K. **USE LIMITATIONS** *

1. Use of the site shall be limited as follows:

Land Area	50.00 acres
Total Floor Area	217,800 square feet
Maximum Floor Area	10%
Electrical Production	50 mega watt maximum
Fuel Yard	35 acre max. net land area

2. Prior to site plan certification, the site plan shall be amended to indicate the location of a truck/vehicle wash facility. This wash facility shall utilize a 100% water recycling system. (ZONING - BUILDING)

3. There shall be no repair or maintenance of vehicles on site. (CODE ENF)

4. No outside storage of disassembled vehicles, or parts thereof, shall be permitted on site. (CODE ENF)

5. The maximum height, from grade to highest point, for all fuel storage areas shall not exceed fifty (50) feet. (BUILDING)

6. Onsite storage shall be contained within the area designated on Exhibit 48 and shall be processed and stored in a manner which controls fugitive and dust particulate emissions. (CODE ENF)

7. All vehicles utilizing public rights-of-way to carry biomass waste (i.e. vegetative matter) to the site shall be equipped, at a minimum, with covering or screens over top of the open bed of the vehicle to prevent the loss of material during transportation to the facility. (CODE ENF)

ATTACHMENT A

REVISED BACT ANALYSIS INCORPORATING HIGHER SULFUR FUEL OPTIONS

REVISED BACT ANALYSIS
INCORPORATING HIGHER SULFUR FUEL OPTIONS

INTRODUCTION

The following BACT analysis study has been performed in response to the request from Palm Beach County to investigate various alternatives of using coals with higher sulfur content than the 0.7 percent sulfur coal proposed for use at the Flo-Energy and Sol-Energy cogeneration projects. A range of sulfur content in coal was considered, consisting of 1.5 percent, 2.5 percent and 3.5 percent sulfur by weight. These three sulfur content levels, along with 0.7 percent sulfur coal, are typical grades of coal delivered to utilities in Florida.

This BACT analysis for medium and high sulfur coals evaluates technologies identified previously (i.e., lime spray dryer, limestone wet scrubber, and dry sorbent injection) to control the emissions of SO_2 and other acid gases. Dry sorbent injection (DSI) processes includes the lime furnace injection, lime duct injection, and dry-sodium duct injection. A general description of each control technology has been described in previous submittals.

The BACT "top-down" hierarchy for controlling SO_2 emissions for the Flo-Energy and Sol-Energy facilities is presented in Table A. As shown in the table, an SO_2 emission level in pounds per million Btu (lb/MMBtu) has been calculated and assigned to all eleven options. The baseline SO_2 emission level chosen is the proposed uncontrolled emissions from firing 0.7 percent sulfur coal, which is equivalent to 1.2 lb/MMBtu.

The alternative of using the DSI process with the high sulfur coal (i.e., 2.5 percent and 3.5 percent sulfur) options has been eliminated because these options result in higher SO_2 emissions than the proposed BACT of firing 0.7 percent sulfur coal. The uncontrolled SO_2 emissions from 0.7 percent sulfur coal are 1,623.6 TPY and 988.2 TPY for the Flo-Energy and the Sol-Energy facilities, respectively. These uncontrolled SO_2 emissions are based on the proposed coal-firing limit of a maximum 25 percent of the total heat input for the boilers at each facility. By comparison would the DSI-controlled SO_2 emission level of 1.9 lb SO2/MMBtu for 2.5 percent sulfur coal result in annual SO_2 emissions of approximately 2,570 TPY for Flo-Energy and 1,565 TPY for Sol-Energy. These annual emission estimates were calculated from the maximum heat inputs due to coal-firing (i.e., 2.706 x 10^6 MMBtu/yr for Flo-Energy and 1.647 x 10^6 MMBtu/yr for Sol-Energy). The DSI-controlled SO_2 emissions from firing 3.5 percent sulfur coal would even be greater than the 2.5 percent sulfur coal option.

BACT EVALUATION OF MEDIUM AND HIGH SULFUR COAL OPTIONS

The following evaluation of the lime spray dryer, the limestone wet scrubber, and the DSI processes as the SO_2 control technologies for various grades of coal will focus on four main aspects of the BACT analysis: technical issues, environmental effects, energy requirements and impacts, and economics.

TECHNICAL ISSUES

Both the lime spray dryer and the limestone wet scrubber are well-demonstrated SO_2 control technologies for medium and high sulfur coal. There are no additional technical issues other than those presented previously concerning the application of either process to the low-sulfur coal (0.7 percent) option. Since coal will be fired only as supplemental fuel, the main technical issue with the low-sulfur coal option was the maintenance of the auxiliary equipment (i.e., lime slaker, slurry tanks, dewatering equipment for sludge, etc.) frequent start-ups, and frequent periods of prolonged inactivity. For the medium to high sulfur coal, the auxiliary equipment will be even larger than similar equipment specified for the low-sulfur coal. Thus, the maintenance problems will be magnified to an even greater extent.

The DSI processes can be used with the 1.5 percent sulfur coal. However, the dry-sodium injection process has only been demonstrated on low-sulfur coal in the mid-west region. There is little historical experience with the DSI processes for the type of boiler (spreader stoker) proposed for the project, but a higher normalized sorbent to SO_2 ratio is expected to be required with the higher sulfur coal.

For all SO_2 control processes, larger ESP units and ash handling systems will be required for handling the additional particulate loading. Even larger ESP units must be designed for the DSI processes, compared to the wet scrubber or spray dryer, because of the lower operating efficiency of the ESP at the higher temperature range.

As in the previous BACT analysis, the limestone wet scrubber can achieve an SO_2 removal efficiency of 95 percent, and the lime spray dryer can achieve an SO_2 removal efficiency of 92 percent. The DSI processes can achieve a maximum SO_2 removal efficiency of 50 percent.

ENVIRONMENTAL EFFECTS

There are several adverse environmental impacts when using medium and high sulfur coals with flue gas controls, compared to the proposed low-sulfur (0.7 percent sulfur) coal option with no add-on controls. These adverse impacts include higher water demand for makeup water, lower plume rise and hence less dispersion of emissions, higher plume visibility, and higher volume of discharged solid wastes.

The amount of fly ash and solid waste generated from controlling the SO_2 emissions from medium and high sulfur coals will be at least twice as great as the fly ash produced by the combustion process. The total solid waste produced will be increased by approximately 115 percent, 260 percent, and 400 percent for any one of the mentioned control technologies when the sulfur content increases to 1.5 percent, 2.5 percent, and 3.5 percent, respectively. For the wet scrubber and the spray dryer processes, SO_2 emission control of 0.7, 1.5, 2.5, and 3.5 percent sulfur coals will produce approximately 11,000, 18,000, 27,000, and 36,000 tons per year, respectively, of additional solid waste for the Flo-Energy facility. Similarly, the total amount of solid waste produced at the Sol-Energy facility will be 6,400, 11,000, 16,500, and 22,000 tons per year, respectively.

Compared to the air quality impacts of the uncontrolled low-sulfur coal option, the predicted ground-level ambient impacts for all other control options will be greater for every pound of pollutant emitted. The higher predicted ground level impact is caused by the lowered stack exhaust temperature due to the humidification step associated with all SO_2 control methods with the exception of the dry-sodium duct injection process. In general, the humidification step cools down the gas stream and enhances the reaction between the SO_2 gas and the sorbent materials. The differential temperature drops are approximately 80 degrees for the lime-based DSI processes and for the spray dryer control, and 220 degrees for the wet scrubber process from the 350°F stack temperature of the uncontrolled exhaust gas.

Compared to the negligible visible emissions associated with firing low-sulfur coal, the low stack temperature and the saturated condition of the flue gas from the wet scrubbing process has been shown to generate a wet plume that is highly visible. The dry-sodium sorbent injection process has been shown to generate a brownish plume due to the formation of NO_2 when the system would be operated at above 30 percent SO_2 removal efficiency. For the proposed projects, the system would be operated above this level. Therefore, visible emissions will be expected to increase when either the wet scrubber or the dry-sodium injection system is used.

Water demands for humidifying the SO_2 emission control processes are about 16 million gallons per year for Sol-Energy and about 40 million gallons for Flo-Energy. Those water demands are just for the humidification system. Additional water will be required for the raw sorbent material processing (i.e., preparing the lime or limestone). Approximately 2 to 12 million gallons of water will be required as make-up water for the spray dryer or limestone wet scrubber.

ENERGY REQUIREMENT AND IMPACTS

The lime spray dryer, the limestone wet scrubber, and the DSI processes require additional electricity to drive mechanical equipment, including agitators, slurry pumps, air sparge systems, waste dewatering systems, conveyor belts, feeder systems, pulverizers, and air handlers. Additional electricity is required for operating larger ESP units and larger ash handling equipment.

For Flo-Energy, the estimated additional energy requirements to control SO_2 emissions from firing low (0.7 percent) to high (3.5 percent) sulfur coals are as follows: 1,700 to 2,500 megawatt-hours per year (MW-hr/yr) for using the wet scrubbing process; and 1,150 to 1,700 MW-hr/yr for using the spray dryer process. Approximately 360 to 500 MW-hr/yr would be required for using the DSI process on low and medium (1.5 percent) sulfur coals. These estimated energy requirements are calculated by assuming the maximum proposed coal-firing for the Flo-Energy facility, and that the SO_2 removal equipment or process would only be operating during coal-firing.

For the Sol-Energy facility, the estimated additional energy requirements to control SO_2 emissions from firing low to high sulfur coals are as follows: 720 to 1,000 MW-hr/yr for using the wet scrubbing process; 480 to 660 MW-hr/yr for using the spray dryer process; and 220 to 310 MW-hr/yr for using the DSI process on low and medium sulfur coals.

ECONOMIC ANALYSIS

The cost analysis has been revised to include the medium and high sulfur coals. The detailed cost analysis of the capital cost, operating cost and annualized cost estimates for the Flo-Energy facility are presented in the attached Tables 1 to 10. Similarly, the detailed cost analysis of the capital cost, operating cost and annualized cost estimates for the Sol-Energy facility are presented in the attached Tables 11 to 20. Each set of 10 tables consists of a cost matrix that includes four grades of sulfur coal (i.e., 0.7, 1.5, 2.5, and 3.5 percent sulfur) and five SO_2 control technologies (i.e., spray dryer, wet scrubber, furnace injection, lime duct injection, and dry-sodium duct injection).

The cost analysis follows EPA's recommended procedure that is outlined in the OAQPS Control Cost Manual. The basic equipment costs were obtained from various budgetary pricing quoted by equipment vendors. Operating and material unit costs were either obtained from current trade publications or suppliers. The cost analysis reflects the differential economic burdens to achieve the emission reductions below the proposed SO_2 emission levels for both Flo-Energy and Sol-Energy facilities.

In the revised cost analysis, the basic equipment costs for the spray dryer and the direct installed cost for the DSI processes are the same as presented in the previous submittal. The direct capital cost estimates for the spray dryer and the DSI processes include the costs of upsizing both the ESP units and the ash handling systems. The indirect installation costs have been added to the cost analysis for the DSI processes, since these were previously omitted. The direct installed cost previously provided for the wet scrubber option was based upon general cost factors. These costs have now been replaced with actual vendor estimates provided by ABB Environmental Systems. This includes the sludge handling system. Other cost factors are similar to those presented in previous submittals.

A summary of the cost analysis for the SO_2 emission reduction options for Flo-Energy, Sol-Energy, and for the combined facilities is presented in the following sections.

Cost Analysis Summary for Flo-Energy

The total capital investment (TCI) to control the SO_2 emissions from firing 0.7 percent sulfur coal are estimated at \$11.24 million for the spray dryer, \$29.90 million for the wet scrubber, and \$9.04 million for the dry-sodium duct injection process. (The dry-sodium injection process represents all three DSI processes in the BACT analysis because of its lowest cost effectiveness among the group). The annualized costs (AC) are estimated at \$5.19 million for the spray dryer, \$9.19 million for the wet scrubber, and \$3.78 million for the dry-sodium duct injection process.

The TCI to control the SO_2 emissions from firing 1.5 percent sulfur coal are estimated at \$12.04 million for the spray dryer, \$30.37 million for the wet scrubber, and \$9.88 million for the dry-sodium duct injection process. The AC are \$6.29 million for the spray dryer, \$9.49 million for the wet scrubber, and \$4.97 million for the dry-sodium duct injection process.

The TCI to control the SO_2 emissions from firing 2.5 percent sulfur coal are estimated at \$13.43 million for the spray dryer, and \$30.86 million for the wet scrubber. The AC are estimated at \$7.91 million for the spray dryer, and \$9.90 million for the wet scrubber.

The TCI to control the SO_2 emissions from firing 3.5 percent sulfur coal are estimated at \$14.49 million for the spray dryer, and \$31.52 million for the wet scrubber. The AC are estimated at \$9.37 million for the spray dryer, \$10.53 million for the wet scrubber.

It should be noted that the annualized costs for each of the control options include the annual cost savings due to the lower cost of higher sulfur coal versus the cost of 0.7 percent sulfur coal. However, in all cases, this savings is offset by the increased costs of controlling the dirtier fuel.

A revised summary of the BACT analysis for the proposed Flo-Energy facility is presented in Table B, and incorporates all of the FGD technologies considered in the order of their "top-down" ranking. Based on the EPA's "top-down" BACT guidance document (EPA, 1990), the cost effectiveness was calculated for each technology by dividing the total annualized cost by the emission reduction over the baseline. As shown in Table B, the cost effectiveness values over the baseline for the proposed Flo-Energy facility range from \$3,477 to \$73,498 per ton of SO_2 removed for all 10 options considered.

The incremental cost effectiveness was calculated by comparing options and eliminating the inferior control options which provide less total SO_2 reduction for a higher total AC. By this method, all of the control options involving the medium and high sulfur coals are considered as inferior options since they cost more but result in higher SO_2 emissions than the low sulfur coal case. Thus, the incremental cost effectiveness values were calculated using the remaining options (with the low-sulfur fuel only), as presented in Table B. As shown, the incremental cost effectiveness for the dry-sodium duct injection process with low sulfur coal is approximately $4,661 per ton of SO_2 removed for the proposed Flo-Energy facility. This cost is considered unreasonable for the Flo-Energy facility, and well above cost effectiveness values which have been deemed reasonable (i.e., $2,000 per ton). The spray dryer and wet scrubber options with low sulfur coal have similar high cost effectiveness values of $3,477/ton and $5,959/ton. Each of these are likewise considered unreasonable.

The removal of other acid gases (i.e., sulfuric acid mist, hydrofluoric acid mist, and hydrochloric acid mist) was also included in the cost analysis for the Flo-Energy facility. The results, presented in Table C, show sightly lower cost effectiveness values. The cost effectiveness values over baseline range from $3,147 to $25,290 per ton of SO_2 removed for all 10 options considered. As discussed above, these values are well above the $2,000/ton value deemed reasonable.

Cost Analysis Summary for Sol-Energy

The TCI to control the SO_2 emissions from firing 0.7 percent sulfur coal at the Sol-Energy facility are estimated at $7.48 million for the spray dryer, $19.93 million for the wet scrubber, and $5.70 million for the dry-sodium duct injection process (dry-sodium injection process represents all three DSI processes in the BACT analysis because of its lowest cost effectiveness among the group). The AC are estimated at $3.36 million for the spray dryer, $6.05 million for the wet scrubber, and $2.37 million for the dry-sodium duct injection process.

The TCI to control the SO_2 emissions from firing 1.5 percent sulfur coal are estimated at $8.01 million for the spray dryer, $20.24 million for the wet scrubber, and $6.24 million for the dry-sodium duct injection process. The AC are $4.05 million for the spray dryer, $6.24 million for the wet scrubber, and $3.10 million for the dry-sodium duct injection process.

The TCI to control the SO_2 emissions from firing 2.5 percent sulfur coal are estimated at $9.46 million for the spray dryer, and $20.57 million for the wet scrubber. The AC are estimated at $5.20 million for the spray dryer, and $6.50 million for the wet scrubber.

The TCI to control the SO_2 emissions from firing 3.5 percent sulfur coal are estimated at $9.64 million for the spray dryer, and $21.0 million for the wet scrubber. The AC are estimated at $5.96 million for the spray dryer, $6.88 million for the wet scrubber.

As for the Flo-Energy analysis, the cost savings of the higher sulfur coal has been considered in the analysis.

A revised summary of the BACT analysis for the proposed Sol-Energy facility is presented in Table D, and incorporates all of the FGD technologies considered in order of their "top-down" ranking. Based on the EPA's "top-down" BACT guidance document, the cost effectiveness was calculated for each technology by dividing the total annualized cost by the emission reduction over baseline. As shown in Table D, the cost effectiveness values for the proposed Sol-Energy facility range from $3,698 to $75,305 for all 10 options considered. As in the case of Flo-Energy, these costs are well above $2,000 per ton and are considered unreasonable.

The removal of other acid gases was also included in the cost analysis for the Sol-Energy facility, and is presented in Table E. The cost effectiveness values over baseline range from $3,346 to $25,865 per ton of SO_2 removed for all 10 options considered. Even considering the additional acid gases removed, the cost effectiveness values remain well above the level considered reasonable.

Cost Analysis Summary for the Combined Flo-Energy and Sol-Energy Facilities

The cost effectiveness over baseline and the incremental cost effectiveness for the combined Flo-Energy and Sol-Energy facilities are presented in Tables B, C, D, and E. These cost effectiveness values are based on the proposed SO_2 emission limit of 1,000 TPY for both facilities, equivalent to a total annual amount of 69,444 tons of coal fired. The cost effectiveness for both facilities range from $9,167 to $193,753 per ton of SO_2 removed. When the removal of other acid gases is also included, the cost effectiveness for both facilities ranges from $8,308 to

$66,701 per ton of acid gases removed. These cost effectiveness values are very high and support the economic infeasibility of add-on flue gas controls when coal will only be burned for a maximum of 25 percent on an annual basis.

BACT CONCLUSIONS

The limestone wet scrubber and the lime spray dryer can reduce the SO_2 emissions from medium to high sulfur coals to levels that are less than the proposed uncontrolled SO_2 emissions from firing low-sulfur coal. Dry-sorbent injection (representing the DSI processes), was only evaluated for the medium-sulfur (1.5 percent sulfur) coal because higher sulfur coal with this option produces SO_2 emissions greater than the proposed emissions. Thus, 10 applicable control scenarios were included in the BACT analysis.

Concerning the technical issues, all control options present more complicated operating and maintenance problems than the proposed low-sulfur coal case because of the expected frequent start-up and shutdown of the control devices. The operating and maintenance problems will be increasingly worse for options with higher sulfur coals and the associated larger auxiliary systems. The proposed firing of low-sulfur coal can avoid this complication; therefore, it is the most preferred option when evaluating the technical aspect of BACT.

In the areas of environmental effects and energy impacts, all 10 options with add-on SO_2 controls have greater solid waste disposal amounts, increased visible emissions, less dispersion of pollutants, and higher water and energy demands. For medium and higher sulfur coal, these impacts are greater than the impacts projected for the options using low-sulfur coal. Some of the concerns are listed as follows:

- The amount of additional solid waste generated from add-on SO_2 controls and low-sulfur coal will be approximately 11,000 TPY for the Flo-Energy facility and 6,400 TPY for the Sol-Energy. For the medium and high sulfur coal options with 1.5 percent, 2.5 percent, and 3.5 percent sulfur, the amount of solid waste will increase about 115 percent, 260 percent, and 400 percent, respectively, over the amount estimated for the low-sulfur coal option for each proposed facility.

- Brown plume visible emissions will be associated with the low and medium sulfur coal options incorporating dry-sorbent injection. A visible steam plume will be associated with all fuel options incorporating the wet scrubbing process.
- Additional water demands will result from all control options, with the exception of the dry-sodium injection process. Water is needed for humidifying the boiler flue gas for better SO_2 removal efficiency in the lime furnace injection process, the lime duct injection process, the lime spray dryer, and the limestone wet scrubber. Higher water demands are associated with the lime spray dryer and the limestone wet scrubber. About 18 million and 42 million gallons per year of water are needed for the spray dryer for Sol-Energy and Flo-Energy, respectively. Also, about 18 million and 52 million gallons per year of water are needed for the wet scrubber for Sol-Energy and Flo-Energy, respectively. Disposal of the wet scrubber effluent is also a concern in the south Florida area, where all potable water requirements are met through the use of groundwater.
- Additional electrical power is needed for all control options. These additional electrical demands are required for powering mechanical devices associated with the auxiliary systems, for the larger ESP, and for handling the additional ash collected in the ESP. For Flo-Energy, the estimated additional energy requirements are 1,700 to 2,500 MW-hr/yr for the wet scrubbing process; 1,150 to 1,700 MW-hr/yr for the spray dryer process; and 360 to 500 MW-hr/yr for the DSI process on low and medium (1.5 percent) sulfur coals. For the Sol-Energy facility, the estimated additional energy requirements to control SO_2 emissions are 720 to 1,000 MW-hr/yr for the wet scrubbing process; 480 to 660 MW-hr/yr for the spray dryer process; and 220 to 310 MW-hr/yr for the DSI process on low and medium sulfur coals.

The use of low-sulfur coal does not have any adverse environmental effects or require any additional energy. Therefore, the proposed use of low-sulfur coal is the most preferred option when evaluating the environmental effects and the energy aspects of BACT.

The economic analysis shows that the cost effectiveness over baseline for all 10 control options are at least $3,400 per ton of SO_2 removed for the worst-case year for the Flo-Energy and Sol-Energy facilities. The incremental cost effectiveness values for the alternative control options are at least \$4,600 per ton of SO_2 removed. These cost effectiveness values have been deemed unreasonable in previous BACT determinations for SO_2. In the case of the combined Flo-Energy and Sol-Energy facilities, which are subject to Palm Beach County's SO_2 emission limit of 1,000 TPY, both the cost effectiveness over baseline and the incremental cost effectiveness are approximately 2.5 times higher than the values for the worst-case year for each individual facility.

When the removal of additional acid gases is included in the cost analysis, the cost effectiveness and the incremental cost effectiveness values are lowered only slightly. As shown in Tables C and E, the lowest combined (SO_2 and other acid gas) cost effectiveness value is greater than \$3,100 per ton. Similarly, the incremental cost effectiveness values for the alternative control options are at least \$3,900 for the Flo-Energy and Sol-Energy facilities. These cost effectiveness values are unreasonable.

In summary, these high cost effectiveness values render all of the SO_2 control alternatives economically infeasible. The environmental and energy impacts of these alternatives also are significant. Therefore, the use of low-sulfur coal is BACT for the proposed Flo-Energy and Sol-Energy cogeneration facilities.

Table A. BACT "Top-down" Hierarchy of SO2 Reduction Methods for Flo-Energy and Sol-Energy Cogeneration Facilities.

Top-down Ranking	Technology	Range of Control Effectiveness (%)	Control level for BACT Analysis (%)	SO2 Emission Level (lb/MMBtu)	Annual Emissions for Flo-Energy* (TPY)	Annual Emissions for Sol-Energy** (TPY)
1st	Wet Limestone Scrubber (0.7% S Coal)	80-95	95	0.06	81.2	49.4
2nd	Spray Dryer (0.7% S Coal)	80-92	92	0.10	129.9	79.1
3rd	Wet Limestone Scrubber (1.5% S Coal)	80-95	95	0.12	155.6	94.7
4th	Spray Dryer (1.5% S Coal)	80-92	92	0.18	249.0	151.5
5th	Wet Limestone Scrubber (2.5% S Coal)	80-95	95	0.19	257.1	156.5
6th	Wet Limestone Scrubber (3.5% S Coal)	80-95	95	0.28	372.1	226.5
7th	Spray Dryer (2.5% S Coal)	80-92	92	0.30	411.3	250.3
8th	Spray Dryer (3.5% S Coal)	80-92	92	0.44	595.3	362.3
9th	Dry-Sodium Duct Injection (0.7% S Coal)	40-50	50	0.60	811.8	494.1
10th	Dry-Sodium Duct Injection (1.5% S Coal)	40-50	50	1.15	1,555.9	947.0
Baseline	Baseline (0.7 Wt% S Coal Uncontrolled)	--	--	1.20	1,623.6	988.2

* Total for the three boilers based on maximum heat input rate of 2.706 E+12 Btu/yr from coal firing only.

** Total for the two boilers based on maximum heat input rate of 1.647 E+12 Btu/yr from coal firing only.

Table B. Summary of Top–Down BACT Impact Analysis Results for SO2, Flo–Energy Cogeneration Facility (Revised 02/05/93).

				Environmental Impacts		Energy Impacts	Economic Impacts			
Control Alternative	Total SO2 Emissions (TPY)	Emissions Reduction Over Baseline (TPY)	Incremental Emission Reduction (TPY)	Air toxics impact? (Yes/No)	Adverse enviromental impacts? (Yes/No)	Additional Energy Requirements Electricity (MW–hr/yr)	Total Annualized Cost ($/yr)	Incremental Annualized Cost ($/yr)	Cost Effectiveness Over Baseline ($/ton)	Incremental Cost Effectiveness ($/ton)
Flo–Energy Facility Only – Worst Case Year										
Wet Limestone Scrubber (0.7% S Coal)	81.2	1,542.4	48.7	Yes	Yes	1,700	$9,191,457	$3,997,473	$5,959	$82,070
Spray Dryer (0.7% S Coal)	129.9	1,493.7	681.9	No	Yes	1,150	$5,193,984	$1,410,197	$3,477	$2,068
Wet Limestone Scrubber (1.5% S Coal)	155.6	1,468.0	--	Yes	Yes	1,950	$9,494,589	--	$6,468	--
Spray Dryer (1.5% S Coal)	249.0	1,374.6	--	No	Yes	1,300	$6,286,174	--	$4,573	--
Wet Limestone Scrubber (2.5% S Coal)	257.1	1,366.5	--	Yes	Yes	2,250	$9,899,057	--	$7,244	--
Wet Limestone Scrubber (3.5% S Coal)	372.1	1,251.5	--	Yes	Yes	2,500	$10,525,717	--	$8,410	--
Spray Dryer (2.5% S Coal)	411.3	1,212.3	--	No	Yes	1,500	$7,912,756	--	$6,527	--
Spray Dryer (3.5% S Coal)	595.3	1,028.3	--	No	Yes	1,700	$9,371,187	--	$9.113	--
Dry–Sodium Duct Injection (0.7% S Coal)	811.8	811.8	811.8	No	Yes	360	$3,783,787	$3,783.787	$4.661	$4.661
Dry–Sodium Duct Injection (1.5% S Coal)	1,555.9	67.7	--	No	Yes	500	$4,972,120	--	$73,498	--
Baseline (0.7 Wt% S Coal Uncontrolled)	1,623.6	--	--	--	--	--	--	--	--	--
Flo–Energy and Sol–Energy Facilities Combined – Average SO2 Emissions										
Wet Limestone Scrubber (0.7% S Coal)	41.7	958.3	25.0	Yes	Yes	2,420	$15,243,000	$6,687,000	$15,906	$267,482
Spray Dryer (0.7% S Coal)	66.7	933.3	433.3	No	Yes	1,630	$8,556,000	$2,401,000	$9,167	$5,541
Wet Limestone Scrubber (1.5% S Coal)	95.8	904.2	--	Yes	Yes	2,760	$15,736,000	--	$17,404	--
Spray Dryer (1.5% S Coal)	153.3	846.7	--	No	Yes	1,840	$10,334,000	--	$12,206	--
Wet Limestone Scrubber (2.5% S Coal)	158.3	841.7	--	Yes	Yes	3,150	$16,389,000	--	$19,472	--
Wet Limestone Scrubber (3.5% S Coal)	229.2	770.8	--	Yes	Yes	3,500	$17,406,000	--	$22,581	--
Spray Dryer (2.5% S Coal)	253.3	746.7	--	No	Yes	2,100	$13,110,000	--	$17,558	--
Spray Dryer (3.5% S Coal)	366.7	633.3	--	No	Yes	2,360	$15,330,000	--	$24,205	--
Dry–Sodium Duct Injection (0.7% S Coal)	500.0	500.0	500.0	No	Yes	580	$6,155,000	$6,155,000	$12,310	$12,310
Dry–Sodium Duct Injection (1.5% S Coal)	958.3	41.7	--	No	Yes	810	$8,073,000	--	$193,753	--
Baseline (0.7 Wt% S Coal Uncontrolled) [1]	1,000.0	--	--	--	--	--	--	--	--	--

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[1] The amount of coal associated with the proposed SO2 emission limit of 1,000 TPY for both Flo–Energy and Sol–Energy is calculated as follows based on 0.7% S coal:

Amount of coal fired = 1,000 TPY SO2 x (1 MMBtu / 1.2 lb SO2) x (1,000,000 lb of Coal / 12,000 MMBtu) = 69,444 TPY of coal

Therefore, the SO2 emissions for each fuel and control technology option will be calculated from the annual heat input rate of 69,444 TPY of coal.

For example, the SO2 emissions from firing 3.5% S coal and with wet limstone scrubber are:

SO2 Emissions = 69,444 TPY of coal x (2000 lb / ton) x (12,000 Btu / lb of coal) + 1,000,000 x (0.275 lb SO2 / MMBtu) + (2000 lb / ton) = 229.2 TPY of SO2.

Table C. Summary of Top-Down BACT Impact Analysis Results for SO2 and Acid Gases, Flo-Energy Cogeneration Facility (Revised 02/05/93).

							Economic Impacts			
Control Alternative	Total SO2 Emissions (TPY)	Total Acid Gas Emissions (TPY)	(a) SO2 Emissions Reduction Over Baseline (TPY)	(b) Acid Gas Emissions Reduction Over Baseline (TPY)	(a)+(b) Combined Emissions Reduction Over Baseline (TPY)	Overall Incremental Emission Reduction (TPY)	Total Annualized Cost ($/yr)	Incremental Annualized Cost ($/yr)	Combined Cost Effectiveness Over Baseline ($/ton)	Incremental Cost Effectiveness ($/ton)
Flo-Energy Facility Only- Worst Case Year										
Wet Limestone Scrubber (0.7% S Coal)	81.2	36.2	1,542.4	151.9	1,694.3	43.8	$9,191,457	$3,997,473	$5,425	$91,187
Spray Dryer (0.7% S Coal)	129.9	31.3	1,493.7	156.8	1,650.5	681.9	$5,193,984	$1,410,197	$3,147	$2.068
Wet Limestone Scrubber (1.5% S Coal)	155.6	69.6	1,468.0	118.5	1,586.5	--	$9,494,589	--	$5,985	--
Spray Dryer (1.5% S Coal)	249.0	59.1	1,374.6	129.0	1,503.6	--	$6,286,174	--	$4,181	--
Wet Limestone Scrubber (2.5% S Coal)	257.1	111.3	1,366.5	76.8	1,443.3	--	$9,899,057	--	$6,859	--
Wet Limestone Scrubber (3.5% S Coal)	372.1	153.1	1,251.5	35.0	1,286.6	--	$10,525,717	--	$8,181	--
Spray Dryer (2.5% S Coal)	411.3	93.9	1,212.3	94.2	1,306.5	--	$7,912,756	--	$6,057	--
Spray Dryer (3.5% S Coal)	595.3	128.7	1,028.3	59.4	1,087.7	--	$9,371,187	--	$8,616	--
Dry-Sodium Duct Injection (0.7% S Coal)	811.8	31.3	811.8	156.8	968.6	968.6	$3,783,787	$3.783.787	$3,907	$3,907
Dry-Sodium Duct Injection (1.5% S Coal)	1,555.9	59.1	67.7	129.0	196.6	--	$4,972,120	--	$25,290	--
Baseline (0.7 Wt% S Coal Uncontrolled)	1,623.6	188.1	--	--	--	--	--	--	--	--
Flo-Energy and Sol-Energy Facilities Combined- Average SO2 Emissions										
Wet Limestone Scrubber (0.7% S Coal)	41.7	22.3	958.3	93.5	1,051.8	22.0	$15,243,000	$6,687,000	$14,492	$303.957
Spray Dryer (0.7% S Coal)	66.7	19.3	933.3	96.5	1,029.8	433.3	$8,556,000	$2,401,000	$8,308	$5.541
Wet Limestone Scrubber (1.5% S Coal)	95.8	42.9	904.2	72.9	977.1	--	$15,973,000	--	$16,347	--
Spray Dryer (1.5% S Coal)	153.3	36.4	846.7	79.4	926.0	--	$10,354,000	--	$11,181	--
Wet Limestone Scrubber (2.5% S Coal)	158.3	68.6	841.7	47.2	888.9	--	$19,227,000	--	$21,630	--
Wet Limestone Scrubber (3.5% S Coal)	229.2	94.3	770.8	21.5	792.3	--	$17,406,000	--	$21,968	--
Spray Dryer (2.5% S Coal)	253.3	57.9	746.7	57.9	804.6	--	$14,379,000	--	$17,871	--
Spray Dryer (3.5% S Coal)	366.7	79.3	633.3	36.5	669.8	--	$15,330,000	--	$22,886	--
Dry-Sodium Duct Injection (0.7% S Coal)	500.0	19.3	500.0	96.5	596.5	596.5	$6,155,000	$6.155.000	$10,318	$10.318
Dry-Sodium Duct Injection (1.5% S Coal)	958.3	36.4	41.7	79.4	121.0	--	$8,073,000	--	$66,701	--
Baseline (0.7 Wt% S Coal Uncontrolled) [1]	1,000.0	115.8	--	--	--	--	--	--	--	--

[1] Equivalent to firing 69,444 TPY of coal for both Flo-Energy and Sol-Energy facilities.

List acid gases

Table D. Summary of Top-Down BACT Impact Analysis Results for SO2, Sol-Energy Cogeneration Facility (Revised 02/05/93).

				Environmental Impacts		Energy Impacts	Economic Impacts			
Control Alternative	Total SO2 Emissions (TPY)	Emissions Reduction Over Baseline (TPY)	Incremental Emission Reduction (TPY)	Air toxics impact? (Yes/No)	Adverse enviromental impacts? (Yes/No)	Additional Energy Requirements Electricity (MW-hr/yr)	Total Annualized Cost ($/yr)	Incremental Annualized Cost ($/yr)	Cost Effectiveness Over Baseline ($/ton)	Incremental Cost Effectiveness ($/ton)
Sol-Energy Facility Only - Worst Case Year										
Wet Limestone Scrubber (0.7% S Coal)	49.4	938.8	29.6	Yes	Yes	720	$6,051,191	$2,688,911	$6,446	$90.701
Spray Dryer (0.7% S Coal)	79.1	909.1	415.0	No	Yes	480	$3,362,280	$990,769	$3,698	$2,387
Wet Limestone Scrubber (1.5% S Coal)	94.7	893.5	--	Yes	Yes	810	$6,241,652	--	$6,986	--
Spray Dryer (1.5% S Coal)	151.5	836.7	--	No	Yes	540	$4,047,867	--	$4,838	--
Wet Limestone Scrubber (2.5% S Coal)	156.5	831.7	--	Yes	Yes	900	$6,489,555	--	$7,802	--
Wet Limestone Scrubber (3.5% S Coal)	226.5	761.7	--	Yes	Yes	1,000	$6,880,054	--	$9,032	--
Spray Dryer (2.5% S Coal)	250.3	737.9	--	No	Yes	600	$5,197,058	--	$7,043	--
Spray Dryer (3.5% S Coal)	362.3	625.9	--	No	Yes	660	$5,959,033	--	$9,521	--
Dry-Sodium Duct Injection (0.7% S Coal)	494.1	494.1	494.1	No	Yes	220	$2,371,512	$2,371,512	$4,800	$4,800
Dry-Sodium Duct Injection (1.5% S Coal)	947.0	41.2	--	No	Yes	310	$3,100,669	--	$75,305	--
Baseline (0.7 Wt% S Coal Uncontrolled)	988.2	--	--	--	--	--	--	--	--	--
Sol-Energy and Flo-Energy Facilities Combined - Average SO2 Emissions										
Wet Limestone Scrubber (0.7% S Coal)	41.7	958.3	25.0	Yes	Yes	2,420	$15,243,000	$6,687,000	$15,906	$267,482
Spray Dryer (0.7% S Coal)	66.7	933.3	433.3	No	Yes	1,630	$8,556,000	$2,401,000	$9,167	$5,541
Wet Limestone Scrubber (1.5% S Coal)	95.8	904.2	--	Yes	Yes	2,760	$15,736,000	--	$17,404	--
Spray Dryer (1.5% S Coal)	153.3	846.7	--	No	Yes	1,840	$10,334,000	--	$12,206	--
Wet Limestone Scrubber (2.5% S Coal)	158.3	841.7	--	Yes	Yes	3,150	$16,389,000	--	$19,472	--
Wet Limestone Scrubber (3.5% S Coal)	229.2	770.8	--	Yes	Yes	3,500	$17,406,000	--	$22,581	--
Spray Dryer (2.5% S Coal)	253.3	746.7	--	No	Yes	2,100	$13,110,000	--	$17,558	--
Spray Dryer (3.5% S Coal)	366.7	633.3	--	No	Yes	2,360	$15,330,000	--	$24,205	--
Dry-Sodium Duct Injection (0.7% S Coal)	500.0	500.0	500.0	No	Yes	580	$6,155,000	$6,155,000	$12,310	$12,310
Dry-Sodium Duct Injection (1.5% S Coal)	958.3	41.7	--	No	Yes	810	$8,073,000	--	$193,753	--
Baseline (0.7 Wt% S Coal Uncontrolled)[1]	1,000.0	--	--	--	--	--	--	--	--	--

[1] The amount of coal associated with the proposed SO2 emission limit of 1,000 TPY for both Flo-Energy and Sol-Energy is calculated as follows based on 0.7% S coal:

Amount of coal fired = 1,000 TPY SO2 x (1 MMBtu / 1.2 lb SO2) x (1,000,000 lb of Coal / 12,000 MMBtu) = 69,444 TPY of coal

Therefore, the SO2 emissions for each fuel and control technology option will be calculated from the annual heat input rate of 69,444 TPY of coal.

For example, the SO2 emissions from firing 3.5% S coal and with wet limstone scrubber are:

SO2 Emissions = 69,444 TPY of coal x (2000 lb / ton) x (12,000 Btu / lb of coal) + 1,000,000 x (0.275 lb SO2 / MMBtu) + (2000 lb / ton) = 229.2 TPY of SO2.

Table E. Summary of Top-Down BACT Impact Analysis Results for SO2 and Acid Gases, Sol-Energy Cogeneration Facility (Revised 02/05/93).

							Economic Impacts			
Control Alternative	Total SO2 Emissions (TPY)	Total Acid Gas Emissions (TPY)	(a) SO2 Emissions Reduction Over Baseline (TPY)	(b) Acid Gas Emissions Reduction Over Baseline (TPY)	(a)+(b) Combined Emissions Reduction Over Baseline (TPY)	Overall Incremental Emission Reduction (TPY)	Total Annualized Cost ($/yr)	Incremental Annualized Cost ($/yr)	Combined Cost Effectiveness Over Baseline ($/ton)	Incremental Cost Effectiveness ($/ton)
<u>Sol-Energy Facility Only- Worst Case Year</u>										
Wet Limestone Scrubber (0.7% S Coal)	49.4	22.3	938.8	92.8	1,031.6	26.6	$6,051,191	$2,688,911	$5,866	$100,912
Spray Dryer (0.7% S Coal)	79.1	19.3	909.1	95.8	1,005.0	415.0	$3,362,280	$990,769	$3,346	$2,387
Wet Limestone Scrubber (1.5% S Coal)	94.7	42.8	893.5	72.3	965.8	--	$6,241,652	--	$6,463	--
Spray Dryer (1.5% S Coal)	151.5	36.4	836.7	78.7	915.4	--	$4,047,867	--	$4,422	--
Wet Limestone Scrubber (2.5% S Coal)	156.5	68.5	831.7	46.6	878.3	--	$6,489,555	--	$7,389	--
Wet Limestone Scrubber (3.5% S Coal)	226.5	94.3	761.7	20.8	782.6	--	$6,880,054	--	$8,791	--
Spray Dryer (2.5% S Coal)	250.3	57.8	737.9	57.3	795.1	--	$5,197,058	--	$6,536	--
Spray Dryer (3.5% S Coal)	362.3	79.3	625.9	35.8	661.7	--	$5,959,033	--	$9,006	--
Dry-Sodium Duct Injection (0.7% S Coal)	494.1	19.3	494.1	95.8	589.9	589.9	$2,371,512	$2,371,512	$4,020	$4,020
Dry-Sodium Duct Injection (1.5% S Coal)	947.0	36.4	41.2	78.7	119.9	--	$3,100,669	--	$25,865	--
Baseline (0.7 Wt% S Coal Uncontrolled)	988.2	115.1	--	--	--	--	--	--	--	--
<u>Sol-Energy and Flo-Energy Facilities Combined- Average SO2 Emissions</u>										
Wet Limestone Scrubber (0.7% S Coal)	41.7	22.3	958.3	93.5	1,051.8	22.0	$15,243,000	$6,687,000	$14,492	$303,957
Spray Dryer (0.7% S Coal)	66.7	19.3	933.3	96.5	1,029.8	433.3	$8,556,000	$2,401,000	$8,308	$5,541
Wet Limestone Scrubber (1.5% S Coal)	95.8	42.9	904.2	72.9	977.1	--	$15,973,000	--	$16,347	--
Spray Dryer (1.5% S Coal)	153.3	36.4	846.7	79.4	926.0	--	$10,354,000	--	$11,181	--
Wet Limestone Scrubber (2.5% S Coal)	158.3	68.6	841.7	47.2	888.9	--	$19,227,000	--	$21,630	--
Wet Limestone Scrubber (3.5% S Coal)	229.2	94.3	770.8	21.5	792.3	--	$17,406,000	--	$21,968	--
Spray Dryer (2.5% S Coal)	253.3	57.9	746.7	57.9	804.6	--	$14,379,000	--	$17,871	--
Spray Dryer (3.5% S Coal)	366.7	79.3	633.3	36.5	669.8	--	$15,330,000	--	$22,886	--
Dry-Sodium Duct Injection (0.7% S Coal)	500.0	19.3	500.0	96.5	596.5	596.5	$6,155,000	$6,155,000	$10,318	$10,318
Dry-Sodium Duct Injection (1.5% S Coal)	958.3	36.4	41.7	79.4	121.0	--	$8,073,000	--	$66,701	--
Baseline (0.7 Wt% S Coal Uncontrolled)[1]	1,000.0	115.8	--	--	--	--	--	--	--	--

[1] Equivalent to firing 69,444 TPY of coal for both Flo-Energy and Sol-Energy facilities.

