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ENGINEERING REPORT COVER SHEET

TITLE: **Big Sandy Unit 2 Flue Gas Desulfurization (FGD)**
Technology Evaluation and Selection

REPORT NUMBER: **BS2-FGDPE-111311**

PROJECT:	Big Sandy Unit 2 – FGD Project
REVISION:	0
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Steam

its generation and use

The Babcock & Wilcox Company
a McDermott company

Edited by J.B. Kitto and S.C. Stultz

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Chapter 33

Particulate Control

As steam is widely generated by the combustion of most fossil fuels, the flue gas carries particulate matter, principally ash, from the furnace. Except for natural gas, all other fossil fuels contain some quantity of non-combustibles which form the majority of the particulate. Unburned carbon also appears as particulate. Particulate control is needed to collect this material and to limit its release to the atmosphere.

All coals contain some amount of ash. The amount of ash varies depending on the type of coal, location, depth of mine and mining method. In the United States (U.S.), eastern bituminous coals typically contain 5 to 15% ash while the western subbituminous coal ash content may range from 5 to 30% ash by weight. Texas lignites also contain up to 30% ash. Mining methods on thin seams of coal may also contribute to higher ash quantities (see Chapter 9).

When coal is burned in conventional boilers, a portion of the ash drops out of the bottom of the furnace (*bottom ash*) while the remainder of the ash is carried out of the furnace in the flue gas. It is this remaining ash (*flyash*) that must be collected after the furnace and before exhausting the flue gas to the atmosphere.

Different combustion methods contribute different proportions of the total coal ash content to the flue gas. With pulverized coal firing, 70 to 90% of the ash is carried out of the boiler with the flue gas. A stoker-fired unit will emit about 40% of its ash in the flue gas along with some amount of unburned carbon. With Cyclone firing, only 15 to 40% of the ash is normally carried by the flue gas. On circulating fluidized-bed boilers, all of the ash, along with the fluidized-bed material, is carried by the flue gas. Therefore, the selection and design of particulate control equipment are closely tied to the type of firing system.

An American Society for Testing and Materials (ASTM) ash composition analysis test (ASTM D-3682-01) of a coal ash sample reveals the major ash components. Proximate and ultimate coal analyses commonly offer additional insight and also provide total ash content. Ash components are typically reported in the oxide form and include silicon dioxide, titanium dioxide, iron oxide, aluminum dioxide, calcium oxide, magnesium oxide, sodium oxide, potassium oxide, sulfur trioxide and diphosphorous pentoxide. Trace quantities of many more elements are also found in ash. The proportion of the major constituents varies significantly between coal type and mine location. The analysis and composition of flyash are discussed in greater detail in Chapters 9 and 21.

Other significant coal ash properties are particle size distribution and shape, both of which are dependent on the type of firing method. Stoker-fired units generally produce the largest particles. Pulverized coal-fired boilers produce smaller, spherical shaped particles of 7 to 12 microns (Fig. 1). Particles from Cyclone-fired units, also mostly spherical, are among the smallest. Fluidized-bed units produce a wide range of particles that are generally less spherical and are shaped more like crystals. Knowledge of ash properties is important in the selection of the correct particulate control equipment.

Regulation of particulate emissions

Particulate control equipment was first used by utilities in the 1920s and before that time in some industrial applications.[1] Prior to 1971, however, controls were installed largely on a best effort basis. In 1971, the first Environmental Protection Agency (EPA) performance standard limited outlet particulate emissions to 0.1 lb/10^6 Btu (123 mg/Nm^3 at 6% O_2) heat input
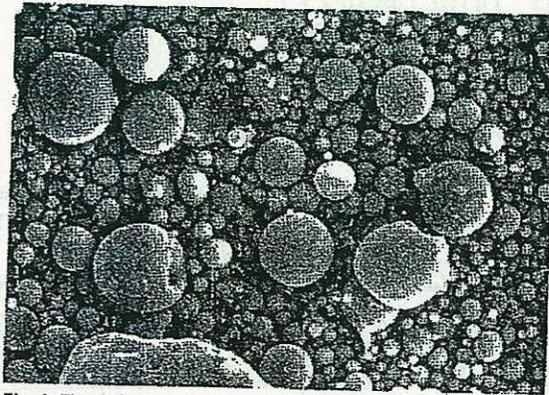

Fig 1 Flyash from pulverized coal (magnified X 1000).

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and stack opacity to 20% for those units larger than 250×10^6 Btu/h (73.3 MW$_t$) heat input. *Opacity*, measured by an instrument called a transmissometer, is the portion of light which is scattered or absorbed by particulate as the source of light passes across a flue gas stream. Therefore, both the amount and appearance of the stack emissions are regulated. Since 1979, the EPA New Source Performance Standards (NSPS) for particulate control permit a maximum of 0.03 lb/10^6 Btu (36.9 mg/Nm^3 at 6% O$_2$) heat input for these units. A 20% opacity is still permissible.

Federal and state EPA regulations set the primary guidelines for particulate emissions. In addition, many local regulatory bodies have generally stricter regulations than those set by the federal EPA. There are separate emissions standards for a variety of combustion processes including steam generators firing coal, oil, refuse and biomass. Currently there are three major classification levels for steam generating units: one for units greater than 250×10^6 Btu/h (73.3 MW$_t$), one for the 100 to 250×10^6 Btu/h (29.3 to 73.3 MW$_t$) units, and a third for those units less than 100×10^6 Btu/h (29.3 MW$_t$) heat input.[2,3] Finally, if a new plant is in a nonattainment area, the permissible particulate emissions and opacity may be significantly reduced from nominal control levels.

Particulate control technologies

Particulate emissions from the combustion process are collected by particulate control equipment (Fig. 2). This equipment must remove the particulate from the flue gas, keep the particulate from re-entering the gas, and discharge the collected material. There are several major types of equipment available including electrostatic precipitators, fabric filters (baghouses), mechanical collectors and venturi scrubbers. Each of these uses a different collection process with different

factors affecting the collection performance. Prior to the 1990s, the technology of choice for large coal-fired utility plants was the dry electrostatic precipitator, with fabric filters a distant second. Today, preference is given to the fabric filter (pulse jet type) for reasons described below.

Dry electrostatic precipitators

A dry electrostatic precipitator (ESP) electrically charges the ash particles in the flue gas to collect and remove them. The unit is normally comprised of a series of parallel, vertical metallic plates (collecting electrodes or CEs) forming ducts or lanes through which the flue gas passes. Centered between the CEs are discharge electrodes (DEs) which provide the particle charging and electric field. Fig. 3 is a plan view of a typical ESP section which indicates the process arrangement.

Charging The CEs are typically electrically grounded and connected to the positive polarity of the high voltage power supply. The DEs are suspended off of electrical insulators in the flue gas stream and are connected to the output (negative polarity) of a high voltage power source, typically 55 to 85 average kV DC. An electric field is established between the DEs and the CEs, and the DEs will exhibit an active glow, or *corona*. As the flue gas passes through the electric field, the particulate takes on a negative charge which, depending on particle size, is accomplished by field charging or diffusion.

Collection The negatively charged particles are attracted toward the grounded CEs and migrate across the gas flow. Some particles are difficult to charge, requiring a strong electric field. Other particles are charged easily and driven toward the plates, but also may lose the charge easily, requiring recharging and recollection. Gas velocity between the plates is also an important factor in the collection process since lower velocities permit more time for the charged particles to move to the CEs and reduce the likelihood of loss back into the gas stream (re-entrainment). In addi-
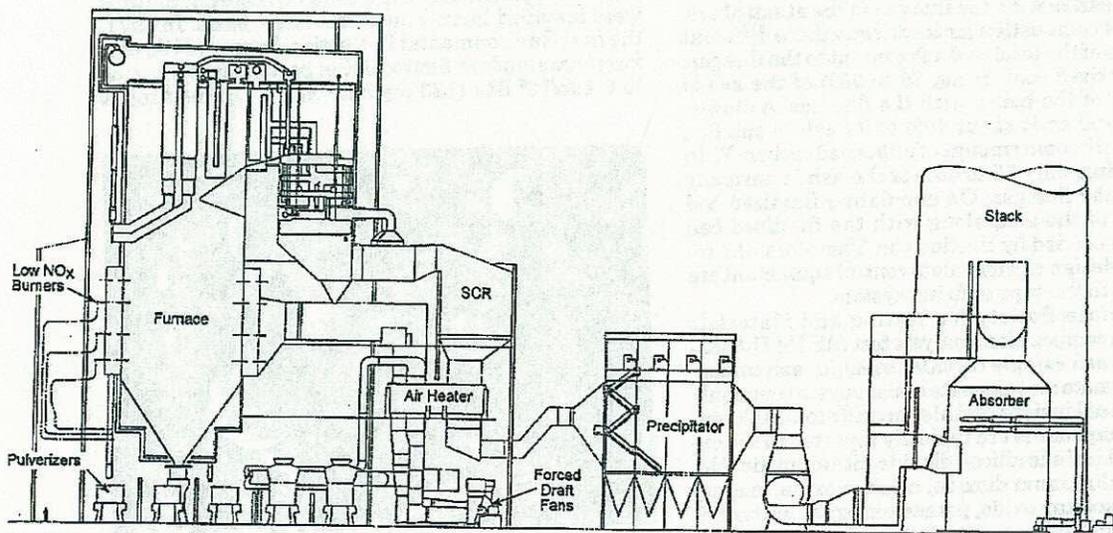
Fig 2 Particulate control equipment – plant side view

Low NOx Burners
Furnace
SCR
Air Heater
Pulverizers
Forced Draft Fans
Precipitator
Absorber
Stack

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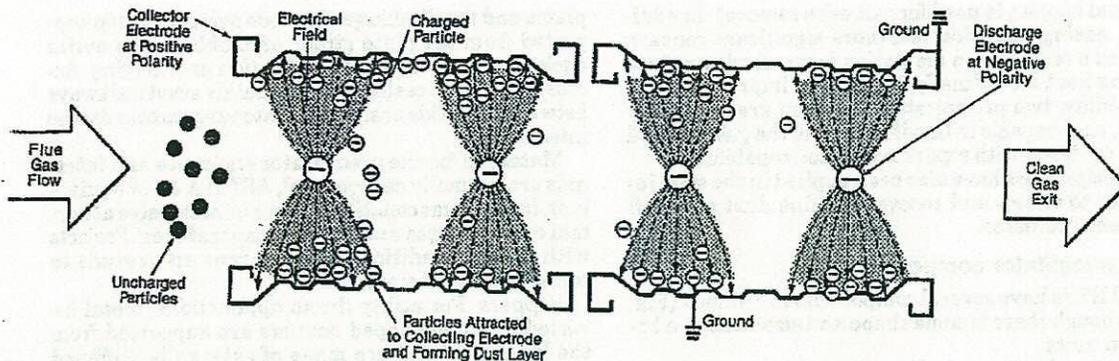BS-02

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Fig. 3 Particle charging and collection within an ESP

tion, a series of CE and DE sections is generally necessary to achieve overall particulate collection requirements. In modern ESPs designed for utility flyash, four or more sections in series are required to achieve design collection levels.

The ash particles form an ash layer as they accumulate on the collection plates. The particles remain on the collection surface due to the forces from the electric field as well as the molecular and mechanical cohesive forces between particles. These forces also tend to make the individual particles agglomerate, or cling together.

Rapping The ash layer must be periodically removed. The most common removal method is rapping which consists of mechanically striking the collection surface; this rapping force dislodges the ash. Because particulate tends to agglomerate, the ash layer is removed in sheets. This sheeting is important to prevent the re-entrainment of individual particles into the flue gas stream, requiring recharging and recollection downstream.

While most of the particles are driven to the CEs, some positively charged particles attach to the DEs. A separate rapping system is therefore used to remove deposits from these electrodes and maintain proper operation.

Ash removal The dislodged particulate falls from the collection surface into hoppers. Once the particulate has reached the hopper it is important to ensure, by proper design, that it remains there in bulk form with minimal re-entrainment until the hopper is emptied. (See Chapter 24 for hopper ash removal methods and equipment.)

Dry ESP applications

Utility Because coal is the most common fuel for steam generation, collection of the coal ash particles is the greatest use of a particulate collector. The electrostatic precipitator has been the most commonly used collector. To meet the particulate control regulations for utility units and considering the resulting high collection efficiency, special attention must be given to details of precipitator sizing, rapping, flow distribution and gas bypass around the collector plates. The result will then be a collector that can be confidently and consistently designed and operated to meet the outlet emissions requirements. Operating collection efficiencies which exceed 99.9% are common on the me-

dium and higher ash coals with outlet emissions levels of 0.01 to 0.03 lb/10⁶ Btu (12.3 to 36.9 mg/Nm³ at 6% O_2) heat input common on all coals. ESPs are also widely installed on utility boilers that fire oil as their principal fuel.

Industrial Other common noncoal-fired industrial units where ESPs are successfully being applied include municipal refuse incinerators and wood, bark, and oil-fired boilers. For these, the ash in the flue gas is typically more easily collected than coal flyash so an ESP of modest size will easily collect the particulate. The moisture content in the refuse, wood and bark is the major contributor to the low resistivity. The carbon content of the residue, ash and unburned combustibles also contributes to low resistivity. (See note below.)

Pulp and paper In the pulp and paper industry, precipitators are used on power boilers and chemical recovery boilers. The power boiler particulate emissions requirements are the same as those for the industrial units using the same fuels. For the recovery boilers, precipitators are used to collect the residual salt cake in the flue gas. Chapter 28 contains further information on the recovery boiler processes and the reuse of the collected material.

A recovery boiler is a unique application for a precipitator due to the small particulate size and the tendency for the cohesive ash particles to stick together. The resistivity of the particulate is low so it is collected easily in the ESP. However, the fine particulate can also cause problems with the generation of effective corona by the DEs due to an effect called space charge. Because the particulate is so small, gas bypass around collector plates and re-entrainment of rapped particulates in the flue gas are more of a design concern. Re-entrainment is minimized by proper gas flow control and by lower gas velocities. Precipitator collection efficiencies are 99.7 to 99.8% to meet the 20% opacity and the local emissions requirements. Due to the characteristics of the salt cake particulates, a drag chain conveyor across a precipitator floor, rather than a

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Note: Resistivity is an inverse measure of a particle's ability to accept and hold a charge. Lower resistivity indicates improved ability to accept a charge and be collected in an ESP.

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normal hopper, is used for salt cake removal. In addition, casing corrosion is a more significant concern and as a result more insulation is required to reduce casing heat loss. Finally, in order to improve system reliability, two precipitator chambers are commonly used, each capable of handling 70% of the gas flow and each equipped with separate isolation capabilities.

Precipitators have also been applied in the steel industry to collect and recover the fine dust given off by some processes.

Dry precipitator components

All ESPs have several components in common (Fig. 4) although there is some shape and size variation between units.

Casing As shown in Fig. 4, the structure forming the sides and roof of an ESP is a gas-tight metal cased enclosure. The structure rests on a lower grid, which serves as a base and is free to move as needed to accommodate thermal expansion All of the collecting plates and the discharge electrode system are top supported from the plate girder assemblies. The entire enclosure is covered with insulation and lagging. Access doors in the casing and adequately sized walkways between the fields assist in maintenance access for the internals.

Materials for the precipitator enclosure and internals are normally carbon steel, ASTM A-36 or equivalent, because gas constituents are noncorrosive at normal operating gas and casing temperatures. Projects with special conditions may warrant an upgrade in some component materials.

Hoppers For utility flyash applications, metal inverted pyramid-shaped hoppers are supported from the lower grid and are made of externally stiffened casing. The hoppers provide the lower portion of the overall enclosure and complete the gas seal. Their sides are designed with an inclination angle of at least 60 deg from horizontal. Hoppers are generally designed as particulate collection devices that can store
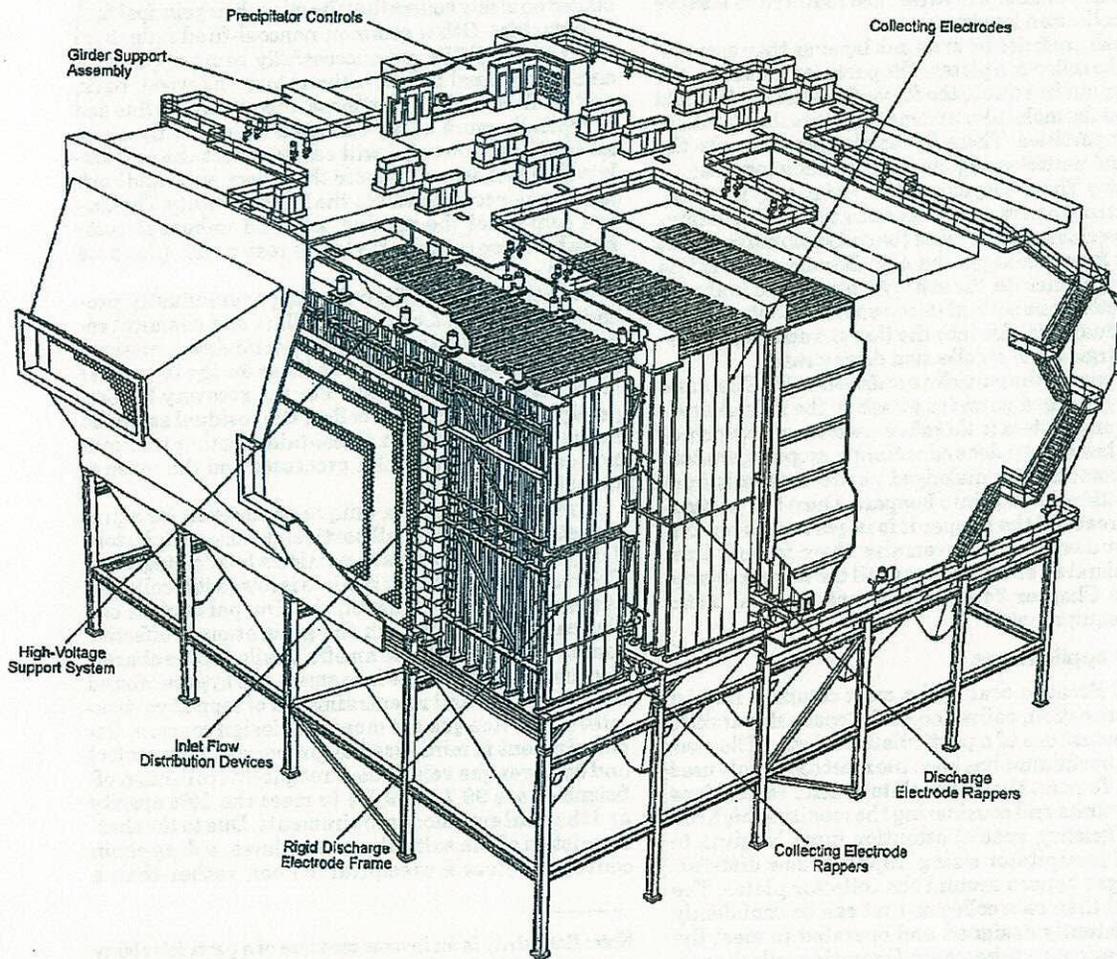

Fig. 4 B&W rigid frame electrostatic precipitator.

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ash for short periods of time when the ash removal system is out of service.

Because many ash removal systems are noncontinuous, the following items are normally supplied with the precipitator hoppers to ensure good particulate removal: hopper heaters, electromagnetic vibrators, poke holes, anvil bars and level detectors. Hot air fluidizing systems are also sometimes supplied to assist in ash removal.

For non-utility flyash applications, the use of alternative ash holding/removal designs are common. These include flat-bottom with drag scrapers, wet-bottom, and trough-type hoppers.

Collecting electrodes As shown in Fig. 4, the CE surface typically consists of a series of roll-formed collector plates assembled into a CE and supported from the top. CEs are also widely referred to as plates or curtains. The CEs are spaced in rows across the width of the precipitator, typically on 12 or 16 in. (305 or 406 mm) centers in modern designs that also use rigid DEs. In older designs, CE spacings of 9 or 10 in. (229 or 254 mm) were generally seen when used with wire (0.109 in./2.8 mm diameter typical) DEs. In the direction of gas flow, the CE sections are arranged into fields which are normally powered by separate and dedicated power supplies. The collection surface area in the Deutsch-Anderson equation (see Equation 1) is the total CE plate area required for particulate collection. For calculating surface area, the CE assembly is treated as a plane and includes both sides of the CE where exposed to gas flow. The rolled plates can be up to 50 ft (15.2 m) in length with a shop straightness tolerance of 0.5 in. (12.7 mm). CEs may also be large flat plates with stiffener bars added to maintain straightness. For optimum performance with a uniform electric field and with minimal inter-electrode sparkover or electrical arcing (high current spark), the alignment of collection surface and electrodes must be maintained within tight tolerances.

Discharge electrodes and insulators As described in the section on charging, the DEs, connected to the high voltage power source, are located in the gas stream and serve as the source of the corona discharge. These electrodes are the central components of the discharge system which is electrically isolated from the grounded portions of the ESP. An electrical clearance gap of 6 to 8 in. (152 to 203 mm), depending on CE spacing, must be maintained throughout the ESP between the DE system and any grounded components.

Discharge electrodes are found in several shapes. Common types include the rigid frame, rigid electrode and weighted wire. The rigid frame, shown in Fig 5, consists of strips of electrode supported between sections of frame tubing. Each frame is attached to a structural carrier, both front and rear. This assembly is supported by insulators forming a four point suspension system. The rigid electrode consists of a member with proprietary shape that is top supported and hangs the full height of the precipitator. The typical rigid electrode top support is also a frame hanging from insulators. The lower ends of the rigid electrode have a guide bar and side to side spacers. The third type of discharge electrode, weighted wire, consists of a round or barbed wire supported at the top and held straight and in tension with a weight at the bottom. The upper frame is supported from insulators and there is a lower steadying frame to guide and space the electrodes.

For highest equipment reliability, either the rigid frame or rigid electrode is the most common configuration. Discharge electrode failure in the form of broken wires has been a recurring problem with the weighted wire electrodes, particularly with lengths of 30 ft (9 1 m) or more, resulting in performance deterioration.

Gas flow control devices ESPs operate most efficiently when the gas flow is distributed evenly across the ESP cross section. Flow control devices such as turning vanes, flow straighteners and perforated plates are frequently installed in the inlet and outlet flues, and flow transition pieces, to provide the desired degree of flow uniformity while optimizing resultant pressure drop.

Rapping systems As shown in Fig. 4, the most effective method of cleaning the collector curtains is to rap each one separately and in the direction of gas flow. This method assures that each curtain receives a rapping force. The rapping system shown is a tumbling hammer type, where the hammer assemblies are mounted on a shaft extending across the ESP in a staggered arrangement. The shaft is turned slowly by an external drive controlled by timers for rapping fre-
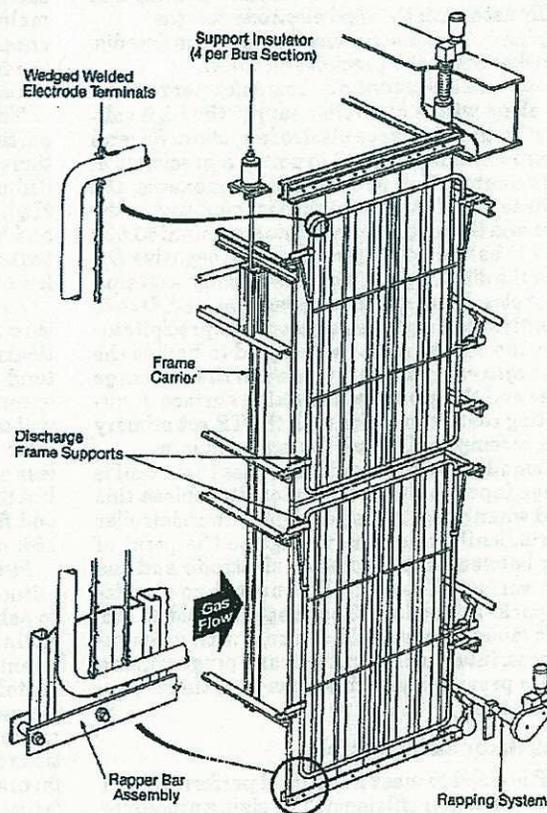
Fig. 5 Rigid frame discharge electrode and rapping system for an ESP.

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quency and optimum cleaning. Hammer size is selected to match the application and size of the collector curtain. External, top-mounted rapping technology has also been widely used, principally in the U.S. Typically, more than one collector curtain is rapped at a time with this method, and the rapping force is in the downward direction on the top edge of the curtains. Both a drop rod and magnetic impulse are the drive mechanisms used.

Due to the difficulty of cleaning high resistivity flyash from the collection surface, considerable tests have been performed to ensure that adequate rapping forces are transmitted across the entire collection surface. A minimum acceleration of 100 g applied at the farthest point from impact has been established as an industry standard.

Typically, rapping of the rigid frame discharge electrodes is accomplished using a tumbling hammer system as shown in Fig. 5. The hammer assemblies are mounted in a staggered arrangement on a shaft across the width of an electrical section. Note that a smaller hammer than that used for the collector system is required for proper cleaning of the discharge electrodes. An external drive unit mounted on the precipitator roof is used to slowly turn the rapper shaft and, because it is attached directly to the carrier frame, the drive shaft must also be electrically isolated with an insulator. As with the collector system, top rapping of the discharge electrodes is another method of cleaning that is typically used with the rigid electrode designs.

A rapping system is sometimes used on the flow distribution devices at the precipitator inlet.

Power supplies and controls A transformer rectifier (TR) set along with a controller supply the high voltage power to the discharge electrode system. Several TR sets are normally needed to power a precipitator. With this combination of electrical components, the single-phase 480 VAC line voltage is regulated in the controller and then transformed into a nominal 55,000 to 75,000 V before being rectified to a negative DC output for the discharge system. Electrically, a precipitator most closely resembles a capacitive load. Due to the capacitive load and the nature of the precipitator internals, the TR set must be designed to handle the current surges caused by arcs between the discharge electrodes and the grounded collection surface. A current-limiting reactor in series with the TR set primary also helps to temporarily limit the current surges.

Traditionally, a voltage controller tries to maximize the voltage input to the precipitator. To achieve this input and when operating as designed, the controller must periodically raise the voltage to the point of sparking between the discharge electrode and the collection surface. The controller must then also detect the sparks and reduce the voltage to avoid an arc.

Today's microprocessor TR controls with quick response times, interface advantages and programming capabilities provide many functions to optimize particulate collection.

Dry precipitator sizing factors

An ESP is sized to meet a required performance or particulate collection efficiency. The sizing procedure determines the amount of collection surface to meet the specified performance. An equation which relates the collection efficiency (E) to the unit size, the particle charging and the collection surface is the Deutsch-Anderson equation:[1]

$$E = 100\% \times \left(\frac{\text{Inlet dust loading} - \text{Outlet dust loading}}{\text{Inlet dust loading}} \right) \quad (1)$$

$$1 - \frac{E}{100} = e^{\frac{-wA}{V}} \quad (2)$$

or

$$A = \left[\ln \left(\frac{1}{1 - (E/100)} \right) \right] \frac{V}{w} \quad (3)$$

where

E = ESP removal efficiency, % of inlet particulate removed
w = migration velocity, ft/min (m/s)
A = collection surface area, ft² (m²)
V = gas flow, ft³/min (m³/s)

Migration velocity is the theoretical average velocity at which the charged particles travel toward the collection surface. This velocity is dependent upon how easily the particulate is charged, and the value is normally selected by empirical means based on experience. The factors which affect migration velocity are the fuel and ash characteristics, the operating conditions, and the effects of gas flow distribution.

These factors also have an effect on the ability of the particulate to accept a charge. A commonly used indication of this effect is resistivity, measured in ohm-cm. Fig. 6 illustrates typical resistivity curves for two fuel ashes. High resistivity ashes result in low migration velocities and large collection surface areas while average resistivity ashes result in moderately sized surface areas, i.e., lower resistivity indicates improved collection.

As previously discussed, the flue gas will pass through a series of collection fields. ESPs collect particulate in geometric fashion meaning each field will tend to collect very roughly 80% of the particulate entering that particular field. Therefore, the inlet field will collect the vast majority of the total mass particulate with each successive field collecting geometrically less of the total. For example, the first field could collect 80% of the total particulate loading while the second field would collect 80% of the remaining 20% or 16% of the total particulate loading, etc.

Fuel and ash characteristics The fuel and ash constituents which reduce resistivity or that are favorable to ash collection in an ESP include moisture, sulfur, sodium and potassium. Applications with sufficient quantities of these components usually result in moderately sized precipitators. Constituents which hamper ash collection and increase outlet emissions include calcium and magnesium. High percentage concentrations of these items without offsetting quantities of the favorable constituents result in poorer collection and larger precipitators. The fuel and ash constituents and their relative quantities must be reviewed in the siz-

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ing process to determine the overall effect on migration velocity. The migration velocity/resistivity can then be altered to some extent by controlling the content of the critical constituents.

Gas temperature, volume and distribution As indicated in Fig. 6, gas temperature has a direct effect on resistivity and on the gas volume passing through the ESP. Gas flow from the boiler also has an effect on sizing as indicated by the Deutsch-Anderson equation. There is an optimum gas velocity range within an ESP for maximum performance which must be considered as part of the design selection. Maximum ESP efficiency is achieved when the gas flow is distributed evenly across the unit cross section. Uniform flow is assumed in the ESP sizing calculations and should be verified during the design stage by using a flow model. These models should include the precipitator as well as the inlet and outlet flues. Flow uniformity is typically achieved by installing distribution devices in the flue transition sections immediately upstream and downstream of the ESP. Hopper design must also prevent high velocity areas to avoid flyash re-entrainment. The industry standard for flow distribution and modeling is the Institute of Clean Air Companies EP-7.[5]

Particle size distribution Particle size also affects ESP design and performance. A particle size distribution versus collection efficiency curve (Fig. 7) indicates that an ESP is less efficient for smaller particles (less than 2 microns) than for larger ones. Therefore, ESP applications with a high percentage of particles less than 2 microns will require more collection surface and/or lower gas velocities.

Sectionalization Sectionalization refers to how many or how few independent high voltage power supplies are installed on an ESP. ESP performance improves with additional, independently energized sections, but improving performance must be weighed against additional capital and operating costs. Proper sectionalization can overcome the effects of a poorly operating section(s) due to adverse process or mechanical/electrical conditions.
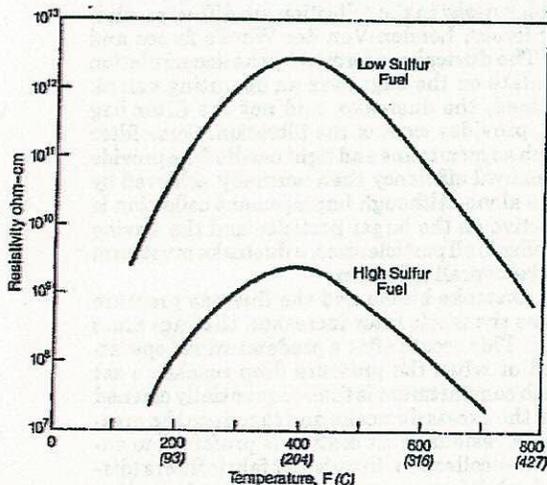

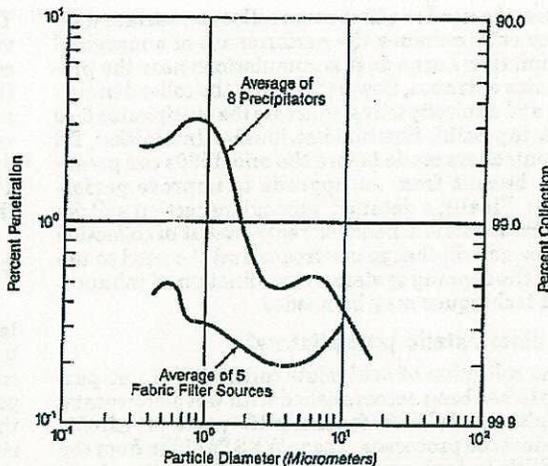**Fig. 6** Typical ash resistivity.

