Table 3
Typical Pulverizer Outlet Temperature

Fuel Type	Volatile Content, %*	Exit Temperature F (C)**
Lignite and subbituminous	—	125 to 140 (52 to 60)
High volatile bituminous	30	150 (66)
Low volatile bituminous	14 to 22	150 to 180 (66 to 82)
Anthracite, coal waste	14	200 to 210 (93 to 99)
Petroleum coke	0 to 8	200 to 250 (93 to 121)

* Volatile content is on a dry, mineral-matter-free basis.
** The capacity of pulverizers is adversely affected with exit temperatures below 125F (52C) when grinding high moisture lignites.

Grindability is not strictly a matter of hardness. Some materials, fibrous in nature, are not hard but are very difficult to grind. Sticky or plastic materials can also defy grinding.

The procedure described in ASTM D 409 was originally developed as a purely empirical method but provides a valid capacity calculation for vertical pulverizers when grinding bituminous coal. When lower rank coals are tested in the Hardgrove apparatus, the correlation between laboratory tests and field operating equipment is often quite poor. There is a strong but unpredictable effect of sample moisture on test results. A more accurate index of grindability for low rank coals is determined in a laboratory sized MPS-32 pulverizer. This equipment has all the features of a field installation and can emulate field conditions. The capacity of the mill is about 800 lb/h (0.10 kg/s) and a coal capacity test requires a 1000 lb (453.6 kg) sample. Fig. 14 is an arrangement diagram of this test apparatus. Results are interpreted as apparent grindability and the values are valid for capacity calculations.

Wear properties Wear in coal pulverizers results from the combined effects of abrasion and erosion. The mechanism which dominates in the wear of a particular machine component depends upon the designed function of the component and on the properties of the coal being ground. Abrasiveness and erosiveness are the important properties to be considered when evaluating the influence of a candidate coal on expected maintenance cost. Unfortunately, these two important properties are inherently difficult to measure, especially with a material as variable as coal.

Abrasiveness One indication of the likely pulverizer wear part life is abrasiveness. It is primarily related to the quantity and size distribution of quartz and pyrite found in the as-received coal, especially for particles larger than 100 mesh (150 microns). While standard and accepted procedures are available to determine the quantity of these two minerals in a sample (ASTM D 2492 and ASTM D 2799 for pyrite and quartz respectively), accepted procedures to establish quartz and pyrite particle sizes are not available.

Unfortunately, direct small-scale laboratory abrasiveness testing and correlation to field wear conditions are not satisfactory because of operating differences between small laboratory equipment and power plant pulverizers. However, satisfactory abrasiveness measurements have been made using the MPS-32 laboratory pulverizer with 3000 lb (1362 kg) samples. Although expensive, wear life testing with this procedure can typically predict field performance with an error of 10% or less. While other abrasiveness tests, such as the Yancey-Geer Price apparatus, are used to provide some insight into relative wear rates, they are of very limited value in predicting actual field wear rates.

Erosiveness Erosion is not the same as abrasion. It can be defined as the progressive removal of material from a target on which a fluid borne stream of solids impinges. For a given combination of particles and target material, wear rate increases with velocity, with particle size, and

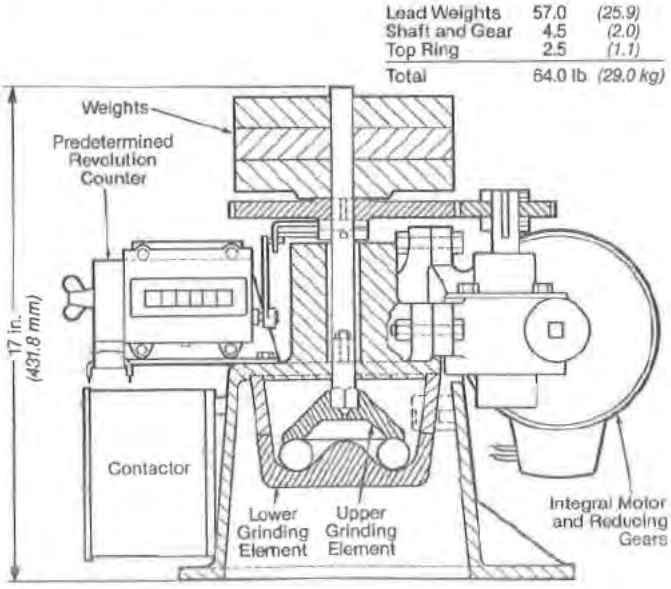


Fig. 13 Hardgrove grindability testing machine.

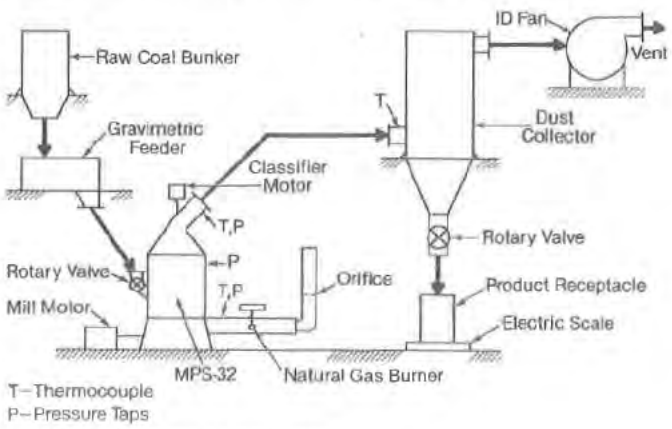


Fig. 14 B&W's MPS-32 test apparatus.

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with the solids content of the stream. For ductile target materials, the wear rate increases up to an impingement angle of approximately 35 to 45 deg, at which point wear begins to decline.[1] For brittle materials, the rate of erosion increases up to the angle of approximately 70 deg, where rebounding material interferes with the impinging stream and measurement becomes erratic.

Moisture Moisture content in the fuel is a key design parameter. This inherent or fuel bound moisture can strongly influence grindability for subbituminous and lignite coals (Fig. 15). In addition, the surface moisture plus inherent moisture strongly influence the amount of drying needed and therefore strongly influence pulverizer air inlet temperature. Coal moisture is highly variable and depends more on coal type than on the amount of water introduced after mining. Moisture may be inherent, that which is present in the geological deposit, or may be surface moisture which is water introduced during handling, transport, processing or storage, or may be an artificial value called equilibrium moisture. This is the stable value reached after thermally dried coal reabsorbs atmospheric moisture. Coals used in the U.S. range from inherent moisture levels of 2% for Appalachian bituminous to near 40% for lignites. Brown coals may range up to 70%. Moisture in coal is determined according to the procedures in ASTM D 3302.

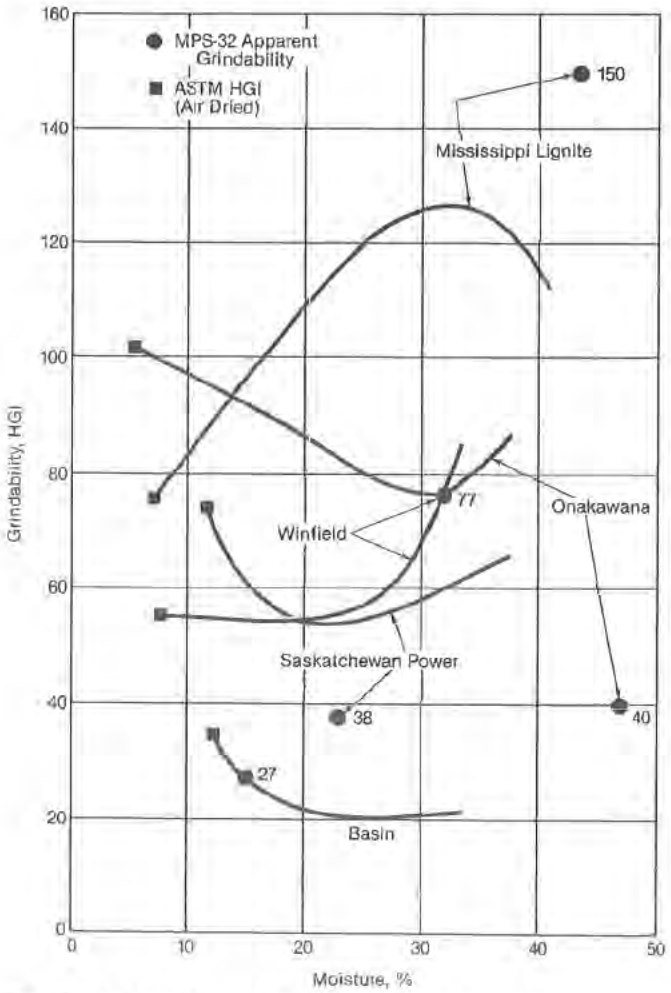


Fig. 15 Hardgrove Grindability Index versus fuel moisture.

Pulverizer size selection

The ultimate task in pulverizing application is the selection of the size and number of mills for the proposed project. The total boiler heat input and coal flow requirements are established from the combustion calculations (see Chapter 9) and the specified boiler steam flow requirements. The coal flow rate is then divided by the capacity correction factor to establish the equivalent required pulverizer capacity. The correction factor not only includes the fineness and grindability factors shown in Fig. 16, but also the appropriate fuel moisture correction. Most roller mills, including the MPS, require a moisture correction for capacity loss while the EL mills require no correction.

The number of pulverizers is then evaluated by dividing the equivalent required capacity by the unit pulverizer rated capacity based upon 70% passing through a 200 mesh (74 microns) screen and a grindability of 50. The unit pulverizer rated capacity is based upon pulverizer size, type and manufacturer. The tradeoff between fewer larger mills and more smaller mills is based upon balancing total capital cost with the proposed operating requirements such as boiler turndown. Frequently an extra pulverizer is specified to permit the boiler to operate at full load while one mill is out of service for maintenance.

Pulverizer selection and sizing are complicated by the frequent need to consider various coal sources and emission requirements. Computer programs can analyze pulverizer performance quickly, even for a large number of candidate coals, to learn which will govern mill selection. If the coal setting the pulverizer size is greatly different from the intended primary use coal, the result may be oversized mills and limits on turndown on the primary coal. In such a case, it may be necessary to reconsider whether this coal should remain on the list of possible fuels.

Performance testing

The most rigorous performance tests are done to assure compliance with the purchase contract. This acceptance testing will usually be guided by the procedures in The American Society of Mechanical Engineers (ASME) Performance Test Code, PTC 4.2 which references other applicable codes and standards. Testing may also be done as part of an overall boiler test, or simply to learn whether maintenance or adjustments are needed.

Fineness testing

Among the more daunting tasks in performance testing is the collection of valid fineness samples. The method of sampling is well described in PTC 4.2 and ASTM D 197. While this procedure has inherent weaknesses, it is the basis for measuring one of the fundamental performance parameters to verify pulverizer capacity. Periodically, alternative sampling procedures or new hardware are proposed to overcome the weaknesses in the ASTM method and apparatus. One such device is the ISO probe, also called the Rotorprobe. This well conceived device can sweep around circles that represent equal concentric areas of the pulverized coal pipe cross section. It is possible that the ISO probe may collect a more representative sample than the ASTM methods. However, the ASTM probe is currently the accepted standard on which capacity correction factors for fineness are based and whatever

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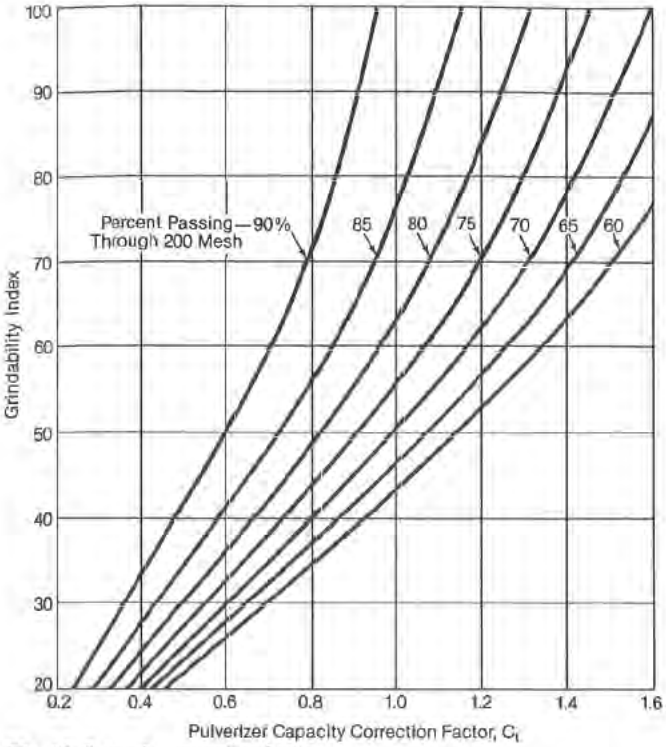


Fig. 16 Capacity correction factor.

errors are inherent in its application are included in its basis for use. It is important that new methods do not introduce new errors or controversies. They should produce results at least as consistent as the ASTM methods.

When tests are run to determine the condition of the pulverizer system or to evaluate a proposed new coal, comparative values between successive tests may be adequate. In such cases, alternatives, such as the ISO probe, may serve well if they provide acceptable repeatability and are not sensitive to variables such as velocity or temperature. Use of panel board instruments, if they can be read with sufficient precision, will serve as well as calibrated instruments for periodic testing.

Evaluating test results

The approach to evaluation depends on the purpose of the tests. If the parties have agreed upon test protocols as prescribed by PTC 4.2, results evaluation is simply a calculation process. When results indicate that the performance requirements have not been met but that the shortfall is slight, reference to the ASTM standards, D 409 and D 197, may show that the shortfall could lie within the limits of laboratory analysis repeatability. The protocols should provide for resolution of small deviations in the test results.

The more usual case of testing to evaluate the equipment or its systems requires a different approach. It is a mistake to try to read excessive precision into test results. It is more important to discover trends as guides to corrective action or the frequency of retesting. In such testing, it is useful to develop checklists or troubleshooting guides to note what each abnormality indicates. For example, acceptable fineness but excessive pressure differential probably indicates a need to adjust the pulverizer load springs. Acceptable fine fraction fineness but poor coarse fraction fineness indicates erosion damage and internal short circuiting of some air-coal flow. Also, it is common to discover that recalibration of primary measuring devices is needed on a regular schedule.

When equipment tests are first undertaken, it is important to collect all data which could possibly prove useful. As test experience grows, extraneous data can be removed. It is usually only a short time before standard data sheets evolve and testing becomes a routine task for an individual plant.

Operations

Power consumption

The main costs of coal pulverizer operation are capital, power consumption and maintenance. The capital cost is usually an annual levelized charge including capital, taxes and several other factors. Power and maintenance are often expressed as costs per ton of coal ground, and are influenced by operating plant practices.

When comparing pulverizer designs, it is common to combine primary air fan and pulverizer power consumption. The power consumption for B&W EL or MPS pulverizers and their respective fans is about 14 kWh/t (15 kWh/t_m) with coal at 50 HGI and 70% fineness, operating at rated output and with new grinding parts. These conditions seldom exist for any length of time during operation. Most modern installations include extra capacity for future coal variation. For most mills, there will be either a decline in capacity or an increase in power consumed as wear progresses. Therefore, because of the combined effect of pulverizer oversizing and wear, a realistic power value might be 20 kWh/t (22 kWh/t_m) for a well sized mill on a base loaded boiler or 22 kWh/t (24 kWh/t_m) on a load following unit.

There are various ways of calculating the cost of auxiliary power. The lowest cost which can be calculated is the cost of coal used to generate the power. This is a heat rate cost and, for a hypothetical boiler using coal with a 13,000 Btu/lb (30,238 kJ/kg) heating value and a thermal heat rate of 10,000 Btu/kWh (34.1% efficiency), 22 kWh/t (24 kWh/t_m) is about 0.85% of the fuel energy input to the boiler. As heating value declines, more power is spent grinding and moving material which does not produce thermal input, and the percentage represented by 22 kWh/t (24 kWh/t_m) increases.

Power can be expected to increase as the grinding parts wear. The depth of the grinding track may play a part in the power increase as well. There is little information to show whether power increases uniformly with time or if it takes place mostly at the end of wear life.

Maintenance

Costs for material and labor depend on wear life which varies with the abrasiveness of the coal. Wear life on MPS mills grinding U.S. coals ranges from under 10,000 to more than 100,000 hours, and normally between 25,000 and 60,000 hours. Therefore, average annual maintenance costs can vary widely. Costs will also vary by mill size, where part costs are proportionately higher for smaller MPS mills. For MPS-67 mills, taking all variables into account, maintenance will cost about 60% of the power costs cited above at a wear life of 25,000 hours and as little as 20% at 60,000 hours. Respective values

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for MPS-89 are 40% and 10%. For a very rough estimate, maintenance costs will be about one half the power costs for small mills and about one third for larger mills.

Maintenance and power costs are also influenced by operating practices. Experienced operators may notice that, for example, an increase in pulverizer pressure differential indicates a need for spring readjustment. There may be cost effective decisions between maintenance and operations regarding rebuild frequency to minimize the costly power increases near the end of wear life.

Ceramics

The use of ceramic tiles for erosion resistance has reduced repair costs. Erosion resistance of ceramic, 97% alumina, is at least eight times greater than an equivalent thickness of carbon steel. Ceramic lined panels are used in classifier cones, in the housing above the throat, and in the discharge turret in MPS mills. Ceramic lining is used extensively in burner pipe elbows and for wear shields on MPS roller brackets. When ceramic tiles are used, they must be carefully fitted to avoid undercutting and loss of the lining (Fig. 17). Ceramic tiles are not suitable for all forms of erosion protection. High angles of impingement are destructive to brittle materials.

Record keeping

Records are useful for pulverizer maintenance. Critical dimensions as recommended by the manufacturer should be taken and recorded during internal inspections. Suspicious wear data should be recorded and given special attention at the next inspection. When lubricants are drained, the amount should be noted and recorded and a sample obtained for analysis. Many owners are now including periodic vibration measurements in their data collection and record keeping. Such data may indicate impending bearing or gear failure.

Fires and explosions

It must always be remembered that a coal pulverizer grinds fuel to a form suitable for good combustion. Combustion can and does occur at unplanned locations with costly results. Pulverizer fires are serious and should be treated with preplanned emergency procedures. The surprising speed with which a fire can develop requires prompt action.

Pulverizer fires can develop in the so-called high temperature areas of mills or in the coal rich low temperature areas. Usually, there should be no coal in the areas upstream of the throat — the high temperature areas. Excessive spillage, because of insufficient air flow or throat wear, can cause coal accumulation in the plenum or windbox of the pulverizer. If this coal is not rejected promptly to the pyrite removal system, it will be ignited by the high temperature primary air. Coal will also enter this area when a mill is tripped or shut down in an abnormally fast mode. Coal in the windbox may burn out without damage and, indeed, this form of fire may go undiscovered. Fires in the cooler, fuel rich zones may develop more slowly, but these fires are fed by enormous amounts of coal and can quickly destroy the internal pulverizer parts.



Fig. 17 Ceramic tile lining.

The most common fire indicator is mill outlet temperature. Temperature indicators are slow in response because they are sheltered by erosion protection, but they are inherently reliable and their indications should be taken seriously.

Control room operators can gain useful information about potential hazards by following consistent operating practices, especially on multiple mill installations. If, for example, all mills are at the same coal feed rate, and if primary air control is in the automatic mode, all mills should have the same air/coal ratio. Under these conditions, a significant temperature difference from inlet to outlet in one mill may indicate a fire and should be investigated immediately. Of course, there are other possible causes of inlet temperature differences such as faulty feeder calibration, faulty air flow calibration, or coal moisture variation between the respective mills. The significant point is that there is important information to be drawn from careful observation of the vital signs of pulverizer system performance. Use of all vital signs as leading indicators of potential hazards is important to safe operations.

Severe fire damage has also been caused by feeding burning coal from silos or bunkers. All pulverizer manufacturers prohibit this practice. Silo fires must be dealt with separately by emptying the fire through special diverter chutes or by extinguishing agents applied to the coal surface.

In mill trips, the sudden loss of air flow causes the collapse of the fluid bed above the throat; the coal then falls into the windbox and contacts the hot metal surfaces. Low rank coals can begin smoldering within a few minutes at temperatures of 450F (232C) or more, and, upon agitation, dust clouds can be created leading to explosions. The agitation may be from mechanical action of the pyrite plows upon restart, or from the start of primary air flow before restart. The danger of explosion is minimized by removing the coal under rigidly controlled conditions. It is first necessary to assure personnel safety by evacuating all workers from places which could be affected by an explosion if the enclosures of the mill, ducts or burner pipes are breached. The mill is isolated from all air flow and an inert internal atmosphere is created by injecting inert gas or vapor. Brief operation of the mill, while isolated, rejects the coal to the pyrite removal system. The inert atmosphere prevents any smoldering coal from igniting the dust cloud raised internally by mill operation.

When a mill explosion occurs it is nearly always during changes in operating status, that is, during startup and shutdown.[2] During startup, the air-coal mixture in

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the mill passes from extremely fuel lean to fuel rich. There is a transitional mixture which is ideal for a dust explosion if a sufficiently strong ignition source exists. One recurring source of ignition is smoldering fires which develop in the residual coal left in the hot zone of a mill following a trip. The coal removal procedure just described was devised to avoid or safely remove this ignition source.

Undiscovered fires can cause dust explosions during otherwise normal shutdowns. These are not common, but experience has led many operators to actuate audible and visible alarms before routine starts and stops. This is a recommended safety precaution for all coal pulverizer operators.

In general, operators have reported a declining rate of incidents since 1985 and those with formal event investigation procedures have reported the greatest decline. The clear message from recent industry surveys is that experience and constant review of safety and emergency procedures will pay dividends in preventing injury and equipment damage from pulverizer fires and explosions. Proper safety systems and procedures are reducing the frequency and damage.

Controls and interlocks

The most widely followed standard for design and application of coal pulverizer systems is the National Fire Protection Association Standard NFPA-8503 (formerly NFPA 85F). This sets the minimum standards listed below for safety interlocks to be used with coal pulverizers:

1. failure of primary air flow trips the pulverizer system,
2. failure of the pulverizer trips the coal feeder and primary air flow,
3. closing of all burner pipe valves trips the pulverizer, the feeder, and the primary air flow, and
4. primary air flow below the manufacturer's minimum trips the pulverizer.

Other interlocks not mandated by NFPA but required by B&W are:

1. loss of the lube oil pump, or low pressure, trips the pulverizer (MPS mill only), and
2. flame safety systems trip the pulverizer if minimum flame detector indications are not met.

Additional indications of malfunctions which may require manually tripping a mill include:

1. excessively high or rapidly rising mill exit temperature probably indicates a fire,
2. inability to maintain an exit temperature of at least 125F (52C) may lead to loss of grinding capacity,
3. loss of seal air pressure or flow will eventually lead to bearing failure, and
4. high lube oil temperature indicates a loss of cooling water flow which may lead to gear drive damage.

The faults indicated by numbers 3 and 4 above are not strictly safety related. They are, however, examples of designers' choices which avoid unnecessary mill trips and the special safety procedures which must be followed to restart tripped mills.

Special safety systems

The procedure for removing coal after a pulverizer trip requires a separate control system. This system must bypass the control logic which prevents operation of the pulverizer drive motor without adequate primary air flow. To prevent misuse of this bypass feature, the coal removal logic (referred to as inert and clear logic) is enabled by actuation of the mill trip logic. A schematic logic diagram is shown in Fig. 18.

Another special system has been developed to deal with a phenomenon which has been observed in ducts after trips. A fine layer of dust is often seen in the clean air portion of the primary air system. This is thought to be due to a turbulent dust cloud which persists after collapse of the fluid bed. An array of water fog nozzles under the throat will intercept this cloud if actuated with

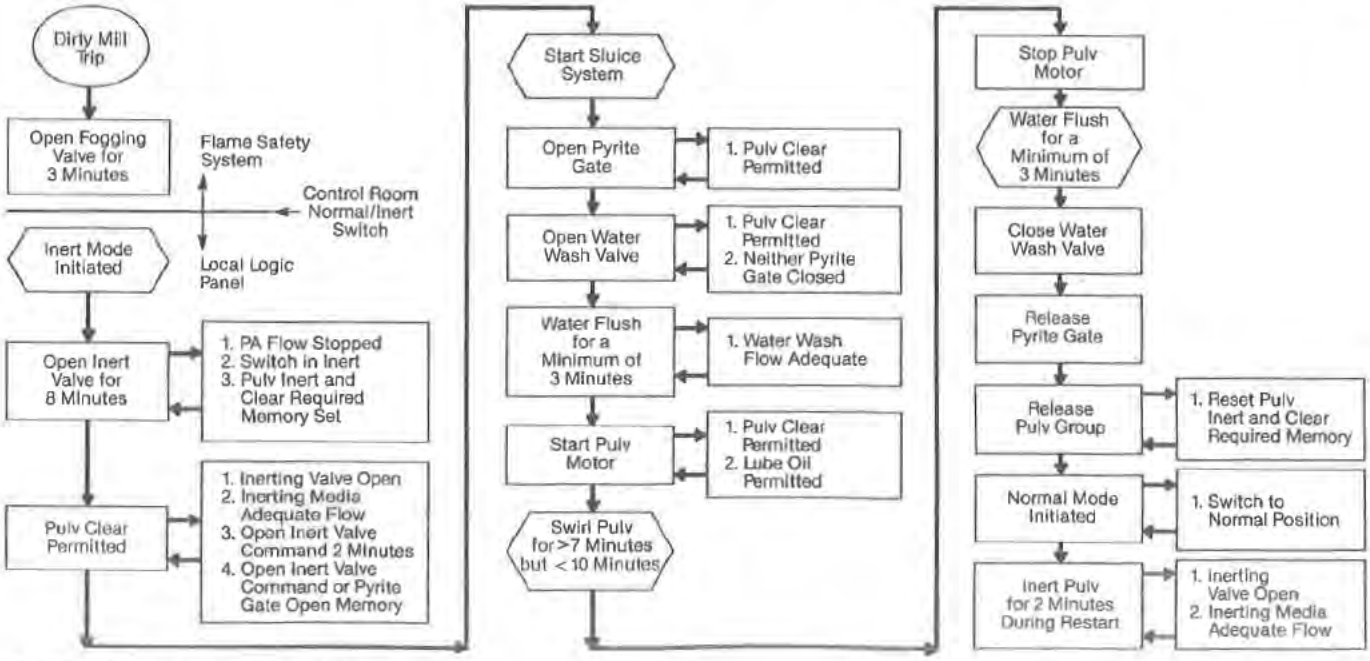


Fig. 18 Safety system logic diagram.

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the mill trip signal. Applied along with the automatic spray system is hardware for tangential wash nozzles to enhance the coal removal action of the pyrite plows during the inert and clear cycle. The wash nozzles are actuated manually at the appropriate time in the cycle.

Advanced designs

Rotating classifiers

As noted under *Application Engineering*, product fineness is under review after being generally standardized at 70% passing through 200 mesh (74 microns) for direct firing of bituminous coal. Raising fineness to 80% will meet most early 1990s emission limits, but more demanding requirements are expected. Fineness values in excess of 95% passing through 200 mesh (74 microns) can be achieved with externally driven rotating classifiers. This is a recent development for coal pulverizers, emerging since the mid 1970s, although rotating classifiers have been applied to large vertical rock grinding mills for many years. The design developed for EL pulverizers is shown in Fig. 19. Interestingly, as shown in Fig. 20, this classifier is more effective at reducing percentages of coarse particles than finer particles. The value shown of 95% passing through 200 mesh (74 microns) may be near the practical limit for large vertical pulverizers. Other configurations for classifier rotors have been developed including squirrel-cage, various vane types and two-stage using a stationary first stage. Each of these represents the designer's solution to problems with retrofit, with particle size distribution, with drive power consumption or with overall pulverizer performance. Classifiers developed for MPS pulverizers have taken forms similar to that shown in Fig. 19 as well as two-stage and squirrel-cage designs.

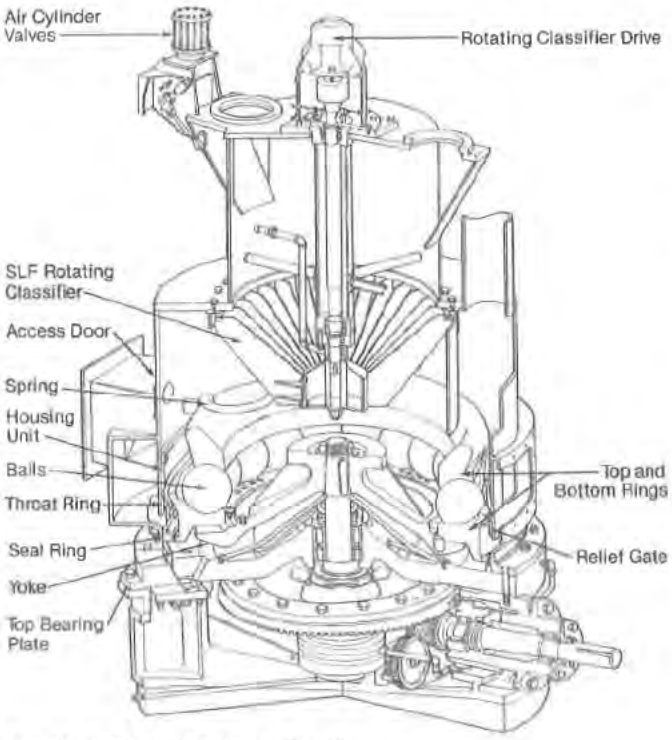


Fig. 19 Rotating classifier for EL mill.

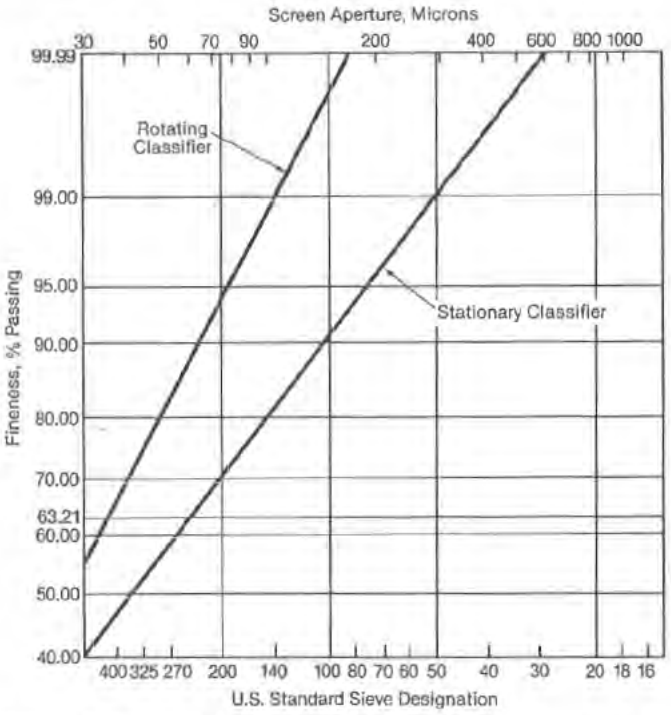


Fig. 20 Particle size distribution.

Product improvements

The 1980s was a period of refinement and product improvement for vertical pulverizers. One major improvement for MPS pulverizers was the introduction of rotating throat rings which improved performance and reduced erosive wear costs. A more subtle improvement was variable spring loading systems. It has been normal practice to operate vertical mill loading systems at constant preload. If, however, the spring preload is caused to follow coal feed rate, as illustrated in Fig. 21, the benefits will be reduced power consumption at low loads, ability to operate at lower output without vibration, and, in some instances, reduced pressure loss at high load by increasing preload above the previous fixed values.

Fine grinding

Vertical roller mills may have a practical limit of about 95% passing through 200 mesh (74 microns) for product fineness. As the product becomes finer, the recirculated material in the grinding chamber also becomes finer and the compressed layer becomes more fluid and less stable.

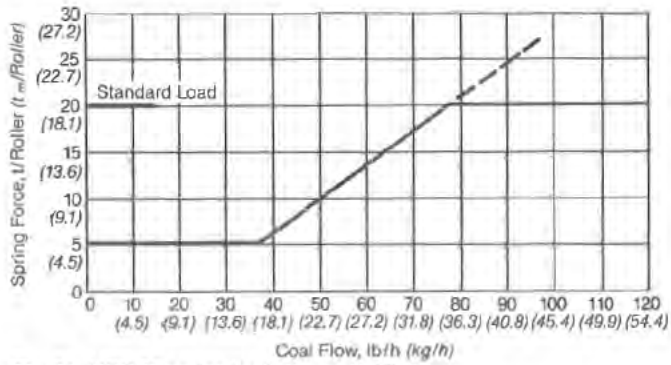


Fig. 21 MPS-89 spring load versus coal flow.

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For very fine grinding other machines are needed. One of these is the ball-and-tube mill, offering the long residence time needed for an extremely fine product. In addition, ball-and-tube mills are built in the large sizes needed for direct firing.

It is not clear that the high capacity of ball-and-tube mill will be needed to meet foreseen fine grinding applications. An interesting application of fine grinding is to provide fuel for coal-fired lighters to reduce oil consumption for startup and low load flame stabilization. Smaller capacity mills could be used for this application.

Mills for fine grinding have been used to grind pigments, food and pharmaceuticals. These are small mills, generally 5 t/h (4.5 t_m/h) or less. Grinding principles include fluid energy, stirred ball, and high speed attrition. Fluid energy mills cause solids bearing fluid streams to impinge; particle impacts cause the size reduction. Stirred ball mills, usually wet grinders, use a charge of small balls in a cylindrical vessel which are agitated by a rotor, grinding the solids flowing through the charge.

High speed attrition mills grind by a combination of first stage impact and final stage attrition. It is these mills which have begun to attract some attention for preparation of oil replacement lighter fuel. Fig. 22 illustrates the grinding chamber of a high speed attrition mill. The rotor is a series of discs, each with impact bars, and each forming the driver for a grinding stage. The grinding stages are separated by diaphragms and coal must move inward against centrifugal force to pass from one stage to the next. Coal captured in each stage is said to provide an effective wearing surface to protect the mill shell from rapid wear. These mills are made in capacities up to about 5 t/h (4.5 t_m/h) and may be once-through or be equipped with external centrifugal classifiers for size control. Manufacturers of these mills refer to the product as micronized coal. When the term was adopted in the early 1980s, micronized coal referred to coal which was 100% finer than a 325 mesh (44 microns). The output of vertical high speed attrition mills is about 98% passing through 200 mesh (74 microns), 86% passing through 325 mesh (44 microns). However, this coal is reported to be fine enough to be easily ignited without the need for preheated primary or combustion air. These mills are high in power consumption and motor power is said to be sufficient to provide the necessary energy for drying most coals. The economic attractiveness of finely ground coal as a stabilizing and startup fuel depends on oil prices.

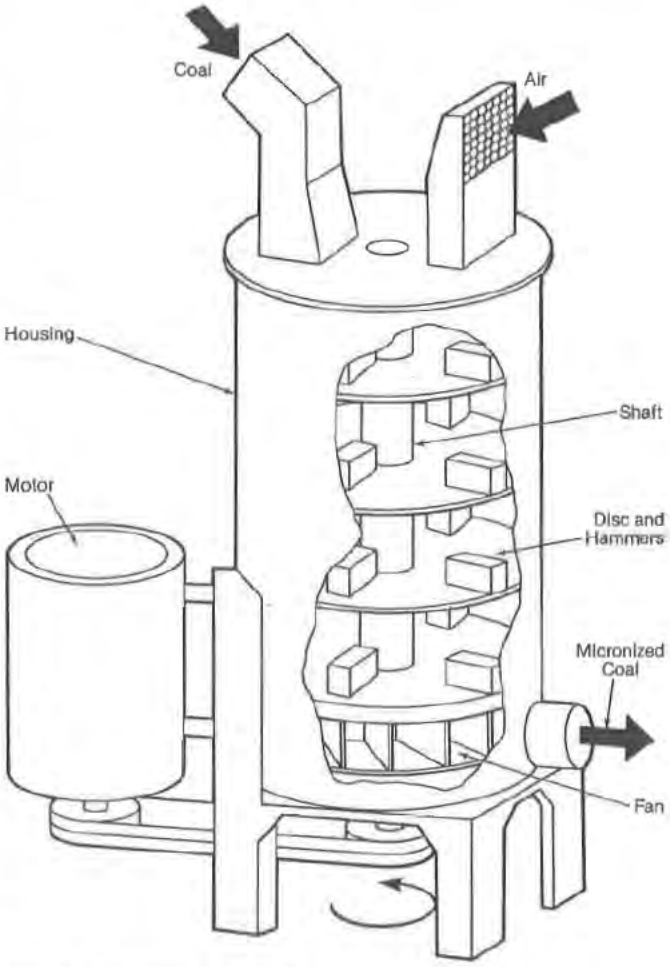


Fig. 22 High speed attrition mill.

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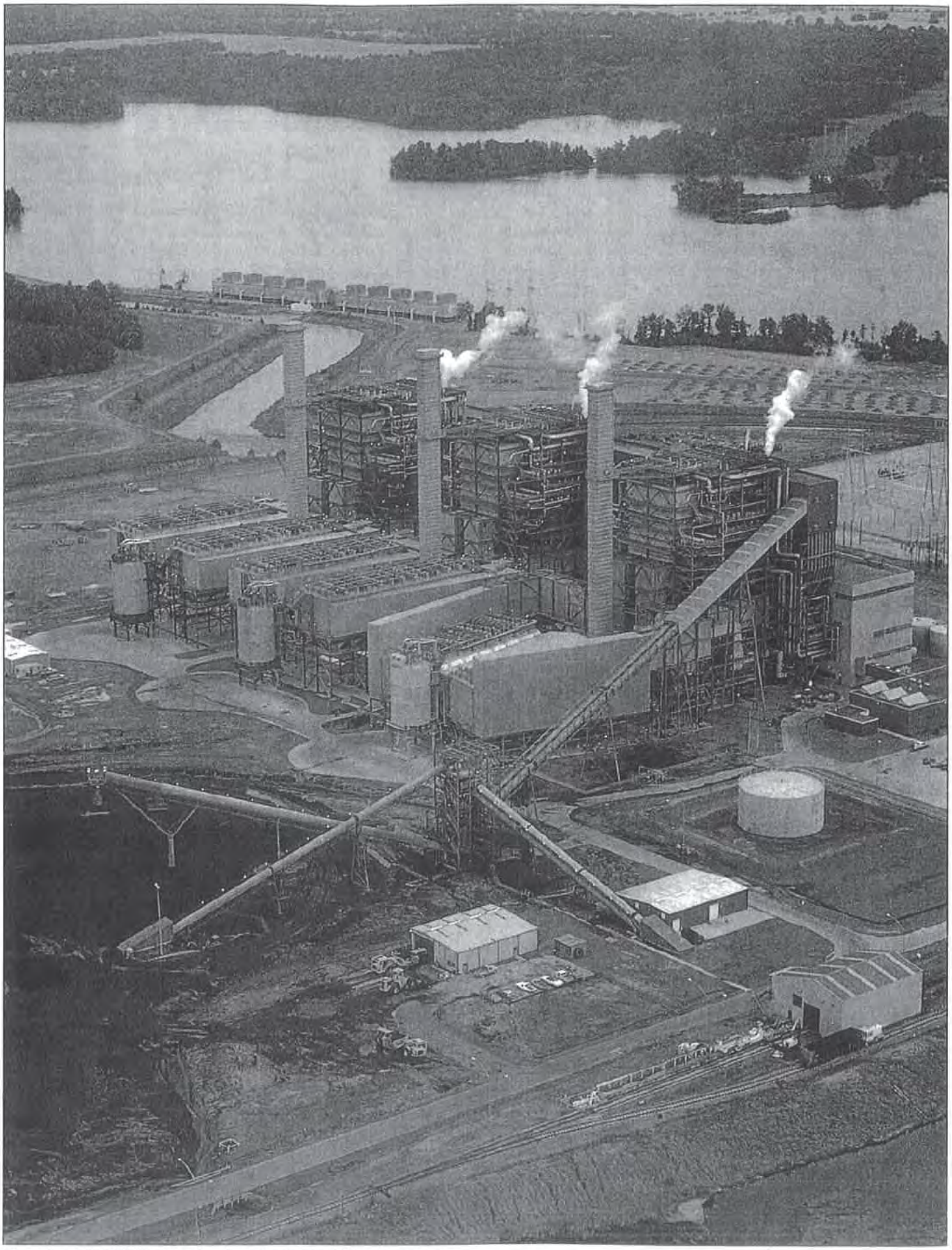
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These three 550 MW B&W boilers fire pulverized subbituminous coal.

Chapter 13
Burners and Combustion Systems for Pulverized Coal

Coal is an abundant and low cost fuel for boilers. It can be burned in a number of ways depending upon the characteristics of the coal and the particular boiler application. Cyclone, stoker, and fluidized bed firing methods are used for a variety of applications as discussed in Chapters 14 through 16. However, pulverized coal (PC) firing — burning coal as a fine powder suspension in an open furnace — is the dominant method in use today. PC firing has made possible the large, efficient utility boilers used as the base load capacity in many utilities worldwide. Modern PC burners provide high combustion efficiency and low emissions through full integration with the entire boiler design.

Pulverized coal firing differs from the other coal combustion technologies primarily through the much smaller particle size used and the resulting high combustion rates. The combustion rate of coal as a solid fuel is, to a large extent, controlled by the total particle surface area. By pulverizing coal to a nominal 50 micron diameter or smaller (see Chapter 12), the coal can be completely burned in approximately one to two seconds. This approaches the rate for oil and gas. In contrast, the other technologies discussed in subsequent chapters use crushed coal of various sizes and provide substantially longer combustion zone residence times (up to 60 seconds or longer).

Pulverized coal was first used in the 1800s as a cost effective fuel for cement kilns.[1] The ash content enhanced the properties of the cement and the low cost resulted in the rapid displacement of oil and gas as a fuel. However, early pulverizing equipment was not highly reliable and, as a result, an indirect bin system was developed to temporarily store the pulverized coal prior to combustion, providing a buffer between pulverization and combustion steps. Use in the steel industry followed closely. In this application, the importance of coal drying, particle size control and uniform coal feed were recognized.

Success for boiler PC firing applications required modification of the boiler furnace geometry to effectively use the PC technology. In the early 1900s, fireboxes and crowded tube banks had to be expanded to accommodate PC firing. By the late 1920s, the first water-cooled furnaces were used with advanced pulverized coal-fired systems, and burner design advanced to improve flame stability, provide better mixing, increase flame temperatures and improve combustion efficiency. Coupled with improvements in pulverizer design and reliability, improved PC burners permitted the use of direct firing with coal transported directly from the pulverizer to the burner. (See Fig. 1.)

The exponential increase in the United States (U.S.) electric power generation and the increase in boiler size from 100 MW to as large as 1300 MW could not have been achieved economically without the development and refinement of pulverized coal firing. Today, nearly all types of coal from anthracite to lignite can be burned through pulverized firing. Combustion efficiencies of most coals approach those of oil and gas. Current research is focusing on continuing reduction in emissions of nitrogen oxides (NO_x) for environmental protection without sacrificing boiler performance or availability. (See also Chapter 34.)

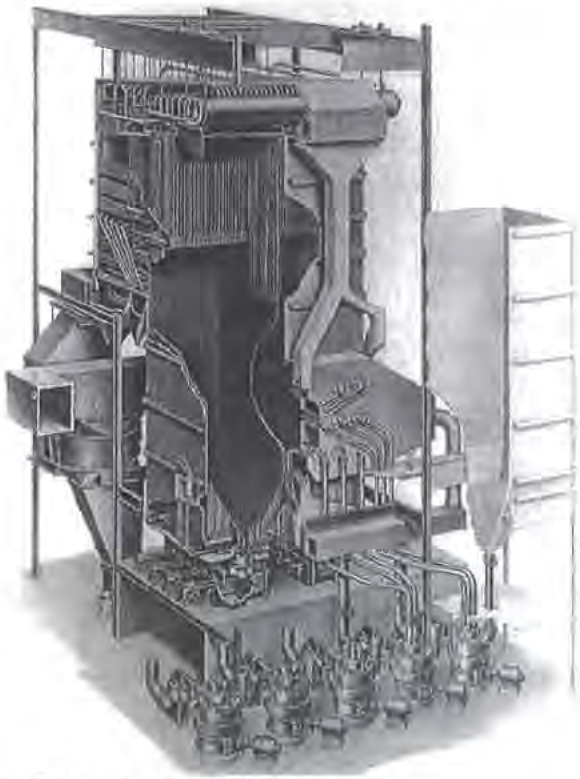


Fig. 1 Early pulverized coal-fired radiant boiler.

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Combustion

The manner in which pulverized coal burns depends on its rank and properties as well as the furnace conditions. As a coal particle enters the furnace (see Fig. 2), its surface temperature increases due to radiative and convective heat transfer from furnace gases and other burning particles. As particle temperature increases, the moisture is vaporized and volatile matter is released. This volatile matter, which ignites and burns almost immediately, further raises the temperature of the char particle, which is primarily composed of carbon and mineral matter. The char particle is then consumed at high temperature leaving the ash content and a small amount of unburned carbon. The volatile matter, fixed carbon (char precursor), moisture and ash content of the fuel are identified on a percentage basis as part of the proximate analysis discussed in Chapter 8.

Volatile matter content

Volatile matter is critical for maintaining flame stability and accelerating char burnout. Coals with minimal volatile matter, such as anthracites and low volatile bituminous, are more difficult to ignite and require specially designed combustion systems. The amount of volatile matter evolved from a coal particle depends on coal composition, the temperature to which it is exposed, and the time of this exposure. The American Society for Testing and Materials (ASTM) Method D 3175 stipulates a temperature of 950 ± 20C for seven minutes for volatile matter content determination.[2] Raising the temperature would increase volatile yield with other factors held constant. Coals with higher volatile matter content also benefit from more effective NO_x control by combustion methods. Ignition is influenced by the quality and the quantity of volatile matter. Volatile matter from bituminous and higher rank coals is rich in hydrocarbons and high in heating value. Volatile matter from lower rank coals includes larger quantities of carbon monoxide and moisture (from thermal decomposition) and consequently has a lower heating value. Volatile matter from higher rank coals can provide twice the heating value per unit weight as that from low grade coals.

Char particles

The speed of the char particle combustion depends on several factors including particle size, porosity, thermal environment, and oxygen partial pressure. Char reactions often begin as the coal particle is heated and devolatilizes, but they continue long after devolatilization is complete. Devolatilization is mostly completed after 0.01 seconds but char-based reactions continue for one to two seconds. The char particle retains a fraction of the hydrocarbons. Small particles, with 10 to 20 micron diameters, benefit from high surface to mass ratios and heat up rapidly, while coarse particles heat more slowly. Many coals go through plastic deformation and swell by 10 to 15% when heated. These changes can significantly impact the porosity of the coal particle.

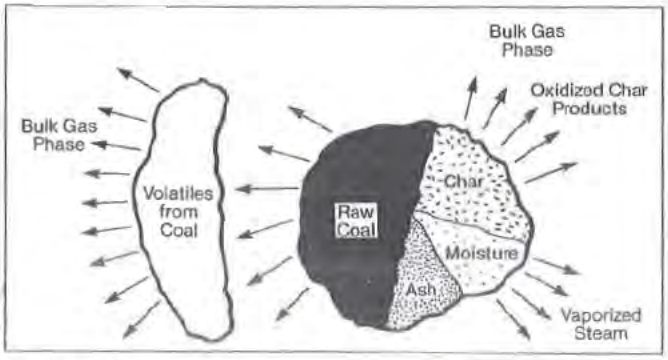


Fig. 2 Coal particle combustion.

Char oxidation requires oxygen to reach the carbon in the particle and the carbon surface area is primarily within the particle interior structure. Char combustion generally begins at relatively low particle temperatures. Reaction rates are primarily dependent upon local temperature as well as oxygen diffusion and char reactivity. For larger particles, the solid mass is reduced as carbon monoxide (CO) and carbon dioxide (CO_2) form, but particle volume is maintained. Coarse particles, more than 100 micron diameter, burn out slowly as a result of their lower surface to mass ratios. Longer burnout times cause these larger char particles to continue reacting downstream where the flame temperature has moderated.

Rapid heat transfer to and combustion of smaller particles lead to higher particle temperatures. Reaction rates increase exponentially with temperature, and oxygen (O_2) diffusion into the particle becomes the controlling parameter. Particle diameter and density change in the process. At higher particle temperatures, char reactions are so fast that oxygen is consumed before it can penetrate the particle surface. The particle shrinks as the outer portions are consumed, and transport of oxygen from the surroundings to the particle is the factor governing combustion rate.

