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**PILOT-SCALE LIMESTONE EMISSION
CONTROL (LEC) PROCESS: A DEVELOPMENT
PROJECT**

**Volume 1: Main Report and
Appendices A, B, C, and D**

OHIO COAL RESEARCH CENTER

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**PILOT-SCALE LIMESTONE EMISSION CONTROL (LEC) PROCESS:
A DEVELOPMENT PROJECT**

[Volume 1: Main Report and Appendices A,B,C, and D]

Final Report

by

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**PILOT-SCALE LIMESTONE EMISSION CONTROL (LEC) PROCESS:
A DEVELOPMENT PROJECT**

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1.0 EXECUTIVE SUMMARY

1.1 INTRODUCTION

ETS, Inc., a pollution consulting firm with headquarters in Roanoke, Virginia, has developed a dry, limestone-based flue gas desulfurization (FGD) system. This SO₂ removal system, called Limestone Emission Control (LEC), can be designed for installation on either new or existing coal-fired boilers. In the LEC process, the SO₂ in the flue gas reacts with wetted granular limestone that is contained in a moving bed. A surface layer of principally calcium sulfate (CaSO₄) is formed on the limestone. Periodic removal of this surface layer by mechanical agitation allows high utilization of the limestone granules.

The primary goal of the current study is the demonstration of the techno/economic capability of the LEC system as a post-combustion FGD process capable of use in both existing and future coal-fired boiler facilities burning high-sulfur coal.

A nominal 5,000 acfm LEC pilot plant has been designed, fabricated and installed on the slipstream of a 70,000 pph stoker boiler providing steam to Ohio University's Athens, Ohio campus. The pilot plant was normally operated on the slipstream of the Ohio University boiler plant flue gas, but also had the capability of operating at higher inlet SO₂ concentrations (typically equivalent to 3-1/2% sulfur coal) than those normally available from the flue gas slipstream. This was accomplished by injecting SO₂ gas into the slipstream inlet. The pilot plant was instrumented to provide around-the-clock operation and was fully outfitted with temperature, SO₂, gas flow and pressure drop monitors.

1.2 SUMMARY OF RESULTS

A total of over 90 experimental trials have been performed using the pilot-scale moving-bed LEC dry scrubber as a part of this research project with run times ranging up to a high of 125 hours. The primary variables studied were:

- Superficial flue gas velocity - 0.5 to 1.6 ft/s.
- Limestone bed velocity - 4.4 to 21.6 ft/hr.
- Inlet SO₂ concentration - 400 to >3000 ppm_{dv}.
- Water addition rate - 0 to 2.0 gallons/minute (includes both over-bed sprays and flue gas humidification water).

SO₂ removal efficiencies as high as 99.9% were achievable for all experimental conditions studied during which sufficient humidification was added to the LEC bed.

1.3 MARKETING AND COMMERCIALIZATION OF THE LEC

Retrofitting existing power plants is the quickest method to reduce power-plant emissions. The potential retrofit market for the LEC would include coal-fired boilers that use a +0.6% sulfur coal in conjunction with a Cold Side ESP with greater than 300 SCA. Using these parameters, the existing boiler population would consist of 243 units creating 70043 MWe of generating capacity (290 MWe on average). Greater than 60% of the unit capacity is owned by 25 utilities located in the Midwest, with more than half of these operating units located in Ohio, Michigan, Indiana, West Virginia and Pennsylvania.

The LEC process and conventional limestone scrubbing have been compared on an equitable basis using flue gas conditions that would be expected at the outlet of the electrostatic precipitator (ESP) of a 500 MW coal-fired power plant. The equipment list for the LEC case was obtained by conceptually scaling up the current LEC moving-bed pilot plant at Ohio University. The equipment list for the wet limestone scrubber case was obtained from an EPRI technical report (Keeth et al., 1986). Equipment costs for both the LEC and the wet limestone scrubber were generally obtained using empirical estimating charts.

The LEC was found to have a definite economic advantage in both direct capital costs and operating costs. The LEC equipment capital cost of \$12,290,000 represented a 48% savings when compared to the wet limestone scrubber process equipment cost of \$23,534,000. On an absorption tower basis, a capital cost advantage of 46% for the LEC occurs due to its simpler mechanical design. Operating costs are also smaller for the LEC process since a slurry is not required, resulting in a lower water consumption. The solids material handling associated with the LEC requires less electricity than that of the slurry handling system required by the wet limestone scrubber.

Based on the success and findings of the present project, the next step in LEC process development will be a full-scale commercial demonstration unit. Sales presentations to utility and industrial sites for a demonstration unit have already been initiated. A commercial demonstration for the LEC has recently been licensed in Taiwan (see Appendix A).

Licensing agreements with two major air pollution control suppliers have already been negotiated. The first agreement was with Proceadair Industries, which is the North American subsidiary of the highly respected industrial furnace manufacturer, Stein Heurtey. Proceadair Industries was granted the right to sell

the LEC system to the North American industrial market. Proceaire Industries is very active in the North American acid gas scrubbing market and has demonstrated market presence and commitment evidenced by multi-year contract values ranging from \$300,000 to \$30,000,000.

The second agreement was granted to U-Tah Industrial Company, Ltd. This agreement granted U-Tah the exclusive rights to sell the LEC process throughout Taiwan, R.O.C. Taiwan's FGD market is expected to reach \$2.8 billion (US) by 1997. U-Tah was reported to have 32% share of Taiwan's industrial air pollution market in 1991.

Copies of press releases reporting the two licensing agreements as well as a copy of an LEC promotional brochure are included in Appendix A.

1.4 WASTE PRODUCTS

The waste product of the LEC process is a dry powder containing the reaction products $\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$ and $\text{MgSO}_4 \cdot 7\text{H}_2\text{O}$, as well as unreacted limestone. Except for water content, the LEC waste product contains compounds identical to those found in the wastes produced by a conventional limestone wet scrubber. Costs and options for the disposal of limestone wet scrubber wastes are given by Keeth et al. (1983, 1991).

1.5 EFFECT ON OHIO COAL USE

At present, high-sulfur Ohio coals are used to satisfy less than 50% of the state's annual utility coal requirements. Clean Air Act regulations will likely reduce the usage of Ohio coals even further. Therefore, there is a near-term need for a practical and economical means of allowing existing utility and industrial boilers to burn high-sulfur Ohio coals.

The results of this project have shown that the Limestone Emission Control (LEC) process, when installed on an industrial or utility boiler, offers the potential for the environmentally sound and inexpensive burning of Ohio coal. The following results from the project are significant:

- (1) The results have successfully demonstrated the feasibility of scaling up the LEC process. The LEC has now been demonstrated to be a highly effective flue gas desulfurization technique at the bench, small pilot and large pilot scales. This is important since the reproducibility of the results at the three different scales will provide the justification and motivation for further scale-up to the commercial level.

- (2) A commercial-scale design of the LEC, based upon the results of this project, has been shown to offer significant cost advantage over conventional spray drying and wet scrubbing while offering equal or superior desulfurization performance.

2.0 INTRODUCTION

2.1 PROBLEM STATEMENT

ETS, Inc., a pollution consulting firm with headquarters in Roanoke, Virginia, has developed a dry, limestone-based flue gas desulfurization (FGD) system. This SO₂ removal system, called Limestone Emission Control (LEC), can be designed for installation on either new or existing coal-fired boilers. In the LEC process, the SO₂ in the flue gas reacts with wetted granular limestone that is contained in a moving bed. A surface layer of principally calcium sulfate (CaSO₄) is formed on the limestone. Periodic removal of this surface layer by mechanical agitation allows high utilization of the limestone granules.

2.2 PROJECT GOAL AND OBJECTIVES

The primary goal of the current study is the demonstration of the techno/economic capability of the LEC system as a post-combustion FGD process capable of use in both existing and future coal-fired boiler facilities burning high-sulfur coal. The specific objectives of the study include:

- Demonstrate the LEC capability of operating at high inlet SO₂ concentrations typical of those encountered in the burning of a 3.5 wt% sulfur coal.
- Generate long-term operational information that would provide optimization parameters for a full-scale demonstration project.
- Develop correlations for on-stream time and performance changes for both SO₂ removal and particulate removal efficiency.
- Demonstrate limestone regeneration techniques.
- Analyze both the waste material and its disposal/resale options.
- Provide an economic analysis/comparison of the LEC process with respect to current FGD systems.

2.3 PROJECT CO-SPONSORS

All large pilot plant scale experimental work was performed using a moving-bed LEC system installed on a slipstream of Ohio University's coal-fired boiler steam plant located in Athens, Ohio. Ohio University was the prime contractor with the OCDO and was responsible for the overall project administration and interface with OCDO. Ohio University also made the equipment and resources of the Department of Chemical Engineering Coal Research Center Laboratory available to the project.

ETS, Incorporated was the prime sub-contractor to Ohio University for the project. ETS provided the moving-bed pilot plant design, test program design, and program implementation. ETS also contributed the pilot plant and all associated instrumentation for the duration of the project. In addition, pilot plant installation, project management, and many technical consultation costs were absorbed by ETS.

2.4 STATEMENT OF WORK SUMMARY

A nominal 5,000 acfm LEC pilot plant will be designed, fabricated and installed on the slipstream of a 70,000 pph stoker boiler providing steam to Ohio University's Athens, Ohio campus. The pilot plant will normally operate on a slipstream of the Ohio University boiler plant flue gas, but will have the capability of operating at higher inlet SO_2 concentrations (typically equivalent to 3-1/2% sulfur coal but up to the equivalent of 6% sulfur coal) than those normally available from the flue gas slipstream. This will be accomplished by injecting SO_2 gas into the slipstream inlet. The pilot plant will be instrumented to provide around-the-clock operation and will be fully outfitted with temperature, SO_2 , gas flow and pressure drop monitors.

The primary objective of this study is to demonstrate the techno/economic capability of the LEC as a post-combustion flue gas desulfurization system suitable for application to both existing and future boiler facilities burning, or considering burning, Ohio coal. Additional objectives will be to obtain long-term operational information on the LEC system and provide system optimization which will be the basis for a full-scale demonstration unit. The program will include the following secondary objectives.

1. Develop correlations between on-stream time and performance changes with respect to SO_2 removal efficiency and particulate removal (both mass and fractional efficiency).

2. Develop maintenance and operating cost records over an extended period of time.
3. Demonstrate limestone renewal techniques.
4. Analyze waste material and conduct disposal alternatives assessment.
5. Economic analysis and comparison with existing FGD systems.

It is anticipated that the design, fabrication, installation and start-up of the pilot plant will take twelve months. During the design and fabrication phase, it is proposed that the small (400 cfm) LEC used in the research program be left in place and operated periodically to provide process optimization support to the demonstration design. A follow-on test and operation program of eight months on the 5,000 acfm pilot plant is proposed. The tasks necessary to complete this program include:

- Task 1 - Program Management.** Overall management of the program will be provided in this task.
- Task 2 - Design.** A decision on LEC process configuration (fixed- versus moving-bed reactor) will be made as part of this task. An LEC pilot unit will be designed based on the selected process configuration. If the moving-bed design appears to be most appropriate, a small moving-bed LEC reactor will be fabricated, installed on a boiler slipstream, and tested prior to final design of the 5,000 acfm unit. Engineering drawings showing the details of the pilot plant layout, duct work, system fan, LEC reactor, stone conveyors, grinders and fill hoppers will be completed during this phase.
- Task 3 - Pilot Plant Fabrication.** Using the detailed design information obtained from Task 2, all necessary equipment will be procured. Any equipment to be fabricated, such as the LEC reactor, will be completed during this task.
- Task 4 - Pilot Plant Installation.** The 5,000 acfm LEC pilot plant will be installed at the Ohio University steam plant.
- Task 5 - Start-Up and Shakedown.** During this task, the large pilot plant will be started up and debugged. All operating equipment will be checked out during this phase. All test methodologies and analytical equipment will be checked. Continuous monitoring instrumentation will be debugged.

Task 6 - Test Program and Data Acquisition. The LEC pilot plant will be operated and tested for a period of eight months. Anticipated variables to be studied include: limestone bed residence time, flue gas throughput velocity, limestone bed temperature, SO₂ inlet concentration, flue gas relative humidity, flue gas temperature, limestone type, limestone regeneration frequency, limestone regeneration duration, et cetera.

Task 7 - Data Acquisition and Final Report. A draft report will be prepared and submitted to OCDO for comment. This report will comprehensively detail the results of the program. Based on OCDO comments, the draft report will be amended and a final report will be issued.

2.5 TECHNICAL PRESENTATIONS

2.5.1 Presentations: This Project

Prudich, M.E.; Appell, K.W.; and J.D. McKenna, "A Final Report on the Moving-Bed Limestone Emission Control (LEC) Pilot Plant Program," *1993 EPRI/EPA/DOE SO₂ Control Symposium*, Boston, MA, August 24-27, 1993.

Visneski, M.J.; Appell, K.W.; and M.E. Prudich, "A Status Report on the Moving-Bed Limestone Emission Control (LEC) Process Pilot Plant," *AIChE 1992 Summer National Meeting*, Minneapolis, MN, August 9-12, 1992.

Prudich, M.E.; Reddy, S.N.; Appell, K.W.; and M.J. Visneski, "A Pilot Demonstration of the Moving-Bed Limestone Emission Control (LEC) Process," *85th Annual Meeting of the Air & Waste Management Association*, Kansas City, MO, June 21-26, 1992.

Prudich, M.E.; Reddy, S.N.; Appell, K.W.; and J.D. McKenna, "A Pilot Demonstration of the Moving Bed Limestone Emission Control (LEC) Process," *1991 EPRI/EPA/DOE SO₂ Control Symposium*, Washington, DC, December 1991.

2.5.2 Presentations: Related Topics

Duespohl, D.; Sampson, K.J.; and M.E. Prudich, "Optimization of a Granular Limestone Moving-Bed Dry Scrubbing Reactor," *Tenth Annual International Pittsburgh Coal Conference*, Pittsburgh, PA, September 20-24, 1993.

Maldei, M.; Prudich, M.E.; and K.J. Sampson, "The Effect of Limestone Composition and Properties on the Performance of a Granular Limestone Flue Gas Scrubbing Process," *Tenth Annual International Pittsburgh Coal Conference*, Pittsburgh, PA, September 20-24, 1993.

Ben-Said, L.; Prudich, M.E.; and K.J. Sampson, "Fundamental Studies of Low Temperature SO₂ Removal by Dry Calcium-Based Sorbents," *Tenth Annual International Pittsburgh Coal Conference*, Pittsburgh, PA, September 20-24, 1993.

Visneski, M.J.; Prudich, M.E.; and K.J. Sampson, "Modeling of the Low Temperature Reaction of Sulfur Dioxide and Limestone Using a Three Resistance Film Theory Instantaneous Reaction Model," *1990 AIChE Annual Meeting*, Chicago, IL, November 1990.

Reddy, S.N.; Prudich, M.E.; and K.J. Sampson, "A Mathematical Simulation of ETS' Limestone Emission Control (LEC) Process Using a Moving Bed Configuration," *1990 AIChE Annual Meeting*, Chicago, IL, November 1990.

Unger, M.E.; Adler, R.J.; and M.E. Prudich, "Recovery of Spent Limestone in LEC Flue Gas Desulfurization," *Seventh Annual International Pittsburgh Coal Conference*, Pittsburgh, PA, September 1990.

3.0 TECHNICAL DISCUSSION

3.1 LEC BASICS

The Limestone Emission Control (LEC) process is a unique system employing standard quarry-sized limestone to remove SO₂ from coal-fired boiler flue gases. In the LEC process, hot flue gases (<350°F) are contacted with a bed of 1/32 to 1/4 inch limestone granules covered with a thin film of water. Sulfur dioxide is absorbed from the flue gas into the water film where it subsequently reacts with dissolved limestone. A layer of reaction products, primarily calcium sulfate and calcium sulfite, forms on the surface of the limestone granules as the reaction proceeds.

3.2 HISTORY OF LEC DEVELOPMENT

In the early 1970s, the U.S. Department of Energy initiated bench-scale testing of dry limestone for flue gas desulfurization. Shale and Stewart (1979) reported that SO₂ removal efficiencies of greater than 90% for extended periods of time were possible, and that the limestone granules were regenerated easily with mild agitation. This early work resulted in U.S. Patent No. 3,976,747 entitled "Modified Dry Limestone Process for Control of Sulfur Dioxide Emissions" (Shale, 1976).

ETS, Inc. began development work on the present LEC system in mid-1982. In June 1983, ETS and the U.S. EPA agreed that the LEC was potentially a simple, low cost FGD system. It was agreed to proceed along three paths: (1) ETS performed a series of bench-scale slip stream tests to verify the basic SO₂ emission reductions. Results showed greater than 90% SO₂ removal; (2) EPA/IERL requested that TVA do a preliminary economic analysis of the LEC system. The resulting analysis was in agreement with preliminary ETS economic studies showing that the LEC process offers significant economic advantage over both spray drying and wet scrubbing systems; and (3) EPA/IERL initiated an in-house laboratory bench-scale study of the LEC process aimed at verification of the ETS slip stream tests. Although Step 3, EPA/IERL's in-house work, was not formally published, ETS arrived at two significant conclusions based on the EPA bench-scale testing program. First, EPA's bench-scale tests using simulated flue gases adequately substantiated ETS' slip stream test results and Shale's early work at DOE. And second, although regeneration of the spent limestone was demonstrated, the regeneration was not accomplished as easily as that reported by Shale and Stewart, i.e., more vigorous grinding was required.

In the fall of 1986 the Ohio Coal Development Office awarded a research grant to Ohio University to design, install and operate a small, semi-batch fixed-bed

LEC pilot plant (400 acfm) on a slip stream of flue gas from the 70,000 lb/hr stoker boiler providing steam to the Ohio University campus. Figure 3-1 shows a simplified flow schematic of the fixed-bed LEC pilot unit. The flue gas first entered a spray chamber where it was conditioned to the desired temperature and humidity. The conditioning was achieved by adding water via an atomizing nozzle and/or by injecting live steam into the chamber. The conditioned gas then entered the LEC reactor, passing downward through a fixed-bed of limestone. The reactor contained a removable basket filled with limestone. The basket was 22 inches square and permitted the limestone bed depth (the bed dimension in the direction of flue gas flow) to be varied between 0 and 24 inches. The scrubbed flue gases were then drawn out from the bottom of the limestone bed by an induced draft fan and returned to the duct leading to the stack.

Figure 3-1. Simplified schematic of the fixed-bed LEC process.

Over 100 experimental runs were conducted on the LEC fixed-bed pilot unit between May 28, 1987 and December 18, 1987 (Prudich et al., 1988; Appell, 1989; Visneski, 1991). Three different Ohio limestones covering the full range of Ohio limestone compositions were used in these tests. The program studied the effects of varying limestone bed depth (from 6 to 18 inches), superficial flue gas velocity (from 1.0 to 2.0 feet per second), flue gas humidity, flue gas SO₂ concentration, and inlet flue gas temperature. All three of the limestones used in the study were able to achieve greater than 90% SO₂ removal in the LEC process. Many tests showed removals of up to 99%.

The fixed-bed LEC unit was shown to perform effectively for inlet SO₂ concentrations in the flue gas ranging from 500 ppm to 3500 ppm. At an inlet SO₂ concentration of 700 ppm, a 6 inch deep LEC bed was able to sustain removals of over 90% for 16 hours and removals of over 50% for more than 24 hours (Figure 3-2). Even at 3500 ppm inlet SO₂, the 6 inch LEC bed (the smallest bed tested) was able to sustain an SO₂ removal efficiency of greater than 90% for about 1 hour and a removal efficiency of greater than 50% for more than 2 hours.

Figure 3-2. LEC fixed-bed experimental results. Continuous flue gas flow/batch limestone bed. LEC bed deactivation due to surface blinding. [Run data from Prudich et al. (1988) with a limestone bed depth of 6 inches and a superficial gas velocity of 1.0 ft/s.]

Two factors were identified which limit LEC operation. These were (1) drying of the limestone surface and (2) build-up of the reaction product on the reactive limestone surface. A dry limestone surface is effectively non-reactive under LEC processing conditions. The build-up of reaction product on the limestone surface results in a decrease in the reaction rate and eventually renders the limestone inactive.

Two operational modes were used to overcome deactivation by drying. Operating the process with a saturated flue gas would halt evaporation and thereby prevent the limestone bed from drying out. An over-bed water spray could also be used to keep the bed wet. In order to keep the bed from drying out, the spray rate would have to be equal to or greater than the rate of evaporation.

Deactivation due to surface blinding cannot be avoided but it can be overcome through the removal of the reaction products. After the limestone became non-reactive due to the blinding of the surface, the solids still retained a core of unreacted limestone. The layer of reaction products could be removed by mild attrition. In the small pilot plant studies the reaction products were removed from the reacted limestone through off-line attrition in a stirred ball mill. After attritting, the stone was returned to service. When the surface of the stone was again blinded this process could be repeated until all of the limestone had been utilized. The reactivated stone showed the same SO_2 removal activity as the virgin limestone.

Three methods were used to supply water to the LEC process. One method was to prewet the bed of limestone before starting the process. This was accomplished by pouring water onto the bed before sealing the reactor. The life of the bed was determined by the amount of conditioning of the incoming flue gas. The higher the relative humidity of the gas the longer the time before the limestone bed became dry and therefore unreactive. The second method was by condensation. Normally, on start up, the limestone bed would be cooler than the incoming flue gas. As the gas contacted the cold bed, water would condense from the gas phase onto the limestone. The amount of water deposited by condensation would depend upon the conditioning of the flue gas. As the relative humidity increased the amount of water deposited onto the limestone would increase. Condensation would slow and eventually cease as the bed was heated by the incoming flue gas. After condensation stopped, the life of the bed was once again determined by the condition of the incoming flue gas. The final method of water introduction was by spraying water directly onto the limestone bed. As shown in Figure 3-1, spray nozzles were positioned directly over the bed. The life of the bed was determined by the rate of the water spray as well as the condition of the flue gas.

Figure 3-3 shows experimental data taken from the LEC pilot unit that illustrates both deactivation due to drying and surface blinding. Runs 870714A', 870716A, 870716B, 870717A, 870720A and 870720B represent consecutive runs using the same 12 inch deep bed of limestone. The superficial gas velocity remained constant from run to run at 1.5 ft/s but the SO₂ concentrations ranged from 620 to 870 ppm. The inlet flue gas relative humidity was approximately 50% but varied slightly from run to run. The variation of SO₂ concentrations and gas humidities was due to the fluctuating conditions of the incoming flue gas. Run 870714A' started with a prewet bed of limestone. After a little over 1.5 hours the bed became unreactive. The same bed of limestone was then rewet before starting run 870716A. The bed regained its initial SO₂ removal efficiency before once again losing its reactivity due to drying. The process of rewetting the bed before starting the next run continued for the remaining four runs. Starting with run 870717A, rewetting the bed failed to return its activity to its original level. This decrease in activity can be attributed to the blinding of the surface due to the build-up of the reaction products on the reactive surface.

Figure 3-3. Deactivation of an LEC bed due to surface drying and surface blinding. [Run data from Prudich et al. (1988) with a limestone bed depth of 12 inches and a superficial gas velocity of 1.5 ft/s.]

Deactivation due solely to surface blinding is illustrated in Figure 3-2. These data show the performance of three different 6 inch deep beds of limestone. The inlet flue gas was saturated and therefore drying of the bed was not a factor in the decline of the SO₂ removal efficiency of the limestone. Run 871217C is a continuation of 871216B and uses the same bed of limestone. Run 871105A continued using the same bed of limestone started by run 871103A. As can be seen from Figure 3-2 each bed lost activity at different times. The life of the bed was inversely proportional to the product of the SO₂ concentration of the incoming flue gas and the flue gas flow rate. Run 871103A/871105A with an inlet SO₂ concentration of 344 ppm lasted the longest. Next was run 871214B with an inlet SO₂ concentration of 702 ppm. Run 871216B/871217C with an inlet SO₂ concentration of 736 ppm had the shortest life. The inlet flue gas flow rate was held constant for all three sets of runs.

Besides the difference in operating conditions, the limestone used for the runs shown in Figure 3-2 differs from the limestone used for the runs shown in Figure 3-3.

The details of LEC development chronology since 1987 are given in Appendix B.

3.3 PILOT PLANT DESIGN

3.3.1 Prepilot Fixed-Bed Study

Summary results of the prepilot fixed-bed LEC study can be found in section 3.2 - HISTORY OF LEC DEVELOPMENT, above.

3.3.2 Prepilot Moving-Bed Study

As a result of the successful completion of the small pilot-scale fixed-bed LEC project, a project proposal was submitted to OCDO for the purpose of scaling up the LEC process to a nearer to commercial scale. As a result of this proposal, in the spring of 1989, the Ohio Coal Development Office (OCDO) granted a development award to Ohio University to design, install and operate a continuous LEC pilot plant.

The continuous LEC pilot plant will have a nominal capacity of about 12-1/2 times (5,000 acfm) the original fixed-bed LEC pilot plant and will be able to operate in a continuous mode (the original fixed-bed LEC pilot plant could only be operated in a semi-batch mode). The pilot plant will be instrumented to provide around-the-

clock operations and will be fully outfitted with temperature, SO_2 , gas flow and pressure drop monitors.

Two process configurations will be considered for the LEC pilot unit: (1) a cycling fixed-bed LEC reactor and (2) a moving-bed LEC reactor. These two configurations are to be evaluated as a part of the design task. The configuration that is found to be the most promising both economically and operationally will be selected for subsequent installation and evaluation. If the moving-bed LEC reactor appears to be the most promising configuration a small moving-bed pilot unit will be fabricated, installed and tested on a boiler slip stream prior to the final design and fabrication of the 5000 acfm unit. This additional experimental work is required since the moving-bed reactor configuration is a radical departure from the original fixed-bed LEC design. The small-scale pilot testing serves to reduce the risk in the demonstration-scale design.

3.3.3 Pilot Plant Design Decision

The three process alternatives to be evaluated include a cycling fixed bed, a chevron moving bed, and a louvered moving bed. These three design alternatives are shown in Figures 3-4, 3-5, and 3-6, respectively. Early in the design phase it was decided to pursue a moving bed design if such a design could be shown to be operationally feasible. Of the two moving bed designs selected, the chevron moving bed (Figure 3-5) seemed to offer the most advantages, a compact design and good moisture level control. However, experimentation with the chevron moving bed design on a 400 acfm pilot unit pointed out a critical disadvantage. The chevron internals in the chevron moving bed provided surfaces from which bed plugging could occur. The louvered moving bed design (Figure 3-6) which is now under investigation at the small pilot-scale avoids the plugging problem by operating the moving-bed in a cross-flow mode without the aid of bed internals. The louvered bed design is, however, less compact than the chevron bed design.

Over 348 hours of experimental results documenting moving-bed LEC performance have been acquired to date. Experimental results on the 400 acfm small moving-bed pilot-unit located in Roanoke, VA are summarized in Appendix C. All experimental runs performed prior to 01/01/90 were made using the chevron moving bed while all experimental runs made after that date were made using the louvered moving bed design.

A quick perusal of the experimental results show that a majority of the runs meet the minimum 90% SO_2 removal efficiency required of the LEC unit.

Figure 3-4. Cycling-bed reactor.

(a) Side view.

(b) Front view.

Figure 3-5. Chevron-bed reactor.

Figure 3-6. Louvered-bed reactor (side view).

Table 3-1

QUALITATIVE EVALUATION OF LEC DESIGN ALTERNATIVES:
ADVANTAGES/DISADVANTAGES

A.   Cycling fixed bed (Figure 3-4)

Advantages

- (i)   Easy to scale up (based on fixed-bed LEC data).

Disadvantages

- (i)   Mechanically complex.
- (ii)   Involves a great deal of solids handling.

B.   Chevron moving bed (Figure 3-5)

Advantages

- (i)   Compact design.
- (ii)   Intrabed water sprays provide good control of bed moisture content.

Disadvantages

- (i)   Bed internals may cause bed plugging.

C.   Louvered moving bed (Figure 3-6)

Advantages

- (i)   Mechanically simple.
- (ii)   No bed internals to promote plugging.

Disadvantages

- (i)   Not as compact as the chevron design.
- (ii)   Water can only be introduced at a bed face. Control of bed moisture may not be as good as chevron design.

3.4 MOVING BED PROCESS DESCRIPTION

A continuous moving-bed LEC pilot plant sized to handle 5000 ACFM has been installed on the slip stream of the 70,000 lb/hr stoker boiler providing steam to the Ohio University campus. A simplified flow scheme of the moving-bed LEC process is shown in Figure 3-7. A detailed schematic of the pilot plant is given in Figure 3-8. Limestone is added to the bed via a 32 foot bucket elevator which is supplied with limestone from the feed hopper, the make-up hopper, the vibrating screen, or the recycle chute by means of a feed screw. The make-up hopper is used to compensate for limestone lost due to reaction with SO_2 and its subsequent removal as reaction product. The limestone bed itself has dimensions of 14 inches x 36 inches x 128 inches (Figure 3-9). The depth of the bed in the direction of the flue gas flow is 14 inches. The LEC reactor internals consist of inlet and outlet louvers, overhead bed sprays and outlet screening. The inlet and outlet louvers (as originally designed) are four inches long and are inclined at a 75° angle with a $3/8$ inch overlap between louvers. The flue gas inlet and outlet plenums on the sides of the reactor are sized such that there is uniform distribution of the gas across the whole bed. The flue gas is drawn from the duct leading to the stack from the electrostatic precipitator. The superficial velocity of the flue gas through the LEC reactor can be varied from 0.3 to 1.6 feet per second. The live hopper at the top of the bed is provided to prevent the leakage of ambient air into the outlet gas plenum. An induced draft fan is used to draw the processed flue gas from the outlet plenum. The processed flue gas is directed to the stack.

Figure 3-7. Simplified schematic of the LEC process.

Limestone moves vertically downward through the bed with a velocity sufficient to prevent bridging of the limestone which may occur when the flue gas dries the bed. The limestone removed from the bottom of the bed is sent to the discharge screw via a 19 foot bucket elevator. The discharge screw has three outlets, any one of which may be opened. The first outlet directs the limestone to the recycle chute which leads directly to the feed screw and back to the reactor. The second outlet leads to the reactivation system. The build-up of reaction products on the surface of the limestone tends to retard the rate of reaction per unit granule surface area. To prevent this surface blinding from rendering the limestone inactive, the limestone granules are passed through

a reactivation device that removes the reaction product layer by either dry or wet abrasion. This abrasion exposes a fresh layer of reactive limestone. This mixture of reactivated limestone and abraded material is then passed through a vibrating screen where the abraded material is removed as a waste. The reactivated limestone is then directed to the feed screw and returned to the top of the LEC bed. The third discharge screw outlet is directed to the waste hopper. Normal LEC operation directs the limestone through the reactivation system.

The SO_2 concentration in the inlet flue gas can be varied from 500 to 3000 ppm. Early experimental runs using this pilot plant have shown that the LEC process has a capability of removing more than 98% of the SO_2 contained in the inlet flue gas. The LEC process is considered to be potentially attractive for enabling coal-fired power plants to meet stringent SO_2 emission standards.

Figure 3-9. Limestone Bed Geometry for Moving-Bed LEC Pilot Plant.

3.5 DISCUSSION OF RESULTS

3.5.1 Initial Moving-Bed Experimental Results

A total of 30 experimental runs were conducted between May 29, 1991 and July 25, 1991 using the LEC moving-bed pilot unit. The length of each experimental run ranged from 30 minutes to 290 minutes. The primary variables studied were:

- Superficial flue gas velocity - 0.5, 1.1, and 1.6 ft/s.
- Limestone bed velocity - 4.4 ft/hr, 13.7 ft/hr, and 21.6 ft/hr (these limestone bed velocities correspond to in-bed limestone residence times of 2.4 hr, 0.8 hr, and 0.5 hr).
- Inlet SO₂ concentration - 500 to 1100 ppm_{dv}.
- Water addition rate - 0 to 2.0 gallons/minute (includes both over-bed sprays and flue gas humidification water).

The limestone used during the LEC demonstration project was a Maxville limestone from Maxville Quarries, Inc., Logan, Perry County, Ohio. An elemental analysis of this limestone [ASTM, Part 05.05, Method D4326-84] is given in Table 3-2. Additional typical analyses and information about this limestone seam are given by Lamborn (1951).

As mentioned earlier, the SO₂/CaCO₃ reaction is a liquid-phase reaction and therefore a thin layer of water must be present on the limestone granules to allow significant SO₂ removal to occur. The run philosophy to date has been to operate the LEC bed so that the limestone exiting the bed is essentially dry. This is accomplished by controlling the position of the drying front that exists in the limestone bed (see also Figure 4-4). As the limestone approaches the bottom of the bed, the drying front moves closer to the outlet side of the reactor. This results in there being less available wet limestone for SO₂ removal. Although bed drying causes relatively poor SO₂ removal values in the lower portion of the bed, the overall average bed removal remains high because of the excellent performance of the top and middle portions of the bed.

The results obtained from the initial experimental trials have demonstrated SO₂ removals in excess of 90% and, often, SO₂ removals in excess of 99%. Performance plots for three experimental trials (910723A, 910724A, and 910725A; all with inlet SO₂ concentrations of about 1000 ppm) are shown in Figures 3-10, 3-11, and 3-12. These Figures also show the influence of the mechanical difficulties that have inhibited long-term operation during the initial

Table 3-2

MAXVILLE LIMESTONE ANALYSIS

<u>Elemental Component*</u>	<u>Weight Percent</u>
Calcium oxide, CaO	61.50
Magnesia, MgO	9.65
Potassium oxide, K ₂ O	0.63
Sodium oxide, Na ₂ O	0.13
Silica, SiO ₂	17.91
Alumina, Al ₂ O ₃	2.73
Ferric oxide, Fe ₂ O ₃	4.64
<u>Moisture (as received)</u>	0.23
<u>LOI** @ 1000°C (dry basis)</u>	38.15
<u>Ca/Mg Mole Ratio</u>	4.6:1

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 * Ignited basis.

** Loss On Ignition.

period of operation. The reactor outlet design of the LEC moving-bed pilot plant relied only on the outlet louver to retain the limestone in the moving bed. As it has turned out, the original outlet louver design did not accomplish this purpose. During operations, limestone was drawn from the bed into the outlet plenum and subsequently into the I.D. fan. In order to prevent this limestone loss from the bed and to protect the I.D. fan, a screen was attached to the inside (limestone bed side) of the outlet louvers. The screen was used to keep the limestone in the bed. However, the presence of the screen introduced another operational problem. Damp fines were deposited on the screen, plugging the screen and significantly increasing the pressure drop across the limestone bed. The blinding of the screen occurred most rapidly at the top of the bed (normally the most active part of the bed). Due to the screen blinding at the top of the bed, the flue gas would begin to flow preferentially through the bottom portions of the bed (normally the least active part of the bed). This action would result in a simultaneous increase in pressure

drop across the bed, decrease in SO₂ removal performance, and decrease in flue gas flow rate. This behavior can be observed in Figures 3-10, 3-11, and 3-12.

Figure 3-10. LEC moving-bed experimental results. Trial number 910723A. AASHTO No. 9 limestone. Inlet SO₂: 1100-1200 ppm_{dv}.

Figure 3-11. LEC moving-bed experimental results. Trial number 910724A. AASHTO No. 9 limestone. Inlet SO₂: 880-1000 ppm_{dv}.

Figure 3-12. LEC moving-bed experimental results. Trial number 910725A. AASHTO No. 9 limestone. Inlet SO₂: 750-850 ppm_{dv}.

After the initial run series, the outlet louvers and outlet plenum were redesigned. The redesign modifications for these systems were made during a September/December 1991 shutdown. The new outlet louvers were eleven inches long with a 3-1/2 inch overlap between louvers. The new and old outlet louver designs are shown in Figure 3-13. Results from shakedown runs performed in February 1992 have indicated that limestone loss from the reactor bed into the outlet plenum was reduced by several orders of magnitude.

It has been observed that significant limestone surface abrasion occurs within the LEC bed itself. In traveling downward through the LEC bed and in transport by the screw conveyors, the limestone particles seem to provide most of the granule on granule abrasion necessary to reactivate the limestone. In cases where this effect may be fully exploited, the external reactivating attritor (equipment item #5 in Figure 3-8) may be eliminated from the process flow sheet.

Figure 3-13. Old and new outlet louver designs used in the LEC moving-bed reactor.

3.5.2 Enlarged Louver Design Experimental Results

A total of 60 experimental runs were conducted from May 5, 1992 through November 24, 1992. These experimental trials ranged up to 134 hours in length. The variables studied, primarily with respect to overall SO₂ removal efficiency, bed pressure drop and stone carry-over, were:

- mean granule diameter.
- superficial gas velocity.
- inlet SO₂ concentration.
- water addition location.

3.5.2.1 *Effect of Mean Granule Diameter.* The four granule diameters that were studied in the LEC moving-bed were:

- AASHTO No.9 (D_p = 0.10 inches).
- AASHTO No.8 (D_p = 0.28 inches).
- 50/50_(volume basis) No.8/No.9 mixture (D_p = 0.20 inches).
- AASHTO No.9 pre-screened with a No.8 mesh screen (D_p = 0.12 inches).

Typical particle size distributions for the two major limestone sizes [AASHTO No. 8 and No. 9] used in LEC processing are shown in Figure 3-14. The No.9 limestone has the smallest mean diameter of the four types used and therefore has the largest surface area available for reaction. Unfortunately, the diameter of the limestone granule is inversely related to its minimum fluidization velocity. At smaller particle diameters this can result in a portion of the limestone becoming fluidized by the flue gas stream and, subsequently, exiting the LEC bed through the outlet louvers. At superficial flue gas velocities greater than 0.75 ft/s, excessive limestone carry-over was observed to occur. In order to reduce the fluidization potential of the stone, limestone sorbent granules with increased mean diameters were studied.

The next larger commercially available size stone is AASHTO No.8 (D_p = 0.28 inches). With the relatively larger particle diameter of the No.8 stone, the carry-over problem was greatly reduced, even at superficial flue gas velocities in excess of 1.2 ft/s. The improved material handling characteristics of the No.8 limestone enabled the first extended trial of the moving-bed system (Figure 3-15). Run 920430 lasted for over 134 hours at an average superficial flue gas velocity of 1.1 ft/s and achieved SO₂ removal efficiencies of up to 85 percent. However, subsequent experimental trials (LEC Runs 920601 and 920602), also using No. 8 limestone, that were operated at lower superficial flue gas velocities (0.3 ft/s) with a 100 percent bed moisture zone yielded SO₂ removals of only 40% (average). With the enlarged mean diameter of the No.8 stone, the loss in available reaction

Figure 3-14. Typical particle size distributions for the AASHTO No. 8 and No. 9 limestones used in LEC processing.

surface area resulted in substantially lower SO_2 removals when compared to No. 9 limestone. The low SO_2 removals resulting from the use of No. 8 limestone are unacceptable in terms of operating a viable commercial system. This low removal more than negates the No. 8 stone carry-over and material handling advantages. Increasing the LEC bed depth beyond 14 inches would add reactive surface area and might compensate for the disadvantages of the No. 8 limestone.

The next stone size used in an attempt to reduce stone carry-over and still achieve +90% SO_2 removal was a 50/50 mixture by volume of No.8 and No.9 limestones. This mixture had an average particle diameter of 0.20 inches (LEC Runs 920604 and 920605). The resulting SO_2 removal was approximately 10% greater than that achieved using No.8 stone alone, but the limestone carry-over was still excessive. Therefore the combination of the two stones yielded a mixture that exhibited the worst traits of both the No. 9 (high limestone carry-over) and No. 8 (low SO_2 removal) stone sizes.

The last method investigated in order to reduce the stone carry-over while at the same time maintaining high SO_2 removals consisted of scalping (removing the

Figure 3-15. Long-term LEC operation using AASHTO No. 8 limestone. Run No. 920430. Inlet SO_2 : 800-1400 ppm_{dv}.

finer from) the No.9 stone with a No.8 mesh screen (0.094 inches). This was done because the theoretical fluidization velocity of a size distribution is weighted towards the higher individual fluidization velocities of the smaller particles within the particle size distribution. Thus, by scalping off the smallest 40% of the No.9 limestone granules, the fluidization velocity of the remaining stone would be increased while at the same time maintaining a larger reaction surface area than that available from the No.8 limestone. The resulting mean particle diameter of the screened stone was 0.12 inches. The screened limestone was used in LEC Runs 920810, 920811 and 920812. These experimental trials yielded poor results since there was no measurable decrease in limestone carry-over when compared to No.9 stone ($D_p = 0.10$ inches) although SO_2 removals were generally reduced to less than 90%.

The overall results of varying the mean particle diameter in order to achieve an optimum stone size that would limit the amount of limestone carry-over into the outlet plenum while maintaining high SO_2 removals was unsuccessful. The carry-over problem must be solved through changes in the mechanical design.

3.5.2.2. *Effect of Superficial Gas Velocity.* The superficial flue gas velocity affects three dependent variables in the LEC system:

- (1) Reaction Theory.
- (2) Stone Carry-Over.
- (3) Bed Pressure Drop.

The reaction kinetics theory with respect to superficial gas velocity in the LEC system has already been extensively dealt with by both Visneski (1991) and Reddy (1992) and their general conclusions have been validated by the moving-bed project.

3.5.2.2.1. *Stone Carry-Over.* It has been empirically determined that flue gas superficial velocities greater than 0.75 ft/s result in excessive limestone carry-over when using No.9 stone ($D_p \approx 0.10$ inches). Fluidization of the limestone particles is believed to take place in the 75° outlet louver slot since the superficial flue gas velocity is more than doubled as it passes through the outlet louvers. The excellent SO₂ capture performance of an experimental trial (LEC Run 920613) operated processing No. 9 stone and at a superficial flue gas velocity of 0.31 ft/s is shown in Figure 3-16. The use of No.8 stone with an average granule diameter of 0.28 inches allows superficial flue gas velocities of up to 1.2 ft/s without excessive stone carry-over at the expense of the loss of reactive surface area (at constant bed depth).

3.5.2.2.2 *Bed Pressure Drop.* Bed pressure drop is related to the square of the superficial gas velocity and can be quantified through the Ergun Equation. This is illustrated by LEC Run 921020, during which the bed pressure drop was observed to be less than 2 in.H₂O at a superficial flue gas velocity of 0.31 ft/s. After approximately 26 hours into the trial, the superficial flue gas velocity was increased to 0.47 ft/s which resulted in an increase in pressure drop to approximately 7 in.H₂O.

Bed pressure drop can also be affected by obstructions within the bed itself, as in either blocked void spaces in the stone or within the louvers themselves. This is shown in Figure 3-17 (LEC Run 921117/921118) where at roughly 30 hours into the trial, the gas flow rate begins to decrease while the bed pressure drop is simultaneously increasing. This phenomenon was caused by fouled limestone bed outlet louvers.