Fig. 7 Summary of fine particulate collection (adapted from Reference 4).

Rapping effects Rapping causes some amount of particulate re-entrainment in the flue gas flow. Given the nature of the precipitation process, adjustment of rapping (how often and how intense) can greatly impact outlet emissions, either positively or negatively. Proper rapping adjustment in the outlet sections is especially important because there are no downstream sections to handle re-entrainment due to over-rapping. There are no set industry standards for rapping adjustments although many suppliers begin with proprietary guidelines. Effective fine-tuning is usually based upon the supplier's field experience or plant personnel familiar with the specific ESP.

Ash removal Proper ESP design will prevent operational issues associated with re-entrainment from the hoppers. However, a malfunctioning ash removal system can also cause problems. High ash levels can permit excessive re-entrainment, and can cause electrical malfunction if the ash pile contacts the discharge electrode system (high resistance grounding). If not corrected, ash removal malfunction can also cause the ash pile to float upward and distort the DEs.

Performance enhancements

A change in fuel, a boiler upgrade, a change in regulation, or performance deterioration may call for a precipitator performance enhancement. Enhancement techniques include additional collection surface, flue gas conditioning, improved/modified gas flow distribution, additional sectionalization, additional rapping, control upgrades and internals replacement. Gas conditioning alters resistivity by adding sulfur trioxide (SO_3), ammonia, moisture, or sodium compounds while the other modifications involve only mechanical and/or electrical hardware changes.

After identifying the causes of current or anticipated performance deterioration, the equipment is surveyed to determine the need for replacement or upgrade. Additional collection surface, in series or in parallel with existing surface, may be needed to meet improved particulate collection needs. Gas condition-

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ing may be used to offset some collection surface deficiency or to enhance the performance of a marginal precipitator. Large dust accumulations near the precipitator entrance, flow patterns on the collection surface, and a velocity traverse across the precipitator face indicate possible flow maldistribution. In addition, TR set controllers made before the mid-1980s can potentially benefit from an upgrade to improve performance. Finally, a detailed internal inspection will determine a possible need for replacement of collection surface and discharge electrodes and the need to upgrade the rapping system. A combination of enhancement techniques may be needed.

Wet electrostatic precipitators[6]

The collection of acid mists consisting of fine particulate has been accomplished with wet electrostatic precipitators (WESPs) for nearly 100 years, principally on industrial processes. These WESPs differ from the dry ESPs in both materials and configuration; however, the collection mechanism is basically the same. Typical operation is at or below the flue gas acid moisture dew point temperature for the gas being filtered, and particulate loading is low compared to normal coal-fired dry ESP applications. Collection efficiencies of 99% have been reported with wet ESPs when precipitator sections or modules are placed in series. (See Chapter 35, Fig. 15.)

Boiler applications Wet ESPs have not historically been used for utility or industrial boiler emissions control when firing coal, oil or gas. However, with the emergence of expanded emission control requirements, reduced emission limits, use of non-traditional fuels and the interactions of other emissions control equipment, there has been renewed interest in the use of WESPs to control selected emissions, especially of sulfuric acid mist and fine dry particulates. Firing of petroleum coke and Orimulsion® fuel can result in elevated levels of SO_3 in the flue gas. In addition, as noted in Chapter 34, SCRs on coal-fired systems have a tendency to increase the concentration of SO_3 in the flue gas due to the oxidation action of the catalyst. Depending upon the catalyst selection, this oxidation can increase the SO_3 significantly on the higher sulfur fuels. As the SO_3 is carried through the cooler backend equipment, it condenses and becomes an acid mist (H_2SO_4). When wet flue gas desulfurization systems are used for sulfur control, significant levels of SO_3/acid mist tend to pass through the system and result in opacity/visibility problems at the stack. (See Chapter 35.) Acid mist has also been noted on some units without the wet FGD systems. WESPs have demonstrated the ability to effectively collect the H_2SO_4 and other aerosols in such environments.

Wet precipitator design The WESP design generally follows that of dry ESPs. However, WESPs differ from dry ESPs in three key areas. First, the physical arrangement of WESPs can be non-standard compared to dry ESPs. The WESP configuration and shapes of the key components can be adapted to integrate the WESP with other system equipment, such as the case of the integration of a WESP in a wet flue gas desulfurization system discussed in Chapter 35.

The components can also be configured in a more conventional stand-alone arrangement. Second, the moist corrosive atmosphere present requires careful selection of material in critical areas. Third, instead of a rapping system to remove the collected particulate, water spray or a water film removes the material deposited on the collection surfaces and discharge system. The physical arrangement of the bottom of the WESP is then adapted to collect the water film discharge.

Fabric filters

A fabric filter, or baghouse, collects the dry particulate matter as the cooled flue gas passes through the filter material. The fabric filter is comprised of a multiple compartment enclosure (see Fig. 8 and chapter frontispiece) with each compartment containing up to several thousand long, vertically supported, small diameter fabric bags. The gas passes through the porous bag material which separates the particulate from the flue gas.

Operating fundamentals

With the typical coal-fired boiler, the particle laden flue gas leaves the boiler and air heater and enters the filter inlet plenum which in turn distributes the gas to each of the compartments for cleaning. An outlet plenum collects the cleaned flue gas from each compartment and directs it toward the induced draft fan and the stack. Inlet and outlet dampers then allow isolation of each compartment for bag cleaning and maintenance. Each compartment has a hopper for inlet gas flow as well as for particulate collection and removal by conventional equipment, as discussed further in Chapter 24. The individual bags are closed at one end and connected to a tubesheet at the other end to permit the gas to pass through the bag assembly. The layer of dust accumulating on the bag is usually referred to as the *dustcake*.

Collection of the particulate on the bag fabric is the heart of the filtering process. The major forces causing this collection include impingement by either direct contact or impaction and dustcake sieving. Minor forces which assist in the collection are diffusion, electrostatic forces, London-Van der Waal's forces and gravity.[7] The dustcake is formed by the accumulation of particulate on the bags over an operating period. Once formed, the dustcake, and not the filter bag material, provides most of the filtration. Some filter media such as membrane and tight needle felts provide higher removal efficiency than normally achieved by filter cake alone. Although impingement collection is most effective on the larger particles and the sieving process collects all particle sizes, a dustcake must form to maximize overall collection.

As the dustcake builds and the flue gas pressure drop across the fabric filter increases, the bags must be cleaned. This occurs after a predetermined operating period or when the pressure drop reaches a set point. Each compartment is then sequentially cleaned to remove the excess dustcake and to reduce the pressure drop. A residual dust coating is preferred to enhance further collection. In pulse jet fabric filters (discussed below), it is common to perform cleaning online which maintains all compartments in operation.

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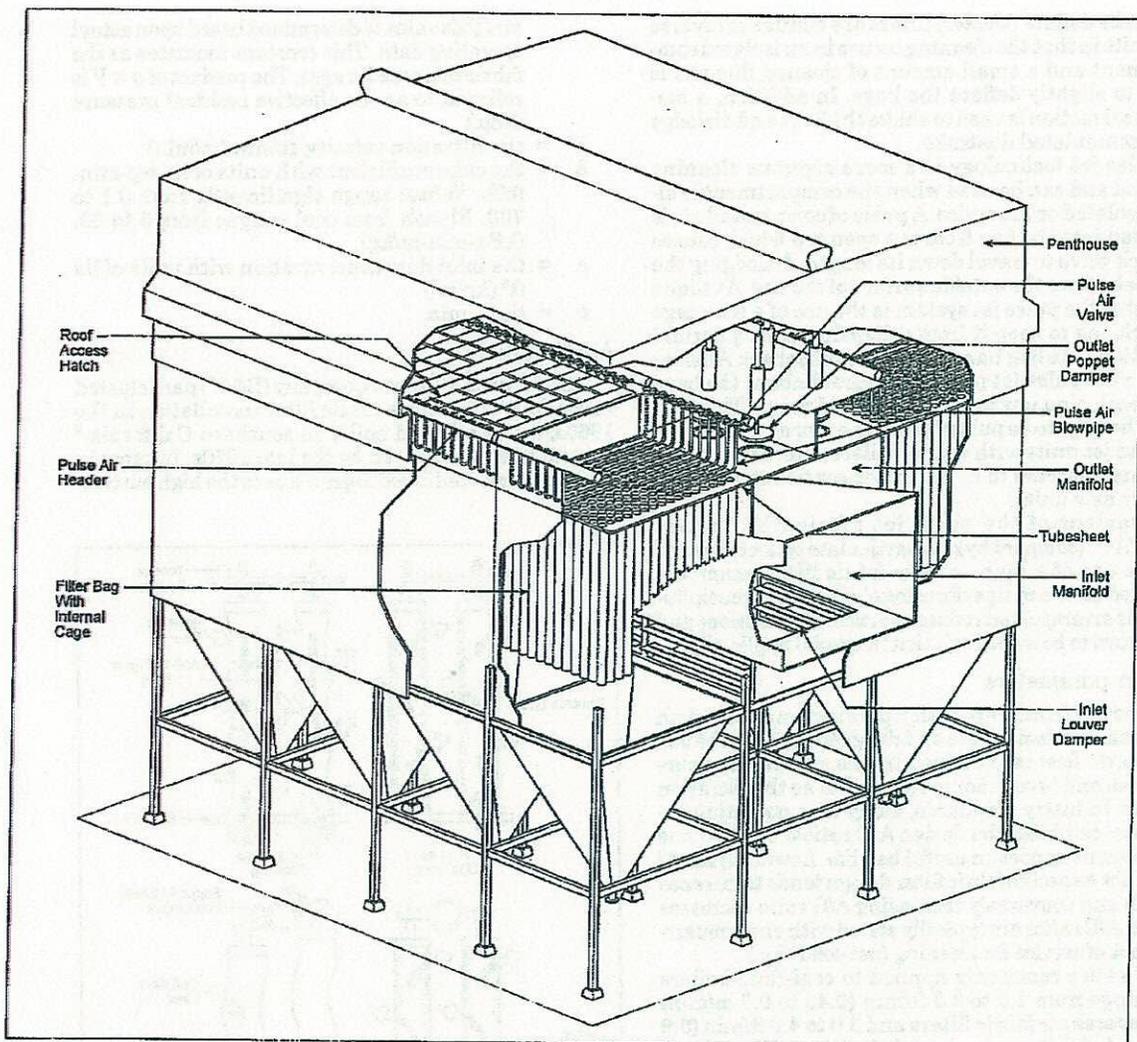BS-02

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Fig. 8 Pulse jet fabric filter.

Initial filter bags have a pre-coat filter media. The pre-coat is an optimum particle size mixture which aids in developing the first filter cake and tends to stay within the outer surface of the filter media.

Types of fabric filters

Bag cleaning methods distinguish the types of fabric filters, with the three most common types being reverse air, shake deflate and pulse jet. The cleaning method also determines the relative size by the air/cloth (A/C) ratio and the filtering side of the bag. Both the reverse air and the shake deflate are inside-the-bag filters with gas flow from inside the bag to outside; the pulse jet is an outside-the-bag filter with flow from outside to inside (Fig. 9). Note that the tubesheet on the inside-bag filtering is located below the bags; for the pulse jet the tubesheet is above the bags.

A reverse air, more correctly termed *reverse gas*, filter reverses the flow of clean gas from the outlet plenum back into the bag compartment to collapse the bags in an isolated compartment and dislodge the dustcake. This is a gentle cleaning motion. Once the dislodged particulate falls to the hopper, the bags are gently reinflated before full gas flow is allowed for filtering. This system requires a reverse gas fan to supply the cleaning gas flow along with additional dampers for flow control. This type of filter system has been used in most large utility power plant fabric filters in the U.S. to date. Experience with this type of fabric filter on some coal flyash applications has demonstrated that reverse gas cleaning alone does not provide an acceptable operating pressure drop. Therefore, some units have added sonic air horns to each compartment to assist in the cleaning.

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Shake deflate (*shaker*) filters are similar to reverse air units in that the cleaning occurs in an isolated compartment and a small amount of cleaned flue gas is used to slightly deflate the bags. In addition, a mechanical motion is used to shake the bags and dislodge the accumulated dustcake.

Pulse jet technology is a more rigorous cleaning method and can be used when the compartment is either isolated or in service. A pulse of compressed air is directed into the bag from the open top which causes a shock wave to travel down its length, dislodging the dustcake from the outside surface of the bag. A unique aspect of the pulse jet system is the use of a wire cage in each bag to keep it from collapsing during normal filtration. The bag hangs from the tubesheet. A series of parallel pulse jet pipes are located above the bags with each pipe row having a solenoid valve. This permits the bags to be pulsed clean one row at a time. Use of pulse jet units with utility boilers is increasing and they are expected to be the major contributor for retrofit or new units.

A variant of the pulse jet fabric filter called COHPAC® (compact hybrid particulate collector) is the unique use of a high velocity fabric filter commonly installed after a malperforming electrostatic precipitator. This arrangement reduces particulate emissions and may prove to be a good solution in special applications.

Design parameters

Three key fabric filter design parameters are air/cloth (A/C) ratio, emission rate and drag. A/C ratio is the gas volumetric flow rate divided by the exposed bag surface area and is commonly referred to as the filtration velocity. Industry standards, along with operating experience, establish the design A/C ratios. The A/C has a significant impact on useful bag life. Lowering the A/C ratio for a specific fabric filter design tends to increase bag life and conversely increasing A/C ratio decreases bag life. A/C ratios are typically stated with one compartment out of service for cleaning (net condition).

A/C ratios commonly applied to coal-fired boilers may range from 1.5 to 2.3 ft/min (0.45 to 0.7 m/min) with reverse air fabric filters and 3.0 to 4.0 ft/min (0.9 to 1.2 m/min) for pulse jet fabric filters. The type of firing and ranges in fuel significantly affect the appropriate filtration velocity.

The emission rates are generally established based upon historical data. Several prediction models have been developed but they are of limited use. Bench tests using collected particulate samples have proven beneficial as an indicator for media selection.

The pressure drop includes the drop across the bags, the dustcake and the attachment of the bag to the tubesheet. The calculation of drag is useful in evaluating performance.

The general equation for drag can be expressed as follows:

$$\Delta P = a \times V + b \times c \times t \times V^2 \qquad (4)$$

where

ΔP = pressure drop, in. wg (kPa)
a = constant with units of in. wg-min/ft (kPa-min/

m) (This value is determined based upon actual operating data. This constant increases as the fabric filter media ages. The product of $a \times V$ is referred to as the effective residual pressure drop.)
V = the filtration velocity, ft/min (m/min)
b = the cake coefficient with units of in. wg-min-ft/lb. Values range significantly from 0.1 to 700. Flyash from coal ranges from 5 to 20. (kPa-min-m/kg)
c = the inlet dust concentration with units of lb/ft³ (kg/m³)
t = time, min.

Applications

The Babcock & Wilcox Company (B&W) participated in the first U.S. utility fabric filter installation in the 1960s, on an oil-fired boiler in southern California.[8] Several utilities followed by the late 1970s. Interest in these systems continues to grow due to the high particu-
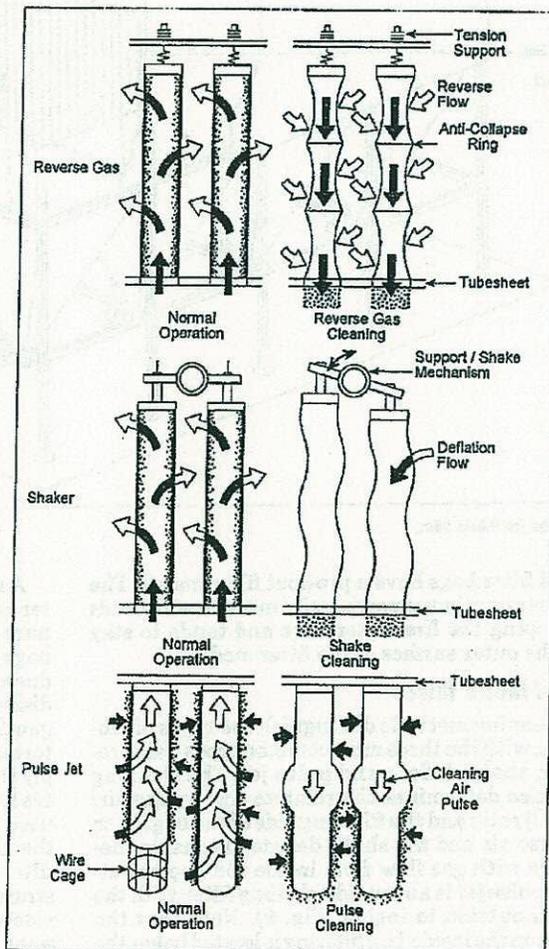

Fig. 9 Fabric filter types.

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late removal efficiency. Well designed filters routinely achieve greater than 99.9% particulate removal, meeting all current U.S. EPA and local regulations.

Besides standard utility coal-fired applications, fabric filters are used on circulating fluidized-bed boilers, industrial pulverized and stoker coal units, refuse-fired units in combination with a dry flue gas scrubber, and in the cement and steel industries. Fabric filters are not currently used on oil-fired units due to the sticky nature of the ash.

A unique advantage with fabric filters is that all of the flue gas passes through the dustcake as it is cleaned. When the dustcake has high alkalinity, it can be used to remove other flue gas constituents and acid gases, such as sulfur dioxide (SO_2). (See Chapter 35.) Addition of activated carbon into the flue gas has been used to control mercury emissions. Fabric filters generally are better collectors of the carbon and mercury components than are electrostatic precipitators.

Fabric filter components

Configurations Pulse jet fabric filters have two general arrangements. The first is called *hatch style*, that has large horizontal covers used to access the area above the tubesheet for bag replacement. The most common arrangement includes a penthouse above the hatch covers. The second type, called *walk-in*, has a gas-tight compartment above the tubesheet with vertical doors to access the area above the tubesheet for bag replacement. Walk-in arrangements are more common for industrial sized units with shorter bags.

Reverse air and shaker units generally have access through vertical doors located at the tubesheet level and also located at a level near the top of the bags.

All styles of fabric filters can be provided with integral compartments sharing common division walls or with separate walls forming the compartments. The latter style is commonly referred to as modular and generally used on small industrial units.

The gas conveying flues of all styles of fabric filters can be of a design integral to the compartment casing walls or totally separated. Most utility size fabric filters utilize common wall construction.

Casing, hoppers and dampers

Enclosure or casing The fabric filter is a metal encased structure with individual bag compartments. The inlet and outlet plenums are typically located between two rows of compartments to provide short inlet and outlet flue connections (see Fig. 8). This enclosure rests on a support steel structure. For reverse air units, interior access is required at both the lower tubesheet and bag support elevations. In a pulse jet filter, access is required above the tubesheet for bag cage removal. This is provided by large roof access doors or by a top plenum and a side manway. Typical enclosure materials are carbon steel ASTM A-36 or equivalent under normal coal-fired boiler conditions. The entire enclosure is covered with insulation and lagging to keep metal temperatures high and to minimize corrosion potential.

Hoppers Each filter compartment has a hopper to collect the dislodged particulate and to channel its flow to the ash removal system. Most filters also use the hopper as part of the flue gas inlet to each compartment. Therefore, the hopper is designed with steep sides for ash removal along with considerations for proper gas flow distribution. Hopper heaters, level detectors, poke holes and an access door are common hopper features.

Dampers Each filter compartment may be provided with both inlet, outlet and on reverse air filters, deflation dampers. These dampers may be manual or automated. Louver and poppet dampers are normally used. Dampers allow isolation of compartments for offline cleaning and online service entry. Some units include bypass dampers.

Bag materials and supports

Substantial research and development on bags and their materials have taken place to lengthen their life and to select bags for various applications. (See Table 1.) The flexing action during cleaning is the major factor affecting bag life. Bag blinding, which occurs when small particulate becomes trapped in the fabric interstices, limits bag life by causing excessive pressure drop in the flue gas. Finishes on the bag surface are also used to make some bags more acid resistant and to enhance cleaning.

The most common bag material in coal-fired utility units with reverse fabric filters is woven fiberglass. Typical bag size is 12 in. (305 mm) diameter with a length of 30 to 36 ft (9.1 to 11.0 m). Bag life of three to five years is common. The shake deflate filters also use mostly fiberglass bags. On both of these units, the fiberglass bag is fastened at the bottom to a thimble in the tubesheet. At the top, a metal cap is fitted into the bag and the bag has a spring loaded support for the reverse air filters. The bags are attached to the tubesheet commonly using a thimble and clampless designed bag. The upper operating temperature limit is 500F (260C) for most fiberglass bags.

The most common bag material in coal-fired utility units with pulse jet fabric filters is polyphenylene sulfide (PPS) needled felt. In addition to polyphenylene sulfide, fiberglass, acrylic, polyester, polypropylene, Nomex®, P84®, special high temperature fiberglass media, membrane covered media, and ceramic are used in various applications. For the pulse jet filters, the typical bag size is 5 or 6 in. (127 or 152 mm) diameter round or oval with a length of 10 to 26 ft (3 to 8 m). Advances in cleaning technology are increasing the ability to provide longer bags. Pulse jet bags are commonly sealed to the tubesheet using a snap-band seal. The bag is supported internally from a metal wire cage to prevent bag collapse during operation. Cages are normally carbon steel and may include a variety of coatings from pre-galvanized to coated wire. Some applications use a stainless steel cage.

Filter media cleaning

Reverse air fabric filters employ the most gentle form of cleaning. The method dislodges collected particulate by closing compartment dampers and, with the use of a reverse air fan, reversing the direction of gas flow. During reverse flow, gas passes from the outside of the bag to the inside, dislodging the collected ash from the in-

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side surface of the bag. Sonic horns have proven successful is certain applications to assist in reverse air cleaning.

The shaker style of cleaning begins with the isolation of a compartment. Next, mechanical energy in the form of sinusoidal acceleration is applied to the top of the bag and travels along the bag to dislodge the material caked on the filter.

Pulse jet cleaning can be accomplished with all compartments operating, which is more normal, or by removing a compartment from service. Pulse air varies from low pressure, high volume to high pressure, low volume. The components of most pulse cleaning systems are similar. Blowers or air compressors are used to supply the cleaning air. Pulse valves are rapid opening and closing valves. Air is discharged from the outlet of the pulse valve along a supply pipe to nozzles or holes for discharge into the top of the bag. This air inflates the bag, dislodging collected filter cake from the outside surface of the bag. The more common style of pulse system has fixed supply pipes called blowpipes. An alternate type of pulse system has a rotating blowpipe.

Mechanical collectors

Mechanical dust collectors, often called cyclones or multiclones, have been used extensively to separate large particles from a flue gas stream. The cyclonic flow of gas within the collector and the centrifugal force on the particulate drive the particulate out of the flue gas (Fig. 10). Hoppers below the cyclones collect the particulate and feed an ash removal system. The mechanical collector is most effective on particles larger than 10 microns. For smaller particles, the collection efficiency drops considerably below 90%.

Mechanical collectors were adequate when the emissions regulations were less stringent and when popular firing techniques produced larger particles. These collectors were frequently used for reinjection to improve unit efficiency on stoker firing of coal and biomass. With stricter emissions regulations, mechanical collectors can no longer be used as the primary control device. However, with the onset of fluidized-bed boilers, there has been a resurgence of mechanical collectors for recirculating the bed material. A high efficiency collector is then used in series with the mechanical one to meet particulate emissions requirements. (See Chapter 17 for more information on fluidized-bed combustion.)

Wet particulate scrubbers

A wet scrubber can be used to collect particulate from a flue gas stream with the intimate contact between a gas stream and the scrubber liquid. The venturi-type wet scrubber (Fig. 11) is used to transfer the suspended particulate from the gas to the liquid. Collection efficiency, dust particle size and gas pressure drop are closely related in the operation of a wet scrubber. The required operating pressure drop varies inversely with the dust particle size for a given collection efficiency; or, for a given dust particle size, collection efficiency increases as operating pressure drop increases.

Due to the excessive pressure drop and the stringent particulate regulations, wet particulate scrubbers are now infrequently used as a primary collection device. However, on most coal-fired applications where wet FGD scrubbers are required in series with a high efficiency collector for control of acid gas emissions, the extra particulate removal of the FGD scrubber is an added benefit.

Equipment selection

Major evaluation factors to consider when selecting particulate control equipment include emissions requirements, boiler operating conditions with resulting particulate quantity and sizing, allowable pressure drop/power consumption, combined pollution control requirements, capital cost, operating cost, and maintenance cost. For new units that must meet the stringent federal, state and local regulations, the selection is reduced to a comparison of electrostatic precipitators and fabric filters because these are the only high efficiency, high reliability choices. For retrofits on operating units, the performance of existing control equipment as well as unique flue gas conditions may require specialized equipment.

The advantages of a well designed ESP are high total collection efficiency, high reliability, low flue gas pressure loss, resistance to moisture and temperature upsets, and low maintenance. Advantages of a fabric filter include high collection efficiency throughout the particle size range, high reliability, resistance to flow upsets, little impact of ash chemical constituents on performance, and good dustcake characteristics for combination with dry acid gas removal equipment.

Table 1
Fabric Filter Typical Media Applications

Fabric Filter Type	Construction	Materials	Typical Maximum Operating Temperature	Chemical Resistance
Reverse gas Shake deflate	Woven	Treated fiberglass	500F (260C) ±	Acid resistant coatings available
Pulse jet	Felted	PPS Nomex® P84® Acrylic Polyester	Ambient to 350F (177C) ± depending on specific media selected	Highly variable — consult manufacturer

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Steam
its generation and use

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Chapter 35

Sulfur Dioxide Control

Sulfur appears in the life cycle of most plants and animals. Most sulfur emitted to the atmosphere originates in the form of hydrogen sulfide from the decay of organic matter. These emissions slowly oxidize to sulfur dioxide (SO_2). Under atmospheric conditions, SO_2 is a reactive, acrid gas that can be rapidly assimilated back to the environment. However, the combustion of fossil fuels, in which large quantities of SO_2 are emitted to relatively small portions of the atmosphere, can stress the ecosystem in the path of these emissions.

Man is responsible for the majority of the SO_2 emitted to the atmosphere. Annual worldwide emissions are approximately 160 million tons, nearly half of which are from industrial sources. The two principal industrial sources are fossil fuel combustion and metallurgical ore refining.

When gaseous SO_2 combines with liquid (ℓ) water, it forms a dilute aqueous solution of sulfurous acid (H_2SO_3). Sulfurous acid can easily oxidize in the atmosphere to form sulfuric acid (H_2SO_4). Dilute sulfuric acid is a major constituent of *acid rain*. (Nitric acid is the other major acidic constituent of acid rain.) The respective reactions are written:

$$SO_2(g) + H_2O(\ell) \rightarrow H_2SO_3(aq) \qquad (1)$$

$$O_2(g) + 2H_2SO_3(aq) \rightarrow 2H_2SO_4(aq) \qquad (2)$$
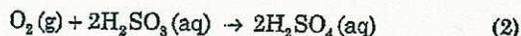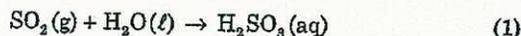
SO_2 can also oxidize in the atmosphere to produce gaseous sulfur trioxide (SO_3). Sulfur trioxide reactions are written:

$$2SO_2(g) + O_2(g) \rightarrow 2SO_3(g) \qquad (3)$$

$$SO_3(g) + H_2O(g) \rightarrow H_2SO_4(\ell) \qquad (4)$$
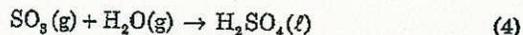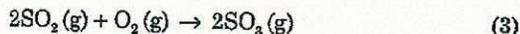
While Equations 1 and 2 describe the mechanism by which SO_2 is converted to sulfuric acid in acid rain, Equations 3 and 4 characterize dry deposition of acidified dust particles and aerosols.

The pH scale, a measure of the degree of acidity or alkalinity, is the method used to quantify the acidity of acid rain.

Pure water has a pH of 7 and is defined as neutral, while lower values are defined as acidic and higher values as alkaline. If rainwater contained no sulfuric or nitric acid, its pH would be approximately 5.7 due to absorption of carbon dioxide (CO_2) from the atmosphere. The contributions of man-made SO_2 and nitrogen oxides (NO_x) further reduce the pH of rainwater. No uniformly accepted definition exists as to what pH constitutes acid rain. Some authorities believe that a pH of about 4.6 is sufficient to cause sustained damage to lakes and forests in the northeastern portion of North America and in the Black Forest region of Europe.

SO₂ emissions regulations

Legislative action has been responsible for most industrial SO_2 controls. Major landmark regulations include the Clean Air Act Amendments of 1970, 1977 and 1990 in the United States (U.S.), the Stationary Emissions Standards of 1970 in Japan, and the 1983 SO_2 Emissions Regulations of the Federal Republic of Germany. Since the mid-1980s, SO_2 emissions regulations have been implemented in most other industrialized nations and many developing nations

SO₂ control

Most utilities have adopted one of two strategies for SO_2 control, either switching to low sulfur coal or installing scrubbers.

A variety of SO_2 control processes and technologies are in use and others are in various stages of development. Commercialized processes include wet, semi-dry (slurry spray with drying) and completely dry processes. The wet flue gas desulfurization (WFGD) scrubber is the dominant worldwide technology for the control of SO_2 from utility power plants, with approximately 85% of the installed capacity, although the dry flue gas desulfurization (DFGD) systems are also used for selected lower sulfur applications.

Total annual SO_2 emissions in the U.S., including electric utility SO_2 emissions, have declined since 1970 as various regulations have been adopted. During the

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same period, electricity generation from coal has almost tripled (see Table 1).[12]

A significant portion of this emissions reduction has been the result of switching to low sulfur coal, predominantly from the western U.S. In 1970 virtually all of the utility coal came from the eastern, higher sulfur coal fields, while by 2000 approximately half of the coal came from western low sulfur sources. Slightly less than two-thirds of SO_2 emission reductions have been attributed to fuel switching while over a third has been through the installation of flue gas desulfurization systems, predominantly wet scrubbers. More than 30% of the U.S. coal-fired capacity already has FGD systems installed and operating. This may come close to doubling over the next decade and a half as existing regulations are implemented and proposed regulations are adopted.

Wet scrubbers

Reagents

Wet scrubbing processes are often categorized by reagent and other process parameters. The primary reagent used in wet scrubbers is limestone. However, any alkaline reagent can be used, especially where site-specific economics provide an advantage. Other common reagents are lime (CaO), magnesium enhanced lime (MgO and CaO), ammonia (NH_3), and sodium carbonate (Na_2CO_3). The first part of this chapter concentrates on limestone-based wet scrubbing systems with forced oxidation to produce gypsum. The next major section focuses on lime-based semi-dry or dry systems.

Reagent regeneration A number of the wet processes are also classified as either non-regenerable or regenerable systems. In non-regenerable systems, the reagent in the scrubber is consumed to directly generate a byproduct containing the sulfur, such as gypsum. In regenerable systems, the spent reagent is regenerated in a separate step to renew the reagent material for further use and to produce a separate byproduct, such as elemental sulfur. The dominant limestone and lime reagent systems used today are non-regenerable. In many cases the regenerable systems have been retrofitted with non-regenerable limestone or lime reagent systems to reduce costs and improve unit availability.

Absorber design

Evolution The first WFGD scrubbers installed in the U.S. were combined particulate collectors and SO_2 absorbers. However, the energy requirements for the venturi scrubbers used for particulate collection proved to be excessive. High efficiency dust collectors, usually electrostatic precipitators (ESPs), replaced venturi scrubbers, and separate, much lower pressure drop absorber towers were used to absorb the SO_2.

The WFGD systems installed in the 1970s and early 1980s were typically sized with multiple modules with spare spray headers and sometimes spare scrubber (absorber) modules. Prior to about 1978, most of the lime and limestone WFGD systems were not designed

with forced oxidation. These absorber internals were subject to severe scaling and plugging of internal components, such as spray nozzles. With the development of forced oxidation, the scaling has effectively been eliminated. This, along with improvements in accessory equipment such as pumps, has eliminated the need for spare scrubber modules. Most wet scrubbers include a spare recirculation pump and spray header. However, the pump and header reliability are such that even spare pumps are not required in some designs.