Effect of moisture content

The moisture content of the coal also influences combustion behavior. Direct pulverized coal-fired systems convey all of the moisture to the burners. This moisture presents a burden to coal ignition; the water must be vaporized and superheated as the particles devolatilize. Further energy is absorbed at elevated temperatures as the water molecules dissociate.

Moisture content increases as rank decreases as discussed in Chapter 8. 15% moisture is common in high volatile bituminous coals, 30% is seen in subbituminous, and more than 40% is common in some lignites. Moisture contents in excess of 40% exceed the ignition capability of conventional PC-fired systems. Alternate systems are then required to boost drying during fuel preparation and/or divert a portion of the evaporated moisture from the burners. Char burnout is impaired by moisture which depresses the flame temperature. This is compensated for in part by the generally higher inherent reactivities and porosities of the higher moisture coals.

Effect of mineral matter content

The mineral matter, or resulting ash, of the coal is inert and dilutes the coal's heating value. Consequently, more fuel by weight is required as ash content increases in order to reach the furnace net heat input.

The ash absorbs heat and interferes with radiative heat transfer to coal particles, inhibiting the combustion

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process noticeably with high ash coals. The forms of mineral matter determine the slagging tendencies of the ash. These impact combustion by influencing the burner/furnace arrangement and the resulting flame temperatures and residence time. The additional effects of ash on overall boiler design are discussed in Chapter 20.

Pulverized coal combustion system

Effect of coal type

Conventional PC Pulverized coal firing is adaptable to most types of coal. Its versatility has made it the most prevalent method of coal firing for power generation worldwide. The majority of coals are PC-fired in conventional combustion systems as shown in Fig. 3. In these systems, the burners are located in the lower portion of the furnace, usually on one or two walls. Primary air typically at 130 to 200F (54 to 93C) conveys pulverized coal directly to the burners at a rate set by the combustion controls based on steam generation requirements. Secondary air is supplied by the forced draft fans and is typically preheated to about 600F (316C). All or most of the secondary air is supplied to the windboxes enclosing the burners. A portion of the secondary air may be diverted from the burners to NO_x ports (discussed later) in order to control the formation of NO_x. The secondary air supplied to the burners is mixed with the pulverized coal in the throat of the burner. This permits the coal to ignite and burn.

The combustion process continues as the gases and unburned fuel move away from the burner and up the furnace shaft. Final burnout of the char depends on the coal properties, particle fineness, excess air, air-fuel mixing, and thermal environment. The products of combustion eventually leave the furnace and enter the convection pass after being cooled sufficiently to minimize convection surface fouling.

Low volatile coals While the majority of bituminous, subbituminous, and lignite PC-fired units are arranged in this manner, alternate designs are required to accommodate other coals. Coals with low volatile matter content, particularly anthracites, are difficult to ignite. Their low levels of rapid burning volatiles limit this critical source of ignition energy. In addition, their advanced coalification (see Chapter 8) has reduced char reactivity, resulting in elevated char ignition and burnout temperatures. These factors require hotter furnaces to sustain combustion. A downshot firing system (see Fig. 4) accomplishes this in combination with an enlarged, refractory-lined furnace. The downshot arrangement causes the flame to travel down and then turn upward to leave the furnace. The char reactions, which build in intensity some distance from the burners, return their heat near the burners as the gases return and leave the furnace. This heat supplies energy to ignite the fuel introduced at the burners. The refractory lining in the furnace inhibits heat transfer and elevates gas temperatures to sustain combustion. The enlarged lower furnace also provides increased residence time to accommodate the slower burning char. In addition, very high coal fineness is used to accelerate char reactions and reduce unburned carbon loss. Growth in unit size and NO_x emission regulations have limited downshot fired units in the U.S. although they are applied internationally where low volatile coals are a fuel source.

High moisture coals The other major category of nonconventional PC-fired units is that used for lignites with very high moisture content. Moisture content in these coals typically ranges from 50 to 70% by weight, which is well beyond the level that air-swept pulverizers can adequately dry. (See Chapter 12.) Instead, hot gases, removed from the upper portion of the furnace, are used in combination with beater mills to dry and prepare the fuel. These hot gases, near 1832F (1000C), are more suitable for coal drying than preheated air, with a practical upper limit of 752F (400C). Furthermore, the low oxygen content of the flue gas provides a relatively inert atmosphere in the grinding equipment, lessening the threat of fire or explosion as the hot gas mixes with these reactive coals. The fuel is conveyed from the mills to the burners located on the walls, and combustion proceeds as secondary air mixes with the fuel in the furnace. Units of this design are prevalent in parts of Europe and Australia where high moisture brown coals are used.

Combustion system and boiler integration

The most fundamental factors that determine the boiler design are the steam production requirements and the coal to be fired. The thermal cycle defines the heat absorption requirements of the boiler which supplies the rated steam flow at design temperature and pressure. Gas side parameters, based on the design coal, are used to estimate boiler efficiency. Heat input requirements from the coal can then be determined, setting the coal firing rate at maximum load. The number and size of pulverizers are then selected. Frequently, the pulverizer

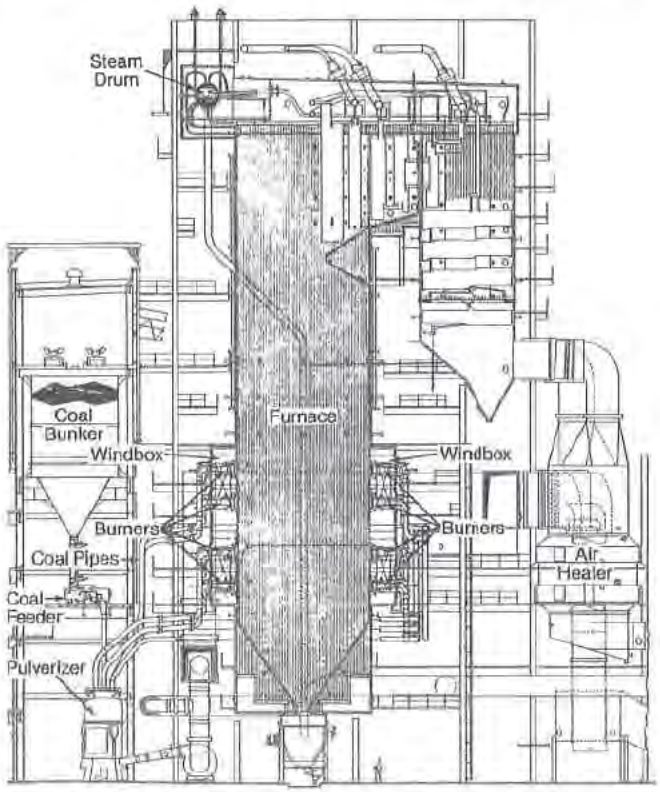


Fig. 3 Conventional pulverized coal-fired system.

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selection is based on meeting maximum requirements with one mill out of operation. This permits maintenance on one pulverizer without limiting boiler load.

The size and configuration of the furnace are designed to accommodate the combustion and slagging/fouling characteristics of the coal as discussed in Chapter 20. NO_x emission control factors are also incorporated into the layouts of modern combustion systems. The number of burners and the heat input per burner are selected to minimize flame impingement on the furnace walls. Multiple burners at moderate inputs, in an opposed-fired arrangement, are favored to provide uniform heat distribution across the furnace. This reduces thermal gradients and helps prevent localized slagging, while improving NO_x emissions control and combustion efficiency. The burner design is based on fuel parameters and NO_x emission control requirements in order to provide complete combustion and minimum pollutant emissions.

Theoretical and excess air quantities are determined for the required coal input rate from combustion air calculations as discussed in Chapter 9. Fans, ductwork, air heaters, windboxes, and burners are sized to satisfy flow, pressure loss, air preheat and velocity requirements for the unit. If needed to meet emission requirements, separate staged combustion can be employed in the furnace through air injection after the burner zone (NO_x ports).

Burner fundamentals

Burners are the central element of effective combustion system designs which incorporate fuel preparation, air-fuel distribution, furnace design and combustion control. The burners provide for the introduction and mixing of the fuel and combustion air as shown in Fig. 5.

Control The quantity of pulverized coal and primary air supplied to the burners is set by the coal feed rate and primary air control dampers. Restrictors are usually included in the piping from the pulverizer to its burners to more evenly distribute coal among several burners. Secondary air flow is controlled by dampers at the forced draft fans to maintain excess air for good combustion and proper boiler operation. The distribution of secondary air among burners is accomplished by duct/windbox dampers and burner adjustments.

Flame stability A burner introduces the primary air and pulverized coal to the secondary air in a manner that establishes a stable flame. This involves producing a flame front close to the burner over the range of operating conditions. An ignitor or lighter is required to initiate combustion as fuel is first introduced to the burner and to sustain combustion when flames would otherwise be unstable. The burner normally sustains a stable flame by using heat from coal combustion to ignite the incoming pulverized coal. A flame safety system, included with modern burners, electronically scans the flame to verify stabilization and triggers corrective action if the flame becomes unstable.

Air-fuel mixing The burner also promotes local air-fuel mixing to completely burn the fuel. The overall mixing is a combination of burner-induced local mixing and the resultant more global furnace-induced mixing. Furnace mixing results from expansion of the air-fuel mixture in the furnace due to rapid temperature increase during combustion, from flow variations introduced by the burners, and from the size and shape of the furnace enclosure.

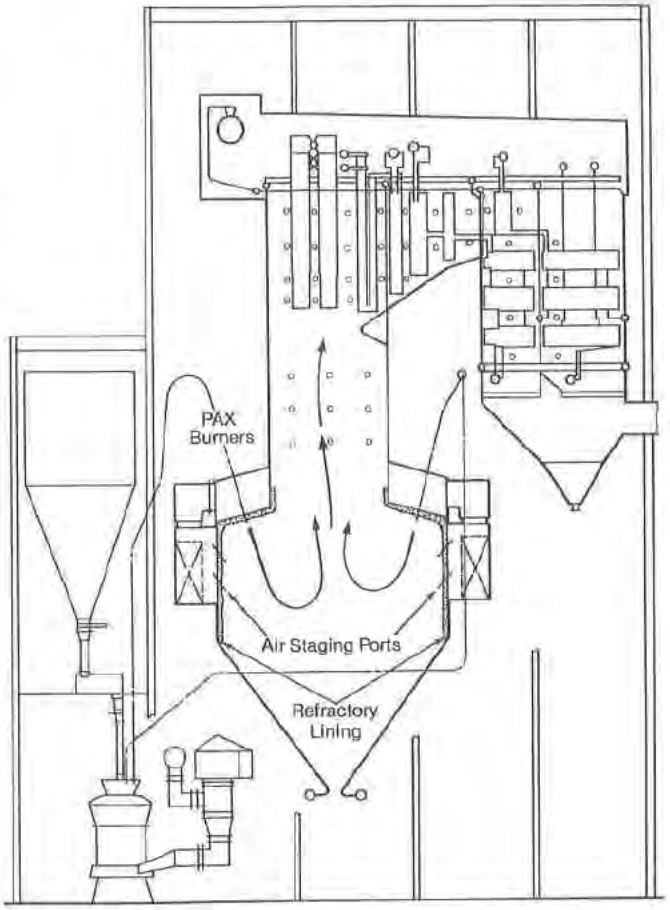


Fig. 4 Downshot-fired unit with refractory-lined furnace for low rank coals and PAX (primary air exchange) burners.

The amount of air-fuel mixing produced by the burner varies considerably with burner type. The simplest burners inject the fuel and air in parallel or concentric streams without the benefit of other burner-induced mixing. Entrainment of adjacent flow streams occurs as the jets develop and due to jet expansion from combustion. Tangentially-fired furnaces operate in this manner, as do some roof-fired designs. Flame length and combustion performance greatly depend on the effectiveness of furnace mixing. Mixing effectiveness decreases with boiler load as air-fuel flow rates drop. This can result in poor combustion performance at reduced boiler loads.

Modern wall-fired burners provide further controlled air-fuel mixing. The amount varies with the firing rate of the burners. Air-fuel mixing is achieved at reduced boiler loads by operating with fewer burners in service and by operating those burners at higher firing rates. Burner mixing can be induced by using the primary air/pulverized coal (PA/PC) stream, by using the secondary air, or by a combination of the two. Looking first at the PA/PC system, the most frequently used burner mixing devices are deflectors, bluff bodies, and swirl generators. Deflectors are frequently installed near the exit of the burner nozzle to cause the PA/PC stream to disperse into the secondary air. These deflectors, or impellers, also reduce axial momentum of the fuel jet, reducing flame length. Impellers may also induce radially-pitched swirl of the fuel jet to further accelerate mixing. Bluff bodies

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are sometimes used in or adjacent to the burner nozzle exit. Flow locally accelerates around the upstream side of the bluff body and recirculates on the downstream side. The recirculation promotes mixing. The bluff body can also be used to increase residence time for a portion of the fuel near the burner, thereby improving flame stability.

Secondary air swirl is the most common way to induce air-fuel mixing in circular throat burners. Swirl generators are used upstream of the burner throat to impart rotating motion to the secondary air. This air leaves the burner throat with tangential, radial and axial velocity components. Radial and axial pressure gradients form in the flow field downstream of the throat, with the lowest pressure being near the center of the throat.[3] As swirl is increased the pressure gradients increase. This causes the flow to reverse and travel along the axis of the flame toward the low pressure zone. A recirculating flow pattern is then generated near the burner.

Primary air and coal flow rates The primary air requirements are determined by the pulverizer, and the burner piping must be accommodated by the burners. Most pulverizers require 40 to 70% of their full load primary air requirements at their minimum output level. (See Chapter 12.) In addition, the PA/PC mixture traveling to the burners must be transported at a minimum of 3,000 ft/min (15 m/s). This velocity serves to prevent the coal particles from dropping out of suspension in horizontal runs of coal pipe. Minimum primary air flow is the greater of the minimum PA flow required for the pulverizer or the minimum required to satisfy burner line velocity limits.

The primary air and coal mixture conveyed to the burners reaches a maximum velocity and solids loading at full pulverizer load and follows the pulverizer output as mill loading is reduced. As the burner nozzle velocity increases, the ignition point gradually moves farther from the burner. At some point, continued increases in nozzle velocity lead to *blowoff* of the flame, a potentially hazardous condition where coal ignition and flame stability are lost. High burner nozzle velocities also result in accelerated erosion of burner hardware. The weight ratio of coal to primary air typically reaches a peak of 0.4 to 0.65 at full load and a minimum of 0.15 to 0.3 for minimum pulverizer load. Solid fuel loading reduction eventually leads to flame instability as the flame temperature drops.

Temperature The temperature of the PA/PC mixture supplied to the burners is based on the raw coal properties. A minimum of 130F (54C) is required with high moisture, reactive coals. A value of 150F (66C) is typical for high volatile bituminous coal. A maximum of 200F (93C), a result of pulverizer mechanical constraints, is used with low volatile coals to improve ignition.

Unburned carbon loss A small portion of the coal is not completely burned in the boiler, typically reducing efficiency by less than 1%. This unburned carbon loss is calculated based on the furnace-combustion system configuration, thermal environment in the furnace, coal reactivity, coal fineness and excess air levels. This unburned carbon level is included in the final determination of total boiler efficiency. (See Chapters 9 and 21.)

Performance requirements

Pulverized coal-fired equipment should meet the following performance conditions:

1. The coal and air feed rates must comply with the load demand over a predetermined operating range. For modern applications with high volatile bituminous or subbituminous coal, flames should be stable without the use of lighters from about 30% to full load. The minimum load depends on the coal, burner design, and pulverizer load; operation below this load is performed in combination with lighters in service.

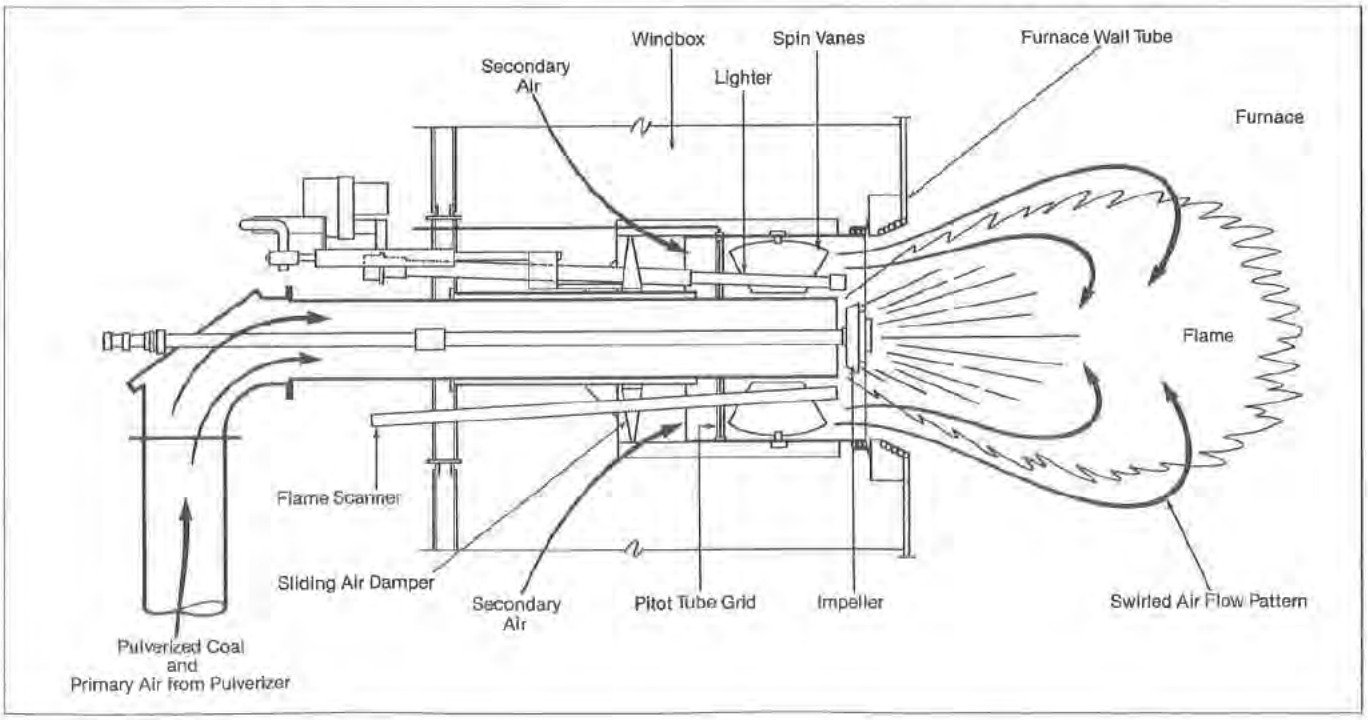


Fig. 5 S-type burner and components.

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2. Unburned combustible loss (UCL) should reduce efficiency by less than 1%. UCL less than 0.2% should be expected with reactive subbituminous or lignite coals. UCL increases with less reactive bituminous coals and in boilers with less than 400,000 lb/h (50.4 kg/s) steam capacity.
3. The burner should not require continual adjustment to maintain performance. The unit should be designed to avoid the formation of localized slag deposits that may interfere with burner performance or damage the boiler.
4. Only minor maintenance should be necessary during the annual outage. To avoid high temperature damage, alloy steel should be used for burner parts exposed to furnace radiative heat transfer. Burner parts subject to PC erosion may require more frequent repair or replacement.
5. Safety must be paramount under all operating conditions. Automated flame safety and combustion control systems are recommended and required in many cases.

Conventional PC burners

Prior to 1971 in the U.S., the primary focus of combustion system development was to permit the design of compact, cost effective boilers. As a result, the burner systems developed focused on maximizing heat input per unit volume with small furnace volumes, rapid mixing burners, and very high flame temperatures. An unintended side effect was the production of high levels of NO_x. Many boilers with this technology remain in operation today. Burners used on such boilers include the conventional circular burner, the cell burner, and a more mechanically reliable version — the S-type burner.

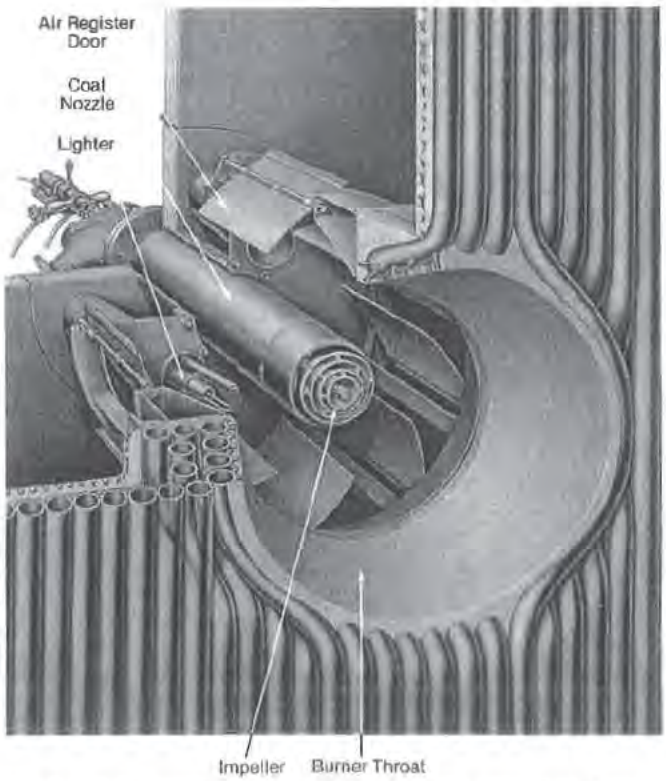


Fig. 6 Circular register pulverized coal burner with water-cooled throat.

Conventional circular burner

The circular burner (Fig. 6) was one of the earliest forms of swirl-stabilized PC-fired burners. Due to its success, this burner has been used for six decades firing a variety of coals in many boiler sizes. The circular burner may still be supplied for conventional use in selected cases. The burner is composed of a central nozzle to which PA/PC is supplied. The nozzle is equipped with an impeller at the tip to disperse the coal into the secondary air. Secondary air is admitted to the burner through a register. The register consists of interlinked doors arranged in a circular pattern between two plates. The doors are closed to cooling position when the burner is out of service, are partially open for lightoff and are more fully open for normal operation.

Opening the register doors allows more air into the burner but reduces swirl. Flame shape can also be adjusted using the registers. However, air distribution among burners is also impacted by register position, and high combustion efficiency depends on uniform air-fuel distribution.

The circular burner can fire natural gas, oil or pulverized coal. Simultaneous firing of more than one fuel is not recommended due to the potential for overfiring the burners and the resulting combustion problems. Variations of the circular burner are in operation at inputs from 50 to 300 $\times 10^6$ Btu/h (15 to 88 MW). Moderate input units provide more uniform heat distribution to the combustion zone, improving operation.

Cell burner

The cell burner combines two or three circular burners into a vertically stacked assembly that operates as a single unit (Fig. 7). In the 1960s and 1970s the cell burner was applied to numerous utility boilers which had compact burner zones. While highly efficient, the cell burners produced high levels of NO_x emissions.

S-type burner

The S-type burner shares the functional attributes of the circular burner in an improved configuration. (See Fig. 5.) The burner nozzle is the same as that in the circular burner. However, secondary air flow and swirl are separately controlled. Secondary air quantity is controlled by a sliding disk as it moves closer to or farther from the burner barrel. Secondary air swirl is provided by adjustable spin vanes positioned in the burner barrel. An air-measuring pitot tube grid is installed in the barrel ahead of the spin vanes. Secondary air can be measured for each burner to distribute the flow uniformly among burners using the sliding disks or other means. Swirl control for flame shaping is controlled separately by spin vanes. The S-type burner provides higher combustion efficiency and mechanical reliability than the circular unit and is a direct plug-in replacement.

Low NO_x combustion systems

NO_x formation

NO_x is an unintended byproduct from the combustion of fossil fuels and its emissions are regulated in the U.S. and many other parts of the world. (See Chapter 32.)

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While a number of options are available to control and reduce NO_x emissions from boilers, as discussed in Chapter 34, the most cost effective means is the use of low NO_x combustion technology either alone or in combination with other techniques. The effectiveness of the pulverized coal NO_x control technology, however, depends upon the fuel characteristics and the overall system design. For pulverized coal wall-fired units, NO_x emissions from conventional combustion systems discussed above typically range from 0.8 to 1.6 lb/10^6 Btu (984 to 1968 mg/Nm^3; see Note below). Low NO_x PC combustion systems are capable of reducing NO_x to 0.2 to 0.7 lb/10^6 Btu (246 to 861 mg/Nm^3).

NO_x control by combustion

More than 75% of the NO_x formed during conventional PC firing is fuel NO_x; the remainder is primarily thermal NO_x. Consequently, the most effective combustion countermeasures are those limiting fuel NO_x formation. Fuel NO_x is formed by oxidation of fuel-bound nitrogen during devolatilization and char burnout. Coal typically contains 0.5 to 2.0% nitrogen bound in its organic matter. High oxygen availability and high flame temperatures during devolatilization encourage the conversion of volatile-released nitrogen to NO_x. Nitrogen retained in the char has a lower conversion efficiency to NO_x, primarily due to lower oxygen availability during char burnout. Reactive coals with high volatile matter and low fixed carbon/volatile matter (FC/VM) ratios have tended to be the most amenable to NO_x control by combustion modification. Coals with higher nitrogen content tend to produce higher NO_x emissions.

The most effective means of reducing fuel-based NO_x formation is to reduce oxygen (air) availability during the critical step of devolatilization. Additional air can then be added later in the process to complete char reactions and maintain high combustion efficiency.

Oxygen availability can be reduced during devolatilization in two ways. One method is to remove a portion of the combustion air from the burners and introduce it elsewhere in the furnace. This method is referred to as *air staging*. PC combustion systems typically operate with 15 to 20% excess air at maximum firing rate. (See Chapter 9.) In some staging situations, reducing the air flow by 10% at the burners is sufficient for a given level of NO_x emissions control. This permits the burners to continue to operate in an excess air condition. Reducing the air flow to less than stoichiometric (theoretically required for complete combustion air flow) conditions can further reduce NO_x emissions. However, further reductions in burner stoichiometry below 70% of the theoretical air requirements can cause NO_x to increase again. This is due largely to the significant fraction of combustion which is deferred and the high air flows and mixing rates necessary to subsequently complete this combustion. Burner operation below theoretical air requirements can also increase unburned carbon loss and slagging in the combustion zone with bituminous coals while increasing the risk of furnace corrosion. Unburned carbon loss

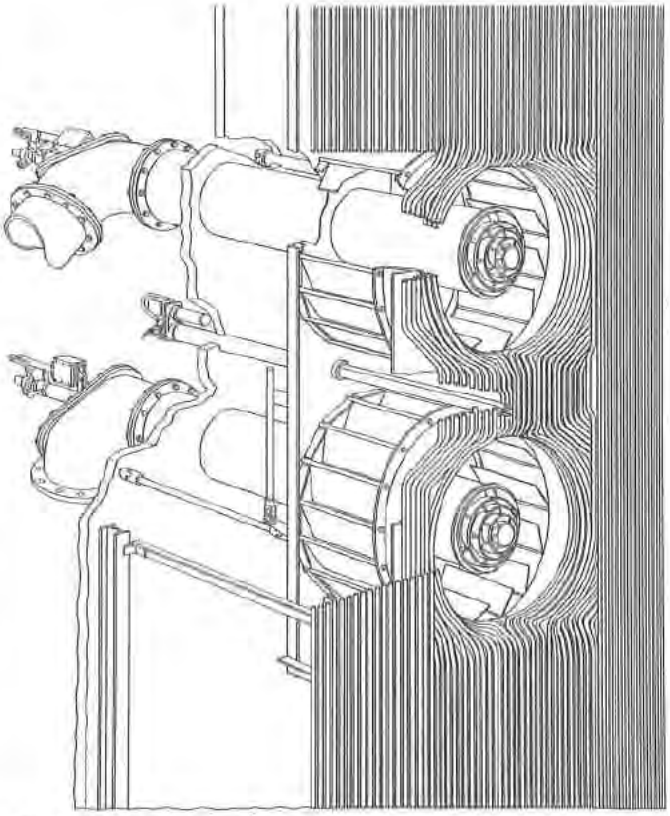


Fig. 7 Conventional cell burner.

can be controlled by increasing coal fineness and by careful management of air-fuel distribution among burners and air staging ports. Slagging can be minimized by proper arrangement of burners and by burner design. Tube corrosion damage by hydrogen sulfide (H_2S) attack is a concern when firing high sulfur coals under substoichiometric conditions. H_2S is formed during the combustion of sulfur bearing coals under oxygen deficient or reducing conditions. The resulting corrosion potential is dependent upon the H_2S concentration, tube material, tube temperature and operating conditions. This corrosion potential can be reduced by switching to low sulfur coal or by coating the burner zone tubes with a corrosion resistant material.

A second method of reducing oxygen availability during coal devolatilization is by burner design. The burner can be designed to supply all the combustion air but to limit its rate of introduction to the flame. Only a fraction of the air is permitted to mix with the coal during devolatilization. The remaining air is then mixed downstream in the flame to complete combustion. However, overall mixing is reduced and the flame envelope is larger compared to rapid mixing conventional burners. NO_x emission reduction using low NO_x burners ranges from 30 to 60% compared to uncontrolled levels. These burners can be used in combination with air staging through NO_x ports (discussed later) to satisfy emission limits. However, from a cost and performance perspective, the use of NO_x ports should be minimized where possible. Advanced low NO_x burners can frequently meet emission control requirements without the use of NO_x ports.

Another method of reducing NO_x emissions is referred to as reburning or fuel staging. This control technique

Note: The International Energy Agency conversion has been adopted for NO_x emission rates: 1230 mg/Nm^3 equals 1 lb/10^6 Btu for dry flue gas, 6% O_2, 350 Nm^3/GJ for coal.

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destroys NO_x after it has formed. Fuel staging involves introducing the fuel into the furnace in steps. The bulk of the fuel is burned in the furnace at near stoichiometric conditions. The balance of the fuel with a limited amount of air is then injected to create a reducing zone part way through the combustion process. The reducing conditions form hydrocarbon radicals which strip the oxygen from previously formed NO_x, thereby reducing overall NO_x emissions. The balance of the air necessary to complete the combustion is then added. Some fuel-staged systems involve separate burners or fuel injectors followed by NO_x ports which supply the remaining combustion air. Some advanced low NO_x burners embody fuel staging while supplying all the fuel to the burners. A portion of the fuel is introduced in a manner to generate hydrocarbon radicals; the balance is mixed into the flame later to reduce previously formed NO_x. These burners can achieve NO_x reductions of 50 to 70% from baseline uncontrolled levels.

Dual register burner

B&W began experimenting with techniques to reduce NO_x from PC-fired burners in the 1950s. A slot burner, which incorporated controlled air-fuel mixing, was first developed.[4] Air introduced through the burner throat mixed rapidly in the flame while air diverted to the external slots mixed gradually. While effective, the design with slots external to the burner throat was difficult to apply to furnaces. In 1972, B&W developed the dual register burner (DRB) for firing pulverized coal. The DRB (Fig. 8) provided two air zones, each controlled by a separate register, around an axially positioned coal nozzle. A portion of the secondary air was admitted through the inner air zone to aid ignition and flame stability. The remainder passed through the outer air zone and was mixed downstream in the flame. Recirculation induced by the swirled outer air also improved flame stability. The burner nozzle was equipped with a venturi to disperse the primary air-coal mixture. The mixture was then injected into the throat without deflection. The results were a stable flame with a fuel-rich core and gradually completed mixing downstream. The first unit retrofit was completed in 1973 and achieved 50% NO_x reduction relative to prior circular burners. This retrofit required no modifications to boiler pressure parts, fans, or pulverizers, and NO_x ports were not used. The U.S. government granted a patent to B&W for the DRB in January 1974. Numerous variations have been used by B&W since its introduction.

The DRB is often used in combination with enlarged furnace combustion zones to reduce thermal NO_x and with a compartmented windbox to better control air distribution (Fig. 8). This design has the burners coupled with each pulverizer situated in an individual windbox compartment. Secondary air is metered and controlled separately for each compartment. This approach corrects much of the flow imbalance among burners that was experienced in open windbox designs. High combustion efficiency with minimal burner zone slagging results.

The DRB typically reduces NO_x 50 to 60% from uncontrolled levels. This efficiency enables its use on new boilers without NO_x ports or other air diversion systems. Utility boiler NO_x emissions with the DRB range from 0.27 to 0.70 lb/10^6 Btu (332 to 861 mg/Nm^3).

DRB-XCL™ burner

The dual register XCL burner is an advanced version of the original DRB. The DRB-XCL™ is reconfigured mechanically as shown in Fig. 9. A DRB-XCL™ burner being prepared for shipment is shown in Fig. 10. Air flow to the burner is regulated by a sliding disk similar to that used on the S-type burner. An impact/suction pitot tube grid is located in the burner barrel to measure secondary air flow. With this information, the secondary air can be uniformly distributed among all burners by adjusting the sliding disks or other means. Balanced air and

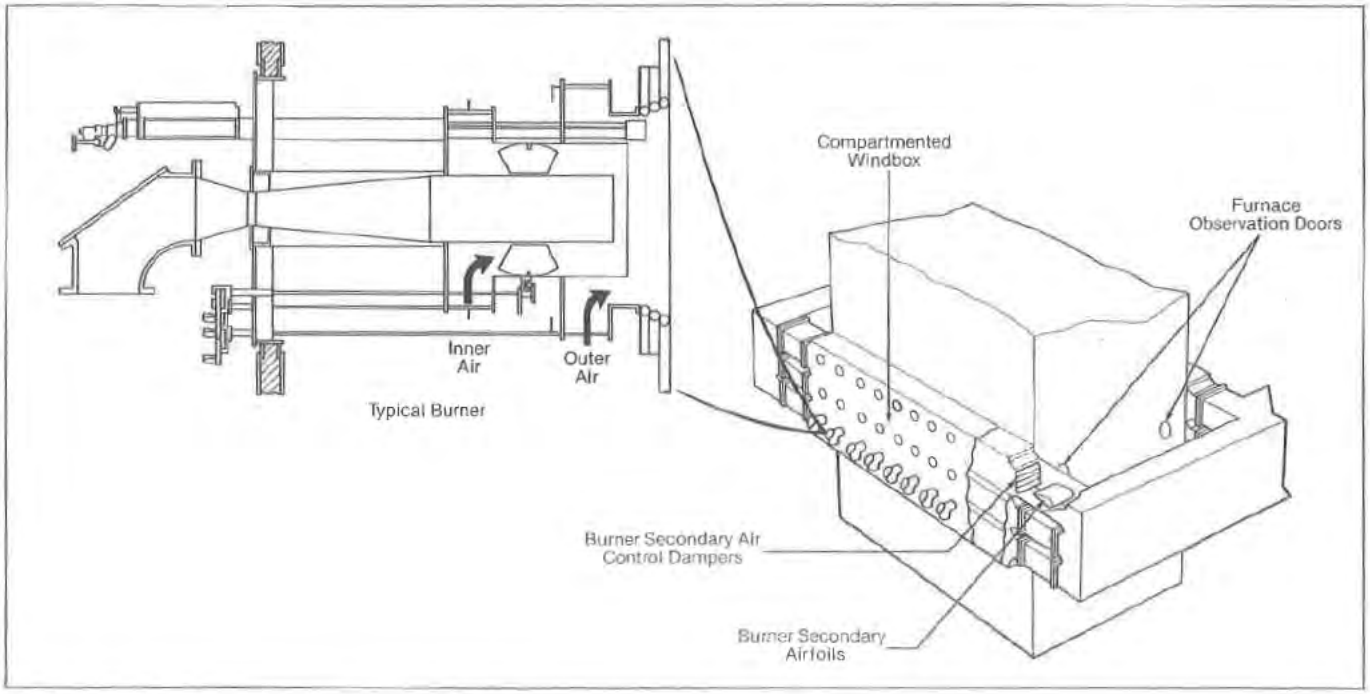


Fig. 8 Dual register burner and compartmented windbox.

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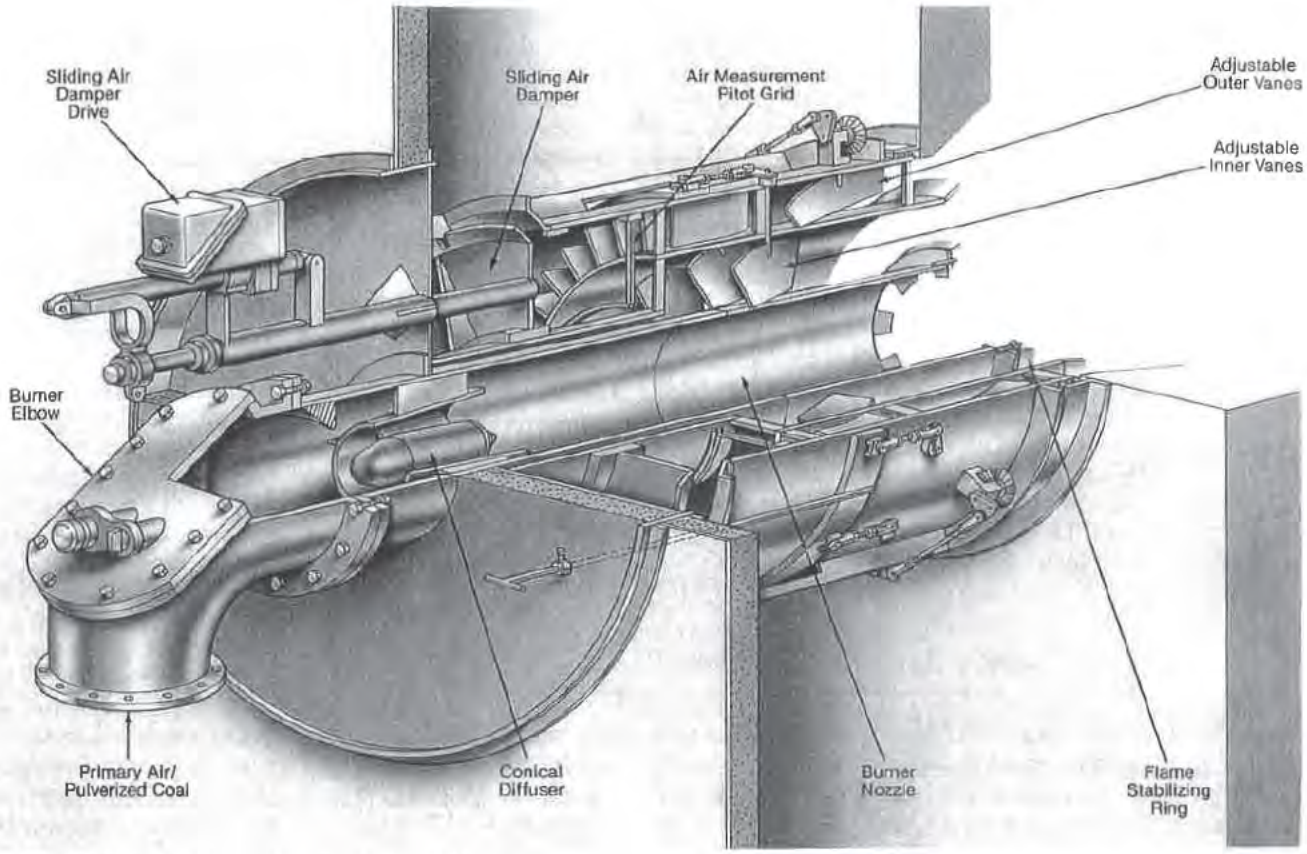


Fig. 9 DRB-XCL™ low NO_x burner for pulverized coal firing.

fuel distribution among all burners is critical to combustion efficiency, particularly with low NO_x systems. Downstream of the pitot tube grid are the inner and outer air zones. Adjustable vanes in the inner zone stabilize ignition at the burner nozzle tip. The outer zone, the main air path, is equipped with two stages of vanes. The upstream set is fixed and improves peripheral air distribution within the burner. The downstream set is adjustable and provides proper mixing of this secondary air into the flame. The burner nozzle is equipped with a conical diffuser and flame stabilizing ring. These combine to improve flame stability while further reducing NO_x by incorporating fuel staging technology. The DRB-XCL™ burner reduces NO_x 50 to 70% from uncontrolled levels without using NO_x ports. This level of NO_x reduction is possible through the creation of the flame combustion zones as shown in Fig. 11. The burner is specifically designed to generate rapid heating and high temperatures in the fuel rich flame core. This causes more of the coal to burn as volatile matter and releases a larger portion of the fuel nitrogen early in the combustion process, leaving less in the char. By limiting available oxygen in the flame core, NO_x formation is minimized. At the same time, reducing species are generated from the volatile materials. These propagate into the flame to aid in reducing NO_x formed in the later zones of the flame. Char oxidation occurs last at reduced temperature and oxygen concentrations, thereby limiting NO_x formation during char burnout. NO_x ports can be added when further NO_x reduction is required. In retrofit applications, the DRB-XCL™ burner can be equipped with modified coal impellers to shape the flame to existing furnaces.

The air and fuel staging technology of the DRB-XCL™ burner is also well suited for firing fuel oil or natural gas and for multi-fuel applications with pulverized coal. Chapter 10 discusses the oil and gas performance.

Low NO_x Cell™ burner

The cell burners discussed earlier produced high NO_x emissions, typically 1.0 to 1.8 lb/10^6 Btu (1230 to 2214 mg/Nm^3). The unique arrangement of the cell burner (Fig. 7) was incompatible with direct replacement by standard

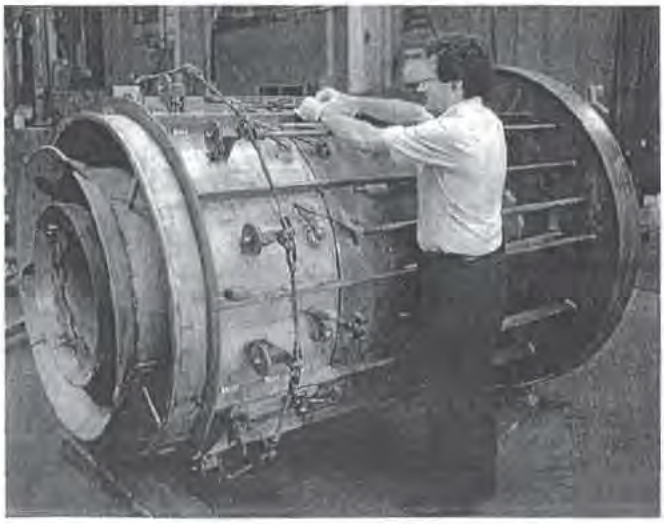


Fig. 10 DRB-XCL™ burner being prepared for shipment.

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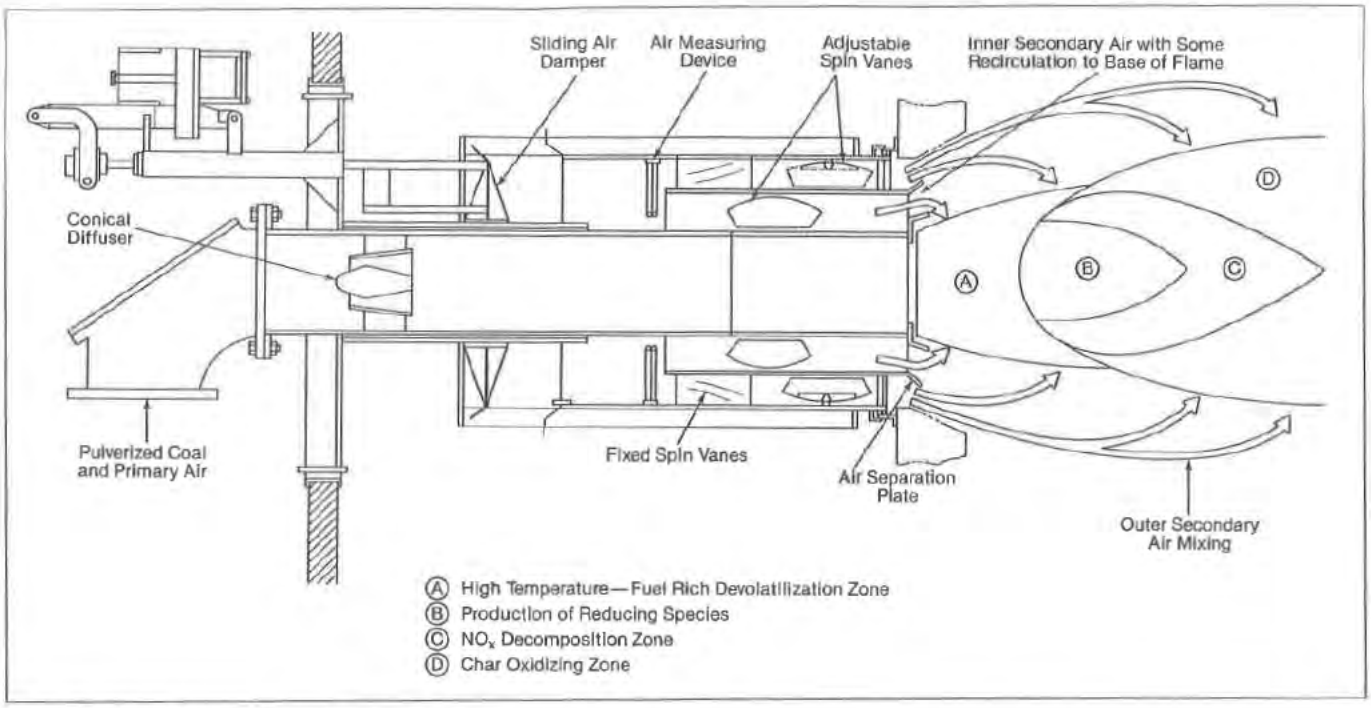


Fig. 11 DRB-XCL™ low NO_x combustion zones.

low NO_x burners. The close spacing of the burner throats prevented direct installation of larger, multi-zone low NO_x burners. One option is to rearrange the burners, increasing space between them and installing the DRB-XCL™ low NO_x burner. This solution is effective but costly. Costs increase due to the need for new boiler tube wall panels, coal piping, burners, ignition system plus windbox alterations. Another alternative is the Low NO_x Cell™ burner (LNCB™). (See Fig. 12.) This burner is designed to fit into existing cell burner locations and reduce NO_x. The LNCB™ was conceived by B&W and developed in cooperation with the Electric Power Research Institute. In essence, all of the coal is supplied to the lower burner throat with a portion of the secondary air. The balance of the secondary air is supplied to the upper burner throat. The upper throat serves as an integral NO_x port for each burner location. The coal piping is altered to supply all the coal to the lower throat, usually by joining the pipes in a Y connection. A larger burner nozzle is installed in the lower throat and can be equipped with an impeller to shape the flame. The lower throat burner assembly is essentially a conventional S-type burner with an oversized nozzle. The upper throat assembly is equipped with a sliding disk and pitot tube grid to set secondary air flow. It is also equipped with vanes to control mixing of this air into the flame. NO_x levels can be reduced by 50% with the LNCB™ while maintaining high combustion efficiency. Retrofit costs are reduced because wall panel and windbox alterations are avoided and because coal piping alterations are reduced.

NO_x ports

Removing a portion of the secondary air from PC-fired burners effectively reduces NO_x emissions as discussed above. The air is diverted to ports which introduce it later in the combustion process. These ports may be located close to the burners, as in the LNCB™, or at some distance from all of the burners. In the majority of applications, the ports are placed above the burner zone in furnaces arranged for gases to travel upward and out. Such ports are sometimes referred to as overfire air (OFA) ports. In some applications, ports are placed beneath or within the burner zone.