Acid gases - list emiss

Table 1. Capital Cost Estimates for Alternative SO2 Control Systems for Flo–Energy Using 0.7% S Coal (Revised 02/05/93).

Cost Items	Cost Factors	Spray Dryer	Wet Scrubber
DIRECT CAPITAL COSTS (DCC):			
(1) Purchased Equipment			
(a) Basic Equipment*	Vendor Quote	$3.990,000	$11,400,000
(b) Auxiliary Equipment	included	included	included
(c) Structure Support	0.10 x (1a)	$399,000	$1.140,000
(d) Instrumentation & Controls	0.12 x (1a)	$478,800	$1,368,000
(e) Freight[1]	0.05 x (1a .. 1d)	$243,390	$695,400
(f) Sales Tax (Florida)	0.06 x (1a .. 1d)	$292,068	$834,480
(g) Subtotal	(1a .. 1f)	$5,403,258	$15,437,880
(2) Direct Installation[1]	0.30 x (1a .. 1f)	$1,620,977	$4,631,364
(3) Upsizing ESP for Particulate Control			
(a) ESP and Auxiliary Equip. Upsizing	See Note 1	$210,000	--
(b) Instrumentation & Controls	0.1 x (3a)	$21,000	--
(c) Freight[1]	0.05 x (3a .. 3b)	$11,550	--
(d) Sales Tax (Florida)	0.06 x (3a .. 3b)	$13,860	--
(e) Direct Installation Costs[2]	0.67 x (3a .. 3d)	$171,795	--
(4) Upsizing Ash Handling System			
(a) Cost of Upsizing for 3 Boilers	See Note 2	$40,000	--
Total DCC:	(1) + (2) + (3) + (4)	$7,492,440	$20,069,244
INDIRECT CAPITAL COSTS (ICC):			
(5) Indirect Installation			
(a) Engineering & Supervision[1]	0.10 x (DCC)	$749,244	$2,006,924
(b) Construction & Field Expenses[1]	0.10 x (DCC)	$749,244	$2,006,924
(c) Contruction Contractor Fee[1]	0.05 x (DCC)	$374,622	$1,003,462
(d) Contigencies[1]	0.20 x (DCC)	$1,498,488	$4,013,849
(6) Other Indirect Costs			
(a) Startup & Testing[1]	0.03 x (DCC)	$224,773	$602,077
(b) Working Capital	30–day DOC**	$146,254	$194,016
Total ICC:	(5) + (6)	$3,742,625	$9,827,254
TOTAL CAPITAL INVESTMENTS (TCI):	DCC + ICC	$11,235,065	$29,896,498

* For the spray dryers, the basic equipment cost for three units are the average of budgetary quotations from ABB–Flakt, Joy Environmental Systems, and United McGill. The basic equipment cost for the limestone wet scrubbers was based on budgetary pricing from ABB Environmental Systems.

** 30 days of direct operating costs, calculated from the annualized cost Table 2 (i.e., total DOC/12 months).

[1] Based on catalytic incinerators, from OAQPS Control Cost Manual, Fourth Edition (1990).

[2] Based on Capital Cost Factors for ESP, from OAQPS Control Cost Manual, Fourth Edition (1990).

Note 1: 10 percent increase in size for the ESP at $7,000 for 1 percent increase, from information supplied by Research–Cottrell.

Note 2: Capital cost estimates of the ash handling system prior to the upsizing in order to handle the additional ash generated from the spray dryer process are approximately $400,000. The upsizing of the ash handling system will add an additional 10% cost or approximately $40,000 to the spray dryer option. The wet FDG option does not involve any upsizing of the ESP ash handling system.

Table 2. Annualized Cost Estimates for Alternative SO2 Control Systems for Flo–Energy Using 0.7% S Coal (Revised 02/05/93).

Cost Items	Basis	Spray Dryer	Wet Scrubber
DIRECT OPERATING COSTS (DOC):			
(1) Labor			
Operator[2]	$22/hr; approx. 4,706 hr/yr for 3 boilers	$103,532	--
	$22/hr; approx. 7,090 hr/yr for 3 boilers	--	$155,980
Supervisor[1]	15% of operator cost	$15,531	$23,386
(2) Maintenance[2]	5% of direct capital cost	$374,622	$1,003,462
(3) Replacement Parts	3% of direct capital cost	$224,773	$602,077
(4) Utilities			
(a) Electricity	$85 per MW–hr; 1,150 and 1,700 MW–hr	$97,750	$144,500
(b) Water	$0.27 / 1,000 gal; 42.0 and 49.6 mil. gal.	$11,340	$13,392
(5) Raw Chemicals			
(a) Limestone (97% purity)	$32 / ton delivered for 3,100 TPY	--	$99,200
(b) Hydrated Lime (74% purity)	$140 / ton delivered for 4,600 TPY	$644,000	--
(6) Differential Fuel Cost or Credit			
(a) 0.7% S Coal	No Cost Adjustment	$0	$0
(7) Solid Disposal	$27 / ton for approx. 10,500 TPY	$283,500	--
(8) Sludge Disposal	$27 / ton for approx. 10,600 TPY	--	$286,200
Total DOC		$1,755,048	$2,328,197
INDIRECT OPERATING COSTS (IOC):			
(7) Overhead[1]	60% of operating labor & maintenance	$1,161,588	$803,240
(8) Property Taxes[1]	1% of total capital investment	$112,351	$298,965
(9) Insurance[1]	1% of total capital investment	$112,351	$298,965
(10) Administration[1]	2% of total capital investment	$224,701	$597,930
Total IOC		$1,610,990	$1,999,100
CAPITAL RECOVERY COSTS (CRC)	CRF of 0.1627 times TCI	$1,827,945	$4,864,160
ANNUALIZED COSTS (AC):	DOC + IOC + CRC	$5,193,984	$9,191,457

[1] Based on catalytic incinerators, from OAQPS Control Cost Manual, Fourth Edition.

[2] Based on maximum of 25% coal firing for the Flo–Energy cogeneration facility.

Table 3. Capital Cost Estimates for Alternative SO2 Control Systems for Flo-Energy Using 1.5% S Coal (Revised 02/05/93).

Cost Items	Cost Factors	Spray Dryer	Wet Scrubber
DIRECT CAPITAL COSTS (DCC):			
(1) Purchased Equipment			
(a) Basic Equipment*	Vendor Quote	$3,990,000	$11,580,000
(b) Auxiliary Equipment	included	included	included
(c) Upsizing Auxiliary Equipment	0.05 x (1a)	$199,500	included
(d) Structure Support	0.10 x (1a)	$399,000	$1,158,000
(e) Instrumentation & Controls	0.12 x (1a)	$478,800	$1,389,600
(f) Freight[1]	0.05 x (1a .. 1d)	$253,365	$706,380
(g) Sales Tax (Florida)	0.06 x (1a .. 1d)	$304,038	$847,656
(h) Subtotal	(1a .. 1g)	$5,624,703	$15,681,636
(2) Direct Installation[1]	0.30 x (1a .. 1g)	$1,687,411	$4,704,491
(3) Upsizing ESP for Particulate Control			
(a) ESP and Auxiliary Equip. Upsizing	See Note 1	$315,000	--
(b) Instrumentation & Controls	0.10 x (3a)	$31,500	--
(c) Freight[2]	0.05 x (3a .. 3b)	$17,325	--
(d) Sales Tax (Florida)	0.06 x (3a .. 3b)	$20,790	--
(e) Direct Installation Costs[2]	0.67 x (3a .. 3d)	$257,692	--
(4) Upsizing Ash Handling Equipment			
(a) Cost of Upzising for 3 Boilers	See Note 2	$40,000	--
Total DCC:	(1) + (2) + (3) + (4)	$7,994,421	$20,386,127
INDIRECT CAPITAL COSTS (ICC):			
(5) Indirect Installation			
(a) Engineering & Supervision[1]	0.10 x (DCC)	$799,442	$2,038,613
(b) Construction & Field Expenses[1]	0.10 x (DCC)	$799,442	$2,038,613
(c) Contruction Contractor Fee[1]	0.05 x (DCC)	$399,721	$1,019,306
(d) Contigencies[1]	0.20 x (DCC)	$1,598,884	$4,077,225
(6) Other Indirect Costs			
(a) Startup & Testing[1]	0.03 x (DCC)	$239,833	$611,584
(b) Working Capital	30-day DOC**	$206,014	$202,051
Total ICC:	(5) + (6)	$4,043,336	$9,987,392
TOTAL CAPITAL INVESTMENTS (TCI):	DCC + ICC	$12,037,757	$30,373,519

* For the spray dryers, the basic equipment cost for three units are the average of budgetary quotations from ABB-Flakt, Joy Environmental Systems, and United McGill. The basic equipment cost for the limestone wet scrubbers was based on budgetary pricing from ABB Environmental Systems.

** 30 days of direct operating costs, calculated from the annualized cost Table 4 (i.e., total DOC/12 months).

[1] Based on catalytic incinerators, from OAQPS Control Cost Manual, Fourth Edition (1990).

[2] Based on Capital Cost Factors for ESP, from OAQPS Control Cost Manual, Fourth Edition (1990).

Note 1: 15 percent increase in size for the ESP at $7,000 for 1 percent increase, from information supplied by Research-Cottrell.

Note 2: Capital cost estimates of the ash handling system prior to the upsizing in order to handle the additional ash generated from the spray dryer process are approximately $400,000. The upsizing of the ash handling system will add an additional 10% cost or approximately $40,000 to the spray dryer option. The wet scrubber option does not involve any upsizing of the ESP ash handling system.

Table 4. Annualized Cost Estimates for Alternative SO2 Control Systems for Flo–Energy Using 1.5% S Coal (Revised 02/05/93).

Cost Items	Basis	Spray Dryer	Wet Scrubber
DIRECT OPERATING COSTS (DOC):			
(1) Labor			
Operator[2]	$22/hr, approx. 7,800 hr/yr for 3 boilers	$171,600	--
	$22/hr, approx. 10,630 hr/yr for 3 boilers	--	$233,860
Supervisor[1]	15% of operator cost	$25,626	$35,079
(2) Maintenance[2]	5% of direct capital cost	$399,721	$1,019,306
(3) Replacement Parts	3% of direct capital cost	$239,833	$611,584
(4) Utilities			
(a) Electricity	$85 per MW–hr; 1,300 and 1,950 MW–hr	$110,500	$165,750
(b) Water	$0.27 / 1,000 gal; 49.050 and 60.322 mil. gal.	$13,244	$16,287
(5) Raw Chemicals			
(a) Limestone (97% purity)	$32 / ton delivered for 6,600 TPY	--	$211,200
(b) Hydrated Lime (74% purity)	$140 / ton delivered for 9,800 TPY	$1,372,000	--
(6) Differential Fuel Cost or Credit			
(a) 1.5% S Coal	$3 credit per ton for 112,750 TPY	–$338,250	–$338,250
(7) Solid Disposal	$27 / ton for approx. 17,700 TPY	$477,900	--
(8) Sludge Disposal	$27 / ton for approx. 17,400 TPY	--	$469,800
Total DOC		$2,472,173	$2,424,616
INDIRECT OPERATING COSTS (IOC):			
(7) Overhead[1]	60% of operating labor & maintenance	$1,373,947	$913,261
(8) Property Taxes[1]	1% of total capital investment	$120,378	$303,735
(9) Insurance[1]	1% of total capital investment	$120,378	$303,735
(10) Administration[1]	2% of total capital investment	$240,755	$607,470
Total IOC		$1,855,457	$2,128,202
CAPITAL RECOVERY COSTS (CRC)	CRF of 0.1627 times TCI	$1,958,543	$4,941,772
ANNUALIZED COSTS (AC):	DOC + IOC + CRC	$6,286,174	$9,494,589

[1] Based on catalytic incinerators, from OAQPS Control Cost Manual, Fourth Edition.

[2] Based on maximum of 25% coal firing for the Flo–Energy cogeneration facility.

Table 5. Capital Cost Estimates for Alternative SO2 Control Systems for Flo-Energy Using 2.5% S Coal (Revised 02/05/93).

Cost Items	Cost Factors	Spray Dryer	Wet Scrubber
DIRECT CAPITAL COSTS (DCC):			
(1) Purchased Equipment			
(a) Basic Equipment*	Vendor Quote	$3,990,000	$11,760,000
(b) Auxiliary Equipment	included	included	included
(c) Upsizing Auxiliary Equipment	0.125 x (1a)	$498,750	included
(d) Structure Support	0.10 x (1a)	$399,000	$1,176,000
(e) Instrumentation & Controls	0.12 x (1a)	$478,800	$1,411,200
(f) Freight[1]	0.05 x (1a .. 1d)	$268,328	$717,360
(g) Sales Tax (Florida)	0.06 x (1a .. 1d)	$321,993	$860,832
(h) Subtotal	(1a .. 1g)	$5,956,871	$15,925,392
(2) Direct Installation[1]	0.30 x (1a .. 1g)	$1,787,061	$4,777,618
(3) Upsizing ESP for Particulate Control			
(a) ESP and Auxiliary Equip. Upsizing	See Note 1	$525,000	--
(b) Instrumentation & Controls	0.10 x (3a)	$52,500	--
(c) Freight[2]	0.05 x (3a .. 3b)	$28,875	--
(d) Sales Tax (Florida)	0.06 x (3a .. 3b)	$34,650	--
(e) Direct Installation Costs[2]	0.67 x (3a .. 3d)	$429,487	--
(4) Upsizing Ash Handling Equipment			
(a) Cost of Upzising for 3 Boilers	See Note 2	$50,000	
Total DCC:	(1) + (2) + (3) + (4)	$8,864,443	$20,703,010
INDIRECT CAPITAL COSTS (ICC):			
(5) Indirect Installation			
(a) Engineering & Supervision[1]	0.10 x (DCC)	$886,444	$2,070,301
(b) Construction & Field Expenses[1]	0.10 x (DCC)	$886,444	$2,070,301
(c) Contruction Contractor Fee[1]	0.05 x (DCC)	$443,222	$1,035,150
(d) Contigencies[1]	0.20 x (DCC)	$1,772,889	$4,140,602
(6) Other Indirect Costs			
(a) Startup & Testing[1]	0.03 x (DCC)	$265,933	$621,090
(b) Working Capital	30-day DOC**	$306,033	$221,899
Total ICC:	(5) + (6)	$4,560,966	$10,159,344
TOTAL CAPITAL INVESTMENTS (TCI):	DCC + ICC	$13,425,410	$30,862,354

* For the spray dryers, the basic equipment cost for three units are the average of budgetary quotations from ABB-Flakt, Joy Environmental Systems, and United McGill. The basic equipment cost for the limestone wet scrubbers was based on budgetary pricing from ABB Environmental Systems.

** 30 days of direct operating costs, calculated from the annualized cost Table 6 (i.e., total DOC/12 months).

[1] Based on catalytic incinerators, from OAQPS Control Cost Manual, Fourth Edition (1990).

[2] Based on Capital Cost Factors for ESP, from OAQPS Control Cost Manual, Fourth Edition (1990).

Note 1: 15 percent increase in size for the ESP at $7,000 for 1 percent increase, from information supplied by Research-Cottrell.

Note 2: Capital cost estimates of the ash handling system prior to the upsizing in order to handle the additional ash generated from the spray dryer process are approximately $400,000. The upsizing of the ash handling system will add an additional 12.5% cost or approximately $50,000 to the spray dryer option. The wet scrubber option does not involve any upsizing of the ESP ash handling system.

the spray dryer option. The wet scrubber option does not involve any upsizing of the ESP ash handling system.

Table 6. Annualized Cost Estimates for Alternative SO2 Control Systems for Flo–Energy Using 2.5% S Coal (Revised 02/05/93).

Cost Items	Basis	Spray Dryer	Wet Scrubber
DIRECT OPERATING COSTS (DOC):			
(1) Labor			
Operator[2]	$22/hr. approx. 9,400 hr/yr for 3 boilers	$206,800	--
	$22/hr. approx. 12,400 hr/yr for 3 boilers	--	$272,800
Supervisor[1]	15% of operator cost	$31,062	$40,925
(2) Maintenance[2]	5% of direct capital cost	$443,222	$1,035,150
(3) Replacement Parts	3% of direct capital cost	$265,933	$621,090
(4) Utilities			
(a) Electricity	$85 per MW–hr; 1,500 and 2,250 MW–hr	$127,500	$191,250
(b) Water	$0.27 / 1,000 gal; 57.9 and 73.8 mil. gal.	$15,633	$19,926
(5) Raw Chemicals			
(a) Limestone (97% purity)	$32 per ton delivered for 10,900 TPY	--	$348,800
(b) Hydrated Lime (74% purity)	$140 per ton delivered for 16,300 TPY	$2,282,000	--
(6) Differential Fuel Cost or Credit			
(a) 2.5% S Coal	$5 credit per ton for 112,750 TPY	–$563,750	–$563,750
(7) Solid Disposal	$27 per ton for approx. 27,000 TPY	$864,000	--
(8) Sludge Disposal	$27 per ton for approx. 25,800 TPY	--	$696,600
Total DOC		$3,672,400	$2,662,792
INDIRECT OPERATING COSTS (IOC):			
(7) Overhead[1]	60% of operating labor & maintenance	$1,519,025	$980,466
(8) Property Taxes[1]	1% of total capital investment	$134,254	$308,624
(9) Insurance[1]	1% of total capital investment	$134,254	$308,624
(10) Administration[1]	2% of total capital investment	$268,508	$617,247
Total IOC		$2,056,042	$2,214,960
CAPITAL RECOVERY COSTS (CRC)	CRF of 0.1627 times TCI	$2,184,314	$5,021,305
ANNUALIZED COSTS (AC):	DOC + IOC + CRC	$7,912,756	$9,899,057

[1] Based on catalytic incinerators, from OAQPS Control Cost Manual, Fourth Edition.

[2] Based on maximum of 25% coal firing for the Flo–Energy cogeneration facility.

Table 8. Annualized Cost Estimates for Alternative SO2 Control Systems for Flo–Energy Using 3.5% S Coal (Revised 02/05/93).

Cost Items	Basis	Spray Dryer	Wet Scrubber
DIRECT OPERATING COSTS (DOC):			
(1) Labor			
Operator[2]	$22/hr, approx. 13,370 hr/yr for 3 boilers	$294,140	--
	$22/hr, approx. 17,720 hr/yr for 3 boilers	--	$389,840
Supervisor[1]	15% of operator cost	$44,108	$58,464
(2) Maintenance[2]	5% of direct capital cost	$476,446	$1,056,276
(3) Replacement Parts	3% of direct capital cost	$285,868	$633,766
(4) Utilities			
(a) Electricity	$85 per MW–hr; 1,700 and 2,500 MW–hr	$144,500	$212,500
(b) Water	$0.27 / 1,000 gal; 66.75 and 87.24 mil. gal.	$18,023	$23,555
(5) Raw Chemicals			
(a) Limestone (97% purity)	$32 per ton delivered for 15,300 TPY	--	$489,600
(b) Hydrated Lime (74% purity)	$140 per ton delivered for 22,800 TPY	$3,192,000	--
(6) Differential Fuel Cost or Credit			
(a) 3.5% S Coal	$7 credit per ton for 112,750 TPY	–$789,250	–$789,250
(7) Solid Disposal	$27 per ton for approx. 35,800 TPY	$966,600	--
(8) Sludge Disposal	$27 per ton for approx. 34,300 TPY	--	$926,100
Total DOC		$4,632,434	$3,000,851
INDIRECT OPERATING COSTS (IOC):			
(7) Overhead[1]	60% of operating labor & maintenance	$1,801,867	$1,136,605
(8) Property Taxes[1]	1% of total capital investment	$144,888	$315,158
(9) Insurance[1]	1% of total capital investment	$144,888	$315,158
(10) Administration[1]	2% of total capital investment	$289,777	$630,317
Total IOC		$2,381,420	$2,397,239
CAPITAL RECOVERY COSTS (CRC)	CRF of 0.1627 times TCI	$2,357,333	$5,127,627
ANNUALIZED COSTS (AC):	DOC + IOC + CRC	$9,371,187	$10,525,717

[1] Based on catalytic incinerators, from OAQPS Control Cost Manual, Fourth Edition.

[2] Based on maximum of 25% coal firing for the Flo–Energy Cogeneration facility.

Table 7. Capital Cost Estimates for Alternative SO2 Control Systems for Flo-Energy Using 3.5% S Coal (Revised 02/05/93).

Cost Items	Cost Factors	Spray Dryer	Wet Scrubber
DIRECT CAPITAL COSTS (DCC):			
(1) Purchased Equipment			
(a) Basic Equipment*	Vendor Quote	$3,990,000	$12,000,000
(b) Auxiliary Equipment	included	included	included
(c) Upsizing Auxiliary Equipment	0.17 x (1a)	$678,300	included
(d) Structure Support	0.10 x (1a)	$399,000	$1,200,000
(e) Instrumentation & Controls	0.12 x (1a)	$478,800	$1,440,000
(f) Freight[1]	0.05 x (1a .. 1d)	$277,305	$732,000
(g) Sales Tax (Florida)	0.06 x (1a .. 1d)	$332,766	$878,400
(h) Subtotal	(1a .. 1g)	$6,156,171	$16,250,400
(2) Direct Installation[1]	0.30 x (1a .. 1g)	$1,846,851	$4,875,120
(3) Upsizing ESP for Particulate Control			
(a) ESP and Auxiliary Equip. Upsizing	See Note 1	$714,000	--
(b) Instrumentation & Controls	0.10 x (3a)	$71,400	--
(c) Freight[2]	0.05 x (3a .. 3b)	$39,270	--
(d) Sales Tax (Florida)	0.06 x (3a .. 3b)	$47,124	--
(e) Direct Installation Costs[2]	0.67 x (3a .. 3d)	$584,102	--
(4) Upsizing Ash Handling Equipment			
(a) Cost of Upzising for 3 Boilers	See Note 2	$70,000	--
Total DCC:	(1) + (2) + (3) + (4)	$9,528,918	$21,125,520
INDIRECT CAPITAL COSTS (ICC):			
(5) Indirect Installation			
(a) Engineering & Supervision[1]	0.10 x (DCC)	$952,892	$2,112,552
(b) Construction & Field Expenses[1]	0.10 x (DCC)	$952,892	$2,112,552
(c) Contruction Contractor Fee[1]	0.05 x (DCC)	$476,446	$1,056,276
(d) Contigencies[1]	0.20 x (DCC)	$1,905,784	$4,225,104
(6) Other Indirect Costs			
(a) Startup & Testing[1]	0.03 x (DCC)	$285,868	$633,766
(b) Working Capital	30-day DOC**	$386,036	$250,071
Total ICC:	(5) + (6)	$4,959,917	$10,390,320
TOTAL CAPITAL INVESTMENTS (TCI):	DCC + ICC	$14,488,835	$31,515,840

* For the spray dryers, the basic equipment cost for three units are the average of budgetary quotations from ABB-Flakt, Joy Environmental Systems, and United McGill. The basic equipment cost for the limestone wet scrubbers was based on budgetary pricing from ABB Environmental Systems.

** 30 days of direct operating costs, calculated from the annualized cost Table 8 (i.e., total DOC/12 months).

[1] Based on catalytic incinerators, from OAQPS Control Cost Manual, Fourth Edition (1990).

[2] Based on Capital Cost Factors for ESP, from OAQPS Control Cost Manual, Fourth Edition (1990).

Note 1: 15 percent increase in size for the ESP at $7,000 for 1 percent increase, from information supplied by Research-Cottrell.

Note 2: Capital cost estimates of the ash handling system prior to the upsizing in order to handle the additional ash generated from the spray dryer process are approximately $400,000. The upsizing of the ash handling system will add an additional 17.5% cost or approximately $70,000 to the spray dryer option. The wet scrubber option does not involve any upsizing of the ESP ash handling system.

Table 9. Cost Estimates for Sorbent Injection Systems for Flo-Energy Using 0.7% S Coal (Revised 02/05/93).

Cost Items	Basis	Furnace Injection Lime Sorbent	Duct Injection Lime Sorbent	Duct Injection Sodium Sorbent
DIRECT CAPITAL COSTS (DCC):				
(1) Dry Sorbent Injection System	See Note 1	$5,880,000	$4,704,000	$4,410,000
(a) Flue Gas Humidification System		Included	Included	--
(2) Upsizing ESP for Particulate Control				
(a) Percent Increase in Size	Based on ESP's Vendor Estimate	55%	60%	25%
(b) Cost of Upsizing for 3 Boilers	$7,000 per 1% Increase per boiler	$1,155,000	$1,260,000	$525,000
(3) Upsizing Ash Handling System				
(a) Cost of Upsizing for 3 Boilers	10% of current system cost. See Note 2.	$40,000	$40,000	$40,000
Subtotal of DCC	(1) + (2b) + (3a)	$7,075,000	$6,004,000	$4,975,000
INDIRECT CAPITAL COSTS (ICC):				
(4) Indirect Installation				
(a) Engineering & Supervision [1]	0.20 x (DCC)	$1,415,000	$1,200,800	$995,000
(b) Construction & Field Expenses [1]	0.20 x (DCC)	$1,415,000	$1,200,800	$995,000
(c) Contruction Contractor Fee [1]	0.10 x (DCC)	$707,500	$600,400	$497,500
(d) Contingencies	0.25 x (DCC)	$1,768,750	$1,501,000	$1,243,750
(5) Other Indirect Costs				
(a) Start-up, Perf. Test & Model Study [1]	0.04 x (DCC)	$283,000	$240,160	$199,000
(b) Working Capital	30-day DOC	$139,634	$139,634	$139,634
Total ICC:	(4) + (5)	$5,728,884	$4,882,794	$4,069,884
TOTAL CAPITAL INVESTMENTS (TCI):	DCC + ICC	$12,803,884	$10,886,794	$9,044,884
DIRECT OPERATING COSTS (DOC):				
(6) Labor				
Operator [2]	$22/hr; 7,200 hr/yr total for 3 boilers	$158,400	$158,400	$158,400
Supervisor [1]	15% of operator cost	$23,760	$23,760	$23,760
(7) Maintenance [2]	0.05 x (DCC)	$353,750	$300,200	$248,750
(8) Replacement Parts	0.03 x (DCC)	$212,250	$180,120	$149,250
(9) Utilities				
(a) Electricity for ESP	$85 per MW-hr	$22,222	$24,242	$10,101
for Auxiliary Equipment	$85 per MW-hr	$44,070	$48,077	$20,032
(b) Humidification Water	$0.27 / 1,000 gal	$12,150	$12,150	--
(c) Steam Lost or Reheat	$6.19/$10^3$ lb	$53,806	--	--
(10) Raw Chemicals				
(a) Hydrated Lime (74% purity)	$140 / ton delivered for approx. 4,600 TPY	$644,000	$644,000	--
(b) Sodium Bicarbonate (98% purity)	$200 / ton delivered for approx. 4,650 TPY	--	--	$930,000
(11) Differential Fuel Cost or Credit				
(a) 0.7% S Coal	None	$0	$0	$0
(12) Solid Disposal	$27 / ton for Lime; $50 /ton for NaHCO3-type	$151,200	$159,300	$151,550
Total DOC		$1,675,608	$1,550,249	$1,691,843
INDIRECT OPERATING COSTS (IOC):				
(13) Overhead [1]	60% of operating labor & maintenance	$321,546	$289,416	$258,546
(14) Property Taxes [1]	1% of total capital investment	$128,039	$108,868	$90,449
(15) Insurance [1]	1% of total capital investment	$128,039	$108,868	$90,449
(16) Administration [1]	2% of total capital investment	$256,078	$217,736	$180,898
Total IOC		$833,701	$724,888	$620,341
CAPITAL RECOVERY COSTS (CRC)	CRF of 0.1627 times TCI	$2,083,192	$1,771,281	$1,471,603
ANNUALIZED COSTS (AC):	DOC + IOC + CRC	$4,592,501	$4,046,418	$3,783,787

[1] Based on Capital Cost Factors for ESP, from OAQPS Control Cost Manual, Fourth Edition (1990).

[2] Based on maximum of 25% coal firing for the Flo-Energy cogeneration facility.

Note 1: Capital cost factors are $4,000/MMBtu of heat input for Furnace Injection and $3,500/MMBtu of heat input for Duct Injection (both using lime) projecting from Babcock and Wilcox's cost analysis summary for Ohio's Edison Power Plant located at Edgewater, Ohio. The capital cost factor for the sodium sorbent injection process is $3,000/MMBtu of heat input based on estimations from Colorado Springs Utilities Company and Public Service Company of Colorado. Total Capital Investment for the dry sorbent injection process was calculated using these cost factors for a total of three 490-MMBtu boilers for the proposed Flo-Energy facility.

Note 2: Capital cost estimates of the ash handling system prior to the upsizing in order to handle the additional ash generated from the DSI process are approximately $400,000.

Table 10. Cost Estimates for Sorbent Injection Systems for Flo-Energy Using 1.5% S Coal (Revised 02/05/93).

Cost Items	Basis	Furnace Injection Lime Sorbent	Duct Injection Lime Sorbent	Duct Injection Sodium Sorbent
DIRECT CAPITAL COSTS (DCC):				
(1) Dry Sorbent Injection System	See Note 1	\$5,880,000	\$4,704,000	\$4,410,000
(a) Flue Gas Humidification System	Included	Included	Included	--
(b) Upsizing Auxiliary Equipment	0.05 x (1)	\$294,000	\$235,200	\$220,500
(2) Upsizing ESP for Particulate Control				
(a) Percent Increase in Size	Based on ESP's Vendor Estimate	60%	65%	35%
(b) Cost of Upsizing for 3 Boilers	\$7,000 per 1% Increase per boiler	\$1,260,000	\$1,365,000	\$735,000
(3) Upsizing Ash Handling System				
(a) Cost of Upsizing for 3 Boilers	12% of current system cost. See Note 2.	\$48,000	\$48,000	\$48,000
Subtotal of DCC	(1) + (2b) + (3a)	\$7,482,000	\$6,352,200	\$5,413,500
INDIRECT CAPITAL COSTS (ICC):				
(4) Indirect Installation				
(a) Engineering & Supervision [1]	0.20 x (DCC)	\$1,496,400	\$1,270,440	\$1,082,700
(b) Construction & Field Expenses [1]	0.20 x (DCC)	\$1,496,400	\$1,270,440	\$1,082,700
(c) Contruction Contractor Fee [1]	0.10 x (DCC)	\$748,200	\$635,220	\$541,350
(d) Contingencies	0.25 x (DCC)	\$1,870,500	\$1,588,050	\$1,353,375
(5) Other Indirect Costs				
(a) Start-up, Perf. Test & Model Study [1]	0.04 x (DCC)	\$299,280	\$254,088	\$216,540
(b) Working Capital	30-day DOC	\$193,661	\$193,661	\$193,661
Total ICC:	(4) + (5)	\$6,104,441	\$5,211,899	\$4,470,326
TOTAL CAPITAL INVESTMENTS (TCI):	DCC + ICC	\$13,586,441	\$11,564,099	\$9,883,826
DIRECT OPERATING COSTS (DOC):				
(6) Labor				
Operator [2]	\$22/hr; 8,640 hr/yr total for 3 boilers	\$190,080	\$190,080	\$190,080
Supervisor [1]	15% of operator cost	\$28,512	\$28,512	\$28,512
(7) Maintenance [2]	0.05 x (DCC)	\$374,100	\$317,610	\$270,675
(8) Replacement Parts	0.03 x (DCC)	\$224,460	\$190,566	\$162,405
(9) Utilities				
(a) Electricity for ESP	\$85 per MW-hr	\$24,242	\$26,262	\$14,141
for Auxiliary Equipment	\$85 per MW-hr	\$48,077	\$52,083	\$28,045
(b) Humidification Water	\$0.27 / 1,000 gal	\$12,150	\$12,150	--
(c) Steam Lost or Reheat	\$6.19/10^3 lb	\$64,567	--	--
(10) Raw Chemicals				
(a) Hydrated Lime (74% purity)	\$140 / ton delivered for approx. 9,800 TPY	\$1,372,000	\$1,372,000	--
(b) Sodium Bicarbonate (98% purity)	\$200 / ton delivered for approx. 9,970 TPY	--	--	\$1,994,000
(11) Differential Fuel Cost or Credit				
(a) 1.5% S Coal	\$3 savings per ton for 112,750 TPY	-\$338,250	-\$338,250	-\$338,250
(12) Solid Disposal	\$27 / ton for Lime; \$50 /ton for NaHCO3-type	\$324,000	\$340,200	\$325,500
Total DOC		\$2,323,938	\$2,191,213	\$2,675,108
INDIRECT OPERATING COSTS (IOC):				
(13) Overhead [1]	60% of operating labor & maintenance	\$355,615	\$321,721	\$293,560
(14) Property Taxes [1]	1% of total capital investment	\$135,864	\$115,641	\$98,838
(15) Insurance [1]	1% of total capital investment	\$135,864	\$115,641	\$98,838
(16) Administration [1]	2% of total capital investment	\$271,729	\$231,282	\$197,677
Total IOC		\$899,073	\$784,285	\$688,913
CAPITAL RECOVERY COSTS (CRC)	CRF of 0.1627 times TCI	\$2,210,514	\$1,881,479	\$1,608,099
ANNUALIZED COSTS (AC):	DOC + IOC + CRC	\$5,433,525	\$4,856,977	\$4,972,120

[1] Based on Capital Cost Factors for ESP, from OAQPS Control Cost Manual, Fourth Edition (1990).

[2] Based on maximum of 25% coal firing for the Flo-Energy cogeneration facility.

Note 1: Capital cost factors are \$4,000/MMBtu of heat input for Furnace Injection and \$3,500/MMBtu of heat input for Duct Injection (both using lime) projecting from Babcock and Wilcox's cost analysis summary for Ohio's Edison Power Plant located at Edgewater, Ohio. The capital cost factor for the sodium sorbent injection process is \$3,000/MMBtu of heat input based on estimations from Colorado Springs Utilities Company and Public Service Company of Colorado. Total Capital Investment for the dry sorbent injection process was calculated using these cost factors for a total of three 490-MMBtu boilers for the proposed Flo-Energy facility.

Note 2: Capital cost estimates of the ash handling system prior to the upsizing in order to handle the additional ash generated from the DSI process are approximately \$260,000.

Table 11. Capital Cost Estimates for Alternative SO2 Control Systems for Sol–Energy Using 0.7% S Coal (Revised 02/05/93).

Cost Items	Cost Factors	Spray Dryer	Wet Scrubber
DIRECT CAPITAL COSTS (DCC):			
(1) Purchased Equipment			
(a) Basic Equipment*	Vendor Quote	$2,660,000	$7,600,000
(b) Auxiliary Equipment	included	included	included
(c) Structure Support	0.10 x (1a)	$266,000	$760,000
(d) Instrumentation & Controls	0.12 x (1a)	$319,200	$912,000
(e) Freight[1]	0.05 x (1a .. 1d)	$162,260	$463,600
(f) Sales Tax (Florida)	0.06 x (1a .. 1d)	$194,712	$556,320
(g) Subtotal	(1a .. 1f)	$3,602,172	$10,291,920
(2) Direct Installation[1]	0.30 x (1a .. 1f)	$1,080,652	$3,087,576
(3) Upsizing ESP for Particulate Control			
(a) ESP and Auxiliary Equip. Upsizing	See Note 1	$140,000	--
(b) Instrumentation & Controls	0.1 x (3a)	$14,000	--
(c) Freight[1]	0.05 x (3a .. 3b)	$7,700	--
(d) Sales Tax (Florida)	0.06 x (3a .. 3b)	$9,240	--
(e) Direct Installation Costs[2]	0.67 x (3a .. 3d)	$114,530	--
(4) Upsizing Ash Handling System			
(a) Cost of Upsizing for 2 Boilers	See Note 2	$26,000	--
Total DCC	(1) + (2) + (3) + (4)	$4,994,293	$13,379,496
INDIRECT CAPITAL COSTS (ICC):			
(5) Indirect Installation			
(a) Engineering & Supervision[1]	0.10 x (DCC)	$499,429	$1,337,950
(b) Construction & Field Expenses[1]	0.10 x (DCC)	$499,429	$1,337,950
(c) Contruction Contractor Fee[1]	0.05 x (DCC)	$249,715	$668,975
(d) Contigencies[1]	0.20 x (DCC)	$998,859	$2,675,899
(6) Other Indirect Costs			
(a) Startup & Testing[1]	0.03 x (DCC)	$149,829	$401,385
(b) Working Capital	30–day DOC**	$90,394	$123,738
Total ICC	(5) + (6)	$2,487,655	$6,545,896
TOTAL CAPITAL INVESTMENTS (TCI):	DCC + ICC	$7,481,949	$19,925,392

* For the spray dryers, the basic equipment cost for three units are the average of budgetary quotations from ABB–Flakt, Joy Environmental Systems, and United McGill. The basic equipment cost for the limestone wet scrubbers was based on budgetary pricing from ABB Environmental Systems.

** 30 days of direct operating costs, calculated from the annualized cost Table 2 (i.e., total DOC/12 months).

[1] Based on catalytic incinerators, from OAQPS Control Cost Manual, Fourth Edition (1990).

[2] Based on Capital Cost Factors for ESP, from OAQPS Control Cost Manual, Fourth Edition (1990).

Note 1: 10 percent increase in size for the ESP at $7,000 for 1 percent increase, from information supplied by Research–Cottrell.

Note 2: Capital cost estimates of the ash handling system prior to the upsizing in order to handle the additional ash generated from the spray dryer process are approximately $260,000. The upsizing of the ash handling system will add an additional 10% cost or approximately $26,000 to the spray dryer option. The wet FDG option does not involve any upsizing of the ESP ash handling system.

Table 12. Annualized Cost Estimates for Alternative SO2 Control Systems for Sol–Energy Using 0.7% S Coal (Revised 02/05/93).

Cost Items	Basis	Spray Dryer	Wet Scrubber
DIRECT OPERATING COSTS (DOC):			
(1) Labor			
Operator[2]	$22/hr; approx. 2,950 hr/yr for 2 boilers	$64,900	--
	$22/hr; approx. 4,450 hr/yr for 2 boilers	--	$97,900
Supervisor[1]	15% of operator cost	$9,720	$14,636
(2) Maintenance[2]	5% of direct capital cost	$249,715	$668,975
(3) Replacement Parts	3% of direct capital cost	$149,829	$401,385
(4) Utilities	480 720		
(a) Electricity	$85 per MW–hr; ~~800~~ and ~~1,450~~ MW–hr	$40,800	$61,200
(b) Water	$0.27 / 1,000 gal; 18.4 and 21.5 mil. gal.	$4,968	$5,805
(5) Raw Chemicals			
(a) Limestone (97% purity)	$32 / ton delivered for 1,900 TPY	--	$60,800
(b) Hydrated Lime (74% purity)	$140 / ton delivered for 2,800 TPY	$392,000	--
(6) Differential Fuel Cost or Credit			
(a) 0.7% S Coal	No Cost Adjustment	$0	$0
(7) Solid Disposal	$27 / ton for approx. 6,400 TPY	$172,800	--
(8) Sludge Disposal	$27 / ton for approx. 6,450 TPY	--	$174,150
Total DOC		$1,084,732	$1,484,851
INDIRECT OPERATING COSTS (IOC):			
(7) Overhead[1]	60% of operating labor & maintenance	$760,958	$527,464
(8) Property Taxes[1]	1% of total capital investment	$74,819	$199,254
(9) Insurance[1]	1% of total capital investment	$74,819	$199,254
(10) Administration[1]	2% of total capital investment	$149,639	$398,508
Total IOC		$1,060,236	$1,324,479
CAPITAL RECOVERY COSTS (CRC)	CRF of 0.1627 times TCI	$1,217,313	$3,241,861
ANNUALIZED COSTS (AC):	DOC + IOC + CRC	$3,362,280	$6,051,191

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[1] Based on catalytic incinerators, from OAQPS Control Cost Manual, Fourth Edition.

[2] Based on maximum of 25% coal firing for the Sol–Energy cogeneration facility.

Table 13. Capital Cost Estimates for Alternative SO2 Control Systems for Sol–Energy Using 1.5% S Coal (Revised 02/05/93).