Since the mid-1990s the use of a single absorber module in WFGD systems has become an accepted design. Single absorber modules have been supplied to handle 1300 MW boilers. Some designs use a single absorber to treat the flue gas from multiple boilers. The availability of most of these systems has been approaching 100%. In most cases, forced outages or load reductions have been due to problems with support systems such as reagent preparation, dewatering and gypsum handling rather than the absorber module.

General arrangement and design Fig. 1 provides a side view of a 660 MW modern coal-fired power system showing the location of the SO_2 scrubber absorber module. Absorber modules are located on either side of the stack though other arrangements are possible. The dust collector, in this case an ESP, is placed upstream of the wet scrubber. In this arrangement, the induced draft (ID) fan is between the dust collector and the scrubber, permitting the fan to operate in a particulate-free, dry flue gas. In cases of retrofit to existing units, it is sometimes necessary to add a booster fan between the ID fan and the scrubber. In some systems, the booster fan is installed after the wet scrubber in what is referred to as the wet position. This is not the preferred method as wet fans require more maintenance and are more expensive.

The most common WFGD absorber module is the spray tower design shown in Fig. 2. The flue gas enters the side of the spray tower at approximately its midpoint and exits through a transition at the top. The upper portion of the module (absorption zone) provides for the scrubbing of the flue gas to remove the SO_2 while the lower portion of the module serves as an integral slurry reaction tank (also frequently referred to as the recirculation tank and oxidation zone) to complete the chemical reactions to produce gypsum. The self-supporting absorber towers typically range in diameter from 20 to 80 ft (6 to 24 m) and can reach 150 ft (46 m) in height. In some designs, the lower reaction tank is flared downward to provide a larger di-

Table 1
U.S. SO₂ Emissions and Coal-Fired Power Generation

	Total U.S. SO_2 10^6 t/yr	Utility SO_2 10^6 t/yr	Coal Fired Utility Generation 10^{12} kWh
1970	31	17	0.7
1980	26	17	1.2
1990	23	16	1.6
2000	16	11	2.0

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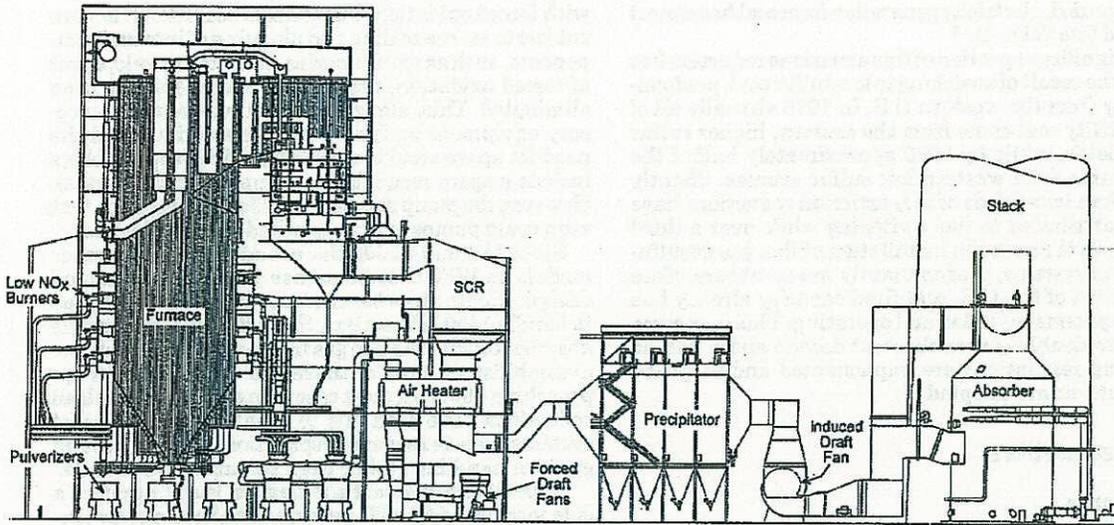BS-03

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Fig. 1 Typical emission control components for a coal-fired utility boiler

ameter tank for larger slurry inventory and longer retention time. Other key components shown include the slurry recirculation pumps, interspatial spray headers and nozzles for slurry injection, moisture separators to minimize moisture carryover, oxidizing air injection system, slurry reaction tank agitators to prevent settling, and the perforated tray to enhance SO$_2$ removal performance. The following sections discuss the gas flow path, slurry flow path, and the materials used in the tower shell design to accommodate the highly corrosive environment.

Gas flow path As noted above, the flue gas enters the absorber at about the midpoint and turns and flows up through an absorption tray, a spray zone, and moisture separators before exiting the absorber and flowing to the stack. Because the flue gas enters the absorber from the side, gas flow non-uniformity in the tower is a potential issue. This non-uniformity reduces overall SO$_2$ removal performance and aggravates moisture separator carryover. The absorber design depicted in Fig. 2 incorporates a perforated plate tray that reduces flue gas flow maldistribution. The pressure drop across the tray is usually between 1 and 3 in. wg (0.2 and 0.7 kPa). The tray provides intimate gas/liquid contacting and increases the slurry residence time in the absorption zone. Some absorbers have two trays providing multiple contact zones for SO$_2$ removal. Absorber modules that do not use a tray are referred to as open spray towers. Flue gas enters the scrubber module at a temperature of 250 to 350F (121 to 177C) and is evaporatively cooled to its adiabatic saturation temperature by a slurry cascading from the absorber tray. A wet-dry interface exists at the scrubber inlet where the inlet fluework at the gas temperature of approximately 300F (149C) merges with the scrubber shell which is at saturation temperature of around 125F (52C). Reagent slurry droplets impinging and drying out on the relatively hot sur-

faces can create growing deposits which can affect scrubber operation. Deposits are minimized by a combination of features that effectively keep the slurry away from the wet-dry interface region.

The fluework from the exit of the wet scrubber to the stack is an important facet of the system design. The potential for severe corrosion and deposition in these flues is well documented. This potential for severe corrosion arises from many factors. The flue gas leaves the moisture separator saturated with water vapor. Some carryover of slurry droplets smaller than 20 microns is inevitable. These droplets will usually be slightly acidic and may contain high concentrations of dissolved chlorides. The flue gases will contain some residual SO$_2$ and ample oxygen to oxidize the SO$_2$ to SO$_3$. Because the flue gas is saturated with water vapor, surface condensation is inevitable. This condensate can become severely acidic (pH less than 1) leading to the formation and accumulation of acidic deposits of sulfuric acid and calcium salts on the walls and floor of the flue.

Two approaches are used to minimize these effects: flue gas reheat and flue/stack lining. The former option involves reheating the flue gas so that no droplets remain. Reheating the flue gas that is leaving the scrubber has been accomplished by various means:
1. steam coil heaters,
2. mixing with some hot flue gas which is bypassed around the scrubber,
3. mixing with hot air, and
4. regenerative heat exchangers that transfer heat from the hot flue gas inlet to the cooler flue gas outlet.

Several problems are associated with each of the reheat methods, primarily corrosion and deposition. As a result, operation without flue gas reheat, i.e., with a wet stack, has become popular in the U.S. Under these conditions, the fluework from the scrubber to the

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stack is made of high alloy materials or lined with corrosion resistant materials, and the stack is lined with acid resistant linings or is made of alloys. Moisture collection devices are installed in the outlet flues and stack, and drains are provided to capture the moisture carryover and condensate collected in these devices. The use of regenerative gas-gas heaters is prevalent internationally because of restrictions on stack plumes.

Slurry flow path The bottom of the absorber is an integral reaction or recirculation tank. Slurry-containing reaction products and unreacted fresh reagent are pumped from the recirculation tank to the spray headers and fall through the spray zone, tray and gas inlet zone to the tank. The slurry is continuously recirculated in this loop. Fresh reagent such as limestone

is added to replenish the alkalinity required to remove SO_2. Reaction products are pumped from the absorber to the dewatering system. Spray nozzles are used in wet scrubbers to atomize the slurry into fine droplets and provide contact surface area for the slurry and flue gas. The operating pressures typically vary between about 5 and 20 psi (34 and 138 kPa). Spray nozzles without internal obstructions are favored to minimize plugging by tramp debris. The arrangement of the spray headers and nozzles is designed to ensure complete coverage of the absorber cross-section and prevent gas bypassing through areas of low slurry spray flux, which reduces overall SO_2 removal efficiency.

A portion of the slurry droplets is entrained by the gas and carried up to the moisture separators. The
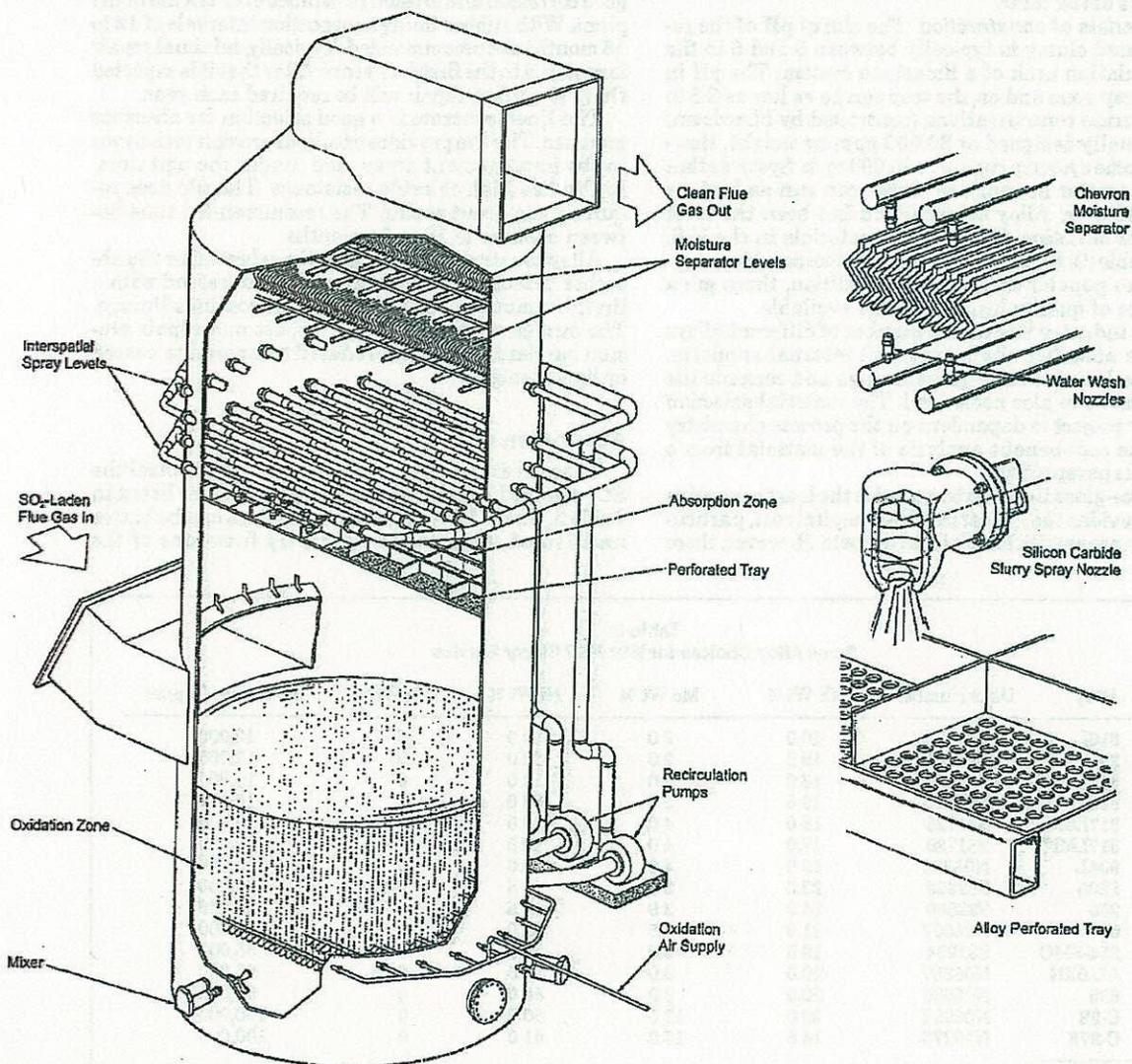

Fig. 2 Wet flue gas desulfurization absorber module.

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moisture separators collect and coalesce these slurry droplets so that they will drain back down into the absorber. Otherwise, slurry droplet carryover would result in the buildup of slurry in outlet flues, fallout in the area around the power plant, and excessive particulate emissions. The moisture separators in most wet scrubbers are of the *chevron* design (see Fig. 2). Chevrons are closely spaced corrugated plates that collect slurry droplets by impaction. They efficiently collect droplets larger than about 20 microns in diameter.

The recirculation tank size is set: 1) to meet the requirements of the recirculation pump suction head, 2) to allow time for the chemical reactions of sulfite oxidation and limestone dissolution, and 3) to provide time and surface area for gypsum crystal growth development. Some of these rate processes are described in more detail later.

Materials of construction The slurry pH of the recirculated slurry is typically between 5 and 6 in the recirculation tank of a limestone system. The pH in the spray zone and on the tray can be as low as 3.5 to 4. Chloride concentrations (controlled by blowdown) are usually designed at 20,000 ppm by weight. However, some systems run up to 50,000 ppm. Systems that use seawater as makeup water can run as high as 100,000 ppm. Alloy construction has been the most popular selection for absorber materials in the U.S. (see Table 2). Rubber-lined, carbon steel scrubber modules are popular in Europe. In addition, there are a number of quality lining systems available.

The industry has used a number of different alloys for the absorber shell, tray and internal supports. Rubber linings, flake-glass linings and ceramic tile systems have also been used. The material selection on any project is dependent on the process chemistry and the cost-benefit analysis of the material from a lifecycle perspective.

Flake-glass lined carbon steel is the least expensive and provides the lowest installed capital cost, particularly in areas with low field labor costs. However, there

are risks to using flake-glass linings and they require the most maintenance. It is expected that these linings would require annual inspections and repair of about 3% of the surface area. This is, therefore, not a good selection for absorbers requiring two to three years between outages. Flake-glass or epoxy resin lining systems are frequently suggested for slurry tanks, process trenches and sumps. The lining in tanks and sumps is less subject to problems because there is less erosion, no internals, a limited number of connections, and no direct contact with the flue gas.

Rubber linings have provided a range of operating issues in the U.S. because of the many failures in the early use of scrubbers. The failures were due to incorrect rubber selection and poor application. The technology has advanced and rubber lining can provide good corrosion and erosion resistance over the life of the plant. With rubber linings, inspection intervals of 12 to 18 months are recommended. Typically, minimal repair is required in the first four years. After that it is expected that 3% surface repair will be required each year.

Tile-lined concrete is a good selection for absorber modules. The tile provides excellent erosion resistance in the impingement areas and under the agitators. It also has high chloride resistance. The tile does require some grout repair. The recommended time between outages is 18 to 24 months.

Alloy construction is frequently selected for the absorber vessel. Areas of wear can be addressed with a limited amount of erosion resistant coatings/linings. The outage time required to inspect and repair erosion barriers (linings) is reduced compared to coated or lined designs.

Absorption of SO₂

There are a number of parameters that control the SO_2 removal capability of a wet absorber as listed in Table 3. The effects of these parameters can be better understood when the two primary functions of the

Table 2
Some Alloy Choices for Wet FGD Slurry Service

Alloy	UNS Number	Cr Wt %	Mo Wt %	Ni Wt %	N Wt %	Maximum Cl ppm
316L	S31603	16.0	2.0	10.0	0	10,000
20	N08020	19.0	2.0	32.0	0	12,000
317L	S31703	18.0	3.0	11.0	0	15,000
825	N08825	19.5	2.5	38.0	0	15,000
317LM	S31725	18.0	4.0	13.5	0	18,000
317LMN	S31726	17.0	4.0	13.5	0.1	20,000
904L	N08904	19.0	4.0	23.0	0	20,000
2205	S32205	22.0	3.0	4.5	0.14	30,000
255	S32550	24.0	2.9	4.5	0.1	45,000
G	N06007	21.0	5.5	36.0	0	50,000
254-SMO	S31254	19.5	6.0	17.5	0.18	55,000
AL-6XN	N08367	20.0	6.0	23.5	0.18	55,000
625	N06625	20.0	8.0	58.0	0	55,000
C-22	N06022	20.0	12.5	50.0	0	100,000
C-276	N10276	14.5	15.0	51.0	0	100,000

Note: Compositions are ASTM minimums.

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absorber, physical and chemical, are considered. The physical parameters set the gas/slurry contacting surface created in the absorber and the chemical factors control the rate that absorbed SO_2 is removed from the liquid phase of the slurry.

The SO_2 removal efficiency of a scrubber is determined by the amount of SO_2 that can be absorbed by a unit volume of slurry. Limestone based systems are reaction-rate limited. That is, the SO_2 absorption is controlled by how fast the absorbed SO_2 can be reacted with limestone in the scrubber. The system must provide sufficient alkalinity to react with the SO_2. The alkalinity can be present as Ca and CO_3 ions or as solid $CaCO_3$. There are three primary methods of increasing the alkalinity in the absorption zone.

The first method is to increase the liquid to gas ratio (L/G) in the absorber. By increasing the flow of slurry, more liquid contact is provided. The increased flow does not change the effective surface in terms of area per unit volume of slurry. It does provide more total surface area and alkalinity to the absorption zone. This does not increase the absorption per unit volume of slurry. However, it increases the total volume of slurry in the spray zone.

The second method is to increase the total and dissolved alkalinity per unit volume of slurry. This is accomplished by increasing the limestone to SO_2 ratio in the slurry and/or by increasing the absorber reaction tank solids residence time. Increasing the stoichiometry increases both the dissolved alkalinity and the solid phase alkalinity. Increasing the tank residence time increases the dissolved alkalinity because there is more time for limestone to dissolve.

The third method is to use one or more trays. A tray is a much more efficient contact device than a slurry spray. This is because a tray creates more surface area between the slurry and the gas. Therefore, the tray generates more surface area per unit volume of slurry. The tray also provides significant holdup time for the slurry. This increases the limestone dissolution in the absorption zone and increases the absorption per unit volume. The limestone dissolution on the tray can be as much as 50% of the dissolution in the entire absorber.

The benefits of the absorber tray compared to an open spray tower include:

1. reduced liquid to gas ratios
2. increased absorption for the same L/G
3. more uniform gas distribution
4. fewer recirculation pumps
5. fewer spray headers
6. reduced pump maintenance
7. acts as a maintenance platform

There are absorber designs based on reagents such as sodium carbonate which are dissolved chemistry absorbers. In these designs, there is sufficient liquid phase alkalinity available so that the SO_2 removal becomes gas phase limited. For these systems, the L/G and tray pressure drop are much lower. Such systems can achieve removal efficiencies as high as 99% at low tray pressure drops and an L/G of about 40 gpm/1000 acfm. Magnesium enhanced lime systems are solid chemistry systems that run like liquid systems because

the magnesium will be a dissolved species. Organic acid can be added to limestone systems to achieve the removal of liquid based (dissolved species) systems.

Table 3
Effect of Various Design Parameters on SO_2 Removal

Parameter	Type	When Parameter Increases
Inlet SO_2	Chemical	Removal decreases
L/G	Physical	Removal increases
Tray ΔP	Physical	Removal increases
Stoichiometry	Chemical	Removal increases
pH	Chemical	Removal increases
Nozzle pressure	Physical	Removal increases
Cl-concentration	Chemical	Removal decreases

Limestone forced oxidized (LSFO) process description

Limestone based wet FGD systems are classified as non-regenerable. This means that the reagent is consumed by the process and must therefore be continually replenished. A generic diagram of non-regenerative processes is shown in Fig. 3. Each consists of four process steps: reagent preparation, SO_2 absorption, slurry dewatering, and final disposal. Within each of these process steps many variations exist. Essentially all wet FGD installations have some unique aspects. The following discussion will consider the limestone forced oxidized system, which is the most common system in use today. The LSFO system produces a gypsum byproduct that is typically sold for use in the manufacture of wallboard. It can also be sold to the cement industry, used as fertilizer, or sent to a landfill.

Reagent preparation

The primary advantages of limestone are its wide availability and cost effectiveness; roughly 5 to 6% of the earth's crust consists of calcium and magnesium carbonates and silicates. Limestone, consisting mostly of calcium carbonate ($CaCO_3$), is easily mined, transported and stored. Its storage and conveying at the plant site are similar to coal handling. A typical analysis of a limestone suitable for use in a wet scrubber is listed in Table 4.

Limestone can be ground dry in an air-swept mill, or wet in an overflow ball mill and can be ground onsite at the power plant, or offsite at a separate location. In North America, most limestone wet FGD systems feature onsite wet grinding for slurry preparation. In most cases, the system of choice is a closed loop ball mill circuit. A typical ball mill circuit is shown in Fig. 4. The energy required to achieve a given grind size is estimated by the Bond relationship:

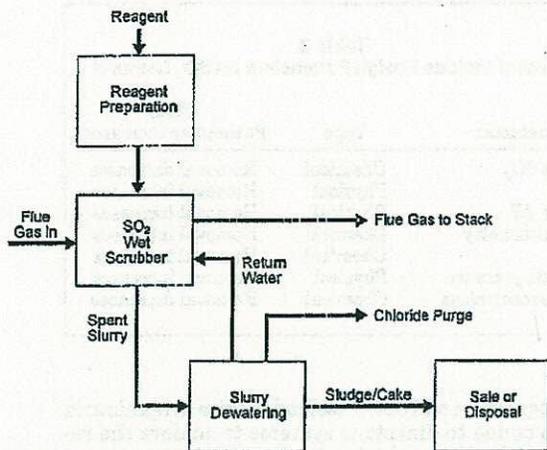$$W = \frac{10W_i}{\sqrt{D_P}} - \frac{10W_i}{\sqrt{D_F}} \tag{5}$$

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Fig. 3 Wet FGD system diagram

Table 4
Typical Limestone Composition

Component	%
CaCO₃	93 – 97
Available CaCO₃	92 – 96
MgCO₃	0.5 – 2
Total inerts	3 – 7
SiO₂	0.5 – 3
Fe₂O₃	0 5 – 3
R₂O₃ (other metal oxides)	0 5 – 4

where

W = power, kWh/t of product
W_i = Bond work index, kWh (micron)$^{1/2}$/t
D_P = diameter in microns for which 80% of product is finer
D_F = diameter in microns for which 80% of feed is finer

The Bond work index for calcitic limestones ranges from about 8 to 15 kWh (micron)$^{1/2}$/t; 10 to 12 is typical.

For onsite grinding systems following Fig. 4, the limestone is received with a maximum diameter of about 0.75 in. (19 mm) or less and is fed through a weigh belt feeder to the ball mill. Either fresh or recycled water is added at the ball mill feed chute in proportion to the feed rate of the limestone. The output from the mill overflows to the mill product tank where it is pumped to a set of hydroclones to separate the fines from the coarse fraction. The coarse fraction (underflow) is returned to the ball mill for further grinding while the fine material is sent to the limestone feed tank. The water balance is maintained to provide 25 to 35% suspended limestone solids in the feed tank.

The limestone grind is usually expressed as a percent passing a certain sieve size. The typical grind is 90 to 95% passing 325 mesh (44 microns). Fine grinding requires a larger ball mill system and higher operating power consumption. Finer material provides greater limestone utilization in a smaller reaction tank and higher dissolution rate due to the higher surface area of finer limestone particles. Stoichiometry, or reagent ratio, is defined as the molar ratio of the reactant, CaCO₃ for limestone systems, to the SO₂ removed.

The limestone grinding mill shown in Fig. 4 is a horizontal ball mill. The mill consists of a cylindrical shell containing steel balls ranging in size from 1 to 4 in. (25 to 102 mm). The cylinder is rotated and the limestone is ground by the tumbling of the balls and limestone in the mill. Other types of wet mills include vertical tower mills and attrition mills.

For systems making marketable gypsum, limestone purity is a primary factor (see Table 4). The concentration of inerts affects the gypsum purity, defined as the percent pure gypsum (CaSO₄ · 2H₂O). In addition there are limits to the amount of SiO₂, Fe₂O₃ and total metal oxides (R₂O₃) in the gypsum.

Limestone can also be dry ground at a central location and transported by truck to the FGD system. At the site, the pulverized limestone is pneumatically conveyed to a storage bunker. It is then fed to water-filled slurry preparation tanks. This system requires less space than onsite grinding facilities and reduces some of the FGD operating burden. However, offsite grinding is typically more expensive than onsite preparation.

Fig. 4 Limestone reagent preparation system

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Slurry dewatering and disposal

The disposal of reaction products from limestone wet scrubbers includes ponding, landfilling, and the production of high grade gypsum for wallboard, cement and fertilizer. In the production of high grade gypsum, the spent slurry consisting of the reaction products, inerts from the limestone, excess reagent, and flyash is dewatered in two stages (primary and secondary dewatering). In addition, a purge stream is used to remove chlorides and fine particles from the WFGD system. A typical system is shown in Fig. 5.

Hydroclones are used for primary dewatering of gypsum. Multiple hydroclones are mounted in a cluster and fed by slurry pumped from the absorber reaction tank to a radial distributor and then to the individual hydroclones. The hydroclone classifies the solids in the feed slurry by particle size. A dilute slurry of fine particles leaves through the *hydroclone overflow* and a concentrated slurry of coarse particles is discharged from the hydroclone underflow. Typically a feed slurry containing 15 to 20% suspended solids will produce an overflow stream containing 3 to 4% suspended solids and an underflow stream containing 50 to 55% suspended solids from the primary cyclone. Most of the hydroclone overflow is returned to the absorber. Recycling of this finer fraction of the solids allows for crystal growth of the gypsum and better utilization of unreacted limestone. A portion of the overflow is sent to a purge system for removal of fine particles and chlorides.
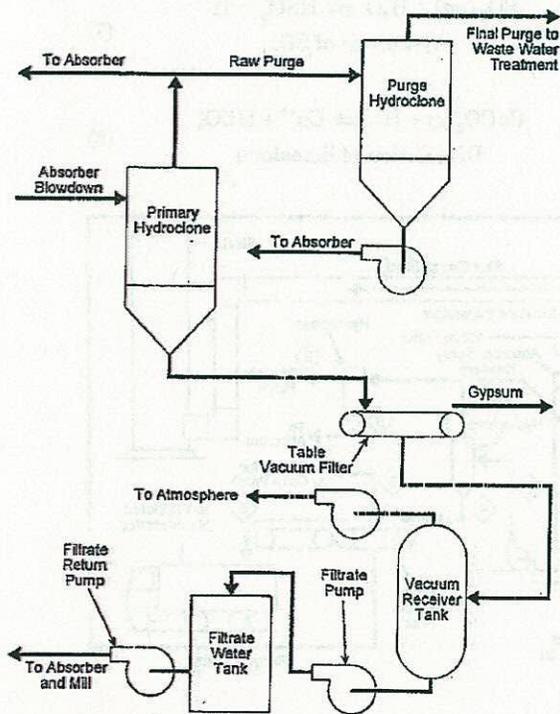
Secondary dewatering of the gypsum is typically accomplished using either horizontal belt or drum vacuum filters. Horizontal belt filters are typically used for wallboard grade gypsum production because the final cake must contain no more than 10% moisture. Drum filters are prevalent when higher moisture fractions are acceptable (20%), such as for the cement industry and fertilizer. However, with recent design improvements, some drum filters can achieve 10% cake moisture. Centrifuges have also been used in secondary dewatering.

The production of wallboard grade gypsum requires that the slurry on the vacuum filter be washed with fresh water to remove total dissolved solids (TDS) and specific ions such as chloride, Cl⁻. The actual specifications will vary based on the wallboard manufacturer's requirements.

Purge system

The hydroclone overflow contains a higher percentage of fine particles of inerts and flyash than the slurry fed to the hydroclones. This stream is the source for the purge stream. The purge stream flow rate can be set by the amount needed to purge dissolved solids from the system or by the need to remove impurities and fines. Typically, the raw purge stream is sent to a second set of (purge) hydroclones to further concentrate fines in the overflow and reduce the final purge stream total suspended solids (TSS) to about 1.5%. The underflow is returned to the absorber. Concentrating the fines in the overflow helps reduce the amount of pure gypsum in the final purge stream.

The purge stream flow is usually set to control chloride concentration, typically to a design concentration of 20,000 ppm or less.

The final purge stream is typically ponded or sent to a waste water treatment system for removal of other chemical species and suspended solids. The treated water can then be returned to a river or other body of water. The solids removed in the treatment system are concentrated or sent to a landfill.

WFGD system water balance

The wet FGD system loses water to evaporation when the incoming gas is quenched. There is also some loss to the gypsum cake. In addition, there is a loss of water in the purge stream. Within the system, water is used for preparing the limestone slurry, washing the moisture separators, cake wash, and as seal water for the vacuum filter seal pumps. Reclaim water (filtrate) from the vacuum filters is collected and used for limestone preparation and level control in the absorber reaction tank.

Cake wash water and vacuum pump seal water must be fresh water from a well, river or lake. For all other uses including moisture separator wash water, the requirements are less stringent. Sources for the remaining water can be cooling tower blowdown, secondary sewage treatment plant effluent, and seawater, among others.

Process flow diagram and mass balance

A typical limestone based wet scrubbing process using in-situ forced oxidation is shown in Fig. 6. In this example, the reagent preparation system includes a

Fig. 5 Slurry dewatering system.

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closed circuit ball mill system using water that is recycled from the slurry dewatering system. The feed slurry is pumped to the absorber reaction tank as required to control the pH in the tank or to achieve the desired SO_2 removal efficiency or emission concentration. The use of pH control can result in excessive limestone consumption at low boiler loads or when firing lower than design sulfur coals without operator intervention. Excess limestone can adversely affect gypsum purity. By contrast, limestone consumption can be minimized by controlling to a removal efficiency or SO_2 emission set point.

Air is also pumped to this reaction tank and distributed by a sparge grid or agitator. Oxidation air can also be introduced into the tank by air lances and dispersed with specially designed agitators. The oxygen in the air reacts with sulfite present in the slurry to produce gypsum ($CaSO_4 \cdot 2H_2O$).

The slurry is pumped from the reaction tank to the spray headers shown in Fig. 2. The slurry is sprayed countercurrently into the flue gas where it absorbs the SO_2. From there, the slurry falls to the sieve tray where additional SO_2 is absorbed into the froth created by the interaction of the flue gas and slurry on the tray. The slurry then drains to the reaction tank.

The reaction products, primarily gypsum, are continuously withdrawn from the wet scrubber. This spent slurry is pumped via gypsum blowdown pumps to the dewatering system. The spent slurry typically contains about 15 to 20% suspended solids. A hydroclone is used to concentrate the slurry. The underflow from the hydroclone is concentrated to 50% solids. The overflow containing 4% suspended solids is sent back to the absorber and to the purge stream.

The underflow from the hydroclone is directed to a vacuum filter where the filtered solids are washed with fresh water and dewatered to form a filter cake containing about 10% free moisture. The cake is then sent by truck to a wallboard manufacturer. A mass balance for this example is presented in Table 5.

A summary of the major power requirements for this limestone to gypsum process is shown in Table 6. The operating power requirement for the example displayed in this table is 1.9% of gross unit output. This percentage will vary with coal sulfur content and the required removal efficiency. Higher overall sulfur removal or high sulfur loading requires higher L/G, more pressure drop, more oxidation air, etc.

Wet scrubber chemistry

SO_2 absorption in a wet scrubber and its subsequent reaction with alkaline earth materials such as limestone is an elementary acid-base reaction which takes place in an aqueous environment. However, the chemical processes involved are complex. SO_2 is a relatively insoluble gas in water. Calcium carbonate ($CaCO_3$) also has a low solubility in water. The principal reaction products are calcium sulfite hemi-hydrate ($CaSO_3 \cdot \frac{1}{2} H_2O$) and calcium sulfate dihydrate ($CaSO_4 \cdot 2H_2O$), or gypsum. These two salts also have low solubility in water.

In a limestone system with forced oxidation, the following reaction model can be used to describe the process using the chemical species in Table 7.