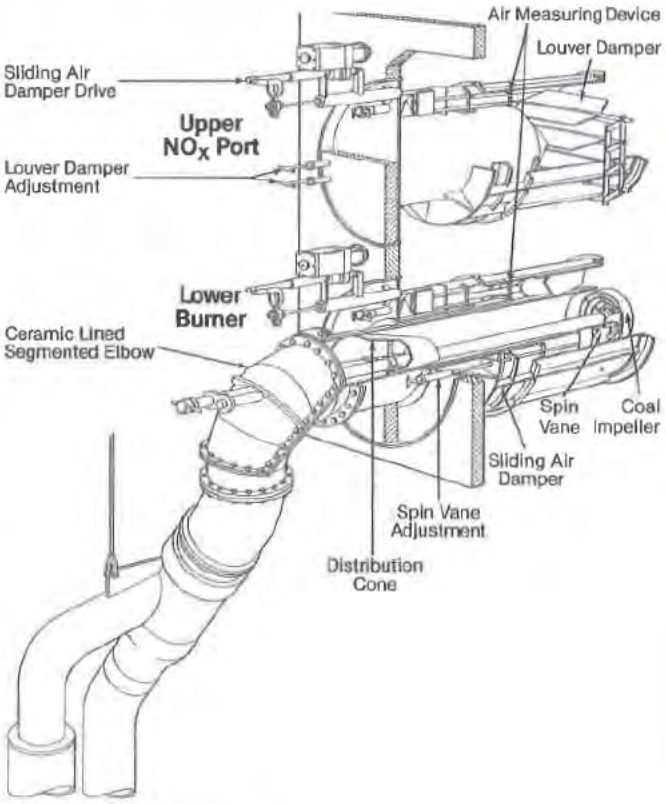


Fig. 12 Low NO_x Cell™ burner.

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B&W pioneered the development and application of NO_x ports and air staging as NO_x emission controls. This work, which began in the 1950s, was initially directed at NO_x reduction for oil- and gas-fired boilers in California. R.M. Hardgrove of B&W filed a patent for a *Method for Burning Fuel* on June 18, 1959, disclosing key aspects of air-staging influences on NO_x emissions. The patent was granted in 1962, and NO_x ports have been used subsequently by B&W.

Despite the proven effectiveness of NO_x ports, they are not always used with pulverized coal. This is due to the potential for increased slagging and corrosion in the furnace and the loss of combustion efficiency. The H_2S corrosion potential is a concern with higher sulfur coals, particularly those exceeding 2 lb/10^6 Btu (860 g/GJ) sulfur content. High and severe slagging bituminous coals are not suitable for NO_x ports due to the potential for wet or plastic slag forming in the furnace. This slag can block burner throats or the furnace hopper and can be corrosive to boiler tubes. Protective coatings of some stainless steels or aluminum can greatly reduce the corrosion potential. However, substoichiometric operation with high sulfur coals is not recommended. Low NO_x burners alone, or those with slight staging to maintain a stoichiometry of at least 1.05 at the burners, are better solutions with such coals.

To maintain high combustion efficiency, NO_x port jet size and velocity must be sufficient to provide penetration across the furnace and mixing with crossflow gases. However, high velocity jets can miss combustion product gases flowing between the ports. B&W developed the Dual Zone NO_x Port (Fig. 13) to address this problem. The inner zone of the port provides a jet to penetrate across the furnace. The outer zone has vanes to deflect air and entrain nearby gases. The ports have adjustable disks and vanes to permit tuning during commissioning. Each port is also equipped with pitot tubes to measure flow and to allow uniform air distribution among ports. Systems intended to operate at low burner zone stoichiometries, e.g., 0.70 to 0.85, benefit from two levels of NO_x ports. This better controls introduction of oxygen in the latter stages of combustion and minimizes associated NO_x formation. NO_x port system performance benefits from the inclusion of equipment to measure and regulate flow to both the burners and the ports through the boiler load range.

An emerging technology which facilitates this integration is numerical modeling. (See Chapters 3 and 4.) Computer programs provide detailed information about mixing effectiveness throughout the furnace. These programs can be used to optimize NO_x port size and placement, burner swirl orientation, and furnace geometry. Major improvements in mixing can be achieved by this parametric evaluation;[5] however, models must be validated with test data.

Burners for difficult coals

A few coals have proven to be considerably more difficult to burn than others, including unreactive coals and reactive coals which would burn readily except for an overabundance of moisture and/or ash. B&W has developed burners specifically for use with these problem coals. The selection of burner type is based on an empirical ignition factor derived from full scale experience with difficult coals. This factor relates variations in fuel ratio (FC/VM), moisture content, ash content, and coal heat-

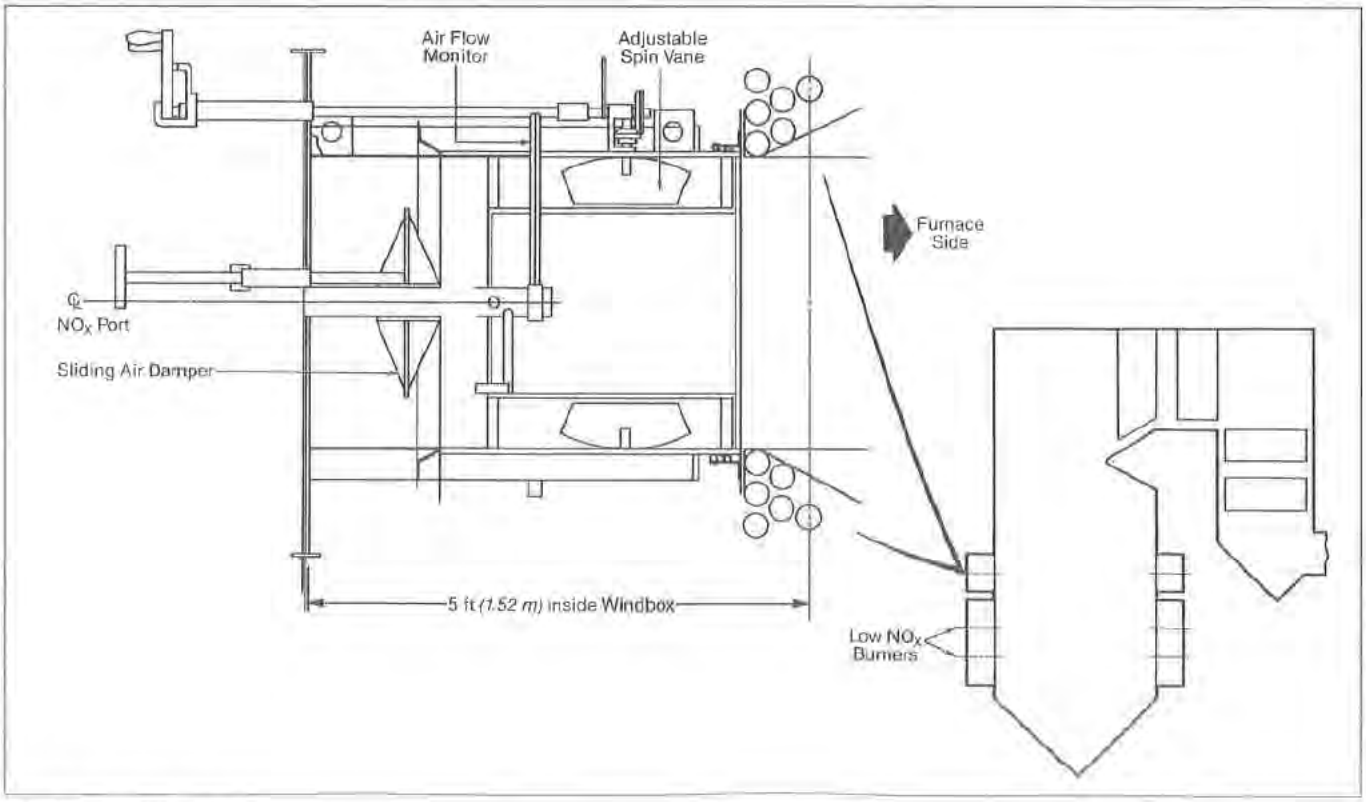


Fig. 13 Dual Zone NO_x Port.

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ing value to ignition behavior. Burner choices for these applications include conventional and low NO_x burners discussed above, the enhanced ignition (EI) burner, and the primary air exchange (PAX) burner.

Enhanced ignition burner

The enhanced ignition (EI) burner resembles the DRB-XCL™ burner. However, the EI burner uses a larger coal nozzle to reduce the primary air and coal mixture velocity as it enters the furnace. The conical diffuser produces a fuel-rich ring and a fuel-lean core; at low velocity, this combination is readily ignited. The dual air zones are adjusted relative to the DRB-XCL™ burner to increase recirculation of hot gases along the fuel jet. The EI burner is well suited for firing high moisture (more than 35%) lignites and low volatile bituminous coal.

Primary air exchange burner

The primary air exchange (PAX) burner is designed to fire anthracite coals. Some anthracites can be wall-fired conventionally with the PAX burner, while those lowest in volatile matter must be fired in a downshot arrangement (Fig. 4). The register design for the PAX burner is similar to that of the EI burner. However, in the case of downshot firing, air staging ports direct air away from the burner to raise flame temperature.

The anthracite fineness is increased to improve combustion efficiency. In addition, mill exit temperature is increased to the maximum for the pulverizer, often limited to 200F (93C) for mechanical reasons. Tests have shown the benefit of higher temperatures on ignition performance. The PAX burner raises the primary air and pulverized coal (PA/PC) temperature by a patented design which separates half of the primary air from the PA/PC mixture and vents it into the furnace. This primary air is replaced by secondary air near 700F (371C), significantly raising fuel temperatures as it enters the furnace. The PAX nozzle tip is also sized for low velocity to further assist ignition. Even with these techniques, good combustion is dependent upon high furnace temperatures. Low load operation still requires lighters for flame stabilization.

Auxiliary equipment

Oil/gas firing equipment

In some cases, PC-fired furnaces are required to burn fuel oil or natural gas up to full load firing rates. These fuels can be used when the PC system is not available for use early in the life of the unit, due to an interruption in coal supply, or as a NO_x control strategy. Some operators avoid installing a spare pulverizer by firing oil or gas when a mill is out of service.

To fire oil, an atomizer is installed axially in the burner nozzle. Erosion protection for the atomizer is recommended. A source of air is needed when firing oil or gas to purge the nozzle and improve combustion. This air system must be sealed off prior to PC firing. Steam-assisted atomizers are recommended for best performance.

Gas elements of several designs can be used in PC-fired burners. The burner gas manifold can be inside or outside of the windbox. Installation in the windbox clears the burner front of considerable hardware but prevents on-line adjustment or repair of gas elements. Gas elements are equipped with flame stabilizing cans to protect the root of the flame at each spud. Retrofitting multiple gas elements into PC-fired burners can disrupt air flow in the throat and impair performance. Compatibility of the gas elements with the PC burner design is important. Retrofit of gas or oil elements increases the forced draft fan load in many cases. Primary air is eliminated which results in higher quantities of secondary air compared to PC. As a result, fan flow and static margins need to be considered.

See Chapter 10 for more information on oil and gas firing equipment.

Lighters

A lighter, or ignitor, is required to initiate combustion as pulverized coal is first introduced to the burner, as the burners are being normally shut down, and as otherwise required for flame stability. Lighters typically use an electrically generated spark to ignite a lighter fuel, which is usually natural gas or No. 2 fuel oil. Most modern units have automated controls to activate, operate, purge, and shut down the lighters. Lighter systems are described in Chapter 10.

Flame safety system

Modern PC-fired boilers are equipped with a flame safety systems (FSS). The FSS uses scanners at each burner to electronically monitor flame conditions. Flame scanners are used to evaluate the lighter and main flames for intensity and frequency. The lack of satisfactory flame signals causes the FSS to automatically take corrective action, which is to shut down the burner and its associated group. This prevents unburned fuel from entering the furnace and significantly reduces the risk of an explosion. See Chapter 41 for additional information on the FSS.

Safety and operation

Uncontrolled ignition of pulverized coal may result in an explosion which can severely damage equipment and injure personnel. Explosions result from ignition of an accumulated combustible mixture within the furnace or associated boiler passes, ducts, and fans.[6,7] The magnitude and intensity of the explosion depend on the quantity of accumulated combustibles and the air-fuel mixture at ignition. Explosions result from improper operation, design, or malfunction. The National Fire Protection Association (NFPA) publishes standards in the following areas: *Pulverized Fuel Systems* (NFPA 85F) and *Prevention of Furnace Explosions in Pulverized Coal-Fired Multiple Burner Boiler-Furnaces* (NFPA 85E). These are excellent sources of information concerning design and equipment issues, operation and control of PC-fired equipment.

Another key source of information is the operating instructions. Operators of PC-fired equipment should expect the instructions to provide specific information concerning the purpose, design, calibration and adjustment, startup, operation and shutdown of the various equipment. The instructions also provide maintenance and troubleshooting information.

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Other PC applications

Pulverized coal in the metals and cement industries

The application of pulverized coal firing to copper- and nickel-ore smelting and refining has been standard practice for many years. With the use of pulverized coal, high purity metal can be obtained because the furnace atmosphere and temperature can be easily controlled. Pulverized coal may be favored over other fuels for several reasons: it may be less expensive, it offers a high rate of smelting and refining, and it readily oxidizes sulfur.

Copper reverberatory furnaces for smelting and refining are fitted with waste heat boilers for steam generation. These boilers supply a substantial portion of power for auxiliary equipment and also provide the means for cooling the gases leaving the furnace.

In producing cement, the fuel cost is a major expense item. Except in locations where the cost favors oil or gas, pulverized coal is widely used in the industry. Direct firing, with a single pulverizer delivering coal to a single burner, is common practice in the majority of U.S. cement plants. The waste heat air from the clinker cooler, taken from the top of the kiln hood through a dust collector, usually serves as the preheated air for drying the coal in the pulverizer.

Pulverized coal has recently been introduced in the steel industry where it is injected into the blast furnace tuyères. The pulverized coal replaces a similar weight of higher priced coke. Theoretically, 40% of the coke could be replaced by pulverized coal.

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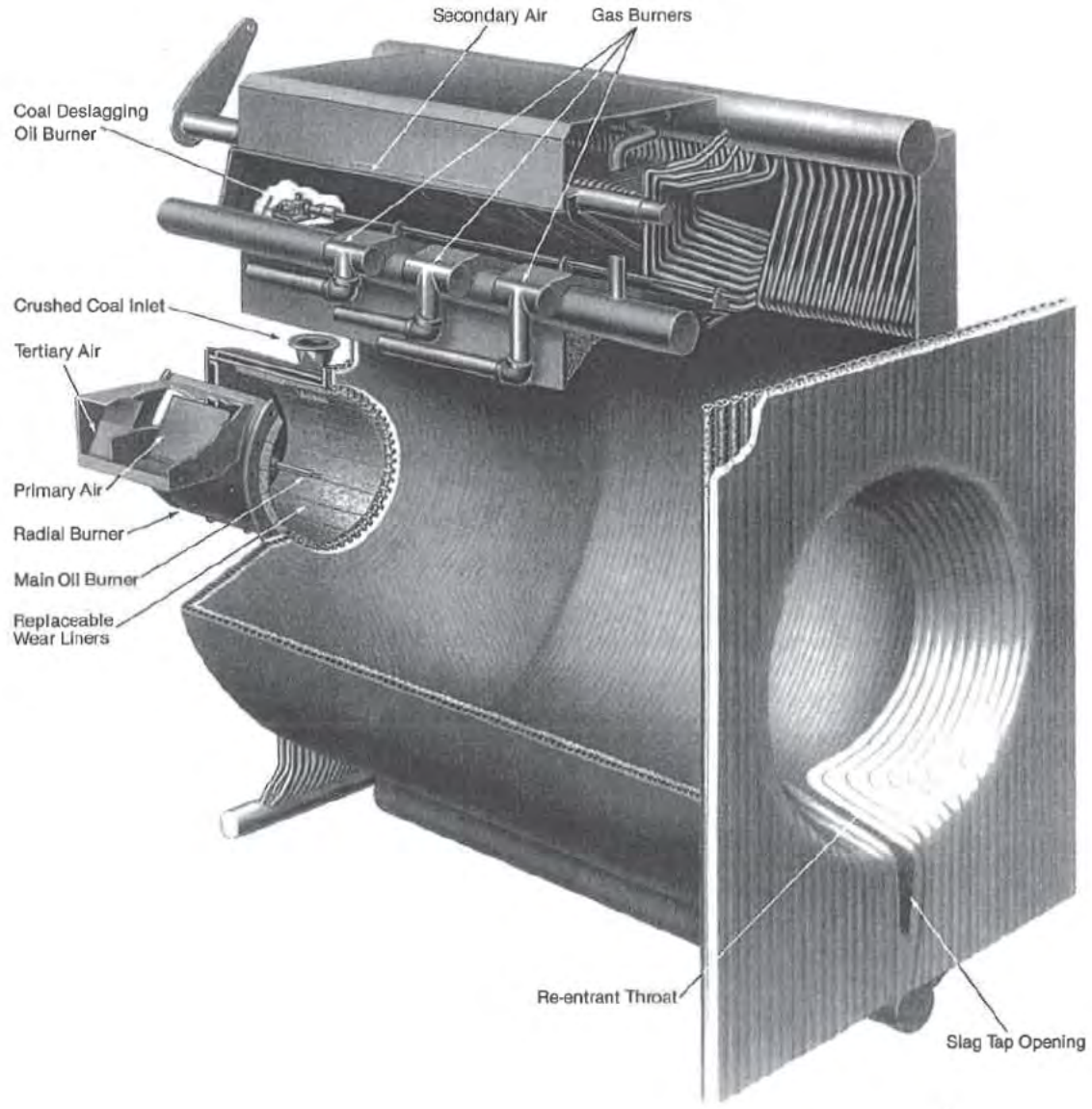
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Cyclone furnace.

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Chapter 16
Atmospheric Pressure Fluidized-Bed Boilers

The fluidized-bed process

Fluidized-bed combustion (FBC) technology has distinct advantages for burning solid fuels and recovering energy to produce steam. The process features a mixture of particles suspended in an upwardly flowing gas stream, the combination of which exhibits fluid-like properties. Combustion takes place in the bed with high heat transfer to the furnace and low combustion temperatures. Key benefits of this process are fuel flexibility and reduced emissions. FBC is technically accepted for industrial sized units and is emerging as an accepted technology for electric power generation.

The following example will help visualize the fluidized-bed process. Fig. 1a shows a container with an air supply plenum at the bottom, a distributor plate that promotes even air flow through the bed, and an upper chamber filled with sand or other granular material.

If a small quantity of air flows through the distributor plate into the sand, it will pass through the voids of an immobile mass of sand. For low velocities, the air does not exert much force on the sand particles and they remain in place. This condition is called a fixed bed and is shown in Fig. 1b.

By increasing the flow rate, the air exerts greater forces on the sand and thereby reduces the contact forces between the sand particles caused by gravity. By increasing the air flow further, a point is reached where the drag forces on the particles counterbalance the gravity forces and sand particles become suspended in the upward flowing air stream. The point where the bed starts to behave as a fluid is called the *minimum fluidization condition*. As shown in Fig. 1c, the increase in bed volume is insignificant when compared with the nonfluidized case.

As the air flow increases further, the bed becomes less uniform, bubbles of air start to form and the bed becomes violent. This is called a *bubbling* fluidized bed which is shown in Fig. 1d. The volume occupied by the air-solids mixture increases substantially. For this case, there is an easily seen bed level and a distinct transition between the bed and the space above.

By increasing the air flow further, the bubbles become larger and begin to coalesce, forming large voids in the bed. The solids are present as interconnected groups of high solids concentrations. This condition is called a *turbulent* fluidized bed.

If the solids are caught, separated from the air and returned to the bed they will circulate around a loop. This type of system is defined as a *circulating* fluidized bed (CFB) and is shown in Fig. 1e. Unlike the bubbling bed, there is no distinct transition between the dense bed in the bottom of the container and the dilute zone above. The solids concentration gradually decreases between these two zones.

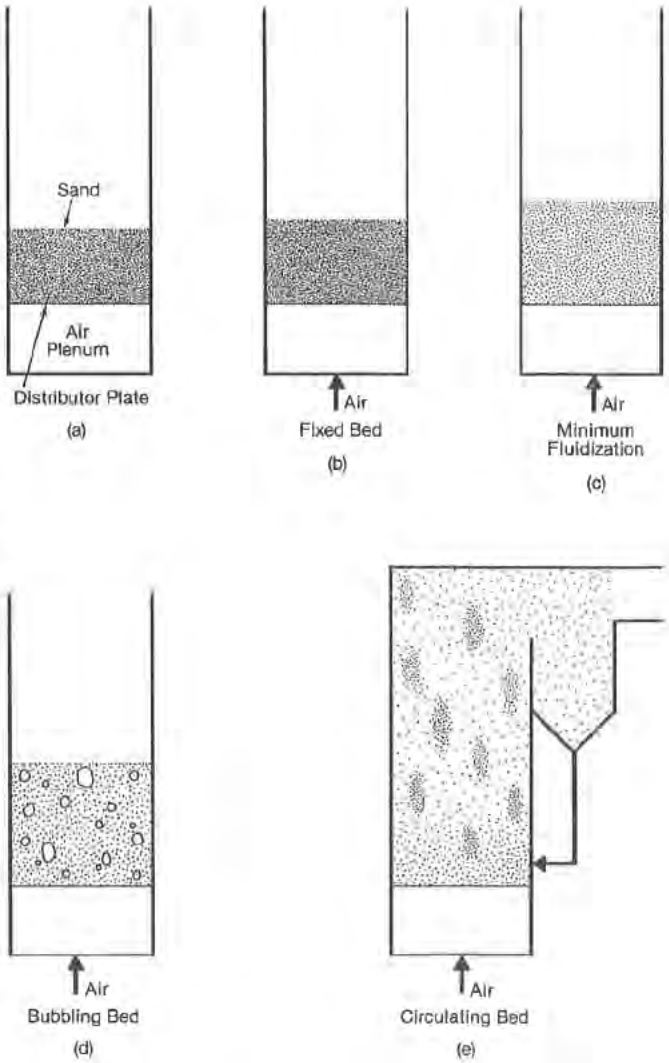


Fig. 1 Typical fluidized-bed conditions.

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The weight of solids recirculated from the outlet back to the bed zone can be hundreds of times the weight of air flowing through the system, and the quantity of solids in the container is proportional to the amount of sand recycled from the collector. As a result, the pressure differential will increase to the value required to support the solids in the container.

The pressure differential between the top and the bottom of the container changes with air flow, as shown in Fig. 2. At low air flow rates, the pressure differential increases with flow until the minimum fluidization velocity is reached. At this point, the sand is supported by the air and the pressure differential is determined only by the mass of sand in the bed. The pressure differential is independent of further increases in air flow until the air velocity becomes high enough to convey the sand out of the container. At higher air flows, the pressure differential starts to decrease as mass is lost from the system.

Of the fluidization conditions just described, only bubbling and circulating beds are currently used by the power industry to generate steam.

Background and evolution

One of the earliest recorded applications of a fluidized bed used coal as the feed stock. In the mid 1920s, the fluidized-bed coal gasification process, invented by Fritz Winkler, was used commercially to generate gas from coal. (See also Chapter 17.) This gas was then used for fuel or as a feed stock for chemical syntheses. While there are a few Winkler gasifiers operating today, industry has found it both easier and less expensive to use natural gas or oil as fuels and for synthesis gas production.

During the late 1930s and early 1940s, a large research and development effort was directed to fluidized beds. This work identified the advantages of a fluidized bed as a solids-gas contacting device and led to development of the fluid catalytic cracker to produce gasoline and other petroleum based products. Today, fluidized beds are used worldwide for a variety of processes in many industries.

In the early 1960s, national attention turned to reducing sulfur dioxide (SO_2) and nitrogen oxides (NO_x) emissions from power plants. The fluidized-bed combustion process offered the potential for reducing these emissions. The effort to develop the coal-fired fluidized-bed boiler began.

In the early 1970s, Babcock & Wilcox (B&W) conducted an extensive study to evaluate the application of atmospheric pressure fluidized-bed combustion to large boilers. In 1977, B&W, in cooperation with the Electric Power Research Institute (EPRI), constructed and placed into operation a 6 × 6 ft (1.8 × 1.8 m) bubbling-bed test unit at the B&W Research Center in Alliance, Ohio. Test results from this facility have contributed significantly to the advancement of bubbling fluidized-bed boiler technology. Recent applications of large, coal-fired bubbling beds have included retrofits to existing boilers.

In the early 1980s, the market for coal-fired fluidized-bed boilers moved toward circulating beds. To meet this changing market, B&W used its knowledge of bubbling-bed technology and the technical concepts of Studsvik AB of Sweden to advance into the circulating-bed field. A schematic of a pilot-scale circulating fluidized bed built at B&W's research center is shown in Fig. 3.

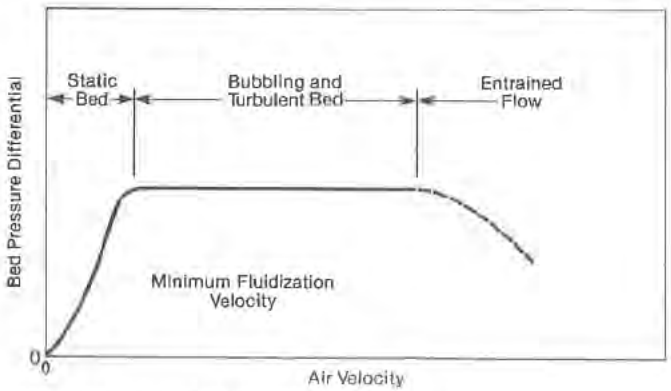


Fig. 2 Effect of velocity on bed pressure drop.

Comparison with other combustion methods

The fluidized-bed combustion process, as with other firing methods, provides a means for mixing fuel with air to convert the chemical heat contained in the fuel into recoverable, sensible heat. Normally, fluidized-bed combustors are used to burn solid fuels.

In a pulverized coal-fired furnace, the combustion process consists of oxidizing fine (70% less than 200 mesh), widely dispersed fuel particles suspended in air and combustion gases. (See Chapter 13.) The volume around the burners is the hottest zone in the furnace with temperatures reaching 3000 to 3500F (1649 to 1927C). Also, the residence time of the particles in the furnace is close to the flue gas residence time.

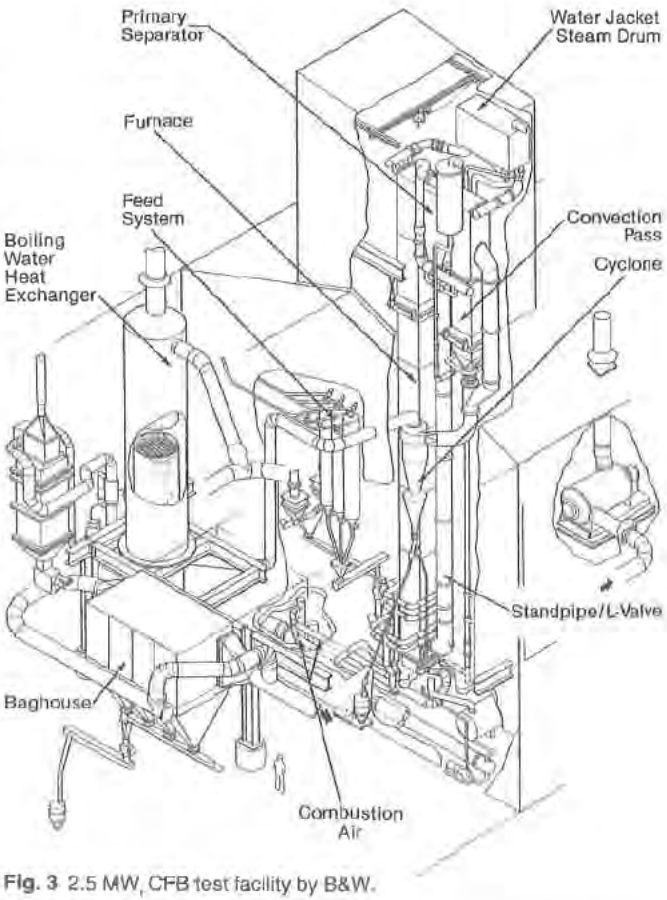


Fig. 3 2.5 MW, CFB test facility by B&W.

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Stoker firing uses considerably larger fuel particles than pulverized coal firing. Fuel sizing is typically 1 to 1.25 in. (25.4 to 31.8 mm) top size for bituminous coal. Most of the fuel is burned as an immobile mass on some type of moving grate, with the air and combustion gas passing through the fixed bed of fuel. (See Chapter 15.) Temperatures in the fuel bed exceed 3000F (1649C) and the fuel residence time is determined by the grate speed.

The fluidized-bed combustion process falls in between pulverized coal and stoker firing with respect to the size of the fuel feed. Coal is typically crushed to less than 0.25 in. (6.4 mm). Depending on coal properties larger coal, 1.25 in. (31.8 mm), or smaller coal, 0.125 in. (3.18 mm), is used. Fuel is fed into the lower portion of a fluidized-bed boiler furnace. The bed has a density of approximately 45 lb/ft^3 (721 kg/m^3) for a bubbling bed and 35 lb/ft^3 (561 kg/m^3) for a circulating bed. The solids are maintained at a temperature of 1500 to 1600F (816 to 871C) in an upwardly moving stream of air and combustion gas.

When fuel is introduced into the bed it is quickly heated above its ignition temperature, ignites and becomes part of the burning mass. The flow of the air and fuel to the dense bed is controlled so that the desired amount of heat is released to the furnace on a continuous basis. Typically, the fuel is burned with 20% excess air. Due to the long fuel residence time and high intensity of the mass transfer process, the fuel can be efficiently burned in the fluidized-bed combustor at temperatures considerably lower than in conventional combustion processes.

The fuel particles remain in the dense bed until they are entrained by combustion gas or removed with the bed drain solids. As the fuel particles burn, their size falls below a given value where the terminal and gas velocities are equal, which allows them to be entrained. Therefore, the residence time is determined by the initial fuel particle size and by the reduction of the initial size resulting from combustion and attrition.

In bubbling fluidized beds, combustion occurs mostly in the bed due to lower gas velocity and coarser fuel feed size. The residence time of the fine fuel particles carried out of the bed with the combustion gas is, in many cases, increased by collecting and recycling the particles to the furnace.

In circulating beds, more particles are blown from the bed (elutriated) than for a bubbling bed. The particles are then collected by a particle separator and recirculated to the furnace. The residence time of the particles is determined by the collection efficiency of the particle separator and the solids circulation rate. As a result of the recirculation process, the effective particle residence time greatly exceeds the gas residence time.

The concentration of fuel in the dense bed is normally quite low. For a reactive fuel such as wood, it is difficult to find a measurable amount of carbon in the bed. Normally, the carbon content in a bed burning bituminous coal is less than 1%. The remaining portion of the bed is made up of fuel ash, lime and calcium sulfate when a sorbent is used for sulfur capture, and sand or other inert material when a sorbent is not used.

Overall carbon conversion efficiencies are near 100% for wood and highly reactive fuels, greater than 98% for bituminous coals, and slightly lower for less reactive fuels and waste coals.

Fluidized-bed combustion advantages

The primary driving force for the development of fluidized-bed combustors in the United States (U.S.) is reduced SO_2 and NO_x emissions. By implementing this technology, it is possible to burn high sulfur coals and achieve low SO_2 emission levels without the need for additional back-end sulfur removal equipment. As the technology developed, it also became apparent that the process could burn low grade fuels that are difficult or impractical to burn with other methods.

Designers of coal burning equipment have known for many years that coal properties vary depending on the fuel type and source and that the firing equipment must be compatible with the specific fuel. While fluidized-bed combustors are more tolerant of these variations, the same basic rules of combustion apply, so fuel type, chemical composition and heating value must be considered on each design.

Fluidized-bed boilers are designed so that the bed operating temperature is between 1500 and 1600F (816 and 871C); the ability to operate at this low temperature results in several operating advantages.

Reduced emissions — SO_2 and NO_x

Because of lower operating temperatures, it is possible to use an inexpensive material, such as limestone or dolomite, as a sorbent to remove SO_2 from flue gas. When limestone or dolomite is added to the bed, a reaction occurs in the furnace between the resulting calcium oxide (CaO) and the SO_2 in the gas. SO_2 emissions can be reduced by 90% or more depending on the sulfur content of the fuel and the amount of sorbent added.

Nitrogen and oxygen will react at high temperatures, above 2700F (1482C), to form nitric oxide. The rate of reaction decreases rapidly as the temperature is reduced from this value. With an operating bed temperature between 1500 and 1600F (816 and 871C), the amount of NO_x formed in the fluidized bed is less than in conventional units which operate at higher temperatures. For some bubbling-bed and all CFB boilers, additional suppression of NO_x formation is achieved by air staging. Fluidized beds can operate in this manner with less impact on combustion efficiency than in pulverized coal furnaces. With the addition of postcombustion reduction techniques, even lower NO_x emissions can be achieved.

Fuel flexibility

Fuel ash properties In addition to reduced emissions, the lower combustion temperatures permit burning high fouling and slagging fuels at temperatures below their ash fusion temperature. As a result, many of the boiler operating problems associated with these fuels are greatly reduced. However, care is still required as high alkali metal concentrations in the bed may cause sintering and in-bed tube bundle fouling. In addition, excessive amounts of above-bed burning may cause a significant increase in the furnace exit gas temperature and cause deposits to form in the superheater. For these reasons, bed temperatures around 1500F (816C) are normally chosen for fuels whose ash is high in alkali metals.

Low Btu fuels The fluidized-bed combustion process can also burn fuels with very low heating values. This

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capability results from the rapid heating of the fuel particles by the large mass of hot bed material and the long residence time that the fuel spends in the bed, both of which offset the effects of lower combustion temperatures. When burning high moisture fuels it is also necessary to consider the additional gas weight, resulting from water vapor, in the design of the convection pass and other related components.

Fuel preparation For high ash coals, the fluidized-bed boiler offers an advantage in fuel preparation over pulverized coal systems. These fuels require greater installed pulverizer capacity and the pulverizers generally require frequent maintenance. The crushed fuel, less than 0.25 in. (6.4 mm), required for a fluidized-bed boiler is easier and less costly to prepare.

A fluidized-bed boiler can be designed to burn a wider range of fuels than alternate firing methods. However, once the boiler is designed for a specified set of fuels there are limitations to the degree of deviation from the specified values that can be fired before design limits are exceeded. A CFB boiler has greater fuel flexibility than a bubbling bed.

Atmospheric pressure fluidized-bed boilers

Bubbling bed

Fig. 4 shows the main features of a bubbling fluidized-bed boiler. The bed itself is usually 4 ft (1.2 m) deep in its expanded or fluidized condition. Fig. 5 shows a typical furnace bulk density profile curve. The sharp drop in density indicates the top of the bed.

Normally, the heat transfer surface, which is in the form of a tube bundle, is placed in the bed to achieve the desired heat balance and bed operating temperature. For fuels with low heating values the amount of surface can

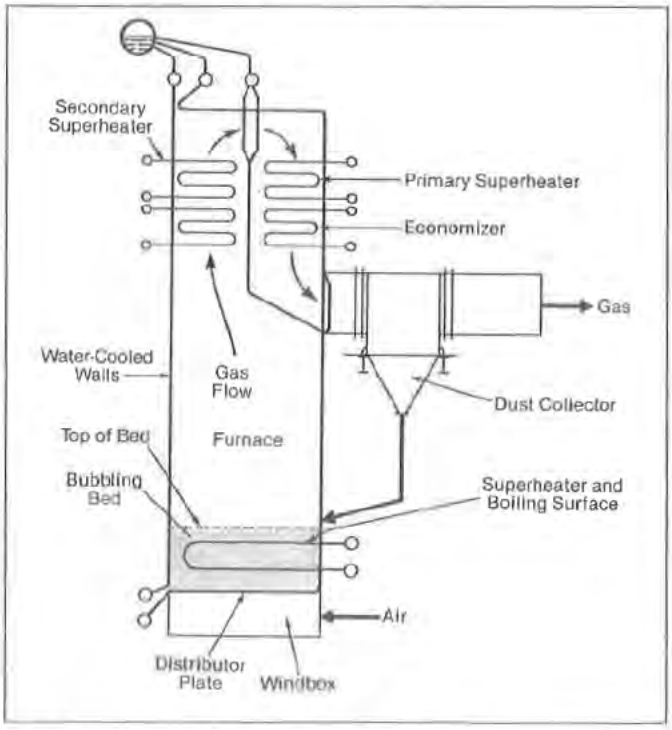


Fig. 4 Typical bubbling fluidized-bed boiler schematic.

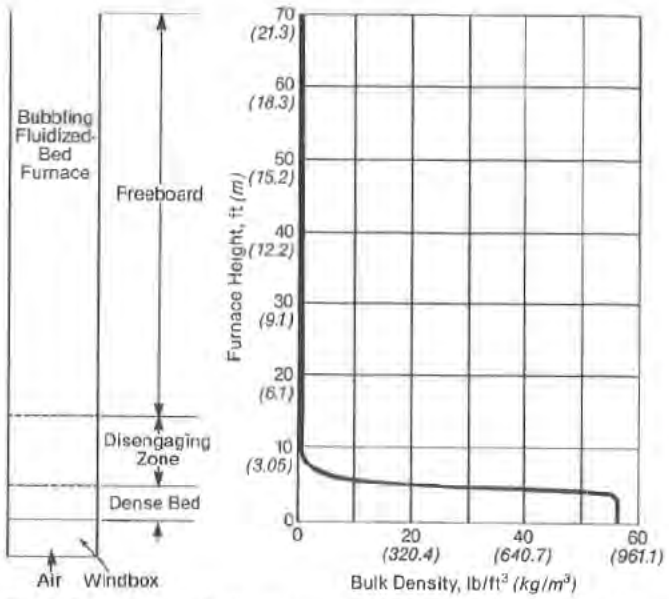


Fig. 5 Typical atmospheric pressure bubbling-bed furnace density profile.

be minimal or absent. In all cases, the bed temperature is uniform, plus or minus 25F (14C), as a result of the vigorous mixing of gas and solids.

Coal-fired bubbling-bed boilers normally incorporate a recycle system that separates the solids leaving the economizer from the gas and recycles them to the bed. This maximizes combustion efficiency and sulfur capture. Normally, the amount of solids recycled is limited to about 25% of the combustion gas weight. For highly reactive fuels this recycle system can be omitted.

Bubbling beds that burn coal usually operate in the range of 8 to 10 ft/s (2.4 to 3 m/s) superficial flue gas velocity at maximum load. The bed material size is 30 mesh (590 microns) and coarser, with a mean size of about 18 to 16 mesh (1000 to 1200 microns).

Circulating bed

Fig. 6 shows the main features of a circulating fluidized-bed boiler and Fig. 7 shows the furnace density profile. The dense bed does not contain any in-bed tube bundle heating surface. The furnace enclosure and internal division wall type surfaces provide the required heat removal. This is possible because of the large quantity of solids that are recycled internally and externally around the furnace. Because the mass flow rate of recycled solids is many times the mass flow rate of the combustion gas, furnace temperatures remain uniform. Also, the heat transferred to the furnace walls is adequate to provide the heat absorption required to maintain the target bed temperature of 1500 to 1600F (816 to 871C). B&W circulating fluidized-bed boilers usually operate at about 20 ft/s (6.1 m/s) superficial flue gas velocity at full load. The size of the solids in the furnace is usually smaller than 30 mesh (590 microns), with the mean particle size in the 150 to 200 micron range.

Emissions

U.S. federal and state governments have imposed limits on emissions from most large boilers and combustion processes. These emissions limits vary by region and gov-

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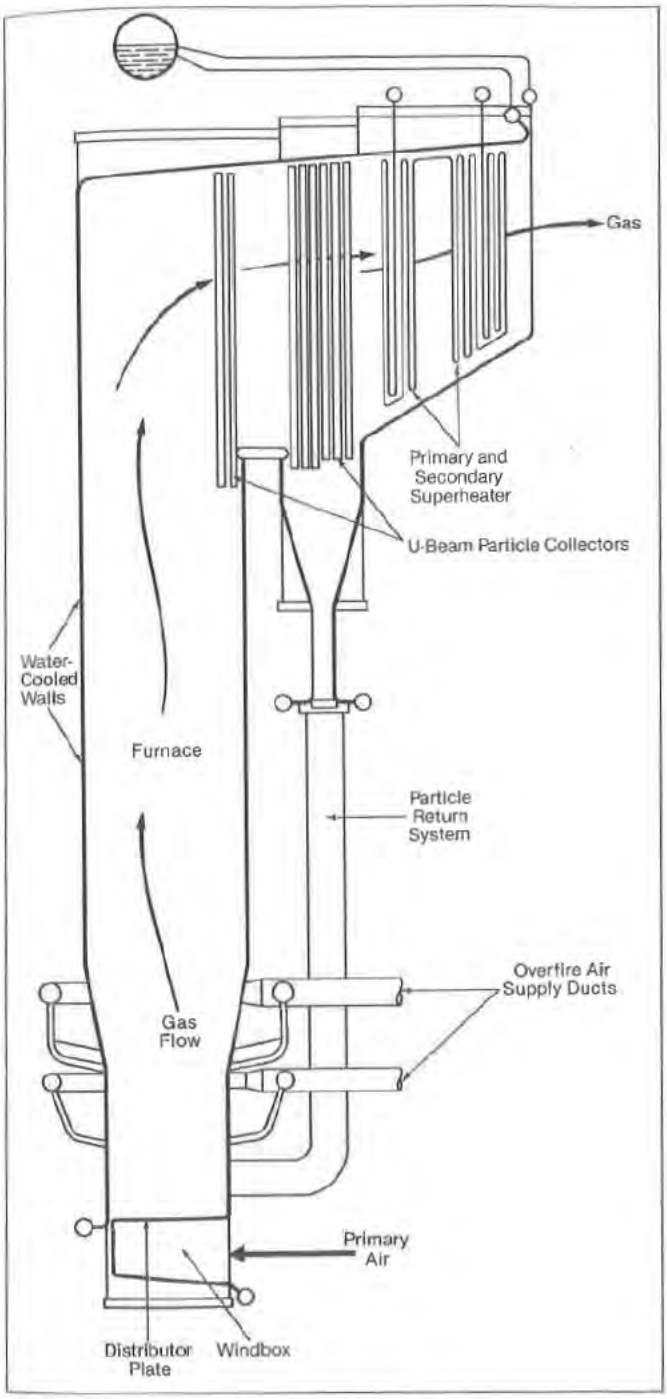


Fig. 6 Typical circulating-bed boiler schematic.

ernment, but the compounds and materials controlled are generally the same. These are SO_2, NO_x, carbon monoxide (CO), hydrocarbons and particulate matter. (See Chapter 32.) Fluidized-bed boilers are designed, primarily, to burn solid fuel while controlling many of these emissions.

Sulfur dioxide

When sulfur bearing fuels burn, most of the sulfur is oxidized to SO_2 which becomes a component of the flue gas. When limestone is added to the bed it undergoes a transformation called *calcination* and then reacts with the SO_2 in the flue gas to form calcium sulfate ($CaSO_4$). The calcining reaction is endothermic and is described by:

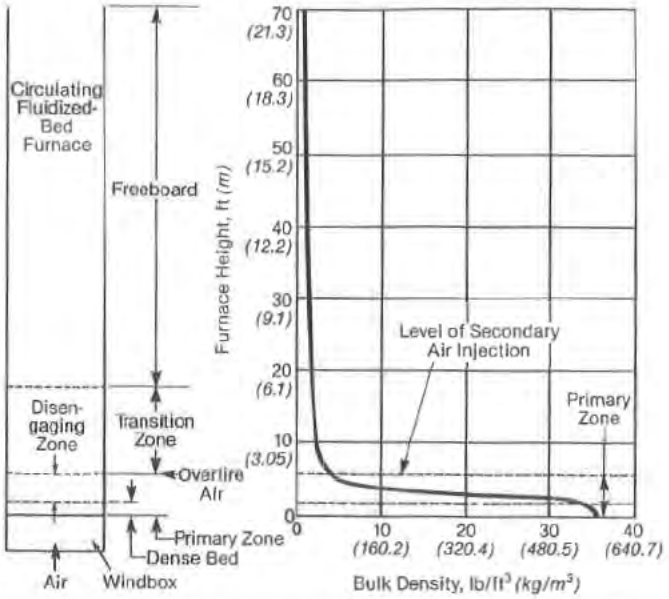


Fig. 7 Typical atmospheric pressure circulating-bed furnace density profile.

$$CaCO_3\,(s) + 766\ \text{Btu/lb (of } CaCO_3) \rightarrow CaO(s) + CO_2\,(g)$$

Once formed, solid CaO (lime) reacts with gaseous SO_2 and oxygen exothermically to form $CaSO_4$ according to the following reaction:

$$SO_2\,(g) + {}^1/_2\,O_2\,(g) + CaO\,(s) \rightarrow CaSO_4\,(s) + 6733\ \text{Btu/lb (of S)}$$

$CaSO_4$ is chemically stable at fluidized-bed operating temperatures and is removed from the system as a solid for disposal.

Early coal-fired fluidized-bed combustion work was carried out on a once-through principle. In this case, the coal and limestone were fed to the combustor, reacted and passed out of the system. Combustion efficiency and sulfur capture fell short of desired values. To overcome this, a portion of the solids leaving the furnace (flyash, $CaSO_4$, carbon and lime) is separated from the gas by a dust collector located between the economizer and the air heater and is recycled to the furnace for further reaction. Fig. 8 shows the effect of solids recycle on sulfur capture for a bubbling bed. Normally, recycle rates are limited to a maximum of 2.5 times the fuel feed rate which is a result of practical considerations of equipment size and arrangement.

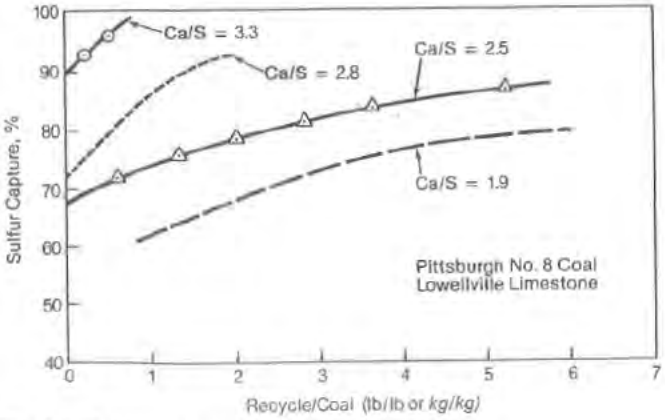


Fig. 8 Sulfur capture as a function of recycle ratio.

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In a CFB, a primary solids collector is located immediately after the furnace and is designed to recycle all of the captured solids. As a result, sulfur capture and combustion efficiency are improved over bubbling beds. Some designs also include a secondary collector after the convection pass which further enhances sulfur capture and combustion efficiency by increasing the concentration of fine particles in the furnace. These improvements result from longer solids-gas contact times and higher specific surface of the fine particles in contact with the gas.

SO_2 reductions of 90% are typically achieved in a circulating bed with calcium to sulfur (Ca/S) mole ratios of 2 to 2.5, depending on the sulfur content of the fuel and the reactivity of the limestone. Slightly higher ratios are required in bubbling beds. The lower the sulfur concentration in the fuel the greater the calcium to sulfur mole ratio must be for a given SO_2 removal level.

For removal requirements greater than 90%, the amount of limestone needed increases rapidly and alternative SO_2 removal methods, such as conventional pulverized coal-fired boilers with scrubbers, may become the economic choice. (See Chapter 35.)

Nitrogen oxides

The nitrogen oxides that are present in the flue gas come from two sources, the oxidation of nitrogen compounds in the fuel and the reaction between nitrogen and oxygen in the combustion air. It is common to refer to the nitrogen oxides in the flue gas as NO_x, where the subscript x implies the presence of several different nitrogen/oxygen compounds. The NO_x formed by oxidizing fuel nitrogen compounds is referred to as *fuel* NO_x and the NO_x formed from nitrogen and oxygen in the combustion air is called *thermal* NO_x, because it is a product of a high temperature process above 2700F (1482C).

As a result of the low temperatures at which a fluidized bed operates, thermal NO_x makes only a minor contribution to overall emissions. A fluidized-bed boiler can also suppress the amount of fuel NO_x formed. This is accomplished by putting less than the theoretical amount of combustion air through the distributor plate and adding the remainder of the combustion air to the furnace above the dense bed. As a result, some of the fuel nitrogen compounds decompose into molecular nitrogen rather than forming NO_x. This process is referred to as *staged combustion* and is also used with other firing methods to reduce NO_x emissions.

Staged combustion is used for bubbling beds that do not contain in-bed surface and for circulating beds. However, it is not used for designs where cooling tube bundles are submerged in the bed. The reason is corrosion. By burning the fuel with less than theoretical air (substoichiometrically) the combustion gas contains many reduced chemical species that cause rapid metal loss from the furnace cooling tubes. When staged combustion is used, the furnace tube walls in the reducing zone are protected with a thin layer of refractory.