3.5.2.3. *Effect of Inlet SO₂ Concentration.* All of the LEC experimental trials were conducted using flue gas obtained from the slip stream of the Ohio University heating plant. The heating plant generally burns a 2wt% sulfur coal. This resulted in the majority of the experimental runs receiving less than 1500 ppm_{dv} of SO₂ at the inlet of the bed. In order to simulate a utility firing its boilers with a 3.5%

Figure 3-16. Lower superficial velocity LEC operation using AASHTO No. 9 limestone. Inlet SO₂: 1000 - 1500 ppm_{dv}.

sulfur coal, the inlet gas stream was spiked with 99.8% SO₂ gas from portable cylinders. This yielded inlet SO₂ concentrations in excess of 3000 ppm_{dv} for LEC Runs 920716, 920718 and 920719. The results of these three trials (Figure 3-18) showed that at SO₂ concentrations of 3000 ppm_{dv}, the 14 inch deep LEC bed was fully capable of removing >95% of the SO₂ entering the scrubber. However, whenever the moisture (damp limestone) zone was reduced in order to allow the exiting stone to remain dry enough for effective regeneration via the vibrating screen (between hours 4 to 5, 6.5 to 8.5, and 11.5 to 13.5), the SO₂ removal efficiency dropped dramatically down to approximately 70%. In terms of sorption theory, this would indicate that at these high inlet SO₂ concentrations the SO₂ breakthrough curve resides within (but only barely) the fully wetted 14 inch limestone bed. This means that the SO₂ breakthrough curve resides close to the outlet side of the bed even at low superficial gas velocities (0.3 ft/s). Therefore,

Figure 3-17. Results of LEC run number 921117/921118. AASHTO No. 9 limestone. Inlet SO_2 : 500-650 ppm_{dv}.

careful water addition is critical in order to ensure a maximum possible moisture zone within the bed when operating at elevated SO_2 concentrations. If high SO_2 levels are to be encountered in a commercial application, then a limestone bed depth in excess of 14 inches would be recommended.

3.5.2.4. Effects of Water Addition Location. The initial location of the water sprays were at the top of the upper pressure seal. It was determined that the bulk of the water added to the system from this location was being entrained in the gas stream and removed from the upper portion of the bed before it could contribute to the SO_2 removal efficiency of the system. To solve this problem, a pair of water nozzles were moved to the inlet plenum adjacent to the top louver. This forced the bulk water phase to pass through the entire bed depth before any of the excess moisture could exit through the outlet louvers and into the outlet plenum. This change also reduced the amount of water required to achieve a sufficient moisture zone since it reduced the amount of liquid water being pulled into the outlet plenum. It also eliminated the thick water film zone that formed under the overhead sprays due to the artificially elevated water addition rate. The thick water layer caused a higher pressure drop in the top portion of the bed, resulting in

Figure 3-18. LEC SO₂ removal efficiency at high inlet SO₂ concentrations. AASHTO No. 9 limestone. Inlet SO₂: 3500 - 3900 ppm_{dv}.

gas-channeling towards the less reactive, lower (dry) portion of the bed resulting in poor SO₂ removal due to the lack of wetted active surface and reduced gas residence time. The gas channeling lead to an increased total bed pressure drop. The combination of changing the water spray location and lowering the water addition rate resulted in a more evenly distributed thin water film in the bed and reduced the potential of gas channeling which are required in order to achieve + 98% SO₂ removal efficiencies.

3.5.3 Regeneration/Attrition/Limestone Utilization

The limestone regeneration (actually limestone surface reactivation through the removal of reaction products) system, as originally conceived for the LEC process, was to consist of an attriting/grinding unit followed by a screen separator (units #4 and #5 in Figure 3-8). All material, including both reaction products and unreacted limestone, that passes through the screen used in the screen separator will be rejected as waste. A maximum theoretical limestone utilization can be estimated based on the assumption that when a limestone particle is reacted down such that its unreacted core is smaller than the screen opening size, its unreacted core will be discarded as waste. The maximum theoretical limestone utilization therefore depends both on the screen opening size used in the screen separator and the particle size distribution of the limestone fed to the LEC process. Tables 3-3 and 3-4 given the estimated maximum theoretical limestone utilization versus separator screen size for both AASHTO #8 and AASHTO #9 limestone particle size distributions. Limestone sample LS920805 was used as the "typical" No. 8 limestone while limestone sample LS920617 was used as the "typical" No. 9 limestone stone in making these calculations. Since a 6 mesh screen was typically used during experimental trials feeding AASHTO No. 8 limestone while a 16 mesh screen was typically used during experimental trials feeding AASHTO No. 9 limestone, maximum theoretical limestone utilizations would be expected to be 66.3% and 74.0%, respectively.

Even though the original process concept included an attritor (item #5 in Figure 3-8) for the removal of reaction products from unreacted limestone surface, actual operation of the pilot unit showed that the limestone received sufficient attrition to reactivate the stone while passing through the process loop. Therefore, the attritor was removed from the limestone flow path. The screen separator was still necessary to remove the powdered reaction products from the more massive unreacted limestone particles. Qualitative observations by operating personnel have indicated that limestone attrition in the screw conveyors was excessive and, as a result, attrited and powdered unreacted limestone as well as the reaction products. These qualitative observations are born out by calculations of limestone utilization based on sulfur content of the collected powder (waste material rejected by the screen separator). Theoretically, this powder should contain 14.59 weight percent sulfur if it is fully reacted. [This calculation is based on a parent limestone with a 4.6:1 $\text{CaCO}_3:\text{MgCO}_3$ mole ratio containing 15.6 wt% inerts and reacting to $\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$ and $\text{MgSO}_4 \cdot 7\text{H}_2\text{O}$.] Typical limestone utilizations for both AASHTO No. 8 and AASHTO No. 9 limestones are presented in Tables 3-5 and 3-6, respectively.

Maxville AASHTO No. 8 limestone showed utilizations ranging from 20.9 to 58.5% which is significantly less than the 74.0% theoretical maximum utilization. The Maxville AASHTO No. 9 limestone showed utilizations ranging from 8.4 to

16.1%, once again much less than the theoretical maximum utilization of 66.3%. This indicates that there was excessive limestone attrition in both cases. It is our opinion that this excessive attrition can be eliminated by reducing the LEC reactor bed speed (increasing the limestone residence time in the LEC reactor) and by replacing the screw conveyors used on the pilot unit with belt conveyors in the commercial design.

Table 3-3

THEORETICAL MAXIMUM LIMESTONE UTILIZATION VERSUS SEPARATOR
SCREEN SIZE FOR AASHTO NO. 8 LIMESTONE (LS920805)

Regeneration Screen Size	Screen Opening Size, mm	Theoretical Maximum Utilization, %
5/16"	8.0	6.4
1/4"	6.3	22.0
3.5M	5.60	33.2
4M	4.75	47.8
5M	4.00	61.5
6M	3.35	74.0
8M	2.36	88.7
10M	2.00	92.6
12M	1.70	95.0
14M	1.40	96.8

Table 3-4

**THEORETICAL MAXIMUM LIMESTONE UTILIZATION VERSUS SEPARATOR
SCREEN SIZE FOR AASHTO NO. 9 LIMESTONE (LS920617)**

Regeneration Screen Size	Screen Opening Size, mm	Theoretical Maximum Utilization, %
5M	4.00	0.76
6M	3.35	2.6
8M	2.36	13.7
10M	2.00	28.7
12M	1.70	42.5
14M	1.40	56.3
16M	1.18	66.3
20M	0.850	77.8
25M	0.710	82.9
30M	0.600	86.4
35M	0.500	89.0
40M	0.425	90.8
45M	0.355	92.3
50M	0.300	93.3
60M	0.250	94.1

Table 3-5

REPRESENTATIVE LIMESTONE UTILIZATIONS FOR AASHTO NO. 8 LIMESTONE

(Separator screen size = 6 mesh; Bed speed = 1460 lb/hr;

Theoretical maximum utilization = 74.0%)

Sample	Sulfur Content, wt%	Limestone Utilization, %
9205010830	3.05	20.9
9205021100	3.75	25.7
9205031100	6.80	46.6
9205041135	8.54	58.5
9205050108	4.70	32.2
9205052148	6.18	42.4

Table 3-6

REPRESENTATIVE LIMESTONE UTILIZATIONS FOR AASHTO NO. 9 LIMESTONE
 (Separator screen size = 16 mesh; Bed speed = 2707 lb/hr;
 Theoretical maximum utilization = 66.3%)

Sample	Sulfur Content, wt%	Limestone Utilization, %
9210221930	1.64	11.2
9210301315	1.77	12.1
9211041730	2.27	15.6
9211091532	2.35	16.1
9211111550	2.03	13.9
9211111652	2.07	14.2
9211171742	1.22	8.4
9211181946	2.04	14.0
9211191400	2.24	15.4
9211241658	2.12	14.5

3.6 CONCLUSIONS

The LEC moving-bed pilot plant project has demonstrated its capability as a high-performance SO₂ removal system. SO₂ removals as high as 99.9% have been achieved for extended periods at inlet flue gas flow rates of up to 3200 acfm. Inlet flue gas SO₂ concentrations ranging up to >3000 ppm have been successfully tested. This shows that the chemistry behind this road-grade limestone-based technology is valid.

Mechanical difficulties, i.e., limestone blow-over and exit screen plugging initially made long-term LEC operations difficult. Outlet louver modifications made over the September/December 1991 shutdown seem to have reduced these mechanical problems.

The next step in the commercialization of the LEC process is the demonstration of the technology on an electric utility boiler as either a stand-alone desulfurization device or as an in-situ insert into the last field of an existing ESP. A conceptual schematic of a generic compact moving-bed LEC is shown in Figure 3-19. A conceptual schematic of a compact moving-bed LEC inserted into the space occupied by the last field of an ESP is shown in Figure 3-20.

Figure 3-19. Schematic of conceptual compact LEC bed. (a) Top view. (b) Cross-sectional face view. Arrows indicate flue gas flows. Shading indicates limestone bed. Flue gas is distributed and collected through louvered plenums.

Figure 3-20. Conceptual schematic of an LEC occupying the space of the last field in an electrostatic precipitator.

4.0 MATHEMATICAL MODEL FOR THE MOVING-BED LEC PROCESS

4.1 PROCESS THEORY

The following section consists of a summary description of the process model for the moving-bed Limestone Emission Control (LEC) process. A detailed description of the balance and correlative equations used in formulating the model as well as the computer techniques used to solve the modeling equations is given by Reddy (1991).

In the LEC process a thin film of water on the limestone particles is required for SO_2 removal. A simplified representation of the sorbent particle is shown in Figure 4-1 (Appell (1989)). In the SO_2 removal process, the SO_2 is absorbed by the thin film of water on the limestone particle and is then transported in the form of ions to the reaction front. The limestone dissolves in the water layer and is transported to the reaction front where it reacts with the absorbed SO_2 . This ionic reaction produces calcium sulfite/sulfate which precipitates onto the limestone particle and then inhibits limestone dissolution.

Visneski (1991) has discussed in detail the film theory representation of the concentration profiles of SO_2 and limestone. For the development of this model it has been assumed that the calcium sulfite/sulfate precipitate layer has no effect on SO_2 removal. The schematic for the concentration profiles of SO_2 and limestone is given in Figure 4-2.

Hence, the film theory model of SO_2 capture kinetics show that the possible rate controlling steps are (i) mass transfer of SO_2 across the gas film in contact with the water layer, (ii) absorption of SO_2 into the water layer, (iii) ionic dissociation of SO_2 in water, (iv) transport of SO_2

Figure 4-1. Conceptual Representation of the Sorbent Particle in the LEC Process.

through the water layer to the reaction front, (v) dissolution of the limestone into the water layer, (vi) ionization of the limestone, (vii) transport of the dissolved limestone through the water layer to the reaction front, and finally (viii) the rate of reaction between the dissolved limestone and the dissociated SO_2 .

Figure 4-2. Film Theory Representation of the Concentration Profiles of Sulfur Dioxide and Limestone During Reaction.

The rate limiting step is determined by various factors such as the relative speeds of the flue gas and the limestone bed, the temperature of reaction, the water content of the limestone bed, and the humidity of the gas stream. The reaction rate is taken to be instantaneous since it is known that ionic reactions occur with great rapidity. It is also assumed that SO_2 and limestone ionize instantaneously. Hence, the rate controlling steps for SO_2 removal are reduced to the mass transfer rate of SO_2 across the gas film (gas-phase control), the diffusion rate of the species through the water layer to the reaction front (liquid-phase control), and the dissolution rate of limestone (solid-phase control).

4.2 MODELING ASSUMPTIONS

In the first generation modeling done by Appell (1989) of the fixed-bed LEC process, one aspect which was not considered is that the water does not necessarily drain off the bed instantaneously. Water trickles down the bed at finite velocities when the water content of the bed exceeds its saturation concentration. This effect is taken into consideration in the development of this model. Even though low flow rates are conducive to partial wetting, it is assumed for the sake of simplicity that the limestone granules are uniformly wetted.

Several of the assumptions used in the development of the flow model for the moving bed LEC process have been used in the fixed-bed LEC process modeling work done by Appell (1989) and Visneski (1991).

The overall assumptions on which the flow model is based are listed below.

- 1) The system is isobaric at atmospheric pressure.
- 2) The ionic dissociation of the species and the ionic reaction of the species is instantaneous.
- 3) All limestone particles are spherical and have the same size. A sphericity constant could be used to describe the sphericity.
- 4) The limestone particles are completely wetted when water is present on the particles.
- 5) The water layer on the limestone bed has a uniform thickness within any given differential element.
- 6) The water flow is in the trickling flow regime.
- 7) The temperatures of the liquid film and the trickling water are equal to the temperature of the solids.
- 8) The superficial gas velocity is constant.
- 9) The flows of the gas phase, solid/liquid phase and the trickling water phase are not deflected by the force of the other flows.
- 10) The gas phase, solid/liquid phase and the trickling water move in plug flow with no axial dispersion.
- 11) Calcium sulfite/sulfate precipitation does not affect the mass and enthalpy balances.
- 12) The heat- and mass-transfer areas are equal.

4.3 MODEL DEVELOPMENT

The procedure followed in developing the steady-state flow model of the moving bed LEC process is similar to that followed by Appell (1989). The flows of the flue gas and the limestone bed are unidirectional with the limestone bed moving vertically down and the flue gas moving horizontally across the limestone bed in a cross-flow pattern. The differential element used to develop the model is shown in Figure 4-3.

Figure 4-3. Differential Element Used In Model Development.

The solid phase (limestone), the liquid phase (water adsorbed on the bed) and the trickling water phase (if present) are lumped into one solid/liquid phase and hence the mass and energy balances are based on only two phases, viz. the gas phase and the lumped solid/liquid phase.

The direction of sulfur dioxide mass transfer is from the gas phase to the solid/liquid phase in the presence of adsorbed water. In the absence of water there is no change of SO_2 concentration in the gas phase. Mass transfer of water can

occur in either direction (from gas phase to solid/liquid phase or vice versa) depending on whether the water is condensing from the gas phase or is evaporating from the solid/liquid phase. The transfer of water between the two phases adds to the sensible heat transfer between the two phases.

4.4 PROCESS MODEL RESULTS

The baseline conditions chosen for the computer simulation of the moving-bed LEC process are based on runs made at the LEC pilot plant installed on the slipstream of the stoker boiler providing steam to the Ohio University campus. These baseline conditions for the computer simulation are listed in Table 4-1.

Table 4-1. Simulation Conditions.

Limestone Bed Geometry: Bed Depth : 14 inches [direction of flue gas flow] Bed Height: 128 inches [direction of limestone flow] Bed Width : 36 inches [perpendicular to both flows]
Flue Gas Flow Rate: 3000 ACFM
Inlet Gas Conditions: 210°F 2 % Water Vapor (by volume) 750 ppm SO ₂
Superficial Limestone Velocity: 14.23 ft/hr
Inlet Limestone Conditions: 77°F 0 % liquid water saturation (Dry Bed) 0.5 GPM overhead spray

The superficial limestone velocity was chosen on the basis of the tachometer reading indicating the speed of the bed twin screws (see Figure 3-8). Low limestone speeds were found to create operational problems such as caking. The maximum limestone flow which the screws and bucket elevators could handle was at the tachometer reading of 20 which corresponds to 14.23 ft/hr of dry limestone.

The primary topics of interest for the modeling study were the limestone bed drying and the average SO_2 removal of the bed. As has been explained earlier the presence of liquid water is essential for the removal of SO_2 . It is also desirable that the limestone be dry at the bottom of the bed since this ameliorates the handling of the stone. Figure 4-4 gives the water content of the bed and the outlet SO_2 concentration profile under the simulation conditions. It can be seen that as the limestone bed moves down, the drying front moves laterally across the bed. Correspondingly, due to the lack of water for SO_2 removal, the outlet SO_2 concentration goes up as the limestone moves down through the reactor. The profile of the drying front appears to be linear in spite of the slightly distorted image produced due to the scale variance between the x-axis and y-axis. The outlet SO_2 concentration profile is not linear. This is due to the fact that the rate of SO_2 removal has a non-linear relationship with the water content of the bed. This is more clearly depicted in Figure 4-5 which gives the contours for various SO_2 concentrations throughout the bed. The gap between the contours is pretty regular until the 150 ppm contour, but there is a larger gap between the 150 ppm and 0 ppm contours. This is due to the fact the SO_2 removal follows an S-shaped curve and at lower SO_2 concentrations the rate of SO_2 removal approaches zero.

Figure 4-4. Example drying front and outlet SO_2 concentration profiles.

Figure 4-5. SO₂ Contours Through the Bed.

It should be noted that although the SO₂ removal at the bottom of the bed is very poor, what matters is the overall SO₂ removal capacity of the bed. The variation of the average SO₂ concentration (over the entire bed height of 128 inches) with the bed depth is given in Figure 4-6. As can be seen from Figure 4-4 the first few inches of the bed in the direction of the gas flow are primarily dry and hence the SO₂ removal is very low. At greater depths of the bed the liquid water concentration increases and henceforth SO₂ removal goes up. The shape of the overall SO₂ concentration curve shown in Figure 4-6 is of the form of a S-shaped curve. This would have been more evident for bed depths greater than 14 inches.

Figure 4-6. Average SO_2 Outlet Concentration Profile.

5.0 MARKETING DISCUSSION

Competition in the utility and industrial FGD market will be fierce in the advent of increasingly stringent SO₂ regulations. The annual global FGD market is expected to exceed \$7 billion by 1996. For a new FGD system to enter this expanding market place, it must have the following four advantages over existing FGD systems in order to be a viable commercial competitor:

- (1) Generate a dry product for disposal.
- (2) Use limestone as the main reactant.
- (3) Minimal overall mechanical complexity.
- (3) Achieve reasonably high SO₂ removal efficiencies with good reactant utilization.

The LEC system meets the above requirements and is suitable for both new utility and industrial boilers or as a retrofit system.

5.1 MARKET OPPORTUNITIES

5.1.1 Global Market

The current global market for air pollution control (APC) systems for stationary sources is approximately \$13 billion, with the U.S. holding almost one-third of this market. The U.S. APC equipment market should increase to 35% or 38% of the estimated \$18 billion that the global market could reach by 1995. These estimates do not include possible eastern European countries just entering the free market place.

The other expanding APC market exists in Asia, where mainland China, in particular, is in the process of constructing significant coal-fired electrical power capacity in order to take advantage of their vast coal reserves.

5.1.2 U.S. Retrofit Market

Retrofitting existing power plants is the quickest method to reduce power-plant emissions. The potential retrofit market for the LEC would include coal-fired boilers that use a +0.6% sulfur coal and in conjunction with a Cold Side ESP with greater than + 300 SCA. Using these parameters, the existing boiler population would consist of 243 units creating 70043 MWe of generating capacity (290 MWe on average). Greater than 60% of the unit capacity is owned by 25 utilities

located in the Midwest, with more than half of these operating units located in Ohio , Michigan, Indiana, West Virginia and Pennsylvania.

5.2 ECONOMIC ANALYSIS OF THE MOVING-BED LEC SYSTEM

5.2.1 Background

The LEC process and conventional limestone scrubbing have been compared on an equitable basis using flue gas conditions that would be expected at the outlet of the electrostatic precipitator (ESP) on a 500 MW coal-fired power plant. The equipment list for the LEC case was obtained by conceptually scaling up the current LEC moving-bed pilot plant at Ohio University. The equipment list for the wet limestone scrubber case was obtained from an EPRI technical report (Keeth et al., 1986). Equipment costs for both the LEC and wet limestone scrubber were obtained using empirical estimating charts. Background information on the LEC/ limestone scrubbing cost comparison can be found in Appendix D.

The LEC was found to have a definite economic advantage in both direct capital costs and operating costs. The LEC equipment capital cost of \$12,290,000 represented a 48% savings when compared to the wet limestone scrubber process equipment cost of \$23,534,000. On an absorption tower basis, a capital cost advantage of 46% for the LEC occurs due to its simpler mechanical design. Operating costs are also smaller for the LEC process since a limestone slurry is not required. This results in a lower water consumption for the LEC process. The solids material handling associated with the LEC requires less electricity than that of the slurry handling system required by the wet limestone scrubber. Table 5-1 lists the operating and equipment costs associated with each process area for both FGD systems along with the total annualized costs. A similar economic analyses has been performed by the Tennessee Valley Authority, the United States Department of Energy, and Babcock & Wilcox. A comparison between the LEC system and four commercial technologies on the basis of cost per ton of SO₂ removed is given in Table 5-2.

Table 5-1: Summary of results of a comparative economic analysis of an LEC and a wet limestone scrubber (LSS) for a 500 MW facility.

RAW MATERIAL AND UTILITY CONSUMPTION	LEC	LSS
Limestone - tons/hr	32.7	33.5
Quicklime - tons/hr	0	2.5
Water - gpm	249	710
Steam - lb/hr	222,000	222,000
Operating Power - hp	8,769	13,058
Waste - tons/hr	86	130
EQUIPMENT COSTS - \$1000 (1989)		
Area 1 - Material Handling	1,316	1,486
Area 2 - Feed Preparation	0	1,805
Area 3 - Gas Handling	3,897	3,326
Area 4 - SO ₂ Absorption	3,716	11,690
Area 5 - Stack Gas Reheat	2,524	2,524
Area 6 - Waste Handling/Reactivation	837	2,703
Total Equipment Costs	12,290	23,534
ANNUALIZED COSTS - \$1000 (1989)		
Total Operating and Maintenance	24,445	34,349
Total Costs	27,872	40,601

Table 5-2: Levelized cost estimate, in \$ per ton SO₂ removed, based on a 4% sulfur coal.

PLANT SIZE	LEC	Commercial Technologies			
		LIMESTONE SCRUBBER	LIME SPRAY DRYING	WELLMAN LORD	LIME DUAL ALKALI
100 MW	\$442	\$736	\$620	\$752	\$659
350 MW	\$317	\$521	\$510	\$499	\$506
500 MW	\$296	\$480	\$487	\$448	\$475
750 MW	\$277	\$439	\$463	\$399	\$444

5.2.2 Competition from Evolving Technologies

Some analysts have projected that the post-combustion FGD market will be dominated by the NOXSO and NaTec systems. The NOXSO system is expected to be used primarily in new installations, while the NaTec process is a retrofit technology. The LEC system can be used in both new installations and as a retrofit system.

The LEC system has the following considerable advantages over the NOXSO and NaTec technologies:

- (1) Smaller system size and is more compact than NOXSO.
- (2) Process is less complex and simpler to operate than NOXSO.
- (3) Lower waste disposal cost than NaTec due to the LEC's non-hazardous waste compared to NaTec's hazardous waste product.
- (4) More easily retrofittable than NOXSO.
- (5) Limestone is more readily available than the Nahcolite used in NaTec.
- (6) Limestone costs \$15/ton compared to Nahcolite at \$220/ton.

The LEC also has a cost advantage when compared to the NOXSO and NaTec systems as shown in Table 5-3. The LEC system capital costs for a range of power plant sizes have been estimated to be from 50 to 170 \$/kw. The capital costs for the NOXSO system for the same power plant sizes have been estimated to be 300 to 500 \$/kw. The control costs for the LEC and the NaTec systems have been estimated to be 160 to 400 \$/kw and 500 to 800 \$/kw, respectively. Thus, the LEC more than favorably compares to the NOXSO and NaTec systems that have been projected to dominate the future FGD market.

Table 5-3: Cost comparison of the LEC to the NaTec and NOXSO systems.

	<u>LEC</u>	<u>NaTec</u>	<u>NOXSO</u>
Capital Costs (\$/kw)	50 - 170	70 - 120	300 - 500
Control Cost (\$/Ton SO ₂)	160 - 400	500 - 800	NA
SO ₂ Removal (%)	+ 95	70	+ 95
Status	Pilot	Field Demo	Pilot

5.3 FUTURE DEVELOPMENT STEPS

Based on the success and findings of the present project, the next step will be a full scale commercial demonstration unit. Sales presentations to utility and industrial sites for a demonstration unit have already been initiated. The commercial demonstration for the LEC is currently under consideration in Taiwan.

Licensing agreements with two major air pollution control suppliers have already been negotiated. The first was Procedaire Industries, which is the North American Subsidiary of the highly respected industrial furnace manufacturer, Stein Heurtey. Procedaire Industries was granted the right to sell the LEC system to the North American industrial market. Procedaire Industries is very active in the North American acid gas scrubbing market and has demonstrated market presence and commitment evidenced by multi-year contract values ranging from \$300,000 to \$30,000,000.

The second agreement was granted to the U-Tah Industrial Company, Ltd. U-Tah obtained the exclusive rights to sell the LEC process throughout Taiwan, R.O.C. Taiwan's FGD market is expected to reach \$2.8 billion (US) by 1997. U-Tah was reported to have 32% share of Taiwan's industrial air pollution market in 1991.

6.0 COMMERCIAL DESIGN

The proposed commercial LEC design incorporates the basic moving-bed reactor design and operational parameters used in the pilot plant project by placing a series of moving-beds in parallel. Each moving-bed reactor consists of a series of vertical inlet and outlet louvers at the bulk gas/solid interface and a variable speed twin screw conveyor located at the bottom of the limestone bed to facilitate stone movement. The moving beds are contained within a single reactor housing with common gas and material handling equipment servicing the individual beds. The key operational parameters for the commercial design were defined based on the pilot plant work done at Ohio University.

6.1 REACTOR DESIGN

The moving-bed reactor is designed to handle a superficial gas velocity of 0.75 fps using AASHTO No. 9 road grade limestone. The bed depth is 15" and is measured in the direction of the gas flow from the inlet louver to the outlet louver.

6.1.1 Reactor Shell

The reactor housing shell that contains the multiple moving-beds will be fabricated in welded, shippable sections of 1/4" 304 stainless steel that can be bolted together on-site. The louvers are then to be welded to the shell. To provide maintenance and inspection to the twin screw conveyors, 12" by 18" access panels are to be located above the conveyors. The shell housing is to be stiffened and flanged as required.

6.1.2 Louver and Slot Design

The inlet and outlet louvers allow the flue gas to pass horizontally across the 15" reactor bed depth while, at the same time, giving the necessary structural support to the packed vertical bed. The louvers are to have a minimum construction of 1/8" width 304 stainless steel. The inlet and outlet louvers have an identical configuration in the commercial design and are to be sloped at 70° to the horizontal and are 11 inches long with 8 inch centers.

The outlet louvers allow a certain amount of stone carry-over. Stone carry-over is the stone that becomes partially fluidized within the outlet louvers and subsequently exits the packed-bed. This carry-over of stone is a function of the superficial gas velocity passing through the bed. The recommended maximum superficial velocity is 0.75 fps. Exceeding this gas rate will cause excessive limestone and entrained water carry-over which will reduce the SO₂ removal efficiency and may damage down-stream process equipment.

Upon exiting the outlet louvers, the carry-over limestone then enters the outlet gas-streams channel or "slot" where the stone falls out of suspension due to reduced gas velocity within the slot. The gas velocity within the slot should be between 500 and 1500 fpm (feet per minute). The slot width is used to regulate the gas velocity through the slot with a minimum slot width of 6". This can be increased to 7" or 8" in order to limit the slot velocity.

At the bottom of each slot section, a single 6" screw conveyor with a rotary air lock is used to remove the stone carry-over. This conveyor should have direct drive motor and a 2 tph (ton per hour) capacity. The stone is transported through the screw conveyor into the rotary air lock; thus the air lock requires the capacity and motor drive to match the screw conveyor.

The conveyor troughs shall be formed of 1/8" thick 304 stainless steel with flanged connections for bolting to the bottom of the reactor shell. The clearance between the screw conveyor and the trough is a minimum of 1/2 inches.

6.1.3 Spray Nozzles

Single phase spray nozzles will be located at the top of the upper pressure seal and if necessary within the inlet plenum. The nozzles will be of Spray Systems manufacture or equal and are required to provide a fine, atomized mist of water over a reasonable pressure range. A sufficient number of sprays must be used to ensure both a full cover of the reactor cross sectional area and water supply. The nozzles and internal piping will be made of stainless steel.

6.2 MATERIAL HANDLING EQUIPMENT

6.2.1 Moving-Bed Twin Screw Conveyors

The twin-screw conveyors are located at the bottom of each reactor unit and provide the means of controlling the extraction of the stone from the vertical packed bed. This conveyor will be equipped with gear reduction boxes and a variable speed control with a 20 to 1 turn down ratio that will allow a maximum throughput capacity of 5 tph. The motor for each set of screw conveyors should be rated at a minimum of 7.5 hp (220VDC).

The conveyor troughs shall be formed of 1/8" thick 304 stainless steel with flanged connections for bolting to the bottom of the reactor shell. The clearance between the screw conveyor and the trough is a minimum of 1/2 inches.

6.2.2 Belt Conveyors

Belt conveyors will be used in material handling when conveying the crushed limestone in a horizontal direction at a maximum rate of 50 tph. The crushed stone has an approximate bulk density of 100 lbs/ft³ and can contain relatively large amounts of surface moisture. A small amount of moisture on the stone can result in very poor material handling characteristics; thus resulting in the need for service walkways and belt wipers at the head pulleys. Dust covers for the belt conveyors are also necessary.

A belt tripper will be used on top of each reactor in order to ensure even loading of stone into the top pressure seal. A belt tripper consists of a manually controlled motorized cart that rides on a pair of rails that traverse the width of the reactor. The cart has an extra set of pulleys that elevates the belt and allows the stone to gravity feed through an attached chute.

6.2.3 Bucket Elevators

Chain-type bucket elevators will be supplied to handle the crushed limestone at a maximum capacity of 50 tph. The elevators will also be equipped with an overlapping arrangement of steel buckets. A top service platform with access via a cages ladder will be provided with each elevator.

6.3 SCREENING AND REGENERATION

Regeneration of the flue-gas exposed limestone is accomplished by the inter-particle abrasion that occurs both during stone movement within the bed and during material handling. To remove the abraded fines that are produced, a SWECO twin deck or equal vibrating screen separator with a 18,000 lbs/hr capacity will be provided. All contact surfaces will be fabricated out of 304 stainless steel except the molded plastic dust cover. The unit will be fitted with market grade screens ranging from 1/4" to 30 mesh and will be equipped with a 2.5 hp, 1200 rpm, 230/460/60 motor.

6.4 GAS HANDLING

The gas handling requirements of the LEC provided by ETS International consists only of the insulated inlet and outlet plenums. Installation is very important in being able to maintain a steady-state moisture zone inside of the bed.

The installation of a plenum sub-girt system is to be welded to the inlet and outlet plenum stiffeners. The sub-girt will consist of 1/8" thick angles and 1/8" by 3" bars for attachment of the insulation board. The insulation consists of 4" thick Owens-Corning Insul-Quick board or equal that is attached to the support bars with

weld pins. The insulation board will be applied in 2" layers with staggered joints. The insulation is then covered with unpainted 0.032 stucco embossed, 4" rib aluminum panels. All of the aluminum will be flashed and will include closure strips in order to prevent water in-leakage. The maximum temperature for this inlet plenum of this design would be 500°F.

7.0 FINAL BUDGET SUMMATION

The Final Budget Summation presented below represents final expenditures by contractor by task for the Limestone Emission Control project. The numbers in **bold** represent the budget numbers taken from Grant Amendment #5 (letter from Kathi S. Jones to Michael Prudich, August 12, 1992). The numbers in *italics* are as spent numbers as calculated by the principal investigator.

7.1 TOTAL BUDGET

7.1.1 Total Project Budget

	Phases (Tasks)							
	<u>1</u>	<u>2</u>	<u>3</u>	<u>4</u>	<u>5</u>	<u>6</u>	<u>7</u>	<u>Total</u>
Personnel	40811 51131	60161 70800	13301 13301	4922 16163	9040 7256	48638 69111	6675 3475	183548 231237
Equipment/ Supplies	12 12	172 172	195720 201549	--- -0-	--- -0-	9934 11070	2346 -0-	208184 212803
Services	--- -0-	--- -0-	--- -0-	66180 51914	--- -0-	61124 79737	--- -0-	127304 131651
Expenses	6744 4798	7150 6063	--- -0-	--- -0-	--- -0-	14790 246	--- -0-	28684 11107
Overhead/ Indirect	62734 <u>83374</u>	110725 <u>132003</u>	26602 <u>26602</u>	7746 <u>30228</u>	16496 <u>12928</u>	110788 <u>128253</u>	9116 <u>2716</u>	344207 <u>416104</u>
Totals	110301 139315	178208 209038	235623 241452	78848 98305	25536 20184	245274 288417	18137 6191	891927 1002902

7.1.2 Total Project Billables

	Phases (Tasks)							<u>Total</u>
	<u>1</u>	<u>2</u>	<u>3</u>	<u>4</u>	<u>5</u>	<u>6</u>	<u>7</u>	
Personnel	5018 5018	38189 38189	13301 13301	1722 1722	2000 2000	26566 38098	5075 3475	91871 101803
Equipment/ Supplies	12 12	172 172	--- -0-	--- -0-	--- -0-	5000 6136	--- -0-	5184 6320
Services	--- -0-	--- -0-	--- -0-	--- -0-	--- -0-	20220 20430	--- -0-	20220 20430
Expenses	144 144	1150 1150	--- -0-	--- -0-	--- -0-	14790 246	--- -0-	16084 1540
Overhead/ Indirect	4355 <u>4355</u>	63371 <u>63371</u>	26602 <u>26602</u>	1346 <u>1346</u>	2416 <u>2416</u>	68802 <u>68389</u>	5916 <u>2716</u>	172808 <u>169195</u>
Totals	9529 9529	102882 102882	39903 39903	3068 3068	4416 4416	135378 133299	10991 6191	306167 299288

7.1.3 Total Project Matching

	Phases (Tasks)							<u>Total</u>
	<u>1</u>	<u>2</u>	<u>3</u>	<u>4</u>	<u>5</u>	<u>6</u>	<u>7</u>	
Personnel	35793 46113	21972 32611	--- -0-	3200 14441	7040 5256	22072 31013	1600 -0-	91677 129434
Equipment/ Supplies	--- -0-	--- -0-	195720 201549	--- -0-	--- -0-	4934 4934	2346 -0-	203000 206483
Services	--- -0-	--- -0-	--- -0-	66180 51914	--- -0-	40904 59307	--- -0-	107084 111221
Expenses	6600 4654	6000 4913	--- -0-	--- -0-	--- -0-	--- -0-	--- -0-	12600 9567
Overhead/ Indirect	58379 <u>79019</u>	47354 <u>68632</u>	--- <u>-0-</u>	6400 <u>28882</u>	14080 <u>10512</u>	41986 <u>59864</u>	3200 <u>-0-</u>	171399 <u>246909</u>
Totals	100772 129786	75326 106156	195720 201549	75780 95237	21120 15768	109896 155118	7146 -0-	585760 703614

7.2 OHIO UNIVERSITY BUDGET

7.2.1 Total Ohio University Budget

	Phases (Tasks)							
	<u>1</u>	<u>2</u>	<u>3</u>	<u>4</u>	<u>5</u>	<u>6</u>	<u>7</u>	<u>Total</u>
Personnel	16811 16811	13570 13570	--- -0-	1722 1722	2000 2000	15147 15380	3475 3475	52725 52958
Equipment/ Supplies	12 12	172 172	--- -0-	--- -0-	--- -0-	5000 6136	--- -0-	5184 6320
Services	--- -0-	--- -0-	--- -0-	--- -0-	--- -0-	--- -0-	--- -0-	0 0
Expenses	144 144	--- -0-	--- -0-	--- -0-	--- -0-	500 246	--- -0-	644 390
Overhead/ Indirect	14734 <u>14734</u>	17543 <u>17543</u>	--- <u>-0-</u>	1346 <u>1346</u>	2416 <u>2416</u>	28785 <u>20791</u>	2716 <u>2716</u>	67540 <u>59546</u>
Totals	31701 31701	31285 31285	0 0	3068 3068	4416 4416	49432 42553	6191 6191	126093 119214

7.2.2 Ohio University Billables

	Phases (Tasks)							<u>Total</u>
	<u>1</u>	<u>2</u>	<u>3</u>	<u>4</u>	<u>5</u>	<u>6</u>	<u>7</u>	
Personnel	5018 5018	9570 9570	--- -0-	1722 1722	2000 2000	9147 9380	3475 3475	30932 31165
Equipment/ Supplies	12 12	172 172	--- -0-	--- -0-	--- -0-	5000 6136	--- -0-	5184 6320
Services	--- -0-	--- -0-	--- -0-	--- -0-	--- -0-	--- -0-	--- -0-	0 0
Expenses	144 144	--- -0-	--- -0-	--- -0-	--- -0-	500 246	--- -0-	644 390
Overhead/ Indirect	4355 <u>4355</u>	6133 <u>6133</u>	--- <u>-0-</u>	1346 <u>1346</u>	2416 <u>2416</u>	18946 <u>10952</u>	2716 <u>2716</u>	35912 <u>27918</u>
Totals	9529 9529	15075 15875	0 0	3068 3068	4416 4416	33593 26714	6191 6191	72672 65793

7.2.3 Ohio University Matching

	Phases (Tasks)							<u>Total</u>
	<u>1</u>	<u>2</u>	<u>3</u>	<u>4</u>	<u>5</u>	<u>6</u>	<u>7</u>	
Personnel	11793 11793	4000 4000	--- -0-	--- -0-	--- -0-	6000 6000	--- -0-	21793 21793
Equipment/ Supplies	--- -0-	--- -0-	--- -0-	--- -0-	--- -0-	--- -0-	--- -0-	0 0
Services	--- -0-	--- -0-	--- -0-	--- -0-	--- -0-	--- -0-	--- -0-	0 0
Expenses	--- -0-	--- -0-	--- -0-	--- -0-	--- -0-	--- -0-	--- -0-	0 0
Overhead/ Indirect	10379 <u>10379</u>	11410 <u>11410</u>	--- <u>-0-</u>	--- <u>-0-</u>	--- <u>-0-</u>	9839 <u>9839</u>	--- <u>-0-</u>	31628 <u>31628</u>
Totals	22172 22172	15410 15410	0 0	0 0	0 0	15839 15839	0 0	53421 53421

7.3 ETS BUDGET

7.3.1 Total ETS Budget

	Phases (Tasks)							
	<u>1</u>	<u>2</u>	<u>3</u>	<u>4</u>	<u>5</u>	<u>6</u>	<u>7</u>	<u>Total</u>
Personnel	24000 34320	46591 57230	13301 13301	3200 14441	7040 5256	33491 53731	3200 -0-	130823 178279
Equipment/ Supplies	--- -0-	--- -0-	195720 201549	--- -0-	--- -0-	4934 4934	2346 -0-	203000 206483
Services	--- -0-	--- -0-	--- -0-	66180 51914	--- -0-	61124 79737	--- -0-	127304 131651
Expenses	6600 4654	7150 6063	--- -0-	--- -0-	--- -0-	14290 -0-	--- -0-	28040 10717
Overhead/ Indirect	48000 <u>68640</u>	93182 <u>114460</u>	26602 <u>26602</u>	6400 <u>28882</u>	14080 <u>10512</u>	82003 <u>107462</u>	6400 <u>-0-</u>	276667 <u>356558</u>
Totals	78600 107614	146923 177753	235623 241452	75780 95237	21120 15768	195842 245864	11946 -0-	765834 883688

7.3.2 ETS Billables

	Phases (Tasks)							<u>Total</u>
	<u>1</u>	<u>2</u>	<u>3</u>	<u>4</u>	<u>5</u>	<u>6</u>	<u>7</u>	
Personnel	--- -0-	28619 28619	13301 13301	--- -0-	--- -0-	17419 28718	1600 -0-	60939 70638
Equipment/ Supplies	--- -0-	--- -0-	--- -0-	--- -0-	--- -0-	--- -0-	--- -0-	0 0
Services	--- -0-	--- -0-	--- -0-	--- -0-	--- -0-	20220 20430	--- -0-	20220 20430
Expenses	--- -0-	1150 1150	--- -0-	--- -0-	--- -0-	14290 -0-	--- -0-	15440 1150
Overhead/ Indirect	--- -0-	57238 57238	26602 26602	--- -0-	--- -0-	49856 57437	3200 -0-	136896 141277
Totals	0 0	87007 87007	39903 39903	0 0	0 0	101785 106585	4800 0	233495 233495

7.3.3 ETS Matching

	Phases (Tasks)							<u>Total</u>
	<u>1</u>	<u>2</u>	<u>3</u>	<u>4</u>	<u>5</u>	<u>6</u>	<u>7</u>	
Personnel	24000 34320	17972 28611	--- -0-	3200 14441	7040 5256	16072 25013	1600 -0-	69884 107641
Equipment/ Supplies	--- -0-	--- -0-	195720 201549	--- -0-	--- -0-	4934 4934	2346 -0-	203000 206483
Services	--- -0-	--- -0-	--- -0-	66180 51914	--- -0-	40904 59307	--- -0-	107084 111221
Expenses	6600 4654	6000 4913	--- -0-	--- -0-	--- -0-	--- -0-	--- -0-	12600 9567
Overhead/ Indirect	48000 68640	35944 57222	--- -0-	6400 28882	14080 10512	32147 50025	3200 -0-	139771 215281
Totals	78600 107614	59916 90746	195720 201549	75780 95237	21120 15768	94057 139279	7146 0	532339 650193

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APPENDIX A - PRESS RELEASES AND PROMOTIONAL BROCHURES

Press Release:

First Commercial LEC Order

Proudly serving industry and government since 1973.

ETS INTERNATIONAL, INC.

Providing Toxic Emission Control Products & Services

FOR IMMEDIATE RELEASE

DATE: MARCH 8, 1994

NO: 67

NEWS RELEASE: ETS INTERNATIONAL, INC. (ETS.EC-ASE)

SUBJECT: BREAKTHROUGH ORDER FOR FIRST COMMERCIAL ACID GAS CONTROL
SYSTEM RECEIVED

ETS International, Inc. today announced that its licensee in Taiwan has received the first ever order for ETS' Limestone Emission Control (LEC) system. The LEC system will be applied to the control of acid gas emissions from a metal alloy production process located in Taiwan R.O.C.

John C. Mycock, executive vice president of ETS International, Inc. stated, "This initial order is of tremendous significance. After ten years of successful system research and development we believe that this first commercial sale was the last major hurdle to be overcome. The beauty of this first order is that because of its modest size it can be up and running in a matter of months, thus it has the potential to facilitate additional sales. The order also confirms management's belief that at this time the foreign market opportunities for the LEC are greater than those currently available in North America."

ETSI granted U-Tah Industrial Company, Ltd. the exclusive right to sell its patented LEC system through Taiwan R.O.C. in February of 1993. ETSI receives a 5 percent royalty fee on the sale of all LEC system components. U-Tah has been very successful in introducing the

1401 Municipal Road, NW • Roanoke, Virginia 24012-1309 • Telephone: 703-265-0004 • FAX: 703-265-0082

ETS International, Inc. is an Equal Opportunity Employer.

system and currently has an additional 29 LEC proposals outstanding. U-Tah specializes in providing air pollution control systems throughout Taiwan R.O.C., Malaysia, Thailand and Mainland China. U-Tah is one of the leading air pollution control suppliers in Taiwan, reporting a 32 percent share of the domestic industrial market in 1991.

The market opportunities in Taiwan are substantial. Taiwan's flue gas desulfurization market is expected to reach \$2.8 billion U.S. by 1997. In addition there are market opportunities derived from waste incineration and industrial acid gas control.