In the gas/liquid contact zone:

$$SO_2(g) \rightleftharpoons SO_2(aq) \tag{6}$$
Dissolving gaseous SO_2

$$SO_2(aq) + H_2O \rightleftharpoons HSO_3^- + H^+ \tag{7}$$
Hydrolysis of SO_2

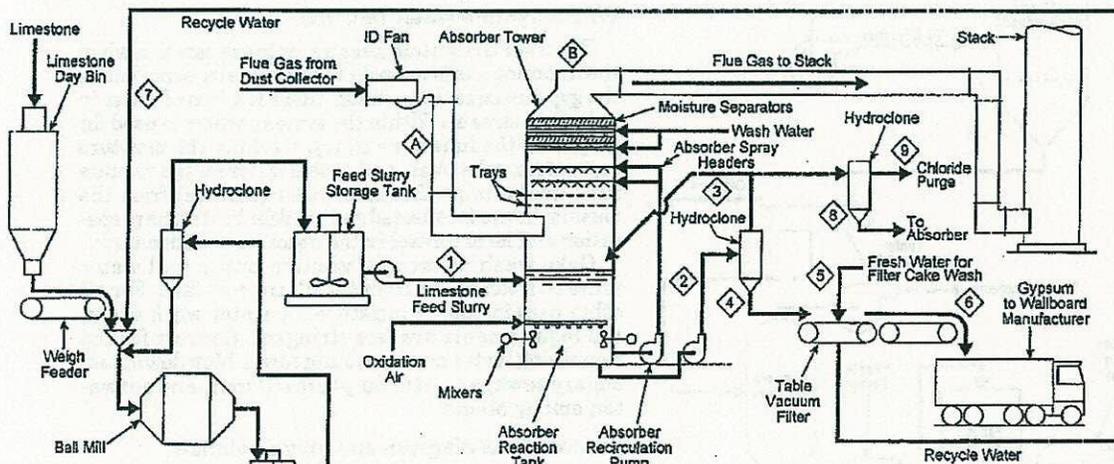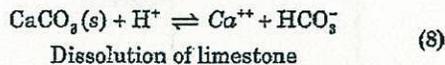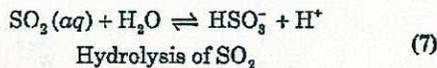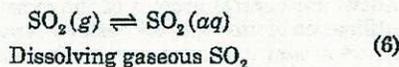$$CaCO_3(s) + H^+ \rightleftharpoons Ca^{++} + HCO_3^- \tag{8}$$
Dissolution of limestone

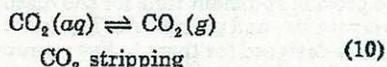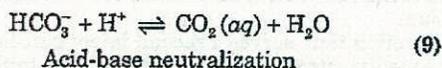Fig. 6 Wet scrubber FGD system flow diagram (see Table 5 for mass balance)

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Table 5
Mass Balance for the Limestone Forced Oxidation System Shown in Fig. 6

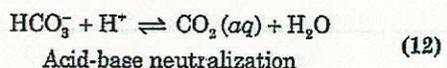| Gas side	Inlet	Exit	
Stream designation	A	B	
Flow rate, ACFM	1,710,000	1,462,000	
Total mass flow rate, lb/h	5,521,000	5,924,175	500 MW boiler burning coal with 4.1% sulfur
Mass flow rate of H_2O, lb/h	240,000	550,000	Limestone stoichiometry = 1.03
Mass flow rate of SO_2, lb/h	34,900	698	SO_2 efficiency = 98%
Mass flow rate of HCl, lb/h	458	5	
Static pressure, in. wg	12	1	
Temperature, F	300	129	

Liquid side	Feed Slurry	To Hydroclone	Hydroclone Overflow	Hydroclone Underflow	Filter Wash Water	Filter Cake	Recycle Water	Hydroclone Underflow	Chloride Purge
Stream designation	1	2	3	4	5	6	7	8	9
Flow rate, GPM	330	961	648	271	53	22	71	11	41
Total flow rate, lb/h	204,225	573,964	354,000	197,435	26,500	107,493	390,338	6,296	20,500
Flow rate of sus. solids, lb/h	61,268	114,793	15,000	98,718	0	96,743	0	944	0
% suspended solids	30	20	4.3	50	0	90	0	15	0
Chloride conc., ppm in liquid	8,000	20,000	20,000	20,000	50	1,000	8,000	20,000	20,000
pH	7.8	5.5	5.5	5.5	7	7	6	5.5	5.5
Temperature, F	100	129	129	129	120	120	120	129	129

$$HCO_3^- + H^+ \rightleftharpoons CO_2(aq) + H_2O \qquad (9)$$
Acid-base neutralization

$$CO_2(aq) \rightleftharpoons CO_2(g) \qquad (10)$$
CO_2 stripping

In the reaction tank:

$$CaCO_3(s) + H^+ \rightleftharpoons Ca^{++} + HCO_3^- \qquad (11)$$
Dissolution of limestone

$$HCO_3^- + H^+ \rightleftharpoons CO_2(aq) + H_2O \qquad (12)$$
Acid-base neutralization

$$CO_2(aq) \rightleftharpoons CO_2(g) \qquad (13)$$
CO_2 stripping

$$O_2(g) + 2HSO_3^- \rightarrow 2SO_4^= + 2H^+ \qquad (14)$$
Sulfite oxidation

$$Ca^{++} + SO_4^= + 2H_2O \rightleftharpoons CaSO_4 \cdot 2H_2O \qquad (15)$$
Precipitation of gypsum
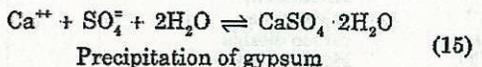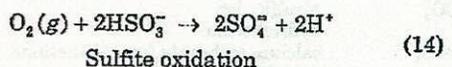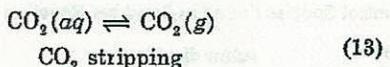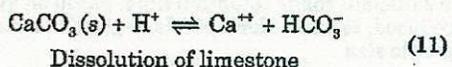
Reaction 6 expresses the mass transfer rate of SO_2 from the gas phase to the liquid or aqueous phase. Its mass transfer rate can be expressed by:

$$\frac{d(Gy)}{dV} = k_g a (y - y^*) \qquad (16a)$$

or

$$N_g = \int \frac{dy}{y - y^*} = \int \frac{k_g a}{G} dV \qquad (16b)$$

where

G = molar gas flow rate, moles/s
y = mole fraction of SO_2 in flue gas
k_g = gas film mass transfer coefficient, moles/m² s
a = interfacial surface area, m²/m³
y^* = equilibrium SO_2 concentration at the gas/liquid interface
V = volume of the gas/liquid regime, m³
N_g = number of gas phase transfer units, dimensionless

Although k_g can be approximated, the interfacial surface area can not. The gas phase mass transfer rate must be determined experimentally. This involves operating the scrubber under conditions in which $y^* \rightarrow 0$. Equation 16 can then be integrated to:

$$N_g = -\ln(1-E) = k_g a V/G \qquad (17)$$

where

N_g = overall number of gas phase transfer units, dimensionless
E = overall SO_2 fractional efficiency

Many factors determine the number of gas phase transfer units (N_g). These include the slurry spray rate, the droplet size and spatial distributions, the gas phase residence time (height of spray zone), the liquid residence time, wall effects, and the gas flow distribution.

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In a limestone based wet scrubber, the rate-limiting reactions in the gas/liquid contact zone are believed to be Reaction 8. The reaction rate for limestone dissolution can be expressed by:

$$\frac{d[CaCO_3]}{dt} = k_c \left([H^+] - [H^+]_{eq} \right) Sp_c [CaCO_3] \quad (18)$$

where

$[CaCO_3]$ = calcium carbonate concentration in the slurry, moles/l
k_c = reaction rate constant
$[H^+]$ = hydrogen ion concentration, moles/l
$[H^+]_{eq}$ = equilibrium (H^+) at the limestone surface, moles/l
Sp_c = specific surface area of limestone in slurry

The reaction tank permits Reactions 12 through 16 to approach completion. In a limestone scrubber, limestone is added directly to the reaction tank. The pH of the slurry returning from the gas/liquid contact zone to the reaction tank can be as low as 3.5. The pH in the reaction tank is usually 5.2 to 6.2. Therefore, the overall reaction in the reaction tank is:

$$CaCO_3(s) + H^+ + HSO_3^- + \tfrac{1}{2} O_2 + H_2O \rightarrow \\ CaSO_4 \cdot 2H_2O(s) + CO_2(g) \quad (19)$$
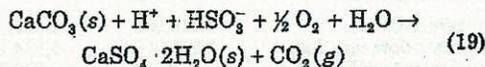
The rate of gypsum crystallization in the reaction tank can be expressed by:

$$\frac{d[CaSO_4 \cdot 2H_2O]}{dt} \\ = k(R-1)Sp_g[CaSO_4 \cdot 2H_2O] \quad (20)$$

where

k = crystallization rate constant
R = $A_{Ca^{++}} A_{SO_4^=} / K_{Sp}$
$A_{Ca^{++}}$ = activity of Ca^{++} ion
$A_{SO_4^=}$ = activity of $SO_4^=$ ion
K_{sp} = solubility product of gypsum
Sp_g = specific surface area of gypsum

R is a measure of the level of supersaturation. If R is greater than 1, the solution is supersaturated with gypsum. If R is less than 1, the solution is subsaturated in gypsum.

The reaction tank serves a second important function in lime/limestone scrubber systems. The tank is sized to provide sufficient time for the dissolved gypsum to crystallize and precipitate. Typically, the reaction tank is designed for three to five minutes of residence time based upon the recirculation rate. An additional consideration in sizing the reaction tank is to provide sufficient solids retention time, based on gypsum produced, to allow the crystals to grow to a suitable particle size.

Table 6
Typical Power Requirements for Limestone Scrubber with Forced Oxidation

Absorber System	Avg. Power (kW)
Oxidation air blower	1300
Absorber recirc. pumps	3250
Absorber recirc. tank agitators	250
Moisture separator wash water pump	50
Misc. pumps and agitators	50
Subtotal	4900

Dewatering Area	Avg. Power (kW)
Vacuum pump for filter	330
Reclaim water pump	50
Filter drive	15
Misc. pumps and agitators	30
Subtotal	425

Reagent Preparation	Avg. Power (kW)
Ball mill	1200
Mill product tank pump	30
Limestone feed tank agitator	80
Misc. pumps and agitators	30
Subtotal	1340

Other Systems	Avg. Power (kW)
General-instrument air	80
Booster fan power	2830
Subtotal	2910
Total	9575

Notes:
500 MW boiler burning coal with 4.1% sulfur
Heating value of coal = 10,950 Btu/lb
Absorber L/G = 130 gal/1000 ACF
Total pressure drop = 12 in. wg
Parasitic power = 9575 kW/500 MW
= 1.9%

Table 7
Chemical Species Found in Scrubber Reaction Model

SO_2	sulfur dioxide
H_2O	water
HSO_3^-	bisulfite ion
H^+	hydrogen ion
$CaCO_3$	calcium carbonate (main limestone constituent)
Ca^{++}	calcium ion
CO_2	carbon dioxide
HCO_3^-	bicarbonate ion
$CaSO_3 \cdot 1/2 H_2O$	calcium sulfite hemihydrate
$CaSO_4 \cdot 2 H_2O$	calcium sulfate dihydrate (gypsum)
(l)	denotes liquid phase
(g)	denotes gaseous phase
(aq)	denotes dissolved specie in water
(s)	denotes solid phase

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Performance enhancing additives

The steady-state hydrogen ion concentration of the slurry in the gas/liquid contact zone is determined by the balance between the rate of H^+ generation in Reaction 7 and the rate of H^+ consumption by Reactions 8 and 9. As the hydrogen ion concentration increases, i.e., as the pH drops, the equilibrium SO_2 vapor pressure increases and SO_2 removal efficiency is reduced. This equilibrium relationship can be expressed by:

$$y^* = \frac{C' \times 10^3 \left[H^+\right]}{k_1 \left(k_2 + \left[H^+\right]\right)} \qquad (21)$$
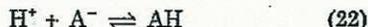
where

k_1 = the equilibrium constant for Reaction 7, moles/l atm

k_2 = the equilibrium constant for Reaction 8, moles/l

C' = total concentration of dissolved SO_2, millimoles/l

= $[SO_2]_{aq} + [HSO_3^-]$

$[H^+]$ = steady-state hydrogen ion concentration

y^* = SO_2 vapor pressure expressed as ppm (assuming barometric pressure = 1 atm)

At 122F (50C):

k_1 = 0.4643 moles/l atm
k_2 = 7.162×10^{-3} moles/l

As the SO_2 vapor pressure rises, the rate of SO_2 absorption diminishes and approaches zero as $y^* \to y$. (See Equation 16.) If a simple means existed to reduce the hydrogen ion concentration in the gas/liquid contact zone, then the SO_2 vapor pressure could be controlled and the SO_2 absorption rate would be maximized. A buffer is a chemical specie which performs this function. A class of weak organic acids has been found suitable for use in limestone scrubbers to control pH and improve overall SO_2 removal. These buffers can be described by:

$$H^+ + A^- \rightleftharpoons AH \qquad (22)$$

where AH is the generalized acid group. When the pH falls in the gas/liquid contact zone, Reaction 22 is driven to the right. Some organic acids which buffer in this range include adipic, formic and succinic acid. Because these are water soluble, the pH buffering is nearly instantaneous compared to the lime or limestone dissolution. Because the SO_2 vapor pressure, i.e., the equilibrium SO_2 concentration of the gas/liquid interface, is proportional to the hydrogen ion concentration, buffers minimize the rise in SO_2 vapor pressure. The buffer concentration required to achieve a given absorption depends upon SO_2 concentration, L/G ratio (the ratio of slurry flow to gas flow), and contact time. Typically, the concentration of these additives which is required to achieve adequate control ranges from 3 to 30 millimoles/l.

A second additive used in wet FGD systems is magnesium oxide, which reacts with SO_2 to form magnesium sulfite. Because magnesium sulfite is highly soluble, the sulfite ion is the primary reactant in the gas/liquid interface as follows:

$$SO_3^= + H^+ \rightleftharpoons HSO_3^- \qquad (23)$$
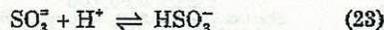
Sodium carbonate can also be added to the lime/limestone system for a similar benefit.

The total concentration of dissolved alkaline species such as $CO_3^=$, HCO_3^-, $SO_3^=$ and OH^- in the slurry is referred to as *dissolved alkalinity*. If the dissolved alkalinity is sufficiently high, the scrubber may become gas-phase diffusion controlled. This is illustrated in Fig. 7.[3] Under these conditions the rate of SO_2 absorption is dependent only upon the amount of interfacial surface area, i.e., the spray droplet surface area plus the tray froth surface area.

Dry scrubbers

Dry scrubbing is more correctly referred to as spray dryer absorption (SDA) to reflect the primary reaction mechanisms involved in the process: drying alkaline reagent slurry atomized into fine drops in the hot flue gas stream and absorption of SO_2 and other acid gases from the gas stream. The process is also called semi-dry scrubbing to distinguish it from injection of a dry solid reagent into the flue gas.

Dry scrubbing is the principal alternative to wet scrubbing for SO_2 control on utility boilers. Since the initial installation in 1980, more than 13,000 MW of dry scrubbers have been installed at U.S. electric utilities. The allowable SO_2 emission permit limits have steadily decreased over time to levels requiring 90 to 95% removal, and the technology has proven to operate reliably at these levels. The application of dry scrubbers to large electric utility boilers is generally limited to those burning low sulfur coals. This is due primarily to the higher unit reagent costs for dry scrubbing. However, for smaller utility and industrial boiler applications, the simplicity and lower capital costs of dry scrubbing make it an attractive alternative for higher sulfur coals. Spray dryer systems are also considered maximum achievable control technology (MACT) for combined HCl and SO_2 control on waste-to-energy units. In the U.S. utility market, dry scrubbers have mainly been applied to units west of the Mississippi River, burning low sulfur fuels, with generating capacities of 90 to 900 MW. However, of the more than 50 current U.S. utility and large coal-fired industrial cogeneration installations, over 40% of the installations are located in the east.

The advantages of dry scrubbing over wet scrubbing include:

1. less costly construction materials (carbon steel),
2. fewer process unit operations,
3. simplicity of control and operation,
4. lower water consumption,
5. lower auxiliary power consumption,
6. use of available alkalinity in the flyash for SO_2 absorption,
7. integral SO_3 emissions control, and
8. production of dry solid byproducts without the need for dewatering.

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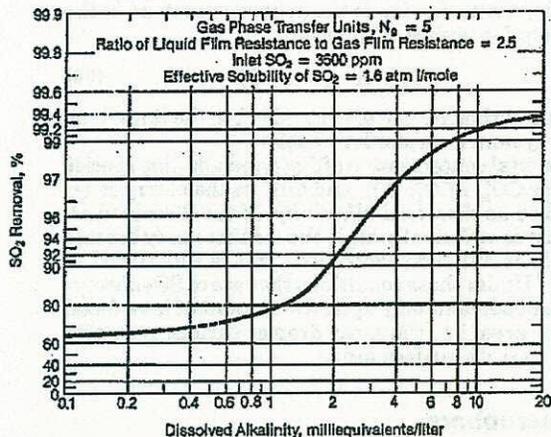BS-03

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Fig. 7 Influence of dissolved alkalinity on SO₂ performance (1 milliequivalent = 50 ppm CaCO₃).

System design

Fig. 8 depicts the equipment orientation for a typical utility SDA installation coupled with a baghouse (BH). Unlike a wet scrubber installation, the SDA is positioned before the dust collector. Flue gases leaving the air heater at a temperature of 250 to 350F (121 to 177C) enter the spray chamber where the reagent slurry is sprayed into the gas stream, cooling the gas to 150 to 170F (66 to 77C). An electrostatic precipitator (ESP) or fabric filter (baghouse) may be used to collect the reagent, flyash and reaction products. Baghouses are the dominant selection for U.S. SDA installations (over 90%) and provide for lower reagent consumption to achieve similar overall system SO₂ emissions reductions. Both reverse-air and pulse-jet baghouse designs are in use (see Chapter 33). The baghouse is an integral part of the process and acts

as a second stage SO₂ removal reactor that can contribute a significant fraction of the overall performance. Operating data from an SDA/baghouse system achieving 90% overall SO₂ emissions reduction has shown that the SDA accounted for 79% of the total and the remaining 21% was obtained across the baghouse. The particulate filter cake on the bags was actually removing 63% of the SO₂ entering the baghouse.

For large utility applications, the flue gases are introduced to the spray chamber through a compound gas disperser (upper and lower) as shown in Fig. 9 for a vertical flow design. For lower gas flow applications, all of the gas flow enters through the roof gas disperser. The gas disperser is designed to assure good mixing of the flue gases with the reagent spray, and to make use of the full volume of the spray chamber. The vertical gas flow configuration dominates utility boiler SDA applications although there are a few horizontal flow units in operation. The SDA chamber is sized to provide sufficient gas-phase residence time for drying of the reagent slurry to produce free-flowing solids in the particulate collector. The required residence time depends on the inlet flue gas temperature, feed slurry solids loading, degree of atomization, and exit gas temperature. In general, for atomization systems producing a Sauter mean drop diameter of 50 to 60 microns, 10 to 12 seconds is sufficient. This residence time is calculated based on the spray chamber volume and the outlet flue gas temperature and pressure. Field studies have shown that the flue gas temperature is within ±10 to 20F (6 to 11C) of the SDA outlet temperature throughout the entire spray chamber.

The reagent slurry is introduced to the reaction chamber as a fine mist of droplets using a single, high-capacity rotary atomizer or multiple dual-fluid (high pressure air/slurry) atomizers. A rotary atomizer assembly for a large utility application is shown in Fig. 10. The assembly shown includes the 900 hp drive motor, a speed-increasing gearbox, the atomizer
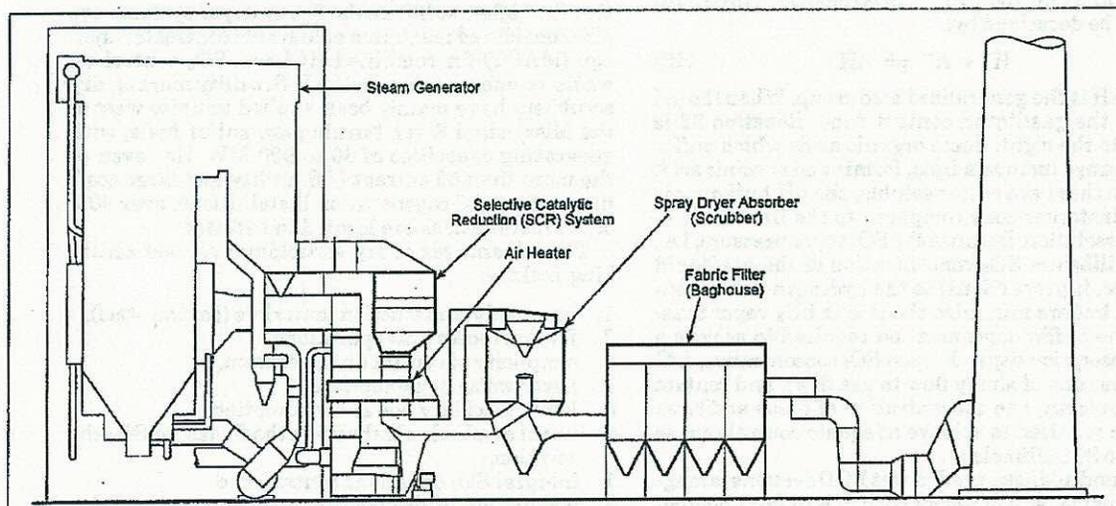

Fig. 8 Utility spray dryer absorption (SDA) system.

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wheel, and associated hardware. Rotary atomization far dominates (90+%) the utility and large industrial installed technology base. This is primarily because the rotary design provides higher slurry flow capacity per atomizer and better performance with higher solids slurries. Both systems have proven reliable and capable of consistently achieving low SO_2 emissions. Key performance requirements for SDA atomizers include:

1. ability to handle variable slurry flow rates,
2. rapid response to flow rate changes,
3. uniform spray coverage of the SDA chamber,
4. low susceptibility to pluggage and build-up, and
5. ease of removal for servicing.

For rotary atomizers, the spray droplet size distribution is a function of the speed of rotation, wheel diameter and wheel design as well as, to a lesser extent, reagent slurry feed properties including the flow rate, viscosity, density, and surface tension.[4] Increasing the wheel speed reduces the droplet size at the cost of higher power consumption. Note that for a rotary unit, atomization quality is not sensitive to the diameter of the exit nozzles and so does not degrade as the nozzles wear from erosion by the reagent slurry. In addition to the slurry properties, key design criteria for dual-fluid nozzle atomizers include the nozzle exit orifice dimensions and orientation, air pressure, and air-to-slurry ratio. Water spray from a typical dual-fluid atomizer with multiple exit orifices is shown in Fig. 11. For this system, minimizing formation of large

(+100 micron) drops from primary atomization or the coalescence of sprays from multiple atomizers is a key criteria for successful operation.

Fig. 12 shows a compact selective catalytic reduction/SDA/BH installation for a 550 MW power plant. At this site, the two SDA chambers, each with a single rotary atomizer, treat approximately 1,800,000 ACFM (850 m³/s) of flue gases from a pulverized coal-fired boiler burning Powder River Basin coal to reduce SO_2 emissions by 94%.

Process description

The quantity of water in the reagent slurry introduced into the SDA is controlled so that it almost completely evaporates in suspension, leaving the solids exiting the spray chamber with a free moisture content of 1 to 2%. SO_2 absorption takes place primarily while the water is evaporating and the flue gas is adiabatically cooled by the spray. The difference between the temperature of flue gas leaving the dry scrubber and the adiabatic saturation temperature is known as the *approach temperature*. Flue gas saturation tem-
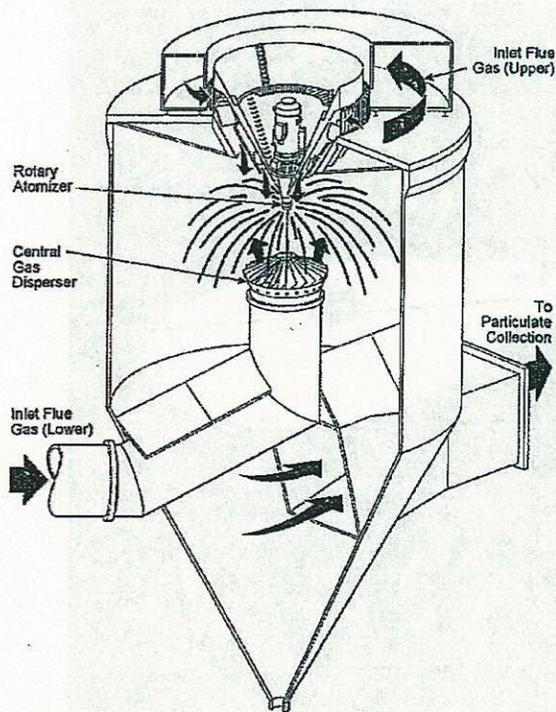

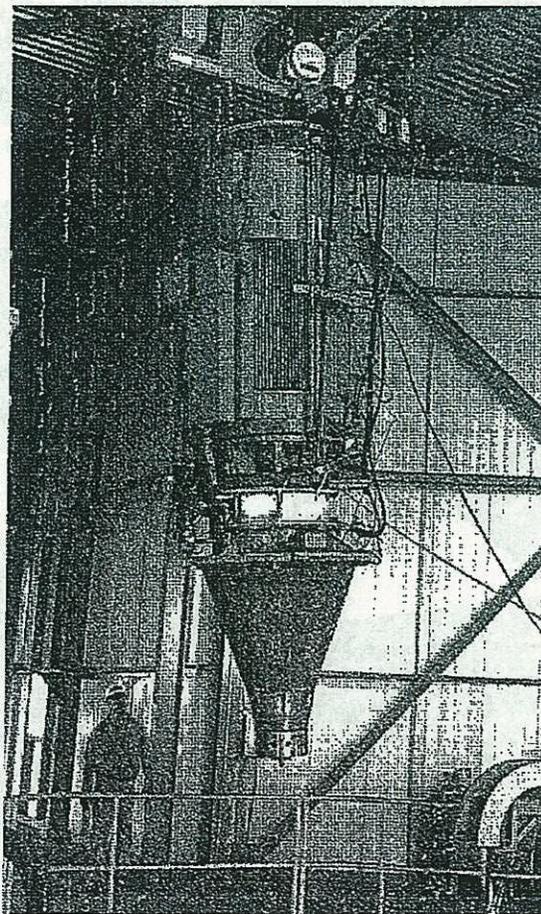Fig. 9 Spray dryer absorber cutaway.

Fig. 10 Rotary atomizer assembly.

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peratures are typically in the range of 115 to 125F (46 to 52C) for low moisture bituminous coals and 125 to 135F (52 to 57C) for high moisture subbituminous coals. The *stoichiometry* is the molar ratio of the reagent consumed to either the inlet SO_2 or the quantity of SO_2 removed in the process. Both definitions are in common use. Reagent stoichiometry and approach temperature are the two primary variables that control the scrubber's SO_2 removal efficiency. The optimal conditions for SO_2 absorption must be balanced with practical drying considerations. The approach temperature selected must be consistent with the atomizer feed slurry solids loading for efficient drying of the reagent solids. The *spraydown*, or difference between the inlet and outlet flue gas temperatures, and the solids loading of the atomizer feed slurry, determine how much reagent may be introduced to the process.

The predominant reagent used in dry scrubbers is lime slurry produced by slaking a high-calcium pebble lime. The slaking process can use a ball mill or a simple detention slaker. SDA systems that use only lime slurry as the reagent are known as *single pass* systems. Some of the lime remains unreacted following an initial pass through the spray chamber and is potentially available for further SO_2 collection. Solids collected in the ESP or baghouse may be mixed with water and reinjected in the spray chamber. This solids *recycle system* is used at most utility SDA installations to make use of the available alkalinity inherent in the flyash and to minimize reagent use. Some of the recycled
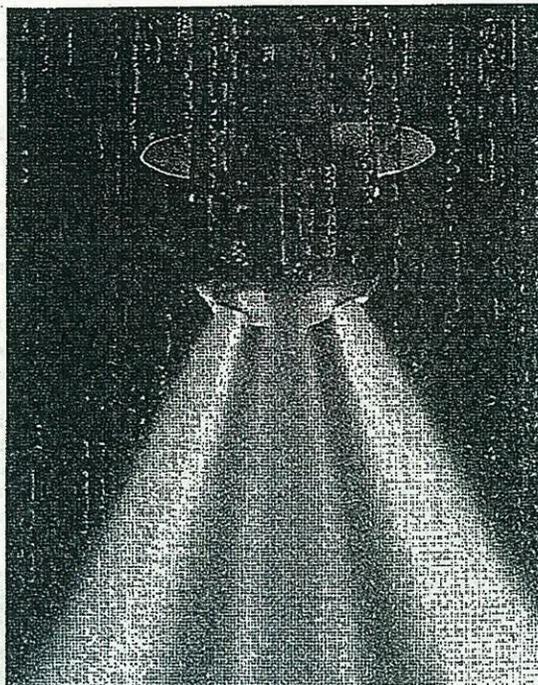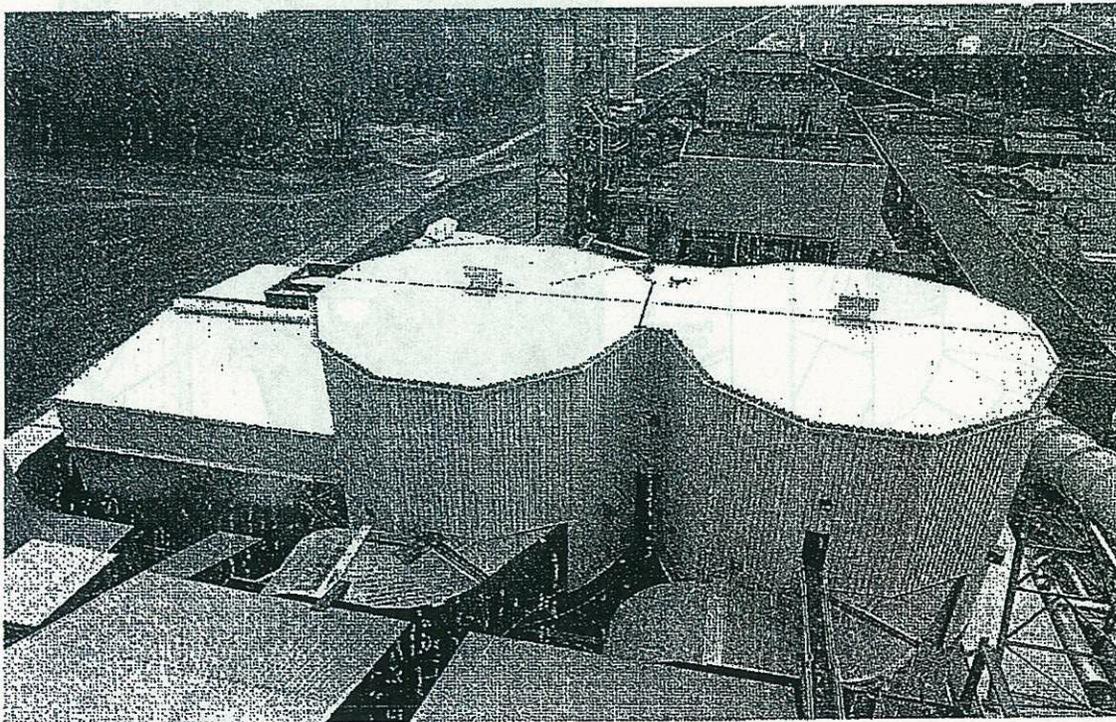

Fig. 11 Dual-fluid atomizer

Fig. 12 B&W's SDAs and fabric filter baghouse provide an integrated SO_2 and particulate removal process.

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solids form an inert core supporting the finer lime solids on the surface, thereby exposing more lime surface area for SO_2 absorption. The flyash may be removed from the flue gas before the SDA by a pre-collector. This is common practice in Europe where marketing of the flyash from coal-fired utility boilers is a well-established industry practice. Even with flyash pre-collection, a recycle system results in more efficient lime utilization.