The combination of low temperatures and staged combustion permits fluidized-bed boilers to operate with significantly lower NO_x emissions. Typical values for NO_x emissions are within the 100 to 200 ppmdv (parts per million dry volume) range for a CFB boiler burning coal.

Carbon monoxide and hydrocarbons

When designing a boiler, it is necessary to maximize combustion efficiency by minimizing unburned carbon and the quantity of CO and hydrocarbons in the flue gas. This is done by choosing the proper number of fuel feed points, by proper design of the overfire air system, if used, and by providing sufficient furnace residence time for mixing and complete combustion (burnout).

Typical flue gas concentrations are less than 200 ppmdv for CO and 20 ppmdv for hydrocarbons in a CFB boiler burning coal.

Particulates

The ash contained in solid fuel is released during the combustion process. Some of this ash remains in the fluidized bed and is discharged by the bed material removal or *drain* system. This ash is normally larger than 140 mesh (105 microns) and is easy to handle and transport. The remaining ash leaves the boiler in the flue gas. This material is typically less than 325 mesh (44 microns) and requires a high efficiency collection device. Normally, a fabric filter is used with atmospheric pressure fluidized-bed boilers because it is less sensitive to the ash properties, such as size, concentration and resistivity, than electrostatic precipitators (ESP). (See Chapter 33 for a discussion of particulate collection.)

Fluidized-bed boiler furnace design

Numerous factors and parameters affect fluidized-bed furnace design. Some are specified by the boiler owner/operator. Many are selected by the designer and can only be derived from empirical data. Some are interrelated, making the design task more challenging.

The following list is typical of the type of information used when establishing the design of a fluidized-bed boiler.

Owner specified

1. unit capacity — steam flow requirements,
2. fuel — type, ash and moisture content, size, reactivity, chemical analyses, attrition characteristics, and fouling and sintering properties,
3. limestone — type, reactivity, size and attrition characteristics,
4. sulfur capture requirements,
5. NO_x emission limits, and
6. load turndown range.

Designer specified

1. type of fuel feed system — in-bed, under-bed or above-bed,
2. number and location of fuel feed points,
3. fuel combustion efficiency,
4. number and location of sorbent injection points,
5. primary/secondary air split when used and the location of the overfire air nozzles,
6. bed operating temperature,
7. bed operating velocity,
8. size of bed particles,
9. amount of solids that leaves through the bed drain and the amount that leaves with combustion gas to the final solids collector, and
10. amount, temperature and location of solids recycled to the furnace from particle separators and bed drain classifiers.

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The following discusses some of the major areas involved with the functional design of fluidized-bed furnaces. The design of the convection pass surface is similar to a conventional boiler design process as discussed in Chapter 18.

Importance of bed material

Combustion and inventory requirements For a fluidized bed to operate properly it must be continuously supplied with a sufficient quantity of particles of the proper size distribution. If the particles are too coarse, the bed will defluidize and become fixed. If the particles are too fine, they will blow out of the furnace making it impossible to maintain an adequate bed inventory. There is a range of bed particle size which is needed to maintain a stable fluidized-bed process. The supply and retention of these particles must be controlled to provide the required inventory.

In the case of a bubbling bed, it is easy to visualize that a properly fluidized bed must be present to receive, suspend, mix and burn the fuel particles. If the bed density is too lean or if the bed is too shallow, an accumulation of fuel particles is possible in a localized area of the bed. This may lead to high temperature zones where the fuel and ash will turn into a sintered mass. At the other extreme, if too many large particles are fed into the bed it will defluidize with a similar result.

In circulating beds, even though the solids inventory is distributed over the furnace height, a dense bed is still required in the lower furnace to support and mix the fuel during combustion to avoid the problems just mentioned.

Normally, when coal is burned, most of the ash is liberated from the burning pieces as fine particles. These are so fine that they quickly blow out of the bed and are carried out of the furnace by the flue gas.

In bubbling beds, this material does not contribute to bed inventory. In circulating beds, this material is caught and returned to the furnace where it becomes part of the circulating mass. However, the fines do not make a significant contribution to the dense bed itself.

Because of the wide variation in fuel ash properties, the ash is not usually depended upon to form a stable bed, and a second inert material of the desired size distribution such as sand is added to the system. When SO_2 capture is required, limestone replaces the sand to achieve in-bed SO_2 capture.

Ideally, the size of the sorbent fed to the boiler would simply be that required to form a stable bed. However, in the process of heating, calcining and sulfating, the sorbent's size and physical properties change. The final size of the sorbent-derived bed material, after these chemical and physical changes, can not, in many cases, be reliably predicted. In addition, some limestones are softer than others and wear faster. For these reasons, limestone characteristics and some trial and error testing during initial operation are needed to establish the proper limestone feed size and makeup flow rate.

The above discussion assumes a typical bituminous coal. In the case of low ash fuels like wood, sand is used for the bed. While sand properties are important, sand does not break down as rapidly as limestone, its makeup rates are lower and the size of the bed material is more predictable.

High ash waste coals require special consideration. Typically, these fuels contain large amounts of ash not tied to organic matter. This ash frequently consists of rock that has been generated by coal beneficiation and is called *tramp ash*. Because tramp ash does not break down into fine particles, it forms a large percentage of the bed material. For this reason, the size of the fuel feed must be chosen carefully so that the ash complements the bed material rather than causing fluidization problems. There are fluidized-bed boilers in operation where the size and consistency of the fuel ash will form a stable bed without the need for additional bed material makeup.

Bed drain classification systems are often used in circulating-bed units in addition to proper fuel and/or sorbent sizing to control the size and inventory of the bed material. These systems help to remove oversized noncirculating material while maintaining the required inventory of circulating particles.

Particle characterization and measurement In fluidized-bed boiler furnaces, particle motion is affected by the combined influence of gravitational and aerodynamic forces, as well as the impact with other particles and the walls of the boiler.

For an individual particle, the three most important characteristics are its size, density and shape. Particles of the same size but with different densities or with the same density but varying sizes do not act the same. The particle's shape, from spherical to flat, determines how it will react to the forces present in the furnace.

From an analytical standpoint, the ideal particle is a homogenous sphere and the ideal mixture consists of many homogenous spheres of equal diameter. In practice, mixtures that consist of a variety of particles with different sizes, densities and shapes are encountered.

Fluidized-bed heat transfer and pressure drop calculations assume that the mixtures of particles can be characterized by a mean particle diameter, a mean particle density and a mixture bulk density. The formulas for calculating mean diameters of mixtures assume spherical particles, and a sphericity correction term is applied to the calculated mean diameter to approximate the influence of the nonspherical nature of the actual material. Mixtures containing a significant quantity of particles that could be described as rods or flakes are difficult to characterize.

The equations that are used to calculate mean diameters are based on the weight fractions of incremental size groups. The first step of the determination is sieving a representative sample of the mixture through a stack of screens having a successively finer mesh and weighing the amount of solids retained on each screen. For calculation purposes, the average size of the particles is assumed to be the average of the mesh opening of the screen the particles were retained on and the next larger screen through which they all passed.

Two characteristic diameters are used in fluid-bed work, the weight mean diameter and the Sauter mean diameter. The Sauter mean diameter is also called the volume-surface mean diameter or the harmonic mean diameter. The particle size is typically given in microns. The weight mean diameter is calculated from the weight fraction of particles in each size cut as follows:

$$D_{WM} = D_1 X_1 + D_2 X_2 + \ldots + D_N X_N \qquad (1)$$

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which can be rewritten as:

$$D_{WM} = \sum_{1}^{N} D_N X_N \qquad (2)$$

where

D_{WM} = weight mean particle diameter, microns
D_1 to D_N = average diameter of first to last size cut, microns
X_1 to X_N = weight fraction of first to last size cut

The Sauter mean diameter for a mixture of particles is calculated from the ratio of the average volume to the average surface area. This diameter is used when predicting the hydrodynamic performance of particle mixtures. This is calculated from the weight fractions of particle mixtures using the following equation:

$$D_{VS} = \frac{X_1 + X_2 + \ldots + X_N}{\frac{X_1}{D_1} + \frac{X_2}{D_2} + \ldots + \frac{X_N}{D_N}} = \frac{1}{\sum_{1}^{N} \frac{X_N}{D_N}} \qquad (3)$$

where

D_{VS} = Sauter mean particle diameter, microns

When all the particles have the same diameter, the weight mean and the Sauter mean diameters are equal. For mixtures with a fairly narrow size distribution of particles, the weight mean and Sauter mean diameters will be similar with the Sauter diameter being smaller. For particle mixtures with a wide range of diameters, the Sauter mean diameter will be considerably smaller than the weight mean diameter.

Pressure loss

Pressure loss in a fluidized-bed boiler furnace is of great importance because it defines the total amount of solids inventory in the furnace, which is a major variable that influences heat transfer. Because the concentration of solids and the pressure profile in a fluidized-bed furnace are closely related, the determination of the pressure loss is a primary task when establishing furnace performance.

Bubbling bed In the case of bubbling-bed boilers, pressure loss is of special interest only in the bed itself. For the remainder of the boiler, the pressure loss is calculated by conventional boiler furnace fluid flow equations. (See Chapter 21.) Fig. 5 shows the furnace density profile and identifies the different zones used in establishing pressure drop and heat transfer.

The following equation is used to calculate the dense bed pressure drop in a bubbling bed.

$$\Delta P = (C)(1-e)(\rho_s - \rho_g)(L) \qquad (4)$$

where

ΔP = pressure loss
C = units conversion constant
e = bed void fraction
L = bed height
ρ_s = particle density
ρ_g = gas density at bed conditions

e is primarily a function of particle size, particle density, bed gas velocity and gas viscosity.

Various methods are used to predict bed voidage, including those proposed by Leva,[1] Babu, *et al.*,[2] and Staub and Canada.[3]

Circulating bed The density profile of a CFB boiler furnace is more complex than in a bubbling bed. It is normal practice to establish a dense bed, bubbling or turbulent, in the bottom of the furnace. This is achieved by staging the admission of air to the furnace and supplying between 50 and 70% of the total air flow to the distributor plate. This reduces the gas velocity in the primary zone below the overfire air ports and allows the maintenance of the dense bed with comparatively low solids recirculation rates. Injection of overfire air converts the solids-gas flow regime to a circulating bed above the overfire air ports. The upward flow of solids decreases with increased furnace height, resulting in a reduction of local furnace density. Fig. 7 shows the density profile and the location of the various density zones in the furnace.

Pressure loss in a CFB furnace conforms to the basic equation below:

$$\Delta P = (C)(\rho_b)(L) \qquad (5)$$

where

ΔP = pressure loss
C = units conversion constant
ρ_b = average bulk density in the furnace or furnace section associated with L
L = height of the furnace or furnace part of interest

To use the equation above, a density profile as shown in Fig. 7 is developed. This curve is a function of many variables and has been derived from empirical data. The important variables are:

D_p = average particle size above the dense bed zone
D_{DB} = average particle size in the dense bed zone
V = nominal gas velocity
T = nominal furnace temperature
W_S = external solids flux, lb/h ft^2 (kg/s m^2)
ρ_s = particle density
ϕ = particle shape factor
D_e = furnace equivalent diameter

Because the mixture bulk density in a furnace varies significantly with height, the furnace is divided into a number of zones (see Fig. 7 for definitions): dense bed, disengaging, transition and freeboard, wherein an average density value is calculated for each based on experimental data. The pressure loss equation, above, is then applied to each zone and summed to get the total furnace pressure loss.

Heat transfer

In conventional furnaces, the combustion gas carries a portion of the fuel ash with it as it goes through and out of the furnace. In general, this fuel ash represents less than 10 lb (4.54 kg) of inert solids per 1000 lb (454 kg) of gas. Also, the heat transfer from the gas to the furnace enclosure walls is predominately by radiation.

In a circulating fluidized-bed furnace, the amount of solids in the gas leaving the furnace may exceed 5000 lb (2268 kg) of solids per 1000 lb (454 kg) of gas. As a result of this high solids content, additional heat transfer mechanisms must be considered in the design. Heat transfer to the in-bed tubes of a bubbling bed and to the

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walls of a circulating bed includes solids and gas convection and solids and gas radiation in decreasing order of importance. In a conventional boiler furnace, gas radiation is the most important and solids convection is the least important to overall heat transfer.

The influence of the high solids concentration is significant. For equal temperatures, the heat transfer coefficients in a fluidized-bed boiler furnace are considerably higher than those in a conventional furnace. However, because the temperatures in the fluidized bed are between 1500 and 1600F (816 and 871C) the overall heat fluxes in the two systems are similar. Typical values for overall heat transfer coefficients in a fluidized-bed boiler furnace vary between 15 and 60 Btu/h ft² F (85 and 341 W/m² K).

Bubbling-bed heat transfer For heat transfer purposes, a bubbling-bed boiler is divided into three zones: bubbling bed or dense bed, disengaging, and upper furnace or freeboard. (See Fig. 5.)

Dense-bed heat transfer to tube banks The equation for the overall heat transfer coefficient for any tube is:

$$U_o = \frac{1}{\frac{1}{h_c + h_r} + R_m + R_{fi}} \qquad (6)$$

where

U_o = overall heat transfer coefficient, Btu/h ft² F (W/m² K)
h_c = convection heat transfer coefficient for the tube bank, Btu/h ft² F (W/m² K)
h_r = radiation heat transfer coefficient for the tube bank and walls, Btu/h ft² F (W/m² K)
R_m = metal wall resistance, h ft² F/Btu (m² K/W)
R_{fi} = tube fluid film resistance, h ft² F/Btu (m² K/W)

The convection heat transfer coefficient h_c is given by Equation 7. Two single tube equations that are used are shown as follows. Equation 8 is a modified Vreedenberg form[4] and applies primarily to beds with particles less than 800 micron average. Equation 9, a Glicksman-Decker[5] type, applies well when the average particle size in the bed exceeds 800 microns.

$$h_c = (h_{st})(FAB) \qquad (7)$$

$$h_{st} = 900\,(1-e)\left(\frac{k}{d_t}\right)\left[\left(\frac{G d_t \rho_s}{\rho_g \mu}\right)\left(\frac{\mu^2}{D_p^3 \rho_s^2 g}\right)\right]^{0.326} (\mathrm{Pr})^{0.3} \qquad (8)$$

for $D_p < 800$ microns

$$h_{st} = \frac{k\,(1-e)}{D_p}\left[C_1 + (C_2)\left(\frac{3600\, D_p\, \rho_g C_p\, V}{k}\right)\right] \qquad (9)$$

for $D_p > 800$ microns

where

h_{st} = convection heat transfer coefficient for a single tube, Btu/h ft² F
e = bed voidage, dimensionless
k = gas thermal conductivity, Btu/h ft F
d_t = tube outside diameter, ft
G = mass velocity or flux of the gas, lb/s ft²
ρ_s = particle density, lb/ft³
μ = gas viscosity, lb/ft s
ρ_g = gas density, lb/ft³
D_p = average particle diameter, ft
g = acceleration constant, 32.2 ft/s²
Pr = Prandtl number, dimensionless
C_1 = experimental constant, dimensionless
C_2 = experimental constant, dimensionless
C_p = gas specific heat, Btu/lb F
V = nominal bed gas velocity, ft/s

To convert the single tube heat transfer coefficients to those suitable for tube banks, the equations below are applied:

$$FAB = \left[1 - \left(\frac{D_o}{S_n}\right)\left(\frac{2D_o + S_p}{D_o + S_p}\right)\right]^{0.25} \qquad (10)$$

where

FAB = bank arrangement factor (staggered arrangement only), dimensionless
D_o = tube outside diameter, in. (mm)
S_n = tube spacing normal to flow, in. (mm)
S_p = tube spacing parallel to flow, in. (mm)

Other variables are as defined previously. The equation for FAB is as derived by Gel'perin, *et al.*[6]

For the radiation heat transfer component, h_r, the following equation may be used:

$$h_r = (\sigma)(\epsilon)[(T_b)^4 - (T_w)^4]/(T_b - T_w) \qquad (11)$$

where

ϵ = average overall emissivity, dimensionless
σ = 0.1713×10^{-8} Btu/h ft² R⁴
T_b = absolute bed gas temperature, R
T_w = absolute wall temperature, R

The average overall emissivity in bubbling beds will be about 0.8 depending on wall emissivity and particle size. Typically, the overall heat transfer coefficient for an in-bed tube bundle is between 40 and 60 Btu/h ft² F (227 and 341 W/m² K).

Dense-bed heat transfer to wall Many equations for the vertical wall convection heat transfer coefficient have been proposed. An example is one proposed by Mickley,[7] which takes the following form:

$$h_{cw} = (C_3)\left[\frac{3600\,(\rho_s)(1-e)(\rho_g)(V)}{D_p^3}\right]^{0.263} \qquad (12)$$

where C_3 is an experimental constant and the other variables are as previously defined. If the walls in the bubbling bed zone are coated with refractory, that resistance must be added to the other resistances in the equation for U_o. Refractory will also impact the calculation of h_r, but will not affect h_{cw} significantly.

Disengaging zone heat transfer to tube banks During periods of low bed level operation, the topmost tubes of the in-bed tube banks will be uncovered. While this portion of the bank is actually in the disengaging zone (called the splash zone in its lower portion), it is convenient to handle the heat transfer calculation as a special part of the bed. The solids content of the gas stream is much less

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than in the bed and decreases almost exponentially with height. As a result, the heat transfer rate to the exposed portion of the bank drops off rapidly. The heat transfer rates for this surface are based on experimental results.

As an example, Tang, *et al.*[8] developed the following empirical equation:

$$\frac{h}{h_m} = \exp\left(-\left(\frac{10+H_L}{25.8}\right)^{2.2}\right) \quad (13)$$

where

h = outside film heat transfer coefficient for an unsubmerged tube
h_m = outside film heat transfer coefficient for a fully submerged tube
H_L = unsubmerged tube height above the bed level, in.

Disengaging zone heat transfer to walls For the vertical walls in this zone, the convection rate conforms to an equation of the type shown below:

$$h_{cw} = C_4\,[1 - C_5(1-e)] + h_{cg} \quad (14)$$

where C_4 and C_5 are experimental constants, e is the dense bed voidage, and h_{cg} is given by the following equation:

$$h_{cg} = 0.023\frac{k}{D_e}\left(\frac{3600\,C_p\mu}{k}\right)^{0.3}\left(\frac{D_e G}{\mu\, e}\right)^{0.8} \quad (15)$$

where

k = gas thermal conductivity, Btu/h ft F
D_e = equivalent diameter of vessel, ft
C_p = gas specific heat, Btu/lb F
μ = gas viscosity, lb/ft s
G = mass velocity or flux of the gas, lb/s ft^2

Note that the voidage, e, is for the dense bed and not the disengaging zone.

As a general rule, the height of this zone may be assumed to extend for a residence time of one second.

Upper furnace heat transfer (freeboard) This portion is handled similarly to conventional boilers. The main difference is the emissivity of the solids-gas mixture. The solids content of the gas will be quite high compared to conventional boilers that burn high ash coals and may alter the radiating properties.

Circulating-bed heat transfer Circulating fluidized-bed boilers do not incorporate tube bank surface and rely entirely on heat absorption of the containment walls and internal partitions, such as division walls and wingwalls. The heat transfer in CFB boiler furnaces is handled by breaking the furnace into two distinct regions. One comprises the dense bed and the other the remainder of the furnace.

Dense-bed heat transfer Dense bed heat transfer is similar to that described for the vertical walls of bubbling-bed boilers. There is some difference because the flow regime will usually be turbulent instead of bubbling.

Disengaging zone and upper furnace heat transfer The zone just above the dense bed, but below the point where secondary air is added to produce a circulating fluid bed, is defined as the disengaging zone. The upper furnace includes the transition zone and the freeboard as indicated in Fig. 7. Bed-to-wall heat transfer in these zones can be predicted by considering three parallel processes: particle convection, gas convection and radiation.

In particle convection, heat is removed from particles at the cooling wall surfaces by conduction. The energy loss is replenished by the material and energy exchange with the upwardly flowing central core of solids and combustion gases.

Gas convection is the dominant mode of heat transfer over the fractions of the surface not in contact with particles. This gas component is of minor consequence where the solids content is high. Even in the upper portions of a furnace, where solids content is relatively low, gas convection is usually small compared to the radiative component.

Radiant heat transfer occurs in a manner similar to that in conventional furnaces. Procedures are available to account for the combined emissivities and scattering effects of solids mixed with gases, as long as the void fraction is near or above 0.8. From a practical standpoint, this applies to the furnace above the dense bed.

The overall effective emissivity is a function of the pertinent radiative properties of the gases and solids found in a given mixture, as well as that of the heat absorbing surface. Typically, overall emissivities are about 0.5.

Various equations have been proposed to predict particle convection heat transfer coefficients. Some are complex and include many parameters. Two parameters have the most influence, particle size and mixture bulk density. Reliable convection equations have been developed considering only these two variables.

At full load conditions, where the solids circulation rate is high, a CFB furnace operates at approximately isothermal conditions from top to bottom. The overall heat transfer, in this case, is determined from the vertical bulk density distribution and a proper mean particle size. When the solids circulation rate is reduced, as occurs at low loads, the furnace becomes less isothermal, and more complex procedures are needed to calculate furnace heat absorption. One such procedure involves dividing the furnace into a larger number of vertically arranged zones. This permits the variation in temperature and bulk density to be properly handled in each zone wherein the variation is small.

The equations that follow are typical of those that predict furnace heat transfer:

$$U_o = \frac{1}{\dfrac{1}{h_c + h_r} + R_{ref} + R_m + R_{fi}} \quad (16)$$

where

$h_c = h_{cp} + h_{cg}$
$h_{cp} = C_6\,(\rho_b)^m/(D_p)^n$
$h_r = (\sigma)\,(\epsilon)\,[(T_g)^4 - (T_w)^4]\,/\,(T_g - T_w)$

and where

T_g = absolute local bulk gas temperature, R
h_{cp} = particle heat transfer coefficient, Btu/h ft^2 F
C_6, m, n = experimentally derived constants

The value of h_{cg} is obtained with Equation 15 and all other variables are as previously defined.

Published literature provides a great deal of theoretical information, laboratory experimental data and test results from operating furnaces. Typical of the latter

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category is the paper of Kobro and Brereton.[9] Figs. 9 and 10 are from that paper. They show measured overall heat transfer rates for a range of bulk densities for two specific average particle sizes. Similar curves are made from commercial boiler field test data and are used to help make empirical corrections to the basic equations.

Heat and material balance

One of the first steps in designing a boiler is to calculate the overall heat and material balances. This establishes the flow quantities, compositions and temperatures of all streams entering and leaving the system. The boiler system is then subdivided and balances are calculated for the pieces. This process is repeated until all systems are defined.

The sensible heat of the solids, as a fraction of gas sensible heat, is quite large in fluidized-bed boilers. It represents the major difference between the heat and material balance for a fluidized-bed boiler and a conventional boiler. (See Chapter 9.) The information that follows describes the many flow streams that are considered when preparing the heat and material balance for a fluidized-bed boiler.

Material balance When limestone is used as a sorbent, the following chemical reactions which affect the solids balance occur in the furnace:

$$CaCO_3 \rightarrow CaO + CO_2$$
$$MgCO_3 \rightarrow MgO + CO_2$$
$$CaO + SO_2 + 1/2\ O_2 \rightarrow CaSO_4$$

There is a net solids weight loss during calcination as CO_2 in the carbonate is driven off endothermically and a solids weight gain as a result of the exothermic sulfation reaction. The Ca/S ratio selected, the amount of sulfur reacted and the ultimate analysis of the limestone will determine the net influence on boiler performance.

For the purpose of illustration, the material balance for an atmospheric pressure CFB boiler is shown in Fig. 11.

From an overall material balance standpoint, the solids that enter the furnace leave as relatively coarse material through the bed drain or as fine material collected after the convection pass. There is a significant amount of internal solids recirculation that takes place within the system, and these streams and their effects must be considered. Also, as discussed in the heat transfer section, the distribution of the solids in the furnace influences overall furnace thermal performance.

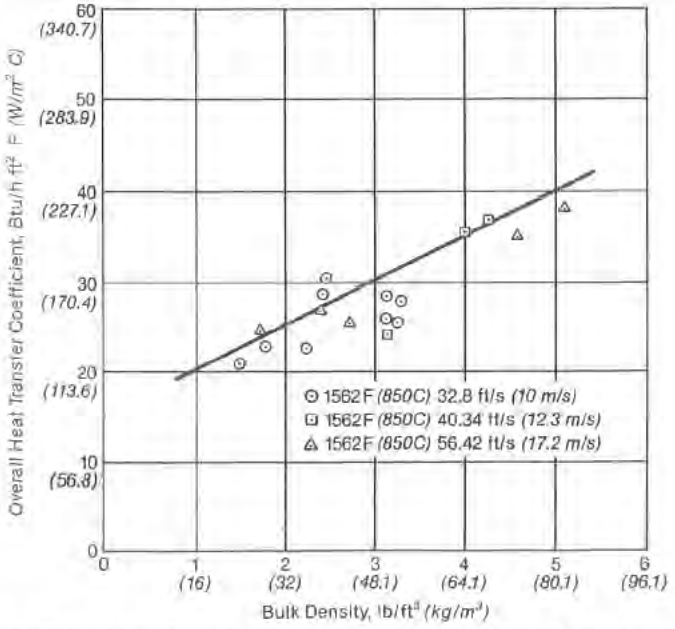


Fig. 9 Heat transfer coefficient versus density in a circulating fluidized bed, 250 micron mean diameter sand.

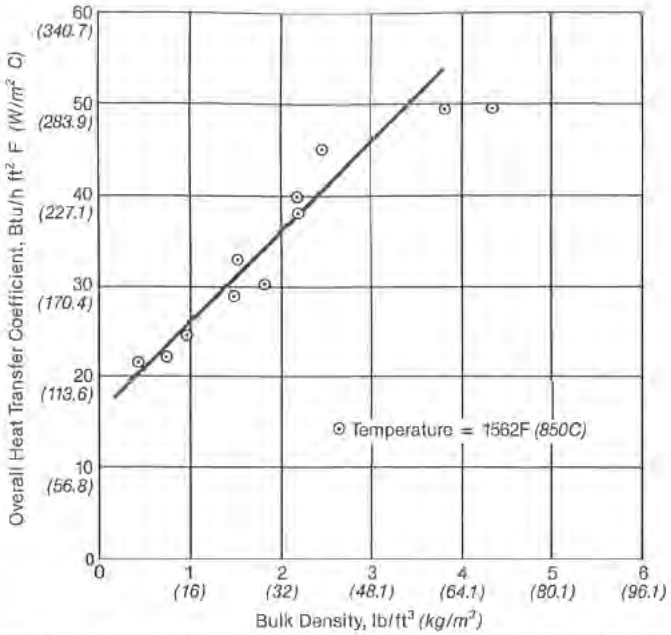


Fig. 10 Heat transfer coefficient versus density in a circulating fluidized bed, 170 micron mean diameter sand.

The fine particles that leave the bed and circulate are called *elutriated solids*. These solids ultimately leave the system after the convection pass with the gas leaving the dust collector or as purged from the multi-clone, if required. Changes in this fraction only change the split between the quantity of solids that leave through the bed drain and the quantity that leave the back end. In the discussions that follow, the sorbent input is considered to have already undergone calcination and sulfation reactions, and the fuel has completely combusted, leaving only ash. These flows should not be confused with the nonreacted sorbent and fuel fed into the boiler.

Definitions of the terms shown in Fig. 11 and the equations which are used to define the mass balance of a B&W CFB boiler are listed in Table 1.

The elutriated fractions Ea, Es and Ei depend on the specific fuel ash, sorbent and inert material in the system. Numerical values are determined from laboratory testing or from operating boilers. N_{UB} and N_{MC} depend on the mechanical arrangement of the collectors plus the size distribution and physical properties of solids circulating (ash, sorbent and inert) in the system. The values are determined empirically. The values of N_{UB}, N_{MC} and E for each input component and for the total solids are interdependent as shown in the following material balance equation.

$$ESF(1-N_{UB})(1-N_{MC}) + MCP = E(ISF) \qquad (17)$$

The external mass flux, W_S, is a variable controlled by the designer and is cited under the section on heat balance.

The outlines that follow describe the steps carried out when establishing the material balances for fluidized-bed boilers. For CFB boilers:

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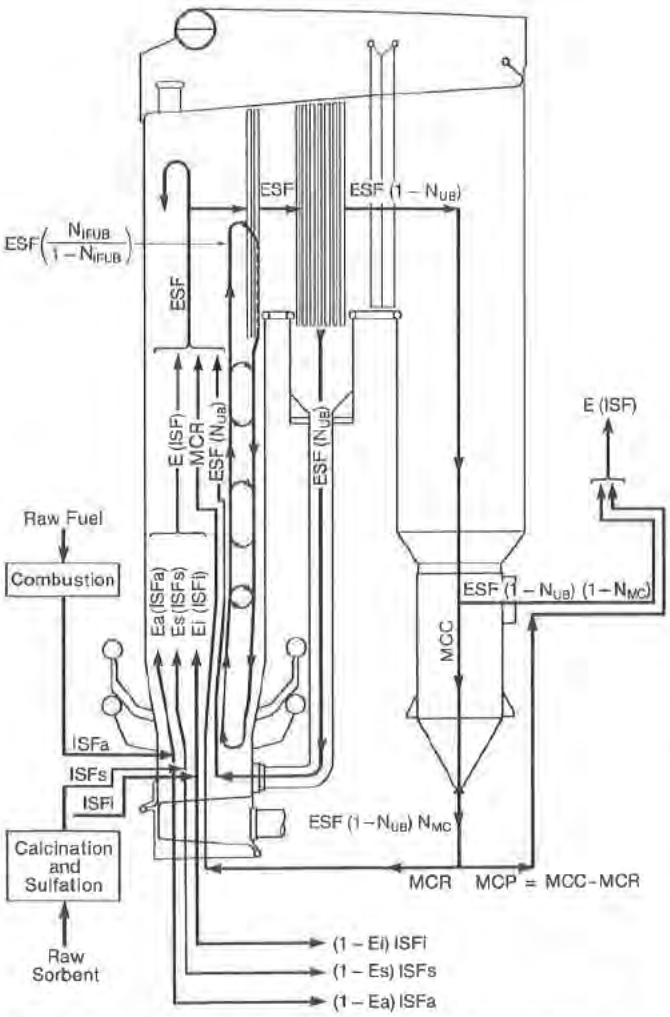


Fig. 11 Circulating fluidized-bed boiler material balance schematic.

1. ISF is determined from the required fuel and limestone inputs. Normally, the limestone input is proportional to the fuel input and is set to provide the required sulfur retention.
2. E is selected based on empirical data.
3. ESF is selected based on furnace design considerations.
4. N_{MC} and N_{UB} values are selected based on empirical data. It should be verified that the selected efficiency values are tied by Equation 17. All other solids flow, external and internal, can be determined from the information above.
5. BDF = (1-E) ISF
6. LVF = ESF (N_{UB})
7. CPSF = ESF (1-N_{UB})
8. BHC = ESF (1 - N_{UB}) (1 - N_{MC})
9. MCC = ESF (1-N_{UB}) N_{MC}
10. MCP = E(ISF) - BHC = MCC - MCR
11. MCR = ESF (1 - N_{UB}) - E(ISF) = CPSF - E(ISF)

Note that the baghouse loss is considered negligible. With properly selected E, N_{MC} and N_{UB} values, the CPSF will not exceed its maximum value corresponding to the allowable solids loading in the convection pass. In most cases, a circulating fluidized-bed boiler can be designed so the multi-clone purge (MCP) is zero. For bubbling-bed boilers:

1. ISF is as above
2. E is selected based on empirical data.
3. BDF = (1-E) ISF
4. Capacity of multi-clone recycle system (MCR_c) is selected based on empirical data in proportion to fuel input.
5. N_{MC} is selected based on empirical data and desired sorbent utilization and carbon conversion.
6. CPSF = ISF (E) + MCR_c
7. MCC = CPSF (N_{MC})
8. BHC = CPSF (1-N_{MC}) (assumes baghouse loss is negligible)
9. MCP = MCC - MCR_c
10. MCR = MCR_C
11. If MCP < 0, then set MCP = 0 and MCR = MCC.
12. BHC = E(ISF)
13. $$MCR = \frac{E(ISF)N_{MC}}{1-N_{MC}}$$
14. $$CPSF = \frac{ISF\ (E)}{1-N_{MC}}$$

Heat balance In furnaces fired by pulverized coal burners, Cyclones and stokers, little heat is removed from the combustion gases in the zone of maximum heat release. As a result, the flue gas reaches a high temperature before it is subsequently cooled by the furnace enclosure surface.

In fluidized-bed combustion, heat is removed from the zone of maximum heat release at a much higher rate. As a result, the maximum temperature of the flue gas is limited to a predetermined level.

Table 1
Mass Balance Terms

Input solids flow (ISF), lb/h (kg/s)

ISF	= ISFa + ISFs + ISFi
ISFa	= Fuel ash and unburned carbon
ISFs	= Postcalcination and postsulfation sorbent
ISFi	= Inert bed material (SiO_2 and H_2O)

Elutriated fractions leaving the boiler, dimensionless

Ea	= Ash
Es	= Sorbent
Ei	= Inert material
E	= [Ea(ISFa) + Es(ISFs) + Ei(ISFi)]/ISF
N_{IFUB}	= In-furnace U-beam efficiency

Additional terms

BDF	= Bed drain flow, lb/h (kg/s)
BHC	= Baghouse catch flow, lb/h (kg/s)
CPSF	= Convection pass solids flow, lb/h (kg/s)
ESF	= External solids flow, lb/h (kg/s)
LVF	= L-valve flow, lb/h (kg/s)
MCC	= Multi-clone catch flow, lb/h (kg/s)
MCP	= Multi-clone recycle purge flow, lb/h (kg/s)
MCR	= Multi-clone recycle flow, lb/h (kg/s)
MCR_c	= Multi-clone recycle system capacity, lb/h (kg/s)
N_{IFUB}	= In-furnace U-beam efficiency
N_{MC}	= Multi-clone dust collector efficiency
N_{UB}	= Hot particle collector efficiency
W_S	= External mass flux based on plan area of the upper furnace shaft (ESF/A), lb/h ft^2 (kg/m^2 s)

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In bubbling-bed systems, the heat removal process is accomplished by having the cooling surface immersed in the active bed of hot solids and burning fuel.

In circulating beds, the large quantity of solids circulating in the system removes heat from the active combustion zone and transfers it to the heating surface throughout the furnace.

The ability to change the amount of heat removed from the combustion process to achieve the desired bed temperature provides flexibility when designing a fluidized-bed boiler. In a bubbling bed, the heat balance around the bed itself determines the bed temperature. The excess heat, which is the difference between the quantity of heat entering the bed with the fuel plus air and the amount of heat leaving the bed in the combustion products and through radiation to the rest of the furnace, must be absorbed by the in-bed tubes and the enclosure surrounding the bed. If the heating value of the fuel decreases, less in-bed surface is required to achieve the same bed temperature. If the heating value of the fuel increases, more in-bed surface is required. At the lower extreme of fuel heating value, no in-bed surface is used and the enclosure is often covered with refractory to minimize heat loss. This latter example is typical when high moisture sludges are burned.

In the CFB designs, all furnace heat transfer surfaces must be considered when making a heat balance. The amount of material circulating from the furnace to the primary collector and back to the furnace determines the inventory or average density of solids in the furnace. The heat transfer coefficient in the furnace is proportional to bulk density. Therefore, in the case of CFB boilers, the furnace heat absorption depends on the total furnace surface and the external recycle rate (which is a direct function of W_S, the external mass flux).

The furnace heat balance is also influenced by the way a particular fuel burns. When fuel is introduced into a fluidized bed, the majority of it burns in the dense bed. However, some of the fuel burns above the dense bed and the heat release pattern of the above-bed burning must be considered when calculating the furnace exit gas temperature. The actual split between in-bed and above-bed burning depends on fuel properties such as type, volatility, size and feed system. Because of the large amount of recirculated solids, the heat release pattern in a CFB furnace does not have a strong influence on the temperature distribution. For this reason, CFB furnaces are more tolerant of fuel changes than bubbling beds.

The heat balance around the convection surface is calculated similarly to conventional boilers. However, it is necessary to include the effects of the solids in the gas because they can have a substantial influence on the convection pass heat balance.

Fluidized-bed boiler arrangements

Boiler subsystems

The unique features discussed include the distributor plate, the bubble caps and the overfire air system. Also discussed are some of the features that are common to all water-cooled boilers designed by B&W.

Distributor plate and bubble caps The distributor plate is located at the bottom of the furnace and separates the windbox from the furnace. The distributor plate is fitted with bubble caps to provide a uniform distribution of combustion air to the entire furnace cross-section over the boiler load range. (See Fig. 12.) For this reason, distributor plates designed by B&W have a pressure drop of about 16 in. (406 mm) of water column across the bubble caps at full load and a minimum drop of 4 in. (102 mm) of water column at minimum load. Typically, 50 to 70% of the combustion air flows through the distributor plate of a CFB boiler at full load. A bubbling fluidized-bed boiler distributor plate is designed for 85 to 100% of the combustion air when staged combustion is not used.

The distributor plate must form an air-tight seal, other than the bubble caps between the furnace and the windbox and must support the weight of a slumped bed and resist the uplift generated from the air pressure drop across it during operation. In general, most distributor plates used in boilers are constructed from water-cooled membraned tubes.

In addition to the items discussed above, the bubble caps are designed to split the air into small streams for good flow distribution, avoid the formation of large bubbles in the bed, minimize erosion and prevent back sifting of bed solids into the windbox.

Overfire air system The overfire, or secondary, air system is part of the larger combustion and emission control system. As previously described under *Emissions*, all CFB boilers and some bubbling fluidized-bed boilers, when no tube bundle is present, use a staged combustion process to control the emission of NO_x. Overfire air is sometimes used in bubbling-bed boilers to improve free-

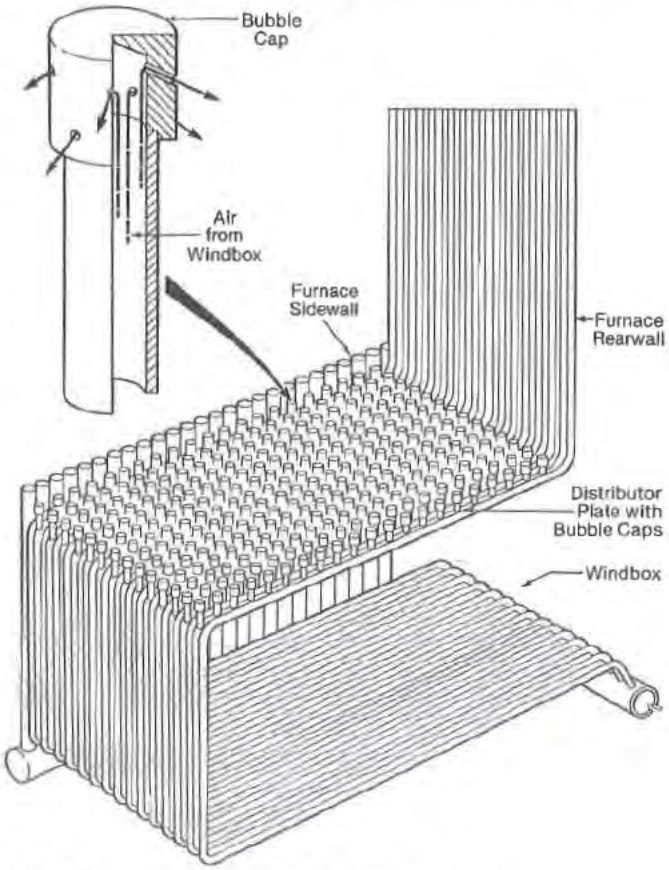


Fig. 12 Furnace distributor plate and bubble caps.

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board mixing and burning. This technique is used when fuel is fed over-bed or when there are excessive fuel fines.

With overfire air, the combustion air that does not flow through the bubble caps is injected into the combustion gases shortly after initial coal ignition to complete combustion. The overfire air system must provide adequate penetration and thorough mixing of the overfire air with the combustion gases to achieve complete fuel burnout and minimize the amount of CO discharged to the atmosphere.

Overfire air penetration depends on the size of the overfire nozzles and the air and gas velocities and densities. In circulating beds, the solids density at the point of injection must also be considered.

Boiler enclosure The furnace and convection pass enclosures are conventional. (See Chapter 22.) They are constructed of water-cooled membraned tubes welded together to form a gas-tight enclosure. This enclosure is also used to support the fluidized bed, windbox, superheater and other components because the boilers are usually top supported.

Auxiliary equipment

Fuel feed systems The fuel feed system has had a greater impact on the evolution of fluidized-bed boilers than any other auxiliary or support system. As a result, three classifications of fuel feed systems have been developed and are in use today. These are the under-bed, over-bed and in-bed systems.

The under-bed feed system is usually applied to bubbling-bed boilers burning bituminous coal. An under-bed feed system is essentially a pneumatic transport system that moves the coal from a storage silo to the bed.

When burning bituminous coal and less reactive fuels, it is necessary to spread the coal as evenly as possible throughout the bed for good fuel-air mixing. To do this, fuel feed pipes are spaced on about 4 ft (1.2 m) centers throughout the bed.

When compared with the over-bed feed system described below, a well designed under-bed feed system can provide between 2 and 4% better combustion efficiency with bituminous coals.

On the negative side, an under-bed feed system is complicated. The fuel, typically, must be crushed to less than 0.25 in. (6.4 mm); the fuel must be dried to less than about 6% moisture or the transport pipes will plug; transport piping erosion can be a problem; the equipment needed to pressurize the coal for transport is a high maintenance item; and costs are generally higher than for an over-bed feed system.

Over-bed feed systems are also used on bubbling beds. They are used for reactive fuels and bituminous coals where the simplicity of the system can justify any reduction in carbon conversion.

The feeders are located above the bed where the furnace gas pressure is slightly below atmospheric. This location simplifies feeding since the coal stream does not have to be pressurized. An over-bed system uses the same fuel feed equipment as a spreader stoker. (See Chapter 15.)

While over-bed feed systems are simpler than under-bed systems, they still have requirements and limitations. The coal for an over-bed feed system is crushed to a top size of 1.25 in. (31.8 mm), and the amount of fine coal (less than 30 mesh) is limited to prevent excessive burning in the freeboard zone immediately above the bed. Also, the amount of fines in the fuel must be relatively consistent.

Fig. 13 compares the combustion efficiency for under-bed and over-bed coal feed systems for two coals, Sarpy Creek and Kentucky No. 9. Sarpy Creek is a subbituminous coal and Kentucky No. 9 is a bituminous coal. It can be seen from Fig. 13 that the Sarpy Creek coal is much more reactive than the Kentucky coal. Also, under-bed feed results in higher combustion efficiency than over-bed. Recycling solids collected in a multi-clone to the furnace also increases combustion efficiency.

In-bed feed systems are used for CFB boilers and some sludge-burning bubbling beds. Typically, some form of a feed screw or air assisted chute is used. The fuel is injected through an enclosure wall just above the distributor plate. Because the furnace pressure at this point can be as high as 50 in. (1270 mm) of water gauge, the feed system must seal against this pressure to allow the fuel to be introduced to the furnace. The pressure seal is accomplished by the use of a rotary seal or a head of coal between the coal silo and a belt feeder.

A CFB with its higher fluidizing velocity is a better fuel mixing process than a bubbling bed with in-bed tubes. As a result, fewer feed points are required.

Sorbent feed systems As previously discussed under *Emissions*, limestone or dolomite is added to a fluidized-bed boiler to capture SO_2. In B&W designs, the sorbent is usually fed into the dense bed area of the furnace. While good limestone distribution is desired to reduce sorbent requirements, it is not nearly as critical as distribution in the case of under-bed coal feed.

The design of the sorbent feed system must consider where the material will be injected into the furnace, the furnace gas pressure at that point and the method of addition. In the case of under-bed and in-bed feed, the sorbent can be mixed and injected with the fuel. Sorbent can not be fed properly with a spreader due to its fineness. Sorbent has been blown into the furnace through separate pneumatic feed points and by gravity from a storage silo.

Bed ash removal systems When an ash bearing fuel is burned in a fluidized-bed boiler, the ash is liberated from the coal in the boiler furnace. Also, sorbent or inerts are fed to the boiler. Therefore, means must be provided to remove the solids from the system to prevent accumula-

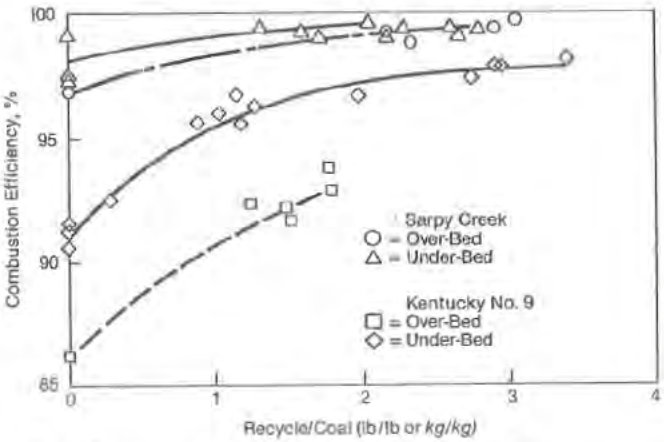


Fig. 13 Comparison of under-bed and over-bed coal feed.

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tion. There are two major places in a fluidized-bed boiler system where solids are removed, the bed drain and the baghouse or electrostatic precipitator.

Normally, the design of the bed ash removal system includes a large safety margin because of the uncertainty as to where the solids will exit the system and to handle changes in fuels. Operating results have shown that the bed drain flow can range from 0 to more than 50% of the total solids output.

In the case of fuels with a high alkali content in the ash, it may be necessary to drain the bed material from the boiler at a rate higher than the solids buildup rate to prevent a concentration buildup of the alkali in the bed.

Under these conditions, it is also necessary to make up any excessive bed material lost from the system. Typically, when the concentration of alkali exceeds 5 to 6% of the bed weight, the probability of forming agglomerates increases significantly.

A second case that requires greater than normal bed drain flow rates is created when the fuel contains a high percentage of large rocks and ash, greater than 0.5 in. (12.7 mm). There is a natural tendency for these large pieces of noncombustible material to accumulate in the bed and cause defluidization. To avoid this, it is necessary to move the oversize material to the bed drain for removal. This is done by maintaining the bed drain flow rate above a minimum experimentally-determined value. It is necessary to make up any excess bed material lost in the rock removal process.

Sootblowers Chapter 23 discusses the application of sootblowers to various types of firing systems. Because the fuel in a fluidized-bed boiler is burned at temperatures below the ash softening point, the flyash never reaches the plastic state. As a result, it forms a dry powder that is easy to remove from the heating surfaces.

There are fluidized-bed boilers in operation that have no sootblowers. As long as the heat transfer surface is properly spaced and the dusting that does occur is accounted for in the design of the unit, the boiler will perform satisfactorily, achieving the desired values of superheat temperature and boiler efficiency. However, when fuels with high sodium and/or potassium ash contents (low ash fusion temperatures) are fired, especially with an over-bed feed system, there is a higher likelihood for convection pass fouling and, therefore, sootblowers may be required.

Typical fluidized-bed boiler designs

At the time of this writing, the majority of applications for bubbling fluidized beds had been directed at installations of less than 200,000 lb/h (25.2 kg/s) of steam. Most of the new, large capacity, coal-fired fluidized-bed boilers have been of the circulating type. However, several large boilers have been retrofitted with bubbling beds because the bubbling bed was physically more compatible with the changes, as compared with circulating beds. In some of these cases, bubbling beds were used to reduce SO_2 emissions. In other cases, fuel-related operating problems have provided the incentive for change.

Another area where bubbling beds, both new and retrofitted, are being used is for burning sewage sludge and pulp mill and recycled de-inking paper sludge.

The design of a boiler and its auxiliaries is a complex task. Because other chapters describe the conventional aspects of boiler design, the following descriptions are limited to unique fluidized-bed boiler features.

Bubbling-bed retrofit During the winter of 1986, the moving grates were removed from Unit No. 2 at Montana-Dakota Utilities' R. M. Heskett station and replaced with a bubbling fluidized bed. (See Fig. 14.) This boiler was originally designed to generate 650,000 lb/h (81.9 kg/s) steam at 1300 psig (89.6 bar gauge) and 950F (510C) superheat burning Beulah lignite. However, high sodium content in the fuel ash caused severe furnace slagging and superheater fouling. Before the bubbling-bed modification, long term output was limited to about 50 MW of the rated capacity of 72 MW. The fluidized bed was installed to reduce furnace operating temperatures, to avoid slagging and fouling problems, and to achieve full power operation.

