# WET ELECTROSTATIC PRECIPITATOR SYSTEM STUDY

# RESEARCH REPORTING SERIES

# WET

# ELECTROSTATIC PRECIPITATOR

# SYSTEM STUDY

by

John P. Gooch and Alan H. Dean

Southern Research Institute
2000 Ninth Avenue South
Birmingham, Alabama 35205

# ABSTRACT

This report describes a study of wet electrostatic precipitators
(WESP) through laboratory experiments, a field test of a full-
scale unit, interviews with manufacturers and users, and a
literature survey. Experiments were conducted with the objec-
tive of determining collection efficiency as a function of
particle size at varying current densities and specific collec-
tion areas. The results obtained were in reasonable agreement
with those predicted by simulating the experimental conditions
with a mathematical model. The feasibility of collecting $SO_2$
and particulate in a WESP was examined. As would be expected,
calculation of ion transport rates indicate that only an in-
significant amount of $SO_2$ can be removed by selective ioniza-
tion of $SO_2$ molecules. Mass transfer calculations indicated
that irrigated electrode surfaces would not provide sufficient
interfacial area for desired $SO_2$ removal levels. Based on
conventional geometry and stainless steel construction, a
WESP-scrubber device would be more costly than a scrubber-only
system because of high WESP capital costs. The effectiveness
of a WESP as a mist eliminator following a scrubber was shown
by calculation to be a function of the particle size distribu-
tion and the concentration of the mist to be collected. The
field test and literature survey indicated that WESPs have
been used effectively to control fine particle emissions in the
aluminum, iron and steel, chemical, and petrochemical industries.

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# SECTION I
## SUMMARY AND CONCLUSIONS

A flexible laboratory scale electrostatic precipitator was
designed and constructed to (1) obtain performance data on
fine particle collection under idealized conditions and (2)
to evaluate the feasibility of incorporating a wet precipitator
scrubber-device in a gas cleaning system for coal-fired boilers.
Experiments were conducted with the objective of determining
collection efficiency as a function of particle size at varying
current densities and specific collection areas using dioctyl-
phthalate aerosol. The results obtained were in reasonable
agreement with those predicted by simulating the experimental
conditions with a mathematical model of electrostatic precipi-
tation.

Calculations were performed to determine the maximum amount of
$SO_2$ which could be transported to the collection electrode by
selective ionization of $SO_2$ molecules. As would be expected,
the results indicated that an insignificant fraction of the
$SO_2$ can be removed by this mechanism. Mass transfer calculations
were also performed to estimate the $SO_2$ which could be absorbed
by a basic solution irrigating the collection electrodes. It
was found that the interfacial area provided by the collecting
electrodes was insufficient to obtain significant $SO_2$ removal levels.
Experiments conducted in the laboratory test unit with NaOH
solution irrigating the collecting electrodes indicated that
no significant increase in $SO_2$ removal rates occurred as a re-
sult of energizing the precipitator power supplies. Simultaneous
collection of $SO_2$ and DOP aerosol in the laboratory test unit
indicated a decrease in $SO_2$ removal, presumably caused by a
decrease in electrode wettability.

Cost estimates were made for a full-scale wet ESP-scrubber device
for collection of $SO_2$ and particulate in a power plant flue gas
cleaning system. The estimates indicated that, although the
WESP system would use less energy, a scrubber-only system is
more economical to operate because of the high capital costs
(using stainless steel construction and conventional electrode
geometries) of a WESP. Electrode scaling with the WESP is a
potentially serious problem which would require pilot plant
trials for an adequate evaluation.

The effectiveness of a wet electrostatic precipitator as a mist
eliminator following a scrubber was shown by calculation to be
a function of the particle size distribution and the concentration
of the mist to be collected. High droplet concentrations, such
as those existing at the outlet of a spray scrubber, may limit
the performance of a WESP through space charge suppression of
the corona current.

A field test of a full scale wet electrostatic precipitator was conducted, and an analysis of the results indicated that the collection efficiency as a function of particle size relationship is in fair agreement with the predictions of the mathematical model over the particle diameter range 0.2 μm to 1.3 μm. For larger particles, the measured collection efficiencies were significantly lower than the theoretical predictions, possibly because of liquor carryover from the electrode irrigation system. The WESP was collecting fume from an aluminum pot line. Overall mass collection efficiencies on the fume, in which about 65% of the mass consisted of sub-micron particles, ranged from 95.0 to 98.0% at a specific collecting area of about 62 $m^2/(m^3/sec)$. Corrosion and deposit formation in the liquor supply system have occurred at this installation.

An evaluation of current WESP technology was conducted by a literature survey and by contacts with manufacturers and users. Both plate and pipe collecting electrodes have been effectively utilized in collecting fumes and mists from metallurgical and chemical processes. The largest recent application of WESP has been in the collection of fume from aluminum pot lines. Duct-type collecting electrodes are generally used for the application in which relatively large gas volume flows are to be cleaned. The advantages of plate electrodes as opposed to pipe electrodes are less expensive construction and the flexibility of allowing variations in electrical sectionalization. Pipe electrodes can be more uniformly irrigated than plates, and gas by-passage (sneakage) is not a factor. Wire-pipe WESP's usually consist of only one electrical section in the direction of gas flow.

# SECTION II
## INTRODUCTION


This report describes a study of wet electrostatic precipitator technology which was conducted with the following objectives: (1) develop methods for calculating performance of WESP's, (2) evaluate the potential of a wet electrostatic precipitator as a device for collecting particulate and sulfur oxides, or mists exiting a scrubber, in a gas cleaning system for a utility power station, and (3) evaluate the status of wet precipitator technology. Wet electrostatic precipitators, for the purposes of this study, are defined as conventional electrostatic precipitator geometries, consisting of either wire-pipe or wire duct design, in which either a liquid or solid phase particulate is collected with electrode irrigation, or in which a liquid phase aerosol is collected without electrode irrigation.

An electrostatic precipitator collecting a liquid aerosol under laboratory conditions offers an opportunity to study performance in the absence of unmodelled processes such as particle reentrainment and back corona. Therefore, a laboratory-scale wet ESP can be used to evaluate mathematical procedures for modelling particle collection. The following sections describe experiments performed with a laboratory-scale WESP, and compare the experimental results with those obtained from a mathematical model developed under other EPA contracts. The model is then used to evaluate potential WESP applications in utility power plant gas cleaning systems.

The status of existing wet ESP technology was evaluated by conducting a performance test on a recently installed WESP in the aluminum industry, and by conducting a literature survey. Information was also obtained from WESP manufactures and users.

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# SECTION III
# LABORATORY STUDIES


## DESIGN OF BENCH-SCALE TEST UNIT

A flexible test unit capable of treating about 0.0737 m$^3$/sec of synthetic flue gas (156 ft$^3$/min)* was designed with the following objectives in mind.

> To obtain fundamental performance data for
> particulate collection, especially in the
> diameter range of 2.0 µm and smaller.

> To evaluate the feasibility of incorporating
> a wet precipitator scrubber device in a gas
> cleaning system for coal-fired boilers.

The general arrangement of the test facility and a side view with dimensions are given in Figure 1. Plate-to-plate spacing is 12.7 cm (5 in.) and the collecting electrode area is 2.32 m$^2$ (25 ft$^2$). A gas velocity of 1.52 m/sec (5 ft/sec) gives a flow rate of 0.0737 m$^3$/sec which results in a Reynolds number of about 16000 (well within the turbulent flow region) and an A/V ratio of 31.5 m$^2$/ (m$^3$/sec) (160 ft$^2$/1000 cfm). The test unit can also be assembled with plate-to-plate spacings of 18.75 cm and 25.4 cm.

The test unit has sufficient space available to accommodate an assembly consisting of alternating scrubbing and collecting sections, with the spray sections comprising one-half the length of the collecting sections. Under this contract, the spray sections were not utilized. Figure 2 is a detailed cross-section of the unit at a corona wire in a precipitation section. The numbers in circles are call-out numbers which refer to details in detail drawings that are not shown.

The high-voltage insulators for the test unit are designed to isolate the discharge electrodes in the presence of a wet environment. This was a difficult design problem for a test unit designed to obtain fundamental performance data, since bleed air to dry the insulators and large open spaces around the insulators are very undesirable from the standpoint of interpreting experimental data. Originally, the plan was to isolate Teflon feed-through insulators at the top of the assembly in a plexiglas conduit purged with heated air. However, water drops from the spray nozzles could deposit on the plexiglas, and the conduit

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*Environmental Protection Agency policy is to express all measurements in agency documents in metric units. When implementing this practice will result in undue costs or lack of clarity, conversion factors are provided for the non-metric units used in the report. Conversion factors are given in Appendix A.

4 SPACES @ 15"+
4 SPACES @ 30" = 180" = 15'-0"

30"
PRECIPITATOR SECTION

15"

48

7½

47¾

44

7'-8¼

7'-6½
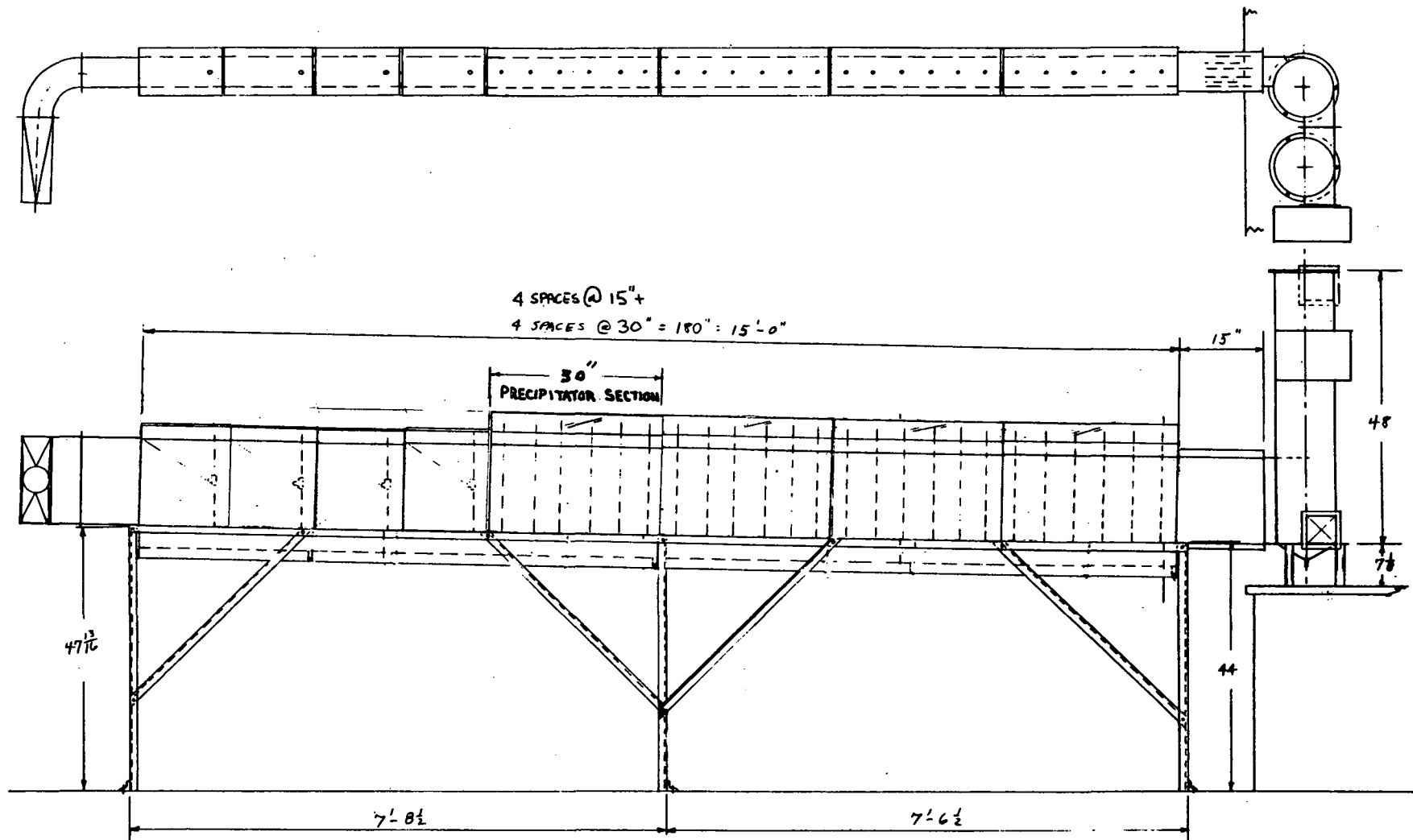
Figure 1.  Plan and Side Views of Pilot ESP Test Facility

could not be maintained at a temperature level sufficient to insure quick evaporation. As a result of the above potential hazard, an alternative concept of using a heated epoxy insulator was designed. The epoxy is a formulation which has a high thermal conductivity and a low electrical conductivity. The heater embedded in the epoxy was designed to maintain the temperature of the insulator at about 250°F. Figure 2 shows a cross-section of the cast epoxy insulator, which is used to jacket the Teflon feed-through insulator surrounding the corona wire. This design was only partially successful because of problems with air bubbles in the casting and cracks forming in the epoxy during operation of the internal heaters. The test unit can at present be operated for short periods of time with electrode irrigation with this design, but long-term operation is not possible due to spray droplet accumulation on the insulator surfaces. A solution to the problem would require a design that would enable higher insulator surface temperature to be maintained.

Although Figure 2 shows the plates irrigated by sprays and weirs, only the sprays were used due to non-uniform flow along the length of the weirs. Experiments with the collecting electrode spray assembly indicated the expected difficulty in maintaining uniform sheet flow on the electrode surfaces. Polishing the plate surface with emery cloth was found to be beneficial, but repolishing was necessary immediately prior to spraying the plates with water, apparently because of a change in the wettability of the surfaces upon exposure to the atmosphere. In an attempt to eliminate the time variation of the wettability, a sample plate was sandblasted with medium grit sand. This procedure resulted in much improved plate wetting, and the wettability did not decrease upon exposure to the atmosphere. All plates were therefore sandblasted prior to their installation in the test unit.

Photographs of the completed installation are shown in Figures 3 and 4. Three electrical sections are employed, the first two of which each consists of 0.581 m² (6.25 ft²); the third power supply feeds 1.162 m² (12.50 ft²). Figure 4 shows a side view of the test unit, including the heater assembly, spray towers, a mixing chamber for studying condensation growth (constructed as part of another project), the water piping and drainage system, the power supplies, inlet sampling ports, and the electrified sections. Figure 4 shows a front close-up view of the gas-fired burner assembly, including a hopper for feeding solid particulate, and a steam venturi-ejector device which can be used for redistribution of particulate. The burner assembly has been equipped with a commercial hot water controller with proportional band control on a pressure reducer in the gas supply line. The pressure is controlled by a temperature bulb in the warm water outlet of the scrubber. The controller also has a thermostatically controlled pilot and main gas valve to shut down in case of flame failure.

Voltage-current curves were obtained with and without water
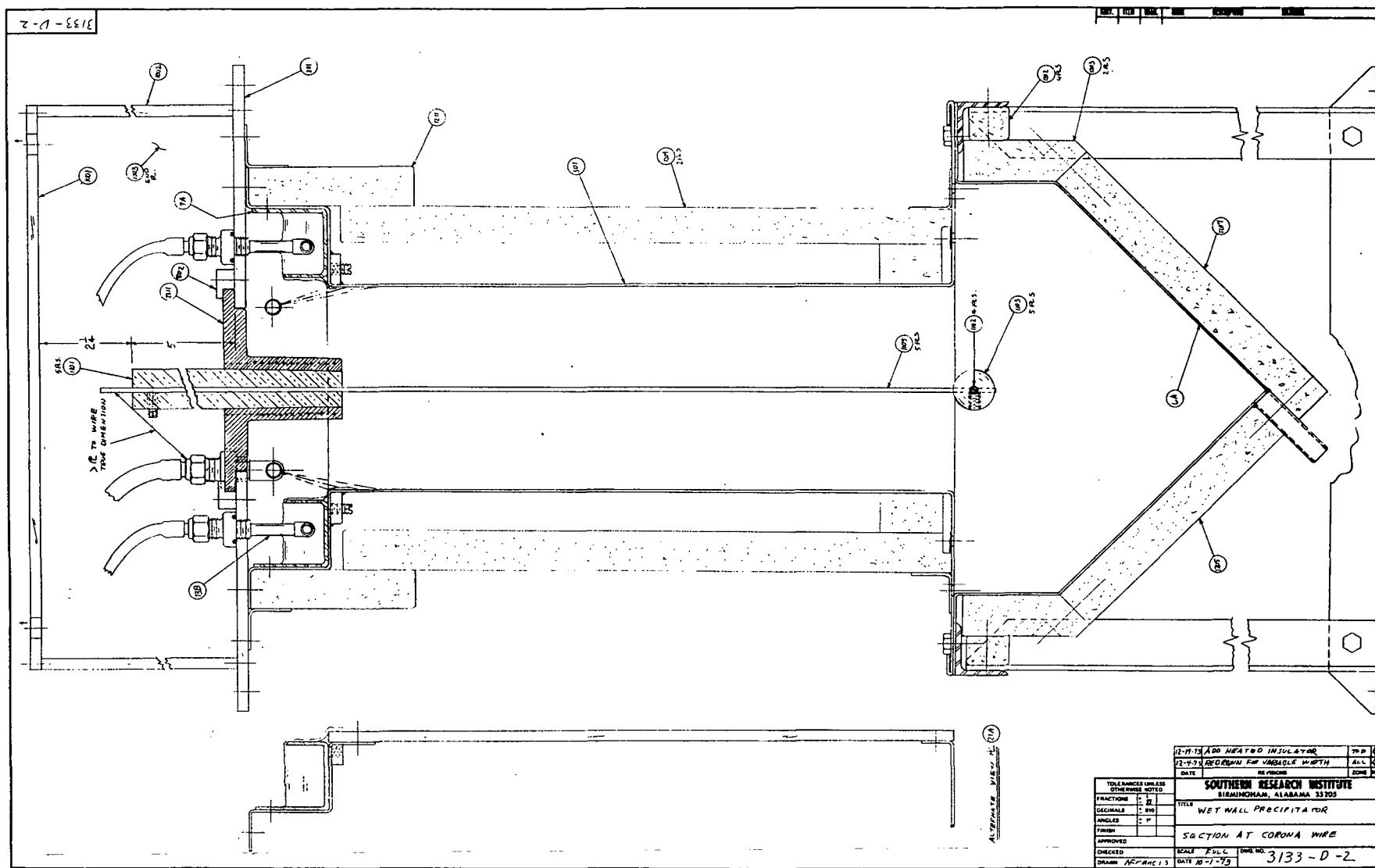
Figure 2.  Wet Wall Precipitator
Section at Corona Wire

Figure 3. Completed wet ESP test assembly

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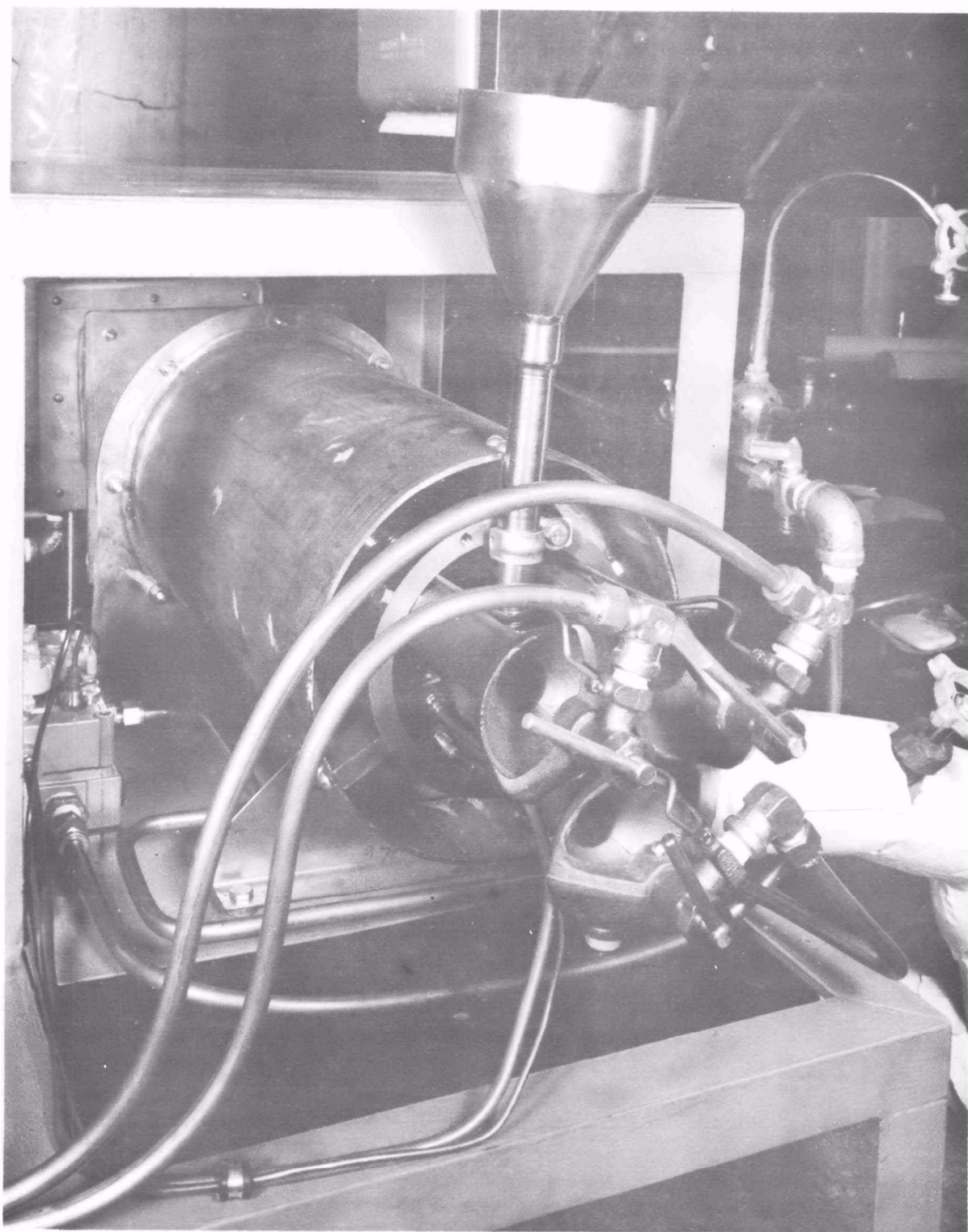
Figure 4.   Gas Burner Assembly

irrigating the collecting electrodes, and with and without the
insulator heaters energized. Normal current-voltage relationships
were obtained with and without electrode irrigation when the
insulator heaters were connected to the AC power source through
an isolation transformer. Without the transformer, sparking
occurred to the epoxy insulator surfaces at relatively low
voltage. A typical voltage-current curve is shown in Figure 5.
Electrode irrigation causes the V-I curve to shift to the left.
Normally, increasing the water vapor content of an air-water
vapor mixture would be expected to shift the V-I curve to the
right. The shift toward higher current for a given applied volt-
age indicated for wet operation in Figure 5 may be caused by
leakage paths along the surface of the insulators due to moisture
accumulation, or an effective decrease in the wire to plate
spacing due to the presence of the electrode irrigation stream in
the upper portion of the inter-electrode region.

## PARTICULATE COLLECTION EXPERIMENTS

### Procedure

A series of particulate collection experiments was conducted for
the purpose of determining collection efficiency as a function
of particle size under idealized conditions. A liquid-phase
aerosol (dioctyl-phthalate or DOP) was chosen for the experiments
to eliminate the possibility of reentrainment. The particulate
generator consisted of a single atomizer which produced a stable
polydisperse aerosol. Electrode irrigation and humidification
were not used, and the collected oil droplets drained from the
collection electrodes by gravity. The carrier gas was ambient
air.

Particle size and concentration measurements at the inlet and
outlet of the test unit can be made with either an optical par-
ticle counter or with inertial impactors. In order to obtain
reliable particle size information with an optical particle
counter, however, it is necessary to obtain a calibration curve
for the instrument with the aerosol under investigation. An
attempt was made to obtain such a curve using a vibrating or-
ifice particle generator. However, due to problems with the
particle generator, we were unsuccessful in obtaining a useful
calibration curve over the particle size range of interest.
Therefore, it was decided to conduct the majority of the ex-
periments using impactors to determine particle size distributions
and fractional efficiencies. A small number of experiments were
conducted using a Climet-Ortec particle size analyzer and a
calibration curve obtained with polystyrene latex beads. These
measurements are compared with those made with the impactors
in a subsequent portion of this section.
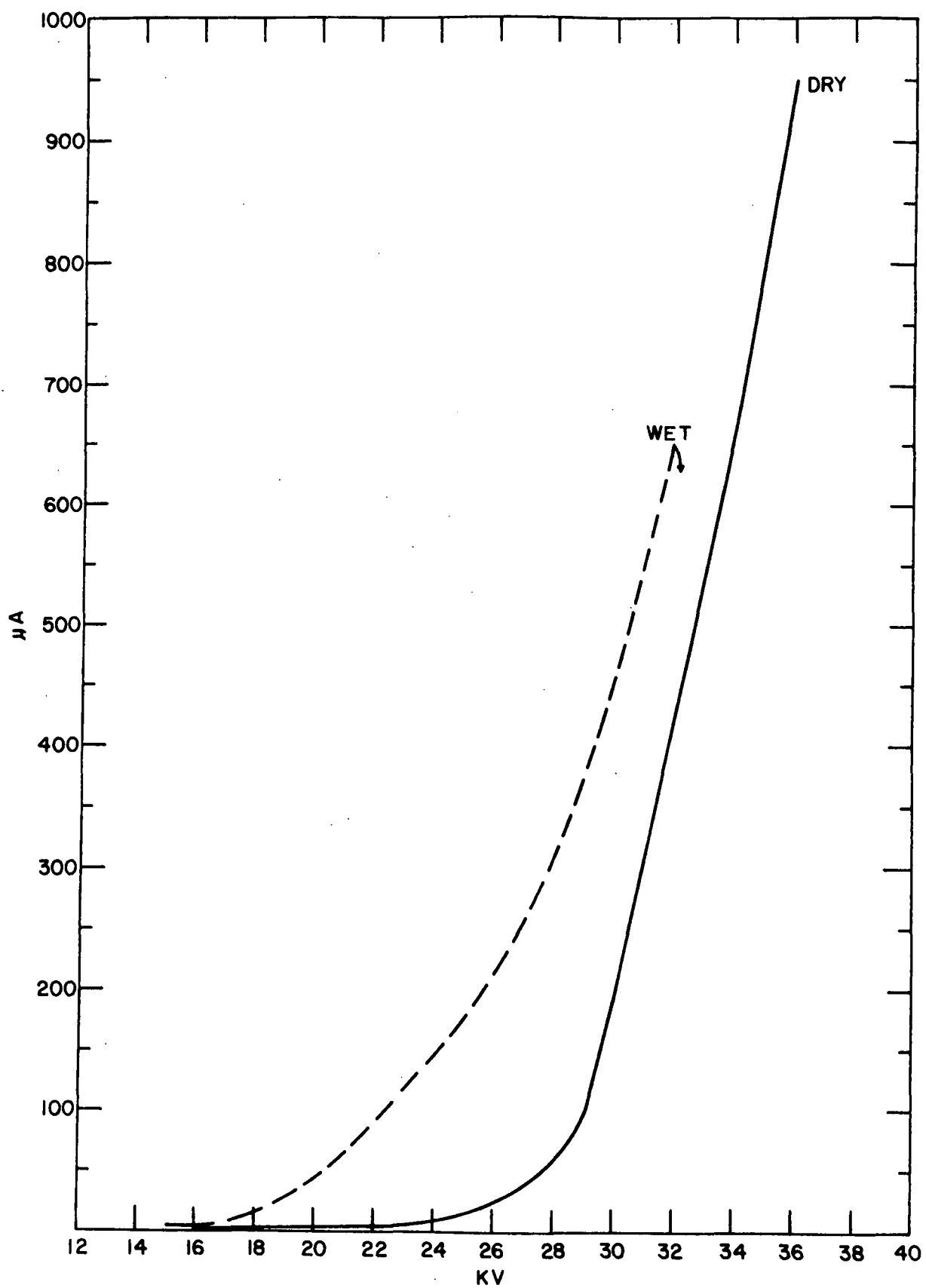
Figure 5.  Voltage-Current Relationship for Section 2
of Pilot Scale Wet ESP

A five-stage Brink impactor was used at the precipitator inlet and outlet. Aluminum foil substrates were employed to facilitate weighing of the small masses collected on the individual stages, and the mass determinations were made with a Cahn Electrobalance. For the experiments conducted at the lower gas velocity, a sampling time of six hours was required at the precipitator outlet to obtain weighable quantities on the impactor substrates. Six and four point traverses were conducted with the impactor at the inlet and outlet, respectively, in order to insure that a representative size distribution was obtained at each location.

Since the particulate from an electrostatic precipitator will have an electrical charge, it was necessary to employ a charge neutralizer at the inlet of the sampling line for all measurements made at the precipitator outlet. Measurements with the optical particle counter indicated that, if metal substrates were used in the impactor, the charge acquired by the particulate substantially influenced the size distribution obtained with the impactor. Since the impactor was calibrated with an uncharged aerosol, the use of the charge neutralizer minimized sizing errors due to electrostatic attraction.

Table 1 gives a typical set of inlet and outlet data obtained from duplicate experiments at the indicated conditions. Good reproducibility was obtained for all stages at the inlet, and for stages 2 through 5 at the outlet. Isokinetic sampling was not employed, since the effect of anisokinetic conditions on particles with diameters less than 2.0 µm is expected to be negligible. The effect of changing the gas flow rate through the impactor on the indicated size distribution was checked by reducing the flow rate from 56.6 $cm^3$/sec to 28.3 $cm^3$/sec. The higher flow rate, which was used for all of the fractional efficiency determinations, resulted in a gas velocity of 0.79 m/sec at the sample point. Figure 6 shows the inlet size distributions from Table 1 and the distribution obtained at the same location with the reduced impactor flow rate. These data indicate that the same size distribution is obtained at either flow rate. The higher flow rate is more desirable because of the smaller cut points and the increased amount of gas that may be sampled per unit time.

Results

Fractional efficiency measurements were performed with the Brink impactor and DOP aerosol at current densities of 26.9, 53.8, and 107.5 $nA/cm^2$ at a gas velocity of 1.37 m/sec, and at current densities of 26.9 and 53.8 $nA/cm^2$ at a gas velocity of 0.72 m/sec. The results obtained from these experiments, expressed as collection efficiency as a function of particle diameter, are presented in Figures 7, 8, and 9. Computed collection efficiencies obtained with a mathematical model are also shown.[1] In view of the difficulties in making these types of measurements, the agreement between measured and calculated efficiencies shown in Figure 7 is considered good. Figures 8 and 9 suggest that

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TABLE I

TYPICAL INLET AND OUTLET PARTICLE SIZE DATA FOR WET ESP EXPERIMENTS
(DOP Sprayer, Gas Velocity = 1.37 m/sec, Current Density = 55 na/cm$^2$)

Mass Loading, mg/am$^3$

| Test No. Stage | Lower Diameter Limit, μm | Inlet 2ai | 2bi | 2ci | 2di | Average | Outlet 2ao | 2bo | Average | Geometric Mean Diameter, μm | Collection Efficiency, % |
|---|---|---|---|---|---|---|---|---|---|---|---|
| 1 | 3.04 | 26.341 | 28.956 | 24.596 | 24.088 | 25.995 | 0.0414 | ~0 | 0.207 | - | - |
| 2 | 1.80 | 18.659 | 13.526 | 14.128 | 15.226 | 15.385 | 0.1494 | 0.1103 | 0.1299 | 2.34 | 99.16 |
| 3 | 1.24 | 7.249 | 6.285 | 7.238 | 8.226 | 7.250 | 0.2575 | 0.2115 | 0.2345 | 1.49 | 96.77 |
| 4 | 0.65 | 7.448 | 6.984 | 6.743 | 7.419 | 7.148 | 0.5724 | 0.4782 | 0.5253 | 0.90 | 92.65 |
| 5 | 0.46 | 2.954 | 1.775 | 2.514 | 2.654 | 2.474 | 0.3862 | 0.3770 | 0.3816 | 0.55 | 84.58 |
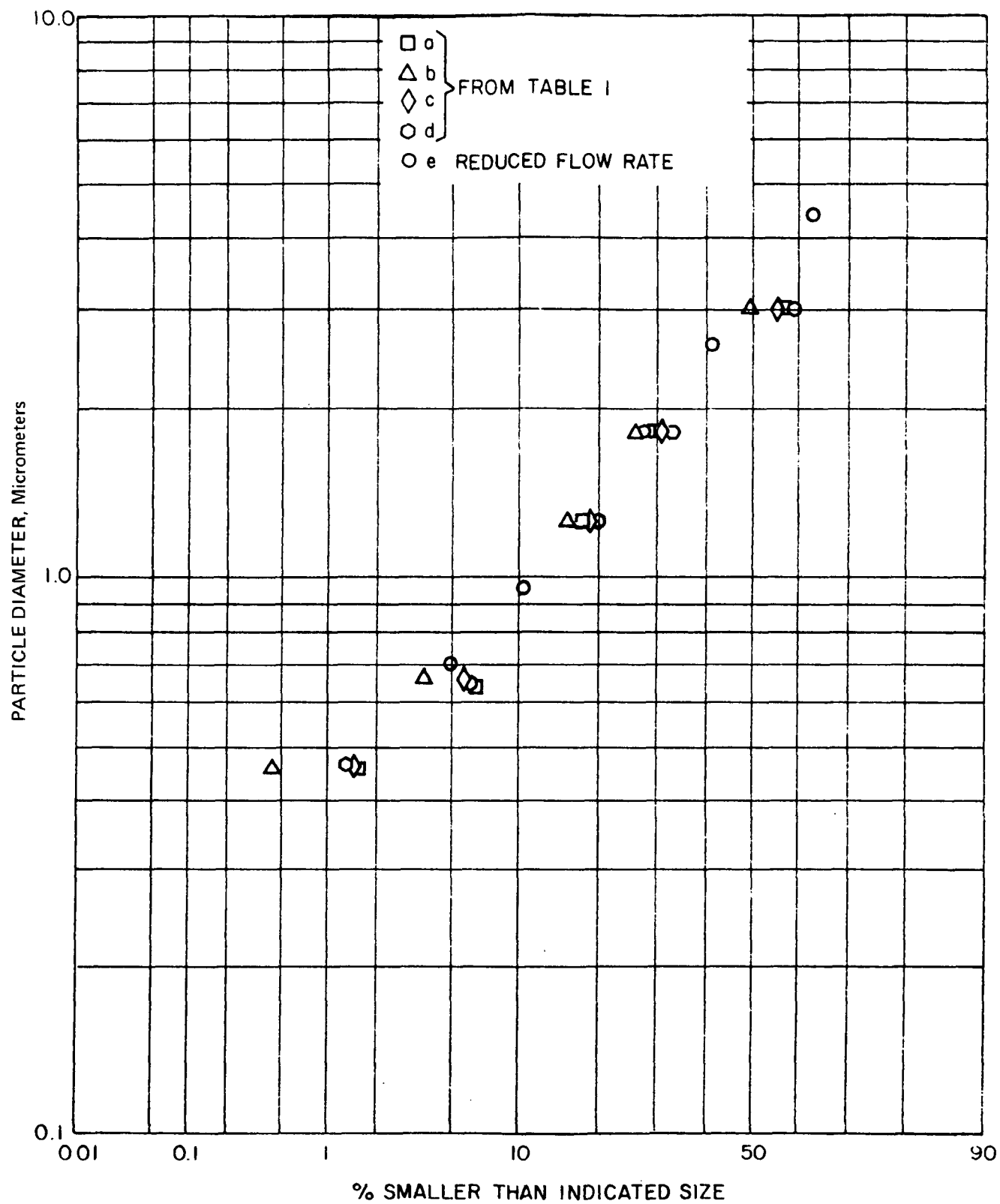
Figure 6. Inlet Particle Size Distributions from
Brink Impactor Data

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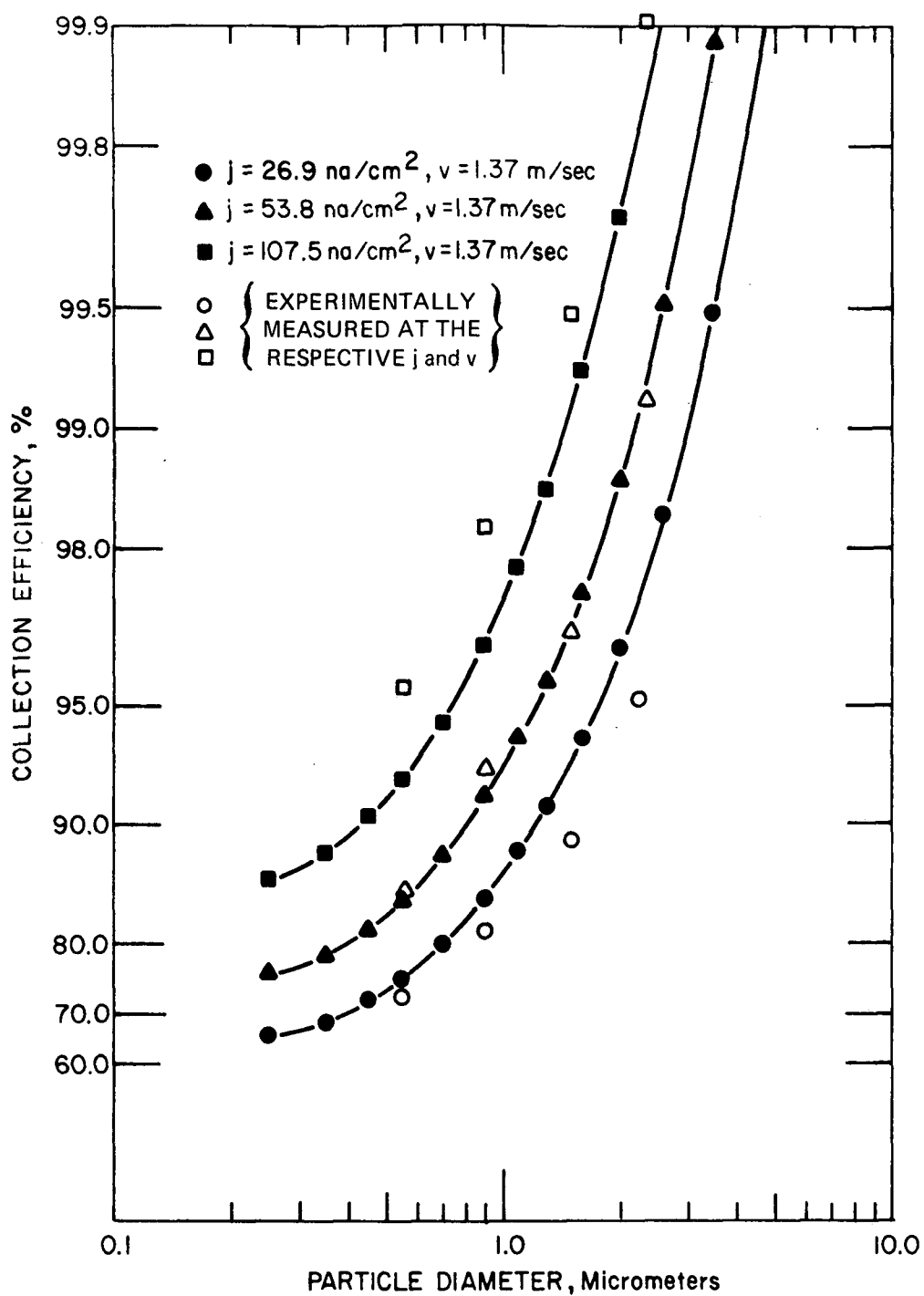
Figure 7. Experimentally Measured and Ideally Calculated Fractional Collection Efficiencies in the Laboratory Precipitator

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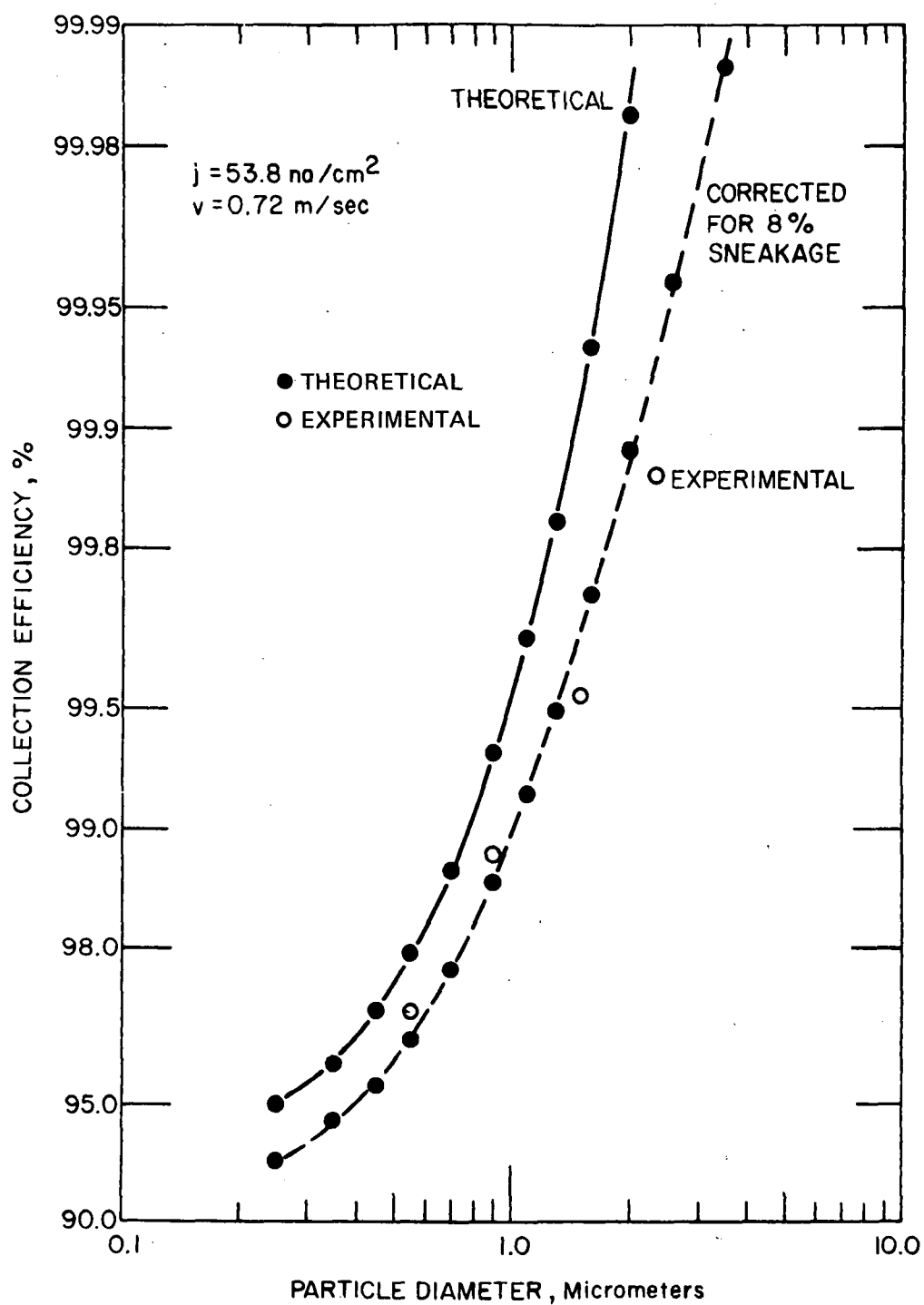
Figure 8. Experimentally Measured, Ideally
Calculated, and Sneakage-Corrected
Fractional Collection Efficiencies
for a Current Density of 53.8 $nA/cm^2$
and a Gas Velocity of 0.72 m/sec in the
Laboratory Precipitator

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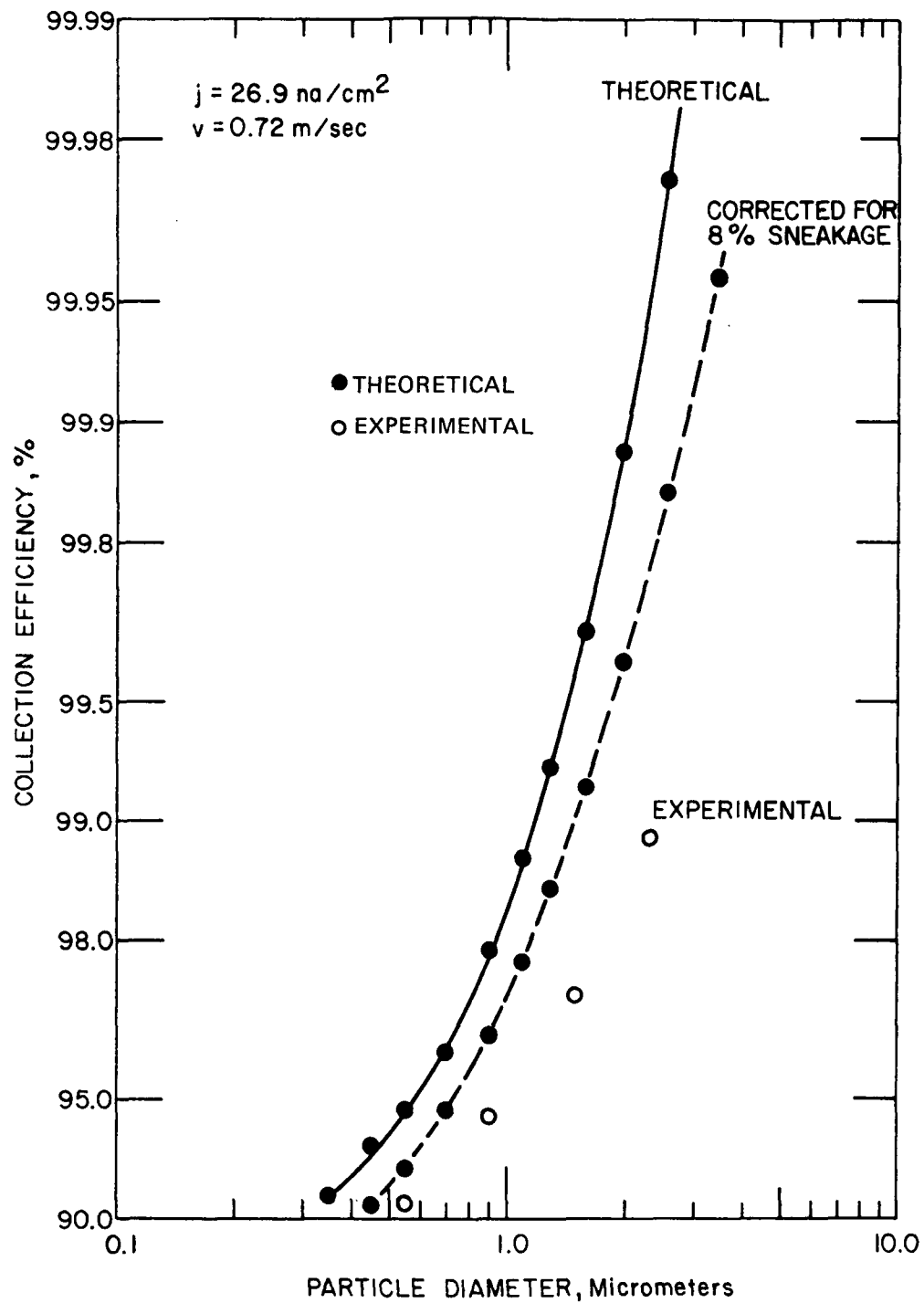
Figure 9.  Experimentally Measured, Ideally
Calculated, and Sneakage-Corrected
Fractional Collection Efficiencies for
a Current Density of 26.9 $nA/cm^2$ and a
Gas Velocity of 0.72 m/sec in the
Laboratory Precipitator

the agreement obtained between computed and measured results worsens with decreasing current density. The assumption of 8% gas sneakage over four stages improves agreement between computed and measured results at a gas velocity of 0.72 m/sec. However, the data obtained at this lower gas velocity with 26.9 $nA/cm^2$ are considerably below the computer projections. Possible causes for this lack of agreement are unmodelled effects such as non-uniform current density and electric field, and particle concentration gradients in the interelectrode space.

The performance of an electrostatic precipitator for monodisperse particles may be expressed in terms of an effective migration velocity, defined by

$$w_e = \frac{Q}{A} \ln \left(\frac{100}{100-n_i}\right) ,$$

where

$w_e$ = effective migration velocity of the particle size under consideration, cm/sec

$Q/A$ = ratio of volume flow to plate area, $m^3/sec/m^2$

$n_i$ = collection fraction of the particle size under consideration.

Figures 10, 11, and 12 give effective migration velocities obtained from the efficiency data in Figures 7, 8, and 9.

Figure 10 indicates good agreement between the computed and experimentally derived values of effective migration velocity. The assumption of 8% sneakage over four stages improves agreement between the computed and experimental results only for the lowest current density. Figure 11 indicates that, at the lower gas velocity of 0.72 m/sec, the correction for 8% by-passage gives agreement with the experimental results. At 26.9 $nA/cm^2$ and 0.72 m/sec, however, Figure 12) the experimentally derived values for effective migration velocity are between 7.7 and 25.5% lower than the corrected computed values. Due to the degree of uncertainty in the data, additional experiments are needed for definite conclusions regarding confidence limits for the model predictions. However, these comparisons suggest that, although the model closely approximates the experimental results, it tends to overpredict performance at lower values of current density and gas velocity, and underpredict performance for the highest value (107.5 $nA/cm^2$) of current density.

Additional work is planned under EPA support to evaluate the model predictions with varying electrode geometries and operating conditions.