Cost Items	Cost Factors	Spray Dryer	Wet Scrubber
DIRECT CAPITAL COSTS (DCC):			
(1) Purchased Equipment			
(a) Basic Equipment*	Vendor Quote	$2,660,000	$7,720,000
(b) Auxiliary Equipment	included	included	included
(c) Upsizing Auxiliary Equipment	0.05 x (1a)	$133,000	included
(d) Structure Support	0.10 x (1a)	$266,000	$772,000
(e) Instrumentation & Controls	0.12 x (1a)	$319,200	$926,400
(f) Freight[1]	0.05 x (1a .. 1d)	$168,910	$470,920
(g) Sales Tax (Florida)	0.06 x (1a .. 1d)	$202,692	$565,104
(h) Subtotal	(1a .. 1g)	$3,749,802	$10,454,424
(2) Direct Installation[1]	0.30 x (1a .. 1g)	$1,124,941	$3,136,327
(3) Upsizing ESP for Particulate Control			
(a) ESP and Auxiliary Equip. Upsizing	See Note 1	$210,000	--
(b) Instrumentation & Controls	0.10 x (3a)	$21,000	--
(c) Freight[1]	0.05 x (3a .. 3b)	$11,550	--
(d) Sales Tax (Florida)	0.06 x (3a .. 3b)	$13,860	--
(e) Direct Installation Costs[2]	0.67 x (3a .. 3d)	$171,795	--
(4) Upsizing Ash Handling Equipment			
(a) Cost of Upzising for 2 Boilers	See Note 2	$26,000	--
Total DCC:	(1) + (2) + (3) + (4)	$5,328,947	$13,590,751
INDIRECT CAPITAL COSTS (ICC):			
(5) Indirect Installation			
(a) Engineering & Supervision[1]	0.10 x (DCC)	$532,895	$1,359,075
(b) Construction & Field Expenses[1]	0.10 x (DCC)	$532,895	$1,359,075
(c) Contruction Contractor Fee[1]	0.05 x (DCC)	$266,447	$679,538
(d) Contigencies[1]	0.20 x (DCC)	$1,065,789	$2,718,150
(6) Other Indirect Costs			
(a) Startup & Testing[1]	0.03 x (DCC)	$159,868	$407,723
(b) Working Capital	30–day DOC**	$127,326	$128,441
Total ICC:	(5) + (6)	$2,685,220	$6,652,002
TOTAL CAPITAL INVESTMENTS (TCI):	DCC + ICC	$8,014,168	$20,242,753

* For the spray dryers, the basic equipment cost for three units are the average of budgetary quotations from ABB–Flakt, Joy Environmental Systems, and United McGill. The basic equipment cost for the limestone wet scrubbers was based on budgetary pricing from ABB Environmental Systems.

** 30 days of direct operating costs, calculated from the annualized cost Table 4 (i.e., total DOC/12 months).

[1] Based on catalytic incinerators, from OAQPS Control Cost Manual, Fourth Edition (1990).

[2] Based on Capital Cost Factors for ESP, from OAQPS Control Cost Manual, Fourth Edition (1990).

Note 1: 15 percent increase in size for the ESP at $7,000 for 1 percent increase, from information supplied by Research–Cottrell.

Note 2: Capital cost estimates of the ash handling system prior to the upsizing in order to handle the additional ash generated from the spray dryer process are approximately $260,000. The upsizing of the ash handling system will add an additional 10% cost or approximately $26,000 to the spray dryer option. The wet scrubber option does not involve any upsizing of the ESP ash handling system.

Table 14. Annualized Cost Estimates for Alternative SO2 Control Systems for Sol−Energy Using 1.5% S Coal (Revised 02/05/93).

Cost Items	Basis	Spray Dryer	Wet Scrubber
DIRECT OPERATING COSTS (DOC):			
(1) Labor			
Operator[2]	$22/hr, approx. 4,900 hr/yr for 2 boilers	$107,800	--
	$22/hr, approx. 6,700 hr/yr for 2 boilers	--	$147,400
Supervisor[1]	15% of operator cost	$16,038	$21,954
(2) Maintenance[2]	5% of direct capital cost	$266,447	$679,538
(3) Replacement Parts	3% of direct capital cost	$159,868	$407,723
(4) Utilities			
(a) Electricity	$85 per MW−hr; 540 and 810 MW−hr	$45,900	$68,850
(b) Water	$0.27 / 1,000 gal; 22.7 and 27.8 mil. gal.	$6,129	$7,506
(5) Raw Chemicals			
(a) Limestone (97% purity)	$32 / ton delivered for 4,000 TPY	--	$128,000
(b) Hydrated Lime (74% purity)	$140 / ton delivered for 6,000 TPY	$840,000	--
(6) Differential Fuel Cost or Credit			
(a) 1.5% S Coal	$3 credit per ton for 68,625 TPY	−$205,875	−$205,875
(7) Solid Disposal	$27 / ton for approx. 10,800 TPY	$291,600	--
(8) Sludge Disposal	$27 / ton for approx. 10,600 TPY	--	$286,200
Total DOC		$1,527,908	$1,541,295
INDIRECT OPERATING COSTS (IOC):			
(7) Overhead[1]	60% of operating labor & maintenance	$895,488	$597,151
(8) Property Taxes[1]	1% of total capital investment	$80,142	$202,428
(9) Insurance[1]	1% of total capital investment	$80,142	$202,428
(10) Administration[1]	2% of total capital investment	$160,283	$404,855
Total IOC		$1,216,054	$1,406,861
CAPITAL RECOVERY COSTS (CRC)	CRF of 0.1627 times TCI	$1,303,905	$3,293,496
ANNUALIZED COSTS (AC):	DOC + IOC + CRC	$4,047,867	$6,241,652

[1] Based on catalytic incinerators, from OAQPS Control Cost Manual, Fourth Edition.

[2] Based on maximum of 25% coal firing for the Sol−Energy cogeneration facility.

Table 15. Capital Cost Estimates for Alternative SO2 Control Systems for Sol–Energy Using 2.5% S Coal (Revised 02/05/93).

Cost Items	Cost Factors	Spray Dryer	Wet Scrubber
DIRECT CAPITAL COSTS (DCC):			
(1) Purchased Equipment			
(a) Basic Equipment*	Vendor Quote	$2,660,000	$7,840,000
(b) Auxiliary Equipment	included	included	included
(c) Upsizing Auxiliary Equipment	0.125 x (1a)	$332,500	included
(d) Structure Support	0.10 x (1a)	$266,000	$784,000
(e) Instrumentation & Controls	0.12 x (1a)	$319,200	$940,800
(f) Freight[1]	0.05 x (1a .. 1d)	$178,885	$478,240
(g) Sales Tax (Florida)	0.06 x (1a .. 1d)	$214,662	$573,888
(h) Subtotal	(1a .. 1g)	$3,971,247	$10,616,928
(2) Direct Installation[1]	0.30 x (1a .. 1g)	$1,191,374	$3,185,078
(3) Upsizing ESP for Particulate Control			
(a) ESP and Auxiliary Equip. Upsizing	See Note 1	$525,000	--
(b) Instrumentation & Controls	0.10 x (3a)	$52,500	--
(c) Freight[1]	0.05 x (3a .. 3b)	$28,875	--
(d) Sales Tax (Florida)	0.06 x (3a .. 3b)	$34,650	--
(e) Direct Installation Costs[2]	0.67 x (3a .. 3d)	$429,487	--
(4) Upsizing Ash Handling Equipment			
(a) Cost of Upzising for 2 Boilers	See Note 2	$32,500	--
Total DCC:	(1) + (2) + (3) + (4)	$6,265,633	$13,802,006
INDIRECT CAPITAL COSTS (ICC):			
(5) Indirect Installation			
(a) Engineering & Supervision[1]	0.10 x (DCC)	$626,563	$1,380,201
(b) Construction & Field Expenses[1]	0.10 x (DCC)	$626,563	$1,380,201
(c) Contruction Contractor Fee[1]	0.05 x (DCC)	$313,282	$690,100
(d) Contigencies[1]	0.20 x (DCC)	$1,253,127	$2,760,401
(6) Other Indirect Costs			
(a) Startup & Testing[1]	0.03 x (DCC)	$187,969	$414,060
(b) Working Capital	30–day DOC**	$190,046	$140,093
Total ICC:	(5) + (6)	$3,197,550	$6,765,056
TOTAL CAPITAL INVESTMENTS (TCI):	DCC + ICC	$9,463,183	$20,567,062

* For the spray dryers, the basic equipment cost for three units are the average of budgetary quotations from ABB–Flakt, Joy Environmental Systems, and United McGill. The basic equipment cost for the limestone wet scrubbers was based on budgetary pricing from ABB Environmental Systems.

** 30 days of direct operating costs, calculated from the annualized cost Table 6 (i.e., total DOC/12 months).

[1] Based on catalytic incinerators, from OAQPS Control Cost Manual, Fourth Edition (1990).

[2] Based on Capital Cost Factors for ESP, from OAQPS Control Cost Manual, Fourth Edition (1990).

Note 1: 15 percent increase in size for the ESP at $7,000 for 1 percent increase, from information supplied by Research–Cottrell.

Note 2: Capital cost estimates of the ash handling system prior to the upsizing in order to handle the additional ash generated from the spray dryer process are approximately $260,000. The upsizing of the ash handling system will add an additional 12.5% cost or approximately $32,500 to the spray dryer option. The wet scrubber option does not involve any upsizing of the ESP ash handling system.

Table 16. Annualized Cost Estimates for Alternative SO2 Control Systems for Sol–Energy Using 2.5% S Coal (Revised 02/05/93).

Cost Items	Basis	Spray Dryer	Wet Scrubber
DIRECT OPERATING COSTS (DOC):			
(1) Labor			
Operator[2]	$22/hr, approx. 5,900 hr/yr for 2 boilers	$129,800	--
	$22/hr, approx. 7,800 hr/yr for 2 boilers	--	$171,600
Supervisor[1]	15% of operator cost	$19,440	$25,613
(2) Maintenance[2]	5% of direct capital cost	$313,282	$690,100
(3) Replacement Parts	3% of direct capital cost	$187,969	$414,060
(4) Utilities			
(a) Electricity	$85 per MW–hr; 600 and 900 MW–hr	$51,000	$76,500
(b) Water	$0.27 / 1,000 gal; 28.1 and 35.8 mil. gal.	$7,587	$9,666
(5) Raw Chemicals			
(a) Limestone (97% purity)	$32 per ton delivered for 6,650 TPY	--	$212,800
(b) Hydrated Lime (74% purity)	$140 per ton delivered for 9,950 TPY	$1,393,000	--
(6) Differential Fuel Cost or Credit			
(a) 2.5% S Coal	$5 credit per ton for 68,625 TPY	–$343,125	–$343,125
(7) Solid Disposal	$27 per ton for approx. 16,300 TPY	$521,600	--
(8) Sludge Disposal	$27 per ton for approx. 15,700 TPY	--	$423,900
Total DOC		$2,280,553	$1,681,114
INDIRECT OPERATING COSTS (IOC):			
(7) Overhead[1]	60% of operating labor & maintenance	$998,318	$639,497
(8) Property Taxes[1]	1% of total capital investment	$94,804	$206,131
(9) Insurance[1]	1% of total capital investment	$94,804	$206,131
(10) Administration[1]	2% of total capital investment	$189,608	$412,261
Total IOC		$1,377,534	$1,464,019
CAPITAL RECOVERY COSTS (CRC)	CRF of 0.1627 times TCI	$1,542,460	$3,353,746
ANNUALIZED COSTS (AC):	DOC + IOC + CRC	$5,200,546	$6,498,880

[1] Based on catalytic incinerators, from OAQPS Control Cost Manual, Fourth Edition.

[2] Based on maximum of 25% coal firing for the Sol–Energy cogeneration facility.

Table 17. Capital Cost Estimates for Alternative SO2 Control Systems for Sol-Energy Using 3.5% S Coal (Revised 02/05/93).

Cost Items	Cost Factors	Spray Dryer	Wet Scrubber
DIRECT CAPITAL COSTS (DCC):			
(1) Purchased Equipment			
(a) Basic Equipment*	Vendor Quote	$2,660,000	$8,000,000
(b) Auxiliary Equipment	included	included	included
(c) Upsizing Auxiliary Equipment	0.17 x (1a)	$452,200	included
(d) Structure Support	0.10 x (1a)	$266,000	$800,000
(e) Instrumentation & Controls	0.12 x (1a)	$319,200	$960,000
(f) Freight[1]	0.05 x (1a .. 1d)	$184,870	$488,000
(g) Sales Tax (Florida)	0.06 x (1a .. 1d)	$221,844	$585,600
(h) Subtotal	(1a .. 1g)	$4,104,114	$10,833,600
(2) Direct Installation[1]	0.30 x (1a .. 1g)	$1,231,234	$3,250,080
(3) Upsizing ESP for Particulate Control			
(a) ESP and Auxiliary Equip. Upsizing	See Note 1	$476,000	--
(b) Instrumentation & Controls	0.10 x (3a)	$47,600	--
(c) Freight[1]	0.05 x (3a .. 3b)	$26,180	--
(d) Sales Tax (Florida)	0.06 x (3a .. 3b)	$31,416	--
(e) Direct Installation Costs[2]	0.67 x (3a .. 3d)	$389,401	--
(4) Upsizing Ash Handling Equipment			
(a) Cost of Upzising for 2 Boilers	See Note 2	$45,500	--
Total DCC:	(1) + (2) + (3) + (4)	$6,351,446	$14,083,680
INDIRECT CAPITAL COSTS (ICC):			
(5) Indirect Installation			
(a) Engineering & Supervision[1]	0.10 x (DCC)	$635,145	$1,408,368
(b) Construction & Field Expenses[1]	0.10 x (DCC)	$635,145	$1,408,368
(c) Contruction Contractor Fee[1]	0.05 x (DCC)	$317,572	$704,184
(d) Contigencies[1]	0.20 x (DCC)	$1,270,289	$2,816,736
(6) Other Indirect Costs			
(a) Startup & Testing[1]	0.03 x (DCC)	$190,543	$422,510
(b) Working Capital	30-day DOC**	$236,658	$157,151
Total ICC:	(5) + (6)	$3,285,351	$6,917,317
TOTAL CAPITAL INVESTMENTS (TCI):	DCC + ICC	$9,636,797	$21,000,997

* For the spray dryers, the basic equipment cost for three units are the average of budgetary quotations from ABB-Flakt, Joy Environmental Systems, and United McGill. The basic equipment cost for the limestone wet scrubbers was based on budgetary pricing from ABB Environmental Systems.

** 30 days of direct operating costs, calculated from the annualized cost Table 8 (i.e., total DOC/12 months).

[1] Based on catalytic incinerators, from OAQPS Control Cost Manual, Fourth Edition (1990).

[2] Based on Capital Cost Factors for ESP, from OAQPS Control Cost Manual, Fourth Edition (1990).

Note 1: 15 percent increase in size for the ESP at $7,000 for 1 percent increase, from information supplied by Research-Cottrell.

Note 2: Capital cost estimates of the ash handling system prior to the upsizing in order to handle the additional ash generated from the spray dryer process are approximately $260,000. The upsizing of the ash handling system will add an additional 17.5% cost or approximately $45,500 to the spray dryer option. The wet scrubber option does not involve any upsizing of the ESP ash handling system.

the spray dryer option. The wet scrubber option does not involve any upsizing of the ESP ash handling system.

Table 18. Annualized Cost Estimates for Alternative SO2 Control Systems for Sol–Energy Using 3.5% S Coal (Revised 02/05/93).

Cost Items	Basis	Spray Dryer	Wet Scrubber
DIRECT OPERATING COSTS (DOC):			
(1) Labor			
Operator[2]	$22/hr, approx. 8,400 hr/yr for 2 boilers	$184,800	--
	$22/hr, approx. 11,100 hr/yr for 2 boilers	--	$244,200
Supervisor[1]	15% of operator cost	$27,605	$36,590
(2) Maintenance[2]	5% of direct capital cost	$317,572	$704,184
(3) Replacement Parts	3% of direct capital cost	$190,543	$422,510
(4) Utilities			
(a) Electricity	$85 per MW–hr; 660 and 1,000 MW–hr	$56,100	$85,000
(b) Water	$0.27 / 1,000 gal; 33.5 and 43.7 mil. gal.	$9,045	$11,799
(5) Raw Chemicals			
(a) Limestone (97% purity)	$32 per ton delivered for 9,300 TPY	--	$297,600
(b) Hydrated Lime (74% purity)	$140 per ton delivered for 13,900 TPY	$1,946,000	--
(6) Differential Fuel Cost or Credit			
(a) 3.5% S Coal	$7 credit per ton for 68,625 TPY	–$480,375	–$480,375
(7) Solid Disposal	$27 per ton for approx. 21,800 TPY	$588,600	--
(8) Sludge Disposal	$27 per ton for approx. 20,900 TPY	--	$564,300
Total DOC		$2,839,891	$1,885,808
INDIRECT OPERATING COSTS (IOC):			
(7) Overhead[1]	60% of operating labor & maintenance	$1,165,763	$737,344
(8) Property Taxes[1]	1% of total capital investment	$96,368	$210,010
(9) Insurance[1]	1% of total capital investment	$96,368	$210,010
(10) Administration[1]	2% of total capital investment	$192,736	$420,020
Total IOC		$1,551,235	$1,577,384
CAPITAL RECOVERY COSTS (CRC)	CRF of 0.1627 times TCI	1,567,907	3,416,862
ANNUALIZED COSTS (AC):	DOC + IOC + CRC	$5,959,033	$6,880,054

[1] Based on catalytic incinerators, from OAQPS Control Cost Manual, Fourth Edition.

[2] Based on maximum of 25% coal firing for the Sol–Energy Cogeneration facility.

Table 19. Cost Estimates for Sorbent Injection Systems for Sol–Energy Using 0.7% S Coal (Revised 02/05/93).

Cost Items	Basis	Furnace Injection Lime Sorbent	Duct Injection Lime Sorbent	Duct Injection Sodium Sorbent
DIRECT CAPITAL COSTS (DCC):				
(1) Dry Sorbent Injection System	See Note 1	$3,680,000	$2,944,000	$2,760,000
(a) Flue Gas Humidification System		Included	Included	--
(2) Upsizing ESP for Particulate Control				
(a) Percent Increase in Size	Based on ESP's Vendor Estimate	55%	60%	25%
(b) Cost of Upsizing for 2 Boilers	$7,000 per 1% Increase per boiler	$770,000	$840,000	$350,000
(3) Upsizing Ash Handling System				
(a) Cost of Upsizing for 2 Boilers	10% of current system cost. See Note 2.	$26,000	$26,000	$26,000
Subtotal of DCC	(1) + (2b) + (3a)	$4,476,000	$3,810,000	$3,136,000
INDIRECT CAPITAL COSTS (ICC):				
(4) Indirect Installation				
(a) Engineering & Supervision [1]	0.20 x (DCC)	$895,200	$762,000	$627,200
(b) Construction & Field Expenses [1]	0.20 x (DCC)	$895,200	$762,000	$627,200
(c) Contruction Contractor Fee [1]	0.10 x (DCC)	$447,600	$381,000	$313,600
(d) Contingencies	0.25 x (DCC)	$1,119,000	$952,500	$784,000
(5) Other Indirect Costs				
(a) Start–up, Perf. Test & Model Study [1]	0.04 x (DCC)	$179,040	$152,400	$125,440
(b) Working Capital	30–day DOC	$86,802	$86,802	$86,802
Total ICC:	(4) + (5)	$3,622,842	$3,096,702	$2,564,242
TOTAL CAPITAL INVESTMENTS (TCI):	DCC + ICC	$8,098,842	$6,906,702	$5,700,242
DIRECT OPERATING COSTS (DOC):				
(6) Labor				
Operator [2]	$22/hr; 4,800 hr/yr total for 2 boilers	$105,600	$105,600	$105,600
Supervisor [1]	15% of operator cost	$15,840	$15,840	$15,840
(7) Maintenance [2]	0.05 x (DCC)	$223,800	$190,500	$156,800
(8) Replacement Parts	0.03 x (DCC)	$134,280	$114,300	$94,080
(9) Utilities				
(a) Electricity for ESP	$85 per MW–hr	$13,589	$14,825	$6,177
for Auxiliary Equipment	$85 per MW–hr	$26,909	$29,356	$12,232
(b) Humidification Water	$0.27 / 1,000 gal	$5,063	$5,063	--
(c) Steam Lost or Reheat	$6.19/$10^3$ lb	$32,749	--	--
(10) Raw Chemicals				
(a) Hydrated Lime (74% purity)	$140 / ton delivered for approx. 2,800 TPY	$392,000	$392,000	--
(b) Sodium Bicarbonate (98% purity)	$200 / ton delivered for approx. 2,830 TPY	--	--	$566,000
(11) Differential Fuel Cost or Credit				
(a) 0.7% S Coal	None	$0	$0	$0
(12) Solid Disposal	$27 / ton for Lime; $50 /ton for NaHCO3–type	$91,800	$97,200	$92,400
Total DOC		$1,041,630	$964,683	$1,049,129
INDIRECT OPERATING COSTS (IOC):				
(13) Overhead [1]	60% of operating labor & maintenance	$207,144	$187,164	$166,944
(14) Property Taxes [1]	1% of total capital investment	$80,988	$69,067	$57,002
(15) Insurance [1]	1% of total capital investment	$80,988	$69,067	$57,002
(16) Administration [1]	2% of total capital investment	$161,977	$138,134	$114,005
Total IOC		$531,098	$463,432	$394,954
CAPITAL RECOVERY COSTS (CRC)	CRF of 0.1627 times TCI	$1,317,682	$1,123,720	$927,429
ANNUALIZED COSTS (AC):	DOC + IOC + CRC	$2,890,409	$2,551,836	$2,371,512

[1] Based on Capital Cost Factors for ESP, from OAQPS Control Cost Manual, Fourth Edition (1990).

[2] Based on maximum of 25% coal firing for the Sol–Energy cogeneration facility.

Note 1: Capital cost factors are $4,000/MMBtu of heat input for Furnace Injection and $3,500/MMBtu of heat input for Duct Injection (both using lime) projecting from Babcock and Wilcox's cost analysis summary for Ohio's Edison Power Plant located at Edgewater, Ohio. The capital cost factor for the sodium sorbent injection process is $3,000/MMBtu of heat input based on estimations from Colorado Springs Utilities Company and Publlic Service Company of Colorado. Total Capital Investment for the dry sorbent injection process was calculated using these cost factors for a total of two 460–MMBtu boilers for the proposed Sol–Energy facility.

Note 2: Capital cost estimates of the ash handling system prior to the upsizing in order to handle the additional ash generated from the DSI process are approximately $260,000.

Table 20. Cost Estimates for Sorbent Injection Systems for Sol-Energy Using 1.5% S Coal (Revised 02/05/93).

Cost Items	Basis	Furnace Injection Lime Sorbent	Duct Injection Lime Sorbent	Duct Injection Sodium Sorbent
DIRECT CAPITAL COSTS (DCC):				
(1) Dry Sorbent Injection System	See Note 1	$3,680,000	$2,944,000	$2,760,000
(a) Flue Gas Humidification System	Included	Included	Included	--
(b) Upsizing Auxiliary Equipment	0.05 x (1)	$184,000	$147,200	$138,000
(2) Upsizing ESP for Particulate Control				
(a) Percent Increase in Size	Based on ESP's Vendor Estimate	60%	65%	35%
(b) Cost of Upsizing for 2 Boilers	$7,000 per 1% Increase per boiler	$840,000	$910,000	$490,000
(3) Upsizing Ash Handling System				
(a) Cost of Upsizing for 2 Boilers	12% of current system cost. See Note 2.	$31,200	$31,200	$31,200
Subtotal of DCC	(1) + (2b) + (3a)	$4,735,200	$4,032,400	$3,419,200
INDIRECT CAPITAL COSTS (ICC):				
(4) Indirect Installation				
(a) Engineering & Supervision [1]	0.20 x (DCC)	$947,040	$806,480	$683,840
(b) Construction & Field Expenses [1]	0.20 x (DCC)	$947,040	$806,480	$683,840
(c) Contruction Contractor Fee [1]	0.10 x (DCC)	$473,520	$403,240	$341,920
(d) Contingencies	0.25 x (DCC)	$1,183,800	$1,008,100	$854,800
(5) Other Indirect Costs				
(a) Start-up, Perf. Test & Model Study [1]	0.04 x (DCC)	$189,408	$161,296	$136,768
(b) Working Capital	30-day DOC	$120,359	$120,359	$120,359
Total ICC:	(4) + (5)	$3,861,167	$3,305,955	$2,821,527
TOTAL CAPITAL INVESTMENTS (TCI):	DCC + ICC	$8,596,367	$7,338,355	$6,240,727
DIRECT OPERATING COSTS (DOC):				
(6) Labor				
Operator [2]	$22/hr; 5,760 hr/yr total for 2 boilers	$126,720	$126,720	$126,720
Supervisor [1]	15% of operator cost	$19,008	$19,008	$19,008
(7) Maintenance [2]	0.05 x (DCC)	$236,760	$201,620	$170,960
(8) Replacement Parts	0.03 x (DCC)	$142,056	$120,972	$102,576
(9) Utilities				
(a) Electricity for ESP	$85 per MW-hr	$14,825	$16,060	$8,648
for Auxiliary Equipment	$85 per MW-hr	$29,356	$31,802	$17,124
(b) Humidification Water	$0.27 / 1,000 gal	$5,063	$5,063	--
(c) Steam Lost or Reheat	\$6.19/10^3 lb	$39,298	--	--
(10) Raw Chemicals				
(a) Hydrated Lime (74% purity)	$140 / ton delivered for approx. 6,000 TPY	$840,000	$840,000	--
(b) Sodium Bicarbonate (98% purity)	$200 / ton delivered for approx. 6,070 TPY	--	--	$1,214,000
(11) Differential Fuel Cost or Credit				
(a) 1.5% S Coal	$3 savings per ton for 68,625 TPY	-$205,875	-$205,875	-$205,875
(12) Solid Disposal	$27 / ton for Lime; $50 /ton for NaHCO3-type	$197,100	$207,900	$192,500
Total DOC		$1,444,310	$1,363,270	$1,645,661
INDIRECT OPERATING COSTS (IOC):				
(13) Overhead [1]	60% of operating labor & maintenance	$229,493	$208,409	$190,013
(14) Property Taxes [1]	1% of total capital investment	$85,964	$73,384	$62,407
(15) Insurance [1]	1% of total capital investment	$85,964	$73,384	$62,407
(16) Administration [1]	2% of total capital investment	$171,927	$146,767	$124,815
Total IOC		$573,347	$501,943	$439,642
CAPITAL RECOVERY COSTS (CRC)	CRF of 0.1627 times TCI	$1,398,629	$1,193,950	$1,015,366
ANNUALIZED COSTS (AC):	DOC + IOC + CRC	$3,416,287	$3,059,163	$3,100,669

[1] Based on Capital Cost Factors for ESP, from OAQPS Control Cost Manual, Fourth Edition (1990).

[2] Based on maximum of 25% coal firing for the Sol-Energy cogeneration facility.

Note 1: Capital cost factors are $4,000/MMBtu of heat input for Furnace Injection and $3,500/MMBtu of heat input for Duct Injection (both using lime) projecting from Babcock and Wilcox's cost analysis summary for Ohio's Edison Power Plant located at Edgewater, Ohio. The capital cost factor for the sodium sorbent injection process is $3,000/MMBtu of heat input based on estimations from Colorado Springs Utilities Company and Public Service Company of Colorado. Total Capital Investment for the dry sorbent injection process was calculated using these cost factors for a total of two 460-MMBtu boilers for the proposed Sol-Energy facility.

Note 2: Capital cost estimates of the ash handling system prior to the upsizing in order to handle the additional ash generated from the DSI process are approximately $260,000.

ATTACHMENT B

ANSI STANDARD K-61.1-1989

ANSI
K-61.1-1989
Revision of
K61.1-1981

American National Standard Safety Requirements for the Storage and Handling of Anhydrous Ammonia

Secretariat
Compressed Gas Association, Inc.
1235 Jefferson Davis Highway
Arlington, VA 22202

CGA Pamphlet G-2.1 — 1989

Approved March 17, 1989
American National Standards Institute, Inc.

2.24 **Pressure Relief Valve.** A device designed to open to prevent an increase in internal fluid pressure in excess of a specified value due to an emergency or abnormal condition, and to close and prevent further flow after normal conditions have been restored. Refer to ANSI B95.1, *Terminology for Pressure Relief Devices.* [12]

2.25 **Protective Gloves, Boots, and Suits.** Items made of rubber or other material impervious to ammonia. Gloves refer to gauntlet-style of sufficient length to allow for cuffing, and which provide thermal protection suitable for ammonia exposure.

2.26 **psig and psia.** Refers to pounds per square inch gauge and pounds per square inch absolute, respectively.

2.27 **Repair.** The work necessary to restore a container or system to a safe and satisfactory operating condition, provided that in all cases the container or system design shall continue to comply with the requirements of this standard or the standard in effect at the time of installation. In addition, the original design of the container or system shall not be altered by the repair. Repairs include the addition or replacement of pressure or nonpressure parts which do not change the design temperature or pressure of the container or system.

2.28 **Semi-trailer.** Any highway motor vehicle with or without motive power designed to be drawn by another motor vehicle, and so constructed that some part of its weight and that of its load rests upon or is carried by the towing vehicle.

2.29 **Shall or Must.** A mandatory requirement.

2.30 **Should.** A recommendation or that which is advised, but not required.

2.31 **Short Term Exposure Limit.** A 15-minute time weighted average exposure to an air contaminant which should not be exceeded at any time during a work day and which should not be repeated more than four times a day. Exposures at the STEL should not occur at less than 60-minute intervals.

2.32 **System.** Refers to an assembly of equipment consisting essentially of the container or containers, hoses, appurtenances, pumps, compressors, and interconnecting piping.

2.33 **Trailer.** Any highway motor vehicle with or without motive power designed to be drawn by another motor vehicle and so constructed that no part of its weight except the towing device rests upon the towing vehicle. Normally called a "full trailer."

2.34 **Transfer, Fill, and Charge.** These terms may be used interchangeably and mean movement of a quantity of ammonia from one container to another container or cylinder, as contrasted to feeding ammonia to a use or application device.

3. SAFETY

It is important that personnel understand the properties of ammonia and that they be thoroughly trained in safe practices for its storage and handling. Some of the important physical properties of ammonia are listed in Table 1.

3.1 Training

3.1.1 Any person required to handle, transfer, transport, or otherwise work with ammonia shall be trained to understand the properties of ammonia, to become competent in safe operating practices, and to take appropriate actions in the event of a leak or an emergency.

3.2 Normal Conditions

3.2.1 Any person making, breaking, or testing any ammonia connection, transferring ammonia, or performing maintenance or repair on an ammonia system under pressure, shall wear protective gloves, and chemical splash goggles. A full faceshield may be worn over the goggles. However, a faceshield shall not be worn as a substitute for a primary eye protection device (goggles).

3.3 Emergency Conditions

3.3.1 If a leak occurs in a permanent storage installation, the personnel trained for and designated to act in such emergencies shall:

(1) See that persons not required to deal with an emergency are evacuated from the contaminated area and limit access to the contaminated area
(2) Put on suitable respiratory protection
(3) Wear protective gloves, suits and boots in contaminated areas
(4) Shut off the appropriate valves
(5) Notify local, state, or federal governmental regulatory authorities as may be appropriate and required by law

3.4 Permanent Storage Installations

3.4.1 All permanent storage installations shall have on hand, as a minimum, the following equipment for emergency and rescue purposes:

3.4.1.1 Two full face gas masks, each with one spare ammonia canister in a readily accessible location for use in ammonia concentrations less than the IDLH. See 2.19.

NOTE: A full facepiece ammonia gas mask will provide effective respiratory protection in concentrations of ammonia in air that are not immediately dangerous to life or health for short periods of time. A gas mask is not recommended for respiratory protection in concentrations exceeding the IDLH except for escape purposes only. Facepiece fitting should be used to determine the ability of each individual gas mask wearer to obtain a satisfactory fit. If ammonia vapor is detected within the gas mask facepiece, the facepiece fit is improper, the ambient concentration is excessive, or the canister is exhausted, the wearer should return to fresh air immediately to take appropriate corrective measures. The life of a canister in service is controlled by many factors including the concentration of ammonia vapor to which it is exposed.

Canisters should not be opened until ready for use and should be discarded after use. Canisters should be discarded and replaced when the shelf life expiration date marked on the canister is exceeded. When canisters include an end-of-service indicator, the manufacturer's expiration instructions are to be followed. In addition to this protection, an independent air-supplied, positive-pressure, self-contained breathing apparatus, approved by NIOSH/MSHA, should be used for entry into concentrations of ammonia vapor that are unknown or immediately dangerous to life or health. The American National Standard Z88.2, *Practices for Respiratory Protection*, should be referred to wherever respirators may be used. [13]

3.4.1.2 One pair of protective gloves impervious to ammonia.

3.4.1.3 One pair of protective boots impervious to ammonia.

3.4.1.4 One protective slicker and/or protective pants and jacket, all impervious to ammonia.

3.4.1.5 Easily accessible emergency shower and a plumbed eyewash unit or in lieu of these, at least 150 gallons (570 L) of clean water in an open top container.

NOTE: It is recommended that the distance from the point of greatest potential exposure to ammonia to the emergency water supply should not exceed 10 seconds travel time or 100 feet (30 m).

3.4.1.6 Chemical splash goggles or chemical splash goggles with full faceshield to be worn over the device (goggles).

NOTE: A full faceshield, if used, shall only be worn as secondary eye protection supplementing the primary eye protection afforded by the chemical splash goggles. A faceshield is not to be worn as a substitute for a proper primary eye protection device (goggles).

3.5 Cargo Tanks

3.5.1 Each cargo tank transporting ammonia, except an implement of husbandry, shall carry:

3.5.1.1 For first aid purposes, at least 5 gallons (20 L) of clean water in a container designed to provide ready access to the water for flushing any area of the body contacted by ammonia.

3.5.1.2 One pair of protective gloves impervious to ammonia.

3.5.1.3 A full facepiece gas mask with an ammonia canister and at least one spare canister.

3.5.1.4 Chemical splash goggles, or chemical splash goggles with full faceshield, to be worn over the goggles.

NOTE: A full faceshield, if used, shall only be worn as secondary eye protection supplementing the primary eye protection afforded by the chemical splash goggles. A faceshield is not to be worn as a substitute for a proper primary eye protection device (goggles).

3.6 Leaks in Transportation Equipment

3.6.1 If a leak occurs in transportation equipment and it is not practical to stop the leak, the driver should make every effort possible, including moving the vehicle to an isolated location downwind from populated communities or heavily traveled highways, to transfer the contents to another approved ammonia container. Local authorities should be notified and assistance requested as needed.

3.7 Cylinder and DOT Portable Tank Installations

3.7.1 At ammonia installations comprising cylinders and DOT portable tanks, the employer shall provide ready access to a supply of clean, running water for emergency use, including provision for flushing of the eyes by an employee in the event of contact with ammonia, or a self-contained eyewash unit with clean water.

4. USE OF WATER IN EMERGENCIES

4.1 Human Exposure

4.1.1 If liquid ammonia contacts the skin or eyes, the affected area should be promptly and thoroughly flushed with clean water for at least 15 minutes total, with the eyes receiving first attention. Eyelids must be held open during flushing. Skin irrigation should include the ears, chin, neck, armpit, and groin areas as appropriate. Contaminated clothing should be removed only after it is thawed. Do not use neutralizing solutions or ointments on the affected areas. Water used for flushing should be within a temperature range and at a controlled flow rate to avoid causing the patient additional injury or discomfort. [10] A physician should treat all cases of exposure to liquid ammonia.

An opthalmologist should be consulted immediately after flushing in the event of eye exposure.

4.1.2 **Nose and Throat.** If ammonia has entered the nose or throat and the patient can swallow, have him drink large quantities of water. Never give anything by mouth to an unconscious person.

4.2 Accidental Release

4.2.1 In the event of an accidental release, the concentration of ammonia vapor in the air can be reduced effectively by the use of adequate volumes of water applied through spray or fog nozzles. Downwind control should be achieved by directing water fog nozzles toward the point of ammonia release from a downwind position. See 3.3.1.

4.2.2 Water should not be used on liquid ammonia spills. Water should only be directed in the form of fog or spray at the cloud emanating from the liquid pool. See 3.3.1.

4.2.3 In the event of a large vapor release from a container, the tank should not be sprayed with water. Under these circumstances, water fog or spray should be applied to the vapor following the procedures outlined in 4.2.1.

4.3 Fire Exposure

4.3.1 If an ammonia container is exposed to fire and cannot be moved, water fog or spray (preferably 500 gal/min [2m^3/min] or more) should be used to cool it. Use caution if flame impinges on the vapor space of the container; violent rupture of the container is possible. Water fog or spray should always be applied to the tank from the sides, as the heads usually are first to rupture. If the fire cannot be controlled and it appears the tank may rupture, the surrounding area should be evacuated to a minimum distance of 2000 feet (600 m) in all directions.

4.4 Absorption in Water

4.4.1 If ammonia is leaking from a container, the safest, practical means should be taken to stop or abate the leak. If the leak cannot be stopped, the ammonia should be fed to the point of use or transferred to another suitable ammonia container. Small quantities of ammonia from a leaking container may be discharged into a vessel containing sufficient water to absorb it. Sufficient water may be taken to be ten parts of water to one part of ammonia. The ammonia should be injected into the water as near the bottom of the vessel as practical. If a hose is used to inject ammonia into water, the hose must be weighted or secured so that the end of the hose will remain near the bottom of the vessel.

4.4.2 Runoff of ammonia contaminated water into streams or other bodies of water should be avoided when possible. Releases of ammonia shall be reported to environmental protection and/or other regulatory authorities as may be appropriate and required by law.

5. BASIC RULES

This Section applies to all sections of this standard unless otherwise noted.

5.1 Equipment and Systems

5.1.1 The provisions of 5.2 shall not be construed as prohibiting the continued use or reinstallation of containers constructed and maintained in accordance with the 1949, 1950, 1952, 1956, 1959, 1965, 1968, 1971, 1974, 1977, 1980, 1983, and 1986 editions of the ASME Code, or any revisions thereof, in effect at the time of fabrication.

5.1.2 Systems and components which were fabricated, installed and maintained in accordance with the American National Standard K61.1, *Safety Requirements for the Storage and Handling of Anhydrous Ammonia*, or the Compressed Gas Association *Standards for the Storage and Handling of Anhydrous Ammonia and Ammonia Solutions—Part 1 Anhydrous Ammonia*, or The Agricultural Nitrogen Institute, Standard M-1, *Standard for Storage and Handling of Agricultural Ammonia*, in effect at the time of installation, are acceptable for continued use.

NOTE: The latter two standards are no longer published.

5.2 Requirements for New Construction and Original Test of Containers (Including DOT Portable Tanks), Other Than Refrigerated Storage Tanks (see exception in 7.1.3).

5.2.1 Containers used with systems covered in Sections 6, 9, 11, and 12 shall be made of steel or other material compatible with ammonia, and tested in accordance with the current ASME Code. An exception to the ASME Code requirements is that construction under Table UW 12 at a basic joint efficiency of under 80% is not authorized. Containers shall not be inspected and tested under the provisions of UG-90(c)(2) of the ASME Code.

5.2.2 Containers designed and constructed in accordance with the ASME Code, other than refrigerated storage containers, shall comply with the following additional requirements:

5.2.2.1 The entire container shall be post-weld heat treated after completion of all welds in and/or to the shells and heads. The method employed shall be as prescribed in the ASME Code. It is recommended that post-weld heat treatment be performed in a furnace of a size sufficient to accommodate the entire container. Welded attachments to pads may be made after post-weld heat treatment. [8]

5.2.2.2 Steels used in fabricating pressure-containing parts of a container shall have a tensile strength no greater than a nominal 70 000 psi (480 MPa) (does not apply to Sections 8, 9, and 10).

5.2.3 All containers, except refrigerated storage tanks with a design pressure of 15 psig (100 kPa) and less, and cylinders and containers covered in Section 8, shall be inspected by a person who holds a valid National Board Commission as an Authorized Inspector or as an Owner-User Inspector as defined in the National Board Inspection Code. [11]

5.2.4 Welding for the repair or alteration of pressure-containing parts of a container shall be performed in compliance with the applicable provisions of the current edition of the National Board Inspection Code. [11] Where specific procedures are not given, it is intended that subject to acceptance of the Inspector, all repair or alteration shall conform insofar as possible to the ASME Code section and edition to which the container was constructed.

5.3 Location of Containers

5.3.1 Selection of a location for a storage container shall be made considering the potential physiological and environmental effects of ammonia on the surroundings adjacent to the proposed site. Containers shall be located outside of buildings except in buildings or sections thereof especially approved for the purpose.

5.3.2 Containers shall be located at least 50 feet (15 m) from a dug well or other sources of potable water supply, unless the container is a part of a water treatment installation.

5.3.3 The minimum distance of a storage container to dwellings or to population centers shall be in accordance with the requirements of the local jurisdiction having authority.

5.3.4 Container locations shall comply with Table 3.

5.3.5 Container storage areas shall be accessible to emergency vehicles and personnel.

5.3.6 Areas within 10 feet (3 m) of a storage container shall be maintained clear of dry grass and weeds and other combustible materials.

5.4 Markings of Non-Refrigerated Containers and Systems Other than DOT Containers

5.4.1 Each system nameplate, when required, shall be made of a noncorroding metal permanently attached to the system by continuous welding around its perimeter, and located so as to be readily accessible for inspection. Nameplates shall be maintained in legible condition and include markings as prescribed in 5.4.2.