The new fluidized bed measured 40 ×26 ft (12.2 ×7.9 m) and was fitted under the boiler with only minor pressure part changes to the furnace division walls. The distributor plate and enclosure walls are water-cooled. In this unit, both superheat and boiling surface are placed in the bed to meet steam generation and turbine superheat temperature requirements and to limit the bed operating temperature to 1500F (816C). The superficial bed velocity is 12 ft/s (3.7 m/s) and the bed is 54 in. (1372 mm) deep in the expanded condition. Eight windbox compartments are used for unit turndown and startup. Because Beulah lignite is a highly reactive fuel, no flyash recycle was installed. Additionally, because of the low sulfur and high alkali content of this fuel, sand is used as the bed material.

The boiler was restarted in May 1987. This unit is now producing 80 MW and avoiding fuel related slagging and fouling problems. However, operators must avoid high bed and furnace temperatures because of the fuel's fouling tendencies.

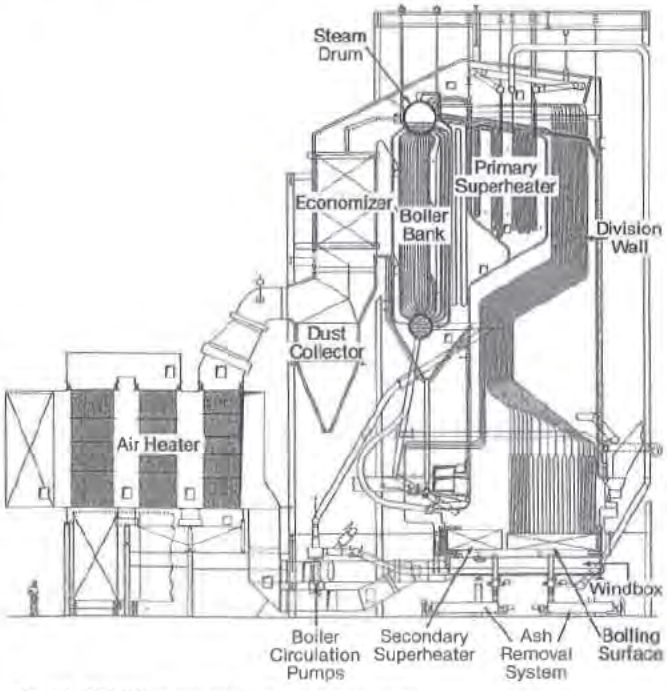


Fig. 14 80 MW bubbling fluidized-bed retrofit.

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Circulating fluidized-bed boiler The B&W circulating fluidized-bed boiler design is based on a completely water-cooled setting. This feature provides a modern gas-tight enclosure suitable for operating with a positive pressure in the furnace. It has no high temperature refractory lined flues in the vicinity of the primary particle collector and therefore requires minimal building space and reduces furnace refractory maintenance. This construction is possible due to the use of an impingement-type primary solids separator (U-beams) which can easily be integrated into the boiler enclosure.

Fig. 15 shows the side view of a 55 MW CFB boiler designed to burn low volatile bituminous coal. The unit produces 465,000 lb/h (58.6 kg/s) steam at 1550 psig (107 bar gauge) and 955F (513C). The furnace is 30 ft (9.1 m) wide, 15 ft (4.6 m) deep, 85 ft (25.9 m) high and contains full height, water-cooled division walls and steam-cooled wingwalls in the upper furnace.

Fuel and sorbent are fed to the bed through the lower furnace frontwall. The ash and spent sorbent are removed through drain pipes in the floor. The solids collected by the U-beams and multi-clone are returned to the lower furnace through the rearwall.

The primary air enters the furnace through the distributor plate and secondary air is injected at elevations approximately 6 and 12 ft (1.8 and 3.7 m) above the distributor plate.

The entire lower furnace, up to 22 ft (6.7 m) above the distributor plate, is covered by a thin layer of highly conductive refractory held to the water tubes by pin studs. Refractory is used in the lower furnace to protect the tubes from corrosion and erosion. The remaining portion of the furnace enclosure consists of bare tubes.

The B&W CFB boiler design uses an impact separator to collect and recycle solids to the furnace. The primary solids separation system consists of staggered rows of U-shaped channel members, or U-beams, suspended from the boiler roof. (See insert, Fig. 15.) Material strik-

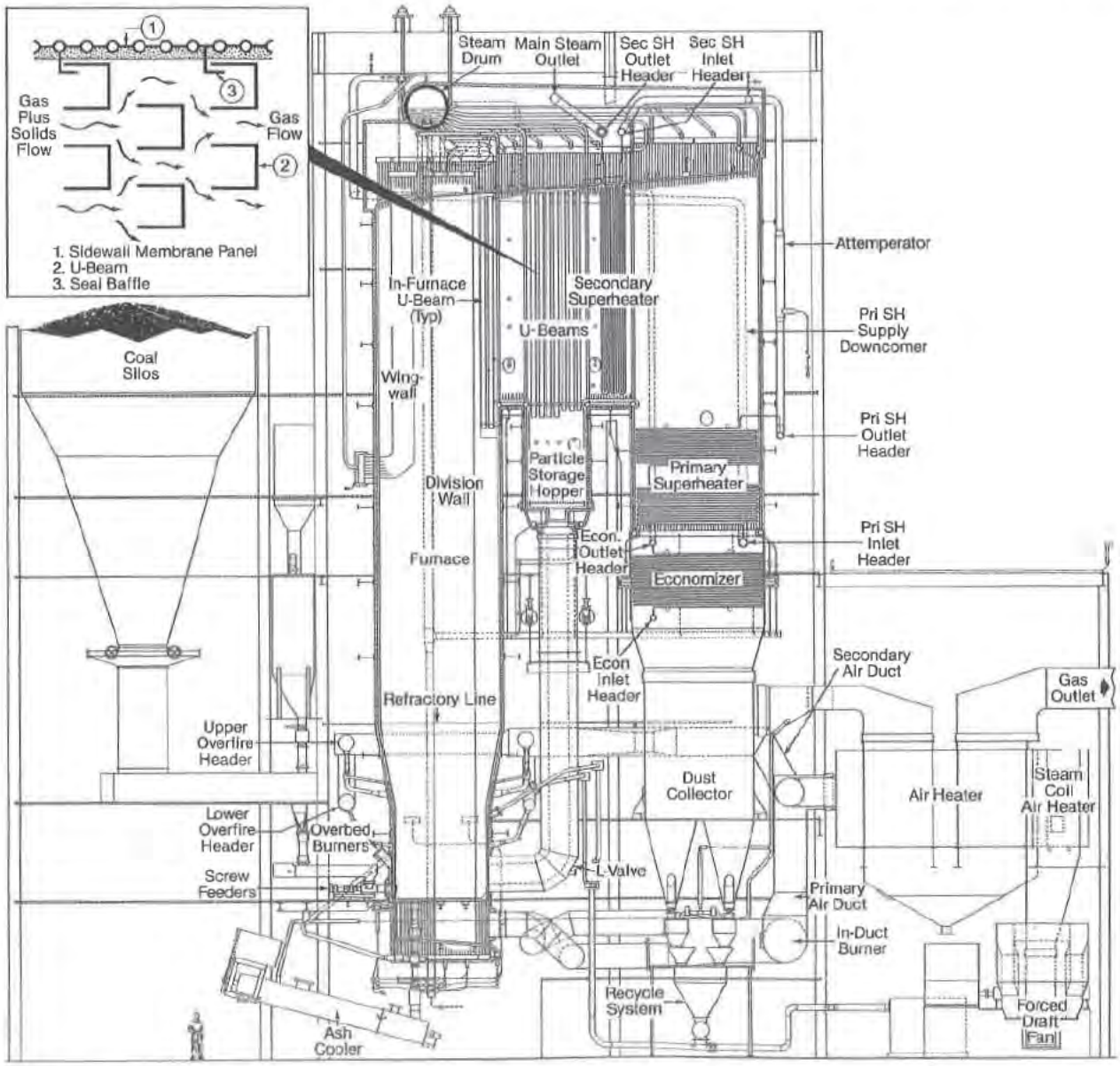


Fig. 15 CFB boiler sectional side view.

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ing the U-beams is separated from the gas, flows down the U-channel and discharges from the bottom. The most recent designs use two stages of U-beam collectors which result in better overall collection efficiency than was provided in the first generation of boilers. The first stage is located in the upper furnace and returns the solids directly to the lower furnace. The second stage is located after the furnace and above a particle storage hopper. Bed material collected by the second stage U-beams is recycled to the lower furnace by flow-controlling L-valves.

The L-valve is a nonmechanical device for returning solids to the furnace (Fig. 16). Solids collected in the particle storage hopper flow to the standpipes and serve as a source of inventory for the furnace. Solids flow is induced by injecting a small amount of aeration into the L-valve. With this arrangement, hundreds of thousands of pounds (kg) of solids per hour can be circulated with air flows on the order of 10 ACFM (4.7×10^{-3} m^3/s).

Startup and operation

Startup

Fluidized-bed boilers are brought into operation by first establishing air flow through the unit, heating the bed material to a temperature that is above the fuel's autoignition temperature, and then introducing the fuel into the hot solids and air. The fact that the bed solids are above the fuel's autoignition temperature ensures that combustion is initiated safely.

The heat for raising the bed material to the desired temperature is normally supplied by a burner in the air supply duct that is capable of heating the air and solids to the required temperature, a burner located above the bed that fires into the bed, or a combination of both methods.

The above-bed type of burner is generally used for circulating beds and bubbling beds that do not have in-bed surface. The duct type of burner is used for bubbling beds that contain in-bed cooling surface. Both the above-bed and duct burners must have complete flame safety systems which include independent flame detection and burner safety trip control circuits.

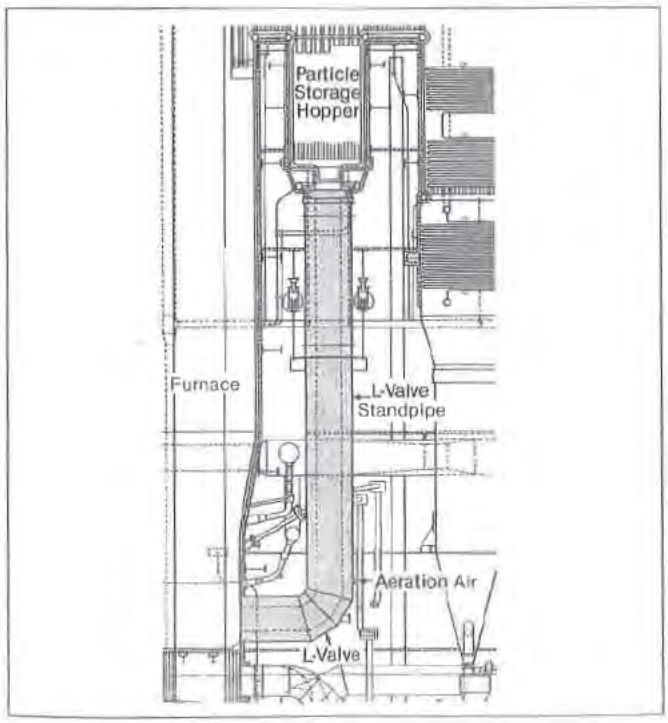


Fig. 16 Particle recirculation for a CFB.

Different fuels have varying ignition temperatures. The following table lists several fuels and the minimum bed temperature that must be attained before the fuel can be fed into the bed.

Fuel	Minimum Bed Temperature, F (C)
Bituminous coal	900 to 950 (482 to 510)
Lignite	900 (482)
Anthracite	1000 to 1050 (538 to 566)
Wet wood	1200 to 1250 (649 to 677)
Oil	1400 (760)
Natural gas	1400 (760)

The actual bed temperature for a particular fuel must be determined by demonstration. Once the minimum safe bed temperature is established, it is only necessary to measure bed temperature to make sure it is above the minimum value whenever feeding fuel to the bed. As fuel is introduced into the bed, it ignites, releases its energy and increases the bed temperature above the minimum value. Further increases in fuel feed will result in higher bed temperatures. This process is continued until the desired boiler load and bed temperature are achieved.

System control

Many of the operating and control features of fluidized-bed and conventional boilers are the same. There are two areas where fluidized-bed and conventional boilers vary significantly.

One is the need for the fluidized-bed boiler control system to monitor and control the transport of large quantities of solids. The other is the need to control the primary and secondary air ratio to achieve minimum emissions and unburned carbon loss. With the ability to control solids and air flows, the unique fluidized-bed parameters of bed temperature and bed inventory can be controlled.

Bed temperature control

Bubbling bed The temperature of a fluidized bed is controlled by manipulating the heat removal process. In a bubbling-bed boiler, the bed itself is limited to the lowermost 4 ft (1.2 m) of the furnace.

Because the heat absorbing surface is located within the bed and solids recycle has only a minor influence on bed temperature, the heat balance around the bed establishes the bed operating temperature. As the load (fuel flow) is changed, it is necessary to change the bed heat balance to achieve the desired bed operating temperature. The heat balance can be altered by the following methods.

For the case of a load reduction, bed depth can be reduced by removing inventory from the system and/or reducing the velocity of the combustion gas through the bed. By decreasing the bed depth, less heat transfer surface will be submerged in the solids. As a result, the amount of heat removed decreases and the bed temperature is maintained at the desired value.

There are limits to the degree of turndown achievable by reducing the bed level and a point is reached where a portion of the bed must be removed from service, or

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slumped. Under these conditions, little heat is removed from the bed in the slumped area. This control requirement dictates that the windbox be designed so that portions of the bed can be isolated and removed from service.

Bed temperatures can also be controlled by excess air and air preheat. By increasing the excess air that flows through the bed, the amount of heat required to heat the air increases and the temperature of the bed decreases. By decreasing the air preheat temperature, the temperature of the bed can also be made to decrease.

Circulating bed In CFB boilers, the solids inventory is distributed more uniformly throughout the furnace than in a bubbling bed. (See Figs. 5 and 7.) The combination of this condition and the large amount of solids that recirculate at high boiler loads results in uniform furnace gas and solids temperatures. In the section on heat transfer, it was pointed out that the heat transfer coefficient in a CFB furnace is greatly dependent on the solids inventory in the furnace. Therefore, by changing furnace inventory, the heat transfer from the gas and solids to the cooling surface can be varied. The net result of this process is the ability to control furnace bed temperature. Therefore, as load changes or as fuel varies, it is possible to increase or decrease furnace temperatures by controlling furnace inventory. In some CFB designs, furnace temperature is moderated with an external heat exchanger. As in bubbling beds, the bed temperature may also be changed by varying excess air and air preheat temperature.

Bed inventory control

The ability to maintain the required bed inventory, as described above, is highly dependent on the properties of the solids that make up the inventory as well as the performance characteristics of the solids collection and recycle system. Furnace inventory is made up of sorbent and/or inert bed material and fuel ash, but each must be properly controlled.

For the system to be under control, the quantity of solids in the furnace must equal or exceed the amount required to support stable combustion and for bed temperature control. If too many solids leave the boiler with the flue gas, furnace inventory will decrease, combustion may become unacceptable and furnace temperatures will increase. This condition must be corrected by modifying the particle size distribution of the solids fed to the boiler or by providing some kind of inventory control system.

Bubbling bed Bubbling-bed inventory is dependent upon the balance between feed solids and the bed drain purge flow, plus the solids that are elutriated and lost from the system through back-end collectors. The amount of fine material leaving the boiler with the flue gas (elutriated fraction) depends on particle size and fluidizing velocity. When solids are recycled from a dust collector, the flow of solids in the freeboard increases significantly. However, bed inventory, which consists mostly of nonentrainable particles, is not affected.

From a practical operating and control standpoint, the minimum solids feed flow rate must provide enough material to maintain the bed drain flow required to remove oversized particles plus make up the back-end loss. With solids input exceeding the minimum value, the rate of bed drain purge is adjusted to maintain the desired bed inventory. In practice, bed inventory is controlled by measuring the pressure drop across the bed and maintaining it at a value corresponding to the desired inventory.

Circulating bed In a circulating fluidized bed, the amount of solids leaving the furnace with the flue gas and the bed inventory are interdependent. Solids leaving the furnace must be collected and returned to the furnace to maintain bed inventory. Overall inventory control is determined by the solids input-output balance similar to a bubbling bed and is controlled in the same way.

In B&W's circulating fluidized-bed system, the inventory consists of solids in the furnace plus those in the particle storage hopper and L-valve standpipes. The inventory distribution between the furnace and storage is established by the external recycle rate which is controlled by nonmechanical L-valves. For a constant total inventory, a reduction in L-valve flow rate transfers inventory from the furnace to the particle storage hopper and L-valve standpipes and vice versa.

Because the solids flow rate circulating through the L-valves is many times larger than the solids input rate, inventory can be exchanged between the furnace and particle storage hopper at the rate of 10 to 20% per minute. This feature allows rapid control response to changes in fuel flow or boiler load.

The inventory split between the dense bed and the upper furnace is dependent upon the particle size distribution of the solids in the furnace and gas velocities in the primary zone and upper furnace. To provide sufficient inventory in the upper furnace and satisfy heat transfer requirements, the total system inventory is controlled by the bed drain purge rate. On the other hand, the bed drain purge rate should be sufficient to prevent accumulation of coarse material in the dense bed.

If the bed drain purge rate required for coarse material

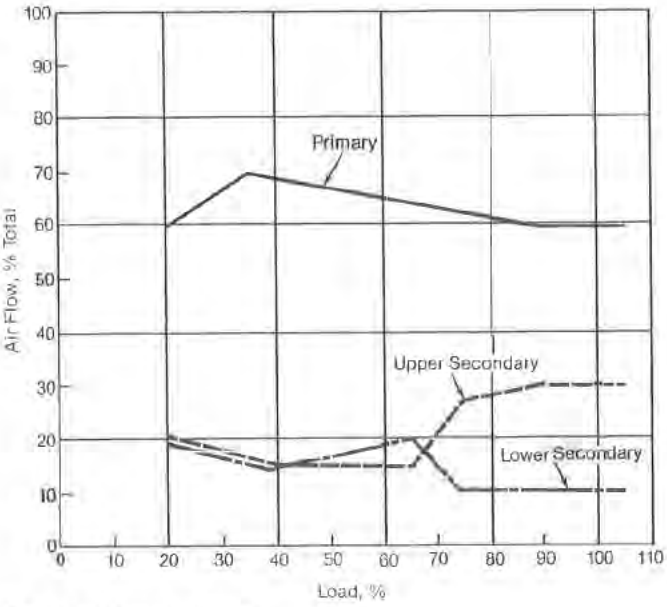


Fig. 17 Air flow distribution.

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removal exceeds that needed for total inventory control, a depletion of upper furnace inventory will result. In this case, means must be provided to bring the system back into balance. Furnace inventory size distribution can be corrected by changing the particle size distribution of the input solids or by classifying and recirculating a portion of the bed drain flow. Like the bubbling bed, furnace pressure drop measurements provide the required control signals.

Overfire air control

The split between primary air and overfire air is established, primarily, to optimize fuel burnout and CO and NO_x emissions. Fig. 17 shows a typical curve of air flow versus load for a bituminous coal-fired boiler. As a result of variations among fuels, final air splits are established as a result of unit operation and control system tuning for optimum combustion and furnace inventory distribution.

Summary

By traditional power industry development standards, fluidized-bed boilers represent an emerging technology. However, during the past 20 years the efforts have brought the technology from concept to commercial status. Activities are continuing to improve existing designs and extend unit size to larger capacities.

In the quest to reduce electric power costs and emissions, pressurized fluidized-bed boiler plants are being built and operated. (See Chapter 29.) In addition, investigators are re-examining fluidized-bed gasification to determine if improved power generating systems can be designed. (See Chapter 17.)

Today thousands of fluidized-bed units are in operation and are providing industry with a technique capable of burning a wide variety of fuels in an environmentally improved manner.

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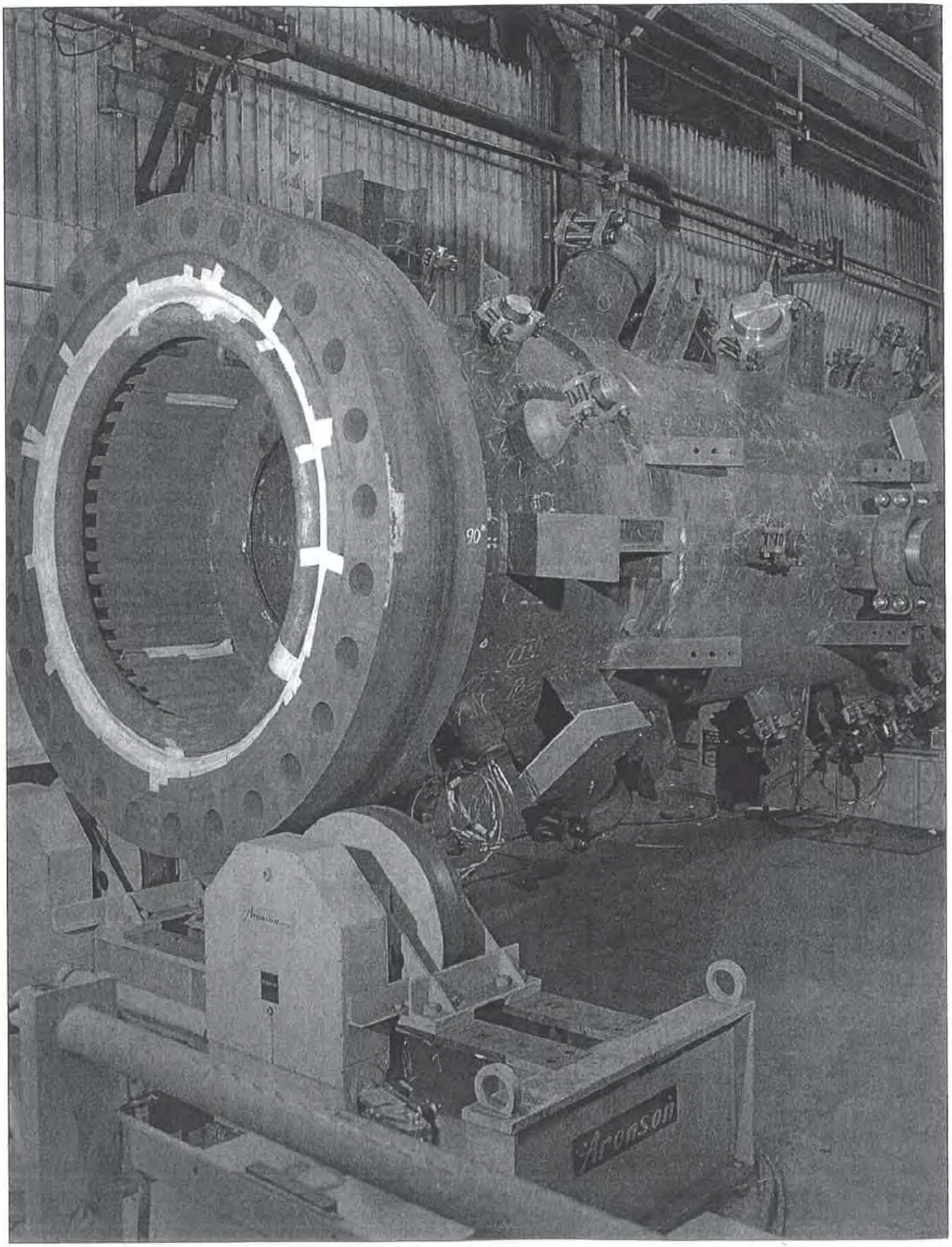
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Portion of an entrained-flow gasifier under fabrication at B&W.

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Chapter 18
Boilers, Superheaters and Reheaters

In a modern steam generator, various components are arranged to efficiently absorb heat from the products of combustion and provide steam at the rated temperature, pressure and capacity. These components include the boiler, superheater, reheater, economizer and air heater. They are supplemented by systems for steam-water separation (see Chapter 5) and the control of steam outlet temperature. The entire boiler system can be divided into two general sections, the furnace and convection pass. The furnace provides a large open volume with water-cooled enclosure walls where combustion takes place and the combustion products are cooled to an appropriate furnace exit gas temperature (FEGT). The convection pass contains tube bundles which consist of the superheater, reheater, boiler bank and economizer. The convection pass is usually followed by the air heater. The boiler or evaporation-circulation system usually includes the furnace enclosure walls, steam drum and steam-water separation equipment, boiler bank tube bundle, and associated connecting piping (downcomers, supplies and risers as discussed in Chapter 1). This chapter focuses on the boiler, superheater and reheater components plus systems for steam temperature control, steam bypass and unit start-up. Chapter 19 discusses economizers and air heaters.

Boilers

Boiler surface is defined as the tubes, drums and shells which are part of the steam-water circulation system and which are in contact with the hot gases. Although the term *boiler* now frequently refers to the overall steam generating system, the term *boiler surface* excludes the economizer, superheater, reheater or any component other than the steam-water circulation system itself.

While boilers can be broadly classified as shell, fire tube and water tube types, as discussed in the introduction to *Steam*, modern high capacity boilers are of the water tube type. In the water tube boiler, the water and steam flow inside the tubes and the hot gases flow over the outside surfaces. The boiler circulation system is constructed of tubes, headers and drums joined in such a way that water flow is provided to generate steam while cooling all parts. The water tube construction allows greater boiler capacity and higher pressure than shell or fire tube designs. Also, the water tube boiler offers greater versatility in arrangement; this permits the most efficient use of the furnace, superheater, reheater and other heat recovery components.

Boiler configurations

Modern high capacity boilers come in a variety of designs, sizes and configurations to suit a broad range of applications. Sizes range from 1000 to 10,000,000 lb/h (0.13 to 1260 kg/s) and pressures range from one atmosphere to above the critical pressure.

The boiler configuration is largely determined by the combustion system, fuel, ash characteristics, operating pressure and total capacity. The diversity in configurations is illustrated in Figs. 1 through 5.

Typical industrial and small power boilers The Integral Furnace boiler shown in Fig. 1 is an oil- and gas- fired, low pressure two drum package boiler. In small capacities it can be entirely shop assembled and shipped to the site. (See Fig. 1, Chapter 1.) Because it burns a clean fuel, provision has not been made for flyash collection or surface cleaning, and a small furnace volume can be used. A boiler bank, closely spaced tubes between a steam drum and lower drum, provides the heat transfer surface necessary for the rated steaming capacity. Fig. 2 illustrates a two drum Stirling® power boiler (SPB) with a controlled combustion zone (CCZ™) furnace specially designed for effective firing of high moisture wood and biomass. (See also Chapter 28.) A spreader-stoker firing system is supplied and a boiler bank is provided for sufficient steam generating surface.

Additional unique designs include fluidized-bed boilers, process recovery boilers and waste-to-energy boilers discussed in Chapters 16, 26 and 27.

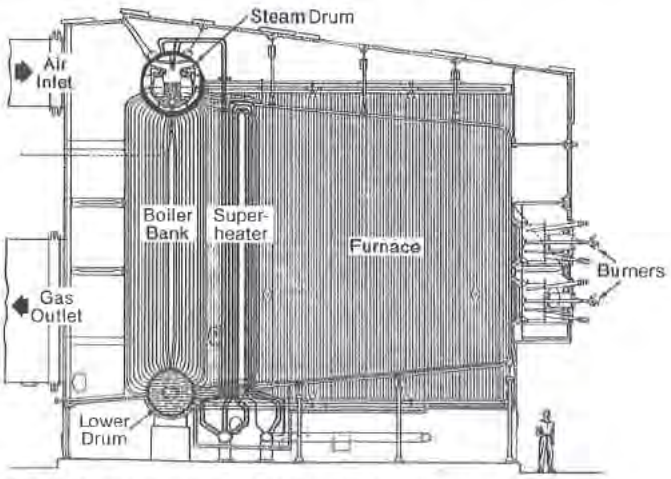


Fig. 1 Integral Furnace industrial boiler for oil and gas firing.

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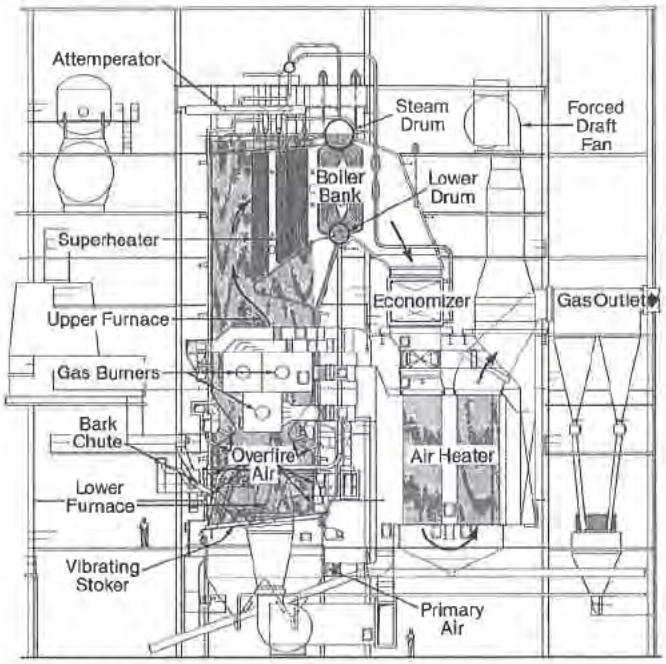


Fig. 2 Two drum Stirling® power boiler — CCZ™ stoker furnace configuration for bark firing.

Large utility boiler designs Figs. 3 and 4 illustrate two variations of the Babcock & Wilcox (B&W) Radiant boiler (RB) for natural circulation drum type steam generating systems. Fig. 5 illustrates one version of the B&W Universal Pressure (UP) boiler designed for once-through flow at supercritical or subcritical pressures. As discussed in Chapter 24, all of these units feature gas-tight, fully water-cooled furnace enclosure walls and floors made of all-welded membrane panel construction. Each design normally includes a single reheat section, although the supercritical once-through boiler has also been supplied as a double reheat unit. Fig. 3 shows an RB unit for coal firing. This is the Carolina-type design (RBC) with downflow convection backpass which minimizes overall steam generator height. Provision is made for sootblower surface cleaning and for flyash collection. A Tower-type (RBT) configuration is also available which features fully drainable convection pass surfaces and a minimum plan area. (See Chapter 24.) Fig. 4 is a Radiant boiler of the El Paso (RBE) configuration for oil and gas firing. This unit is very compact because of the relatively clean fuels being used. The compact design minimizes the boiler footprint (plan area) and support steel. Selected provision may be made for cleaning equipment.

Fig. 5 is a 1300 MW pulverized coal unit for supercritical pressures using vertical furnace wall tubes. As discussed in Chapter 24, the once-through circuitry design eliminates the need for a steam drum. Multiple fluid passes through the furnace minimize the imbalances in steam temperature around the furnace periphery which could result from nonuniform heat input. Other UP designs are available for subcritical pressure operation with spiral circuitry furnaces. (See Fig. 11 of Chapter 24.) Inclined tubes wrapping around the furnace make multiple fluid passes through the furnace enclosure unnecessary.

Boiler design

Regardless of the size or configuration, modern boiler design remains driven by four key factors: 1) efficiency (boiler and cycle), 2) reliability, 3) capital and operating cost and 4) environmental protection. These factors, combined with specific applications, produce the diversity of designs presented above and discussed at length in Chap-

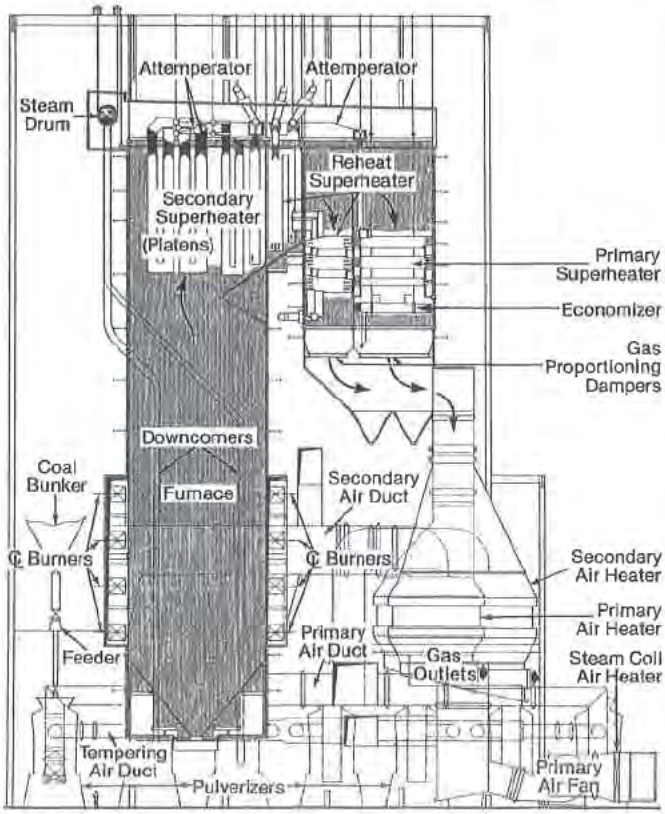


Fig. 3 Carolina-type Radiant boiler for pulverized coal firing. Design pressure 2950 psig (203.4 bar gauge); primary and reheat steam temperatures 1005F (541C); capacity 4,900,000 lb steam/h (617 kg steam/s).

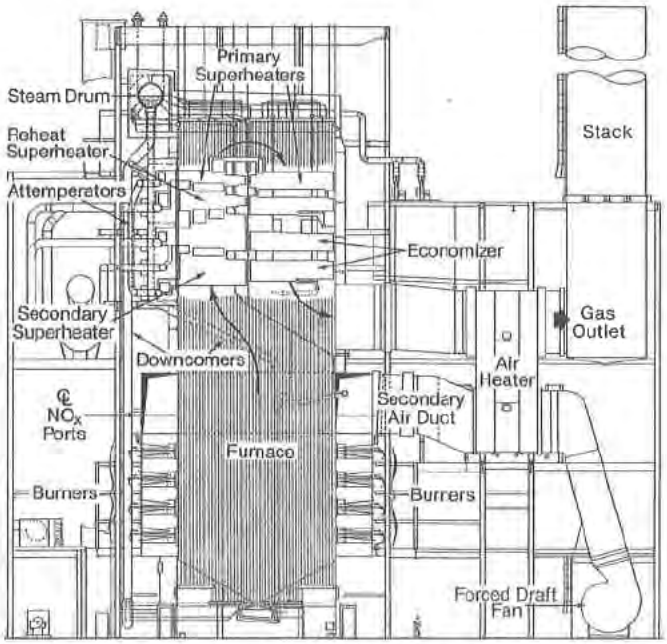


Fig. 4 El Paso-type Radiant boiler for gas firing. Design pressure 2550 psig (175.8 bar gauge); primary and reheat superheat temperatures 955F (513C); capacity 3,825,000 lb steam/h (482 kg steam/s).

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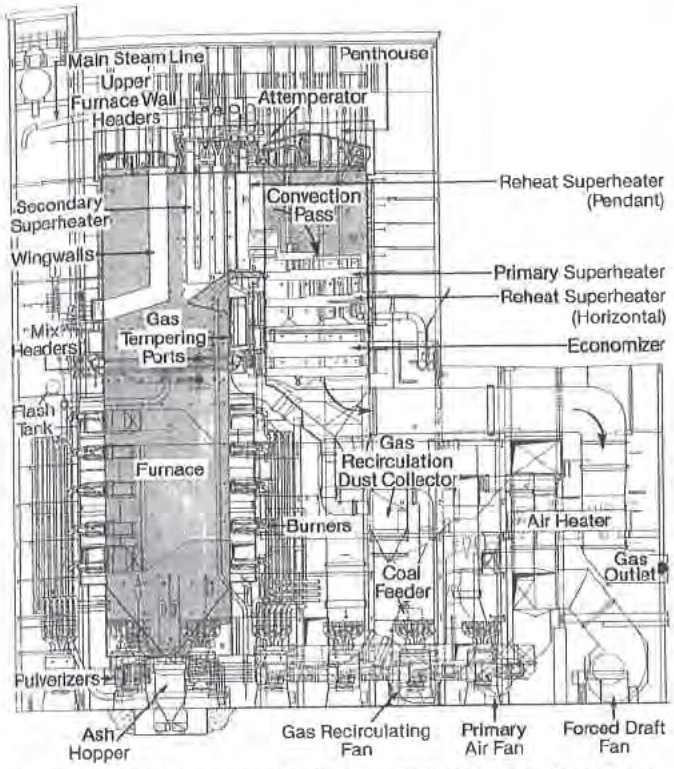


Fig. 5 Universal Pressure boiler for pulverized coal firing. Superheater outlet pressure 3845 psig (265.1 bar gauge); primary and reheat steam temperatures 1010F (543C); capacity 9,775,000 lb steam/h (1232 kg steam/s).

ters 24 through 31. However, all of these units share a number of fundamental elements upon which the site- and application-specific design is based.

The boiler evaluation begins by identifying the overall application requirements specified in Table 1. These are generally selected in an iterative process balancing initial capital cost, operating costs (especially fuel), steam process needs and operating experience. The selection of these parameters can dramatically impact cost and thermal efficiency. These are addressed further in Chapters 1 and 37.

From a boiler evaluation perspective, the temperature-enthalpy diagram shown in Fig. 6 (for a typical high pressure, single reheat unit) provides important design information about the unit configuration. In this example, the relative heat absorption for water preheating, evaporation and superheating are 30%, 32% and 38%, respectively. Reheating the steam increases the total heat absorption by approximately 20%. For cycles at supercritical operating pressures, a second stage of reheat may be added. For process applications, only the preheating and evaporation steps may be required.

Boilers can be designed for subcritical or supercritical pressure operation. At subcritical pressures, the furnace enclosure is cooled by constant-temperature boiling water, and the flow circuits must be designed to accommodate the two-phase steam-water flow and boiling phenomena addressed in Chapter 5. At supercritical pressures, the water acts as a single-phase fluid with a continuous increase in temperature as it passes through the boiler. These designs require special consideration to avoid excessive unbalances in metal temperatures due

Table 1
Use-Derived Specifications for Boiler Design

Specified Parameter	Comments
Steam use	Flow rates, pressures, temperatures — for utility boilers, the particular power cycle and turbine heat balance.
Fuel type and analysis	Combustion characteristics, fouling and slagging characteristics, ash analysis, etc.
Feedwater supply	Water source, analysis and economizer inlet temperatures.
Pressure drop limits	Gas side and steam side.
Government regulations	Including emission control requirements.
Site-specific factors	Geographical and seasonal characteristics.
Steam generator use	Base load, cycling, etc.
Customer preferences	Specific design guidelines such as flow conditions, equipment preferences and steam generator efficiency.

to variations in heat pickup in different flow circuits. In addition, special heat transfer phenomena must also be addressed. (See Chapter 5.)

Two basic fluid circulation systems are used — natural circulation and once-through. In natural circulation systems which operate at subcritical pressures, water is only partially evaporated in the boiler circuits producing a steam-water mixture at the tube outlets. Steam-water separation equipment is provided to separate the steam and water, supply saturated (dry) steam to the superheater and recirculate water back to the boiler circuits. Natural circulation is the result of the density difference between the hot and cold legs of the loop. Indus-

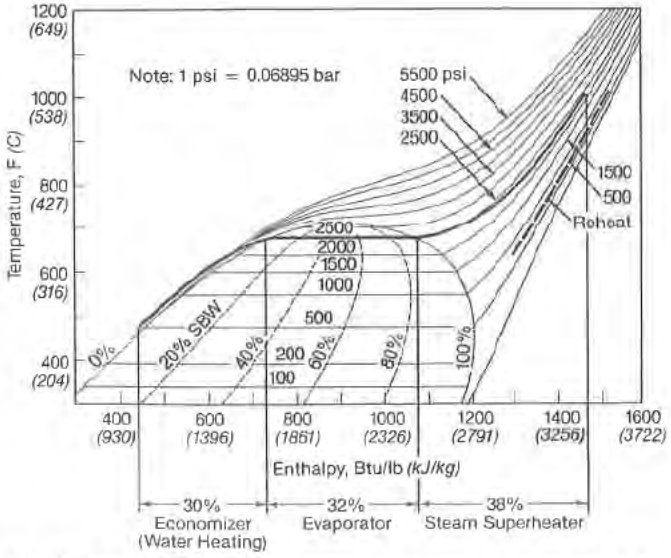


Fig. 6 Temperature-enthalpy diagram for subcritical pressure boiler absorption — one reheat section.

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try accepted water chemistry limits are typically less stringent, especially at low pressures, and steam-water pressure drop inside the tubes is less of an issue. In B&W UP once-through designs, the steam drum and internal steam separation equipment are eliminated and a separate startup system added. UP boilers have been designed for both subcritical and supercritical operation. At supercritical pressures, the system can increase overall power cycle efficiency but at a higher initial capital cost. More precise operation is needed (see Chapter 43) and more stringent water treatment is required. (See Chapter 42.) Hybrid recirculating and once-through systems are presented in Chapter 5.

Design criteria Within the preceding framework, the important items which must be accomplished in boiler design are the following:

1. Define the energy input based upon the steam flow requirements, feedwater temperature, and an assumed or specified boiler thermal efficiency.
2. Evaluate the energy absorption needed in the boiler and other heat transfer components.
3. Perform combustion calculations to establish fuel, air and gas flow requirements. (See Chapter 9.)
4. Determine the size and shape of the furnace, considering the location and space requirements of the burners or other combustion system, and incorporating sufficient furnace volume for complete combustion and low emissions. Provision must be made for handling the ash contained in the fuel and cooling the flue gas so that the furnace exit gas temperature (FEGT) meets design requirements.
5. Determine the placement and configuration of convection heating surfaces. The superheater and reheater, when provided, must be placed where the gas temperature is high enough to produce effective heat transfer, yet not so high as to cause excessive tube temperatures or ash fouling. All convection surfaces must be designed to minimize the impact of slag or ash buildup and permit surface cleaning without erosion of the pressure parts.
6. Provide sufficient saturated boiler surface to generate the remainder of the steam not generated in the furnace walls. This can be accomplished with or without an economizer.
7. Design pressure parts in accordance with applicable codes using approved materials.
8. Provide a gas-tight boiler setting or enclosure around the furnace, boiler, superheater, reheater and economizer.
9. Design pressure part supports and the setting for expansion and local conditions, including wind and earthquake loading.

Fuel selection and specification are particularly important. Boiler systems are designed for specific fuels and frequently will encounter combustion, slagging, fouling or ash handling problems if a fuel with characteristics other than those originally specified is fired. All potential boiler fuels must be assessed to determine the most demanding fuel.

Procedures for optimizing steam pressure and temperature, and for evaluating the value of specific auxiliary equipment in a given application are outlined in Chapter 37.

As discussed in Chapter 9, the boiler or combustion efficiency is usually evaluated as 100 minus the sum of the heat losses expressed as a percentage. Chapter 9 provides procedures for calculating the corresponding fuel, air and gas flow rates.

Enclosure surface design

The furnace of a large pulverized coal-, oil- or gas-fired boiler is essentially a large enclosed volume where fuel combustion and cooling of the combustion products take place prior to their entry into the convection pass tube bundles. Excessive gas temperatures entering these tube banks could lead to elevated metal temperatures or unacceptable fouling and slagging. Heat transfer to the furnace enclosure walls is basically controlled by radiation. The walls can be cooled by either boiling water (subcritical pressure) or high velocity supercritical pressure water.

The convection pass enclosure contains the horizontal and vertical downflow gas passes, where most of the superheater, reheater and economizer surfaces are located. (See Fig. 3.) These enclosure surfaces can be water or steam cooled.

The furnace enclosures and convection pass enclosures are usually made of water-cooled tubes in an all-welded membrane construction. These enclosures have also been made from tangent tube construction or closely spaced tubes with an exterior gas-tight seal. For a membrane construction, the tube wall and membrane surfaces are exposed on the furnace side to the combustion process while insulation and lagging (sheet metal) on the outside protect the boiler, minimize heat loss and protect operating personnel. (See Chapter 22.)

Furnace size versus cycle requirements Besides providing the volume necessary for complete combustion and a means to cool the gas to an acceptable FEGT, the furnace enclosure also provides much of the steam generating surface in a boiler. These roles may not perfectly match one another. In coal-fired units, the minimum furnace volume is usually set to provide the fuel ash specific FEGT. Frequently this results in too much evaporator surface in high pressure boilers and too little surface in low pressure units to meet the thermodynamic requirements for the desired steam exit temperature.

Fig. 7 illustrates the effect of the steam cycle and the operating pressure and temperature on the relative energy absorption between the boiler/economizer and superheater/reheater. As the pressure and temperature increase, the total unit absorption for a given power production progressively declines because of increased cycle efficiency. The boiler and economizer absorption represents the relative amount of heat added to the entering feedwater to produce saturated steam or reach the critical point in a supercritical pressure UP boiler. As the operating pressure increases, the amount of heat required to produce saturated steam declines. Conversely, the amount of heat required for the superheater and reheater increases. The change in required boiler/economizer absorption may not seem significant. However, a 1% shift in absorption is equivalent to approximately 10F (6C) of superheat or reheat temperature.

On a drum unit, the furnace and water-cooled convection pass enclosure walls are the boiler surface. On low pressure units, the amount of heat absorbed in the fur-

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nace is usually not adequate to produce all the saturated steam required, and a boiler bank is installed after the superheater. (See Figs. 1 and 2.) On a high pressure unit, the heat absorbed by the furnace and economizer is adequate to produce all the saturated steam required. As the furnace size increases, the economizer can be made smaller to produce the same amount of steam. It can be envisioned that as the furnace is made larger, some point will be reached where an economizer will not be required. However, as the furnace is enlarged to reduce the FEGT, too much steam would be produced, leaving insufficient energy in the flue gas to meet design superheat and reheat temperatures. This situation requires special design features to meet all of the design goals.

Furnace design criteria The furnace is basically a large open volume enclosed by water-cooled walls for combustion. Its shape and volume are established by the selection of fuel and combustion system. For wall firing using circular burners (see Chapter 13), minimum clearances between individual burners, between burners and walls, as well as between burners and the furnace floor are established based upon physical clearances and functional criteria for complete combustion. These clearances prevent fuel stream and flame interaction, assure complete combustion, avoid unacceptable flame impingement on the walls (which could lead to tube overheating or excessive deposits) and minimize the formation of nitrogen oxides (NO_x). The maximum fuel input rate, number of burners and the associated clearances establish: 1) the furnace cross-sectional area, 2) the height of the combustion zone, 3) the height of the overfire air injection zone (if used), and 4) the distance between the burner zone and the furnace floor. Where the fuel is burned on stokers, the furnace cross-sectional area is established by a specified heat release rate per unit bed area. (See Chapter 15.)

As discussed in detail in Chapter 13, combustion system design and its impact on furnace volume and shape have become complex and critical as emissions limits have been reduced. Not only are low NO_x burners such as the B&W DRB-XCL™ used to reduce NO_x emissions, but other techniques, such as in-furnace staging with overfire air (NO_x ports), fuel reburning and reactant injection, can also be considered for NO_x reduction. In addition, some techniques for furnace sorbent injection to reduce sulfur dioxide (SO_2) emissions have also been developed. (See Chapter 35.) Each technique can have an impact on the furnace size and configuration.

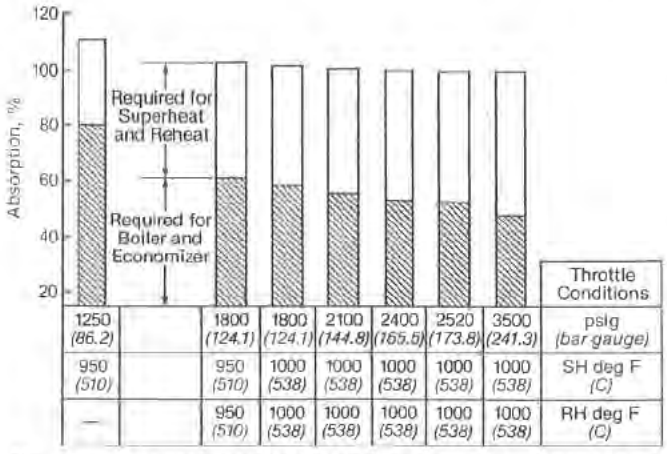


Fig. 7 Relative heat absorption by operating pressure and steam temperature.