Figure 13 is a comparison of results obtained with the Brink impactor and with a Climet-Ortec particle size analyzer. The particle size information obtained with the Climet-Ortec is based on a calibration curve obtained with polystyrene latex beads (PSL). Since the measurements obtained with the Climet

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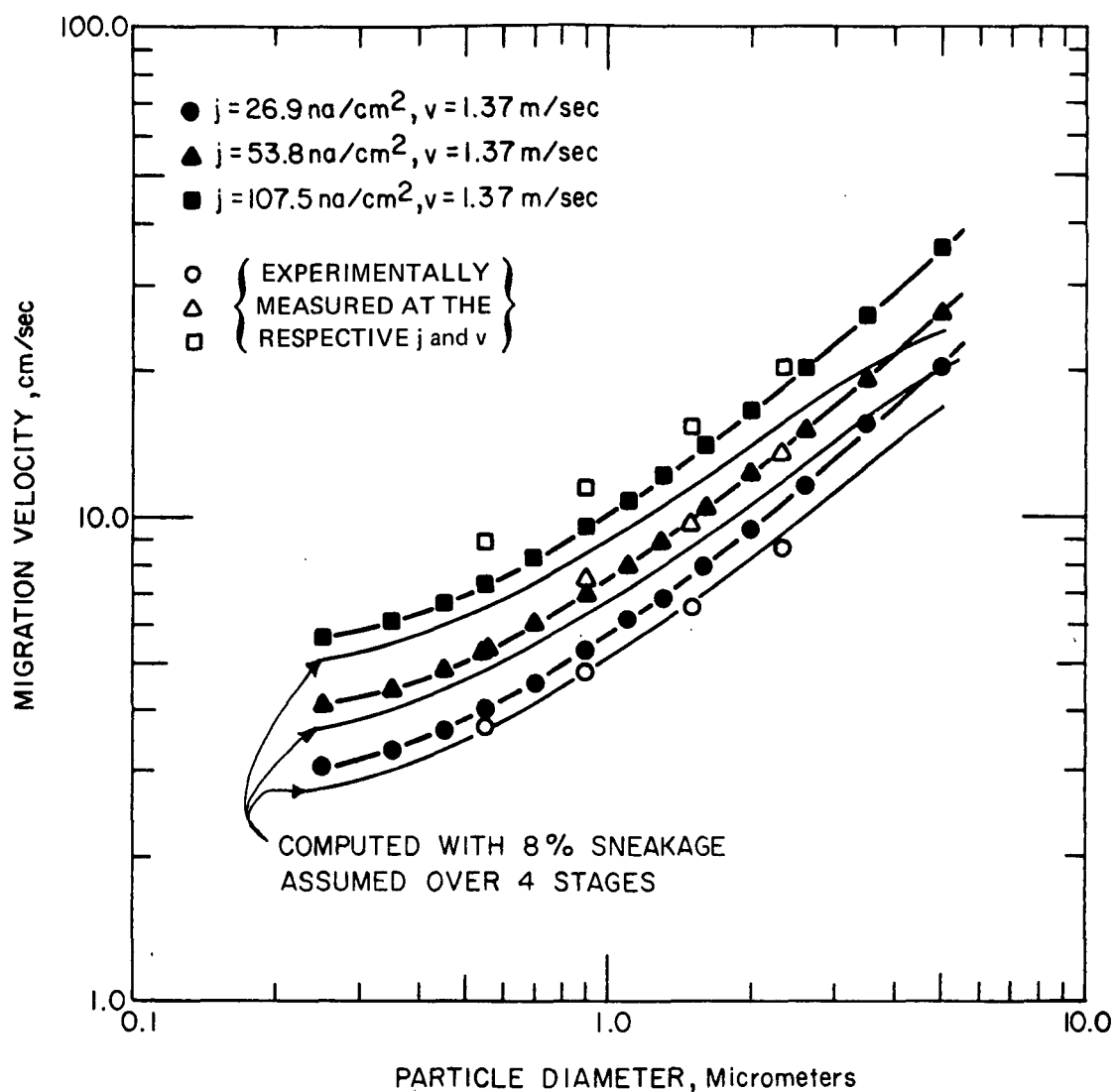
Figure 10. Experimentally Measured, Ideally
Calculated, and Sneakage-Corrected
Migration Velocities as a Function of
Particle Size in the Laboratory
Precipitator

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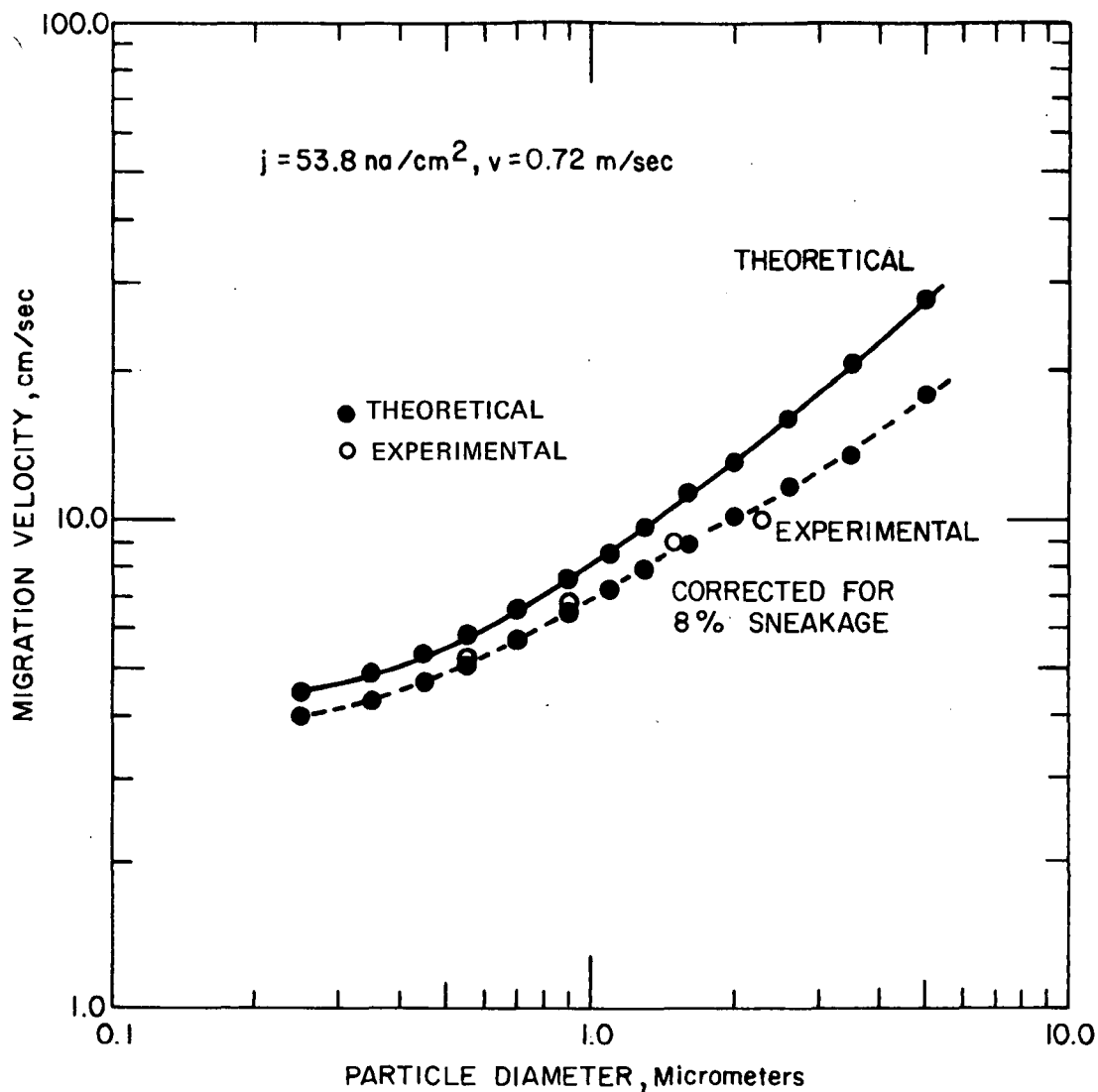
Figure 11. Experimentally Measured, Ideally Calculated, and Sneakage-Corrected Migration Velocities as a Function of Particle Size for a Current Density of 53.8 nA/cm² and a Gas Velocity of 0.72 m/sec in the Laboratory Precipitator
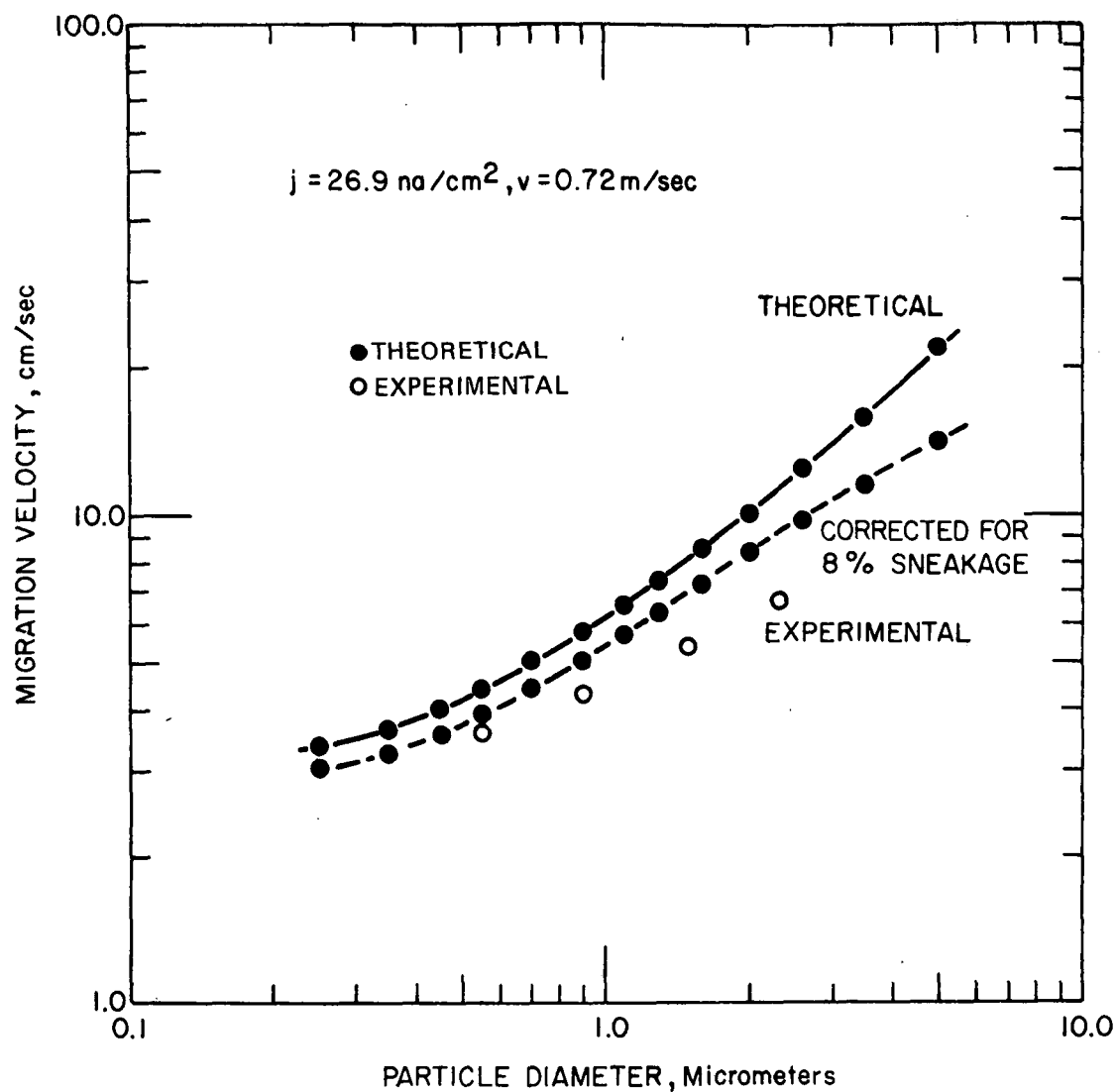
Figure 12. Experimentally Measured, Ideally
Calculated, and Sneakage-Corrected
Migration Velocities as a Function of
Particle Size for a Current Density
of 26.9 $nA/cm^2$ and a Gas Velocity of
0.72 m/sec in the Laboratory Precipitator
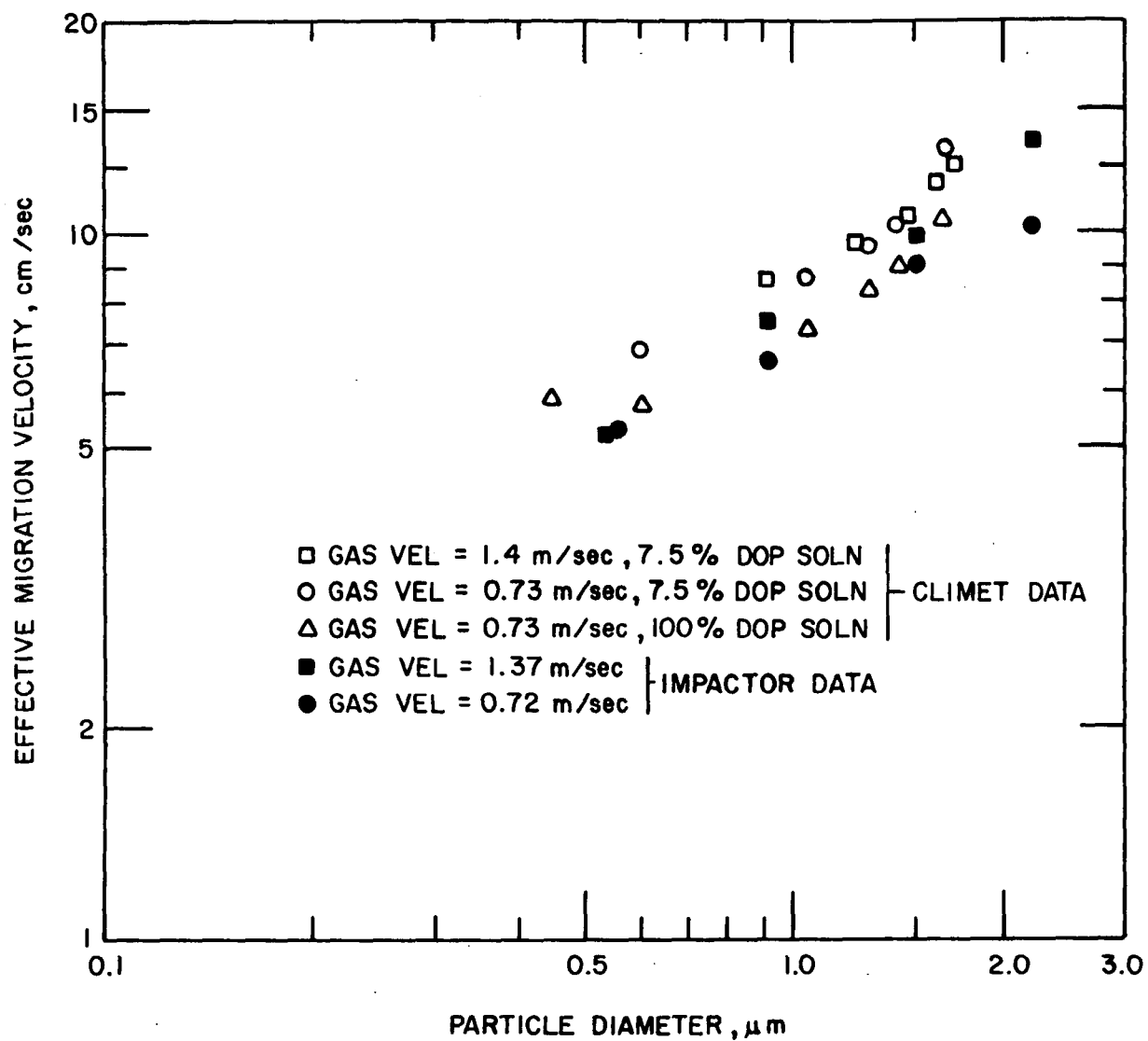
Figure 13. Effective Migration Velocities from Climet and Brink Efficiency Determination with DOP Particulate at 50 na/cm².

were accomplished with the use of a diluter with an intake several seconds in residence time downstream from the aerosol generator, the concentration of DOP in the solution used for the particle generator should not have an influence on the indicated particle sizes. The comparison in Figure 13 indicates fair agreement between effective migration velocities derived from measured efficiencies obtained with the Climet-Ortec and with the impactor. This suggests that the calibration curve obtained for the optical instrument with PSL is also reasonably accurate for a DOP aerosol.

SULFUR OXIDE AND PARTICULATE COLLECTION EXPERIMENTS

Since wet precipitator-scrubber systems have been proposed for removing particulate and sulfur oxides from stack gases, it is of interest to examine the simultaneous removal of DOP particulate and $SO_2$ in the laboratory test unit. Section V will present calculations which indicate that no significant additional $SO_2$ removal should be anticipated in a wet ESP due to the presence of a corona current and an electric field. The experiments reported in this section were limited in scope and were conducted primarily for the purpose of determining the validity of this conclusion.

The wet precipitator test unit was equipped with a positive displacement metering pump and a static mixer for metering and mixing a concentrated solution of sodium hydroxide into the water used for irrigating the collection electrodes. Sulfur dioxide was metered from a gas cylinder into the test unit upstream from the mixing chamber. Ambient air was used as the carrier gas at a flow velocity of 0.76 m/sec and the $SO_2$ was fed at a rate which resulted in a concentration of about 2000 parts per million by volume. The liquid to gas ratio was held constant at 2.8 $l/m^3$ (20.7 gal/1000 $ft^3$), and the residence time of the gas in the irrigated and electrified region was about four seconds.

The following groups of experiments were conducted:

1. The $SO_2$ removal efficiency of the device was determined with only $H_2O$ irrigating the electrodes with the electrified sets deenergized.

2. The $SO_2$ removal efficiency was determined with NaOH solution irrigating the electrodes with the electrical sets deenergized.

3. The $SO_2$ and particulate removal efficiency was determined using DOP particulate with NaOH solution irrigating the electrodes with the electrical sets energized.

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The $SO_2$ removal efficiencies obtained in these experiments are given in Table 2. The molar ratio of NaOH to $SO_2$ entering the test unit was 2.68 to 1 for experimental groups 2 and 3; there-fore, the degree of removal in these experiments is governed by mass transfer dynamics instead of chemical equilibria. Mass transfer calculations that will be discussed in Section V indicate that only a small portion of the absorption shown in Table 2 can be accounted for by the area of the collecting electrodes, and that electrical transport of $SO_2$ would be expected to have a negligible effect on the overall $SO_2$ removal efficiency. The major portion of the $SO_2$ absorption in these experiments was apparently achieved by droplets generated by the electrode irrigation manifolds. Table 2 shows that the $SO_2$ absorption decreased for the experiments conducted with DOP particulate undergoing collection. It is hypothesized that the indicated decrease of $SO_2$ uptake was caused by a noticeable decrease in the wettability of the collection electrodes resulting from the collection of oil droplets on the electrodes.

These limited experiments tend to confirm the hypothesis that electric fields and corona current do not significantly improve the $SO_2$ uptake capability of a wet ESP simultaneously collecting particulate and $SO_2$. Unfortunately, the conclusion is clouded somewhat by the decrease in absorption area caused by the DOP particulate. It may be argued, however, that plate wettability is also likely to decrease under practical operating conditions due to residual material remaining on the plates.

Particulate collection efficiencies were measured with the Climet-Ortec particle size analyzer during the $SO_2$-DOP particulate experiments. Effective migration velocities calculated from these efficiencies are presented in Figure 14, and the applied voltages for the "dry" and "wet" particulate collection experiments at a current density of 50 $\mu A/ft^2$ are given in Table 3. A comparison of the data in Figures 13 and 14 indicate that the particulate collection performance of the test unit was not affected, within the uncertainty of the measurement techniques, by the presence of $SO_2$ and the use of electrode irrigation.

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## TABLE 2

### SO$_2$ REMOVAL EXPERIMENTS

| Experiment Group No. (see text) | SO$_2$ Outlet Concentration ppm by volume | [1]SO$_2$ Removal Efficiency % |
|:---:|:---:|:---:|
| (1) | 1300 | 34.3 |
| | 1379 | 30.3 |
| | 1326 | 33.0 |
| (2) | 1015 | 48.7 |
| | 960 | 51.5 |
| | 1024 | 48.3 |
| (3) | 1172 | 40.8 |
| | 1109 | 44.0 |

1.  Average inlet concentration = 1979 ppm

## TABLE 3

### APPLIED VOLTAGE IN WET ESP , TEST UNIT AT
### 50 $\mu A/ft^2$ CURRENT DENSITY

kV

| Power Supply No. | Dry Walls, Ambient Air, DOP Particulate | Wet Walls, Ambient Air & $SO_2$, DOP Particulate |
|---|---|---|
| 1 | 33 | 25 |
| 2 | 32 | 33 |
| 3 | 31.5 | 31 |
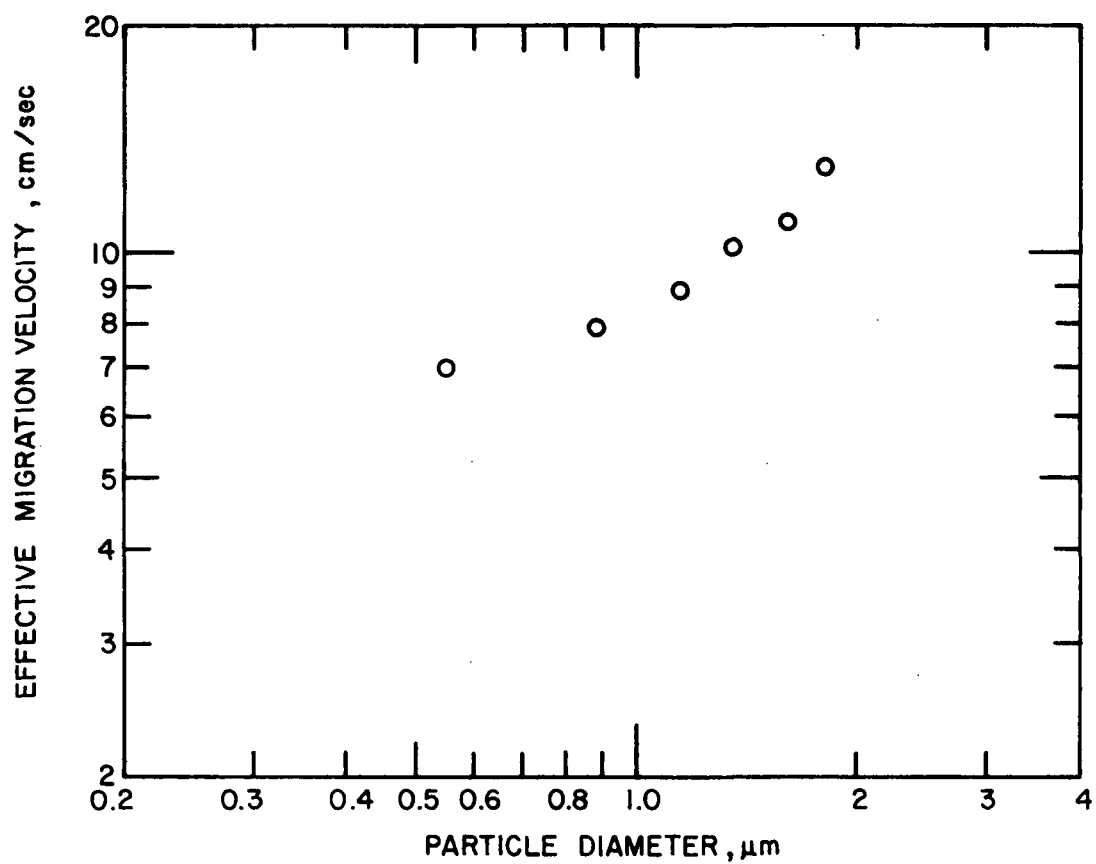
Figure 14.  Effective Migration Velocities for $SO_2$ – DOP Particulate Experiment (gas Velocity = 0.76 m/sec).

SECTION IV
EVALUATION OF FULL-SCALE WET
ELECTROSTATIC PRECIPITATOR INSTALLATIONS

FIELD TEST MEASUREMENTS

Introduction

Fractional and overall particulate collection efficiency measurements were made on a plate-type wet electrostatic precipitator collecting fume from an aluminum pot line consisting of horizontal stud self-baking aluminum reduction cells. These measurements were conducted under another contract and reported in detail elsewhere.[2] Comparisons were made between measured (with Andersen impactors) and predicted collection efficiencies obtained from a mathematical model of an electrostatic precipitator.

Figure 15 shows the arrangement of the unit on which the test series was conducted, and Figure 16 is a schematic of the liquor flow through the system as given by Bakke.[3]

Measurement Techniques

Particle Sizing -

Particle size and concentration measurements were conducted using the following methods:

1. Diffusional techniques using condensation nuclei counters and diffusion batteries for determining concentration and size distribution on a number basis for particles having diameters less than approximately 0.2 µm.

2. Optical techniques for determining concentrations and size distributions for particles having diameters between approximately 0.3 µm and 1.5 µm. Figure 17 shows the optical and diffusional sizing system.

3. Inertial techniques using cascade impactors for determining concentrations and size distributions on a mass basis for particles having diameters between approximately 0.25 µm and 5.0 µm. Andersen impactors were used simultaneously at the precipitator inlet and outlet. Extremely low mass loadings at the outlet made necessary the operation of the impactors for approximately 16 hours to obtain weighable quantities of particulate.

Mass Loading Measurements -

A modified EPA sampling train with an in-stack filter holder was

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SCRUBBERS

PLATFORM,
VERT.
SAMPLING

50 FT.

HUT, HORIZ. SAMPLING

30 FT.

WET
ELECTROSTATIC
PRECIPITATOR
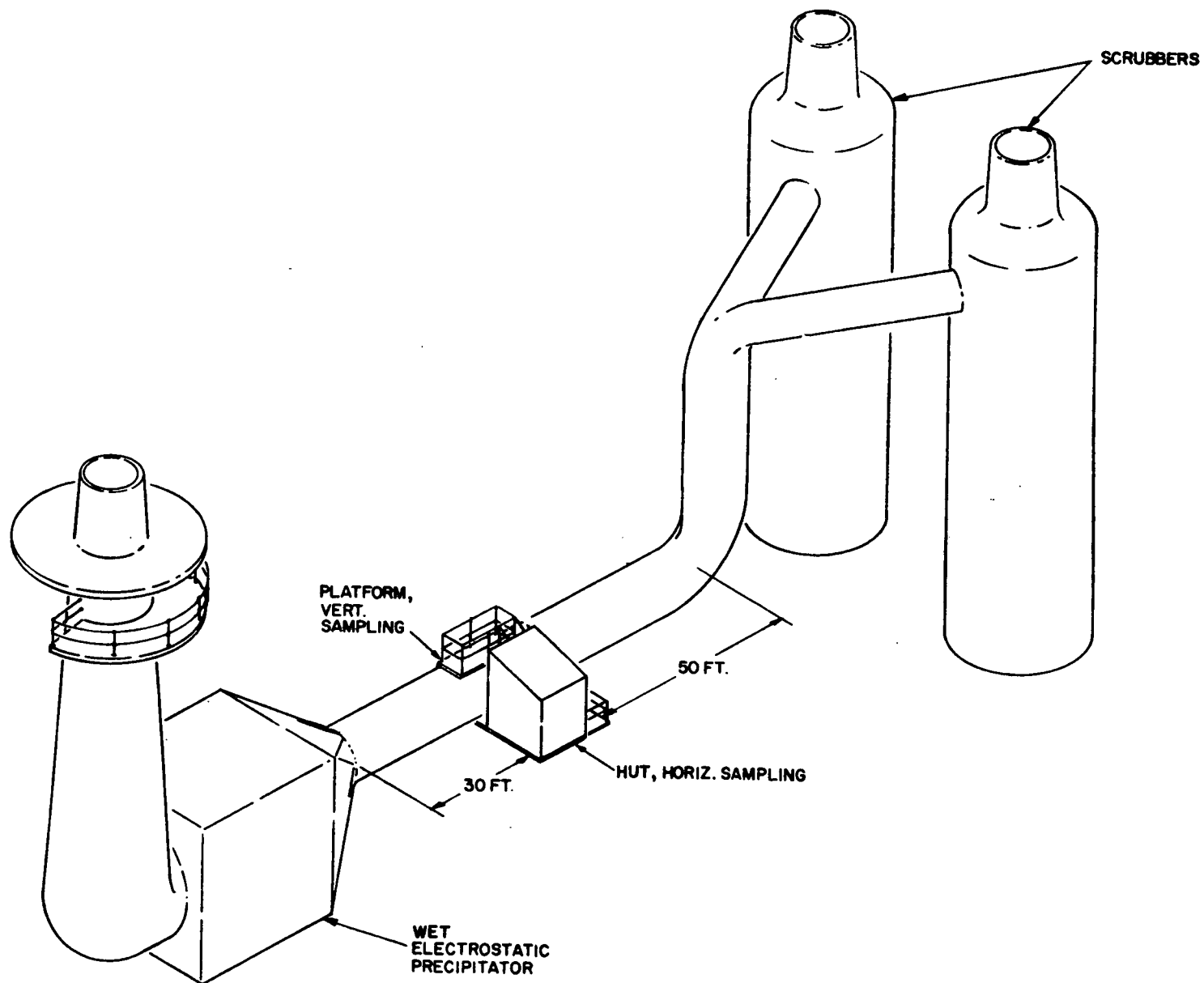
Figure 15. Schematic of scrubber-precipitator system and sampling locations
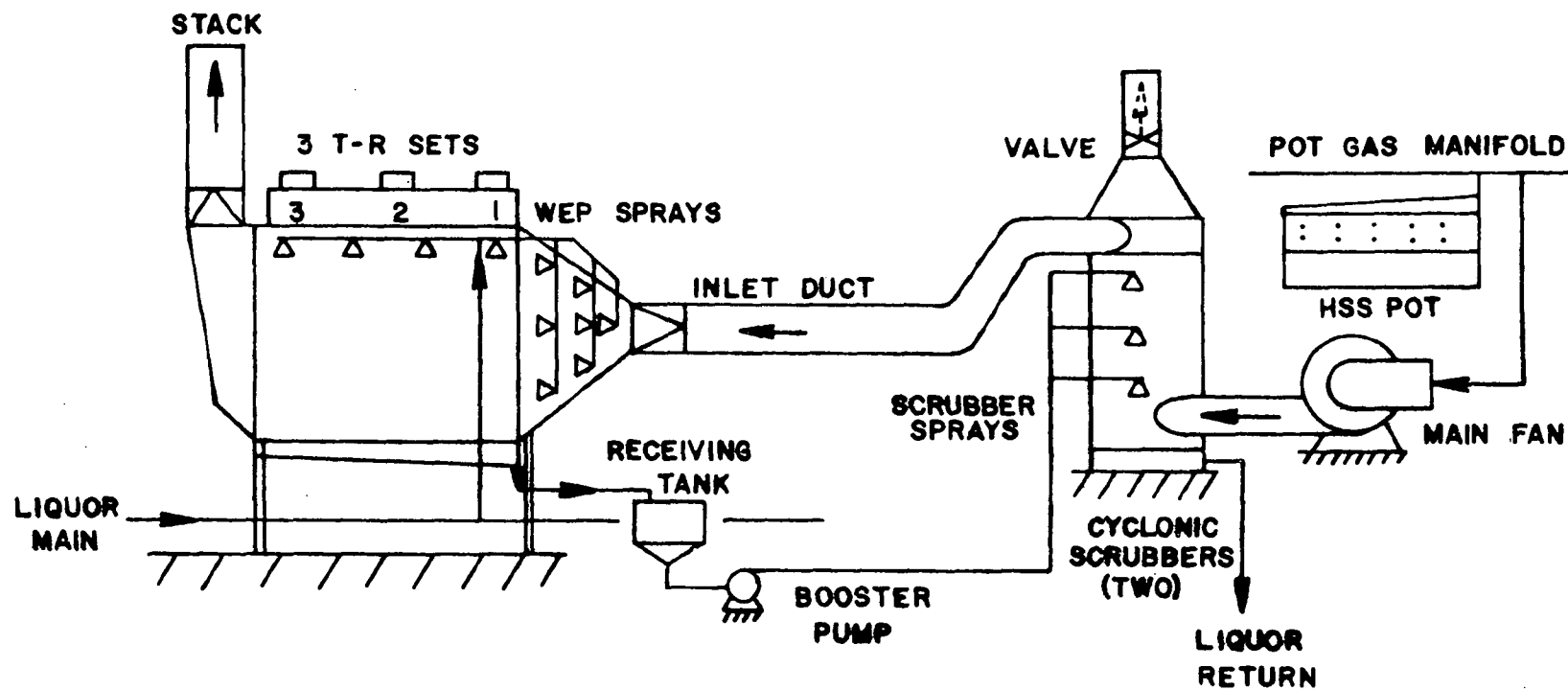
Figure 16. Schematic of primary emission control system [1]
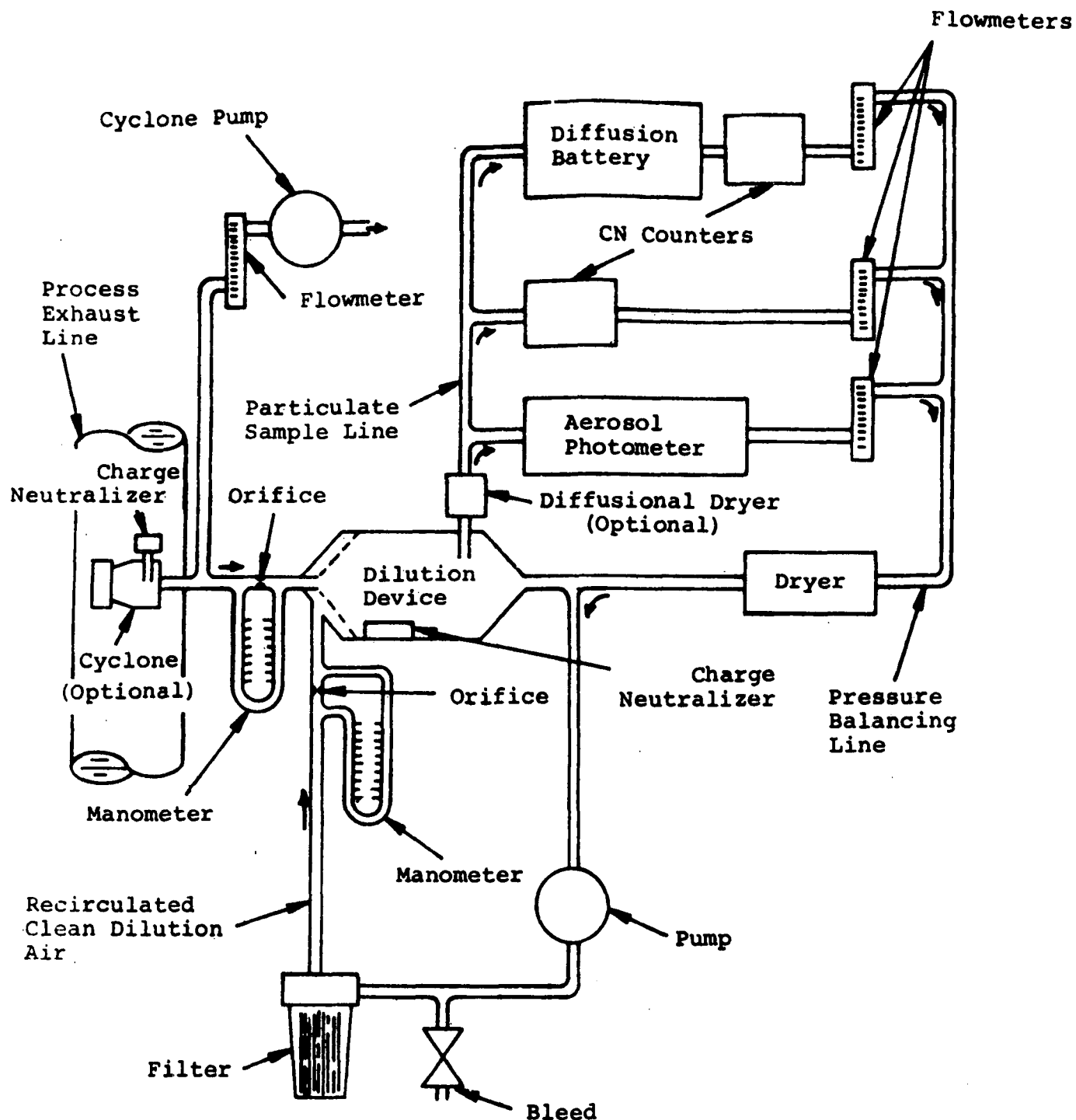
Figure 17. Optical and diffusional sizing system

used for the mass loading measurements, which were conducted at the inlet and outlet simultaneously with the impactor runs. An isokinetic traverse across the stack was conducted at both the precipitator inlet and outlet through a single sampling port at each location except for the last day, when a single point mass determination was performed at the outlet.

## Results

The mass median diameters of both inlet and outlet distributions as determined by the Andersen impactors are less than 1.0 μm. Figure 18 presents the Andersen data on log probability coordinates.

Figure 19 gives the cumulative size distribution on a number basis for this test series and various other industrial particulate sources as measured by optical and diffusional methods. Fractional efficiencies were computed from the optical and diffusional data, based on inlet measurements conducted on August 20 and 21, and outlet measurements conducted on August 22 and 23, 1974. Figure 20 gives the results of these calculations, together with the inertially determined fractional efficiencies. A theoretically calculated fractional efficiency curve is also given, and this will be discussed later.

### Mass Loading Measurements -

Mass train measurements were performed by Guardian Systems, Inc. of Birmingham, Alabama, under subcontract to Southern Research Institute and these results are given in Table 4. Fair agreement was obtained between the total mass loading with Andersen impactors and the mass train at the inlet, but severe disagreement occurred at the outlet.

The total mass obtained with a traverse using the mass train at the outlet was greater than that collected with the impactors by a ratio of approximately 5 to 1. When the mass filter was operated near the center of the stack and near the sampling location used for the impactors, the disagreement was reduced to a ratio of about 3 to 1. A comparison of outlet loadings however, indicates that the mass train results obtained during this test series are in fair agreement with those obtained previously by a local pollution control agency. Both the Andersen data and the mass data showed good reproducibility.

It is our conclusion that the most probable cause of the mass loading discrepancy is the collection of large water droplets containing solids by the mass filter. Such droplets would be subject to stratification in the stack, and this is qualitatively indicated by the decrease in loading which occurred when the mass train was operated at a single point. Additional work with a

32
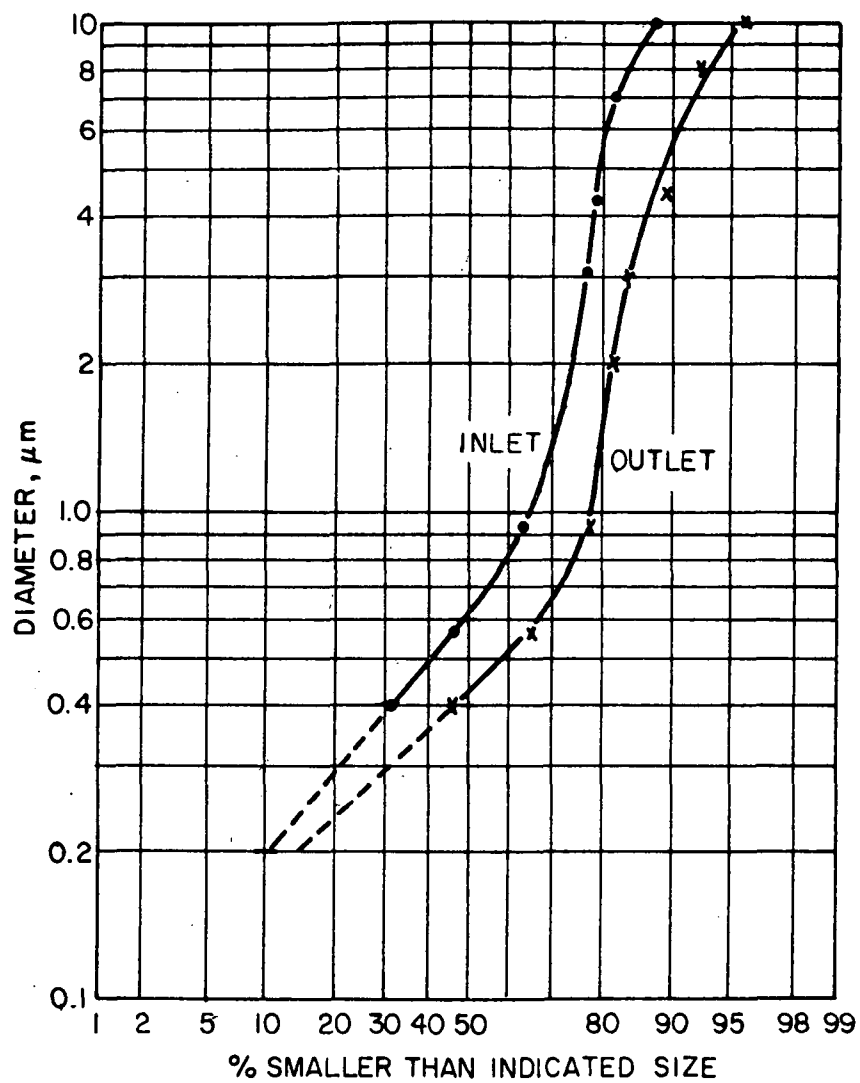
Figure 18. Andersen data on log probability
co-ordinates

33
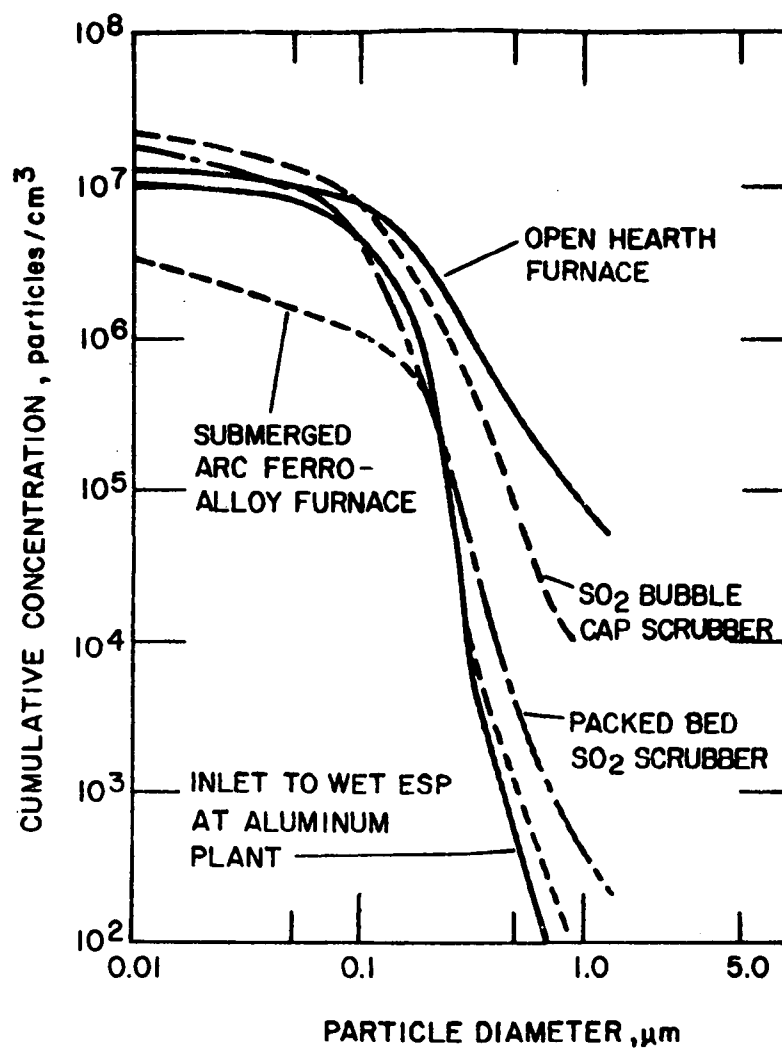
Figure 19. Cumulative size distributions on
a number basis for various
industrial particulate sources
as measured by optical and
diffusional methods

34
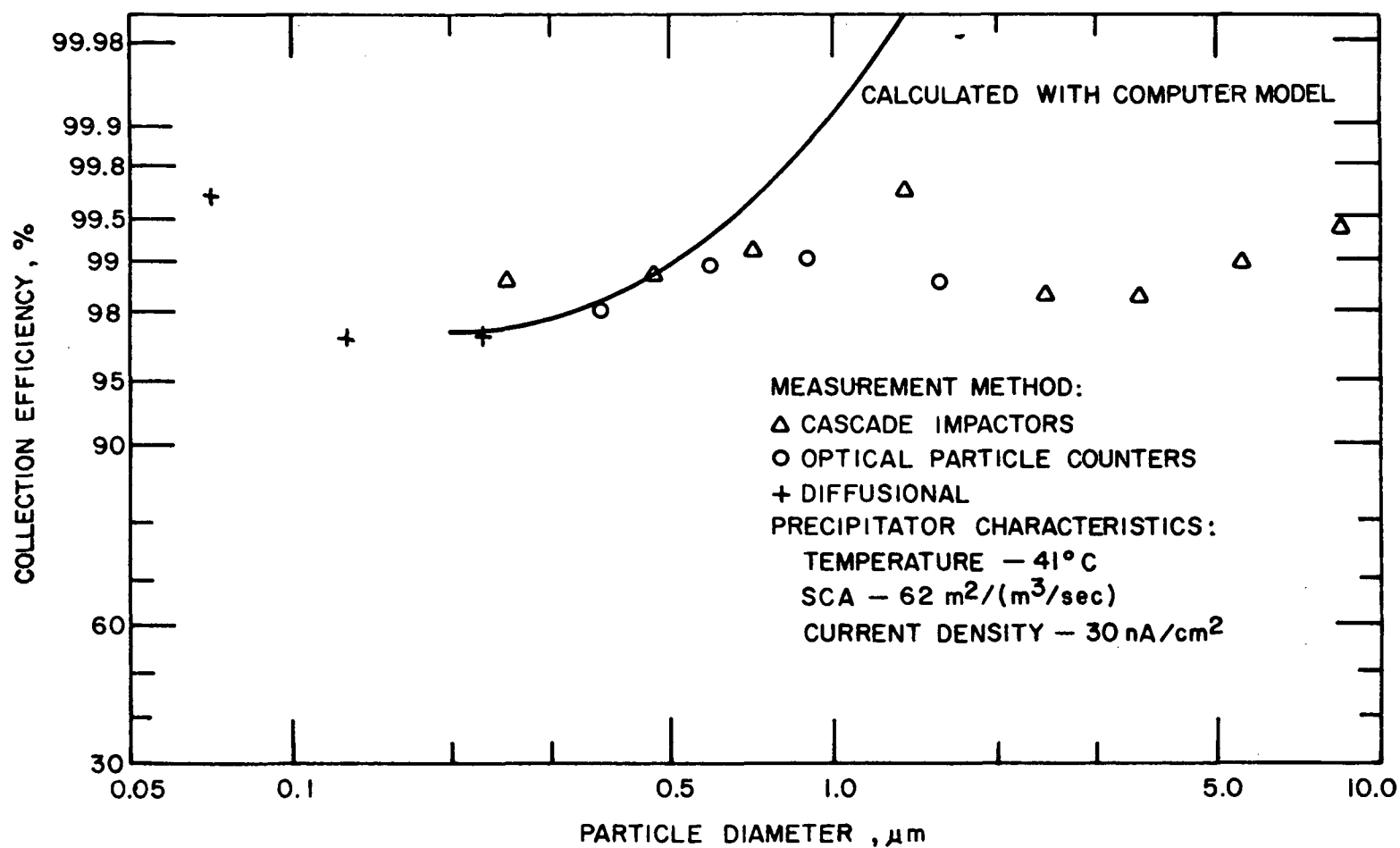
Figure 20. Measured fractional efficiencies for a wet electrostatic precipitator with the operating parameters as indicated, installed downstream of a spray type scrubber on an aluminum reduction pot line.

# Table 4. MASS TRAIN TEST RESULTS - 553 WET ESP

| Run No. | Inlet | | | | Outlet | | | |
|---|---|---|---|---|---|---|---|---|
| | 1 | 2 | 3 | 4 | 1 | 2 | 3 | 4[4] |
| Date | 8/20 | 8/21 | 8/22 | 8/23 | 8/20 | 8/21 | 8/22 | 8/23 |
| Sampling Time, min. | 300 | 250 | 280 | 310 | 376 | 375 | 375 | 360 |
| $H_2O$, % by vol in gas | 5.09 | 4.91 | 5.45 | 6.03 | 5.19 | 4.96 | 5.22 | 5.95 |
| Avg. Gas Temp., °C | 40.9 | 40.9 | 40.9 | 40.9 | [2]34.2 | 38.1 | 38.1 | 38.1 |
| [1]Flow, am$^3$/sec | 67.2 | 62.5 | 54.9 | 62.5 | 44.6 | 43.5 | 44.4 | 43.9 |
| [3]Flow, DN m$^3$/sec | 55.4 | 51.7 | 45.1 | 51.1 | 37.6 | 36.3 | 36.9 | 36.3 |
| mg/am$^3$ | 89.0 | 94.5 | 95.7 | 100.9 | 4.58 | 4.26 | 3.57 | 1.97 |
| gr/acf | 0.0389 | 0.0413 | 0.0418 | 0.0441 | 0.00200 | 0.00186 | 0.00156 | 0.00086 |
| gr/Dscf | 0.0443 | 0.0449 | 0.0476 | 0.0494 | 0.00220 | 0.00209 | 0.00166 | 0.00098 |
| Efficiency, % | | | | | 95.03 | 95.34 | 96.51 | 98.02 |

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Notes:
1. Based on traverse across one sampling port and area of 3.05 m$^2$ (32.85 ft$^2$)
2. Based on traverse across one sampling port and area of 3.54 m$^2$ (38.10 ft$^2$).
3. 0°C and 760 mm Hg.
4. Obtained at a single point near the center of the stack.

traverse using a sampling device designed to provide sizing information above 10 μm diameter would be required to resolve the problem.

## Analysis of Results

The wet electrostatic precipitator on which this test series was conducted is a wire and plate design with three electrical sections in series in the direction of gas flow. Plate-to-plate spacing is 30.5 cm (1 ft), and each collecting electrode is 1.83 m long (6 ft) and 7.52 m high (25 ft). Thus, the total parallel plate collecting electrode length is 5.48 m, or 18 ft. Each electrical set powers 28 gas passages. The total parallel plate collecting area is 2342 $m^2$ (25,200 $ft^2$), and the "transverse baffles", which are perpendicular to the gas flow, provide additional collecting electrode area. The effective collecting area provided by these baffles was estimated as 390 $m^2$ (4200 $ft^2$), resulting in a total collection area of 2732 $m^2$ (29,400 $ft^2$). Average specific collecting area during the test series was therefore 62 $m^2/(m^3/sec)$, or 315 $ft^2/(100 cfm)$.

Electrode irrigation is provided by sprays at the precipitator inlet and above the collection plates. The sprays provide a mist which is collected along with the particulates in the flue gas, and the electrode cleaning is accomplished by the coalescence and subsequent downward flow of the collected spray droplets. The sprays are operated continuously, except for those installed near the precipitator outlet, which are operated only periodically. These spray nozzles were not in operation during the test program. The irrigating fluid is a high pH sodium-based liquid which is returned to clarifiers and a cryolite recovery plant. Plant personnel reported that the cryolite recovery system is essentially a closed liquid loop, which results in a solids content of about 5% by weight being returned to the wet ESP - scrubber system. Liquor flow through the wet ESP during the test program was constant at 31.5 l/sec (500 gal/min), which gives a liquid to gas ratio of about 0.7 $l/m^3$ (5.3 gal/1000 $ft^3$). Liquor temperature, based on measurements reported by plant personnel, ranges from 90 to 104°F, and is usually 94 to 95°F. No significant temperature drop has been observed in the liquor loop across the precipitator.