TABLE 3

MINIMUM DISTANCES FOR LOCATION OF AMMONIA STORAGE CONTAINERS
(Customary Units and SI Units)

Nominal Capacity of Container (Gallons or Cubic Meters)	Minimum Distances (in feet or meters) from Each Container to:		
	Line of Adjoining Property which may be built upon, Highways & Mainline of Railroad	Place of Public Assembly	Institution Occupancy
*Over 500 to 2000 gals	25 ft	150 ft	250 ft
Over 2000 to 30 000 gals	50 ft	300 ft	500 ft
Over 30 000 to 100 000 gals	50 ft	450 ft	750 ft
Over 100 000 gals	50 ft	600 ft	1 000 ft
Over 2 to 8 m^3	8 m	45 m	75 m
Over 8 to 110 m^3	15 m	90 m	150 m
Over 110 to 400 m^3	15 m	140 m	230 m
Over 400 m^3	15 m	180 m	300 m

*NOTE: For 500 gallons (2m^3) or less, see 5.3.1 and 5.3.3.

5.4.2 Each container or system covered in Sections 6, 9, 10 (except "ton containers" and cylinders), 11, and 12 shall be marked as specified in the following:

5.4.2.1 With a marking as required by paragraph UG-116 of the ASME Code and identifying compliance with the rules of the ASME Code under which the container is constructed

5.4.2.2 With National Board of Boiler and Pressure Vessel Inspectors stamping to indicate registration of the container with this organization

5.4.2.3 With a notation on the container and system nameplate to indicate whether the system is designed for aboveground or underground installation or both

5.4.2.4 With the minimum and maximum temperatures in degrees Fahrenheit (°F) or degrees Celsius (°C) for which the container is designed

5.4.2.5 With the wall thickness of the container shell and heads in inches or millimeters (mm)

5.4.2.6 With the water capacity of the container in pounds or kilograms (kg), or U.S. standard gallons or cubic meters (m^3) at 60°F (15.6°C)

5.4.2.7 With the outside surface area of the container in square feet or square meters (m^2)

5.4.2.8 Marking required by paragraph UG-116 of the ASME Code shall be arranged in accordance with the requirements of UG-118(b). Marking required by 5.4.2.2 through and including 5.4.2.7 must be stamped on the nameplate required in 5.4.1, following the marking arrangement specified by UG-118(b) on a separate nameplate immediately adjacent to the ASME Code nameplate. Requirements of 5.4.1 shall also apply to the separate nameplate.

5.4.3 Each container or system covered in Sections 6, 9, 10 (except cylinders), 11, and 12 shall be fitted with a liquid level gauge indicating the maximum level to which the container may be filled with liquid anhydrous ammonia at temperatures between 20°F (-7°C) and 100°F (38°C), except on containers provided with fixed maximum level indicators, such as fixed length dip tubes or containers that are filled by weight. Marks shall be in increments of not more than 20°F (10°C). (See 5.9.3 regarding requirement for thermometer well and thermometer.)

5.4.4 All container openings except for pressure relief valves, pressure indicating devices, thermometer wells, or liquid level indicators shall be marked, stenciled, tagged, or decaled to indicate whether the opening is in contact with the liquid or vapor phase when the container is filled to the maximum allowable filling density.

5.5 Container Appurtenances

5.5.1 All appurtenances of each system shall be approved. See 2.3.

5.5.2 All appurtenances shall be designed for not less than the maximum working pressure of that portion of the system on which they are installed. All appurtenances shall be fabricated from materials proved suitable for anhydrous ammonia service.

5.5.3 All connections to containers except those for pressure relief devices, thermometer wells, liquid level gauging devices, or those fitted with a No. 54 (0.055 in or 1.40 mm) drill size orifice, or those plugged, shall have shut-off valves located as close to the container as practical.

5.5.4 Excess flow valves shall close automatically at the rated flows of vapor or liquid as specified by the manufacturer. The connections and line, including valves and fittings being protected by an excess flow valve, shall have a greater capacity than the rated flow of the excess flow valve.

5.5.5 Liquid level gauging devices that require bleeding of the product to the atmosphere, and which are so constructed that outward flow will not exceed that passed by a No. 54 (0.055 in or 1.40 mm) drill size opening, need not be equipped with excess flow valves.

5.5.6 An opening in a container to which a pressure gauge connection is made need not be equipped with an excess flow valve, if such an opening is not larger than No. 54 (0.055 in or 1.40 mm) drill size.

5.5.7 An excess flow or back pressure check valve, where required by this standard, shall be installed directly in the container opening.

5.5.8 Excess flow valves shall be designed with a by-pass, not to exceed a No. 60 (0.040 in or 1.02 mm) drill size opening, to allow equalization of pressures.

5.5.9 Shut-off valves with an integral excess flow valve shall be designed for proper installation in a container opening so that the excess flow valve will close in the event that the valve body, extending above the coupling, is sheared or broken off.

5.5.10 All excess flow valves shall be plainly and permanently marked with the name or trade-mark of the manufacturer, the catalog number, and the rated capacity.

5.5.11 Each liquid filling connection shall have a positive shut-off valve in conjunction with either an internal back-pressure check valve or an internal excess flow valve. Vapor connections shall have a positive shut-off valve together with an internal excess flow valve.

5.5.12 Quick opening (1/4 turn) valves are not recommended for use on transfer lines.

5.6 Piping, Tubing and Fittings

5.6.1 All piping, tubing, and fittings shall be made of steel or other material suitable for anhydrous ammonia service.

5.6.2 All piping, tubing, and fittings shall be designed for a pressure not less than the maximum pressure to which they may be subjected in service.

5.6.3 All piping shall be supported in accordance with good piping practices and provisions shall be made as necessary for expansion, contraction, impact, vibration, and for settling. All non-refrigerated ammonia piping shall conform to ANSI/ ASME B31.3, *American National Standard for Chemical Plant and Petroleum Refinery Piping;* all refrigerated ammonia piping used with refrigerated systems shall conform to ANSI/ASME B31.5, *American National Standard for Refrigeration Piping*, sections of the American Standard Code for Pressure Piping, as they apply to ammonia. [14] and [15]

5.6.4 Piping used on non-refrigerated systems shall be at least ASTM A-53 Grade B seamless or Electric Resistance Welded Pipe. [16] Pipe joints shall be threaded, welded or flanged. Pipe shall be at least Schedule 40 when joints are welded, or welded and flanged. Pipe shall be at least Schedule 80 when joints are threaded. Brass, copper, or galvanized steel pipe or tubing shall not be used. Threaded nipples shall be seamless. Welding shall be done by a welder certified in accordance with the ASME Code, Section IX, "Welding Qualifications." [17] Tubing joints shall be flared and made up with flare type fittings complying with ANSI/SAE J513f. [18]

5.6.5 All metal flexible connections for permanent installations shall have a minimum working pressure of 250 psig (1700 kPa) (safety factor of 4). For temporary installations, hose meeting the requirement of 5.7 may be used.

5.6.6 Cast iron fittings shall not be used. Those parts of valves which are subjected to gas pressure should be made of steel, ductile (nodular) iron, or malleable iron. Valves in this case include shut-off valves, excess flow valves, back check valves, emergency shut-off valves, and remotely controlled valves. Ductile iron shall meet the requirements of ANSI/ ASTM A395 and malleable iron the requirements of ANSI/ASTM A47. [19]

5.6.7 Adequate provisions shall be made to protect all exposed piping from physical damage that might result from impact by moving machinery, automobiles or trucks, or any other equipment at the facility.

5.6.8 Joint compounds shall be resistant to ammonia at the maximum pressure and temperature to which they may be subjected in service.

5.6.9 After assembly, all piping, hose, and tubing shall be tested and proved to be free from leaks at a pressure not less than the normal operating pressure of the system.

5.7 Hose Specification

5.7.1 Hose used in ammonia service and subject to container pressure shall conform to the American National Standard RMA IP-14, *Specifications for Anhydrous Ammonia Hose* (see Appendix A).

5.7.2 Hose subject to container pressure shall be designed for a minimum working pressure of 350 psig (2400 kPa) and a minimum burst pressure of 1750 psig (12 000 kPa). Hose assemblies, when made up, shall be capable of withstanding a test pressure of 500 psig (3400 kPa).

5.7.3 Hose and hose connections located on the low pressure side of flow control, or pressure reducing valves on devices discharging to atmospheric pressure, shall be designed for the maximum low side working pressure. All connections shall be designed, constructed, and installed so that there will be no leakage when connected. Shut-off valves on the end of liquid and vapor transfer hoses shall be equipped with bleed valves to enable the operator to bleed off pressure prior to disconnecting the hoses.

5.7.4 Where liquid transfer hose is not drained of liquid upon completion of transfer operations, such hose shall be equipped with an approved shut-off valve at the discharge end. Provision shall be made to prevent excessive hydrostatic pressure in the hose. See 5.8.11.

5.7.5 On all hose one-half inch (13 mm) O.D. and larger, used in ammonia service and subject to container pressure, there shall be etched, cast, or impressed at five foot (1.5 m) intervals on the outer hose cover the following information:

Anhydrous Ammonia
XXX psig (Maximum Working Pressure)
Manufacturer's Name or Trademark
Year of Manufacture

5.7.6 Hose in service shall be requalified periodically in accordance with requirements specified in CGA P-7, *Standard for Requalification of Cargo Tank Hose Used in the Transfer of Compressed Gases.* [20]

5.8 Pressure Relief Devices

5.8.1 Every container used in systems covered by Sections 6, 11, and 12 shall be provided with one or more pressure relief valves of the spring-loaded type conforming with the applicable requirements of UL 132, *Standard on Safety Relief Valves for Anhydrous Ammonia and LP-Gas*, or other equivalent pressure relief valve standard. [21]

5.8.2 Pressure relief valves shall be in direct communication with the vapor space of the container.

5.8.3 The discharge from pressure relief valves shall be vented away from the container, upward and unobstructed to the atmosphere. All pressure relief valve discharge openings shall have suitable rain caps that will allow free discharge of the vapor and prevent the entrance of water. Provision shall be made for draining condensate which may accumulate. The rate of the discharge shall be in accordance with the provisions of Appendix B.

5.8.4 Container pressure relief valves with relation to the design pressure of the container shall be set to start-to-discharge as follows:

Containers	Minimum	Maximum*
ASME U-68, U-69	110%	125%
ASME U-200, U-201	95%	100%
ASME 1952, 1956, 1959, 1962, 1965, 1968, 1971, 1974, 1977, 1980, 1983, 1986, and 1989.	95%	100%
API-ASME	95%	100%
U.S. Coast Guard (As required by USCG regulations)		
DOT (As required by DOT regulations)		

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*A relief valve manufacturer's tolerance of plus 10% is permitted.

5.8.5 Pressure relief valves used on containers covered by Sections 6, 11, and 12 shall be constructed to discharge at not less than the rates required in 5.8.3 before the pressure is in excess of 120% (not including the 10% tolerance referred to in 5.8.4) of the maximum permitted start-to-discharge pressure setting of the device.

5.8.6 Pressure relief valves shall be so arranged that the possibility of tampering will be minimized. If the pressure setting adjustment is external, the relief valves shall be provided with means for sealing the adjustment.

5.8.7 Shut-off valves shall not be installed between the pressure relief valves and the containers or systems covered by Sections 6, 11, and 12, except that a shut-off valve may be used where the arrangement of the shut-off valve is such as always to afford the full capacity flow specified in 5.8.3 through a nonisolated pressure relief valve(s) which shall remain operative.

NOTE: The above exception is made to cover such cases as a three-way valve installed under two pressure relief valves, each of which has the required rate of discharge and is so installed as to allow either of the pressure relief valves to be closed off, but does not allow both pressure relief valves to be closed off at the same time. Another exception to this may be where two separate pressure relief valves are installed with individual shut-off valves. In this case, the two shut-off valve stems shall be mechanically interconnected in a manner which will allow full required flow of one pressure relief valve at all times. Still another exception is a pressure relief valve manifold which allows one valve to be closed off with the remaining unblocked valve or valves providing not less than the rate of discharge shown on the manifold nameplate.

5.8.8 Each pressure relief valve used with systems covered by Sections 6, 11, and 12 shall be plainly and permanently marked as follows:

(1) With the letters "AA" or the symbol "NH_3"

(2) The pressure in pounds per square inch gauge (psig) at which the valve is set to start-to-discharge

(3) The rate of discharge of the valve in cubic feet per minute of air at 60°F (15.6°C) and atmospheric pressure

(4) The manufacturer's name and catalog number

For example, a pressure relief valve marked AA-250-4200 (air) would mean that this valve is suitable for use on an anhydrous ammonia container; that it is set to start-to-discharge at 250 psig (1700 kPa); and that its rate of discharge (see 5.8.1, 5.8.3, and 5.8.4) is 4200 cubic feet per minute (120m^3/min) of air.

5.8.9 The flow capacity of the pressure relief valve shall not be restricted by any connection to it on either the upstream or downstream side.

5.8.10 The manufacturer or supplier of a pressure relief valve manifold shall publish complete data showing the flow rating through the combined assembly of the manifold with pressure relief valves installed. The manifold flow rating shall be determined by testing the manifold with all but one valve discharging. If one or more openings have restrictions not present in the remaining openings, the restricted opening or openings, or those having the lowest flow, shall be used to establish the flow rate marked on the manifold nameplate. The marking shall be similar to that required in 5.8.8 for individual valves.

5.8.11 A hydrostatic relief valve or equivalent shall be installed in each section of piping (including hose) in which liquid ammonia can be isolated between shut-off valves to relieve the pressure which could develop from the trapped liquid. If an equivalent pressure relieving device is used, the maximum accumulative pressure possible within the system shall not exceed the limits of the system.

5.8.12 The discharge opening from any pressure relief valve shall not terminate inside any building or below the highest roof line of any such building.

5.8.13 A pressure relief device shall be subject to a systematic, periodic visual external inspection at least annually, to determine that it:

(1) Meets the applicable requirements specified in 5.8

(2) Is free of evidence of tampering, damage, corrosion, or foreign matter that might prevent proper operation

(3) Is free of leakage when subject to pressures below the minimum allowable start-to-discharge setting

(4) Has a properly installed rain cap or other device to avoid entry of moisture or other matter into the relief valve outlet

(5) Has an open weep hole to permit moisture to escape

5.8.14 Any deficiency as may be found in 5.8.13 shall require immediate corrective action, replacement, or repair of the pressure relief device as may be appropriate.

5.8.15 No container pressure relief device shall be used after the replacement date as specified by the manufacturer of the device. If no date is specified, a pressure relief valve shall be replaced no later than five years following the date of its manufacture or last repair unless it has first been disassembled, inspected, repaired, and tested by the manufacturer, or by a qualified repair organization in a manner such that the valve's condition and performance is certified as being equivalent to the standards for the original valve.

5.9 Filling Densities. (See 2.15)

5.9.1 The filling densities for non-refrigerated containers shall not exceed the following:

	Aboveground	Underground
(1) Uninsulated	56%*	58%
(2) Insulated	57%	

(3) DOT containers and cylinders shall be filled in accordance with DOT regulations.

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*NOTE: This corresponds to 82% by volume at -28°F (-33.3°C), 85% by volume at 5°F (-15°C), 87.5% by volume at 30°F (-1.1°C), and 90.6% by volume at 60°F (15.6°C).

5.9.2 The filling density for refrigerated storage tanks shall be such that the tanks will not be liquid full at a liquid temperature corresponding to the vapor pressure at the start-to-discharge pressure setting of the pressure relief valve.

5.9.3 If containers are to be filled according to liquid level by any gauging method other than a fixed length dip tube gauge, each container should have a thermometer well and thermometer so that the internal liquid temperature can be easily determined and the amount of liquid and vapor in the container corrected to a 60°F (15.6°C) basis.

5.10 Transfer of Liquids

5.10.1 Anhydrous ammonia shall always be at a temperature suitable for the material of construction and design of the receiving containers. Certain steels are not suitable for refrigerated ammonia. See Appendix R of API Standard 620, *Recommended Rules for Design and Construction of Large Welded Low-Pressure Storage Tanks*, for materials for low temperature service. [22]

5.10.2 At least one qualified operator experienced in the procedures shall monitor the transfer of ammonia from the time the connections are first made until they are finally disconnected. Such monitoring may be performed by a person on site, or from a remote location, or by electronic means. Capability shall be provided to halt the transfer in the event of an emergency.

5.10.3 Cargo tanks and tank cars shall not be unloaded with gas pressure other than from an ammonia source.

5.10.4 Containers and cylinders shall be filled or used only upon the owner's authorization.

5.10.5 Containers and cylinders shall be gauged and charged only in the open atmosphere or in buildings provided for that purpose.

5.10.6 Pumps used for transferring ammonia shall be recommended and labeled for ammonia service by the manufacturer.

5.10.6.1 Positive displacement pumps shall be equipped with a pressure actuated by-pass valve on the discharge side of the pump. This valve shall operate to limit the pressure developed by the pump to the maximum for which the pump is rated. Piping or tubing sized to carry the full capacity of the pump at the actuation pressure of this valve shall connect the discharge of this valve with the container from which ammonia is being pumped. If this line is capable of being closed off by a valve, an additional by-pass device shall be incorporated in the pump to by-pass back to the suction port. The pressure actuated by-pass valve and the return piping or tubing shall be installed in accordance with the pump manufacturer's recommendations.

5.10.6.2 On the discharge side of the pump, before the by-pass valve line, install a pressure gauge graduated from 0 to 400 psig (0-2800 kPa).

5.10.6.3 Plant piping shall contain shut-off valves located as close as practical to pump connections.

5.10.7 Compressors used for transferring or refrigerating ammonia shall be recommended and labeled for ammonia service by the manufacturer.

5.10.7.1 Compressors, except those used for refrigeration, shall be designed for at least 250 psig (1700 kPa) working pressure. Crank cases of compressors not designed to withstand system pressure shall be protected with a suitable pressure relief valve.

5.10.7.2 Plant piping shall contain shut-off valves located as close as practical to compressor connections.

5.10.7.3 A pressure relief valve large enough to discharge the full capacity of the compressor shall be connected to the discharge side before any shut-off valve.

5.10.7.4 Compressors shall have pressure gauges at suction and discharge graduated to at least one and one-half times the maximum pressure that can be developed.

5.10.7.5 Adequate means, such as a drainable liquid trap, shall be provided on the compressor suction to minimize the entry of liquid into the compressor.

5.10.7.6 Where necessary to prevent contamination, an oil separator shall be provided on the discharge side of the compressor.

5.10.8 Loading lines shall be protected by a suitable backflow check valve, and unloading lines shall be protected by a suitable in-line excess flow valve. Piping shall be sized so as not to restrict flow rates to the extent that protective devices will not function.

5.10.8.1 By December 31, 1990, all stationary storage installations shall have an approved emergency shut-off valve installed in the fixed piping of the transfer system within 5 lineal feet (1.5 m) of where the hose or swivel piping is attached to the fixed piping. This requirement does not apply to a line feeding a fixed process system. The emergency shut-off valve shall be installed in the facility piping so that any break will occur on the side of the hose or swivel connection.

NOTE: This may be accomplished by concrete bulkheads or equivalent anchorage, or by the use of a weakness or shear fitting. Such anchorage is not required for tank car unloading.

5.10.8.2 An approved emergency shut-off valve shall incorporate a manually activated shut-off from a remote location, and a manually activated shut-off at the installed location.

5.10.9 Meters used for the measurement of liquid anhydrous ammonia shall be recommended and labeled for ammonia service by the manufacturer.

5.10.9.1 Liquid meters shall be designed for minimum working pressure of 250 psig (1700 kPa).

5.10.9.2 The metering system shall incorporate devices that will prevent the inadvertent measurement of vapor.

5.11 Liquid Level Gauging Devices

5.11.1 Each container except those filled by weight shall be equipped with an approved liquid level gauging device.

5.11.2 All gauging devices shall be arranged so that the maximum liquid level to which the container is filled is readily determined. See 5.4.3

5.11.3 Gauging devices that require bleeding of the product to the atmosphere such as the rotary tube, fixed tube, and slip tube devices, shall be designed so that the maximum opening of the bleed valve is not larger than No. 54 (0.055 in or 1.40 mm) drill size unless provided with an excess flow valve. (This requirement does not apply to farm vehicles used for the application of ammonia as covered in Section 12)

5.11.4 Gauging devices shall have a design pressure equal to or greater than the design pressure of the container on which they are installed.

5.11.5 Fixed maximum liquid level gauges shall be designed and installed to indicate a volumetric level not to exceed 85% of the container's water capacity. Refer to 5.9.1 NOTE in regard to volumetric limits at various temperatures.

NOTE: This does not apply to refrigerated storage.

5.11.6 Gauge glasses of the columnar type shall be restricted to stationary non-refrigerated storage installations. They shall be equipped with shut-off valves having metallic hand wheels, with excess-flow valves, and with extra heavy glass adequately protected with a metal housing applied by the gauge manufacturer. They shall be shielded against the direct rays of the sun.

5.12 Painting of Containers

5.12.1 Aboveground uninsulated containers should have a reflective surface maintained in good condition. White is recommended for painted surfaces, but other colors having similar reflecting characteristics are acceptable.

NOTE: Caution should be exercised to ensure that graphic designs, company logos, etc. do not significantly reduce the necessary reflective characteristics of the container surface.

5.13 Electrical Equipment and Wiring

5.13.1 Electrical equipment and wiring for use in ammonia installations shall be general purpose or weather resistant as appropriate.

5.13.2 Where concentrations of ammonia in air in excess of 16% by volume are likely to be encountered, electrical equipment, and wiring shall be installed to comply with the requirements specified for use in hazardous locations, Class I, Group D, of NFPA 70, *National Electrical Code*, Articles 500 and 501. [23]

6. SYSTEMS UTILIZING STATIONARY, PIER-MOUNTED OR SKID-MOUNTED ABOVEGROUND OR UNDERGROUND, NON-REFRIGERATED STORAGE

This section applies to stationary, pier-mounted, skid-mounted, aboveground or underground, non-refrigerated storage installations utilizing containers other than those constructed in accordance with U.S. Department of Transportation Specifications. All Basic Rules of Section 5 apply to this section unless otherwise noted.

6.1 Design Pressure and Construction of Containers

6.1.1 The minimum design pressure for non-refrigerated containers shall be 250 psig (1700 kPa). See 5.1.2.

NOTE: Existing U-68 and U-69 ASME Code containers with a design pressure of 200 psig (1400 kPa) are acceptable for reinstallation if re-certified to 250 psig (1700 kPa) in accordance with National Board Inspection Code procedures and if approved by the local jurisdictional authority. [11]

6.2 Container Valves and Accessories, and Discharge Connections

6.2.1 All vapor and liquid connections, except for pressure relief valves and those specifically exempted in 5.5.5 and 5.5.6, shall be equipped with approved excess flow valves; or in lieu thereof, may be fitted with approved quick-closing internal valves which, except during operating periods, shall remain closed.

6.2.2 Each storage container shall be provided with a pressure gauge graduated from 0 to 400 psig (0-2800 kPa). Gauges shall be designated for use in ammonia service.

6.2.3 All containers shall be equipped with a suitable vapor equalizing connection.

6.2.4 All containers shall be equipped with a fixed maximum liquid level gauge.

6.3 Pressure Relief Devices

6.3.1 Every container shall be provided with one or more pressure relief valves of spring-loaded or equivalent type which shall comply with the following:

6.3.1.1 Relief valves shall be installed in a manifold or other suitable device so that they can be replaced while the container remains pressurized. See 5.8.7 NOTE.

6.3.1.2 The discharge from pressure relief valves shall be vented away from the container, upward and unobstructed to the open air to an area such that persons, property, and the environment will not be harmed. Vent pipes shall not be restrictive or smaller in size than the pressure relief valve outlet connection. All pressure relief valves shall have suitable rain caps that will allow free discharge of the vapor and prevent the entrance of water. Suitable provision shall be made for draining condensate which may accumulate.

6.3.1.3 If desired, vent pipes from two or more pressure relief devices located on the same unit, or similar lines from one or more different units, may be run into a common header, provided the cross-sectional area of such header is at least equal to the sum of the cross-sectional areas of the individual vent pipes.

6.3.2 The rate of discharge of spring-loaded pressure relief valves installed on underground containers may be reduced by not more than 30 percent of the rate of discharge specified in Appendix B. Containers so protected shall not be uncovered after installation until the liquid ammonia has been removed. Containers which may contain liquid ammonia before being installed underground, and before being completely covered with earth, are to be considered aboveground containers when determining the rate of discharge requirements of the pressure relief valves.

6.3.3 On underground installations where there is a probability of the manhole or housing becoming flooded, the discharge from vent lines shall be located above the high water level. All manholes or housings shall be provided with ventilated louvers or their equivalent, the area of such openings equalling or exceeding the combined discharge areas of the pressure relief valves and vent lines which discharge their content into the manhole housing.

6.4 Installation of Storage Containers

6.4.1 Containers installed aboveground shall be provided with substantial reinforced concrete footings and foundations, or structural steel supports mounted on reinforced concrete foundations. In either case, the reinforced concrete foundations or footings shall extend below the established frost line and shall be of sufficient width and thickness to support the total weight of the containers and contents adequately. Where required by local codes, seismic loads shall be considered in the design of the footings and foundations. The foundation shall maintain the lowest point of the tank not less than 18 inches (0.5 m) above the ground. Floating type foundations shall also be acceptable providing the foundations are designed to adequately support the tank, contents, and piping. See 5.6.

6.4.2 Horizontal aboveground containers shall be mounted on foundations in such a manner as to permit expansion and contractions. Every container shall be supported so as to prevent the concentration of excessive loads. The bearing afforded by the saddles shall extend over at least one third of the circumference of the shell. Suitable means for preventing corrosion shall be provided on that portion of the container in contact with the foundations or saddles.

6.4.3 The location and installation of an underground container, and the type of corrosion control employed must have approval of the appropriate jurisdictional authority. Containers buried underground shall be placed so that the top of the container is at least one foot (0.3 m) below the surface. Should ground conditions make compliance with these requirements impractical, precautions shall be taken to prevent physical damage to the container. It is not necessary to cover the portion of the container to which a manhole and other connections are affixed. When necessary to prevent floating, containers shall be securely anchored or weighted.

6.4.4 As a minimum, underground containers shall be set on firm foundations (firm earth may be used) and be surrounded by at least six inches of noncorrosive, inert materials, such as soft earth, sand, or gravel well compacted into place. As a further means of resisting corrosion, the container and its piping, prior to placement in the ground, shall be provided with the following:

(1) a suitable protective coating applied after proper surface preparation in accordance with the coating manufacturer's recommendations

(2) cathodic protection

(3) electrical isolation of the container from ancillary equipment

Corrosion-resistant materials of construction may be used as an option. A container which has been coated shall be lowered into place in a manner to prevent abrasion or damage to the coating. Selection of the type of protection should be based upon the judgment of a qualified engineer having knowledge of the corrosion history of the area.

6.4.5 The horizontal distance between aboveground and underground containers of over 1200 gallons (4.5 m^3) capacity shall be at least 5 feet (1.5 m).

6.4.6 Secure anchorage or adequate pier height shall be provided against container flotation wherever sufficiently high flood water might occur.

6.4.7 A groundwater monitoring program meeting local, state, or federal regulatory requirements shall be established at the storage site by the owner of the underground storage system.

6.5 Reinstallation of Containers

6.5.1 Containers, once installed underground, shall not later be reinstalled aboveground or underground, unless they successfully withstand hydrostatic pressure retests at the pressure specified for the original hydrostatic test as required by the ASME Code under which the tank was constructed, and show no evidence of serious corrosion..

6.5.2 Where containers are reinstalled underground, the corrosion resistant coating, if used, shall be put in good condition; see 6.4.4. Where containers are reinstalled aboveground, pressure relief devices or gauging devices shall comply with 5.8, 5.11, and 6.3 as applicable to aboveground containers.

6.6 Marking Containers

6.6.1 Each container or group of containers shall be marked on at least two sides that are visible with the words, "ANHYDROUS AMMONIA," or "CAUTION — AMMONIA," in sharply contrasting colors with letters not less than 4 inches high (100 mm).

6.6.2 Each container or group of containers shall be conspicuously marked with a hazard warning label complying with 29 CFR 1910.1200. [5]

6.6.3 Each container or group of containers which are installed underground shall have a sign bearing marks and labeling as required in 6.6.1 and 6.6.2 located adjacent to the cover described in 6.7.2.

6.7 Protection of Container and Appurtenances

6.7.1 Containers and appurtenances shall be located or protected by suitable barriers so as to avoid damage by trucks or other vehicles. Main container shut-off valves shall be kept closed and locked when the installation is unattended. If the facility is protected against tampering by fencing or other suitable means, valve locks are not required.

6.7.2 All connections to underground containers should be located within a suitable dome, housing, or manhole fitted with a substantial removable cover.

6.7.3 Storage containers need not be electrically grounded.

6.8 Identification

6.8.1 A legible sign shall be displayed on the premises at which a storage system is located, so as to be readily visible to emergency response personnel, stating the name, address, and telephone number of the nearest representative, agent, or owner of the storage system.

7. REFRIGERATED STORAGE

This section applies specifically to systems utilizing tanks for the storage of anhydrous ammonia under refrigerated conditions. All Basic Rules of Section 5 apply to this section unless otherwise stated.

7.1 Design of Tanks

7.1.1 Tanks may be designed for any storage pressure desired as determined by economical design of the refrigerated system.

7.1.2 The design temperature shall be the minimum temperature to which the container will be refrigerated and shall be so designated.

7.1.3 Containers with a design pressure exceeding 15 psig (100 kPa) shall be constructed in accordance with 5.2 and the material shall be selected from those listed in API Standard 620, *Recommended Rules for Design and Construction of Large, Welded, Low-Pressure Storage Tanks*, Tables 2.02, R.2.2, R.2.3, or R.2.4. [22]

7.1.4 Tanks with a design pressure of 15 psig (100 kPa) and less shall be constructed in accordance with the general requirements of API Standard 620, including Appendix R. [22]

7.1.5 When austenitic stainless steels or nonferrous metals are used, the ASME Code shall be used as a guide in selection of materials for use at the design temperature. [15]

7.2 Installation of Storage Tanks (Aboveground)

7.2.1 Tanks shall be supported on suitable noncombustible foundations designed to accommodate the type of tank being used.

7.2.2 Adequate protection against flotation or other water damage shall be provided wherever high flood water might occur.

7.2.3 Tanks storing product which is at less than 32°F (0°C) shall be supported in such a way, or heat shall be supplied, to prevent the effects of freezing and subsequent frost heaving of the soil.

7.2.4 The area surrounding a refrigerated tank or group of such tanks shall be provided with drainage, or shall be diked or provided with other secondary containment systems to prevent accidental discharge of liquid from spreading to uncontrolled areas.

7.2.5 When drainage is employed, a slope of not less than one percent shall be provided. The drainage system shall terminate in an impounding basin having a capacity as large as the largest tank served.

7.2.6 Provision shall be made for the drainage of rain water from the dike or impounding area. Such drainage shall be provided with a positive means to stop the flow.

7.2.7 Where a dike is employed, the capacity of the diked enclosure shall be 110% of the capacity of the largest tank served. When computing the volume of the dike, allowance shall be made for the volume displaced by all other containers in the diked area.

ATTACHMENT C

SELECTED LITERATURE ON NOXSO AND ADVACATE PROCESSES

PILOT PLANT TEST FOR THE NOXSO FLUE GAS TREATMENT SYSTEM

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ABSTRACT

The NOXSO process is a FGT system that employs a reusable sorbent. A fluidized bed of sorbent simultaneously removes SO_2 and NO_x from flue gas. The spent sorbent is regenerated by treatment at high temperature with a reducing gas. Adsorbed NO_x is evolved on heating the sorbent to regeneration temperature. The concentrated stream of NO_x produced is returned to the boiler with the combustion air.

NOXSO Corporation, MK-Ferguson, W.R. Grace & Co., and Ohio Edison will conduct a pilot test of the NOXSO system at Ohio Edison's Toronto station. The plant treats 12,000 SCFM of flue gas containing 2300 ppm SO_2 and 350 ppm NO_x, which is roughly 1/20 the size of a commercial module. The paper summarizes the system design.

An additional test of the NO_x recycle concept will be conducted at the Babcock & Wilcox Research Center in Alliance, Ohio. The test apparatus is a 6 million Btu/hr small boiler simulator. It is a scaled-down version of B&W's single cyclone front wall fired boiler design. The proposed test plan and the data from previously reported NO_x reduction tests using a pc-fired system at the Pittsburgh Energy Technology Center (PETC) are included.

from: EPA/EPRI 1991 Joint Symposium on Stationary Combustion NO_x Control
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INTRODUCTION

The NOXSO Process simultaneously removes SO_2 and NO_x from the flue gas of coal-fired boilers using a dry, regenerable sorbent. Three previous tests of the NOXSO Process have been conducted. The first was a bench-scale program conducted at TVA's Shawnee Steam Plant for the purposes of establishing process chemistry and kinetics, quantifying sorbent attrition rates, and establishing the corrosion properties of different metals for use in specific applications within the NOXSO Process. The kinetic tests were all performed in a fixed bed reactor (1,2). Funding was provided by NOXSO and by the U.S. Department of Energy's (DOE) Pittsburgh Energy Technology Center (PETC). The second and third test programs were funded and conducted by DOE at PETC with technical guidance provided by NOXSO Corporation. The second test program was designed to test laboratory data in a scaled-up process, 3/4 MW in size (3). The third test program was a life-cycle test to determine sorbent physical and chemical performance over repeated cycles of adsorption and regeneration (4). The current test program is a 5 MW pilot plant that will provide the data necessary to scale up to a full size (100 MW) module (5). The pilot plant is currently under construction at Ohio Edison's Toronto Station and is scheduled to begin operation in May 1991. NOXSO Corporation is responsible for operation of the pilot plant while funding comes from DOE, the Ohio Coal Development Office, NOXSO Corporation, W.R. Grace & Co., and MK-Ferguson Co. A brief comparison of these four test programs is given in Table 1. Detailed information on test facility design, test results, and data analysis can be obtained from the previously referenced reports.

PROCESS DESCRIPTION

The NOXSO Process is a post-combustion flue gas treatment technology that simultaneously removes both SO_2 and NO_x from the flue gas generated by coal-fired utility boilers. The process utilizes a high surface area γ-alumina substrate impregnated with sodium to achieve removal efficiencies of 90% for SO_2 and 70%-90% for NO_x. A process flow diagram is shown in Figure 1, and a description of the process is given below.

Flue gas leaving the boiler passes through the combustion air preheater, the electrostatic precipitator, and into the NOXSO flue gas treatment system. The flue gas is first cooled to 120°C by vaporizing a water stream sprayed directly in the ductwork. The cooled flue gas is then passed through a fluidized bed of sorbent where the SO_2 and NO_x are simultaneously adsorbed. The clean flue gas flows through a cyclone where attrited sorbent is separated and returned to the adsorber bed. Finally, the flue gas is returned to the power plant duct and exhausted through the stack.

After the sorbent is loaded with SO_2 and NO_x, it is removed from the adsorbers and pneumatically conveyed to a sorbent heater. The sorbent heater is a three-stage fluidized bed where a hot air stream is used to heat the sorbent to 660°C. During the heating process, NO_x and loosely bound SO_2 are desorbed and transported away in the heating gas stream. The hot air stream exiting the sorbent heater is recycled back to the boiler replacing a portion of the combustion air while providing an energy credit to the NOXSO Process. At normal boiler operating conditions, the recycled NO_x will either be reduced by hydrocarbon fuel or suppressed by the formation of additional NO_x so that a steady-state equilibrium concentration of NO_x is attained.

Once the sorbent reaches a regeneration temperature of 660°C, it is transported from the sorbent heater to a moving bed regenerator. In the regenerator, sorbent is contacted with natural gas in a countercurrent fashion. The natural gas reduces sulfur compounds on the sorbent (mainly sodium sulfate) to primarily SO_2 and H_2S with some COS also formed (less than 2% of total inlet sulfur). Approximately 48% of the sodium sulfate is reduced to sodium sulfide which must subsequently be hydrolyzed in the steam treatment vessel. The moving bed steam treatment is obtained from the reaction of steam with Na_2S. The regenerator off-gasses are sent to a Claus plant where SO_2 and H_2S are reacted to form elemental sulfur. The sulfur is sold as a by-product of the NOXSO Process.

From the steam treatment vessel, the sorbent is fed to a sorbent cooler. The cooler is a three-stage fluidized bed where the sorbent is cooled to 120°C using an ambient air stream. The warm air exiting the cooler is further heated in a natural gas fired heater before being used to heat the sorbent in the fluidized bed heater. The cooled sorbent is returned to the adsorber completing one full cycle.

PROCESS CHEMISTRY

The NOXSO sorbent is prepared by spraying Na_2CO_3 solution on the surface of γ-alumina sphere (1.6 nominal diameter). Both sodium and alumina contribute to the NOXSO sorbent's capacity to adsorb SO_2 and NO_x from flue gas. Our laboratory tests show that the presence of steam in the flue gas helps the SO_2 sorption. The reaction of the sodium can be described as follows:

$$Na_2CO_3 + Al_2O_3 \xrightarrow{\Delta} 2NaAlO_2 + CO_2 \tag{1}$$

$$2NaAlO_2 + H_2O \longleftrightarrow 2NaOH + Al_2O_3 \tag{2}$$

$$2NaOH + SO_2 + \frac{1}{2}O_2 \longrightarrow Na_2SO_4 + H_2O \tag{3}$$

$$2NaOH + 2NO + \frac{3}{2}O_2 \longleftrightarrow 2NaNO_3 + H_2O \tag{4}$$

$$2NaOH + 2NO_2 + \frac{1}{2}O_2 \longrightarrow 2NaNO_3 + H_2O \tag{5}$$

Adsorbed nitrogen oxides are decomposed and evolved on heating the spent sorbent to regeneration temperature. The concentrated stream of NO_x produced on heat-up is returned to the boiler with the combustion air. This results in no significant increase of NO_x concentration in the boiler flue gas because of the reversibility of NO_x formation in the boiler (1,2).

The spent sorbent can be regenerated at high temperature with many kinds of reducing gases, such as H_2S, CO, H_2, natural gas, etc. The regeneration reaction, for example, using natural gas at 610°C is described below:

$$4Na_2SO_4 + CH_4 \longrightarrow 4Na_2SO_3 + CO_2 + 2H_2O \tag{6}$$

$$4Na_2SO_3 + 3CH_4 \longrightarrow 4Na_2S + 3CO_2 + 6H_2O \tag{7}$$

$$Al_2O_3 + Na_2SO_3 \longleftrightarrow 2NaAlO_2 + SO_2 \tag{8}$$

$$Al_2O_3 + Na_2S + H_2O \longleftrightarrow 2NaAlO_2 + H_2S \tag{9}$$

Although sulfite has not been identified in our studies, it is a likely intermediate in sulfate reduction. A detailed discussion on the existence of sulfide during regeneration had been given by Gavalas it.al. (6) who used CO to study the regeneration of alkali-alumina. The SO_2 and H_2S produced from regeneration are then converted to elemental sulfur in a Claus-type reactor.

$$SO_2 + 2H_2S \longleftrightarrow xS_{3/x} + 2H_2O \tag{10}$$

The sulfur produced is a marketable by-product of the process.

PROOF-OF-CONCEPT PILOT TEST

Background

On May 10, 1989, a consortium assembled by NOXSO Corporation signed a cost-shared contract with the DOE/PETC to conduct a POC test of the NOXSO process. The consortium consists of NOXSO, MK-Ferguson, W.R. Grace & Co., Ohio Edison and the Ohio Coal Development Office. The POC project will take approximately three years to complete, and the test will be conducted at a coal-fired Ohio Edison plant in Toronto, Ohio.

POC Test Site

The POC unit will treat flue gas from either Boiler #10 or Boiler #11 at Ohio Edison's Toronto Station. Two sources of flue gas will be tapped so that the POC test can continue as long as one of the boilers is operating. A slipstream of flue gas will be taken downstream of the Toronto plant's electrostatic precipitators. The Toronto boilers are pc-fired and burn Ohio coal containing 3.7% sulfur. The flue gas typically contains 2300 ppm SO_2 and 350 ppm NO_x.

POC Test Schedule

Detailed design engineering has been completed and the major pieces of equipment have been ordered. Construction began in April 1990 and will be completed in May 1991. The test will run through February 1992.

POC Process

The process flow diagram for the POC has shown in Figure 1. The system differs from a commercial application of the NOXSO technology in two important areas. First, the POC facility does not include a Claus plant, which in the commercial design would be used to produce a sulfur by-product from the concentrated stream of SO_2 and H_2S produced in the regenerator. This is because Claus technology is commercially available and therefore does not require testing at pilot scale. Second, the POC does not include NO_x recycle to the coal combustor. In the commercial design, NO_x in the air leaving the sorbent heater is recycled to the combustor as part of the combustion air. Since NO_x formation in the coal combustor is a reversible reaction, addition of NO_x to the combustion air suppresses the formation of NO_x in the combustor. However, NO_x recycle is impractical in the POC test since the POC treats less than 10% of the flue gas produced by Toronto Unit 10 or 11.

POC Test Unit Design

Data from three previous tests of the NOXSO process were used to design the POC test facility. A comparison of the three previous test programs was given in Table 1. The design specifications for the major equipment in the POC test facility are listed in Table 2.

Materials of Construction

During development of the NOXSO process, some corrosion problems were encountered, particularly in the regenerator. Different materials of construction were tested to withstand the high temperature environment of SO_2, H_2S, elemental sulfur, and sulfated sorbent. Corrosion results were documented in an earlier report (2), the practical results of the test program are discussed here.