The overall furnace height is established by several criteria. For clean fuels, such as natural gas, the furnace volume and height are generally set to cool the combustion products to an FEGT which will avoid superheater tube overheating. For fuels such as coal and some oils which contain significant levels of ash, the furnace volume and height are established to cool the products of combustion to an FEGT which will prevent excessive fouling of the convection surfaces. The relationships between furnace volume and FEGT are explored in Chapter 4. The specific effects of ash on furnace design are discussed below and in Chapter 20. The furnace height must also be set to provide at least the minimum time to complete combustion and to meet minimum clearance requirements from the burners and NO_x ports to the arch and convective surface.

Ash effects In the case of coal and to a lesser extent with oil, an extremely important consideration is ash in the fuel. If this ash is not properly considered in the unit design and operation, it can deposit on the furnace walls, sloping surfaces and throughout the convection pass tube banks. Ash not only reduces the heat absorbed by the unit, but it increases draft loss, erodes pressure parts and eventually can result in unit outages for cleaning and repairs. (See Chapter 20.)

In coal-fired furnaces the ash problems are the most severe. There are two general approaches to ash handling: the dry ash or dry-bottom furnace and the slag-tap or wet-bottom furnace.

In the dry ash furnace, which is particularly applicable to coals with high ash fusion temperatures, a hopper bottom (Figs. 3 and 5) and sufficient cooling surface are provided so that the ash impinging on furnace walls or the hopper bottom is solid and dry; it can be removed essentially as dry particles. When pulverized coal is burned in a dry ash furnace, about 80% of the ash is carried through the convection banks. The chemistry of the ash can have a dramatic impact on the furnace volume necessary for satisfactory dry-bottom unit operation. This is illustrated in Fig. 8 which compares the furnace volume required for a nominal 500 MW boiler burning a low slagging bituminous or subbituminous coal to that for a high slagging lignite coal. The relationships defining the furnace size requirements are discussed in detail in Chapter 20.

With many coals having low ash fusion temperatures, it is difficult to use a dry-bottom furnace because the slag is molten or sticky; it clings and builds up on the furnace walls and hopper bottom. The slag-tap or wet-bottom furnace has been developed to handle these coals. The most successful form of the slag-tap furnace is that used with Cyclone furnace firing. (See Figs. 7 and 8, Chapter 14.) The furnace comprises a two-stage arrangement. In the lower part of the furnace, sufficient gas temperature is maintained so that the slag drops onto the floor in liquid form. Here, a pool of liquid slag is maintained and tapped into a slag tank containing water. In the upper part of the furnace, the gases are cooled below the ash fusion point, so ash carried over into the convection banks is dry and does not cause excessive fouling. Because of high NO_x emissions, slag-tap designs are used infrequently on new boilers.

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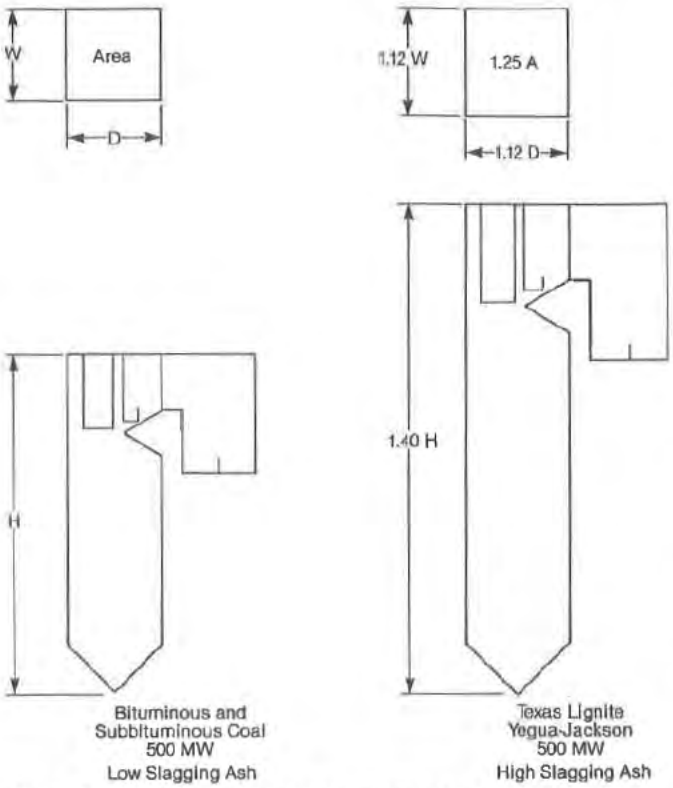


Fig. 8 Boiler size comparison for alternate coal types.

Water-cooled walls Most boiler furnaces have all water-cooled membrane walls. This reduces maintenance on the furnace walls and reduces the temperature of the gas entering the convection bank to the point where slag deposits and superheater corrosion can be controlled by sootblowing.

Furnace wall tubes are spaced on close centers to obtain maximum heat absorption while maintaining tube and membrane temperatures and thermal stresses within limits. The membrane panels (see Fig. 9) are composed of tube rows spaced on centers wider than a tube diameter and joined by a membrane bar securely welded to the adjacent tubes. This results in a continuous wall surface of rugged, pressure-tight construction capable of transferring the required heat from the furnace gas to the steam-water mixture in the tubes. The width and length of individual panels are suitable for economical manufacture and assembly, with bottom and top headers shop-attached prior to shipment for field assembly. Size limitations are frequently governed by shipping clearances and erection constraints. Membrane construction with refractory lining is used in the lower furnace walls of Cyclone-fired, refuse and fluidized-bed units.

Convection boiler surface

Some designs include boiler tubes as the first few rows of tubes in the convection bank. The tubes are spaced to provide gas lanes wide enough to prevent ash and slag pluggage and to facilitate cleaning for dirty fuels. These widely spaced boiler tubes, known as the *slag screen* or *boiler screen*, receive heat by radiation from the furnace and by radiation and convection from the combustion gases passing through them. Another option is the use of water- or steam-cooled wingwalls in the upper furnace. This provides additional radiant boiler surface in the furnace while allowing the furnace size to be optimized.

In the larger high pressure units, superheater surface generally forms the furnace outlet plane. The gas temperature entering the superheater must be high enough to give the desired superheat temperature with a reasonable amount of heating surface and the use of economical materials. The arrangements in Figs. 1 through 5 illustrate various configurations of superheater surface at the furnace outlet. Also, to optimize superheater design, widely spaced steam-cooled platens or wingwalls may be incorpo-

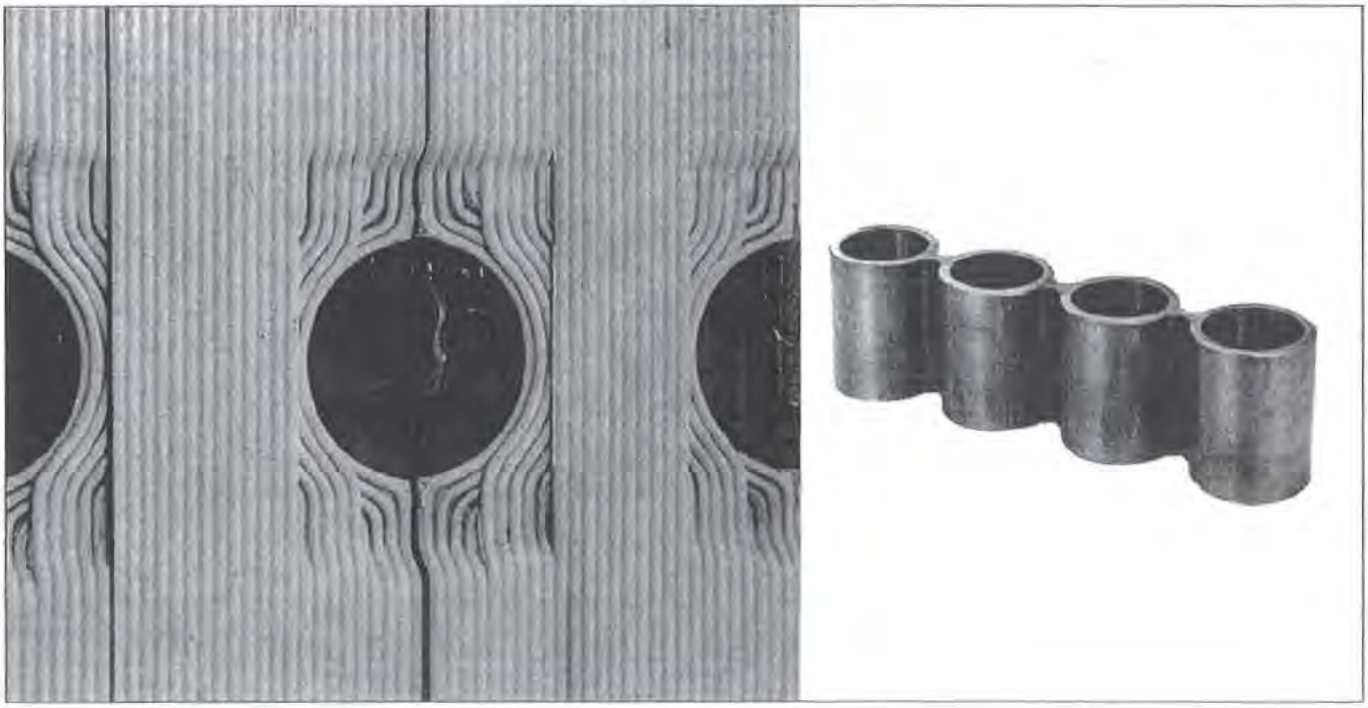


Fig. 9 Membrane wall construction at burner openings.

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rated in the upper furnace as shown in Figs. 3 and 5. Fig. 10 shows a plan arrangement of convection surface and the change in average gas temperature.

Design of boiler surface after the superheater depends on the type of unit (industrial or utility), desired gas temperature drop and acceptable gas side resistance (draft loss) through the boiler surface. Typical arrangements of boiler and superheater surface for various types of boilers are illustrated in the preceding section on boilers. In designing convection heating surfaces, the objective is to establish the proper combination of several parameters to provide the desired gas temperature drop with allowable flue gas resistance. These parameters are tube diameter and length, tube spacing, number and orientation of tubes, and gas baffling.

In the gas side design of convection surfaces, the quantity of heating surface (ft^2 or m^2) needed for a given load or duty (Btu/h or W) is generally inversely related to gas side flow resistance or pressure loss. Design changes which increase resistance, such as tightening tube spacing perpendicular to flow, result in higher heat transfer rates (Btu/h ft^2 or W/m^2). This in turn reduces the amount of heating surface needed to carry the desired total thermal load. An optimal gas mass flux and design result from balancing the capital cost of the heating surface against the operating cost of fan power needed to overcome the resistance.

For a given gas flow rate, a considerably higher gas film heat transfer coefficient, heat absorption and draft loss result when the gases flow at right angles to the tubes (crossflow) compared to flow parallel to the tubes (long flow). Gas turns between tube banks generally add draft loss and maldistribution with little benefit to heat absorption. Turns leaving the upstream bank and entering the downstream bank of tubes should therefore be designed for minimum resistance and optimum distribution.

Determination of the amount of radiation and convection boiler surface required for a specified heat transfer is illustrated in Chapter 21.

Numerical analysis

When one dimensional analysis and experimental data are insufficient for design alone, numerical computer modeling can be used to determine local heat transfer and flue gas conditions in boilers, superheaters and reheaters. These numerical tools are complex multi-dimensional computer codes which solve conservation equations for mass, momentum, energy and species. (See Chapters 3 and 4.)

A numerical model of the superheater and/or reheater regions generally includes the boiler enclosure, burners and any other geometric aspects which may affect the flow. Downstream and upstream components are sometimes added to provide the best boundary conditions to the area of interest. In applying these computer codes, model size and complexity must be balanced against the associated computer calculation time required. Required numerical inputs include:

1. geometric data — boiler and heat exchanger surface configurations,
2. performance data — heat exchanger pressure drops and heat absorptions,
3. operating data — stoichiometry and furnace exit gas temperature,

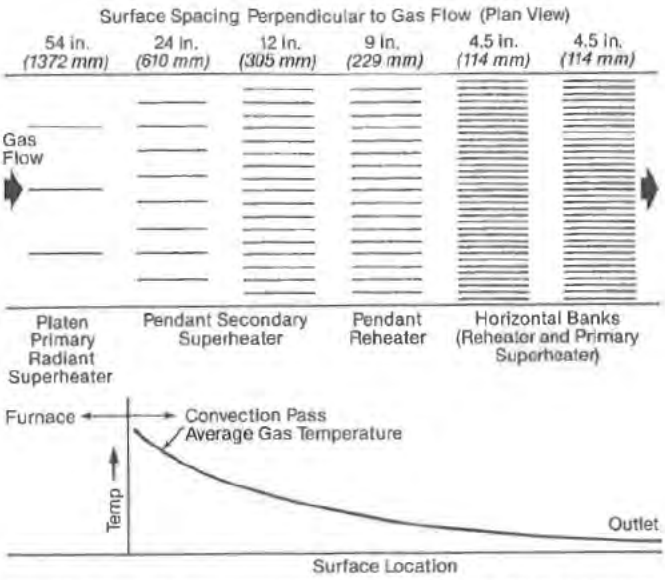


Fig. 10 Schematic plan arrangement of convection surface and change in average gas temperature.

4. burner setup — vane angles and turbulence intensity, and
5. fuel properties — fuel and flue gas quantities, composition and coal fineness, if applicable.

Use of operating data along with predictions based on company standards improves the accuracy of the model by tuning it to actual conditions. This is a necessary step because actual local conditions, e.g., ash thickness and property variation, are generally not known. Often, published data and sound engineering judgment are relied upon to obtain otherwise unavailable information. A numerical model can then be run once the boundary and inlet conditions are calculated and fluid properties are known.

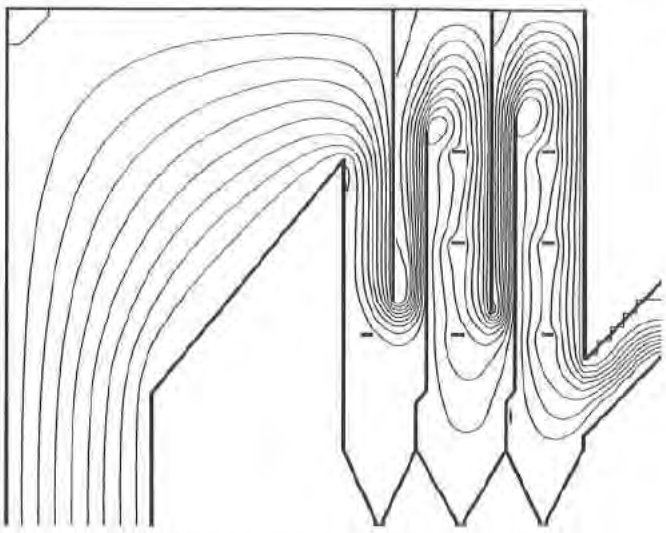
After the model has been successfully run, the results can be postprocessed to determine the dominant physical characteristics. Further postprocessing can be used to obtain derived quantities such as slagging resistance or mixing effectiveness.

Numerical analyses fall into three general categories: 1) verification of proposed designs, 2) evaluation of design modifications, and 3) investigation of localized problems. Figs. 11, 12 and 13 provide sample numerical computer model results for illustration purposes. Figs. 11a and 11b are typical streamline plots containing curves that are tangent to the gas flow in a proposed boiler design. The design was modified as shown because the amount of heat transfer surface being bypassed by the gas flow lowered the expected heat absorption. Such evaluation of design modifications can be performed prior to fabrication to reduce long term costs and improve unit performance.

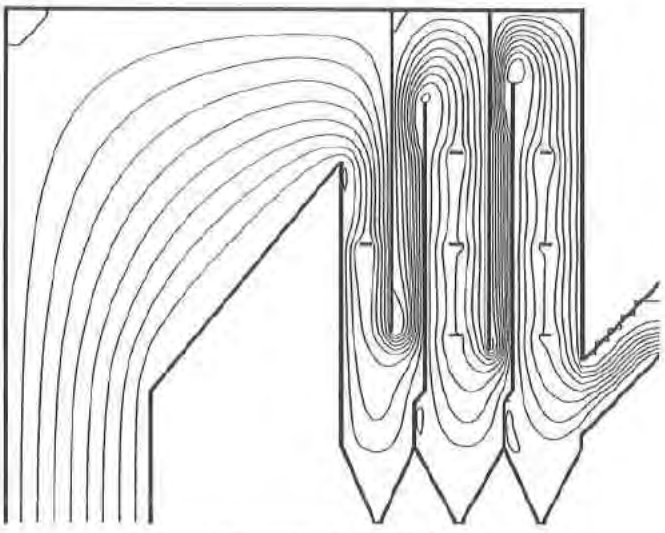
A second numerical model application includes an analysis to determine the optimum NO_x port arrangement in a boiler system. Such an analysis can use mixing effectiveness to rank the individual arrangements. Fig. 12 contains plots of fuel-air stoichiometry at an elevation 15 ft (4.6 m) above the NO_x ports before (a) and after (b) NO_x port adjustment. The more uniform distribution with improved mixing shown in (b) results in less NO_x formation.

Numerical modeling is also well suited to investigate boiler problems. The model takes into account the actual boiler geometry and performance characteristics, i.e., heat

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a) Original Convection Pass Baffle Arrangement



b) Improved Convection Pass Baffle Arrangement

Fig. 11 Numerical modeling results — flow streamlines.

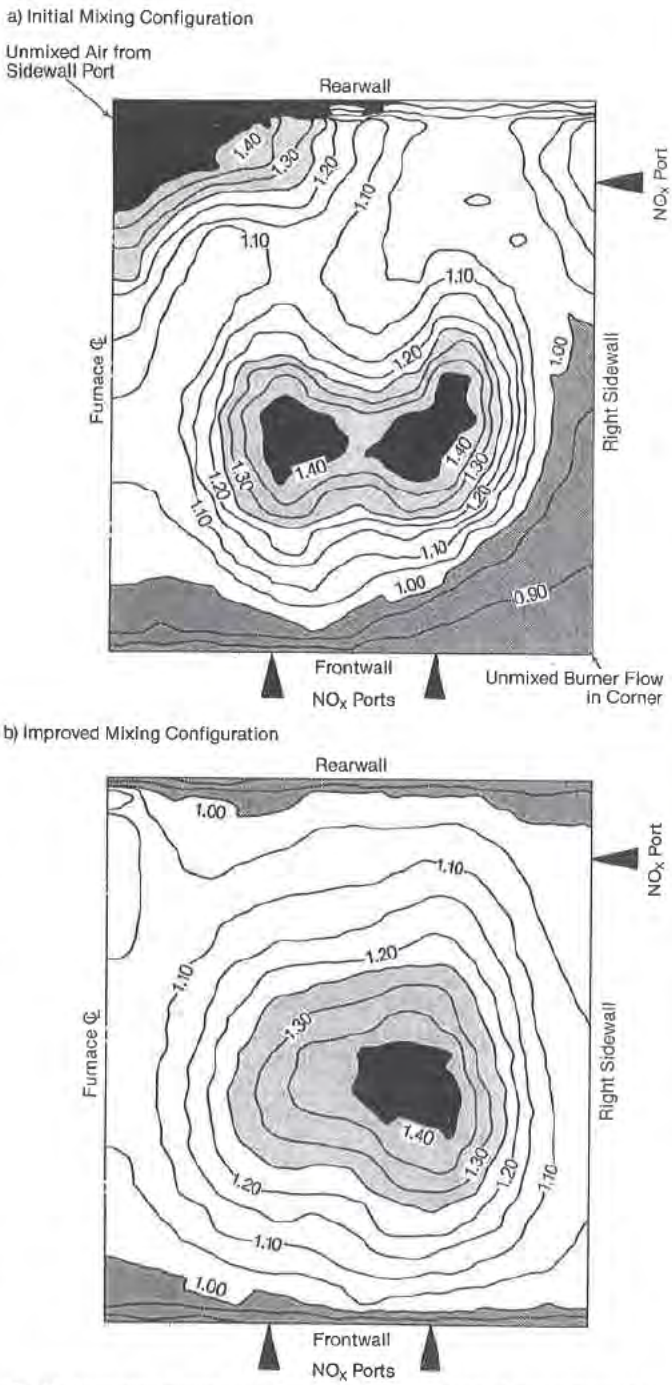


Fig. 12 Numerical modeling results — local combustion fuel-air stoichiometry.

absorption, temperature and pressure drop. The model also provides a more detailed and clearer performance picture than can be attained using analytical, field or laboratory methods. As an example, a unit was experiencing excessive slag accumulation on the lower leading edge of the secondary superheater inlet bank. Numerical modeling was used to determine the cause of the slagging and to evaluate several design modifications to reduce it. High velocities and associated high ash loadings were found to occur on the lower surfaces with a bias toward the sidewalls. Temperature contour plots for the as-built nominal case and for a larger furnace arch case (shown in Fig. 13) indicated that increasing the size of the arch moved the hotter furnace gases away from the secondary superheater problem area, virtually eliminating the hot ash impaction. This reduction of about 100F (56C) in the maximum temperature entering the secondary superheater enhanced the unit's resistance to slagging.

Accurate modeling of boilers and boiler components depends heavily upon the availability of information to determine the proper inlet conditions, boundary conditions and fluid properties. Careful qualification of the model is required and is augmented by known global characteristics such as pressure drop and heat absorption. Correct application of accurate information to develop numerical models will improve designs and determine the cause of localized problems.

Design of pressure parts

Boilers have achieved today's level of safety and reliability through the use of sound materials and safe practices for determining acceptable stresses in drums, headers, tubes and other pressure parts. Boilers must be designed to applicable codes. Stationary boilers in the United States

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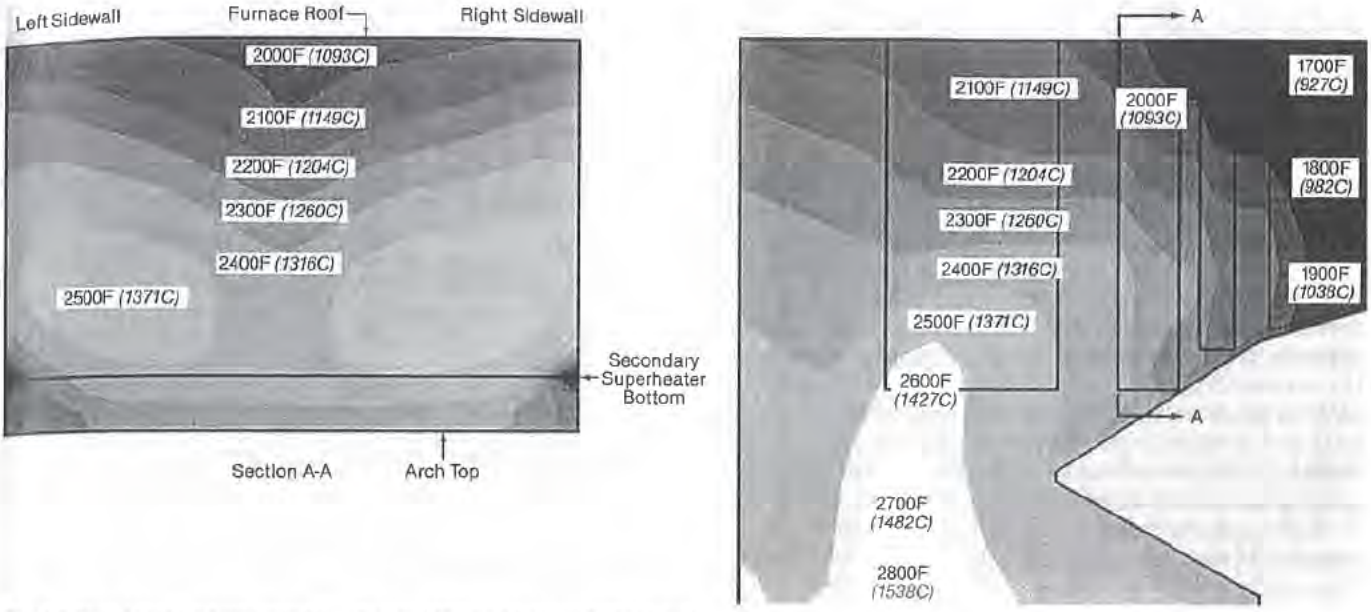


Fig. 13 Numerical modeling results — flue gas temperatures at furnace exit.

(U.S.) are designed to the American Society of Mechanical Engineers (ASME) Boiler and Pressure Vessel Code. (See Chapter 7 and Appendix 2.) The allowable design stress depends on the maximum temperature to which the part is subjected and, therefore, it is important that pressure part design temperatures are known and not exceeded in operation. Drum boiler enclosure material temperatures are a function of upset spot heat flux, design pressure, metal conductivity and the saturation temperature corresponding to the maximum boiler operating pressure. These parameters are also used to determine each tube outside diameter and thickness. For boiler tubes, temperatures are maintained at known levels by providing a sufficient flow of water to prevent the occurrence of CHF, or critical heat flux phenomena. (See Chapter 5.) An adequate saturated water velocity must exist for each tube and particular attention must be given to high heat flux zones and sloped tubes with heat on top.

Because steam drums have thick walls, it is necessary to limit the heat flow through them to avoid excessive thermal gradients during startup, shutdown and normal operation. This is particularly important where the drum is exposed to flue gas. Where the drum is penetrated by a number of tube holes, the flow of water through these holes serves to cool the drum wall. Where the heat input through a drum would be too high because of high gas temperature or velocity, insulation may be provided on the outside of the drum or the drum relocated out of the heat.

In a drum-type boiler, steam separation equipment is provided to maintain steam moisture and solids at acceptable levels. (See Chapter 5.) In once-through boilers, all moisture is evaporated in the tubes, so that boiling and superheating occur sequentially. In boilers of this type, steam purity depends on maintaining adequate feedwater purity. (See Chapter 42.)

Boiler safety valves constitute very important protection items. (See Chapter 23.) The ASME Code stipulates that the boiler design pressure must not be less than the high-set safety valve relief pressure. As a practical matter, to avoid unnecessary losses and maintenance from frequent popping of the safety valves, the first valve should be set to relieve at not less than the boiler operating pressure plus 5%. The operating pressure in the steam drum, in turn, depends on the pressure required at the point of use and the intervening pressure drop. As an example, where the steam is used in a turbine, the boiler operating pressure is determined by adding the turbine throttle pressure and the pressure drop through the steam piping, nonreturn valve, superheater and drum internals at maximum unit steam flow.

Boiler enclosure

The methods used to provide a tight boiler setting and a tight enclosure around the superheater, reheater, economizer and air heater are described in Chapter 22.

Boiler supports

Furnace wall tubes are usually supported by the headers to which they are attached, and generating bank tubes and screens are supported by the drum or headers to which they are connected. As discussed in Chapter 7, the following considerations for proper support design are important:

1. The tubes must be arranged and aligned so that they are not subjected to excessive bending-moment stresses in supporting the weight of the tubes, headers, drums, attachments and fluid within. When the unit is bottom supported, the tubes must satisfy column buckling requirements.
2. The holding strength of the tube seats must not be exceeded.
3. Provision must be made to accommodate the expansion of the pressure parts. For a top supported unit, the hanger rods which tie the pressure parts to structural steel must be designed to swing at the proper angle. They must be long enough to withstand the movement without excessive stresses in the rods or the pressure parts. Bottom supported boilers should be anchored only at one point, guided along one line and allowed to expand freely in all other directions. To reduce the frictional forces and resultant stresses in the pressure parts, roller saddles or mountings are desirable for bottom supported heavy loads.

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Superheaters and reheaters

Advantages of superheat and reheat

When saturated steam is used in a turbine, the work done results in a loss of energy by the steam and subsequent condensation of a portion of the steam, even though there is a drop in pressure. The amount of work that can be done by the turbine is limited by the amount of moisture that it can handle without excessive turbine blade wear. This is normally between 10 and 15% moisture. It is possible to increase the amount of work done with moisture separation between turbine stages, but this is economical only in special cases. Even with moisture separation, the total energy that can be transformed to work in the turbine is small compared to the amount of heat required to raise the water from feedwater temperature to saturation and then evaporate it. Therefore, moisture content constitutes a basic limitation in turbine design.

Because a turbine generally transforms the energy of superheat into work without forming moisture, this energy is essentially all recoverable in the turbine. This is illustrated in the temperature-entropy diagram of the ideal Rankine cycle (shown in Fig. 6, Chapter 2). While this is not always entirely correct, the Rankine cycle diagrams in Chapter 2 indicate that this is essentially true in practical cycles.

The foregoing discussion is not specifically applicable at steam pressures at or above the critical pressure. In fact, the term *superheat* is not truly accurate in defining the temperature of the working fluid in this region. However, even at pressures exceeding 3208 psi (221 bar), heat added at temperatures above 705F (374C) is essentially all recoverable in a turbine.

The benefit of superheat is indicated by the reduction of cycle heat rate when steam temperatures entering the turbine are raised. For example, in a simple calculation of a 2400 psig (165.5 bar gauge) ideal Rankine cycle with a single reheat stage, an increase in superheat temperature from 900 to 1100F (482 to 593C) reduces the gross heat rate from approximately 7550 to 7200 Btu/kWh. This is more than a 4.5% efficiency improvement attributable to the superheat.

Superheater types

Two basic types of superheaters are available depending upon the mode of heat transfer from the flue gas. The original type was the *convection superheater*, for gas temperatures where the portion of heat transfer by radiation from the flue gas is small. With a unit of this design, the steam temperature leaving the superheater increases with boiler output because of the decreasing percentage of unit heat input that is absorbed in the furnace. This results in more heat available for superheater absorption. Because convection heat transfer rates are almost a direct function of gas flow rate and therefore boiler output, the total absorption in the superheater per pound of steam, and therefore steam temperature, increase with boiler output. (See Fig. 14.) This effect is increasingly pronounced the farther the superheater is located from the furnace and the lower the gas temperature entering the superheater.

A *radiant superheater* receives energy primarily by thermal radiation from the furnace with little energy from convective heat transfer. It usually takes the form of widely spaced [24 in. (609.6 mm) or larger side spacing] steam-cooled wingwalls or pendant superheat platens located in the furnace. It is sometimes incorporated into the furnace enclosure curtain walls. Because the heat absorption by furnace surfaces does not increase as rapidly as boiler output, the radiant superheater outlet temperature declines with an increasing boiler output, as shown in Fig. 14.

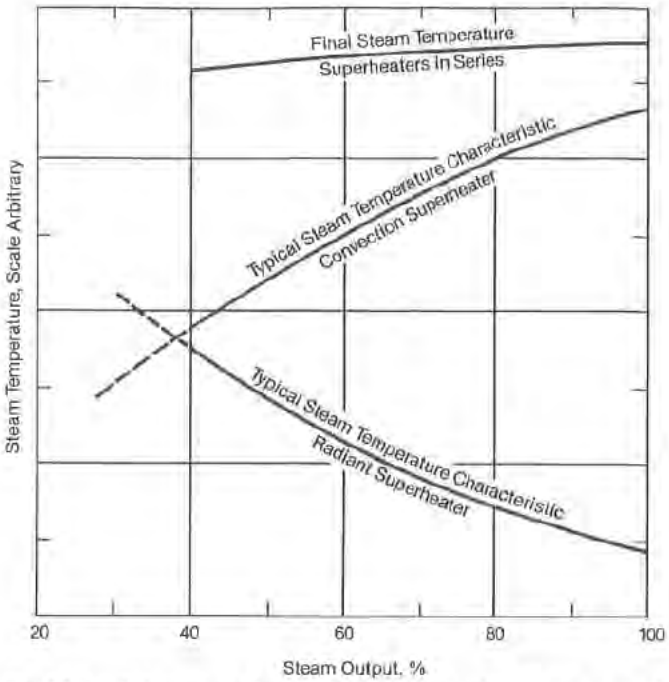


Fig. 14 A substantially uniform final steam temperature over a range of output can be attained by a series arrangement of radiant and convection superheater components.

In certain cases the two opposite sloping curves have been coordinated by the series combination of radiant and convection superheaters to give a flat superheat curve over a wide load range, as indicated in Fig. 14. A separately-fired superheater can also be used to produce a flat superheat curve.

The design of radiant and convective superheaters requires extra care to avoid steam and flue gas distribution differences which could lead to tube overheating. Superheaters generally have steam mass fluxes of 100,000 to 1,000,000 lb/h ft^2 (136 to 1356 kg/m^2 s) or higher. These are set to provide adequate tube cooling while meeting allowable pressure drop limits. The mass flux selected depends upon the steam pressure and temperature as well as superheater thermal duty. In addition, the higher pressure loss associated with higher velocities improves the steam side flow distribution.

The fundamental considerations governing superheater design also apply to reheater design. However, the pressure drop in reheaters is critical because the gain in heat rate with the reheat cycle can be nullified by too much pressure loss through the reheater system. Therefore, steam mass fluxes are generally somewhat lower in the reheater.

Tube sizes

Bare cylindrical tubes of 1.75 to 2.75 in. (44.5 to 69.9 mm) outside diameter are typical in current superheaters and reheaters. Steam pressure drop is higher and alignment is more difficult with the smaller diameters, while larger diameters result in higher pressure stresses.

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Recent designs have called for greater spans between supports for horizontal superheater tubes and for wider tube spacing or fewer tubes per row to avoid slag accumulation. The 2.5 in. (63.5 mm) tube has met these new conditions with minimum sacrifice of the smaller tube advantages; 2.75 or 3 in. (69.9 or 76.2 mm) tubes are used to advantage in some cases. When steam temperatures increase, the allowable stresses may force a return to the smaller diameter, thinner-walled tubes.

Bare tubes are used almost exclusively in superheaters. Extended surface on superheater tubes in the form of fins, rings or studs makes gas side cleaning difficult, and the added thickness can increase metal temperature and thermal stress beyond tolerable limits.

Relationships in superheater design

Effective superheater design must consider several parameters including:

1. the steam temperature specified,
2. the range of boiler load over which steam temperature is to be controlled,
3. the superheater surface required to give this steam temperature,
4. the gas temperature zone in which the surface is to be located,
5. the type of steel, alloy or other material best suited for the surface and supports,
6. the rate of steam flow through the tubes (mass flux or velocity), which is limited by the permissible steam pressure drop but which, in turn, exerts a dominant control over tube metal temperatures,
7. the arrangement of surface to meet the characteristics of the anticipated fuels, with particular reference to the spacing of the tubes to prevent accumulations of ash and slag or to provide for easy removal of these formations in their early stages, and
8. the physical design and type of superheater as a structure or component.

A change in any of these items may require a counterbalancing change in some or all of the other items.

The steam temperature desired in advanced power station design is typically the maximum for which the superheater designer can produce an economical component. Economics in this case requires the assessment of two interrelated costs — initial investment and the subsequent cost of upkeep to minimize operating problems, outages and replacements. The steam temperature desired is, therefore, based upon an iterative evaluation of variations in factors 3 through 7, past operating experience and the requirements of the particular project. Operating experience in recent years has resulted in the use of 1000F (538C) or 1050F (566C) steam temperatures for the superheat and reheat in nearly all new large utility boilers.

After the steam temperature is specified, the amount of surface necessary to provide this superheat must be established. This is dependent on parameters 5 through 8. Because there is no single correlation, this quantity must be determined by trial and error, locating the superheater in a zone of gas temperature that satisfies the design criteria. In standard boilers, the zone is fairly well established by unit arrangement and by the space designated for superheater surface.

After the amount of surface is established for the optimum location and tube spacing, the steam mass flux or velocity, steam pressure drop, and superheater tube metal temperatures are calculated considering material options and thickness requirements. The material determination is then made based on economic and functional optimization of the tubes, headers and other components. It may be necessary to compare several arrangements to obtain an optimum combination that:

1. requires an alloy of less cost,
2. gives a more reasonable steam pressure drop without jeopardizing the tube temperatures,
3. gives a higher steam mass flux or velocity to lower tube temperatures,
4. gives the tube spacings that minimize ash accumulations with various types of fuel,
5. permits closer spacing of the tubes, thereby making a more economical arrangement for a favorable fuel supply,
6. gives an arrangement of tubes which reduces the draft loss for an installation where this parameter evaluation is crucial, and
7. permits the superheater surface to be located in a zone of higher gas temperature, with a subsequent saving in surface.

It is possible to achieve a practical design with optimum economic and operational characteristics and with all criteria reasonably satisfied, but a large measure of experience and the application of sound physical principles are required for best results. The calculation methods for superheater performance are given in Chapter 21.

Relationships in reheater design

Superheaters and reheaters are similar in design, but the reheater is limited by the permissible steam pressure drop. Steam mass fluxes or velocities in reheater tubes should be sufficient to keep the difference between the bulk steam and metal surface temperature below 150F (83C). Ordinarily this is done with a pressure drop through the reheater tubes that is 4 to 5% of reheater inlet pressure. This allows another 4 to 5% pressure drop for the reheat piping and valves without exceeding the usual 8 to 10% total allowable pressure loss. The pressure drop allocated for piping is usually distributed with one-third to the cold (inlet) reheat piping and two-thirds to the hot (outlet) reheat piping.

Tube metals

Oxidation resistance, allowable stress and economics determine the materials used for superheater and reheater tubes. The use of carbon steel is maximized. However, carefully selected alloy steels are used where required. Additional information on these materials including maximum metal temperatures are given in Chapter 6.

Variations in basic superheater heat transfer surface arrangements are shown in Fig. 15. These variations in surface arrangement permit the economic tradeoff between material unit costs and differences in the quantity of surface required due to thermal-hydraulic considerations. The main purpose of the counterflow versus parallel flow comparisons is to show basic relationships between heat transfer surface requirements and the corresponding tube metal temperatures and requirements.

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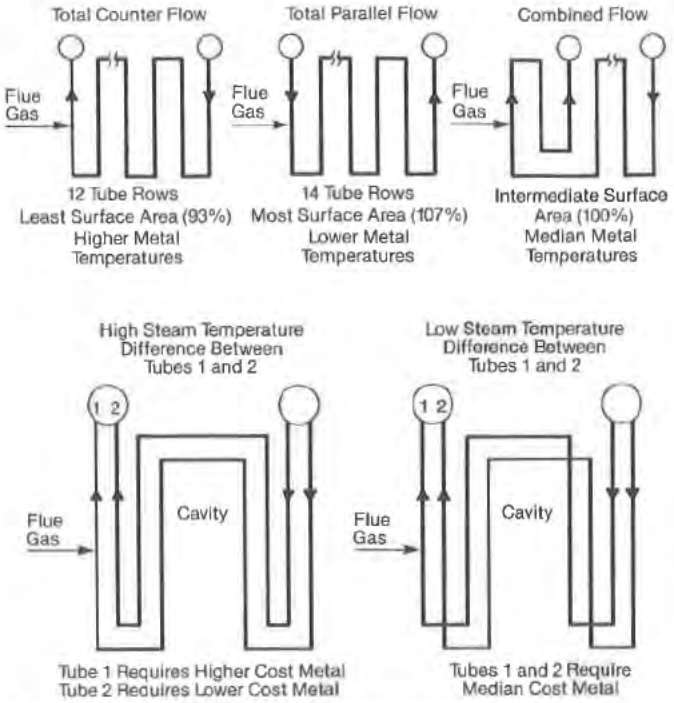


Fig. 15 Typical superheater heat transfer surface arrangements.

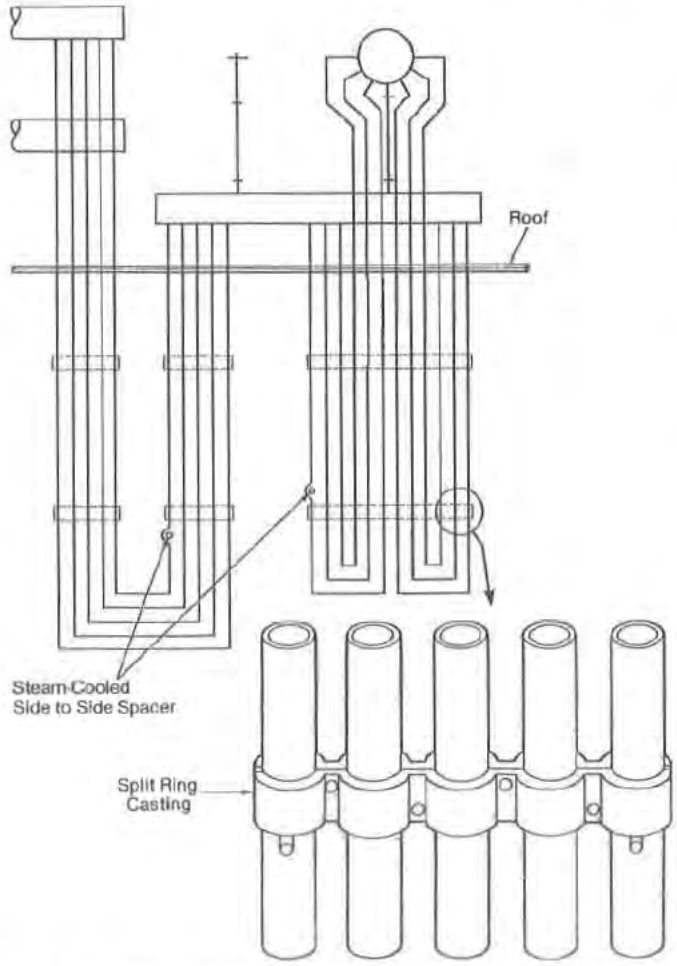


Fig. 16a Pendant superheater section with split ring casting supports.

Supports for superheaters and reheaters

Because superheaters and reheaters are located in zones of relatively high gas temperature, it is preferable to have the major support loads carried by the tubes themselves. For pendant superheaters, the major support points are located outside of the gas stream, with the pendant loops supporting themselves in simple tension. Figs. 16a and 16b illustrate standard support arrangements for a pendant superheater outlet section with major section supports above the roof line. Where adequate side spacing is available and the ash cleaning is not abrasive, steam-cooled wraparound guides are used. Where abrasive ash cleaning is anticipated (coal firing), high chromium-nickel alloy ring-type guides are used. In addition, in the higher gas temperature zones, steam-cooled side to side ties are used to maintain side spacings. For closer side spaced elements, steam-cooled wraparounds are not practical and mechanical ties, such as D-links, are used to maintain alignment as shown in Fig. 17. In this case the clear backspacing between tubes and the size of attachments have been kept to a minimum. This serves to reduce the thermal stresses imposed on the tube wall. Fig. 17 also shows a typical arrangement of a pendant reheater section and illustrates the support of a separated bank by a special loop of reheater, which permits all major supports to be kept above the roof and out of the gas stream.

In horizontal superheaters, the support load is usually transferred to boiler or steam-cooled enclosure tubes or economizer stringer tubes by lugs, one welded to the support tubes and the other to the superheater tubes as indicated in Fig. 18. These lugs are made of carbon steel through high chromium-nickel alloy. Depending upon design considerations, they must slide on one another to provide relative movement between the boiler tubes and the superheater tubes. Saddle-type supports, also shown in Fig. 18, provide for relative movement between adjacent tubes of the superheater.

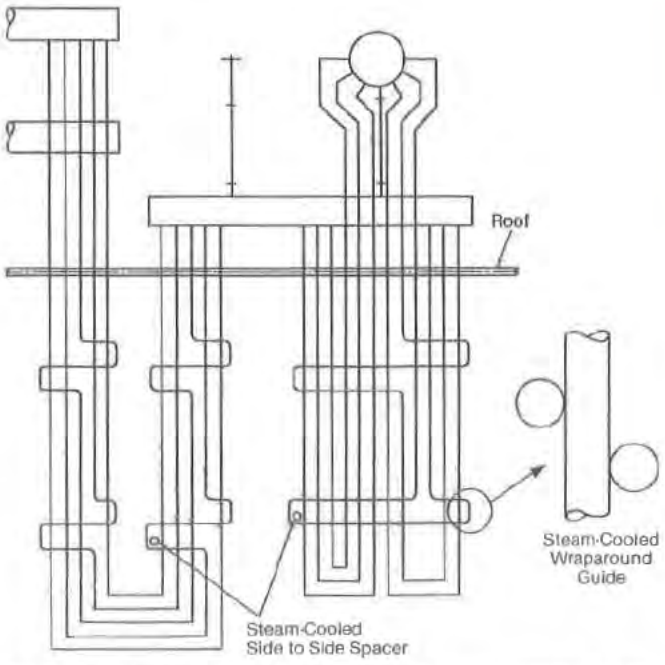


Fig. 16b Pendant superheater section with steam-cooled wraparound guide.

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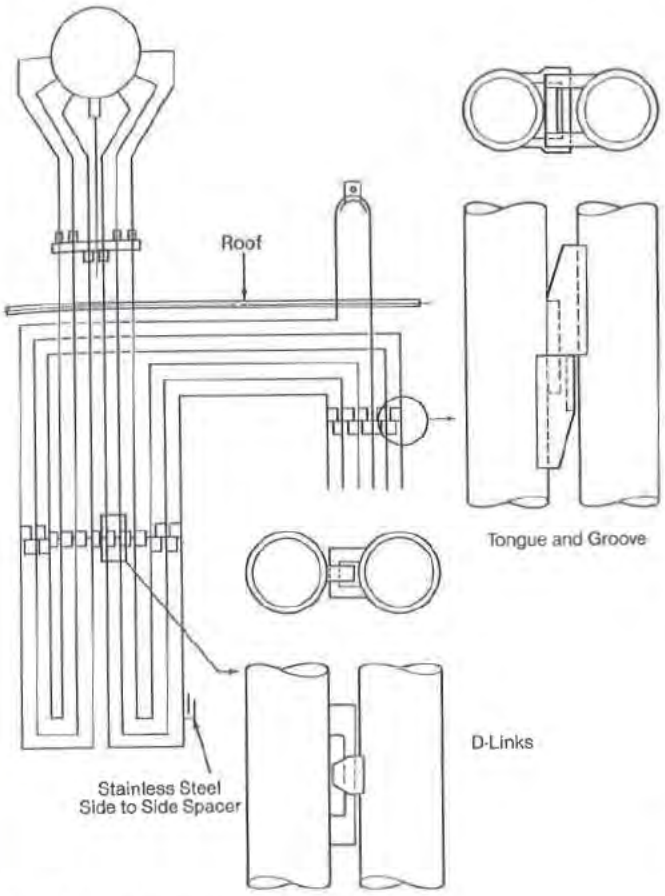


Fig. 17 Pendant reheater section with supports.

As units grow in size, the span of the superheater tubes may become so great that it is impossible to end-support these tubes. Most of the larger units use stringer tubes, generally hung from the economizer outlet, to support the superheater tubes. Tube spacing within the section is maintained by saddle-type supports.

Internal cleaning

The internal cleaning of superheater and/or reheater surfaces is not normally required; however, under certain circumstances these surfaces have been chemically cleaned. This is discussed in depth in Chapter 42. During startup, steam line blowing is used to remove residual scale, oils and residual debris.

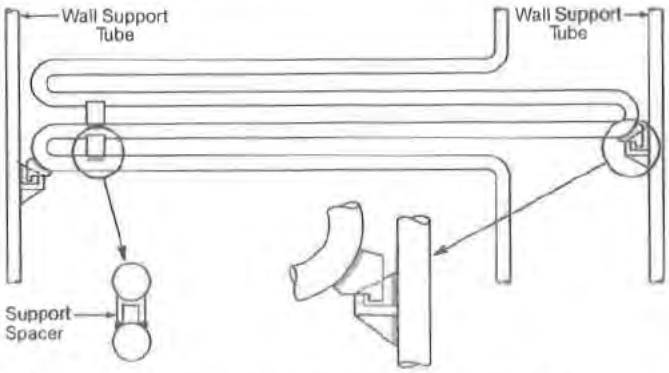


Fig. 18 Horizontal superheater section end supports on walls.

External cleaning and surface spacing

Units purchased today are designed for continuous operation, in some cases for 18 to 24 months between outages, so gas side cleanability is critical. Usually one year between outages is considered acceptable. To enhance cleanability, the superheater sections (platens) of modern utility boilers are spaced according to the gas temperature and the fuel fired. Fig. 10 illustrates the spacing for pulverized coal-fired units. The backspacing in the direction of gas flow is usually set at 0.50 to 0.75 in. (12.7 to 19.1 mm) clear space between tubes in the high temperature zones, with increased spacing allowable in the horizontal surface zones of less than 1500F (816C) entering gas temperature. These spacings are empirical; they are based on tube fouling and erosion experience and on manufacturing requirements. Surface arrangement and cleanability are discussed in Chapter 23 under *Sootblowers*.