Voltage and current readings were obtained from the panel meters of the 553 precipitator periodically during the test program. At the conclusion of the test program, voltage-current curves were obtained for the unit with the spray system operating normally. The secondary voltage-current relationships are given in Figure 21, along with the range of operation that was observed for each electrical set during the test program. The difference
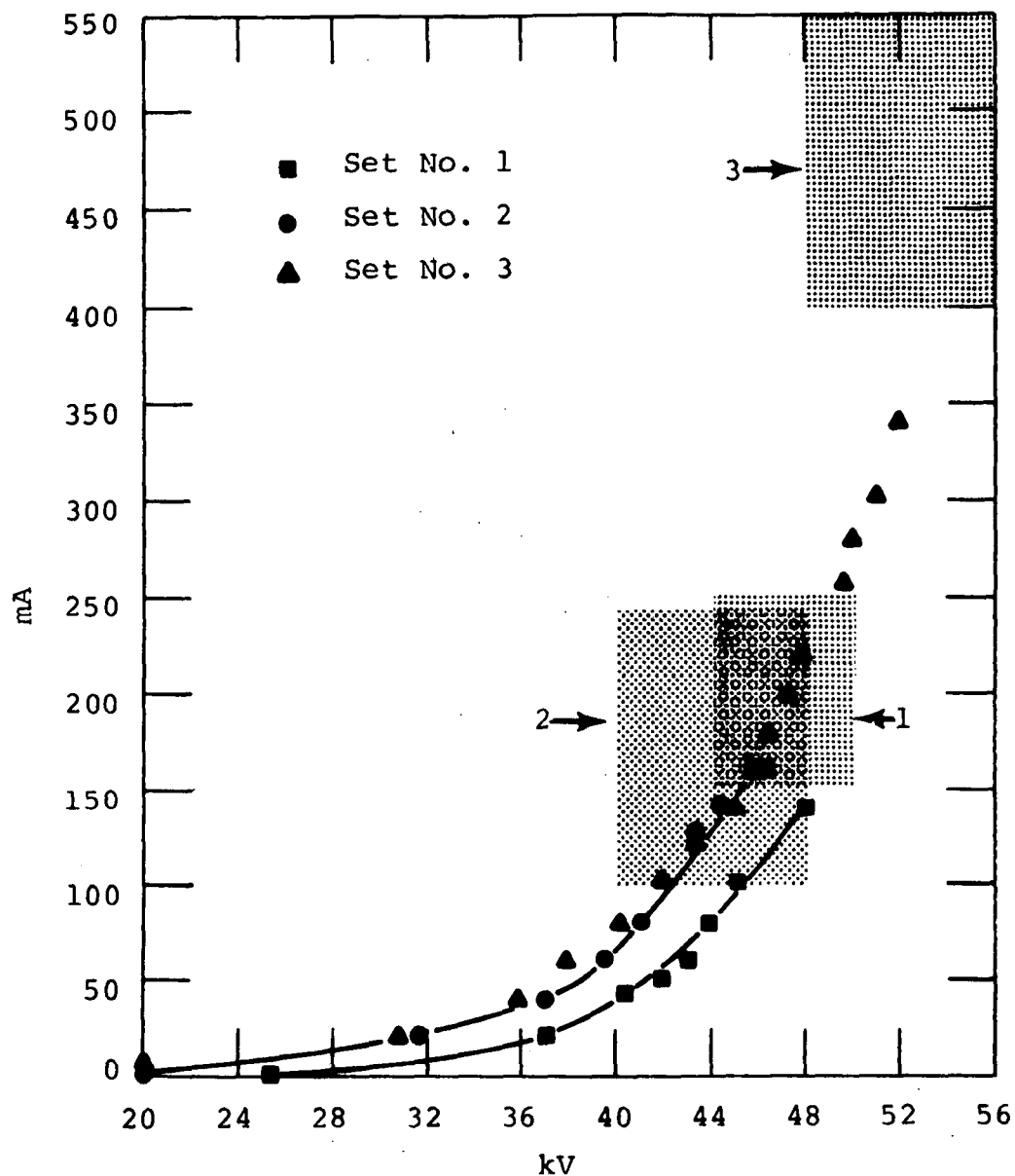
Figure 21. Voltage Current Relationship (Manual Control)
and Operating Ranges (Automatic Control)
Shaded Areas Designate Observed Range of Operation
For Each Electrical Set During the Test Program

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between the voltage-current curves and the operating ranges is a result of the fact that, in normal operation, the power supplies are operating under automatic control with a certain spark rate, whereas the V-I curves were obtained by manually increasing the applied voltage until sparking occurred. The plant personnel were operating the power supplies at a spark rate which was believed to maximize the time-averaged electric field.

The V-I curve for the first electrical set is shifted toward high voltages for a given current when compared with readings from the other electrical sets. This behavior is often observed and is a reflection of the higher space charge density contributed by the higher particulate loadings which exist in the inlet field. Although the third field operates at a relatively high current, the average current density for all three sets was only about 30 na/cm². The resistivity of the particulate is not a factor in the wet mode of operation.

Figure 20 presented fractional efficiencies and a predicted curve obtained from a theoretically based computer model of an electrostatic precipitator[1]. The mathematical model calculates theoretically expected collection efficiencies for representative particle diameters as a function of precipitator operating conditions. Predicted collection efficiencies for each particle diameter are a function of the electric field, the charge on the particle, and the ratio of collection area to gas volume flow rate. Fair agreement is obtained between the measured efficiencies over the particle diameter range 0.25-1.3 µm, but the measured values depart drastically from the predictions and from the trends indicated by the previously discussed laboratory measurements, at diameters larger than 1.5 µm. This apparent departure from the expected functional form may be caused by the generation of particles within the device, possibly originating from the liquid sprays or from reentrained liquid that is not captured by the outlet transverse baffles, which are considered by the manufacturer to function as an electrostatically augmented mist eliminator. It should be noted that the diameter band 0.25-1.3 µm, based on the Andersen measurements, represents 54% of the mass at the inlet and 56% of the mass at the outlet.

Electrostatic precipitator performance is often described by an empirical performance parameter termed the precipitation rate parameter. The parameter is obtained by evaluating the Deutsch equation using the overall mass efficiency and the ratio of volume flow to plate area:

$$w_p = \frac{V}{A} \ln \left(\frac{100}{100-\eta}\right)$$

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Evaluation of this relationship using the data in Table 4 gives the results presented in Table 5. A predicted precipitation rate parameter may be obtained from the computer model based on the inlet size distribution obtained from the Andersen impactor measurements. Based on the predicted efficiencies indicated by Figure 20, numerical integration over the inlet size distribution gives a total predicted penetration of 1.1% (98.9% efficiency), and predicted precipitation rate parameter of 7.3 cm/sec, which shows fair agreement with the data in Table 5. Figure 20 shows, however, that the model underpredicts fine particle collection efficiencies, and overpredicts collection for particles larger than about 0.60 μm.

SUMMARIES OF TRIP REPORTS TO WET ESP MANUFACTURERS AND USERS

Dr. Charles E. Bates, head of the metallurgy section, has met with representatives of Lurgi in Frankfurt, West Germany, and British Steel Corporation in Moorgate, England. The following paragraphs summarize the more important aspects of these meetings.

Visit to British Steel Corporation

British Steel Corporation began applying wet electrostatic precipitators to electric arc furnaces about ten years ago to eliminate a fluorine problem in the effluent which, for the case of two plants, resulted in the development of fluorosis in cattle grazing nearby. Effluent particles from the electric arc furnaces are predominantly under one micron in diameter. Therefore, the only choices for emission control were electrostatic precipitators, bag filters, and venturi scrubbers. The difficulties with bag filters were that the temperatures of the gas streams were quite high, the particles quite small, and the bags would tend to plug. The presence of moisture and fluorine eliminated the use of fiberglass bags, and the operating costs of venturi scrubbers were prohibitive. After considering the possibilities, British Steel decided to install wet precipitators. Initially, the design did not permit 100% saturation of the gas before entering the precipitator, and after four years of labor and much cost, it became apparent that 100% saturation was essential at all times. This saturation must be achieved in a conditioning chamber because of the time required to fully saturate the gas stream. British Steel found that if the gas entering the precipitator were not cooled and completely saturated, the warm, moist iron oxide would stick to the baffle, wires, and plates of the precipitator like cement, but if the effluent were cooled and wet, the iron oxide would not stick. Since water carry-over during flushing caused sparkover and could possibly knock out a complete power supply, individual power supplies were installed on each stage of the precipitator to enable sections being cleaned to have the

TABLE 5

PRECIPITATION RATE PARAMETERS

| Run No. | Gas Flow, $m^3$/sec | Mass Efficiency, % | Precipitation Rate Parameter, cm/sec |
|---------|---------------------|---------------------|--------------------------------------|
| 1 | 44.6 | 95.03 | 4.90 |
| 2 | 43.5 | 95.34 | 4.88 |
| 3 | 44.4 | 96.51 | 5.45 |
| 4 | 43.9 | 98.02 | 6.30 |

power turned off. The first installation had two fields in series, but while the plates in one section were being washed, only 50% of the cleaning capacity was being utilized. In the second installation, the arrangement was three fields, so that two fields could operate while the third was being cleaned. All three fields operate expect when washing and all precipitators are washed intermittently for 3-4 minutes during each 30 minutes of operation with high pressure water sprays. Constant irrigation proved unsatisfactory because of irregularities in either the water flow or plate shape.

The first precipitators installed by British Steel were designed based on a precipitation rate parameter of 10 to 12 cm per second. Experience showed that better results were obtained if the design precipitation rate parameters were reduced to 7 to 8 cm/sec. Specifications required by British Steel for wet precipitators recently constructed include those given in Table 6.

Several types of discharge electrodes were tested prior to installation of the electrostatic precipitators at British Steel. Collection electrodes are of the plate type. Initially a mild steel barbed wire was used, but dust bonded to the wire, leading to corrosion. This same experience occurred when a galvanized wire was used. Differently designed wires were found to be satisfactory, but the most successful design proved to be a stainless rod with stainless spikes drilled or welded into the rod's surface. The stainless steel rod gave the necessary rigidity, and the spikes provided the necessary field strength (shown in Figure 22).

The maintenance costs of these wet precipitators are now quite low and no unscheduled maintenance is experience. The operating costs consist of costs for power, water supplies, and maintenance, with about 1/3 of this cost due to maintenance. The power supplies, with automatic voltage control, are rated at 60 kV and 450 milliamps. Silicon rectifiers are employed and the primary power supply is 415 volts, 50 cycles, + 6%. The precipitators are supplied by W. C. Holmes of Huddersfield in Yorkshire.

Visit to Lurgi Apparate - Technik GmbH, Frankfurt, Germany


Lurgi is of the opinion that wet precipitators are applicable where the gas is already wet or the resistivity of the dust is so high that water must be added to saturate the air, thus lowering resistivity. If large amounts of water must be added, or if the gas volumes are not too high, they prefer to use a scrubber and avoid some of the corrosion and maintenance problems and high capital costs associated with the wet precipitators. To capture

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## TABLE 6

### PERFORMANCE CHARACTERISTICS OF WET PRECIPITATORS SPECIFIED BY BRITISH STEEL

| | |
|---|---|
| Gas Flow | 43 m$^3$/sec at 60°C |
| Particle Size Distribution of Inlet Dust | Particles range from 5 μm to .1 μm diameter, 95% less than 0.5 μm |
| Precipitation Rate Parameter | 8.5 cm/sec maximum |
| Collection Efficiency | 97.5% |
| Inlet Dust Loading | 4.6 g/m$^3$ (2.0 gr/ft$^3$) at N.T.P. dry (maximum) |
| Outlet Dust Loading | 0.115 g/m$^3$ (0.05 gr/ft$^3$) at N.T.P. dry (maximum) |

Studs 12.7 mm x 3.175 mm
dia are drilled or welded on

34 pitches @ 73 mm = 2482 mm

2751 mm
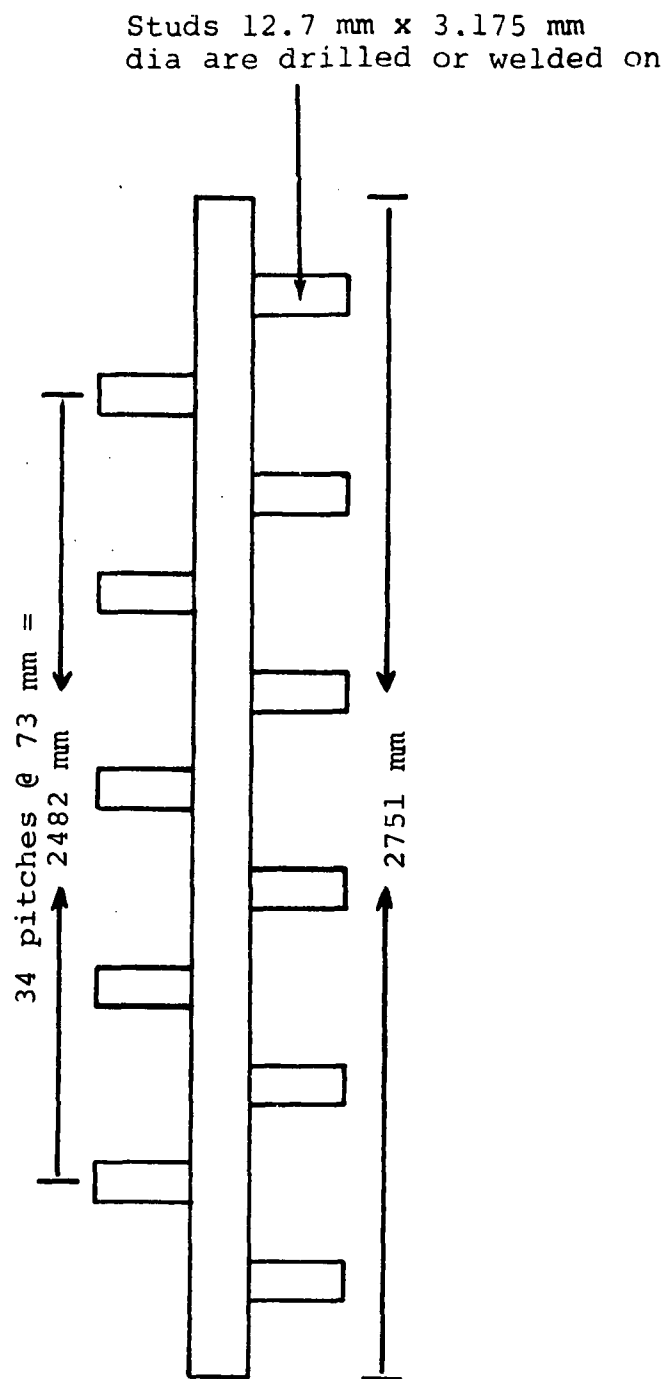
Figure 22.   Discharge electrodes for wet ESP units
at British Steel.

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particulate and gases, one could use a scrubber to remove chloride and sulfur oxides. Lurgi has never looked at wet precipitators for power plant applications because they consider them too costly and complicated. Lurgi has installed wet precipitators on cupolas, arc furnaces, basic oxygen furnaces, blast furnaces, and scarfers.

Basic Oxygen Furnaces -

Basic oxygen furnaces commonly use 40-60" pressure drop scrubbers for particulate control, but Lurgi has built about five wet electrostatic precipitators for this application. These wet cleaning plants require considerable maintenance due to the large amount of oxygen, fluorides, and sulfides in the gas, which lead to corrosion of the internal parts of the control system. Corrosion could be avoided by using stainless steel but this would be prohibitively expensive. The market for wet precipitators for this application does not appear favorable to Lurgi.

Scarfers -

Scarfers commonly use water sprays to cool the grit produced by the scarfing operation, so a wet precipitator becomes a natural choice. Saturation of the gas before entering the precipitator is essential as is maintaining an adequate flow of water over the plates to prevent buildup. Fully saturated effluent does not bond, and in fact, is easily washed off. The gas volumes are relatively low, so one can afford stainless steel internals, thereby minimizing maintenance costs. The wet precipitator provides good performance, low operating cost, and can be small in size when applied to scarfers. Lurgi has sold seven or eight wet precipitators for effluent control of scarfing operations over the past few years.

Blast Furnaces -

Electrostatic precipitators were very popular on older blast furnaces where the top pressure was not particularly high. At the present time, however, all blast furnaces being built have top pressures of 15 to 20 psi and part of this pressure is available to do work. With a high top pressure, sufficient energy is available to simply pass the gas through a venturi and allow the gas expansion to do much of the work required. Therefore, high energy wet scrubbers, using this gas energy, can be employed just as economically with a lower capital investment compared to precipitators. The wet precipitator does work well on blast furnaces and corrosion is not a problem because of the very low oxygen partial pressure, and hence, low corrosion potential, of the effluent gas. Lurgi has about 250 wet precipitators in operation and a partial list of these was

furnished to Southern Research. Demand for wet precipitators for blast furnace cleaning is not expected to increase except in undeveloped countries where sophisticated furnaces will not be used.

Cupolas -

Small gas volumes are involved in cupola operations compared to most processing equipment, and wet precipitator usage is possible. Lurgi has built three or four wet precipitators and two or three dry precipitators in the last few years for foundries, but because of the small profit margin involved for supplying pollution equipment for small foundry operations, Lurgi does not expect to pursue it.

Electric Arc Furnaces -

The most common control devices for arc furnace effluents are bag houses and electrostatic precipitators. At this time Lurgi leans toward dry precipitators since the dust resistivity is reasonable when the gas is diluted with plant air and cooled to about 70°F. Above a gas volume of about 500,000 cu. meters/hr, the precipitator is less expensive than a bag house in total operating costs.

Design Details -

Lurgi wet precipitators normally use plates and two types of water supply. The continuous water sprays (about 45-50 psi) employ nozzles to achieve as fine a mist as possible, using clean water. The second system of sprays provides a high volume of water to the plates every three or four hours to give the plates a good cleansing. During this flushing operation the power is left on, but with voltage reduced to 10-15 kV instead of the normal 40 kV. The precipitator normally has three fields, sometimes four, and one field is washed at a time.

When the gas going through the precipitators is incompletely saturated, buildup and corrosion can be serious problems. Basic water will cause buildup and acid must be added to get a release. Acidic water will cause corrosion in a precipitator.

There are very few, if any, gravity water flows in use because of difficulty in maintaining uniform water flow down the plates (the weirs cannot easily be kept level and the plates or tubes cannot easily be kept straight). Precipitation rate parameter values of 7 to 10 cm/sec in wet precipitators and 3 to 5 cm/sec in dry precipitators are obtained with the same distance between wires and plates. The precipitation rate parameter in dry precipitators at 70 to 90°F using plant dilution air will sometimes run as high as 10 cm/sec, but will vary widely with dust resistivity.

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On old blast furnaces, dust loads of 0.78 to 1.3 grams per
cubic meter are common in the raw effluent from the furnace.
A cyclone then drops the dust loading to 0.65 to 0.90 grams/m$^3$
With a center charge going into the blast furnace, the dust
coming out is typically 0.5 to 0.65 grams/m$^3$. The gas is then
cleaned by a wet scrubber which cleans the effluent to 30-100
mg/m$^3$, and then finally by a precipitator which cleans the gas
to about 5 mg/m$^3$.

If the gas is to be sent to a turbine, the dust loading must
not be higher than 0.5-1.0 mg/m$^3$, which can usually be achieved
with a secondary precipitator. The discharge electrodes of such
a precipitator are schematically illustrated in Figure 23. Com-
monly three fields are used. In the first field there will be
8 mm diameter stainless steel rod with welded studs spaced on
the rod. The rod in the second stage is the same size but with
square studs welded on it. In the third stage, a bare wire is
used with water flowing over it, the water stream itself pro-
viding enough asperities to cause corona for charging the dust.

## Mikropul Division of United States Filter Corporation

Mikropul has recently constructed large-scale wet precipitators
on aluminum pot lines, and the company has also been actively
pursuing wet precipitator usage in other application areas.
During the course of this contract, a trip was made to the
Mikropul Plant in Summit, New Jersey, and discussions were held
concerning Mikropul's experience and pilot plant test program.

At the time of the plant trip, the company was running pilot
plant tests of a wet precipitator designed for collection of
sulfur oxides and fly ash from a coal-fired power plant. The
pilot plant is equipped with a propane burner to produce syn-
thetic flue gas. Collected fly ash from a coal-fired boiler
is injected with a blower, and $SO_2$ is added from a gas cylinder.
At the time the visit was made, $SO_2$ and particulate removal
trails were being performed with a calcium-based process. Liquid
to gas ratios were said to be of the order of 100 gal/1000 ft$^3$,
and $SO_2$ removal efficiencies as a function of liquid to gas
ratio were reportedly typical of data reported by other investi-
gators. Scaling caused by precipitation of calcium salts on the
electrodes was viewed as the major problem area to be overcome
for power plant applications.

Mikropul's objective is to produce a single "box" for removing
both $SO_2$ and particulate. The pilot plant design essentially
consists of a scrubber, a particle precharger, a discharge
electrode and parallel plate collector, followed by an electro-
static demister section which can be periodically washed. In
addition to the calcium-based process, a sodium-based process

8 mm                                    8 mm



4 mm diameter
stainless steel
studs attached
by either
welding or
drilling and
welding

4 mm

1st Stage Discharge          2nd Stage Discharge
      Electrode                    Electrode
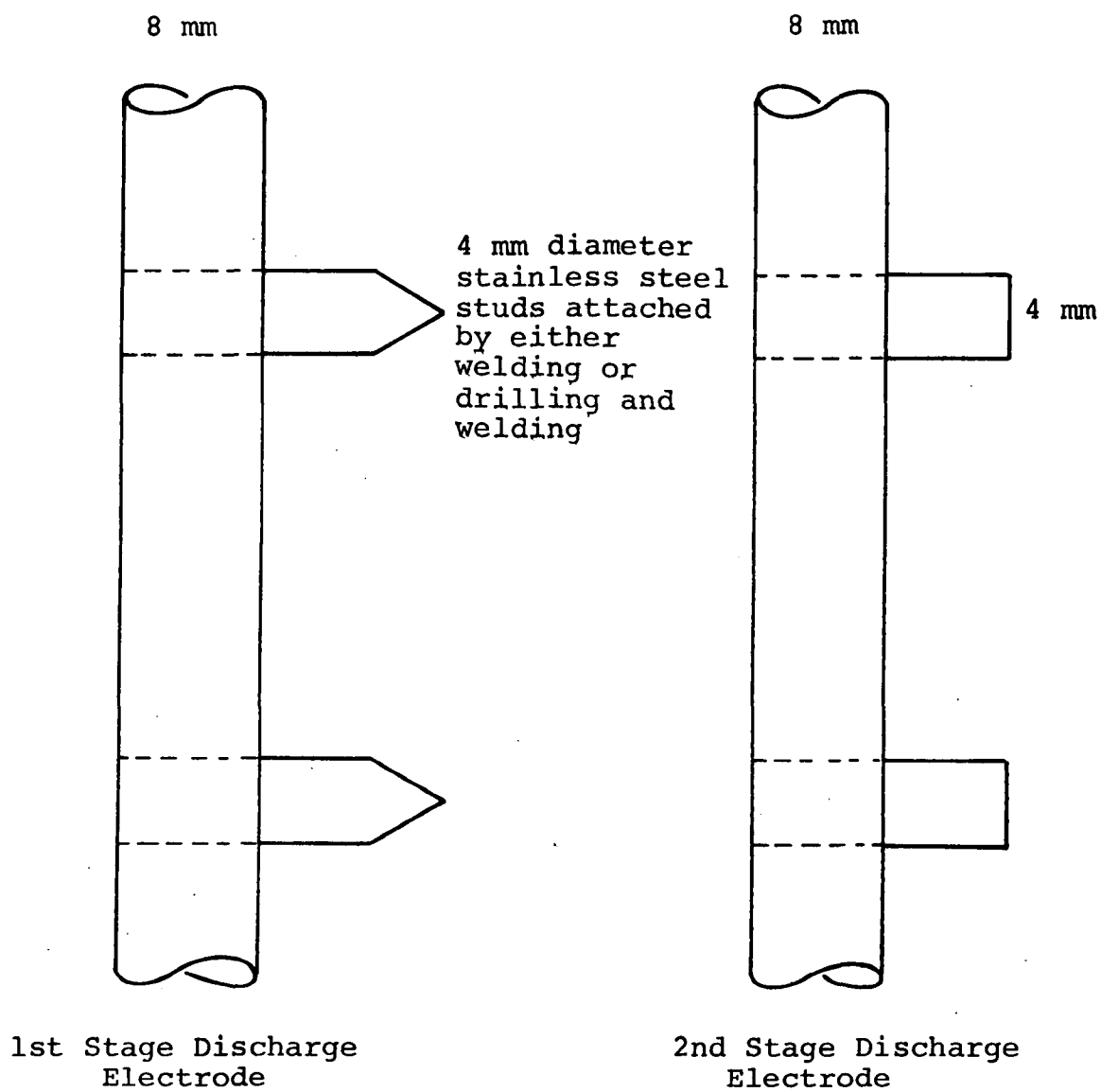

Figure 23.  Lurgi Discharge Electrodes

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is also under consideration for use with the wet precipitator system.

Mikropul reported that they have obtained pilot plant data from the following application areas:

Tire-cord curing ovens
Glass melting furnaces
Fiberglass forming lines - curing ovens, cooling section
Coke ovens
Sinter plant
Anode baking furnaces
Secondary brass melting furnaces
Phosphorus reduction furnace
Borax dust.

Table 7 gives a list, furnished by Mikropul, of their wet precipitator installations.

## Research Cottrell

A telephone conversation held with a staff member of Research Cottrell concerning their current interests in the wet precipitator market. He reported that his company has no recently constructed wet precipitator installations. Most of their wet units have been used on blast furnaces, and many of these are being replaced by high energy scrubbers.

## Western Precipitator Division of Joy Manufacturing Company

Discussions were held with representatives of Western to determine their opinions concerning the future application of wet precipitators. Western manufactures wire-pipe wet ESP's, typically with 96 ten-inch diameter, 15 ft long pipes in each module. These modules have recently been used on scarfing machines, with each unit treating about 25,000 cfm. No test data was available for these units.

Although Western has considered the use of wire-pipe wet ESP's for incorporation into an $SO_2$ removal system, they do not consider it practical, because of corrosion and resultant high materials cost, to use a wet ESP as the primary particle remover. In trials of a scrubber-wet precipitator system that have been conducted by Western, the wet ESP was said to function as a mist eliminator following a low energy scrubber. Western recommends "hot" precipitators for particulate collecting in coal-fired power plants.

## Blast Furnace Wet Precipitator Installation

A visit was made to a steel mill which uses wire and tube-type

## TABLE 7

### MIKROPUL WET ELECTROSTATIC PRECIPITATOR INSTALLATIONS

| Location | Description | Capacity, cfm | Installation Date |
|---|---|---|---|
| **Aluminum Potlines** | | | |
| Martin Marietta Goldendale, Wash. | 20 units | 7,500 cfm | 1971 |
| Martin Marietta The Dalles, Oregon | 7 units 4 units | 12,000 cfm 6,000 cfm | 1972 |
| Reynolds Metals Co. Longview, Wash. | pilot unit | 50,000 cfm | Oct. 1971 |
| Phase I | 1 unit 4 units | 50,000 cfm 100,000 cfm | June 1973 |
| Phase II | 4 units 12 units | 50,000 cfm 100,000 cfm | |
| Phase III | 10 units | 100,000 cfm | 1974 |
| **Fiberglas Forming Lines** | | | |
| Certain-Teed Prod. Kansas City, Kansas | 2 units | 100,000 cfm | 1973 |
| **Carbon Anode Baking Furnace** | | | |
| Airco Speer Carbon Graphite Niagara Falls, N.Y. | 4 units | 9,000 cfm | 1973 |
| Reynolds Metals Co. | 1 unit | 26,500 cfm | 1973 |
| **Oil Mist** | | | |
| General Motors Corp. Buick Division Flint, Michigan | 1 unit | 15,000 cfm | 1972 |
| **Sodium Sulfite Mist** | | | |
| Shattuck Chemical Co. Denver, Colorado | 1 unit | 3,500 cfm | 1972 |
| **Phosphorus Rock Dust** | | | |
| W.R. Grace & Co. Bartow, Florida | 2 units | 115,000 cfm | 1974 |
| **Coke Oven Emissions** | | | |
| Dominion Foundry & Steel Hamilton, Ontario | 2 units | 150,000 cfm | 1974-1975 |

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wet precipitators for cleaning blast furnace gas. Two precipitators are employed, each of which has two units. Each unit consists of 126 pipes 8 in. in diameter and 15 ft long, giving a total of 504 tubes with a collecting area of 15833 ft$^2$. These units were installed in the 1941-1942 time period on a large blast furnace. They are now being used on a smaller blast furnace, and a high energy venturi has replaced the wet precipitators on the large blast furnace. The current plan is to phase out the precipitators in about two years, and use high energy scrubbers exclusively.

Performance tests on the wet precipitators were conducted by Research Cottrell in April of 1960, and Table 8 gives the data reported to the user. At the time these tests were conducted, the units were powered by mechanical rectifiers. These have since been replaced with two tube-type rectifiers, each of which powers one precipitator. During our visit, the panel meters indicated a total DC current of 0.5 amps, which would result in an average current density of about 30 μa/ft$^2$. The precipitators are preceded by wet scrubbers which cool the furnace gas and remove the relatively large particles. The gas flow and efficiency data in Table 8 result in precipitation rate parameters between 9.5 and 11.5 cm/sec. These values are approximately within the range of design values reported by Oglesby and Nichols[4] for blast furnace wet precipitators.

Although the management at the blast furnace wet ESP installation that we visited was planning to phase out the units, their performance was said in general to be satisfactory. The tubes require cleaning once or twice a year, and the precipitators have been retubed at least once since their original installation. The general reasoning seemed to be that operating a blast furnace at relatively high pressure was advantageous, and this pressure could be effectively utilized in a scrubber for particulate removal.

TABLE 8

TEST RESULTS OBTAINED BY RESEARCH COTTRELL ON A BLAST FURNACE
WET PRECIPITATOR INSTALLATION

| Test No. | *1 | *2 | *3 | **4 | **5 |
|---|---|---|---|---|---|
| Date | 4/22/60 | 4/25/60 | 4/25/60 | 4/26/60 | 4/26/60 |
| Furnace Wind, cfm | 73,000 | 71,000 | 71,000 | 64,000 | 64,000 |
| Blast Pressure, psig | 24.0 | 22.0 | 22.0 | 19.0 | 18.5 |
| Top Gas Pressure, "$H_2O$ | 40 | 40 | 30 | 43 | 44 |
| Hot Blast Temperature, °F | 1050 | 1050 | 1050 | 950 | 890 |
| Precipitator Inlet Gas Temp.,°F | 88 | 90 | 90 | 90 | 90 |
| Gas Volume, cfm | 99,000 | 103,500 | 103,500 | 86,000 | 86,000 |
| Precipitator Rating, % | 90 | 94 | 94 | 156 | 156 |
| Dust Concentrations, Grains per cu. ft. Dry Gas @ 60°F & 30" Hg. | | | | | |
| Inlet | 0.133 | 0.0915 | 0.128 | 0.0675 | 0.0698 |
| Outlet | *** | 0.0051 | 0.00705 | 0.00861 | 0.0103 |
| Precipitator Efficiency, % | *** | 94.4 | 95.0 | 87.2 | 85.3 |

*Tests made with gas from No. 2 Furnace passing through both precipitators.
**Tests made with gas from No. 2 Furnace passing through one precipitator.
***Contaminated outlet sample was discarded.

# SECTION V
## ANALYSIS OF POTENTIAL WET ESP APPLICATIONS

## SIMULTANEOUS REMOVAL OF PARTICULATE AND SULFUR OXIDES

The recent emphasis on reduction of sulfur oxide emissions
has resulted in a number of processes which involve wet
scrubbing of flue gas to remove both particulate and sulfur
oxides. High efficiency particulate removal in scrubbers
generally requires pressure drops of 25 to 50 cm of water or
more, which for utility power plants involves a considerable
energy loss. The removal of sulfur oxides, however, can be
accomplished with less energy in a scrubber which is designed
primarily for gas absorption. The electrostatic precipitation
process can collect particulate with less energy expenditure
than with wet scrubbers, but the sensitivity of dry precip-
itators to fuel composition changes and resultant dust resis-
tivity variations is a serious drawback. It is therefore of
interest to consider the use of a wet precipitator, which is
not sensitive to dust resistivity, as the primary particulate
collection device for small particle removal.

Figure 24 illustrates possible configurations of a combination
scrubber-wet precipitator. As indicated in the illustration,
the device may consist of a separate low energy scrubber and
wet-wall precipitator, or the scrubber and precipitator can
be combined into one unit with alternating scrubbing and
collecting plate sections. In either arrangement, it is ob-
vious that simultaneous particulate collection and gas absorption
will occur in the precipitator sections. Consideration will
therefore be given to the potential mechanisms of sulfur oxide
removal which exist in the electrical sections. These mechanisms
are:

1. Migration of ionized $SO_2$ molecules to the collecting
   electrode under the influence of an electric field.

2. Diffusion and absorption of $SO_2$ into a basic solution
   irrigating the collecting electrodes.

3. Absorption of $SO_2$ by droplets of basic solution which
   may be sprayed into the interelectrode region.

An estimation of the maximum amount of $SO_2$ which might be removed
by electrical transport can be calculated by assuming a current
density and a plate area to volume flow ratio (specific collecting
area). As an example, we will consider a wet precipitator with a
specific collecting area of 59.1 $m^2/(m^3/sec)$ (300 $ft^2/1000$ cfm),
which operates with a current density of 100 $nA/cm^2$. The corona

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ABSORBING SOLUTION

```
                    ┌──────────────┐      ┌──────────────┐      ┌──────────────────┐
                    │ PRE-         │      │              │      │                  │
FLUE GAS   ────────▶│ CONDITIONING │ ───▶ │ LOW ENERGY   │ ───▶ │ WET-WALL         │ ──▶ FLUE GAS
INLET               │ CHAMBER      │      │ SCRUBBER     │      │ PRECIPITATOR     │     TO STACK
                    └──────────────┘      └──────────────┘      └──────────────────┘
```

SPENT ABSORBENT AND FLY ASH
TO DISPOSAL OR REGENERATION SYSTEM

OPTION A

ABSORBING SOLUTION

```
                ┌──────────────┐   ┌───────┬──────────┬───────┬──────────┬───────┬──────────┐
                │ PRE-         │   │ SPRAY │COLLECTING│ SPRAY │COLLECTING│ SPRAY │COLLECTING│
FLUE GAS ──────▶│ CONDITIONING │──▶│VOLUME │ PLATES   │VOLUME │ PLATES   │VOLUME │ PLATES   │──▶ FLUE GAS
INLET           │ CHAMBER      │   │       │          │       │          │       │          │   TO STACK
                └──────────────┘   └───────┴──────────┴───────┴──────────┴───────┴──────────┘
```

SPENT ABSORBENT AND FLY ASH
TO DISPOSAL OR REGENERATION SYSTEM

OPTION B

Figure 24.  Schematic Diagram of Scrubber-Wet
Precipitator System.

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current density is given by

$$100 \times 10^{-9} \ \frac{amp}{cm^2} \times 10^4 \ \frac{cm^2}{m^2} \times 1 \ \frac{coul}{amp-sec} = 10^{-3} \frac{coul}{m^2-sec}$$

If we assume that the current is carried by singly charged ionized gas molecules, the rate at which such molecules are transported to the collecting electrode is

$$10^{-3} \ \frac{coul}{m^2-sec} \times \frac{1 \ ion}{1.6 \ x10^{-19} coul} = \frac{6.25 \times 10^{15} \ ions}{m^2-sec}$$

From the specific collecting area and the removal rate of ions, it can be seen that the number of ions removed per cubic meter of gas is given by

$$\frac{6.25 \times 10^{15} ions}{m^2-sec} \times \frac{59.1 \ m^2}{m^3/sec} = 3.69 \times 10^{17} \ \frac{ions}{m^3}$$

At 120°F, it can be shown that the number of gas molecules in a cubic meter of gas is equal to $2.28 \times 10^{25}$. Thus, the fraction of the total number of gas molecules which could be electrically transported to the collecting electrodes is equal to

$$\frac{3.69 \times 10^{17} \ ions/m^3}{2.28 \times 10^{25} \ molecules/m^3} = 1.62 \times 10^{-8}, \ or \ 0.0162 \ ppm$$

It is therefore apparent that the corona current cannot be expected to directly contribute significantly to the amount of sulfur oxides which might be removed in a wet precipitator.

The second potential mechanism is absorption of $SO_2$ into a basic solution irrigating the collection electrode as a result of the reduced partial pressure of $SO_2$ above such a solution. An estimate of the gas phase mass transfer coefficient applicable to

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these conditions may be obtained from the Sherwood equation,[5] which was obtained for wetted wall columns with the fluid descending in laminar flow and the gas ascending in well developed turbulent flow. The equation is

$$\frac{K_g d}{D_m} = 0.023 \left(\frac{du\rho}{\mu}\right)^{0.83} \left(\frac{\mu}{\rho D_v}\right)^{0.33}$$

where

$$
\begin{aligned}
d &= \text{equivalent diameter of gas passage} \\
D_m, D_v &= \text{molal and volumetric diffusivity of gas being} \\
&\quad \text{absorbed } (SO_2) \\
u &= \text{gas velocity} \\
\rho &= \text{density of gas phase} \\
\mu &= \text{viscosity of gas phase} \\
K_g &= \text{gas film coefficient}
\end{aligned}
$$

For a precipitator with 25.4 cm plate spacing and a gas velocity of 1.52 m/sec (5 ft/sec), the above correlation gives $K_g$ as 0.183 g mole/[m² (sec) (mole frac.)]. If we make the assumptions that the gas phase resistance is the rate controlling step and that the partial pressure of $SO_2$ above the solution is negligible, the absorption rate may be estimated from

$$N = K_g (A) (Y_{ln})$$

where

$$
\begin{aligned}
N &= \text{moles/sec of } SO_2 \text{ removed} \\
A &= \text{absorption area} \\
Y_{ln} &= \text{log mean mole fraction of } SO_2 \text{ in gas phase} \\
&= (Y_{in} - Y_{out})/\ln\left(\frac{Y_{in}}{Y_{out}}\right)
\end{aligned}
$$

For the previous calculation of the amount of $SO_2$ which could be electrically transported, we assumed an A/V ratio of 59.1 m²/(m³/sec). If we further assume a log mean $SO_2$ concentration of 1000 ppm, the $SO_2$ removal rate for 59.1 m² is given by

$$N = 0.183 \ \frac{\text{gmole}}{(m^2)(\text{sec})(\text{mole frac.})} \ \times \ 59.1 \ m^2 \times 1 \times 10^{-3} \ \text{mole frac.}$$

$$= 0.0108 \ \text{gmole/sec}$$

The total gas flow passing each 59.1 $m^2$ of plate area is 1 $m^3$/sec, or 37.9 gmoles/sec at 120°F. The quantity of $SO_2$ removed, expressed as a fraction of the total gas, is therefore

$$\frac{0.0108 \ \text{gmole} \ (SO_2)/\text{sec}}{37.9 \ \text{gmole} \ (\text{gas})/\text{sec}} = 285 \times 10^{-6}$$

These calculations indicate that the solution irrigating the walls will provide sufficient interfacial area for only a minor portion of the required $SO_2$ removal, although it is possible that turbulence created by the electric wind in the interelectrode space may increase the value of $K_g$ to some extent. The principal $SO_2$ removal, however, must be accomplished by sprays in order to increase the interfacial areas.

The use of a wet ESP in an SOx-particulate removal system must therefore be decided upon the basis of the WESP's performance as a particulate collection device. Figures 25 and 26 give computed projection of particulate collection performance of a WESP on a typical power station fly ash. These projections were obtained from the SRI-EPA precipitator computer model,[6] and the voltage-current relationships are based upon the data contained in Figure 21. A gas velocity distribution with a standard deviation of 25% of the mean velocity was used in the model for both sets of indicated electrical conditions. Gas by-passage was assumed to be 5% over two and three stages. Figure 25 indicates the expected collection efficiency relationship using what we consider to be optimistic electrical operating conditions, whereas the efficiency relation-ships in Figure 26 are based upon the average electrical conditions which were obtained during the test program described in Section IV. These projections suggest that a specific collecting area of 39.37 $m^2/(m^3/\text{sec})$, or 200 $ft^2/1000$ cfm, would constitute a con-servative estimate of the collection area required for achieving 99.6% or better collection efficiency of particulate.

Although in principal $SO_2$ removal can be accomplished by intro-duction of spray droplets in the interelectrode region, it is our
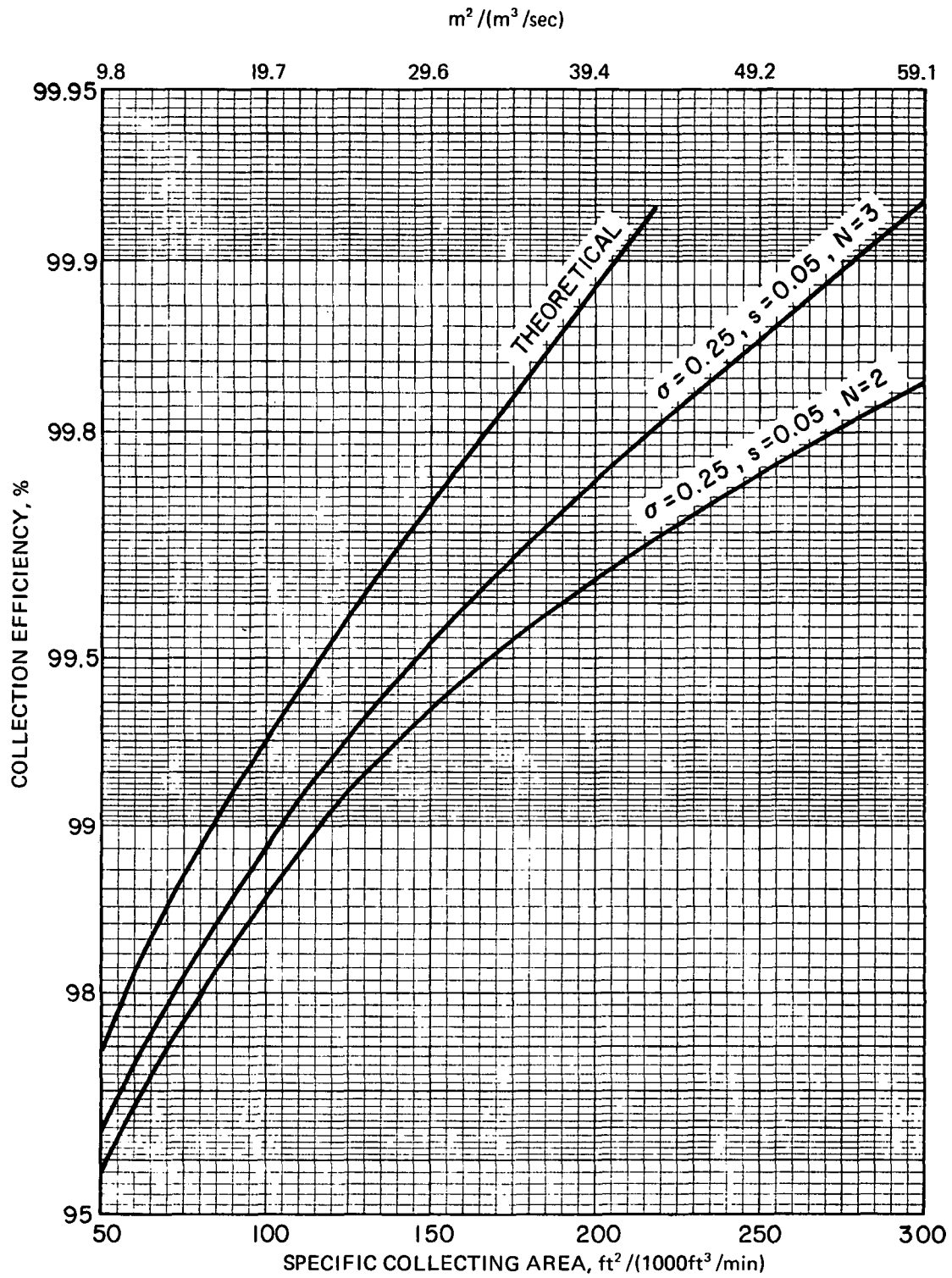
Figure 25. Collection Efficiencies Calculated With Computer Model for 50 x $10^{-9}$ amps/cm$^2$ and 50 KV

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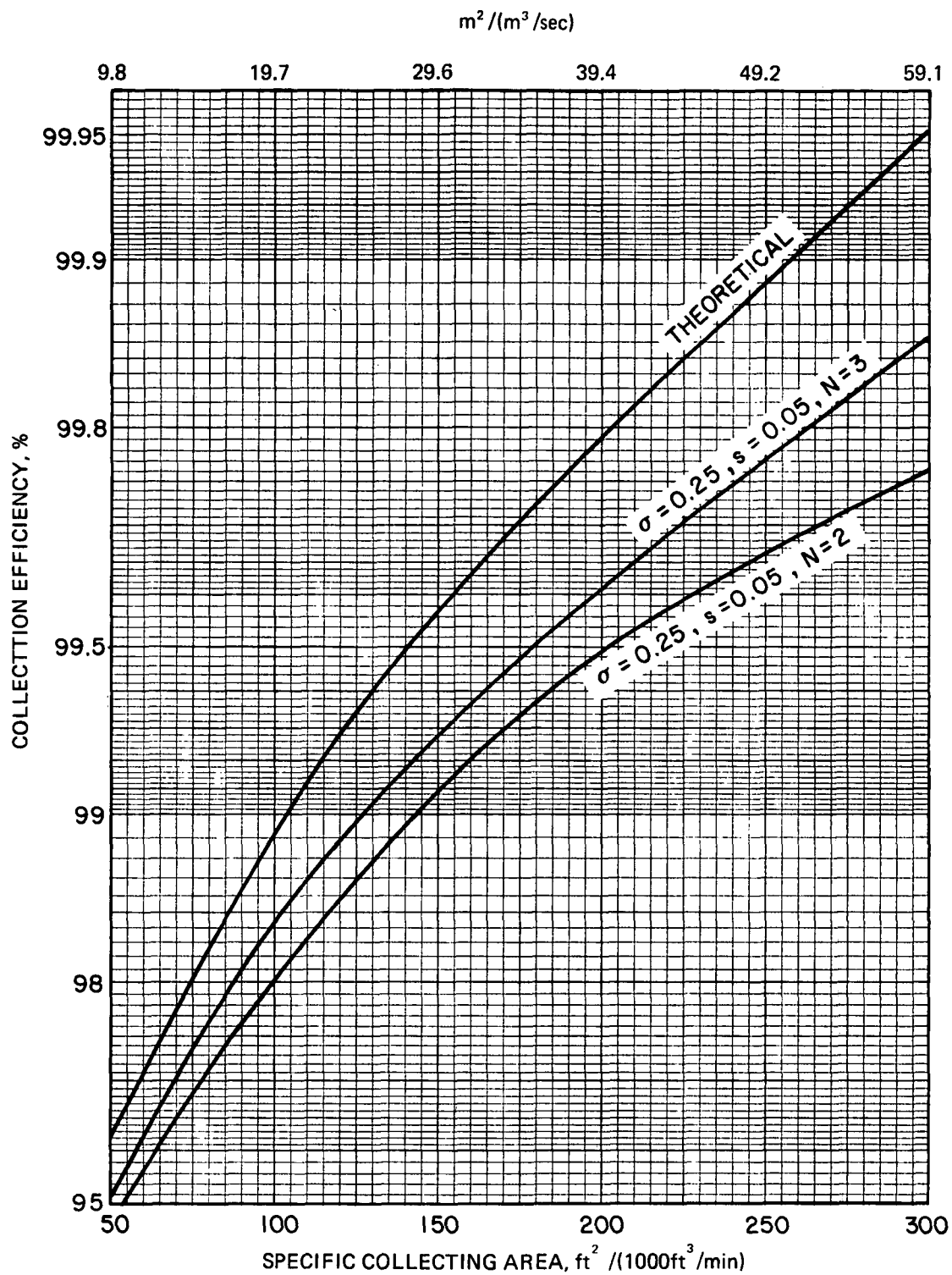
Figure 26.  Collection Efficiencies Calculated With
Computer Model for 33 x $10^{-9}$ amps/cm$^2$ and
47.3 KV

opinion that introduction of the sprays into the electrified
region in the required L/G ratios would limit electrical operating
conditions for full-scale units with practical values of plate
area for each T-R set. Therefore, $SO_2$ removal should be accom-
plished in a relatively inexpensive spray tower preceding a wet
ESP. Perhaps the most favorable circumstances for use of a WESP
would be in a sodium-based absorption system in which $SO_2$ removal
can be accomplished with moderate L/G ratios.

Table 9 gives a comparison of power requirements for two particulate
control methods in an SOx-particulate removal gas cleaning system.
Scrubber operating assumptions are based on data obtained at the
EPA Alkali Scrubbing Test Facility.[7] Based on the indicated assump-
tions, the WESP system has lower power requirements, primarily as
a result of the low pressure drop through the system. Table 10
and 11 give capital and operating cost estimates. The WESP capital
costs are based on the assumption of stainless steel construction
with installed costs of $377/$m^2$, or $35/$ft^2$, and a specific col-
lecting area of 39.37 $m^2/(m^3/sec)$. These comparisons indicate that
the WESP costs for particulate removal are about two times as
great as the scrubber due to the high capital costs. The wet ESP
system costs would have to be reduced by about 60% in order to
have the same annual operating costs as the TCA scrubber system.
Maintenance costs have not been considered quantitatively in this
analysis, but the possibility of scale formation on the electrodes
in the WESP is potentially a much more serious problem for the WESP
than is scale formation in a scrubber vessel.

Another consideration which is not reflected in these comparisons
is the performance of a dry ESP. Collection efficiencies of dry
ESP's under favorable operating conditions with high sulfur coals
have been measured as 99.7% with specific collecting areas less
than 49.2 $m^2/(m^3/sec)$, or 250 $ft^2/1000$ cfm. Therefore, unless
WESP's can be demonstrated to achieve 99.5% or better collection
efficiency with SCA's considerably less than 39.4 $m^2/(m^3/sec)$,
or 200 $ft^2/1000$ cfm, there is no economic justification for sub-
stituting a WESP for a conventional dry electrostatic precipitator
in an SOx-particulate removal system with flue gas produced from
high sulfur coals. For such instances, an economic analysis should
be conducted which compares the operating costs of a scrubber-only
system with those of a system consisting of a low energy scrubber
and a dry ESP.

COLLECTION OF HIGH RESISTIVITY DUST

Utilities in some parts of the United States are faced with re-
quirements for SOx removal from flue gas in which the dust resis-
tivity is sufficiently high to limit the electrical operating
conditions in the precipitator. Under these conditions, the
plate area required for 99.5% or greater particulate collection

## TABLE 9

### POWER REQUIREMENT COMPARISON FOR PARTICULATE CONTROL IN SOx REMOVAL SYSTEMS

| | kw/kacfm | |
| --- | --- | --- |
| Item | Scrubber-Wet ESP | Scrubber Only (TCA) |
| Pump | 1.080 | 2.03 |
| Fan | 0.468 | 2.30 |
| Power for TR Sets | 0.405 | |
| Insulator Power | 0.180 | |
| Total | 2.133 | 4.33 |

Assumptions:

1. Fan and pump efficiency 50%.

2. 5.08 cm $H_2O$ $\Delta P$ in gas for scrubber-WESP system.

3. Fluid head for pumping calc. = 28.1 m.

4. L/G for scrubber-wet ESP System
   Scrubber (spray tower) = $3.4\ell/m^3$ (25.8 gal/1000 $ft^3$)
   WESP = $0.7\ell/m^3$ (5 gal/1000 $ft^3$)

5. L/G for scrubber only system = $7.8\ell/m^3$ (58.2 gal/1000 $ft^3$).

6. 24.9 cm $H_2O$ $\Delta P$ in gas for scrubber-only system.

7. WESP has SCA of 39.37 $m^2/$ ($m^3$/sec).

8. TCA particulate collection efficiency = 99.8%.

9. WESP particulate collection efficiency = 99.75%.

10. $SO_2$ removal = 90%.

## TABLE 10

### CAPITAL COST ESTIMATES FOR PARTICULATE CONTROL
### IN PARTICULATE-SOx REMOVAL SYSTEM

150 MW Boiler, 183 $m^3$/sec (394,000 acfm) at 50°C

| Item (a) | Scrubber-WESP | Scrubber Only (TCA) |
|---|---|---|
| Wet ESP | 2,760,000 | |
| Spray Scrubber | 105,000[10] | 85,000 (c) |
| TCA Vessel | | 608,000[9] |
| Scrubber Fans & Motors | | 255,000 |
| Total | 2,865,000 (b) | 948,000 (b) |

(a)  Elements common to both systems not included.

(b)  Cost scaled to February 1975 dollars with chemical plant cost index.

(c)  Presaturator.

TABLE 11

OPERATING COST ESTIMATES FOR PARTICULATE CONTROL
IN PARTICULATE-SOx REMOVAL SYSTEM, $/yr

| Item (a) | Scrubber WESP | Scrubber Only |
|---|---|---|
| Cap charges at 15%/yr | $430,000 | $142,200 |
| Energy cost (b) | 40,300 | 81,900 |
| | $470,300 | $224,100 |

(a)  Elements common to both systems not included.