In tests performed at the Shawnee Steam Plant, sorbent was heated with electrical resistance heaters made of Inconel 600, Monel 400, type 316 and type 316L stainless steel (SS). All these materials exhibited severe corrosion in areas of sorbent contact attributed to hot sulfation of nickel. It should be noted that the temperature of the heating elements themselves were substantially higher than the bed temperature of 600°C. The reactor, made out of either type 316 or type 316L SS, showed scale on the inside surfaces after use. When the reactor was made of type 446 SS or alonized type 316L SS, there was no scale and only a slight discoloration of the metal surfaces observed.

In the LCTU, the regenerator was made of alonized type 304 SS and showed no visible evidence of corrosion at the end of 330 regeneration cycles. Based on these results, it was felt that either 446 SS or alonized 304 or 316L SS would be satisfactory for the POC regenerator.

The sorbent heater also encounters hot sulfated sorbent and will therefore be made of type 304 SS. The bottom bed of the sorbent heater where the temperature is 660°C will be alonized. All other vessels will be made of standard A-285 or A-283 grade C carbon steel, as no corrosion problems are anticipated.

The other area in the NOXSO process that requires special consideration for materials is between the flue gas cooler and the adsorber. In this area, sub-acid dewpoint corrosion can occur. All previous NOXSO tests have cooled the flue gas indirectly while at the POC the flue gas will be cooled by a direct water spray in the ductwork. The flue gas temperature in this portion of the system will be 112°F so that an acid-resistant epoxy coating will be used to line the ductwork from the cooler to and including the bottom of the adsorber. This epoxy has not been tested previously by NOXSO, but there exists ample literature that supports its use as an acid resistant material in other similar applications.

NO_x RECYCLE

NO_x recycle will be implemented at the full-scale commercial demonstration plant. The concept of NO_x recycle has been tested previously using the 500 lb/hr coal combustor used for the 3/4 MW tests and also using a tunnel furnace capable of being fired with a variety of fuels including gas, fuel oil, coal, and coal-water mixtures.

Previous NO_x Recycle Results

NO_x recycle was tested by spiking the combustion air with varying concentrations of bottled NO_x and measuring the outlet NO_x concentration from the combustor. The net NO_x production rate was determined by a material balance on the combustor as shown schematically in Figure 2. The NO_x flow rate at the exit of the combustor minus the NO_x feedrate into the combustor equals the rate that NO_x is produced in the combustor, which is defined as the net NO_x production rate (R). For data reduction purposes, the NO_x production rate (R) and the NO_x feedrate (F) were normalized with respect to conditions at zero NO_x feed according to $R^*=R/R_o$ and $F^*=F/R_o$ where R_o is the NO_x production rate at $F = O$. Results from the 500 lb/hr combustor are compiled in Table 3. The measured data are NO_x concentration at the exit of the combustion system and the flow rate of NO_x fed into the combustor with the combustion air. Data provided in the other columns were calculated.

A plot of R* versus F* is shown in Figures 3 and 4 for both the 500 lb/hr combustor and the tunnel furnace, respectively. In each case, the data fall in a straight line, but with different slopes. The two lines are described by the equation $R^* = 1 - aF^*$. The parameter "a" is the slope of the line and also represents the fraction of NO_x fed to the combustor that is destroyed, The value of "a" is 0.65 for the 500 lb/hr combustor and 0.75 for the tunnel furnace. The data for the tunnel furnace include both natural gas combustion and coal-water slurry combustion.

These results demonstrate that the nature of the fuel has no affect on the effectiveness of the combustion system to reduce NO_x fed through the combustion air. Also, the NO_x reduction capability of a combustion system is independent of the amount of NO_x fed with the combustion air. Finally, the most important variables are those associated with the combustor design. NO_x recycle will be extensively studied at the Babcock & Wilcox Research Center in Alliance, Ohio.

Pilot-Scale NO_x Recycle Test

The power plant selected for the NOXSO full-scale demonstration (Ohio Edison's Niles Station, Niles, Ohio) uses cyclone burners. Since the destruction efficiency of NO_x recycle has not previously been tested with cyclone type burners, a demonstration of NO_x recycle with this type of coal combustor is necessary for the proper design of the NOXSO full-scale plant.

Pilot-scale NO_x recycle tests will be done using Babcock & Wilcox's 6 million Btu/hr Small Boiler Simulator (SBS) shown in Figure 5. The water-cooled furnace is a scaled-down version of B&W's single-cyclone, front-wall fired boiler design. The cyclone has been in operation since February 1985. The SBS cyclone furnace simulates a large cyclone unit very well. A comparison between the SBS cyclone furnace and commercial units is given in Table 4.

The NO_x recycle tests will begin with three loads and three excess air levels to establish the baseline of the NO_x emission from the SBS furnace. NO will then be injected in multiples of the baseline NO_x production levels. The NO concentration at the air inlet duct to the cyclone will be measured to document the inlet level. Stack NO_x will be measured to determine NO_x destruction occurring in the flame. The series of tests with different NO injection rates will also be performed at three furnace loads and three excess air levels. This test result will assist the determination of a second injection location for the next series of tests.

In the second series of tests, NO and NO_2 will be injected separately for two furnace loads and two excess air levels. Volumetric flowrate of the injected NO and NO_2 will be based on the proportion of these gasses that are present in the NOXSO sorbent heater off-gas. The addition of methane to the air stream to assist the NO_x destruction (7) will also be tested. The NO_x recycle test will be finalized by burning the coal from the Niles plant in the SBS furnace. Since the coal-ash slagging characteristics are important to the power plant operation, the use of Niles plant coal will assess the change of the coal ash's "flowability" in the Niles plant when the NO_x recycle stream is installed.

FUTURE WORK

On December 21, 1989, NOXSO Corporation, in association with MK-Ferguson Company, W.R. Grace & Co., and Ohio Edison, received an award from DOE's Clean Coal Technology Program to conduct a $66 million, full-scale commercial demonstration of the NOXSO technology. The U.S. DOE will provide $33 million and the remaining funds will be provided by the Ohio Coal Development Office, the Electric Power Research Institute, the Gas Research Institute, the East Ohio Gas Company, and the aforementioned NOXSO development team. The 115 MW demonstration plant will be installed at Ohio Edison's Niles Power Plant in northeastern Ohio. Construction is scheduled to begin in early 1993 with plant startup scheduled in May 1994. This project is the final step in the program to commercialize the NOXSO technology.

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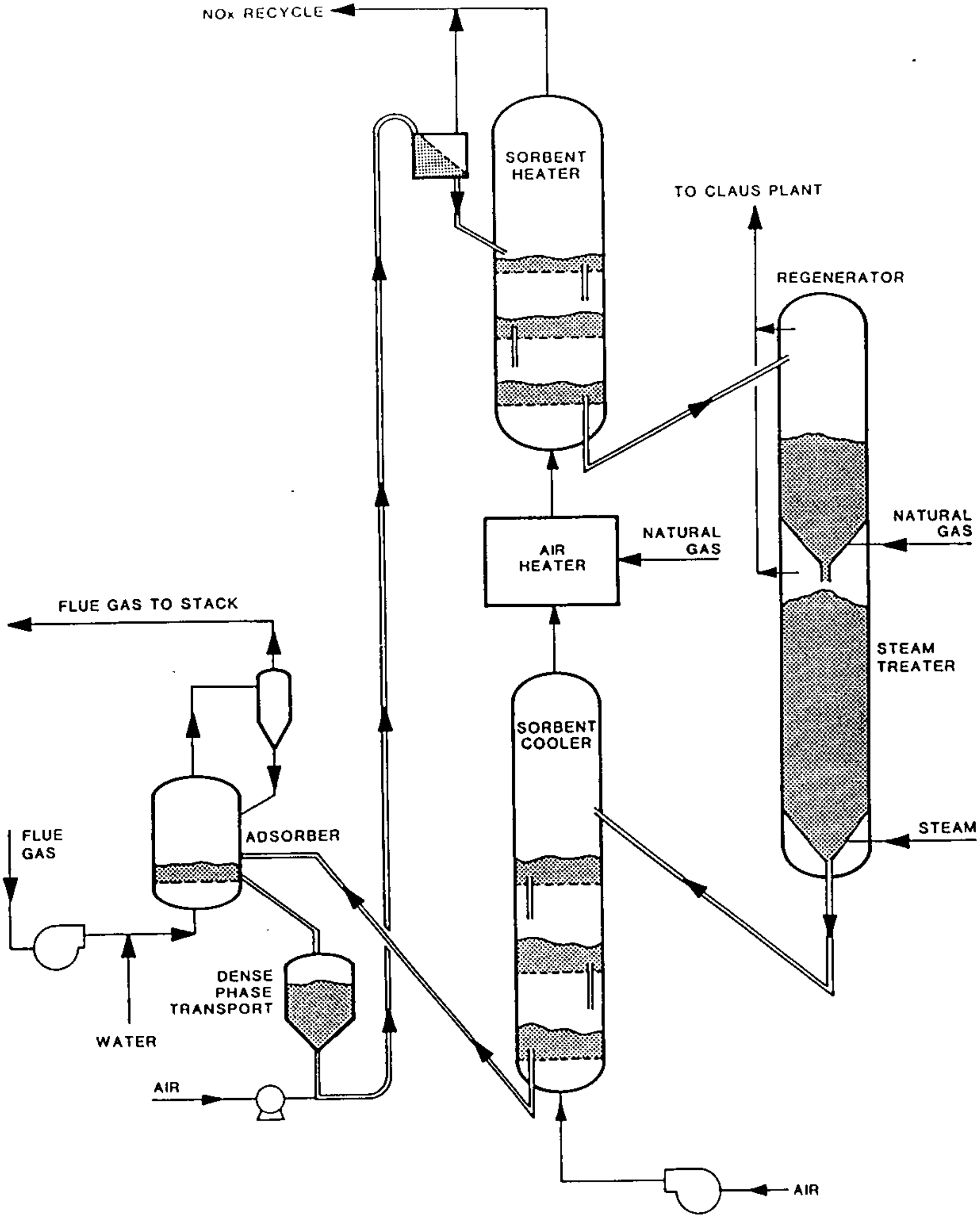

NOXSO PROCESS FLOW DIAGRAM

FIGURE 1

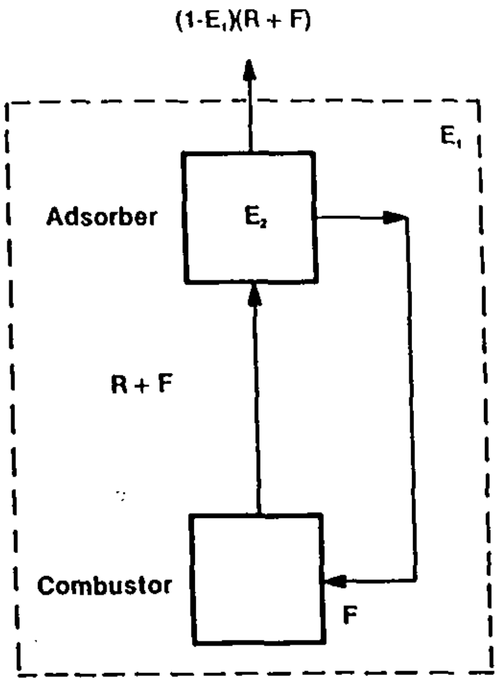

FIGURE 2. SCHEMATIC DIAGRAM OF NITROGEN OXIDE RECYCLE.

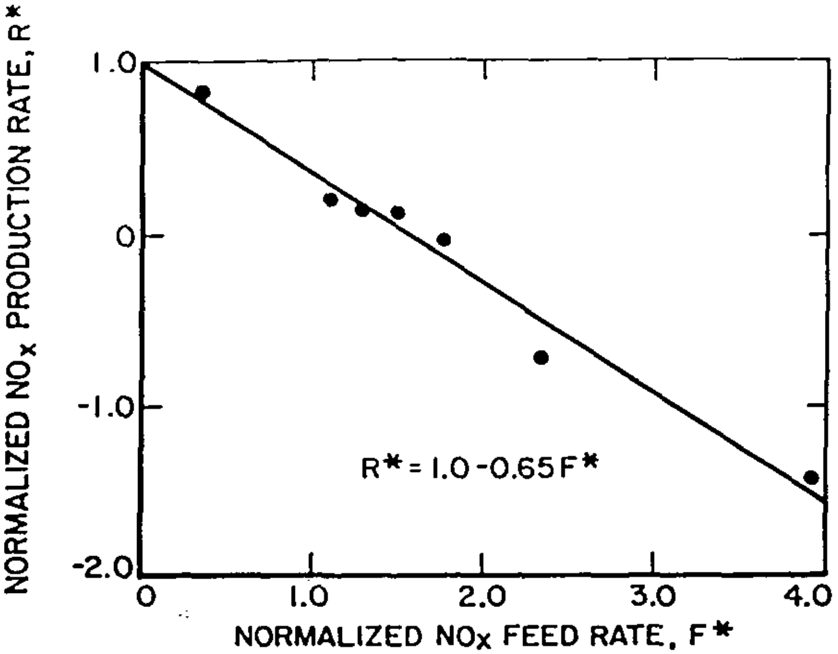

FIGURE 3. NORMALIZED NO$_x$ REDUCTION DATA-PC COMBUSTOR.

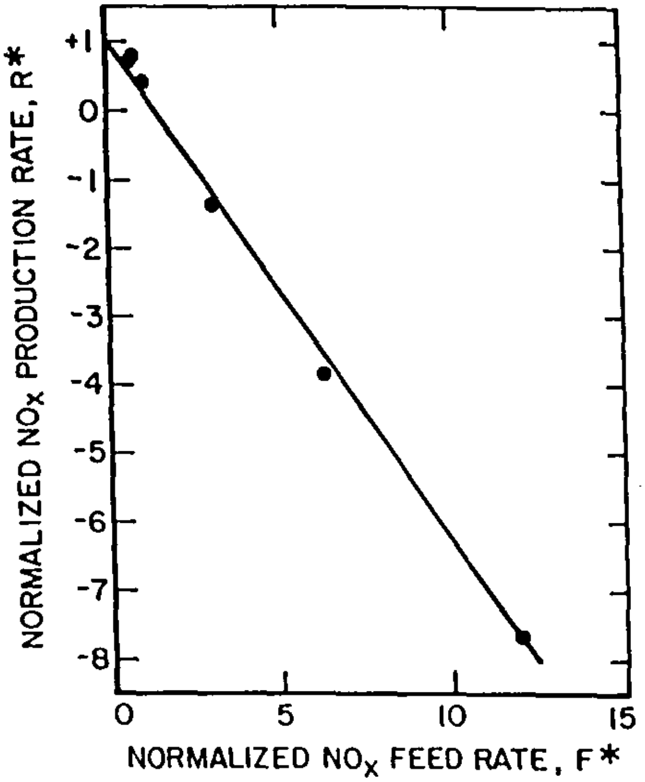

FIGURE 4. NORMALIZED NO$_x$ REDUCTION DATA-TUNNEL FURNACE.

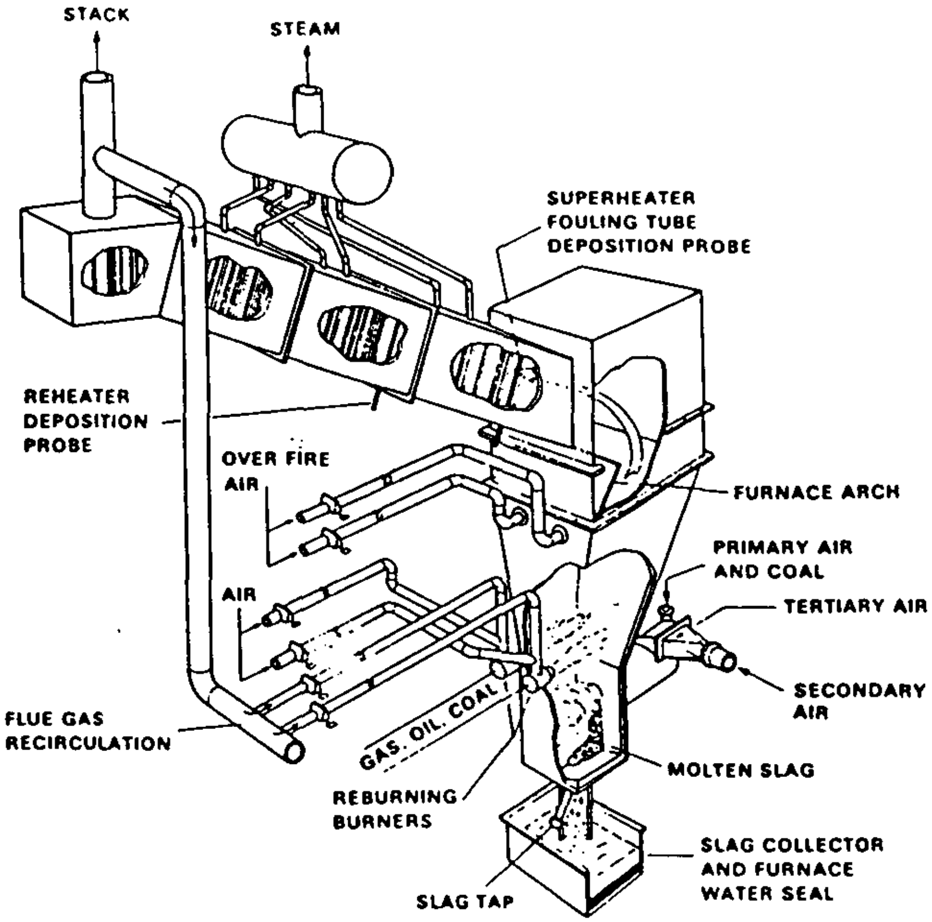

FIGURE 5. SMALL BOILER SIMULATOR (SBS) SCHEMATIC

Table 1. Comparison of NOXSO Test Programs

	Test Program	
Operating Parameter	TVA	3/4 MW
Coal Burned, lbs/hr	NA	500
Flue Gas Volume, SCFM	0.35	1200
Adsorber Type	Fixed Bed	Fluid Bed
SO_2 Inlet Concentration, ppm	2300	1465-5000
NO_x Inlet Concentration, ppm	600	470-720
SO_2 Removal Efficiency, %	90	90-99*
NO_x Removal Efficiency, %	90	80-92*
Reducing Gas for Regeneration	H_2S, H_2, CO	H_2, H_2+CO, CH_4
Operating Mode	Batch	Batch

	Test Program	
Operating Parameter	LCTU	POC
Coal Burned, lbs/hr	40	NA
Flue Gas Volume, SCFM	120	12000
Adsorber Type	Fluid Bed	Fluid Bed
SO_2 Inlet Concentration, ppm	1450-3000	2300
NO_x Inlet Concentration, ppm	240-800	350
SO_2 Removal Efficiency, %	60-90*	**
NO_x Removal Efficiency, %	60-90*	**
Reducing Gas for Regeneration	H_2, CH_4	Natural Gas
Operating Mode	Continuous	Continuous

NA = Not applicable, i.e., small slipstream was drawn from the power plant ductwork.

* = In the 3/4 MW and LCTU tests, removal efficiencies cover a wide range since operating conditions were intentionally varied to study their effect on process performance.

** = Pilot plant is under construction.

Table 2. POC Major Equipment Specifications*

Fluidized Bed Adsorber	
Diameter	10.5 ft
Temperature	120°C
Settled Bed Height	2 ft
Sorbent Residence Time	45 min
Superficial Gas Velocity	3 ft/s
Transport Disengaging Height	7.7 ft
Material of Construction	Carbon Steel
Fluidized Bed Sorbent Heater	
Number of Stages	3
Diameter	7.7 ft
Settled Bed Height	0.9 ft
Sorbent Residence Time	30 min
Superficial Gas Velocity	3 ft/s
Transport Disengaging Height	2.8 ft
Material of Construction	Type 304 SS
Fluidized Bed Sorbent Cooler	
Number of Stages	3
Diameter	5.7 ft
Settled Bed Height	1.2 ft
Sorbent Residence Time	20 min
Superficial Gas Velocity	3 ft/s
Transport Disengaging Height	4.3 ft
Material of Construction	Carbon Steel
Moving Bed Regenerator/Steam Treater	
Diameter	4 ft
Bed Height	10.3 ft/6.8 ft
Sorbent Residence Time	30 min/20 min
Material of Construction	Alonized Type 304H SS
Air Heater	
Design Flow (Air)	14,300 lbs/hr
Temperature Rise	330°C
Type	Natural gas fired in duct burners
Pneumatic Conveyor	
Sorbent Circulation Rate	9,994 lbs/hr
Lift Distance	80 ft
Adsorber Cyclone	
D-50	20 μm
D-100	100 μm
Gas Flowrate	16,257 ACFM @ 120°C

* At base case operating conditions.

Table 3. NO_x Reduction Data; 500 lb/hr Combustor (3)

Test No.#	NO_x Exit ppm	NO_x Exit lb/hr	F NO_x Fed lb/hr	R lb/hr	R*	F*
1	550	3.59	0	+3.59	1.0	0
2	1370	8.94	14.09	-5.15	-1.43	3.92
3	875	5.71	8.29	-2.58	-0.72	2.31
4	650	4.24	0	+4.24	1.0	0
5	850	5.55	4.66	+0.89	0.21	1.10
6	930	6.07	5.49	+0.58	0.14	1.29
7	700	4.56	0	+4.56	1.0	0
8	1100	7.17	6.64	+0.53	0.12	1.46
9	1200	7.82	7.98	-0.16	-0.04	1.75
10	820	5.34	1.60	+3.74	0.82	0.35

Tests 1 - 3. Coal feedrate = 223 lbs/hr, Flue gas flowrate = 122.1 moles/hr (dry), and Temperature = 2500°F.

Tests 4 - 6. Coal feedrate = 352 lbs/hr, Flue gas flowrate = 160.0 moles/hr (dry), and Temperature = 2500°F.

Tests 7 - 10. Coal feedrate = 431 lbs/hr, Flue gas flowrate = 180.4 moles/hr (dry), and Temperature = 2500°F.

Table 4. Comparison of Baseline Conditions Between the SBS Facility and Commercial Units

	SBS	Typical Cyclone-Fired Boilers
Cyclone Temperature	>3000°F	>3000°F
Residence Time at full load	1.4 sec	0.7-2.0 sec
Furnace Exit Gas Temperature	2265°F	2200°-2350°F
NOx Level	900-1200 ppm	600-1400 ppm
Ash Retention	80%-85%	60%-80%
Unburned Carbon	<1% in Ash	1%-20%
Ash Particle Size (MMD; Bahco)	6-8 microns	6-11 microns

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Current Status of ADVACATE Process for Flue Gas Desulfurization (Journal Article)

Acurex Corp., Research Triangle Park, NC

Prepared for:

Environmental Protection Agency, Research Triangle Park, NC

1992

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16. ABSTRACT The paper discusses current bench- and pilot-plant advances in preparation of ADVAnced siliCATE (ADVACATE) calcium silicate sorbents for flue gas desulfurization. It also discusses current bench- and pilot-plant advances in sorbent preparation. Fly ash was ground in a laboratory-scale grinder prior to slurrying in order to decrease the slurrying time needed for the sorbent to be reactive with SO_2. Reactivity of ADVACATE sorbents with SO_2 in the bench-scale reactor correlated with their surface area. ADVACATE sorbents produced with ground fly ash were evaluated in the 1.42 cu m/min pilot plant, providing 2 s duct residence time. ADVACATE sorbent was produced by slurrying ground fly ash (4.3 micrometer median particle size) with $Ca(OH)_2$ at a weight ratio of 3:1 at 90 C for 3 h to yield solids with 30 wt % of initial free moisture. When this sorbent was injected into the duct with 1500 ppm SO_2 and at 11 C approach to saturation, the measured SO_2 removal was about 60% at a Ca/S stoichiometric ratio of 2.0. Previously, when ADVACATE sorbent was produced at 90 C and at the same fly-ash-to-$Ca(OH)_2$ weight ratio using unground fly ash, removal under the same conditions in the duct was about 50% following 12 h slurrying. The report gives results of pilot-scale recycle tests at the recycle ratio of 2.0. Finally, the report discusses EPA's plans to commercialize ADVACATE.

17\. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Pollution	Pollution Control	13B
Flue Gases	Stationary Sources	21B
Desulfurization	ADVACATE Process	07A, 07D
Calcium Silicates		07B
Fly Ash		

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The following report discusses current bench- and pilot-plant advances in preparation of ADVAnced siliCATE (ADVACATE) calcium silicate sorbents for flue gas desulfurization. It also discusses current bench- and pilot-plant advances in sorbent preparation. Fly ash was ground in a laboratory scale grinder prior to slurrying in order to decrease the slurrying time needed for the sorbent to be reactive with SO_2. Reactivity of ADVACATE sorbents with SO_2 in the bench-scale reactor correlated with their surface area.

ADVACATE sorbents produced with ground fly ash were evaluated in the 50 cfm (85 m^3/h) pilot plant providing 2 s duct residence time. ADVACATE sorbent was produced by slurrying ground fly ash (median particle size of 4.3 μm) with $Ca(OH)_2$ at the weight ratio of 3:1 at 90°C (194°F) for 3 h to yield solids with 30 weight percent of initial free moisture. When this sorbent was injected into the duct with 1500 ppm SO_2 and at 11°C (20°F) approach to saturation, the measured SO_2 removal was approximately 60 percent at a Ca/S stoichiometric ratio of 2. Previously, when ADVACATE sorbent was produced at 90°C (194°F) and at the same fly-ash-to-$Ca(OH)_2$ weight ratio using unground fly ash, removal under the same conditions in the duct was approximately 50 percent following 12 h slurrying. The report presents the results of pilot-scale recycle tests at the recycle ratio of 2. Finally, the report discusses future U.S. Environmental Protection Agency plans for commercialization of ADVACATE.

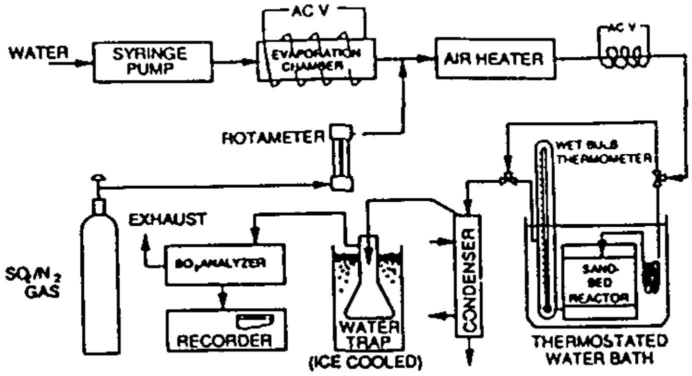

Figure 1. Schematic of sand-bed reactor bench-scale facilities.

Under U.S. Environmental Protection Agency (EPA) sponsorship, Jozewicz and Rochelle[1] studied the reaction of fly ash with calcium hydroxide [$Ca(OH)_2$] for the production of reactive sorbents for use in flue gas desulfurization (FGD) processes. They found that solids produced by slurrying fly ash and $Ca(OH)_2$ were significantly more reactive than $Ca(OH)_2$ alone. This work demonstrated that silica was the most reactive compound of fly ash and that reactivity increased with slurrying time/temperature. It also postulated that the rate limiting step of the reaction of fly ash with $Ca(OH)_2$ was dissolution of silica from fly ash. Three different approaches were tested that could increase the dissolution rate of silica: additives (sodium hydroxide, ammonium phosphate, phosphoric acid), more reactive form of silica (diatomaceous earth, bentonitic clays), and an increased slurrying temperature (pressure hydration).[2,3] The use of additives, or more reactive form of silica significantly increased the cost of sorbent preparation. High capital cost of pressure hydration and decreased conversion of $Ca(OH)_2$ above 150°C likely due to formation of crystalline calcium silicates indicated that the commercial competitiveness of calcium silicate sorbents could be increased by the elimination/modification of this hydration step. Another means to increase the rate of formation of calcium silicates is to increase the availability of silica by fly ash grinding. To test this approach, fly ash was ground in an attritor prior to slurrying with $Ca(OH)_2$ throughout the work described here.

Current bench- and pilot-plant research has been directed toward optimal sorbent preparation. Fly ash grinding before slurrying was investigated on the bench-scale to decrease the slurrying time needed for the sorbent to be reactive with sulfur dioxide (SO_2). Results indicate that the slurrying time can be decreased by a factor of 4, from 12 to 3 h, and will yield the same or better reactivity results than with unground fly ash. The reactivity of these sorbents with SO_2 can be correlated with their surface area, which is also a function of the particle size of the ground fly ash. High reactivities measured can be attributed to the large surface area of the calcium silicate as well as its water-retaining capability. Time and temperature studies were also conducted to determine the optimum temperature for slurrying the ground fly ash and $Ca(OH)_2$. Results suggest that 90°C (194°F) is optimum.

Pilot-plant work investigated using ground fly ash as the basis for calcium silicate formation. ADVAnced siliCATE (ADVACATE) sorbents produced using ground fly ash were evaluated in the 50 cfm (85 m^3/h) pilot plant. Fly ash with a median particle size of 16 μm was ground to a median particle size of 4.3 μm and slurried with $Ca(OH)_2$ for 3 h at a weight ratio of 3:1 and 90°C (194°F). The resultant slurry was dried in an oven to form free-flowing solids containing approximately 30 percent free moisture. This sorbent was injected into the duct with 1500 ppm SO_2 and at 11°C (20°F) approach to adiabatic saturation temperature (ΔT_s). The measured SO_2 removal was about 60 percent at a

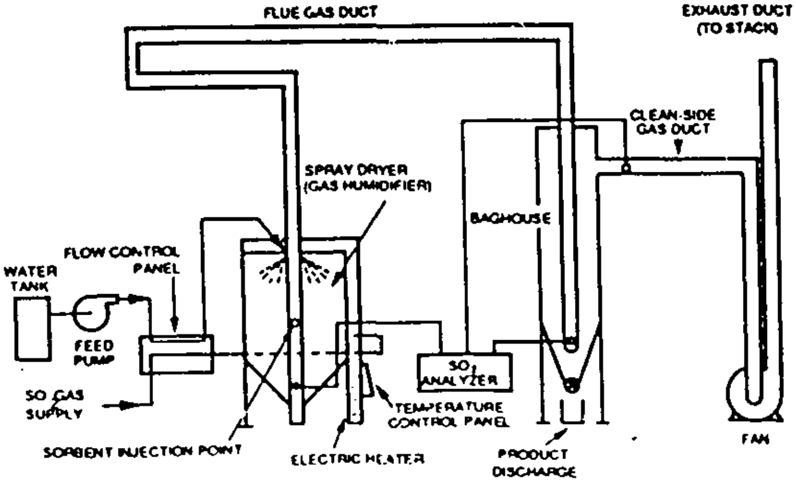

Figure 2. Schematic of duct injection pilot-plant facility.

stoichiometric ratio of 2.0. Previous tests[4] using ADVACATE sorbent prepared with unground fly ash, at the same conditions but following 12 h slurrying, yielded approximately 60 percent in-duct SO_2 removal. Pilot-scale recycle tests at a recycle ratio of 2.0 were also conducted using the 60 cfm pilot plant. The results of these tests, through the third pass of recycle, are shown below.

Future plans for developing the ADVACATE process include a 10 MW_e field demonstration at the Shawnee facility of Tennessee Valley Authority. A complete factorial test plan has been developed to evaluate in-duct SO_2 removal along with long-term, steady-state operation.

Experimental

Bench-Scale Facilities

Sorbent reactivity was evaluated in a bench-scale sand-bed reactor designed to simulate baghouse conditions. The reactor was a fixed bed normally using 1 g of sorbent dispersed in 40 g of sand. Each sample was exposed for 1 h to an SO_2-containing nitrogen stream controlled to the desired conditions of temperature and relative humidity.

Figure 3. SEM photomicrograph of "as received" Clinch River fly ash (CR) sample, ×2000.

The experimental apparatus, shown in Figure 1, has been discussed in more detail elsewhere.[5,6] Experiments were performed at a temperature of 64°C (147°F) and at ΔT_s of 10°C (18°F) (corresponding to a relative humidity of 60 percent). Flue gas was simulated with 1000 ppm SO_2 in dry nitrogen fed at 280 L/h (10 cfh) at 20°C (68°F).

SO_2 removal was determined by integrating the 1-h concentration curve recorded by a Thermoelectron Pulsed Fluorescent SO_2 Analyzer. Conversion has been defined as the moles of SO_2 removed from the flue gas divided by the number of moles of $Ca(OH)_2$ in the sorbent (0.0034 moles/g at an ash-to-$Ca(OH)_2$ ratio of 3:1).

Pilot Plant

The pilot-plant facility, shown in Figure 2, consists of humidification to control the gas temperature and humidity. The normal gas flow for the facility is 50 acfm at 177°C (350°F), and SO_2 is injected at a rate of 0.4–1.0 lb/h (0.18–0.45 kg/h) to provide an SO_2 concentration of 1000–2500 ppm. Ambient air is drawn through an air filter, heated to 177°C (350°F), and then humidified by evaporative cooling to 54°C (130°F) in the spray dryer vessel and in the duct. This corresponds to a ΔT_s of 11°C (20°F) through the system. The cooled gas stream then flows through an 80-ft (24.5-m) heat-traced and insulated section of duct to a pulse-jet baghouse for collection of solids. The cleaned gas is exhausted into the atmosphere. The calcium silicate sorbents are injected by a screw feeder into the duct downstream of the humidifier. The 5 cm (2 in.) diameter, 24.5 m (80 ft) long duct provides for a flue gas residence time of approximately 2 s. The collected solids are then discharged from the baghouse hopper and stored for later use. A detailed description of the pilot plant is provided elsewhere.[4].

The SO_2 concentrations throughout the system were monitored using a Dupont 411 Photometric Analyzer. Sampling points are at the spray dryer outlet (system inlet), baghouse inlet (duct), and baghouse outlet (total system).

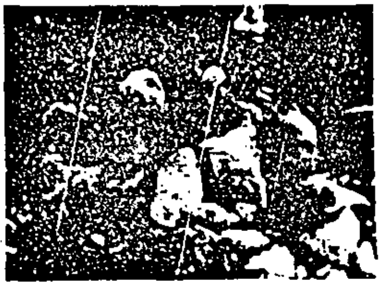

Figure 4. SEM photomicrograph of ground Clinch River fly ash (UPCR) sample, ×5000.

The SO_2 concentration was set at the system inlet and held constant by a calibrated rotameter. To ensure no dilution or air leakage, the concentration was checked at the baghouse inlet and outlet before the test. Data were collected every 0.5 h during a 4–8 h test run at fairly constant conditions.

Sorbents

Clinch River Ash. The fly ash from the Appalachian Power Company's Clinch River Plant in Carbo, Virginia, has been used throughout this work. The Clinch River fly ash (CR) was produced by burning eastern bituminous coal (fly ash CaO content of 5.4 percent). Fly ash morphology was characterized using a Scanning Electron Microscope (SEM). The view of a typical microscope field is presented in Figure 3, where spherical particles with smooth surfaces can be seen. Fly ash density was measured to be 2.1 g/cm^3. Mass median particle size was found to be 16.5 μm by sedimentation method (Microtrak). The nitrogen adsorption/desorption Brunauer, Emmett, and Teller (BET) surface area was measured to be 1.8 m^2/g, and pore volume was 0.00366 cm^3/g.

$Ca(OH)_2$. Reagent grade $Ca(OH)_2$ from Fisher (Fisher C-97) was used throughout bench-scale experiments. It was determined by a thermogravimetric analyzer (TGA) to be 93 percent $Ca(OH)_2$ (the remaining 7 percent was surface water, calcium carbonate, and impurities). Structural properties of a reagent grade $Ca(OH)_2$ were measured to be: 15 m^2/g surface area, 0.100 cm^3/g pore volume, and 5.6 μm mass median particle size.

Commercial Mississippi $Ca(OH)_2$ was used throughout the pilot-scale tests. It was determined by TGA to be 94 percent $Ca(OH)_2$ and structural properties were: 19 m^2/g surface area, 0.100 cm^3/g pore volume, and 4.0 μm median particle size.

Bench Scale. Samples of CR were ground in a Union Process laboratory-scale 1-S Batch Attritor. The attritor is a vertically oriented mill with a stationary stainless-steel tank loaded with 0.31-cm (0.125-in.) stainless-steel balls agitated by a rotating internal shaft with arms turning at 300 rpm. Fly ash was ground as a 30 weight percent slurry at room temperature for 1 h, then dried before further use. Fly ash morphology was characterized using SEM. A typical microscope field is presented in Figure 4. Most of the

Figure 5. SEM photomicrograph of sorbent made from ground Clinch River fly ash (BCR) hydrated with $Ca(OH)_2$, at a 3:1 weight ratio, for 3 h at 90°C (194°F), ×5000.

spherical particles (as shown in Figure 3) are broken and of irregular shape. Mass median particle size by Microtrak method of CR fly ash ground by Union Process (UPCR) was found to be 2.93 μm. The BET surface area and porosity have both increased to 5.4 m^2/g and 0.017246 cm^3/g, respectively.

Samples of CR were also ground in a laboratory ball mill for times ranging from 8 to 56 h. The mill consists of a 23.5 cm (10 in.) dia. horizontal steel cylinder with 1.23 cm (0.5-in.) stainless-steel balls rotating at 50 rpm. CR was loaded dry to half fill the cylinder. Grinding was continuous with pauses only to take periodic grab samples. At the end of 56 h of ball mill grinding, BET surface area of the fly ash ground in the ball mill (BCR) had increased to 3.4 m^2/g.

Sorbents were prepared from the above fly ash in a two-step process: hydration and drying. Hydration was performed atmospherically in a beaker placed in a hot water bath controlled at 60°C (140°F) to boiling. Reagents were added to hot water (1 part solid:15 parts water) and stirred for the entire hydration time, varied from 1 to 6 h. After hydration, the slurry was vacuum-filtered, and the filter cake was microwaved for 6 min to prevent any further hydration of damp sorbents.

Typical sorbents were made with 3 parts fly ash and 1 part $Ca(OH)_2$ (fly-ash-to-$Ca(OH)_2$ weight ratio of 3:1). Several sorbents were characterized by SEM. A typical, well-developed sorbent is presented in Figure 5. The pictured sorbent's BET surface area has increased to 26.3 m^2/g from an original fly ash surface area of 3.4 m^2/g. The surface area is very irregular with few planes visible.

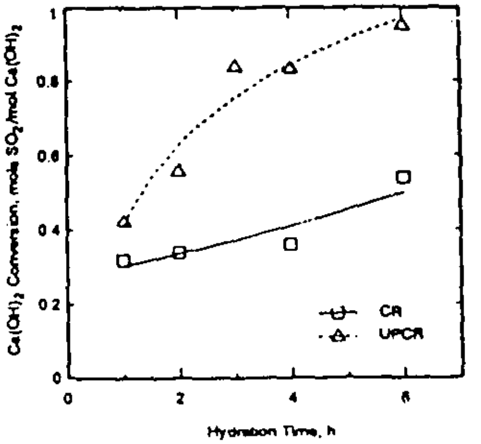

Figure 6. Conversion as a function of hydration time at 90°C (194°F), fly ash to $Ca(OH)_2$ weight ratio of 3:1 (conditions in the sand-bed reactor: 60 percent relative humidity, 1000 ppm SO_2 concentration)

Pilot Scale. Baseline tests were performed with unground CR that had been atmospherically hydrated with $Ca(OH)_2$ at 90°C (194°F) and at a fly-ash-to-$Ca(OH)_2$ ratio of 3:1 for 12 h. This sorbent was termed unground CR ADVACATE (UCA). The resultant slurry, approximately 70 percent water, was then dried in an oven at 65–93°C (150–200°F) until the bulk residual moisture of the solids was about 30 percent. Other tests involved using ground CR ash that had been wet ground in a Union Process Model 15-S Attritor. The fly ash was ground in a 15-S Attritor from an initial mass median particle size of 16.5 μm to a final mass median particle size of 4.3 μm. The ground ash was also hydrated with $Ca(OH)_2$ at a weight ratio of 3:1, at 90°C (194°C), but only for 3 h. This sorbent was termed ground CR ADVACATE (GCA). The resultant slurry, for the first pass, was also dried in the oven to approximately 30 percent residual moisture. All other passes of ground material, namely recycle tests, were prepared by backmixing the dry spent silicates collected in the baghouse with the fresh ADVACATE slurry. All solids were then passed through a Vibro-Energy separator to break down all agglomerates to suitable size for feeding. The screen that the solids passed through was 18 mesh, which resulted in moist solids with a free-flowing consistency. The fresh ADVACATE slurry consisted of 2 parts recycle (spent silicates) to 1 part fresh $Ca(OH)_2$, corresponding to a recycle ratio (RR) of 2.0. It also contained 1.5 parts fresh ground fly ash to 1 part fresh $Ca(OH)_2$, which simulated the inherent ash in the system from the boiler. All recycle work was performed at a bulk solid moisture content of 35 percent.

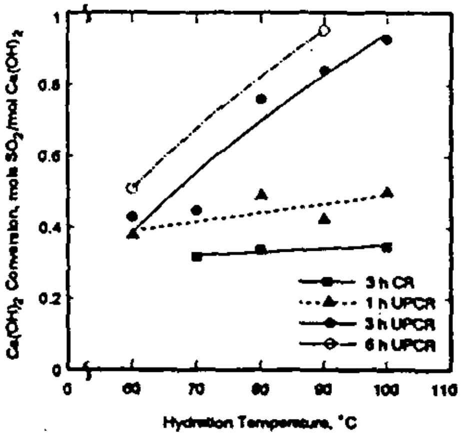

Figure 7. The effect of hydration temperature and time on the conversion of $Ca(OH)_2$. (conditions in the sand-bed reactor: 60 percent relative humidity, 1000 ppm SO_2 concentration).