The Limestone Emission Control system is a proprietary acid gas control system which has demonstrated high levels of acid gas removal (99 percent) at very competitive costs. A ten year R&D program included scrutiny by the Ohio Coal Development Office, the U.S. EPA and DOE as well as major U.S. industry.

ETS International, Inc. is a technology based firm which provides pollution control products and services. The firm specializes in toxic emission measurement and control. ETSI is headquartered in Virginia and has offices in Taiwan.

On behalf of the Board,

John C. Mycock
Executive Vice President
ETS International, Inc.

For further information:
Marc McKenna: (703) 265-0004
Richard Rozzi: (212) 564-7791 ext16

Press Release:

U-Tah Industrial Company, Ltd License

Proudly serving industry and government since 1973.

ETS INTERNATIONAL, INC.

Providing: Toxic Emission Control Products & Services

FOR IMMEDIATE RELEASE

DATE: FEBRUARY 4, 1993

NO: 52

NEWS RELEASE: ETS INTERNATIONAL, INC. (ETS.EC-ASE; ETS.V-VSE)

SUBJECT: ETS LICENSES LIMESTONE EMISSION CONTROL SYSTEM IN
TAIWAN R.O.C.

ETS International is pleased to announce that ETS has granted U-Tah Industrial Company, Ltd. the exclusive rights to sell its patented Limestone Emission Control System (LEC) throughout Taiwan, R.O.C. The base period of the license is five years. The contract calls for U-Tah to pay ETS an upfront license fee followed by a 5% royalty fee on the sale price of all LEC system components.

The Limestone Emission Control System (LEC) is a proprietary air pollution control system developed by ETS for the cost effective removal of sulfur dioxide. The LEC addresses the acid rain problem by controlling sulfur oxide emissions from power plants and industry.

Ten years of research and development have culminated in the following results:

1. The LEC is capable of greater than 99% sulfur dioxide removal.
2. The LEC has consistently demonstrated significant economic advantage over conventional Flue Gas Desulfurization technologies in use.

John Mycock, Executive Vice President of ETS International, Inc., stated, "Taiwan's Flue Gas Desulfurization (FGD) market is expected to reach \$2.8 billion U.S. by 1997. U-Tah selected the LEC because it offers a low cost yet highly effective approach to FGD. In Taiwan, for products of high quality, the lowest price wins! Also, the LEC has unlimited market share potential because it is suitable for use with both new and existing boilers. In addition to its economic advantage and superior performance, the LEC offers operational simplicity, compactness of size and operational safety over competing technologies."

U-Tah Industrial Company, Ltd. specializes in providing air pollution control systems throughout Taiwan R.O.C., Malaysia, Thailand and Mainland China. U-Tah, one of the leading air pollution control suppliers in Taiwan, reported a 32% share of Taiwan's industrial air

Page 2

pollution market in 1991. This market is expected to increase at an annual rate of 25% through 1997 in part because of the aggressive infrastructure program enacted by the Taiwan Government. U-Tah's headquarters and manufacturing facilities are located in Taipei Hsien, Taiwan R.O.C. They also have manufacturing facilities in Southern Taiwan.

ETS International, Inc. is a technology based firm which provides pollution control products and services. The firm specializes in toxic emission measurement and control.

On behalf of the Board,

John Mycock
Executive Vice President
ETS International, Inc.

For further information:
Marc McKenna: (703) 265-0004
Richard Rozzi: (212) 564-7791

THE VANCOUVER STOCK EXCHANGE HAS NOT REVIEWED AND DOES NOT ACCEPT RESPONSIBILITY FOR THE ADEQUACY OR ACCURACY OF THIS RELEASE.

Press Release:
Procedair Industries License

Proudly serving industry and government since 1973.

ETS INTERNATIONAL, INC.

Providing: Toxic Emission Control Products & Services

FOR IMMEDIATE RELEASE

DATE: June 24, 1992

NO: 47

NEWS RELEASE: ETS INTERNATIONAL, INC. (ETS.V)

SUBJECT: LIMESTONE EMISSION CONTROL SYSTEM LICENSE

Dr. John McKenna, President and Chairman of the Board of ETS International, Inc. is pleased to announce that ETS has licensed Procedair Industries the exclusive rights to sell its patented Limestone Emission Control System for the North American industrial market. The base period of the license is five years and Procedair Industries will pay ETS a 5% royalty fee on the sale price of all LEC system components.

As part of a large international corporation, Procedair Industries has both the technical and commercial resources to handle air pollution control projects of any size for most industrial applications. Procedair has completed many multi-million dollar, multi-year contracts. Contract values have ranged anywhere between \$300,000 and \$30,000,000. Few of the companies working in the field of pollution control possess the in-house disciplines found within Procedair Industries.

Procedair Industries is the North American Subsidiary of the highly respected industrial furnace manufacturer, Stein Heurtey. It is also a sister company of Procedair, S.A. a French air pollution control firm, and with EGCI Pillard, a French engineering firm specializing in combustion. Stein Heurtey is the leading firm of the major French financial holding company, the Compagnie de Fives-Lille.

Though Procedair Industries is a subsidiary of a French company, all its engineering is performed in the USA or in Canada while fabrication of the equipment is supplied from North American sources as are all required equipment spares.

Procedair Industries is very active in the North American acid gas scrubbing market with demonstrated market presence and commitment. It has recently complemented its in-house capability and substantial financial resources by collaborating with the German based industrial supplier, Thyssen via a former business arrangement.

Procedair Industries has offices in the United States, Canada, and Mexico, including offices in Montreal, Kentucky, Connecticut, Ohio, Pennsylvania, California, New Jersey and Mexico City.

Page 2

The Limestone Emission Control System (LEC) is a proprietary air pollution control system developed by ETS for the cost effective removal of sulfur dioxide. The LEC addresses the acid rain problem by controlling sulfur oxide emissions from power plants and industry.

Ten years of research and development has culminated in the following results:

1. The LEC is capable of greater than 99% sulfur dioxide removal.
2. The LEC has consistently demonstrated significant economic advantage over conventional Flue Gas Desulfurization technologies in use.

The LEC system is suitable for controlling sulfur dioxide emissions (i.e., Flue Gas Desulfurization (FGD)) from both electric utility and industrial boilers. The market for such system is expected to be a multi-billion dollar market throughout the 1990's. Numerous market researchers have forecast this market. For example, Bob McIlvaine of the McIlvaine Company (Northbrook, IL) has estimated that the current global market for APC systems for stationary sources is about \$13 billion. The U.S. represents about 30% of the market, or \$4 billion this year. Flue gas desulfurization (FGD) is the largest contributor to the U.S. market, representing an estimated \$1.7 billion in sales in the U.S. in 1991.

This is further confirmed by the EPA estimates that the Acid Rain portion of the Clean Air Act of 1990 will cost industry \$3.8 billion annually.

ETS International, Inc. is a technology based firm which provides pollution control products and services. The firm specializes in toxic emission measurement and control.

On behalf of the Board,

John McKenna, Ph.D.
President & Chairman
ETS International, Inc.

For further information:

Marc McKenna: (703) 265-0004
Richard Rozzi: (212) 564-7791

THE VANCOUVER STOCK EXCHANGE HAS NOT REVIEWED AND DOES NOT ACCEPT RESPONSIBILITY FOR THE ADEQUACY OR ACCURACY OF THIS RELEASE.

LEC Promotional Brochure

Benefits of the LEC System

✓ Lowest Cost —

- ✧ The key to maximum profits is minimizing costs.
- ✧ The U.S. Department of Energy, The Tennessee Valley Authority, and Babcock and Wilcox all were impressed with the low cost of ETS's limestone emission control technology.

✓ Extremely Retrofittable —

- ✧ Easily adapted to new construction or existing facilities.
- ✧ The LEC has a small footprint, making it easy to locate.
- ✧ The LEC handles high and low amounts of SO₂ in flue gas.
- ✧ The LEC is adaptable either for utility or industrial use.

✓ Capable of B.A.C.T. Performance —

- ✧ The LEC is one of the few wet/dry flue gas desulfurization processes that can achieve high SO₂ removal efficiencies.
- ✧ Pilot studies have demonstrated that SO₂ removal is over 90%, and very often up to 99%.

✓ Operational Simplicity —

- ✧ The LEC has few moving parts.
- ✧ Parts are simple to operate and repair.
- ✧ Advanced education is not required for the LEC operators.
- ✧ The system uses very little energy.

✓ Operational Safety —

- ✧ Raw limestone is utilized, which is inexpensive and safe.
- ✧ The system operates at low temperatures.
- ✧ The solid byproduct is gypsum, a safe and valuable substance.
- ✧ The process uses 80% less water than wet scrubbers.

Process Description:

The LEC system includes three major components:

- (1) Limestone Handling;
- (2) SO₂ Absorption; and
- (3) Limestone Regeneration.

Hot flue gases (<350 °F) are contacted with a bed of 1/32 to 1/3 inch limestone granules covered with a thin film of water. Sulfur dioxide is absorbed from the flue gas into the water film where it subsequently reacts with dissolved limestone. A layer of reaction products (gypsum) forms on the limestone. Next, the reacted limestone is passed through the regeneration system. The re-active limestone is then recycled back to the reactor to continuously react with SO₂.

Evaluations of the ETS' LEC System from Independent Organizations:

The Limestone Emission Control Process has the potential of being the most cost effective SO₂ abatement method for boilers firing high-sulfur Ohio coal, and possibly the only emerging technology that will, alone, meet New Source Performance Standards (NSPS).

A Major U.S. Boiler Manufacture

Cost Comparison of LEC System

By a government agency

PLANT SIZE	LEC	Commercial Technologies			
		LIMESTONE SCRUBBER	LIME SPRAY DRYING	WELLMAN LORD	LIME DUAL ALKALI
100 MW	\$442	\$736	\$620	\$752	\$659
350 MW	\$317	\$521	\$510	\$499	\$506
500 MW	\$296	\$480	\$487	\$448	\$475
750 MW	\$277	\$439	\$463	\$399	\$444

**Levelized Cost Estimate, \$ per Ton SO₂ Removed
Based on 4% Sulfur Coal**

For more information about the LEC or other ETS products and services contact:

ETS International, Inc.
1401 Municipal Road
Roanoke, VA 24012-1390
Phone # (703) 265-0004
Fax # (703) 265-0131

Other ETS Products and Services:

- **BPM/BPMES:**
An instrument and software which serves as an early warning system of baghouse dust collector malfunction.
- **Dry Reactor:**
A system which controls acid gases and particulate emissions from municipal and hospital incinerators.
- **Services**
  - Testing
  - Training
  - Design

Distributed by:

"DEVELOPMENTS at ETS" (Vol. 1, No. 2)

DEVELOPMENTS

AT ETS

Vol. 1, No. 2

Conventional wisdom says the solution to pollution is dilution. Such may be true at the beginning of a pollution cycle in a pristine environment, but it is hardly the case in the modern world where, literally, what goes around comes around.

The Pollution

In ancient Egypt the granite obelisk known as Cleopatra's Needle was carved for the ages. For 35 centuries it stood intact in Egypt. Then it was shipped to New York City's Central Park about 75 years ago. In the span of one lifetime the windward side of Cleopatra's Needle has been almost completely defaced by acid air pollution. This phenomenon is not unique to New York City: Around the world classical monuments in Athens, Rome, Venice and elsewhere are crumbling before the noxious fumes of industrial activity.

A trip to Sudbury, Ontario will reward the curious with a view of the world's largest smokestack, taller even than the Empire State Building. Most of that smoke is pushed up into the higher atmosphere and sent on its way to Europe. Meanwhile, back in the Midwest smokestacks are kicking their pollution into the air, where much of it finds its way to Canada! This is a

classic example of "out of sight, out of mind," but not out of our lives.

Thousands of lakes in North America and elsewhere are profoundly wounded from acid rain. These acids mostly origi-

Cleopatra's Needle before it was taken to New York and defaced by pollutants.

nate from coal burning utilities in the Midwest that are belching thousands of tons of sulfur dioxide which end up on earth as sulfuric acid.

The Solution To Pollution

The Solution

The world has exhausted the option of dilution as the solution to acid rain. That is why Congress enacted the strong provisions in the 1990 Clean Air Act. The Act did not mandate which technology would be preferred. It only mandated reductions in air pollution, with the market determining what technologies, old or new, would be purchased. Congress' strategy was to encourage competition among technologies, leading to the least expensive, best available technology.

The problem that utilities and major industries face in cleaning up their air pollution has become a major profit opportunity for the emerging air pollution control industry. Technologies that promise the cleanest air for the least cost will take center stage in the 1990s.

and beyond. One of the most promising technologies has been developed at ETS International, Inc. This is the Limestone Emission Control System, also known as the LEC.

How the LEC Was Developed

ETS has been developing a proprietary sulfur dioxide removal system since 1982. As a new technology, the development of this system has necessarily progressed from bench scale studies, through small pilot scale units, up to the current large pilot plant being operated on the slipstream of a coal-fired boiler at Ohio University.

Simply, the LEC is an application of scrubbing. Hot flue gases (greater than 350° F) are contacted with a bed of 1/32 inch to 1/3 inch diameter limestone granules that are covered with a thin film of water. Sulfur dioxide gas is absorbed from the flue gas into the water film where it subsequently reacts with dissolved limestone, forming a layer of non-hazardous gypsum. This layer is periodically removed by abrasion for sale, and the renewed limestone is reintroduced to the flue gas.

The LEC pilot plant testing is scheduled to be completed by the spring of 1992, with a full-scale commercial demonstration on a 10 megawatt electric utility boiler to follow. Even though older scrubbing technologies are much more expensive, and usually less effective in removing sulfur dioxide, the LEC needs to go through this process of proving itself on progressively larger boilers so that utilities can

confidently invest the millions needed for each retrofit.

How has the university pilot plant performed to date? As with earlier and smaller models of the LEC, this large pilot plant has demonstrated sulfur dioxide removal of greater than 90%, and often in excess of 99%. The Environmental Protection Agency, the Department of Energy, and Babcock & Wilcox have previously performed economic evaluations of the LEC systems. These independent studies have all confirmed the cost-effectiveness of this new scrubbing technology.

Licensing arrangements with a major air pollution control supplier are currently being negotiated. ETS is looking to team up with a major partner that can help the LEC capture about ten percent of this emerging market.

What is the Market For the LEC?

The Environmental Business Journal reported in November 1991 that the U.S. air pollution control systems market will reach between \$6 billion and \$7 billion each year by 1995 as the new law takes effect. Flue gas desulfurization (FGD) is the largest dollar contributor to the U.S. market, already representing an estimated \$1.7 billion in sales for 1991.

According to government sources, 311 electric utility power plants in America will have to install some type of flue gas desulfurization system. There are also hundreds more large

industrial plants with their own coal-fired boilers generating electricity and steam, constituting another market roughly equal in size to the electric utility market. Frank Princiotta, director of EPA's AAER Laboratory said in May of 1990 that the "EPA estimates that the Acid Rain portion of the Clean Air Act of 1990 will cost industry \$3.8 billion annually."

Where will ETS' patented LEC fit into this emerging picture? In October of 1989 Babcock & Wilcox wrote the Ohio Coal Development Office as follows: "The Limestone Emission Control Process has the potential of being the most cost effective SO₂ abatement method for boilers firing high-sulfur Ohio coal, and possibly the only emerging technology that will, alone, meet New Source Performance Standards."

Coal is the only secure long-term fuel, and it is by far the most economical. America doesn't have to import coal from sensitive regions of the globe. We are the OPEC of coal, so to speak. However, if we want to use coal we must clean up the by-products of coal combustion. This is where ETS' proprietary LEC technology will lead the way into the 21st century.

For Further Information, Contact:

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APPENDIX B - LEC. DEVELOPMENT CHRONOLOGY

APPENDIX B - LEC DEVELOPMENT CHRONOLOGY

The present commercial design of the LEC moving bed reactor is the end product of over 6 years of research and development by ETS International.

B.1 1987

B.1.1 Commercial Design Work

The 400 acfm fixed-bed pilot plant project located at Ohio University produced the following initial commercial design parameters:

- (1) a bed superficial velocity of 1.5 fps.
- (2) a bed depth of 6 inches.
- (3) a operational time of 8 hours between regenerations of the bed.

B.1.2 Pilot-Plant Study

The 400 acfm fixed-bed pilot plant located at Ohio University, which operated in a batch mode, completed a total of 106 experimental runs which successfully demonstrated the capabilities of the LEC desulfurization system. The following results provided the basis for future pilot plant projects:

- (1) All three Ohio limestones used in the study were able to achieve over 90% SO₂ removal. Many tests showed removals as high as 99%.
- (2) The LEC process was able to sustain high SO₂ removal efficiencies for multi-hour periods.
- (3) SO₂ removal performance for the LEC was effective for inlet SO₂ concentrations ranging from 500 ppm_{dv} to 3500 ppm_{dv}.
- (4) Two factors that limit the LEC operation were identified as:
  - (i) Drying of the reactive surface of the limestone. A dry limestone surface is unreactive.
  - (ii) Build-up of reaction products produces a blinding effect of active calcium sites.
- (5) Two methods to produce long cycle times were the humidification of the inlet gas stream and the use of overhead bed sprays. Both these

methods created a constant liquid water phase on the surface of the limestone particle.

- (6) The reaction product build-up on the surface of the stone can be removed by attrition; thus exposing active reaction sites on the limestone particle.
- (7) LEC on stream cycle was observed to increase with increasing bed depth and decreasing gas superficial velocity.

B.2 1988

B.2.1 Commercial Design Work

Using the design parameters generated by the 1987 fixed-bed pilot plant project, a LEC was designed for a 100 MW power plant using 33 stacked horizontal beds that used vibration as a means to convey the limestone granules across the bed.

B.2.2 Pilot-Plant Study

No pilot plant activity.

B.3 1989

B.3.1 Commercial Design Work

B.3.1.1 Preliminary Search. A search was conducted for existing moving-bed and similar filtration system technologies that could be adapted to the LEC design scheme. These can be classified into either vertical or horizontal moving-beds.

B.3.1.2 Vertical Moving-Beds. Nine possible moving-bed designs from seven different companies that were used for either gas removal or filtration were located. The system showing the most promise was the Hellmich Fluorine Cascade Absorber.

The Hellmich design incorporates the use of a cascade system consisting of inverted chevrons interspersed within a moving-bed of dry 1/4" limestone granules. This design also incorporates the following unique items:

- (1) Pneumatic conveying.
- (2) One month storage hopper on top of the absorber.
- (3) A surface regeneration device known as the "peeler".

B.3.1.3 Horizontal Moving-Beds. A commercially-used horizontal moving-bed called the Puritreat process manufactured by Air Resources uses 1/4" coke granules as the reactant was located at the Pinetree Power Plant in New Hampshire. This type of system closely matches the configuration of the fixed-bed project in that the gas flow passes downward over a horizontal bed of approximately 1/4" granules. A fresh bed section of reactant is conveyed into place by an indexing motor when the bed pressure drop across the bed becomes great enough to increase the current draw of the induced draft fan. This type of horizontal bed configuration looked very promising, so a preliminary LEC design using this scheme was completed.

The horizontal bed was eliminated as a potential LEC design in April, 1989 due to the following:

- (1) gas seal problems on commercial units.
- (2) complex and high maintenance of the index-drive mechanism.
- (3) obtaining even stone feed to all the stacked-beds would have been a potential problem.

B.3.2 Experimentation

B.3.2.1 Vibrating-Bed Flow Tests. In order to test the horizontal flow capabilities of wet limestone across a vibrating bed, a vibrating test-bed driven by a pneumatic rapper was constructed. This testing showed that little or no stone movement occurred using this configuration. No stone movement was observed even when the bed was inclined at up to 30°. No further investigations into the horizontal vibrating bed were made.

B.3.2.2 Dry-Method Regeneration Studies. Grinding tests on dry, spent limestone were conducted in order to determine the most effective method for removing the $\text{SO}_2/\text{CaCO}_3$ reaction by-product from the surface of the granule. The testing consisted of attriting spent limestone using either grinders, ball mills or vibrating screens separately or in combination from five companies.

The results of the testing showed that none of the commercially available grinders were acceptable for the LEC system for one or all of the following reasons:

- (1) Too low a percentage of the reaction by-product was removed from the granule.

- (2) Extreme abrasion of the granule yielded a high percentage of lost CaCO_3 in the undersize, i.e., the resulting percent calcium utilization was economically very unattractive.

The potential cost of developing a combination mill/screen in-house or in-conjunction with Combustion Engineering that would be able to remove only the outside surface area of the limestone granule is not within the budget.

B.3.2.3 Pilot-Plant Studies. From the three possible LEC design scenarios for a pilot plant being; (1) a cycling fixed-bed; (2) a Hellmich-inspired chevron moving-bed; or (3) a louvered moving-bed, the chevron moving-bed was selected due to its compact size and intra-bed moisture control capability. Table 1 in Reference test # 90-1 lists the advantages and disadvantages of each design scheme.

Subsequently, a 400 acfm moving-bed reactor with internal chevrons was installed at the Norfolk and Southern (N&S) Roanoke shops in September 1989. Between September 25, 1989 and November 7, 1989, seven experimental runs were conducted on the unit. Although the internal chevron design allowed for a small footprint with excellent stone moisture control, the reactor internals caused excessive bridging of the limestone granules between and around the chevrons. The internal chevrons were removed for the further moving-bed experimentation conducted in early 1990. Run summary data for these runs can be found in Appendix C.

B.4 1990

B.4.1 Commercial Design Work

Negotiations with the Hellmich firm from Germany ended and no further commercial designs were proposed at this time.

B.4.2 Pilot-Plant Study

B.4.2.1 Norfolk and Southern Pilot Unit. The moving-bed pilot plant at Norfolk and Southern (N&S) was converted from the Hellmich-inspired chevron/cascade design to the louvered moving-bed design during the Fall of 1989. The louvered-bed design has no internal cascades that cause the reactor bridging problems as experienced with the chevrons.

The louvered pilot unit testing consisted of 56 experimental runs that were conducted at N&S between January 8, 1990 and March 12, 1990. The majority of

these runs produced greater than 90% SO₂ removal efficiency without an increase in bed pressure drop, relative to the chevron/cascade design. The one disadvantage is that some stone leakage into the inlet and outlet plenums through the louvers was experienced.

A SO₂ removal efficiency as a function of temperature and water addition rate generally showed that higher efficiencies were produced at water rates less than 0.15 gpm and at inlet gas temperatures greater than 300°F. This was determined from the fresh stone runs only, where in most cases mechanical steady-state was not achieved before the end of the run; therefore this finding was labeled as only preliminary.

B.4.2.2 LEC-Ohio Moving-Bed Pilot Plant. Based on the N&S pilot unit study results, a 5000 acfm louver moving-bed pilot plant was designed. The pilot unit was installed on the slip stream of the 70,000 lb/hr stoker boiler providing steam to the Ohio University campus, located in Athens, Ohio. Start-up and testing did not commence until 1991.

B.4.2.2.1 Material Handling Equipment. The pneumatic material handling equipment as used in the Hellmich design was not used on the LEC-Ohio project. Instead, conventional stone handling equipment such as bucket elevators and screw conveyors were used.

A twin screw conveyor with a width of 15 inches was designed for removal of reacted stone from the bottom of the bed. A twin screw was selected due to the following;

- (1) A twin screw conveyor will have more even stone removal from the bed.
- (2) The twin augers can more easily crush the expected blocks of fused granules than a single auger-type.
- (3) A single auger would require tapering of the bottom of the bed; thus causing possible bridging problems.
- (4) The 15 inch bed depth is in the range used during the 1987 fixed-bed project.

B.4.2.2.2 Regeneration Equipment. Three limestone regeneration techniques were scheduled to be tested at Ohio University;

- (1) The triple cascade impeller fluidizer as a wet method regeneration technique that was developed through Case Western University and

Acirion Technologies. A single-deck SWECO 24" vibrating screen would be used as a dewatering device for the oversized regenerated granules.

- (2) A SWECO finishing mill as a light abrasion attritor was scheduled to be used as a dry method grinder that would employ several different types of finishing media. A twin-deck SWECO 24" vibrating screen would be used as a classifier.
- (3) A single-deck SWECO 24" vibrating screen would be used as a stand-alone regeneration device.

B.5 1991

B.5.1 Commercial Design Work

A commercial LEC louvered bed design was developed based upon operating parameters used at the time of the LEC-Ohio project. These were;

- (1) 15 inch bed depth.
- (2) Superficial gas velocity equal to 1.5 fps.
- (3) Maximum stone velocity of 2.4 in/min.

The proposed commercial design incorporates the basic moving-bed reactor design by placing a series of these louvered moving-beds in parallel. Each individual moving-bed reactor is separated by the adjacent bed by a series of inverted inlet and outlet chevrons located at the bulk gas/solid interface. These chevrons are not of the cascade-type found in the Hellmich design, but are simply 8 inch louvers, angled at 75°, that are joined at the apex. A variable speed twin screw conveyor is located at the bottom to facilitate stone movement. The moving-beds are contained within a single reactor housing with common gas and material handling equipment servicing the individual beds as shown in Drawing # 792-44. The chevrons are shown in detail in Drawing # LEC-Study.

B.5.2 Pilot-Plant Study

A total of 30 experimental runs were conducted on the LEC pilot unit between May 29, 1991 and July 25, 1991. The length of experimental runs ranged from 30 to 290 minutes. The performance of the LEC system yielded SO₂ removals greater than 99% up to gas flow rates of 3200 acfm. The two limestone regeneration techniques studied were the Case fluidizer and the vibrating screen.

The experimental runs were halted after July 25, 1991 in order to retro-fit the LEC-Ohio pilot plant with new design changes that would solve mechanical problems encountered with limestone carry-over and the material handling system. The unit was reinstalled during the Winter of 1991/1992.

B.5.2.1 Limestone Carry-Over. The limestone carry-over before a wire mesh screen was installed on outlet louvers limited experimental runs to 30 minutes due to the excessive amount of whole limestone entering the outlet plenum. The screen mesh stopped the limestone from exiting the bed through the louvers, but the outlet screen would become fouled with fines after approximately four hours of operation.

This problem was partially solved by increasing the length and overlap of the outlet louvers and installing a drop-out leg in the outlet plenum. The original louvers were four inches long with a half inch overlap. The outlet louvers installed in the Fall of 1991 measured 11 inches long with a 3-1/2 inch overlap. The inlet louvers were not changed.

B.5.2.2 Material Handling. The following items were retro-fitted for the LEC-Ohio during the 1991 shut-down:

- (1) Access doors to all chutes.
- (2) Motor up-grades for the twin screw and feed conveyors.
- (3) Bucket elevator discharge chute extensions.

B.5.2.3 Limestone Regeneration. Experimental runs with the Case fluidizer were halted when it became clear that its stone capacity had been extremely undersized and that the amount of water required to operate the cascade was economically unfeasible. The vibrating screen as a stand-alone regeneration device was adopted.

B.6 1992

B.6.1 Commercial Design Work

Due to the continued limestone carry-over problem exhibited at the LEC-Ohio project even after the enlarged outlet louvers were installed, the commercial chevron design was abandoned for the current slot and louver design.

The slot and louver design consists of inserting a 6 inch outlet gas channel or "slot" between each of the moving beds. At the bottom of each slot there is a constant speed single screw conveyor with rotary air lock. The chevrons are

replaces with 11 inch long louvers included at a 70° angle from the horizontal. This design more closely mirrors the LEC-Ohio configuration, since it enables the stone carry-over to drop down into the slot and be conveyed to the regeneration section. A 10 MW slot and louver design are shown in Drawing #100-93.

B.6.2 Pilot-Plant Studies

The LEC-Ohio restarted in 1992 after the retro-fit of 1991 with new 11 inch outlet louvers and a stone drop-out leg. This enabled the LEC-Ohio project to obtain long-term run data. The experimental results for LEC Runs 920526 through LEC Run 920719 can be summarized as follows;

- (1) SO₂ removal is not a function of inlet SO₂ concentration over the range studied (300 ppm_{dv} to 3500 ppm_{dv}) at inlet gas rates less than 1000 acfm.
- (2) SO₂ removal is not a function of bed speed at "low" (< 1500 ppm_{dv}) inlet concentrations. Higher bed speeds (4040 lb/hr as compared to 2300 lb/hr) may be required in order to remove the sulfate layer at higher SO₂ concentrations (> 3000 ppm_{dv}).
- (3) SO₂ removal is a function of thin film water addition, i.e., SO₂ removal increases with increasing water addition to a maximum removal rate, but then decreases rapidly after this point. The reasoning for this is two fold: (i) The thicker the water film layer, the more time it takes SO₂/water diffusion to occur. (ii) Excess water will occupy stone void space resulting in both increased pressure drop and gas channelling.
- (4) At bed speeds below 2300 lb/hr, stone bridging problems occur.
- (5) No. 9 Maxville limestone yields consistent +99% SO₂ removal values over the range of inlet gas flow rates tested (400 acfm to 1500 acfm).
- (6) No. 8 Maxville limestone has produced an average SO₂ removal of 47% for 388 minutes (LEC Run 920619).

B.6.2.1 Limestone Carry-Over. The inlet gas flow rate during this phase of testing was again limited by the amount of limestone carry-over encountered when the superficial bed velocity exceeded 0.5 fps. The new louvers greatly decreased the amount of stone being carried over the outlet louvers, but it did not entirely solve the problem. This was due primarily to the space restrictions on the size of the drop-out leg. This resulted in labor-intensive manual unloading of stone from the drop-out leg.

B.6.2.2 Limestone Regeneration. The regeneration of the limestone was proved possible with long-term experimental runs using the vibrating screen as a stand-alone regeneration device for both No. 8 and No. 9 limestone.

APPENDIX C - PREPILOT MOVING-BED LEC EXPERIMENTAL RESULTS

**LEC PILOT-SCALE MOVING-BED PERFORMANCE RUNS:
DATA SUMMARY**

C-1

<u>Run No.</u>	<u>Elapsed Run Time (hr)</u>	<u>Inlet Temperature (F)</u>	<u>Outlet Temperature (F)</u>	<u>Inlet Gas Rate (ACFM)</u>	<u>H2O Rate (gal/min)</u>	<u>Screw Speed (%)</u>	<u>Stone</u>	<u>Bed</u>	<u>Bed Pressure Drop (in.H2O)</u>	<u>Inlet SO2 Conc. (ppm)</u>	<u>Outlet SO2 Conc. (ppm)</u>	<u>SO2 Removal (%)</u>
890925	179.5	372	107	406	14.0	3	N	a	1.3	273	45	83.5
891011	8.4	412	110	387	20.0	13	N	a	1.2	1088	180	83.5
891025	29.3	382	108	295	17.0	13	N	a	3.5	366	68	81.4
891031	1.0	347	113	256	0.0	13	N	a	1.7	1474	365	75.2
891102	2.1	304	96	170	18.0	8	N	a	1.6	1382	494	64.3
891103	1.0	399	99	188	23.0	13	N	a	1.7	1530	683	55.4
891107	5.3	383	108	273	17.0	8	N	a	2.9	1271	160	87.4
900109a	0.8	309	88	176	0.85	100	N	b	1.6	260	17	93.5
900109b	0.6	312	97	166	0.21	100	N	b	1.7	1608	89	94.5
900109c	0.8	313	92	165	0.23	50	N	b	1.6	1408	105	92.5
900110a	0.7	317	93	179	0.30	50	N	b	2.0	1121	110	90.2
900110b	1.6	306	92	162	0.22	50	N	b	2.0	817	99	87.9
900116a	2.7	329	96	233	0.22	50	N	b	2.0	1473	98	93.3
900116b	2.9	238	80	102	0.23	50	N	b	1.0	1380	10	99.3
900116c	1.9	346	97	254	0.23	50	N	b	2.6	1713	92	94.6
900117a	2.4	346	94	247	0.25	50	N	b	2.5	1211	44	96.4
900117b	3.0	248	83	101	0.22	50	N	b	1.4	1525	3	99.8
900117c	0.9	334	91	208	0.35	50	N	b	2.8	1690	52	96.9
900122a	0.7	298	77	203	0.23	50	N	b	2.3	261	3	98.7
900122b	2.4	315	91	230	0.16	50	N	b	2.3	1575	108	93.1
900122c	2.3	316	92	217	0.22	50	N	b	2.8	3029	408	86.5
900123a	2.1	327	96	168	0.11	50	N	b	2.8	1468	24	98.3
900123b	3.4	311	95	160	0.05	50	N	b	3.1	1550	44	97.2
900123c	2.4	326	98	164	0.05	50	N	b	3.3	1561	57	96.3
900124a	2.3	328	97	149	0.22	50	N	b	3.2	1334	30	97.8
900124b	3.2	337	99	164	0.06	50	N	b	3.2	1325	23	98.3
900124c	1.5	326	98	161	0.04	50	N	b	3.4	2727	225	91.7

**LEC PILOT-SCALE MOVING-BED PERFORMANCE RUNS:
DATA SUMMARY (continued)**

<u>Run No.</u>	<u>Elapsed Run Time (hr)</u>	<u>Inlet Temperature (F)</u>	<u>Outlet Temperature (F)</u>	<u>Inlet Gas Rate (ACFM)</u>	<u>H2O Rate (gal/min)</u>	<u>Screw Speed (%)</u>	<u>Stone</u>	<u>Bed</u>	<u>Bed Pressure Drop (in.H2O)</u>	<u>Inlet SO2 Conc. (ppm)</u>	<u>Outlet SO2 Conc. (ppm)</u>	<u>SO2 Removal (%)</u>
900126a	0.8	283	90	135	0.09	50	N	b	2.8	1516	149	90.2
900126b	1.5	256	89	128	0.07	50	N	b	3.6	2047	445	78.3
900126c	0.9	200	81	94	0.03	50	N	b	3.4	1989	240	87.9
900129a	1.3	306	99	194	0.08	50	N	b	1.9	1469	90	93.9
900129b	3.0	292	97	166	0.08	50	N	b	1.9	2952	340	88.5
900130a	3.3	264	89	98	0.04	50	N	b	1.0	2780	132	95.3
900130b	4.3	293	95	143	0.23	50	N	b	3.0	2899	416	85.7
900131a	1.4	264	87	117	0.11	50	N	b	2.3	3161	233	92.6
900131b	2.0	275	88	123	0.09	100	N	b	2.7	3087	235	92.4
900202a	4.8	279	94	121	0.09	50	N	b	2.9	2923	168	94.3
900202b	1.0	277	97	138	0.06	50	N	b	4.5	2945	184	93.8
900214	7.8	344	100	251	0.04	50	U	b	2.4	2875	508	82.3
900219a	4.3	354	104	303	0.06	50	U	b	3.3	1314	414	68.5
900219b	1.8	229	87	250	0.09	100	U	b	2.8	1496	479	68.0
900220	2.0	190	91	238	0.12	25	U	b	3.6	2387	1109	53.5
900222a	1.8	133	68	243	0.11	45	U	b	2.9	1348	511	62.1
900222b	3.6	173	81	252	0.11	50	N	b	2.9	1573	490	68.8
900223a	1.0	167	74	234	0.13	50	N	b	2.6	1181	214	81.9
900223b	0.7	286	92	259	0.11	50	N	b	2.5	1468	150	89.8
900223c	0.9	145	85	227	0.15	50	N	b	2.9	1253	253	79.8
900223d	1.4	284	93	262	0.09	50	N	b	3.0	1648	310	81.2
900223e	0.4	168	93	240	0.12	50	N	b	3.4	1385	262	81.1
900302a	2.1	279	92	289	0.11	100	U	b	4.2	1470	341	76.8
900302b	1.6	277	97	264	0.13	100	U	b	3.7	2433	838	65.6

C-2

**LEC PILOT-SCALE MOVING-BED PERFORMANCE RUNS:
DATA SUMMARY (continued)**

<u>Run No.</u>	<u>Elapsed Run Time (hr)</u>	<u>Inlet Temperature (F)</u>	<u>Outlet Temperature (F)</u>	<u>Inlet Gas Rate (ACFM)</u>	<u>H2O Rate (gal/min)</u>	<u>Screw Speed (%)</u>	<u>Stone</u>	<u>Bed</u>	<u>Bed Pressure Drop (in.H2O)</u>	<u>Inlet SO2 Conc. (ppm)</u>	<u>Outlet SO2 Conc. (ppm)</u>	<u>SO2 Removal (%)</u>
900305a	2.1	271	97	256	0.15	100	U	b	3.9	1453	395	72.8
900305b	0.5	174	78	253	0.22	50	U	b	5.5	1561	445	71.5
900305c	1.9	245	93	254	0.13	100	U	b	5.5	1540	447	71.0
900306a	5.3	276	93	254	0.09	100	U	b	4.1	1919	420	78.1
900306b	1.4	273	96	240	0.10	50	U	b	4.4	1512	136	91.0
900307	5.1	266	90	240	0.21	50	U/N	b	4.7	1393	295	78.8
900308	6.1	273	89	253	0.15	100	U/N	b	4.8	1328	159	88.0
900309a	1.0	285	96	290	0.19	50	N	b	6.0	1287	333	74.1
900309b	0.7	235	95	271	0.15	50	N	b	6.0	1400	524	62.6
900309c	1.4	291	102	271	0.15	50	N	b	5.0	1517	538	64.5
900309d	0.8	250	7	264	0.17	27	N	b	6.0	1506	217	85.6
900312	3.6	287	107	233	0.0	0	N	b	7.9	2469	888	64.0

Stone: N = new.
U = used.

Bed: a = chevron bed.
b = louvered bed.

APPENDIX D - ECONOMIC CALCULATIONS

AN ECONOMIC COMPARISON
OF THE LEC WITH A
LIMESTONE SCRUBBING SYSTEM ON A
500 MW COAL FIRED POWER PLANT

JANUARY, 1991

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Specializing In
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EXECUTIVE SUMMARY

The cost effective removal of sulfur dioxide from coal-fired combustor flue gases is rapidly becoming more important to electric power utilities as emission limits for SO₂ are lowered due to recent amendments to the Clean Air Act. The Limestone Emission Control (LEC) process is currently under development by ETS, Inc. as a retrofit technology to fill this need of greater SO₂ removal, while achieving a cost advantage over currently used flue gas desulfurization (FGD) systems. To illustrate the economic potential of the LEC, this study compared the annualized capital and operating costs of the LEC process to that of a conventional limestone scrubber. The resulting annualized capital and operating costs for the LEC process were significantly lower; thus clearly making the LEC a more economically viable FGD process.

In this economic study, the two FGD systems are compared on an equatable basis using flue gas conditions that would be expected on the outlet of an electrostatic precipitator (ESP) of a 500 MW coal-fired power plant. The equipment list for the LEC case was obtained from scaling up the current LEC moving bed pilot project located at Ohio University. The equipment list for the limestone scrubber case were obtained from an Electrical Power Research Institute (EPRI) technical report. Equipment costs for both the LEC and limestone scrubber were generally obtained through empirical estimating charts.

The LEC was found to have a definite economic advantage in both direct capital costs and operating costs. The LEC equipment capital costs of \$15,104,000 represented a 36% savings when compared to the limestone scrubber process equipment costs of \$23,534,000. On a absorption tower basis, a capital cost advantage of 46% for the LEC occurs due to its simpler mechanical design. Operating costs are also less with the LEC process since a slurry is not required, resulting in a lower water consumption. The solid material handling associated with the LEC requires less electricity than that of the slurry pumping required by the limestone scrubbing system. Table X-1 lists the operating and equipment costs per process area for each FGD system along with the total annualized costs.

TABLE X-1

SUMMARY OF RESULTS FOR A COMPARATIVE ECONOMIC ANALYSIS
OF A LEC AND A LIMESTONE SCRUBBER PROCESS
FOR A 500 MW FACILITY

<u>RAW MATERIAL AND UTILITY CONSUMPTION</u>	<u>LEC</u>	<u>LSS</u>
Limestone - Tons/hr	32.7	33.5
Quicklime - Tons/hr	0	2.5
Water - gpm	513	710
Steam - lb/hr	222,000	222,000
Operating Power - hp	11,019	13,058
Waste - Tons/hr	86	130
 <u>EQUIPMENT COSTS - \$1000 (1989)</u>		
Area 1 - Material Handling	1316	1486
Area 2 - Feed Preparation	0	1805
Area 3 - Gas Handling	3897	3326
Area 4 - SO ₂ Absorption	3716	11690
Area 5 - Stack Gas Reheat	2524	2524
<u>Area 6 - Waste Handling/Regeneration</u>	<u>3651</u>	<u>2703</u>
Total Equipment Costs	15104	23534
 <u>ANNUALIZED COSTS - \$1000 (1989)</u>		
Total Operating and Maintenance	25142	34349
Total Costs	29323	40601

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1.0 INTRODUCTION

The removal of sulfur dioxide from coal-fired combustor flue gases is becoming increasingly important as SO₂ emission limits are lowered due to acid rain legislation and amendments to the Clean Air Act. In order to burn high-sulfur coals economically, low cost SO₂ control technologies for retrofit applications need to be developed. The Limestone Emission Control (LEC) process is currently under development to fulfill this requirement. Experimentation and modeling of the LEC system is continuing, while the economic favorability of the LEC with respect to current Flue Gas Desulfurization (FGD) technologies is undertaken in this study.

The LEC process is a unique system employing standard quarry-sized limestone to remove SO₂ from coal-fired boiler flue-gases. In the LEC process, hot flue gases (< 350°F) are contacted with a bed of 1/32 to 1/4 inch limestone granules covered with a thin film of water. Sulfur dioxide is absorbed from the flue gas into the water film where it subsequently reacts with dissolved limestone. A layer of reaction products, primarily calcium sulfate and calcium sulfite, forms on the surface of the limestone granules as the reaction proceeds.

Initial experiments at the small pilot-scale (400 acfm) unit using a semi-batch fixed-bed LEC reactor have indicated that >90% removal can be achieved using this process (Prudich et al, 1988). A larger (5000 acfm) continuous moving-bed LEC pilot plant is currently installed on a slipstream of the 70,000 lb/hr stoker boiler providing steam to the Ohio University campus in Athens, Ohio.

The LEC process in order to be employed in a full scale commercial operation must not only be able to meet or exceed required government emission levels, but must also be able to compete on an economic basis with current FGD processes. To this, the LEC process is compared on an economic basis to a conventional limestone scrubbing system. These two FGD systems were evaluated for high sulfur coal applications at a hypothetical 500 MW power plant. The conventional limestone scrubbing process equipment list came from an Electrical Power Research Institute (EPRI) technical report titled "Economic Evaluations of FGD Systems; Vol 5" (Keeth et al, 1986). The LEC equipment list was derived from upscaling the LEC moving-bed pilot plant design that is installed at Ohio University to the same inlet and outlet gas flow and composition as used in the development of the limestone scrubbing design. Capital costs for process equipment were estimated from empirical costing charts found in Chemical Engineering Economics (Garrett, 1989). Operating costs were estimated based on expected reagent and utility consumption of each process.

2.0 DESIGN AND ECONOMIC PREMISE

In this study, the economics of two limestone based FGD processes are compared on an equitable basis using conditions that would be expected on the outlet side of an electrostatic precipitator (ESP) of a 500 MW coal-fired power plant. The process design premises used were those developed by an EPRI technical report (Keeth et al, 1986) for a high sulfur coal and were formulated to establish efficiencies, flow rates, SO₂ load, and other operating conditions. The economic premises are used to establish a consistent cost estimation scheme in comparing the two FGD systems and were developed in the final fixed-bed LEC report (Prudich, et al, 1988).

2.1 PROCESS DESIGN PREMISE

The design premises used were developed in the EPRI technical report (Keeth et al, 1986) and are imposed equally upon both the LEC and limestone scrubbing processes.