(b)  Based on 6000 hr/yr $0.008/kwh

efficiency can be 98.4 $m^2/(m^3/sec)$, [500 $ft^2$/1000 cfm], or greater if a conventional ESP is used at operating temperature in the 150°C range. The use of a WESP could lower the plate area requirements by a factor of two or more, but presently available information indicates that the increased materials costs required by wet operation would partially or completely offset the cost savings resulting from plate area reduction. The possibility of electrode scaling in WESP operation is a factor which would require pilot plant trials for an adequate evaluation.

COLLECTION OF MISTS FROM SCRUBBER SYSTEMS

The laboratory studies described in Section III indicate that the mathematical model should provide a basis for evaluating the potential effectiveness of WESP's for collecting mists from scrubber systems if a size distribution of the mist is available. Table 12 gives a size distribution, provided by EPA, of droplets at a spray scrubber exit.[8] The total droplet concentration is 11000 grams of liquid per cubic meter of gas, which is equivalent to about 80 gal/1000 $ft^3$. This extremely high concentration of large droplets causes two problems in using the model to project overall mass collection performance:

1. The model is based on the Deutsch equation for individual particle sizes, and the Deutsch collection mechanism does not apply to large particles (7-10 µm diam.).

2. The high droplet concentration results in high values of particulate space charge, and the effects of this space charge are not adequately represented in the model.

In view of these difficulties, results obtained from the model with this size distribution should be considered as only very rough approximations. Figure 27 shows overall collection efficiency projected by the model with the size distribution of Table 12. Wire-plate geometry was assumed, with a current density of $33 \times 10^{-9}$ amps/$cm^2$ and an applied voltage of 47.3 kv. Model correction parameters for the overall mass efficiency were:

1. A gas velocity distribution with a standard deviation of 25% of the mean.

2. Gas by-passage assumed to be equivalent to 5% over two stages.

Figure 28 shows the theoretical collection efficiencies for the same current density and applied voltage, but with space charge effects due to large droplets ignored, as a function of particle diameter.

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TABLE 12

SIZE DISTRIBUTION OF DROPLETS AT A SPRAY SCRUBBER EXIT

| Average Droplet Diameter, cm | Cumulative Mass Loading, grams/m$^3$ |
|---|---|
| 1.5 | 1.6 |
| 2.6 | 8.8 |
| 6.6 | 26 |
| 18 | 590 |
| 36 | 1,400 |
| 100 | 11,000 |
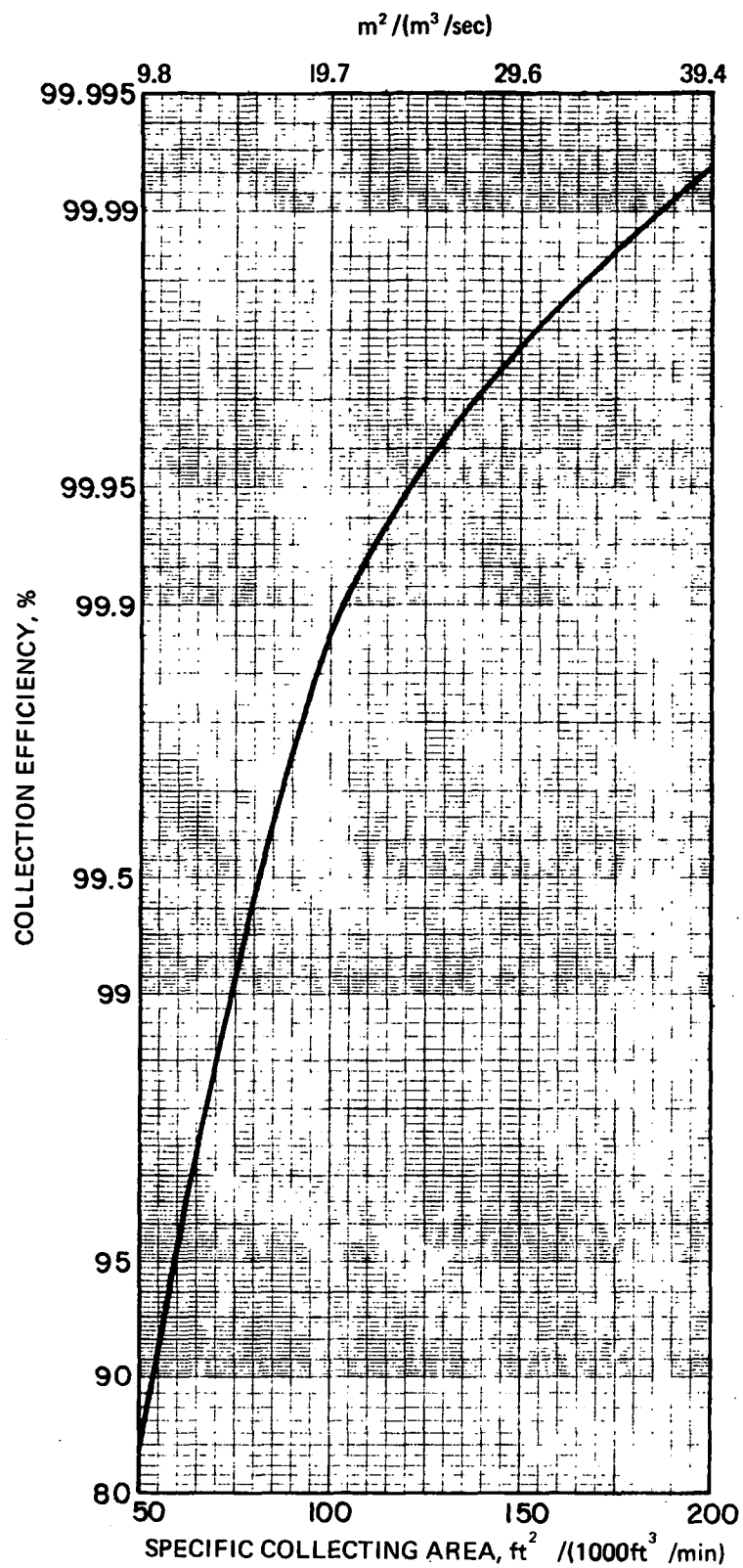
Figure 27.   Collection Efficiency Calculated With
            Droplet Size Distribution for 33 x 10⁻⁹
            amps/cm², 47.3 KV


σ = 0.25, S = 0.05, N = 2

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These results indicate that a WESP would require about 39.4 m$^2$/ (m$^3$/sec) to achieve a theoretical collection efficiency of about 99.5% on 2.5 μm diameter particles (again ignoring space charge due to the large drops), even though the overall mass collection efficiency at this SCA is projected to exceed 99.99%. If the large droplet space charge is allowed to retard collection efficiency using an approximate method for reducing free ion densities, the theoretical collection efficiencies at 39.4 m$^2$/(m$^3$/sec) for the droplet size bands are reduced as shown in Figure 28. The overall mass collection efficiency of 99.992% indicated in Figure 27 would result in an outlet mass loading of 0.88 grams/m$^3$, which would reduce to ~0.09 grams/m$^3$ if 10% of the droplets consisted of non-volatile solids. Thus, the model projections indicate that a WESP would require an excessive plate area to collect the distribution given in Table 12 with an efficiency such that the outlet loading from the WESP would be less than .0458 g/m$^3$ (.02 gr/acf). This is a result of the excessive inlet loading of 11,000 grams/m$^3$.

For other size distributions, it is recommended that the theoretical fractional efficiencies curves of Figure 28 be used to obtain an estimate of overall collection efficiency. These curves are based on wire-plate geometry and are valid only if particulate space charge is not sufficient to disrupt the electrical conditions. EPA is currently supporting studies to better define conditions under which space charge suppression of corona current will occur.
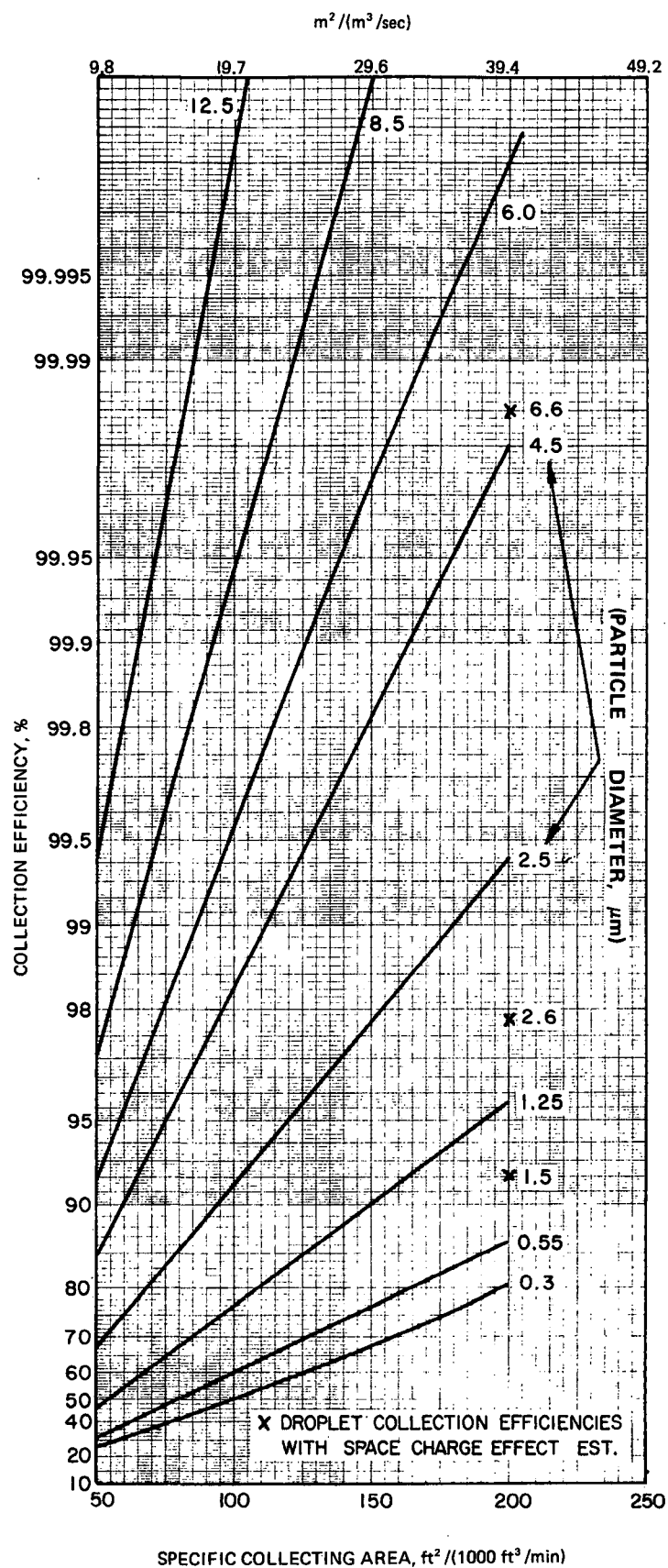
Figure 28. Theoretical Collection Efficiencies
as a Function of Particle Diameter
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# SECTION VI
## RECOMMENDED RESEARCH AND DEVELOPMENT

The laboratory studies conducted under this contract have shown that wet electrostatic precipitators can be effective collectors of fine particulate under idealized conditions. The effectiveness of WESP's used under field conditions is below the potential performance due to (1) space charge effects, which limit allowable current density, caused by incoming particulate and the electrode irrigation system, (2) deposits forming on the electrodes which cause sparkover at low applied voltage levels, (3) corrosion problems peculiar to the application area and the chosen materials of construction. Recommended research and development programs for these problem areas are given below.

| PROBLEM | SUGGESTED EFFORT |
|---|---|
| 1. Space charge limitation of current | a) Develop more accurate methods of calculating the effects of space on collection efficiency. |
| | b) Conduct laboratory studies to better define conditions under which corona quenching will occur when high concentrations of fine particulate are present. |
| | c) Use theoretical calculations and available data, and attempt to determine relationship between collection efficiency and sectionalization under space charge conditions. Also investigate effect of electrode geometry. Objective is to optimize fine particulate collection. |

Total Effort — 2.0 man-years

2. Deposit formation

    a) Survey types of electrode irrigation systems used by different manufacturers in each application area.

    b) Determine location and nature of deposit formation problems.

    c) Develop recommendations for implementing electrode irrigation in major application areas.

Total Effort - 1.5 man-years

3. Corrosion

    a) Survey corrosion problems in each application area.

    b) Establish causes of corrosion and recommend materials of construction for the various existing and potential applications.

Total Effort - 1.0 man-year

With regard to the possible use of a WESP in an $SO_x$-particulate removal system, the previous section has shown that, with existing WESP technology, a scrubber system appears to be more economical than a WESP-scrubber combination. In view of the potential energy saving associated with the use of the electrostatic precipitation process in the collection of particulate, there is some justification for exploring means of reducing the capital cost of a WESP for power plant applications. At present, a WESP appears to have potential value for such an application only when $SO_x$ removal must be accomplished in the presence of high resistivity dust which results in excessive dry ESP plate areas. The following research and development efforts are suggested.

1) Determine which $SO_x$ removal systems are feasible for use in power plants burning relatively low S Western coals (0.5 man-year).

2) Determine whether the candidate processes have the capability of providing a liquor stream which is likely to be suitable for plate irrigation (0.5 man-year).

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3) If the above item gives a positive result, design a pilot-plant WESP test program for the power plant application to examine the scaling problem and to study performance under various conditions (0.25 man-year). The pilot program would also explore means of reducing materials costs.

# SECTION VII
## LITERATURE STUDY


Wet electrostatic precipitators have been used in the metallurgical and chemical industries for many years. Information on the design and application of WESP's in the industries with the highest usage was accumulated by SRI under Contract CPA 22-69-73. This information has been updated and expanded where appropriate, and is presented in the following sub-sections.

## IRON AND STEEL INDUSTRY

### Introduction[11,12]

On the basis of existing evidence, the deliberate smelting of ore to produce iron began between 1350 B.C. and 1100 B.C. over a wide geographic area. In America an iron works was established in Virginia on the James River about 1619; this was destroyed in an Indian raid in 1622 and never rebuilt. The Hammersmith (now Saugus), Massachusetts iron works was begun in 1645 and was the first successful iron works in this country. Iron was then produced by the reduction of iron ore in charcoal furnaces. The blast furnace was introduced into England about 1500 A.D. Coke was first used as a blast-furnace fuel in England in 1619. About 200 years later - again in England - the principle of heating the air before it was blown into the furnace was introduced, hence the term "hot blast". The Bessemer steelmaking process eventually supplanted all techniques before it. Closely following the Bessemer process was the development of the regenerative-type furnace that, now known as the open-hearth furnace, became adapted to steel-making and evolved into the principal means for producing steel throughout the world. Although the Bessemer converter production exceeded that of the open-hearth until 1908, the latter grew rapidly until, in the 1950's, 90% of the steel produced in this country was by the open-hearth process. A third process, the electric furnance, is a relative newcomer to the field of steelmaking and is gradually finding more and more applications in the quantity production of quality steels. The most recently developed steelmaking method is a pneumatic process that involves blowing high-purity oxygen onto the surface of a bath of molten pig iron, a method known as the basic oxygen furnace (BOF), or LD process. Estimates for 1980 show about 25-30% of the steelmaking will be carried out in electric furnaces, 65-70% by the BOF process, and 5% by the open-hearth process. The Bessemer converter has essentially disappeared as a steelmaking method. A typical steel plant operation in which there are about 10 major areas requiring dust or fume control equipment is shown as a flow chart in Figure 29. Each of these areas will be discussed in detail in subsequent

IRON ORE          COKE          LIMESTONE          RAW MATERIAL
                                                   PREPARATION

                              SCRAP

FINES

SINTERING                                          PIG AND CAST IRON
                                                   PRODUCTION

BLAST FURNACE

                              CUPOLA
         PIG IRON

                              CAST IRON

STEEL SCRAP

              1. BASIC OXYGEN FURNACE              STEEL PRODUCTION
              2. OPEN HEARTH FURNACE
              3. ELECTRIC FURNACE
              4. BESSEMER CONVERTER

              STEEL INGOTS

              SCARFING                             SURFACE PREPARTION

              TO ROLLING MILL
              OR PURCHASER
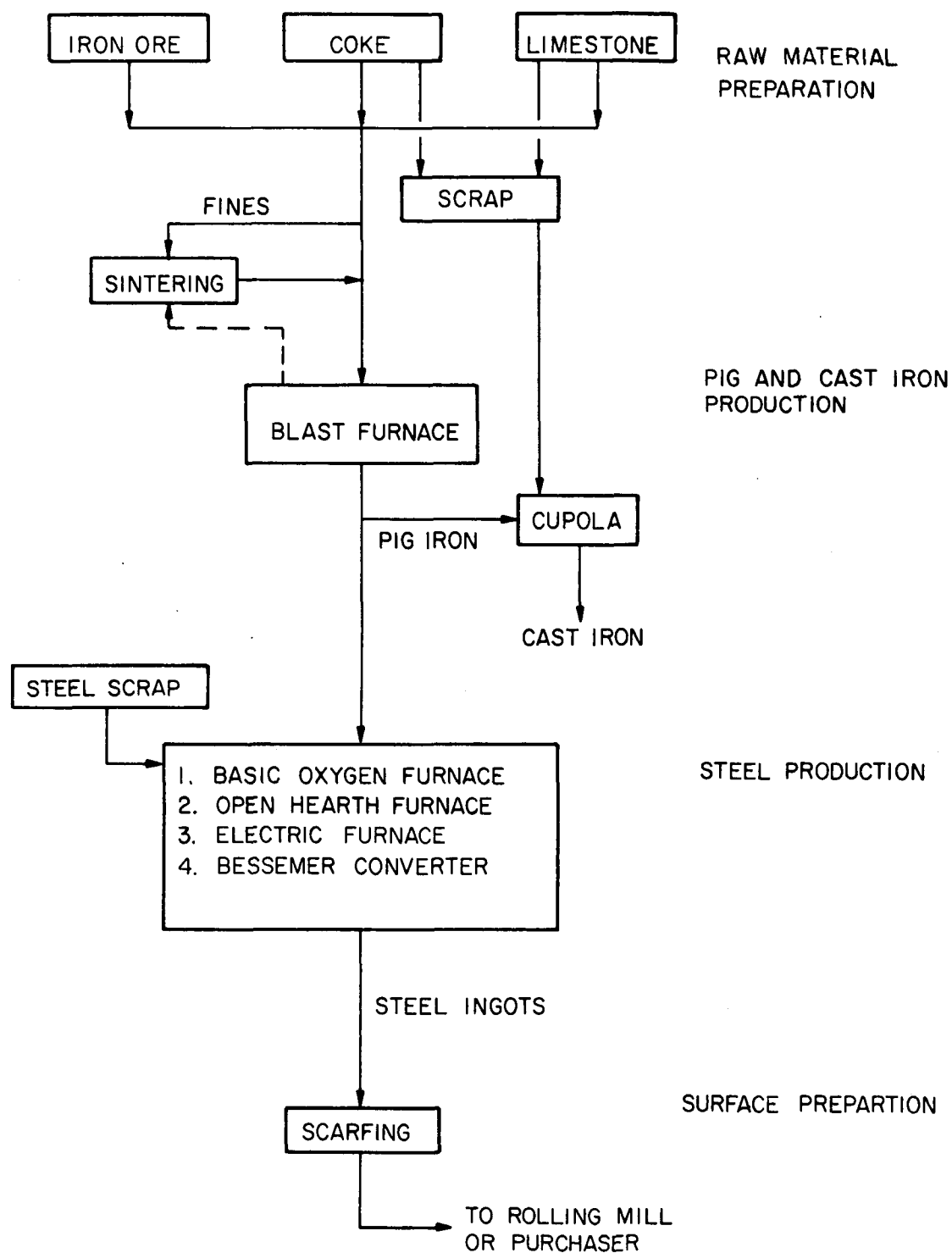
Figure 29.   A Flow Chart Showing the Steps in the Steelmaking
             Process From the Basic Raw Material and Scrap
             Input to the Finished Product.[4]

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sections with emphasis on process parameters relating to the use of wet electrostatic precipitators. Coke oven gas detarring will be discussed in a section describing collection of mists.

## Blast Furnace

The blast furnace is a cylindrically-shaped structure made of steel and lined with refractory brick, usually about 100 feet tall and 25-35 feet in diameter. Near the bottom are nozzles or "tuyeres" through which preheated air is blown into the furnace under pressure. The raw materials used in the production of pig iron are semi-continuously charged into the blast furnace. The charge consists of iron-bearing materials (iron ore, sinter, pellets, mill scale, open-hearth or Bessemer slag, iron or steel scrap, etc.), fuel (coke), and flux (limestone and/or dolomite), and is introduced into the top of the furnace through two successive cone-shaped valves, at least one of which is always closed. The shape and location of the lower valve (or bell) is such that the charge is deposited uniformly around the circumference of the furnace near the walls. As the charge melts, or is burned in the lower region of the furnace, the stock column gradually settles, thus leaving room for additional charges at the top. The reducing atmosphere necessary for reduction of iron ore is generated by a reaction between the precharged coke and preheated air (1000-1700°F). The air may be enriched with natural gas, fuel oil, or oxygen. As the blast rises through the burden, it reacts exothermically with the coke to produce the high temperature reducing gases which react with the iron oxide to produce molten pig iron. The reduction reactions can be summarized as a single reversible reaction as follows:

$$Fe_2O_3 + 3CO \rightleftharpoons 2Fe + 3CO_2$$
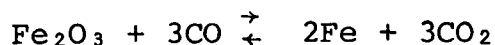
The presence of an excess of carbon monoxide keeps the equilibrium shifted to the right. The molten iron and slag collect in two layers at the bottom of the furnace, the less dense slag floating on the iron and protecting it against oxidation. The slag is occasionally withdrawn from the furnace, and several times a day, the molten iron is withdrawn through a tap hole into a large ladle lined with fire brick that holds several hundred tons of the hot metal.

The gases which pass through the charge leave the top of the furnace through four "off takes" arranged around the dome, and pass through a brick-lined downcomer to the dustcatcher, which is a large settling chamber. The coarser dust particles settle in the low gas velocity regions of the settling chamber.

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The particulate material emitted during normal operation of the
blast furnace originates from several sources, including dirt and
other fines in the charged ore, the dust caused by the downward
abrasive action of the burden through the furnace, and coke and
limestone dust. These particulates are carried out of the furnace
by the gas stream, and dust loadings range from 7 to 17 gr/scf.

A troublesome source of blast furnace emissions is that occurring
during a burden "slip". Slips are sudden movements of the burden
in the furnace occurring when a portion of the burden forms a
bridge within the furnace. The underlying burden continues to
melt and settles under the bridge and the gas pressure increases.
When the bridge opens or slips, a large volume of gas suddenly
passes through the furnace system and carries with it great
quantities of the lighter portions of the stock.

Two emissions come from the top of a blast furnace; top gas and
the dust which it entrains. After passing through the charge
the gases which leave the furnace at a temperature of around 300°F,
have a composition typically as follows:

| Constituent | Volume % |
|:-----------:|:--------:|
| CO | 26.2 |
| $CO_2$ | 13.0 |
| $H_2$ | 1.9 |
| $CH_4$ | 0.2 |
| $N_2$ | 58.7 |

The particulate material emitted during normal operation of the
blast furnace comes from several sources, including dirt and other
fines in the charged ore, the dust developed by the downward abra-
sive action of the burden through the furnace, and coke and lime-
stone dust.[13]

The output of particulate material from blast furnace operations
is primarily dependent on the physical characteristics of the charge.
For example, the burden can range from unscreened ore to that con-
sisting mainly of sinter, pellets, or mixtures of sinter and pellets.
Different combinations of burden, fuel, and air can be used to
effect different output characteristics. Sinter and pelletized
ore, when used as a charge material, naturally effect a marked re-
duction in dust product as compared to either screened or unscreened
ore, because of the improved physical characteristics. A good
illustration of this is shown in Figure 30, where the amount of
dust from a furnace is shown as a function of the amount of sinter
in the charge.[14]

One of the major reasons industry cleans blast furnace gas is to
render it sufficiently clean for heating coke ovens, boilers,
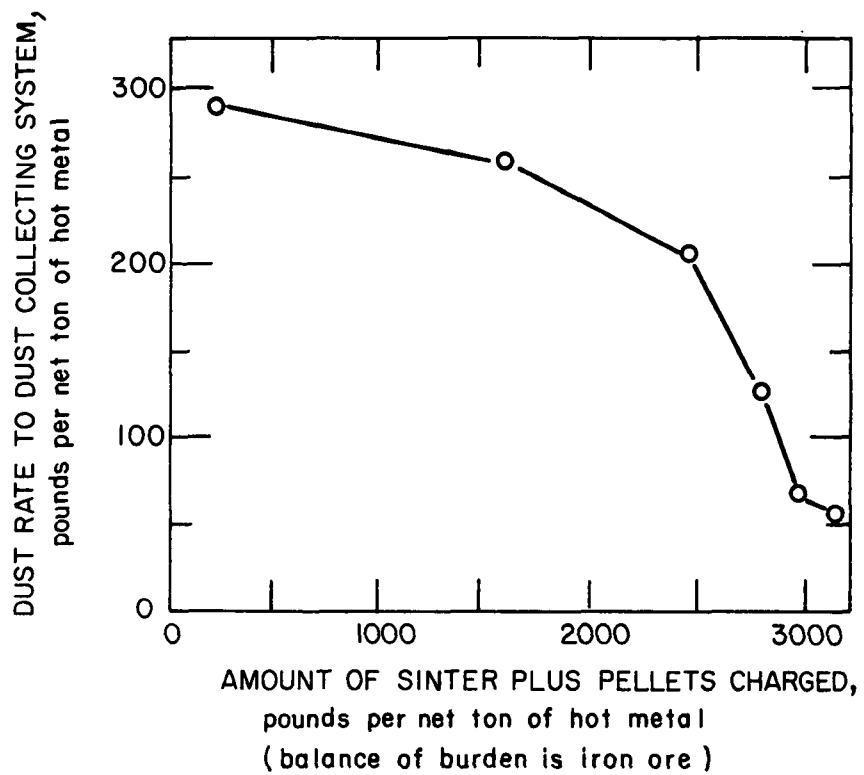
Figure 30.   Effect of Burden Improvement on Dust Rate From
A Blast Furnace to its Dust Collecting System[14]

stoves, soaking pits, and gas engines. The gas has a heating value of about 100 Btu/cf, but must be cleaned before being used successfully.

During normal blast furnace operations, the dust loading of the gas leaving the furnace is in the 7-10 gr/scf range although loadings as high as 17 gr/scf have been reported. For each 1000 tons of hot metal produced, about 100 tons of particulate is expected. From a 1400 ton-per-day furnace, Bishop reports 68% coarse flue dust and 32% fine.[15] The coarse dust to which he referred was removed by greatly reducing the velocity of the gas, and at the same time, suddenly changing the direction of the gas stream. The dust for the above mentioned plant contained around 30% iron, 15% carbon, 10% silicon dioxide, and small amounts of aluminum oxide, manganese oxide, calcium oxide, and other materials.

A typical chemical composition of blast furnace effluent is difficult to supply since the composition varies so greatly as shown in Table 13.[16] However, the major constituents and the usual composition range is as follows: 35-50% iron, 8-13% silica, 2-5% alumina, 3-4% calcium oxide, 3-10% carbon, and small amounts of alkali elements, phosphorus, zinc, lead, sulfur, and other trace elements. Some additional data on the chemical composition of dust samples taken from the dust catchers, wet scrubbers, electrostatic precipitators, and dust leaving the precipitator are given in Table 14.[17] The material collected in flue dust catchers and washers has a high iron content, and after agglomeration, becomes a suitable material for recharging. Precipitator dust contains less iron and several objectionable impurities. Particle size distributions of blast furnace dust also varies depending on the type of ore being charged into the furnace and prior beneficiation. Table 15 presents a range of observed particle size distributions.

As stated earlier, blast furnace gas must be cleaned prior to utilizing its heating value in order to prevent clogging of burners, gas mains, and other maintenance problems. In a typical installation, gas from a blast furnace is generally cleaned as follows. Upon leaving the furnace, gas is passed through a dust catcher which removes particles larger than a few hundred microns (50-70% by weight), leaving a dust concentration in the gas of 3-6 gr/scf. From the dust catcher, the gas is further cleaned in a two-stage plant. The first stage consists of a primary cleaner to separate the coarse fractions and a secondary cleaner to separate the fine dust. A typical combination of a two-stage cleaner consists of a wet scrubber and a wet electrostatic precipitator in series. Typical dust concentrations for the gas at various points in a modern gas-cleaning system are:

Table 13[16]

Chemical Analysis of Blast Furnace Flue Dusts

| Components | Weight Percent | | |
|---|---|---|---|
| | Range for Several U. S. Plants | Midwest Plant | Range for Several European Plants |
| Fe | 36.5 - 50.3 | 47.10 | 5.0 - 40.0 |
| FeO | n.a. | 11.87 | n.a. |
| SiO$_2$ | 8.9 - 13.4 | 8.17 | 9.0 - 30.0 |
| Al$_2$O$_3$ | 2.2 - 5.3 | 1.88 | 4.0 - 15.0 |
| MgO | 0.9 - 1.6 | 0.22 | 1.0 - 5.0 |
| CaO | 3.8 - 4.5 | 4.10 | 7.0 - 28.0 |
| Na$_2$O | n.a. | 0.24 | - |
| K$_2$O | n.a. | 1.01 | - |
| ZnO | n.a. | 0.60 | - |
| P | 0.1 - 0.2 | 0.03 | 0.3 - 1.2 |
| S | 0.2 - 0.4 | n.a. | - 0.1 |
| Mn | 0.5 - 0.9 | 0.70 | 0.3 - 1.5 |
| C | 3.7 - 13.9 | n.a. | 5.0 - 10.0 |
| Pb | n.a. | n.a. | 0 - 15.0 |
| Cu | n.a. | n.a. | trace |
| Alkali | - | - | 0 - 20.0 |
| Zn | n.a. | n.a. | 0 - 35.0 |

Table 14[17]

## Weight Percent Composition of
## Dust Samples from Blast Furnace Gas Cleaning Plant

| Sample Components | Deposit from primary dust catcher | Deposit from secondary dust catcher | Deposit from washers | Dust in gas leaving washers | Deposit from precipitator | Dust in gas leaving precipitator | Deposit in power station burners |
|---|---|---|---|---|---|---|---|
| Insol. | 8.50 | 11.60 | 9.40 | 13.68 | 22.60 | 11.44 | |
| $Fe_2O_3$ | 75.43 | 53.00 | 61.30 | 14.40 | 19.30 | 8.64 | 3.52 |
| $Al_2O_3$ | 1.80 | 1.55 | 4.35 | 6.91 | 15.58 | 6.70 | |
| MnO | 0.62 | 0.60 | 0.63 | 2.16 | 1.10 | 1.49 | |
| CaO | 1.60 | 2.40 | 3.56 | 6.36 | 7.12 | 4.68 | 2.80 |
| MgO | 0.80 | 1.09 | 1.67 | 8.04 | 9.90 | 9.28 | 3.04 |
| P | 0.24 | 0.20 | 0.41 | 0.54 | 0.71 | 0.57 | 0.14 |
| S | - | - | 0.41 | 1.89 | 1.33 | 1.37 | 0.36 |
| Cl | trace | trace | nil | 1.32 | 0.20 | 3.44 | 29.22 |
| Zn | nil | nil | nil | 1.20 | 0.90 | 1.20 | 0.70 |
| $Na_2O$ | 0.28 | 0.32 | 0.25 | 9.25 | 1.55 | 9.75 | 5.80 |
| $K_2O$ | 0.56 | 0.92 | 0.55 | 20.90 | 2.70 | 22.40 | 36.15 |
| Loss on ignition | 5.90 | 23.50 | 15.65 | - | 15.95 | - | 9.06 |

Table 15[17]

Size Analysis of Blast Furnace Flue Dust

| Screen Size | | Dust Loading |
|---|---|---|
| Mesh | Microns | Range, % |
| 20 | 833 | 2. 5 - 20. 2 |
| 30 | 589 | 3. 9 - 10. 6 |
| 40 | 414 | 7. 0 - 11. 7 |
| 50 | 295 | 10. 7 - 12. 4 |
| 70 | 208 | 10. 0 - 15. 0 |
| 100 | 147 | 10. 2 - 16. 8 |
| 140 | 104 | 7. 7 - 12. 5 |
| 200 | 74 | 5. 3 - 8. 8 |
| -200 | -74 | 15. 4 - 22. 6 |

| | |
|---|---|
| at top of furnace | 7-13 gr/acf |
| dust catcher outlet | 3-6 gr/acf |
| after primary washer | 0.05-0.7 gr/acf |
| after electrostatic precipitator | 0.005-0.01 gr/acf |

Wet electrostatic precipitators are normally used for secondary blast furnace gas cleaning since the gas delivered to the blast furnace precipitator has been cooled to saturation. The type most widely used is a vertical flow employing pipes of 8-inch or 12-inch diameter as collecting electrodes. Plate-type collecting electrodes with horizontal flow have also been used. A rather unique plate-type with vertical flow example is located in the Usinor-Dunkirk steel plant in France.[18]

In this plant the primary gas cleaning consists of dust catcher with a diffusor cone where deposition occurs by gravity and two cyclones in parallel with a rectangular and tangential gas outlet. The wet secondary cleaning consists of three secondary cleaning lines from the two cyclones. Each line carries a scrubber and wet electrostatic precipitator. The scrubber is a simple tower in which the atomized water travels in the opposite direction of the gas stream and its essential task is to lower the gas temperature from about 200°C to 30°C. Each precipitator consists of a vertical cylinder, divided into two parts which can be isolated in operation by a damper. Each section contains a group of vertical plates at a distance of 25 cm and in the form of a rectangle. Between each plate are suspended ionizing filaments fixed at the top to heated high voltage insulators. The plates are suspended at the top from tubes through which the water, which streams down the plates, is supplied. Intensive periodic spraying reinforces the cleaning operation. For the three lines, and with maximum feed rate, the grain loading after cleaning is 0.0028 grains/cu. ft.

A fairly new innovation by Lurgi known as the "Venturion Electro-Precipitator"[19], combines venturi scrubbers and the wet precipitator in a common housing to meet the demand for additional space economy. Other advantages are less consumption of water resulting in lower investment cost for the water treatment equipment, and elimination of scrubbing tower foundations. Precipitators of this type have been installed at Yawata's Kukioka (Japan) blast furnace and at the Tobata (Japan) No. 3 blast furnace with excellent results. With inlet gas concentrations between 2.51 and 5.02 grams/$Nm^3$, collection efficiencies averaged greater than 99.9%.[20] Compared in cost to the dry Cottrell and the normal wet-precipitator-spray tower combination, the "Venturion-Precipitator" costs considerably

less (about 30%) than the Cottrell and slightly less (about 10%) than the normal arrangement.

The more typical wet type pipe electrostatic precipitator is shown in Figure 31. In some designs the pipes are suspended from a header sheet at the top, with the lower end unsupported, allowing the gas to surround the pipes. In other designs, a lower header sheet is used to prevent the gas from surrounding the pipes. Discharge electrodes are usually vertically hung, twisted square bars (3/16 - 1/4 in.) spanning the height of the collecting electrodes. Rapping the electrodes to dislodge dust is not required since the collectors are flushed with liquid. An overall flow diagram of wet cleaning blast furnace gas with a wet precipitator is given in Figure 32. Of the 60 blast furnaces at present on basic iron production in the British Steel Corporation, all but seven use this three-stage cleaning system.[21]

Trends in the application of electrostatic precipitators for cleaning blast furnace gas are indicated by the number of installations from 1931 to 1968 as shown in Figure 33, which presents the number of furnaces and the number of furnaces with electrostatic precipitators installed. Precipitator inlet gas temperatures and inlet particulate loading are shown in Figures 34 and 35. Distribution of precipitator gas velocity, field strength, and power are shown in Figures 36, 37, and 38, and design efficiency trends are given in Figure 39.

Figure 40 is a plot of the efficiency of blast furnace precipitators as a function of the collection plate area to gas volume ratio. The range of design values is given by the curves, and test data are plotted individually as points. The design data would give precipitation rate parameters ranging from about 9-11 cm/sec (0.3-0.37 ft/sec). Test data show a fairly wide scatter band. Design values of precipitation rate parameter w for pipe-type blast furnace precipitators are usually in the range of 0.25 to 0.45 ft/sec, while performance values of w tend to be somewhat higher. Figure 41 gives more detail on the comparison of design and actual performance by using as a basis of comparison the ratio (R) of actual performance w to the anticipated performance w. For instance, a value of R = 1.20 means that the precipitator size is 20% larger than required to meet the expected performance for that particular unit.

Electric Arc Furnaces

Electric arc furnaces employed for melting and refining steel are cylindrical, refractory-lined structures. They range in size from a diameter of about seven feet, with a hot metal capacity of about four tons, to a diameter of 24.5 feet with a capacity of 200 tons. The refractory lining used inside the steel structure may be either

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Gas
Outlet

Gas
Inlets

Insulator
Compartment

Steam
Coils

Discharge
Electrode System

Water Pond
and Weirs

Gas
Deflector

Approximate Scale:

5 ft = 1 in.
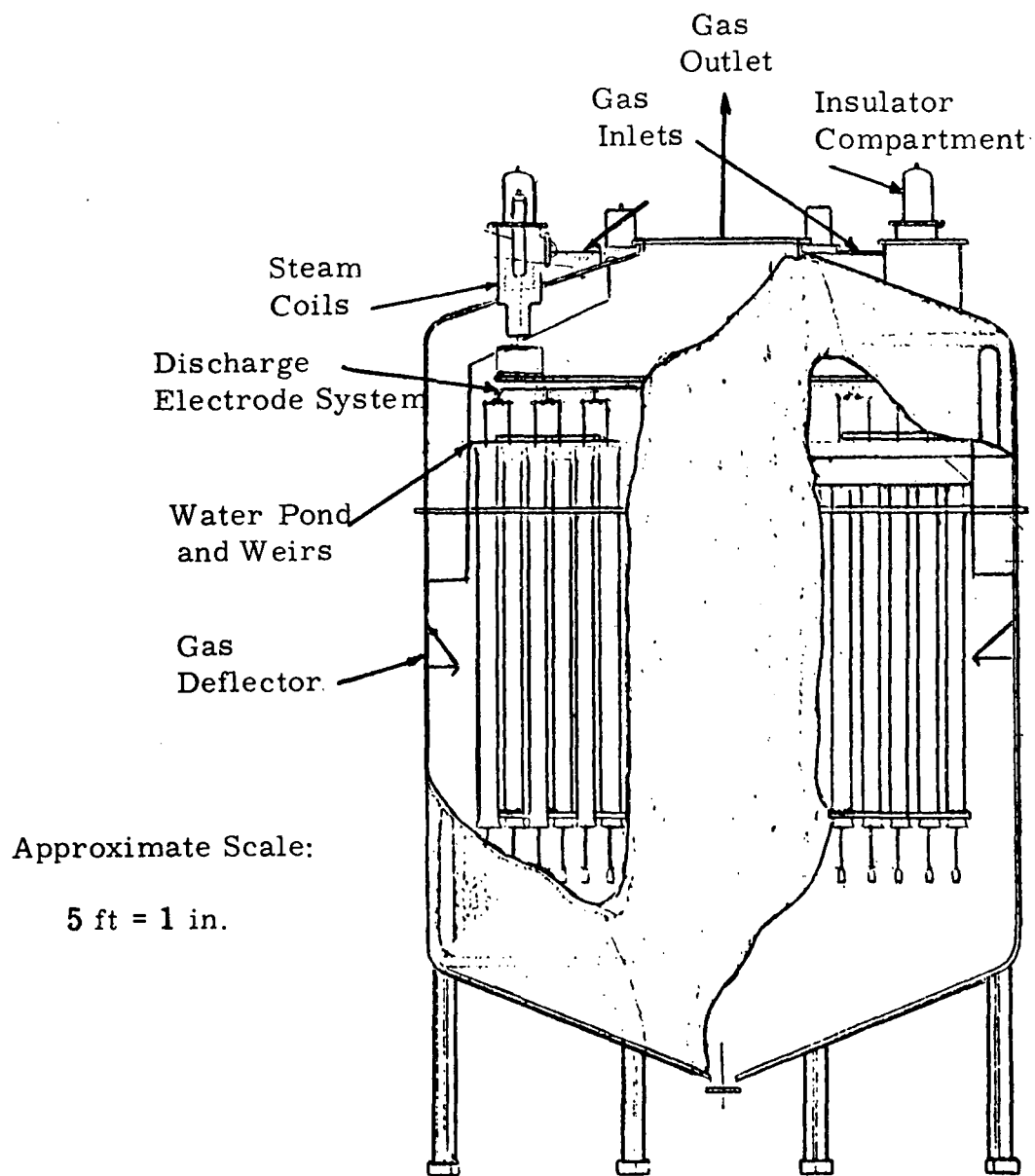
Figure 31. Typical Wet Type Pipe Precipitator for Cleaning
Blast Furnace Gases. [17]
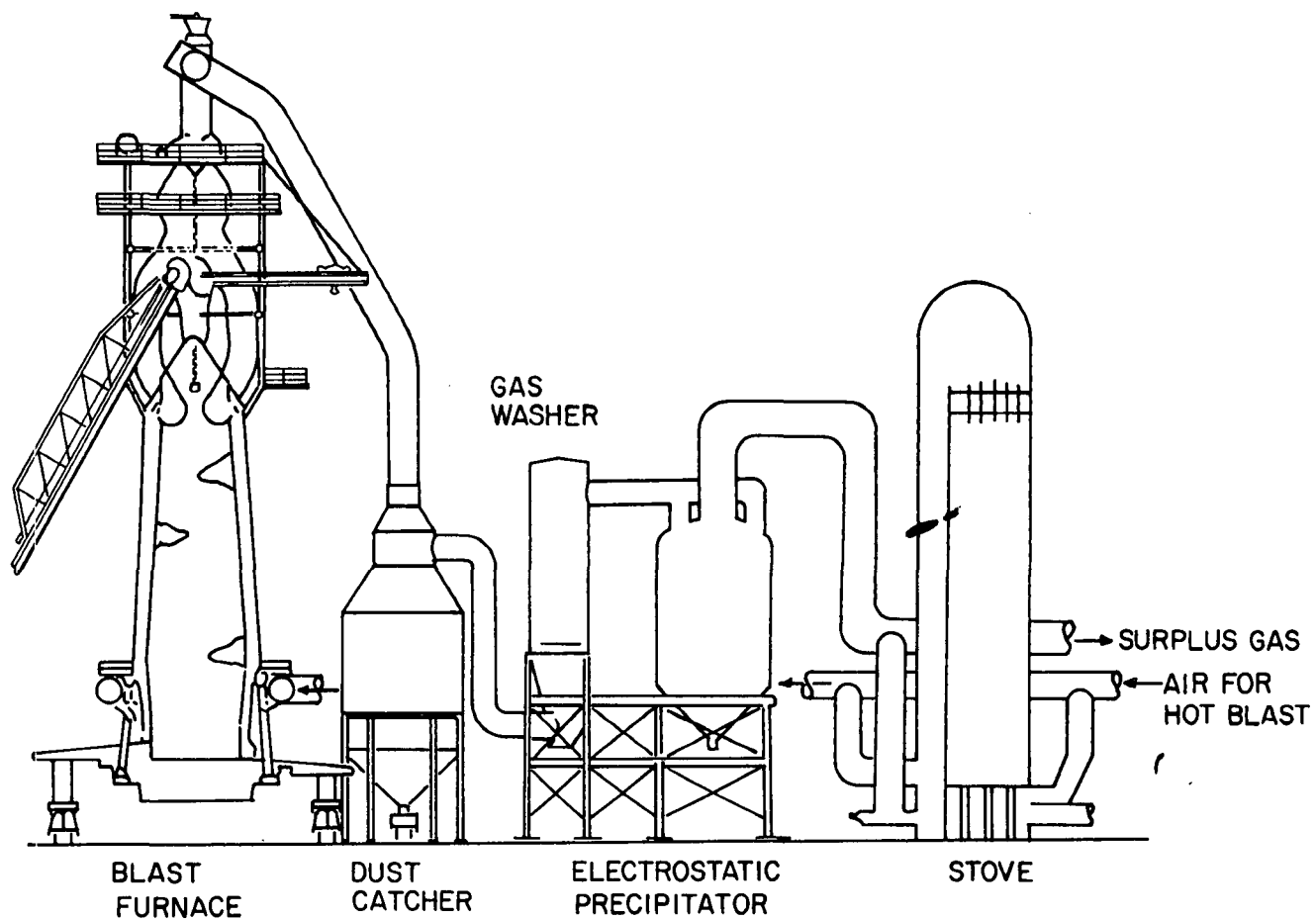
Figure 32. Flow Diagram for Wet Cleaning Iron Blast
Furnace Gas with Electrostatic Precipitator[21]
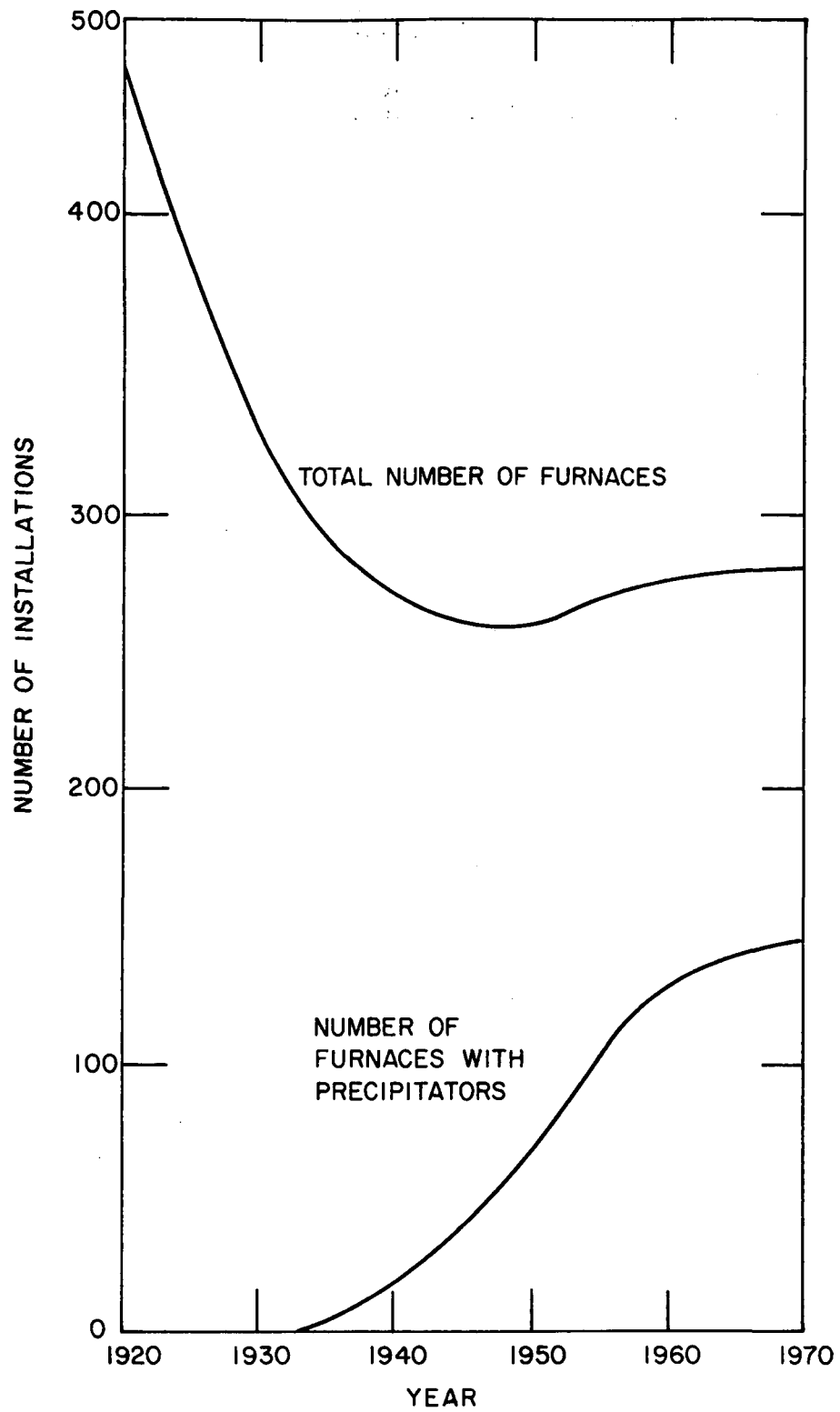
Figure 33.  Blast Furnace Statistics for Period 1920-1969.[38]
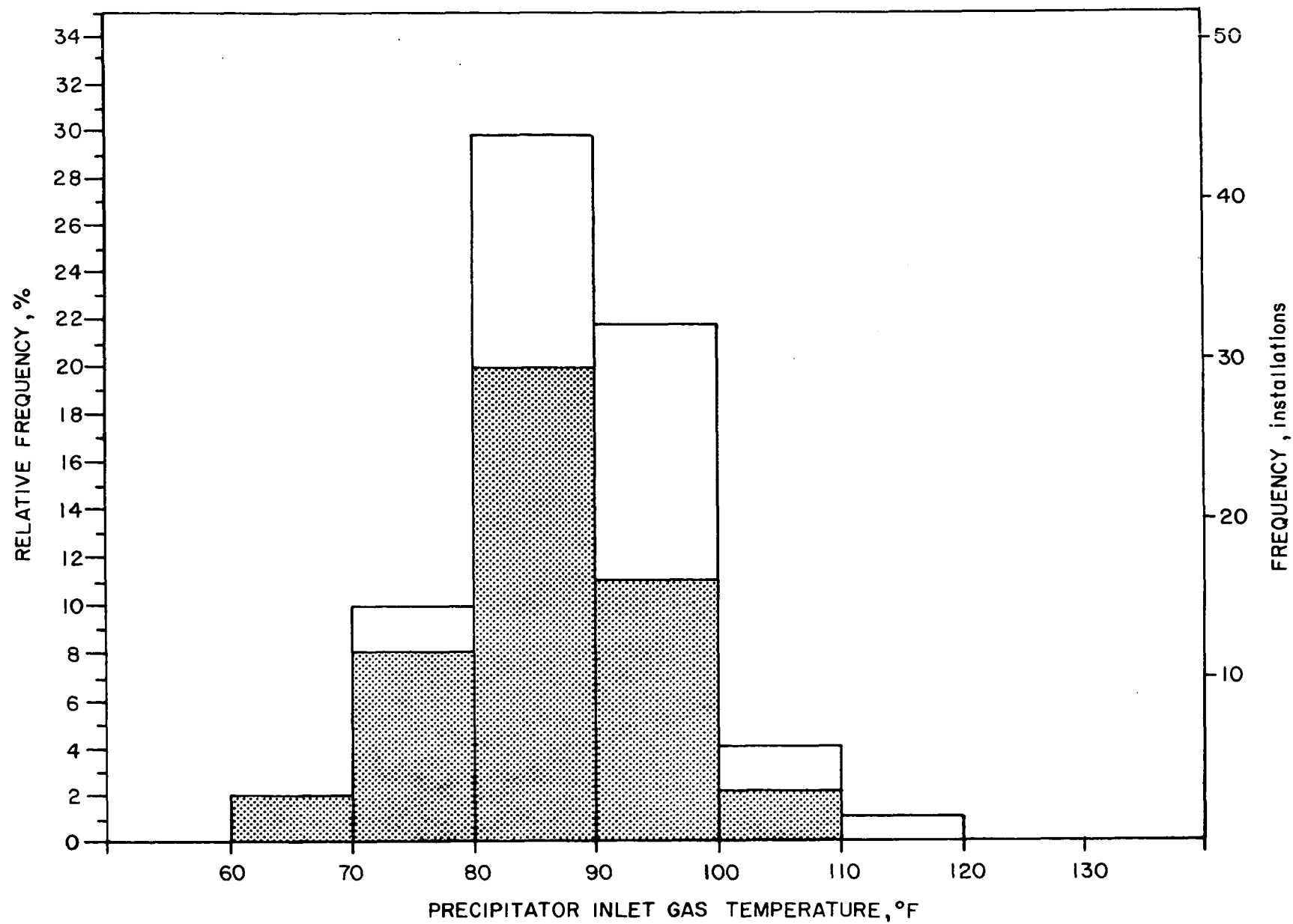
Figure 34.    Distribution of Precipitator Inlet Gas Temperature.
Blast Furnace Installations.[38]
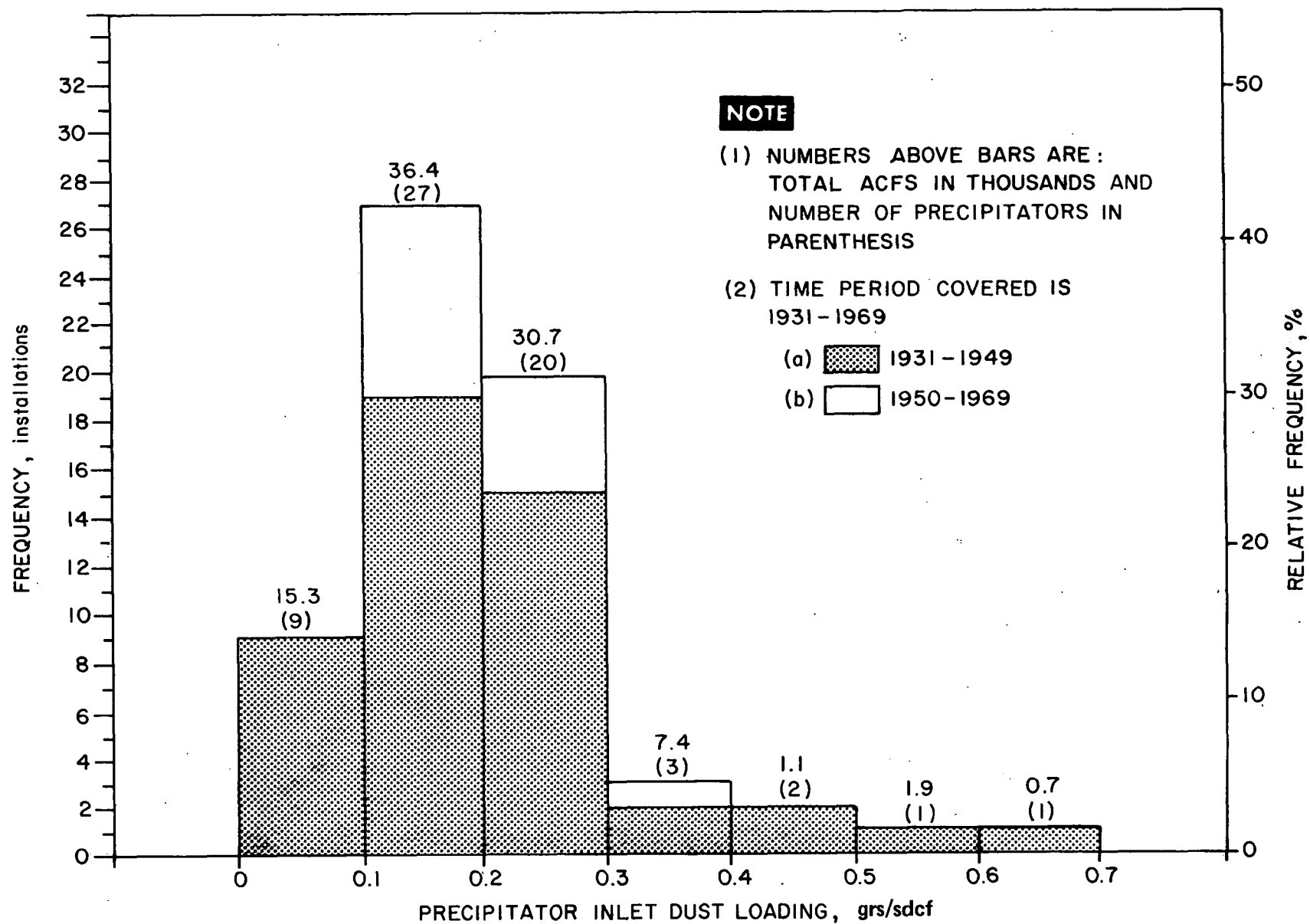
Figure 35. Distribution of Precipitator Inlet Dust Loading.
Blast Furnace Installations.[38]
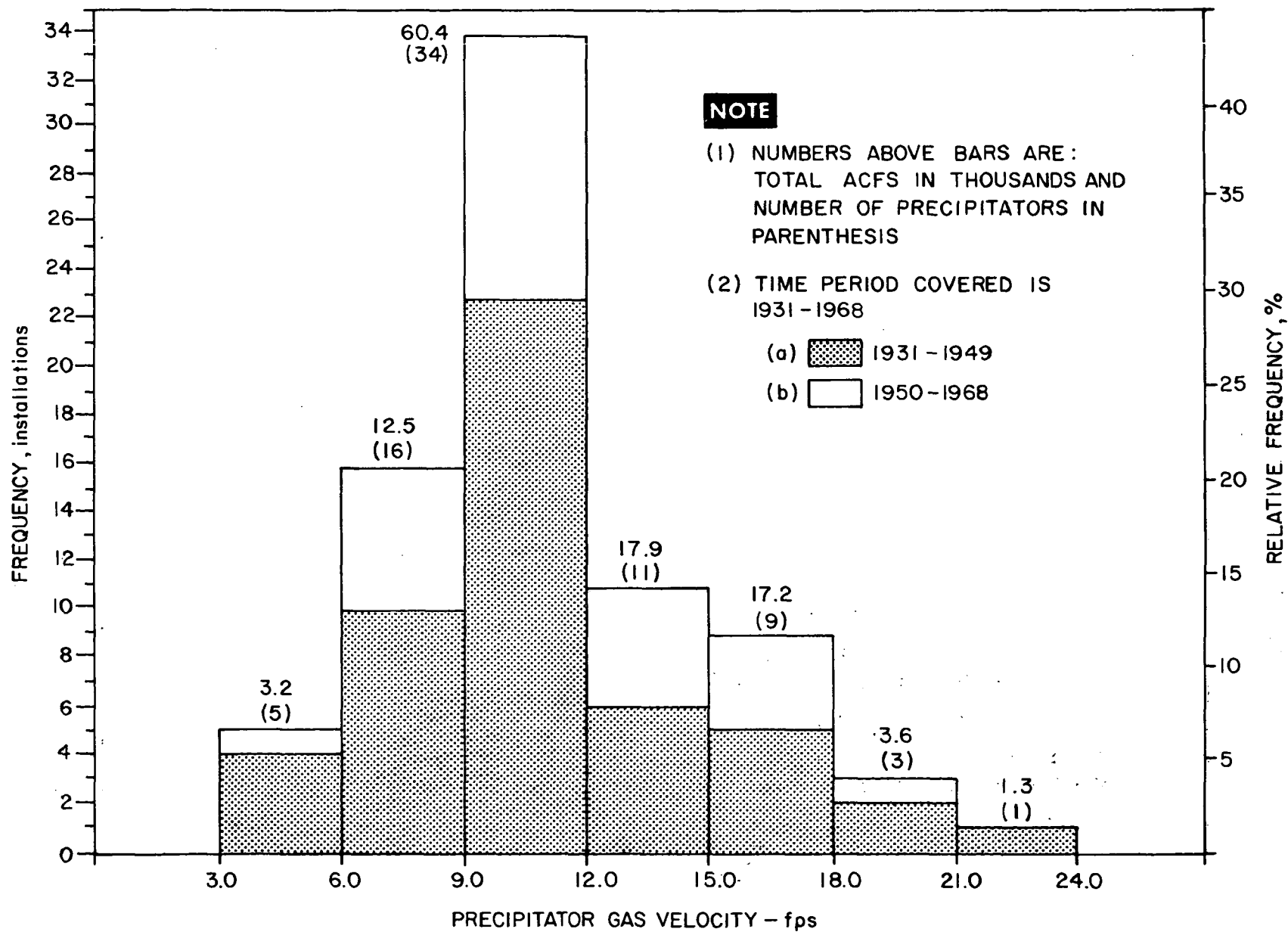
Figure 36.   Distribution of Precipitator Gas Velocity.
Blast Furnace Installations.[38]
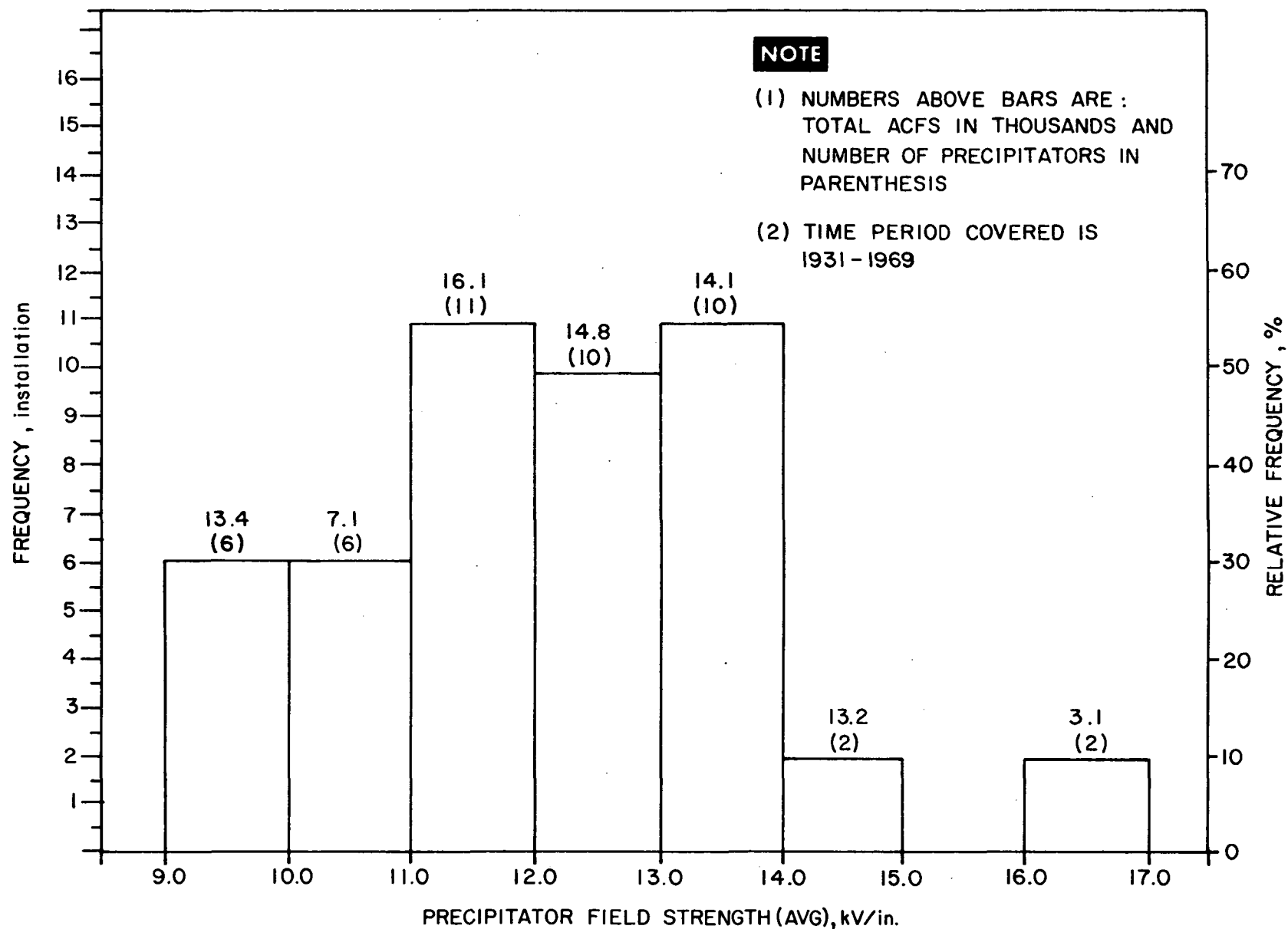
Figure 37. Distribution of Precipitator Field Strength. Blast Furnace Installations.[38]
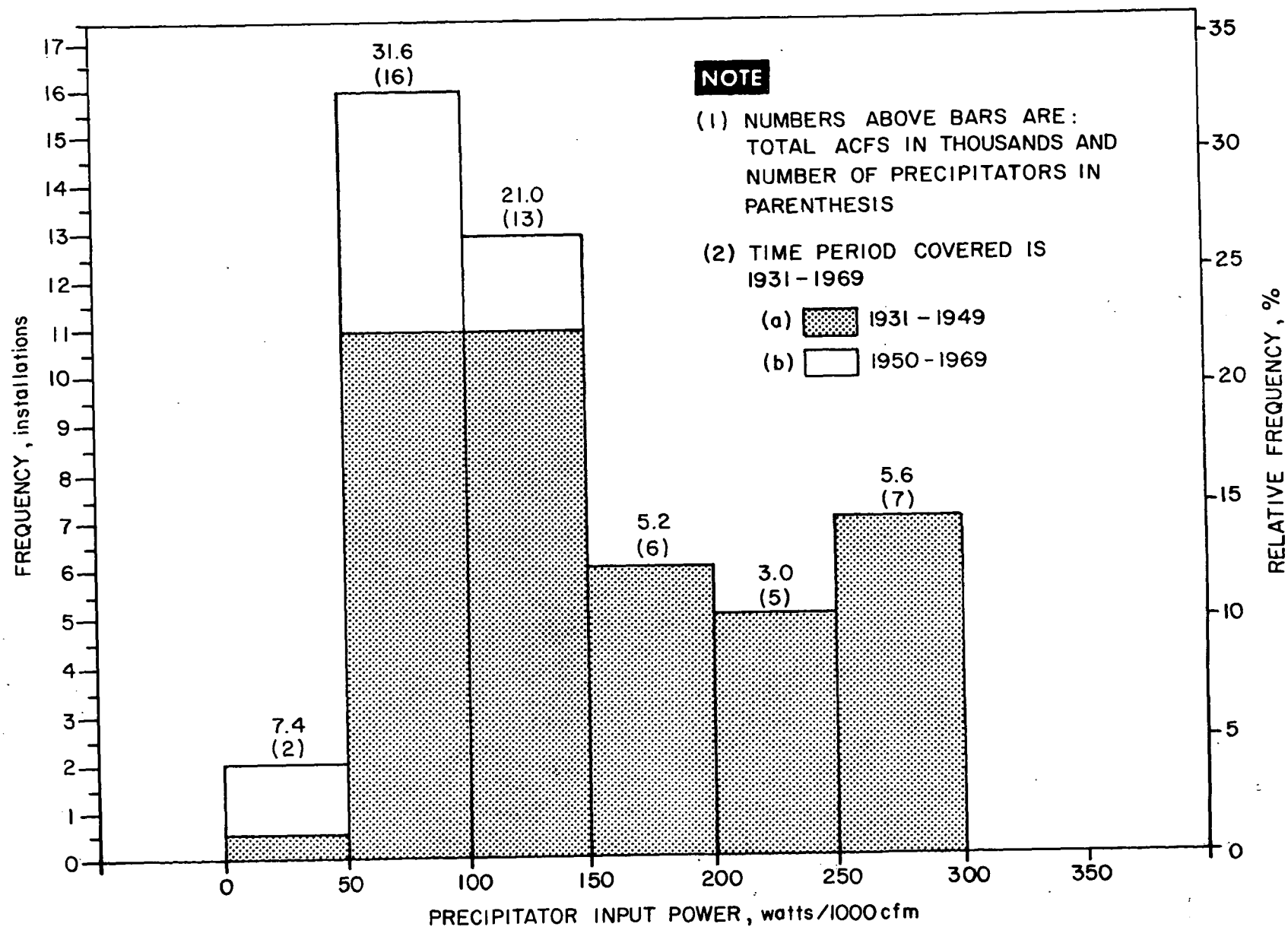
Figure 38.   Distribution of Precipitator Input Power.
Blast Furnace Installations.[38]
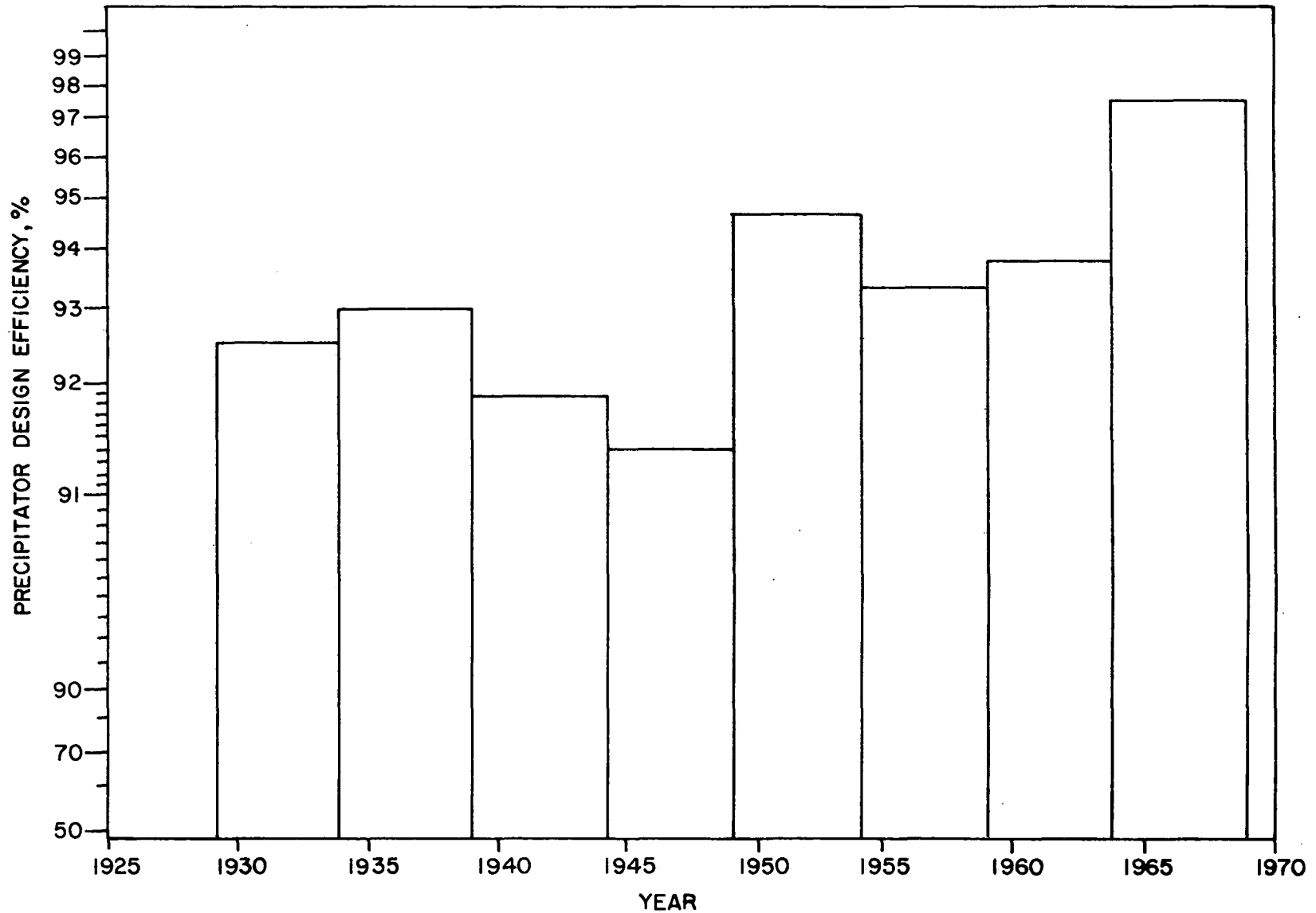
Figure 39.   Design Efficiency Trends Over the Period 1939-1969
             Prorated on acfm Basis for Blast Furnace Installations.[38]
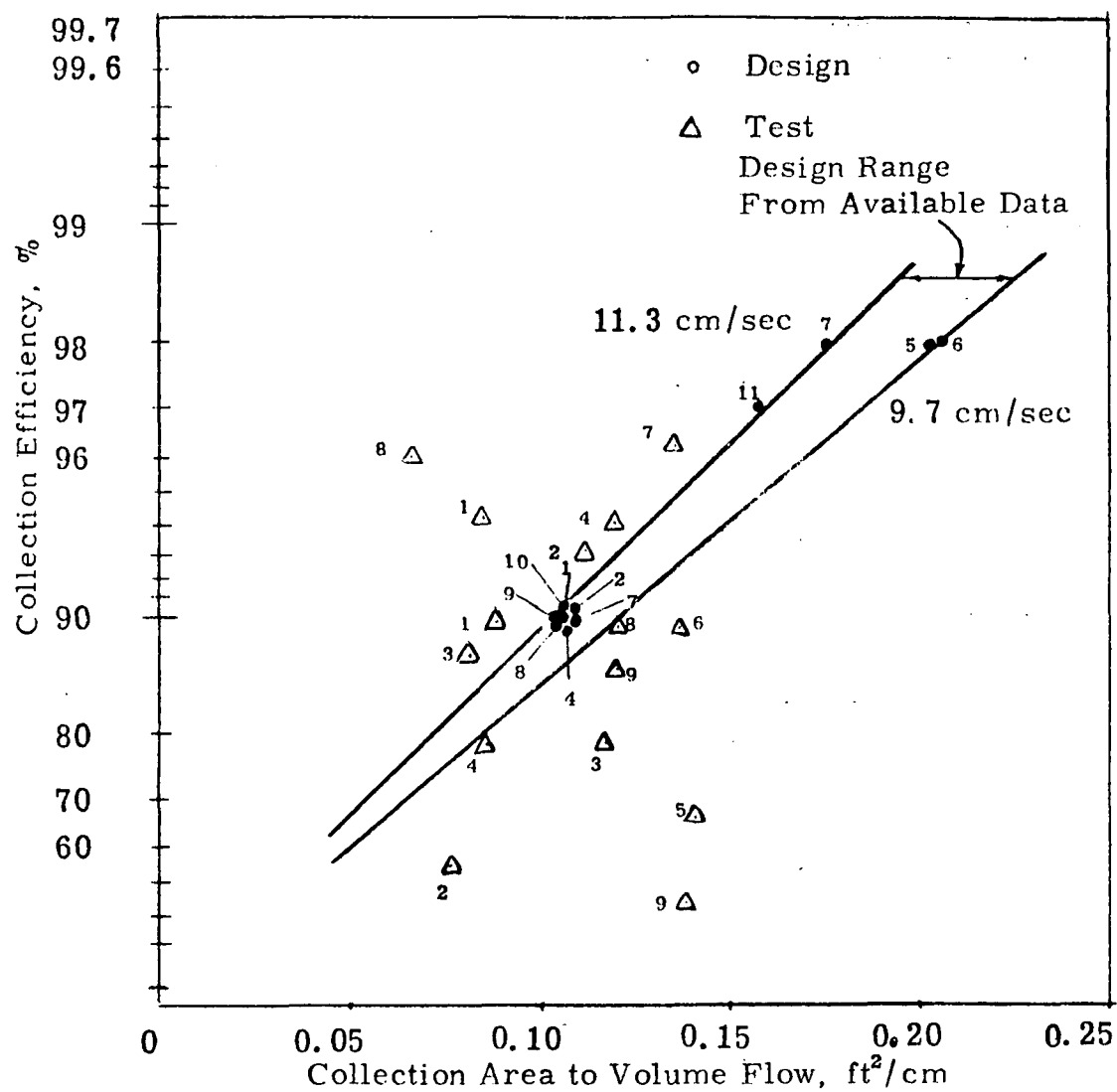
Figure 40. Relationship Between Collection Efficiency
and Specific Collection Area for Electrostatic
Precipitators Operating on Blast Furnace
Installations. [17]
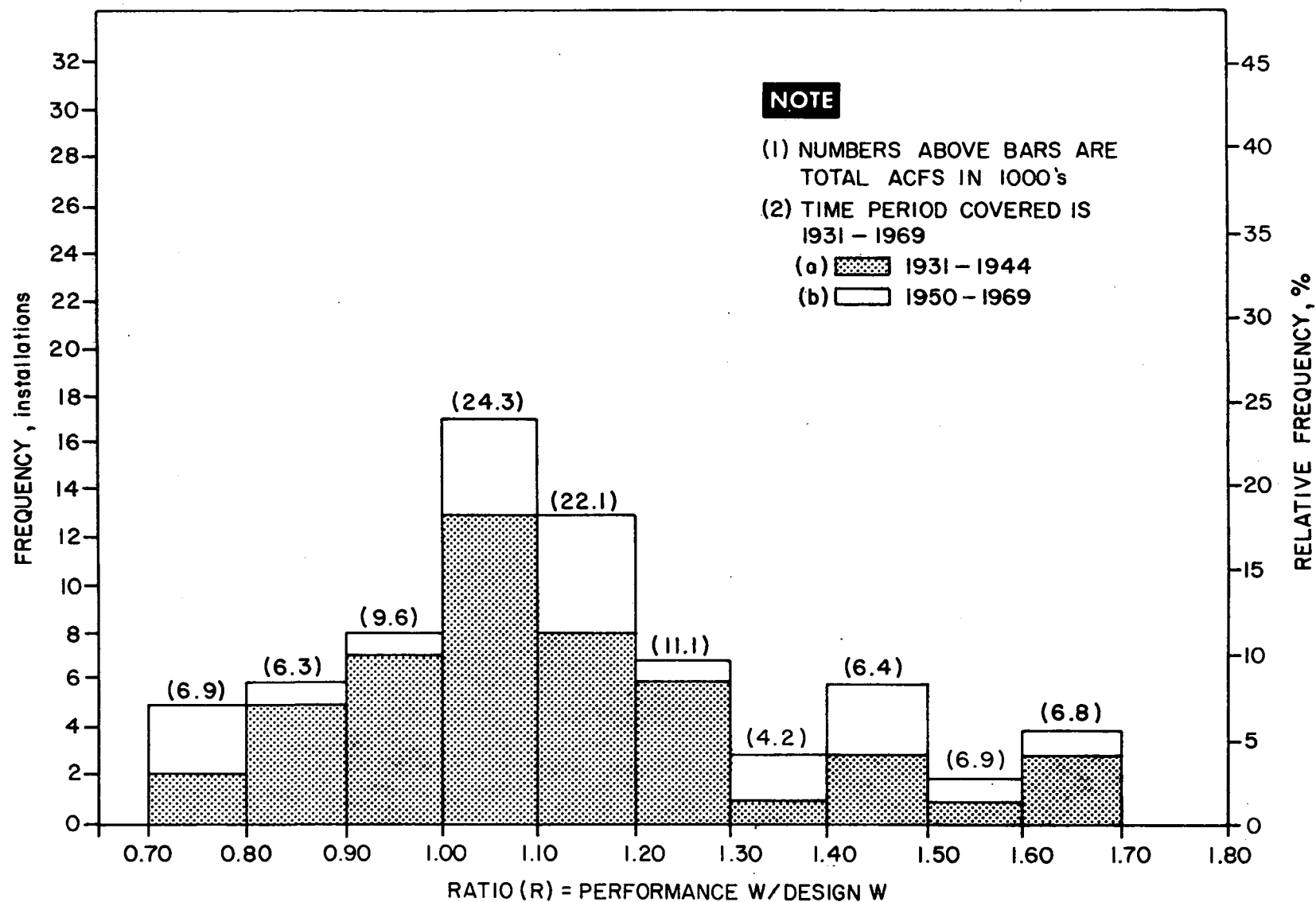
Figure 41. Comparison of Actual Performance to Design
Performance on Blast Furnaces (Basis: ratio
of precipitation rate parameters using
Deutsch equation).[38]

acid or basic, but basic refractories are more commonly used. Because of the flexibility of operation, the electric arc furnace is used to produce steels with a wide range of composition, including carbon, alloy, and stainless grades. In 1970, 15.3% of the steel produced in the U.S. was produced in electric arc furnaces.[22]

In practice, steel scrap, and perhaps hot metal from a blast furnace, are charged into the furnace, the electrodes lowered, and the power turned on. Power for melting and super-heating the charged material is supplied to the furnace through externally supported carbon electrodes which are automatically raised or lowered through holes in the furnace roof. Intense heat is produced by the current arcing between the electrodes and the metal charge, and this, coupled with the resistance heating that occurs as the current flows through the charge materials, results in melting and super-heating of the charge.

Very little air is admitted to the electric arc furnace during the meltdown period, and during this time the furnace operates with an atmosphere containing a small amount of carbon monoxide.[23] The particulate emissions from the electric arc furnaces during this meltdown period are composed primarily of volatile matter from the charged scrap including oil, grease, and oxides of metals with high vapor pressures. Zinc oxide from galvanized scrap is probably the most common metallic oxide evolved during this period. When lancing occurs after meltdown, the oxygen combines with carbon, silicon, and manganese in the melt, to produce a furnace atmosphere containing primarily carbon monoxide (80%), free oxygen, carbon dioxide, and hydrogen. Reported exit temperatures vary from 1800-2900°F. Since the flue gases contain large carbon monoxide concentrations when released from the electric furnace, there is an explosion hazard and the gases must be burned or diluted with air before entering dust cleaning devices. The flue gas volume to be treated by a dust collecting device ranges from a minimum of 12 to perhaps 200 times the volume of oxygen lanced into the furnace, depending on the method used to cool and collect the waste gases.

Wide variations of emissions generated during electric furnace steelmaking are attributable to the physical nature of scrap used, the cleanliness of the scrap, the nature of the melting operation, and oxygen lancing. A compilation of data on total particulate emissions from various electric arc furnaces is given in Table 16.[24]

Just as the amount of particulate matter from the electric furnace varies with the charge and operating conditions, so does the composition of the particulate. Table 17 presents reported compositions of dust emissions.[25,26]

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Table 16[24]

Electric Arc Steel Furnace Fume Emission Data

| Case | Rated Furnace Size, tons | Average Melting Rate, tons/hr | Cycle Time, hr | Fume emission/ton Melted, lb/ton | Furnace Process |
|------|------|------|------|------|------|
| A | 50 | 18.3 | 4 | 9.3[3] | |
| | 75 | 23.5 | 4 | 18.6[4] | Basic, single slag |
| | 75 | 23.5 | 4 | | |
| | | | | | |
| B | 50[1] | 14.4 | 4 | 7.6 | |
| | 50[1] | 13.6 | 4 | 6.9 | Basic, single slag |
| | 75[1] | 21.9 | 4 | 12.3 | |
| | 3 | 1.5 | 2 | 12.6 | Acid, oxygen blow |
| | 3 | 1.1 | 2 | 7.6 | Acid, oxygen blow |
| | 6 | 3.1 | | 10.4 | |
| | 10 | 6.6 | 2 | 5.5 | Basic, oxygen blow |
| | 10 | 5.4 | 2 | 5.2 | Basic, oxygen blow |
| | 2[2] | 1.5[2] | | 13.4 | |
| | | | | | |
| C | 3 | 1.9 | 2 | 4.5 | Acid |
| | 3 | 1.6 | 2 | 5.8 | Acid |
| | 3 | 1.9 | 2 | 5.7 | Acid |
| | 6 | 2.6 | 2.3 | 15.3 | |
| | 6 | 3.0 | 2 | 12.8 | |
| | | | | | |
| D | 18 | 5.4 | 3 | 6.1 | |
| | | | | | |
| E | 6 | 4.1 | 1.2 | 29.4 | Acid, single slag |
| | | | | | |
| F | 3 | 1.8 | 1.8 | 12.7 | Acid, single slag |

[1]Refer to same furnaces as Case A.

[2]Two 2-ton furnaces operating in parallel.

[3]Average for one 50-ton and two 75-ton furnaces processing normal scrap.

[4]Average for one 50-ton and two 75-ton furnaces processing dirty, subquality scrap.

Table 17[25,26]

Electric Furnace Dust Composition

| Component | Weight, % |
|-----------|-----------|
| ZnO | 37 |
| Fe | 25 |
| CaO | 6 |
| MnO | 4 |
| $Al_2O_3$ | 3 |
| $SO_3$ | 3 |
| $SiO_2$ | 2 |
| MgO | 2 |
| CuO | 0.2 |
| $P_2O_5$ | 0.2 |

Dust from electric arc furnaces tends to be extremely fine as indicated in Table 18. These data, from several sources, indicate that 70 to 100% of the fume is below 5.0 microns in size.

Because of the small particle size of arc furnace fume, only three types of cleaning equipment, bag filter, venturi scrubber, and electrostatic precipitator, are considered. A British Steel Corporation survey in 1969 indicates the distribution of utilization of these collectors for arc furnace fume.[29] As shown in Table 19, there is a strong preference in the United Kingdom to use wet precipitators on the larger arc furnaces. High energy wet scrubbers are increasingly used for 20-30 ton furnaces while bag filters are used mostly on small furnaces. A paper by Holland and Whitwam[30] deals with the design problems involved in providing a suitable technique (in this case, wet precipitation) to overcome the difficulties of providing direct roof extraction for a large (100 ton) electric furnace using high rates of tonnage oxygen. Pilot plant tests at the plant described in the paper left a choice of two types of cleaning equipment as best for a large plant - venturi scrubber and wet electrostatic precipitator. The venturi scrubber was eliminated because of the large pressure drop (30 in. wg) and the necessity for varying the throat to maintain the pressure drop at the varying volumes. The operating conditions which had to be met are listed below:

| | |
|---|---|
| maximum oxygen usage | 2300 cfm NTP |
| maximum gas volume | 35,000 cfm NTP |
| maximum dust burden | 6.5 grains/cf NTP |
| particle size | 90% < 1 $\mu$m |
| maximum gas temperature | 1250°C during lancing |
| | 1000°C during refining and melting |
| volumes at NTP | lancing 35,000 cfm |
| | melting 15,000 cfm |
| | refining 0-5000 cfm |

The wet precipitator is of the horizontal gas-flow type with two fields in series connected to a high-voltage supply from one transformer selenium rectifier equipment with transductor control. If an electrical fault develops in the precipitator, either of the two fields can be isolated from the high voltage supply, which enables partial gas cleaning to be maintained until the fault is remedied. Stainless steel was used for the collecting and discharge electrodes, the casing was tile lined, and other mild steel surfaces in contact with the gases were protected with acid-resistant paint. Flushing of the precipitated materials is provided by means

<div align="center">

Table [18]

Electric Steel Furnace Fume Particle
Size Data

</div>

Percent by Weight below Given Particle Diameter

| 0.5 micron | 1.0 micron | 3.0 micron | 5.0 micron |
|:---:|:---:|:---:|:---:|
| | 90[1] | | |
| 95 | | 100[1] | |
| 90[1] | | | |
| | | | 70[1] |
| | | | 71.9[2] |
| | | | 67.9[2] |
| | | | 72.5[2] |

[1]ESP Manual[27]

[2]Muhlrad, W., "Dust Extraction from the
Fume of Electric Arc Furnaces", Iron and
Coal 183, 669-675 (1961).[28]

Table 19[29]

FUME CLEANING FOR STEELMAKING PROCESSES 1969 (BSC)

| Steelmaking Process | Electrostatic Precipitators | | High Energy Wet Scrubbers | Bag Filters |
|---|---|---|---|---|
| | Dry | Wet | | |
| LD, Kaldo, Rotor Converters | 17 | - | 6 | - |
| Open Hearth $O_2$ Lanced | 32 | - | - | - |
| Electric Arc Furnaces, $O_2$ Lanced | | | | |
| up to 60 tons capacity | - | 1 | 7 | 8 |
| over 60 tons capacity | 4 | 14 | - | - |

of oscillating water jets located above each field and their introduction is automatically sequenced to occur during the fettling and charging period on the furnace, when the precipitator is de-energized. The wash water is introduced at the rate of 1000 gal/min for a period of 7 minutes.

A fairly recent innovation in wet precipitator technology for cleaning electric furnace fume is described by Fraunfelder.[31,32] This system consists of a hydro-scrubber (marble-bed scrubber) with a wet electrical precipitator arranged above it in the same housing. The fumes are pre-cooled at the scrubber inlet by direct water sprays and passed through a bed of marbles which is sprayed with water from below. The fumes then pass through the precipitator where final cleaning takes place. At predetermined intervals the plates are flushed for about three minutes. The power, supplied by standard selenium rectifiers, 55 kV, 750 mA, is maintained during flushing.

The pilot system having yielded good results, the ferro-silicon works of Valmoesa (Fraubunden, Switzerland) placed an order for an industrial collection system of this type to cope with a gas volume of 108,000 $m^3$/hr. With an initial dust loading ranging from 0.5 to 5.0 g/$m^3$ NTP, cleaning efficiency was measured as 98.4%. According to the findings of the Swiss Materials Testing Institute, the dust content of the clean gas ranged from 30-60 mg/$m^3$ NTP, averaging 37 mg/$m^3$ NTP.

Water requirements are 1400 l/min continuously for the Hydro-washer (scrubber) and an additional 1000 l/min during the three min per hour the nozzles for the precipitator are used. This gives an overall water consumption of 90 $m^3$/hr or 0.83 l/$m^3$ of gas treated.

## Scarfing

Before steel can be rolled, surface defects in the bloom, ingot, and billets must be removed by scarfing, in which the skin of the steel slab is removed. In this operation, slabs from the slabbing mill are conveyed to the stationary scarfing machine, where they are preheated with oxygen and acetylene for about three seconds, after which the scarfing operation begins. The slabs pass the cutting torches at 80 to 120 fpm where a cut of about 1/16 inch is made on two sides of the slab. The sparks and fume are blown downward by compressed air toward a target plate which is continuously sprayed with water.

During the cutting operation a coarse, high pressure spray of water is introduced immediately after the gas nozzles to granulate the larger chips and wet the fume. Because of the high temperatures generated during the scarfing operation, some of the steel is vaporized and subsequently oxidized. The iron oxide fume is usually quite fine.

There are many references in the literature to wet precipitator usage for cleaning scarfing effluent, but little detail is provided concerning precipitator design and performance. A paper by Elliott and Lafreniere[33] is the best example of wet precipitator usage uncovered thus far. The Steel Company of Canada, Limited, Hamilton, Ontario, investigated various collecting equipment used elsewhere for similar applications and selected a tubular-type wet electrostatic precipitator. Since the gases entering the precipitator are saturated with water vapor, it was felt that the collected material could be more readily removed by water sprays than by mechanical shakers. The choice of a pipe-type rather than the plate-type precipitator was made to insure adequate water distribution during the washing cycle. Distribution weirs were installed as a precautionary measure to insure that the tubes were kept clean.

The precipitator installation consists of twin units in parallel designed to handle 70,000 cfm at dust loadings of 1.0 gr. per cu. ft. and maximum temperature of 150°F. Actual operating conditions indicate a flow of 56,000 cfm and temperatures varying between 40°F and 70°F. The precipitator consists of a group of 192, ten-inch diameter collecting pipes and 0.1055 inch diameter discharge electrode wires (#30 stainless) enclosed in two rectangular casings of 5/16 inch steel plate. All interior surfaces of the structure are epoxy coated and internal parts are made of stainless steel. Each electrode is supported from a centering clip attached to the electrode support framework, and the lower end of each discharge electrode is held taut by a 15 lb cast iron weight. A frame is provided to hold the bottom end of the electrodes in position. The upper framework is suspended from four insulators which are located in the respective insulator compartments. The entire framework including the electrodes is insulated from the grounded shell by these support insulators. To prevent dust and moisture from accumulating on the insulators, a heating and ventilating system is provided which consists of a fan and heater and the necessary ducts to the insulator compartments. Warm, dry air is blown into the compartments, and a thermostat maintains the temperature at about 200°F.

The collecting pipes are 10" in diameter by 15' long and constructed of 12 gauge #304 stainless steel. Each pipe is suspended from a stainless steel header sheet and held vertically in position by an alignment grid located at the lower end of the pipes.

The washing system consists of eight stainless steel full cone pattern nozzles located at the top of the casing. The power is on the precipitator at all times. After twenty blooms have been scarfed, the bloom scarf counter energizes a relay timer which energizes a wash timer which opens the water valve to the spray nozzles and closes the louvers at the entry to the exhaust fan. The louvers remain closed during the washing to reduce the gas flow to the precipitator and thus prevent carry-over of wash water out the stacks. The complete wash cycle is accomplished between scarfs and takes approximately 42 seconds to complete.

## Cupolas[34]

The iron cupola is a moderately low cost, efficient method of continuously providing molten iron at the desired temperature and chemistry for foundry operations. The cupola is a cylindrical furnace lined with fireclay or firestone refractories which resemble a miniature blast furnace. It differs primarily in that pig iron and steel scrap replace iron ore in the charge. Since, in most installations, no water cooling is utilized in the melting zone, the lining has to be repaired between periods of operation. The charging door is located on the side of the cupola near the top. Legs support the cupola, thus permitting the use of a drop bottom which facilitates the removal of the remaining burden after the last charge has been tapped.

The charge is composed of coke, steel scrap, iron scrap, and pig iron in alternate layers of metal and coke with sufficient limestone being added to flux the ash from the coke to form slag. When properly mixed with the above materials and with sufficient combustion air, a self-sustaining exothermic reaction takes place due to burning of the coke, which provides the heat necessary to melt the charge. Blast air for combustion is blown in through tuyeres, located near the bottom of the furnace. The combustion air can be handled in one of two ways, depending upon the type of blast system used:

1.  In the cold blast system, the cupola operates with preheating of the combustion air. Temperatures at the cupola exit usually range from 1200 to 2000°F.

2.  In the hot blast system, air preheaters preheat the blast air for the cupola. Changes in production rate can be obtained by altering the preheat temperature and air blast volume.

Cupola operation is usually cyclic because of the limited life of the refractory linings. Considerable quantities of effluent, both gaseous and particulate, are evolved, and the amount is seldom constant except for short periods of time. The effluent rate varies with blast rate, coke consumption, physical properties of coke, type and cleanliness of metal scrap in the charge, coke-to-iron ratio, bed height, burden height, and preheat temperature. The effluent rate also changes at intervals when the furnace is charged with iron, steel scrap, coke, and flux.

Stack gases consist mainly of carbon dioxide, carbon monoxide, sulfur dioxide, nitrogen, and oxygen. The sulfur dioxide is probably the most objectionable, both from a standpoint of nuisance odor and as a cause of corrosion to the gas cleaning plant by the formation of sulfurous acid. Carbon monoxide is usually oxidized to carbon dioxide, though explosion can occur if this oxidation is not complete. Among the remaining constituents of cupola waste

gas, only fluorine has some occasional importance. The fluorine
emission can be attributed to the addition of fluorspar which is
sometimes used in basic cupolas to create favorable conditions for
desulfurizing by decreasing the slag viscosity. Only a few cupolas
are operated with a basic slag.

Particulate material emitted from the cupola stack consists pri-
marily of the oxides of silicon, iron, calcium, aluminum, magnesium,
and manganese, but may also contain small amounts of coke dust,
coke ash, limestone, zinc oxide, and smoldering products from paint,
grease, oil, rubber, and other combustibles. In Table 20 is sum-
marized a mean range of composition of cupola dust as gathered by
H. Pacyna from available literature and reported by Engels. Engels
also summarizes work by W. Patterson, H. Siepman, and H. Pacyna
which gives particulate concentration ranges and scatter values
(Table 21).

The particle size distribution ranges between wide limits, depending
on melt rate, coke usage, scrap formulation, and furnace operating
variables. Older data indicate that the proportion of dust below
10 microns, as a rule, does not exceed 10%, but later studies in-
dicate that particle sizes under 1 μ may constitute 40% and more
of the total weight.

There appears to be a relationship between the coke ratio (coke/scrap)
and the melting rate per unit area of furnace and the emissions from
a cupola stack, as illustrated in Figure 42. While a substantial
amount of scatter exists, increasing amounts of coke per unit of
iron melted in basic hot blast cupolas generally increase the amount
of particulate material. Higher melting rates tend to decrease the
amount of particulate when expressed in terms of weight per unit of
iron melted. Although the data indicate a wide range of dust
emission, most of the data points fall within the range of 4 to
12 lb/t iron.

Particle size distribution data for both hot and cold blast cupolas
are presented in Figure 43.

Though the literature search has not revealed extensive utilization
of wet precipitators for collection of cupola dusts in the United
States, there are a number of such installations overseas, some
of the more recent of which are described by Engels and Weber, and
Ussleber.[35,36]

One of the first wet-type electrostatic precipitators for dust
removal in a hot blast cupola went into operation in early 1964
in a foundry in Velbert, Germany. Figure 44 illustrates the hot
blast cupola plant. The comparatively small space available and
the fact that a settling tank with pump station was already on hand
were decisive in the selection of the wet-type precipitator. The
gases not needed in the recuperator are piped directly to the gas

## Table 20[34]
### CHEMICAL COMPOSITION OF CUPOLA DUST

|  | Mean Range | Scatter Values |
|---|---|---|
| $SiO_2$ | 20-40 % | 10-45 % |
| CaO | 3-6 % | 2-18 % |
| $Al_2O_3$ | 2-4 % | 0.5-25 % |
| MgO | 1-3 % | 0.5-5 % |
| FeO ($Fe_2O_3$,Fe) | 12-16 % | 5-26 % |
| MnO | 1-2 % | 0.5-9 % |
| Ignition Loss (C, S, $CO_2$) | 20-50% | 10-64 % |

## Table 21[34]

### DUST CONTENT OF CUPOLA WASTE GASES

| | Main Range | Scatter Values |
|---|---|---|
| **Cold blast cupola:** | | |
| Top gas, undiluted, $g/Nm^3$ | 6 - 11 | 2 - 15 |
| Waste gas from cupola stack, $g/Nm^3$ | 2 - 6 | 1 - 8 |
| Total dust emission, kg/t iron | 5 - 10 | 2 - 12 |
| | | |
| **Hot blast cupola (acid)** | | |
| Top gas, undiluted, $g/Nm^3$ | 6 - 14 | 3 - 25 |
| Waste gas from recuperator, $g/Nm^3$ | 3 - 7 | 1 - 10 |
| Residual gas from cupola stack, $g/Nm^3$ | 0-5 - 3 | 0.2 - 7 |
| Total dust emission at cupola furnace, kg/t iron | 8 - 12 | 2 - 20 |
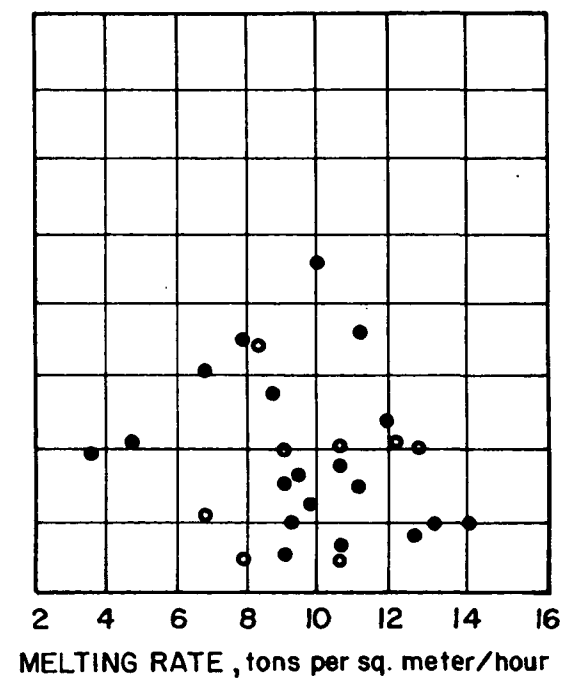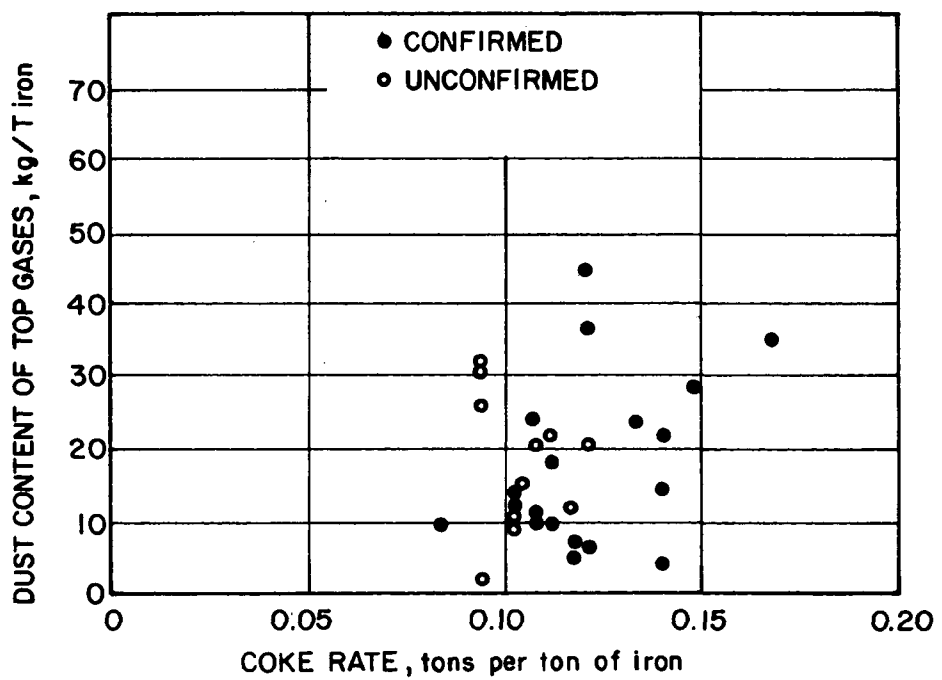
Figure 42. Dust Output Survey as a function of Coke Rate and Melt Rate.[34]
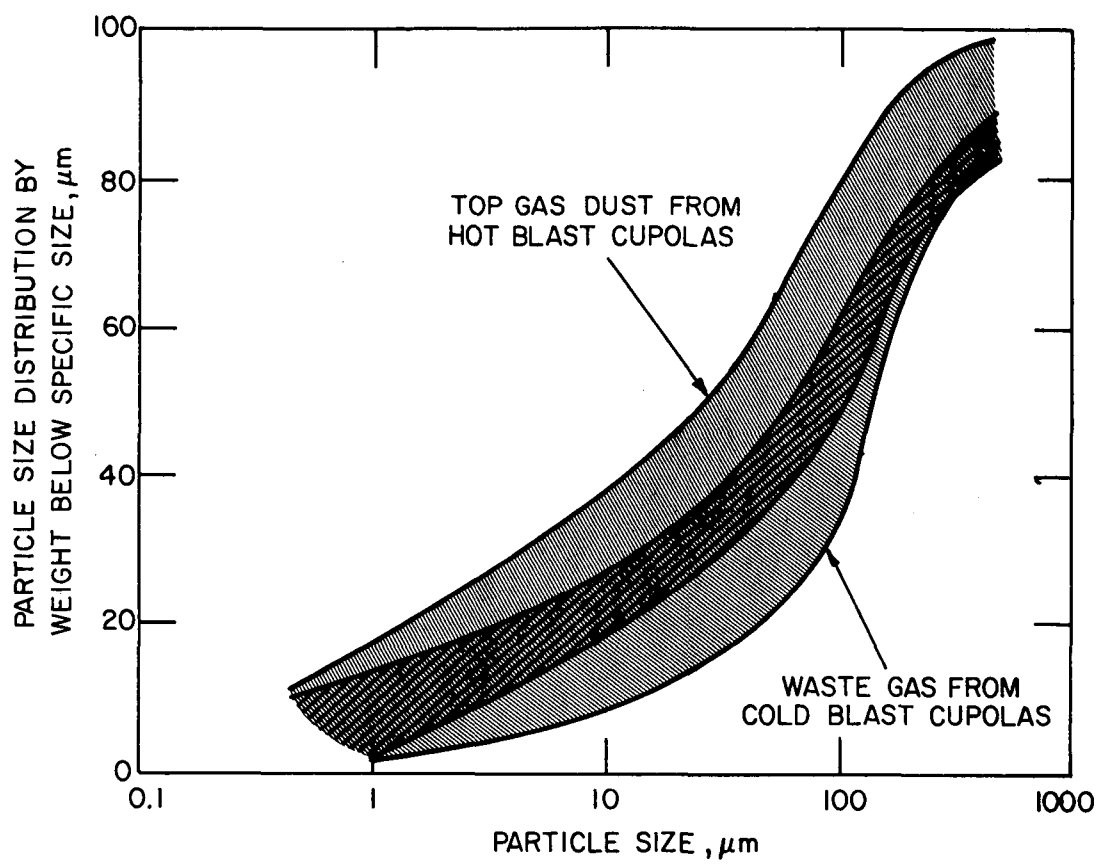
Figure 43.   Particle Size Studies Made with
             Dusts from Hot and Cold Blast Cupolas.[34]
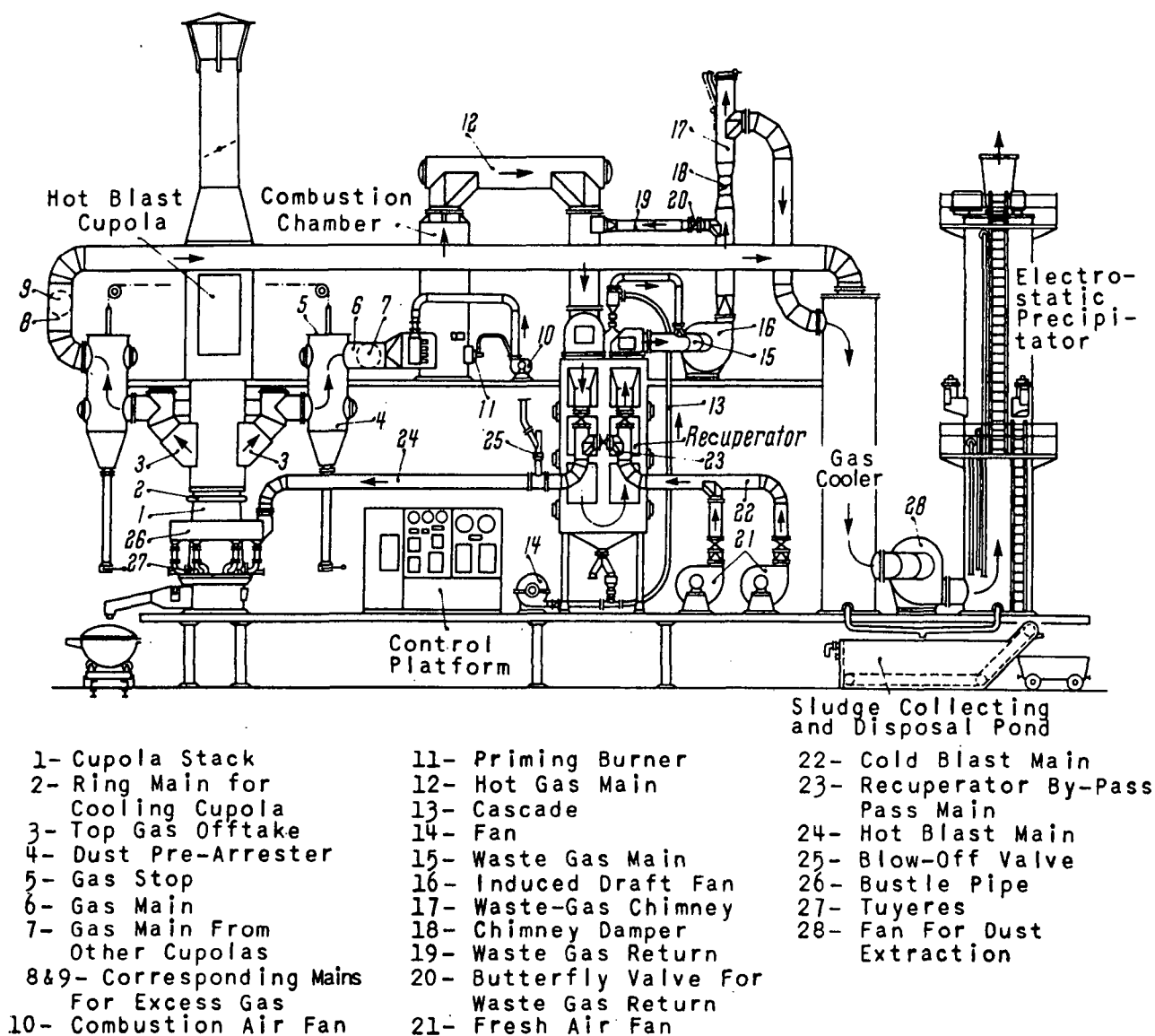
**Figure 44.** Hot-Blast Cupola Plant (melting rate: 6 m.ton/hr) with Wet Electrostatic Precipitator for Cleaning the Waste Gases.[36]

1- Cupola Stack
2- Ring Main for Cooling Cupola
3- Top Gas Offtake
4- Dust Pre-Arrester
5- Gas Stop
6- Gas Main
7- Gas Main From Other Cupolas
8&9- Corresponding Mains For Excess Gas
10- Combustion Air Fan

11- Priming Burner
12- Hot Gas Main
13- Cascade
14- Fan
15- Waste Gas Main
16- Induced Draft Fan
17- Waste-Gas Chimney
18- Chimney Damper
19- Waste Gas Return
20- Butterfly Valve For Waste Gas Return
21- Fresh Air Fan

22- Cold Blast Main
23- Recuperator By-Pass Pass Main
24- Hot Blast Main
25- Blow-Off Valve
26- Bustle Pipe
27- Tuyeres
28- Fan For Dust Extraction

cleaning plant through a by-pass piping system. The following values were measured:

| | |
|---|---|
| Dust content of the mixed gas prior to wet precipitator | 2.8 g/Nm$^3$ |
| Clean gas content behind wet precipitator | 0.0474 g/Nm$^3$ |
| Collection efficiency | 98.3% |
| Dust emission | 0.384 kg/h = 0.064 kg/t iron |

The gas cleaning equipment consists principally of a cooling tower (which cools gases to about 60°C), an induced draft ventilator, the vanes of which are sprayed to prevent scaling, and the vertical wet electrostatic precipitator. Water consumption is 1.5 m$^3$/h, consumption of neutralizing agent is 16.6 kg/h, and power consumption is 23 KW.