The process is controlled by monitoring the flue gas temperature at the SDA outlet to control the amount of water or recycle slurry added to the atomizer feed. SO_2 emissions at the stack are compared to a desired operating set point and the lime slurry flow to the atomizers is adjusted to reduce the outlet SO_2 as necessary. For a rotary atomizer, the atomizer spindle, which connects the motor to the atomizer wheel, is monitored continuously for vibrations to provide an early indication of unbalance or pluggage of the wheel.

Process flow diagram and mass balance

A typical SDA process flow diagram is shown in Fig. 13. The flue gases, with or without pre-collection of the flyash, enter the spray dryer absorber where the gas stream is cooled by the reagent slurry spray. The mixture then passes on to the baghouse for removal of particulate before entering the ID fan and passing up the stack. Pebble lime (CaO) is mixed with water at a controlled rate to maintain a high slaking temperature that helps generate fine hydrated lime ($Ca(OH)_2$) particles with high surface area in the hydrated lime slurry (18 to 25% solids). A portion of the flyash, unreacted lime and reaction products collected in the baghouse is mixed with water and returned to the process as a high solids (35 to 45% typical) slurry. The remaining solids are directed to a storage silo for byproduct utilization or disposal. The fresh lime and recycle slurries are combined just prior to the atomizer(s) to enable fast response to changes in gas flow, inlet SO_2 concentrations, and SO_2 emissions as well as to minimize the potential for scaling.

Process mass balances for recycle and single pass SDA system designs for a typical application are presented in Table 8 while the power requirements are provided in Table 9. In this example, use of the recycle system reduces lime consumption by 3141 lb/h (1424.7 kg/h), or 42% over that expected with single-pass operation. The stoichiometry for the single pass system is 1.78 on an inlet SO_2 basis compared to 1.04 for the recycle system. Note the difference in acceptable approach temperatures for long-term operation of the two systems. The 7% solids feed slurry with the single pass system requires operation at a 45F (25C) approach temperature or above for adequate drying while a 30F (17C) approach is used with the recycle system.

Dry scrubber chemistry

The mechanism of SO_2 absorption in an SDA is similar to that attained by wet scrubbing when viewed on the level of what takes place in the individual slurry droplets. Most of the reactions take place in the aqueous phase; the SO_2 and the alkaline constituents dissolve into the liquid phase where ionic reactions produce relatively insoluble products. The reaction path can be described as follows:
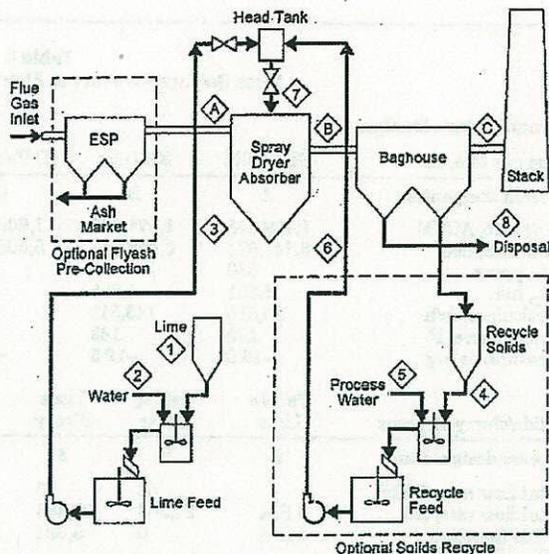
Fig. 13 Dry scrubber FGD system flow diagram (see Table 8 for mass balance).

$$SO_2(g) \rightleftharpoons SO_2(aq) \tag{24}$$
Dissolving gaseous SO_2

$$Ca(OH)_2(s) \rightarrow Ca^{++} + 2OH^- \tag{25}$$
Dissolution of lime

$$SO_2(aq) + H_2O \rightleftharpoons HSO_3^- + H^+ \tag{26}$$
Hydrolysis of SO_2

$$SO_2(aq) + OH^- \rightleftharpoons HSO_3^- \tag{27}$$

$$OH^- + H^+ \rightleftharpoons H_2O \tag{28}$$

$$HSO_3^- + OH^- \rightleftharpoons SO_3^= + H_2O \tag{29}$$
Neutralization

$$Ca^{++} + SO_3^= + \tfrac{1}{2}H_2O \rightarrow CaSO_3 \cdot \tfrac{1}{2}H_2O(s) \tag{30}$$
Precipitation
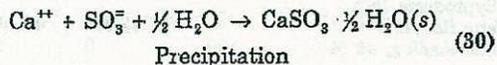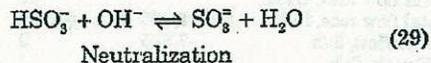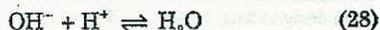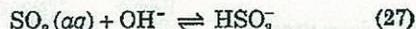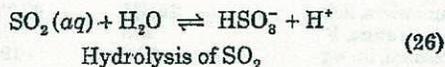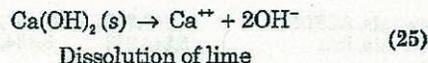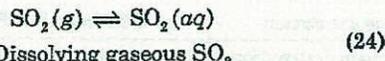
These reactions generally describe activity that takes place as heat transfer from the flue gas to the slurry droplet is evaporating water from the droplet. The rate-determining reaction may vary at different stages of the drying and absorption process. The high pH in the droplet environment (10 to 12.5) helps maintain a low concentration of acid in the liquid phase that enhances SO_2 absorption from the gas stream. Rapid SO_2 absorption occurs when liquid water is present. The drying rate can be slowed down to prolong this period of efficient SO_2 removal by adding deliquescent salts to the reagent feed slurry. Salts such

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Table 8
Mass Balance for Process Flowsheet Shown in Fig. 13

Recycle System Design

Flue gas stream	SDA Inlet	BH Inlet	ID Fan Inlet
Stream designation	A	B	C
Flow rate, ACFM	1,729,848	1,571,675	1,609,250
Flow rate, lb/h	5,244,973	5,454,664	5,508,400
SO_2, ppmv	380	89	21
SO_2, lb/h	4,381	1,086	263
Particulate, lb/h	26,835	148,511	76
Temperature, F	275	163	159
Pressure, in. wg	−16.0	−19.5	−27.0

500 MW boiler burning 0.44% sulfur coal with high calcium flyash
Available flyash alkalinity = 0.40 milliequivalents/gram
Flue gas saturation temperature = 133F
SO_2 emissions reduction = 94%
SO_2 emissions rate = 0.05 lb/10^6 Btu
Particulate emissions rate = 0.015 lb/10^6 Btu

Solids/slurry streams	Pebble Lime	Slaking Water	Lime Slurry	Recycle Solids	Process Water	Recycle Slurry	Atomizer Feed	Byproduct Solids
Stream designation	1	2	3	4	5	6	7	8
Total flow rate, GPM		48	50		269	363	413	
Total flow rate, lb/h	4,394	24,069	28,463	114,384	134,747	249,132	277,594	34,861
Solids flow, lb/h	4,394	0	5,693	112,096	0	112,096	117,789	34,167
Flyash, lb/h								24,273
Byproducts, lb/h								9,894
Water flow, lb/h		24,069	22,770	2,288	134,747	137,036	159,805	694
Solids loading, wt. %	100	0	20	98	0	45	42	98

Single Pass System Design

Flue gas stream	SDA Inlet	BH Inlet	ID Fan Inlet
Stream designation	A	B	C
Flow rate, ACFM	1,729,848	1,594,410	1,632,843
Flow rate, lb/h	5,244,973	5,424,712	5,478,350
SO_2, ppmv	380	73	21
SO_2, lb/h	4,381	880	263
Particulate, lb/h	26,835	40,936	76
Temperature, F	275	177	173
Pressure, in. wg	−16.0	−19.5	−27.0

Solids/slurry streams	Pebble Lime	Slaking Water	Lime Slurry	Recycle Solids	Process Water*	Recycle Slurry	Atomizer Feed	Byproduct Solids
Stream designation	1	2	3	4	5	6	7	8
Total flow rate, GPM		83	86		183	0	269	
Total flow rate, lb/h	7,535	41,278	48,813	0	91,254	0	140,067	41,467
Solids flow, lb/h	7,535	0	9,763	0	0	0	9,763	40,638
Flyash, lb/h								26,787
Byproducts, lb/h								13,851
Water flow, lb/h		41,278	39,050	0	91,254	0	130,304	829
Solids loading, wt. %	100	0	20	0	0	0	7	98

*Process water addition to lime slurry for atomizer feed to achieve desired SDA outlet temperature.

as calcium chloride also increase the equilibrium moisture content of the end product. However, since the use of these additives alters the drying performance of the system, the operating conditions must be adjusted (generally increasing the approach temperature) to provide for good long-term operability of the SDA and the ash handling system. Ammonia injection upstream of a dry scrubber also increases SO_2 removal performance. SO_2 absorption continues at a slower rate by reaction with the solids in the downstream particulate collector.

An SDA/baghouse combination also provides efficient control of HCl, HF and SO_3 emissions by the summary reactions of:

$$Ca(OH)_2 + 2HCl \rightarrow CaCl_2 + 2H_2O \qquad (31)$$

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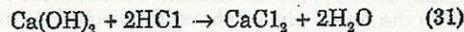
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Table 9 Typical Power Requirements for 500 MW SDA/Baghouse System in Table 8		
	Recycle System Design	Single Pass System Design
Atomizers and accessories, kW	1190	640
Reagent preparation, kW	230	125
Baghouse accessories, kW	445	445
ID fan power	6780	6880
Total	8645	8090
% of net 500 MW output	1.73	1.62

$$Ca(OH)_2 + 2HF \rightarrow CaF_2 + 2H_2O \quad (32)$$

$$Ca(OH)_2 + SO_3 \rightarrow CaSO_4 + H_2O \quad (33)$$
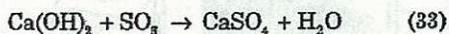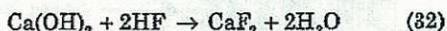
Proper accounting of the reagent consumption must include these side reactions, in addition to the SO_2 removed in the process.

The SDA process can make use of a variety of water sources for preparation of the reagent slurry and cooling of the flue gas. River water, well water, cooling tower blowdown, municipal waste water treatment effluent, and seawater have all been used successfully in SDA applications. In general, lime slaking requires good quality water with limitations on the sulfite, sulfate and TDS concentrations being the primary concerns for producing good quality fresh reagent slurry. Preparation of the recycle solids slurry and humidification of the flue gas to the desired operating temperature can be achieved with lower quality water sources. Selection of the process operating conditions must also consider the chloride content of the water sources to avoid problems with drying the reagent slurry and potential corrosion.

Sulfur trioxide (SO₃)

SO₃ and acid mist formation

Sulfur trioxide (SO_3) is formed directly in the combustion of sulfur-containing fuels, and indirectly by the conversion of small quantities of flue gas SO_2 to SO_3 in the presence of iron, some ash constituents such as vanadium, and some selective catalytic reduction (SCR) catalysts. Under normal operating conditions in coal-fired boilers without SCR systems, 0.75% of the SO_2 typically is converted to SO_3. High vanadium containing ashes can increase this range to 2%. Conversion of SO_2 to SO_3 by SCR catalysts is dependent upon SO_2 concentration, gas temperature, and catalyst formulation with a typical range of an additional 0.7 to 1.5%. Catalysts have been formulated with relatively low SO_2 to SO_3 oxidation potential.

As the flue gas temperature falls below 1000F (538C), SO_3 begins to react with water in the flue gas to form sulfuric acid vapor, H_2SO_4. By the time the flue gas exits the air heater, virtually all of the SO_3 has reacted to form H_2SO_4.

Sulfuric acid condenses on metal surfaces beginning in the temperature range of 200 to 300F (93 to 149C) and results in corrosion of metal surfaces in flues, air heaters, dry ESPs and fans. Acid dew point curves are presented in Fig. 14. In addition, in units with wet scrubbers, any significant residual sulfuric acid mist can cause visible plume or opacity problems (blue plume), visible several miles from the power plant. Wet scrubbers are not effective in removing acid mists because the rapid flue gas quenching in the scrubber produces submicron aerosol mists which are not easily collected (20 to 30%). Dry scrubbers in combination with downstream baghouses are effective in removing H_2SO_4 (> 95% typical). Finally, in units equipped with SCR systems, excess ammonia (NH_3) leaving the SCR can react with SO_3 to form sticky, acidic particles of ammonium bisulfate (NH_4HSO_4) which can foul air heaters.

Control methods

Sorbent injection The injection of wet or dry alkali compounds (usually calcium or magnesium oxide compounds), typically in the furnace, after the SCR, or upstream of the particulate collection device has been shown to be an effective means of controlling SO_3/H_2SO_4. Sorbent injection has the advantage of low capital cost, but has very high reagent cost due to the high stoichiometric ratio (4 to 8) that is needed to achieve 80 to 90% SO_3/H_2SO_4 control. Dry ESP performance may be adversely affected by removing the SO_3 from the flue gas (see Chapter 33).

Humidification Cooling the flue gas by humidification upstream of an existing dry ESP will allow the dry ESP to collect some sulfuric acid. It is possible to operate a dry ESP at 20F (11C) below the acid dew point without encountering excessive corrosion. 20 to 40% sulfuric acid removal can be achieved in some instances. This is possible because the acid is absorbed on the flyash rather than on the metal surfaces of the dry ESP. Careful attention must be paid to thermal insulation integrity and in-leakage of seal air streams.

Ammonia injection Ammonia injection upstream of the dry ESP has been used to control acid mist. However, there can be problems with ash handling, since the ash tends to be more cohesive. It is also possible that excess ammonia will react with the SO_3 to form
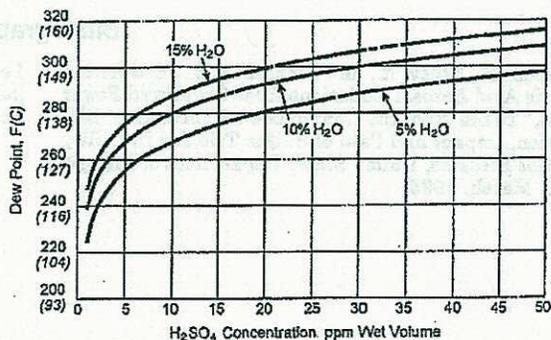
Fig. 14 Dew point versus concentration

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submicron aerosols that are difficult to collect. Capital costs are low, but reagent costs are high.

Wet precipitator To achieve very high removals of acid mist, a wet ESP can be used. For a utility boiler, the wet ESP would be used downstream of the wet FGD system.

Either a stand-alone arrangement or an integrated close-coupled wet ESP can be used to control SO_3; however, the standalone arrangement is typically more costly than the integrated design and consumes considerable additional space. Fig 15 shows the integrated design.

The submicron nature of the acid mist can only be treated in a multi-field unit. Acid mist removal efficiencies of 80 to 90% can be achieved. For example, if the inlet is 34 ppmdv, less than 5 ppmdv at the stack can be achieved. Wet ESP can also remove 80 to 90% of any solid particulate that is present, such as residual flyash from the dry ESP and the small amounts of gypsum that result from the wet FGD moisture separator wash system. Even if a wet ESP is not included in a wet FGD project today, careful thought should be given to future needs. The addition of a wet ESP to an existing wet FGD system may not be possible if provisions are not made before the initial wet FGD system installation.
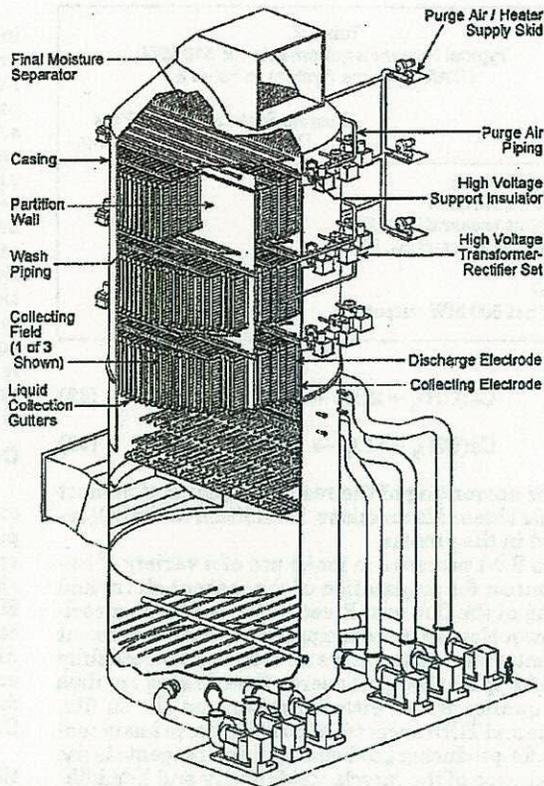
Fig. 15 Integrated wet ESP and wet FGD system

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Lesley, S., "Trends in Emissions Standards," CCC/77 Report, IEA Clean Coal Centre, International Energy Agency, London, England, United Kingdom, November, 2003

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RE Consulting

Wet Scrubber O&M Issues

Ron Richard

Senior Consultant

RE Consulting

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Wet Scrubber O&M Issues

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Ball Mills

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Wet Scrubber O&M Issues

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Limestone Fineness

- Some of the FGD vendors have stated that one of the biggest causes of performance issues they find is improper limestone fineness in the limestone slurry feed to the absorber tower.

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Effect of Limestone Grind

(Different Shaped Symbols Represent Different Limestone Tested)
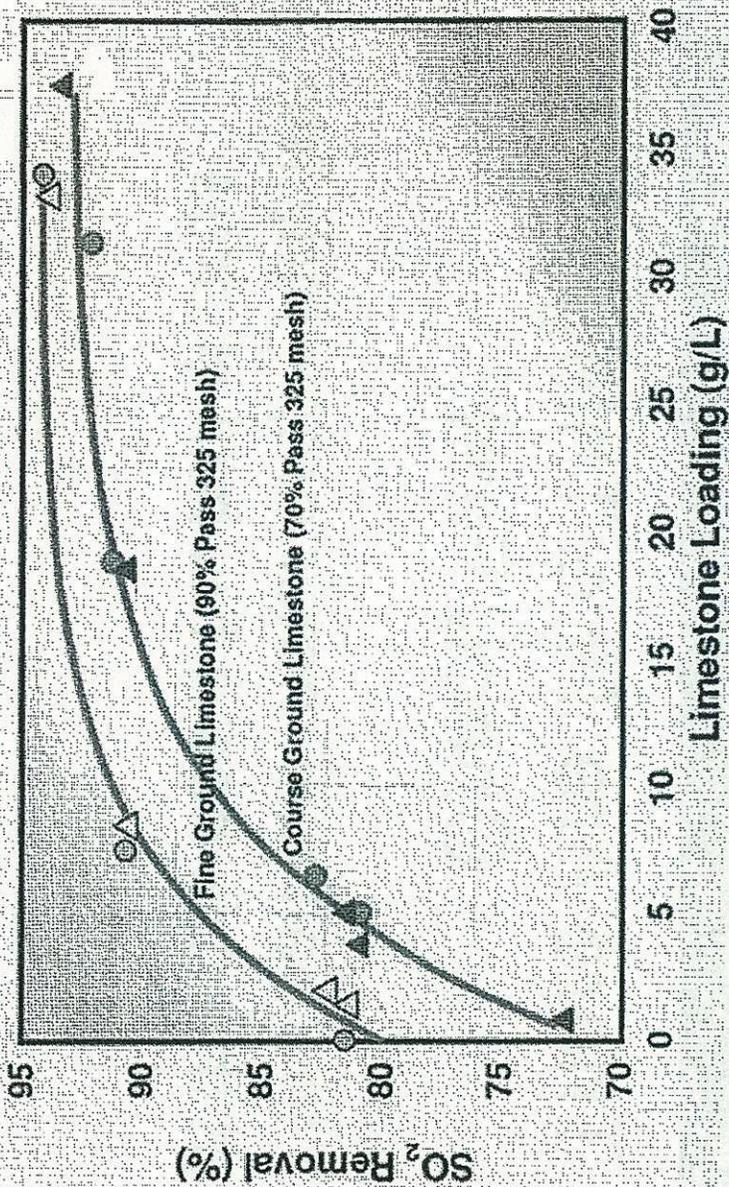
Fine Ground Limestone (90% Pass 325 mesh)

Course Ground Limestone (70% Pass 325 mesh)

y-axis: SO$_2$ Removal (%) — 70, 75, 80, 85, 90, 95

x-axis: Limestone Loading (g/L) — 0, 5, 10, 15, 20, 25, 30, 35, 40

Source: URS

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There Are Two Things To Check First

- Do the ball mill motor amps indicate that the mill has a full ball charge.

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Normal Ball Level

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RE Consulting
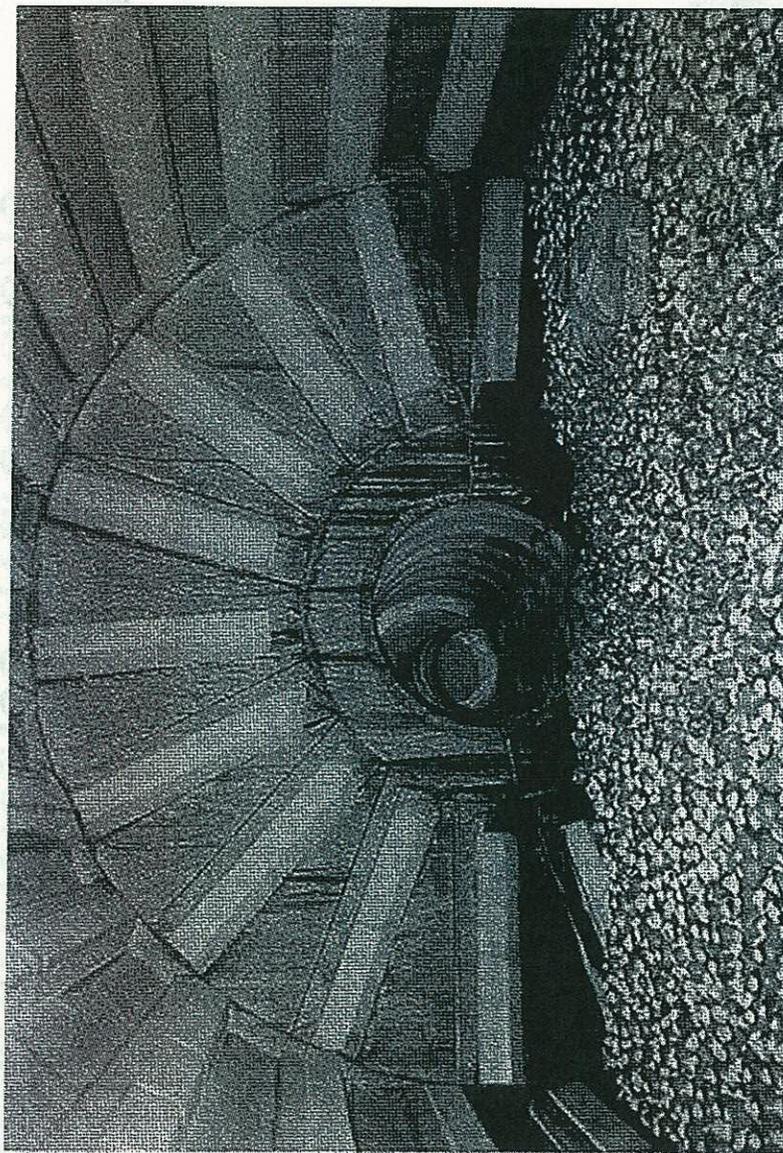
There Are Two Things To Check First

- Do the ball mill motor amps indicate that the mill has a full ball charge.

- Is there grit in a hydrocyclone overflow line that indicates hydrocyclone puggage.

Wet Scrubber O&M Issues

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CYCLONE CUTAWAY

OVERFLOW

FEED

Cylindrical Feed Chamber

Vortex Finder

Cylinder Section

Conical Section

Apex

UNDERFLOW

KREBS ENGINEERS

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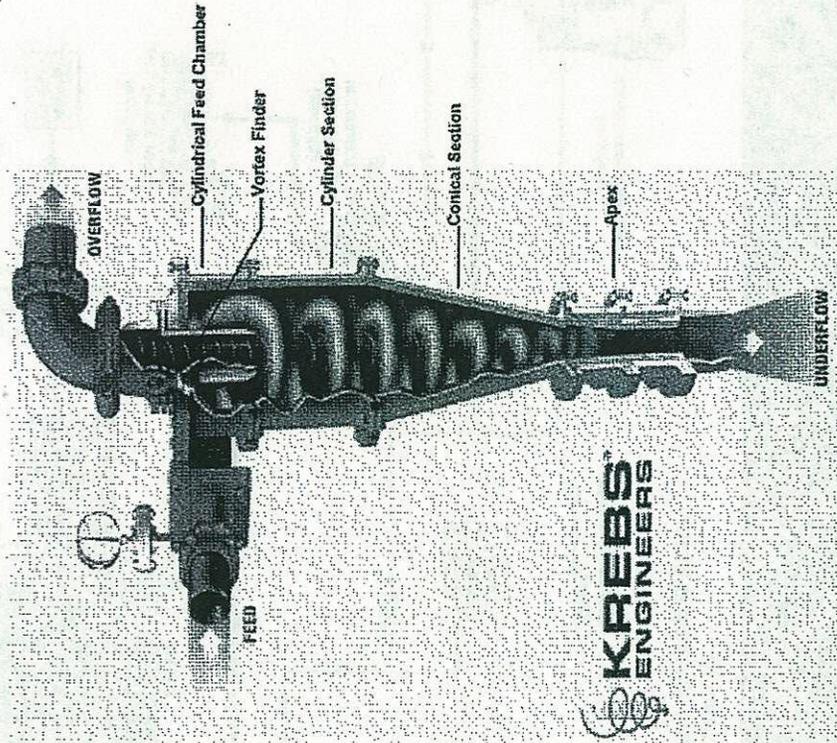
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Typical Flow Diagram for
Ball Mill Limestone Grinding System

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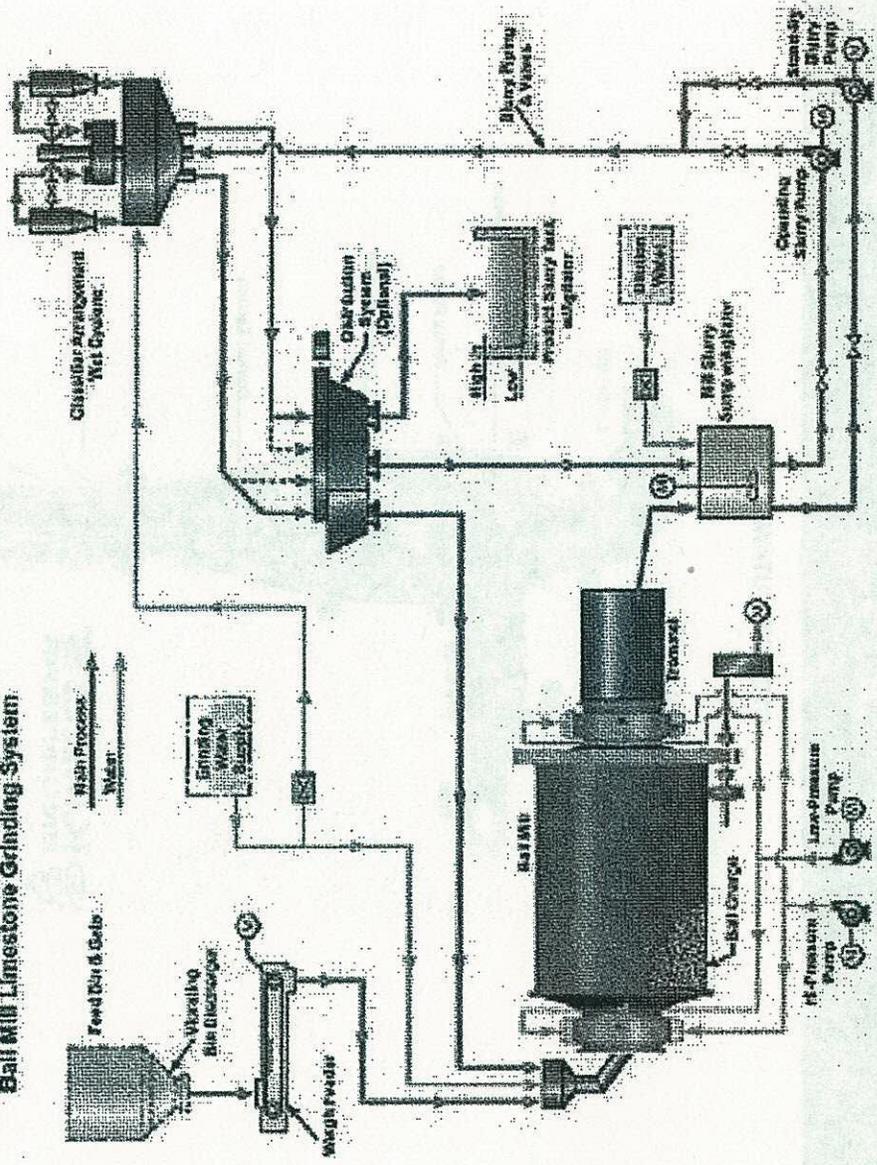
Ball Mill Design

$$P = T \times W$$

P = Power Draw (kW) [Motor Size]

T = Limestone Feed Rate (ton/hour)

W = Work Input (kWh/ton)

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Work Input

$$W = \frac{10\, W_i}{\sqrt{D_P}} - \frac{10\, W_i}{\sqrt{D_F}}$$

W_i = Bond Work Index (kWh/ton)

D_P = Diameter of 80% of Product (microns)

D_F = Diameter of 80% of Feed (microns)

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Typical Generating Station Work Index Versus Capacity

Capacity—DSTPH

Work Index
Ore Hardness Analysis

■ Denotes 95% Passing 325 Mesh
■ Denotes 90% Passing 325 Mesh

Typical Generating Station Work Index Versus Product Fineness

Cumulative Percent Passing 325 Mesh—(44 Microns)

Work Index
Ore Hardness Analysis

Source: METSO

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Mist Eliminators

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Achilles' Heel??

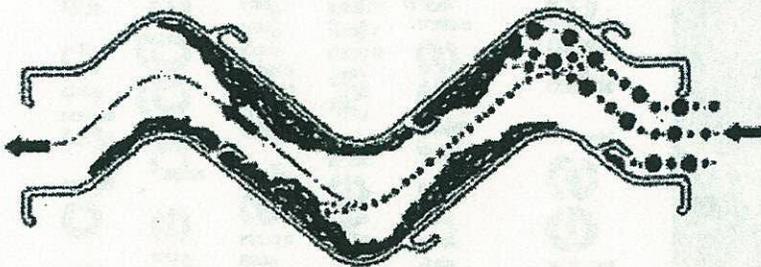- With the current design of a large single FGD tower with no bypass, I feel that the mist eliminators may pose one of the biggest risks for a forced outage of the unit.

Wet Scrubber O&M Issues

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The Problem – Scale Formation

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The Solution

- All FGD vendors supply a mist eliminator wash system as part of their design.