2.1.1 Emission Standards

New source performance standards (NSPS) established by the EPA in 1979 require that new plants must reduce uncontrolled SO₂ emissions by 89.6% for high sulfur eastern coals. The SO₂ reduction used in this study is 90% of uncontrolled emissions based on a 30-day rolling average.

2.1.2 Power Plant

The power plant used in this study is a new, single unit, pulverized coal-fired producing a nominal 500 MW. The operating parameters of the boiler are assumed to be typical of a new power plant. The particulate control device assumed used is an ESP conforming to current NSPS of 0.03 lb/10⁶ Btu. The capital and operating costs of the power plant and ESP are not included in the analysis. The amount of fly ash produced is equal to 31.6 tons/hr.

2.1.3 Power Plant Operation

A total operating lifetime of the power plant is 165,000 hours over a 15 year period, which equates to 5,500 hr/yr assuming a boiler capacity factor of 68.2%.

2.1.4 Coal and Flue Gas

The coal used in this study is a deep-mined Illinois high sulfur bituminous coal with a sulfur content of 3.96%. The coal analysis is given in Table A-1.

The raw material and utility consumptions are based upon the generating plant operating at 100% load and are shown in Table IV.

2.1.5 General Equipment

2.1.5.1 Stack Gas Reheaters

It is assumed that desulfurized flue gas must be reheated to 50 °F above its wet bulb temperature using steam-heated ambient air. A wet-bulb temperature of 127 °F for the FGD units saturated outlet gas.

2.1.5.2 Duct Work

A 100% flue gas by-pass duct was provided for the FGD process. The by-pass and reheat duct work was sized for a gas velocity of 4200 ft/min with the remainder of the ducting being proportioned to handle a gas velocity of 3,600 ft/min. A pressure drop of 9 inches of water is used as the resistance of the duct work. This does not included the pressure drop caused by the FGD absorption towers or reactors.

2.1.5.3 Equipment Redundancy

Equipment redundancy as been included for equipment that would require immediate operational shutdown due to its on-stream loss. A spare SO₂ absorption train, i.e. the SO₂ absorber, the absorber internals and pumps for the internal sprays has also been added.

2.2 ECONOMIC DESIGN PREMISE

The economic premises used were partially based on those used in "Small Pilot Plant Demonstration of ETS' Limestone Emission Control System - Volumes 1 and 2" (Prudich et al, 1988) and are imposed equally on both the LEC and limestone scrubbing processes. Table A-5 lists the components used in the costs estimations.

2.2.1 Capital Cost Estimating

Capital costs are divided into direct and indirect costs and were developed as summarized in the following sections.

2.2.1.1 Direct Capital Cost

Direct capital costs is the sum of the individual process equipment costs plus the equipment installation cost. The equipment cost is determined by the type and size of the equipment needed by each FGD system. The equipment size and type for the 500 MW limestone scrubber process was generally taken from an EPRI report (Keeth et al, 1986). The I.D. fan, pumps and conveyors of the limestone scrubber were resized using the same technique used to derive the proper sizing of the LEC process equipment. The type of equipment needed for the LEC was that being used in the moving bed pilot project at Ohio University. The size of LEC process conveying equipment was derived from the Chemical Engineers Handbook (Perry, 1985). The LEC reactor dimensions were determined from scaling up the process parameters of the moving bed pilot project.

The estimated installed equipment costs for each piece of process equipment except the SO₂ absorbers and internals were obtained from empirical cost estimating charts in Chemical Engineering Economics (Garrett, 1989). Costs of the SO₂ absorber shells were obtained on a weight basis (\$1.38/lb of carbon steel), then multiplied by a material factor of 2.0 for stainless steel or 1.5 for rubber lined carbon steel. Absorber internals costs for the LEC were obtained from Garrett and manufactures catalogs for the piping and nozzle costs, respectively. The limestone scrubber internal costs for a 28 ft diameter spray tower were proved by Research-Cottrell and were subsequently readjusted for a 38 ft I.D. tower by taking the cross-sectional area ratio to the 0.6 power (See Appendix D).

2.2.1.2 Indirect Capital Cost

Indirect capital costs consist of engineering, construction and file expenses, construction fees, start-up costs, and performance costs. These are defined as percentages of the total direct capital costs, which are listed in Table A-5 along with the contingencies and working capital requirements used.

2.2.2 Operating and Maintenance Costs

2.2.2.1 Direct Operating Costs

Direct operating costs consists of the manpower, utilities and raw material costs that must be outlaid in order for ensure the day to day operation of the process. The labor costs were obtained through the same technique used in the EPA report "Industrial Boiler SO₂ Cost Report" (EPA,1984). It was then assumed that the same labor requirements would be needed by both the LEC and limestone scrubbing processes. Utility costs for water and electricity for each FGD process were obtained through their respective mass balances and electric motor power consumptions. The steam required by the reheaters is equal in both processes since the outlet gas condition of each FGD unit is assumed to be identical. Raw material costs consist of limestone for the LEC process, while the limestone scrubbing process requires limestone and quicklime. The quicklime is used as a waste sludge fixative. Limestone usage rates for the two FGD processes differ due to the different CaCO₃ utilization used in each mass balance. The limestone scrubbing process used a 87% utilization compared to the 90% CaCO₃ utilization employed by the LEC process. Mass rates for each process are given in Table A-2 and A-3 with the unit costs used in the analysis provided in Table A-4.

2.2.2.2 Indirect Operating Costs

Indirect operating costs are incurred by the process in the form of payroll and plant overhead. These are listed in Table A-5.

2.2.3 Annualized Cost

The total annualized cost is the sum of the annual operating and maintenance costs with the capital charges. The parameters used were an interest rate of 6.10% with a plant life expectancy of 15 years with an annual plant operation of 5,500 hours. Table A-5 lists the complete derivation of this term.

3.0 PROCESS DESCRIPTIONS

3.1 Conventional Limestone Scrubber

The design and process parameters used for the 500 MW limestone scrubber were developed by EPRI (Keeth et al, 1986).

3.1.1 Chemistry

The absorption reagent in the limestone scrubber process is CaCO_3 , which contacts the SO_2 laden flue gas in a water slurry. The SO_2 forms the bisulfite ion upon dissolution into the liquid causing the pH to drop. The pH drop causes the solid CaCO_3 to dissolve; thus forming the calcium sulfate/sulfite waste sludge product. This reaction occurs in an open spray tower with a recirculation tank located below it. This design was used by EPRI and subsequently in this analysis, for it is an established FGD technique widely accepted by the utility industry. The overall reaction scheme is as follows:

The mole ratio of sulfite to sulfate produced is usually 9 to 1. Unfortunately, the sulfate crystals are more desirable since their larger size means less entrained water and a faster settling rate which results in a smaller required thickener and vacuum filter system (Keeth et al, 1986). $\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$ is also known as gypsum, a resalable raw material for the dry-wall industry.

3.1.2 Process Design

The limestone scrubbing process is partitioned into following 6 areas of operation:

Area 1	-	Material Handling
Area 2	-	Feed Preparation
Area 3	-	Gas Handling
Area 4	-	SO_2 Absorption
Area 5	-	Stack Gas Reheat

Area 6 - Waste Handling

The process equipment list showing size, cost and power consumption is included in Appendix B.

3.1.2.1 Area 1 - Material Handling

This area includes all of the facilities required for the receiving and storage of raw limestone and quicklime delivered by railcar to a 30-day storage pile for limestone and a 60-day quicklime storage silo (EPA, 1981).

3.1.2.2 Area 2 - Feed Preparation

The limestone day bins supply one ball mill that grinds the limestone to 90 percent less than 200 mesh screen. A 60% solids slurry exits the ball mill and is subsequently diluted to 20% in the slurry feed tank (Keeth et al, 1986).

3.1.2.3 Area 3 - Gas Handling

The flue gas handling system is sized to handle 1,884,000 acfm at 283°F at a flow velocity of 3,600 ft/min. Outlet gas from the absorbers is assumed to be saturated at 127 °F. One fan for each absorption train (3 operating, 1 spare) are located after the ESP and are sized to handle a 15 inch pressure drop. Inlet and outlet manifolds connect common duct work to the four absorber trains. Guillotine dampers provide in-duct isolation of individual absorbers while a FGD by-pass duct is also included (Keeth et al, 1986).

3.1.2.4 Area 4 - SO₂ Absorption

The absorption process occurs within three operating and one spare vertical, countercurrent absorption towers. Each tower is designed to handle 628,000 acfm with a superficial gas velocity of 10 ft/s and a gas residence time of 10 seconds at a pressure drop through the tower of 6 in H₂O. The internal sprays provide a liquid to gas ratio of 80 gpm/1000 saturated acfm and also act as a gas distributor. A mist eliminator is also provided in each absorption tower. The towers shells are constructed of rubber-lined carbon steel to prevent corrosion.

Located at the bottom of each absorption tower is a reaction mix tank into which the exposed scrubber slurry drains. This agitated tank is sized to provide 6 minutes of slurry retention time to enhance reaction completeness and allow time for crystalline growth. Make-up water along with fresh limestone slurry from the slurry feed tank are added here in order to maintain the proper 1.15 stoichiometric ratio. A purge stream from the reaction mix tank is pumped to the waste handling area (Keeth et al, 1986).

3.1.2.5 Area 5 - Stack Gas Reheat

A stack gas reheater is used to increase the cleansed flue gas dry-bulb temperature to 50°F above its wet-bulb temperature using steam (Keeth et al, 1986). This is done in order to give the stack gas its required plume and also prevents in-chimney condensation. The reheater is constructed of Inconel® and Corten® (EPA, 1981).

3.1.2.6 Area 6 - Waste Handling

The purge streams from the three operating reaction tanks is pumped into a single thickener system which includes a clarifier, under and over-flow tanks, a vacuum filter, and the required pumping equipment. The clarifier was designed on 18 ft² per ton of dry solids per day with flocculent being added to aid in solids settling. Supernatant is returned to the absorption trains from the over-flow tanks, while the clarification underflow is pumped to the filter feed tank prior to mechanical dewatering by a rotary vacuum filter (Keeth et al, 1986).

The filter cake from the vacuum filter is combined with pneumatically conveyed fly ash and quicklime in a pug mill to form a fixed sludge. The sludge is then belt conveyed to a truck loading site for eventual landfill disposal.

3.2 Limestone Emission Control

The design and process parameters used for the 500 MW limestone scrubber were developed by ETS, Inc.

3.2.1 Chemistry

The absorption reagent used in the LEC process is CaCO₃ in the form of standard quarry limestone in the size range of 1/32 to 1/4 inches. Hot SO₂ laden flue gas passes through the cross section of a vertical gravity fed bed of limestone that is covered with a thin water film. The source of the water may be from the condensation of flue gas moisture or from direct water spray.

Sulfur dioxide diffuses through the water film forming the bisulfite ion which reduces the pH and dissolves the CaCO₃. The products of the ionic reaction are calcium sulfite/sulfate and follow the same reaction mechanism as in the limestone scrubber process as shown in Equations (3-1) and (3-2).

A layer of sulfite/sulfate reaction product forms on the surface of the limestone granules as the absorption process proceeds that retards the reaction rate due to active CaCO₃ site blinding. The limestone granule is then reactivated by mild abrasion to remove the solid calcium sulfite/sulfate product layer subsequently exposing active calcium carbonate sites.

3.2.2 Process Design

The LEC process is partitioned into following 6 areas of operation:

- Area 1 - Material Handling
- Area 2 - Feed Preparation
- Area 3 - Gas Handling
- Area 4 - SO₂ Absorption
- Area 5 - Stack Gas Reheat
- Area 6 - Regeneration

The process equipment list showing size, cost and power consumption is included in Appendix C.

3.2.2.1 Area 1 - Material Handling

This area includes all of the facilities required for the receiving and storage of raw limestone delivered by railcar to a 30-day storage pile for limestone (EPA, 1981).

3.2.2.2 Area 2 - Feed Preparation

The LEC process uses standard quarry sized limestone; therefore no feed preparation is necessary.

3.2.2.3 Area 3 - Gas Handling

The flue gas handling system is sized to handle 1,884,000 acfm at 283°F at a flow velocity of 3,600 ft/min. Outlet gas from the absorbers is assumed to be saturated at 127 °F. One fan for each absorption train (3 operating, 1 spare) are located after the ESP and are sized to handle a 19 inch pressure drop. Inlet and outlet manifolds connect common duct work to the four absorber trains. Guillotine dampers provide in-duct isolation of individual absorbers while a FGD by-pass duct is also included (Keeth et al, 1986).

Inlet and outlet plenums are required to redirect the flue gas evenly over the 100 ft tall by 52 ft wide face area of the four LEC reactors. The inlet ducting to the reactor is made of carbon steel, while the outlet plenums are composed of stainless steel to prevent corrosion. Appendix F contains the LEC plenum design.

3.1.2.4 Area 4 - SO₂ Absorption

The absorption process occurs within three operating and one spare LEC moving bed reactors. Each reactor is designed to handle 628,000 acfm with a superficial gas velocity of 2 ft/s and a pressure drop through the reactor of 10 in H₂O. Internal overhead

sprays provide the water necessary to keep a liquid film on the limestone bed and to replace the water evaporated by the flue gas. A live limestone hopper is positioned at the top of each reactor to act as an air seal, while 70° angle finned louvers juxtaposition the plenums and reactor core to prevent limestone from entering the flue gas duct work. The reactor box, live hopper and louvers are constructed of stainless steel to prevent corrosion.

Continuous bucket elevators lift the limestone discharged from weigh belt conveyors to the top of the live hoppers. A twin screw conveyor located the bottom of each reactor regulates stone flow through the bed. Centrifugal pumps are used to pump the water to the overhead spray nozzles located at the top of the live hopper.

3.2.2.5 Area 5 - Stack Gas Reheat

A stack gas reheater is used to increase the cleansed flue gas dry-bulb temperature to 50°F above its wet-bulb temperature using steam (Keeth et al, 1986). This is done in order to give the stack gas its required plume and also prevents in-chimney condensation. The reheater is constructed of Inconel® and Corten® (EPA, 1981).

3.2.2.6 Area 6 - Regeneration

Bucket elevators for each LEC reactor (3 operating, 1 spare) lift the reacted limestone to a horizontal screw conveyor which is positioned overhead of the regeneration equipment in order to use a gravity feed system for the limestone.

The regeneration system for each LEC reactor consists of a Marcy® ball mill with a double-deck vibrating screen that abrades the limestone and separates the reactivated limestone from the removed sulfate/sulfite reaction products, respectively.

The sulfite/sulfate reaction products are then belt conveyed to a truck loading site for eventual landfill disposal. The regenerated limestone is returned to the top of the live hopper for further flue gas exposure.

4.0 Discussion of Results

The economics of the LEC process as compared to a conventional limestone scrubber can be readily analyzed through the comparison of the capital and operating costs of the equipment used in both processes.

4.1 Capital Costs

A comparison of the capital costs of the two FGD systems by equipment area is given in Table II.

4.1.1 Area 1 - Material Handling

The material handling sections of both the LEC and limestone scrubbing process have identical limestone storage and transfer equipment which cost approximately \$1,239,000. The limestone scrubbing process requires a quicklime pneumatic conveying system and storage silo. The LEC process uses no waste product fixative and therefore has no need for the quicklime processing equipment. This results in a LEC cost advantage of \$170,000.

4.1.2 Area 2 - Feed Preparation

The limestone scrubber uses a ball mill, two tanks, and six pumps in this process area to produce a usable limestone slurry at a cost of \$1,805,000. The LEC process uses standard quarry sized limestone; therefore no feed preparation is necessary. This results in a LEC cost advantage of \$1,805,000.

4.1.3 Area 3 - Gas Handling

The four induced draft fans for both the LEC and limestone scrubber are sized to handle 628,000 acfm with a 19 in. H₂O system pressure drop for the LEC and a 15 in. H₂O system pressure drop for the limestone scrubber. This results in a capital cost disadvantage for the LEC of \$15,600.

The duct-work for the LEC and limestone scrubber are identical except for the inlet and outlet plenums required to distribute the flue gases evenly across the face of the LEC reactor. The outlet plenum has an added cost of being made of stainless steel. The cost disadvantage to the LEC in added duct-work is \$556,000 which represents 17% more in capital costs for duct-work over the limestone scrubbers duct-work.

The total gas handling disadvantage for the LEC is \$571,600 which represents an additional 17% capital outlay for the LEC over the limestone scrubber process.

4.1.4 Area 4 - SO₂ Absorption

The SO₂ absorption area's capital costs for the two FGD systems diverge significantly. This results in the SO₂ absorption area being the most economically critical. Using three operating absorption trains with one spare results in a LEC cost advantage of \$7,974,000, which is slightly less than the limestone scrubber capital costs for equipment in areas 1, 2 and 3.

Most of the added cost of the spray tower is in its internals, i.e. the mist eliminator, spray nozzles and piping. This is due to the cross sectional area of the spray tower (1134 ft²) being greater than the cross section area of the LEC (65 ft²); thus resulting in more piping and spray nozzles required assuming a standard spray coverage area. Another factor is that the piping for the LEC is in a single deck of sprays while a standard scrubber spray tower

generally has four decks of spray piping (Keeth et al, 1986).

The LEC internals consists of only the stainless steel louvers. The cost difference between the louvers and the internals of the spray tower is approximately \$500,000 per absorber.

4.1.5 Stack Gas Reheat

The inline steam reheaters used in the LEC and limestone scrubber processes are identical.

4.1.6 Waste Handling/Regeneration

The LEC has regeneration area has a cost disadvantage of approximately \$948,000 compared to the limestone scrubbers waste handling section. This is due to each LEC train has its own complete regeneration area, i.e. three operating with one spare regeneration section. Each regeneration section contains one Marcy® ball mill, one double-deck vibrating screen and one waste conveyor, while the limestone scrubber has a single waste handling section treating the scrubber sludge produced by the three operating absorption trains.

4.2 Operating Costs

A comparison of the operating costs of the two FGD systems by equipment area is given in Table III. Table IV lists the raw material costs and power consumption a 500 MW power plant.

4.2.1 Reagent Costs

The reagent costs for the LEC are lower for two reasons. First, the limestone costs for the LEC are lower since the utilization of the CaCO_3 is assumed to be greater (90% as compared to 87%) for the LEC process. Second, the LEC process does not require a fixative for its waste product since it is in solid form, while the limestone scrubber employs quicklime as a sludge fixative.

4.2.2 Utility Consumption

The utility costs for the LEC are lower than that of the same size limestone scrubber for two reasons. First, the LEC process requires less water since only a thin water film is necessary for the reaction, while the scrubber process requires a 20% by weight limestone slurry. Even with a water recycle loop, the limestone scrubber process still requires 197 gpm more water than the LEC process. Secondly, the power required to operate the fluid handling pumps for the scrubber process is greater than that required for the solid material handling equipment used in the LEC process by 1579 hp as shown in Table IV.

5.0 Conclusions

The removal of sulfur dioxide from coal-fired combustor flue gases is becoming increasingly important as SO₂ emission limits are becoming stricter in the advent of recent amendments to the Clean Air Act. In order to burn high-sulfur coals economically, low cost SO₂ control technologies available for retrofit applications need to be developed. The Limestone Emission Control (LEC) process is such a technology as shown in this economic comparison between it and a conventional limestone scrubbing process when applied to a 500 MW coal-fired utility.

The capital costs of the two FGD systems is \$15,104,000 and \$23,534,000 for the LEC and limestone scrubber processes, respectively. This represents a capital cost difference of 36%. The annual operating cost of the LEC are also very favorable as compared to a limestone scrubber. The annual costs are \$25,147,000 and \$34,342,000 for the LEC and limestone scrubber processes, respectively and represents an operational savings of \$9,195,000 per year. The annualized cost comparison for the LEC and limestone scrubber processes are shown in Table I.

The LEC process is mechanically very simple, i.e. limited reactor internals and a low operational temperature (stone temperature will not exceed the wet-bulb temperature of the flue-gas, < 160°F); thus resulting in low maintenance costs. The LEC has a small footprint when compared to other FGD systems, which makes the LEC process suitable for easy retrofitting.

The LEC FGD process is shown in this economic analysis to have savings over the limestone scrubber process in both the areas of capital and operational expenditures along with the added advantage of having a lower projected maintenance cost due to its mechanical simplicity. For the three reasons of lower capital, operational and maintenance costs, the LEC process is clearly a more economically viable process than a conventional limestone scrubbing system.

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TABLE I

ANNUALIZED COST COMPONENTS FOR A 500 MW FACILITY

\$1000 (1989)

<u>Capital Costs</u>		<u>LEC</u>	<u>LSS</u>
(1)	Total Direct Cost - TDC (Equipment + Installation)	15104	23534
(2)	Indirect Costs - ID (33% TDC)	4984	7766
(3)	Contingences - Cont. (20% of TDC + ID)	4018	6260
(4)	Total Turnkey Cost - TK (TDC + ID + Cont.)	24106	37560
(5)	Working Capital - WC (25% total O&M)	6285	8587
<u>Annual Operating Costs</u>			
(1)	Reactant Cost - RC Limestone Quicklime	1680 0	1723 600
(2)	Operating and Maintenance (excluding RC) Direct Labor Supervisor Labor Maintenance (8% of TDC) Electricity Water Steam FGD Waste Disposal Fly Ash Disposal	187 37 1208 2199 42 9186 6195 4407	187 37 1883 2613 59 9186 18061 0
(3)	Total Operating Costs - Total O&M (RC + O&M)	25142	34349
<u>Total Annualized Costs</u>			
(1)	Capital Recovery - CR (10.36% TK)	2497	3891
(2)	Working Capital Charges - WCC (10% WC)	629	859
(3)	Miscellaneous - Misc. (4% TK)	964	1502
(4)	Total Capital Charges - TCC (CR + WCC + Misc.)	4090	6252
(5)	Total Annualized Costs - TAC (TCC + Total O&M)	29232	40601

TABLE II
COST COMPARISON BY EQUIPMENT AREA
FOR A 500 MW FACILITY

	<u>\$1000 (1989)</u>	
	<u>LEC</u>	<u>LSS</u>
Area 1 - Material Handling	1316	1486
Area 2 - Feed Preparation	0	1805
Area 3 - Gas Handling	3897	3326
Area 4 - SO2 Absorption	3716	11690
Area 5 - Stack Gas Reheat	2524	2524
<u>Area 6 - Waste Handling/Regeneration</u>	<u>3651</u>	<u>2703</u>
Total	15104	23534

TABLE III

ANNUALIZED OPERATIONAL COSTS BY AREA
(EXCLUDING REAGENT COSTS)

		<u>\$1000 (1989)</u>	
		<u>LEC</u>	<u>LSS</u>
<u>Area 1 - Material Handling</u>			
Water		0	0
Electricity		28	77
Steam		0	0
<u>Area 2 - Feed Preparation</u>			
Water		0	7
Electricity		0	166
Steam		0	0
<u>Area 3 - Gas Handling</u>			
Water		0	0
Electricity	(15 inch system pressure drop)	1268	1268
	(19 inch system pressure drop)*	1606	N/A
Steam		0	0
<u>Area 4 - SO2 Absorption</u>			
Water		42	2
Electricity		61	1038
Steam		0	0
<u>Area 5 - Stack Gas Reheat</u>			
Water		0	0
Electricity		0	0
Steam		9186	9186
<u>Area 6 - Waste Handling</u>			
Water		0	50
Electricity		504	64
Steam		0	0
<u>Total Process</u>			
Water		42	59
Electricity	(15 inch system pressure drop)	1861	2613
	(19 inch system pressure drop)*	2199	N/A
Steam		9186	9186
<u>Total Operating Costs (excluding reagent costs)</u>			
	(15 inch system pressure drop)	11089	11858
	(19 inch system pressure drop)*	11427	N/A

* The 19 inch system pressure drop was used in this analysis for the LEC.

TABLE IV
RAW MATERIAL AND UTILITY CONSUMPTION
FOR A 500 MW FACILITY

<u>ITEM</u>	<u>LEC</u>	<u>LSS</u>
Limestone - Tons/hr	32.7	33.5
Quicklime - Tons/hr	0	2.5
Water - gpm	513	710
Steam - lb/hr	222,000	222,000
Operating Power - hp		
Area 1 - Material Handling	141	382
Area 2 - Feed Preparation	0	826.5
Area 3 - Gas Handling	8,028	6,339
Area 4 - SO ₂ Absorption	327	5,190
Area 5 - Stack Gas Reheat	0	0
Area 6 - Waste Handling/Regeneration	2,523	320
Total Power - hp	11,019	13,057.5
Waste Solids - Tons/hr	54.4	0
Waste Fixed Sludge - Tons/hr	0	129.5
Fly Ash - Tons/hr	31.6	0

TABLE V

INSTALLED EQUIPMENT COSTS AS PERCENT OF TOTAL COST BY AREA
FOR A 500 MW FACILITY

	<u>PERCENT (%)</u>	
	<u>LEC</u>	<u>LSS</u>
Area 1 - Material Handling	8.7	6.3
Area 2 - Feed Preparation	0.0	7.7
Area 3 - Gas Handling	25.8	14.1
Area 4 - SO ₂ Absorption	24.6	49.7
Area 5 - Stack Gas Reheat	16.7	10.7
<u>Area 6 - Waste Handling/Regeneration</u>	<u>24.2</u>	<u>11.5</u>
Total - \$1000 (1989)	15104	23534

TABLE VI

OPERATING AND MAINTANANCE COST AS A PERCENT OF TOTAL O&M
FOR A 500 MW FACILITY

		<u>PERCENT (%)</u>	
		<u>LEC</u>	<u>LSS</u>
<u>Annual Operating Costs</u>			
(1)	Reactant Cost - RC		
	Limestone	6.7	5.0
	Quicklime	0.0	1.7
(2)	Operating and Maintenance (excluding RC)		
	Direct Labor	0.7	0.5
	Supervisor Labor	0.1	0.1
	Maintenance (8% of TDC)	4.8	5.5
	Electricity	8.7	7.6
	Water	0.2	0.2
	Steam	36.5	26.7
	FGD Waste Disposal	24.6	52.6
	Fly Ash Disposal	17.5	0.0
(3)	Total Operating Costs - Total O&M (RC + O&M) in \$1000 (1989)	25142	34349

Figure 1a Conventional Limestone PFC SO₂ Removal and Reagent Handling Systems

Figure 16 Conventional Limestone PFD Waste Handling System

Limestone Emission Control Process. Flow Diagram

Figure 2

APPENDIX A
MISCELLANEOUS TABLES

TABLE A-1
HIGH SULFUR COAL COMPOSITION*

Proximate Analysis - %	
Moisture	12.0
Ash	16.0
Fixed Carbon	39.0
Volatile Matter	33.0
Ultimate Analysis - %	
Moisture	12.0
Carbon	57.5
Hydrogen	3.7
Nitrogen	0.9
Sulfur	4.0
Chloride	0.1
Ash	16.0
Oxygen	5.8
Btu/lb	10.100
Sulfur Content - lb/10 ⁶ Btu	3.96

* "Economic Evaluation of FGD Systems, Volume 5 (Keeth, 1989)

TABLE A-2

CONVENTIONAL LIMESTONE SCRUBBER PROCESS MATERIAL BALANCE

STREAM NUMBER	1	2	3	4	5	6	7	8	9
Temperature, °F	277	283	127	127	127	177	60	382	572
Pressure, "H ₂ O	-15	+10	+9	+4	+4	+1	0	+4	-
10 ³ ACFM	1990	628	510	516	1032	2123	338	550	-
CO ₂ , 10 ³ lb/hr	1043.0	347.7	347.7	355.7	712.0	1068.0	-	-	-
H ₂ O, 10 ³ lb/hr	296.0	98.7	176.7	176.7	354.0	551.0	20.0	20.0	111.0
N ₂ , 10 ³ lb/hr	4196.0	1398.7	1398.7	1398.7	2798.0	5399.0	1160.0	1160.0	-
SO ₂ , 10 ³ lb/hr	39.2	13.1	13.1	1.3	2.6	3.9	-	-	-
O ₂ , 10 ³ lb/hr	377.0	125.7	125.7	125.7	252.0	698.0	352.0	352.0	-
Total, 10 ³ lb/hr	5951.2	1983.9	2061.9	2058.1	4118.6	7719.9	1532.0	1532.0	111.0

STREAM NUMBER	10	11	12	13	14	15	16	17	18	19	20	21
Temperature, °F	60	90	70	126	126	60	60	90	70	125	60	60
GPM	170	195	500	41300	540	-	85	590	450	1620	605	10
CaSO ₃ ·½H ₂ O, 10 ³ lb/hr	-	-	-	1636.0	21.4	-	-	-	-	64.1	-	-
CaSO ₄ ·2H ₂ O, 10 ³ lb/hr	-	-	-	244.0	3.2	-	-	-	-	9.5	-	-
Inerts, 10 ³ lb/hr	-	1.3	-	100.0	1.3	4.0	-	4.0	-	4.0	-	-
CaCO ₃ , 10 ³ lb/hr	-	21.0	-	216.0	2.8	63.0	-	63.0	-	8.3	-	-
H ₂ O, lb/hr	84.0	89.2	250.0	19764.0	258.0	-	42.6	267.6	225.0	773.0	303.0	4.0
Total, lb/hr	84.0	111.5	250.0	21960.0	286.7	67.0	42.6	334.6	225.0	858.9	303.0	4.0

TABLE A-2 (Continued)

CONVENTIONAL LIMESTONE SCRUBBER PROCESS MATERIAL BALANCE

STREAM NUMBER	22	23	24	25	26	27	28	29	30	31	32
Temperature, °F	60	70	70	70	70	70	70	60	60	80	80
GPM	-	10	1344	568	1950	1500	318	305	-	-	-
$\text{CaSO}_3 \cdot \frac{1}{2}\text{H}_2\text{O}$, 10^3 lb/hr	-	-	-	64.1	-	-	64.1	-	-	-	64.1
$\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$, 10^3 lb/hr	-	-	-	9.5	-	-	9.5	-	-	-	9.5
Inerts, 10^3 lb/hr	0.02	0.62	-	4.0	-	-	4.0	-	0.5	63.0	67.5
CaCO_3 , 10^3 lb/hr	-	-	-	8.3	-	-	8.3	-	-	-	8.3
CaO , 10^3 lb/hr	-	-	-	-	-	-	-	-	4.5	-	4.5
H_2O , 10^3 lb/hr	-	4.0	672.0	258.0	975.0	750.0	105.0	153.0	-	-	105.0
Total, lb/hr	0.02	4.02	672.0	343.9	975.0	750.0	190.9	153.0	5.0	63.0	258.9

TABLE A-3

LIMESTONE EMISSION CONTROL PROCESS MATERIAL BALANCE

STREAM NUMBER	1	2	3	4	5	6	7	8	9
Temperature, °F	277	283	127	127	127	177	60	382	572
Pressure, "H ₂ O	-15	+10	+9	+4	+4	+1	0	+4	-
10 ³ ACFM	1990	628	510	516	1032	2123	338	550	-
CO ₂ , 10 ³ lb/hr	1043.0	347.7	347.7	355.7	712.0	1068.0	-	-	-
H ₂ O, 10 ³ lb/hr	296.0	98.7	176.7	176.7	354.0	551.0	20.0	20.0	111.0
N ₂ , 10 ³ lb/hr	4196.0	1398.7	1398.7	1398.7	2798.0	5399.0	1160.0	1160.0	-
SO ₂ , 10 ³ lb/hr	39.2	13.1	13.1	1.3	2.6	3.9	-	-	-
O ₂ , 10 ³ lb/hr	377.0	125.7	125.7	125.7	252.0	698.0	352.0	352.0	-
Total, 10 ³ lb/hr	5951.2	1983.9	2061.9	2058.1	4118.6	7719.9	1532.0	1532.0	111.0

STREAM NUMBER	10	11	12	13	14	15	16	17	18	19
Temperature, °F	60	60	60	60	127	60	60	60	60	60
GPM	-	-	-	-	-	171	-	-	-	-
CaSO ₄ ·2H ₂ O, 10 ³ lb/hr	-	-	-	-	31.7	-	-	31.7	63.4	95.1
Inerts, 10 ³ lb/hr	3.9	2.6	1.3	11.8	11.8	-	10.5	1.3	2.6	3.9
CaCO ₃ , 10 ³ lb/hr	61.4	40.9	20.5	184.0	166.0	-	164.0	2.1	4.1	6.1
H ₂ O, 10 ³ lb/hr	-	-	-	8.3	8.6	85.5	7.4	1.2	2.4	3.6
Total, 10 ³ lb/hr	65.3	43.5	21.8	204.1	218.1	85.5	181.9	36.3	72.5	108.7

TABLE A-4

UNIT COSTS USED IN COST CALCULATIONS¹Utilities

Electricity	0.0438/kWh
Water	\$0.25/10 ³ gal

Raw Materials

Limestone	\$9.35/ton
Quicklime	\$48.00/ton

Labor

Direct Labor	\$21.40/man-hour
Supervision	\$27.82/man-hour

Waste Disposal

Solids (Ash, Spray Dried Solids)	\$25/ton
Sludge	\$25/ton
Liquid	\$0.67/10 ³ gal

Construction

Fabricated Carbon Steel	\$1.38/lb

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¹All costs in 1989 \$.

TABLE A-5

COMPONENTS USED IN COST ESTIMATIONSA. Capital Cost Components

(1) Direct Costs

Equipment
 + Installation
 = Total Direct Costs

(2) Indirect Costs

Engineering (10% of total direct costs)
 + Construction and Field Expenses (10% of total direct costs)
 + Construction Fees (10% of total direct costs)
 + Start-up Costs (2% of total direct costs)
 + Performance Costs (1% of total direct costs)

 = Total Indirect Costs

(3) Contingencies = 20% of (Total Indirect + Total Direct Costs)

(4) Total Turnkey Cost = Total Indirect Cost + Total Direct Cost +
 Contingencies

(5) Working Capital = 25% of Total Operating and Maintenance Costs

(6) Total Capital Cost = Total Turnkey + Working Capital

B. Operating and Maintenance

(1) Direct Operating Costs

Operating Labor
 + Supervision
 + Maintenance Labor, Replacement Parts and Supplies
 + Electricity
 + Water
 + Steam
 + Waste Disposal
   Solids (Fly ash and bottom ash)
   Sludge
   Liquid
 + Reactants
 = Total Direct Operating Costs

TABLE A-5 (continued)

- (2) Indirect Operating Costs (Overhead)
  - + Payroll (30% Operating Labor)
  - + Plant (26% of Operating Labor + Supervision + Maintenance Costs + Replacement Parts)
- (3) Total Annual Operating and Maintenance Costs = Total Direct + Total Indirect Costs

C. Annualized Cost Components

- (1) Total Annualized Cost = Annual O&M Costs + Capital Charges
- (2) Capital Charges = Capital Recovery + Interest on Working Capital + Miscellaneous (G&A, taxes and insurance)
- (3) Calculation of Capital Charges Components

A. Capital Recovery = Capital Recovery Factor (CRF) x Total Turnkey Cost

$$CRF = \frac{i(1+i)^n}{(1+i)^n - 1}$$

i = interest rate

n = number of years of useful life of boiler or control system

$\frac{n}{15}$	$\frac{i}{6.10\%}$	$\frac{CRF}{0.1036}$

B. Interest on Working Capital = 10% of Working Capital

C. G&A, taxes and insurance = 4% of total turnkey cost

APPENDIX B
CONVENTIONAL LIMESTONE SCRUBBER PROCESS EQUIPMENT LIST

LIMESTONE SCRUBBER PROCESS
EQUIPMENT LIST, DESCRIPTION, AND COST
(500 MW FACILITY)

ID	ITEM	No.	Description	Total Power (hp)	Installed Cost (\$)	Annual Utilities Required (\$)	Ref. No.
Area 1 - Material Handling							
1-1	Car Shaker	1	Top mounted with crane	0.0	96337	0	E
1-2	Car Puller	1	25 hp puller, 5 hp return	30.0	93841	6002	E
1-3	Hopper, L/S Unloading	1	16 ft dia X 10 ft high, 6 in. square grating	0.0	16924	0	E
1-4	Feeder, L/S Unloading	1	Vibrating pan, 42 in. wide x 20 ft long, 250 tons/hr	3.0	36788	600	E
1-5	Conveyor, L/S Unloading	1	Belt, 36 in. wide X 20 ft long, 250 tons/hr, 130 ft/min	5.0	30429	1000	E
1-6	Conveyor, L/S Stocking (Incline)	1	Belt, 36 in. wide X 310 ft long, 250 tons/hr, 130 ft/min	50.0	221695	10003	E
1-7	Dust Collecting System	1	Bag Filter, polypropylene bag, 2200 acfm, auto shaker system, with dust hood	7.5	31844	1500	E
1-8	Unloading Pit Sump Pump	1	Centrifugal, 60 gpm, 70 ft head stainless steel	2.0	5679	400	E
1-9	Conveyor, L/S Stocking	1	Belt, 36 in. wide X 200 ft long, 250 tons/hr, 130 ft/min	5.0	165185	1000	E
1-10	Tripper	1	30 ft/min	1.0	41190	200	E
1-11	Mobile Equip.	1	Scraper Tractor	0.0	161016	0	E
1-12	Hopper, reclaim	2	7 ft X 4.25 ft X 2 ft deep, 419 gal, carbon steel	0.0	1922	0	E
1-13	Feeder, L/S live Storage	2	Vibrating Pan	3.0	36788	1200	E
1-14	Tunnel Sump Pump	2	Vertical, 60 gpm, 70 ft head, stainless steel (1 operating, 1 spare)	5.0	11358	1000	E
1-15	Conveyor, Live L/S Feed	1	Belt, 16 in. wide X 200 ft long 100 tons/hr, 60 ft/min	2.0	101546	400	E
1-16	Conveyor, Live L/S Feed (Incline)	1	Belt, 16 in. wide X 200 ft long, 35 ft lift, 100 tons/hr, 60 ft/min	3.0	96201	600	E
1-17	Elevator, live L/S Feed	1	Continuous Bucket, 10 in. X 7 in. X 11.75 in., 90 ft lift, 100 tons/hr, 160 ft/min	10.0	39709	2001	E

LIMESTONE SCRUBBER PROCESS

EQUIPMENT LIST, DESCRIPTION, AND COST (500 MW FACILITY)

ID	ITEM	No.	Description	Total Power (hp)	Installed Cost (\$)	Annual Utilities Required (\$)	Ref. No.
1-18	Bin, L/S Day	3	6,017 gal, 30 hr capacity, w/ cover, carbon steel	0.0	9087	0	E
1-19	Conveyor, Feed Belt	1	Belt, 16 in. wide X 60 ft long, 100 tons/hr, 60 ft/min, including tripper	3.0	36998	600	E
1-20	Dust Collecting System	1	Bag Filter, polypropylene bag, 2200 acfm, auto shaker system	7.5	4409	1500	E
1-21	Positive Pressure Blower	1	Rotary, used for pneumatic lime conveying, 3170 scfm	225.0	80836	45012	E
1-22	Fixative Reagent Storage Tank	1	Quicklime storage, 56,400 gal, carbon steel	0.0	80390	0	E
1-23	Fixative Storage Baghouse	1	10,000 acfm	20.0	85735	4001	E
Subtotal: Area 1					1485907	77021	
Area 2 - Feed Preparation							
2-1	Feeder, Crusher	1	Weigh belt, 16 in. wide x 14 ft long, 34 tons/hr	2.0	27791	400	G
2-2	Ball Mill	2	Wet, open system, 34 tons/hr (1 operating, 1 spare)	700.0	653032	140038	G
2-3	Tank, Mills	1	21058 gal, open top, carbon steel, glass lined	0.0	187058	0	G
2-4	Agitator, Mills Product Tank	1	36 in. dia, stainless steel	15.0	12720	3001	G
2-5	Pump, Mills Product Tank	3	Centrifugal, 51 gpm, 60 ft head, stainless steel (2 operating, 1 spare)	2.0	16836	1200	G
2-6	Tank, L/S Slurry Feed	1	329,000 gal, glass lined, open top, carbon steel	0.0	851061	0	G
2-7	Agitator, Feed Slurry Tank	1	83 in. dia, stainless steel	80.0	29150	16004	G
2-9	Pump, Slurry Feed Tank	3	Centrifugal, 25 gpm, 60 ft head, stainless steel	1.0	14028	600	G
2-10	Dust Collecting System, Ball Mill	3	Bag Filter, polypropylene bag, 2200 acfm, auto shaker system	7.5	13227	4501	G
Subtotal: Area 2					1804903	165745	

LIMESTONE SCRUBBER PROCESS

EQUIPMENT LIST, DESCRIPTION, AND COST (500 MW FACILITY)

ID	ITEM	No.	Description	Total Power (hp)	Installed Cost (89\$)	Annual Utilities Required (89\$)	Re N
Area 3 - Gas Handling							
3-1	Fans	4	Induced draft, 628,000 acfm @ 283 F, 15 in. H2O static head, axial fan type (3 operating, 1 spare)	2113	212272	1268147	
3-2	Inlet Ducting from Plant to ESP	1	Flue gas service, rectangular with external stiffeners, 14,570 ft2 of carbon steel	0.0	195275	0	
3-3	Outlet Ducting from ESP to fan	1	Flue gas service, rectangular with external stiffeners, 13,990 ft2 of carbon steel	0.0	187501	0	
3-4	Inlet Ducting from I.D. to Absorber Manifold	1	Flue gas service, rectangular with external stiffeners, 8,015 ft2 of carbon steel	0.0	107421	0	
3-5	Inlet Ducting, Absorber Manifold	1	Flue gas service, rectangular with external stiffeners, 13,210 ft2 of carbon steel	0.0	177047	0	
3-6	Outlet Ducting from Absorber (including manifold)	1	Flue gas service, rectangular with external stiffeners, 23,860 ft2 of stainless steel	0.0	639569	0	
3-7	FGD Bypass	1	Flue gas service, rectangular with external stiffeners, 17960 ft2 of carbon steel	0.0	240709	0	
3-8	Chimney Transition	1	Flue gas service, rectangular with external stiffeners, 1890 ft2 of carbon steel	0.0	25331	0	
3-9	I.D. Fan Inlet Isolation Damper	2	Spade Guillotine w/ purge air Duct size: 10' x 18'	0.0	186093	0	
3-10	Scrubber Inlet Isolation Damper	4	Double Guillotine w/purge air Duct Size: 14' x 18'	0.0	594596	0	
3-11	Scrubber Outlet Isolation Damper	4	Double-Louvre, opposed and parallel blade w/purge air Duct Size: 10' x 17'	0.0	669027	0	
3-12	FGD System Bypass Damper	1	Double-Louvre, opposed and parallel blade w/purge air Duct Size: 10' x 17'	0.0	91004	0	
Subtotal: Area 3					3325845	1268147	

LIMESTONE SCRUBBER PROCESS

EQUIPMENT LIST, DESCRIPTION, AND COST (500 MW FACILITY)