Engels states that the withdrawal of top gases below the top opening is advisable in newly constructed plants where the heat content of the gases is not utilized in the recuperator. The externally-heated hot blast stove is independent of the top gas supply. The top gases are drawn off below the top opening without prior dilution with outside air for cleaning and are first introduced into an Imatra-Venturi scrubber which provides cooling, saturation, and preliminary cleaning. The scrubber is in combination with a vertical wet-type electrostatic precipitator to form an integrated unit. The measured clean gas dust content of 0.084 g/Nm$^3$ corresponds to a collection efficiency of 98.85%.

A wet precipitator with an Imatra-Venturi scrubber as preliminary filter and saturator is employed for cleaning a mixture of top gas and recuperator waste gas in an acid-lined 12 ton hot blast cupola (Figure 45). In this plant the top gases are completely drawn off below the charge opening, and only part of the gases are introduced into the combustion chamber. With a raw dust content of 4.65 g/Nm$^3$ and a gas volume of 15,700 Nm$^3$/h, the system cleans the gases to a dust content of 0.140 g/Nm$^3$ (97% collection efficiency).

Table 22 presents a cost comparison for various dust collectors for cleaning waste gases of a hot blast cupola. The individual data were taken from various sources and converted by Engels so that costs for the venturi tube dust removal device were fixed at 1 and all other collectors were expressed in relation to the venturi scrubber.

TOP GAS
COLLECTING PIPE

RESIDUAL
TOP GAS

COMBUSTION
CHAMBER

VENTURI TUBE

WASH WATER

COLD BLAST

WET
PRECI-
PITATOR

HOT
BLAST

RECUPERATOR
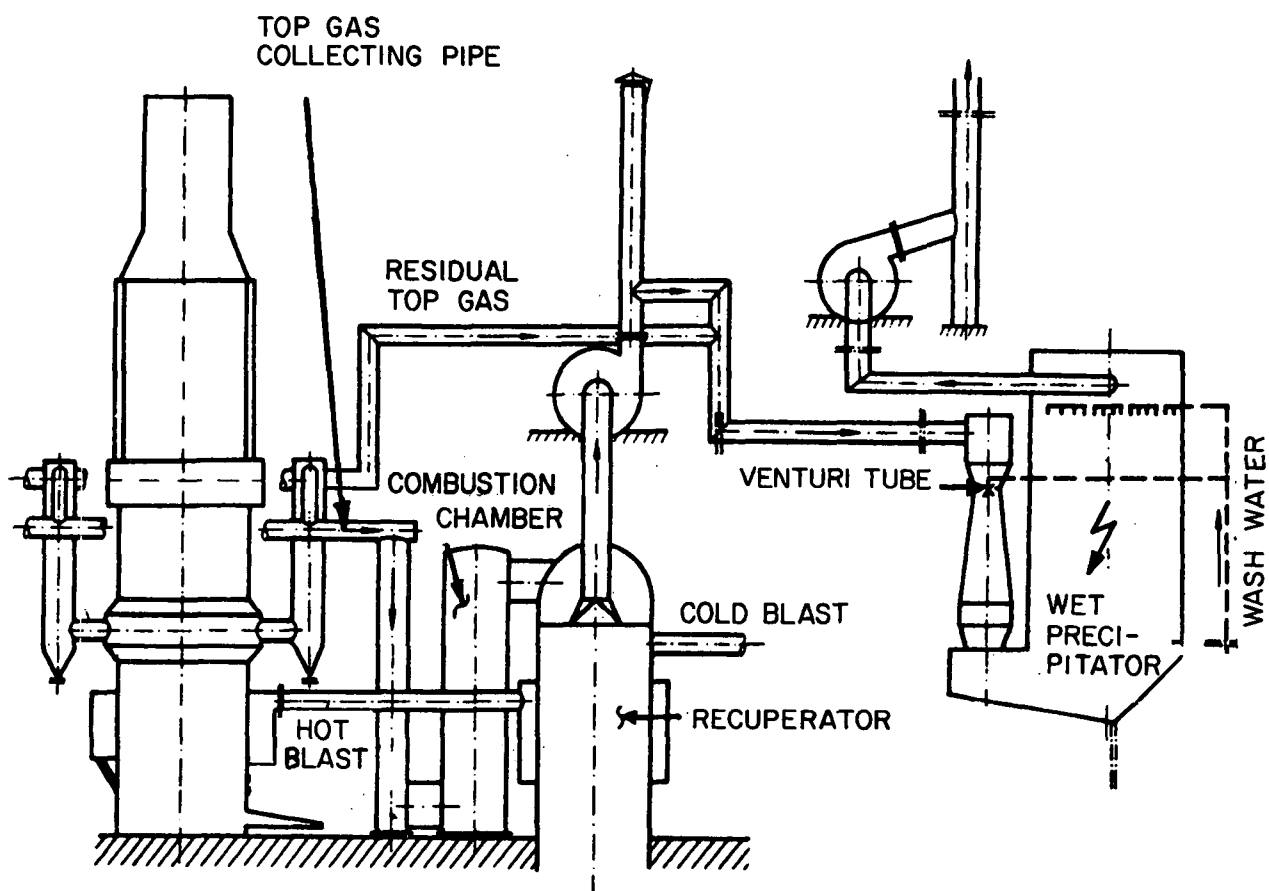
Figure 45. Waste and Top Gas Cleaning with
Wet Electrostatic Precipitator and
Venturi in Series.[34]

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Table 22. [34] Cost Comparison for Various Dust Collectors for Cleaning Waste Gases of a Hot Blast Cupola
(Index value 1 for venturi tube dust separation)
Waste gas quantity to be cleaned: 12500 $Nm^3/h$   Depreciation: 20% annum

| Plant | Venturi Tube Dust Collector | Wet Dust Collector, Simple Design | Dry Electro-precipitator Offer 1 | Offer 2 | Wet Electro-Precipitator | Fabric Dust Collector Offer 1 | Offer 2 |
|---|---|---|---|---|---|---|---|
| Guaranteed clean dust content: | | | | | | | |
| kg/h | 3.50 | 3.50 | 1.75 | 2.60 | 1.75 | 2.40 | 1.50 |
| g/$Nm^3$ | 0.280 | 0.280 | 0.140 | 0.2 | 0.140 | 0.192 | 0.120 |
| Initial investment costs (Index value) | 1.00* | 1.04* | 3.24 | 1.68 | 1.59* | 1.45 | 2.36 |
| Depreciation and interest (Index value) | 1.00* | 1.04* | 3.24 | 1.68 | 1.59* | 1.45 | 2.36 |
| Power consumption (Index value) | 1.00 | 0.62 | 0.23 | 0.36 | 0.24 | 0.65 | 0.43 |
| Maintenance, servicing, water consumption, neutralization (Index value) | 1.00 | 1.06 | 0.62 | 0.536 | 0.83 | 0.55 | 0.54 |
| Costs per operation hour (Index value) | 1.00* | 0.88* | 1.67 | 1.01 | 0.96 | 1.14 | 1.50 |

* without facilities for water supply and settling tank

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# MIST COLLECTION IN THE IRON AND STEEL AND IN THE PETROCHEMICAL INDUSTRY

Electrostatic precipitators are used as collectors of tars and oils in the iron and steel and petrochemical industries. These types of units do not usually employ forced irrigation of the collecting electrodes, but they are "wet" in the sense that the collected material is a fluid, and no rapping is employed. The literature survey has not revealed any significant new developments in this application area since the preparation of "A Manual of Electrostatic Precipitator Technology, Part II" in 1970.[37] Therefore, the following discussion is taken from the Manual and from Research Cottrell reports submitted to SRI under Subcontract No. H-6228.[38,39,63]

The use of mist precipitators is most prevalent in the iron and steel industry for collection of coke oven emissions. Coke, the chief fuel used in blast furnaces, is the residue after distillation of certain grades of bituminous coal. The two methods of producing metallurgical coke are the beehive and by-product processes. The beehive process was used for coke production in the U.S. until about 1917, and since there were just a few beehive oven installations where the waste heat from the products of combustion was used in steam generating units, no attempts at gas cleaning were made. The by-product process is presently used in the production of 98% of all coke, primarily because of its favorable economic aspects. Gas cleaning equipment is necessary in the by-product process since means are provided for full utilization and recovery of the gas and coal chemicals. In short, all the products of combustion and destructive distillation of the coal involved in the beehive process are vented directly to the atmosphere while those of the by-product process are withdrawn from the oven and processed to produce useful by-products. A coke oven flow diagram is shown in Figure 46.

In either type of oven, the beehive or by-product, the coke producing processing consists mainly of driving off certain volatile matter, leaving in the residue a high percentage of carbon mixed with relatively small amounts of impurities. The beehive oven, seldom used anymore, will not be discussed in this report. In the by-product coking process, the coal is heated in the absence of air, and the volatile material is piped to equipment which extracts the valuable ingredients. After the extraction process, some of the gas (heating value, 550 Btu/ft$^3$) returns to the ovens for use in heating the coking chambers and other plant processes. These ovens are rectangular in shape, usually 30 to 40 feet long, 6 to 14 feet high, and 11 to 22 inches wide. As many as 100 of them may be set together in a battery for ease in charging
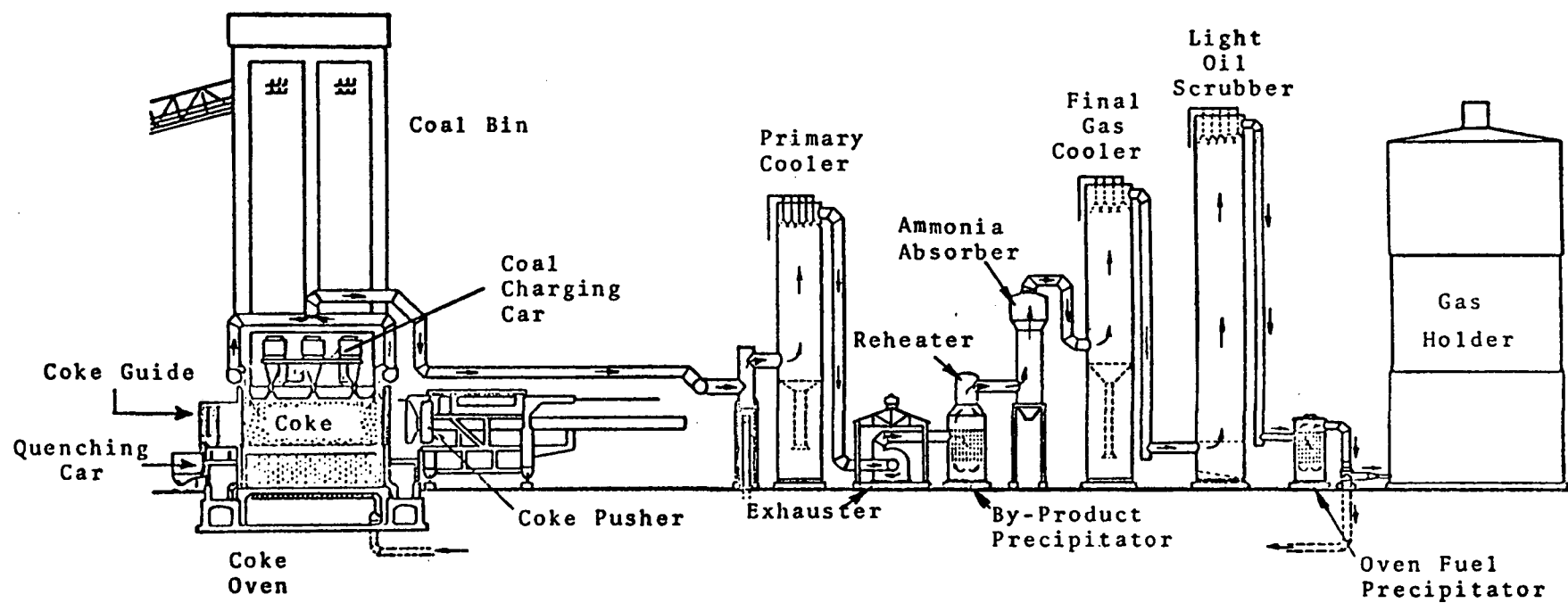
Figure 46.  Coke Oven Flow Diagram[38]

and discharging the coal and coke. Ports at the top of the oven can receive 16 to 20 tons of coal. The ports are then sealed and coal begins to fuse, starting at the walls of the oven, which may generate heat from 1600° to 2100°F. When coking is finished (18 to 20 hours), doors at the ends of the oven chamber are opened, and a pusher ram shoves the entire charge of the coke into railway cars.

As stated above, the products of destructive distillation in the by-product process are withdrawn from the oven and processed to produce useful by-products. Carbonization of one one of coal yields the following products:

| | |
|---|---|
| coke | 1200-1400 lbs |
| coke breeze | 100-200 lbs |
| coke oven gas (550 Btu/cf) | 9500-11,500 cf |
| crude light oil | 2-4 gallons |
| tar | 8-12 gallons |
| ammonium sulfate | 20-28 lbs |
| ammoniacal sulfate | 13-35 gallons |

All of the above products except the coke leave the oven in association with the gas. The analysis of the gas coming from the coke is approximately as follows:

| | |
|---|---|
| $CO_2$ | 2.2% |
| $O_2$ | 0.8% |
| $N_2$ | 8.1% |
| CO | 6.3% |
| $H_2$ | 46.5% |
| $CH_4$ | 32.1% |
| $C_2H_4$ | 3.5% |
| $C_6H_6$ | 0.5% |

The concentration of suspended matter at the precipitator inlet varies from 1 to 15 grains/scf. Water comprises about 50% of total precipitate, the remainder being tar, oil, etc.

Control Technology

As early as 1916 pilot scale electrostatic precipitators were demonstrating the technical feasibility of removing tar from coke oven gas, but fear of explosions caused by sparking caused reluctance on the part of industry. Successful operation eventually eliminated resistance, and in 1927 commercial installations in coke oven gas were first made and rapidly became standard equipment. The arrangement of equipment developed to remove and recover coke oven emissions is shown in Figure 46. The gas leaves the ovens at 1100 to 1300°F, after which it is cooled to about 95°F, thus condensing out a considerable amount of tar and ammoniacal liquor. These condensed products are processed further in other equipment for separation and refinement. The gas then passes through the exhauster and

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enters the electrostatic precipitator where the suspended oil and tar are collected with an efficiency normally between 95 and 99%. The gas then moves through reheaters, ammonia absorbers or saturators, gas coolers, and light oil scrubbers, before entering the gas holder. If the gas is to be used for underfiring the coke ovens, the gas must be further cleaned in another electrostatic precipitator referred to as a fuel gas precipitator, which cleans the gas sufficiently to prevent the deposition of tar in the very small apertures of the coke oven burners. Some of this gas is also used to ventilate the insulator compartments of the primary precipitator. The gas volume used for underfiring is usually a small fraction of the total gas, and consequently, only relatively small precipitators are required.

The typical electrostatic precipitator used for detarring consists of a group of grounded pipes six inches to eight inches in diameter and six feet to nine feet in length, suspended from a header plate in a round shell. Figure 47 shows the typical design for this type precipitator. The early coke oven precipitators were energized by electrical equipment using mechanical rectifiers, thus requiring a separate substation to house the electrical controls, transformers, and rectifiers. The more recent utilization of tube and silicon rectifiers permits the location of the rectifiers and transformers in weatherproof cabinets, thereby eliminating the necessity for substations. Figure 48 shows a self-contained tar precipitator attached to the shell of which is a compartment containing a 35 kV electrical set with tube rectifier. The collecting pipes of this unit are six inches in diameter and six feet long.

Another precipitator design used for cleaning coke oven gas is shown in Figure 49. The discharge electrode wires are centrally located in the annulus between the concentric tube-collection electrodes. Both surfaces of the tubes serve as collecting surface.

Based on data from a group of installations covering a period of 1931 to 1961, design efficiencies are generally about 95-99%. Typical design parameters for coke oven gas precipitators are given in Table 23.

TABLE 23[38]

TYPICAL COKE OVEN GAS PRECIPITATOR DESIGN PARAMETERS

148 pipes, 8 in. dia. x 9 ft. long
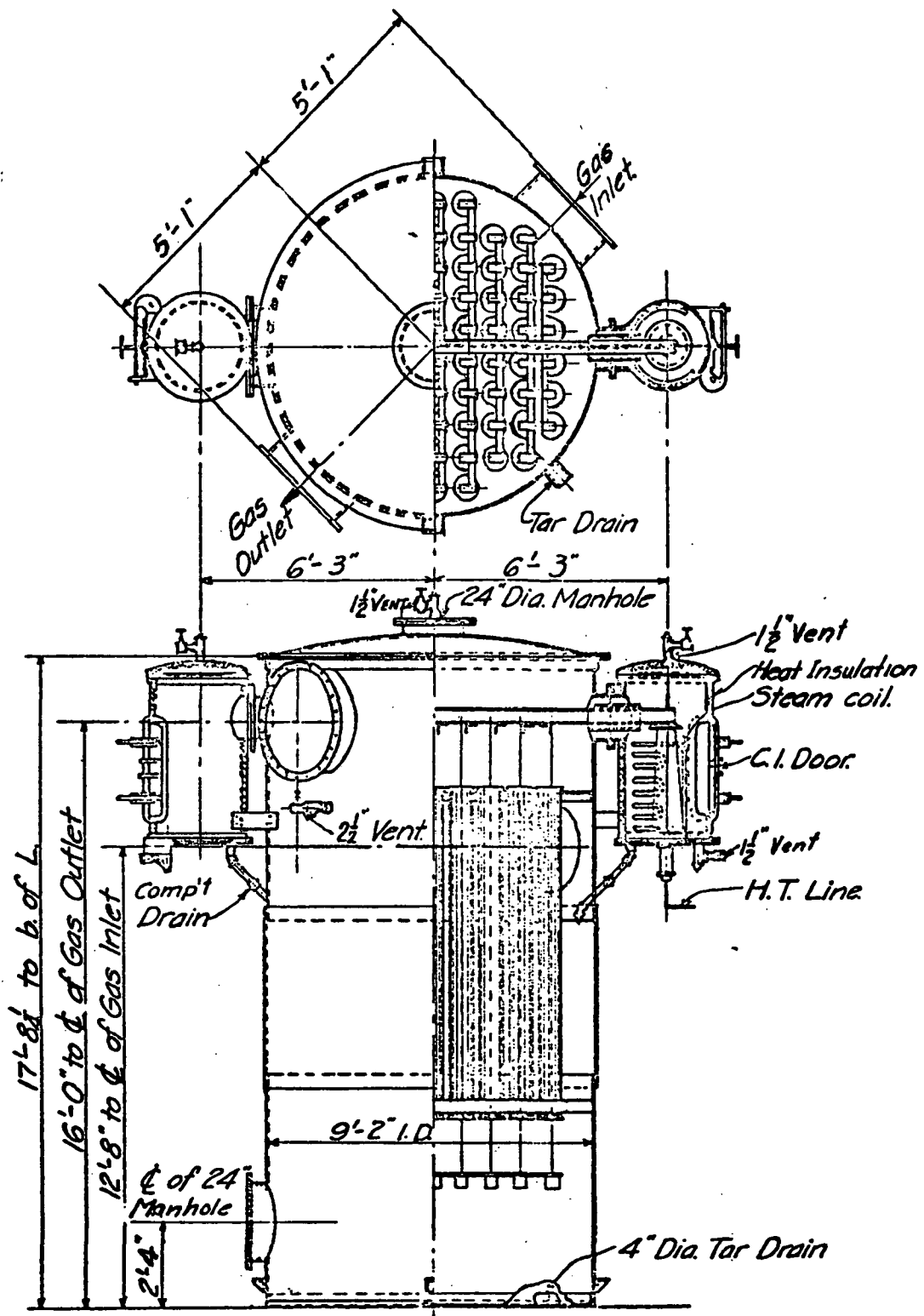
Figure 47. Typical Pipe Type Electrostatic
Precipitator for Collection of Tar.[38]

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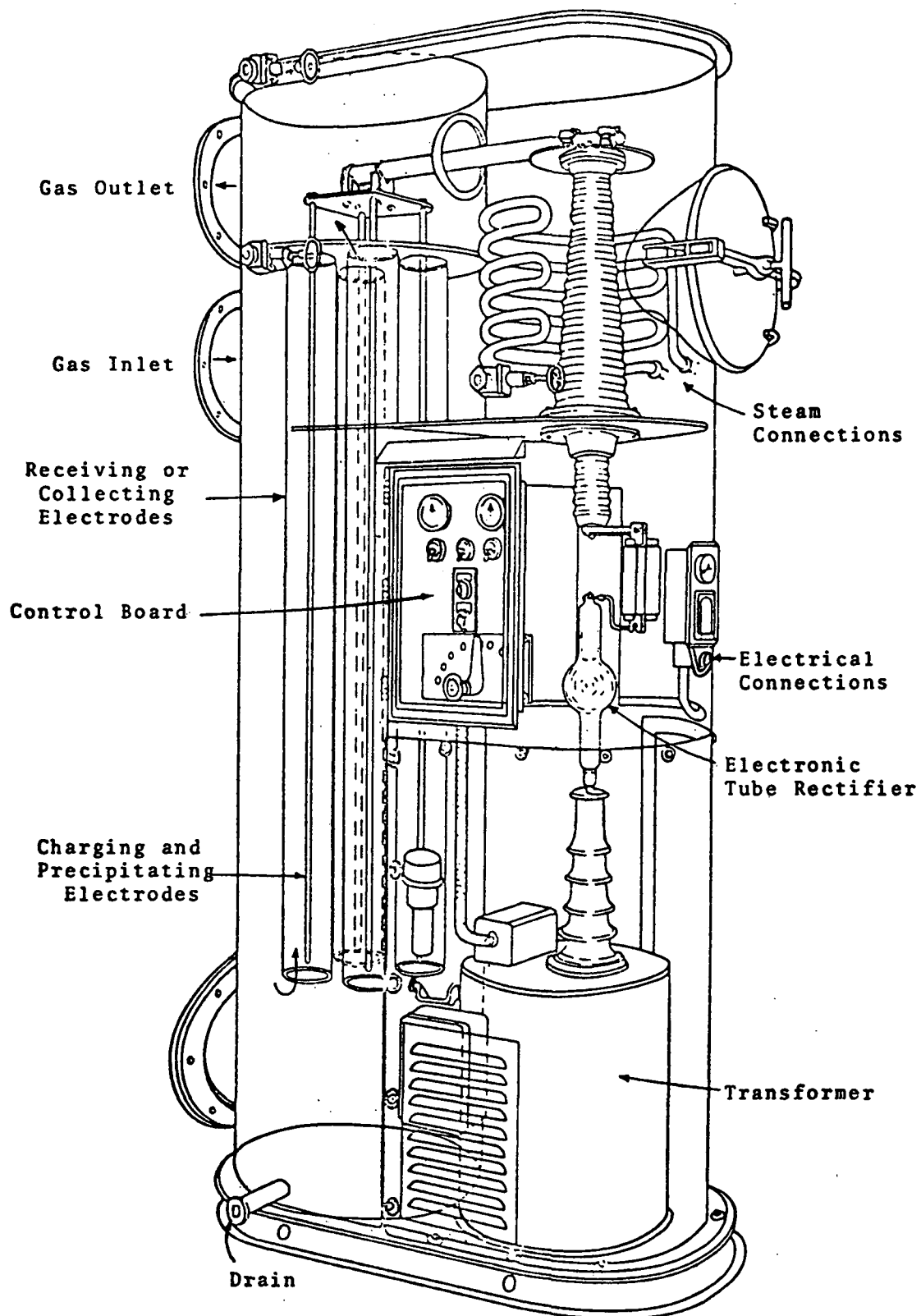
Figure 48. An Integral Tar Collecting Pipe
Type Precipitator.[38]

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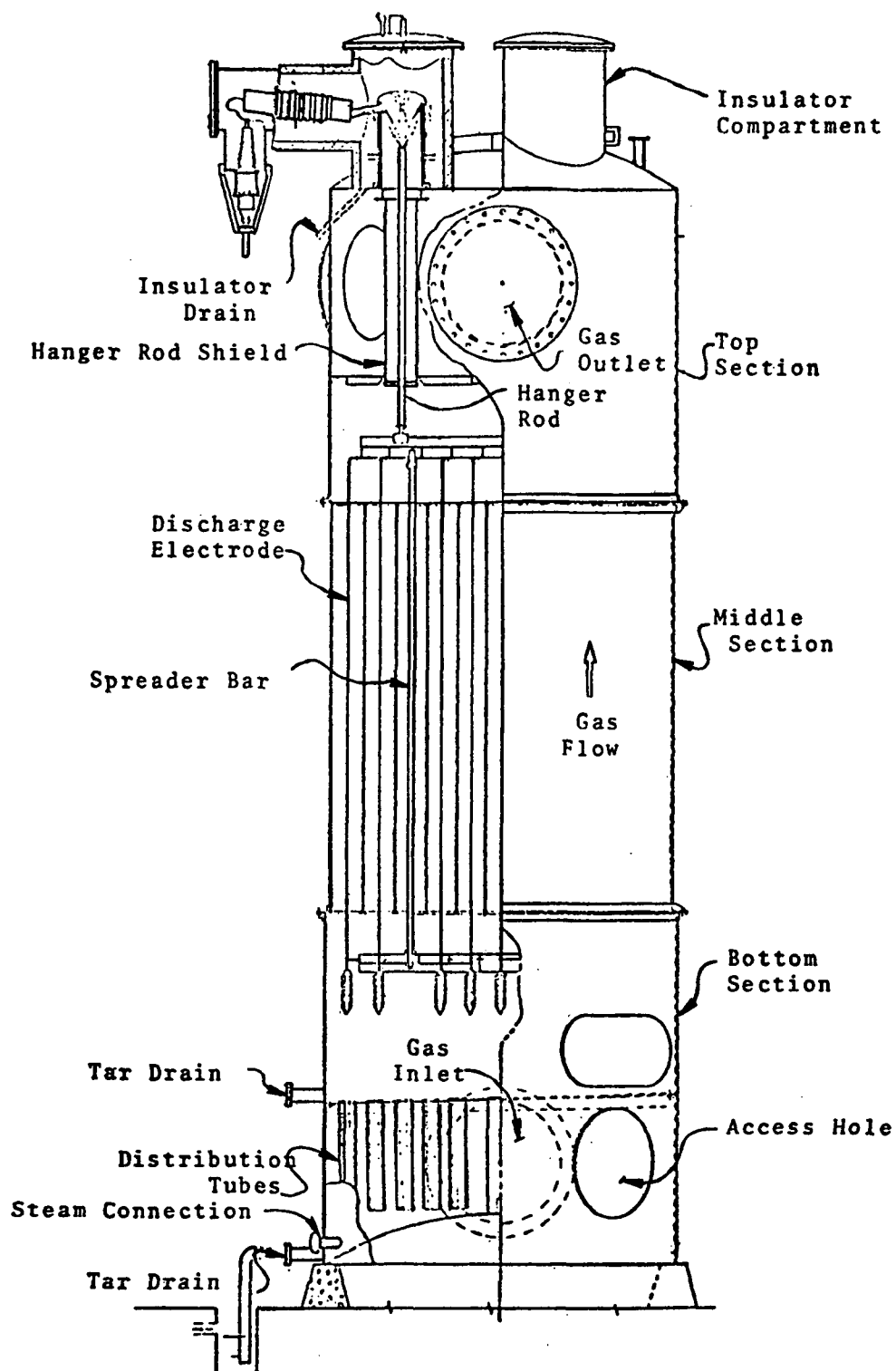
Figure 49. Concentric Ring Detarrer[38]

Gas flow = 16,500 cfm at 100°F
Inlet concentration = 0.68 gr/cf
Outlet concentration = 0.0068 gr/cf
Efficiency = 99%
Precipitation rate parameter = 13.7 cm/sec
Corona power = 190 watts/1000 cfm


Table 24 gives an average of electrostatic precipitator design parameters for detarring coke ovens. The ratio (R) is the ratio of performance to design precipitation rate parameters.

TABLE 24[38]

AVERAGE DESIGN PARAMETERS FOR COKE OVEN PRECIPITATORS

| No. of Installations | Avg. Gas Vel.(Ft/sec) | Avg. Ppt. Field Strength (kV/in.) | Avg. Ppt. Input Power (watts/1000 cfm) |
|---|---|---|---|
| 12 | 8.3 | 11.2 | 100 |

| Avg. Ppt. Gas Temp, °F | Avg. Ppt. Inlet Loading (gr/scfd) | Average Ratio (R) |
|---|---|---|
| 113 | 0.3 | 1.11 |


Average precipitator cost data both on an FOB and erected basis are presented in terms of cost per ACFM for the efficiency and gas volume indicated in Table 25. Numbers in parentheses are numbers of installations used in determining the costs indicated.

TABLE 25[38]

PRECIPITATOR ECONOMICS FOR DETARRING OF COKE OVEN GAS
(Time Period 1959-1969)

| Type Precipitator | Gas Volume Range - 100's ACFM | Eff. Range - % | Avg. Cost in Dollars/ACFM Erected | FOB |
|---|---|---|---|---|
| Pipe-type | 0-50 | 95 | 3.55(2) | 2.52(3) |
| (vertical flow) | 50-100 | 95 | 1.43(1) | 1.10(1) |

Because of the limited number of installations, the above data is only a rough indication of costs.

Detarring in the petrochemical industry is primarily concerned with the cleaning of water gas, oil gas, and producer gas. As in the case of tar removal from coke oven gas, the precipitation of tars from the gases of the petroleum industry is technically straightforward inasmuch as the precipitated tars are usually free flowing.

A relatively new application is the use of electrostatic precipitators to remove tar, fine carbon, and oil mist from the acetylene gas manufactured from naphtha or crude oil. Another relatively new application in the petrochemical field is the utilization of precipitators for the removal of tars from oil shale distillation gases. This application may become more important with the recent acceleration of oil production from shale. The purpose of the precipitator in this application is to clean the gases which are subsequently condensed and treated to produce petrochemical products.

In the process for manufacturing carburetted water or producer gas, the initial gas known as "blue gas" is made by passing steam through a bed of incandescent carbon, which may originate from either coke or anthracite coal. There are two cycles, designated in plant parlance as "make" and "blow". The coal or coke is fed to the gas generator and air is admitted under the bed during the "blow" cycle. The air is shut off and steam is admitted for the "make" cycle. The residue ash is then withdrawn. Carburetted blue gas is a mixture of blue gas (or water gas, as it is often termed), made as above, and oil gas formed by the cracking of oil in a chamber through which the blue gas passes.

The carburetting process enriches the blue gas to as much as 700 Btu per cubic foot of gas, depending upon the amount of oil used. With the present day use of natural gas, many companies mix natural gas with the above gases, the mixture being automatically controlled by means of calorimetric equipment. Prior to entering the precipitator the gas is passed through a direct contact water cooler where the temperature is reduced to around 100°F and saturated with water vapor. In the cooler some of the tars and oils contained

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in the gas are scrubbed out by the water sprays, and sulfur compounds and other gaseous material are removed from the system in purifiers located after the electrostatic precipitator. A process flow diagram of production of carburetted water gas is shown in Figure 50.

The process for separating oil from shale is shown in the flow diagram, Figure 51. The shale to be processed is fed into a retort where heat is applied to drive off the various components such as gas, oil, and carbonaceous residue. From the retort the gas is passed through indirect water coolers where most of the condensable material condenses out as a submicron sized mist. The mist-laden gas in then introduced to the precipitator where the condensed oil and carbonaceous residue are separated from the gas.

Natural gas is used as the primary raw material in the production of acetylene. The gas is burned in an oxygen-limited atmosphere under controlled temperature and pressure. The gaseous products consist of from 50 to 55% hydrogen, from 30 to 35% carbon monoxide, and from 7 to 9% acetylene. Because of the explosion hazards involved, the oxygen content is rigidly controlled to a fraction of 1%. In addition to the gases, submicron size particulate carbon is also produced and carried along by the gas as a smoke. Prior to entering the precipitator, the gas is passed through a direct-contact water cooler where the temperature is reduced to about 100°F. The gas then enters the wet-type precipitators where practically all of the remaining carbon is removed from the gas.

The Cottrell precipitator for detarring manufactured gases was first developed at the plant of the Ann Arbor Gas Company at Ann Arbor, Michigan, about 1914, while the first commercial application recorded was on producer gas in 1916 at the Minnesota Steel Company at Duluth, Minnesota.[40] In initial applications the gas was usually treated following the wash boxes or scrubbers at low temperature, so that the tar was collected in the presence of water. Early experimental and semi-commercial operations soon led to regular commercial installations. In 1924 there were five such installations in the United States cleaning about 70 million cubic feet per day. By 1962 there were 600 tar precipitators treating 5 million cfm of fuel gas. Since 1962 only about 25 additional tar precipitators have been installed with some of these being replacements for old precipitators.

The electrostatic precipitator most commonly utilized for detarring and cleaning of gaseous products is a single-stage vertical wire and pipe unit, as illustrated in Figure 52. A plate type unit is shown in Figure 53.[41]
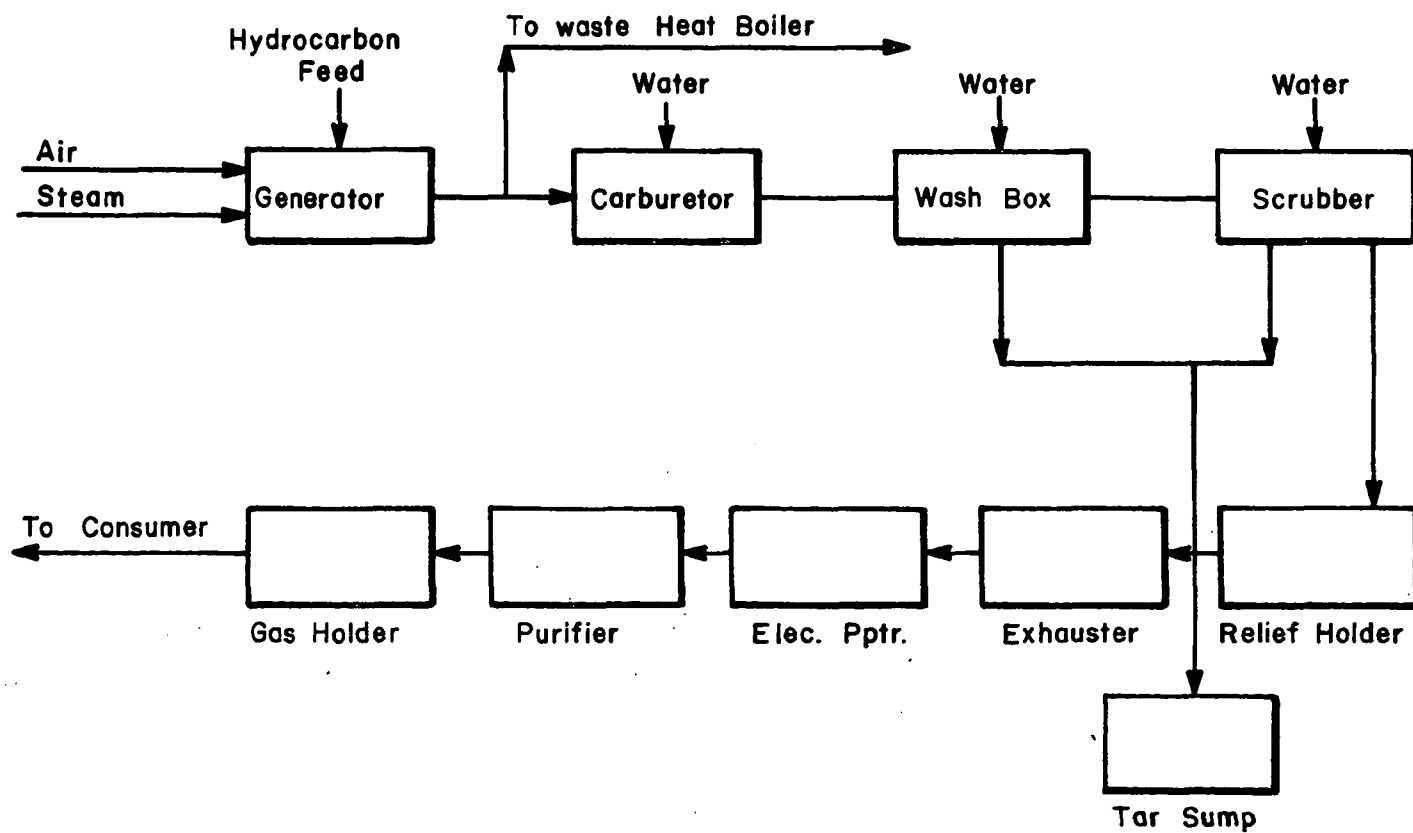
Figure 50. A Process Flow Diagram of Production of Carburetted Water Gas[63]
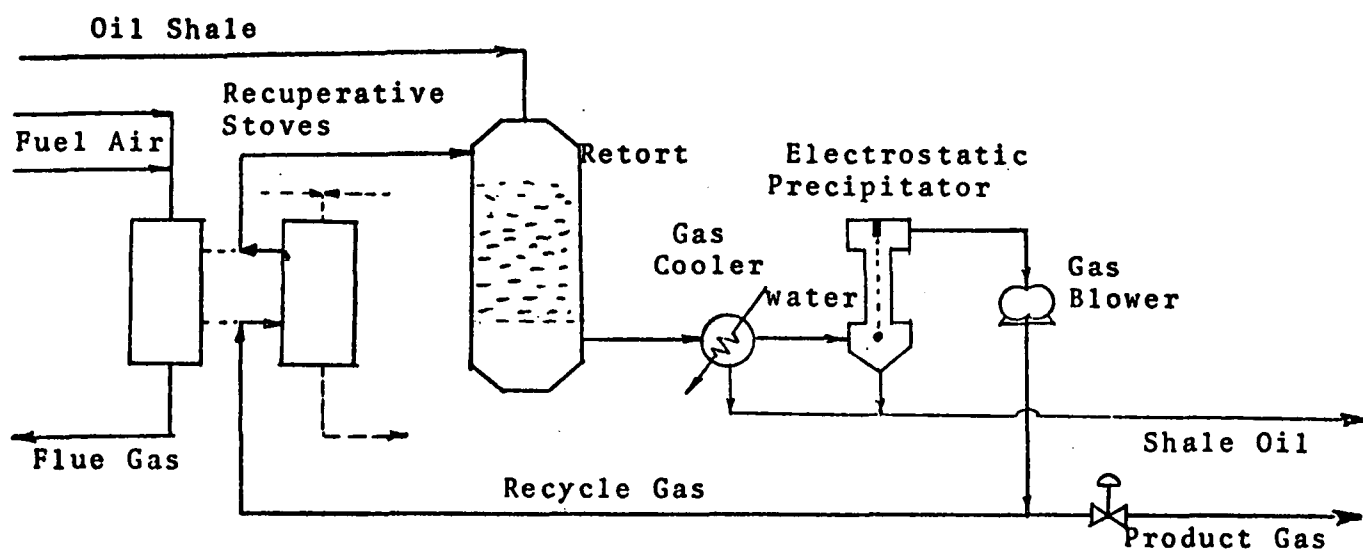
Figure 51. Oil Shale Retorting Process[63]
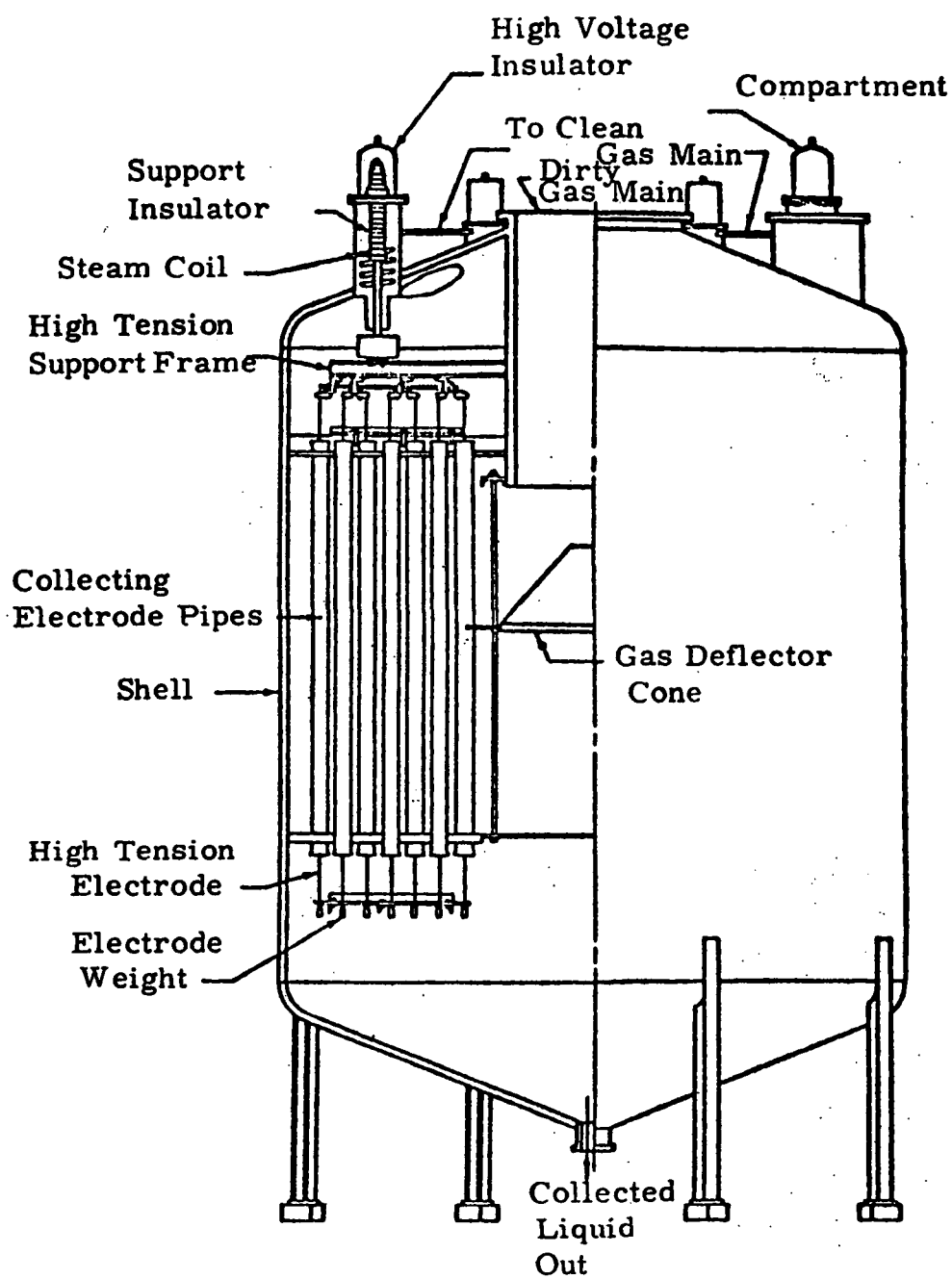
Figure 52. A Single-Stage Vertical Wire and Pipe Unit[38]

Central electrode centering gear

Chamber filled with penetrol oil

Interlocked earthing switch

H.T. cable with styrenated terminations

Steam connection

Interlocked hand hold

Steam pipe to trap

Gas outlet branch

Light hole

Central electrode (serrated)

Steam trap

Bottom electrode framework and gas distributor

Purge connection

Tar drain and seal pot

Electrode centering insulator

Main H.T. insulator

Thermometer

Steam coil

Gas baffel

Electrode flushing system

Top manhole

Flushing connection

Sight hole

Compression springs

Steel plate collecting electrodes

Stainless steel wire discharge electrodes

Gas inlet branch

Bottom manhole

Figure 53. Plate-Type Precipitator Used for Detarring[41]

The precipitator universally used for the removal of
tars and oils from carburetted water gas is a single stage
vertical pipe type unit. The design consists of the collecting
pipes, supported from a top header and immersed in the
incoming gas. The high voltage support insulators are
mounted in heated turrets attached to the side of the unit,
which is cylindrical in shape. The precipitator is provided
with gas purge connections at all of the high points in the
system. This type of unit is often referred to as a
"low-volume tar" precipitator. The discharge electrode used
in the low-volume tar precipitator is known as the "stiff rod
and prong" electrode. The most widely used form of this
type electrode consists of pointed star-shaped washers spaced
about 1 1/2 inches apart along a rigid rod for its full
length. Collecting electrodes are of the pipe type, the sizes
of which vary from 6" to 15" diameter. The discharge
electrode assembly is supported directly on the insulators
which are mounted on brackets inside of the precipitator
directly in the gas stream.

The collected tars and oil are free flowing and are
removed from the vessel sump to an external tar pot through
a liquid trap. From the tar pot the liquid is pumped to
its final point of reuse or disposal.

The precipitator most commonly used for cleaning
acetylene gas is a single-stage, vertical flow type unit.
To withstand the gas pressure the configuration of the
shell is cylindrical. The collecting pipes are supported
from a top header. The outer surfaces of the collecting pipes
are exposed to the incoming gas, while the inside surfaces
are continuously flushed by a film of water which overflows
weirs attached to the top of the pipe. The normal overflow
is augmented by periodic spraying from sprays located under
the roof.

Customarily there is only one type of discharge electrode
used in the treatment of acetylene gas, the "weighted twisted
squares" or weighted bar electrodes. These electrodes consist
of vertically hung square bars spanning the full height of the
collecting electrodes. They are usually 3/16" or 1/4" square
and are twisted longitudinally, held by weights at the bottom.
Because of the moisture content of the gas treated in the
precipitator, the discharge electrode support system consists
of heated outboard compartments. In this arrangement the
discharge electrode frame is supported by a horizontal beam
running across the top of the precipitator. The beam is
supported by two insulators, one on either side, located
in heated compartments attached to the outside of the precipi-
tator shell.