Wash Nozzles

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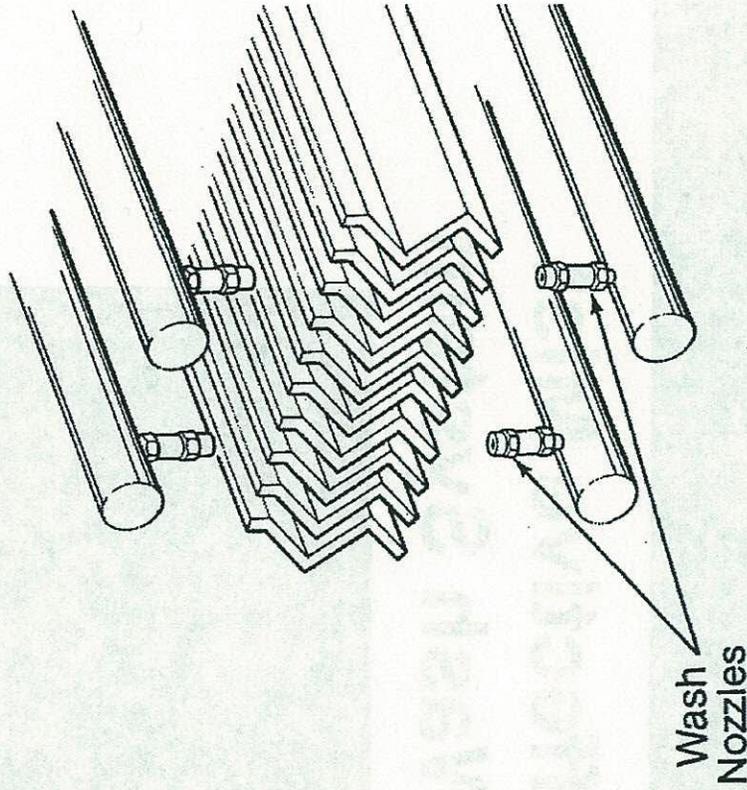
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The Only Effective Mist Eliminator Wash System

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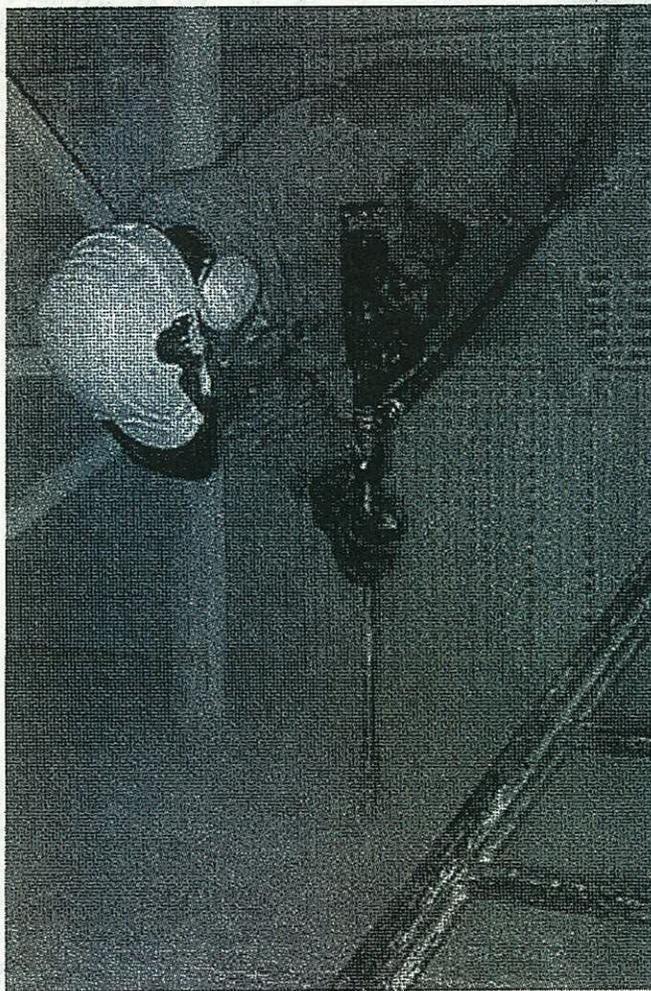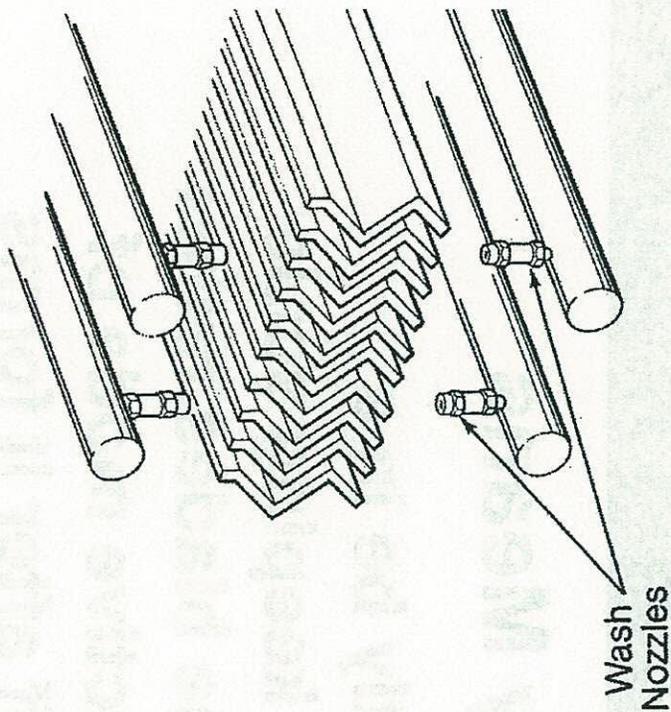
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What You Really Have

- A "keep the blades wet and the liquid collecting on them diluted below calcium saturation so scale doesn't form" system.

Wash Nozzles

Wet Scrubber O&M Issues

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What That Really Means

- You must constantly be in a proactive mode of keeping scale from forming on the blades rather than being in a reactive mode of trying to wash it off after it forms because you can't wash all of it off with the system you have.

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What That Means

- The wash water needs to be as fresh (unsaturated with calcium) as possible.

- Each blade needs to be washed as often as possible.

- An "in-situ forced oxidized" tower has less problems.

- Adding a scale inhibitor to the wash water may need to be considered.

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Wet Scrubber O&M Issues

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ME Wash Water

- Use as much "service water" or other fresh water as the water balance will allow.

- Add "process water" only as needed to maintain the level in the ME Wash Water Tank.

Wet Scrubber O&M Issues

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Wash Cycles

- Most designs only wash one portion of one side of a mist eliminator at a time.

- If possible, always be washing something.

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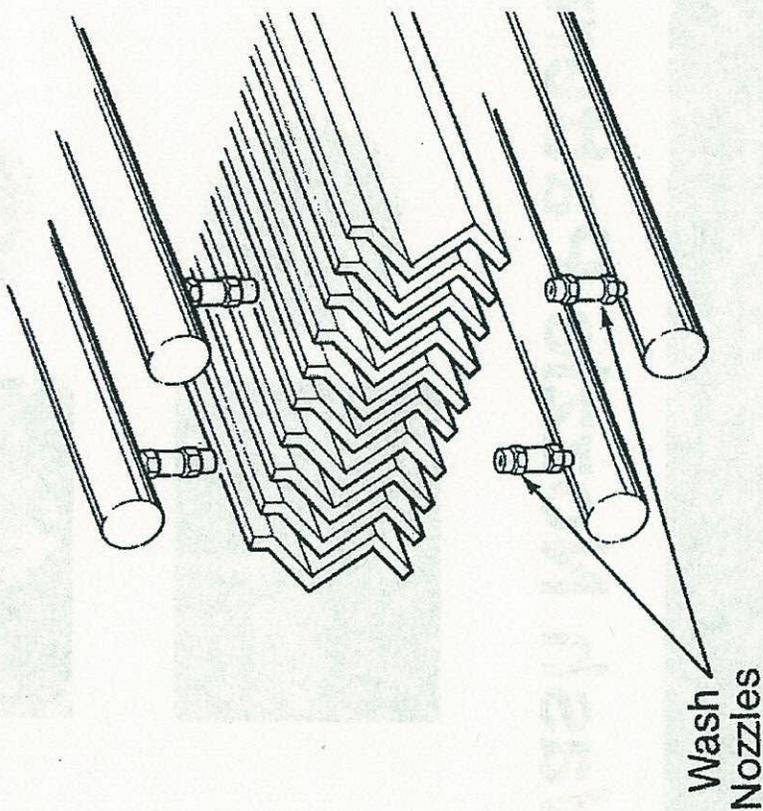
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Wash Cycles

- If the water balance will only allow you to wash each section for 8 minutes every hour:

  - It is better to wash for 1 minute every 8 minutes

  - Than to wash for 2 minutes every 15 minutes
  - Than to wash for 4 minutes every 30 minutes
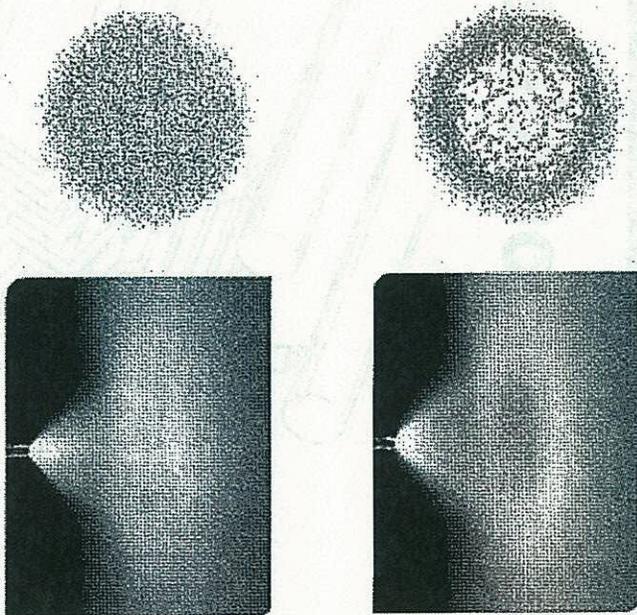 - Than to wash for 8 minutes every 60 minutes

RE Consulting

Wet Scrubber O&M Issues   WPCA/Duke Seminar

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Wash Nozzle Considerations

Wash Nozzles

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Wash Nozzle Pattern

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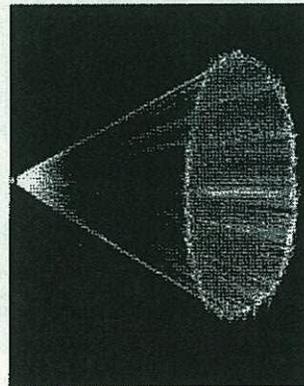Item No. 5
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Spray Pattern Angle
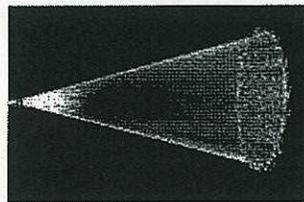
Full Cone 30° (NN)

Full Cone 60° (N)
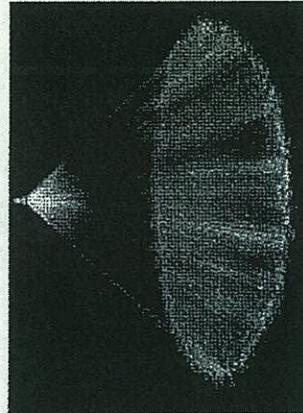
Full Cone 90° (M)

Full Cone 120° (W)
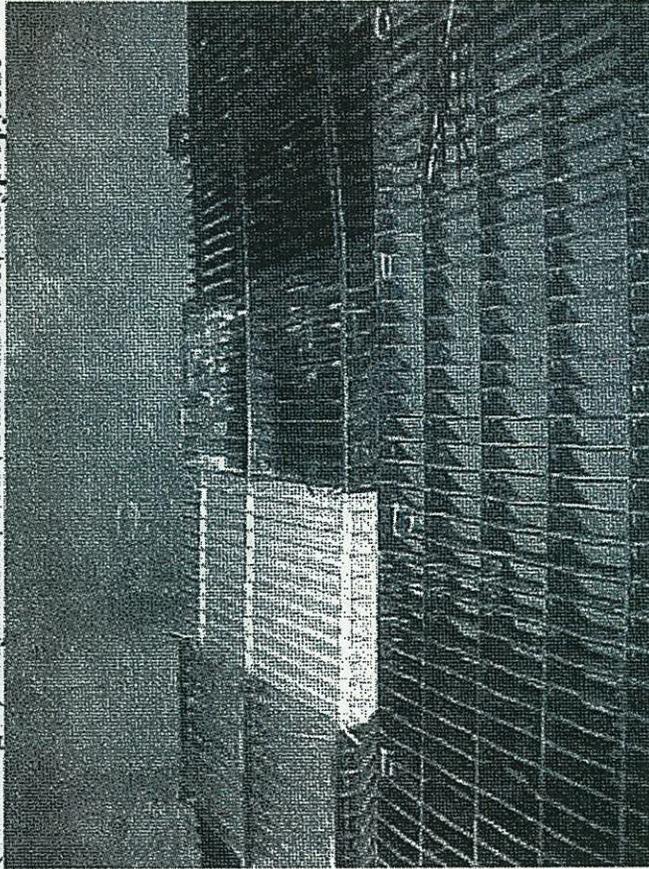
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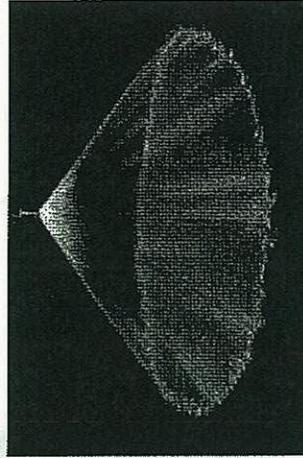
APS Side by Side Comparison

Figure 2. Test Installation at APS's Cholla Station, Unit 2: Materials
(left to right) Polysulfone, FRP, Stainless Steel, Polypropylene

Source: KOCH-Glitsch

Wet Scrubber O&M Issues

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APS Side by Side Comparison

Figure 3. Polypropylene mist eliminator
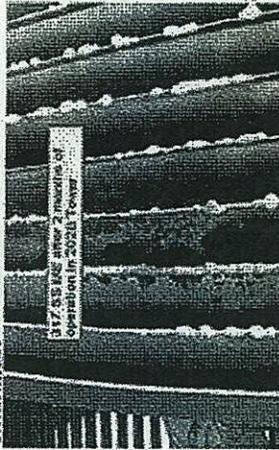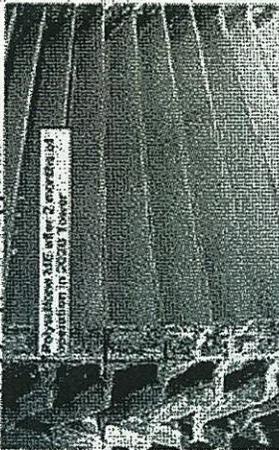
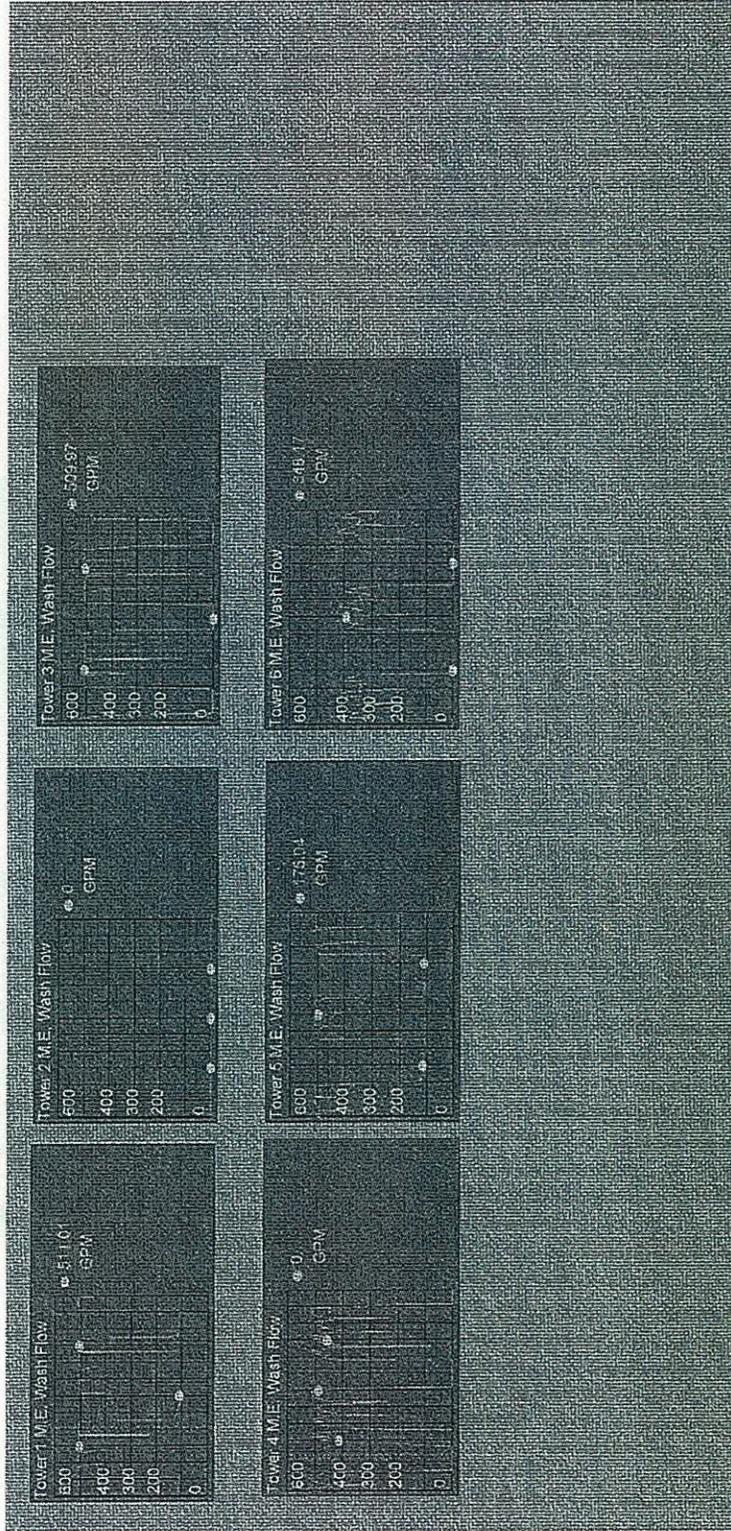Figure 5. Stainless steel mist eliminator

Figure 4. FRP mist eliminator from

Figure 6. Polysulfone mist eliminator with polypropylene supports

Source: KOCH-Glitsch

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Quadrant Valves Not Opening

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Leaking ME Wash Valves?

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Asiatic Clams Plugging ME Wash Pump Suction Strainer – Part 1

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Asiatic Clams Plugging ME Wash Pump Suction Strainer – Part 2

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ME Plugged 15 Days After Previous Cleaning

FGD-FIAL-J650
66.812
GPM
FGD-PDI-B364
Bad Input
IN H2O

Plot-0

600 6

500

400

300

200

100

0   -1

7/26/00 1:07:15 PM

15:00 Day(s)

8/10/00 1:07:15 PM

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Wash Failure Resulting In Tower Forced Outage

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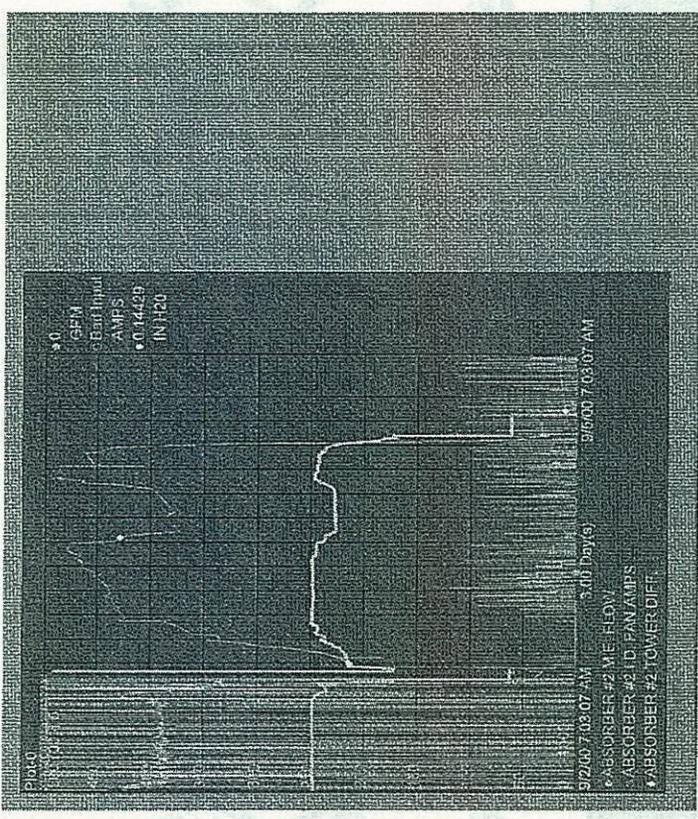
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Cleaning Effectiveness

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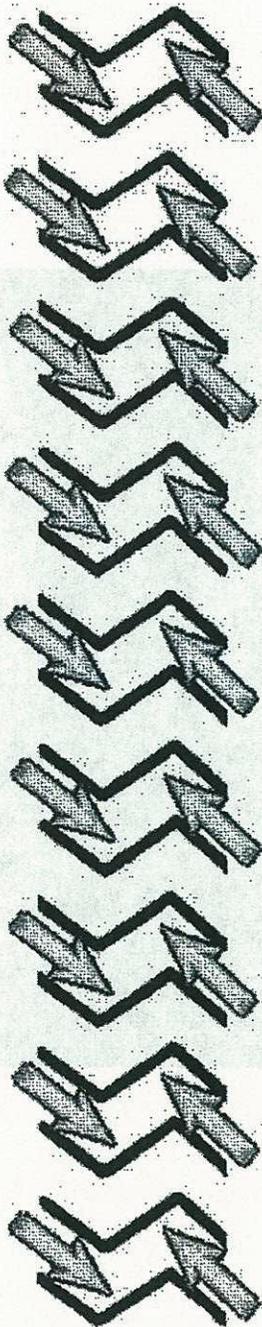
Preventing Damage During High Pressure Washing

- Don't use any more nozzle pressure than necessary to remove the deposit.

- A fan pattern nozzle does less cutting of the blade than a straight pattern nozzle.

- Keep the nozzle several inches from the blade.

- Keep the nozzle as parallel as possible with the blades.

RE Consulting

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Wet Scrubber O&M Issues

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RE Consulting

I hope that something you saw here today will allow you to experience less problems with your mist eliminators.

Wet Scrubber O&M Issues

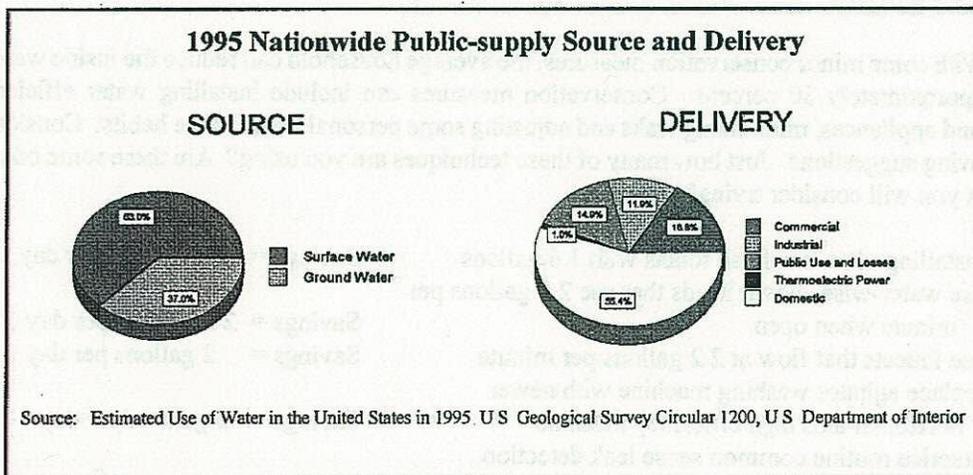WPCA/Duke Seminar

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BS-05

ENRI-117

COOPERATIVE EXTENSION SERVICE
UNIVERSITY OF KENTUCKY • COLLEGE OF AGRICULTURE
UK

Water Usage

Water has always been a very precious resource. However, many of us do not fully understand or appreciate water's value until there is a shortage. According to the U.S. Geological Survey (USGS), "the United States as a Nation possesses abundant water resources and has developed and used those resources extensively." As consumers and users, we must be good managers today, and teach future generations the importance of water-use efficiency.

Just how are we doing? According to recent water statistics compiled by the USGS, the United States is using *less water* (402 billion gallons per day of fresh and saline water) for all uses. This 1995 data represents a 2 percent drop since 1990, and almost a 10 percent decrease since 1980 despite a continuous gain in the number of users (population). The USGS has been compiling statistics since 1950 and reports findings in five year intervals. The drop in water usage is believed to be the result of enhanced citizen awareness of the value of water, conservation programs undertaken in many communities, improved irrigation techniques, and more efficient use of water by industry. Irrigation is the leading freshwater use category at 134 billion gallons a day, down 2% since 1990. Other categories include public supply (up 4%), rural domestic and livestock (up 13%), industrial (down 3%), mining, and thermoelectric (down 3%). In a state-by-state comparison, California leads the nation in largest total water use followed by Texas, Illinois and Florida.

According to the survey, public suppliers served approximately 225 million people during 1995. The term *public supply* refers to water used by public and private water suppliers and delivered to multiple users. The chart below reflects water sources and the delivery of water for public supply.

1995 Nationwide Public-supply Source and Delivery

SOURCE

DELIVERY

63.0%

37.0%

Surface Water
Ground Water

11.9%

14.9%

1.0%

16.8%

55.4%

Commercial
Industrial
Public Use and Losses
Thermoelectric Power
Domestic

Source: Estimated Use of Water in the United States in 1995. U S Geological Survey Circular 1200, U S Department of Interior

AGRICULTURE & NATURAL RESOURCES • FAMILY & CONSUMER SCIENCES
4-H/YOUTH DEVELOPMENT • RURAL & ECONOMIC DEVELOPMENT

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In terms of publically supplied water, the major water-use categories are domestic, commercial, industrial, public use and losses, and thermoelectric power (representing less than 1 percent). Domestic water use includes water used for normal household purposes such as drinking, food preparation, bathing and personal hygiene, dish washing, laundry, general household cleaning, watering lawns and gardens. The average consumption per person per day is estimated at 80 gallons. This figure has stayed the same or slightly declined in some areas over the past five years. Kentucky reported an average use of 50 gallons per person per day by individuals using private water supplies (wells, cisterns, springs, etc.), and 70 gallons per person per day by individuals using public water supplies.

Water use inside the home varies from household to household, individual to individual. The American Water Works Association provides us with the following statistics for water use in a *single family home* without the use of any type of conservation measures:

Activity	Gallons Used per day	Percent of Total
Toilets	19.3	26
Clothes Washer	16.8	22.7
Showers	13.2	17.8
Faucets	11.4	15.4
Leaks	9.4	12.7
Other domestic use	1.6	2.1
Baths	1.3	1.8
Dish washer	1.0	1.4

With some minor conservation measures, the average household can reduce the inside water use by approximately 30 percent. Conservation measures can include installing water efficient fixtures and appliances, minimizing leaks and adjusting some personal water usage habits. Consider the following suggestions. Just how many of these techniques are you using? Are there some other ideas that you will consider trying?

- installing ultra-low flush toilets with 1.6 gallons   Savings = 9.3 gallons per day
- use water-wise showerheads that use 2.5 gallons per
    minute when open   Savings = 2.1 gallons per day
- use faucets that flow at 2.2 gallons per minute   Savings = 2 gallons per day
- replace agitator washing machine with newer
    horizontal-axis high efficiency machine   Savings = 5 gallons per day
- practice routine common sense leak detection
    periodically "zero read" water meter for leaks
    replace worn valves, faucet washers, O rings   Savings = 4.7 gallons per day

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- take shorter showers --15 minutes instead of 20
  (showers using 5 gallon per minute)   Savings = 25 gallons per shower
- take a 10 minute shower instead of a bath
  (full tub bath uses 35 gallons)   Savings = 15 gallons per bathing
- turn water off in shower to soap up/shave   Savings = 15 - 20 gallons per shower
- on older/conventional toilets, displace water in
  tank using plastic jug   Savings = 15 gallons per day
- turn water off to brush teeth, shave   Savings = 10 gallons per day
- run a full load of dishes in dishwasher   Savings = 15 gallons per load
- peel and clean vegetables in a bowl of water
  instead of letting water run   Savings = 15 gallons per meal
- hand washing dishes--soak dirty dishes in a basin
  then rinsing them off   Savings = 15 gallons per meal
- use "small load" designation on washing
  machine for less than full load   Savings = 20 to 30 gallons per load
- wait and launder clothing when you have a
  full load; however, DON'T over load
  washing machine as clothing will not be
  clean and re-washing will be necessary!   Savings = 30 to 60 gallons per load

Even though we have seen a decrease in the amount of water used, there remains heightened concern for the quality of water. The quantity of water is a given; water quality is not! With increases in population comes increased demands for water-based recreation activities in addition to good quality drinking water. As a household, a community, a state and a nation, we must continue to make good decisions and good use of our precious water resource.

References:
American Water Works Association, *Water Use Inside the Home* 1998 Water Use Summary.
United States Environmental Protection Agency, Office of Water, *Fact Sheet 21 Water Conservation Measures for Everybody.*
United States Geological Survey, U.S. Department of the Interior, *Estimated Use of Water in the United States in 1995*, U.S. Geological Survey Circular 1200.
United States Geological Survey, USGS web site, *Trends in Domestic Water Use*, Water Science for Schools.

Written by: Linda Heaton, Ph.D., Extension Professor for Textiles & Environment
water usage.wpd [Disk: Water #2]

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Energy Information Administration
Official Energy Statistics from the U.S. Government

Glossary | search

Home > Electricity > Electric Sales, Revenue, and Price > Average Monthly Bill by Sector Census Division and State

Table 5. U.S. Average Monthly Bill by Sector, Census Division, and State 2007

Date of Latest Data: 2007
Report Released: January 2009
Next Release Date: December 2009

| Residential | Commercial | Industrial |

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Residential

Census Division State	Number of Consumers	Average Monthly Consumption (kWh)	Average Retail Price (Cents per Kilowatthour)	Average Monthly Bill (Dollar and cents)
New England	6,122,590	649	16 70	$108.46
CT	1,459,141	764	19 11	$145.95
MA	2,642,144	635	16.23	$103 09
ME	694,235	530	16.52	$87.53
NH	592,238	632	14.88	$94 06
RI	429,504	608	14.05	$85 36
VT	305,328	592	14 15	$83 77
Middle Atlantic	15,535,148	722	13 95	$100 71
NJ	3,394,474	730	14 14	$103.28
NY	6,933,304	604	17 10	$103.25
PA	5,207,370	874	10.95	$95.65
East North Central	19,558,522	830	9 74	$80.84
IL	5,065,123	790	10.12	$80 01
IN	2,727,752	1,058	8 26	$87 44
MI	4,298,455	686	10 21	$70.02
OH	4,894,635	926	9 57	$88 60
WI	2,572,557	725	10 87	$78 75
West North Central	8,969,792	970	8 31	$80.62
IA	1,323,176	886	9.45	$83 65
KS	1,250,241	920	8 19	$75 40
MN	2,267,167	832	9 18	$76.40
MO	2,666,181	1,121	7 69	$86.22
ND	314,466	1,078	7 30	$78 73
NE	790,165	1,028	7.59	$78.04
SD	358,396	991	8.07	$79.98
South Atlantic	25,336,801	1,156	10 03	$115 95
DC	212,389	773	11 18	$86 42
DE	387,923	960	13 16	$126 40
FL	8,445,561	1,163	11.22	$130 47
GA	4,001,957	1,171	9 10	$106.48
MD	2,163,625	1,086	11 89	$129 15
NC	4,090,510	1,143	9.40	$107 38
SC	2,035,713	1,210	9.19	$111 20
VA	3,139,055	1,207	8 74	$105.55
WV	860,068	1,138	6 73	$76 63
East South Central	7,895,454	1,290	8 35	$107.79
AL	2,093,093	1,305	9.32	$121 70
KY	1,917,257	1,217	7 34	$89 35
MS	1,227,001	1,261	9 36	$118 00
TN	2,658,103	1,344	7 84	$105 42
West South Central	13,968,942	1,149	11 15	$128.14
AR	1,298,492	1,118	8 73	$97 58
LA	1,885,657	1,276	9 37	$119 62
OK	1,617,944	1,100	8.58	$94.45
TX	9,166,849	1,136	12 34	$140.17
Mountain	8,722,741	908	9 31	$84.54

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RIVERS OF LIFE — Inquiry Projects

Rivers Through Time · Steamer Trunk · Energy Odyssey · Chasing the Flood · Teachers Guide

Compare Cargo Capacity

Huge quantities of grain, corn, coal, petroleum, and other products make their way up and downstream on the Mississippi River each year. How huge? Study the comparisons on this page--then do the simple math problems at the bottom of the page, and you'll be amazed!