Steam temperature adjustment and control

Improvement in the heat rate of modern boiler units and turbines results in large part from the high cycle efficiency possible with high steam temperatures. The importance of regulating steam temperatures within narrow limits is illustrated by the fact that a change of 35 to 40F (19 to 22C) corresponds to a change of about 1% in heat rate at pressures above 1800 psi (124.1 bar).

Other important reasons for accurate steam temperature regulation are to prevent failures due to excessive metal temperatures in the superheater, reheater or turbine; to prevent thermal expansion from dangerously reducing turbine clearances; and to avoid erosion from excessive moisture in the last stages of the turbine.

The control of temperature fluctuations from variables of operation, such as slag or ash accumulation, is important. However, superheater and reheat steam temperatures in steam generation are mainly affected by changes in steam output.

With drum-type boilers, steam output and pressure are maintained constant by firing rate, while the resulting superheat and reheat steam temperatures depend on basic design factors, such as total surface quantity and the ratio of convection to radiant heat absorbing surface. Steam temperatures are also affected by other important operating variables such as excess air, feedwater temperature, changes in fuel that affect burning characteristics and ash deposits on the heating surfaces, and the specific burner combination in service. In the Universal Pressure once-through boiler, which has a variable steam-water transition zone, steam output, pressure and temperature are controlled by the coordination of the firing rate and the boiler feedwater flow rate, leaving reheat steam temperature as a dependent variable. (See Chapter 41.) Standard performance practice for steam generating equipment usually permits a tolerance of ± 10F (6C) in a specific steam outlet temperature.

Definitions of terms

Adjustment is a change in the arrangement of heating surface which affects steam temperature but can not be used to vary steam temperature during operation. This could be the addition or deletion of component surface or boiler surface ahead of the superheater and/or reheater.

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Control is the regulation of steam temperature during operation without changing the arrangement of surface. Examples are operation of attemperators, gas proportioning dampers and gas recirculation fans.

The *attemperator* is an apparatus for reducing and controlling the temperature of a superheated fluid passing through it. This is accomplished by spraying high purity water into an interconnecting steam pipe usually between superheater stages or upstream of a reheater inlet.

Effect of operating variables

Many operating variables affect steam temperatures in drum-type units. To maintain constant steam temperature, means must be provided to compensate for the effect of these variables.

Load As load increases, the quantity and temperature of the combustion gases increase. In a convection superheater (see *Superheater Types* above), steam temperature increases with load, the rate of increase being less the closer the superheater surface is to the furnace. In a radiant superheater (see *Superheater Types* above), steam temperature decreases as load increases. Normally a proportioned combination of radiant and convection superheater surface is installed in series in a steam generating unit to maintain substantially constant steam temperature over the control range of the unit. (See Fig. 14.)

Excess air For a change in the amount of excess air entering the burner zone there is a corresponding change in the quantity of gas flowing over the convection superheater; therefore an increase in excess air generally raises the steam temperature.

Feedwater temperature An increase in feedwater temperature causes a reduction in superheat because, for a given steam flow, less fuel is fired, less gas flows over the superheater and the gas temperatures are lower.

Heating surface cleanliness Removal of ash deposits from heat absorbing surfaces ahead of the superheater reduces the temperature of the gas entering the superheater and, subsequently, the steam temperature. Removal of deposits from the superheater surface increases superheater absorption and raises steam temperature.

Use of saturated steam If saturated steam from the boiler is used for sootblowers or auxiliaries, such as pump or fan drives, an increased firing rate is required to maintain constant main steam output; this raises the steam temperature.

Blowdown The effect of blowdown is similar to the use of saturated steam but is to a lesser degree because of the low enthalpy of water as compared to steam.

Burner operation The distribution of heat input among burners at different locations or a change in burner adjustment usually has an effect on steam temperature due to changes in furnace heat absorption rate.

Fuel Variations in steam temperature may result from changing the type of fuel burned or from day to day changes in the characteristics of a given fuel.

Adjustment

A power generating unit represents a large capital investment, and means should be provided at a reasonable cost for adjusting steam temperature to meet changing conditions. For instance, if a long term fuel change will have a considerable effect on steam temperature, it is good engineering practice to design for compensating physical alterations of the equipment.

Adjustment for regulating steam temperature is required when the operating conditions depart from the conditions on which the design is based. To provide the ultimate adjustment to meet such variations in operation, the design should accommodate anticipated changes at minimum expense.

The basic method of adjustment for regulating steam temperature is the addition or removal of superheater and/or reheater surface. A good design will provide an economical way of doing so.

On certain unit designs, adjustment is also possible by a reduction or increase in the amount of saturated surface ahead of the superheater surface. Such alterations modify the gas temperature at the inlet to these superheaters. If saturated surface is removed to increase steam temperature, this type of adjustment is relatively simple and, in general, costs less than the addition of superheater and/or reheater surface. However, the addition of saturated surface to decrease steam temperature can be difficult and expensive, or even impractical.

Limited use of a refractory coating on selected areas of water-cooled furnace surfaces is also permissible to increase gas temperatures entering the superheater surfaces. This can have a favorable effect on combustion and carbon loss, but refractory should not be added in areas where undesirable ash would deposit. This may also add to maintenance costs.

One of the simplest, least expensive and most effective means of adjustment in regulating steam temperature is to change the mass velocity of the gas flowing over the superheater elements by baffle modifications, if the unit design permits. Several standardized boilers, especially in the smaller sizes, have an adjustable baffle suitable for steam temperature regulation. This feature permits varying the steam temperature control range as much as 20%. The limit of variation is the effect on draft loss and efficiency. A 10F (6C) increase in the temperature of the gas at the boiler outlet reduces efficiency by about 0.25%.

Control

Control is necessary to regulate steam temperature within required limits in order to correct fluctuations caused by operating variables — particularly boiler load. Besides boiler load variation, ash deposition on heat transfer surfaces is the most frequent cause of steam temperature fluctuations. This condition can usually be corrected by changes in the sequence or frequency of sootblower operation. Selective operation of furnace wall blowers or implementation of unit load reductions to induce slag shedding from the furnace walls can reduce gas temperatures entering the superheater surfaces.

The time in which a turbine may be brought to full load is established by its manufacturer in accordance with a safe steam temperature-time curve. Because the temperature of the steam is directly related to the degree of expansion of the turbine elements and consequently the maintenance of safe clearances, this temperature must be regulated within permissible limits by an accurate control device.

The removal of feedwater heaters or pulverizers from service can impact steam temperature and require steam temperature control.

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Among the means of control for regulating steam temperature are: attemperation, gas proportioning dampers, gas recirculation, excess air, burner selection, movable burners, divided furnace with differential firing, and separately-fired superheaters. With attemperation, steam temperature is regulated by diluting high temperature steam with low temperature water or by removing heat from the steam. By comparison, the other methods of control are based on varying the amount of heat absorbed by the steam superheating surfaces.

Attemperation Attemperators may be classified as two types — direct contact and surface. The direct contact design is exemplified by the spray type, where the steam and the cooling medium (water and saturated steam) are mixed. In the surface design, which includes the shell type and the drum type, the steam is isolated from the cooling medium by the heat exchanger surface. Surface-type attemperators are rarely used on current utility boiler designs, while spray attemperators are generally used on all units which have attemperation requirements.

The superheater attemperator may be located in one of two places: at some intermediate point between two sections of the superheater or at the superheater outlet. The ideal location for the superheater attemperator for *process control* would be at the superheater outlet. Control would be direct and there would be no time lag. However, problems with this location include that: 1) water may carry over into the turbine, and 2) the spray does not protect the superheater metals from overheating. The superheater attemperator, located between superheater stages, addresses these concerns and is therefore generally preferred. The steam temperature leaving the superheater does not exceed the maximum temperature desired. In addition, steam from the elements of the first stage superheater is so thoroughly mixed that it enters the second stage superheater at a uniform temperature. For reheat superheater applications, the attemperator is located at the reheater inlet.

The superheater spray attemperator, illustrated in Fig. 19, has proven most satisfactory for regulating steam temperature. High purity water is introduced into the superheated steam line through a spray nozzle at the throat of a venturi section within the line. Because of the spray action at the nozzle and the high velocity of the steam passing through the venturi throat, the water vaporizes, mixes with and cools the superheated steam. An important construction feature is the continuation of the venturi section into a thermal sleeve downstream from the spray nozzle; this protects the high temperature piping from thermal shock. This shock could result from nonevaporated water droplets striking the hot surface of the piping. The reheat attemperator is similar but does not have a venturi section.

The spray attemperator provides a quick acting and sensitive control for regulating steam temperature. It is important that the spray water be of highest purity, because solids entrained in the water enter the steam and may cause troublesome deposits on superheater tubes, piping or turbine blades. High pressure heater drains are a source of extremely pure water but require a separate high pressure corrosion resistant pump if used for attemperator supply. Normally, boiler feedwater is satisfactory, provided condenser leakage and makeup do not introduce too much contamination. The total solids concentration in the spray water should not exceed 2.5 ppm and the spray water should not add more than 40 ppb solids to the steam flow.

Three attemperator arrangements are possible depending on the boiler performance requirements:

1. *Single-stage attemperator* A single attemperator may be installed in each of the connecting pipes between two stages of superheat.
2. *Tandem attemperator* A single-stage attemperator with two spray water nozzles may be installed in series in the connecting pipe between two stages of superheat. This arrangement is used where the spray quantity exceeds the capacity of a single nozzle or where the required turndown can not be achieved with a single spray water nozzle. The usual application requires a spray control valve for each spray nozzle. The operation is sequential with the control valve for the downstream spray nozzle opening first and closing last.
3. *Two-stage attemperator* Two single-stage attemperators are used. The first unit is located in the connecting pipes between the first and second stages of superheat and the second is between the second and third stages of superheat. A spray control valve is required for each stage of spray. The first stage spray attemperator is used first, with the maximum spray flow based on a minimum allowable difference between the temperature of the steam leaving the attemperator and the saturation temperature. The second stage attemperator is used after the flow limit is reached on the first stage unit.

In most instances, the pressure loss from the boiler feed pump through the feedwater heaters, piping and boiler to the attemperator location in the superheater results in sufficient boiler feed pump discharge pressure to provide the required pressure differential in the attemperator system. In certain cases, however, the feed pump discharge pressure is not sufficient due to the low pressure loss in the boiler and feedwater system or due to the high pressure differential the spray systems require to suit particular boiler characteristics. In these cases, a booster stage in the boiler feed pump is desirable to raise the spray water to the required pressure. Other options are a separate spray water booster pump or an additional feed line valve to increase boiler side resistance at the required loads.

Over the past years there has been an increasing frequency of problems in industrial steam power cycles, involving deposits in the superheater and on turbine blading, which has resulted in failures. Many of these problems are due to impurities in the attemperator spray water. To as-

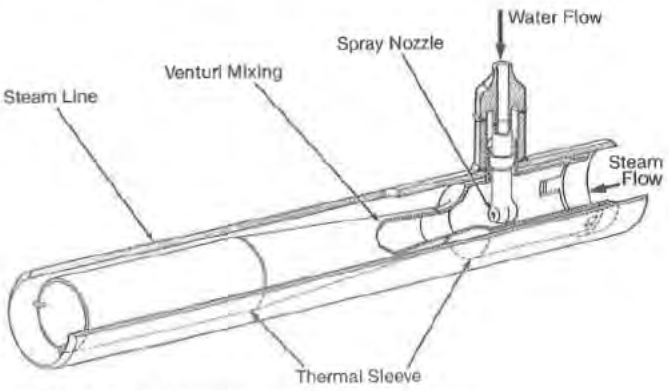


Fig. 19 Spray attemperator showing thermal sleeve.

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sure high quality spray water, additional cleanup equipment can be installed on the condensate return and feedwater makeup.

For low pressure units with feedwater purity less than that required for spray attemperation, the B&W condenser attemperator system is an economical and reliable system used to produce high quality spray water. (See Chapter 25.)

Steam attemperation may be used on drum and once-through boilers as described in the *Bypass and Startup* section of this chapter.

UP boiler attemperator applications UP boilers are supplied without superheater spray attemperators. However, they can be supplied when specified to reduce variations in superheater outlet temperatures during transients. Spray attemperators are installed between the primary superheater outlet and the secondary superheater inlet, upstream of the high pressure superheater stop valve. Spray water is supplied from the UP boiler economizer inlet. Spray attemperation corrects main steam temperature deviations only on a temporary basis and must not be the primary means of steam temperature control. Under steady-state conditions, steam temperature is determined by the ratio of firing rate to feedwater flow. Special rules apply to surface and boiler design for spray attemperators supplied with a UP boiler.

Gas proportioning dampers As shown in Fig. 3, the horizontal convection tube banks in the back end of a boiler can be divided into two or more separate gas passes separated by a baffle wall. The use of dampers in these gas passes then permits proportioning of the gas over the heat transfer surfaces and the control of reheat and superheat temperatures.

Design considerations for such systems include the following:

1. Dampers must be placed in a cool gas zone to assure maximum reliability (typically downstream of all boiler heat transfer surfaces).
2. Draft loss through the unit could increase for some designs, particularly with alternate fuels, so this parameter must be optimized.
3. Control system design and tuning are critical because damper control response is slower than with spray attemperators. Therefore, spray attemperators are used for transient control.
4. Under maximum bias conditions, the gas temperatures at the dampers and the heat transfer surfaces nearest the dampers will be at their highest. These temperatures then set the metal design requirements.

Gas proportioning dampers are combined with spray attemperation for overall optimal steam temperature control systems. Spray attemperators provide for short term transient temperature control. The gas proportioning dampers provide longer term control and adjustment between superheat and reheat temperatures with a minimum impact on overall unit efficiency.

Gas recirculation Another method of controlling superheat or reheat is gas recirculation. As the name implies, gas from the boiler, economizer or air heater outlet is reintroduced to the furnace by fans and flues. For the sake of clarity, recirculated gas introduced in the immediate vicinity of the initial burning zone of the furnace and used for steam temperature control is referred to as *gas recirculation* and recirculated gas introduced near the furnace outlet and used for control of gas temperature is referred to as *gas tempering*. Fig. 20 shows an application of gas recirculation through the hopper bottom and gas tempering in the upper furnace of a Radiant boiler. In most instances the gas is obtained from the economizer outlet. The recirculated gas must be introduced into the furnace in a manner that avoids interference with the fuel combustion. The amount of recirculated gas is expressed as a percentage of the gas that remains downstream of its withdrawal point.

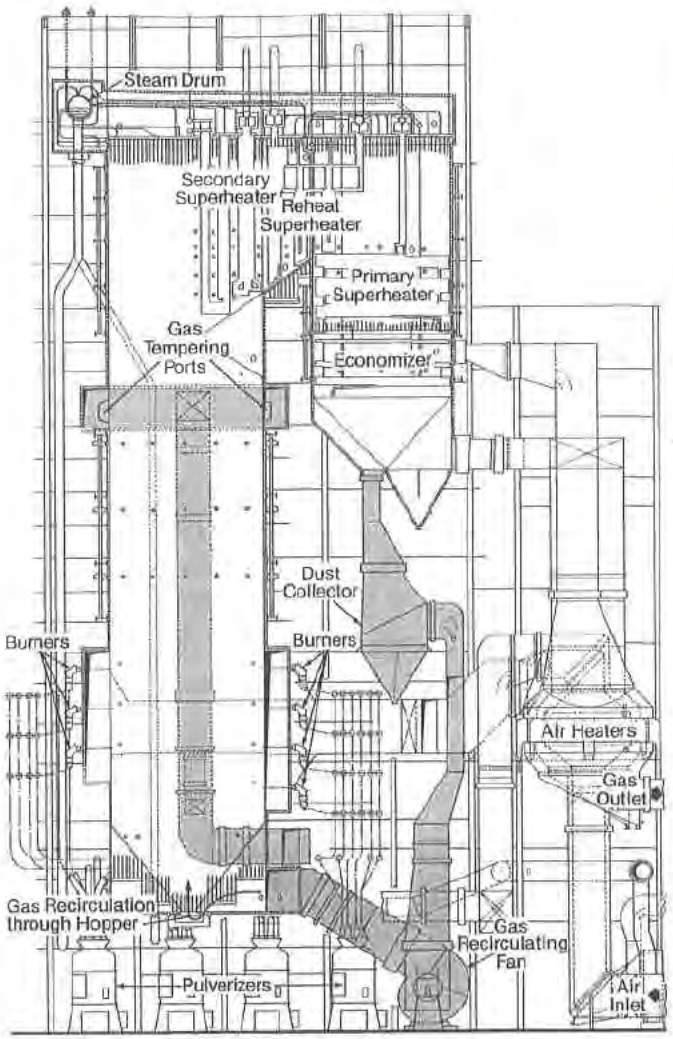


Fig. 20 Radiant boiler with gas tempering for gas temperature control and gas recirculation for control of furnace absorption and reheat temperature.

While recirculated gas may be used for several purposes, its basic function is to alter the heat absorption pattern within a steam generating unit. Recirculated gas has the special advantage of providing heat absorption adjustment that may be used as a design factor in initial surface arrangement and as a method of controlling the heat absorption pattern under varying operating conditions.

An important feature of recirculated gas is that its use changes only the pattern of heat absorption through a boiler; it has a negligible effect on the total boiler heat absorption and the weight of the gas sent up the stack. The thermal effect of recirculated gas depends on the amount of gas recirculated, the location of gas introduction and the furnace heat release rate.

Fig. 21 shows the variation in heat absorption with gas recirculation into the hopper. Introduction of gas at this

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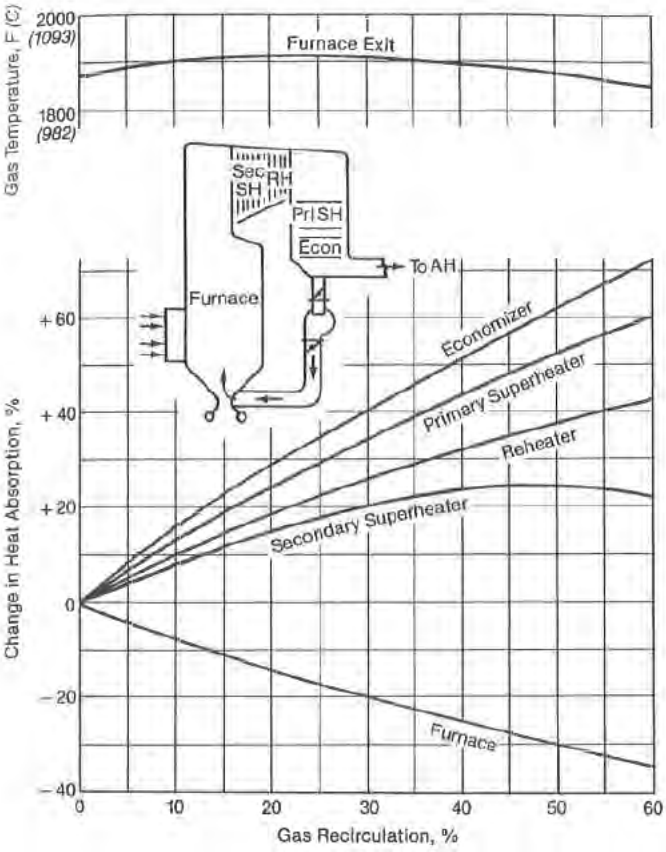


Fig. 21 Effect of gas recirculation on heat absorption pattern at a constant firing rate.

location produces a marked reduction in furnace absorption and increases the absorption of the convection section. Furnace heat absorption is primarily a function of the gas temperatures and gas temperature patterns throughout the furnace, because the heat is mainly transferred by radiation. Therefore, the introduction of gas recirculation into the furnace hopper reduces furnace absorption by altering the gas temperature pattern.

The major portion of the heat absorbed in the superheater, reheater and economizer is transferred by convection, which depends on gas temperature and gas mass flow rate. Both parameters are affected by gas recirculation. Therefore, when the mass velocity of gas flowing through a convection bank is increased by gas recirculation, the amount of heat transferred may increase, decrease or remain unchanged depending on changes in the relationship between the temperature and weight of the gas entering the bank. Fig. 21 illustrates a condition in which the gas temperature entering the secondary superheater (FEGT) is relatively unchanged by gas recirculation. Increasing the amount of gas recirculation, therefore, increases the heat absorption in the secondary superheater. The heat absorption in the reheater, primary superheater and economizer is also increased, with the greatest increase occurring at the cold end of the unit. This is a typical example of the variation in convection pass heat absorption pattern by gas recirculation.

While gas recirculation into the hopper always reduces furnace heat absorption, its effect on the FEGT depends mainly on furnace rating. This exit gas temperature may increase, decrease or, as shown in Fig. 21, be essentially unchanged by gas recirculation. In general, recirculated gas introduced in the hopper decreases the FEGT of a unit operating at high furnace loading and increases this gas temperature at low loading.

Fig. 22 illustrates the effect of introducing tempering gas at a point near the furnace exit. Because the portion of the furnace in which the bulk of heat absorption occurs is unaffected by the recirculated gas, the furnace heat absorption is decreased only slightly. There is, however, a large decrease in FEGT caused by dilution of the hot combustion gases with cooler recirculated gas.

In the case of tempering gas introduced near the furnace outlet, the reduction in FEGT is usually sufficient to overbalance the effect of gas weight increase, and the heat absorption in the secondary superheater is decreased. The effect of gas tempering on the primary superheater and economizer follows the pattern that was shown in Fig. 21, with the greatest change in heat absorption again occurring at the cold end. Because of the location of the reheater in Fig. 22, its absorption remains constant regardless of the percentage of gas tempering.

Figs. 21 and 22 illustrate the effect of introducing recirculated gas into the hopper or at a point adjacent to the furnace exit. Introduction of gas at intermediate points results in heat absorptions and gas temperatures between those shown. To show the effect of recirculated gas only, Figs. 21 and 22 have been based on constant firing rates.

Excess air Boiler operators have long known that the steam outlet temperature of a convection superheater on a drum type or separator type unit can be increased at fractional loads by decreasing the furnace heat absorption through an increase in the amount of excess combustion

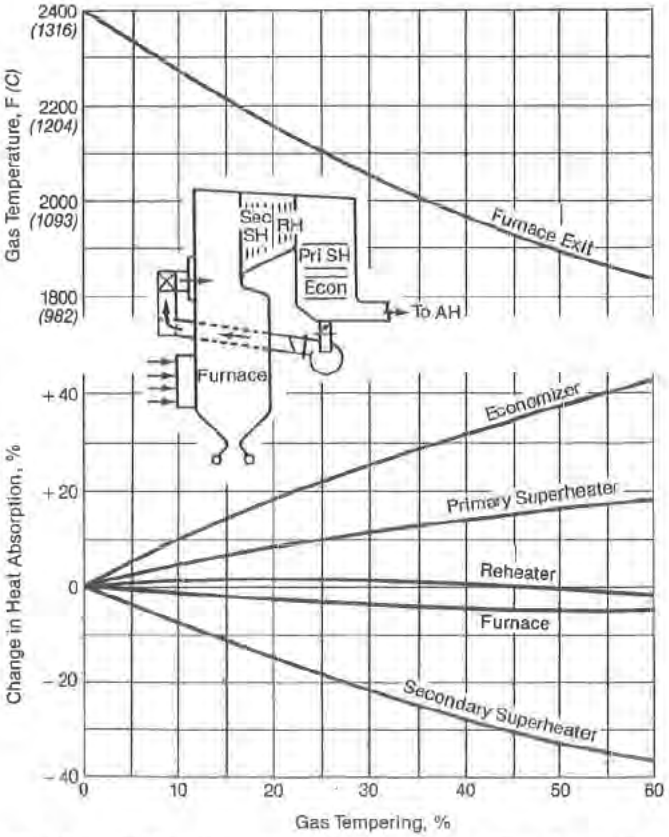


Fig. 22 Effect of gas tempering on heat absorption pattern at a constant firing rate.

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air. The resulting greater weight of gas sent to the stack increases the stack loss. However, the drop in boiler efficiency can be offset by the increase in turbine efficiency.

Burner selection It is often possible to regulate steam temperature by selective burner operation. Higher steam temperatures may be obtained at less than full load by operating only the burners giving the highest furnace outlet temperature. When steam temperature reduction is required, firing may be shifted to the lower burners. This method of control can be improved by distributing the burners over an extended height of the burner wall or by installing a special burner near the furnace outlet.

Movable burners Regulation of steam temperature by changing the furnace absorption pattern can also be effected by using movable burners to raise or lower the main combustion zone in the furnace. Tilting burners are used for this purpose.

Differentially-fired divided furnaces In some divided furnaces, the superheater receives heat from one section of the furnace only, while the other section of the furnace generates only saturated steam or may include a reheater. Steam temperature is regulated by changing the proportion of fuel input between the two furnaces. This arrangement, similar in principle to the separately-fired superheater, was formerly widely used in marine practice. The differentially-fired divided furnace method of steam temperature control is no longer used today.

Separately-fired superheaters A superheater that is completely separate from the steam generating unit and independently fired may serve one or several saturated steam boilers. This arrangement is not generally economical for power generation, where a large quantity of high temperature steam is needed.

Reheat steam temperature

The need for regulating the temperature of reheated steam and the methods of adjustment and control to do so are, in general, the same as for superheated steam. However, the designer does not have the same freedom of action as when only a superheater is used. For instance, in a drum-type unit, the removal of boiler heating surface ahead of the superheater to increase superheat temperature or the removal of superheat surface itself to reduce superheat steam temperature results in increased reheat steam temperature, which may be undesirable. Furthermore, to reduce the gas temperature below slagging limits for convection tube banks and to give the desired steam temperature, such a large proportion of the total input must be absorbed in the furnace and in the superheater and reheater, so there is no boiler surface ahead of the superheater available for adjustment purposes. Some boilers have furnace division walls or water-cooled wingwalls, which may provide adjustment surface if furnace exit gas temperature permits.

Most of the control methods described for regulating steam temperature also affect the reheat temperature if a reheater is incorporated.

Bypass and startup systems

High pressure drum and once-through Universal Pressure boilers must respond to the change in required operating modes of large fossil fueled generating plants. This requires rapid, frequent and reliable unit startups and load changes to meet demand with economical electrical production. During startup or low load (less than 20%) conditions, regular water attemperator steam temperature control systems are usually ineffective as the outlet steam temperature tends to follow the flue gas temperature because the gas flow is substantially higher than the steam flow. However, provision must still be made to match the differing flow, pressure and temperature needs of the steam turbine and boiler. To address these special requirements, three bypass and startup systems have been developed for drum and once-through UP boilers.

Drum boiler bypass system

The B&W Radiant boiler bypass system minimizes startup time, controls shutdowns in anticipation of restarts, provides control of steam temperature to match turbine metal temperature, and allows dual pressure operation of boiler and turbine for better load response. These features reduce stresses in the turbine for improved turbine availability and reduced maintenance costs. The drum bypass system consists of a control system and carefully engineered steam valves and piping as shown in Fig. 23. Probes monitor gas temperatures at the superheater and reheater outlet tubes to permit control of firing rates and gas temperatures during startup.

Operational benefits

Decreased unit startup time The bypass system substantially reduces the time required for a cold startup because it controls the temperature differences of the saturated boiler surface, superheater surface and turbine.

This is accomplished by providing direct control of the steam temperature by mixing saturated steam with superheat and reheat outlet steam as shown in Fig. 23. This arrangement provides the desired steam temperature for the turbine without restricting the startup firing rate of the boiler.

Rapid load changing The system has a set of superheater stop and bypass control valves, which allow dual pressure operation with the throttle pressure controlled separately from drum pressure. This control system permits constant pressure operation of the major boiler components and variable pressure operation of the turbine during load changes. Dual pressure operation minimizes thermal stresses in the boiler and turbine. A dual pressure shutdown keeps the boiler near full pressure and the turbine metal near maximum temperature in preparation for a quick restart. In addition, it allows more rapid load changes than full variable boiler pressure operation.

A superheater bypass diverts excess steam from the boiler to the condenser, thereby separating firing rate from drum pressure control during shutdown and startup. This feature is used during shutdown and next morning restart to keep the turbine and steam piping near full temperature for quick starts and minimum stresses.

Operation The flexible bypass system can conform to most operating changes. For example, it can adopt to a turbine malfunction by adjusting the steam temperature and flow through any load range from synchronization to load points. Although the bypass system enhances unit operation, the boiler can be operated as a conventional nonbypass unit at any time. The following valves (see Fig. 23), with their associated functions, provide the required system flexibility:

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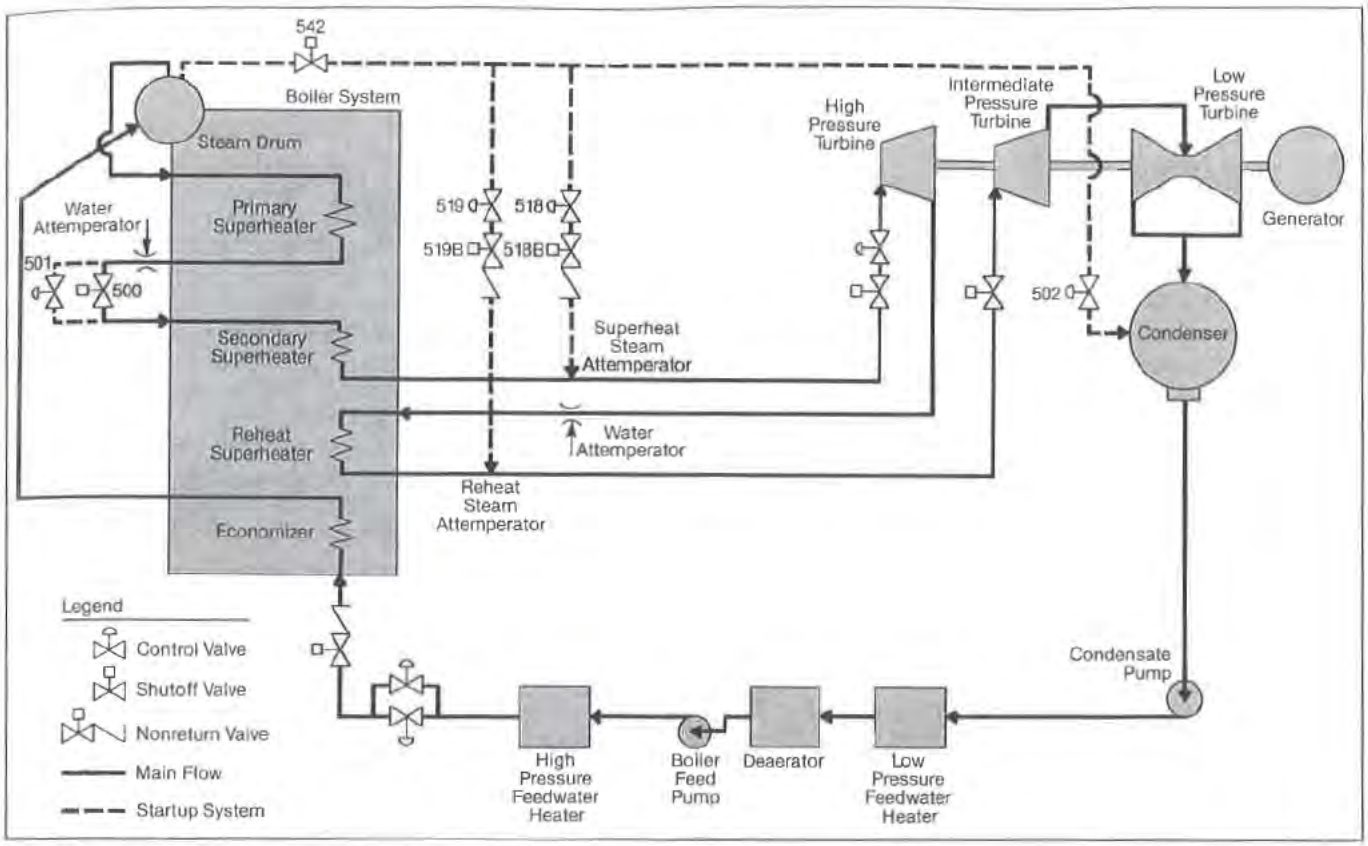


Fig. 23 Drum boiler bypass system schematic.

Valve No.	Function
500	*Secondary superheater stop valve* Separates the secondary superheater from the rest of the steam generator. Closed when using bypass system. Usually designed to open above 70% load.
501	*Secondary superheater stop valve bypass valve* Controls steam pressure leaving the secondary superheater below drum pressure for startup, variable pressure operation and steam attemperation. Design capacity is usually 70% of boiler design flow.
502	*Primary superheater bypass to condenser control valve* Permits firing for control of gas temperature entering the superheater to achieve high steam temperature during startup. Used primarily during hot starts, hot restarts and shutdowns. Closed at loads above 20%.
518	*Main steam outlet attemperator control valve* Reduces main steam temperature to match first-stage turbine metal temperature during cold starts and starts following weekend outages. Closed at loads above approximately 20%. The 518B shutoff valve provides isolation when the 518 valve is closed.
519	*Reheat outlet steam attemperator control valve* Reduces reheat steam temperature to match reheat bowl temperature during cold starts and postweekend starts. Closed at loads above approximately 20%. The 519B shutoff valve provides isolation when the 519 valve is closed.
542	*Primary superheater bypass system shutoff valve* Used to isolate bypass system during normal operation. Closed when the 518, 519 and 502 valves are closed.

Design considerations

Drum pressure and superheater outlet pressure control On restarts following overnight shutdowns, the gas temperature leaving the furnace has to be kept high. This maintains high main steam and reheat temperatures to match high turbine metal temperatures. The turbine must be rolled at a low pressure to prevent a large adiabatic throttling temperature drop when steam is admitted.

The saturation surface can be isolated from the secondary superheater surface by means of the secondary superheater stop valve (500) and bypass valve (501). The boiler can then be fired at the desired rate to raise steam temperatures or drum pressure while maintaining low pressures in the secondary superheater and entering the high pressure turbine. If the drum pressure increases too rapidly or reaches its limit, the primary superheater bypass valve (502) relieves pressure from the drum to the condenser, avoiding condensate waste. A better turbine metal temperature match is obtained with maximum heat input to the boiler and the firing limit dictated by metal protection of the convection pass surface.

Reheat steam temperature control Control of reheat steam temperature is maintained by the steam attemperator valve (519) during startups and shutdowns. Before steam is taken to the turbine, the reheater is without flow. The reheat metal absorbs heat from the flue gas and eventually reaches the flue gas temperature, which can approach

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1000F (538C). When steam is first admitted to the turbine and passes through the reheater, the reheater outlet steam temperature rises very rapidly to the gas temperature level, resulting in a poor match with reheat bowl temperatures in the intermediate pressure turbine for cold starts and starts after a weekend shutdown. Water attemperators are not effective at low loads where reheater steam temperatures are controlled by flue gas temperatures in the reheater banks.

Reheat steam attemperation with saturated steam from the drum limits the rise of reheat outlet steam temperature when steam is first admitted to the turbine and offers positive reheat steam temperature control up to about 20% of full load.

Main steam temperature control The main steam attemperator valve (518) controls main steam temperature. Water attemperators are not effective during low boiler loads, where superheater steam temperatures are determined by the flue gas temperatures in the superheater banks. Therefore, the main steam temperature is controlled by a steam attemperator valve, using saturated steam from the drum to lower temperatures.

Design conditions

Cold start A cold start is defined as a unit startup from no drum pressure or from ambient furnace gas temperature. Turbine metal temperatures are less than 300F (149C) and prewarming is required.

Warm start A warm start is defined as a unit startup after a two day shutdown, such as a weekend shutdown. Turbine metal temperatures are at least 300F (149C). Remaining drum pressure can be as high as 500 psig (34.5 bar gauge).

Hot start A hot start is defined as a unit startup following a six to eight hour outage. The boiler is closed up to retain the maximum internal energy. Boiler drum pressure is quite high and vacuum will have normally been maintained. Turbine metal temperatures of 900 to 925F (482 to 496C) may exist following a controlled shutdown and a decay in turbine metal temperatures of 100F (56C) may occur during the shutdown. Therefore, minimum steam temperatures of 800 to 825F (427 to 441C) are needed during the restart. The steam temperature is usually less than the temperature limit for gas entering the superheater during a no-flow condition.

Hot restart If the unit is tripped but ready for a restart very quickly, turbine metal temperatures may be 1000F (538C) and drum pressure may be near the pressure existing at the time of the trip. To achieve the high gas temperatures needed, it may be necessary to continue unit firing and dump excess steam to the condenser.

Controlled shutdown When shutting the unit down, the bypass system may be used to facilitate the operation that follows the shutdown. If maintenance on the unit is scheduled, the turbine metal temperature and boiler pressure should be kept as low as possible while still carrying load. The throttle pressure is 501 valve controlled and necessary sprays to reduce steam and metal temperatures per the cooldown curves are controlled by the 518 and 519 valves. After the unit is tripped, the fans may remain on to cool the boiler.

Variable pressure operation Under most circumstances, the unit heat rate can be increased at partial loads if boiler pressure is reduced with load. This mode of operation is normally referred to as *variable pressure operation*.

The secondary superheater stop valve (500) and the stop valve bypass valve (501), used with the B&W bypass system, permit operating with the drum and primary superheater pressures at constant levels while the secondary superheater outlet pressure is varied with load. Operating in this mode maintains turbine steam temperatures at the design level over a greater load range than is possible under constant throttle pressure operation.

Maintaining drum and primary superheater pressure relatively constant at reduced loads permits rapid load pickup. The savings in pump power, which could be achieved by permitting drum pressure to vary with load, are eliminated when the rapid load pickup feature is used.

Startup without bypass system The unit can be started and operated without use of the bypass system. Fully opening the 500 valve allows starting up the unit like a conventional drum boiler.

UP boiler startup system

A key requirement of UP startup and bypass systems is the need for minimum design circulation flows in high heat absorbing circuits for cooling before the unit can be fired. Additional important features include providing a turbine bypass until steam pressure and temperatures are matched, reducing the bypass flow pressure and temperature before condenser and auxiliary equipment steam admission, recovering heat during startup, providing clean water for full startup, accelerating the startup processes, and providing greater unit flexibility through dual pressure boiler operation. UP boiler plants, while originally designed for base load operation, have been adapted to load cycling, including daily operation on the bypass system.

Two different UP systems are available — one for constant furnace pressure operation over the load range and a second for variable furnace pressure operation over the load range.

Constant furnace pressure startup system For constant furnace pressure the startup system has the steam separator (*flash tank*) located in a bypass that can be isolated from the boiler during normal operation. The general arrangement is shown in Fig. 24.

The boiler feed pump supplies the minimum required flow of feedwater during startup and low load operation to protect the furnace circuitry. Included in the startup circuitry are the economizer, furnace convection pass enclosure and primary superheater.

The fluid leaving the primary superheater (at full pressure) bypasses the secondary superheater through the pressure reducing valve (207) to the flash tank, where the steam-water mixture is separated during startup.

Water level in the flash tank is controlled by drain valves 230 and 241, with the 230 valve controlling the flow to the deaerator for maximum heat recovery. Excess water (above the capability of the deaerator) is discharged to the condenser through valve 241. If the drains are not within water quality limits (see Chapter 42), all of the flow is through valve 241 to the condenser and polishing system.

The 242 block valve remains closed until a level is established in the flash tank to assure that water does not enter the steam lines. Once a level is established, the 242 valve is opened so the deaerator steam line from the flash tank can be used to hold pressure in the deaerator (controlled by the 231 valve). This permits returning all of the

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drains to the condenser through the 241 valve during a hot cleanup without using an auxiliary steam source for maintaining deaerator pressure, and also serves to recover the heat in the flash tank steam during cleanup.

The low pressure steam line, the low pressure superheater nonreturn valve (205), and the downstream high pressure steam line connect the flash tank to the secondary superheater inlet downstream of the high pressure superheater block valve (200) and stop/control valve (401). After a water level is established in the flash tank, 205 valves normally open at 300 psig (20.7 bar gauge) and dry steam flows to the secondary superheater for warming steam lines. The turbine bypass valve (210) normally opens at 300 psig (20.7 bar gauge) main steam pressure to assist with warming and boiling out the superheater during the initial stage of startup. When sufficient steam is available, the turbine is rolled and placed on line.

Steam separated in the flash tank in excess of that required is relieved through the 240 valve to the condenser. This valve also acts as an overpressure relief valve to avoid tripping spring loaded safety valves on the flash tank. The 240 valve has an adjustable set point, which can be set to hold various flash tank pressures at particular load points during startup.

The entire bypass system is sized to handle minimum required flow during startup and to permit operating at minimum load on the flash tank.

The transition from operation on the flash tank to once-through flow is made at minimum load. As the steam entering and leaving the flash tank at this time is dry and superheated, the transition from flow through the bypass to once-through flow is accomplished, with minimum fluctuation in steam temperature, by opening the 200 block valve and the 401 combination stop/control valve and by closing the 207 and 205 bypass valves.

Variable throttle pressure Above minimum load the unit can be operated at constant or variable throttle pressure, with the 401 valve controlling the throttle pressure from minimum to 100% load, while maintaining full pressure in the upstream circuits.

The variable throttle pressure feature permits operating the unit with the throttle valves essentially wide open. This eliminates turbine metal temperature changes resulting from valve throttling and permits rapid load changes without being limited by turbine heating or cooling rates. Shutdown with variable throttle pressure maintains high temperatures in the turbine metals and permits rapid hot restarting.

Steam temperature control The means for controlling main steam and reheat temperatures at normal operating loads are not effective during startup or at very low loads.

The startup system shown in Fig. 24 includes provision for steam attemperation from the flash tank to the main and reheat steam outlet headers for precise control of steam conditions during startup to meet the turbine metal temperature requirements.

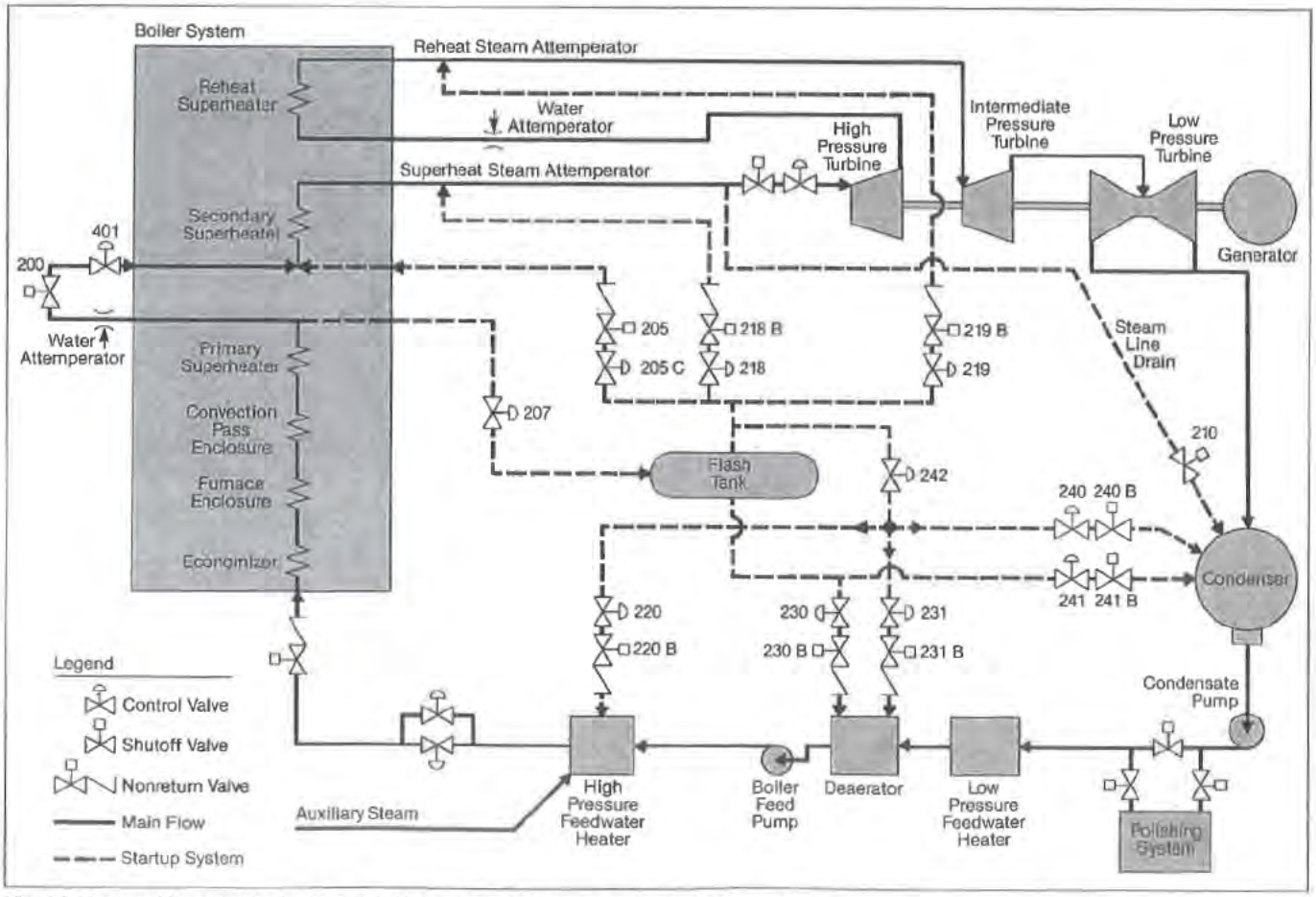


Fig. 24 Universal Pressure boiler startup system — constant pressure furnace operation.

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The superheater outlet steam attemperator valve (218) is used at loads less than 20% to introduce saturated steam from the flash tank to the superheater outlet header. Initial rolling of the turbine, for a cold start, may be done with saturated steam passing from the flash tank through the 218 valve. This steam may be mixed with a limited quantity of steam passing through the 205 valves and the secondary superheater to control the high pressure turbine inlet temperature down to about 550F (288C). The 205C control valve achieves the necessary pressure drop between the flash tank and the secondary outlet header for attemperation.

The reheat outlet steam attemperator valve (219) is used at loads below about 20% of full load to introduce flash tank steam to the reheat outlet header. The ratio of flow through the attemperator valve to that through the high pressure turbine is limited mainly for turbine and turbine control considerations.

Overpressure relief The bypass system is also used to relieve excessive pressure in the boiler during a load trip. This is accomplished by the use of the 207 valve, which sends excess steam to the flash tank.

Variable furnace pressure startup system For units capable of operating with variable furnace pressure over the load range, the startup system has the steam separator located in the main flow path upstream of the primary superheater. The general arrangement is shown in Fig. 25.

The boiler feed pump supplies the minimum required flow of feedwater during startup and low load operation to protect the furnace circuits. Included in the startup circuitry are the economizer, the furnace enclosure, the wingwall pass (on units where the wingwalls are upstream of the primary superheater), and the steam separator.

The steam-water mixture is separated in the vertical steam separator during startup and low load operation. The unit is started with low furnace pressures to obtain the maximum amount of steam early in the startup; the 330 and 341 valves control the drains from the separator. The 330 valve also controls the flow to the deaerator for maximum heat recovery.

The 342 block valve remains closed until there is a steam-water mixture entering the separator, and it is closed any time a high level is indicated in the separator to prevent water from entering the superheater and reheater through the steam attemperators. Once the 342 valve is opened, the deaerator steam line holds pressure in the deaerator (controlled by the 331 valve). This permits returning all of the drains to the condenser through the 341 valve during a hot cleanup without using an auxiliary steam source for maintaining deaerator pressure; it also serves to recover the heat in the separated steam during cleanup.

During the initial stage of startup, the steam flow from the superheater is through the main steam line drains and 310 valve to warm the steam lines.