The precipitator used to collect the oil and carbonaceous
material from shale oil manufacture consists of a single stage
vertical flow unit using pipes for the collecting surface.
The vessel construction is cylindrical. Since the temperature
of the gas being treated approaches atmospheric, no thermal
insulation is used. The collected liquid accumulates in a sump
provided at the bottom of the shell. The discharge electrodes
for shale oil application are of the weighted wire and stiff
rod and prong types, and the support frame used is the heated
outboard compartments method. The use of pipe type collecting
electrodes is common for the small sized precipitators normally
used in shale oil production.

Table 26 presents a summary of design data for electrostatic
precipitator installations in fuel gas purification from 1940
to 1963.

COLLECTION OF ACID MISTS FROM SULFURIC AND PHOSPHORIC ACID
PRODUCTION

Sulfuric Acid Production

Basically, the production of sulfuric acid involves the
oxidation of generated sulfur dioxide to sulfur trioxide and
its hydration to sulfuric acid. Sulfuric acid capacity in
1970 amounted to 38 million short tons produced in 250 plants
of which 215 are of the contact type, the remaining 35 being
the older chamber type.

The chamber process uses the reduction of $NO_2$ to NO as the
oxidizing mechanism to convert the $SO_2$ to $SO_3$. The oxidation takes
place in lead-lined chambers. However, the process produces
sulfuric acid with a concentration of only about 78% and this
low concentration together with high operational costs and $NO_x$
emissions, have caused an ever diminishing use of the chamber
process. For these reasons, this discussion will deal
exclusively with the contact process of sulfuric acid manu-
facture.

The chemistry of the contact process is very simple.
First, the sulfur in the feedstock is burned in air, giving
sulfur dioxide:

$$S + O_2 \rightarrow SO_2$$

Then the sulfur dioxide is catalytically oxidized to sulfur
trioxide:

$$2SO_2 + O_2 \rightarrow 2SO_3$$
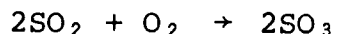
TABLE 26[63]

A SUMMARY OF DESIGN DATA OF ELECTROSTATIC PRECIPITATORS ON VARIOUS APPLICATIONS
FOR REMOVING TAR AND OIL MIST OVER PERIOD 1940 to 1963

| Type of Applications | No. of Install. | Suspended Matter | Total Gas Vol. ACFM | Temperature °F | Collecting Efficiency % |
|---|---|---|---|---|---|
| Carburetted Water Gas | 55 | Tar and Oil Mist | 315,800 | 70-110 | 95 |
| Oil Gas | 3 | Tar and Oil Mist | 18,200 | 80-100 | 95-98 |
| Reformed Gas | 3 | Tar and Oil Mist | 3,200 | 80-100 | 95-98.5 |
| Shale Oil Gas | 2 | Tar and Oil Mist | 20,900 | 100-200 | 95-97.5 |
| Acetylene | 1 | Tar and Oil Mist | 42,100 | 100 | 99-92 |

Finally, the sulfur trioxide is absorbed in a strong, aqueous solution of sulfuric acid:

$$SO_3 + H_2O \rightarrow H_2SO_4$$

Contact plants are often classified according to the raw materials charged to them; e.g., elemental sulfur, spent acid, and hydrogen sulfide, and sulfide ores and smelter gas. The sulfur burning plants are sometimes called hot-gas purification plants, while those plants which utilize sulfide ores are called metallurgical or cold gas purification acid plants. Plants burning hydrogen sulfide may be of the hot-gas or true wet-gas purification type, the latter case being where the sulfur dioxide gas is not dried, thus allowing the moisture to pass through the conversion system.

Sulfur burning plants require no gas purification (except for hot filtration) and are the most inexpensive of the contact type plants. No heat is required for the $SO_2$ gases, only cooling, so that all heat evolved may be recovered as relatively high pressure steam. A flow chart for this system is shown in Figure 54, and can be divided into the following sequences:[42]

Transportion of sulfur to the plant

Melting of sulfur

Pumping and atomizing of melted sulfur

Burning of sulfur

Recovery of heat from or cooling of hot $SO_2$ gas

Purification of $SO_2$ gas (by hot filtration)

Oxidation of $SO_2$ to $SO_3$ in converters

Temperature control to secure good yields of $SO_3$

Absorption of $SO_3$ in strong acid

Cooling of acid from absorbers

Pumping of acid over absorption towers

Two types of plants are used to process sulfuric acid produced by burning hydrogen sulfide and spent-acid. In one, the sulfur dioxide and the combustion products are passed through gas-cleaning and mist removal equipment. Mist removal is usually accomplished by electrostatic precipitation and moisture removal by absorption in concentrated sulfuric acid. The gas stream then passes through

Exit gas stack

DRYING
TOWER

Air intake
silencer
or filter

To boiler

ABSORBING
TOWER

Steam

Air

Air

Feedwater

SULFUR
BURNER

Air

Turbo-    Starting
blower    fan

WASTE HEAT
BOILER

HOT GAS
FILTER

HEAT
EXCHANGER

4-PASS
CONVERTER

ECONOMIZER

OLEUM
TOWER

98-99% acid to storage

Absorbing acid
cooler

Oleum
to storage

Sulfur

Molten sulfur

Drying acid
cooler

Melter-
settler

Steam

Oleum
cooler

Circulating
tank

Oleum pump

| | | |
|---|---|---|
| Sulfur | 0.337 tons | |
| Water | 4,400 gal. | |
| Electricity | 10 kwh | Per ton of 100% acid in plant |
| Direct operating labor | 0.16 man-hr. | of 200 tons daily capacity |
| Steam credit | 2000 lbs. | |
| Repairs (labor and materials) | 60¢ | |

Figure 54. Typical flowchart for sulfur-burning contact plant.[42]

a drying tower from which a blower draws the gas and discharges
it to the sulfur trioxide converter. A schematic of the contact
process sulfuric acid plant burning spent acid is shown in Figure 55.[43]
In the other type plant, known as a "wet-gas plant", the wet gases
from the combustion process are charged directly into the converter
after heat recovery with no intermediate treatment. A highly
efficient mist recovery system following the absorber is required
since absorption is not highly efficient due to the excess moisture
and acid mist content of the gas.

The configuration of sulfide ore and smelter gas plants
is very similar to that of a spent-acid plant except that
a roaster is used in place of the combustion furnace. The
raw material used in these plants is smelter gas available
from ore roasting, smelting, and refining operations associated
with nonferrous metals production. The sulfur dioxide in the
gas is contaminated with dust, acid mist, and gaseous impurities,
the removal of which is accomplished by cyclone dust collectors,
scrubbers and electrostatic precipitators (wet, dry, or mist
type). After the gases are cooled and cleaned and the excess
water vapor removed, they are scrubbed with 98 percent acid
in a drying tower. Beginning with the drying tower stage,
the process is nearly identical to that of sulfur burning
plants. When sulfide ores are the source of sulfur, the cost
may be as much as three times that of the sulfur burning plant.


Effluent from Contact Plants

Exit gas from the absorber accounts for the major portion
of emissions from contact sulfuric acid plants. This gas
contains unreacted sulfur dioxide, sulfuric acid spray and mist,
and unabsorbed sulfur trioxide. Tail gases containing $SO_3$
hydrate form a finely divided mist upon contact with atmospheric
moisture. The particle size of these mists can range from
submicron to 10 microns and larger. Smaller particles have
greater light-scattering effect, so as particle size decreases,
the plume becomes more dense. Contact plant emissions range
from 7 to 95 wt. % less than 3 μ. A report by G. R. Gillespie[44]
gives particle size distribution data of sulfuric acid from
the waste gases of a contact acid plant, and also of sulfuric
acid made from $SO_3$ and water under laboratory conditions,
as measured with a jet impactor. The field work was conducted
at the Krummrich Contact Plant in Monsanto, Illinois, and the
gases were sampled from a point in the exit stack of two units
approximately three feet from the absorption towers. Two
samples were extracted from each of the two units by a sampling
train consisting of a cyclone, a four-stage impactor, and a fine
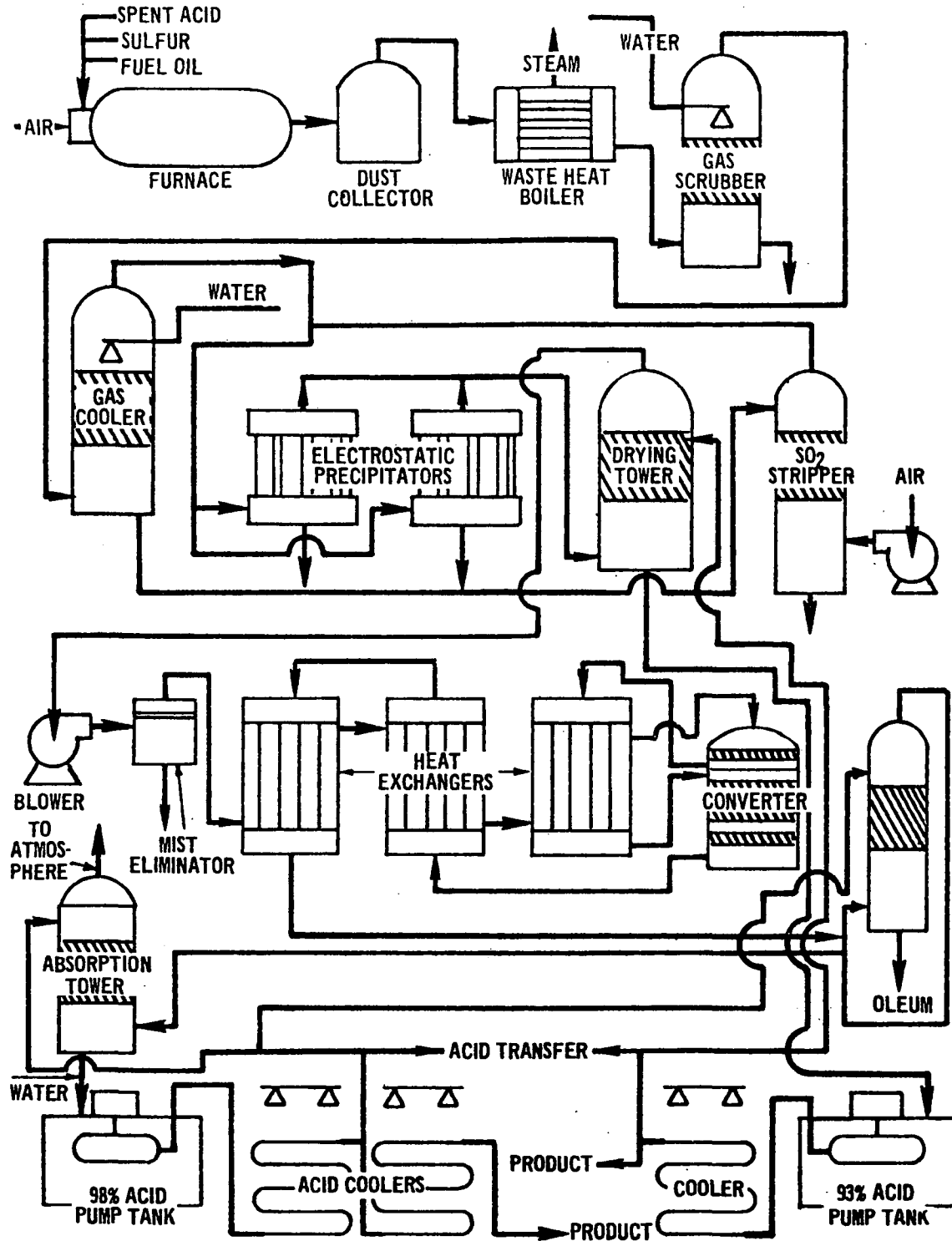
Figure 55. Basic flow diagram of contact-process sulfuric acid plant burning spent acid. [43]

glass wool filter. The size fractionations of the impactor were
1.36, 0.77, 0.39, and 0.21 microns (diameter) and the size
distributions obtained are shown in Figure 56. The flow rates
(at 1 atmos. and 150°F) in the two stacks were 6800 cfm and
15,700 cfm with respective mist loadings of 0.73 mg/l and 0.62
mg/l.

## Precipitator Design for Sulfuric Acid Plants

Electrostatic precipitators are highly effective when used
for collection of acid mist. Typically, the design for an acid
mist precipitator is of the wire-in-tube type shown in Figure 57
though the wire-and-plate design is common in Europe. The wire-
in-tube electrostatic precipitator shown in the figure contains
lead alloy tubes mounted between lead-covered tube plates. The
discharge electrodes consist of lead-covered steel cable in which
the cross-section of the sheath forms a six-pointed star to give
sharp edges to enhance corona discharge. The ionizing electrodes
are suspended from metal bars attached to lead-covered cross-
pieces supported by the heavy high-tension leads. Because of
possible shorting-out of the high voltage system, the sealing
of the high tension leads is of paramount importance. This
can be accomplished by oil seals, but, because of the risk of
fire, these have been superseded in many cases by a closed
chamber at the top of the precipitator which houses the insulators
carrying the lead-in wire and electrode connection. Condensation
of moisture on the insulator surface is prevented by feeding
dried and purified air into the chamber. Steam coils with a
water-repellant silicone film can be fitted around the insulators
to accomplish the same purpose.

Gas flows for acid mist precipitators range from about 10,000
to 30,000 cfm at gas temperatures of 100°F to 180°F. To meet
these requirements, together with a high collection efficiency,
the following basic design parameters are typically used:

| Basic Parameter | Range |
|---|---|
| Precipitation rate | 0.20-0.30 ft/sec |
| Specific collection surface | 200-400 sq ft/1000 cfm |
| Pipe diameter | 10 in. |
| Gas velocity | 3-6 ft/sec |
| Corona power | 100-500 watts/1000 cfm |
| Corona power density | 0.5-2.0 watts/sq ft |
| Rectifier voltage | 75-100 kV peak |

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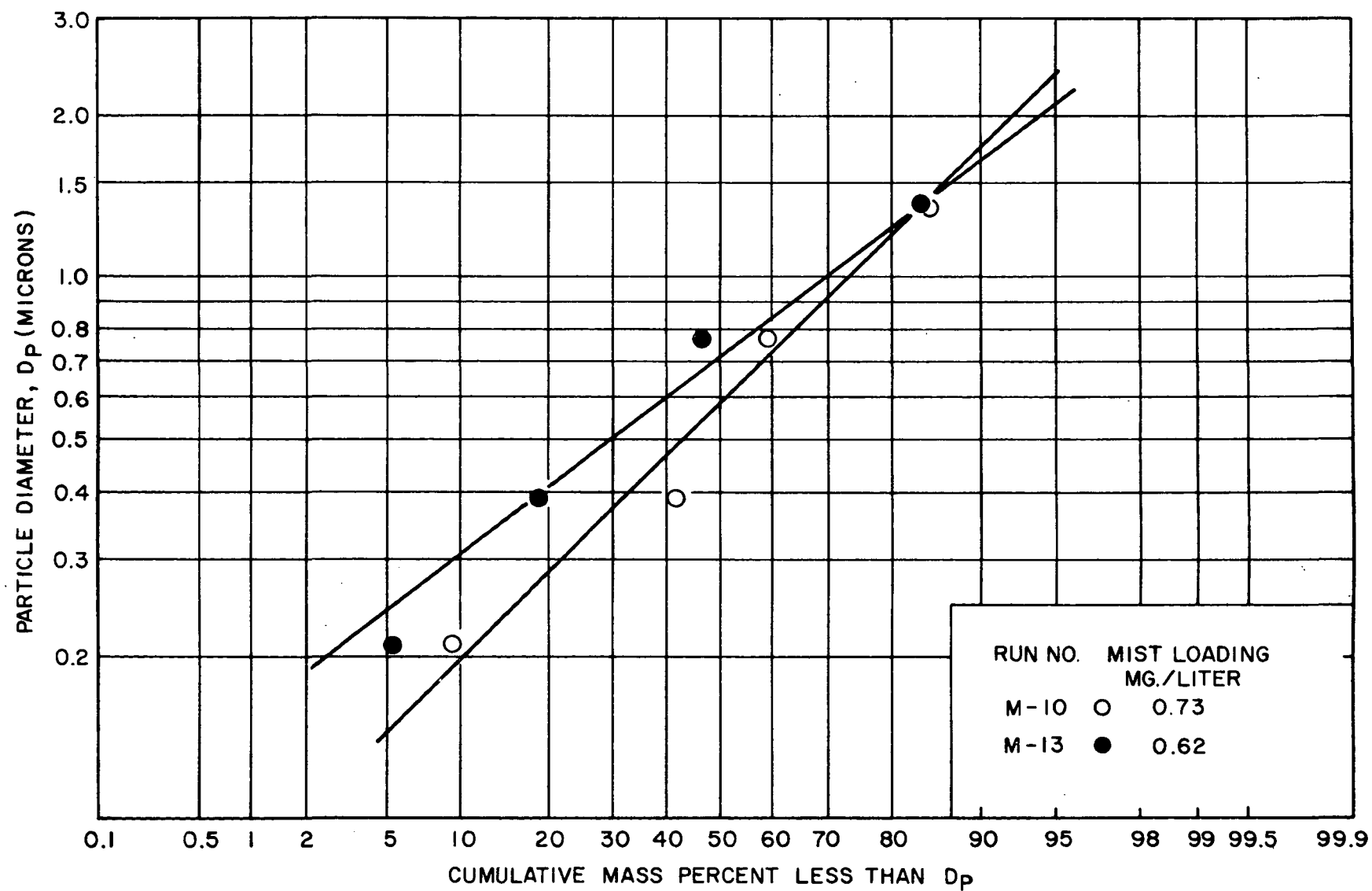
Figure 56. Particle size distribution of sulfuric acid mist from commercial contact plants.[44]
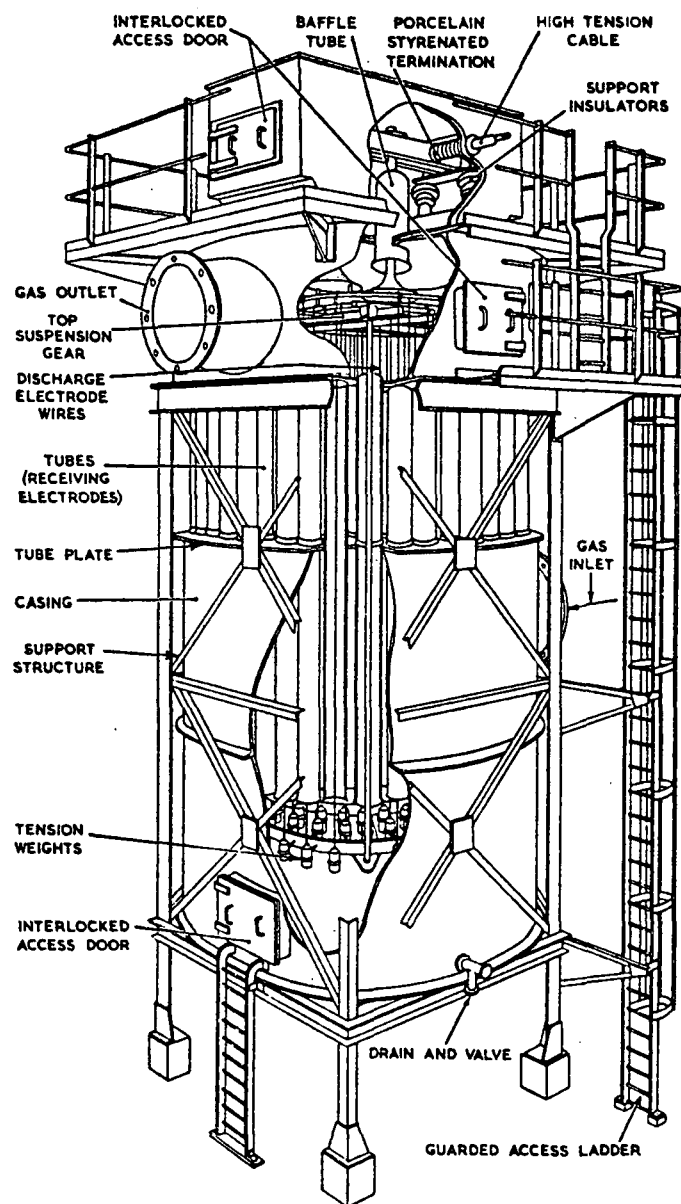
Figure 57.  Wire-in-tube acid mist precipitator.[39]

Very little design data concerning plate-type mist precipitators has been uncovered in the literature search. However, the Air Pollution Engineering Manual describes briefly a lead-lined, two-stage precipitator designed to handle 20,000 cfm tail gas from a 300 ton/day contact sulfuric acid plant.[45] Conditioned gas flows to the precipitator ionizing section, which consists of about 75 grounded curtain electrodes and 100 electrode wire extensions. The gas then flows to the precipitation section where there are twelve 14' x 14' lead plates and 375 electrode wires supplied 75,000 volts DC by a battery of silicon rectifiers. All structural material in contact with the acid mist is lead clad, and electrical wires are stainless steel cores with lead cladding.

## Performance Statistics and Cost Data for Sulfuric Acid Mist Precipitator Installations

Research Cottrell has summarized critical process and electrostatic precipitator operating parameters, and these data are given in the following graphs and tables.[46] Tables 27 through 31 summarize for the years 1945-1969 the precipitator gas velocity, inlet mist loading, inlet gas temperature, and input power and field strength. The number of installations and precipitators, and total gas volumes involved in the statistical analyses are also presented.

Precipitator cost data on an installed basis are presented in terms of cost versus gas flow rate for several efficiency levels for the period 1960 to 1969. Limited data on FOB precipitator cost are included. The data are summarized in Table 32 and Figure 58. The spread in data is indicated by the arrows on the figure. The scatter in cost data, at a given efficiency and gas rate, can be attributed to many factors, the most important ones being the size of the precipitator as related to mist characteristics, primarily particle size, and gas conditions; the geographical location of the installation (particularly for erected cost since labor cost can vary considerably across the country); the pricing-profit policy of various corporations bidding on the job; whether the installation is a backfit or upgrading of an existing installation which may require additional improving to "shoehorn" the precipitator into the overall installation; and the type and degree of sophistication of the electrical system.

The performance of sulfuric acid precipitators has been reported in terms of the ratio of the actual value to the design value of precipitation rate parameter (w). The ratio (R) is defined as the ratio of performance w to the design w and is therefore the ratio of actual performance to the anticipated performance. For instance, a value of R = 1.20 indicates that the precipitator size is 20% larger than required to meet the guarantee efficiency for that particular unit. Table 33 summarizes the data used in

# TABLE 27[39]

## PRECIPITATOR INLET MIST LOADING

### Summary of Performance Statistics
### Sulfuric Acid Mist Precipitator
### (1945 - 1969)

| Pptr. Inlet Load gr/SCFD | No. of Install. | No. of Pptr. | Pptr. Capacity Thousand ACFM In Interval | % of Install. In Loading Interval |
|---|---|---|---|---|
| 0-0.19 | 1 | 1 | 16.5 | 4.3 |
| 0.2-0.39 | 3 | 4 | 31.9 | 13.0 |
| 0.4-0.59 | 4 | 8 | 122.3 | 17.4 |
| 0.6-0.79 | 3 | 5 | 88.5 | 13.0 |
| 0.8-0.99 | 3 | 6 | 79.6 | 13.0 |
| 1.0-1.19 | 0 | 0 | 0 | 0 |
| 1.2-1.39 | 0 | 0 | 0 | 0 |
| 1.4-1.59 | 3 | 4 | 57.9 | 13.0 |
| 1.6-1.79 | 1 | 2 | 36 | 4.3 |
| 1.8-1.99 | 1 | 2 | 28.1 | 4.3 |
| 2.0-2.19 | 1 | 2 | 33 | 4.3 |
| 2.2-2.39 | 0 | 0 | 0 | 0 |
| 2.4-2.59 | 2 | 2 | 27 | 8.7 |
| 4.0-4.19 | 1 | 1 | 10.2 | 4.3 |
| Totals | 23 | 37 | 531 | 100 |

TABLE 28[39]

Precipitator Inlet Gas Temperature

Summary of Performance Statistics
Sulfuric Acid Mist Precipitator
(1945-1969)

| Inlet Gas<br>Temp. °F | No. of<br>Install. | No. of<br>Pptr. | Pptr. Capacity<br>Thousand ACFM<br>In Interval | % of<br>Install.<br>In<br>Temperature<br>Interval |
|---|---|---|---|---|
| 60-79.9 | 0 | 0 | 0 | 0 |
| 80-99.9 | 6 | 11 | 139 | 26.1 |
| 100-109.9 | 5 | 5 | 37.9 | 21.7 |
| 120-139-9 | 1 | 2 | 12.0 | 4.3 |
| 140-159.9 | 4 | 9 | 95.1 | 17.4 |
| 160-179.9 | 7 | 8 | 221.8 | 30.4 |
| 180-199.9 | 0 | 0 | 0 | 0 |
| Totals | 23 | 35 | 505.8 | 100 |

## TABLE 29[39]

### Precipitator Input Power

Summary of Performance Statistics
Sulfuric Acid Mist Precipitator
(1945 - 1969)

| Input Power, Watts/1000 ACFM | No. of Install. | No. of Pptrs. | Pptr. Capacity Thousand ACFM In Interval | % of Install. In Power Interval |
|---|---|---|---|---|
| 0-99.9 | 0 | 0 | 0 | 0 |
| 100-199.9 | 2 | 2 | 20.7 | 15.4 |
| 200-299.9 | 2 | 4 | 41.0 | 15.4 |
| 300-399.9 | 3 | 5 | 101.0 | 23.1 |
| 400-499.9 | 2 | 3 | 40.1 | 15.4 |
| 500-599.9 | 1 | 1 | 10.2 | 7.7 |
| 600-699.9 | 2 | 2 | 47.8 | 15.4 |
| 700-799.9 | 0 | 0 | 0 | 0 |
| 800-899.9 | 0 | 0 | 0 | 0 |
| 900-999.9 | 1 | 1 | 8.0 | 7.7 |
| Totals | 13 | 18 | 268.8 | 100 |

## TABLE 30.39

### PRECIPITATOR FIELD STRENGTH

Summary of Performance Statistics
Sulfuric Acid Mist Precipitator
(1945-1969)

| Field Strength KV/inch | No. of Install. | No. of Pptrs. | Pptr. Capacity Thousand ACFM In Interval | % of Install. In Interval |
|---|---|---|---|---|
| 8-8.4 | 0 | 0 | 0 | 0 |
| 8.5-8.9 | 1 | 2 | 17.0 | 7.1 |
| 9-9.4 | 1 | 3 | 42 | 7.1 |
| 9.5-10.9 | 0 | 0 | - | 0 |
| 10-10.4 | 3 | 6 | 127.8 | 21.4 |
| 10.5-10.9 | 0 | 0 | - | 0 |
| 11-11.4 | 2 | 3 | 25.8 | 14.3 |
| 11.5-11.9 | 2 | 4 | 61 | 14.3 |
| 12-12.9 | 3 | 4 | 59.1 | 21.4 |
| 12.5-12.9 | 2 | 3 | 43.5 | 14.3 |
| Totals | 14 | 25 | 376.2 | 100 |

# TABLE 31[39]

## Precipitator Gas Velocity

### Summary of Performance Statistics
Sulfuric Acid Mist Precipitator
(1945-1969)

| Velocity Interval FPS | No. of Install. | No. of Pptrs. | Pptr. Capacity Thousand ACFM In Velocity Interval | % of Install. in Velocity Interval |
|---|---|---|---|---|
| 1.5-1.9 | 0 | 0 | 0 | 0 |
| 2-2.4 | 1 | 1 | 28.75 | 4.8 |
| 2.5-2.9 | 1 | 11 | 8 | 4.8 |
| 3-3.4 | 6 | 2 | 143.3 | 28.6 |
| 3.5-3.9 | 2 | 4 | 23.2 | 9.5 |
| 4-4.4 | 2 | 8 | 58.2 | 9.5 |
| 4.5-4.9 | 5 | 8 | 127.9 | 23.8 |
| 5-5.4 | 0 | 0 | 0 | 0 |
| 5.5-5.9 | 1 | 2 | 33 | 4.8 |
| 6-6.4 | 1 | 2 | 25 | 4.8 |
| 6.5-6.9 | 1 | 2 | 25 | 4.8 |
| 7-7.4 | 0 | 0 | 0 | 0 |
| 7.5-7.9 | 1 | 1 | 16.5 | 4.8 |
| Totals | 21 | 42 | 488.9 | 100 |

TABLE 32[39]

Summary of Sulfuric Acid Mist
Precipitator Costs 1960-1969

| Efficiency Range | 0 - 10,000 cfm | | | | 10,000 - 25,000 cfm | | | | 25,000 - 50,000 cfm | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | 1965-1969 | | 1960-1965 | | 1965-1969 | | 1960-1965 | | 1965-1969 | | 1960-1965 | |
| | FOB* | Erected* | FOB | Erected | FOB | Erected | FOB | Erected | FOB | Erected | FOB | Erected |
| 90 - 95 | - | - | - | - | - | - | - | - | - | 2.21 (1) | - | - |
| 95 - 99 | - | - | 14.5 (1) | 6.00 (1) | 3.77 (1) | 5.84 (4) | - | - | - | 4.65 (1) | - | 3.94 (2) |
| 99+ | - | - | - | - | - | - | - | - | - | - | - | - |

| Efficiency Range | 50,000 - 100,000 cfm | | | | > 100,000 cfm | | | |
|---|---|---|---|---|---|---|---|---|
| | 1965-1969 | | 1960-1965 | | 1965-1969 | | 1960-1965 | |
| | FOB | Erected | FOB | Erected | FOB | Erected | FOB | Erected |
| 90 - 95 | - | - | - | - | - | - | - | - |
| 95 - 99 | - | 6.22 (1) | 2.93 (1) | - | - | 2.53 (5) | - | - |
| 99+ | - | 3.71 (3) | - | - | - | - | - | - |

*Note: Costs are $/acfm. Number in parentheses is number of installations on which contract prices were averaged.
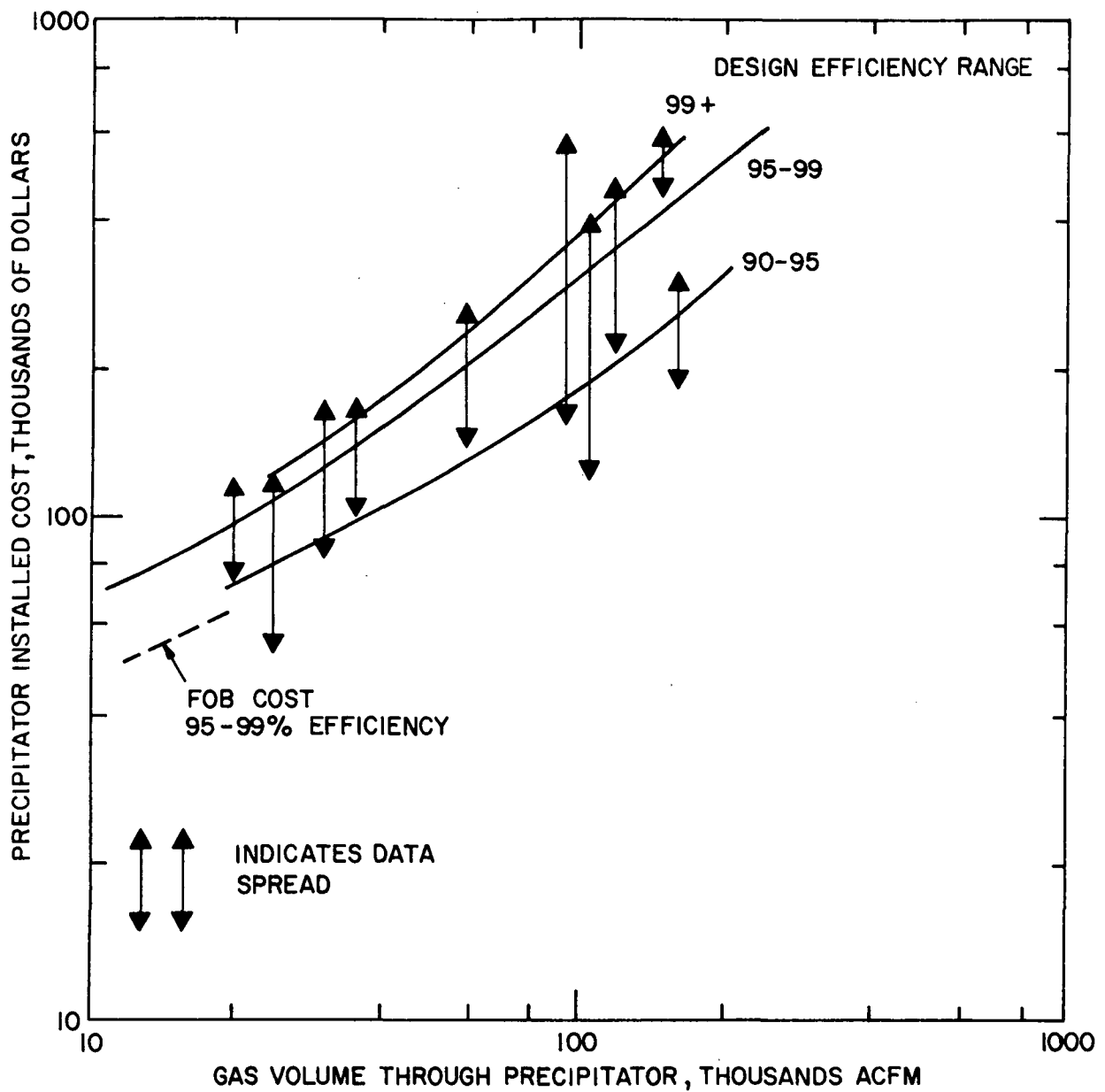
Figure 58. Installed Cost of Sulfuric Acid Mist Precipitators, 1965-1969.[39]

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TABLE 33[39]

DISTRIBUTION OF RATIO (R) FOR SULFURIC
ACID MIST PRECIPITATORS
(1945 - 1969)

| $R = W_p/W_D$ Interval | No. of Install. | No. of Pptrs. | Pptr. Capacity Thousand ACFM In Interval | % of Installations In Interval |
|---|---|---|---|---|
| 0.5-0.59 | 1 | 2 | 36 | 4.5 |
| 0.6-0.69 | 2 | 3 | 53.8 | 9.1 |
| 0.7-0.79 | 2 | 3 | 41 | 9.1 |
| 0.8-0.89 | 1 | 2 | 57.5 | 4.5 |
| 0.9-0.99 | 2 | 4 | 50.2 | 9.1 |
| 1.0-1.09 | 4 | 6 | 76.1 | 18.2 |
| 1.1-1.19 | 2 | 2 | 20.4 | 9.1 |
| 1.2-1.29 | 3 | 5 | 57.2 | 13.6 |
| 1.3-1.39 | 4 | 7 | 96.8 | 18.2 |
| 1.4-1.49 | 1 | 2 | 36.0 | 4.5 |
| 1.5-1.59 | 0 | 0 | 0 | 0 |
| TOTALS | 22 | 36 | 525.0 | 100 |

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the statistical analysis in which a total of 22 installations consisting of 36 acid mist precipitators were examined. The ratio for all the installations was distributed between 0.5 and 1.50.

## Phosphoric Acid Production

Phosphoric acid is manufactured by two processes: (1) the thermal process, involving the burning of elemental phosphorus to the pentoxide, followed by hydration, and (2) the wet process, involving the treatment of phosphate rock with sulfuric acid. Since the literature search thus far has revealed extensive use of mist electrostatic precipitators for acid mist collection for the thermal process only, the following discussion will not include information on the wet process.

In its simplest form, manufacture of phosphoric acid by the thermal process involves three steps:

1. Oxidizing liquid elemental phosphorus in a cylindrical combustion chamber to produce phosphorus pentoxide,

$$P_4 + 5O_2 \rightarrow P_4O_{10}$$
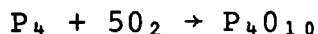
2. Hydrating the phosphorus pentoxide with dilute acid or water to produce phosphoric acid liquid and mist,

$$P_4O_{10} + 6H_2O \rightarrow 4H_3PO_4$$
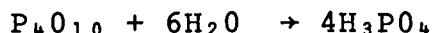
3. Removal of the phosphoric acid mist from the gas stream, usually accomplished by scrubber, mist eliminators, or electrostatic precipitators.

A typical flow diagram for thermal process manufacture of phosphoric acid is given in Figure 59.[47] The phosphorus is transferred from the liquid phosphorus feed tank to the burner tower when the phosphorus is mixed with air and oxidized. The resulting phosphorus pentoxide vapor and excess air then pass into a hydration tower from which the product acid discharges. The weak acid collected in the abatement equipment is recycled within the process.

The principal emission from the manufacture of phosphoric acid by the thermal process is acid mist in the absorber discharge gas. The particle size of the acid mist ranges from 0.2 to 2.6 microns with a mass median diameter of 1.6 microns.[48]

## Collection of Phosphoric Acid Mists

Packed and open tower scrubbers, venturi scrubbers, cyclonic separators with wire mesh eliminators, fiber mist eliminators,

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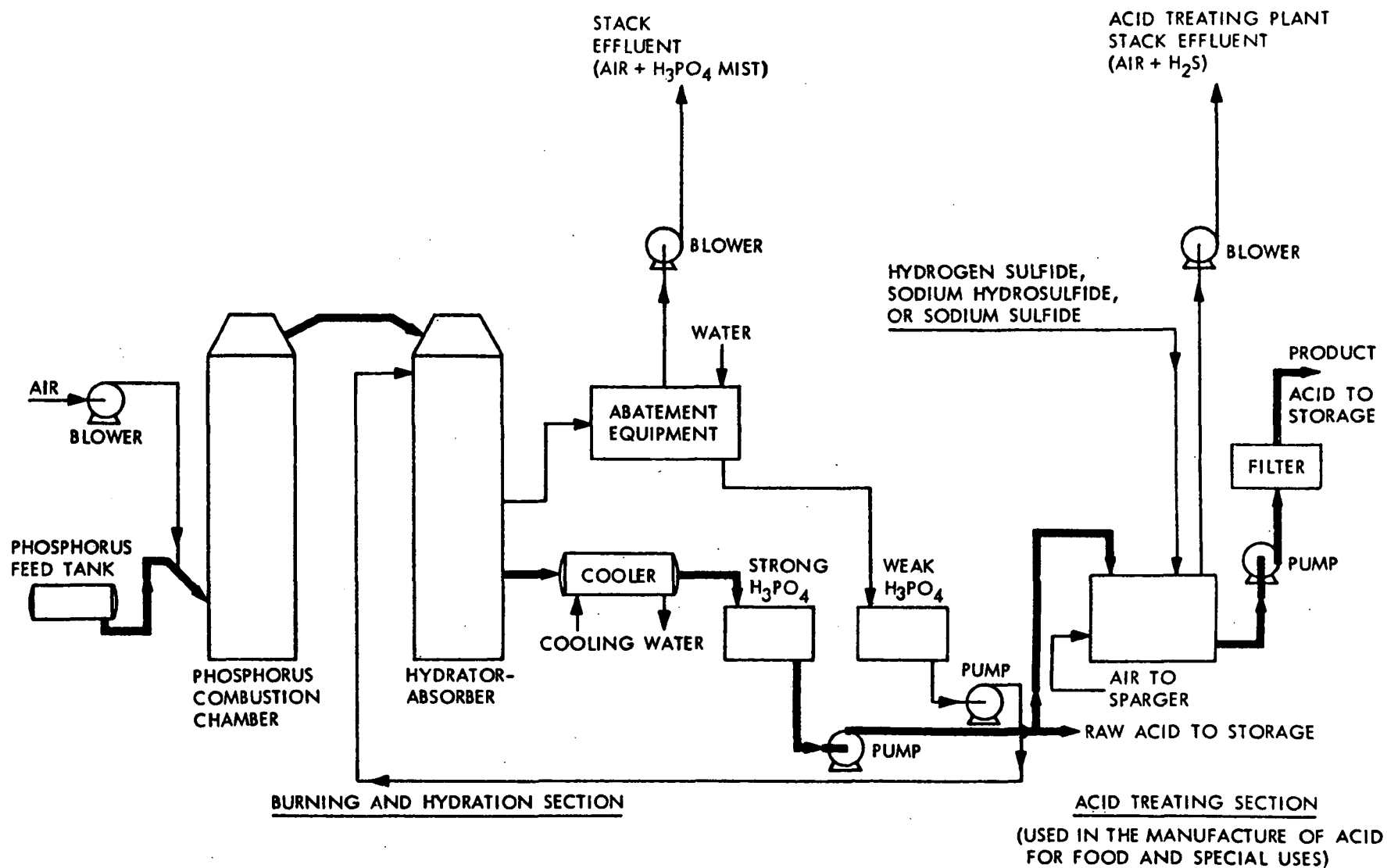
Figure 59. Flow Diagram for Typical Thermal-Process Phosphoric Acid Plant[47]

high energy wire mesh contactors, and mist electrostatic pre-
cipitators have all been used to collect phosphoric acid mist
from thermal-process phosphoric acid manufacture.

The typical design for a mist precipitator for phosphoric
acid collection consists of a single-stage, vertical flow,
pipe-type unit. Stainless steel is the conventional material
of construction instead of lead because of temperatures which
may fluctuate above the softening point of lead. The collecting
electrodes, normally stainless, are of variable diameter and
require no cleaning because of the free-flowing nature of the
precipitate. Inlet temperatures are usually below 200°F with
grain loadings, which vary considerably, typically being 10-60
gr/scf ($H_3PO_4$ basis). Typical voltage requirements are
about 75 kV supplied by various electrical energization systems.
TVA has used precipitators for many years to reduce phosphoric
acid emissions, and the construction details of a large precipi-
tator operated by the TVA are shown in Figure 60.[49] All
stainless steel used on this unit is American Iron and Steel
Institute type 316. Electric current is supplied from a
half-wave rectifier by electrical leads entering the precipi-
tator through lead-lined oil seals which contain insulators.
A stainless steel fan pulls 9.4 $m^3$/sec of gases at 106°C
through the three parallel sections of the precipitator,
which collects 99% of the fine mist. The precipitator is
followed by a water scrubber.

Since high concentrations of phosphoric acid mist are
often present in process streams and since the particle size
of the mist is small, large precipitators are usually required.
Two mist electrostatic precipitators are often used in series
to meet the high efficiency requirements rather than the single
section as depicted in the figure. In general, phosphoric
acid losses from a mist precipitator are affected only
slightly by the rate of gas flow or by temperature so long
as the gas flow is below design rate. However, cleanliness
of the equipment and electrical conditions employed affect
such losses materially.


ALUMINUM INDUSTRY

Primary aluminum production, as used in this report, will
include the smelting and refining of aluminum as opposed to
secondary aluminum production which is primarily engaged in the
recovery of aluminum from scrap. There are 27 primary aluminum
smelters in the United States with plant capacities ranging from
80,000 to 275,000 tons per year, the locations, capacities,
and cell types of which are shown in Table 34.[50] Aluminum, pro-
duced from alumina ($Al_2O_3$) by electrolytic reduction in fused
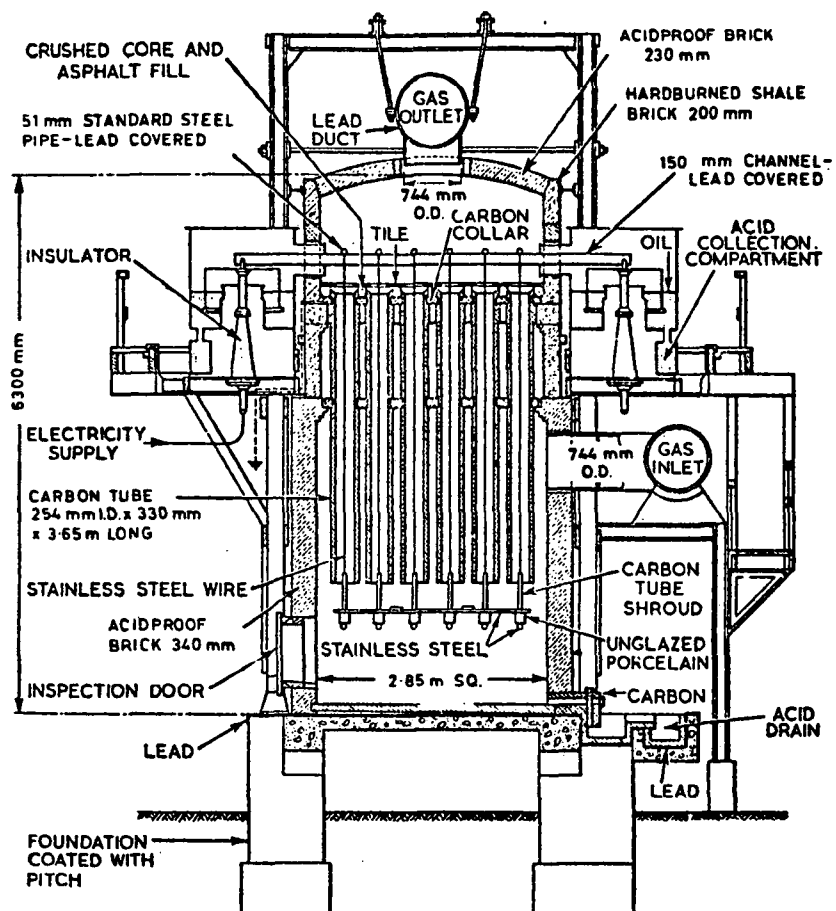
Figure 60. Electrostatic precipitator used for $P_2O_5$ mist removal.[49]

TABLE 34[50]

PLANT CAPACITY FOR MANUFACTURING PRIMARY ALUMINUM

| State and City | Annual Capacity, Tons | Company | Reduction Cell Types* |
|---|---|---|---|
| **Alabama** | | | |
| Sheffield | 221,000 | Reynolds | HS |
| Scottsboro | 210,000 | Revere | PB |
| **Arkansas** | | | |
| Arkadelphia | 63,000 | Reynolds | HS |
| Jones Mills | 122,000 | Reynolds | PB |
| **Indiana** | | | |
| Evansville | 175,000 | Alcoa | PB |
| **Kentucky** | | | |
| Hawesville | 45,000 | National-Southwire Aluminum | PB |
| **Louisiana** | | | |
| Chalmette | 260,000 | Kaiser | HS |
| Lake Charles | 35,000 | Gulf Coast Aluminum | PB |
| **Maryland** | | | |
| Frederick | 90,000 | Eastalco | PB |
| **Missouri** | | | |
| New Madrid | 110,000 | Noranda | PB |
| **Montana** | | | |
| Columbia Falls | 175,000 | Anaconda | VS |
| **New York** | | | |
| Messena | 128,000 | Reynolds | HS |
| | 125,000 | Alcoa | PB |
| **North Carolina** | | | |
| Badin | 100,000 | Alcoa | PB |
| **Ohio** | | | |
| Hannibal | 240,000 | Ormet | PB |
| **Oregon** | | | |
| The Dalles | 87,000 | Martin Marietta | VS |
| Trontdale | 100,000 | Reynolds | PB |
| **Tennessee** | | | |
| Alcoa | 200,000 | Alcoa | all |
| New Johnsonville | 140,000 | Conalco | PB |
| **Texas** | | | |
| Corpus Christi | 110,000 | Reynolds | HS |
| Point Comfort | 175,000 | Alcoa | VS |
| Rockdale | 275,000 | Alcoa | PB |
| **Washington** | | | |
| Bellingham | 265,000 | -------- | PB |
| Longview | 190,000 | Reynolds | HS |
| Tacoma | 81,000 | Kaiser | HS |
| Spokane | 206,000 | -------- | PB |
| Vancouver | 100,000 | Alcoa | PB |
| Wenatchee | 175,000 | Alcoa | PB |
| **West Virginia** | | | |
| Ravenswood | 163,000 | Kaiser | PB |

* PB = Prebaked Anode
  HS = Horizontal Stud Soderberg
  VS = Vertical Stud Soderberg

cryolite (AlF₃·3NaF), passes through three stages from ore to
metal: bauxite, alumina, and primary aluminum. In most cases
the Bayer process, shown in Figure 61, is used to refine alumina
from bauxite.[51] The alumina is next sent to a primary aluminum
processing plant where it is converted to the metal by electrolysis.

Process Description

The Hall-Heroult process is used by all domestic primary aluminum
producers for the electrolytic reduction of purified alumina
to aluminum metal. The process involves the dissolving of
alumina in a bath of molten cryolite. The bath is less dense
than molten aluminum and is maintained in a molten state in a
carbon crucible known as a "cell" or "pot". The carbon crucible
serves as the cathode, and a carbon block serves as the anode.
Electrolysis decomposes the alumina into aluminum and oxygen, and
because of its greater density, the aluminum sinks to the bottom
of the cell. Periodically the aluminum metal is withdrawn, and
fresh alumina feed and bath chemicals are added by piercing the
frozen crust of the bath. This "crust breaking" agitates the bath
and is a time of increased particulate and gaseous emissions.

Classes of Electrolytic Cells Employed in Hall-Heroult Process

The variations of cell construction are based on the method of
anode manufacture, and the effluent characteristics from each
cell differ in type and quantity. Reduction cells are of three
basic types: prebaked cells which use prebaked carbon anodes,
and two types of Soderberg cells which use large single anodes
continuously baked in place over the bath. In all cell types
the anode is composed of coke bound with pitch.

Prebaked or Niagara Cell (Illustrated in Figure 62)[52] -

Prebaked potlines employ multiple blocks of preformed carbon,
supported by copper and aluminum hangers to the electrical bus-
bars overhead. The bus is adjustable to compensate for anode
consumption.

Vertical Stud Soderberg (Illustrated in Figure 63)[53] -

The vertical stud Soderberg cell, because of the placement of
the anode studs in a vertical position, allows gases liberated
by the cell to be hooded most easily of the cell types. Carbon
paste is added to the top of the cell and slowly carbonized as
it moves toward the bath. The vertically positioned metal studs
carry electric current to the carbonized portion through the top.