One Barge
1500 Ton
52,500 Bushels
453,600 Gallons

One 15 Barge Tow
22,500 Ton
787,500 Bushels
6,804,000 Gallons

Jumbo Hopper Car
100 Ton
3,500 Bushels
30,240 Gallons

100 Car Train Unit
10,000 Ton
350,000 Bushels
3,024,000 Gallons

Large Semi
26 Ton
910 Bushels
7,865 Gallons

Equivalent Units

One Barge carries the same quantity as:

15 Jumbo Hopper Cars (above) & **58 Large Semis** (right)

and subsiquently ..

One 15 Barge Tow carries the same quantity as:

2.25 100-Car Unit Trains (above) &

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Rivers of Life: Rivers Through Time - Compare Cargo Capacity   Page 2 of 2

870 Large Semis (right)
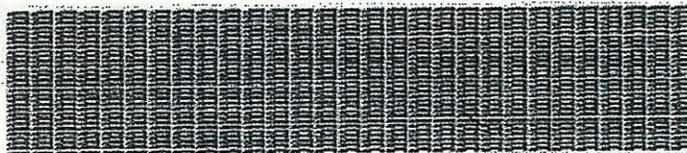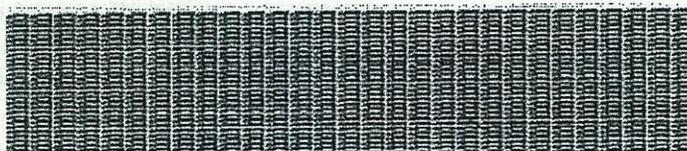
Equivalent Lengths
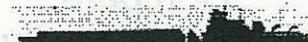
One 15 Barge Tow

is .25 miles long

2.25 100-Car Train Units is
2.75 miles long & the **870 Large
Semis** are 11.5 miles long
(bumper to bumper)

**In 1997, 38.8 million tons of cargo was transported
on the Upper Mississippi River.
With the above information, figure out the answers to these questions.**

1) If all the cargo transported in 1997 was done by semis, how many semis would it have taken?

2) If all the cargo transported in 1997 was done on barges, how many 15-barge tows would it have taken?

2) How many bushels of cargo is in the 11.5 mile long string of large semis?

3) EXTRA CREDIT..... .How long is a string of 58 large semis going down the highway, bumper to bumper?

Return to Rivers Through Time resources page

SITE MAP   RIVERS OF LIFE HOME PAGE   CONFERENCE CENTER   CONTACT US
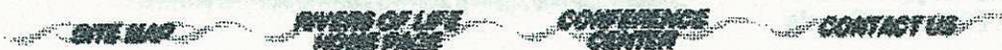
Center for Global Environmental Education
Hamline University Graduate School of Education
1536 Hewitt Avenue, St. Paul, MN 55104-1284
Phone: 651-523-2480 Fax: 651-523-2987

BS-08

Technical Paper
BR-1815

An Emissions Approach to SO₃ Mitigation

Authors:
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A L. Moretti
R E. Snyder

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Power Generation Group, Inc.
Barberton, Ohio, U.S.A.

A Lau
AES Deepwater Inc.
Pasadena, Texas, U.S.A.

B&W babcock & wilcox power generation group

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An Emissions Approach to SO₃ Mitigation

D.P. Tonn, A.L. Moretti and R.E. Snyder
Babcock & Wilcox Power Generation Group, Inc.
Barberton, Ohio, U.S.A.

A. Lau
AES Deepwater Inc.
Pasadena, Texas, U.S.A.

Abstract/Introduction

Total particulate emissions are being set near detection limits on new or revised power plant operating permits. Acid gas emissions, such as SO_3, which form particulate and make up a portion of total particulate, must be controlled to even lower values. Thus, the selected acid gas control technologies must be based on meeting the required emissions. This paper provides the results of investigations that have been performed and demonstrates achievable SO_3 emissions utilizing an emissions control system approach for a range of fuel sulfur contents. Available total particulate emission results on both high and low sulfur fuels with multiple control technologies are presented as demonstration of SO_3 mitigation. Consideration of multiple SO_3 mitigation technologies and reagents is included. The conclusions of this analysis can be utilized in the economic selection of the proper control technologies for the selected primary fuel and sulfur content to achieve the total particulate emissions.

SO₃ emissions and total particulate

When fuel is combusted in a utility boiler, the sulfur in the fuel combines with oxygen to form gaseous sulfur dioxide (SO_2). Some of the SO_2 is further oxidized to SO_3. The SO_3 hydrates to form H_2SO_4 vapor in the flue gas stream. Finally, the H_2SO_4 vapor condenses to form a sub-micron aerosol. The acid aerosols that avoid capture or collection in the downstream flue gas path (air heaters) and in the air pollution control equipment (electrostatic precipitators, baghouses and FGD systems) exit the stack and contribute to total particulate emissions.

When establishing or selecting the required technologies to achieve the total particulate emissions, it is important to understand SO_3 formation as well as the equipment and technologies for SO_3 mitigation. SO_3 forms in two areas of a fossil fuel fired power plant, the boiler and across the catalyst on those units utilizing SCR for NO_x reduction. Combustion of fuel liberates elemental sulfur, which oxides to form SO_2. At typical combustion temperatures very little SO_2 is converted to SO_3, approximately 0.1%, because the thermodynamic equilibrium at boiler combustion temperatures favor SO_2 formation. As heat is absorbed by the boiler surfaces and the flue gas temperature lowers, the equilibrium changes so that more SO_3 can be formed. The formation of SO_3 is dependent on the existence of atomic oxygen which is present at temperatures of 1500F (815C) and greater. SO_3 is therefore formed directly from SO_2 as long as there is sufficient atomic oxygen available. SO_3 formation is increased by catalytic effect of metal oxides on the boiler tube surfaces. The percent of SO_2 that is oxidized to SO_3 in the boiler furnace and convection pass is defined as the boiler SO_2 to SO_3 conversion rate. Furnace geometry (surface area), oxygen concentration (excess air), fuel vanadium levels, fuel sulfur levels and ash composition all have an effect on the boiler conversion rate. Metal oxides in the ash can increase SO_2 conversion and alkaline ash constituents can react with SO_3 to lower the measured conversion rate.

SCR effects on SO₃ formation

When SCRs are installed in the typical high dust gas stream at the boiler exit, this NO_x reduction technology

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contributes additional SO_3 to the flue gas resulting in greater SO_3 that will contribute to total emissions and requires further SO_3 reduction. The SCR catalyst consists primarily of titanium dioxide (TiO_2), vanadium pentoxide (V_2O_5) and tungsten trioxide (WO_3) or molybdenum oxide (MoO_3). Vanadium is one of the active catalyst constituents and is the primary constituent responsible for the conversion of SO_2 to SO_3 as the flue gas passes across the catalyst surface. The rate of SO_2 conversion will vary depending on the vanadium concentration; however, there is an offset on catalyst NO_x reduction capability that also must be considered when selecting the catalyst chemistry. Thus a reduction in the catalyst SO_2 conversion rate results in a requirement to increase catalyst volume.

Catalyst conversion rate is also dependent on flue gas temperature and boiler load. The greater the flue gas temperature the greater the conversion rate. A greater temperature than normal or typical will occur as the boiler convections pass heating surfaces season or become fouled with an ash layer. At lower loads and reduced gas velocities there is also an increase in conversion rate due to the greater exposure time to the catalyst surface although this increase is somewhat offset by the lower flue gas temperature. Typical SO_2 conversion rates across the catalyst can vary from less than 0.5% to greater than 2% in a reactor with multiple catalyst layers and with the ammonia injection system in service. For those units that do not have an SCR bypass and utilize the SCR only during the summer ozone season, there will be an increase in SO_2 oxidation when the ammonia is not present. In summary, the SCR SO_2 conversion rate can vary considerably depending on catalyst chemistry and operating conditions. The total SO_3 concentration in the flue gas is dependent on these factors as well as the fuel sulfur content. Fuel sulfur content becomes the dominant factor when evaluating the SO_3 effects on particulate emissions and the necessary mitigation requirements.

Plant arrangements and mitigation technologies

Other than the boiler and SCR which generate SO_3, the typical power plant equipment in the downstream flue gas stream can affect the SO_3 content in the flue gas and the resulting particulate emissions at the stack. Refer to Figure 1 for the most typical power plant equipment arrangements.

Air heaters The air heater is typically the first piece of equipment the flue gas passes after the SCR and is common to all utility power plants. While the air heater exit gas temperature will be above the acid dew point, the surfaces within the air heater will expose the flue gas to temperatures below the acid dew point. This results in the collection or capture of a portion of the SO_3 present in the entering flue gas. This reduction in SO_3 will vary from almost no reduction to as great as 50% depending on many variables including operating gas temperatures, fuel type, ash characteristics and cold end metal temperatures.
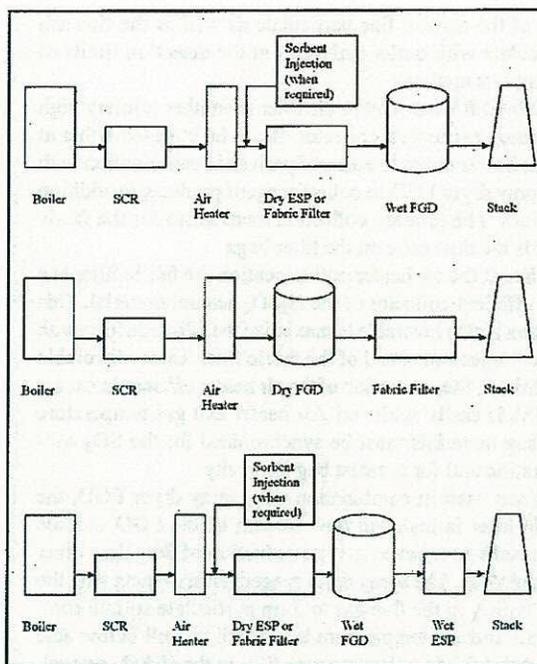

Fig. 1 Typical power plant arrangements.

Electrostatic precipitators The dry electrostatic precipitator (ESP) remains the most common high efficiency particulate collection device in use today. When installed after the air heater the ESP depends on the presence of SO_3 on the surface of the ash particles for optimum operation and particulate removal. SO_3 conditioning of the ash is utilized on some units burning lower sulfur coals to reduce the ash resistivity to optimize particulate collection. ESPs installed ahead of the air heater (hot side ESPs) are in a gas temperature zone where the SO_3 is all vapor and does not attach to the ash particles; therefore the hot side ESP depends on bulk ash resistivity for effective particulate collection. Thus, ESPs installed after the air heater result in some SO_3 reduction in the flue gas. This reduction may vary from 20% to 80% and is enhanced to the upper end of the range by the injection of sorbent materials ahead of the ESP, being careful to not upset the basic ash resistivity in a negative manner and cause a reduction in particulate collection.

Wet electrostatic precipitators (Wet ESPs) are effective collectors of both ash particles and acid mists such as H_2SO_4. The Wet ESP operating principles are similar to the dry ESP with a couple of major differences. First, the Wet ESP operates at or near saturated flue gas conditions so that moisture plays an important role in the Wet ESP operation. At these conditions all of the SO_3 is present as H_2SO_4 in aerosol form for ease of collection. The second major difference is the cleaning of internal surfaces where the Wet ESP utilizes water sprays instead of rapping systems. This difference is critical in the elimination of particulate re-entrainment by the Wet ESP. Wet ESPs can be designed to collect over

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90% of the aerosol fine particulate as well as the fine ash particulate with outlet emissions at the detection limits of current test methods.

Fabric filters The fabric filter is another primary high efficiency particulate collector. It can be installed either at the air heater outlet to collect flyash or in combination with the spray dryer FGD to collect reagent products in addition to flyash. The primary collection mechanism for the fabric filter is the dust cake on the filter bags.

Thus at the air heater outlet location the fabric filter is a very efficient collector of the H_2SO_4 aerosol material. This location is also favorable to maximize the SO_3 reduction with sorbent injection ahead of the fabric filter. Under favorable conditions, SO_3 reduction of the air heater effluent in excess of 90% is easily achieved. Air heater exit gas temperature and bag materials must be synchronized for the SO_3 concentration and for greatest bag longevity.

When used in combination with spray dryer FGD, the fabric filter is installed downstream of the FGD module and results in an even greater reduction of SO_3 than other technologies. The spray dryer reagent slurry reacts with the SO_3/H_2SO_4 in the flue gas to form particulate sulfate compounds and the temperature is reduced to well below acid dew point driving all remaining SO_3 to the H_2SO_4 aerosol. The highly efficient filter cake drives the fabric filter effluent H_2SO_4 concentration to detection limits. This combination of technologies is most applicable to the lower sulfur fuels.

Wet FGD The primary purpose of the Wet Flue Gas Desulfurization (Wet FGD) system is the capture of SO_2. The Wet FGD will collect some particulate as the flue gas passes through. The intimate contact between a flue gas stream and the scrubber liquid will transfer some particulate in the flue gas to the liquid in removal from the gas stream. Particulate collection efficiency in the Wet FGD is dependent on the particle size with the finer particulate being more difficult to remove. Therefore, since the H_2SO_4 aerosol is present in the flue gas only as a fine particulate, the Wet FGD has a limited ability to be an efficient collector of this material. The measured and reported capture of H_2SO_4 across the Wet FGD varies from 30% to greater than 50% with limited knowledge of how to maximize the removal.

Dry FGD The spray dryer FGD (Dry FGD) technology injects lime slurry into the flue gas stream downstream of the air heater for environmental control of SO_x. The slurry brings the flue gas temperature down to a temperature that is approximately 25F (14C) above saturation and allows the lime to react with SO_2 and SO_3 forming particulate calcium compounds. Dry FGD systems are installed in combination with a high efficiency particulate collector which is typically a fabric filter for greatest SO_x reduction. In the case of high alkaline coal ash, the particulate collected by the fabric filter is recycled for injection with the lime slurry for greater utilization of the ash and lime alkalinity. Due to the reduced temperature operating conditions and the slurry alkalinity, this combination of technologies is very effective in the capture of H_2SO_4, resulting in stack H_2SO_4 emissions of 1 ppm or less when burning a low sulfur PRB coal.

Injection sorbents As referenced in both the dry ESP and fabric filter sections above, injection sorbents can be used to enhance the capture of SO_3/H_2SO_4. Most injection sorbents are injected dry, although not exclusively. With this technology the dry sorbent, frequently alkaline in nature, reacts with the SO_3/H_2SO_4 vapor for collection as a particulate. Popular sorbents for SO_3 capture include magnesium hydroxide, hydrated lime, high surface area lime, Trona, sodium bi-sulfite, ammonia and other proprietary products. Considerations on sorbent usage and injection include suitable gas temperature for optimum reaction, effects on ash resistivity and ESP performance, available residence time for drying and reaction, affects on ash sales and disposal, air heater fouling, and reagent storage and handling issues.

This technology is suited for existing units that may have a tendency for a visible or blue plume, particularly after a SCR system is added and the level of SO_3 in the flue gas is increased due to the catalyst oxidation of SO_2. This technology can also be used on a new unit to achieve the low total particulate requirements at the stack which includes contribution from H_2SO_4. It may be logical to install a sorbent injection system and avoid the use of a Wet ESP depending on fuel sulfur content. Sorbents for SO_3 may be required to enable the use of injection technologies for mercury capture. An example of sorbent technology operating data for the Gavin plant with Trona injection is given in Table 1. The Gavin plant burns high sulfur coals and demonstrates the high capture rates of SO_3 by dry sorbent technologies. These capture rates can be used in the determination of total particulate emissions.

Investigations and studies

One purpose of this paper is to assemble a sampling of the demonstrated and available technologies which have achieved low particulate stack emissions including the effects of SO_3/H_2SO_4. The following provides stack emission results for total particulate and includes the filterable and condensable portions on a range of applications.

Table 1 SO₃ Reduction Across ESP with Trona Injection at AEP Gavin Unit 2				
	Injection Rate Moles Trona : Moles SO_3			
SCR Out of Service	0	2.0	3.0	3.5
SO_3 Reduction, %	20	67	80	84
	Injection Rate Moles Trona : Moles SO_3			
SCR In Service	0.5	1.0	1.5	2.0
SO_3 Reduction, %	66	74	76	78

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Low sulfur coals

Hawthorn Kansas City Power & Light's Hawthorn Unit 5 is a nominal 550 MW PC fired boiler burning PRB coals. The equipment train at Hawthorn includes the opposed wall fired boiler, the two reactor SCR system, the air heaters, the two module dry FGD system, the ten compartment pulse jet fabric filter and the ID fans. Total particulate emission test results for two different fabric filter bag materials are provided in Table 2. This data indicates that the total particulate emission is consistently at a very low emission value and that the condensable portion typically makes up less than half of the total with some dependability on type of fabric filter bag. The available data indicates the inorganic portion of the condensable catch ranges from 25% to 85%.

600 MW Unit The flue gas train for this operating project includes the boiler, SCR, air heater, fabric filter and Wet FGD. This unit burns PRB coals. Particulate test data results for this unit are provided in Table 3. While not achieving as low total particulate emissions, this data also demonstrates the small condensable portion but consisting of a high portion of inorganic material

550 MW Unit This operating PRB coal fired project has a flue gas train that includes the boiler, air heater, fabric filter and Wet FGD. Particulate test data results for this unit are provided in Table 4. This data indicates a low total particulate emission level with the condensables about equal to the filterable emissions and with the condensables consisting of about 50% inorganic material.

Medium and high sulfur coals

500 MW Unit The flue gas train for this operating unit includes the boiler, SCR, air heater, dry ESP and Wet FGD. The boiler burns a 3.3% sulfur coal. This data, Table 5, represents a low total particulate emissions with over half

consisting of the condensable portion. Further lab testing indicated the condensable particulate consisted mostly of sulfates in the inorganic portion.

Petroleum coke/heavy oil

Deepwater AES Deepwater is a petroleum coke fired cogeneration plant located in Pasadena, Texas on the Houston Ship Channel in a non-attainment area. This plant is designed to burn petroleum coke with a sulfur value up to 6% content. To limit the emissions of flyash, SO_2 and sulfuric acid mist, the plant chose a three-stage flue gas cleaning system consisting of 1) dry electrostatic precipitator for particulate capture, 2) limestone sorbent-based wet flue gas desulfurization (FGD) system, and 3) wet electrostatic precipitator for removal of sulfuric acid mist and residual particulate fines.

The Deepwater unit has been in commercial operation since 1986 and added an SCR system in 2007. Upgrades to the Wet ESP modules were initiated in 1999 and 2005 and modifications to the collection surface and discharge electrodes have been completed in six of the twelve parallel flow modules. These modifications have resulted in improved performance compared to the original Wet ESP system performance (without SCR) which had successfully met the design stack particulate emissions of 0.005 grains/

Table 3
Low Sulfur PRB Coal Application A
Particulate Emissions

Test ID	Particulate Emissions – lb/million Btu			
	Total	Filterable	Condensable	Inorganic/Organic %
1	0.0406	0.0351	0.0055	91.8/8.2
2	0.0480	0.0387	0.0099	93.0/7.0
3	0.0124	0.00389	0.0085	94.1/5.9
Avg. Results	0.0337	0.0259	0.0080	93.0/7.0
Test Method		Method 5B	Method 202	

Table 2
Hawthorn Particulate Emissions on PRB Coal

Test	Unit Load	Particulate Emissions – gr/acf			Stack Opacity
	MW	Total	Filterable	Condensable	%
Needle Felt Bags					
Nov. 2001	585	0.00638	0.00443	0.00195	NA
Aug. 2002	NA	0.00426	0.00345	0.00082	NA
Apr. 2003	515	0.00438	0.00299	0.00138	1.25
May 2004	545	0.00555	0.00346	0.00210	1.16
Membrane Bags					
May 2005	515	0.00545	0.00080	0.00465	2.10
Apr. 2006	530	0.00433	0.00093	0.00341	1.54
Test Method			Method 5B	Method 202	

Table 4
Low Sulfur PRB Coal Application B
Particulate Emissions

Test ID	Particulate Emissions – lb/million Btu			
	Total	Filterable	Condensable	Inorganic/Organic %
1	0.0019	0.00076	0.0012	42.4/57.6
2	0.0019	0.00113	0.0008	57.7/42.3
3	0.0010	0.00069	NA	0/100
Avg. Results	0.0016	0.00086	0.0010	50.0/50.0
Test Method		Method 5B	Method 202	

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Table 5
High Sulfur Coal Application C
Particulate Emissions

Test ID	Particulate Emissions – lb/million Btu			
	Total	Filterable	Condensable	Inorganic/Organic %
1	0.0084	0.00191	0.0065	89.5/10.5
2	0.0073	0.00242	0.0048	77.4/22.6
3	0.0123	0.00243	0.0099	92.1/7.9
Avg. Results	0.0093	0.00225	0.0070	86.3/13.7
Test Method		Method 5B	Method 202	

scfd for the combined filterable flyash and sulfuric acid mist concentrations

Recent emissions data is presented in Table 6. The condensable particulate is made up of virtually all sulfate as indicated in the inorganic portion.

Colson Cove In 2004 New Brunswick Power placed high efficiency Wet ESPs into service following two new limestone-based Wet FGD scrubbers at their 1050 MW Coleson Cove station. This was part of a plant wide effort to reduce the cost of electricity generation by switching to lower cost fuels and to upgrade emission controls to reduce SO_x and particulate. New Brunswick Power's decision to install Wet ESPs was to assure control of sulfuric acid emissions below 5 ppmvd @ 3% O_2 and limit flyash particulates below 0.015 lb/million Btu. To achieve this level of control of sulfuric acid at all times, design collection efficiency exceeded 90%.

The Wet ESP consists of a three field upflow design constructed in an upward extension of each of the two Wet FGD modules. Thus the Wet ESPs were installed within the circular absorber module. Scrubbed gas enters the close coupled inlet field of the Wet ESP after exiting the Wet FGD mist eliminator. The gas exits the top of the third field of the Wet ESP through a final mist eliminator section which captures any re-entrained droplets that may be present during flushing cycles. This design simplifies the arrangement. Materials of construction for the inlet field of the Wet ESP collecting plates are C-276. Stainless steel made of 6% Mo was utilized for the high voltage system and the balance of the collection plates.

This application again demonstrates the successful installation and operation of Wet ESP technology in the utility atmosphere to control the total particulate emissions from a high sulfur fuel.

Impact on plant equipment

Aside from stack opacity/particulate issues, SO_3 can have a major impact on the operation of the plant equipment. These impacts include such items as an increase in the acid dew point temperature which can limit heat recovery from the air heater and cause backend corrosion. The SO_3 can react with ammonia present from the SCR or SNCR system resulting in potential fouling of the air heater and catalyst. SO_3 also impacts the activated carbon injection system's ability to reduce mercury when the SO_3 competes with mercury for the active adsorption sites on carbon particles. Thus, removal of SO_3 can produce significant benefits for utility boilers.

The acid dew point temperature is a function of the SO_3 concentration and moisture content of the flue gas. The acid dew point temperature is the temperature at which acid will condense at or below this temperature. When the SCR is placed in service, the acid dew point temperature can substantially increase due to the increase in SO_3 concentrations leaving the SCR. If the air heater outlet temperature is not selected appropriately for new units or is not analyzed properly for SCR retrofits, there is an increased risk of acid corrosion on the downstream equipment. However, when the air heater outlet temperature is increased, the heat rate of the boiler is negatively affected.

Air heater fouling is an issue when acid condensation occurs. The ash combines with the sulfuric acid under certain conditions and forms a moist and sticky deposit. SCRs increase the SO_3 concentration. However, even if the temperature leaving the air heater is increased to maintain a similar margin as existed prior to SCR installation, some units will experience increased air heater fouling due to a change in the cohesiveness of the material.

SO_3 and NH_3 react under certain conditions and locations to form ammonia salt deposits. If SO_3 is present in larger molar concentration than that of the molar concentration of NH_3, Ammonium Bisulfate (ABS) will form. ABS is a sticky substance which can foul the air heater which results in increased pressure drop through the air heater, and over time the air heater will require an outage for cleaning. Air heater surface profile design, basket depth, number of basket sections and cleaning devices can be utilized to minimize ABS effects.

SO_3 has an impact on mercury absorption through the ash or via activated carbon. SO_3 competes with mercury for the active adsorption sites on the flyash/carbon. This reduces the

Table 6
AES Deepwater Petcoke
Particulate Emissions

Test ID	Particulate Emissions – gr/dscf			
	Total	Filterable	Condensable	Inorganic/Organic %
1	0.0078	0.0016	0.0062	81/19
2	0.0057	0.0005	0.0052	100/0 (as reported)
3	0.0047	0.0004	0.0043	100/0 (as reported)
Avg. Results	0.0061	0.0008	0.0052	
Test Method		Method 5B	Method 202	

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capability of the ash to capture mercury The same effect is true when using activated carbon to capture mercury Thus, the presence of SO_3 can have a significant impact on the amount of carbon required to remove mercury and the selection of the appropriate mercury mitigation technology.

Economic considerations

When a plant is considering the use of SO_3 mitigation technologies, several areas of the boiler island and air quality control system need to be investigated to optimize the SO_3 mitigation philosophy employed. For example, most users plan to use higher sulfur coals after the retrofit of a Wet FGD system. This can impact boiler conversion of SO_2 to SO_3 and economizer outlet temperature. Both of these issues could substantially affect stack H_2SO_4 emissions by increasing values. Air heater operation (temperature and operating conditions) has a large effect on the amount of SO_3 exiting the air heater and even the Dry ESP. Operating conditions and equipment sizing are important considerations in SO_3 mitigation. Wet FGDs do collect some H_2SO_4 but not to the levels that provide optimal stack condition on higher sulfur coals. Selection of a low SO_2 to SO_3 conversion catalyst could reduce both the cost of mitigation equipment as well as sorbent costs.

Several sorbents are effective in removing H_2SO_4 from the flue gas stream but careful consideration is required on location of injection, cost and reliability of the handling and injection system, and sorbent costs. In some cases, use of alternate sorbents should be considered in design of the sorbent handling system allowing flexibility to switch to a different sorbent if supply interruptions occur and/or price increases occur. In summary, the approach to SO_3 mitigation based on plant-specific conditions may reduce the overall cost of mitigation equipment.

Conclusions

SO_3 emissions play an important role in total particulate emissions and contribute to stack opacity and blue plume issues. SO_3 formation is based on the combustion of sulfur laden fuels and is enhanced by ash constituents on tube surfaces and by the addition of SCR catalyst. Multiple pieces of typical power plant equipment downstream of the boiler contribute to the capture or reduction of SO_3. There is operating experience with stack test results demonstrating that low SO_3/H_2SO_4 emissions are possible. However, the ultimate allowable SO_3/H_2SO_4 emission is based on the allowed total particulate emission and the portion that is allowed from SO_3/H_2SO_4. Based on the findings of this paper, this portion is dependent on the fuel sulfur content and ash constituents as well as the selection of back end equipment and any sorbent injection needs. For the determination of the selected SO_3/H_2SO_4 technology to achieve the final allowable emission, some economic evaluation of the optimum technology (for instance sorbent injection vs. Wet ESP) will be required for the higher sulfur fuels.

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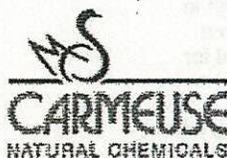
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Our customer service department answers many questions regarding our lime products for all the industries we serve. The questions range from lime applications and use to storage, shipping and handling. As a resource to our customers and website viewers, the listing below represents the most frequently asked questions.

- Are dry FGD systems only suited for smaller industrial applications?
- What is the difference between wet and dry lime-based FGD systems?
- I require FGD, but I have no space on my site for disposal of FGD by-products. Are there lime-based FGD systems that produce by-products which I can sell or provide for beneficial uses?
- What references are available that describe lime-based and other options for flue gas desulfurization (FGD) and compare costs of leading systems?
- I have other pollutants in my flue gas that I need to remove in addition to sulfur dioxide. Will a lime-based FGD system remove other pollutants?

Are dry FGD systems only suited for smaller industrial applications?
Dry lime-based FGD systems are suitable for both small industrial applications and large applications including coal and oil-fired power generating systems.
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What is the difference between wet and dry lime-based FGD systems?
Wet and dry systems are distinguished mainly by the type of by-product produced immediately after capture of sulfur dioxide. In dry lime FGD, the product of reaction of lime with sulfur dioxide is a dry powder. In wet lime FGD, the reaction product is a liquid. In dry FGD, milk of lime is sprayed directly into hot flue gas containing sulfur dioxide. Sulfur dioxide reacts directly with lime, and the heat of the flue gas evaporates water in the milk of lime, leaving a dry by-product. The by-product is removed from the gas using either fabric filters in a so-called baghouse or in an electrostatic precipitator. In wet lime FGD, sulfur dioxide is removed from flue gas using a two-step chemical process, compared with dry FGD which is a one-step process flue gas containing sulfur dioxide flows into the side of and up through a cylindrical vessel called an absorber (see figure) A liquid mixture of water, products of previous reaction of sulfur dioxide and lime, including an alkaline salt (magnesium sulfite), is sprayed into the gas from above. Sulfur dioxide is grabbed by the alkaline salt and is absorbed into the liquid mixture (step 1). The liquid mixture containing sulfur dioxide falls into the bottom of the absorber, where a pool of the liquid mixture is held. The sulfur dioxide, which is acidic, lowers the pH of the liquid pool. Milk of lime is then added to the pool to raise pH to about 6. Calcium hydroxide in the lime reacts with most of the sulfur dioxide to form a solid by-product, and magnesium hydroxide reacts with the remainder of the sulfur dioxide to replenish the alkaline salt (step 2). The replenished liquid mixture is then available to be sprayed again into the top of the absorber tower to remove sulfur dioxide from additional flue gas.
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I require FGD, but I have no space on my site for disposal of FGD by-products. Are there lime-based FGD systems that produce by-products which I can sell or provide for beneficial uses?
Yes. Lime-based systems can produce by-products both for sale and for a number of beneficial uses. Wet lime systems produce gypsum (calcium sulfate), a

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beachsand-like product which is purchased by building products manufacturers for production of plasterboard. Gypsum is also used by cement manufacturers as a set retarder. In some areas of the US, agricultural soils are deficient in sulfur or high in alkali, and gypsum is used as a soil amendment to increase crop yields. Dry lime FGD systems produce dry by-products rich in alkali which have also been used for soil amendment. Additional beneficial uses for lime FGD by-products include structural fill, mining mortars, and production of lightweight aggregate.

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What references are available that describe lime-based and other options for flue gas desulfurization (FGD) and compare costs of leading systems?
A good general reference produced by US EPA, Controlling SO2 Emissions: A Review of Technologies, is available through the National Technical Information Service ([EPA/600/R-00/093, Order No. PB2001-101224; Telephone: 703-605-6000; 800-553-6847 U.S. only]. Leading FGD systems for large industrial or power generating plants are wet lime FGD, dry lime FGD, and wet limestone FGD. Lime-based systems have lower equipment costs than limestone-based systems. Wet lime systems are suited for flue gases with all ranges of sulfur dioxide content up to 10000 parts per million by volume of sulfur dioxide and where up to 99% sulfur removal is required. Dry lime systems have the lowest equipment costs and are suited for flue gases with less than 1500 parts per million by volume of sulfur dioxide and where up to 94% sulfur removal is required.
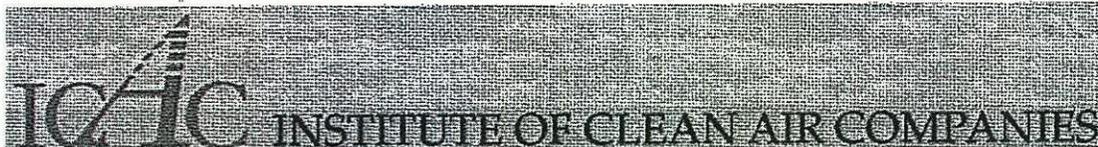
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I have other pollutants in my flue gas that I need to remove in addition to sulfur dioxide. Will a lime-based FGD system remove other pollutants?
Lime will remove additional acid gases such as hydrogen chloride (HCl), hydrogen fluoride (HF) and sulfur trioxide (SO_3). Wet lime and dry lime systems achieve over 95% removal of HCl and HF, and dry lime systems also achieve over 95% removal of SO3. Sulfur trioxide in combination with moisture produces acid mist, which produces a dark brown or bluish plume and greatly increases visual opacity of flue gas. Mercury is also efficiently removed by lime-based systems. Dry lime FGD in combination with activated carbon is a proven technology for removal of more than 70% of mercury from flue gases produced by incineration and other industrial facilities. Wet lime scrubbers efficiently remove water-soluble forms of mercury.