ID	ITEM	No.	Description	Total Power (hp)	Installed Cost (\$)	Annual Utilities Required (\$)	Ref. No.
Area 4 - SO2 Absorption							
4-1a	SO2 Absorber Shell	4	Vertical countercurrent spray tower, 38' I.D. x 100' straight side, 1/4 in. carbon steel, rubber lined (3 operating, 1 spares) Design Basis: Superfiscial gas velocity =10 ft/s Gas Residence time = 10 s	0.0	2284824	0	D
4-1b	SO2 Absorber Mist Eliminator	4	Horizontal Chevron Cost provided by R.C. (3 operating, 1 spares)	0.0	418360	0	D
4-1c	SO2 Absorber Internal Sprays	4	Cost provided by R.C. (3 operating, 1 spares)	0.0	4183584	0	D
4-2	Reaction Mix Tank	4	Agitated carbon steel tank, Open Top, 259,000 gallon, 30' normal level, 10 wt% slurry (3 operating, 1 spares)	0.0	2999548	0	G
4-3	Agitator, Reaction Mix Tank	4	Stainless Steel (3 operating, 1 spares)	50.0	93280	30008	G
4-4	Spray Pump	16	Centrifugal, Stainless Steel (12 Operating, 4 Spare)	420.0	1710240	1008276	G
Subtotal: Area 4					11689836	1038284	
Area 5 - Stack Gas Reheat							
5-1	Reheater	4	Inline steam type, 2,776 ft2, half of tubes Inconel, other half made of Corten	0.0	2524147	0	EPA
Area 6 - Waste Handling							
6-1	Clarifier plus Flocculant system	1	86,000 lb hr dry solids, 18,500 ft2	0.0	1135706	0	G
6-2	Tank, Thickener Overflow	1	Unagitated, carbon steel, 44,000 gal, open top	0.0	77941	0	G
6-3	Pump, Thickener Overflow	2	Centrifugal, 1900 gpm, 100 psi, 1.0 sp.gr., stainless steel. (1 operating, 1 spare)	45.0	33404	9002	G
6-4	Pump, Thickener Underflow	2	Reciprocating, 600 gpm, 50 ft head, stainless steel	15.0	246916	6002	G
6-5	Tank, Filter Feed	1	Agitated, 110,000 gal, open top, carbon steel w/glass lining	0.0	476195	0	G

LIMESTONE SCRUBBER PROCESS

**EQUIPMENT LIST, DESCRIPTION, AND COST
(500 MW FACILITY)**

ID	ITEM	No.	Description	Total Power (hp)	Installed Cost (89\$)	Annual Utilities Required (89\$)	Ref No
6-6	Agitator, Filter Feed Tank	1	stainless steel	20.0	14575	4001	
6-7	Pump, Filter Feed Tank	2	Centrifugal, 900 gpm, 50 psi, stainless steel (1 operating, 1 spare)	35	36744	7002	
6-8	Vacuum Filter	2	Rotary drum, 860 ft ² filter area, 86,000 dry lb/hr (1 operating, 1 spare)	150	405292	30008	
6-9	Fixative Day Bin/Tank	1	Holds granular quicklime, 29,400 gal, carbon steel	0.0	63466	0	
6-9	Fixative Day	1	Holds fly ash to be used as sludge fixant, 15,800 gal, carbon steel	0.0	17793	0	
6-10	Process Mixer	2	Pug mill extruder, 26 tons/hr (1 operating, 1 spare)	30.0	101710	6002	
6-11	Conveyor, Waste solids	1	30 in. Belt conveyor, 100 ft long, 140 Tons/hr	10.0	93517	2001	
Subtotal: Area 6					2703259	64018	
LIMESTONE SCRUBBER TOTALS:					23533896	2613215	

APPENDIX C

LIMESTONE EMISSION CONTROL PROCESS EQUIPMENT LIST

LIMESTONE EMISSION CONTROL PROCESS
EQUIPMENT LIST, DESCRIPTION, AND COST
(500 MW FACILITY)

ID	ITEM	No.	Description	Total Power (hp)	Installed Cost (\$)	Annual Utilities Required (\$)	Ref. No.
Area 1 - Material Handling							
1-1	Car Shaker	1	Top mounted with crane	0.0	96337	0	
1-2	Car Puller	1	25 hp puller, 5 hp return	30.0	93841	6002	
1-3	Hopper, L/S Unloading	1	16 ft dia X 10 ft high. 6 in. square grating	0.0	16924	0	
1-4	Feeder, L/S Unloading	1	Vibrating pan, 42 in. wide x 20 ft long, 250 tons/hr	3.0	36788	600	
1-5	Conveyor, L/S Unloading	1	Belt, 36 in. wide X 20 ft long, 250 tons/hr, 130 ft/min	5.0	30429	1000	
1-6	Conveyor, L/S Stocking (Incline)	1	Belt, 36 in. wide X 310 ft long, 250 tons/hr, 130 ft/min	50.0	221695	10003	
1-7	Dust Collecting System	1	Bag Filter, polypropylene bag, 2200 acfm, auto shaker system, with dust hood	7.5	31844	1500	
1-8	Unloading Pit Sump Pump	1	Centrifugal, 60 gpm, 70 ft head stainless steel	2.0	5679	400	
1-9	Conveyor, L/S Stocking	1	Belt, 36 in. wide X 200 ft long, 250 tons/hr, 130 ft/min	5.0	165185	1000	
1-10	Tripper	1	30 ft/min	1.0	41190	200	
1-11	Mobile Equip.	1	Scraper Tractor	0.0	161016	0	
1-12	Hopper, reclaim	2	7 ft X 4.25 ft X 2 ft deep, 419 gal, carbon steel	0.0	1922	0	
1-13	Feeder, L/S live Storage	2	Vibrating Pan	3.0	36788	1200	
1-14	Tunnel Sump Pump	2	Vertical, 60 gpm, 70 ft head, stainless steel (1 operating, 1 spare)	5.0	11358	1000	
1-15	Conveyor, Live L/S Feed	1	Belt, 16 in. wide X 200 ft long 100 tons/hr, 60 ft/min	2.0	101546	400	
1-16	Conveyor, Live L/S Feed (Incline)	1	Belt, 16 in. wide X 200 ft long, 35 ft lift, 100 tons/hr, 60 ft/min	3.0	96201	600	
1-17	Elevator, live L/S Feed	1	Continuous Bucket, 10 in. X 7 in. X 11.75 in., 90 ft lift, 100 tons/hr, 160 ft/min	10.0	39709	2001	

LIMESTONE EMISSION CONTROL PROCESS

EQUIPMENT LIST, DESCRIPTION, AND COST (500 MW FACILITY)

ID	ITEM	No.	Description	Total Power (hp)	Installed Cost (\$)	Annual Utilities Required (\$)	Ref. No.
1-18	Bin, L/S Day	3	6,017 gal, 30 hr capacity, w/ cover, carbon steel	0.0	9087	0	E
1-19	Conveyor, Feed Belt	1	Belt, 16 in. wide X 60 ft long, 100 tons/hr, 60 ft/min, including tripper	3.0	36998	600	E
1-20	Dust Collecting System	1	Bag Filter, polypropylene bag, 2200 acfm, auto shaker system	7.5	4409	1500	E
1-21	Feeder, Make-up L/S	4	Weigh belt, 14 in. wide x 10 ft long, 11 tons/hr (3 operating, 1 spare)	1.0	76956	200	E
Subtotal: Area 1					1315902	28208	
Area 2 - Feed Preparation				0	0	0	
Area 3 - Gas Handling							
3-1	Fans	4	Induced draft, 628,000 acfm @ 283 F, 19 in. H2O static head, axial fan type (3 operating, 1 spare)	2676	227868	1606039	F
3-2	Inlet Ducting from Plant to ESP	1	Flue gas service, rectangular with external stiffeners, 14,570 ft2 of carbon steel	0.0	195275	0	F
3-3	Outlet Ducting from ESP to fan	1	Flue gas service, rectangular with external stiffeners, 13,990 ft2 of carbon steel	0.0	187501	0	F
3-4	Inlet Ducting from I.D. to Absorber Manifold	1	Flue gas service, rectangular with external stiffeners, 8,015 ft2 of carbon steel	0.0	107421	0	F
3-5	Inlet Ducting, Absorber Manifold	1	Flue gas service, rectangular with external stiffeners, 13,210 ft2 of carbon steel	0.0	177047	0	F
3-6	Outlet Ducting from Absorber (including manifold)	1	Flue gas service, rectangular with external stiffeners, 23,860 ft2 of stainless steel	0.0	639569	0	F
3-7	FGD Bypass	1	Flue gas service, rectangular with external stiffeners, 17960 ft2 of carbon steel	0.0	240709	0	F
3-8	Chimney Transition	1	Flue gas service, rectangular with external stiffeners, 1890 ft2 of carbon steel	0.0	25331	0	F

LIMESTONE EMISSION CONTROL PROCESS
EQUIPMENT LIST, DESCRIPTION, AND COST
(500 MW FACILITY)

ID	ITEM	No.	Description	Total Power (hp)	Installed Cost (\$)	Annual Utilities Required (\$)	Ref. No.
3-9	I.D. Fan Inlet Isolation Damper	2	Spade Guillotine w/ purge air Duct size: 10' x 18'	0.0	186093	0	
3-10	Scrubber Inlet Isolation Damper	4	Double Guillotine w/purge air Duct Size: 14' x 18'	0.0	594596	0	
3-11	Scrubber Outlet Isolation Damper	4	Double-Louvre, opposed and parallel blade w/purge air Duct Size: 10' x 17'	0.0	669027	0	
3-12	FGD System Bypass Damper	1	Double-Louvre, opposed and parallel blade w/purge air Duct Size: 10' x 17'	0.0	91004	0	
3-13	Plenum, Inlet	4	13,820 ft ² , fits 100 ft x 52 ft LEC reactor's face area, carbon steel	0.0	185233	0	
3-14	Plenum, Outlet	4	13,820 ft ² , fits 100 ft x 52 ft LEC reactor's face area, stainless steel	0.0	370446	0	
Subtotal: Area 3					3897120	1606039	
Area 4 - SO ₂ Absorption							
4-1a	SO ₂ Absorber Box	4	LEC Moving Bed Reactor, 100 ft tall x 52 ft wide x 2 ft deep, 1/4 in. 304 stainless steel (3 operating, 1 spare) Design Basis: Superficial Gas Velocity of 2 ft/s Superficial Residence Time of 1.6 s	0	108088	0	
4-1b	SO ₂ Absorber Live L/S Hopper	4	Fits on top of the LEC reactor box for air seal, 6 ft tall x 52 ft wide x 2 ft deep, 1/4 inch stainless steel	0	161264	0	
4-1c	SO ₂ Absorber Louvers	4	Inlet and outlet louvers, 100 ft tall x 52 ft wide, 1/8 inch stainless steel	0	2601908	0	
4-1d	SO ₂ Absorber Overhead Spray Piping	4	Two 2 inch piping runs of 52 ft, stainless steel, includes fittings Design Basis: 100 gpm per pipe run water velocity of 5 ft/s	0	34200	0	
4-1e	SO ₂ Absorber Spray Nozzles	4	+ 50 degree full cone spray nozzles, stainless steel, 1.2 gpm, 166 per reactor	0	128392	0	

LIMESTONE EMISSION CONTROL PROCESS

EQUIPMENT LIST, DESCRIPTION, AND COST (500 MW FACILITY)

ID	ITEM	No.	Description	Total Power (hp)	Installed Cost (\$)	Annual Utilities Required (\$)	Ref. No.
4-2	Elevator, Bulk Limestone Reactor Feed	4	Continuous bucket, 16 in. x 8 in. x 11 3/4 in. - 12, 102 tons/hr, 110 ft high (3 operating, 1 spare)	30.0	238292	18005	H
4-3	Screw Conveyor Bulk Limestone (Recycle)	4	16 in. screw x 60 ft long, 91 Tons/hr (3 operating, 1 spare)	24.0	136804	14404	H
4-4	Screw Conveyor, Live Bottom	4	Twin screw w/14 in screws, 30 ft long, 102 Tons/hr (3 operating, 1 spare)	48.0	273608	28808	H
4-5	Overhead Bed Spray Pumps	4	Centrifugal, 200 gpm, 110 ft of head, stainless steel (3 operating, 1 spare)	7.0	33404	4201	H
Subtotal: Area 4					3715960	61217	
Area 5 - Stack Gas Reheat							
5-1	Reheater	4	Inline steam type, 2,776 ft ² , half of tubes Inconel, other half made of Corten	0.0	2524147	0	EPA
Area 6 - Regeneration							
6-1	Elevator, Reacted L/S	4	Continuous bucket, 16 in. x 8 in. x 11 3/4 in. - 12, 109 tons/hr, 50 ft high (3 operating, 1 spare)	10.0	156876	6002	H
6-2	Screw Conveyor, Horizontal	4	Twin screw 14 in screws, 30 ft long, 54 Tons/hr (3 operating, 1 spare)	48.0	273608	28808	H
6-3	Mill, Ball	4	Dry/wet Marcy ball mill, 10 ft X 10 ft, 109 tons/hr (3 operating, 1 spare)	750	2813832	450123	H
6-4	Screen, (L/S)/(L/S)	4	Vibrating, 6 ft ² , stainless steel, 109 tons/hr L/S (excluding balls), No. 30 and 40 mesh double deck screen (3 operating, 1 spare)	30.0	169600	18005	H
6-6	Conveyor, Solid Waste Disposal	4	Belt, 14 in. wide x 100 ft long, 100 ft inclined, 18 tons/hr, 100 ft/min (3 operating, 1 spare)	3.0	236784	600	H
Subtotal: Area 6					3650700	503538	
LIMESTONE EMISSION CONTROL TOTALS:					15103829	2199002	

APPENDIX D
REACTOR SYSTEMS SIZING AND COSTING

Costs of Reactor Systems (500 tW)

LEC (1 Reactor)

4-1a Reactor Shell/Box:

$$\text{Cost} = 6752 \text{ lb } (\$1.38/\text{lb}) \cdot 2.0 \cdot 1.45 = \$27,022$$

4-1b Live Hopper:

$$\text{Cost} = 10,074 \text{ lb } (\$1.38/\text{lb}) \cdot 2.0 \cdot 1.45 = \$40,316$$

4-1c Louvers:

$$\text{Cost} = 162538 \text{ lb } (\$1.38/\text{lb}) \cdot 2.0 \cdot 1.45 = \$650,477$$

3-13+3-14 Plenums:

$$\text{Cost} = \$177,075 \text{ (see duct work section for design)}$$

4-1e Nozzles - Piping:

$$\text{Cost} = \$32,098 + 8550 = \$40,648$$

$$\text{Total LEC Reactor Cost} = \$935,538$$

$$\frac{\text{LSS Reactor Cost}}{\text{LEC Reactor Cost}} = \frac{1721692}{935,538} = 1.84$$

$$\% \text{ Difference} = \frac{1721692 - 935538}{1721692} \times 100 = 45.7\%$$

4-1 LEC Reactor Design

The LEC reactor design is the product of the experimental work done on a fixed-bed configuration in 1987 at Ohio University and a moving bed design tested in 1990 at Roanoke, VA. From these past pilot units, it was found that at a superficial gas velocity of 2 ft/s required a 15 inch bed depth to achieve greater than 90% SO₂ removal from the flue gas.

500 MW

If the LEC reactor is 100 ft tall,

$$\text{width } (t) = 628,000 \frac{\text{ft}^3}{\text{min}} \left(\frac{\text{s}}{24} \right) \left(\frac{\text{min}}{60 \text{ s}} \right) \left(\frac{1}{100 \text{ ft}} \right) = 52.4$$

The bed depth plus the lower* space (4") yields a reactor box depth = $4" + 15" + 4" = 24"$

• LEC reactor: 100' tall x 52' wide x 2' deep (not included in volume)

The louvers^{*} are made of $\frac{1}{8}$ " stainless steel and have the following dimensions:

* All lower designs, are taken directly from

The louvers cover 3" of vertical space each due to overlap; therefore the weight for louvers is:

$$\text{weight of louvers} = 2 \text{ sides} \cdot 100' \text{ side} \left(\frac{12 \text{ in}}{144} \right) \left(\frac{1 \text{ louver}}{3 \text{ in}} \right) \left(\frac{7 \text{ in}}{1 \text{ louver}} \right) \left(\frac{1}{8 \text{ in}} \right) \left(\frac{1 \text{ lb}}{144 \text{ in}^2} \right) (52.4) \\ \left(10.3 \frac{\text{Carbon steel } \frac{1}{4} \text{ in}}{\text{Water } \frac{1}{4} \text{ in}^2} \right) (62.4278 \frac{\text{lb}}{\text{ft}^3}) = 162,538 \text{ lb}$$

The reactor shell/box for cost estimating purposes will be composed of a bottom plate measuring 2' x 52' and two sides measuring 2' x 100'. The remaining two shell/box sides are covered by the plenums.

$$\text{weight of box} = \left(\frac{1}{4} \text{ in} \right) \left(\frac{144}{12 \text{ in}} \right) (10.3) (62.4278 \frac{\text{lb}}{\text{ft}^3}) [2 \times 52' + 2 (2' \times 100')] \\ = 6752 \text{ lb}$$

The live hopper that fits on top of the reactor is also made from 1/4" steel and measures 52' wide x 2' deep x 6' tall.

$$\text{weight of live hopper} = \left(\frac{1}{4} \text{ in} \right) \left(\frac{144}{12 \text{ in}} \right) (10.3) (62.4278) [2 (52' \times 6') + 2 (2' \times 6') + (52' \times 2')] \\ = 10,074 \text{ lb}$$

Using a material factor of S.S. = 2.0

Installation factor = 1.45

cost of carbon steel = \$1.32/lb of carbon steel

LEC Pumps + Piping

500 MW

Using the same hp and utilities equations used on the LSS pumps, the LEC requires ~200 gpm/reactor;

4-5 Overhead Bed Spray Pump, 200 gpm, centrifugal, 0% solids, stainless steel, head ~110 ft

$$hp = \frac{(200 \text{ gpm})(1.0 \text{ sp. gr.})(110 \text{ ft})}{3960 (0.80)} = 7 \text{ hp}$$

Using the pump cost chart in Garrett.

$$\text{cost} = \$2,500 \cdot 1.5 \cdot 2.0 \left(\frac{356.3}{320} \right) = \$8351$$

Internal Piping

Since the LEC bed is 15" wide and 52' across:

assuming a 5 ft/s water velocity through the piping: $100 \text{ gal/min} \left(\frac{7.48 \text{ gal}}{\text{ft}^3} \right) \left(\frac{1 \text{ s}}{5 \text{ ft}} \right) = 5.35 \text{ ft}^2 = \pi r^2; r = 0.92$

rounding piping size to nearest inch = 2" diameter pipe.

Using Garrett's pipe cost chart for 2" pipe yields a purchase cost of \$1.32/ft (L.E. = 320)

Since the piping will be in an acidic environment, use stainless steel cost factor of 2.05 for bare pipe. To include the cost of fittings (tee's, elbows and other peripheral plumbing), use the fittings cost factor of 1.8.

Thus the cost of internal piping not including spray nozzles is:

$$\text{cost} = \$138/\text{ft} (104 \text{ ft}) \cdot 2.05 \cdot 1.8 \left(\frac{356.3}{320} \right) \cdot 1.45 = \$8550$$

Nozzles

Using the Spraying Systems Co. Industrial Catalog 27 for nozzle description + Catalog 19 as a price list, it was determined from the Spray Coverage Information Table that a 40° spray angle was required (12' coverage, - 7.5" ft for one coverage). Assuming each nozzle will cover 7.5", therefore

$$\text{number of nozzles required} = 2 \left[(52') \left(\frac{\text{nozzle}}{7.5''} \right) \left(\frac{12'}{7.5'} \right) - 1 \right] \approx 166$$

$$\text{gpm per nozzle} = 200 \text{ gpm} / 166 = 1.20$$

From p. 17 in Catalog 27, chose 2RR5045

From p. 7 in Catalog 19, for a stainless steel nozzle,
\$111.16 / nozzle

$$\text{nozzle cost} = [\$111.16 / \text{nozzle} \cdot 166 \text{ nozzles}] \left(\frac{356.3}{297.0} \right) = \$22,137$$

$$+ 11.0 \cdot 1.45 \cdot 1.8 \cdot \frac{356.3}{297.0} = \$27,098$$

Limestone Scrubber Tower Design.

From EPRI CS-3342, the design basis for the absorption tower is a superficial gas velocity of 10 ft/s and a gas residence time of 10 s .

To achieve these design parameters, the spray tower must 100 ft tall with a diameter of 38 ft for an inlet gas flow rate of $628,000 \text{ acfm}$.

Using a cylinder to get the surface area of the spray tower and using 4 steel yields:

$$\text{weight of LSS shell: } (4)(\frac{1}{2})(10.3)(52.4278) [2\pi(19)^2 + 2\pi(19)100] \\ = 140,307 \text{ lb}$$

From Garrett's Tank Cost chart, the material factor for a rubber-lined carbon steel tank is 1.5

Cost of LSS spray tower shell:

$$140,307 \text{ lb } (\$1.38/\text{lb of carbon steel}) \cdot 1.5 \cdot 1.45 = \$5571,206$$

Cost of mist eliminator: ($\$50,000$ for $28'$ diameter tower, R.C)

$$\$50,000 \left[\frac{\pi(19)^2}{\pi(14)^2} \right]^{0.6} 1.45 = \$104,540$$

Cost of the internal sprays: ($\$500,000$ for $28'$ diameter tower, R.C)

$$\$500,000 \left[\frac{\pi(19)^2}{\pi(14)^2} \right]^{0.6} 1.45 = \$1,045,896$$

Total cost of LSS Tower = $\$1,721,642$

FAX (201) 685-4834

TELECOPY COVER SHEET

DATE: 11/14/90

TELECOPY NUMBER: 614 593 4684

PLEASE DELIVER THE FOLLOWING PAGES TO:

NAME: KEN APPELL

LOCATION: OHIO UNIV. CH. C. DEPT.

FROM: DENNIS HELFRICH

LOCATION: R - C

TOTAL NUMBER OF PAGES INCLUDING COVER SHEET: 1

COMMENTS:

For a 28' diameter tower, the
wind eliminator would cost \$50,000
(not installed) and the spray delivery
system would cost \$500,000 (not
installed)

[notes]

[initials]

Glen

IF YOU DO NOT RECEIVE ALL OF THE PAGES, PLEASE NOTIFY US AS SOON AS POSSIBLE. THANK YOU.

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TELEX 833413

NOZZLE Engineering Data

SPRAY COVERAGE INFORMATION

This table lists the theoretical coverage of spray patterns as calculated from the included spray angle of the spray and the distance from the nozzle orifice. These values are based on the assumption that the spray angle remains the same throughout entire spray distance. In actual practice, the tabulated spray angle does not hold for long spray distances. Write for Data Sheets on actual spray coverage.

Included Spray Angle	THEORETICAL COVERAGE AT VARIOUS DISTANCES (IN INCHES) FROM NOZZLE ORIFICE											
	2"	4"	5"	8"	10"	12"	15"	18"	24"	30"	36"	48"
5°	0.2"	0.4"	0.5"	0.7"	0.9"	1.1"	1.3"	1.6"	2.1"	2.6"	3.1"	4.2"
10°	0.4"	0.7"	1.1"	1.4"	1.8"	2.1"	2.6"	3.1"	4.2"	5.2"	6.3"	8.4"
15°	0.5"	1.1"	1.5"	2.1"	2.6"	3.2"	3.9"	4.7"	6.3"	7.9"	9.5"	12.6"
20°	0.7"	1.4"	2.1"	2.8"	3.5"	4.2"	5.3"	6.4"	8.5"	10.6"	12.7"	16.9"
25°	0.9"	1.8"	2.7"	3.5"	4.4"	5.3"	6.6"	8.0"	10.6"	13.3"	15.9"	21.2"
30°	1.1"	2.1"	3.2"	4.3"	5.4"	6.4"	8.1"	9.7"	12.8"	16.1"	19.3"	25.7"
35°	1.3"	2.5"	3.8"	5.0"	6.3"	7.6"	9.5"	11.3"	15.5"	18.9"	22.7"	30.3"
40°	1.5"	2.9"	4.4"	5.8"	7.3"	8.7"	10.9"	13.1"	17.5"	21.8"	26.2"	34.9"
45°	1.7"	3.3"	5.0"	6.6"	8.3"	9.9"	12.4"	14.9"	19.9"	24.8"	29.8"	39.7"
50°	1.9"	3.7"	5.6"	7.5"	9.3"	11.2"	14.0"	16.8"	22.4"	28.0"	33.6"	44.8"
55°	2.1"	4.2"	6.3"	8.3"	10.3"	12.5"	15.6"	18.7"	25.0"	31.2"	37.5"	50.0"
60°	2.3"	4.6"	6.9"	9.2"	11.5"	13.8"	17.3"	20.6"	27.7"	34.6"	41.6"	55.4"
65°	2.5"	5.1"	7.5"	10.2"	12.7"	15.3"	19.2"	22.9"	30.5"	38.2"	45.8"	61.2"
70°	2.8"	5.6"	8.4"	11.2"	14.0"	16.8"	21.0"	25.2"	33.6"	42.0"	50.4"	67.2"
75°	3.1"	6.1"	9.2"	12.3"	15.3"	18.4"	23.0"	27.6"	36.8"	46.0"	55.2"	73.6"
80°	3.4"	6.7"	10.1"	13.4"	16.8"	20.2"	25.2"	30.3"	40.3"	50.4"	60.4"	80.6"
85°	3.7"	7.3"	11.0"	14.7"	18.3"	22.0"	27.5"	33.0"	44.0"	55.0"	66.0"	88.0"
90°	4.0"	8.0"	12.0"	16.0"	20.0"	24.0"	30.0"	36.0"	48.0"	60.0"	72.0"	96.0"
95°	4.4"	8.7"	13.1"	17.5"	21.8"	26.2"	32.8"	39.3"	52.4"	65.5"	78.6"	105"
100°	4.8"	9.5"	14.3"	19.1"	23.8"	28.6"	35.8"	43.0"	57.2"	71.6"	85.9"	114"
110°	5.7"	11.4"	17.1"	22.8"	28.5"	34.3"	42.8"	51.4"	68.5"	85.6"	103"	
120°	6.9"	13.9"	20.8"	27.7"	34.6"	41.6"	52.0"	62.4"	83.2"	104"		
130°	8.5"	17.2"	25.7"	34.3"	42.9"	51.5"	64.4"	77.3"	103"			
140°	10.9"	21.9"	32.3"	43.8"	54.8"	65.7"	82.2"	98.6"				
150°	14.9"	29.8"	44.7"	59.6"	74.5"	89.5"	112"					
160°	22.7"	45.4"	68.0"	90.6"	113"							
170°	45.8"	91.6"										

FLOW OF WATER THROUGH SCHEDULE 40 STEEL PIPE

Recommended capacity range for each size is shown in shaded areas.

Flow in G.P.M.	Pressure Drop in p.s.i. for Various Pipe Sizes (in 10 Ft. Length)							
	1/4"	1/2"	3/4"	1"	1 1/4"	1 1/2"	2"	3"
0.3								
0.4								
0.5	1.1							
0.6	1.5							
0.8	2.5							
1.0	3.7	83						
1.5	9.0	1.8						
2.0	13.4	3.0	66					
2.5		4.5	1.0					
3.0		6.4	1.4					
4.0		11.1	2.4	74				
5.0			3.7	1.1				
6.0			5.2	1.5				
8.0			9.1	2.3	66			
10				4.2	1.0			
15					2.2	54		
20					3.3	1.1		
25						1.7	42	
30						2.4	59	

Flow in G.P.M.	Pressure Drop in p.s.i. for Various Pipe Sizes (in 10 Ft. Length)							
	2"	2 1/2"	3"	3 1/2"	4"	5"	6"	8"
35								
40								
45								
50								
60	.29							
70	.38							
80	.50							
90	.62							
100	.76	.31						
125	1.2	.47						
150	1.7	.67						
200	2.9	1.2	.39					
250			.59	.28				
300			.84	.40				
400				.70	.37			
500					.57	.18		
750						.39	.16	.04
1000						.68	.27	.07
2000							1.0	.25

APPROXIMATE FRICTION LOSS IN PIPE FITTINGS

in terms of equivalent feet of straight pipe.

Pipe Size Std. Wt.	Actual inside diam. in.	Gate Valve FULL OPEN	Globe Valve FULL OPEN	45° Elbow	Run of Std. tee	Std. elbow or run of tee reduced 1/2	Std. tee thru side outlet
1/4	.259	15	8	.35	.40	.75	1.4
1/2	.364	20	11	.50	.65	1.1	2.2
3/4	.522	35	18.6	.78	1.1	1.7	3.3
1	.824	44	23.1	.97	1.4	2.1	4.2
1 1/4	1.049	56	29.4	1.2	1.8	2.6	5.3
1 1/2	1.380	.74	38.6	1.6	2.3	3.5	7.0
2	1.610	86	45.2	1.9	2.7	4.1	8.1
2 1/2	2.067	1.1	58	2.4	3.5	5.2	10.4
3	2.469	1.3	69	2.9	4.2	6.2	12.4
4	3.068	1.6	86	3.6	5.2	7.7	15.5
5	4.026	2.1	113	4.7	6.8	10.2	20.3
6	5.047	2.7	142	5.9	8.5	12.7	25.4
8	6.065	3.2	170	7.1	10.2	15.3	31

MISCELLANEOUS EQUIVALENTS AND FORMULAS

UNIT	EQUIVALENT	UNIT	FORMULA
Ounce	28.35 Gr.	Fahrenheit (F°)	= 9/5 C° + 32
Pound	0.4536 Kg.	Celsius (C°)	= 5/9 (F° - 32)
Horse-Power	0.746 Kw.	Circumference of Circle	= 3.1416 x Diameter
British Thermal Unit	0.2520 Kg.-Cal.	Area of a Circle	= .7854 x Square of the Diameter
Square Inch	6.452 Sq. Cm.	Volume of a Sphere	= .5236 x Cube of the Diameter
Square Foot	0.09290 Sq. M.	Area of a Sphere	= 3.1416 x Square of the Diameter
Acre	0.4047 Hectare		
Acre	43,560 Sq. Ft.		

NOZZLE *Engineering Data*

The capacities, spray angles, dimensions, and weights shown in this catalog are nominal values.

If any of these values are critical to your application, please contact us for specific data.

SPRAY CHARACTERISTICS

DESCRIPTION

SPRAY CHARACTERISTICS																						
SPRAY PATTERNS																						
	Hollow Cone	Full Cone	Flat Spray	Solid Stream	Air Atomizing																	
CAPACITY (Flow Rate)	All capacity tabulations in this catalog are based on water. Since the specific gravity of a liquid affects its flow rate, tabulated catalog capacities must be multiplied by the conversion factor that applies to the specific gravity of the liquid sprayed . . . as follows:			Nozzle capacity varies with spraying pressure. In general, the relationship between GPM and pressure is as follows:																		
	<table><tr><td>Specific Gravity</td><td>.84</td><td>.96</td><td>1.00</td><td>1.08</td><td>1.20</td><td>1.32</td><td>1.44</td></tr><tr><td>Multiply by Factor</td><td>1.09</td><td>1.02</td><td>1.00</td><td>.96</td><td>.91</td><td>.87</td><td>.83</td></tr></table>			Specific Gravity	.84	.96	1.00	1.08	1.20	1.32	1.44	Multiply by Factor	1.09	1.02	1.00	.96	.91	.87	.83	$\frac{GPM_1}{GPM_2} = \frac{\sqrt{P.S.I._1}}{\sqrt{P.S.I._2}}$		
Specific Gravity	.84	.96	1.00	1.08	1.20	1.32	1.44															
Multiply by Factor	1.09	1.02	1.00	.96	.91	.87	.83															
				See capacity tabulations on following pages.																		
SPRAY ANGLE	Tabulated spray angles indicate approximate spray coverages based on water. In actual spraying, the effective spray angle varies with spray distance. If the spray coverage requirement is critical write for specific spray coverage data. See next page for "Spray Coverage Information." Liquids more			viscous than water form relatively smaller spray angles (or even a solid stream), depending upon viscosity, nozzle capacity and spraying pressure. Liquids with surface tensions lower than water will produce relatively wider spray angles than those listed for water.																		
SPRAY ATOMIZATION	Fine atomization is most easily obtained with air atomizing nozzles because compressed air is used to break up the liquid into fine particles. Above this level, hydraulic spray nozzles produce a larger particle size spray range depending upon such factors as nozzle type, capacity, spray angle and pressure. Wide angle hollow cone, low capacity hy-			draulic spray nozzles operating at higher pressures produce smaller spray particles. Spray particle sizes become coarser as nozzle capacities increase, spray angles become narrower as pressures decrease. The largest particles are obtained with the largest capacity full cone nozzles spraying at the lowest pressure. For specific spray nozzle particle size data . . . write for information.																		
	• 800 Microns • 1,200 Microns ● 5,500 Microns																					
IMPACT	The total theoretical impact of a spray depends primarily upon the GPM and spray pressure. The highest impact per square inch is produced by solid stream and flat spray patterns . . . using large capacity			nozzles operating at higher pressures. Lowest impact efficiency is provided by wide spray full cone and wide spray hollow cone spray nozzles.																		
VELOCITY	The theoretical velocity of a spray leaving the orifice of a nozzle, depends upon the spraying pressure. Solid stream and flat spray nozzles have the highest velocity effi-			ciencies . . . while wide angle full cone nozzle and wide angle hollow cone nozzles have the lowest velocity efficiencies.																		

Note: The catalog tabulations show orifice dimensions as "Nom." (nominal). Specific dimensions are available on request.

Spray performance data and nozzle dimensions are also available in Metric Units. Write for information.

Linear Unit	LINEAR UNITS—EQUIVALENTS						
	Micron	Mil	Milli-meter	Centi-meter	Inch	Foot	Meter
Micron	•	0.039	0.001	1.0x10 ⁻⁴	3.94x10 ⁻⁵		
Mil	25.4	•	2.54x10 ⁻²	2.54x10 ⁻³	0.001	8.33x10 ⁻⁵	
Millimeter	1000	39.4	•	0.10	0.0394	3.28x10 ⁻³	0.001
Centimeter	10000	394	10	•	0.394	0.033	0.01
Inch	2.54x10 ⁴	1000	25.4	2.54	•	0.083	0.0254
Foot	3.05x10 ⁴	1.2x10 ⁴	305	30.5	12	•	0.305
Meter	1.0x10 ⁴	3.94x10 ⁴	1000	100	39.4	3.28	•

Volumetric Unit	VOLUMETRIC UNITS—EQUIVALENTS						
	Cubic Centi-meter	Fluid Ounce	Pound of Water	Liter	US Gallon	Cubic Foot	Cubic Meter
Cubic Centimeter	•	0.034	2.2x10 ⁻³	0.001	2.64x10 ⁻⁴	3.53x10 ⁻⁵	1.0x10 ⁻⁶
Fluid Ounce	29.6	•	0.065	0.030	7.81x10 ⁻³	1.04x10 ⁻³	2.96x10 ⁻⁵
Pound of Water	454	15.4	•	0.454	0.12	0.016	4.54x10 ⁻⁴
Liter	1000	33.8	2.2	•	0.264	0.035	0.001
US Gallon	3785	128	8.34	3.785	•	0.134	3.78x10 ⁻³
Cubic Foot	28320	958	62.4	28.3	7.48	•	0.028
Cubic Meter	1.0x10 ⁶	3.38x10 ⁴	2202	1000	264	35.3	•

Liquid Pressure	LIQUID PRESSURES—EQUIVALENTS						
	Lb./In ² (p.s.i.)	Ft Water	Kg./Cm ²	Atmosphere	Bar	Inch Mercury	kPa (kilopascal)
Lb./In ² (p.s.i.)	•	2.31	0.070	0.068	0.069	2.04	6.895
Ft Water	0.433	•	0.030	0.029	0.030	0.882	2.99
Kg./Cm ²	14.2	32.8	•	0.968	0.981	25.0	98
Atmosphere	14.7	33.9	1.03	•	1.01	29.9	101
Bar	14.5	33.5	1.02	0.987	•	29.5	100
Inch Mercury	0.491	1.13	0.035	0.033	0.034	•	3.4
kPa (kilopascal)	0.145	0.335	0.01	0.009	0.01	0.296	•

FullJet. NOZZLES Distribofet. SERIES—65° and 50°

Spray Characteristics—Large capacity, narrow angle full flow, deluge-type, full cone spray pattern.

Construction—One piece cast type with non-removable internal vane. Grooved orifice for accurate control of capacity and spray angle.

Materials—Available from stock in cast brass, cast iron or cast 316 stainless steel. See page 3 for list of other materials.

DISTRIBUTION

HOW TO ORDER: Specify Nozzle No., Material. Example: 2R6560 Cast Iron.

Nozzle No.			Pipe Size Inches	Inner Orifice Diam. Nom. Inches	CAPACITY GPM (gallons per minute) at p.s.i. (pounds per square inch)											SPRAY ANGLE** at 3 p.s.i.
Female NPT Pipe Conn.	Male NPT Pipe Conn.	Flange Conn.*			1 p.s.i.	3 p.s.i.	5 p.s.i.	7 p.s.i.	10 p.s.i.	15 p.s.i.	20 p.s.i.	30 p.s.i.	40 p.s.i.	60 p.s.i.		
2R5045	2RR5045		2	1	27	45	57	57	78	94	108	130	148	179	50°	
2R6545	2RR6545			1 1/16	27	45	57	67	78	94	108	130	148	179	65°	
2R6560	2RR6560			1 1/4	36	60	76	89	104	126	144	174	199	239	65°	
2 1/2 R5070	2 1/2 RR5070		2 1/2	1 13/16	42	70	88	103	122	147	168	202	232	278	50°	
2 1/2 R6570	2 1/2 RR6570			1 11/16	42	70	88	103	122	147	168	202	232	278	65°	
2 1/2 R6590	2 1/2 RR6590			1 1/2	54	90	113	133	157	189	216	260	297	358	65°	
3R50110	3RR50110		3	1 3/4	66	110	139	163	192	231	263	318	363	439	50°	
3R65110	3RR65110			1 13/16	66	110	139	163	192	231	263	318	363	439	65°	
3R65140	3RR65140			1 1/2	84	140	177	207	243	293	335	404	463	557	65°	
4R50160	4RR50160	4RF50160	4	1 1/2	97	160	203	237	279	337	384	466	532	642	50°	
4R65160	4RR65160	4RF65160		2 1/4	97	160	203	237	279	337	384	466	532	642	65°	
4R50190	4RR50190	4RF50190		2 1/4	115	190	240	280	331	399	456	549	628	756	50°	
4R65190	4RR65190	4RF65190		2 13/16	115	190	240	280	331	399	456	549	628	756	65°	
4R65250	4RR65250	4RF65250		2 3/4	151	250	317	369	436	524	600	723	827	997	65°	
5R50250	5RR50250	5RF50250	5	2 3/4	151	250	317	369	436	524	600	723	827	997	50°	
5R65250	5RR65250	5RF65250		2 3/4	151	250	317	369	436	524	600	723	827	997	65°	
5R50280	5RR50280	5RF50280		2 15/16	168	280	354	413	487	587	671	809	927	1120	50°	
5R65280	5RR65280	5RF65280		2 15/16	168	280	354	413	487	587	671	809	927	1120	65°	
5R65380	5RR65380	5RF65380		3 1/2	229	380	480	560	660	795	909	1100	1260	1520	65°	
6R50360	6RR50360	6RF50360	6	2 3/4	217	360	456	532	624	755	863	1040	1190	1430	50°	
6R65360	6RR65360	6RF65360		3 1/4	217	360	456	532	624	755	863	1040	1190	1430	65°	
6R50400	6RR50400	6RF50400		2 31/32	241	400	505	588	696	837	957	1160	1320	1590	50°	
6R65400	6RR65400	6RF65400		3 1/4	241	400	505	588	696	837	957	1160	1320	1590	65°	
6R65560	6RR65560	6RF65560		3 3/4	337	550	710	827	975	1170	1340	1620	1850	2230	65°	
8R50650	8RR50650	8RF50650	8	3 11/16	393	550	823	962	1130	1370	1570	1880	2150	2600	50°	
8R65650	8RR65650	8RF65650		4 1/16	393	550	823	962	1130	1370	1570	1880	2150	2600	65°	
8R50750	8RR50750	8RF50750		4 1/4	453	750	949	1120	1310	1570	1800	2170	2480	2990	50°	
8R65750	8RR65750	8RF65750		4 13/16	453	750	949	1120	1310	1570	1800	2170	2480	2990	65°	
8R65850	8RR65850	8RF65850		4 3/4	512	850	1070	1260	1480	1780	2030	2450	2810	3390	65°	

*Bolt circle and hole diameters are in accordance with American Standard class 125 cast iron pipe flanges.

**See page 5 for spray coverage information and page 25 for DIMENSIONS AND WEIGHTS.

FULL CONE SPRAY PATTERN

Type R female connection

Type RR male connection

Type RF flange connection

Patent No. 3,275,248

ACCESSORIES

Liquid Strainers

See pages 82-84

Adjustable Ball Fittings
See page 73

Junction Boxes
See page 71

TYPICAL APPLICATIONS

- WATER COOLING
- DELUGING
- QUENCHING
- DEAERATING
- FIRE PROTECTION
- DRENCHING
- CHEMICAL PROCESSING

NET PRICE LIST
No. 19
for use with Catalog 27

WING SYSTEMS CO.

SPRAY NOZZLES AND ACCESSORIES

TERMS 1%, 10 days . . . Net 30 days.
Prices effective September 1, 1981
All prices . . . F.O.B. Wheelon, IL 60187.
MINIMUM CHARGE \$20.00 NET.

Prices are subject to change without notice and are also subject to the addition of any tax levied by any federal, state, county or municipal government.

For prices as listed in this price list apply only to nozzles of the same type, pipe connection, capacity and material shipped to a single destination.

D-60

Prices will be quoted upon request for nozzles made of materials other than those listed.