Results and Discussion

Bench Scale

Sorbent made from unground CR hydrated at 90°C (194°F) was used as a baseline for bench-scale sorbent reactivity. No significant improvement in conversion was seen by increasing hydration time from 1 to 4 h; however, an increase in conversion was observed from 36 percent at 4 h to 54 percent at a 6 h hydration (see Figure 6). Because hydration time is directly related to hydrator size, which affects capital cost for commercial application, hydration

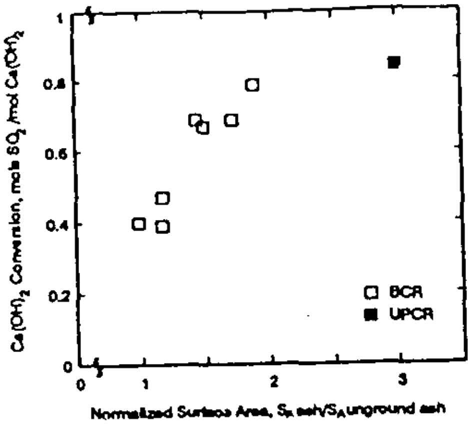

Figure 8. Effect of fly ash initial surface area on conversion of sorbents hydrated at 90°C (194°F) for 3 h at a fly-ash-to-Ca(OH)$_2$ weight ratio of 3:1 (conditions in the sand-bed reactor: 60 percent relative humidity, 1000 ppm SO$_2$ concentration).

time must be minimized. In order to test the potential for decreased hydration time, sorbent was made from UPCR and Ca(OH)$_2$ at the weight ratio of 3:1 by hydrating at 90°C (194°F) for periods ranging from 1 to 6 h. All UPCR sorbents tested had significantly higher reactivity than similar sorbents made from CR, when exposed to SO$_2$ in the sand-bed reactor operated at a ΔT_s of 10°C (18°F). Shorter hydration time was needed to obtain a given conversion with UPCR than with CR sorbents. For example, only 2 h was needed to give about 50 percent conversion for UPCR sorbent, whereas 6 h was needed for similar conversion with CR sorbent. UPCR sorbent conversion increased with increased hydration time. Conversion increased significantly from 1 to 3 h hydration, but little improvement occurred from 3 to 6 h hydration.

Increased temperature of hydration increases the cost of sorbent. To test the possibility of using lower temperatures, sorbents were prepared by hydrating from 60 to 100°C (140 to 212°F). Figure 7 shows reactivity of these sorbents prepared from CR and UPCR. Reactivity of CR sorbents remained at a conversion of about 0.33 mol SO$_2$/mol Ca(OH)$_2$ over the hydration temperature range of 70 to 100°C for 3 h hydrations. Sorbents prepared from UPCR decreased in reactivity as hydration temperature decreased from 100 to 60°C. There seemed to be a step change in conversion for UPCR sorbents hydrated 3 h from about 0.42 mol SO$_2$/mol Ca(OH)$_2$ below 70°C to 0.80 mol SO$_2$/mol Ca(OH)$_2$ or better above 80°C. Reactivity for UPCR sorbent prepared at 60°C increased with hydration time, but not as dramatically as UPCR sorbent prepared at higher temperatures.

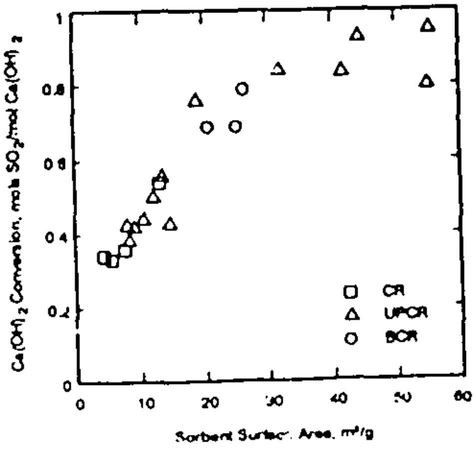

Figure 9. Effect of sorbent surface area on conversion for sorbents prepared at a fly ash to Ca(OH)$_2$ weight ratio of 3:1 (conditions in the sand bed reactor: 60 percent relative humidity, 1000 ppm SO$_2$ concentration).

As shown in Figures 6 and 7, grinding fly ash before hydration with Ca(OH)$_2$ to produce ADVACATE sorbents yielded substantial improvement in conversion over similarly prepared sorbents using unground fly ash for conditions tested. However, the amount of grinding needed will influence mill size and power requirements. To test the effect of the amount of fly ash grinding on the sorbents' reactivity, sorbents were made with CR ground in a ball mill (BCR) from 8 to 56 h. The ratio of BCR surface area to CR surface area (normalized surface area) was used to indicate the amount of grinding performed. BCR sorbents were produced at the BCR-to-Ca(OH)$_2$ weight ratio of 3:1 and at 90°C (194°C) for 3 h and were subsequently tested on the sand-bed reactor. Figure 8 shows the effect of grinding on sorbent reactivity. Conversion increased steadily as the normalized surface area of BCR increased from 1 to 2. The data point at a normalized surface area of 3 reflects UPCR sorbent.

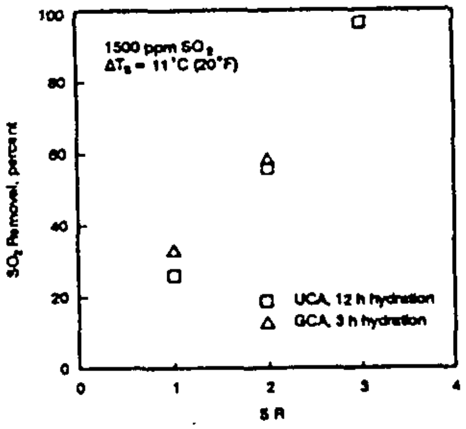

Figure 10. Comparison of CR ADVACATE (UCA and GCA).

For the same hydration conditions, surface area of sorbents increased as the fly ash surface area increased due to grinding. In Figure 9, sorbent surface area is correlated with conversion for sorbents hydrated under different conditions using CR, UPCR, BCR. Sorbents were hydrated from 1 to 6 h at temperatures from 60 to 90°C. Conversion in the sand-bed reactor increased as sorbent surface area increased.

Pilot Scale

Verification of Grinding Effect. Pilot-plant tests were conducted with UCA to provide a baseline for reactivity with SO$_2$. These results were compared with data from injection of the moist GCA. Each sorbent was injected at approximately 30 to 35 percent residual moisture and at Ca/S stoichiometric ratios (SR) of 1, 2, and 3. All tests were conducted at a humidifier inlet temperature of 350°F at a ΔT_s of 11°C (20°F) throughout the system. The inlet concentration of SO$_2$ was 1500 ppm, and only the baghouse inlet data were reported to simulate duct SO$_2$ removal in a field operation.

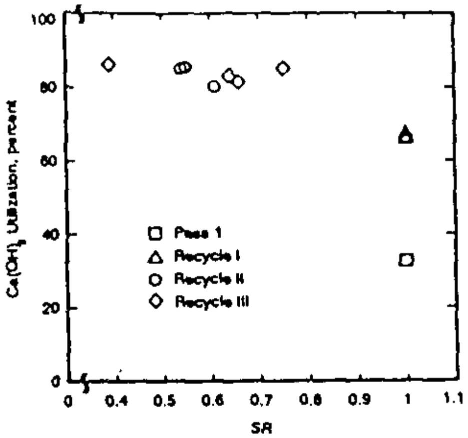

Figure 11. Calculated utilization of $Ca(OH)_2$ in ground CR (GCA) sorbent.

The UCA and GCA results are shown in Figure 10. Previous data showed that $Ca(OH)_2$ alone yielded approximately 28 percent in-duct SO_2 removal at a SR of 2.0 and a concentration of 2000 ppm. At 1500 ppm SO_2 and a SR of 2.0, the unground calcium silicate solids yielded approximately 56 percent in-duct removal. This demonstrates the much higher reactivity associated with the silicate-based solids. Even at a SR of 1.0, the UCA yielded 26 percent removal. The effective calcium utilization in Figure 10 increases gradually with increasing stoichiometric ratio. At a SR of 1.0, the UCA indicated an in-duct utilization of 26 percent; whereas, at SRs of 2.0 and 3.0, utilizations of 28 and 32.2 percent, respectively, were noted. One possible explanation for this slight increase in utilization is the increase in humidity of the gas stream owing to the increased grain loadings. Normally, an increase in humidity would correspond to a decrease in gas temperature but, because of the duct heat trace, no change in temperature was noted. As indicated above, all sorbents were fed at 30 to 35 percent residual moisture, but analysis of the baghouse catch indicated that all material had dried to approximately 5 to 10 percent. This suggests that, even at a SR of 3.0, the sorbent was releasing all possible moisture.

Previous data reported[4] indicated that in-duct SO_2 removals of 20 to 35 percent were achieved when unground oven-dried sorbent was injected at a SR range of 1.0 to 3.0 for a ΔT_s of about 11°C (20°F). Current data, shown in Figure 10, indicate SO_2 removals of 25 to 95 percent for comparable conditions. The lower removals associated with previous data can be mostly attributed to the initial inherent solids moisture content. Solids used in previous tests[4] were oven dried to a "crisp" product that could be easily crushed into a fine, dry powder suitable for dry injection. These solids normally possessed 5 percent residual moisture before injection. The UCA solids recently tested contained approximately 35 percent moisture. Because the solids were identically prepared, the large difference in SO_2 removal can be attributed to the additional inherent moisture contained by the UCA. This inherent moisture provides the media for the reaction between the sorbent particle and SO_2; whereas, the low-moisture sorbent must rely on the moisture in the gas stream to provide a basis for reaction to occur.

When comparing UCA and GCA, it is apparent that only a marginal increase in reactivity is noted for the ground solids. Although the increase in reactivity is very small, the important underlying factor is that, by grinding the fly ash prior to hydration with $Ca(OH)_2$, its hydration time could be reduced from 12 h to an optimum of 3 h. This confirms the bench-scale results discussed earlier (Figure 6).

Recycle Tests. The sorbents used in the above dry-injection tests were prepared by oven drying the ADVACATE slurry to a desired moisture content. However, on a commercial scale, a continuous operation, such as recycle mixing, would be a more practical and efficient way to produce a manageable, free-flowing product for feeding into a duct. Spent material from the first pass of GCA testing was used to prepare fresh solids for subsequent recycle tests.

One part $Ca(OH)_2$, 2 parts recycle product (RR = 2.0), and 1.5 parts ground CR were hydrated at 90°C (194°F) for 3 h. The resultant slurry was then mixed with sufficient recycle product to obtain a moist, free-flowing ADVACATE product of approximately 30 percent residual moisture. The mixture appeared sufficiently dry to be handled by a dry-solids feeder. Injection of these solids resulted in an immediate lowering of the flue gas temperature. For this reason, tests were conducted with the temperatures initially set sufficiently high to allow for the decrease in bulk gas temperature to the desired ΔT_s of 11°C (20°F). Moist solids were injected into the simulated flue gas at approximately 93°C (200°F), and they consequently lowered the duct temperature to 54°C (130°F).

Three consecutive recycle passes were tested in the pilot

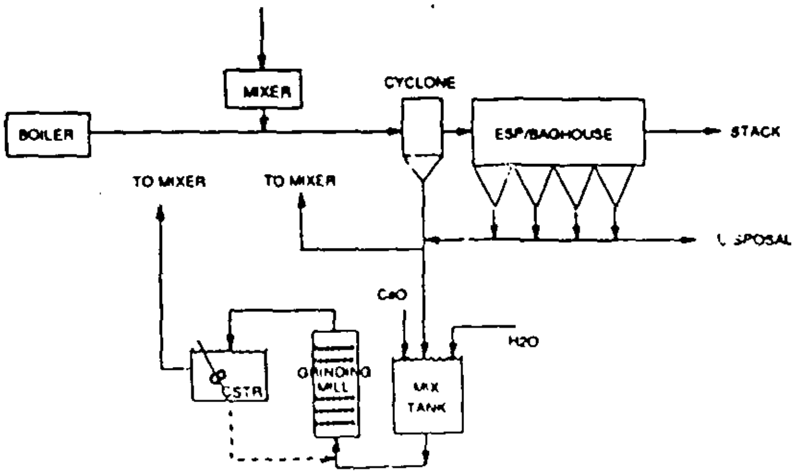

Figure 12. Commercial scheme of the ADVACATE process

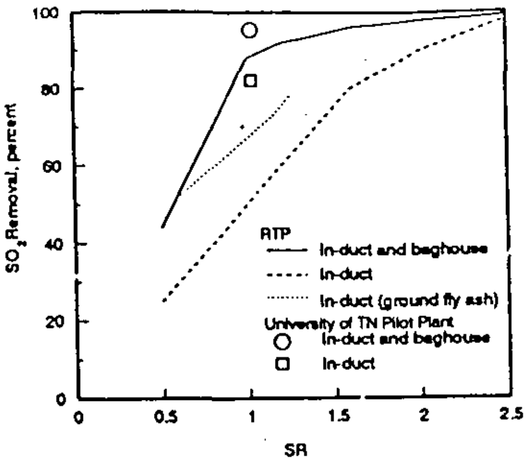

Figure 13. The effect of SR on SO_2 removal during calcium silicate damp injection at near equilibrium conditions.

plant to demonstrate the effect recycle has on overall sorbent reactivity and utilization. The results of these three passes, in addition to the baseline pass, are shown in Figure 11. Results indicate that the first pass of recycle (Recycle I) yielded 66 percent utilization at a SR of 1.0 which is better than previous data reported for unground fly ash.[4] One possible explanation is that using unground ash as the basis for ADVACATE sorbents provides much less overall surface area per gram of sorbent for calcium silicate formation than if ground ash is used. The recycle solids that were used in the current testing emanated from ground fly ash, which provided higher initial surface area for calcium silicate formation.

Figure 11 also shows the overall results of the second and third passes of recycle testing (Recycle II and III, respectively). These passes were tested at a SR of less than 1 owing to the increased grain loadings and the consequent difficulty in maintaining stable conditions. This difficulty in maintaining stable conditions can be attributed greatly to small (5 cm or 2 in. I.D.) duct size used with an 85 m^3/h (50 ft^3/min) pilot system. This thermal and mass transfer instability region has not been encountered in larger, and therefore more turbulent flow, systems under comparable duct conditions, and should not prove a problem in commercial applications. Several tests were conducted for each pass to ensure consistent operation of the pilot plant. Recycle II material was injected at a moisture content of 35 percent and yielded 80 percent utilization at a SR of 0.61. Recycle III material, at 38 percent moisture, was injected at a SR of 0.75 and yielded 85 percent utilization. When compared with the first pass and Recycle I in Figure 11, these two data points suggest that equilibrium, or steady-state operation, is being approached owing to the diminishing increase in utilization. The above results demonstrated that utilization of $Ca(OH)_2$ increased using recycle solids as a basis and indicate that SO_2 removals of up to 90 percent in-duct are possible at a SR of 1.2 and complete equilibrium.

Commercial Activities[7]

Figure 12 represents the commercial scheme of the ADVACATE process as envisioned for the 10 MW_e pilot evaluation in 1991-1992. Key process features include:

- In-duct sorbent drying and SO_2 absorption
- One-step sorbent mixing/injection
- Sorbent preparation with two slurry tanks and one mill (attritor)

As envisioned, spent sorbent and boiler fly ash are collected in the electrostatic precipitator (ESP) or baghouse, which has been modified mechanically for higher loadings. A portion of the waste material is fed to a mix tank along with fresh CaO, a larger portion is diverted directly to the mixer, and the remainder discarded as waste. From the mix tank, CaO/fly ash slurry is pumped to a constant stir tank reactor (CSTR), where the slurry is constantly recirculated through a mill. The combination of particle abrasion and heated water promotes the lime/silica reaction to form a calcium silicate phase on fly ash surfaces. The product silicate slurry is then routed to the mixer and blended with dry recycle solids to form a damp, free-flowing powder that is then injected into the duct at least 0.75 sec upstream of the dust collector. Duct velocities greater than 15 m/sec are necessary to ensure good gas/sorbent contact and to keep duct walls swept clean of agglomerates. Although no studies have been made of silicate waste disposal, the waste is physically and chemically similar to spray dryer FGD waste except that it is more hygroscopic and pozzolanic. Results of one study show capital costs of \$60–85/kW and operating costs of 6 to 7 mills/kWh, assuming 2.7 percent sulfur Midwestern coal and a SR of 1.2.[7]

Optimism that 90 percent SO_2 at a SR of 1.2 or less will be achieved on a commercial scale is based on in-house pilot and two previous field evaluations. EPA's in-house pilot (85 m^3/h) plant has consistently shown 80 percent utilization with mill-ground sorbent (GCA) (Figure 11). The first field evaluation on a 1700 m^3/h slipstream at the University of Tennessee pilot plant showed 80+ and 90+ percent calcium utilization in-duct and in-duct/baghouse, respectively, using unground sorbent, when ADVACATE was integrated with the Moist Dust Injection process (Figure 13).[7] When the Limestone Injection Multistage Burner (LIMB) process was augmented by ADVACATE injection on a 3400 m^3/h slipstream with ESP control, up to 97 percent additional SO_2 removal was realized.[6] Field tests on a 10 MW_e slipstream pilot plant at TVA's Shawnee facility are currently scheduled to begin in January 1992. From this effort, design and operating data for a full-scale system should be derived.

To support these efforts, Acurex Corporation and EPA/AEERL will continue research in the following process areas with federal funds and funding from appropriate vendors:

- Optimization of particle attrition
- Optimization of slurry/solids mixing
- Integrated small pilot evaluation
- ESP performance
- Evaluation of fly ashes and alternative calcium sources
- Effects of additives and impurities

Concurrent with the above efforts, Acurex and EPA will be developing non-fly-ash ADVACATE technology for possible use on non-coal-fired combustion sources of acid gases, including waste combustors, smelters, and industrial boilers.

Conclusions

1. Hydration time and temperature have significant effects on sorbent reactivity for sorbents prepared from ground fly ash and lime. Sorbent reactivity increases with increasing hydration time and temperature. Sorbents prepared from ground fly ash were always more reactive than sorbents made from unground fly ash at the same hydration conditions.
2. For sorbents hydrated at 90 °C (194 °F) for 3 h, reactivity tends to increase as the amount of fly ash grinding increases prior to sorbent preparation.

3. Reactivity tends to increase as sorbent surface area increases for sorbents hydrated from 1 to 6 h at 60 to 100°C (140–212°F).
4. Pilot-plant results verify that initially grinding the fly ash before hydration with $Ca(OH)_2$ decreases the optimal hydration time from 12 to 3 h.
5. The inherent moisture content of the ADVACATE sorbent greatly affects its overall reactivity toward SO_2.
6. Initially using ground rather than unground fly ash provides for a more reactive sorbent in subsequent passes, owing to the higher initial surface area associated with the ground ash.
7. Pilot-scale data suggest that possible equilibrium, or steady-state operation, can be achieved by the third or fourth pass of recycle material. Recycle III material demonstrated $Ca(OH)_2$ utilization of up to 86 percent.

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Current Status of ADVACATE Process
for
Flue Gas Desulfurization

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INTRODUCTION

Under U.S. Environmental Protection Agency (EPA) sponsorship, Jozewicz and Rochelle[1] studied the reaction of fly ash with calcium hydroxide [$Ca(OH)_2$] for the production of reactive sorbents for use in flue gas desulfurization (FGD) processes. They found that solids produced by slurrying fly ash and $Ca(OH)_2$ were significantly more reactive than $Ca(OH)_2$ alone. This work demonstrated that silica was the most reactive compound of fly ash and that reactivity increased with slurrying time/temperature. It also postulated that the rate limiting step of the reaction of fly ash with $Ca(OH)_2$ was dissolution of silica from fly ash. Three different approaches were tested that could increase the dissolution rate of silica: additives (sodium hydroxide, ammonium phosphate, phosphoric acid), more reactive form of silica (diatomaceous earth, bentonitic clays), and an increased slurrying temperature (pressure hydration).[2,3] The use of additives, or more reactive form of silica significantly increased the cost of sorbent preparation. High capital cost of pressure hydration and decreased conversion of $Ca(OH)_2$ above 150 °C likely due to formation of crystalline calcium silicates indicated that the commercial competitiveness of calcium silicate sorbents could be increased by the elimination/modification of this hydration step. Another means to increase the rate of formation of calcium silicates is to increase the availability of silica by fly ash grinding. To test this approach, fly ash was ground in a bench-scale attritor prior to slurrying with $Ca(OH)_2$ throughout the work described here.

Current bench- and pilot-plant research has been directed toward optimal sorbent preparation. Fly ash grinding before slurrying was investigated on the bench scale to decrease the slurrying time needed for the sorbent to be reactive with sulfur dioxide (SO_2). Results indicate that the slurrying time can be decreased by a factor of 4, from 12 to 3 h, and will yield the same or better reactivity results than with unground fly ash. The reactivity of these sorbents with SO_2 can be correlated with their surface area, which is also a function of the particle size of the ground fly ash. High reactivities measured can be attributed to the large surface area of the calcium silicate as well as with its water-retaining capability. Time and temperature studies were also conducted to determine the optimum temperature for slurrying the ground fly ash and $Ca(OH)_2$. Results suggest that 90 °C (194 °F) is optimum.

Pilot-plant work investigated using ground fly ash as the basis for calcium silicate formation. ADVAnced siliCATE (ADVACATE) sorbents produced using ground fly ash were evaluated in the 50 cfm (85 m^3/h) pilot plant. Fly ash with a median particle size of 16 µm was ground to a median particle size of 4.3 µm and slurried with $Ca(OH)_2$ for 3 h at a weight ratio of 3:1 and 90 °C (194 °F). The resultant slurry was dried in the oven to form free-flowing solids containing approximately 30 percent free moisture. This sorbent was injected into the duct with 1500 ppm SO_2 and at 11 °C (20 °F) approach to adiabatic saturation temperature (ΔT_s). The measured SO_2 removal was about 60 percent at a stoichiometric ratio of 2.0. Previous tests[4] using ADVACATE sorbent prepared with unground fly ash, at the same conditions but following 12 h slurrying, yielded approximately 50 percent in-duct SO_2 removal. Pilot-scale recycle tests at a recycle ratio of 2.0 were also conducted using the 50 cfm pilot plant. The results of these tests, through the third pass of recycle, are shown below.

Future plans for developing the ADVACATE process include a 10 MW_e field demonstration at the Shawnee facility of Tennessee Valley Authority. A complete factorial test plan has been developed to evaluate in-duct SO_2 removal along with long-term, steady-state operation.

EXPERIMENTAL

Bench-Scale Facilities

Sorbent reactivity was evaluated in a bench-scale sand-bed reactor designed to simulate baghouse conditions. The reactor was a fixed bed normally using 1 g of sorbent dispersed in 40 g of sand. Each sample was exposed for 1 h to an SO_2-containing nitrogen stream controlled to the desired conditions of temperature and relative humidity. The experimental apparatus, shown in Figure 1, has been discussed in more detail elsewhere.[5,6] Experiments were performed at a temperature of 64 °C (147 °F) and at ΔT_s of 10 °C (18 °F) (corresponding to a relative humidity of 60 percent). Flue gas was simulated with 1000 ppm SO_2 in dry nitrogen fed at 280 L/h (10 cfh) at 20 °C (68 °F).

SO_2 removal was determined by integrating the 1-h concentration curve recorded by a Thermoelectron Pulsed Fluorescent SO_2 Analyzer. Conversion has been defined as the moles of SO_2 removed from the flue gas divided by the number of moles of $Ca(OH)_2$ in the sorbent (0.0034 moles/g at an ash-to-lime ratio of 3:1).

Pilot Plant

The pilot-plant facility, shown in Figure 2, consists of humidification to control the gas temperature and humidity. The normal gas flow for the facility is 50 acfm at 177 °C (350 °F), and SO_2 is injected at a rate of 0.4–1.0 lb/h (0.18-0.45 kg/h) to provide an SO_2 concentration of 1000–2500 ppm at 20 °C (68 °F). Ambient air is drawn through an air filter, heated to 177 °C (350 °F), and then humidified by evaporative cooling to 54 °C (130 °F) in the spray dryer vessel and in the duct. This corresponds to a ΔT_s of 11 °C (20 °F) through the system. The cooled gas stream then flows through an 80-ft (24.5-m) heat-traced and insulated section of duct to a pulse-jet baghouse for collection of solids. The cleaned gas is exhausted into the atmosphere. The calcium silicate sorbents are injected by a screw feeder into the duct downstream of the humidifier. The 5 cm (2 in.) diameter, 24.5 m (80 ft) long duct provides for a flue gas residence time of approximately 2 s. The collected solids are then discharged from the baghouse hopper and stored for later use. A detailed description of the pilot plant is provided elsewhere[4].

The SO_2 concentrations throughout the system were monitored using a Dupont 411 Photometric Analyzer. Sampling points are at the spray dryer outlet (system inlet), baghouse inlet (duct), and baghouse outlet (total system). The SO_2 concentration was set at the system inlet and held constant by a calibrated rotameter. To ensure no dilution or air leakage, the concentration was checked at the baghouse inlet and outlet before the test. Data were collected every 0.5 h during a 4–8 h test run at fairly constant conditions.

Sorbents

Clinch River Ash. The fly ash from Appalachian Power Company's Clinch River Plant in Carbo, VA, has been used throughout this work. The Clinch River fly ash (CR) was produced by burning eastern bituminous coal (CaO content of 5.4 percent). Fly ash morphology was characterized using a Scanning Electron Microscope (SEM). The view of a typical microscope field is presented in Figure 3, where spherical particles with smooth surfaces can be seen. Fly ash density was measured to be 2.1 g/cm^3. Median particle size was found to be 16.0 μm by sedimentation method (Microtrak). The nitrogen adsorption/desorption Brunauer, Emmett, and Teller (BET) surface area was measured to be 1.8 m^2/g, and pore volume was 0.00366 cm^3/g.

$Ca(OH)_2$. Reagent grade $Ca(OH)_2$ from Fisher (Fisher C-97) was used throughout bench-scale experiments. It was determined by a thermogravimetric analyzer (TGA) to be 93 percent $Ca(OH)_2$ (the remaining 7 percent was surface water, calcium carbonate, and impurities). Structural properties of a reagent grade $Ca(OH)_2$ were measured to be: 15 m^2/g surface area, 0.100 cm^3/g pore volume, and 5.6 μm median particle size.

Commercial Mississippi $Ca(OH)_2$ was used throughout the pilot-scale tests. It was determined by TGA to be 94 percent $Ca(OH)_2$ and structural properties were: 19 m^2/g surface area, 0.100 cm^3/g pore volume, and 4.0 μm median particle size.

Bench Scale. Samples of CR fly ash were ground in a Union Process laboratory-scale 1-S Batch Attritor. The attritor is a vertically oriented mill with a stationary stainless-steel tank loaded with 0.31-cm (0.125-in.) stainless-steel balls agitated by a rotating internal shaft with arms turning at 300 rpm. Fly ash was ground as a 30 weight percent slurry at room temperature for 1 h, then dried before further use. Fly ash morphology was characterized using SEM. A typical microscope field is presented in Figure 4. Most of the spherical particles (as shown in Figure 3) are broken and of irregular shape. Median particle size by Microtrak method of CR fly ash ground by Union Processs (UPCR) was found to be 2.93 µm. The BET surface area and porosity have both increased to 5.4 m^2/g and 0.017246 cm^3/g, respectively.

Samples of CR were also ground in a laboratory ball mill for times ranging from 8 to 56 h. The mill consists of a 23.5 cm (10 in.) dia. horizontal steel cylinder with 1.23 cm (0.5-in.) stainless-steel balls rotating at 50 rpm. CR was loaded dry to half fill the cylinder. Grinding was continuous with pauses only to take periodic grab samples. At the end of 56 h of ball mill grinding, BET surface area of the fly ash had increased to 3.4 m^2/g.

Sorbents were prepared from the above fly ash in a two-step process: hydration and drying. Hydration was performed atmospherically in a beaker placed in a hot water bath controlled at 60 °C to boiling. Reagents were added to hot water (1 part solid:15 parts water) and stirred for the entire hydration time, varied from 1 to 6 h. After hydration, the slurry was vacuum-filtered, and the filter cake was microwaved for 6 min to prevent any further hydration of damp sorbents.

Typical sorbents were made with 3 parts fly ash and 1 part $Ca(OH)_2$ (fly-ash-to-$Ca(OH)_2$ weight ratio of 3:1). Several sorbents were characterized by SEM. A typical, well-developed sorbent is presented in Figure 5. The pictured sorbent's BET surface area has increased to 26.3 m^2/g from an original fly ash surface area of 3.4 m^2/g. The surface area is very irregular with few planes visible.

Pilot Scale. Baseline tests were performed with unground CR that had been atmospherically hydrated with $Ca(OH)_2$ at 90 °C (194 °F) and at a fly-ash-to-$Ca(OH)_2$ ratio of 3:1 for 12 h. This sorbent was termed unground CR ADVACATE (UCA). The resultant slurry, approximately 70 percent water, was then dried in an oven at 65–93 °C (150-200 °F) until the bulk residual moisture of the solids was about 30 percent. Other tests involved using ground CR ash that had been wet ground in a Union Process Model 15-S Attritor. The fly ash was ground in a 15-S Attritor from an initial particle size of 16.0 µm to a final median particle size of 4.3 µm. The ground ash was also hydrated with $Ca(OH)_2$ at a weight ratio of 3:1, at 90 °C (194 °C), but only for 3 h. This sorbent was termed ground CR ADVACATE (GCA). The resultant slurry, for the first pass, was also dried in the oven to approximately 30 percent residual moisture. All other passes of ground material, namely recycle tests, were prepared by backmixing the dry spent silicates collected in the baghouse with the fresh ADVACATE slurry. All solids were then passed through a Vibro-Energy separator to break down all agglomerates to suitable size for feeding. The screen that the solids passed through was 18 mesh, which resulted in solids with a moist, free-flowing consistency. The fresh ADVACATE slurry consisted of 2 parts recycle (spent silicates) to 1 part fresh $Ca(OH)_2$, corresponding to a recycle ratio (RR) of 2.0. It also contained 1.5 parts fresh ground fly ash to 1 part fresh $Ca(OH)_2$, which simulated the inherent ash in the system from the boiler. All recycle work was performed at a bulk solid moisture content of 35 percent.

RESULTS AND DISCUSSION

Bench Scale

Sorbent made from unground CR hydrated at 90 °C (194 °F) was used as a baseline for bench-scale sorbent reactivity. No significant improvement in conversion was seen by increasing hydration time from 1 to 4 h; however, an increase in conversion was observed from 36 percent at 4 h to 54 percent at a 6-h hydration (see Figure 6). Because hydration time is directly related to hydrator size, which affects capital cost for commercial application, hydration time must be minimized. In order to test the potential for decreased hydration time, sorbent was made from UPCR and $Ca(OH)_2$ at the weight ratio of 3:1 by hydrating at 90 °C (194 °F) for periods ranging from 1 to 6 h. All UPCR sorbents tested had significantly higher reactivity than similar sorbents made from CR, when exposed to SO_2 in the sand-bed reactor operated at a ΔT_s of 10 °C (18 °F). Shorter hydration time was needed to obtain a given conversion with UPCR than with CR sorbents. For example, only 2 h was needed to give about 50 percent conversion for UPCR sorbent, whereas 6 h was needed for CR sorbent. UPCR sorbent conversion increased with increased hydration time. Conversion increased significantly from 1 to 3 h hydration, but little improvement occurred from 3 to 6 h hydration.

Increased temperature of hydration increases the cost of sorbent. To test the possibility of using lower temperatures, sorbents were prepared by hydrating from 60 to 100 °C. Figure 7 shows reactivity of these sorbents prepared from CR and UPCR hydrated for 3 h. Reactivity of CR sorbents remained at a conversion of about 0.33 mol SO_2/mol $Ca(OH)_2$ over the hydraion temperature range of 70 to 100 °C. Sorbents prepared from UPCR decreased in reactivity as hydration temperature decreased from 100 to 60 °C. There seemed to be a step change in conversion for UPCR sorbents hydrated 3 h from about 0.42 mol SO_2/mol $Ca(OH)_2$ below 70 °C to 0.80 mol SO_2/mol $Ca(OH)_2$ or better above 80 °C. Figure 8 further illustrates the effect of hydration temperature on UPCR in more detail. Reactivity for UPCR sorbent prepared at 60 °C increased with hydration time, but not as rapidly as UPCR sorbent prepared at 90 °C (194 °F).

As shown in Figures 6 and 7, grinding fly ash before hydration with $Ca(OH)_2$ to produce ADVACATE sorbents yielded substantial improvement in conversion over similarly prepared sorbents using unground fly ash for conditions tested. However, the amount of grinding needed will influence mill size and power requirements. To test the effect of the amount of fly ash grinding on the sorbents' reactivity, sorbents were made with CR ground in a ball mill (BCR) from 8 to 56 h. The ratio of BCR surface area to CR surface area (normalized surface area) was used to indicate the amount of grinding performed. BCR sorbents were produced at the BCR-to-$Ca(OH)_2$ weight ratio of 3:1 and at 90 °C (194 °C) for 3 h and were subsequently tested on the sand-bed reactor. Figure 9 shows the effect of grinding on sorbent reactivity. Conversion increased steadily as the normalized surface area of BCR increased from 1 to 2. The data point at a normalized surface area of 3 reflects UPCR sorbent.

Figure 10 shows the relationship between ground fly ash surface area and the surface area of the resultant sorbent at two distinct slurrying conditions. For both 3 and 6 h hydrations at 90 °C (194 °F), sorbent surface area increased with increasing ground fly ash surface area. Although the two data points at 5.5 m^2/g fly ash surface area correspond to UPCR, they are consistent with other data points, which correspond to BCR. Figure 10 demonstrates the importance of ground fly ash surface area for the development of surface area by the sorbent. In Figure 11, sorbent surface area is correlated with conversion for sorbents hydrated under different conditions using CR, UPCR, and moist ADVACATE solids produced in the pilot plant. Sorbents were hydrated from 1 to 6 h at temperatures from 60 to 90 °C. Conversion in the sand-bed reactor increased as sorbent surface area increased.

Pilot Scale

Verification of Grinding Effect. Pilot-plant tests were conducted with UCA to provide a baseline for reactivity with SO_2. These results were compared with data from injection of the moist GCA. Each sorbent was injected at approximately 30 to 35 percent residual moisture and at Ca/S stoichiometric ratios (SR) of 1, 2, and 3. All tests were conducted at a humidifier inlet temperature of 350 °F at a ΔT_s of 11 °C (20 °F) throughout the system. The inlet concentration of SO_2 was 1500 ppm, and only the baghouse inlet data were reported to simulate duct SO_2 removal in a field operation.

The UCA and GCA results are shown in Figure 12. Previous data showed that $Ca(OH)_2$ alone yielded approximately 28 percent in-duct SO_2 removal at a SR of 2.0 and a concentration of 2000 ppm. At 1500 ppm SO_2 and a SR of 2.0, the unground calcium silicate solids yielded approximately 56 percent in-duct removal. This demonstrates the much higher reactivity associated with the silicate-based solids. Even at a SR of 1.0, the UCA yielded 26 percent removal. It is apparent from Figure 12 that the effective calcium utilization increases gradually with increasing stoichiometric ratio. At a SR of 1.0, the UCA indicated an in-duct utilization of 26 percent; whereas, at SRs of 2.0 and 3.0, utilizations of 28 and 32.2 percent, respectively, were noted. One possible explanation for this slight increase in utilization is the increase in humidity of the gas stream owing to the increased grain loadings. Normally, an increase in humidity would correspond to a decrease in gas temperature but, because of the duct heat trace, no change in temperature was noted. As indicated above, all sorbents were fed at 30 to 35 percent residual moisture, but analysis of the baghouse catch indicated that all material had dried to approximately 5 to 10 percent. This suggests that, even at a SR of 3.0, the sorbent was releasing all possible moisture.

Previous data reported[4] indicated that in-duct SO_2 removals of 20 to 35 percent were achieved when unground oven-dried sorbent was injected at a SR range of 1.0 to 3.0 for a ΔT_s of about 11 °C (20 °F). Current data, shown in Figure 12, indicate SO_2 removals of 25 to 95 percent for comparable conditions. The lower removals associated with previous data can be mostly attributed to the initial inherent solids moisture content. Solids used in previous tests[4] were oven dried to a "crisp" product that could be easily crushed into a fine, dry powder suitable for dry injection. These solids normally possessed 5 percent residual moisture before injection. The UCA solids recently tested contained approximately 35 percent moisture. Because the solids were identically prepared, the large difference in SO_2 removal can be attributed to the additional inherent moisture contained by the UCA. This inherent moisture provides the media for the reaction between the sorbent particle and SO_2 to take place; whereas, the low-moisture sorbent must rely on the moisture in the gas stream to provide a basis for any significant reaction to occur.

When comparing UCA and GCA, it is apparent that only a marginal increase in reactivity is noted for the ground solids. Although the increase in reactivity is very small, the important underlying factor is that, by grinding the fly ash prior to hydration with $Ca(OH)_2$, its hydration time could be reduced from 12 h to an optimum of 3 h. This confirms the bench-scale results discussed earlier (Figure 6).

Recycle Tests. The sorbents used in the above dry-injection tests were prepared by oven drying the ADVACATE slurry to a desired moisture content. However, on a commercial scale, a continuous operation, such as recycle mixing, would be a more practical and efficient way to produce a manageable, free-flowing product for feeding into a duct. Spent material from the first pass of GCA testing was used to prepare fresh solids for subsequent recycle tests.

One part $Ca(OH)_2$, 2 parts recycle product (RR=2.0), and 1.5 parts ground CR were hydrated at 90 °C (194 °F) for 3 h. The resultant slurry was then mixed with sufficient recycle product to obtain a moist, free-flowing ADVACATE product of approximately 30 percent residual moisture. The mixture appeared sufficiently dry to be handled by a dry-solids feeder. Injection of these solids resulted in an immediate lowering of the flue gas temperature. For this reason, tests were conducted with the temperatures initially set sufficiently high to allow for the decrease in bulk gas temperature to the desired ΔT_s of 11 °C (20 °F). Moist solids were injected into the simulated flue gas at approximately 93 °C (200 °F), and they consequently lowered the duct temperature to 54 °C (130 °F).

Three consecutive recycle passes were tested in the pilot plant to demonstrate the effect recycle has on overall sorbent reactivity and utilization. The results of these three passes, in addition to the baseline pass, are shown in Figure 13. Results indicate that the first pass of recycle (Recycle I) yielded 66 percent SO_2 removal at a SR of 1.0. Previous pilot-plant data[4] for unground fly ash reported that there was little enhancement, if any, from using recycle solids to enhance $Ca(OH)_2$ reactivity during slurrying instead of using fly ash only. One possible explanation is that using unground ash as the basis for ADVACATE sorbents provides much less overall surface area per gram of sorbent for calcium silicate formation than if ground ash is used. The recycle solids that were used in the current testing emanated from ground fly ash, which provided higher initial surface area for calcium silicate formation.

Figure 13 also shows the overall results of the second and third passes of recycle testing (Recycle II and III, respectively). These passes were tested at a SR of less than 1 owing to the increased grain loadings and the consequent difficulty in maintaining stable conditions. This difficulty in maintaining stable conditions can be attributed greatly to small (2 in. I.D.) duct size used with an 85 m^3/h (50 ft^3/min) pilot system. This terminal and mass transfer instability region has not been encountered in larger, and therefore more turbulent flow, systems under comparable duct conditions, and should not prove a problem in commercial applications. The results given for any recycle pass signify the highest achievable SR for that particular pass. Several tests were conducted for each pass to ensure consistent operation of the pilot plant. Recycle II material was injected at a moisture content of 35 percent and yielded 49 percent SO_2 removal at a SR of 0.61. Recycle III material, at 38 percent moisture, was injected at a SR of 0.75 and yielded 64 percent SO_2 removal. When compared with the first pass and Recycle I in Figure 13, these two data points suggest that equilibrium, or steady-state operation, is being approached owing to the diminishing increase in SO_2 removal.

Figure 14 presents the calculated external utilization of $Ca(OH)_2$ for each pass of material. It is even further suggested that equilibrium is being approached by comparing Recycles II and III to the previous passes. Utilization between Pass 1 and Recycle I increased from 33 to approximately 66 percent. Recycle II and III passes continued the upward trend to an average of 80 to 86 percent utilization, respectively. It can be assumed that steady-state operation has been achieved. The above results demonstrated that utilization of $Ca(OH)_2$ increased using recycle solids as a basis and indicate that SO_2 removals of up to 90 percent in-duct are possible at a SR of 1.0 and complete equilibrium.

COMMERCIAL ACTIVITIES[7]

Figure 15 represents the commercial scheme of the ADVACATE process as envisioned for the 10 MW_e pilot evaluation in 1991-1992. Key process features include:

- In-duct sorbent drying and SO_2 absorption
- One-step sorbent mixing/injection
- Sorbent preparation with two slurry tanks and one tower mill (attritor)

As envisioned, spent sorbent and boiler fly ash are collected in the electrostatic precipitator (ESP) or baghouse, which have been modified mechanically for higher loadings. A portion of the waste material is fed to a slurry tank along with fresh CaO, a larger portion is diverted directly to the mixer, and the remainder discarded as waste. From the first tank, CaO/fly ash slurry is pumped to a constant stir tank reactor (CSTR), where the slurry is constantly recirculated through a tower mill.