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Acid Gas/SO2 Controls - Institute of Clean Air Companies

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Wet Scrubbers

Scrubber is a general term that describes air pollution control devices or systems that use absorption, both physical and chemical, to remove pollutants from the process gas stream. Scrubber systems rely on a chemical reaction with a sorbent to remove a wide range of pollutants, including sulfur dioxide (SO2), acid gases, and air toxics, from flue gases. When used to remove or "scrub" SO2 from the flue gas, these devices are commonly called flue gas desulfurization (FGD) or scrubber systems. FGD or scrubber systems are generally classified as either "wet" or "dry" Wet scrubbers are increasingly recognized as an important part of a multi-pollutant control program.

In a wet scrubber, a liquid sorbent is sprayed into the flue gas in an absorber vessel. The gas phase or particulate pollutant comes into direct contact with a sorbent liquid and is dissolved or diffused (scrubbed) into the liquid. The liquid interface for gas and particle absorption include liquid sheets, wetted walls, bubbles and droplets. In the wet processes, a wet slurry waste or by-product is produced. Most wet FGD systems use alkaline slurries of limestone or slaked lime as sorbents. Sulfur oxides react with the sorbent to form calcium sulfite and calcium sulfate. Uptake of the pollutant by the sorbent results in the formation of a wet solid by-product that may require additional treatment, or when oxidized, results in a gypsum by-product that can be sold.

Scrubber technologies for wet scrubbing of gaseous pollutants can achieve extremely high levels of multi-pollutant control, including acid gases, SO2, fine particulates and heavy metals (i.e., mercury) from utility and industrial coal-fired boilers, waste-to-energy systems, and other industrial processes. Wet scrubber technology can be applied to difficult processes such as gas absorption and particle collection, treating combustible particles, and removal of wet, sticky or corrosive particles. Wet scrubbers are used in industrial process mercury removal and to remove ionic forms of mercury from the gas stream of coal-fired power generation facilities

Wet scrubbers generally have relatively small space requirements, low capital cost, and are able to process high temperature, high acidity, and high humidity flue gas streams. Scrubber costs have continued to decrease, largely because of technical innovations. Scrubber energy requirements have also continued to decrease, helping to lower operating costs. FGD systems are an increasingly significant part of a multi-pollutant control approach, even as the energy requirements of these systems are decreasing to where these systems now consume only about 1% of total boiler output. Where low-cost high-sulfur fuels are available, or where the required reductions are very high, scrubbing is often a viable control option.

New wet scrubbers routinely achieve SO2 removal efficiencies of 95%, with some scrubbers achieving removal efficiencies of up to 99%. Scrubbers have been used in the EPA Acid Rain Program on coal-fired boilers, which are significant sources of hydrochloric acid (HCl) and hydrofluoric acid (HF). According to the EPA and others, both wet and dry scrubbers have been shown to reduce HCl emissions by 95% and more, and wet scrubbers have been shown to reduce HF emissions by more than one-third. Others have reported ranges of 87-94% removal of chlorine and 43-97% removal of fluorine by both wet and dry scrubbers. In addition, wet scrubbers also provide significant removal of arsenic, beryllium, cadmium, chromium, lead, manganese, and

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mercury from flue gas.

Wet scrubbers can be generally grouped by geometric designs and method for gas-liquid contact. Several groupings include packed-bed, counter-flow, cross-flow, bubble-plate, open spray (single and double loop) tower, dual-flow tray, cyclonic, and venturi designs. However, there are many proprietary systems designed around specific industry needs. Design and operating parameters include scrubber geometrical shape, liquid spray or injection locations, gas residence time, gas velocities, gas and liquid temperatures, gas and liquid pressure drop, and, liquid/gas flow rate ratio.

Dry Scrubbers

In a dry scrubber or FGD process, particles of an alkaline sorbent are injected into a flue gas, producing a dry solid by-product. In dry FGD scrubbing, the flue gas leaving the absorber is not saturated (the major distinction between wet and dry scrubbers). Dry scrubbers generally involve simple designs and low capital and maintenance costs. Scrubber costs have continued to decrease, largely because of technical innovations. However, dry scrubbers are increasingly being recognized as an important part of a multi-pollutant control program. Dry scrubber energy requirements, while less than wet FGD processes, continue to decrease which helps to lower operating costs.

Generally, the fate of absorbent in dry FGD is characterized as either once-through or regenerable. Dry scrubbers systems can be grouped into three categories: spray dryers, circulating spray dryers, and dry injection systems. All these systems offer multi-pollutant control opportunities by combining acid gas, SO2, particulate control, and air toxics, including mercury.

In a spray dryer, a slurry of alkaline reagent, typically lime or sodium based, is atomized into the hot flue gas to absorb the pollutants. The resulting dry material, including fly ash, is collected in a downstream particulate control device, typically an electrostatic precipitator or fabric filter. In some cases, a portion of the dry material is recycled into the lime slurry mixture. Oxidation of the mixture results in a gypsum by-product that can be sold. A lime spray dryer can typically be installed on industrial and utility boilers, and on municipal and hazardous waste incinerators. Spray dryers commonly are designed for SO2 removal efficiencies of 70-95%.

A circulating dry scrubber uses an entrained fluidized bed reactor for contacting the reagent, usually hydrated lime, with sulfur dioxide and particulate laden flue gas. The intensive gas-solid mixing that occurs in the reactor promotes the reaction of sulfur oxides in the flue gas with the dry lime particles. The mixture of reaction products (calcium sulfite/sulfate), unreacted lime, and fly ash is carried to a downstream particulate collector that is separated from the gas stream. Part of the dry waste product is removed for disposal, but most of the waste product is mixed with fresh calcium hydroxide for use in the reactor. Water spray is introduced into the fluidized bed separately to enhance performance (for maximum SO2 capture with minimum lime utilization) by optimizing the surface moisture content of the lime. Circulating dry scrubbers can provide removal efficiencies of more than 90%.

Dry injection systems involve the injection of a dry sorbent (normally lime or limestone) into the flue gas in the upper reaches of the boiler, or in the ductwork following the boiler. Sulfur oxides react directly with the dry sorbent, which are collected in a downstream particulate control device. Because a separate scrubber vessel is not needed, capital costs are minimized. Low capital costs are partially offset by lower reagent utilization, which result in higher operating costs for equivalent SO2 removal rates. Dry injection systems are generally applied when lower removal efficiencies are required, or on small plants where the capital cost for other scrubber types may not be justified. Dry injection systems typically have removal efficiencies ranging from 50-70%.

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Designing wet duct/stack systems for coal-fired plants

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Designing wet duct/stack systems for coal-fired plants

As limits on SO_2 emissions become tighter, many coal-fired utility power plants are adding new flue gas desulfurization (FGD) systems or upgrading existing ones. The majority of these systems employ wet FGD technologies. Many plants with wet FGD systems use reheated flue gas to dry the saturated gases exiting the system's absorber before they are sent to the stack.

However, due to the increasingly high cost of energy and/or the need to decrease total plant SO_2 emissions, flue gas reheating typically is no longer used on new plants and is being eliminated at plants already in service. Without reheat, saturated flue gases exiting the absorber enter the stack directly. Along with droplets that are carried over from the mist eliminator, water vapor condensing from the saturated flue gases create a liquid film that covers the walls of the absorber outlet ducting and stack liner. This film must be collected and drained from the system by a process called wet stack operation. If the ductwork, stack liner, and liquid collection system are not properly designed, unacceptable amounts of liquid droplets may be discharged from the top of the stack.

Trouble-free operation of an FGD unit with wet ducts and stacks requires investigation of several potential problem areas related to the handling and discharge of wet flue gas. This must be done during the design of the stack and duct system. This article outlines the important design aspects that must be addressed if liquid discharge from a new or retrofitted wet stack installation is to be minimized.

Wetter is better . . . but

Although wet FGD systems have been used for decades, most operated with dry stacks. In the late 1970s, some utilities began using wet ducts and stacks to reduce their operating costs, but in many cases the (unforeseen) result of this switch was unacceptably high levels of stack liquid discharge (SLD). To improve understanding of the physical processes taking place in wet FGD systems, in the 1980s, EPRI sponsored a number of programs to determine the key variables contributing to liquid re-entrainment. The results of one of these studies were summarized in EPRI Report No. CS-2520, "Entrainment in Wet Stacks."

By the late 1990s many utility plants were operating with wet duct/stack systems—most equipped with liquid collectors—and a sufficient body of experience had been developed for EPRI to sponsor another program to develop practical guidelines for wet stack design and operation. The results of this study were summarized in EPRI Report No. TR-107099, "Wet Stacks Design Guide."

As the two reports detailed, the development of an effective wet stack installation requires a thorough fluid dynamic design of the wet duct/stack system and of the liquid collectors and drains. The design process, usually performed by a flow modeling laboratory with experience in this area, typically consists of five distinct phases:

- Phase 1—Initial review of the proposed system design
- Phase 2—Condensation calculations
- Phase 3—Design and development of the liquid collection system
- Phase 4—Study of the plume downwash
- Phase 5—Field installation and operational inspections

Using the results of the first four phases, utility engineers can finalize the design of the wet FGD system and write the specifications for supply and construction bids. The results of the fifth phase are used to support the installation of the liquid collection system.

Phase 1—Reviewing the initial design

Experience suggests that an early review of the proposed geometry of an absorber outlet duct and stack breach/inlet often produces recommended modifications for making the geometry more amenable to wet operation. This review should be performed by a consultant with broad experience in wet stack system design and operation. Key system design variables such as gas velocities, breach height and width, and liner diameter should be compared with values that have proven favorable for wet operation at other plants. Simple changes in system geometry, such as adjusting the breach aspect ratio, often can significantly improve the efficacy of the liquid collection system—by improving the flow patterns in the lower liner, minimizing the potential for liquid re-entrainment, reducing the total number of liquid collectors required, and/or reducing system complexity.

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Many economic factors play a role in the design of a stack liner, including minimizing its diameter. But the primary controlling parameter for effective wet stack operation is the liner gas velocity. Different liner materials and construction techniques have different velocities considered favorable for wet operation. According to the EPRI "Wet Stacks Design Guide " well-constructed C276 and fiberglass-reinforced plastic (FRP) liners should be operated at maximum velocities of 50 to 60 ft/sec, whereas brick liners—with rougher surfaces—should be operated at gas velocities no higher than 35 to 45 ft/sec. If properly installed, modern materials, such as borosilicate block liners, can operate effectively at velocities up to 65 ft/sec (Figure 1, p 22).
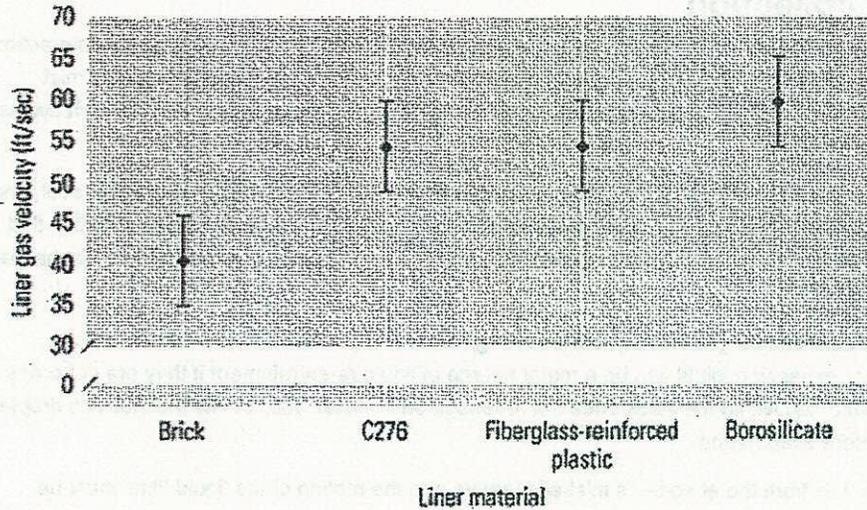
graph
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Source: Alden Research Laboratory

1. Faster is better. Maximum stack liner gas velocities for four materials.

Sometimes the maximum recommended liner velocity must be reduced if there are horizontal discontinuities in the liner such as weld beads or FRP joints. These discontinuities can act as sites for liquid re-entrainment. Special care must also be taken to design the stack entrance area, where the gas flow is three-dimensional and very nonuniform, with velocities two or more times higher than the area-average value. Expansion joint locations in this area are particularly critical and, if at all possible, should not be located within one to two liner diameters of the top of the breach.

Phase 2—Making condensation calculations

Much of the liquid that must be removed from a wet stack is condensation on the duct and stack liner walls. It results from thermal condensation on the walls (the result of heat transfer from the flue gas to the outside air through the liner, insulation, annulus air, and concrete shell) and from adiabatic condensation of the saturated flue gas as it flows up the stack and expands as pressure falls with elevation

The amount of thermal condensation on the duct and liner surfaces is a function of the stack's geometry and thermal conductivity, its internal flow conditions, ambient temperatures and wind velocity, and—in some instances—wind direction. Thermal insulation can reduce the amount of thermal condensation on a stack liner. In practice, the addition of 2 inches of liner insulation has proven capable of reducing the quantity of thermal condensation by a factor of four

As for adiabatic condensation, industry experience indicates that the relatively small pressure drop from the breaching duct to the top of the stack may produce an appreciable amount of liquid. A small fraction of this liquid will deposit on the liner surface due to turbulent diffusion, and the rest will be discharged from the liner as part of the bulk gas stream in the form of very small droplets.

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These droplets do not present a problem because they are so small that they will evaporate before reaching the ground

Customized heat-transfer computer programs are used to calculate the total amount of liquid deposited on the stack liner walls due to thermal and adiabatic condensation. Such applications also can determine the rate of condensation, which is a function of ambient air temperature, wind velocity, and liner design variables Both sets of calculations are needed to design and develop an effective liquid collection system, because the results will determine the number, size, and location of its gutters and drains

Phase 3—Modeling liquid collection

The problem of SLD from a duct and stack system operating in "all-scrubbed" mode is exacerbated by the two-phase flow interaction of liquid and gas. The liquid enters the ducts and stack as droplets as well as condensed water vapor carried over from the mist eliminators of the absorbers. Rates of liquid carryover can increase significantly over "as designed" rates during normal wash cycles or if mist eliminators become fouled with normal use

How much liquid and condensate is deposited is a function of flue gas velocity, duct and stack geometry, the liquid loading level, and the distribution of droplet size. The liquid films that form can move under the influence of gravity and gas shear forces Any film that does move either will accumulate elsewhere or be re-entrained from liner walls, internal struts, dampers, or vanes and be discharged from the top of the stack by the high-velocity gases

This re-entrainment process, in turn, is dependent on flue gas velocity, surface roughness, and surface discontinuities such as duct/liner weld seams or FRP liner joints Liner expansion joints can be a major source of liquid re-entrainment if they are not properly located and arranged Most of the liquid re-entrained within the stack does not redeposit on the liner wall; it exits the stack as droplets that are large enough to reach the ground before evaporating.

The behavior of droplets entrained in the gas flow from the absorber's mist eliminators and the motion of the liquid films must be evaluated by a physical flow model of the unit Such models facilitate the design and development of liquid collection devices capable of maximizing primary droplet deposition and liquid collection while minimizing the potential for re-entrainment of droplets from liquid pools and films. Computational fluid dynamics (CFD) models cannot be used for this purpose, because these computer codes, though effective at predicting droplet trajectories and droplet collection patterns, are currently incapable of accurately simulating the development and motion of liquid films on the duct and liner walls

Physical flow models of wet duct and stack installations are usually built to a scale of 1:12 to 1:16. They typically encompass the system from the outlet of the absorber mist eliminator to a point in the stack liner approximately three to four liner diameters above the roof of the breaching duct. Typical single and multiple absorber wet stack flow models are shown in Figures 2 and 3, respectively To the extent possible, models are fabricated from clear Plexiglss to allow detailed visualization of internal gas and liquid flows. To ensure that the liquid film motion in the primary collection zone of the lower liner is accurately simulated, care must be taken to ensure that the surface of the material used in this area has wetting properties similar to those found in the field
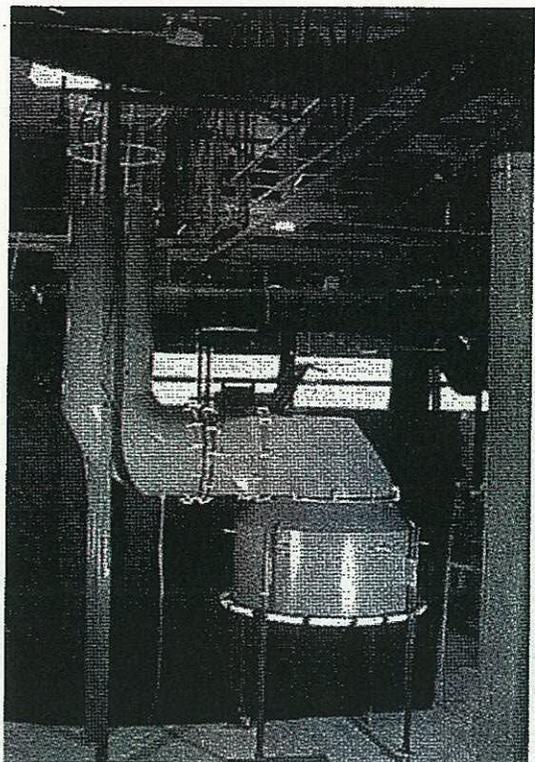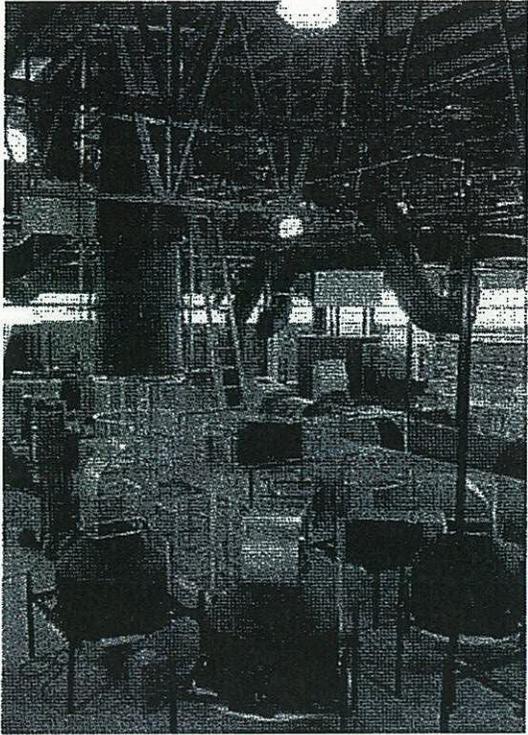
Courtesy: Alden Research Laboratory Inc

2 One is good. A typical single absorber wet stack flow model.

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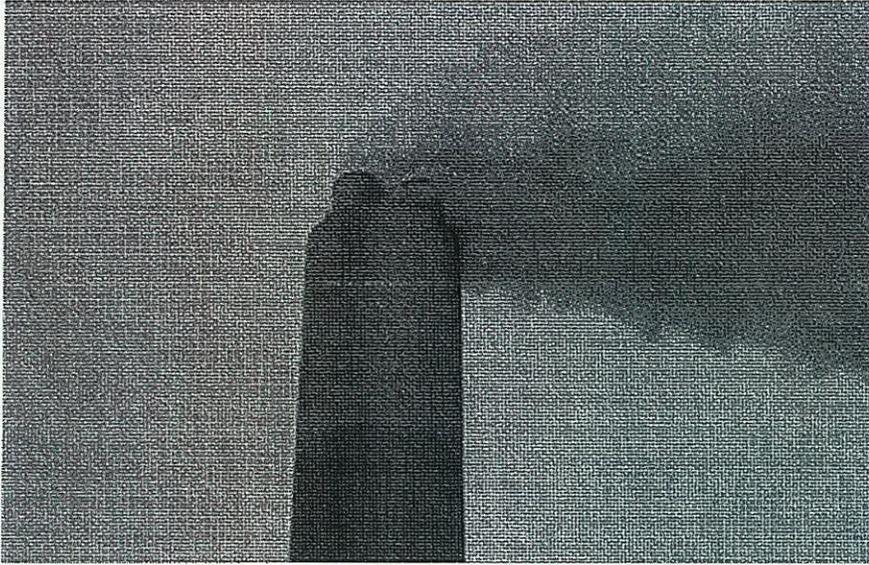Attachment 179--Redacted
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Courtesy: Alden Research Laboratory Inc.

3 More is better A typical multiple absorber wet stack flow model

Use of a physical flow model enables droplet trajectories and the movement of liquid films to be observed as flow patterns. These patterns, in turn, can be used to design, develop, and optimize a system of vanes, gutters, ring collectors, dams, baffles, and drains capable of working effectively over a range of expected boiler loads and operating scenarios. To the extent possible, the liquid collectors and gutters are fabricated from commercially available structural shapes made of noncorrosive materials such as C276 or FRP

Phase 4—Evaluating plume downwash

A crosswind at the top of a power plant stack will give the plume exiting it a horizontal component As the ratio of vertical plume momentum to horizontal wind momentum decreases, the plume may become partially entrained in the wake formed on the downwind side of the liner and stack shell At lower momentum ratios, the reduced static pressure in the wake can draw the flue gas into a downwash pattern along the downwind side of stack shell (Figure 4)

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Courtesy: Alden Research Laboratory Inc

4   Result of reduced static pressure. Plume downwash from the stack of a coal-fired power plant
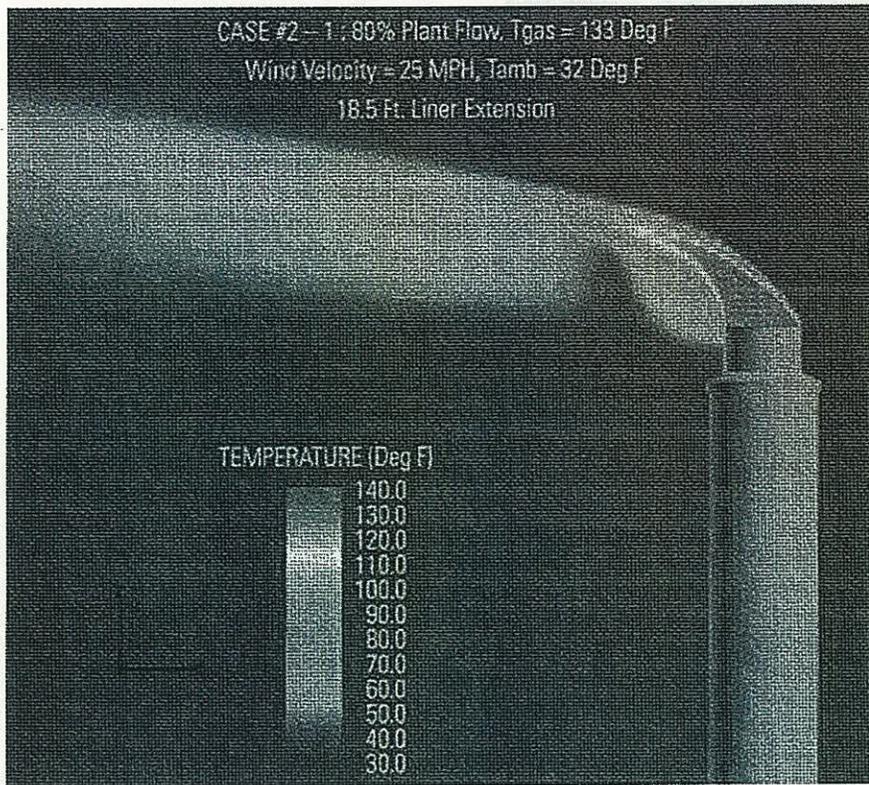
If the saturated flue gas drawn into this wake comes into contact with the roof and sides of the stack liners and the shell, problems involving metal corrosion, concrete deterioration, and ice buildup (during winter months) can arise. Plume downwash is particularly important to address for stacks with multiple, interacting discharge plumes. Severe downwash situations may even produce plume contact with lower surrounding plant structures. In the most extreme cases, plumes may touch down at ground level near the stack.

The interactions between the prevailing wind and the individual plumes also play a significant role in the propensity and extent of downwash for stacks with multiple flues. The wind direction relative to the plumes can produce different degrees of downwash. For a given ratio of vertical to horizontal momentum, plume downwash will be greater in a dual-liner stack if the direction of the prevailing wind is perpendicular to the axis of the two liners, compared with a condition in which the wind direction is parallel to that axis. Evaluating the downwash potential for two (or more) wind directions allows definition of the maximum downwash potential and development of appropriate liner extensions and stack top geometry to mitigate the problem.

Ultimately, all stacks can enter into downwash mode at low unit load and at high wind velocities. The only questions are:

- Under what conditions and how frequently will downwash occur, given the expected boiler load schedule and wind frequency profile?
- How much downwash is considered acceptable?

In contrast to simulations of liquid collection systems, CFD modeling is ideally suited for evaluating the variables of downwash, a phenomenon resulting almost entirely from gas-gas interactions. Figure 5 presents a typical output from such a model showing good stack operation without downwash. Various liner velocity/atmospheric conditions can be evaluated to determine the liner extension height or the exit choke size reduction required to eliminate downwash or the extent to which the top of the stack shell should be covered with an acid-resistant coating. The potential for stack top icing also can be evaluated by including heat transfer in the model and by predicting how quickly the plume will cool after exiting the stack.

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CASE #2 – 1 : 80% Plant Flow, Tgas = 133 Deg F
Wind Velocity = 25 MPH, Tamb = 32 Deg F
18.5 Ft. Liner Extension

TEMPERATURE (Deg F)
140.0
130.0
120.0
110.0
100.0
90.0
80.0
70.0
60.0
50.0
40.0
30.0

Courtesy: Alden Research Laboratory Inc

5. Very predictable Typical results of a computational fluid dynamics model of plume downwash.

Phase 5—Supporting installation and operation

The results of this phase are used to support field installation of the liquid collection system Field construction and installation drawings of the system should be reviewed by the engineers who designed and developed it, to ensure that their recommendations have been interpreted correctly and to evaluate any changes made by the installation company to facilitate fabrication assembly, installation, or structural support.

During field installation, deviations from the specified design are often required due to unanticipated interferences or installation issues To ensure that the liquid collection system has been installed properly, an on-site inspection of the installation by its designer is highly recommended and should be scheduled for a day when the installation is 80% to 90% complete This way, errors can be identified, on-the-spot modifications can be defined if necessary, and corrections can be made while the construction crew is on-site before the unit start-up.

Inspecting the liquid collection system after several months of operation is also highly recommended If any stack emission incidents occur during normal operation, the need for inspection is obvious However, even if SLD is not experienced, it still is important to inspect the liquid collectors as a way to ensure their satisfactory long-term operation These inspections may reveal incipient deposition and drainage problems that can then be corrected during future plant outages.
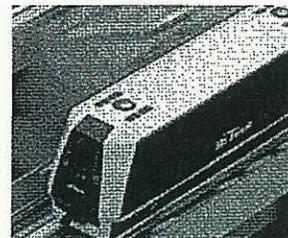
— David K. Anderson is director of Air/Gas Flow Modeling at Alden Research Laboratory Inc He can be reached at danderson@aldenlab com or 508-829-6000, ext 464 Lewis A Maroti is a senior consultant at Alden; he can be reached at

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lamaroti@aldenlab.com or 508-829-6000, ext 496

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MHIA Environmental Systems

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Also our patented post-combustion capture systems help industrial plants and power plants grab and retain up to 90% of their CO_2 emissions, helping place plant operations at the forefront of green technology utilization. It's how we help make the world a better place. We can help you do the same.

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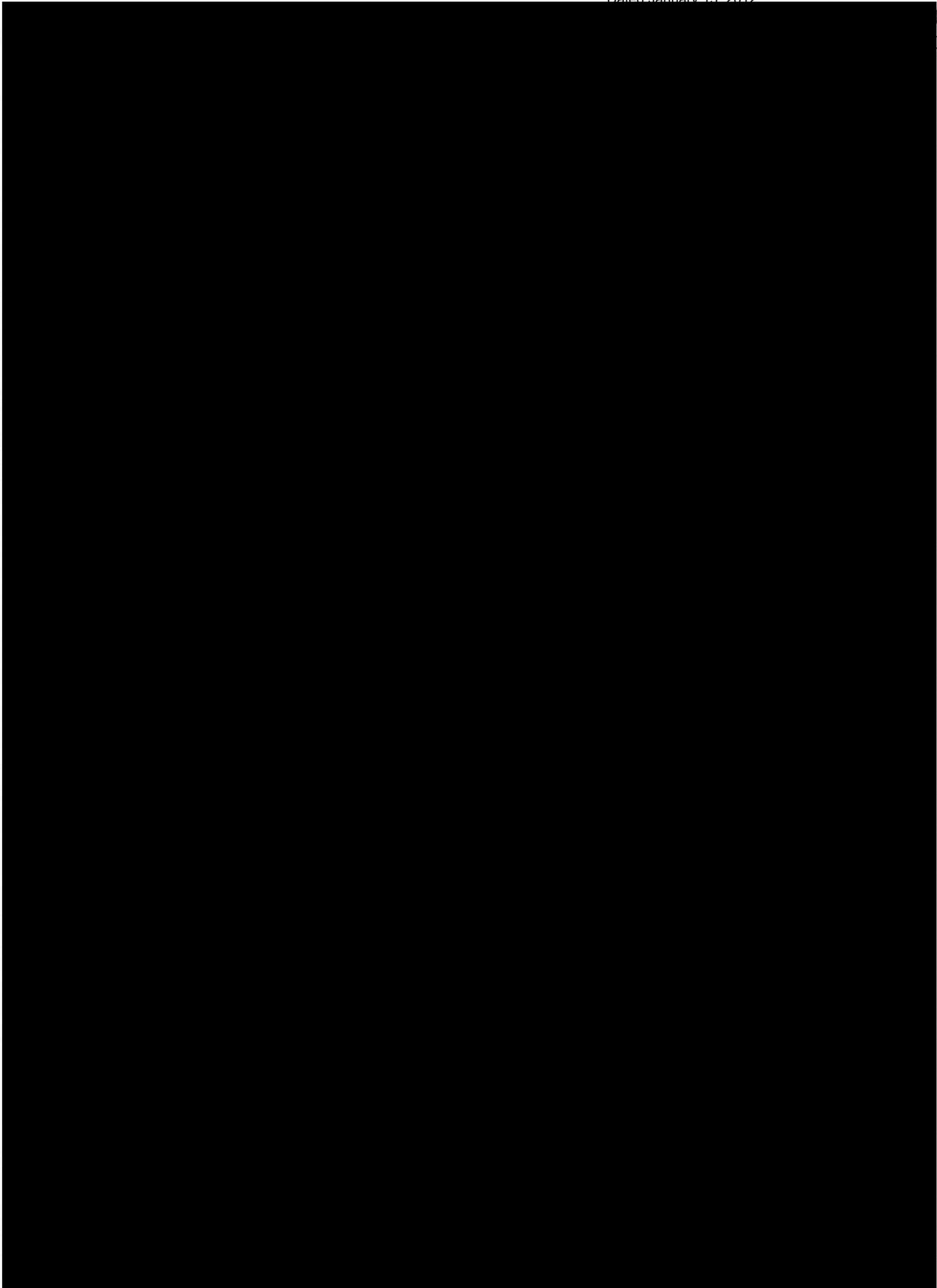Tips and Facts from Wakamaru

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