TYPES G-W, GG-W,
H-W and HH-W

FullJet NOZZLES

wide spray

FULL CONE SPRAY PATTERN

Material	NOZZLE NO.		PRICES EACH IN LOTS OF				
	Types G and H Female Conn.	Types GG and HH Male Conn.	1- 11	12- 35	36- 99	100- 499	500- up
Brass	1/4 G-W	1/4 GG-W	\$4.58	\$3.89	\$3.66	\$3.44	\$3.21
	1/4 G-W	1/4 GG-W	5.05	4.29	4.04	3.79	3.54
	1/4 G-W	1/4 HH-W	3.02	2.57	2.42	2.27	2.11
	1/4 G-W	1/4 GG-W	5.97	5.07	4.78	4.48	4.18
	1/4 G-W	1/4 HH-W	3.63	3.09	2.90	2.72	2.54
	1/4 G-W	1/4 GG-W	7.02	5.97	5.62	5.27	4.91
	1/4 G-W	1/4 HH-W	4.36	3.71	3.49	3.27	3.05
	1/4 G-W	1/4 HH-W	8.70	7.40	6.95	6.53	6.09
	1/4 H-W	1/4 HH-W	12.18	10.35	9.74	9.14	8.53
	1H-W	1HH-W	11.48	9.76	9.18	8.61	8.04
Cast Brass	1 1/4 H-W		20.40	18.36			
	1 1/4 H-W		25.45	23.81			
	2H-W		40.56	36.50			
	2 1/2 H-W		63.67	57.30			
	3H-W		85.27	76.74			
Mild Steel	1/4 G-I W	1/4 GG-I W	5.92	5.88	5.54	5.19	4.84
	1/4 G-I W	1/4 GG-I W	7.52	6.48	6.10	5.72	5.33
	1/4 G-I W	1/4 HH-I W	5.36	4.56	4.29	4.02	3.75
	1/4 G-I W	1/4 GG-I W	10.10	8.59	8.08	7.58	7.07
	1/4 G-I W	1/4 HH-I W	6.33	5.38	5.06	4.75	4.43
	1/4 G-I W	1/4 GG-I W	10.93	9.29	8.74	8.20	7.65
	1/4 G-I W	1/4 HH-I W	7.52	6.48	6.10	5.72	5.33
	1/4 G-I W	1/4 HH-I W	12.55	10.67	10.04	9.41	8.79
	1H-I W	1HH-I W	16.69	14.19	13.35	12.52	11.68
	1H-I W	1HH-I W	16.69	14.19	13.35	12.52	11.68
Cast Iron	1 1/4 H-I W		21.41	19.27			
	1 1/4 H-I W		23.39	21.05			
	2H-I W		37.43	33.69			
	2 1/2 H-I W		51.84	46.66			
	3H-I W		71.97	64.77			
Type 303 Stainless Steel	1/4 G-SS W	1/4 GG-SS W	12.92	10.98	10.34	9.69	9.04
	1/4 G-SS W	1/4 GG-SS W	13.58	11.54	10.86	10.19	9.51
	1/4 G-SS W	1/4 HH-SS W	9.79	8.32	7.83	7.34	6.85
	1/4 G-SS W	1/4 GG-SS W	17.94	15.35	14.35	13.46	12.56
	1/4 G-SS W	1/4 HH-SS W	11.61	9.87	9.28	8.70	8.13
	1/4 G-SS W	1/4 GG-SS W	21.49	18.27	17.19	16.10	15.04
	1/4 G-SS W	1/4 HH-SS W	13.95	11.87	11.17	10.47	9.77
	1/4 G-SS W	1/4 HH-SS W	25.46	21.54	20.37	19.19	17.92
	1H-SS W	1HH-SS W	45.82	38.95	36.66	34.37	32.07
	1H-SS W	1HH-SS W	33.55	28.53	26.85	25.17	23.49
Type 316 Stainless Steel	1/4 G-316SS W	1/4 GG-316SS W	18.09	15.33	14.47	13.67	12.86
	1/4 G-316SS W	1/4 GG-316SS W	19.02	16.17	15.22	14.37	13.51
	1/4 G-316SS W	1/4 HH-316SS W	13.57	11.52	10.94	10.25	9.57
	1/4 G-316SS W	1/4 GG-316SS W	23.14	21.31	20.11	18.86	17.60
	1/4 G-316SS W	1/4 HH-316SS W	15.10	12.84	12.08	11.33	10.57
	1/4 G-316SS W	1/4 GG-316SS W	30.04	25.53	24.03	22.53	21.03
	1/4 G-316SS W	1/4 HH-316SS W	19.54	16.51	15.53	14.66	13.58
	1/4 G-316SS W	1/4 HH-316SS W	35.59	30.25	28.47	26.69	24.91
	1H-316SS W	1HH-316SS W	63.05	53.59	50.44	47.29	44.14
	1H-316SS W	1HH-316SS W	48.00	40.80	38.40	36.00	33.60
Cast Type 316 Stainless Steel	1 1/4 H-SS W		87.44	73.70			
	1 1/4 H-SS W		110.84	99.75			
	2H-SS W		158.50	142.55			
	2 1/2 H-SS W		208.60	187.74			
	3H-SS W		277.06	249.35			
PVC		1/4 GG-PVC W	12.15	10.34	9.73	9.12	8.51

Prices
on
Application

APPENDIX E

MATERIAL HANDLING SECTION FOR THE FGD PROCESSES

Area 1 - Material Handling

The material handling section comprises of all the equipment necessary to move fresh limestone from a railroad spur adjacent to the electric utility upto the ball mill in the LSS or upto the bulk L/S screw conveyor in the LEC system. The equipment size and cost for both the LEC + LSS processes are identical in Area 1.

	100 MW		500 MW	
	LEC	LSS	LEC	LSS
Make-up L/S requirement ^(lb/hr)	13,100	13400	65,300	67,000
^(T/hr)	6.6	6.70	32.7	33.5

500 MW + 100 MW

1-1 Car Shaker, Top mounted w/ crane,
cost taken from EPA-600/7-81-014 (Ref 1)
and converted from 1982 dollars to 1989 dollars
cost = \$84900 $\left(\frac{356.3}{314.0}\right)$ = \$96,337

1-2 Car Puller and return, 30 hp;
cost taken from (Ref 1) + converted to 1989 dollars
cost = \$82,700 $\left(\frac{356.3}{314.0}\right)$ = \$93,841

1-3 Limestone unloading Hopper / Tank
 16 ft diameter x 10 ft height $\approx 15,000$ gal, carbon steel
 Using the Tank cost chart from Garrett, p. 303

$$\text{Cost} = \$8,000 \cdot 1.9 \left(\frac{356.3}{320} \right) = \$16,924$$

1-4 Feeder, L/S unloading, vibrating pan,
 42 in. wide x 20 ft long, 3 hp
 Using cost chart for conveyors from Garrett, p. 273

$$\text{Cost} = \$11,800 \cdot 2.4 \left(\frac{42 \text{ in.}}{36 \text{ in.}} \right)^{1.0} \left(\frac{356.3}{320} \right) = \$36,788$$

1-5 Limestone unloading belt conveyor, 250 tons/hr,
 36 in. belt width x 20 ft long, 5 hp, 130 $\frac{\text{ft}}{\text{min}}$
 Using cost chart for conveyors from Garrett, p. 273

$$\text{Cost} = \$7,000 \cdot 2.4 \left(\frac{36 \text{ in.}}{16 \text{ in.}} \right)^{0.6} \left(\frac{356.3}{320} \right) = \$30,429$$

$$\text{hp} = \underbrace{20 \text{ ft}}_{\text{run}} \left(\frac{1.59 \text{ hp}}{100 \text{ ft}} \right) + \underbrace{20 \text{ ft}}_{\text{lift}} \left(\frac{2.44 \text{ hp}}{10 \text{ ft} \cdot 4.1 \text{ ft}} \right) \approx 5 \text{ hp}$$

Power calculation taken from Table 7-7 in Perry's.

1-6 Limestone stacking ^{belt} conveyor, 36 in. wide,
 250 tons/hr, 310 ft long, 185 ft $\cdot$ 4 $\frac{\text{ft}}{\text{min}}$

$$\text{hp} = 310 \text{ ft run} \left(\frac{1.59 \text{ hp}}{100 \text{ ft}} \right) + 185 \left(\frac{2.44 \text{ hp}}{10 \text{ ft} \cdot 4.1 \text{ ft}} \right) \approx 50 \text{ hp}$$

$$\text{Cost} = \$51,000 \left(\frac{36 \text{ in.}}{16 \text{ in.}} \right)^{0.6} \cdot 2.4 \left(\frac{356.3}{320} \right) = \$221,695$$

1-7 Dust Collecting System; bag filter,
2200 acfm, 7.5 hp

Using the Dust Collector Cost chart; Garrett, p. 279

$$\text{Cost} = \$13,000 \cdot 2.2 \left(\frac{356.3}{320} \right) = \$31,844$$

1-8 Unloading Pit Sump Pump, 60 gpm, 70' head

$$\text{hp} = \frac{Q \cdot H \cdot \text{SpGr}}{3960 \cdot \text{eff}_p} = \frac{(60 \text{ gpm} \times 70 \text{ ft}) (1.0)}{3960 (0.80)} \approx 2.0 \text{ hp}$$

Using the Pump Cost Chart from Garrett, p. 269
centrifugal composed of stainless steel

$$\text{Cost} = \$1,700 \cdot 1.5 \cdot 2.0 \left(\frac{356.3}{320} \right) = \$5679$$

1-9 Belt conveyor limestone stacking, 36 in. wide
 $\times 200 \text{ ft}$ long, 250 T/hr, hp ≈ 5.0

$$\text{Cost} = \$38,000 \cdot 2.4 \left(\frac{36''}{16''} \right)^{0.6} \left(\frac{356.3}{320} \right) = \$165,185$$

1-10 Tripper for Belt conveyor (1-9), 2 hp

cost taken directly from Ref. 1 + converted to 1989 \$.

$$\text{Cost} = \$36,300 \left(\frac{356.3}{314} \right) = \$41190$$

1-11 Scraper Tractor (why not?)

Cost taken directly from Ref. 1 + converted to 1989 \$.

$$\text{Cost} = \$141,900 \left(\frac{356.3}{314.0} \right) = \$161,016$$

1-12 Reclaim Hopper/Tank, carbon steel,
 $7' \times 4' \times 2' = 419 \text{ gal.}$

Using the small storage tank line from the
Tank Cost Chart by Garrett, p. 303

$$\text{cost} = \$1000 \cdot 1.6 \left(\frac{419+1}{1000 \text{ gal}} \right)^{0.71} \left(\frac{356.3}{320} \right) = \$961$$

1-13 Vibrating Pan, 10 ft long, 42 in. wide., 3 hp

$$\text{cost}^* = \$11,800 \cdot 2.4 \left(\frac{42 \text{ in}}{36 \text{ in}} \right)^{1.0} \left(\frac{104}{204} \right)^{1.0} \left(\frac{356.3}{320} \right) = \$18,394$$

* same cost technique as 1-4.

1-14 Vertical Tunnel Sump Pump, 60 gpm, 70' head,
stainless steel, approx. 2 hp., same cost as 1-8.

$$\text{cost} = \$5679$$

The following Area 1 equipment takes the limestone from the storage area and delivers it to the LEC or LSS units. The 500 MW and 100 MW remaining portions of Area 1's equipment list differs in size and cost due to the different mass rates of stone needed by each.

$$\text{L/S required (T/hr)} \quad \frac{500 \text{ MW}}{34} \quad \frac{100 \text{ MW}}{7}$$

1-15 Live Limestone Feed Belt Conveyor, 200 ft long

Using Table 7-7 in Perry's CHE Handbook

500 MW: 16 in belt, $hp = 200 \left(\frac{0.56}{100} \right) \approx 2 \text{ hp}$

Garrett, p. 273 $\left\{ \text{Cost} = \$38,000 \cdot 2.4 \left(\frac{356.3}{320} \right) = \$101,546 \right.$

100 MW: 14 in belt, $hp \approx 1 \text{ hp}$

$\text{Cost} = \$38,000 \cdot 2.4 \left(\frac{14 \text{ in}}{16 \text{ in}} \right)^{0.6} \left(\frac{356.3}{320} \right) = \$43,727$

1-16 Live Limestone Feed Inclined Belt Conveyor

140 ft long, 35 ft lift

500 MW: 16 in. belt, $hp = 140 \left(\frac{0.56}{100} \right) + 35 \left(\frac{0.46}{10} \right) \approx 3 \text{ hp}$

$\text{Cost} = \$36,000 \cdot 2.4 \left(\frac{356.3}{320} \right) = \$96,201$

100 MW: 14 in. belt, $hp = 140 \left(\frac{0.44}{100} \right) + 35 \left(\frac{0.34}{10} \right) = 2 \text{ hp}$

$\text{Cost} = \$36,000 \cdot 2.4 \left(\frac{14 \text{ in}}{16 \text{ in}} \right)^{0.6} \left(\frac{356.3}{320} \right) = \$88,744$

1-17 Live Limestone Feed Bucket Elevator

that will feed the L/S clay bins, 90' lift

Using the Continuous Bucket Elevator Spec. (Table 7-4)

chart from Perry's to size + Garrett's cost chart yields:

500 MW: Oversize to $10 \times 7 \times 11 \frac{3}{4}$, $hp = 8 + (90 - 75) 0.1 = 10 \text{ hp}$

$\text{Cost} = \$10,000 \cdot 2.4 \left(\frac{70}{241} \right)^{0.37} \left(\frac{356.3}{320} \right) = \$39,709$

100 MW: $8 \times 5 \frac{1}{2} \times 7 \frac{3}{4}$, $hp = 5 + (90 - 75) 0.06 = 6 \text{ hp}$

$\text{Cost} = \$10,000 \cdot 2.4 \left(\frac{44}{241} \right)^{0.37} \left(\frac{356.3}{320} \right) = \$33,441$

1-18 Limestone clay bin / Tank, carbon steel,

Design Basis = 30 hrs, bulk density of L/S $\approx 1.47 = 92 \text{ lb/ft}^3$

$$\underline{500 \text{ MW}}: 37 \text{ T/hr} \left(\frac{2000 \text{ lb}}{\text{T}} \right) \left(\frac{1}{42.15} \right) \left(\frac{7.481 \text{ gal}}{1 \text{ ft}^3} \right) = 6,017 \text{ gal}$$

Since there are 3 tanks in this system; vol/tank $\approx 2000 \text{ gal}$

Using cost chart in Garrett, p. 303 (small storage tank)

$$\text{cost} = \$1,700 \cdot 1.6 \left(\frac{356.3}{320} \right) = \$3,029$$

$$\underline{100 \text{ MW}}: 7 \text{ T/hr} \left(\frac{2000 \text{ lb}}{\text{T}} \right) \left(\frac{1}{42.15} \right) \left(\frac{7.481 \text{ gal}}{1 \text{ ft}^3} \right) = 1138 \text{ gal} \approx 1100 \text{ gal}$$

Use the same cost chart; 1 tank only

$$\text{cost} = \$1,200 \cdot 1.6 \left(\frac{356.3}{320} \right) = \$2,138$$

1-19 Limestone Feed Belt conveyor, 60 ft long

$$\underline{500 \text{ MW}}: 16 \text{ in belt } \text{hp} = (1 \text{ hp})_{\text{belt}} + (2 \text{ hp})_{\text{tr. app.}} = 3 \text{ hp}$$

$$\text{cost} = \$15,000 \cdot 2.4 \left(\frac{356.3}{320} \right) = \$40,084$$

$$\underline{100 \text{ MW}}: 14 \text{ in. belt } \text{hp} = 2 \text{ hp (includes tr. app.)}$$

$$\text{cost} = \$15,000 \cdot 2.4 \left(\frac{14 \text{ in}}{16 \text{ in}} \right)^{0.0} \left(\frac{356.3}{320} \right) = \$36,998$$

1-20 Dust Collecting System, Bag Filter, 2,200 ft^3/min

Use same system for both 500 MW + 100 MW units

Using the Dust Collector Cost Chart; Garrett, p. 279

$$\text{cost} = \$1,800 \cdot 2.2 \left(\frac{356.3}{320} \right) = \$4,409$$

1-21 Weigh Belt that for the LEC system feeds into the pre-reactor bucket elev. or for the LSS, it feeds into the ball mill crusher. (Use a factor of 2 for a weigh belt)

LSS { 500 MW: 16 in. x 10' long

$$\text{cost} = \$3,900 \cdot 2.4 \cdot 2.0 \cdot \underbrace{\left(\frac{356.3}{320}\right)}_{\text{weigh belt}} = \$20,844$$

100 MW: 14 in x 10' long

$$\text{cost} = \$3,900 \cdot 2.4 \cdot 2.0 \left(\frac{14 \text{ in}}{16 \text{ in}}\right)^{0.6} \left(\frac{356.3}{320}\right) = \$19,239$$

Use 3 operating weigh belts (1 spare) for the LEC 11 Tons/hr wet clay, 4 tons L/S.

Using Table 7-7 from Perrys, use a 14" belt at 100 ft/min (may over-sized), 10' long, 0' 1.4

$$\text{hp} = ((0.44) / 100 \text{ ft}) \cdot 10' \approx 1 \text{ hp}$$

Use cost chart from Garrett

EC { $\text{cost} = \$3,900 \cdot 2.4 \cdot 2.0 \left(\frac{14 \text{ in}}{16 \text{ in}}\right)^{0.6} \left(\frac{356.3}{320}\right) = \$19,239$

APPENDIX F

DUCT WORK FOR THE FGD PROCESSES

Duct Work: 500 MW

Duct work for the 500 MW case is taken from Economic Evaluation of FGD Systems, Vol. 5, EPRI CS-3342, October 1986.

$$\text{weight of material} = A_0 \cdot \frac{1}{8}'' \left(\frac{144}{12} \right) \cdot \rho_{\text{carbon steel}} = W_D$$

$A_0 \equiv$ area of duct work material required
 $\rho_{\text{carbon steel}} = 643 \text{ lb/ft}^3$

$$\text{cost} = W_D (\text{lb}) (\$1.38/\text{lb}) \cdot I_i \cdot M_f$$

$I_i \equiv$ installation factor = 1.45
 $M_f = 1.0$ for carbon steel + 2.0 for stainless steel

3-2 Inlet Ducting from Plant to ESP
14,570 ft², carbon steel
cost = \$195,675

3-3 Outlet Ducting from ESP to I.D. Fan
13,440 ft², carbon steel
cost = \$181,501

3-4 Inlet Ducting from I.D. Fan to absorber manifold
8,015 ft², carbon steel
cost = \$107,421

Duct Work: 50019W

3-5 Inlet duct, absorber manifold
13,210 ft², carbon steel
cost = \$177,047

3-6 Outlet duct, absorber manifold
23860 ft², stainless steel
cost = \$639,569

3-7 Bypass, carbon steel, 17960 ft²
cost = \$240,709

3-8 Chimney Transition, carbon steel, 1890 ft²
cost = \$25,331

3-9 I.D. Fan inlet selection damper

* Damper costs are taken directly from ERT-05-3242 and are then converted to 1984 dollars from 1982 dollars

cost = 2 dampers $\cdot$ \$82,000 $\left(\frac{356.3}{314.0} \right)$ = \$186,093

3-10 Scrubber Inlet Isolation damper

cost = 4 dampers $\cdot$ \$131,000 $\left(\frac{356.3}{314.0} \right)$ = \$594,596

3-11 Scrubber Outlet Isolation Damper

cost = 4 dampers $\cdot$ \$147,400 $\left(\frac{356.3}{314.0} \right)$ = \$669,027

Duct Work: 500 MW

3-12 FGD System Bypass Damper

Double-louver, opposed blade - parallel
blade w/ purge air, carbon steel

$$\text{cost} = \$80,200 \left(\frac{356.3}{314.0} \right) = \$91,004$$

The above is for a 500 MW LSS system.

The LEC system requires a plenum to distribute the inlet-outlet gas properly. Using the duct size given by EPRI-CS-3342 for the inlet isolation damper of 14' by 18' yields a flow rate of 42 ft³/s. If the LEC reactor is ~~100~~ tall, it needs to be

$$\text{width (ft)} = 628,000 \frac{\text{ft}^3}{\text{min}} \left(\frac{\text{sec}}{264} \right) \left(\frac{\text{min}}{60 \text{ s}} \right) \left(\frac{1}{100 \text{ ft}} \right) = 52.4$$

where the 2 ft/sec is the required ^{superficial} gas velocity to achieve the proper residence time for a 15" bed depth. This bed depth is the same used at the Ohio University pilot plant. This, of course, is a worst case scenario for face area.

$$\begin{aligned} \text{Plenum shell} &= 2(\text{Area 1} + \text{Area 2}) \\ &= 2(1585 \text{ ft}^2 + 1871 \text{ ft}^2) = 6910 \text{ ft}^2 \end{aligned}$$

assume that this area is needed for turning vane material

3-13 Plenum = (6910 ft²) 2 = 13,820 ft²

$$\text{cost} = \$185,223 \quad \text{for inlet (carbon steel)}$$

$$\text{cost} = \$370,446 \quad \text{for outlet (stainless steel)}$$

Duct Work:
100 MW*

If the same length of duct runs and gas velocity are assumed to be equal in both the 500 MW and 100 MW cases, then the surface area for each piece of duct work is reduced by the same ratio as the gas flow rate is reduced within the duct.

3-2 Inlet Ducting from Plant to ESP, carbon steel
Area = $\frac{1}{5} (14570 \text{ ft}^2)$; cost = \$39,055

3-3 Outlet Ducting from ESP to I.D. fan, carbon steel
Area = $\frac{1}{5} (13,490 \text{ ft}^2)$; cost = \$37,500

3-4 Inlet Ducting from I.D. fan to Absorber, carbon steel
Area = $\frac{1}{5} (8060 \text{ ft}^2)$; cost = \$21,605

3-5 Outlet Ducting from Absorber, stainless steel
Area = $\frac{1}{5} (10,650 \text{ ft}^2)$; cost = \$57,095

3-6 By Pass Ducting, carbon steel
Area = $\frac{1}{5} (17,460 \text{ ft}^2)$; cost = \$48,142

* The same cost equations used to calculate the duct work for the 500 MW case are used in the 100 MW case.

Duct Work: 100 MW

3-7 Chimney Transition, stainless steel
Area = $\frac{1}{5}$ (1890 ft²); cost = \$10,132

3-8 I.D. Fan Inlet Isolation Damper; 6.3' x 11.4'
* Convert damper costs from 500 MW units
to 100 MW units using a 0.6 power factor.
cost = \$164,000 ($\frac{1}{5}$)^{0.6} ($\frac{356.3}{314.0}$) = \$70,851

3-9 Scrubber Inlet Isolation Damper; 12.5' x 16.1'
cost = \$524,000 ($\frac{1}{5}$)^{0.6} ($\frac{356.3}{314.0}$) = \$226,379

3-10 Scrubber Outlet Isolation Damper; 8.9' x 15.2'
cost = \$589,600 ($\frac{1}{5}$)^{0.6} ($\frac{356.3}{314.0}$) = \$254,719

3-11 FGD System Bypass Damper; 4.5' x 7.6'
cost = \$80,200 ($\frac{1}{5}$)^{0.6} ($\frac{356.3}{314.0}$) = \$34,648

3-12 LEC Plenum:

The 100 MW LEC reactor face area is 3140 ft²
(78' tall x 40' wide), resulting in the plenum shell
shown in Figure 4.

Plenum shell = 4404 ft² } does this include
turning vanes?

Inlet carbon steel Plenum = \$59,025

Outlet stainless steel Plenum = \$118,050

Total = \$177,075

Figure 3
Plenum for 500 MW LEC units

Figure 4.

Plenum for 100 MW LEC units

Induced Draft Fans

LSS: 500 MW case

$$\Delta P_{\text{reactor}} \approx 6 \text{ in H}_2\text{O}$$

$$\Delta P_{\text{duct}} = 9 \text{ in H}_2\text{O}$$

$$\Delta P_{\text{total}} = 15 \text{ in H}_2\text{O}$$

$$Q_{\text{min}} = 628,000 \text{ acfm} \quad T = 283^\circ\text{F}$$

$$\text{Air hp} = 0.000157 Q \cdot \Delta P \quad \text{Perry's 5th, p 621}$$

$$\text{shaft hp} = \text{Air hp} / \eta_{\text{iso}} \quad \eta_{\text{iso}} \approx 70\%$$

$$\text{hp} = \frac{0.000157 (628,000) 15}{0.70} \approx 2113 \text{ hp}$$

$$\text{Cost}^* = \$14,000 \cdot 2.2 \cdot \left(\frac{15}{3.5}\right)^{0.3} \left(\frac{356.3}{320}\right) = \$53,068$$

* I.D. fan costs from Garrett, p. 281

LEE: 500 MW case

$$\Delta P_{\text{reactor}} = 6 \text{ to } 10 \text{ in H}_2\text{O}$$

$$\Delta P_{\text{duct}} = 9 \text{ in H}_2\text{O}$$

$$\Delta P_{\text{total}} = 15 \text{ in H}_2\text{O to } 19 \text{ in H}_2\text{O}$$

$$\text{hp}_{15 \text{ in H}_2\text{O}} = 2113 \text{ hp} \quad \text{hp}_{19 \text{ in H}_2\text{O}} = 2676 \text{ hp}$$

$$\text{Cost}_{15 \text{ in H}_2\text{O}} = \$53,068$$

$$\% \text{ diff} = \frac{19 - 15}{15} \times 100 = 26.7\% \text{ cost}$$

$$\text{Cost}_{19 \text{ in H}_2\text{O}} = \$56,967$$

APPENDIX G
LIMESTONE SCRUBBER EQUIPMENT SIZING AND COST

Lime Material Handling^{*1}
(LSS)

500 MW

1-21 Positive Pressure Blower, rotary, for
pneumatic conveying of lime; 3170 scfm, HP=225
Using Blower cost chart in Garrett, p. 261
$$\text{cost} = \$33,000 \cdot 2.2 \left(\frac{356.3}{320} \right) = \$80,836$$

1-22 Fixative (lime) Reagent Storage Tank,
carbon steel, 56,400 gal.
Using Tank cost chart in Garrett, p. 303
$$\text{cost} = \$38,000 \cdot 1.4 \left(\frac{356.3}{320} \right) = \$80,390$$

1-23 Fixative Storage Silo Baghouse, 10,000 acfm, 20 hp
Using cost chart from Garrett, p. 279
$$\text{cost} = \$35,000 \cdot 2.2 \left(\frac{356.3}{320.0} \right) = \$85,735$$

*2 Lime and fly ash is combined in a pug mill
with scrubber waste to form a "fixed" sludge
waste product.

6-9 Fixative Day Bin/Tank
carbon steel, $4000 \text{ ft}^3 = 29,400 \text{ gal}$
Using the Tank Cost Chart in Garrett, p. 303
$$\text{cost} = \$30,000^{*2} \cdot 1.4 \left(\frac{356.3}{320} \right) = \$63,466$$

*2 Large Tank (Field Erected)

LSS (500 MW):

Agitated Tanks and Agitators costs
derived from cost charts from Garrett.

2-4 Mills Product Tank, 21058 gal, glass lined, open top
cost = \$42,000 · 2.5 · 1.6 $\left(\frac{356.3}{320}\right)$ = \$187,058

2-5 Mill Product Tank Agitator, 15 hp, stainless steel
cost = \$4,800 · 2.0 · 1.19 $\left(\frac{356.3}{320}\right)$ = \$12,720

2-7 Limestone Slurry Tank, 329,000 gal, glass lined, open top
cost = \$30,000 $\left(\frac{329,000 \text{ gal}}{10,000 \text{ gal}}\right)^{0.53} \cdot 2.5 \cdot 1.6 \left(\frac{356.3}{320}\right)$ = \$851,061

2-8 Limestone Slurry Tank Agitator, 80 hp, stainless steel
cost = \$11,000 · 2.0 · 1.19 $\left(\frac{356.3}{320}\right)$ = \$29,150

— Reaction Mix Tank, 259,000 gal, open top?, glass lined
cost = \$30,000 $\left(\frac{259,000 \text{ gal}}{10,000 \text{ gal}}\right)^{0.53} \cdot 2.5 \cdot 1.6 \left(\frac{356.3}{320}\right)$ = \$749,88

— Reaction Mix Tank Agitator, 50 hp, stainless steel
cost = \$8,800 · 2.0 · 1.19 $\left(\frac{356.3}{320}\right)$ = \$23,320

LSS (500 Mw):

_____ Filter Feed Tank; 110,000 gal, open top, glass lin

$$\text{Cost} = \$30,000 \cdot 2.5 \cdot 1.6 \left(\frac{110,000 \text{ gal}}{10,000 \text{ gal}} \right)^{0.53} \left(\frac{356.3}{320} \right) = \$476,195$$

_____ Filter Feed Tank Agitator, 20 hp, stainless steel

$$\text{Cost} = \$5,500 \cdot 2.0 \cdot 1.19 \left(\frac{356.3}{320} \right) = \$14,575$$

_____ Thickener Overflow Tank; * Unagitated, carbon
44,000 gal, open top

$$\text{Cost} = \$35,000 \cdot 2.0 \left(\frac{356.3}{320} \right) = \$77,941$$

* Cost derived from Tank cost chart, Garrett, p. 30

LSS (500 MW)

Pumps.

$$bhp = \frac{Q \cdot H \cdot S_p G}{3960 \cdot \text{eff}_p}$$

Q = flow rate - gpm

H = head - ft

$S_p G$ = specific gravity.

$$\text{Annual Utilities} = (bhp) \left(5500 \frac{\text{operating hrs}}{\text{yr}} \right) (0.7457 \frac{\text{kwh}}{\text{hp-hr}}) \left(\frac{1}{\text{eff}_m} \right) (0.0439 \frac{\$}{\text{kwh}})$$

eff_p = efficiency for centrifugal pumps ~ 80%.

eff_m efficiency for motors ~ 90%

4- Slurry Spray Pump; 11,000 gpm, 20 wt% solids, stainless
head ~ 100', $bhp = \frac{11,000 \cdot 100 \cdot 1.2}{3960} = 336.3 \text{ hp}$.

$$\text{cost} = \$ 2,000 \cdot 1.5 \cdot 2.0 \left(\frac{356.3}{320} \right) = \$ 106,250$$

Thickener Overflow pump; 1900 gpm, H = 100 psi ($\frac{4 \text{ Hic}}{2.309 \text{ ft}}$)
Head = 43', $bhp = 40 \text{ hp}$, Centrifugal, stainless

$$\text{cost} = \$ 5,000 \cdot 1.5 \cdot 2.0 \left(\frac{356.3}{320} \right) = \$ 16,702 \text{ } \$ 6,702$$

Thickener Underflow Pump; Reciprocating, stainless
15hp

$$\text{cost} = \$ 14,000 \cdot 3.3 \cdot 2.4 \left(\frac{356.3}{320} \right) = \$ 123,458$$

Filter Feed pump, 900 gpm, 50 psi, 25 hp, stainless

$$\text{cost} = \$ 5,500 \cdot 1.5 \cdot 2.0 \left(\frac{356.3}{320} \right) = \$ 18,372$$

Vacuum Filter System (LSS)

Design Basis: 100 lb/hr/ft² - dry solids

500 MW

6-8 Rotary drum vacuum filter, 86,000 $\frac{\text{lb (dry)}}{\text{hr}}$
Filter area = $86000 \frac{\text{lb/hr}}{[100 \frac{\text{lb}}{\text{hr} \cdot \text{ft}^2}]}$ = 860 ft²
Using filter cost chart from Garrett, p. 282
cost = \$130,000 · 2.8 $\left(\frac{356.3}{320}\right)$ = \$405,291

100 MW (not used in this study)

6-8 Rotary drum vacuum filter, 17,200 $\frac{\text{lb}}{\text{hr}}$
Filter area = 172 ft²
cost = \$65,000 · 2.8 $\left(\frac{356.3}{320}\right)$ = \$202,646

Thickener System (LSS)

The thickener system consists of one clarifier plus a flocculant system which is assumed to be included by the clarifier module factor of 3.0. The flocculant system would include mixing tanks, agitator, metering pumps, water booster pumps and local controls.

500 MW

Design basis: approx. $18 \text{ ft}^2 / \text{ton of dry solids/day}$
 area of clarifier = $(858.9 - 773.0) \times 10^3 \left(\frac{2.59 \times 10^{-16}}{\text{day}} \right) \times 18 \frac{\text{ft}^2}{\text{ton}} \approx 18,500 \text{ ft}^2$

* Flow rates are from stream no. 19 of the material balance
 $\text{Dia. (ft)} = \left(\frac{4}{\pi} (18,500 \text{ ft}^2) \right)^{1/2} = 153'$

Using the Thickener/Clarifier cost chart (sheet P. 30)

$$\text{cost} = \$340,000 \cdot 3.0 \left(\frac{32}{320} \right) = \$1,135,706$$

$$\frac{100 \text{ MW}}{\text{area of clarifier}} = (17.2 \frac{1}{\text{hr}}) \times 10^3 \left(\frac{2.59 \times 10^{-16}}{\text{day}} \right) \cdot 18 = 3711 \text{ ft}^2$$

$$\text{Dia.} = 68.7'$$

$$\text{cost} = \$160,000 \cdot 3.0 \left(\frac{32}{320} \right) = \$534,450$$

Process Mixer System (LSS)

D-103   D-104   D-105   D-106   D-107   D-108   D-109   D-110

The pug mill extruder mixes the Ash, sludge, sludge water, and fixative lime together to form waste fixed sludge for disposal at a landfill.

500 MW*

6-10 Pug Mill Extruder, $300,000 \text{ lb/hr} = 150 \text{ T/hr}$, 200
Using the Size Enlargement cost chart
from Garrett, p. 302
$$\text{cost} = \$18,000 \left(\frac{150 \text{ T/hr}}{10 \text{ T/hr}} \right)^{0.15} 2.2 \left(\frac{356.3}{320} \right) = \$66,187$$

100 MW*

6-10 Pug Mill Extruder, 26 T/hr or 30 T/hr max. cap, 30
$$\text{cost} = \$18,000 \left(\frac{26 \text{ T/hr}}{10 \text{ T/hr}} \right)^{0.15} 2.2 \left(\frac{356.3}{320} \right) = \$50,855$$

* Horsepower rating are from the Chemical Engineers Handbook, 6th Ed., Table 8-64, p. 8-68

APPENDIX H
LEC PROCESS EQUIPMENT SIZING AND COST

LEC - SO₂ Absorption
500 MW

4-2 Continuous Bulk Elevator, 204257 ^{lb}/hr, 1104 ft.
 $204257 \frac{\text{lb}}{\text{hr}} \left(\frac{4^3}{1.47 \cdot 62.32 \cdot 15} \right) = 2230 \frac{\text{ft}^3}{\text{hr}}, 102 \frac{\text{T}}{\text{hr}}$

Using Table 7-9 in Perry's $\rightarrow 16 \times 8 \times 11\frac{3}{4}$, hp = 30 hp.

$$\text{Cost} = \$12,000 \cdot 2.4 \left(\frac{128 \frac{\text{ft}^3}{\text{hr}}}{24 \text{ in}^2} \right)^{0.37} \left(\frac{356.3}{320} \right) = \$59,573$$

4-3 Bulk limestone screw conveyor (Recycle),
181,562 ^{lb}/hr = 91 T/hr, 60 ft long, $\rho_{\text{LCS}} = 92 \frac{\text{lb}}{\text{ft}^3}$

Using Table 7-6 for a screw conveyor, convert
T/hr into a density of $50 \frac{\text{lb}}{\text{ft}^3}$.

$$G = 91 \frac{\text{T}}{\text{hr}} \left(\frac{50 \frac{\text{lb}}{\text{ft}^3}}{92 \frac{\text{lb}}{\text{ft}^3}} \right) = 49 \frac{\text{T}}{\text{hr}}$$

Select the largest available: 16" screw, hp = 13 $\left(\frac{92}{50} \right) = 24$

$$\text{Cost} = \$7000 \cdot 2.2 \cdot \left(\frac{16}{9} \right)^{1.2} \left(\frac{356.3}{320} \right) = \$34,201$$

4-4 Live Btm Screw conveyor (Twin Screw)

51 T/hr screw, double hp = cost of 4-3

16" twin screw, 48 hp, cost = \$68,402

LEC - Solids Separation / Regeneration

6-1 Continuous Bucket Elevator (Post-Reactor)
 $217,806 \frac{lb}{hr} = 109 \frac{Thr}{hr} = 2,400 \frac{43}{hr}$, 50' high

Using Table 7-9 in Perry's - $16 \times 8 \times 11\frac{3}{4} - 12$

$$\text{cost} = \$7,900 \cdot 2.4 \cdot \left(\frac{128.1^2}{24.1^2}\right)^{0.37} \left(\frac{356.3}{320}\right) = \$39,219$$

6-2 Horizontal Screw Conveyor, 50' long, Twin Screw
 $109 \frac{Thr}{hr} = 2,400 \frac{43}{hr}$

Same Price and cost as 4-4

$$\text{cost} = \$68,402, \text{ hp} = 48$$

6-3 } see ball mill & screen section
6-4 }

6-5 Solid Waste Disposal Belt Conveyor
 $36,244 \frac{lb}{hr} = 18.1 \frac{Thr}{hr}$, 100 ft long, 50' incline
from Table 7-7 in Perry's, choose a 14" belt
 $\text{hp} = 100 (0.44 \frac{\text{hp}}{100 \text{ ft}}) + 50 (0.34 \frac{\text{hp}}{10}) = 3 \text{ hp}$

$$\text{cost} = \$24,000 \cdot 2.4 \left(\frac{14.2}{16.2}\right)^{0.6} \left(\frac{356.3}{320}\right) = \$59,196$$

6-6 Ball Mill Bag Filter, 2,200 acfm

Using Garrett's Dust Collectors Cost Chart,

$$\text{cost} = \$1,800 \cdot 2.2 \left(\frac{356.3}{320}\right) = \$4409$$

LEC Ball Mills

6-3 (500 MW)

The mass flow rate is: $217806 \frac{\text{lb}}{\text{hr}} \approx 2600 \frac{\text{T}}{\text{day}}$

Using a No. 35 sieve yields a 10'x10' Marcy Ball mill with a 750 hp drive from Table 8-18 in Perry's CHE Handbook.

Using Garrett's Mill cost chart, p. 292

$$\text{Cost} = \$50,000 \left(\frac{1089 \frac{\text{T}}{\text{hr}}}{10 \frac{\text{T}}{\text{hr}}} \right)^{0.62} \cdot 2.3 \cdot 1.25 \left(\frac{356.3}{320} \right) = \$703,45$$

6-3 (100 MW)

Same method as 500 MW unit

$$121562 \frac{\text{lb}}{\text{hr}} \left(\frac{\text{T}}{2000 \text{ lb}} \right) = 61 \frac{\text{T}}{\text{hr}} \text{ or } 1459 \frac{\text{T}}{\text{day}}$$

5'x4' Marcy Ball Mill, hp = 50 hp

$$\text{Cost} = \$52,000 \left(\frac{61 \frac{\text{T}}{\text{hr}}}{10 \frac{\text{T}}{\text{hr}}} \right)^{0.62} \cdot 2.3 \cdot 1.25 \left(\frac{356.3}{320} \right) = \$491,109$$

Screens LEC - 500 MW

Using the screen sizing technique given in Perry's Chemical Engineering Handbook:

From Table 21-6: No. 30 screen $\Rightarrow a = 0.0234$, $d = 0.0154$

Figure 21-15:

$$\frac{0.4}{0.05} = \frac{C_u}{0.0234}$$

$$C_u \equiv \text{unit capacity} = 0.1872 \text{ Tons/hr. ft}^2$$

Figure 21-16

$$\text{for square holes: } F_{ca} = 100 \cdot C_u^2 \cdot m^2, \quad m = \frac{1}{a+d}$$

$$F_{ca} = 100 \cdot (0.0234)^2 \left(\frac{1}{0.0234 + 0.0154} \right)^2 = 36.37$$

Table 21-7
 $F_s = 1.0$

Using Equation 21-3: Screen area = $C_u \cdot G / (C_u \cdot F_{ca} \cdot F_s)$

$$\text{Area} = 0.4 (217806 \text{ lb/hr}) \left(\frac{2000 \text{ lb}}{1 \text{ Tons}} \right) [0.1872 \text{ Tons/hr. ft}^2 \cdot 36.37 \cdot 1.0]$$

$$\text{Area} = 6 \text{ ft}^2$$

Using the double deck screen cost chart by Gairrett:

$$\text{Cost} = \$6,800 \cdot 2.8 \cdot 1.25 \cdot 1.6 \left(\frac{356.3}{320} \right) = \$42,400$$

factor   still   double deck

APPENDIX I
THE CHEMICAL ENGINEERING HANDBOOK SIZING CHARTS

TABLE 7-5 Screw-Conveyor Capacities and Loading Conditions*

Material class†	Screw diam. in.	Max. lump size, in.		Capacity, cu. ft./hr.		Approx. area occupied by material‡
		25%	100%	At 1 r.p.m.	At max. r.p.m.	
A, B, C, D, and H 16, 26, 36	6	3/4	3/4	2.27	375	
	9	1 1/4	1 1/4	8.0	1,200	
	12	2	2	19.3	2,700	
	14	2 1/4	2 1/4	30.8	4,000	
	16	3	3	46.6	5,600	
	18	3 1/2	3 1/2	66.1	7,600	
A, B, C, D, and H 17, 27, 37	6	3/4	3/4	1.5	75	
	9	1 1/4	1 1/4	5.6	280	
	12	2	2	13.3	663	
	14	2 1/4	2 1/4	21.1	1,055	
	16	3	3	31.4	1,570	
	18	3 1/2	3 1/2	45.4	2,270	
A, B, C, D, and H 18, 28, 38	6	3/4	3/4	0.75	25	
	9	1 1/4	1 1/4	2.8	90	
	12	2	2	6.7	200	
	14	2 1/4	2 1/4	10.5	300	
	16	3	3	15.7	425	
	18	3 1/2	3 1/2	22.7	590	

*FMC Corporation, Material Handling Systems Division. To convert cubic feet per hour to cubic meters per hour, multiply by 0.02832; to convert screw diameter in inches to the nearest screw size in centimeters, multiply by 2.5. See Table 21-6 for the conversion of particle sizes from one measurement system to another.

†These classifications cover a broad list of materials that generally can be handled in a screw conveyor. Special consideration must be given to applications handling materials with the following characteristics:

Highly corrosive. Class N

Degradable, affecting use or salability. Class T

Interlocks or mats. Class X

Highly aerated or of fluid nature. Class Y

‡Capacity for horizontal conveyor uniformly fed. Volumetric capacity is based on material slightly agitated or duffed. Material highly duffed or aerated will decrease in weight and increase in volume.

§Maximum capacity for economical service.

¶Percentages higher than those indicated will result in excessive wear on hanger bearings and couplings.

enclosed dust-tight quick-acting valve for free-flowing materials. The rotary-vane feeder (Fig. 7-1c) delivers a uniform predetermined volume of material and may be driven from the screw or independently by constant- or variable-speed drive. Rack-and-pinion gates (Fig. 7-1d) are well suited to free-flowing materials in bins, hoppers,

tanks, or silos and are also used as side inlet gates (Fig. 7-1e) for heavy or lumpy materials.

Typical discharge arrangements are shown in Fig. 7-2. Plain discharge openings (Fig. 7-2a) equipped with a discharge spout (Fig. 7-2b) are most common, although the open-end trough (Fig. 7-2c) is frequently used, as is the discharge-trough end (Fig. 7-2e). Open-bottom troughs (Fig. 7-2g) are often used for spreading material uniformly over a storage area. Flat-bottomed rack-and-pinion gates (Fig. 7-2f) allow selective discharge, as do hand slide gates (Fig. 7-2d). However, for perishable materials, the curved slide gate (Fig. 7-2h) eliminates the dead-storage pocket. Enclosed rack-and-pinion gates (Fig. 7-2j) give dust-tight operation, and rotary cutoff valves (Fig. 7-2i) allow quick shutoff and are readily adaptable to remote control. Air-cylinder-actuated gates have become more and more prominent because of the low investment required and the ease of connecting to automatic process-control centers.

BELT CONVEYORS

The belt conveyor is almost universal in application. It can travel for miles at speeds up to 5.08 m/s (1000 ft/min) and handle up to 4539 metric tons/h (5000 tons/h). It can also operate over short distances at speeds slow enough for manual picking, with a capacity of only a few kilograms per hour. However, it is not normally applicable to processing operations, except under unusual conditions.

Belt-conveyor slopes are limited to a maximum of about 30°, with those in the 18 to 20° range more common. Direction changes can occur only in the vertical plane of the belt path and must be carefully designed as vertical curves or relatively flat bends. Belt conveyors inside the plant may have higher initial cost than some other types of conveyors and, depending on idler design, may or may not require more maintenance. However, a belt conveyor given good routine maintenance can be expected to outlast almost any other type of conveyor. Thus, in terms of cost per ton handled, outstanding economy records have been established by belt conveyors.