The combination of particle abrasion and heated water promotes the lime/silica reaction to form a calcium silicate phase on fly ash surfaces. The product silicate slurry is then routed to the mixer and blended with dry recycle solids to form a damp, free-flowing powder that is then injected into the duct at least 0.75 sec upstream of the dust collector. Duct velocities greater than 15 m/sec are necessary to ensure good gas/sorbent contact and to keep duct walls swept clean of agglomerates.

Although no studies have been made of silicate waste disposal, the waste is physically and chemically similar to spray dryer FGD waste except that it is more hygroscopic and pozzolanic. We speculate that, at the worst, the ADVACATE waste may be disposed in a similar manner as spray dryer wastes, but that by-product use as an insulation material or as an aggregate for construction materials is likely.

A number of cost studies for ADVACATE have been performed.[7] Results of a study performed by Radian for EPA show capital costs of $60-85/kW and operating costs of 6 to 7 mills/kWh. Table 1 summarizes a case involving retrofit of a 300 MW_e utility boiler burning 2.7 percent sulfur Midwestern coal, whereas Table 2 estimates costs for the same plant on a new (NSPS) construction basis. EPRI is currently estimating

ADVACATE costs on a par basis with other FGD technologies.

All the above presumes that ADVACATE will achieve 90 percent SO_2 removal at a SR of 1.2 or less. Optimism that these performance goals will be achieved on a commercial scale is based on in-house pilot and two previous field evaluations. EPA's in-house pilot (85 m^3/hr) plant has consistently shown 50 percent utilization of $Ca(OH)_2$ in-duct with unground sorbent (UCA) and 70 to 80 percent utilization with tower mill-ground sorbent (GCA) (Figure 16). The first field evaluation on a 1700 m^3/h slipstream at the University of Tennessee power plant showed 80+ and 90+ percent calcium utilization in-duct and in-duct/baghouse, respectively, using unground sorbent, when ADVACATE was integrated with the Moist Dust Injection process (Figure 16).[7] When the Limestone Injection Multistage Burner (LIMB) process was augmented by ADVACATE injection on a 3400 m^3/hr slipstream with ESP control, up to 97 percent additional SO_2 removal was realized (Table 3).[7]

Based on these performances, EPA and the University of Texas are currently licensing the ADVACATE technology to the Tennessee Valley Authority (TVA) for use in their system and to Asea Brown Boveri (ABB) Environmental Systems for marketing worldwide. The initial effort to prepare ADVACATE for commercialization is a joint venture of TVA, ABB, EPA, and EPRI at the Shawnee facility in Paducah, Kentucky, on a 10 MW_e slipstream. The current schedule calls for installation in mid-1991 and operation by fall 1991. From this effort, design and operating data for a full-scale system should be derived. Some consideration is being made at this time for the resulting full-scale ADVACATE demonstration to be proposed for a clean coal demonstration.

To support these efforts, Acurex Corporation and AEERL/EPA will continue research in the following process areas with federal funds and funding from ABB:

- Optimization of particle attrition
- Optimization of slurry/solids mixing
- Integrated small pilot evaluation
- ESP performance
- Evaluation of fly ashes and alternative calcium sources
- Effects of additives and impurities

Concurrent with the above efforts, Acurex and EPA will be developing non-fly-ash ADVACATE technology for possible use on non-coal-fired combustion sources of acid gases, including waste combustors, smelters, and industrial boilers.

CONCLUSIONS

1. Hydration time and temperature have significant effects on sorbent reactivity for sorbents prepared from ground fly ash and lime. Sorbent reactivity increases with increasing hydration time and temperature. Sorbents prepared from ground fly ash were always more reactive than sorbents made from unground fly ash at the same hydration conditions.

2. For sorbents hydrated at 90 °C (194 °F) for 3 h, reactivity tends to increase as the amount of fly ash grinding increases prior to sorbent preparation.

3. Reactivity tends to increase as sorbent surface area increases for sorbents hydrated from 1 to 6 h at 60 to 100 °C.

4. Pilot-plant results verify that initially grinding the fly ash before hydration with $Ca(OH)_2$ decreases the optimal hydration time from 12 to 3 h.

5. The inherent moisture content of the ADVACATE sorbent greatly affects its overall reactivity toward SO_2.

6. Initially using ground rather than unground fly ash provides for a more reactive sorbent in subsequent passes, owing to the higher initial surface area associated with the ground ash.

7. Pilot-scale data suggest that possible equilibrium, or steady-state operation, can be achieved by the third or fourth pass of recycle material. Recycle III material demonstrated $Ca(OH)_2$ utilization of up to 86 percent.

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TABLE 1. ADVACATE cost estimates for retrofit 300 MW_e plant

Cost Accounting Area	Description	$/kW Lime
10	Reagent Feed System	18.9
20	SO_2 Removal System	1.1
30	Flue Gas Handling System	8.7
60	Waste Handling System	3.4
70	General	1.3
	Installation	9.5
	Engineering & Start-up[a]	2.9
	Total Process Capital	45.8
General Facilities		4.6
Engineering and Home Office Fees[b]		4.6
Project Contingency		11.1
Process Contingency		12.2
Total Plant Cost		78.3
Allowance for Funds During Construction		1.9
Total Plant Investment		80.2
Royalty Allowance		0.2
Preproduction Costs		3.4
Inventory Capital		1.7
Initial Catalyst and Chemicals		0.0
Total Capital Requirement		85.5
Total Levelized Busbar Cost, mills/kWh		7.0

[a] Engineering & Start-up covers fees paid to the process vendor for hardware design and fabrication management.

[b] Engineering and Home Office Fees covers additional charges paid by the utility for plant engineering and project administration.

TABLE 2. ADVACATE cost estimates for new 300 MW_e plant.

Cost Accounting Area	Description	$/kW Lime	Baghouse[a]
10	Reagent Feed System	14.5	--
20	SO_2 Removal System	0.9	--
30	Flue Gas Handling System	4.5	--
60	Waste Handling System	2.6	--
70	General	1.0	--
	Installation	7.3	--
	Engineering & Start-up[b]	2.2	--
	Total Process Capital	33.0	35.9
General Facilities		3.3	3.6
Engineering and Home Office Fees[c]		3.3	3.6
Project Contingency		8.2	9.5
Process Contingency		8.8	0.0
Total Plant Cost		56.6	52.6
Allowance for Funds During Construction		1.4	1.3
Total Plant Investment		58.0	53.9
Royalty Allowance		0.2	0.0
Preproduction Costs		2.9	2.4
Inventory Capital		1.7	1.3
Initial Catalyst and Chemicals		0.0	0.0
Total Capital Requirement		62.8	57.6
Total Levelized Busbar Cost, mills/kWh		6.0	2.7

[a] Baghouse costs not included as SO_2 control costs.
[b] Engineering & Start-up covers fees paid to the process vendor for hardware design and fabrication management.
[c] Engineering and Home Office Fees covers additional charges paid by the utility for plant engineering and project administration.

TABLE 3. LIMB[a]/ADVACATE (Edgewater) SO_2 removal at a SR of 2.0, 63 °C

Dry ADVACATE Injection Rate (g/dscm)	Total SO_2 Removal (percent)	SO_2 Removal ADVACATE Only (percent)
0	65	0
10	84	54
20	93	80
30	97	91
40	99	97

[a] Limestone Injection Multistage Burners

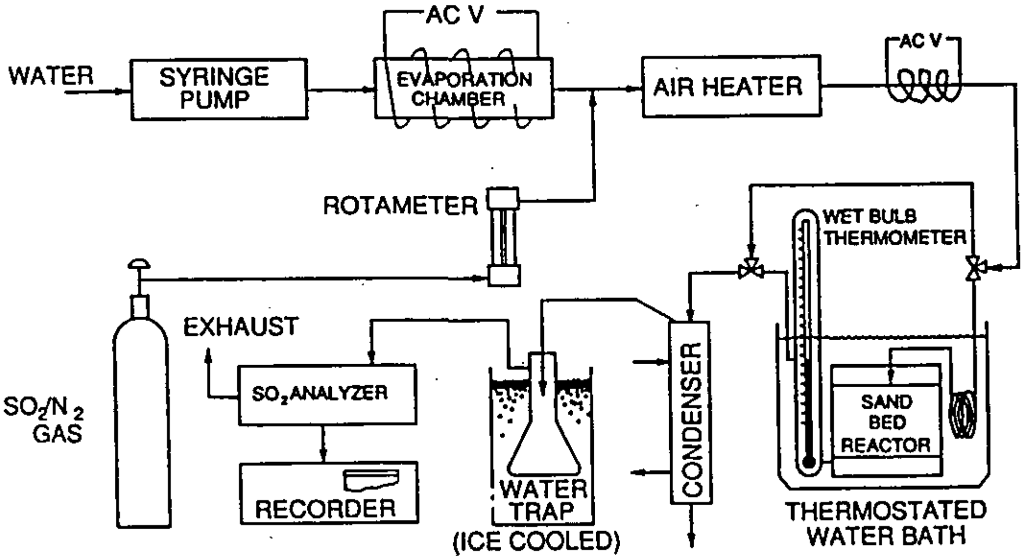

FIGURE 1. Schematic of sand-bed reactor bench-scale facilities.

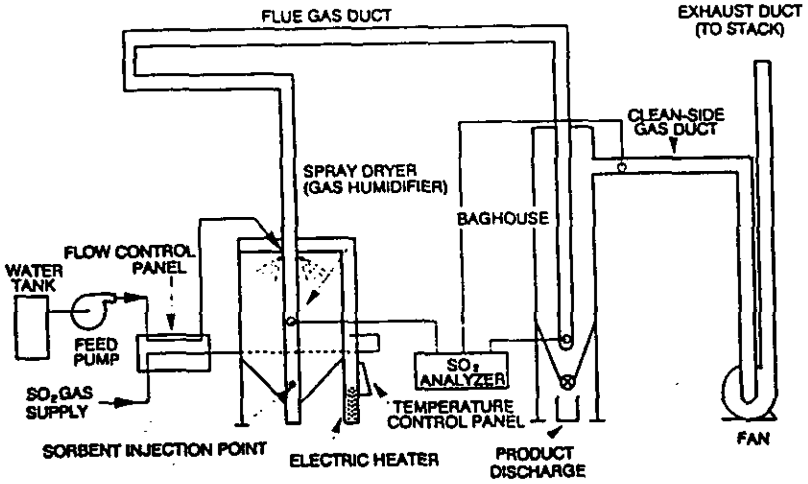

FIGURE 2. Schematic of duct injection pilot-plant facility.

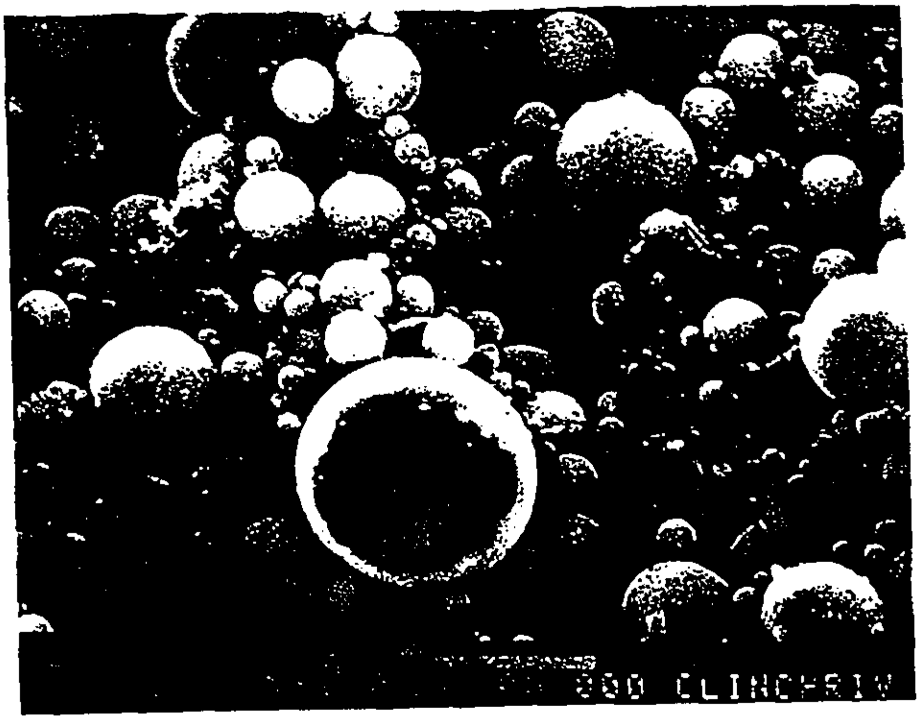

FIGURE 3. SEM photomicrograph of "as received" Clinch River fly ash (CR) sample, x2000.

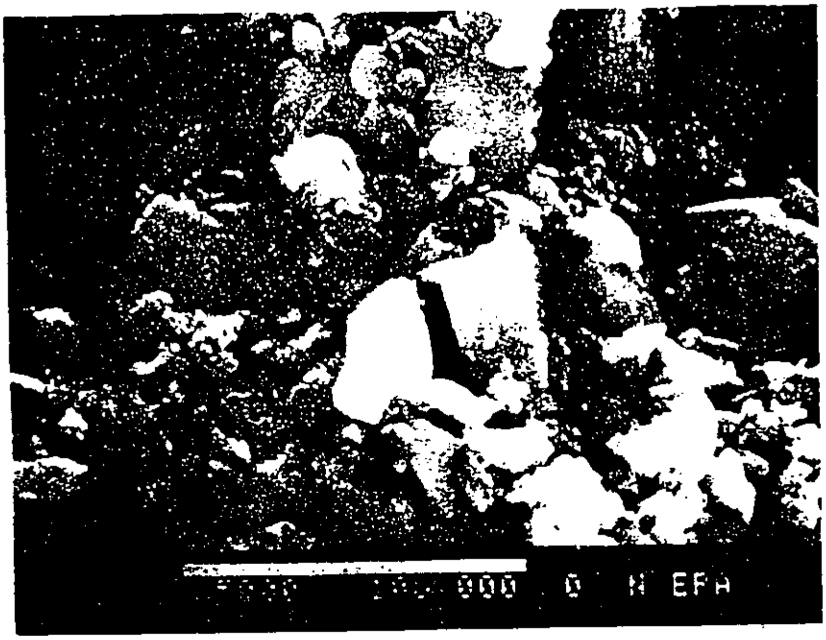

FIGURE 4. SEM photomicrograph of ground Clinch River fly ash (UPCR) sample, x5000.

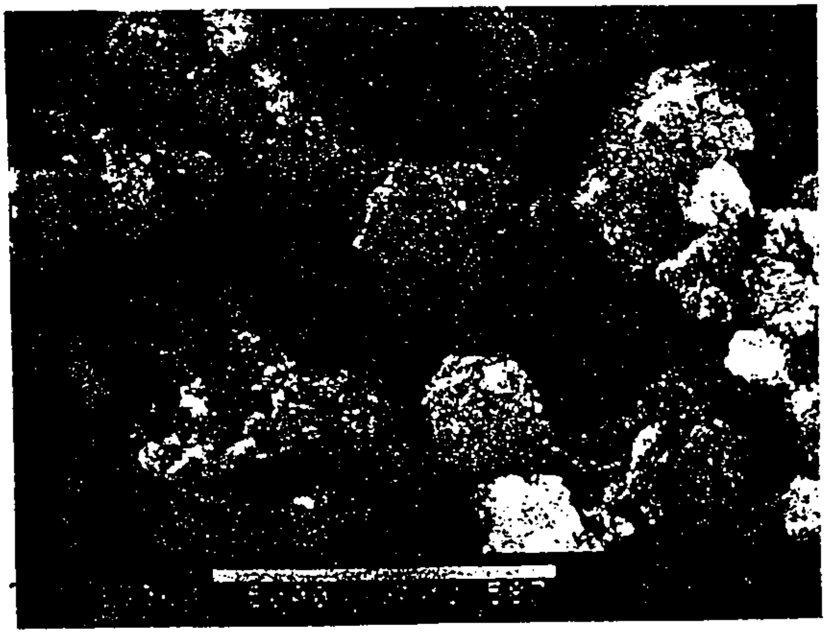

FIGURE 5. SEM photomicrograph of sorbent made from ground Clinch River fly ash (BCR) hydrated with $Ca(OH)_2$, at a 3:1 weight ratio, for 3 h at 90 °C (194 °F).

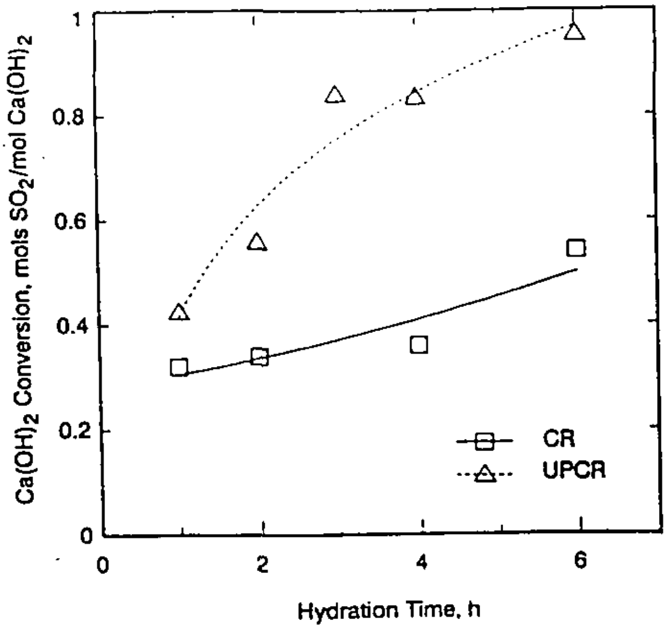

FIGURE 6. Conversion as a function of hydration time at 90 °C (194 °F), fly ash to $Ca(OH)_2$ weight ratio of 3:1 (conditions in the sand-bed reactor: 60 percent realtive humidity, 1000 ppm SO_2 concentration).

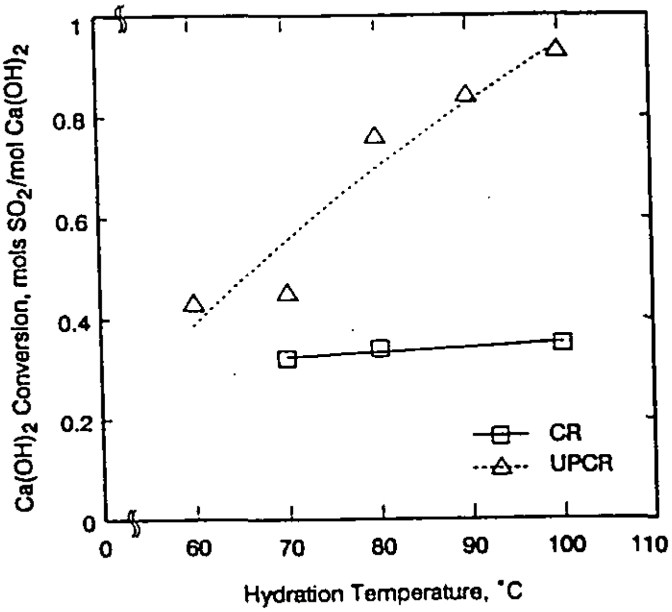

FIGURE 7. The effect of hydration temperature on the conversion of $Ca(OH)_2$, 3 h hydration (conditions in the sand-bed reactor: 60 percent relative humidity, 1000 ppm SO_2 concentration).

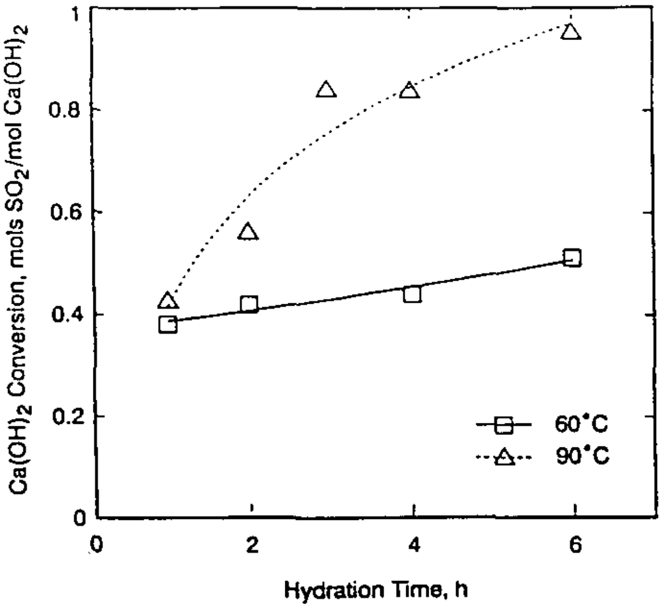

FIGURE 8. Conversion as a function of hydration time at 60 °C (140 °F), UPCR-to-Ca(OH)$_2$ weight ratio of 3:1 (conditions in the sand-bed reactor: 60 percent relative humidity, 1000 ppm SO$_2$ concentration).

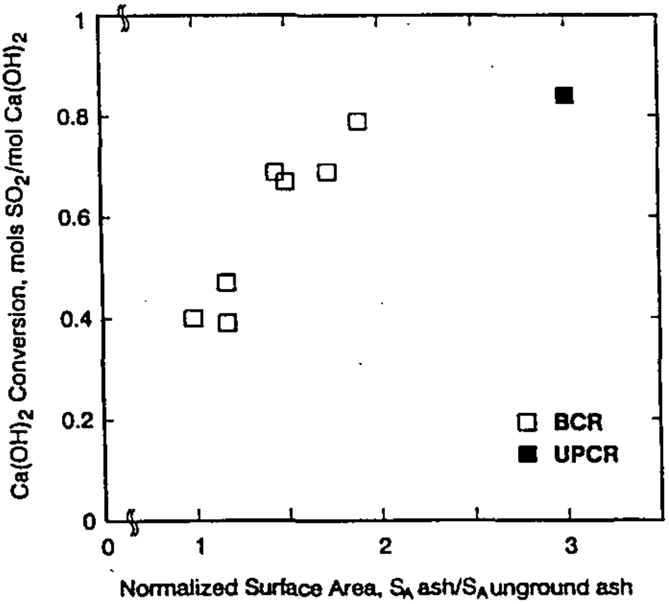

FIGURE 9. Effect of fly ash initial surface area on conversion of sorbents hydrated at 90 °C (194 °F) for 3 h at a fly ash-to-Ca(OH)$_2$ weight ratio of 3:1 (conditions in the sand-bed reactor: 60 percent relative humidity, 1000 ppm SO$_2$ concentration).

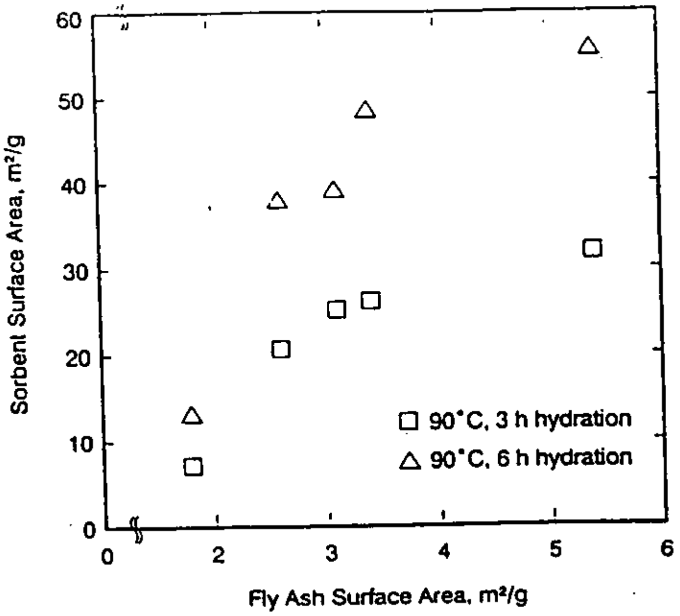

FIGURE 10. Effect of initial fly ash surface area on sorbent surface area for sorbents hydrated for 3 and 6 h at 90 °C (194 °F), fly ash-to-Ca(OH)$_2$ weight ratio of 3:1.

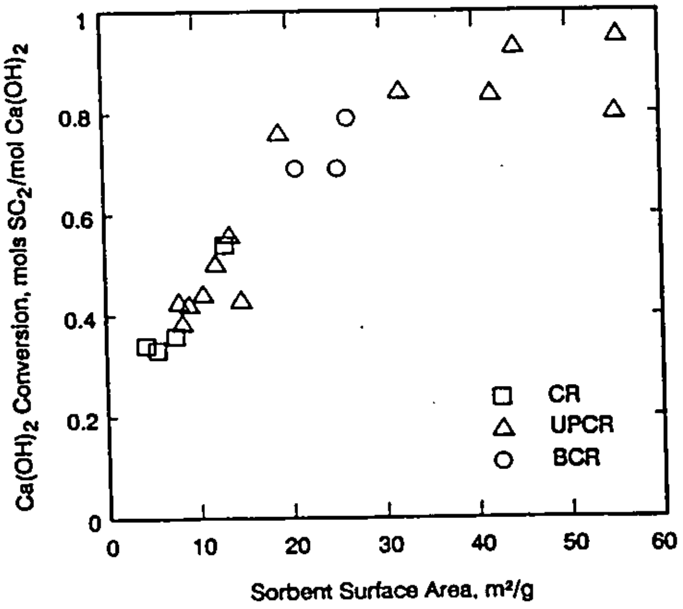

FIGURE 11. Effect of sorbent surface area on conversion for sorbents prepared at a fly ash-to-Ca(OH)$_2$ weight ratio of 3:1 (conditions in the sand-bed reactor: 60 percent relative humidity, 1000 ppm SO$_2$ concentration).

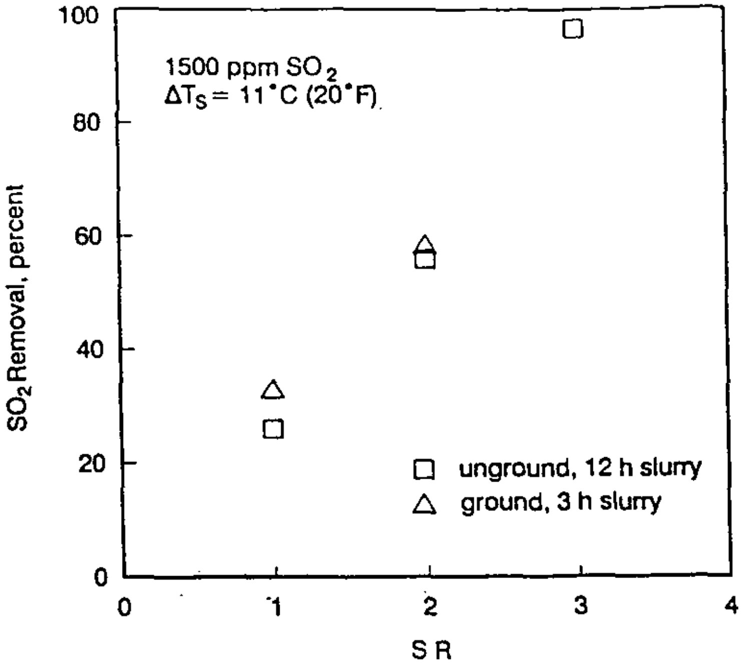

Figure 12. Comparison of ground and unground Clinch River ADVACATE at SR=1.0.

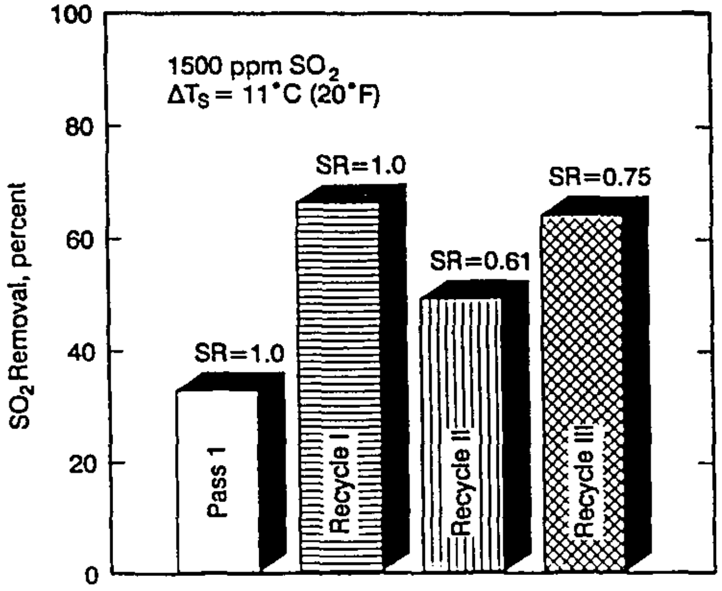

Figure 13. Comparative effect of recycle solids on in-duct SO_2 removal.

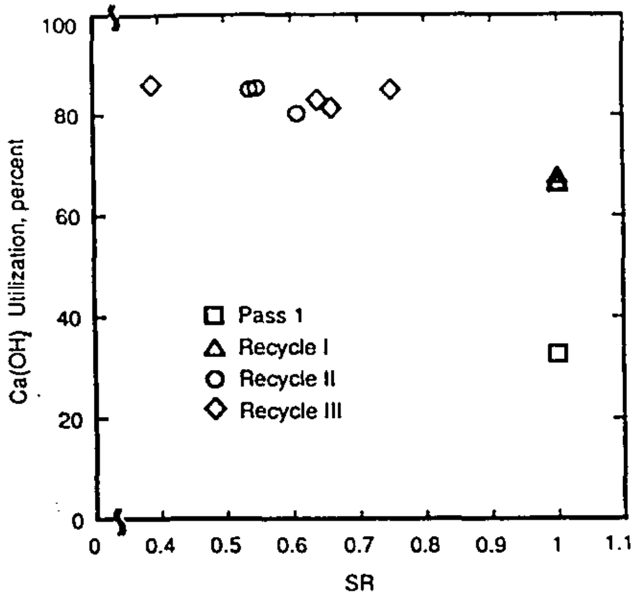

FIGURE 14. Calculated utilization of $Ca(OH)_2$ in ground CR (GCA) sorbent.

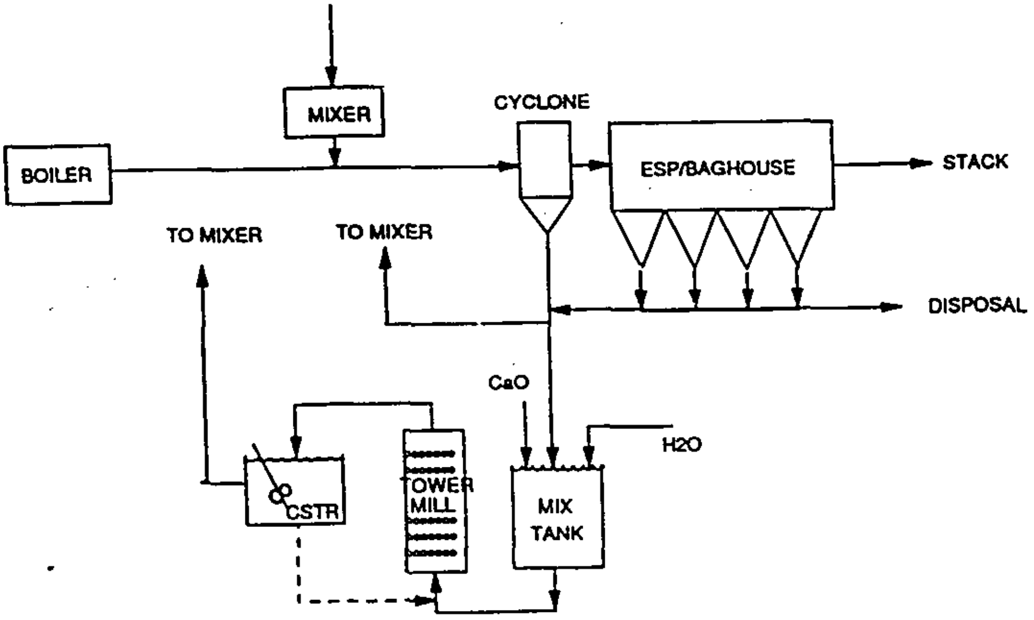

FIGURE 15. Commercial scheme of the ADVACATE process.

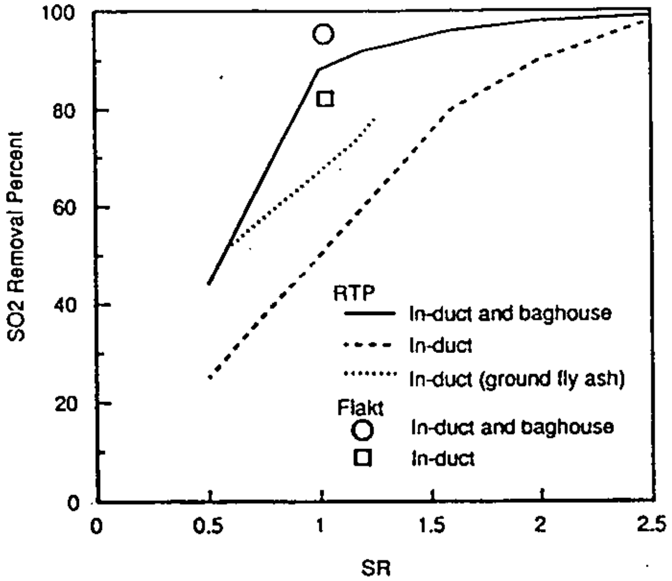

FIGURE 16. The effect of SR on SO_2 removal during calcium silicate damp injection. and it presents ADVACATE commercialization plans.

PB90-263930

COMMERCIAL DEVELOPMENT OF THE ADVACATE PROCESS FOR FLUE GAS DESULFURIZATION

Acurex Corporation
Research Triangle Park, NC

1990

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16. ABSTRACT The paper discusses the commercial development of the advanced silicate (ADVACATE) process for flue gas desulfurization (FGD). This commercialization follows several years of bench-scale sorbent development, pilot plant process optimization, and field pilot evaluation. Results are presented showing the importance of silica dissolution in preparation of calcium silicates. Pilot and field optimization and evaluation shows excellent potential for over 90% SO_2 control at half the cost of conventional wet FGD processes. Projected costs and commercialization plans are presented.

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Silicon Dioxide	Stationary Sources	07B
Flue Gases	ADVACATE Process	21B
Desulfurization	Advanced Silicate Process	07A, 07D
Sulfur Dioxide	Silica	

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COMMERCIAL DEVELOPMENT OF THE ADVACATE PROCESS FOR FLUE GAS DESULFURIZATION

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ABSTRACT

Commercialization of the ADVACATE process for flue gas SO_2 removal follows several years of bench-scale sorbent development, pilot plant process optimization, and field pilot evaluation. Results are presented showing the importance of silica dissolution in preparation of calcium silicates. Pilot and field optimization and evaluation shows excellent potential for over 90 percent SO_2 control at half the cost of conventional wet flue gas desulfurization (FGD) processes. Projected costs and commercialization plans are presented.

SORBENT DEVELOPMENT

Under U.S. Environmental Protection Agency sponsorship, Jozewicz and Rochelle [1] studied the reaction of flyash with calcium hydroxide [$Ca(OH)_2$] for the production of reactive sorbents for use in FGD. They used the bench-scale packed bed reactor shown in Figure 1 which is capable of simulating conditions in the baghouse. They found that solids produced by slurrying flyash and $Ca(OH)_2$ were significantly more reactive than $Ca(OH)_2$ alone. The conversion of $Ca(OH)_2$ [defined as the number of mols of SO_2 reacted with 1 mol of $Ca(OH)_2$] was proportional to the weight ratio of flyash to $Ca(OH)_2$. The results of experiments at 54 percent relative humidity are presented in Figure 2. Increasing the flyash-to-$Ca(OH)_2$ weight ratio from 0.5 to 20 increased $Ca(OH)_2$ conversion from 0.17 to 0.78. For a fixed flyash-to-$Ca(OH)_2$ weight ratio, the slurrying temperature was the decisive parameter affecting the process. There was a critical slurrying time for every temperature tested, for which conversion reached a maximum value. The time needed to reach this maximum shortened with increasing temperature. For example, at a flyash-to-$Ca(OH)_2$ weight ratio of 16, $Ca(OH)_2$ conversion converged on 0.4 after 16 hr of slurrying at 25 °C and on 0.8 after 5 hr at 92 °C. This was a dramatic improvement of $Ca(OH)_2$ conversion, compared with 0.12 for $Ca(OH)_2$ alone at 54 percent relative humidity. The above work concluded that silica was the most reactive compound of flyash and that the reactivity increased with slurrying time and temperature. It was postulated that the rate limiting step of the reaction of flyash with $Ca(OH)_2$ was the dissolution of silica from flyash.

Several approaches were taken to increase the dissolution rate of silica. In one series of experiments, different chemical species were added in varying concentrations to a solid mixture at a 3:1 flyash-to-$Ca(OH)_2$ ratio. The most significant enhancement was seen when 0.25 N NaOH was added to the sample slurried for 8 hr at 90 °C where the conversion in the sand bed reactor was 0.5 compared with 0.35 without NaOH addition [2]. These test results support the hypothesis that dissolution of silica in the flyash is the limiting step for the reaction between flyash and lime during slurrying. Similar gains in sorbent reactivity were reported when ammonium phosphate or phosphoric acid were used as additives for flyash dissolution [3].

An alternative way of increasing the reaction of $Ca(OH)_2$ with silica is to use a more reactive form of silica. Diatomaceous earth and bentonitic clays were selected for the screening study [4]. Diatomaceous earth or bentonitic clay was reacted with lime by slurrying, and the prepared sorbents were tested in the sand bed reactor. Significant enhancement of reactivity of $Ca(OH)_2$ was possible by using either one of these siliceous materials, with diatomaceous earth being the better silica source. Experimental data indicated that a mixture of diatomaceous earth (MN-53) and CaO at a weight ratio of around 1:1 gives the optimum performance with respect to sorbent conversion with SO_2. This ratio also translates into a molar ratio of SiO_2/CaO of around 1:1.

Another way of increasing the flyash dissolution rate is to increase slurrying temperature. Jozewicz et al. [5] investigated the effects of pressure hydration on the reactivity of the flyash/$Ca(OH)_2$ solids towards SO_2. Pressure-hydrated sorbents with flyash-to-$Ca(OH)_2$ weight ratios ranging from 1:2 to 5:1 were investigated. Results were observed as follows:

- $Ca(OH)_2$ conversion of 0.46 percent at 150 °C, was the maximum observed.
- Conversion of $Ca(OH)_2$ decreased above 150 °C, likely due to formation of cryptocrystalline silicates.
- Conversion of $Ca(OH)_2$ decreased less dramatically as temperatures were lowered from 150 °C to 100 °C.

Another approach to increase the rate of formation of calcium silicates by reacting flyash with $Ca(OH)_2$ is to increase the availability of silica by means of flyash grinding. A laboratory scale grinder using stainless steel balls as a grinding medium was used for grinding. The ash was ground for 1 hr at room temperature and atmospheric pressure. As a result, median particle sized decreased from 12 to 5 µm.

The importance of the weight ratio of siliceous material to $Ca(OH)_2$, hydration time, and hydration temperature, known to have the most significant effects on the reactivity with SO_2 of sorbents prepared with unground flyash and $Ca(OH)_2$, was verified for sorbents produced with ground flyash and $Ca(OH)_2$. The time/temperature effect was tested for hydration time ranging from 1 to 6 hr. Initially, 90 °C hydration temperature was evaluated, and the results are shown in Figure 3. The conversion of $Ca(OH)_2$ in the sand bed reactor increased with increasing hydration time. Conversion increased significantly for the hydration time extended from 1 to 3 hr. However, no significant increase of conversion took place for further increased hydration time from 3 to 6 hr. Reactivities of sorbents produced with the unground flyash are shown for comparison in Figure 3. For the time/temperature conditions tested, higher conversions of $Ca(OH)_2$ were measured in the sand bed reactor for sorbents produced with ground than with unground flyash.

The attractiveness of the furnace sorbent injection process for the removal of SO_2 could be further increased if higher levels of overall sorbent utilization were attainable [6]. One option is to use the unreacted sorbent leaving the boiler to further remove SO_2 downstream. If this sorbent, which under normal conditions is relatively unreactive, could be reactivated to remove SO_2 during its residence time in the duct and particulate control device, it would improve SO_2 capture of the furnace sorbent injection process.

Two sorbents are considered as candidates for injection into the furnace: $CaCO_3$ and $Ca(OH)_2$. Injection of either could result in a mixture of flyash/$CaSO_4$/CaO leaving the boiler. The ratio of $CaSO_4$ to CaO may vary for $Ca(OH)_2$ and $CaCO_3$ injection, depending on their reactivity. To test the potential for reactivation of post-furnace injection solids (PLIS) resulting from the injection of $CaCO_3$ [$CaCO_3$-based PLIS] or $Ca(OH)_2$ [$Ca(OH)_2$-based PLIS], samples were tested for their reactivity with SO_2 following slurrying of "as received" as well as ground solids. The results are presented in Figure 4 giving the reactivity in the form of calcium normalized SO_2 capture in mmol SO_2/g of calcium. Each set of reactivation conditions was tested on $CaCO_3$- and $Ca(OH)_2$-based PLIS. For the reactivation conditions used, there was no significant difference in reactivity of reactivated solids. Grinding with subsequent slurrying at high temperature almost tripled the SO_2 capture, compared to "as received" solids. Grinding appears to be a necessary step in order to reactivate PLIS. Significantly less reactivation of PLIS could be achieved by slurrying of "as received" material.