Horizontal Stud Soderberg (Illustrated in Figure 64)[54] -
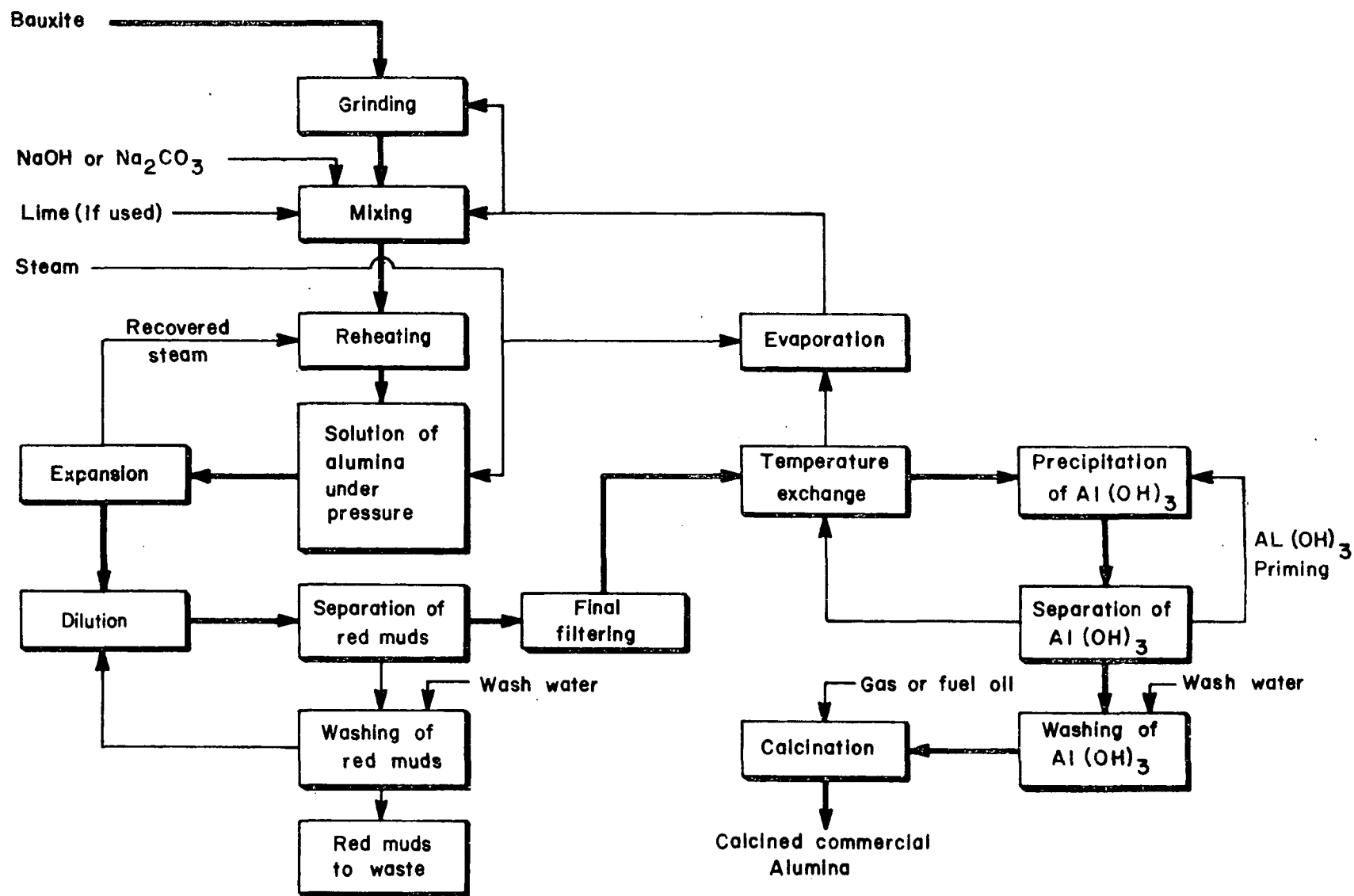
Figure 61. Diagram of the Bayer Process[51]
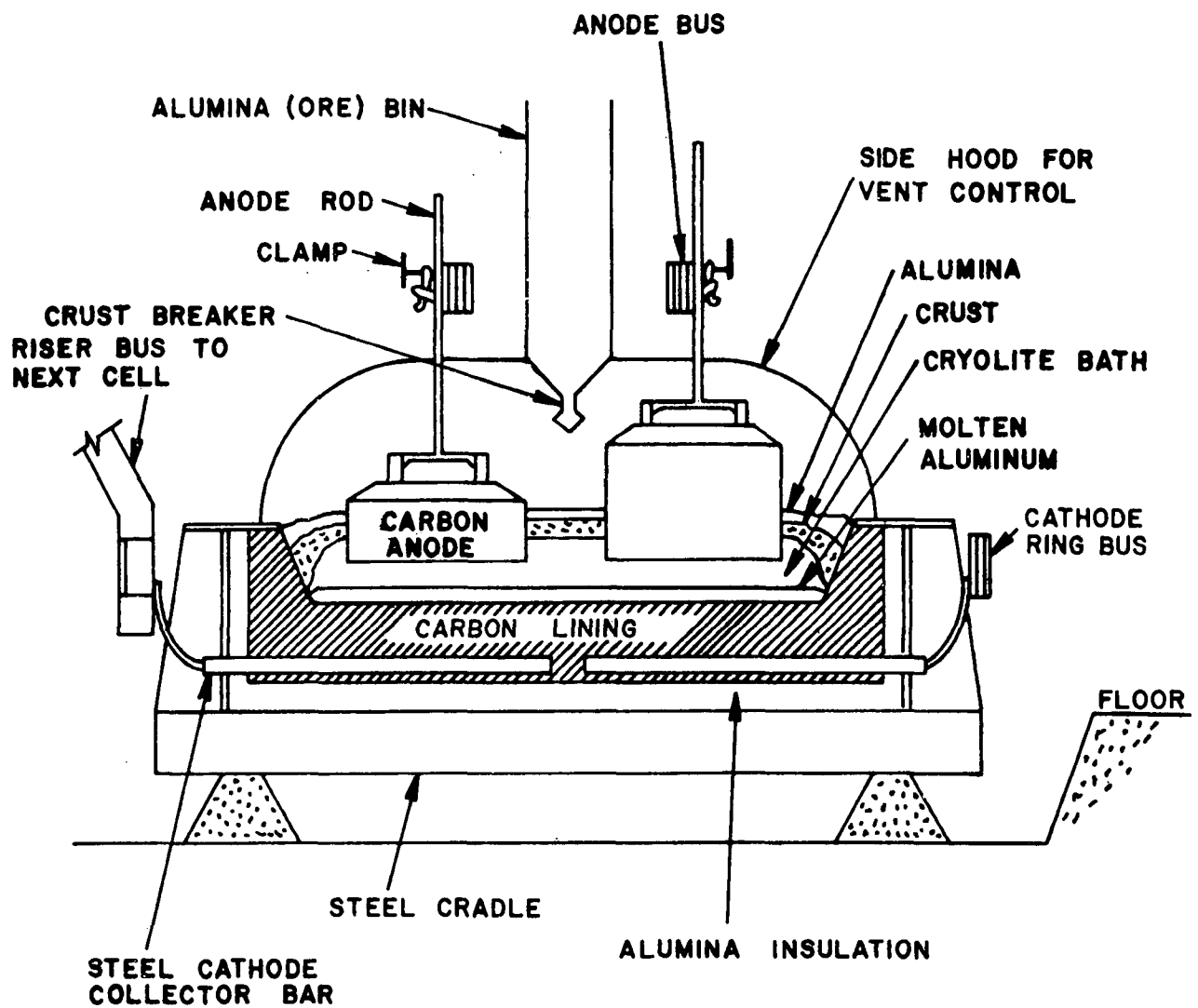
Figure 62.   Prebake Reduction Cell Schematic Arrangement[52]
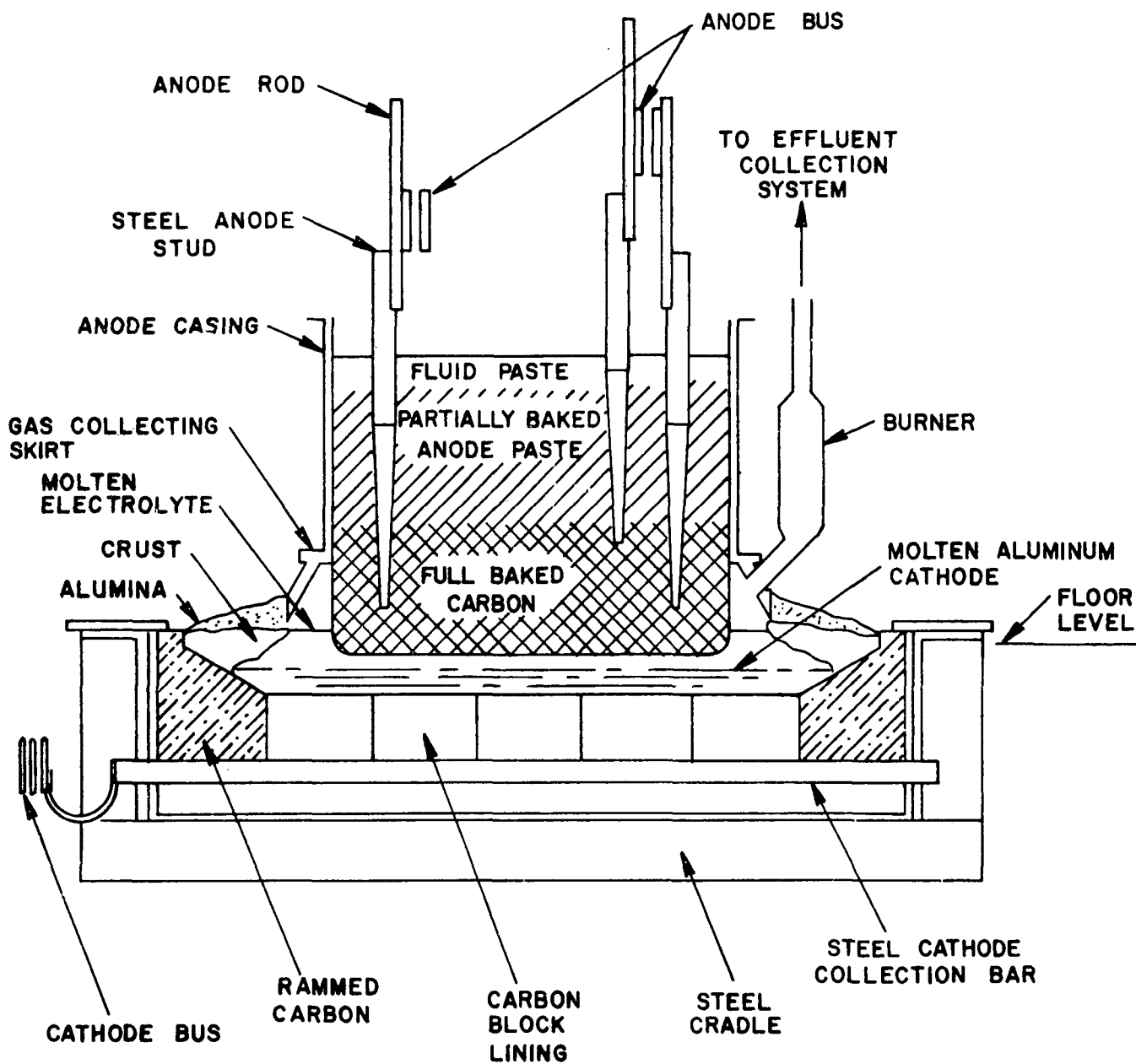
Figure 63. Vertical Stud Soderberg Cell Schematic Arrangement[53]
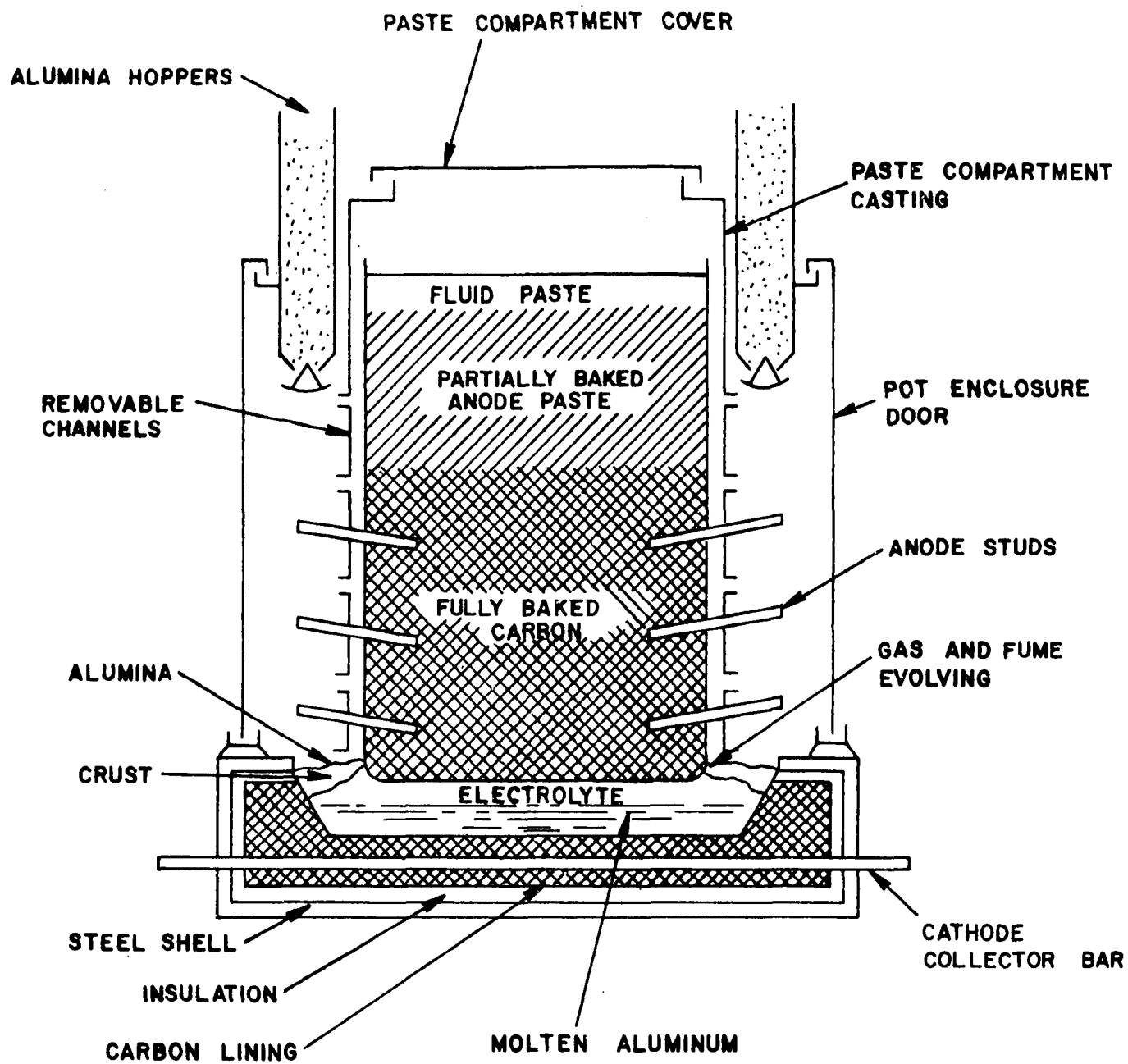
Figure 64. Horizontal Stud Soderberg Cell Schematic Arrangement[54]

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Hood requirements for the horizontal stud Soderberg are much more
difficult because of the location of the studs at the side of
the electrode rather than at the top.  Since access to the cell is
necessary for electrode pin removal and replacement, burning of
the gases and tars is not possible, and tar-fouling becomes a
problem.  Therefore, when electrostatic precipitator usage is
attempted, the plates require water flushing to prevent fouling
by tars.

Effluents originate at several processing operations in an
aluminum smelter.  Since the most difficult to control come from
the potlines, this report will emphasize those effluents originating
from the prebaked and Soderberg cells.  The airborne effluents
from the electrolytic process consist mainly of:

> CO and $CO_2$, formed when the released oxygen reacts
> with the carbon anode and escapes as CO or $CO_2$;
>
> $SO_x$, from sulfur present in the petroleum coke and
> pitch used to make the anodes;
>
> Particulates of vaporized bath materials;
>
> HF gas, resulting from the hydrolysis of fluoride
> salts;
>
> Alumina, cryolite, and aluminum fluoride dusts,
> entrained from the bath crust;
>
> Hydrocarbons, volatilized from the binders used to
> make the anode.

The quantity and type of fume emitted from reduction cells
depend greatly on the cell design and operating techniques used.
Reported determinations of particle size distributions of the dust
and fume emissions collected from prebaked and horizontal Soderberg
cells are shown in Figure 65.[55] Two plots are shown for prebake
potlines, one the average of four samples of pot emissions, the
other the average of five samples of electrostatic precipitator
intake.  Another size distribution of horizontal spike Soderberg
cell emissions is presented in Figure 66.[56] This typical size
distribution was determined with a cascade impactor at the exit
of the cell hood on potline 4 at the Kaiser Aluminum and Chemical
Corporation plant in Tacoma, Washington.  Microscopic analyses were
performed to verify the theoretical predictions of the University
of Washington cascade impactor.  The size distribution is the result
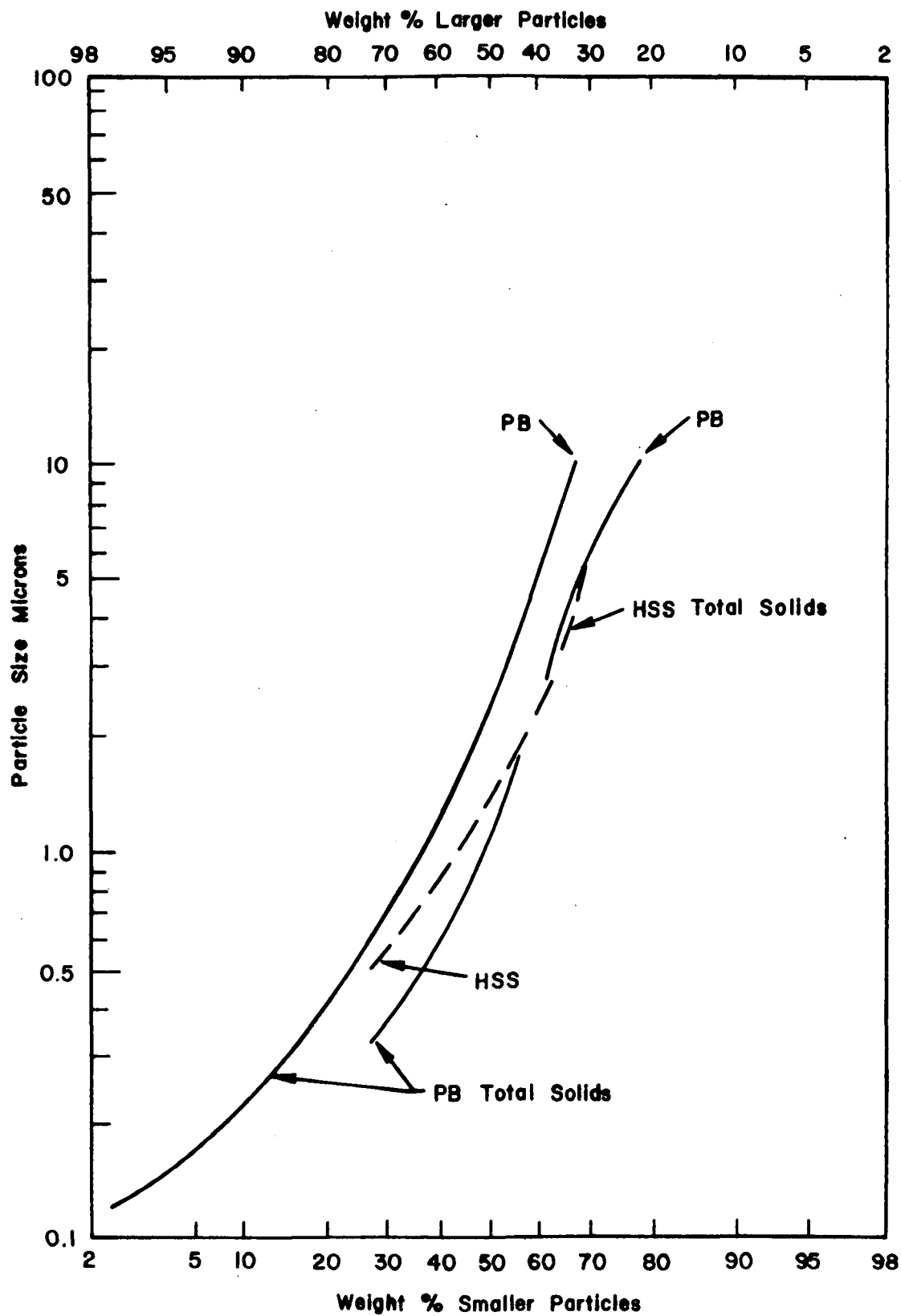
Figure 65. Particle Size Weight Distribution,
Potline Primary Effluent[55]
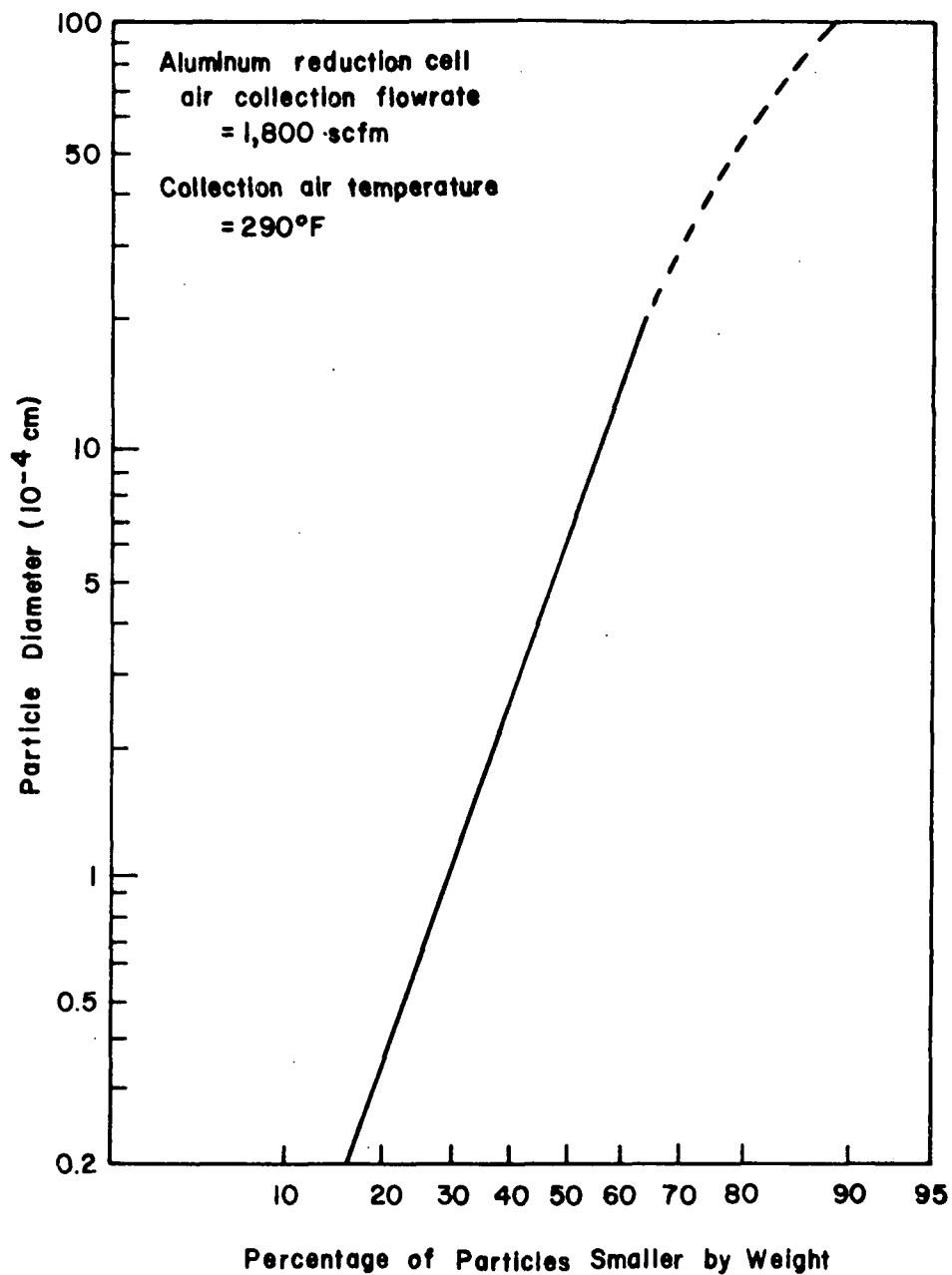
Figure 66. Composite Particle Size Distribution by Weight for Aluminum Reduction Cell Air Emissions (Kaiser Aluminum and Chemical Corporation Plant at Tacoma, Wash., Potline 4) at the Exit of the Reduction Cell.[56]

of combining all the size distribution data (140 tests) and taking into account the percentage of time each cell operation takes during a day. Particle concentration and size distribution were found to change significantly with changes in cell process operation.

Effective control of aluminum potline effluents involves three types of collection systems which can be classed as: 1) capture of all effluents by potline hooding, 2) subsequent collection of these captured pollutants by existing control devices, and 3) scrubbing of roof "monitor" emissions. This report will deal with those collection devices utilized for collection of potline effluents which have been hooded.

While the overall pollution potential of a primary aluminum facility is not significantly affected by the anode configuration, the choice of the associated air pollution control systems is closely related to the cell type. Therefore controls will be discussed separately for the prebake and Soderberg cells.

Control devices for vertical stud Soderberg cells have included multicyclones and spray-type scrubbers. Exhausts may be treated by bag filters coated with lime or alumina, or by electrostatic precipitators, but the residual tar creates a fouling problem in the collection system. A collection system for Montecatini Edison, a plant in Milan which incorporates dry cyclones, a dry electrostatic precipitator, and two scrubbers, is shown in Figure 67.[57]

As stated earlier, hooding for the horizontal stud Soderberg cell is inherently less complete and the large volumes of air entrained during replacement and readjustment of studs creates a dilute mixture of hydrocarbons which encourages a tar-fouling problem in ducts and control equipment. Existing controls have consisted of scrubbers, but where electrostatic precipitators are attempted, the plates require water flushing to prevent fouling by tars.

Particulate emissions from prebake pots contain none of the tar found in the horizontal stud Soderberg; instead "dusting" of the carbon anode produces carbon particles in addition to alumina, etc. Current controls have consisted of dry-type cyclones or electrostatic precipitators followed by wet scrubbers.
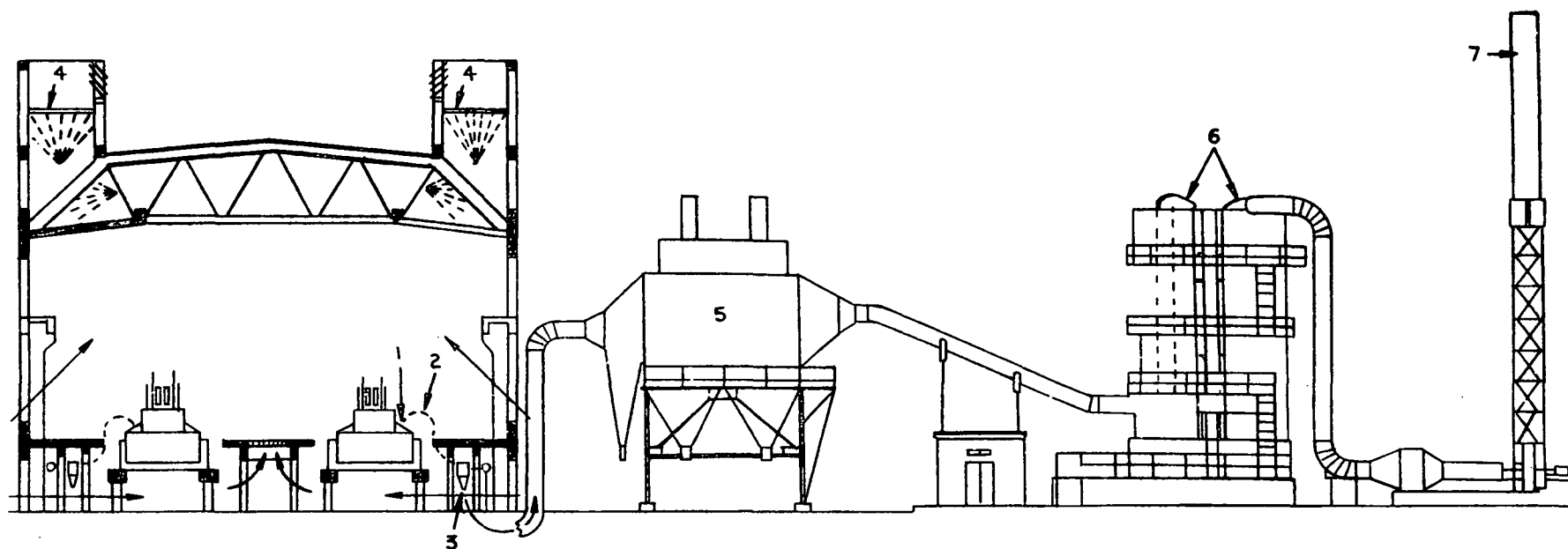
1. Gas suction skirt
2. Burners
3. Cyclones
4. Filter screen

5. Electrofilter
6. Scrubber
7. Chimney

Figure 67.   Anodic Gas Purification Plant Flowsheet[57]

Table 36 summarizes the removal equipment considered for emission control of the different types of effluent streams in the aluminum industry.[58]

Table 37 presents source sampling test results of a number of plants which were selected on the basis of representation of best control technology. The program included testing of potline installations at two vertical stud plants, three pre-bake plants, and one horizontal stud plant. A legend is given in Table 38 for the various abbreviations used.[59]

Cost data for control equipment in the primary aluminum industry are sparse, but Singmaster and Breyer have presented data derived from purchase costs reported by equipment manufacturers. Of specific interest is cost data for wet electrostatic precipitators used for control of vertical and horizontal stud cell types (shown in Figure 68).[60] It can be seen that the recent data of Bakke indicates much higher costs than the data obtained by Singmaster and Breyer.[61]

## Use of Wet Precipitators for Collection of Fume from Electrolytic Reduction Cells

Martin Marietta Aluminum -

Several wet-wall precipitator units of the Mikropul design are used to collect fume from vertical spike Soderberg cells at the Dalles, Oregon. Two sizes are employed, which are capable of treating 2.83 $m^3$/sec (6,000 acfm) and 5.66 $m^3$/sec (12000 acfm). The precipitators have two plate sections in series with sprays at the entrance of each section and above the plates.

A burner is installed in the stack, but it does not remove all of the condensable hydrocarbons. A scrubber precedes the precipitator, which reportedly removes about 95% of the gaseous contaminants. Performance data from this installation, as reported by a representative of Martin Marietta to SRI personnel, is given in Table 35.

TABLE 35[64]

TYPICAL PERFORMANCE DATA OF WET
PRECIPITATOR AT THE DALLES

|  | gr/scf | mg/$m^3$ | Efficiency Range |
|---|---|---|---|
| Inlet | 0.13-0.20 | 298-458 | |
| Outlet | 0.0024-0.003 | 5.49-6.87 | 97.7-98.8 |

# TABLE 36[58]

## CONTROL EQUIPMENT CONSIDERED FOR THE PRIMARY ALUMINUM INDUSTRY

| | PREBAKE POTLINES | | | VS POTLINES | | HS POTLINES | | ANCILLARY PROCESSES | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|
| | Prim. no Sec. | Sec. with Prim. | Sec. no Prim. | Prim. no Sec. | Sec. with Prim. | Prim. no Sec. | Sec. with Prim. | Bake Plant | Dry Mtls. | Paste Mix | Cast House |
| Burner | - | - | - | A | - | - | - | - | - | - | - |
| Incinerator | - | - | - | - | - | - | - | - | - | - | - |
| Multiple Cyclone | A | - | - | A | - | - | - | - | A | - | - |
| Baghouse Filter | - | - | - | - | - | - | - | - | A | - | - |
| Fluid Bed Dry Scrubber | A | - | - | A* | - | - | - | - | - | - | - |
| Coated Filter Dry Scrubber | A | - | - | - | - | - | - | - | - | - | - |
| Injected Alumina Dry Scrubber | A* | - | - | A* | - | - | - | - | - | - | - |
| Dry Electrostatic Precipitator | A | - | - | A* | - | - | - | A | A | - | - |
| Wet Electrostatic Precipitator | - | - | - | A | - | C | - | C | - | B | B |
| Spray Tower | A | - | - | A | - | A | - | A | - | A | - |
| Spray Screen | - | A | A | - | A | B | B | B | - | - | - |
| High Pressure Spray Screen | B | - | - | B | - | D | - | B | - | B | - |
| Wet Centrifugal Scrubber | - | - | - | D | - | - | - | - | - | - | - |
| Venturi | - | - | - | A | - | D | - | C | - | A | B |
| Chamber Scrubber | A | - | - | B | - | B | - | - | - | - | - |
| Wet Impingement Scrubber | - | - | - | - | - | - | - | - | - | B | - |
| Cross Flow Packed Bed | B | - | A | B | - | C | - | - | - | - | - |
| Floating Bed (Bouncing Ball) | A | - | B | A* | A*1/ | A | - | B | - | B | A |
| Sieve Plate Tower | - | - | - | A* | - | - | - | - | - | B | - |
| Self-Induced Spray (Bubbler) | B | - | - | A | - | B | - | A | - | B | - |
| Vertical Flow Packed Bed | A | - | - | A | - | - | - | - | - | B | - |

Prim.   Primary collection system.
Sec.    Secondary or potroom system.
A       In current use in the United States.
A*      In current use outside the United States.
A*1/    Used in one foreign plant. Not considered economically feasible in the United States.

B   Considered feasible but not known to be in use.
C   In development stage.
D   Superseded by other equipment.
-   Considered not feasible, economically and/or technically.

## TABLE 37[59]

RESULTS OF EPA SOURCE TESTS FLUORIDE AND PARTICULATE
PRIMARY ALUMINUM INDUSTRY
lbs/Ton Aluminum *

| Plant | A | $A_1$ | $A_2$ | B | $B_1$ | C | D | $D_1$ | E | F |
|---|---|---|---|---|---|---|---|---|---|---|
| Cell Type | VS | VS | VS | PB | PB | PB | PB | PB | HS | VS |
| Primary Control Device | BS-ST | BS-WESP | BS-WESP | FBDS | FBDS | ST | FBDS | FBDS | ST-WESP | ST-WESP |
| Secondary Control Device | SS | SS | SS | None | None | SS | None | None | None | None |

**PARTICULATES**

| | A | $A_1$ | $A_2$ | B | $B_1$ | C | D | $D_1$ | E | F |
|---|---|---|---|---|---|---|---|---|---|---|
| Primary Collection | NS | 91.26(67.60)[2] | NS | 110.10 | 100.30 | NS | 72.26 | NS | 81.80 | 38.20 |
| Secondary Collection | – | 27.12(26.68) | – | NS | NS | – | NS | – | NS | NS |
| Primary Emission | – | 0.12(0.11) | – | 13.79 | 1.80 | – | 4.00 | – | 5.95 | 1.34 |
| Secondary Emission | – | 9.53(5.83) | – | NS | NS | – | NS | – | NS | NS |
| Total Emission | – | 9.65(5.94) | – | – | – | – | – | – | NS | NS |
| Primary Efficiency(%) | – | 99.86(99.84) | – | 87.46 | 98.19 | – | 94.43 | – | 92.73 | 96.58 |
| Secondary Efficiency(%) | – | 64.85(78.15) | – | NC | NC | – | NC | – | NC | NC |
| Overall Control Efficiency(%) | – | 91.85(93.70) | – | – | – | – | – | – | – | – |

**TOTAL FLUORIDE** (Particulates & Gaseous)

| | A | $A_1$ | $A_2$ | B | $B_1$ | C | D | $D_1$ | E | F |
|---|---|---|---|---|---|---|---|---|---|---|
| Primary Collection | 53.50 | 37.69 | NS | NS | 37.80(48.05) | 28.10 | 61.40 | NS | 46.53 | NS |
| Secondary Collection | 3.65 | 3.02 | NS | – | 1.20 | 9.06 | 1.17 | NS | 2.06 | – |
| Primary Emission | 1.65 | 0.01 | 0.02 | – | 0.14 | 69.64[3] | 0.87(0.49)[4] | 0.35 | 0.41 | – |
| Secondary Emission | 1.65[1] | 0.81 | 2.05 | – | 1.20 | 7.30 | 1.17 | 1.17 | 2.06 | – |
| Total Emission | 3.30 | 0.82 | 2.07 | – | 1.34 | 76.94 | 2.04(1.66) | 1.52 | 2.47 | – |
| Primary Efficiency(%) | 96.91 | 99.97 | 99.94 | – | 99.62(99.54) | – | 98.64(99.23) | 99.11 | 99.11 | – |
| Secondary Efficiency(%) | 54.79 | 73.17 | 32.12 | – | NC | 19.42 | NC | NC | NC | – |
| Overall Control Efficiency(%) | 94.22 | 97.98 | 94.92 | – | 96.89(97.46) | – | 96.89(97.46) | 94.95 | 94.95 | – |

| Plant | G | H |
|---|---|---|
| Control | ESP | ESP |

**ANODE BAKE PLANTS[5]**

| | G | H |
|---|---|---|
| Particulate Emissions | 1.56 | 3.96 |
| Total Fluoride Emissions | 0.88 | 1.25 |

*Refer to legend on following page.

TABLE 38[59]

LEGEND FOR TABLES AND FIGURES

VS - Vertical stud Söderberg cell

PB - Prebake cell

HS - Horizontal stud Soderberg cell

BS - "Bubbler" scrubber

ST - Spray tower

WESP - Wet electrostatic precipitator

SS - Spray screen

FBDS - Fluid bed dry scrubber

NS - No sampling

NC - No Control

(1) - Average of two tests; one test suspected to be contaminated and deleted.

(2) - Average of two tests; one test deleted due to stud blow during test.

(3) - Samples suspected to be contaminated during sampling; these plant data are suspect.

(4) - Data with two tests deleted due to suspected control system malfunction.

(5) - Anode plant emission units are lb/ton anode produced.
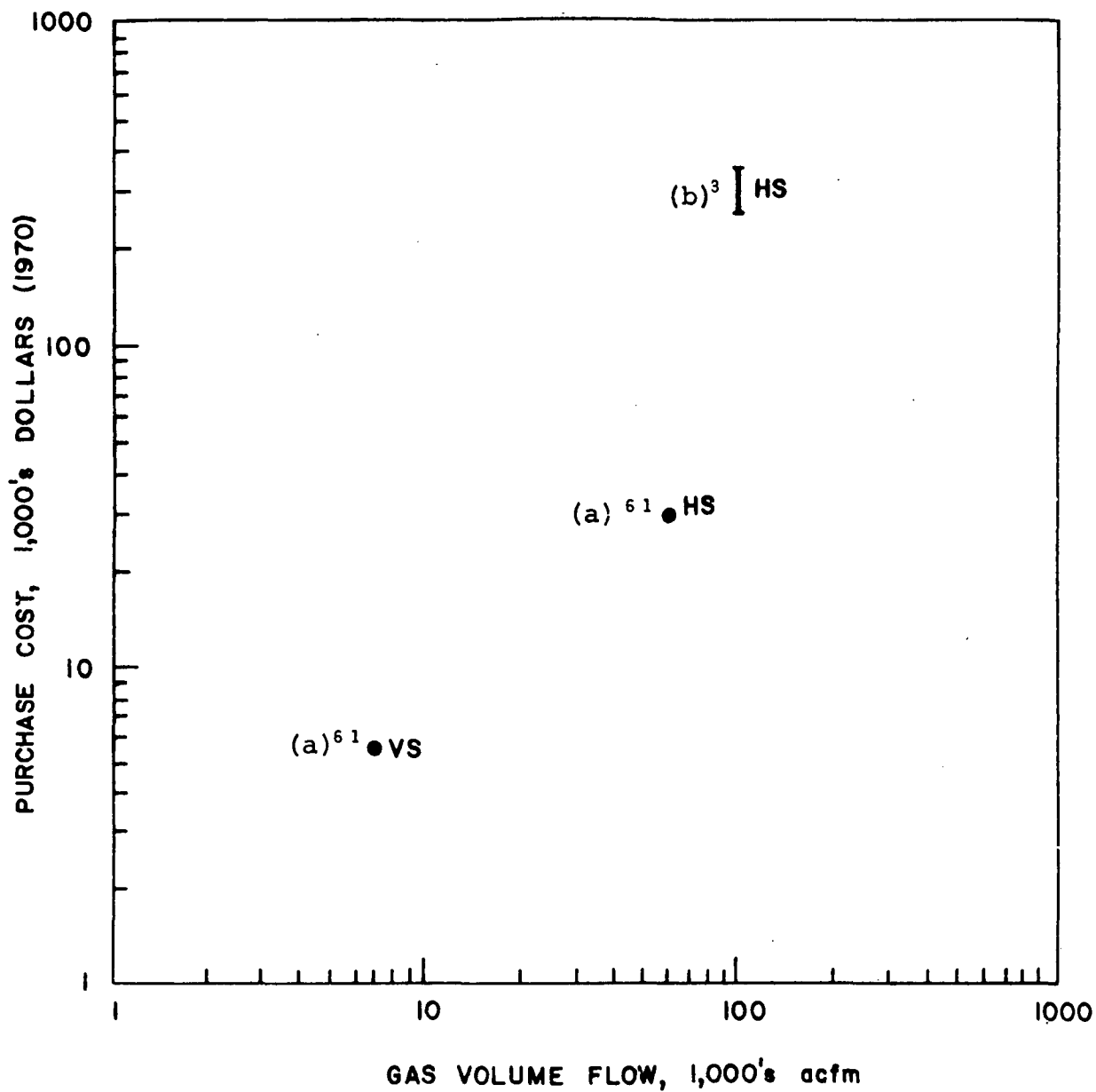
Figure 68. Cost Data for Wet Precipitators

The specific collecting area of these units, not including the exit baffles is about 53.2 $m^2/(m^3/sec)$, or 270 $ft^2/1000$ cfm. Typical electrical operating conditions are given in Table 39 The liquid to gas ratio used in the precipitator sprays was about 1.0 1 of liquid per $m^3$ of gas (7.5 gal/1000 cf).

TABLE 39[64]

TYPICAL ELECTRICAL CONDITIONS FOR WET PRECIPITATOR
AT THE DALLES

|  | Voltage, kV | Current Density, na/cm$^2$ |
|---|---|---|
| Inlet Section | 56 | 45.7 |
| Outlet Section | 49 | 56.5 |

The precipitator shell bottom has been replaced with fiberglass because of corrosion due to the acidic environment. Discharge electrodes and the precipitator plates are stainless steel and have held up fairly well, although there have been some problems with plate bending. Insulators have also caused some problems, but the overall down time of the precipitators was said to be less than 1%. A routine cleaning of the precipitators is performed every six weeks, and no problems with deposits have been encountered.

Plant personnel stated that local pollution control authorities were pleased with the gas effluent from the wet precipitators, which comprises only about 3% of the total particulate emissions from the plant. However, water pollution from the wet cleaning system is a problem, and some consideration is being given to a dry removal system. The pH of the discharge water from the system is about 2.5.

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A similar wet precipitator installation is employed by
Martin Marietta at Goldendale, Washington.  Table 40  gives outlet
data obtained by the State of Washington in October of 1972 at
this location on the No. 1 precipitator, and Table 41   compares
the wet precipitator emissions with other sources of emissions at
the plant.[62]

TABLE 40[62]

EMISSION MEASUREMENTS ON ELECTROSTATIC PRECIPITATOR
NO. 1, MARTIN MARIETTA PLANT AT GOLDENDALE, WASHINGTON

| | Mass Loading | | | | | | Sampling |
| | Benzene Sol. | | Benzene Insol. | | Total | | Time |
| $ft^3$ Sampled | $mg/m^3$ | $gr/ft^3$ | $mg/m^3$ | $gr/ft^3$ | $mg/m^3$ | $gr/ft^3$ | Minutes |
|---|---|---|---|---|---|---|---|
| 1437 | 1.00 | 0.00044 | 3.02 | 0.00132 | 4.03 | 0.00176 | 222 |

## TABLE 41[62]

### SUMMARY OF MEASURED EMISSIONS

| SOURCE | SAMPLING LOCATION | POLLUTANT MEASURED | CONCENTRATION | EMISSION RATE | ALLOWABLE EMISSION | EMISSION* FACTOR | METHOD |
|---|---|---|---|---|---|---|---|
| Pot room | Roof monitor | Particulate | .00063 gr/ft$^3$ | 69.1 lb/hr | | 5.38 lb/ton | Intermediate Volume Sampler |
| Hoods around anode bases | Electrostatic precipitator | Particulate | .00176 gr/ft$^3$ | 2.54 lb/hr | | .20 lb/ton | |
| Pot room | Roof monitor | Extracted Particulate | .00026 gr/ft$^3$ | 28.5 lb/hr | 15 lb/ton total | 2.22 lb/ton | Intermediate Volume Sampler (benzene extract) |
| Hoods around anode bases | Electrostatic Precipitator | Extracted Particulate | .00132 gr/ft$^3$ | 1.91 lb/hr | | .15 lb/ton | |
| Pot room | Roof monitor | Fluorides (as HF) | 203 µg/m$^3$ | 10.15 lb/hr | To meet WAC 18-48 ambient & forage standard | .79 lb/ton | NaOH absorption & specific ion electrode |
| Hoods around anode bases | Electrostatic Precipitator | Fluorides (as HF) | 98 µg/m$^3$ | .62 lb/hr | | .05 lb/ton | |
| Hoods around anode bases | Electrostatic Precipitator | SO$_2$ | 26 ppm | 47.6 lb/hr | | 3.65 lb/ton | NaOH absorption & turbidometric |
| Pot room | Roof monitor Line B, Sec. 1 | CO | 24 ppm | 1496 lb/hr | | 117 lb/ton | Grab sample & non-dispersive infra-red |
| Hoods around anode bases | Electrostatic Precipitator No. 1 | CO | 12750 ppm | 9523 lb/hr | | 742 lb/ton | |
| Pot room | Roof monitor Line B, Sec. 1 | Hydrocarbons (as CH$_4$) | 6 ppm | 214 lb/hr | | 16.7 lb/ton | Grab sample & flame ionization |
| Hoods around anode bases | Electrostatic Precipitator No. 1 | Hydrocarbons | 75 ppm | 32 lb/hr | | 2.5 lb/ton | |
| Pot room | Roof monitor | 3 - 4 Benzopyrene | 158 Nanograms/M$^3$ | .0083 lb/hr | | .00065 lb/ton | I.V. Sampler + Benzene Extraction + Adsorbtion Chromatography + Colorimetric |

* Based on average daily production of 308 tons aluminum

# SECTION VIII

## SUMMARY OF STATUS OF WESP TECHNOLOGY AND DESIGN METHODS

Wet electrostatic precipitators have been effectively used to collect particulate matter in the aluminum, iron and steel, chemical, and petrochemical industries. Duct-type collecting electrodes are generally used for the applications in which relatively large gas volumes are to be cleaned. The advantages of plate electrodes as opposed to pipe electrodes are less expensive construction and the flexibility of allowing variations in electrical sectionalization. Pipe electrodes can be more uniformly irrigated than plates, and gas by-passage (sneakage) is not a factor. Saturation of the gas stream with water prior to entering a WESP is desirable with all electrode geometries to prevent deposit formation and to maximize effective utilization of the precipitation process for removing particles.

The largest recent application of WESPs has been in the collection of fume from aluminum pot lines. A field test of a full scale duct-type wet ESP installed on an aluminum pot line indicated that collection efficiencies of 95 to 98% were achieved on a particle size distribution in which about 65% of the mass consisted of sub-micron particles. The specific collection area during the test period was approximately 62 $m^2/(m^3/sec)$. Corrosion and deposit formation in the liquor supply system have occurred at this installation.

In the iron and steel industry, WESPs are used on electric arc furnaces, blast furnaces, basic oxygen furnaces, scarfers, and cupolas. In some of the electric arc furnace WESP installations, intermittent washing of the plate-type electrodes was found to be preferable to continuous irrigation. These units have either two or three fields in series, and the field undergoing washing is de-energized during the washing period. This is in contrast to the aluminum pot line installations, which employ continuous irrigation. Wet precipitators of the wire and pipe design have been widely used for cleaning blast furnace gas, but this is a declining application area because of the recent trend toward the use of blast furnaces with high top pressures. The energy available with the high pressures can be utilized in high energy particulate scrubbers with a lower capital investment compared to precipitators.

Electrostatic precipitators are used as collectors of tars and oils in the steel and petrochemical industry, and for collection of acid mists from sulfuric and phosphoric acid production in the chemical industry. These types of units do not usually employ forced irrigation of the collecting electrodes, but they are "wet" in the sense that collected material is a fluid, and no electrode rapping is employed. Detarring ESPs are typically of the wire and pipe design, with the pipes suspended from a header plate in a round shell. Acid mist precipitators are also typically of the wire in tube type, but the wire and plate design is common in Europe.

Design methods for sizing wet electrostatic precipitators have included pilot plant studies and design by analogy based on the empirical application of the Deutsch equation with a "precipitation rate parameter" used as an indicator of performance. For the blast furnace and sulfuric acid mist applications, the data in the literature concerning design versus installed performance is reported as the term (R) which is the ratio of the precipitation rate parameter obtained from test data to that used in designing the unit. For example, a value of R of 1.20 means that the precipitator plate area is 20% larger than required to meet the design efficiency for that particular unit. The scatter in the R values (typical range, .5 to 1.5) indicates the need for more precise sizing methods.

The agreement obtained between the laboratory data and the predictions of the mathematical model (Section III) suggests that the mathematical model offers a tool which can be used to estimate plate area requirements for various wet precipitator application areas. The computer model in its present state of development is useful for qualitatively predicting performance trends caused by changes in specific collecting area, electrical conditions, and particle size distributions. Current density, applied voltage, and the particle size distribution are the most important variables in the calculation of overall mass collection efficiency for a given specific collection area. A detailed discussion of the mathematical model is given in Reference 1.

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APPENDIX A

CONVERTING UNITS OF MEASURE

| To Convert From | To | Multiply By |
|---|---|---|
| lbs | kg | 0.454 |
| grains/cf | grams/m$^3$ | 2.29 |
| cfm | m$^3$/sec | 0.000472 |
| lbs/in.$^2$ | kg/m$^2$ | 703 |
| °F | °C | ( F - 32) x 5/9 |
| ft$^2$/1000 cfm | m$^2$/(m$^3$/sec) | 0.197 |
| inches w.g. | mm Hg | 1.868 |
| gallon | liter | 3.785 |
| ft | m | 0.3048 |
| inches | m | 0.0254 |
| tons | kg | 908 |
| cubic inches | cubic centimeters | 16.39 |
| cubic feet | cubic meters | 0.028 |
| cubic feet | liters | 28.32 |
| gallon/minute | liters/second | 0.0631 |
| square feet | square meters | 0.0929 |
| square inches | square centimeters | 6.452 |
| gallon/1000 ft$^3$ | liters/m$^3$ | 0.135 |
| grams | grains | 15.43 |

SECTION XII

APPENDIX B

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**16. ABSTRACT** The report describes a study of wet electrostatic precipitators (WESP's) through laboratory experiments, a field test of a full-scale unit, interviews with manufacturers and users, and a literature survey. Experiments were aimed at determining collection efficiency as a function of particle size at varying current densities and specific collection areas. The results agreed reasonably with those predicted by simulating the experimental conditions with a mathematical model. The feasibility of collecting $SO_2$ and particulate in a WESP was examined. As expected, calculation of ion transport rates indicates that only an insignificant amount of $SO_2$ can be removed by selective ionization of $SO_2$ molecules. Mass transfer calculations indicated that irrigated electrode surfaces would not provide sufficient interfacial area for desired $SO_2$ removal levels. Based on conventional geometry and stainless steel construction, a WESP/scrubber device would be more costly than a scrubber-only system because of high WESP capital costs. The effectiveness of a WESP as a mist eliminator following a scrubber was calculated to be a function of the particle size distribution and the concentration of the mist to be collected. The field test and literature survey showed that WESP's have been used effectively to control fine particle emissions in the aluminum, iron and steel, chemical and petrochemical fields.

| 17. | KEY WORDS AND DOCUMENT ANALYSIS | | |
|---|---|---|---|
| **a. DESCRIPTORS** | | **b. IDENTIFIERS/OPEN ENDED TERMS** | **c. COSATI Field/Group** |
| Air Pollution Electrostatic Precipitators Dust Sulfur Dioxide Aluminum Industry | Iron and Steel Industry Chemical Industry Petrochemistry | Air Pollution Control Stationary Sources Wet Electrostatic Precipitators Collection Efficiency Particulates | 13B 07A 11G 08G 07B 11F |
| **18. DISTRIBUTION STATEMENT** Unlimited | | **19. SECURITY CLASS** *(This Report)* Unclassified | **21. NO. OF PAGES** 204 |
| | | **20. SECURITY CLASS** *(This page)* Unclassified | **22. PRICE** |

EPA Form 2220-1 (9-73)