Belt-conveyor design begins with a study of the material to be handled. Since weight per cubic meter or foot is an important factor, it should be accurately determined with the material in an as-handled condition. It is not wise to rely solely on published tables of weight per cubic meter or foot for various materials, since many processing operations will affect this by fluffing or compacting the material. Lump size is important, too. For a 600-mm (24-in) belt, uniform lump size can range up to about 102 mm (4 in). For each 152-mm

TABLE 7-6 Screw-Conveyor Data for 50-lb./ft.³ Material and Pipe-Mounted Sectional Spiral Flights*

Capacity		Diam of flights in.	Diam of pipe, in.	Diam of shaft, in.	Hanger centers ft.	Max. size of lumps				Max. torque capacity lb.-ft.	Feed section diam. in.	hp at motors					Max. hp capacity at speed listed	Cost \$	
Tons /h	ft ³ /h					40 lumps	20 to 25%	10% or less	Speed r.p.m.			15-ft max. length	30-ft max. length	45-ft max. length	60-ft max. length	75-ft max. length			
5	200	3	2½	2	10	1	1½	2½	40	7500	8	0.43	0.85	1.27	1.69	2.11	4.5	2030	
10	400	10	2½	2	10	1	1½	2½	35	7500	9	0.55	1.09	1.69	2.25	3.00	3.75	4.6	2250
15	400	10	2½	2	10	1	1½	2½	40	7500	9	1.27	2.25	3.36	3.94	4.93	9.6	2250	
		12	2½	2	12	1	2	3	45	7500	10	1.27	2.25	3.36	3.94	4.93	9.4	2655	
		12	2½	3						16400		1.27	2.25	3.36	3.94	4.93	11.7	2710	
20	500	12	2½	2	12	1	2	3	40	7500	10	1.69	3.00	3.94	4.97	5.63	7.2	2655	
			2½							16400		1.69	3.00	3.94	4.97	5.63	15.6	2710	
25	1000	12	2½	2	12	1	2	3	75	7500	10	2.12	3.75	4.93	5.63	6.55	9.0	2655	
			3½	3						16400		2.12	3.75	4.93	5.63	6.55	9.0	2710	
		14	2½	3		1½	2½	3½	45	16400	12	2.12	3.75	4.93	5.63	6.55	11.7	3150	
30	1200	14	3½	3	12	1½	2½	3½	35	16400	12	2.25	3.94	5.05	6.75	7.50	14.3	3150	
35	1400	14	2½	3	12	1½	2½	3½	65	16400	12	2.62	4.58	5.90	7.00	7.75	16.9	3625	
40	1600	16	2½	3	12	1½	3	4	50	16400	14	3.00	4.50	6.75	5.00	10.00	13.0	3825	

*Fairfield Engineering Co. data in U.S. customary system. To convert cubic feet per hour to cubic meters per hour, multiply by 0.02832; to convert tons per hour to metric tons per hour, multiply by 0.9078, and to convert screw size in inches to the nearest screw size in centimeters, multiply by 2.5.

†Capacities are based on screws carrying 31 percent of their cross section and, in the case of feed sections with half-pitch flights, based on 100 percent of their cross section.

‡Pipe sizes given are for 1-in. (6.35-mm) flights.

§Horsepowers listed are calculated for average conditions and are of the proper motor size with factors for length of conveyor, momentary overloads, etc., taken into consideration.

¶These are approximate 1980 prices for 30-ft (9-m) lengths, carbon steel construction, dustproof electrical equipment for common three-phase voltages. Drives are included in prices.

7-10 HANDLING OF BULK SOLIDS AND PACKAGING OF SOLIDS AND LIQUIDS

TABLE 7-7 Belt-Conveyor Data for Troughed Antifriction Idlers*

Belt width in (cm)	Cross-sectional area of load ft ² (m ²)	Belt speed, ft./min (m/min)		Belt plies		Maximum lump size, in (mm)		Belt speed, ft./min (m/min)		Capacity and hp for 100-lb./ft. ³ material				Add for tripper hpt
		Normal	Maximum	Mini- mum	Maximum	Sized material, 50% under	Un-sized material, not over 20%	ft./min	(m/min)	Capacity tons/h (metric tons/h)	hp/10-ft (3.05-m) lift	hp/100-ft (30.48-m) centers		
14 (35)	0.11 (.010)	200 (61)	300 (91)	3	5	2.0 (.51)	3.0 (.76)	100 (30.5)	200 (61.0)	32 (.29)	0.34	0.44		2.0
								300 (91.5)		96 (.87)	1.04	1.22		
16 (40)	0.14 (.013)	200 (61)	300 (91)	3	5	2.5 (.64)	4.0 (1.02)	100 (30.5)	200 (61.0)	44 (.40)	0.46	0.56		2.5
								300 (91.5)		132 (1.20)	1.36	1.58		
18 (45)	0.18 (.017)	250 (76)	350 (107)	4	6	3.0 (.76)	5.0 (1.27)	100 (30.5)	250 (76.2)	54 (.49)	0.58	0.70		3.0
								350 (106.7)		190 (1.72)	2.00	2.42		
20 (50)	0.22 (.020)	250 (76)	350 (107)	4	6	3.5 (.89)	6.0 (1.52)	100 (30.5)	250 (76.2)	66 (.60)	0.70	0.84		3.20
								350 (106.7)		230 (2.09)	2.44	2.90		
24 (60)	0.33 (.030)	300 (91)	400 (122)	4	7	4.5 (1.14)	9.0 (2.03)	100 (30.5)	300 (91.5)	98 (.89)	1.02	1.02		3.5
								400 (121.9)		392 (3.56)	4.08	4.04		
30 (75)	0.53 (.049)	300 (91)	450 (137)	4	8	7.0 (1.78)	12.0 (3.05)	100 (30.5)	300 (91.5)	158 (1.43)	1.60	1.50		5.0
								450 (137.2)		710 (6.45)	7.20	6.74		
36 (90)	0.78 (.072)	400 (122)	600 (183)	4	9	9.0 (2.03)	15.0 (3.81)	100 (30.5)	400 (121.9)	230 (2.09)	2.44	1.59		7.0
								600 (182.9)		1380 (12.53)	14.60	9.52		
42 (105)	1.09 (.101)	400 (122)	600 (183)	4	10	10.0 (2.54)	18.0 (4.57)	100 (30.5)	400 (121.9)	330 (3.00)	3.50	2.28		9.5
								600 (182.9)		1980 (17.97)	23.20	13.68		
48 (120)	1.46 (.136)	400 (122)	600 (183)	4	12	12.0 (3.05)	21.0 (5.33)	100 (30.5)	400 (121.9)	440 (3.99)	4.66	3.04		12.5
								600 (182.9)		2640 (23.97)	28.00	18.20		
54 (135)	1.90 (.177)	450 (137)	600 (183)	6	14	14.0 (3.56)	24.0 (6.10)	100 (30.5)	450 (137.2)	570 (5.17)	6.04	3.94		20.0
								600 (182.9)		3420 (31.05)	36.20	23.60		
60 (150)	2.40 (.223)	450 (137)	600 (183)	6	16	16.0 (4.06)	28.0 (7.11)	100 (30.5)	450 (137.2)	720 (6.54)	7.64	4.98		25
								600 (182.9)		4320 (39.21)	45.80	29.90		

* Fairbaird Engineering Co. data in U.S. customary system. Metric conversion is rounded off. For inclined conveyors, add lift horsepower to center horsepower for total horsepower. For terminals multiply horsepower by the following factors: 0-50 ft (15.2 m), 1.20; 51-100 ft (30.3 m), 1.10; 101-150 ft (45.7 m), 1.05. For countershaft drives, multiply horsepower by 1.05 for each reduction (cut gears).

[†] Tripper horsepower is based on material bulk density of 100 lb./ft.³ (1602 kg./m.³) and a belt speed of 300 ft./min (91.4 m./min).

FIG. 7-3 Typical belt-conveyor idler and plate-support arrangements. (a) Flat belt on flat-belt idlers. (b) Flat belt on continuous plate. (c) Troughed belt on 20° idlers. (d) Troughed belt on 45° idlers with rolls of unequal length. (e) Troughed belt on 45° idlers with rolls of equal length. (f) Troughed belt on continuous plate. (FMC Corporation, Material Handling Systems Division.)

moderate several factors. Figure 7-4a shows details for one type of rubber seal on a metal skirt plate. It is particularly important that material be loaded onto the belt in its center and in the direction of its travel, preferably with lumps falling on a layer of fine material. Fines can be delivered to the belt first by notching the feed chute or installing a screen section or grizzly bars. Figure 7-4b shows a heavy-duty loading-section design using not only rubber idler rolls but an

additional short pad belt. Mass-flow bins and/or bin-flow-assisting devices are often used to minimize segregation of fines and to assure a uniform feed from a hopper onto a conveyor belt.

A clean discharge is vital to good belt life. On the return run the carrying side of the belt is in contact with the return rollers, and any material adhering to it is ground in or deposited on the roller. Extremely sticky material may require a belt-cleaning device in the

Belt width, B
(in (cm))

Trough width,
A, in (cm)

Skirt seal D
(in (cm))

FIG. 7-4 (a) roller-bearing

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TABLE 7-9 Bucket-Elevator Specifications for Continuous Buckets on Chain*

Size of bucket and bucket spacing, in (mm)†	Elevator centers, ft‡	Capacity, tons/h (metric tons/h)§	Size of lumps handled, in (mm)¶	r/min, head shaft	hp required at head shaft	Additional hp/ft for intermediate lengths	Shaft diameter, in		H
							Head	Tail	
9 x 5½ x 7½ - 9 (203 x 140 x 197) - (203)	25	35 (31.7)	1 (25.4)	28	1.8	0.06	1½	1½	2
	50	35 (31.7)	1 (25.4)	28	3.4	0.06	2½	1½	2
	75	35 (31.7)	1 (25.4)	28	5.0	0.06	2½	1½	2
10 x 7 x 11½ - 12 (254 x 178 x 298) - (305)	25	60 (54.5)	1½ (38.1)	23	3.0	0.10	2½	1½	2
	50	60 (54.5)	1½ (38.1)	23	5.5	0.10	2½	1½	2
	75	60 (54.5)	1½ (38.1)	23	8.0	0.10	2½	1½	2
12 x 7 x 11½ - 12 (305 x 178 x 298) - (305)	25	70 (63.5)	1½ (38.1)	23	3.5	0.12	2½	1½	2
	50	70 (63.5)	1½ (38.1)	23	6.5	0.12	2½	1½	2
	75	70 (63.5)	1½ (38.1)	23	9.5	0.12	3½	2½	2
14 x 7 x 11½ - 12 (356 x 178 x 298) - (305)	25	90 (81.3)	1½ (44.5)	23	4.0	0.14	2½	2½	2
	50	90 (81.3)	1½ (44.5)	20	7.5	0.14	2½	2½	2
	75	90 (81.3)	1½ (44.5)	20	11	0.14	3½	2½	2
14 x 8 x 11½ - 12 (356 x 203 x 298) - (305)	25	100 (90.8)	2 (50.8)	20	5.0	0.17	2½	2½	2
	50	100 (90.8)	2 (50.8)	20	9.3	0.17	2½	2½	2
	75	100 (90.8)	2 (50.8)	20	13.3	0.17	3½	2½	2
16 x 8 x 11½ - 12 (406 x 203 x 298) - (305)	25	115 (104.4)	2 (50.8)	20	6.0	0.20	2½	2½	2
	50	115 (104.4)	2 (50.8)	20	11	0.20	3½	2½	2
	75	115 (104.4)	2 (50.8)	20	16	0.20	4½	2½	2
18 x 8 x 11½ - 12 (406 x 203 x 298) - (305)	25	130 (118.0)	2 (50.8)	20	7	0.22	2½	2½	2
	50	130 (118.0)	2 (50.8)	20	13	0.22	3½	2½	2
	75	130 (118.0)	2 (50.8)	20	20	0.22	4½	2½	2

* From Stepanens-Adamson Division, Allis-Chalmers Corporation.

† Bucket size given: width x projection x depth. Assumed bucket linear speed is 150 ft/min (45.7 m/min).

‡ Elevator centers to nearest SI equivalent are 25 ft ≈ 8 m, 50 ft ≈ 15 m, and 75 ft ≈ 23 m.

§ Capacities and horsepower are given for materials having bulk densities of 100 lb/ft³ (1602 kg/m³). For other densities these will vary in direct proportion. 50-lb/ft³ material will reduce the capacity and horsepower required by 50 percent.

¶ If the total amount of lump product is less than 15 percent of the total, lump size may be twice that given.

FIG. 7-6 Types of skip hoists and skip-hoist paths. Fairfield Engineering Co.

mitted over the entire support length rather than at a specific point. Regardless of drive type, care should be taken to mount the conveyor properly so that supporting structures will not be damaged by the frequency of vibration of the conveyor should in no case be the natural frequency of the supporting structure.

Mechanical vibrating conveyors are designed to operate at specific frequencies and do not perform well at other frequencies. Carefully designed alterations. Thus they are not adaptable to capacity changes except by varying the depth of the trough. Positive eccentric drives maintain their free magnitude of stroke regardless of load, and serious drive o

FIG. 7-7 Vibration-conveyor classification. (Modern Materials Handling.)

TABLE 8-6 Performance Data of Blake Swing Jaw Crusher*

Crusher size receiving opening, in.	Approx. rough wt., lb.	Approx. speed, r.p.m.	Required hp.	Capacity at setting, tons/hr.	Setting, in.
7 × 10	6,000	275	7	7-10	2½
10 × 24	18,500	275	15	25-30	2½
14 × 24	25,000	275	25	45-55	4
18 × 36	51,000	250	40	70-90	5
24 × 36	80,000	200	75	150-175	5
				180-210	7
30 × 42	115,000	200	100	190-230	6
				240-270	9
36 × 48	160,000	200	125-150	250-270	6
				290-320	9
48 × 60	350,000	175	175-200	400-475	6
				525-600	9

*Kennedy Van Saun Corp. To convert inches to centimeters, multiply by 2.54; to convert pounds to kilograms, multiply by 0.4535; to convert horsepower to kilowatts, multiply by 0.746; and to convert tons per hour to kilograms per hour, multiply by 907.

TABLE 8-7 Performance Data of Dodge Jaw Crusher*

Crusher size, in.	R.p.m.	Jaw motion, in.	Hp.	Capacity, tons/hr.			
				Setting, in.			
4 × 6	275	1½	3	1½	1	1	1
7 × 9	200	1½	5	1	1	2	3
8 × 12	220	1½	10	1½	3	4	4
11 × 15	200	1½	15	2	4	5	5

*Allis-Chalmers Corporation. To convert inches to centimeters, multiply by 2.54; to convert pounds to kilograms, multiply by 0.4535; to convert horsepower to kilowatts, multiply by 0.746; and to convert tons per hour to kilograms per hour, multiply by 907.

reciprocating motion of the jaws causes the opening to vary between close and wide. Specifications usually are based on close settings. The setting is adjustable.

Capacities of Kennedy Van Saun swing jaw crushers (Blake type) are given in Table 8-6. The capacities and power consumption of a Dodge (Allis-Chalmers) type of jaw crusher operating on a tough ore that offers considerable resistance to reduction are shown in Table 8-7.

Performance data of overhead eccentric crushers with straight jaw plates are given in Table 8-8. Power and capacities of similar

crushers with curved jaw plates are 50 percent greater. The jaw crusher (Pennsylvania Crusher Corp.) is a version of head eccentric in which the eccentric is positioned well swinging jaw plate to which it is attached. Consequently the jaw meets the rock firmly and squarely. There is no rubbing or vertical motion, and so abrasion is eliminated and the last longer. Kue-Ken jaw crushers are most advantageous for primary and secondary reduction of hard, abrasive rock to a coarse but uniform product. Depending on machine size, they can be as large as 1.2 m (48 in.), and product top size may be 1.3 cm (½ in.). Capacities range from a few hundred pounds up to 634 Mg/h (700 tons/h) when crushing hard dry quartz.

Jaw crushers are preferred when the crusher gap is more important than the throughput. They can accept a larger size of gyratory crushers of the same capacity. If the required throughput in tons per hour is less than the square of the gap in the crusher is more economical (Taggart, *Handbook of Mining*, Wiley, New York, 1945). Gyratory crushers are preferred for larger throughputs. Jaw crushers can also handle clayey material than other types of primary crushers.

Crusher Product Sizes Table 8-9 relates product size to charge setting of the crusher in terms of the percent smaller size in the product. Size-distribution curves differ for various materials crushed, and a general set of curves is not available.

GYRATORY CRUSHERS

The development of improved supports and drive mechanism allowed gyratory crushers to take over most large hard-ore crushing applications. The largest expense of these crushers is relining them. Operation is intermittent; so power demand and the total power cost is not great.

Design and Operation The gyratory crusher (see Fig. 8-1) consists of a cone-shaped pestle oscillating within a larger cone mortar or bowl. The angles of the cones are such that the passage decreases toward the bottom of the working pestle consists of a mantle which is free to turn on its spindle is driven from an eccentric bearing below. The motion causing attrition can occur only when pieces are simultaneously at the top and bottom of the passage owing to different radii at these points.

The circular geometry of the crusher gives a favorable nip angle in the horizontal direction. The nip angle in the vertical is less favorable and limits feed acceptance. The angle is determined by the shape of the mantle and bowl similar to that of a jaw crusher since the jaw geometry onto the curved profile of the gyratory vertical outline.

Primary crushers have a steep cone angle and a small ratio. Secondary crushers have a wider cone angle; the finer product to be spread over a larger passage area and

TABLE 8-8 Performance Data of Overhead Eccentric Jaw Crushers*

Crusher size, in.	r/min	Jaw motion, in.	Hp.	Capacity, tons/hr.						
				Setting, in.						
				2	3	4	5	6	8	10
10 × 24	300	1	11	32	55					
10 × 26	300	1	11	42	67	90				
12 × 48	250	1	75	70	110	140	150			
15 × 24	300	1	11	37	55	70	59			
15 × 36	275	1	75	60	91	110	125	150		
15 × 48	250	1	100	95	130	150	200	225		
22 × 48	250	1	100		140	165	220	240		
24 × 36	260	1	125		115	150	190	230		
30 × 42	230	1	175		125	190	250	300	400	500
42 × 48	225	1	200			270	335	405	540	675
48 × 60	200	1	300				440	570	645	770

*Kennedy Van Saun Corp. To convert inches to centimeters, multiply by 2.54; to convert pounds to kilograms, multiply by 0.4535; to convert horsepower to kilowatts, multiply by 0.746; and to convert tons per hour to kilograms per hour, multiply by 907.

Fig. 8-9 Relation of Product Size to Discharge Setting of Crusher*

Kind of feed	% of product passing a square opening equal to discharge setting of crusher			
	Limestone	Granite	Trap rock	Ores
Quarry-run	55-90	70-75	65-70	55-90
Prescalped	50-65	65-70	60-65	80-85
Quarry-run	55-90	75-80	65-70	85-90
Prescalped	50-65	70-75	60-65	80-85
Screened	85-90	90-95	75-80	95-90
Screened	70-75	65-70	65-70	70-75

*From "Crushing Theory and Practice," Allis-Chalmers Mfg. Co.
 *These type, or crushers with equivalent speeds and throws: opening measured from tip of corrugations on one jaw plate to bottom of corrugations in opposing jaw.
 *For standard, or reduction, types with non-choking concaves. Single-togge jaw product, on screened feed, will approximate that of gyratory-type secondary crushers with non-choking concaves.
 *For high-speed short-throw crushers, such as Newhouse or Type R.

Fig. 8-22 Primary gyratory crusher with spider suspension. (Nordberg Industrial)

the wear over a wider area. Wear occurs to the greatest extent in the lower, fine-crushing zone.
 The three general types of gyratory crusher are the suspended-spindle, the supported-spindle, and the fixed-spindle types. Primary gyratories are designated by the size of feed opening and secondary

or reduction crushers by the diameter of the head in feet and inches. There is a close opening and a wide opening as the mantle gyrates with respect to the concave ring at the outlet end. The close opening is known as the close setting or the close-side setting and sometimes as the closed-side setting, while the wide opening is known as the wide-side or open-side setting. Specifications usually are based on close settings. The setting is adjustable by raising or lowering the mantle.

The length of the crushing stroke greatly affects the capacity and the screen analysis of the crushed product. A very short stroke will give a very evenly crushed product but will not give the greatest capacity. A very long stroke will give the greatest capacity, but the product will contain some very fine particles, a larger amount of medium-size pieces, and a greater proportion of large fragments.

Performance Crushing occurs through the full cycle in a gyratory crusher, and this produces a higher crushing capacity than a similar-sized jaw crusher, which crushes only in the shutting half of the cycle. Gyratory crushers also tend to be cheaper and are easier to operate. They operate most efficiently when they are fully charged. These crushers are designed to operate with the main shaft fully buried in charge. Power consumption for gyratory crushers is also lower than that of jaw crushers. These are preferred over jaw crushers when capacities of 800 Mg/h (900 tons/h) or higher are required.

The crushing rate of a gyratory generally depends on the hardness of the material being crushed and on the amount of product-size material in the feed. For this reason gyratories are often operated in parallel with a scalping grizzly screen, provided the added cost of the screen is less than the cost of increased crusher capacity.

Primary gyratories will accept feed directly from truck or railcar. Most manufacturers make both mechanical and hydraulically supported types. Figure 8-22 shows a Nordberg primary gyratory crusher with spider suspension. It is available in 1- to 1.5-m (42-, 48-, 54-, and 60-in) feed sizes. Table 8-10 gives capacity data for the Superior gyratory crusher (Allis-Chalmers).

Gearless primary gyratory crushers (Kennecott Van Saun) are driven by a multi-V belt from an electric motor. They are available in a larger series of 10 sizes with feed openings from 0.3 to 1.4 m (12 to 54 in). Capacity ranges from 68 to 1360 Mg/h (75 to 1500 tons/h), while power ranges from 37 to 336 kW (50 to 450 hp) on ordinary rock. A finer series of 10 sizes has feed openings from 4.4 to 43 cm (1 1/4 to 17 in) and capacities from 1 to 363 Mg/h (1 to 400 tons/h). Secondary crushers are driven through a gear from the motor.

The multistage fine-reduction crusher (Traylor) features upper and lower stages. The upper stage functions as a distributing feeder for the lower or finishing stage. The upper stage accomplishes about half of the crushing and provides feed of proper size for the lower stage.

8-34 SIZE REDUCTION AND SIZE ENLARGEMENT

are available in sizes of 1.8 to 4.9-m (6 to 16-ft) diameter and 2.4 to 10 m (8 to 33 ft) in length; wet overflow mills are available in sizes of 1.5- to 4.5-m (5- to 15-ft) diameter and 2.1 to 7.6 m (7 to 25 ft) in length. These mills are powered with 37- to 4500-kW (50- to 6000-hp) motors.

Kennedy air-swept and air-lift grinding systems are used extensively for pulverizing power-plant coal, for firing cement kilns and metallurgical furnaces, and for grinding phosphates and other ores. They simultaneously grind and dry materials in a closed circuit with nonmechanical air separators for kiln feed and kiln firing. This system can be adjusted to produce and maintain 38 to 90 percent passing 200 mesh with control of fine and coarse ends. For firing systems the cheapest fuel obtainable may be used, such as high-ash, high-moisture, low-grindability, and low-Btu coal. Wear and maintenance are low, and foreign material cannot damage the system.

Kennedy rod mills are built in sizes from 0.9 by 1.8 to 4 by 6.1 m (3 by 6 to 13 by 20 ft) for both wet and dry grinding.

Since rod mills avoid packing, they are particularly useful for the reduction of damp or sticky materials. Ordinarily they are used to produce materials in the 6- to 20-mesh range, although both finer and coarser products are readily obtained. They are used to grind ores, cement clinker, and for numerous other materials.

Marcy Ball Mill (Fig. 8-30) This is traditionally a grate-discharge mill, used to give a high throughput rate for a high circulating load in the wet and dry grinding of ores. The data in Table 8-18 must not be used for design but for orientation only. Mill design must be based on pilot experiments or other techniques previously discussed.

Koppers Co., Inc. The Hardinge conical mill is used extensively for both wet and dry grinding in open and closed circuits. The conical mill is similar to the cylindrical mill in that it consists of a drum rotating about its horizontal axis and operating in much the same way, but unlike the cylindrical mill, it has conical ends instead of straight ends.

Hardinge wet-grinding mills are supplied with discharge arrangements illustrated in Fig. 8-32 for high, medium, or low pulp levels, the use of which depends on the particular problem under consideration. For dry grinding, a vertical grate with low-pulp-level discharge lifters is used.

Hardinge mills are available in sizes from 0.9 to 4.3 m (3 to 14 ft) in diameter with a length 1½ to 2½ times the diameter. These are used for the wet grinding of sandstone, quartzite, and granites and the dry grinding of abrasives and coke.

Autogenous Tumbling Mills The principle of the ball mill has been employed in some cases in which coarse lump feed will serve as the grinding medium while it is itself being ground. The Cascade mill (Koppers Co., Inc.) is a wet or dry autogenous mill. It has been built in diameters up to 11 m (36 ft). For all sizes the length-to-diameter ratio is 1:3. A relatively slow speed is used to promote cascading action and prevent segregation of large lumps into the center of the mill. A grate discharges through a trommel screen in the discharge trunnion. The latter scuffs oversize pieces, which are conveyed to the feed end of the mill. The feed can be run-of-mine or primary crusher product, with provision made in the bin and feeder arrangements to

assure uniform, constant feed-size distribution. For an application, see Fig. 8-59.

The Aerofall mill (Aerofall Mills Ltd.) is a dry autogenous mill of a shape similar to the Cascade mill. It is an air-swept mill and therefore does not need a discharge grate, but it does require an air-handling system and cyclone. Run-of-mine feed is reduced in closed circuit to final product sizes. Autogenous mills eliminate wear of ball media, although frequently a 5 percent charge of large balls is included. Their small length-to-diameter ratios make them suitable for very high circulating loads in closed-circuit operations.

Allis-Chalmers Rockcyl mills are used for either dry or wet autogenous grinding.

Rockcyl mills with a length-to-diameter ratio of approximately 1 eliminate all except primary crushing stages, all rod-mill grinding, and part or all of the ball-milling stages in a conventional flow sheet.

Intermediate rock grinding reduces, to a desired size, the minus 100-mesh (150-µm) product from a closed-circuit crushing operation. Size rock from the primary crusher product serves as the medium. Rockcyl mills with length-to-diameter ratios of approximately 1:2 are used in intermediate rock grinding.

Secondary rock grinding is the reduction, to a desired size, of rod-mill or primary rock-mill product, using sized media from either a crushing stage or a primary mill. This is often referred to as rock pebble milling.

Rockpeb mills with a length-to-diameter ratio of approximately 2:1 are used in secondary rock grinding. Because Rockcyl lifters wear faster than liner plates, lifter bars are designed for easy removal and replacement. A large-diameter spout feeder provides free flow of feed into the mill through the short-length trunnion bearing.

NONROTARY BALL OR BEAD MILLS

These include stirred and vibratory types. In the first, a central paddle wheel or impeller armature stirs the media at speeds from 100 to 1500 r/min. In the second an eccentric motion is imparted either to an armature or to the shell at frequencies up to 1800/min. In media oscillate in one or more planes and commonly rotate very slowly. Stirred mills use media 0.6 cm (¼ in) in size or smaller, whereas vibratory mills use larger media for the same power input. Vibratory mills may grind dry, but most stirred mills are restricted to wet milling. Solids vary from 25 to 70 percent, depending on feed size and rheology. Media load varies from 5 to 6 times the charge mass. Unlike in rotary ball mills, some sedimentation occurs.

Although applications overlap, vibrational equipment is generally used for hard-grinding operations (ZrSiO₄, SiO₂, TiO₂, Al₂O₃, etc.) while stirred grinders are mainly used for dispersion and soft grinding (dyes, clays, CaCO₃, etc.). Stirred mills are also called sand mills when Ottawa sand is used as media.

Contamination and grinder body wear may be minimized in these types by the use of resilient coatings.

Stirred Mills The Sweco dispersion mill (Sweco, Inc.) has counterrotating radial armatures to move the grinding medium in a vibrating chamber. The high-amplitude DM-70 has a working

TABLE 8-18 Illustrative Performance of Marcy Ball Mills

Size, ft.	Ball charge, tons	Hp. to run	Mill speed, r.p.m.	Capacity, tons/24 hr. (based on medium-hard ore)							
				No. 8 sieve*	No. 20 sieve	No. 35 sieve	No. 48 sieve	No. 65 sieve	No. 80 sieve	No. 100 sieve	No. 150 sieve
				20% -200	35% -200	50% -200	60% -200	70% -200	80% -200	85% -200	93% -200
3 × 2	0.85	5-7	35	19	15	12	10	8	6½	5	4
4 × 3	2.73	20-24	30	80	64	53	45	36	28	22	18
5 × 4	5.25	44-50	27	180	145	120	102	82	63	51	41
6 × 4½	8.90	85-95	24	375	300	250	210	170	135	105	85
7 × 5	13.10	135-150	22½	640	510	425	360	290	225	180	145
8 × 6	20.2	220-245	21	1100	885	735	625	500	390	310	250
9 × 7	30.0	345-380	20	1800	1450	1200	1020	815	635	505	410
10 × 8	46.50	700-750	18	3680	2960	2450	2100	1700	1325	1050	850
12 × 12	90.5	1260-1345	16.4	7125	5725	4750	4070	3290	2570	2035	1650

* Sieve through which substantially all the material can pass.

NOTE: To convert horsepower to kilowatts, multiply by 0.746; to convert tons to megagrams, multiply by 0.907; and to convert tons per 24 hours to megagrams per day, multiply by 0.907.

8-68 SIZE REDUCTION AND SIZE ENLARGEMENT

which, in turn, is a very effective control variable. For materials containing soluble constituents, such as fertilizer formulations, the total solution phase is the controlling factor, and not simply the amount of water used. In addition, solubility is affected by temperature.

Increased retention time leads to larger and denser agglomerates of higher wet strength. Interaction of agglomerate size, retention time, throughput rate, and moisture content is shown qualitatively in Fig. 8-70. Numerical values vary with the particular application. With the inclined-pan agglomerator retention time can also be increased by increasing pan depth and speed or by decreasing pan slope. Retention time in drum agglomerators can be influenced by internal baffling, as in the multicone drum previously referred to.

FIG. 8-70 Relationship between mean pellet size and residence time (throughput rate) as a function of liquid content for pan agglomerators [Adapted from Pietsch, *Autbereitung. Tech.*, 7, 177 (1966).]

A uniform product with constant properties is obtained only when the agglomeration device is operated uniformly, with a constant feed of raw material of uniform properties as well as constant liquid flow rate or a flow rate in constant proportion to the solids rate.

Mixer Agglomerators Many types of solid-liquid mixers can be used to form agglomerates. Control of the amount of liquid phase and the intensity and duration of mixing determine agglomerate size.

Pug mills and other intensive mixers use agitators and other tools to provide a positive rubbing and shearing action. Advantages claimed for this method over tumbling include harder, stronger agglomerates owing to the kneading action, the ability to process plastic sticky materials, and greater tolerance in accommodating variations in operating conditions. Hignett, in Sauchelli, ed., *Chemistry and Technology of Fertilizers*, p. 259. Because of the greater compaction achieved, less liquid phase is required for agglomeration in a mixer than in a tumbling device [Sherrington, *Chem. Eng. (London)*, no. 120, CE101 (1965)]. Disadvantages include generally higher maintenance and power requirements and an irregular product form which may require further shaping, such as in a tumbling dryer.

Pug mills (blenders, pug mixers, and paddle mixers) consist of a horizontal trough or troughs containing a mixing shaft or shafts. Attached to the shaft are mixing blades or bar, rod, paddle, and other designs. The vessel may be of single- or double-trough design, although the latter configuration is more popular. Twin shafts rotate in opposite directions, throwing the materials forward and to the center as the pitched blades on the shaft pass through the charge. Characteristics of a range of pug mixers available for fertilizer granulation are given in Table 8-64.

In applications such as the preparation of tableting feeds, the manufacture of detergent powders, and the preparation of "instant" food products, the aim is to produce small agglomerates (usually 2-mm diameter and less) with improved flow, wetting, dispersing, or dissolution properties. This is accomplished by superficially wetting the feed powder, often with less than 5 percent of bridging liquid in the form of a spray, steam, mist, etc. The wetting is carried out in a relatively dry state in standard or specialized powder mixers in which the mass becomes moist rather than wet or pasty.

TABLE 8-64 Characteristics of Pug Mixers for Fertilizer Granulation*

Model	Material bulk density, lb ft ⁻³	Approximate capacity, tons/h	Size (width X length), ft	Plate thickness, in.	Shaft diameter, in.	Speed, r/min	Drive hp
A	25	8	2 X 8	1/2	3	56	13
	30	15	2 X 8	1/2	3	56	20
	75	25	2 X 8	1/2	3	56	25
	100	30	2 X 8	1/2	3	56	25
B	25	20	4 X 8	1/2	4	56	30
	30	60	4 X 8	1/2	4	56	30
	75	90	4 X 8	1/2	4	56	75
	100	120	4 X 8	1/2	5	56	100
C	25	20	4 X 12	1/2	5	56	50
	30	60	4 X 12	1/2	5	56	100
	75	90	4 X 12	1/2	5	56	100
	100	120	4 X 12	1/2	5	56	100

*Freco International, Inc. To convert pounds per cubic foot to kilograms per cubic meter, multiply by 16; to convert tons per hour to megagrams per hour, multiply by 0.907; to convert feet to centimeters, multiply by 30.5; to convert inches to centimeters, multiply by 2.54; and to convert horsepower to kilowatts, multiply by 0.746.

A continuous flow-mixing system used to cluster-agglomerate a wide range of food powders is represented in Fig. 8-71. The feed powder is introduced to the moistening zone by means of a pneumatic conveyor and a rotary valve. The dry powder falls in a narrow stream between two jet tubes, which inject the agglomerating fluid in a highly dispersed state. Steam, water, or solvents or a combination of these are used. In addition, air at ambient temperature is introduced through radial wall slots in the moistening chamber to induce a vortex motion. Control of this air flow controls the flow pattern and particle temperature. The reduced temperature serves to condense fluid onto the particles, while the vortexing motion induces particle-particle collisions. The clustered material then drops through an air-heated chamber onto a conditioning conveyor, where it is allowed sufficient time to reach a uniform moisture distribution. The material then passes to an afterdryer, cooler, and sifter, followed by bagging of the selected product.

Other types of mixer agglomerators are described by Capes, *Particle Size Enlargement*, op. cit., p. 531.

THERMAL PROCESSES

Bonding and agglomeration by temperature elevation or reduction are applied either in conjunction with other size-enlargement processes or as a separate process. Agglomeration occurs through one or more of the following mechanisms:

1. Drying of a concentrated slurry or wet mass of fines
2. Fusion
3. High-temperature chemical reaction
4. Solidification and/or crystallization of a melt or concentrated slurry during cooling

Sintering and Heat Hardening In powder metallurgy compacts are sintered with or without the addition of binders. In ore processing the agglomerated mixture is either sintered or indurated. Sintering refers to a process in which fuel is mixed with the ore and burned on a grate. The product is a porous cake. Induration, or heat hardening, is accomplished by combustion of gases passed through the bed. The aim is to harden the pellets without fusing them together, as is done in the sintering process.

Ceramic bond formation and grain growth by diffusion are the two prominent reactions for bonding at the high temperature (1100 to 1370°C, or 2000 to 2500°F, for iron ore) employed.

In addition to agglomeration, other useful processes may occur during sintering and heat hardening. For example, carbonates and sulfates may be decomposed, or sulfur may be eliminated. Although the major application is in ore beneficiation, other applications, such as the preparation of lightweight aggregate from fly ash and the formation of clinker from cement raw meal, are also possible. Nonferrous sinter is produced from oxides and sulfides of manganese, zinc, lead, and nickel. An excellent account of the many possible appli-

TABLE 21-6 U.S. Sieve Series and Tyler Equivalents (ASTM—E-11-61)

Sieve designation		Sieve opening		Nominal wire diam.		Tyler equivalent designation
Standard	Alternate	mm.	in. (approx. equivalent)	mm.	in. (approx. equivalent)	
107.6 mm.	4.24 in.	107.6	4.24	6.40	0.2520	
101.6 mm.	4 in.	101.6	4.00	6.30	0.2480	
90.5 mm.	3 1/2 in.	90.5	3.50	6.08	0.2394	
76.1 mm.	3 in.	76.1	3.00	5.80	0.2283	
64.0 mm.	2 1/2 in.	64.0	2.50	5.50	0.2165	
53.8 mm.	2.12 in.	53.8	2.12	5.15	0.2028	
50.8 mm.	2 in.	50.8	2.00	5.05	0.1988	
45.3 mm.	1 3/4 in.	45.3	1.75	4.85	0.1909	
38.1 mm.	1 1/2 in.	38.1	1.50	4.59	0.1807	
32.0 mm.	1 1/4 in.	32.0	1.25	4.23	0.1665	
26.9 mm.	1.06 in.	26.9	1.06	3.90	0.1535	1.050 in.
25.4 mm.	1 in.	25.4	1.00	3.80	0.1496	
22.6 mm.	7/8 in.	22.6	0.875	3.50	0.1378	0.883 in.
19.0 mm.	3/4 in.	19.0	0.750	3.30	0.1299	0.742 in.
16.0 mm.	5/8 in.	16.0	0.625	3.00	0.1181	0.624 in.
13.5 mm.	0.530 in.	13.5	0.530	2.75	0.1083	0.525 in.
12.7 mm.	1/2 in.	12.7	0.500	2.67	0.1051	
11.2 mm.	7/16 in.	11.2	0.438	2.45	0.0965	0.441 in.
9.51 mm.	3/8 in.	9.51	0.375	2.27	0.0894	0.371 in.
8.00 mm.	5/16 in.	8.00	0.312	2.07	0.0815	0.312 in.
6.73 mm.	0.265 in.	6.73	0.265	1.87	0.0736	0.265 in.
6.35 mm.	1/4 in.	6.35	0.250	1.82	0.0717	
5.66 mm.	No. 31 1/2	5.66	0.223	1.68	0.0661	0.223 in.
4.76 mm.	No. 4	4.76	0.187	1.54	0.0606	0.187 in.
4.00 mm.	No. 5	4.00	0.157	1.37	0.0539	0.157 in.
3.36 mm.	No. 6	3.36	0.132	1.23	0.0484	0.132 in.
2.83 mm.	No. 7	2.83	0.111	1.10	0.0430	0.111 in.
2.38 mm.	No. 8	2.38	0.0937	1.00	0.0394	0.0937 in.
2.00 mm.	No. 10	2.00	0.0787	0.900	0.0354	0.0787 in.
1.68 mm.	No. 12	1.68	0.0661	0.810	0.0319	0.0661 in.
1.41 mm.	No. 14	1.41	0.0555	0.725	0.0285	0.0555 in.
1.19 mm.	No. 16	1.19	0.0469	0.650	0.0256	0.0469 in.
1.00 mm.	No. 18	1.00	0.0394	0.580	0.0228	0.0394 in.
841 micron*	No. 20	0.841	0.0331	0.510	0.0201	0.0331 in.
707 micron*	No. 25	0.707	0.0278	0.450	0.0177	0.0278 in.
595 micron*	No. 30	0.595	0.0234	0.390	0.0154	0.0234 in.
500 micron*	No. 35	0.500	0.0197	0.340	0.0134	0.0197 in.
420 micron*	No. 40	0.420	0.0165	0.290	0.0114	0.0165 in.
354 micron*	No. 45	0.354	0.0139	0.247	0.0097	0.0139 in.
297 micron*	No. 50	0.297	0.0117	0.215	0.0085	0.0117 in.
250 micron*	No. 60	0.250	0.0098	0.180	0.0071	0.0098 in.
210 micron*	No. 70	0.210	0.0083	0.152	0.0060	0.0083 in.
177 micron*	No. 80	0.177	0.0070	0.131	0.0052	0.0070 in.
149 micron*	No. 100	0.149	0.0059	0.110	0.0043	0.0059 in.
125 micron*	No. 120	0.125	0.0049	0.091	0.0036	0.0049 in.
105 micron*	No. 140	0.105	0.0041	0.076	0.0030	0.0041 in.
88 micron*	No. 170	0.088	0.0035	0.064	0.0025	0.0035 in.
74 micron*	No. 200	0.074	0.0029	0.053	0.0021	0.0029 in.
63 micron*	No. 230	0.063	0.0025	0.044	0.0017	0.0025 in.
53 micron*	No. 270	0.053	0.0021	0.037	0.0015	0.0021 in.
44 micron*	No. 325	0.044	0.0017	0.030	0.0012	0.0017 in.
37 micron*	No. 400	0.037	0.0015	0.025	0.0010	0.0015 in.

*These sieves correspond to those proposed as an international I.S.O. standard. It is recommended that wherever possible these sieves be included in all sieve analysis data or reports intended for international publication.
†These sieves are not in the fourth-root-of-2 series, but they have been included because they are in common usage.

Screening machines actuated by rotating unbalanced weights have a symmetrical shaft through the screen body with an unbalanced flywheel on each end. Counterweights on each flywheel, which may be moved in relation to the shaft, permit adjustment of the amplitude of vibration. On some makes of machines the complete shaft assembly is contained in a unit bolted to the top of the screen body.

The horizontal-type screen is actuated by an enclosed mechanism consisting of off-center weights geared together on short horizontal shafts. The mechanism is usually mounted between the side plates and above the screen body (Fig. 21-11).

Electrically Vibrated Screens These screens are particularly useful in the chemical industry. They handle very successfully many light, fine, dry materials and metal powders from approximately 4

FIG. 21-10 Ty-Rock screen with air-seal enclosure. (W. S.

mesh to as fine as 325 mesh. Most of these screens have high-speed (25 to 120 vibrations/s) low-amplitude vibration by means of an electromagnet.

Typical of these is the Hummer screen used in the chemical industry. Figure 21-12 shows one used in the fertilizer industry for handling mixed chemical fertilizers.

Oscillating Screens These screens are characterized by speed oscillations (5 to 7 oscillations per second) (300 in a plane essentially parallel to the screen cloth).

Screens in this group are usually used from 0.013 mesh. Some light free-flowing materials, however, can be at 200 to 300 mesh. Silk cloths are often used.

Reciprocating Screens These screens have many uses in chemical work. An eccentric under the screen supports a gyratory (about 0.05-m (2-in) diameter) to reciprocating motion at the discharge end. Frequent oscillations per second (500 to 600 r/min), and since inclined about 5°, a secondary high-amplitude normal about 0.0025 m (1/8 in) is also set up. Further vibration is caused by balls bouncing against the lower surface of the screen.

These screens are used extensively in the United States as standard equipment in many chemical and process industries for handling fine separations even down to 300 mesh. They handle a variety of chemicals, usually dry, light, or bulky light metal powders, powdered foods, and granular materials. They are not designed for handling heavy tonnages of materials or gravel. Machines of this type are exemplified by Figure 21-13.

Gyratory Screens These are boxlike machines, essentially square, with a series of screen cloths nested atop one another, supplied by eccentrics or counterweights, in a near-circular orbit. In some machines a supplemental action is set up. Most gyratory screens have an auxiliary caused by balls bouncing against the lower surface of the cloth. A typical machine is shown in Fig. 21-14. Machines are operated continuously and can be located in line with

FIG. 21-11 Mechanically vibrated horizontal screen. (Courtesy Concentrator Company, Inc.)

service. Tables of comparative openings are available for selecting a screen specification with a specific opening or for picking a specification having a heavier or lighter wire but the same opening.

Punched Plates These are available in a variety of perforations including round, square, hexagonal, and elongated openings. Punched metal will generally wear longer than wire cloth and has more rigidity, which is an advantage in certain applications. However, it usually does not give the capacity per unit area that wire cloth does and is generally heavier. Its use is normally limited to the coarser separations.

Bar Screens These screens are generally used in handling large and heavy pieces of material. They are formed from rails, rods, or bars, suitably shaped; made from rolled steel or castings; fixed in parallel position and held by crossbars and spacers. Bars, which taper in thickness from top to bottom and may also taper in width from one end to the other, are recommended because they tend to avoid blinding.

Rod decks, composed of spring-steel rods approximately 0.6 m (2 ft) long, sprung into position between molded-rubber blocks, and held in position by means of rubber spacers, are also available.