PROCESS DEVELOPMENT

Bench-scale testing had indicated that lime slurried with silica produced a more reactive sorbent than lime alone. In searching for a way to best utilize this superior sorbent, a series of tests was begun in the 85 m^3/h (50 cfm) pilot facility shown in Figure 5, consisting of a Niro Atomizer spray dryer, a dry sorbent screw feeder, a 5.08 cm (2 in.) I.D. duct varying from 12 to 35 m (40 to 110 ft) in length, a cyclone module, and a pulse-jet baghouse having a 2.3 m^2 (25 ft^2) collection area. Electrically heated air augmented with SO_2 (typically at 1,500-2,000 ppmv) serves as a surrogate flue gas [4].

Initial tests, in March 1986, examined the sorbent slurry prepared by mixing 5 parts flyash, 1 part $Ca(OH)_2$, and 14 parts water in a heated tank at 90 °C for 12-16 hr. The slurry then was injected via a spray dryer into a gas stream. Although 40-50 percent SO_2 removal was observed, severe wall deposits were noted at approaches to saturation of 10-50 °C. Subsequently, tests were conducted using spray dried sorbent prepared similarly to that above, but injected as a powder into a prehumidified gas stream of 11-17 °C approach to saturation. With this approach a 50 percent SO_2 removal at Ca/S=1.0 and 80 percent at Ca/S=2.0 were observed across the baghouse. After this time (April 1986), all subsequent process development focused on a dry sorbent [4,7].

In September 1986, spray dried silicates prepared at 90 °C for 8 hr were shipped to the Electric Power Research Institute's (EPRI's) Arapaho pilot facilities for evaluation. Despite the fact that the "EPA hydrate" fared better than any EPRI calcium-based material—94 percent and 55 percent SO_2 removal at Ca/S = 2.0 and ΔT_{sat} = 11 °C for a fabric filter and electrostatic precipitator collector, respectively—the performance was not as good as EPRI's results for sodium sorbents [8]. Examination of the sorbent shipped to Arapaho showed a significant deterioration (via an unknown mechanism) in surface area (from 46 to 31 m^2/g) due to spray drying prior to shipment or change during transport. Therefore, in October 1986, other methods of converting silicate slurry to a dry sorbent were investigated.

An alternative method to thermal drying was developed wherein a slurry of flyash and hydrated lime was reacted at a ratio of 3:1 for 12 hr, and filtered to obtain a cake of 30-60 percent moisture. A cement mixer was used to blend this cake with dry recycle solids produced from a previous test run. The 3 parts of dry recycle solids combined with 1 part of wet cake produced a damp powder containing 25-33 percent moisture [7]. At this point, a second important property of calcium silicates became evident—that of moisture-carrying ability. The 25-33 percent moisture solids generated by mixing slurry or filter cake with dry recycle solids were generally free-flowing and dry to the touch. Any lumps in the mixture could be easily pulverized by modest shaking or even drying (when injected). With the development of more reactive calcium silicates, this moisture-carrying ability has been shown to increase in proportion to surface area; therefore, the reactivity toward SO_2 increases with increased moisture content.

In 1987, EPA and Acurex began preparing pressure hydrated sorbents in 50 percent solids slurries and mixing 2 parts dry recycle solids with 1 part fresh slurry [9]. The pilot plant results for silicates injected into a humidified gas stream showed typically 50 percent SO_2 removal in-duct (for 2 s residence time) and 90 percent SO_2 removal through the fabric filter at a Ca/S ratio of 1.0 as shown in Figure 6.

In 1988, EPA and Acurex made an extensive set of runs using furnace injected sorbent waste as the starting material for calcium silicate. The results showed that up to 80 percent additional capture of SO_2 could be achieved at a recycle ratio of 2.0 using furnace solids from a furnace injection (LIMB) process that was 25 percent utilized in the furnace (50 percent SO_2 capture at Ca/S = 2.0)[10]. In this evaluation, the newly evolved process was christened ADVACATE for advanced silicate.

In April-May 1989, EPA sponsored a field pilot evaluation of a combined LIMB-ADVACATE process where a 105 MW_e LIMB demonstration (Edgewater Plant, Lorain, OH) fed a 3400 m^3/hr (2,000 cfm) slipstream through a pilot ESP. The results of this evaluation, shown in Figure 7 and Table 1, were so encouraging—up to 98 percent SO_2 removal—that intense efforts regarding possible demonstrations were begun [11]. It is worth noting here that current sorbents are roughly three times more reactive than those prepared and used at Edgewater.

Referring to Figure 6, AEERL was negotiating the first EPA Cooperative Research Agreement under the Federal Technology Transfer Act of 1986 with Flakt, USA, of Knoxville, TN, during 1987-88 when it was learned that Flakt was developing its own dry injection technology, called Moist Dust Injection or MDI. Flakt's concept was to create a dense-phase reaction in-duct using moistened recycle product with fresh lime to simultaneously capture SO_2 and humidify the gas in the duct.

Following the signing of this agreement in November 1988, silicate sorbent was prepared at a 3:1 flyash-to-lime ratio at about 20 percent moisture for Flakt to use in place of lime in their MDI pilot plant at the University of Tennessee in Knoxville. In early December 1988, Flakt conducted runs at 1000, 2000, and 3000 ppmv SO_2 at a ΔT_{sat}=11 °C and realized SO_2 removals of ~80 percent in-duct and 96 percent through the fabric filter at a Ca/S=1.0 as shown in Figure 6. The reason for this improvement in in-duct SO_2 capture—80 percent at Flakt versus 50 percent at the EPA pilot plant located at Research Triangle Park (RTP), NC—is still uncertain. However, in recent months, attempts to simulate MDI at RTP have shown up to 60 percent SO_2 capture at Ca/S=1.0. Currently the effect of gas-phase turbulence (Reynolds number) on SO_2 capture in-duct is being studied as a possible explanation. For the time being, it is sufficient to know that SO_2 capture using MDI/ADVACATE appears to improve with increasing duct dimensions, assuming an adequate duct velocity for good mixing (at least 15 m/s).

As a final note on process development, silicate sorbents of 90 m^2/g are routinely produced at RTP in 2-3 hr and show an ability to hold 60 percent moisture and be free-flowing. Such a remarkable improvement in sorbent characteristic suggests the need to update previous pilot plant data. A more intriguing notion is operation of dust collectors at saturation temperatures. If a silicate sorbent holds up to 60 percent moisture and appears dry, is condensation even feasible for a cooled gas-stream containing dry or partially dried silicates? By the time this paper is presented in August 1990, the authors should have more insight into operation at apparent saturation conditions.

The impact of an ADVACATE, LIMB-ADVACATE, or ADVACATE/MDI process upon particle collection has been ignored up to now because we have little experience with ADVACATE-ESP operation, although it is acknowledged that ESPs do not normally operate at high dust loadings. Conversely, both EPA and Flakt have considerable experience with ADVACATE and MDI processes on fabric filters, and no significant operational problems have occurred.

CURRENT STATUS

Figure 8 represents the diverse process schemes for ADVACATE. The source of calcium for silicates may be furnace-injected lime or limestone, lime added to the silicate slurry tanks, or, in the case of many Western coals, calcium occurring in coal ash. Waste solids from a LIMB or fluid-bed combustor (FBC) are also a potential economic source of calcium.

The way sorbent is injected depends upon vendor preference and calcium source. In LIMB experience, limestone or lime is pneumatically conveyed into the furnace while ADVACATE sorbent has been fed via direct screw feeding, pneumatic conveying, or the proprietary MDI process of Flakt. The flue gas may be cooled/humidified via moisture injection, indirect gas cooling, evaporation of moisture in damp solids injected, or a combination of these techniques. Particle collection is either by fabric filter or ESP, although the latter may require a precollector to protect electrical integrity, as well as segregate the coarser (less reacted) material for recycle.

EPA's current plans include a 10 MW_e field pilot evaluation at the Shawnee No. 9 facility of the Tennessee Valley Authority (TVA) starting in 1991. Several sources of sorbent, including FBC and LIMB waste, will be evaluated on a 2.5-3.0 percent sulfur Midwestern coal.

If successful, the Shawnee effort will be followed by augmentation of the LIMB demonstration at Virginia Power's Yorktown No. 2 (180 MW_e) unit starting in 1992. This effort currently is scheduled for installation of LIMB using a hydrated lime sorbent, but can be converted to a furnace-injected limestone-ADVACATE process upstream of an ESP. A 2.4 percent sulfur Eastern bituminous coal will be fired at Yorktown for this demonstration.

Another interesting process variation is the ability of ADVACATE to use the natural alkalinity of Western coal flyash to remove SO_2. In many cases, such as Powder River subbituminous coal, a variation of ADVACATE can potentially remove all the SO_2 at roughly half the cost as presented later in this paper. This is because half the cost of ADVACATE is for lime or limestone receiving, storage, and handling. The low-sulfur ADVACATE process variation would also incur no additional solid waste (except for the weight of sulfite/sulfate added due to SO_2 absorption). This option is slated for study later in 1990 on a pilot scale at RTP.

COST ESTIMATES

Based on the preliminary design of an ADVACATE/MDI process for the Yorktown demonstration, cost estimates have been prepared by Radian Corporation using EPRI economic premises. Performance predictions based on the Flakt field evaluation for ADVACATE and Edgewater results for LIMB/ADVACATE have been incorporated along with cost estimates. In Figure 9 the total annualized cost estimates for a 300 MW_e retrofit using 2.2 percent sulfur coal and the furnace limestone ADVACATE/MDI system are compared with wet FGD. Although the economic assumptions are too numerous to discuss here, the basic philosophy is that all factors are comparable except that the highest process and project contingency factors have been assigned to ADVACATE due to the unproven nature of the technology.

For utility-scale applications, the estimated cost of ADVACATE is typically 30-35 percent capital and 40-50 percent operating of the costs for wet FGD. Table 2 shows a similar analysis performed on slurry costs of wet FGD vs. lime and limestone ADVACATE for a 350 MW_e, 2.2 percent sulfur coal retrofit. ADVACATE is projected to cost around \$60/kW and, for this case, \$225-250/ton SO_2 removed, compared with \$195/kW and \$520/ton SO_2 removed for wet FGD. In these comparisons, the wet FGD was assumed to have no spare modules, no reheat, and a retrofit factor of 1.5.

PROCESS SUMMARY

The key feature of the ADVACATE process is its unique calcium silicate sorbent which has the following properties:

- Typical surface area of 70-90 m^2/g.
- Reactivity toward SO_2 4-6 times that of lime.
- Ability to carry high (60 percent) moisture content and be free flowing.

High calcium utilization observed in bench, pilot, and field pilot evaluations is likely due to the free calcium's being all surface calcium. Unlike conventional sorbent injection technology, spent ADVACATE sorbent does not have an unreacted core.

The process developed for this sorbent has the following features:

- No absorber vessel.
- No requirement for separate humidification of flue gas.
- Dry waste solids.
- Various plant waste waters, including boiler blowdown, can be used for ADVACATE slurry.
- SO_2 removal comparable to wet FGD at less than half the cost.

Additionally, virtually no potential for in-duct sorbent deposits exists due to the hygroscopic nature of ADVACATE sorbent. Potential operation at apparent saturation temperatures seems feasible.

At this time, the process has not operated in a continuous recycle mode, nor has ESP performance with ADVACATE injection been studied in depth. These and other questions should be resolved at Shawnee during 1991.

REFERENCES

1. Jozewicz, W., and G.T. Rochelle, "Flyash Recycle in Dry Scrubbing," Environ. Progress, 5, 219, 1986.
2. Jozewicz, W., J.C.S. Chang, C.B. Sedman, and T.G. Brna, "Characterization of Advanced Sorbents for Dry SO_2 Control," Reactivity of Solids, 6, 243, 1988.
3. Jozewicz, W., and J.C.S. Chang, "Evaluation of FGD Dry Injection Sorbents and Additives, Vol. 1, Development of High Reactivity Sorbents," EPA-600/7-89-006a (NTIS PB89-208920), May 1989.
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5. Jozewicz, W., J.C.S. Chang, C.B. Sedman, and T.G. Brna, "Silica-Enhanced Sorbents for Dry Injection Removal of SO_2 from Flue Gas," JAPCA, 38, 8, 1027, August 1988.
6. Jozewicz, W., J.C.S. Chang, T.G. Brna, and C.B. Sedman, "Reactivation of Solids from Furnace Injection of Limestone for SO_2 Control," Environ. Sci. Technol., 21, 664, 1987.
7. Chang, J.C.S., and C. Jorgensen, "Evaluation of FGD Dry Injection Sorbents and Additives, Vol. 2, Pilot Plant Evaluation of High Reactivity Sorbents," EPA-600/7-89-006b (NTIS PB89-214134), May 1989.

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9. Chang, J.C.S., and C.B. Sedman, "Scale-up Testing of the ADVACATE Damp Solids Injection Process," In: Proceedings: First Combined FGD and Dry SO2 Control Symposium, Vol. 3, EPA-600/9-89-036c (NTIS PB89-172175), March 1989.

10. Jorgensen, C., J.C.S. Chang, C.B. Sedman, and D.C. Drehmel, "Pilot Plant Evaluation of Post-Combustion LIMB SO2 Capture," In: Proceedings: First Combined FGD and Dry SO2 Control Symposium, Vol. 2, EPA-600/9-89-036b (PB89-172167), March 1989.

11. Chang, J.C.S., B.W. Hall, W. Jozewicz, G.H. Marchant Jr., C.B. Sedman, and M.A. Maxwell, "Reactivation of Edgewater LIMB Solids by the ADVACATE Process for In-Duct SO2 Removal," Presented at 1990 SO2 Control Symposium, Session 4A, New Orleans, LA, May 1990.

Table 1. Edgewater SO_2 Removal at Ca/S = 2.0, H_{out} = 63 °C

Dry ADVACATE Injection Rate (g/dscm)	Total SO_2 Removal (%)	SO_2 Removal ADVACATE Only (%)
0	65	0
10	84	54
20	93	80
30	97	91
40	99	97

Table 2. ADVACATE Retrofit Economic Evaluation
(Basis: 350 MWe, 2.23% S coal, May 1988 dollars, limestone=$9.76/tonne, lime=$67.46/tonne)

	Wet FGD	Limestone/ ADVACATE	Lime/ ADVACATE
Capital cost, $/kW	195	58	59
Cost, $/tonne SO_2	520	224	255

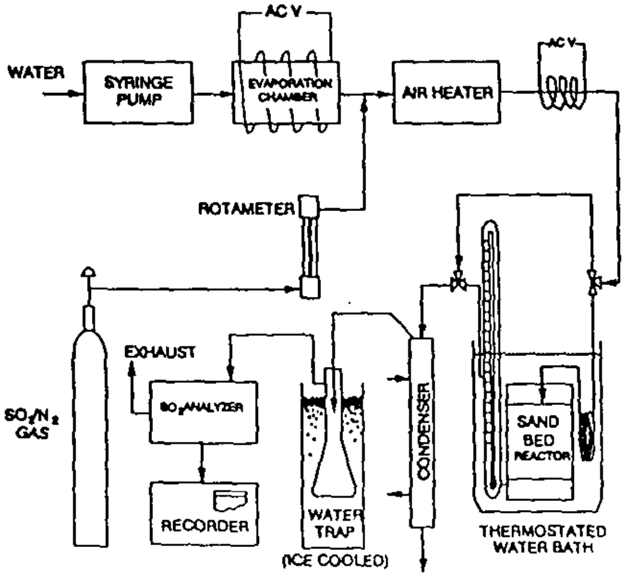


Figure 1. Bench-scale sand bed reactor.

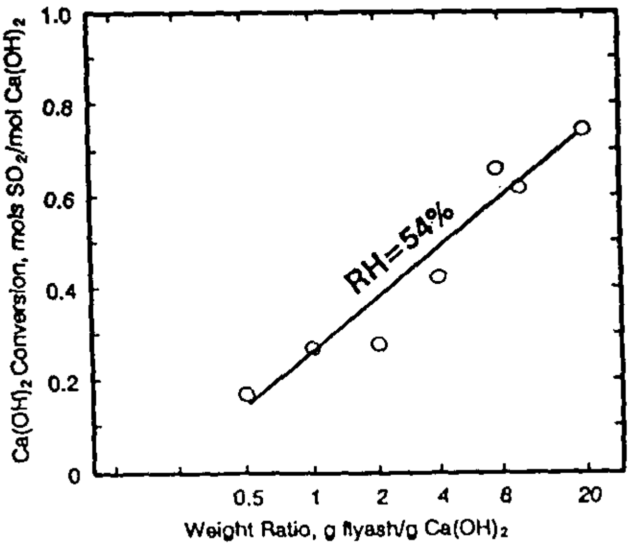


Figure 2. The effect of flyash-to-$Ca(OH)_2$ weight ratio on $Ca(OH)_2$ conversion (54 percent relative humidity, 500 ppm SO_2).

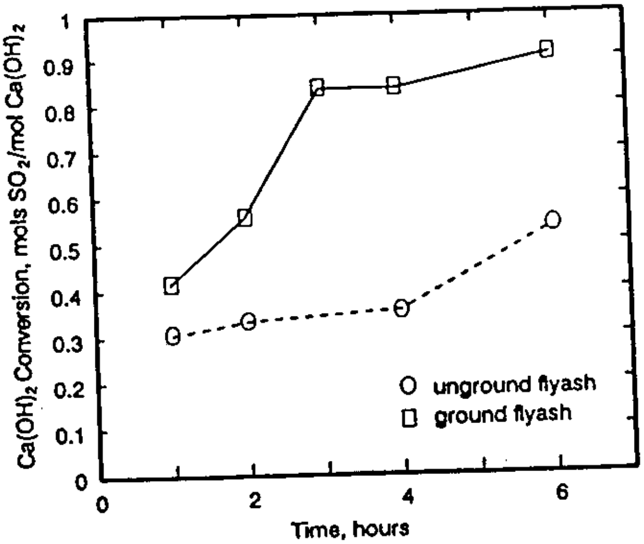


Figure 3. The effect of slurrying time on the conversion of Ca(OH)2 (90 °C slurring, flyash-to-Ca(OH)$_2$ weight ratio of 3:1).

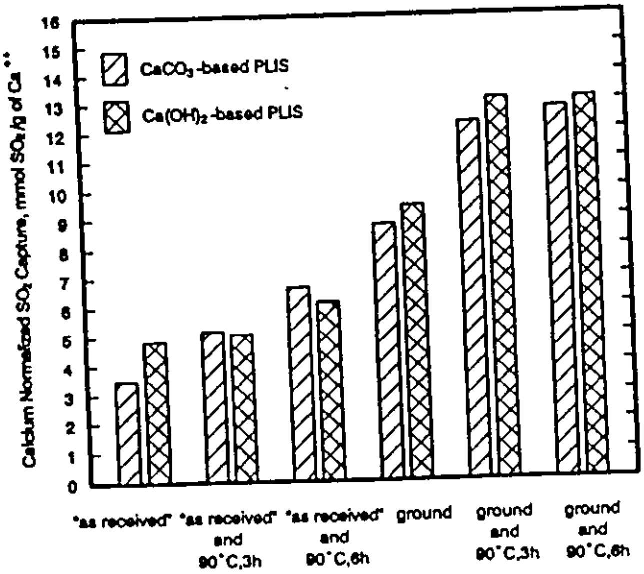


Figure 4. Comparative reactivity evaluation of CaCO3- and Ca(OH)2-based PLIS.

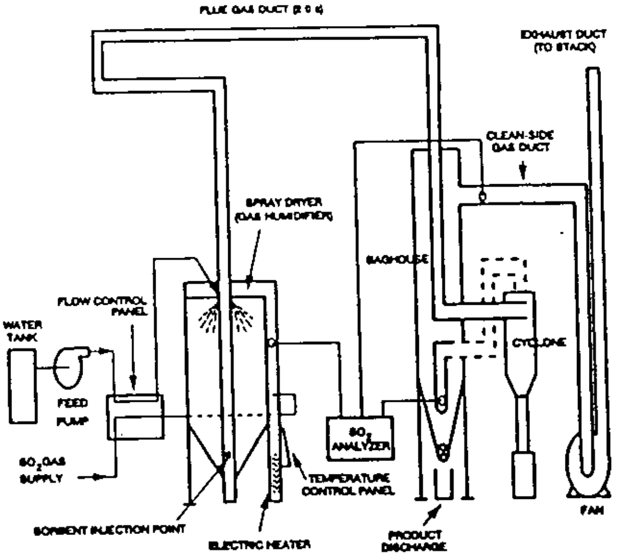


Figure 5. Pilot plant facility.

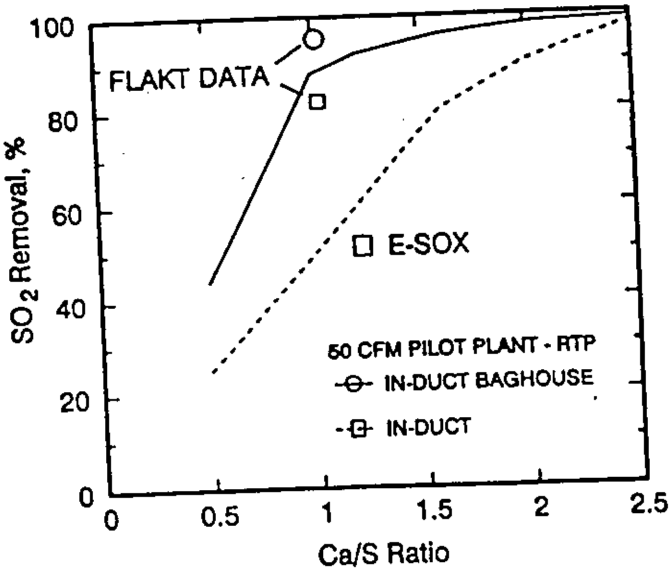


Figure 6. SO$_2$ removal vs. stoichiometric ratio for calcium silicate damp injection.

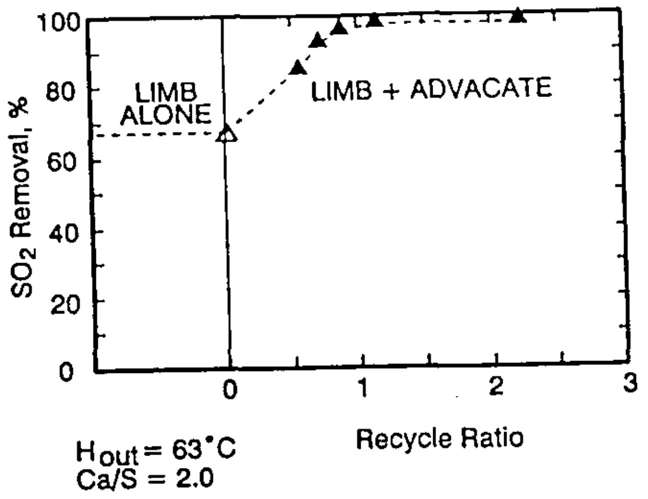
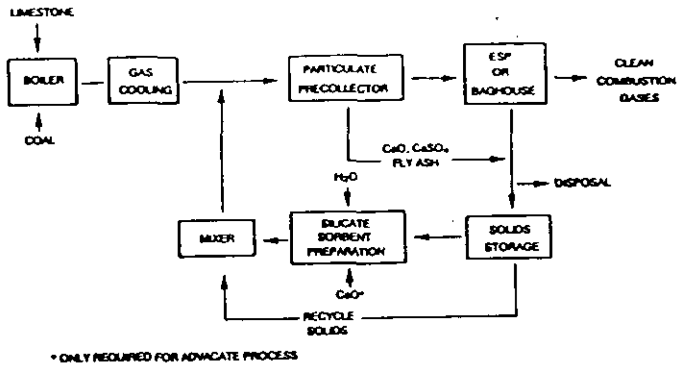


Figure 7. Edgewater LIMB/ADVACATE results.



* ONLY REQUIRED FOR ADVACATE PROCESS

Figure 8. ADVACATE process schematic.

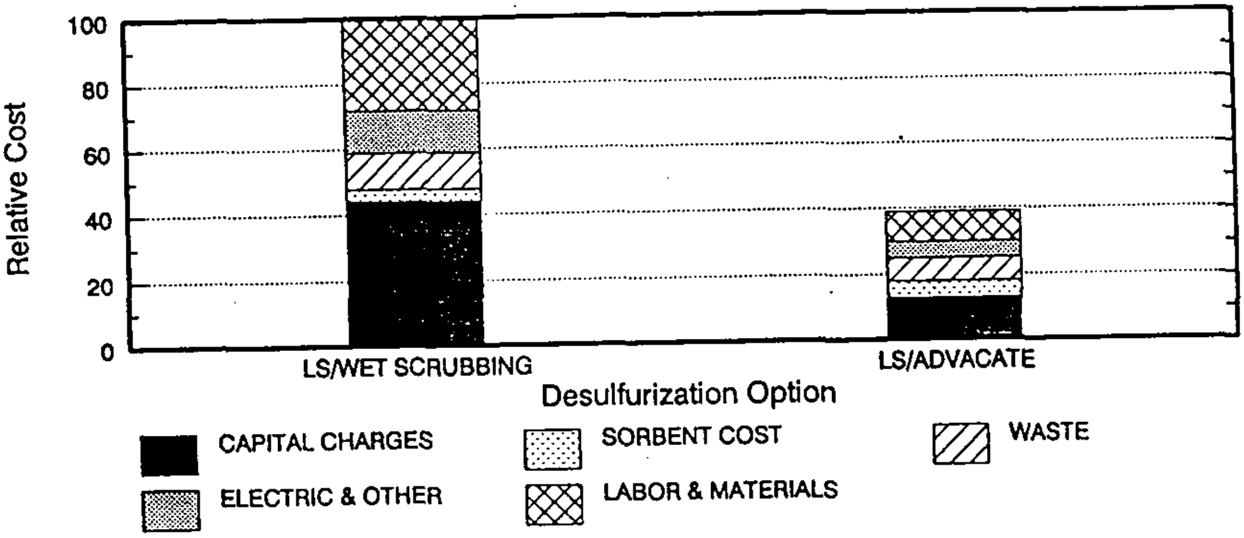


Figure 9. Total cost comparison - limestone wet scrubbing vs. limestone/ADVACATE.

This paper has been reviewed in accordance with the U.S. Environmental Protection Agency's peer and administrative review policies and approved for presentation and publication.

HRS STATE OF FLORIDA
DEPARTMENT OF HEALTH AND REHABILITATIVE SERVICES

ESE-WPB

January 12, 1993

RECEIVED

JAN 15 1993

Division of Air
Resources Management

Mr. Willard Hanks
Florida Department of Environmental Regulation
Division of Air Resources
Bureau of Air Regulation
2600 Blair Stone Road
Twin Towers Office Building
Tallahassee, Florida 32399-2400

Re: Comments on the Letter from Carlton, Fields, Ward, Emanuel, Smith & Cutler, P.A., the Law Firm Representing the Two Cogeneration Projects of the "Flo-Sun" Corporation

Dear Mr. Hanks:

Our office received a copy of the above referenced letter on January 4, 1993 and offers the following comments:

1. The applicant suggests that the BACT analysis should be based on the Palm Beach County imposed long term average SO_2 emission limitation of 1000 TPY on the combined projects. This seems unreasonable since this limitation is based on the average SO_2 emissions over the life of the facility (estimated at 30 years) and is largely not enforceable. The county zoning conditions also provide SO_2 limits based on the biomasss fuel that is supplied by the county. If the county makes at least 200,000 tons of biomass per year available, the applicant may emitt a maximum of 1500 TPY of SO_2, not to exceed an average of 1300 TPY for a five year period. If the county cannot make this quantity of biomass available, the applicant may emitt up to 1700 TPY of SO_2, not to exceed 1500 TPY for a ten year period. These limitations appear to be more enforceable, and therefore, we believe that at least 1500 TPY of SO_2 emissions should be used to determine BACT for these projects.

2. It is mentioned in the above referenced letter that BACT regulations indicate DER and EPA <u>may</u> consider actual operating conditions when determining BACT. From the information provided in the applications, it is difficult to determine what might constitute "actual" operating conditions for these facilities in the future. The applicant provides scenarios of only biomass fuel firing, 25% oil firing and 25% coal firing. The primary factor in predicting coal and oil usage is the availability of the biomass fuel which is entirely unknown, as recognized by the county and the applicant. If the biomass fuel is not available, the facilities will surely burn quantities of coal and/or oil necessary to profitably run these units. Therefore, we could realistically expect the following cases:

- Biomass fuel is readily available and SO_2 emissions are actually reduced
- Only a portion of the required biomass fuel is available and coal is used at only one facility

- Only a portion of the required biomass fuel is available and coal is used at both facilities
- Biomass availability is poor; bagasse is burned during sugar cane season and 25% coal is fired in the off-season.

It is impossible to predict which case will eventually define "actual" operating conditions in the future.

3. Consider the following information found in the applications, which assumes that both sites will operate at 25% Coal Firing:

	HEAT INPUT (BTU x 10^{12})		TPY OF FUEL	
	Biomass	Coal	Biomass	Coal
Sol-Energy	4.941	1.647	581,294	68,625
Flo-Energy	8.118	2.706	955,059	112,750
Total	13.059	4.353	1,536,353	181,375

 1,536,353 TPY of Biomass
-1,130,000 TPY of Biomass from Sugar Mills
- 200,000 TPY of Biomas from PBC County Proposal
 *206,353 TPY of Additional Biomass to Operate at capacity

*(This is about 1,754,000 MMBTU/Year of heat input, or about 10% of the total required heat input, assuming 4250 BTU/lb fuel, similar to bagasse).

In other words, if these projects secured twice the amount of biomass that Palm Beach County is proposing to supply, they would still have to fire 25% coal to come close to the design capacity. This is another reason for assuming that the maximum firing of coal should be used for determining BACT.

Thank you for the opportunity to comment on these applications.

Sincerely,

For the Division Director
Environmental Science and Engineering

Jeffery J. Koerner
Jeffery F. Koerner, Engineer IV
Air Pollution Contorl Section

FJG/JFK/lh

cc: AP-Source File

HRS STATE OF FLORIDA
DEPARTMENT OF HEALTH AND REHABILITATIVE SERVICES

ESE - WPB

December 24, 1992

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DEC 30 1992

Division of Air
Resources Management

Mr. Willard Hanks
Air Permitting and Standards
Florida Department of Environmental Regulation
2600 Blair Stone Road
Tallahassee, FL 32399-2400

Re: AC50-219413, Flo-Energy, Inc.
AC50-219795, Sol Energy, Inc.

Dear Mr. Hanks,

After reviewing the response by KBN Engineering to the additional information requested for the proposed Flo-Energy and Sol-Energy cogeneration projects, we offer the following comments:

1. BACT DETERMINATION FOR SO_2

Response #7 (page RTC-53): Low sulfur coal is BACT for this project due to unreasonable costs associated with add-on controls.

Concern: We are confused as to the type of coal that was assumed for analysis of the three add-on controls:

- Furnace Injection W/Lime
- Duct Injection W/Lime
- Duct Injection W/Sodium

Was the 0.7% sulfur by weight coal used for the determination of the additional costs for these controls, or was a higher sulfur coal used? A higher sulfur coal would seemingly be less expensive and, with the assumed control efficiencies, could still provide less SO2 emissions.

2. EMERGENCY AMMONIA RELEASE

Response #8 (page RTC-66): The pressure relief valve on the ammonia storage tank will open discharge to the atmosphere when specified levels are exceeded. The ensuing ammonia vapor cloud would be controlled with water via a fire hose.

Concern: In 1992, an ice block plant in Pahokee, Florida had just such a release during filling of the ammonia tank. (The facility uses anhydrous ammonia as a refrigerant.) The pressure relief valve released a large cloud of ammonia which slowly dissipated. Several people were exposed to relatively high levels of ammonia, with two persons being admitted to the local hospital.

3. <u>ACTIVATED CARBON INJECTION (ACI)</u>

Response #12 (page RTC-73): The mercury control system will operate at a flue gas temperature of about 350°F.

Concern: Most of the literature on ACI indicates significantly increased control efficiencies when the operating temperature is kept below 300°F. Would it be possible for the applicant to reduce the operating temperature to this level without affecting performance or other control mechanisms?

Thank you for the opportunity to comment on these applications.

Sincerely,

For the Divisional Director
Environmental Science and Engineering

Jeffery F. Koerner

Jeffery F. Koerner, Engineer IV
Air Pollution Control Section

cc: AP-Source File

RECEIVED
DEC 23 1992
Division of Air
Resources Management

CARLTON, FIELDS, WARD, EMMANUEL, SMITH & CUTLER, P.A.
ATTORNEYS AT LAW

ONE HARBOUR PLACE	FIRSTATE TOWER	HARBOURVIEW BUILDING	FIRST FLORIDA BANK BUILDING	ESPERANTE'	BARNETT TOWER
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TAMPA, FLORIDA 33601	ORLANDO, FLORIDA 32802	PENSACOLA, FLORIDA 32582	TALLAHASSEE, FLORIDA 32302	WEST PALM BEACH, FLORIDA 33402	ST. PETERSBURG, FLORIDA 33731
(813) 223-7000	(407) 849-0300	(904) 434-0142	(904) 224-1585	(407) 659-7070	(813) 821-7000
FAX (813) 229-4133	FAX (407) 648-9099	FAX (904) 434-5366	FAX (904) 222-0398	FAX (407) 659-7368	FAX (813) 822-3768

PLEASE REPLY TO :
Tallahassee

VIA HAND DELIVERY

December 23, 1992

Mr. Willard Hanks
Florida Department of Environmental Regulation
2600 Blair Stone Road
Tallahassee, FL 32399-2400

Re: Flo-Energy and Sol-Energy Cogeneration Facilities;
AC50-219413, PSD-FL-196; AC50-219795, PSD-FL-197

Dear Mr. Hanks:

This law firm is assisting Flo-Energy, Inc., and Sol-Energy, Inc., with their efforts to obtain the PSD permits for their proposed cogeneration facilities in Palm Beach County, Florida. It is our understanding that you asked four questions about the permit applications during a recent telephone conversation with Mr. Gus Cepero. At Mr. Cepero's request, other representatives of Flo-Energy and Sol-Energy have prepared the following information for us to submit to you and the Department of Environmental Regulation (DER) in response to your questions.

1. Flue Gas Temperature.

You asked whether the electrostatic precipitators (ESP) could be redesigned to operate at temperatures below 350° fahrenheit because a 325°F temperature might improve the mercury capture efficiency of the ESP: Although it is theoretically possible to redesign the ESP, we believe it is unnecessary and inappropriate.

The boilers could be redesigned to produce a reduced exit flue gas temperature of 325°F, however, it would require an increase in the size of the economizers or the air heaters or both. These changes would involve additional expense. Further, the lower flue gas temperature will increase the potential for metal corrosion and thereby create future maintenance problems in the entire flue gas train. Lower flue gas temperatures also might adversely affect the plume rise, which could reduce dispersion of the flue gases and increase ground level concentrations.

Reducing the flue gas temperature to 325°F is not necessary in this case. The mercury emissions from the proposed facility will be extremely low because the mercury content of biomass fuels is extremely low. Moreover, the mercury control system proposed for the cogeneration facilities can effectively control mercury emissions at temperatures of 350°F.

The United States Environmental Protection Agency recently conducted tests with a carbon injection system at a resource recovery facility in Camden, New Jersey. The Camden facility has a dry scrubber and ESP, but tests were conducted with only the ESP in operation to evaluate the mercury removal capabilities of a carbon injection system at facilities that only have an ESP. EPA's tests demonstrated that mercury can be captured at 350°F at removal rates up to 98%.

EPA's tests indicate that the extremely low levels of mercury that may be emitted from the proposed facilities can be effectively controlled with the proposed carbon injection system. Conversely, we are not aware of any clear evidence to demonstrate that a reduction of 25°F will have any meaningful effect on the potential mercury emissions. Consequently, there is no reason to require the cogeneration facilities to incur additional capital and maintenance costs in an effort to achieve a slightly reduced flue gas temperature.

2. Overpile and Underpile Fuel Reclaimers.

Stored wood chips and bagasse will be reclaimed from the storage piles by means of either an overpile reclaimer or an underpile reclaimer for each boiler feed conveyor. Either reclaimer will be capable of reclaiming 100% of the required fuel. Therefore, it is highly improbable that both reclaimers would be operating at the same time.

The overpile reclaimer is normally a continuous chain conveyor with a series of relatively small slats that skim the fuel from the top of the storage pile and move it to the boiler feed conveyor. The chain conveyor is usually mounted on and supported by a structural arm cantilevered from a point adjacent to the boiler feed conveyor. The arm operates through a defined horizontal arc, commonly of 180° or more.

The underpile reclaimer is contained within a shallow concrete pit and normally consists of a continuous drag chain conveyor with a series of vertical slats that move the fuel horizontally to the boiler feed conveyor. A large pile of fuel is maintained over the reclaimer at all times, usually with the use of a front end loader or dozer.

3. <u>Ash Handling</u>.

Fly ash and siftings ash will be conveyed by means of an enclosed mechanical-type conveyor to an ash conditioner where the ash will be mixed with enough water to keep dust to a minimum at the conditioner outlet. The wetted ash will be discharged into a bin that will be enclosed with a roof and walls on three sides to reduce the exposure to the wind and rain. Ash will be removed through the open side of the bin while still in the wetted state and loaded into trucks that will be covered while en route to the point of proper disposal.

4. <u>Trucks Hauling Wood Chips</u>.

Trucks will be covered when hauling wood chips.

We hope this information satisfactorily answers the questions you have raised. Please call us at your earliest convenience if you need additional information.

We also ask you to distribute this information to any other person or agency that is interested in these issues.

Sincerely,

David S. Dee

DSD/nsh

cc: Mr. Gus Cepero
Mr. Don Schaberg
Mr. David Buff

OKEELANTA CORPORATION
6 MILES SOUTH OF SOUTH BAY
POST OFFICE BOX 86
SOUTH BAY, FLORIDA 33493

TELEPHONE: (407) 996-9072
TELEX: 803444

December 21, 1992

C. H. Fancy, P.E.
Chief, Bureau of Air Regulations
Florida Department of Environmental Regulation
Twin Towers Office Bldg.
2600 Blair Stone Road
Tallahassee, FL 32399-2400

Dear Mr. Fancy:

Enclosed is the Proof of Publication document for the Notice of Application published for the Flo-Energy, Inc. and the Sol-Energy, Inc. cogeneration plants in The Palm Beach Post. The notice was published as received from your office except that two clarifications were added concerning the generating capacities and the burning of No. 2 fuel oil and coal. The clarifications were agreed upon during a telephone conversation between Mr. Willard Hanks of your office and Mr. David Buff of KBN.

Sincerely,

Gus R. Cepero

Gus R. Cepero
Vice President

cc: David Buff, KBN w/ copy of Proof of Publication document.
W. Hanks
C. Holladay
D. Knowles, S Dist
S. Brooks, SE Dist

RECEIVED
DEC 23 1992
Division of Air
Resources Management

THE PALM BEACH POST

Published Daily and Sunday
West Palm Beach, Palm Beach County, Florida

PROOF OF PUBLICATION

STATE OF FLORIDA

COUNTY OF PALM BEACH

Before the undersigned authority personally appeared Chris Bull who on oath says that she/he is Class. Sales Mgr. of The Palm Beach Post, a daily and Sunday newspaper published at West Palm Beach in Palm Beach County, Florida; that the attached copy of advertising, being a Notice in the matter of application in the --- Court, was published in said newspaper in the issues of November 25, 1992

Affiant further says that the said The Post is a newspaper published at West Palm Beach, in said Palm Beach County, Florida, and that the said newspaper has heretofore been continuously published in said Palm Beach County, Florida, daily and Sunday and has been entered as second class mail matter at the post office in West Palm Beach, in said Palm Beach County, Florida, for a period of one year next preceding the first publication of the attached copy of advertisement; and affiant further says that she/he has neither paid nor promised any person, firm or corporation any discount, rebate, commission or refund for the purpose of securing this advertisement for publication in the said newspaper.

Chris Bull

Sworn to and subscribed before me this 25 day of November A.D. 19 92

NO. 595569

Notice of Application

The Department of Environmental Regulation announces receipt of two applications for permits to construct 74.9 megawatts or less cogeneration facilities in Palm Beach County. The facilities will burn bagasse, which is a sugar cane derivative, and clean wood waste as the primary fuels and No. 2 fuel oil and coal, not to exceed 25 percent on a heat input basis, as alternative fuels. One facility, Flo-Energy, Inc., will be located 6 miles south of South Bay off U.S. Highway 27. The other facility, Sol-Energy, Inc., will be located near the intersection of U.S. Highway 98 and Hatton Highway, Pahokee.

The applications are available for public inspection at the following Department offices: Bureau of Air Regulation, 2600 Blair Stone Road, Tallahassee, FL 32399-2400; South District, 2295 Victoria Avenue, Suite 364, Ft. Myers, FL 33901; South District, 1900 S. Congress Avenue, Suite A, West Palm Beach, FL 33406; and at the Palm Beach County Environmental Office located at 901 E. Evernia Street, West Palm Beach, FL 33402.

Anyone may send written comments of these applications to Mr. Preston Lewis at the Department's office in Tallahassee.

PUB: The Palm Beach Post
November 25, 1992