Factors in Selecting Screening Equipment In attempting to pick a screening machine for a specific screening problem it should be emphasized that generalized formulas and charts used to predict screen capacity can give only an approximation because of the many variables which may affect performance. Screen consultants will readily admit that they must depend largely on laboratory tests and field experience. However, two governing factors bear mention: generally width of screen relates to capacity and length of screen relates to efficiency. Width is necessary to reduce the bed thickness to a practical maximum and length to allow the undersize to be removed without an inordinate amount of fines in the oversize. In attempting to choose a screening machine for a particular screening application the customer and manufacturer should consider the following: (1) Full description of material involved, including the name and type of material, bulk density, and physical characteristics such as hardness, particle shape, flow characteristics (free-flowing, sluggish, or sticky), percent of moisture and temperature. (2) Normal and maximum total rate of feed to screen. (3) Complete sieve analysis of screen feed, including maximum lump size, and sieve analysis of desired product. (4) Separation or separations required and the purpose of screening. Can slotted or rectangular openings be used in place of square openings? (5) Is screening to be accomplished dry or wet, and what amount of water is available? (6) Other important factors include method of delivering feed to the screen, open or closed circuit, open or enclosed screens, previous screening experience with the material, flow sheet or description of related equipment, operating hours per day, power available, and space limitations.

Variables in Screening Operations It will readily be seen that many variables in a screening operation can easily be changed in the field, and practical operators will always be trying to improve their operations or adapt them to new products or processes. Capacity and efficiency in screening operations are closely related. Capacity may be large if low efficiency is not objectionable. Usually, as the tonnage to a screen is increased, efficiency is decreased.

Method of Feed The screening machine must be fed properly in order to obtain maximum capacity and efficiency. The feed should be spread evenly over the full width of the screen cloth and approach the screen surface in a direction parallel to the longitudinal axis of the screen and at as low a practical velocity as is possible.

Screening Surfaces It is generally agreed that the most efficient screening results when a series of single-deck screens is used. This is true because lower decks of multiple-deck screens are not fed so that their entire area is used and because each separation requires a different combination of angle, speed, and amplitude of vibration for maximum performance.

Angle of Slope The optimum slope of inclined vibrating screens is that which will handle the greatest volume of oversize and still remove the available undersize required by the standards of the particular operation. To separate a material into coarse and fine fractions the bed thickness must be limited so that vibration can stratify

the load and allow fines to work their way to the screen surface and pass through the opening. Increased slope naturally increases the rate of travel, and at a given rate it reduces the bed thickness.

In the oscillating screen the angle of inclination must be coordinated with the speed and stroke for best results.

Direction of Rotation In circle-throw screens somewhat greater efficiency can be obtained by counterflow rotation, that is, having the material move down the screen against the rotation. Screens rotating with the flow of material will handle greater tonnage and operate at a lower angle.

Vibration Amplitude and Frequency Speed and amplitude of vibration should be designed to convey the material properly and to prevent blinding of the cloth. They are somewhat dependent upon the size and weight of the material being handled and are related to the angle of installation and the type of screen surface. The object, of course, is to see that the feed is properly stratified for the most efficient separation.

Performance Formulas

Screen Efficiency There is confusion concerning the meaning of screen efficiency, as a uniform method for figuring efficiency has never been established. A sound method of evaluating screen performance is given by W. S. Tyler, Inc., Mentor, Ohio, in its *Sieve Handbook*, no. 53. In this formula, when material put through the screen is the desired product, "efficiency" is the ratio of the amount of undersize obtained to the amount of undersize in the feed.

$$E = (R \times d) / b \quad (21-1)$$

where E = efficiency, R = percent of fines through the screen, d = percent finer than the designated size in screen fines, and b = percent finer than the designated size in screen feed.

When the object is to recover an oversize product from the screen, efficiency may be expressed as a ratio of the amount of oversize obtained to the amount of true oversize:

$$E = (O \times c) / a \quad (21-2)$$

where O = percent of oversize over the screen, c = percent coarser than the designated size in screen oversize, and a = percent coarser than the designated size in screen feed.

Other formulas for the derivation of screen efficiency are used. Taggart (*Handbook of Mineral Dressing*) gives the formula

$$E = 100 \times \frac{100(e - v)}{e(100 - v)}$$

where E is the efficiency, e is the percentage of undersize in the feed, and v is the percentage of undersize in the screen oversize.

Graphical methods of evaluating efficiency, using sieve analyses, are also employed and are recommended when serious research on screening is done.

Estimating Screen Capacity Various methods of predicting screening capacity have been proposed, and each has its limitations. The throughflow method of Matthews uses the following equation:

$$A = 0.4 C_u C_o F_o F_s \quad (21-3)$$

where A = screen area
 C_u = throughflow rate
 C_o = unit capacity
 F_o = open-area factor
 F_s = slotted-area factor.

The unit capacity C_u can be determined from Fig. 21-15. Figure 21-16 can be used to determine the open-area factor F_o , and the slotted-opening factor F_s for various screen types is given in Table 21-7.

Probability Screening Principle Probability screening uses the fact that particles moving almost at right angles to a screening surface are not likely to pass through when the particle size is greater than about half of the distance between the screen elements. Screens utilizing the probability principle are manufactured by Dutch State Mines (DSM), Bartles (CTS), and Morgensen. The last-named incorporates multiple decks. Higher throughput, longer screen life, and

FIG. 21-15 Unit capacity (C_u) for square-opening screens. To convert inches to meters, multiply by 0.0254; to convert tons per hour-square foot to kilograms per second-square meter, multiply by 2.7182.

lower capital costs are claimed for these screening systems. The performance of several types of probability screens was reviewed by Moir (op cit.)

Constant Screen Bed Thickness Another screen design approach to improve screening performance was patented (British Patent 124,904, 1970) and is utilized in the HCC-Burstlein system. Rather than the gradually decreasing bed thickness from feed to discharge of the conventional screening system, the Burstlein system breaks up the screen surface into a number of inclined elements that are each inclined at different angles and vibrated differently to thin

Aperture	Formula	
Rectangular openings	$F_{os} = \frac{2(a_1 a_2 + a_1 a_3 + a_1 a_4 + a_2 a_3 + a_2 a_4 + a_3 a_4)}{a_1 + a_2 + a_3 + a_4} \times 100$ F_{os} is open area, %; a_1, a_2, a_3, a_4 are diameter of wire, or horizontal width of bar (for plate), a is clear opening dimension	21-4
Square openings Specified by opening size	$F_{os} = 100 \left(1 - \frac{a}{s} \right)^2$ a is diameter of wire, or horizontal width of bar (for plate), s is clear opening dimension	21-5
Square openings Specified in mesh, m	$F_{os} = 100 \left(1 - \frac{1}{m} \right)^2$ m is mesh number	21-6
Parallel-rod decks	$F_{os} = \frac{2(a_1 a_2 + a_1 a_3 + a_1 a_4 + a_2 a_3 + a_2 a_4 + a_3 a_4)}{a_1 + a_2 + a_3 + a_4} \times 100$	21-7
Special weaves	Assuming $a_3 = a_1$, $a_4 = a_2$ $F_{os} = 100 \left[\frac{2(a_1 a_2 + a_1 a_3 + a_1 a_4 + a_2 a_3 + a_2 a_4 + a_3 a_4)}{a_1 + a_2 + a_3 + a_4} \right] \quad (21-8)$	21-8

FIG. 21-16 Open-area factor (F_{os}) for flow-through screen-capacity calculation.

TABLE 21-7 Slotted-Opening Factors

Screen type	Length-to-width ratio
Square and slightly rectangular openings	Less than 2
Rectangular openings	Equal to or greater than 2 but less than 4
Slotted openings	Equal to or greater than 4 but less than 25
Parallel-rod decks	Equal to or greater than 25

out the feed end and thicken the discharge end, thereby finer particles to penetrate the bed and reach the screen more uniformly. Advantages are claimed in capacity and operating costs.

Noise and Safety Noise is generated in screening by the impact of the feed material on the screen surface. The mechanism also generates noise. Rubber and rubber-clad reduce substantially feed-impact noise with the added longer life of the decks. Noise from the drive mechanism is reduced by enclosing the mechanism in a box or by adding to the side plates to dampen the noise.

Depending on the feed materials, the dust generated in operation may be hazardous because of possible toxicity. These hazards must be carefully evaluated before design of the facility and selecting the apparatus.

Testing Sieves Many product specifications require a test of material in terms of given percentages passing or retained on test sieves. Test sieves are also generally used to determine efficiency of screening devices and the work of crushing machinery.

It is essential that standard sieves, with standard-size openings, be used for sieve analyses. The time of screening and the method of agitating the material on the sieve should also be standardized. Many industries the practice of specifying the test-sieve size and the time and method of sieving is followed. An excellent reference on the theory and use of standard testing sieves is given in *Sieve Handbook*, No. 53, published by W. S. Tyler, Inc. (Ohio).

U.S. Sieve Series The American Society for Testing and Materials in cooperation with the National Bureau of Standards and the American National Standards Institute has further refined the U.S. sieve series, combining the former coarse and fine series into one series with a fourth-root-of-2 ratio (Table 21-6). The openings of individual sieves have remained unchanged except for minimum openings in sieves coarser than 0.00673 m (6.73 mm). In the series, sieves 1 mm and coarser are identified by opening size in millimeters, and those finer than 1 mm by their openings in micrometers.

Tyler Standard Sieve Series Many users have their own standard sieves and tests based on using Tyler standard screen scale tests (Table 21-6). The only difference between the U.S. and Tyler sieves is the identification method. Tyler standard sieves are identified by nominal mesh number and the U.S. sieves are identified by millimeters or micrometers. The Tyler standard sieve scale series has as its base a 200-mesh sieve in which the opening is 7.37×10^{-3} m (0.0029 in.) and the diameter 5.35×10^{-3} m (0.0021 in.).

International Test Sieve Series The International Organization for Standardization (ISO) has been intensifying its efforts to establish an international test sieve series. At a meeting held in The Hague in October 1959, the ISO provisionally recommended adoption as an international standard 19 sieves as shown by Table 21-6. These sieves correspond to every alternating sieve of the fourth-root-of-2 U.S. sieve series from 0.022-m (7/8-in.) opening to 0.075-m (3/4-in.) mesh. The ISO has prepared a manual of sieving procedures available through the American Society for Testing and Materials.

The Ro-Tap testing sieve shaker (Fig. 21-17) manufactured

TABLE 21-8 Sizes, Limitations, and Major Applications of Wet Classification Machines

Figure	Type of classifier	Normal size range, ft.			Normal mesh of separation range*	Normal feed tonnage range	Max. over-size in feed	Normal over-flow, % solids range	Normal feed density, % solids range	Normal sand product, % solids range	Motor range, hp.	Typical applications
		Width	Diam	Max. length								
21-60	Non-mechanical, Cone classifier		2-12		28-325	2-100 tons/hr.	3/4 in.	5-30	Not critical	35-60	None	For desliming and primary dewatering
21-61	Liquid cyclone		10 mm. to 4 ft.	9	35 mesh to 5 μ	35-1500 gal/min	14-325 mesh	5-30	10-60	55-70	Power for pressure head 5-60 lb./sq. in.	For medium or fine separations and closed circuit grinding
21-62	Mechanical, Drag classifier	1-10		Not critical	28-200	5-350 tons/hr.	1 1/2 in.	5-30	Not critical	70-83	1-10	For desliming, conveying, and closed circuit grinding
21-63 and 21-64	Rake and spiral classifiers	1-20		40	20-200	5-350 tons/hr.	1 in.	5-30	Not critical	75-83	1/2-25	Closed circuit grinding, washing and dewatering, desliming, process feed control
21-65	Bowl classifier	1 1/2-20	4-28	40	100-325	5-200 tons/hr.	3/4 in.	5-25	Not critical	75-80	Flow: 1-2 1/2 Rake: 1-25	Closed-circuit grinding usually in secondary circuits
21-66	Bowl desilter	4-16	20-50	40	100-325	5-250 tons/hr.	3/4 in.	1-15	Not critical	75-83	Flow: 1-10 Rake: 5-25	Recovery of fine sand, limestone, coal, and fine phosphate rock from large flow volumes
21-67	Hydroseparator		10-150		100-325	5-700 tons/hr.	1 1/2 in.	1-20	Not critical	30-50	1-15	For fine separation where large feed volumes are involved and drainage not critical
21-68	Solid bowl centrifuge		18-54 in.	70 in.	200 mesh to 1 μ	10-600 gal/min	1/4 in.	1-40	5-50	10-70	15-150	For fine-size fractionating
21-69	Counter-current classifier		1 1/2-10	40	35-100	1-600 tons/hr.	3 in.	5-30	Not critical	75-83	1/2-25	Sand-slime separations, washing, closed circuit grinding
21-70	Hydraulic, Jet sizer	1 1/2-20		5-20	8-150	2-100 tons/hr.	1/4 in.	1-10	30-60	40-60	1-2 for air pressure	Multiproduct unit for exceptionally clean sands fractionated into narrow size ranges. Min. 3 tons hydraulic water per ton sand
21-71	SuperSorter ¹	6		40	8-150	40-150 tons/hr.	3/8 in.	1-10	30-60	40-60	1 to operate pincher valves	Multiproduct unit for exceptionally clean sands fractionated into narrow size ranges. Min. 3 tons hydraulic water per ton sand
21-72	SiphonSizer ²		1-30		14-150	1-100 tons/hr.	1 in.	1-10	30-60	40-60	None	Two product unit efficient for desliming and exceptionally clean sands, washing, closed-circuit grinding. Min. 2 tons hydraulic water per ton sand

NOTE: Equipment can be provided in stainless steel and rubber-covered steel for corrosion resistance. To convert feet to meters, multiply by 0.3048, to convert tons per hour to kilograms per second, multiply by 0.2520, to convert inches to meters, multiply by 0.0254, and to convert horsepower to kilowatts, multiply by 0.7457.

*Size of screen retaining 18% of the overflow solid.

¹Trademark of Deister Concentrator Company, Inc.

²Trademark of Dorr-Oliver Inc.

APPENDIX J

THE CHEMICAL ENGINEERING ECONOMICS COSTING CHARTS

APPENDIX 1

EQUIPMENT COST ESTIMATES

The following section provides very rough cost estimates for a wide variety of process equipment. It must be remembered in using these charts that there is no such thing as an exact, definite, fixed price for any piece of equipment of a given size or capacity. As with buying merchandise, clothing or a car there are many styles, quality differences, optional features and designs to meet specific needs or services. Presumably charts could be made for each of these variations, but the number would be large and confusing, and for many preliminary estimates the engineer would not know exactly what he wanted at that stage of the design, so only average, representative equipment should be more useful. Again, a range of prices could be shown, but usually a single line is more practical, keeping in mind that the price could quite normally vary considerably depending upon the exact design requirements and the company policy on quality, maintenance, and so on.

With these generalities in mind, the following charts have been taken from a number of sources. Most are from cost estimating articles or books, although some are from recent vendor quotations. In case only a single source was available, that reference has been noted. However, often many sources were available and a somewhat biased consensus of opinion curve was selected. In this case the sources were not noted except for inclusion in the reference list at the end of the appendix. In case different variables were used as the sizing parameter, the most logical one in the author's opinion was selected.

All costs were factored to an early 1987 basis, or a chemical engineering index number of 320. When equations were available for the cost relationship they were listed beneath the charts, and when straight line functions existed for the costs on log-log paper a sizing exponent was given:

$$\text{cost size 2} = \text{cost size 1} \left(\frac{\text{size 2}}{\text{size 1}} \right)^{\text{size exponent}}$$

In a number of references various authors have estimated the fraction of the purchased equipment cost that it takes to install the equipment. This generally included freight and shipping costs, foundations, mounting, and simple electric and piping connections, such as switch gear, starters, flange connections, and so on. Unfortunately these numbers often varied widely, so the range and average are both listed when available:

A similar number that also includes all of the adjacent minor equipment and connections is sometimes listed in the literature (principally by Guthrie 1975 and Ulrich 1984) covering the cost of purchase and installation of the major equipment as well as all of the supporting equipment around each major unit. This is called the module factor, and when available is also listed under the charts as the range given by different authors and the average value.

$$\text{cost of the installed module} = \text{purchase price} \times \text{module factor}$$

As a final item under the equipment cost graphs, often a simple factor can be used to estimate the cost of some other material, pressure, size, or other variable for the equipment, than is shown on the graph. For instance, the cost of a stainless steel agitated tank is 1.7 times the cost of a mild steel tank (which is shown on the chart). These factors have also been listed when available, and again, sometimes as a consensus of different authors' estimates.

Equations:

$$\text{Cost} = 15,200 - 1,100W_c^{0.481} \text{ for } W_c > 250, < 10,000 \text{ lb.}$$

$$\text{Cost} = 76,200 - 0.422W_c^{1.2} \text{ for } W_c > 10,000, < 200,000 \text{ lb.}$$

W_c = weight of activated carbon, lb.

f the adjacent minor equipment and connec-
principally by Guthrie 1975 and Ulrich 1984)
on of the major equipment as well as all of
r unit. This is called the module factor, and
s as the range given by different authors and

purchase price × module factor

graphs, often a simple factor can be used to
ssure, size, or other variable for the equip-
ce, the cost of a stainless steel agitated tank
hich is shown on the chart). These factors
ain, sometimes as a consensus of different

ted Carbon
instruments and controls

000 lb.

r $W_p > 250, < 10,000$ lb.
e $> 10,000 < 200,000$ lb.

Agitators

Dual turbine blades: mild steel; 30-45 rpm; motor, gear reduction, shaft
Propeller: mild steel, single blade

Size exponent:	Installation factor:	Factors for:	Range	Average
Turbine: > 30 HP 0.68	Range	Turbine:		
4-30 HP 0.56	Turbine 1: 20-40	Single blade	0.75-0.85	0.82
< 4 HP 0.23	Propeller 1: 12-32	56-100 rpm	0.57-0.70	0.66
Propeller: 3-100 HP 0.51	Module factor 2.0	125-230 rpm	0.37-0.51	0.47
1-3 HP 0.42		316 stainless	1.23-1.87	1.47
		Propeller:		
		Stainless steel		1.19
		With seal (for closed tanks)		1.32

Size exponent 0.33
Module factor 2.5

Installation factor:

	Range	Avg.
Open tank	1.41 - 66	1.58
Low pressure	1.30 - 57	1.44
Autoclave	1.50 - 70	1.60

Material factors:

Stainless steel
1.2 - 2.2, avg. = 1.7
Glass lined
1.2 - 2.0, avg. = 1.6
*See Reactors

D-109

D-110

Compressors, High-Capacity and/or Pressure
1,000 psi; electric motor drive, gear reducer, steel

Size exponent 0.80

Equation:

(isothermal compression)

$$HP = 0.0044 P_2 Q_1 \ln P_1/P_2$$

P_1 = inlet pressure, psi

P_2 = outlet pressure, psi

Q_1 = inlet flow rate, cfm

Installation factor:

1.30-87; avg. 1.49

Module factor:

2.15-3.1; avg. 2.6

Factors:

Turbine drive 1.13

Gas engine 1.41

Pressure $\frac{P}{1000}^{0.18}$

Stainless steel 2.5

Nickle alloy 5.0

Conveyors
Mild steel construction

Size exponent:

Screw conveyor	0.78
Belt conveyor	0.76
Bucket elevator, roll	0.5
Pneumatic conveyor	0.37
Vibrating	1.0

Installation factor:

Range 1.40-2.15 avg. 1.72

Module factors:

Screw, pneumatic, roll	2.2
Belt, bucket, vibrating	2.4

Size factors:

Screw conveyor $\left(\frac{\text{diameter}}{9 \text{ in.}}\right)^{1.2}$

Belt conveyor $\left(\frac{\text{width}}{16 \text{ in.}}\right)^{0.8}$

Bucket elevator $\left(\frac{\text{bucket wd.} \times \text{ht.}}{5 \times 4 = 24 \text{ in.}^2}\right)^{0.27}$

Pneumatic conveyor $\left(\frac{\text{diameter}}{3 \text{ in.}}\right)^{0.55}$

Roll $\left(\frac{\text{width}}{20 \text{ in.}}\right)^{0.55}$; 4 in. spacing $\times 0.84$

Vibrating $\left(\frac{\text{width}}{36 \text{ in.}}\right)^{0.57}$

Drivers
Mild steel construction

Size exponents:

Rotary driver 0.45
Fluid bed 0.48
Spray driver 0.29

Installation factor:

* 25-96, avg. 1.64

Module factor:

Rotary 2.3
Fluid, spray 2.7

Factors:

Rotary to:

Roto-Louvre 1.25
Vacuum sheet 0.35
(sheet area)

Materials:

Nickle alloy 3.7
Brick-lined
stainless steel 2.2

Ducts
Wall thickness 1/8 in.

Size exponent: 1.08
Installation factor: 1.45

Equations: Sift

Mild steel $-2.22 + 1$
Stainless $-6.43 + 5$
Water cooled $179 + 6.78L$

D-114

Installation factor:
1.76-2.00; avg. 1.90

Module factors:

Electrostatic precipitators	1.3
Bag filters	1.2
Venturi scrubber	1.5
Cyclone, multichone	1.0

Material factors for
venturi, cyclone scrubbers:

High temperature with membrane brick lining	1.6
304 Stainless	1.8
316L Stainless	2.1
316L Stainless clad	1.9
Monel	3.0
Monel clad	2.7
Titanium	3.2

Fans
Mild steel motor, starter: 33 in. H₂O/P

Installation factor:
Range: 1.30-2.05; avg. 1.61

Module factor: 2.2

Factors:
Pressure: $\frac{1.2}{3.5}$ 0.3
Fiberglass: 1.8
Stainless steel: 2.5

Size exponents:

Rotary vacuum drum, leaf	0.39
Vacuum table, tilting pan, belt	0.5
Pressure leaf, plate & frame	0.61

Installation factor:

1.19 - 2.21, avg. 1.69

Module factor:

Rotary table, belt, tilting pan
1.4 - 2.8; avg. 2.4
Others 2.8

Factors:

Rotary drum: belt/screw or
string discharge 1.2
General/paper pulp 2.17-3.38
Mild steel/stainless steel 0.6
Vacuum table mild steel/SS 0.4
Vacuum filter auxiliaries (vac. pump
receivers, etc.). Often ~ 50% of total

Mills: Hammer, Jaw, Gyratory, Roll Crushers

Module factor:

Hammer 2.8
Others 2.1

Installation factor:

1.20-2.15, avg. 1.83

Mills: Ball, Rod, Pebble (Wet), Jet, Rubbish
 Reduction ratio 34 (i.e., ~ 1/2 in. - 65 mesh; 3/4 in. - 45 mesh)

Size exponent

Shredder 0.53
 Mills: installed 0.62
 purchased 0.70

Installation factor:

1.30 - 2.15; avg. 1.83

Modular factor:

1.8 - 2.8; avg. 2.3

Factors:

Ball, etc. mills
 Size reduction

$\left(\frac{\text{Reduction}}{34}\right)^{1.2}$

Dry/wet = 1.25

Pipe, Pipelines
Mild steel

Size exponent:
Pipelines 0.99

Factors
304 stainless schedule 10S
Sare pipe 2.05
Traced insulated 3.4
Fittings 18
Valves 34

Pumps, Centrifugal
Cast iron, horizontal, includes motor, coupling, base

Factors:

	Conven- tional	In-line	Axial flow	Mixed flow
Size exponent	1.30	1.27	0.79	0.79
Installation	1.5	1.25	1.58	1.32
Module factor	1.5	1.25	2.05	1.70
Cast steel	1.4	1.3		
316 stainless	2.0	1.6		
Cooper alloy	1.3			
Nickel alloy	3.6			
Titanium	5.7			

Pressure factor:

	to 150 psi	150- 300 psi
In-line	1.00	1.48
Conventional	1.0	1.62

Factors:

Conventional: APS/AVS = 1.6
In-line: vertical/horizontal = 0.89
Mixed, axial flow: vertical/horizontal =

Pumps, Miscellaneous
Mild steel construction

Factors

	Reciprocating	Turbine
Size exponent	0.59	0.47
Installation		1.38
Module factor	3.3	1.80
Cast iron	1.0	
Cast steel	1.8	
Stainless	2.4	
Nickel alloy	5.0	
0-150 psi	1.0	
150-500	1.32	
500-1000	1.53	

Chemical
Injection

0.52
1.58
2.83
1.0
1.25
1.95
1.0
1.27
1.79

General installation factor:
1.25 - 2.40, avg. 1.74

Other size exponents:

Diaphragm	43
Rotary	52
Gear	75
Sump	15

Factors:

Reciprocating: $\Delta P 11,000 - 5,000 / 10 - 1,000 = 3.8$
 Sump: $3600 / 1800 \text{ rpm} = 1.2$
 $1200 / 1800 \text{ in.} = 1.5$
 Chemical injection: Fixed variable speed = 1.67

Reactors*

304 stainless steel; jacketed; no agitation

Module factors:

Stainless	1.8
Glass lined	2.1
Mild steel	2.3

Installation factor:

1.40 - 2.10; avg. 1.70

Material factors:

316 stainless	1.2
Glass lined	.8
Lead lined	.7
Mild steel	.6

*See agitated tanks.

Screens, Vibrating
Mild steel, single deck

Size exponent: 0.75

Installation factor:

Factors:

1.45-2.27, avg. 1.85

Double deck 1.5
Stainless steel 1.25
Nickel alloy 1.8

Module factor: 2.8

Size exponent

Pug mill	0.15
Pellet mill	0.12
Pelletizing rolls	0.58

Installation factor: 2.05

Factors

	Pug mill extruder	Disk drum granulator	CIF
Module factor	2.2	2.8	1.2
Stainless steel	1.2	1.1	1.2
Nickel alloy	1.4	1.3	1.2

Screw/pug mill extruder	5.6
Disk/drum granulator	0.68

Tanks
Mild steel construction unless otherwise noted

	Size exponent	Module factor	Pressure factor
Small cone top	0.51	1.6	
Large cone top	0.51	1.9	
Horizontal, pressure	0.72	2.08	
Sphere	0.62	1.87	
Fiberglass	0.71		
Small storage	0.71		
			200 150 psi
			1.13 1.23
			50 75 100 125 150
			1.08 1.19 1.25 1.39 1.53

Factors:	
Rubber lined	1.5
Lead lined	1.6
Stainless	2.0
Coating roof	
large, field erected	1.3
Installation factor	
1.20 - 2.00, avg. 1.88	

Tanks, (Small)
304 stainless steel

Size exponent

Dished head, 50 psi	0.68
Flanged, dished head	0.48
Cone top, bottom, legs	0.57
Flat top, bottom	0.93

Factor

316/304 stainless steel 1.39

Thickeners, Clarifiers
 Rake mechanism, concrete tank, drive

Size exponent: 1.03
 Installation factor:
 1.63-2.61; avg. 2.12

Module factor: 3.0

Tank factor:
 Concrete/steel 0.7 for
 units under 40 ft diameter

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APPENDIX K
MASS BALANCE T-K SOLVER PROGRAM FOR THE LEC

LEC Moving Bed Mass Balance for the CO_2 -Sulver

Assumptions:

- 1) Steady-State Operation
- 2) Ideal Gas
- 3) Raoult's Law applies
- 4) 4.23 wt% water in sat. L/S

Inlet Gas Streams:

- (1) $VP_{in} = \exp(13.7008 - 9200.73 / (TW_{in} + 460.15)) \text{ atm}$
- (2) $LAT_{in} = 18140.4 \cdot (2.5441 - 0.001802 (TW_{in} + 460.15))^{0.38} \frac{\text{Btu}}{\text{lb}}$
- (3) $SAT_{in} = (VP_{in} / (1 - VP_{in})) \cdot 7.54 (TD_{in} - TW_{in}) / LAT_{in} \frac{\text{lbmol H}_2\text{O}}{\text{lbmol T}}$
- (4) $SO_{2in} = Q_{in} (actm) \left(\frac{C_{in} (\text{ppm vol})}{10^5 \text{ vol}} \right) \left(\frac{\text{lbmol BDA}}{1 + SAT_{in} (\text{lbmol H}_2\text{O})} \right) \left(\frac{\text{lbmol SO}_2}{0.7207 (TD_{in} - 460.15)} \right) \cdot (64.06 \frac{\text{lb}}{\text{lbmol SO}_2}) (60 \frac{\text{min}}{\text{hr}})$
- (5) $H_{2Oin} = Q_{in} (actm) \left(\frac{SAT_{in} (H_2O)}{1 + SAT_{in} (H_2O)} \right) \left(\frac{\text{lbmol H}_2\text{O}}{0.7207 (TD_{in} - 460.15)} \right) \cdot (18.02 \frac{\text{lb}}{\text{lbmol H}_2\text{O}}) (60 \frac{\text{min}}{\text{hr}})$

where VP_{in} = vapor pressure at inlet gas temperature (atm)
 LAT_{in} = inlet gas latent heat of vaporization (Btu/lb)
 SAT_{in} = absolute inlet gas humidity (lbmol H₂O / lbmol T)
 SO_{2in} = Inlet SO₂ mass rate (lb/hr)
 H_{2Oin} = Inlet H₂O mass rate (lb/hr)
 TW_{in} = Inlet gas-phase wet-bulb Temp. (°F)
 TD_{in} = Inlet gas-phase dry-bulb Temp. (°F)
 C_{in} = Inlet SO₂ concentration in the gas-phase (ppm vol)
 Q_{in} = Inlet gas flow rate (actm)

Outlet Gas Stream

- (6) $VP_{out} = \exp(13.7003 - 4200.43 / (TW_{out} + 460.15))$
- (7) $LAT_{out} = 18140.4(2.0991 - 0.001802(TW_{out} + 460.15))^{0.33}$
- (8) $SAT_{out} = (VP_{out} / (1 - VP_{out})) - 7.54(TD_{out} - TW_{out}) / LAT_{out}$
- (9) $H2O_{out} = Q_{out} (act) \left(\frac{SAT_{out} A^{1.3} H_2O}{1 + SAT_{out} act} \right) \frac{1 \text{ lbmol } H_2O}{(0.7302(TD_{out} + 460.15) A^{1.3})}$
 $\cdot (18.02 \text{ lb } H_2O / \text{lbmol}) (60 \text{ min/hr})$
- (10) $Q_{out} = Q_{in} (act) \left(\frac{TD_{out} + 460.15}{TD_{in} + 460.15} \right) - (SO_{2,in} - SO_{2,out}) \left(\frac{\text{lb}}{\text{hr}} \right) \left(\frac{1 \text{ lbmol } SO_2}{64.06 \text{ lb}} \right)$
 $\left(\frac{0.7302(TD_{out} + 460.15) A^{1.3}}{1 \text{ lbmol } SO_2} \right) \left(\frac{1 \text{ hr}}{60 \text{ min}} \right) + \frac{1}{2} \text{ lbmol} \frac{0.7302(TD_{out} + 460.15) A^{1.3}}{60}$
 $- (H_{2O,in} - H_{2O,out}) \left(\frac{\text{lb}}{\text{hr}} \right) \left(\frac{1 \text{ lbmol}}{18.02 \text{ lb}} \right) \frac{0.7302(TD_{out} + 460.15) A^{1.3}}{1 \text{ lbmol } H_2O} \left(\frac{1 \text{ hr}}{60 \text{ min}} \right)$

where VP_{out} = Water vapor pressure of outlet gas (atm)
 LAT_{out} = Latent heat of vaporization for outlet gas (Btu/lbm)
 SAT_{out} = absolute humidity of outlet gas (lbmol H_2O / lbmol B)
 $H2O_{out}$ = water content of outlet gas (lb/hr)
 Q_{out} = Outlet flow rate of gas-phase (act)

Reactor Section

P. 3

X_i = mass rate of $CaCO_3$ in stream: (16/hr)
 Z_i = mass rate of $CaSO_4$ in stream: (16/hr)
 I_i = mass rate of inlets in stream: (16/hr)

The amount of $CaCO_3$ needed to yield a specified percent removal efficiency is:

$$(11) \quad X_{rxn} = \left(\frac{\%RE}{100} \right) \cdot SO_2 \text{ in } \left(\frac{16}{hr} \right) \left(\frac{150.06 \text{ lb}}{150.06 \text{ lb}} \right) \left(\frac{150.06 \text{ lb}}{100.84 \text{ lb}} \right)$$

where %RE = percent SO_2 removal efficiency

The amount of $CaCO_3$ required by the reactor from an assumed "reaction rate" is:

$$(12) \quad X_r = X_{rxn} \left(\frac{16}{150} \right) \left(\frac{\%RXN}{100} \right)$$

where %RXN = percent $CaCO_3$ reaction efficiency

The amount of $CaCO_3$ produced from the $CaCO_3$ reaction is:

$$(13) \quad Z_{rxn} = X_{rxn} \left(\frac{16}{hr} \right) \left(\frac{150.06 \text{ lb}}{100.84 \text{ lb}} \right) \left(\frac{150.06 \text{ lb}}{150.06 \text{ lb}} \right) \left(\frac{150.06 \text{ lb}}{136.14 \text{ lb}} \right)$$

The amount of water for a saturated L/S bed

$$(14) \quad H_2O_r = (X_r + I_r) 0.0423 \quad (16/hr)$$

Stream 1 (Reactor Inlet):

$$(15) \quad X_1 = X_r$$

$$(16) \quad Z_1 = (Z_{rxn} / GF) - Z_{rxn}^*$$

$$(17) \quad I_1 = X_1 \left(\frac{lb}{hr} \right) \left[\frac{(1 - WT\% / 100) (lb \text{ inerts})}{WT\% / 100 (lb CaCO_3)} \right] \\ + Z_1 \left(\frac{lb}{hr} \right) \left(\frac{1 \text{ mol } CaSO_4}{136.14 \text{ lb}} \right) \left(\frac{1 \text{ mol } CaCO_3}{100.09 \text{ lb}} \right) \left[\frac{(1 - WT\% / 100) (lb \text{ inerts})}{WT\% / 100 (lb CaCO_3)} \right]$$

Since the amount of inerts present is a constant,
mass ratio = $(1 - WT\% / 100)$ where $WT\%$ is defined
as the percent $CaCO_3$ (lb) per total lb of $CaCO_3$.
For Maxville L/S $WT\% = 34$.

The amount of water entering the reactor
is given by the overhead trays of the gas stripper:

$$(18) \quad H_2O_1 = H_2O_4 + H_2O_5 \quad (lb/hr)$$

The amount of water needed to keep the
entire L/S bed saturated is

$$(19) \quad SPRAY = H_2O_f + H_2O_{ex} - H_2O_1 - H_2O_2 - H_2O_3 - H_2O_4 - H_2O_5$$

where H_2O_{ex} is the amount of liquid H_2O
exiting the reactor via a drain and is assumed
to be 5% of water in the reactor.

$$(20) \quad H_2O_{ex} = 0.05 \cdot H_2O_2$$

* See the Mathematical Manipulations Section

Stream 2 (Reactor Outlet)

The amount of CaCO_3 exiting the reactor is the difference of the CaCO_3 entering and the CaCO_3 that reacted:

$$(21) \quad X_2 = X_1 - X_{\text{rxn}}$$

$$(22) \quad Z_2 = Z_{\text{rxn}} / G_F^*$$

The amount of inert exiting the reactor is equal to the amount entering:

$$(23) \quad I_2 = I_1$$

From a point-of-view around the grinder:

$$(24) \quad H_2O_2 = H_2O_3 - -204 - H_2O_{\text{gr}}^*$$

Stream 3 (Grinder Outlet)

$$(25) \quad X_3 = (1 - \%U_r / 100) \cdot X_{\text{rxn}} / (100 - 200)^*$$

$$(26) \quad Z_3 = Z_{\text{rxn}}$$

The inerts calculation is the same as Eqn. 7

$$(27) \quad I_3 = X_3 (1 - \text{WT}\% / 100) / (\text{WT}\% / 100) -$$

$$Z_3 \cdot 100.0 \text{ g} \cdot (1 - \text{WT}\% / 100) / (136.14 \cdot \text{WT}\% / 100)$$

The L/S is assumed to be saturated after the reactor section:

$$(28) \quad H_2O_3 = (X_3 + I_3 + Z_3) \cdot 0.0423$$

$$\text{g c. H}_2\text{O} \quad \text{L} \quad \text{D-136} \quad \text{L} \quad \text{L}$$

Stream 4 (Recycle Stream)

From a grinder point balance:

$$X_4 = X_2 - X_3$$

From a point 1 balance knowing that fresh L/S has no CaSO_4 :

$$Z_4 = Z_1$$

Using the same old inerts calculation:

$$I_4 = X_4 \left(\frac{1 - \text{WT}\% / 100}{\text{WT}\% / 100} \right) + Z_4 \left(\frac{100.094}{136.14} \times \frac{1 - \text{WT}\% / 100}{\text{WT}\% / 100} \right)$$

The recycle stream is assumed saturated

$$\text{H}_2\text{O}_4 = (X_4 - Z_4 - I_4) \cdot 0.0423$$

Stream 5 (Make-up L/S)

$$X_5 = X_{\text{res}} / (\% \text{Ca} / 100)^*$$

Since fresh L/S contains no CaSO_4 :

$$Z_5 = 0$$

$$I_5 = X_5 (1 - \text{WT}\% / 100) / (\text{WT}\% / 100)$$

The amount of CaSO_4 in the L/S from storage varies depending on "operational conditions."

$$\text{H}_2\text{O}_5 = (X_5 + I_5) / W_5$$

where W_5 = water in storage L/S $\left(\frac{16.23}{136.14} \right)$

Pre-TK Mathematical Manipulations:

CaCO_3 in Streams 3 and 5:

Since percent L/S utilization, $\% U_r$, is defined as total CaCO_3 reacted per total CaCO_3 introduced into the system; i.e. the CaCO_3 in the waste stream, X_3 , is:

$$X_3 = (1 - \%U_r/100) X_5 \quad (\text{PMM-1})$$

The make-up CaCO_3 , X_5 , must be equal to the sum of the CaCO_3 reacted, X_{rxn} , and CaCO_3 lost in the waste stream.

$$X_5 = X_{rxn} + X_3 \quad (\text{PMM-2})$$

Combining PMM-1 and PMM-2 and simplifying yields:

$$X_3 = (1 - \%U_r/100) \cdot X_{rxn} / (\%U_r/100) \quad (\text{PMM-3})$$

$$X_5 = X_{rxn} / (\%U_r/100) \quad (\text{PMM-4})$$

CaSO₄ in System:

$$\left. \begin{aligned} Z_1 &= Z_4 + Z_5 \\ Z_2 &= Z_1 + Z_{rxn} \\ Z_3 &= GF \cdot Z_2 \\ Z_4 &= Z_2 - Z_3 \\ Z_5 &= 0 \end{aligned} \right\} \begin{array}{l} 5 \text{ equations} \\ 5 \text{ unknowns} \end{array} \quad \begin{array}{l} (PM-5) \\ (PM-6) \\ (PM-7) \\ (PM-8) \\ (PM-9) \end{array}$$

where GF = Grinding Factor is the mass ratio of CaSO₄ in fines to total CaSO₄ exiting the reactor; i.e. an ideal grinder system would be able to remove all CaSO₄ from the L/S, i.e. $GF = 1$ (see 1)

Combining PM-5 - PM-9 yields:

$$Z_4 = Z_1 \quad (PM-10)$$

Inserting PM-10 into PM-3 + the introducing the result for Z_1 into PM-6 yields:

$$Z_3 = Z_{rxn} \quad (PM-11)$$

Solving PM-7 using PM-11 gives:

$$Z_2 = Z_{rxn} / GF \quad (PM-12)$$

Finally, substituting PM-12 into PM-3:

$$Z_1 = Z_4 = \frac{Z_{rxn}}{GF} = Z_{rxn} \quad (PM-13)$$

S Rule

- ----

"INLET GAS STREAMS

```
* VPIn = exp(13.7008-9200.43/(TWin+460.15))
* LATIn = 18140.4*((2.0991-0.001802*(TWin+460.15))^0.38)
* SATIn = (VPIn/(1-VPIn))-7.54*(TDIn-TWin)/LATIn
* SO2In=Qin*Cin*64.06*60/((1+SATIn)*1E6*0.7302*(TDIn+460.15))
* H2OIn=Qin*(SATIn/(1+SATIn))*18.02*60/(0.7302*(TDIn+460.15))
* inwa=(SATIn/(1+SATIn))*100
```

"OUTLET GAS STREAMS

```
* VPout = exp(13.7008-9200.43/(TWout+460.15))
* LATout = 18140.4*((2.0991-0.001802*(TWout+460.15))^0.38)
* SATout = (VPout/(1-VPout))-7.54*(TDout-TWout)/LATout
* H2Oout=Qout*(SATout/(1+SATout))*18.02*60/(0.7302*(TDout+460.15))
* SO2out=SO2in*(1-%RE/100)
* Qout=Qin*((TDout+460.15)/(TDIn+460.15))-0.5*((SO2in-SO2out)*0.7302*(TDout+460.15))
* outwa=(SATout/(1+SATout))*100
```

"REACTOR SECTION

```
* Xrxn=(%RE/100)*SO2in*(100.089/64.06)
* Irxn=Xrxn*(100/WT%)
* Zrxn=Xrxn*136.142/100.089
* H2Orxn=(Xrxn+Irxn)*0.0423
```

"L/S REQUIRED BY REACTOR

```
* Xr=Xrxn/(%RXN/100)
* Ir=Xr*(1-WT%/100)/(WT%/100)
* H2Or=(Xr+Ir)*0.0423
```

"SOLIDS STREAM 1

```
* X1=Xr
* Z1=(Zrxn/GF)-Zrxn
* I1=X1*(1-WT%/100)/(WT%/100)+Z1*100.09*(1-WT%/100)/(136.14*WT%/100)
* H2O1=H2O4+H2O5
* SPRAY=H2Or+H2Oexit-H2O5-H2Oin+H2Oout-H2O4
```

"SOLIDS STREAM 2

```
* X2=X1-Xrxn
* Z2=Zrxn/GF
* I2=I1
* H2O2=H2O3+H2O4+H2Oexit
* H2O2=H2O3+H2O4
* H2Oexit=0.05*H2O2
```

"SOLIDS STREAM 3

```
* X3=(1-%Ut/100)*Xrxn/(%Ut/100)
* Z3=Zrxn
* I3=X3*(1-WT%/100)/(WT%/100)+Z3*100.09*(1-WT%/100)/(136.14*WT%/100)
* H2O3=0.0423*(X3+Z3+I3)
```

"SOLIDS STREAM 4

```
* X4=X2-X3
* Z4=Z1
* I4=X4*(1-WT%/100)/(WT%/100)+Z4*100.09*(1-WT%/100)/(136.14*WT%/100)
* H2O4=(X4+Z4+I4)*0.0423
```

"SOLIDS STREAM 5

$I5 = X_{rxn} / (\%Ut / 100)$
 $I5 = 0$
 $I5 = X5 * (1 - WT\% / 100) / (WT\% / 100)$
 $H2O5 = (X5 + I5) * WS$

STREAM TOTALS (DRY BASIS)

$T1D = X1 + Z1 + I1$
 $T2D = X2 + Z2 + I2$
 $T3D = X3 + Z3 + I3$
 $T4D = X4 + Z4 + I4$
 $T5D = X5 + Z5 + I5$

CHECK POINTS (DRY BASIS) IS OK IFF THEY EQUAL ZERO

$CP1D = T4D + T5D - T1D$
 $CP2D = T2D - T3D - T4D$

STREAM TOTALS (WET BASIS)

$T1W = X1 + Z1 + I1 + H2O1$
 $T2W = X2 + Z2 + I2 + H2O2$
 $T3W = X3 + Z3 + I3 + H2O3$
 $T4W = X4 + Z4 + I4 + H2O4$
 $T5W = X5 + Z5 + I5 + H2O5$

CHECK POINTS (WET BASIS) IS OK IFF THEY EQUAL ZERO

$CP1W = T4W + T5W - T1W$
 $CP2W = T2W - T3W - T4W$

$WATER = SPRAY * (7.48 / 62.32) / 60$

DATE

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END