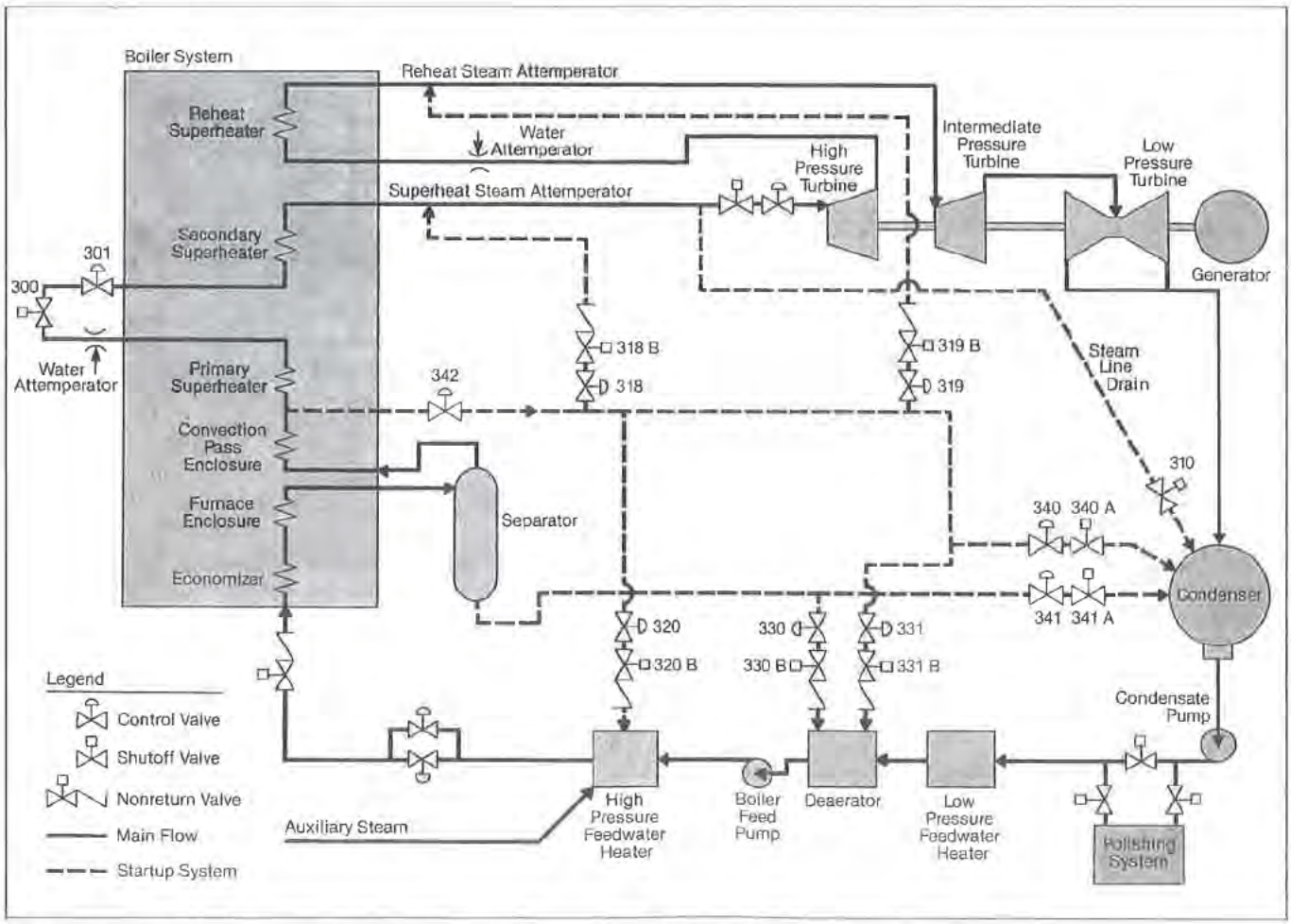


Fig. 25 Universal Pressure boiler startup system -- variable pressure furnace operation.

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As the enthalpy of the fluid entering the separator increases, the drains diminish until there is dry steam entering the separator. The drain valves are closed at this time, the deaerator is being controlled with turbine extraction steam, and the unit is on once-through operation with all the flow going to the turbine.

Division valves (dual pressure capability) On hot starts, including starts following overnight shutdowns, the gas temperature leaving the furnace has to be kept high to maintain high main steam and reheat temperatures. This generally results in an excessively rapid rise in throttle pressure, which is undesirable because of the resulting large throttling temperature drop when admitting steam to the turbine.

By means of the superheater combination stop and control valve (301), the boiler surface can be isolated from the secondary superheater surface. The overfiring required to raise and maintain steam temperatures can be allowed to raise saturation temperature or boiler pressure while maintaining the desirable low pressures in the secondary superheater and entering the high pressure turbine. Dual pressure operation also permits more rapid rates of load change. As the boiler temperature and pressure can be maintained at the higher level, the unit also responds more quickly to a change in load demand.

Superheater bypass to condenser When reaching the maximum desired boiler pressure or when starting up with the superheater stop valve open, the superheater bypass to the condenser valve (340) provides a means to control boiler pressure during hot start conditions. During the transient loading period for the unit following a hot restart, the superheater bypass valve (340) may be opened to permit higher firing rates and therefore sustain raised steam temperature until the boiler control load is reached.

The superheater bypass to the condenser is used to relieve excessive pressure in the boiler during a load trip. It can also be used as an overpressure relief valve to supplement and avoid popping spring loaded safety valves.

Steam temperature control The startup system shown in Fig. 25 includes provision for attemperation with saturated steam. This steam is taken from downstream of the separator and injected into the main and reheat steam outlet headers for precise control of steam conditions during startup in order to meet the turbine metal temperature requirements.

The superheater outlet steam attemperator valve (318) is used in a manner similar to the 218 valve discussed above under *Constant Furnace Pressure, Steam Temperature Control.*

The division valve (301) is used to achieve the necessary pressure drop between the separator and the secondary outlet header for attemperation.

The reheat outlet steam attemperator valve (319) is used at loads below 20% of full load to introduce separator steam to the reheat outlet header. The ratio of flow through the attemperator valve (319) to the flow through the high pressure turbine is limited due to turbine and turbine control considerations.

Permissible rates of load change

The permissible rates of load change, based on the allowable rates of temperature change in the boiler components for various modes of operation, are shown in Fig. 26.

The longer time required for load change with variable pressure operation is due to the greater change in furnace enclosure temperature and the need to restrict the rate of temperature change to avoid tearing casing attachments from the enclosure walls. With dual pressure operation, the furnace enclosure remains at constant pressure and practically constant temperature; however, the secondary superheater inlet header goes through a large temperature change with variable pressure operation, and the time required for load change is based on the limiting rate of temperature change for the outlet header.

If the entire unit is operated at constant pressure, there is virtually no boiler imposed limit on the rate of load change except for the rate of change possible from the firing equipment. With a large load change the expected rate would be a maximum of about 5% of full load per minute. The rate of load change is also restricted by the turbine metals which go through a large temperature swing due to the throttling at low loads and are limited by the design number of cycles.

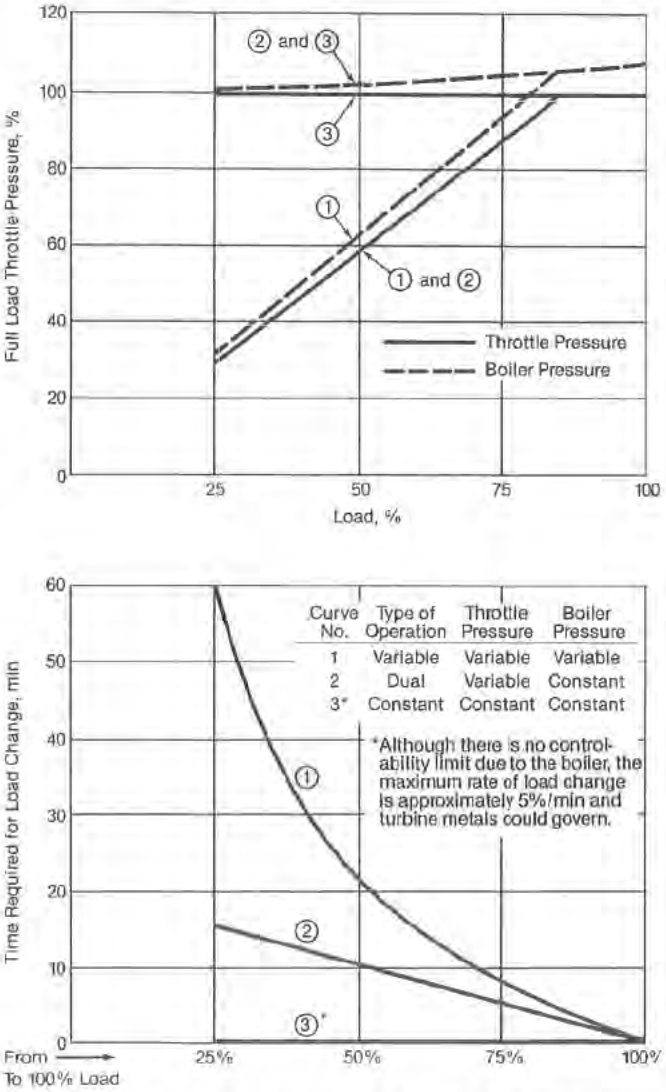
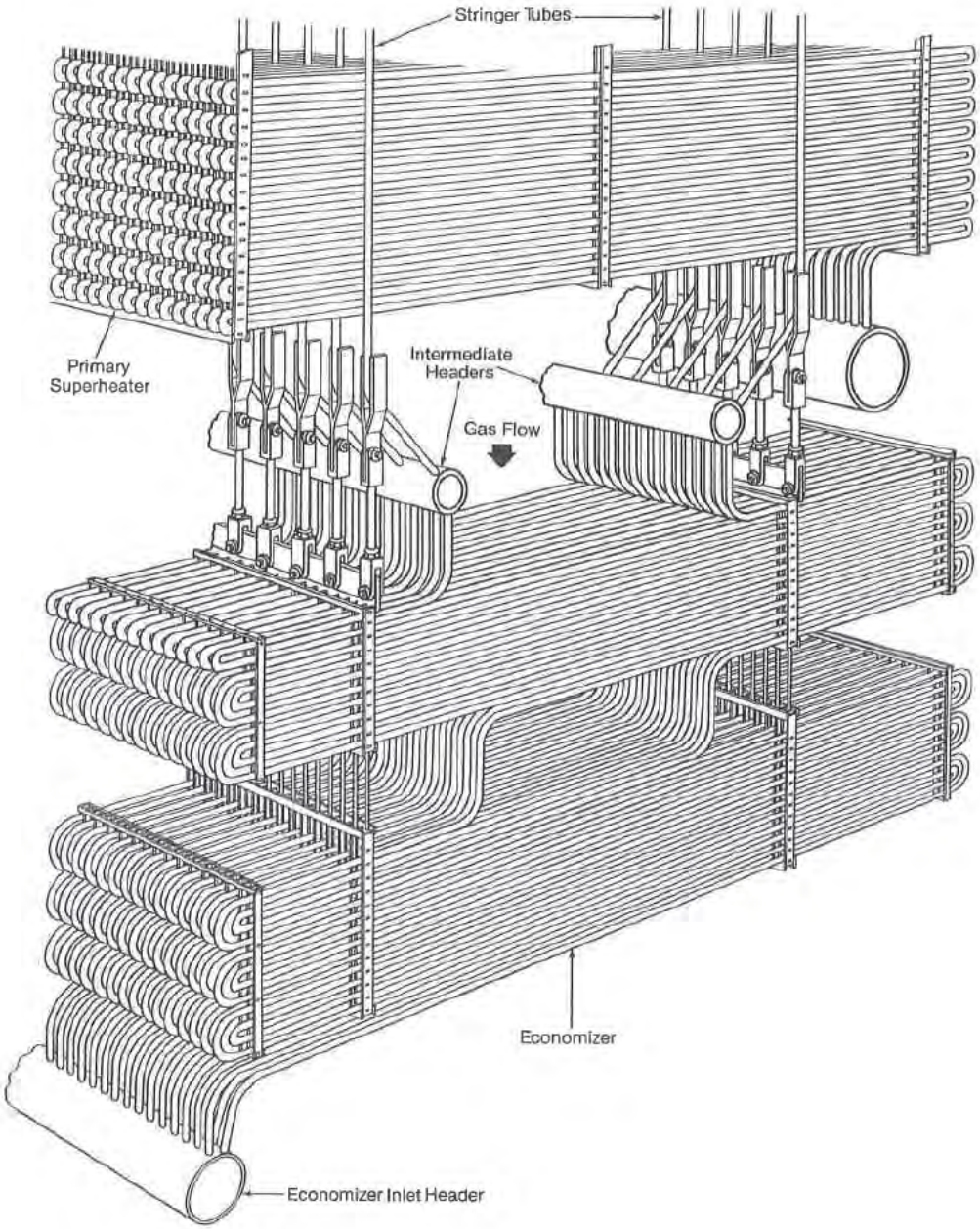


Fig. 26 Permissible rates of load change for three boiler operating configurations.

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Typical utility boiler economizer.

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Chapter 19
Economizers and Air Heaters

Economizers and air heaters perform a key function in providing high overall boiler thermal efficiency by recovering the low level, i.e., low temperature, energy from the flue gas before it is exhausted to the atmosphere. For each 40F (22C) that the flue gas is cooled by an economizer or air heater, the overall boiler efficiency increases by approximately 1% (Fig. 1). Economizers recover the energy by heating the boiler feedwater while air heaters heat the combustion air. Air heating also enhances the combustion of many fuels and is critical for pulverized coal firing for drying the coal and ensuring stable ignition.

In comparison to the furnace waterwalls, superheater and reheater, economizers and air heaters require a large amount of heat transfer surface per unit of heat recovered. This is because of the relatively small difference between the temperature of the flue gas and the temperature of either the feedwater or the combustion air. Use and arrangement of the economizer and/or air heater depend upon the particular fuel, application, boiler operating pressure, power cycle and overall minimum cost configuration.

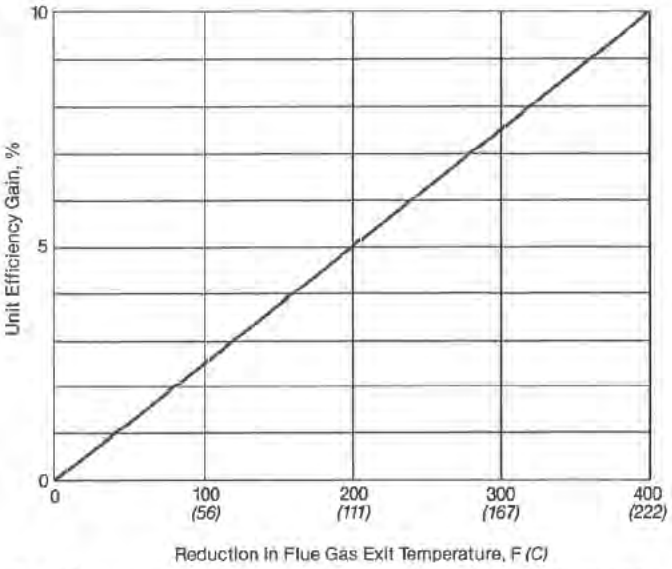


Fig. 1 Approximate unit efficiency increase due to an economizer and air heater.

Economizers

Economizers are basically tubular heat transfer surfaces used to preheat boiler feedwater before it enters the drum (recirculating units) or furnace surfaces (once-through units). The term economizer comes from early use of such heat exchangers to reduce operating costs or economize on fuel by recovering extra energy from the flue gas. Economizers also reduce the potential of thermal shock and strong water temperature fluctuations as the feedwater enters the boiler drum or waterwalls. Fig. 2 shows an economizer location on a coal-fired boiler. The economizer is typically the last water-cooled heat transfer surface upstream of the air heater. (See facing page.)

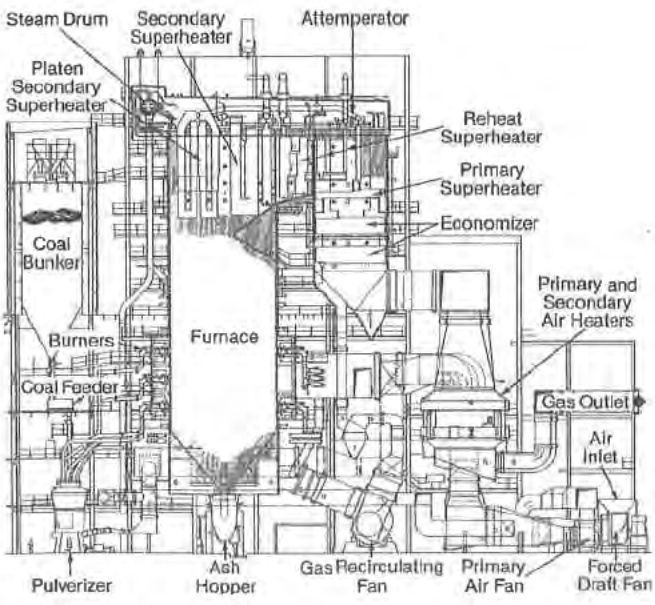


Fig. 2 Economizer and air heater locations in a typical coal-fired boiler.

Economizer surface types

Bare tube

The most common and reliable economizer design is the bare tube, in-line, crossflow type. (See Fig. 3a.) When coal is fired, the flyash creates a high fouling and erosive environment. The bare tube, in-line arrangement minimizes the likelihood of erosion and trapping the ash as compared to a staggered arrangement shown in Fig. 3b.

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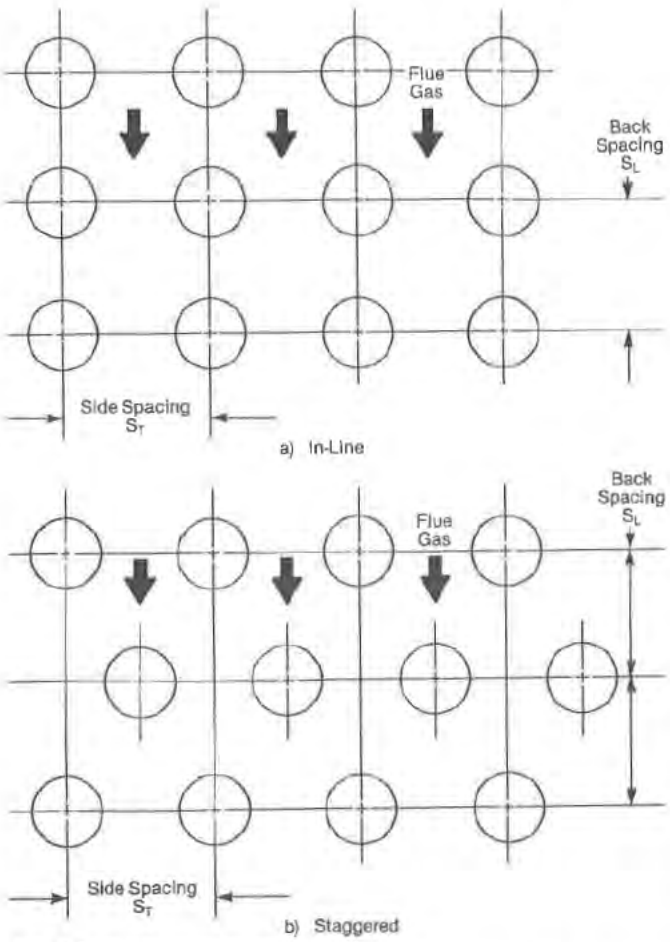


Fig. 3 Bare tube economizer arrangements.

It is also the easiest geometry to be kept clean by sootblowers. However, these benefits must be evaluated against the possible larger weight, volume and cost of this arrangement.

Extended surfaces

To reduce capital costs, most boiler manufacturers have built economizers with a variety of fin types to enhance the controlling gas side heat transfer rate. Fins are inexpensive nonpressure parts which can reduce the overall size and cost of an economizer. However, successful application is very sensitive to the flue gas environment. Surface cleanability is a key concern. In selected boilers, such as Cyclone furnace units (see Chapter 14), extended surface economizers are not recommended because of the coarser flyash characteristics.

Stud fins Stud fins have worked reasonably well in gas-fired boilers. However, stud finned economizers can have higher pressure loss than a comparable unit with helically-finned tubes. Studded fins have performed poorly in coal-fired boilers because of high erosion, loss of heat transfer, increased pressure loss and plugging resulting from flyash deposits.

Longitudinal fins Longitudinally-finned tubes in staggered crossflow arrangements, shown in Fig. 4, have also not performed well over long operating periods. Excessive plugging and erosion in coal-fired boilers have resulted in the replacement of many of these economizers. In oil- and gas-fired boilers, cracks have occurred at the

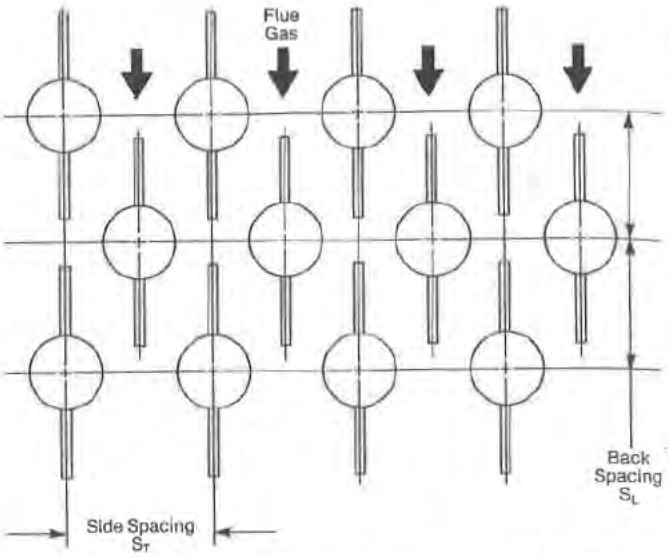


Fig. 4 Longitudinal fins, staggered tube arrangement. (Fin width exaggerated for clarity.)

points where the fins terminate. These cracks have propagated into the tube wall and caused tube failures in some applications.

Helical fins Helically-finned tubes (Fig. 5) have been successfully applied to some coal-, oil- and gas-fired units. The fins can be tightly spaced in the case of gas firing due to the absence of coal flyash or oil ash. Four fins per inch (1 fin per 6.4 mm), a fin thickness of 0.06 to 0.075 in. (1.5 to 1.9 mm) and a height of 0.75 in. (19.1 mm) are typical. For 2 in. (51 mm) outside diameter tubes, these fins provide ten times the effective area of bare tubes per unit tube length. If heavy fuel oil or coal is fired, a wider fin spacing must be used and adequate measures taken to keep the heating surface as clean as possible. Economizers in units fired with heavy fuel oil can be designed with helical fins, spaced at 0.5 in. (13 mm) intervals. Smaller fin spacings promote plugging with oil ash, while greater spacings reduce the amount of heating surface per unit length. Sootblowers are required and the maximum bank height should not exceed 4 to 5 ft (1.2 to 1.5 m) to assure reasonable cleanability of the heating surface. An in-line arrangement also facilitates cleaning and provides a lower gas side resistance.

Rectangular fins The square or rectangular fins, arranged perpendicular to the tube axis on in-line tubes as shown in Fig. 6, have had some success in retrofits. The fin spacing typically varies between 0.5 and 1 in. (13 and 25 mm) and the fins are usually 0.125 in. (3.18 mm) thick. There is a vertical slot down the middle because the two halves of the fin are welded to either side of the tube. Most designs are for gas velocities below 50 ft/s (15.2 m/s).

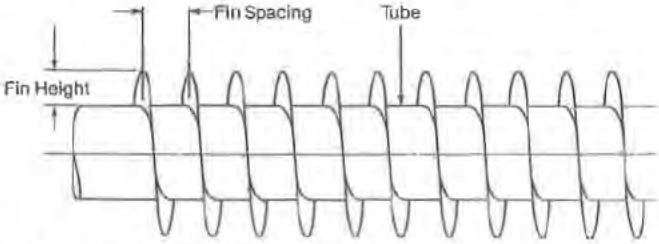


Fig. 5 Helically-finned tube.

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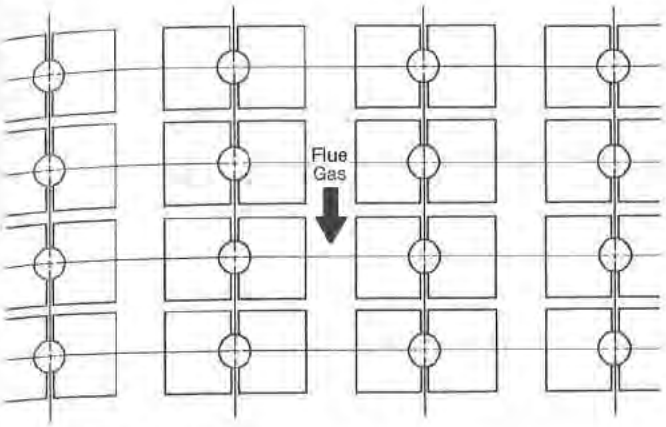


Fig. 6 Rectangular fins, in-line tube arrangement.

Baffles The tube ends should be fully baffled (Fig. 7) to minimize flue gas bypass in finned bundles. Such bypass flow can reduce heat transfer, produce excessive casing temperatures and with coal firing, can lead to tube bend erosion because of very high gas velocities. Baffling is also used with bare tube bundles but is not as important as for finned tube bundles.

Velocity limits

The ultimate goal of economizer design is to achieve the necessary heat transfer at minimum cost. A key design criterion for economizers is the maximum allowable gas velocity (defined at the minimum cross-sectional free flow area in the tube bundle). Higher velocities provide better heat transfer and reduce capital cost. For clean burning fuels, such as gas and low ash oil, velocities are typically set by the maximum economical pressure loss. For high ash oil and coal, gas side velocities are limited by the erosion potential of the flyash. This erosion potential is primarily determined by the percentage of Al_2O_3 and SiO_2 in the ash, the total ash in the fuel and the gas maximum velocity. Experience dictates acceptable velocities. Fig. 8 provides sample base velocity limits as a function of ash characteristics.

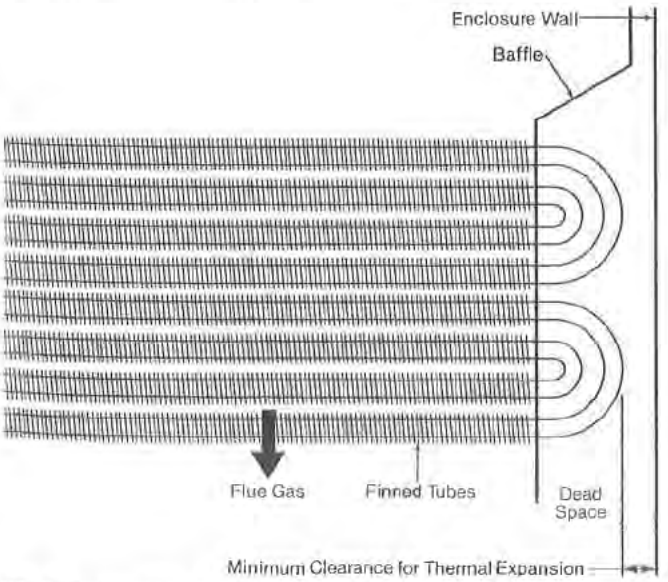


Fig. 7 Baffling of bare tube return bends for finned tube bundles.

Further criteria may also be needed. For example, a 5 ft/s (1.5 m/s) reduction in the base velocity limit is recommended when firing coals with less than 20% volatile matter. In other cases, such as Cyclone boilers, high flue gas velocities can be used because much less flyash is carried into the convection pass, as much of the ash (> 50%) is collected in the bottom of the boiler as slag. Particles that carry over are also less erosive. (See Chapter 14.)

For a given tube arrangement and boiler load, the gas velocity depends on the specific volume of flue gas which falls as the flue gas is cooled in the economizer. To maintain the gas velocity, it can be economical to decrease the free flow gas side cross-section by selecting a larger tube size in the lower bank of a multiple bank design. This achieves better heat transfer and reduces the total heating surface.

Other types of economizers

Fig. 9 depicts an industrial boiler with a long flow economizer, often used in chemical recovery boilers. Such heating surfaces consist of vertical, longitudinally-finned (membraned) tubes through which the feedwater flows upward. The gas flows downward in pure counterflow, outside the tubes and fins. While the heat transfer is less efficient than crossflow banks of tubes, there is minimal gas side resistance and fouling products are removed through hoppers at the bottom of the enclosure.

Steaming economizers

Steaming economizers are defined as meeting the following enthalpy relationships:

$$H_2 - H_1 \geq {}^2/_3 \left(H_f - H_1 \right) \qquad (1)$$

where

H_2 = enthalpy of fluid leaving economizer (to drum)
H_1 = enthalpy of fluid (water) entering economizer
H_f = enthalpy of saturated water at economizer outlet pressure

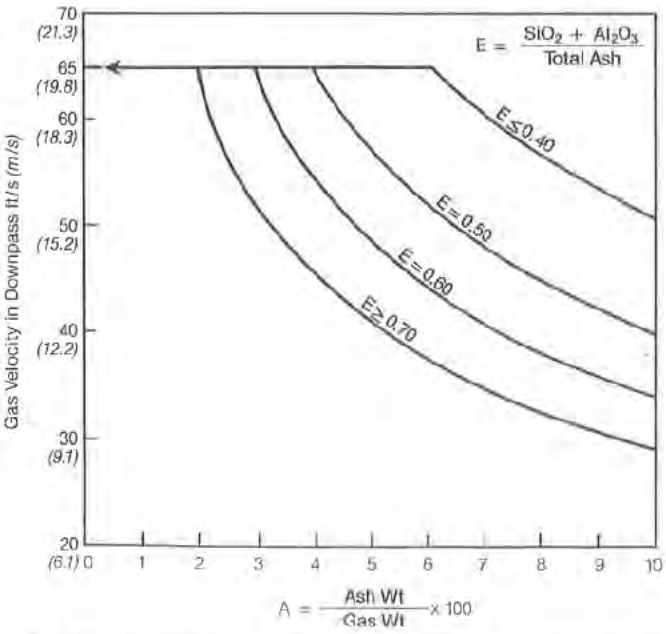


Fig. 8 Base maximum allowable velocity for pulverized coal-fired boiler economizers.

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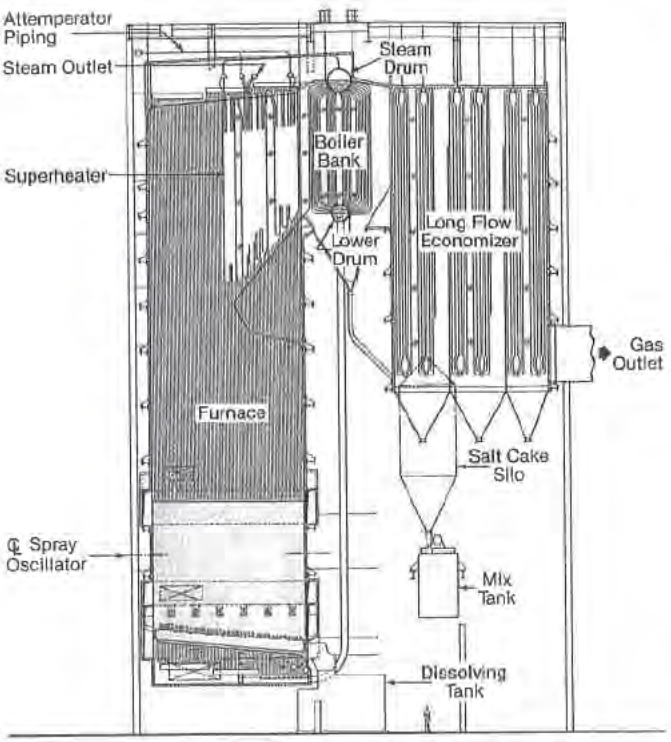


Fig. 9 Long flow economizer for a chemical recovery boiler.

These economizers can be economical in certain boilers. They require careful design and must be oriented so the water flows upward and the outlet is below the level of the drum. This avoids water hammer and excessive flow instabilities. The enthalpy equation accounts for possible steaming due to flow imbalances and differences in individual circuit heat absorptions.

High pressure, drum type units are usually sensitive to feedwater temperatures close to saturation. To enhance circulation, feedwater temperature to the drum should normally be at least 50F (28C) below saturation temperature.

Performance

Heat transfer

Bare tubes The equations discussed in Chapter 4 can be used to evaluate the quantity of surface for an economizer. For the economizer shown in Fig. 2 with the upflow of water and downflow of gas and nonsteaming conditions, the bundle can be treated as an ideal counterflow heat exchanger with the following characteristics:

1. bundle log mean temperature difference correction factor = 1.0,
2. heat absorbed by the tube wall enclosures and heat radiated into the tube banks from various cavities can generally be neglected,
3. all of the energy lost by the flue gas is absorbed by the water, i.e., no casing heat loss,
4. the water side heat transfer coefficient is typically in the range of 2000 Btu/h ft² F (11,357 W/m² K) and has only a small overall impact on the economizer performance, and
5. the effect of gas side ash deposition can be accounted for by a cleanliness factor based upon experience.

In general the heat transfer rate is primarily limited by the gas side heat transfer for in-line bare tube bundles. In this case, the overall heat transfer coefficient (flue gas to feedwater) used in the heat exchanger calculation can be approximated by the following relationship:

$$U = 0.98\ (h_c + h_r)\ k_f \qquad (2)$$

where

U = overall heat transfer coefficient, Btu/h ft² F (W/m² K)
h_c = gas side heat transfer coefficient for a bare tube bundle, Btu/h ft² F (W/m² K) (Chapter 4, Equations 54 and 55)
h_r = inter-tube radiation heat transfer coefficient, Btu/h ft² F (W/m² K) ≈ 1.0 for coal firing
k_f = surface effectiveness factor = 0.7 for coal, 0.8 for oil and 1.0 for gas

Finned tubes Heat transfer performance of finned tube economizers can be evaluated in a similar fashion except that appropriate relationships for extended surfaces should be used. In addition, because the gas side heat transfer has been enhanced, the water side heat transfer coefficient and tube wall thermal resistance are more significant and must be included in the evaluation. (See Chapter 4.) As a general guideline, the overall heat transfer coefficient can be approximated by the following relationship for most types of economizer fins:

$$U = 0.95\ (h_g k_f) \qquad (3)$$

where h_g is the gas side heat transfer coefficient for the heat transfer across finned tube bundles evaluated with the procedures defined in Chapter 4. A calculated example is provided in Chapter 21 for a bare tube economizer tube bundle.

Gas side resistance

The gas side pressure loss across the economizer tube bank can be evaluated using the crossflow correlations presented in Chapter 3. The pressure loss should be adjusted for the number of tube rows using the correction factors provided. The gas side resistance across the in-line finned tube banks is approximately 1.5 times the resistance of the underlying bare tubes.

Water side pressure drop

The water side pressure loss can be evaluated using the procedures in Chapter 3 where the total pressure loss ΔP_T is calculated:

$$\Delta P_T = \Delta P_f + \Delta P_l + \Delta P_z \qquad (4)$$

where

ΔP_f = friction pressure loss	Ch. 3, Eq. 42
ΔP_l = sum of the local losses (entrance, bends and exits)	Ch. 3, Eq. 47
ΔP_z = static head loss	Ch. 5, Eq. 10

The design pressure is then evaluated by the sum of the drum design pressure and the total pressure loss ΔP_T rounded up to the nearest 25 psig (1.7 bar gauge).

If the calculated water side pressure drop is excessive, the number of parallel flow paths must be increased. If the gas velocity can be increased, the water side pres-

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sure drop can also be reduced by increasing the tube size, usually in increments of 0.125 in. (3 mm). As indicated in Chapter 3, the dynamic pressure drop is inversely proportional to the fifth power of the inside tube diameter. This is significant and may be advantageous for retrofit applications. Sometimes, a material upgrade (from SA-210A1 to SA-210C, for example) can permit use of the optimum tube wall thickness.

Economizer support systems

Economizers are located within tube wall enclosures or within casing walls, depending on gas temperatures. In general, casing enclosures are used at or below 850F (454C) and inexpensive carbon steel can be used. If a casing enclosure is used, it must not support the economizer. However, tube wall enclosures may be used as supports.

The number of support points is determined by analyzing the allowable deflection in the tubes and tube assemblies. Deflection is important for tube drainability. Figs. 10 through 12 show typical support arrangements for bare tube economizers.

Wall or end supports are usually chosen for relatively short spans and require bridge castings or individual lugs welded or attached to the tube wall enclosures. (See Fig. 10.) Another possibility exists if enclosure wall (usually primary superheater circuitry) headers are present above the economizer (for example, Fig. 11).

Quarter point stringer supports are used for spans exceeding the limits for end supports (Fig. 12). The stringers are mechanically connected to the economizer sections, which are held up by ladder type supports. The supports exposed to hot inlet gases may be made of stainless steel, while lower grade material is normally used to support the lower bank which is exposed to reduced gas temperatures. In Babcock & Wilcox (B&W) designs, stringer tubes also usually support other horizontal convection surfaces above the economizer. (See frontispiece.) Bottom support is sometimes used if the gas temperature leaving the lowest economizer bank is low enough.

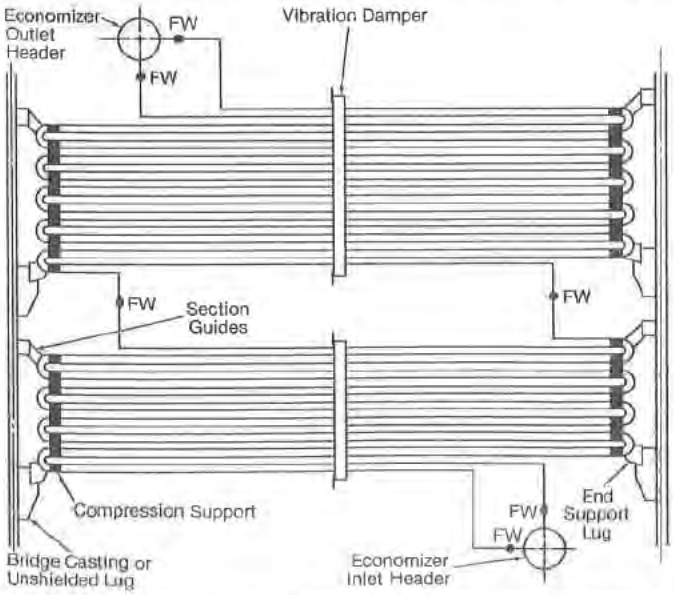


Fig. 10 Economizer supports — sample waterwall support arrangement. (FW = field weld.)

Bank size

The bank size is limited by the following constraints:

1. type of fuel,
2. fabrication limits,
3. sootblower range,
4. maximum shipping dimensions,
5. construction considerations, especially for retrofits, and
6. maintenance.

Bank depths greater than 6 ft (1.82 m) are rare in new boilers, while larger banks can be tolerated in retrofits.

Access requirements

Cavities around the banks are needed for field welding, tube leg maintenance and sootblower clearance. A sufficient number of access doors must be arranged in the enclosure walls to access these cavities. Cavity access can be provided from the outside through individual doors or from the inside through special openings across stringers or collector frames. The minimum cavity height should be 2 ft (0.6 m) of crawl space.

Headers

B&W economizer header designs are based on American Society of Mechanical Engineers (ASME) Code requirements. Inlet headers are frequently located inside the gas stream and may receive feedwater through one or both ends. Regardless of design, it is necessary to properly seal the inlet pipe where it penetrates the enclosure by using brackets and flexible seals. The seal becomes especially important in pressurized (forced draft) units. Other important considerations are tube leg flexibility, differential expansion and potential gas temperature imbalances and upsets.

The outlet headers, not to be confused with intermediate headers seen in larger boilers and stringer supports, receive the heated feedwater and convey it to the drum

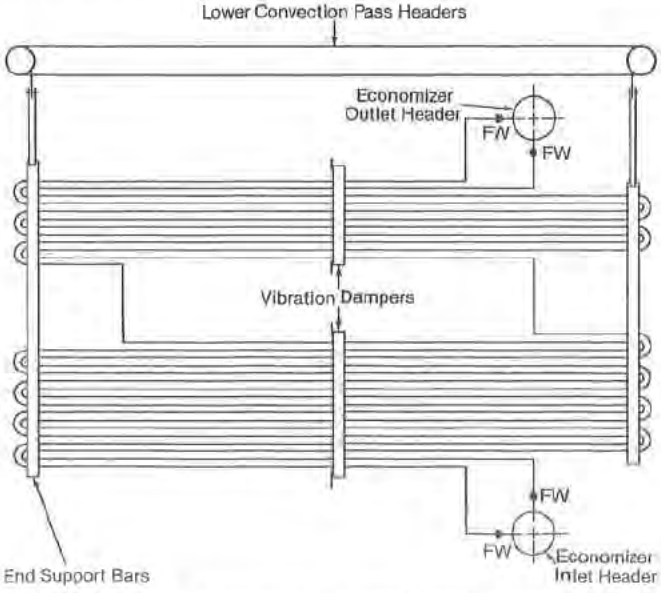


Fig. 11 Economizer supports — sample lower waterwall header arrangement. (FW = field weld.)

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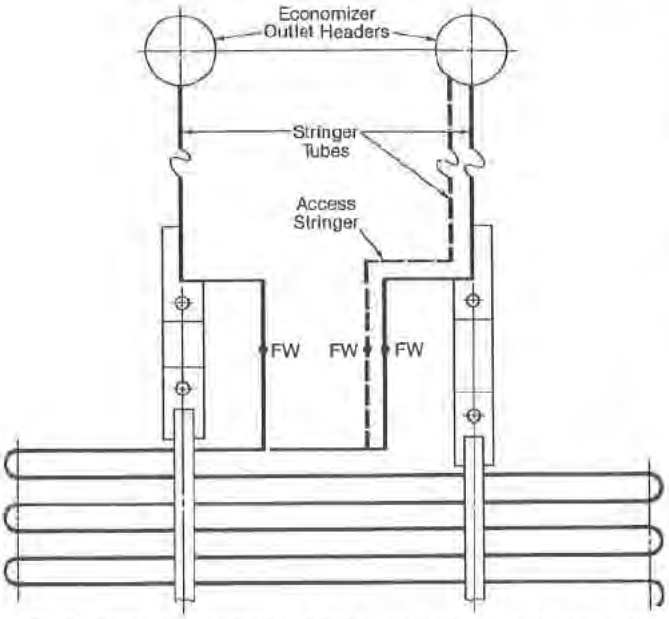


Fig. 12 Economizer supports — sample stringer support arrangement.

or, in the case of once-through boilers, to the downcomer supplying the furnace circuitry. Inlet and outlet headers must be large enough to assure reasonable water flow distribution in the economizer banks. Flow velocities are typically less than 20 ft/s (6 m/s).

Vibration ties

Vibration ties or tube guides are required on some end-supported tube sections. These ties may be needed if the natural frequencies within the boiler load range are in or near resonance with the vortex shedding frequency.

Stringer tubes are also subject to vibration. This vibration is magnified by long unsupported stringer tube lengths near the large cavity below the convection pass roof.

Tube geometry, materials and code requirements

Economizer tube diameters typically range between 1.75 and 2.5 in. (44.5 and 63.5 mm). Tubes outside this range are sometimes used in retrofits. Smaller tubes are normally used in once-through, supercritical boilers where water side pressure drop is less of a consideration. In these units, tube wall thicknesses are minimized.

The ASME Code requires that the design temperature for internal boiler pressure parts is at least 700F (371C). The calculated mean tube wall temperature in economizers seldom reaches this temperature. It usually lies 10 to 20F (6 to 12C) above the fluid temperature, which seldom exceeds 650F (343C) along any economizer circuit.

The minimum tube wall thickness is determined in accordance with procedures outlined in Chapter 7.

In coal-fired boilers, the side spacing is usually determined by the maximum allowable gas velocity and gas side resistance, which are functions of a given tube size. If fins are used, the side and back tube spacings should permit the fin tips to be at least 0.5 in. (13 mm) apart. For bare tubes, a minimum clear spacing of 0.75 in. (19 mm) is desirable.

The minimum back (vertical) spacing of the tubes should be no less than 1.25 times the tube outside diameter. Smaller ratios can reduce heat transfer by as much as 30%. Ratios larger than 1.25 have relatively little effect on heat transfer but increase the gas side resistance and bank depth.

Air heaters

Air heaters are used in most steam generating plants to heat the combustion air and enhance the combustion process. Most frequently, the flue gas is the source of energy and the air heater serves as a heat trap to collect and use waste heat from the flue gas stream. This can increase the overall boiler efficiency by 5 to 10%. Air heaters can also use extraction steam or other sources of energy depending upon the particular application. These units are usually employed to control air and gas temperatures by preheating air entering the main gas-air heaters.

Air heaters are typically located directly behind the boiler, as depicted in Fig. 2, where they receive hot flue gas from the economizer and cold combustion air from the forced draft fan. The hot air produced by air heaters enhances combustion of all fuels and is needed for drying and transporting the fuel in pulverized coal-fired units.

Classification of air heaters

Air heaters are classified according to their principle of operation as recuperative or regenerative.

Recuperative

In a recuperative heat exchanger, heat is transferred continuously through stationary solid heat transfer surfaces which separate the hot flow stream from the cold flow stream. The most common heat transfer surfaces are tubes and parallel plates. Recuperative heat exchangers function with little cross-contamination, or leakage, between streams.

Tubular air heaters In a typical tubular air heater, energy is transferred from the hot flue gas flowing inside many thin walled tubes to the cold combustion air flowing outside the tubes. The unit consists of a nest of straight tubes that are roll expanded or welded into tubesheets and enclosed in a steel casing. The casing serves as the enclosure for the air or gas passing outside of the tubes and has both air and gas inlet and outlet openings. In the vertical type (Fig. 13), tubes are supported from either the upper or lower tubesheet while the other (floating) tubesheet is free to move as tubes expand within the casing. An expansion joint between the floating tubesheet and casing provides an air/gas seal. Intermediate baffle plates parallel to the tubesheets are frequently used to separate the flow paths and eliminate tube damaging flow induced vibration.

Carbon steel or low alloy corrosion resistant tube materials are used in the tubes which range from 1.5 to 4 in. (38 to 100 mm) in diameter and have wall thicknesses of 18 to 11 gauge [0.049 to 0.120 in. (1.24 to 3.05 mm)]. Larger diameter, heavier gauge tubes are used when the potential for tube plugging and corrosion exists. Tube arrangement may be in-line or staggered with the latter being more thermally efficient.

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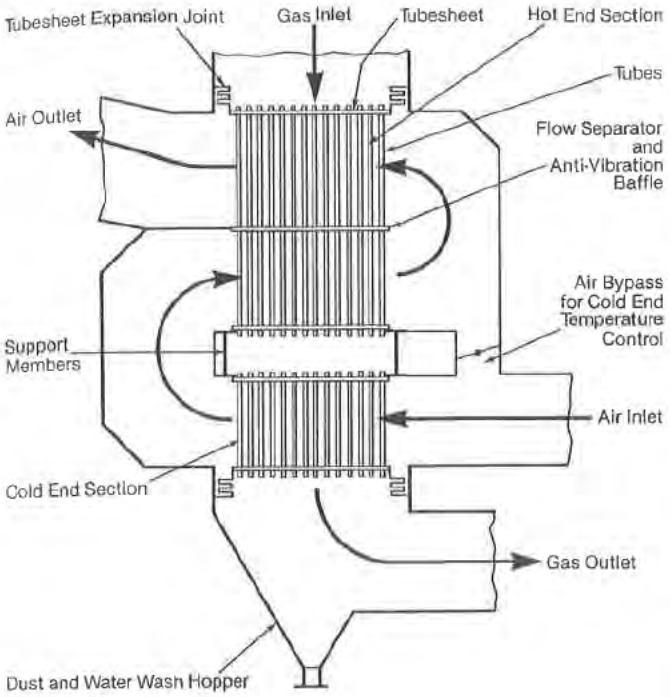


Fig. 13 Vertical type tubular air heater.

The most common flow arrangement is counterflow with gas passing vertically through the tubes and air passing horizontally in one or more passes outside the tubes. A variety of single and multiple gas and air path arrangements are used to accommodate plant layouts. Designs frequently include provisions for cold air bypass or hot air recirculation to control cold end corrosion and ash fouling. Modern tubular air heaters are shop assembled into large, transportable modules. Several arrangements are shown in Fig. 14.

Cast iron air heaters Cast iron tubular air heaters are heavy, large and durable. Their use is mainly limited to the petrochemical industry, but some are used on electric utility units. Cast iron is used because of its superior corrosion resistance. Rectangular, longitudinally split tubes are assembled from two cast iron plates and individual tubes are assembled into air heater sections. Air heaters are usually arranged for a single gas pass and multiple air passes with air flow inside the tubes. Heat transfer is maximized by fins cast into inside and outside tube surfaces.

Plate air heaters This type of air heater transfers heat from hot gas flowing on one side of a plate to cold air flowing on the opposite side, usually in crossflow. Heaters consist of stacks of parallel plates. Sealing between air and gas streams at plate edges is accomplished by weld-

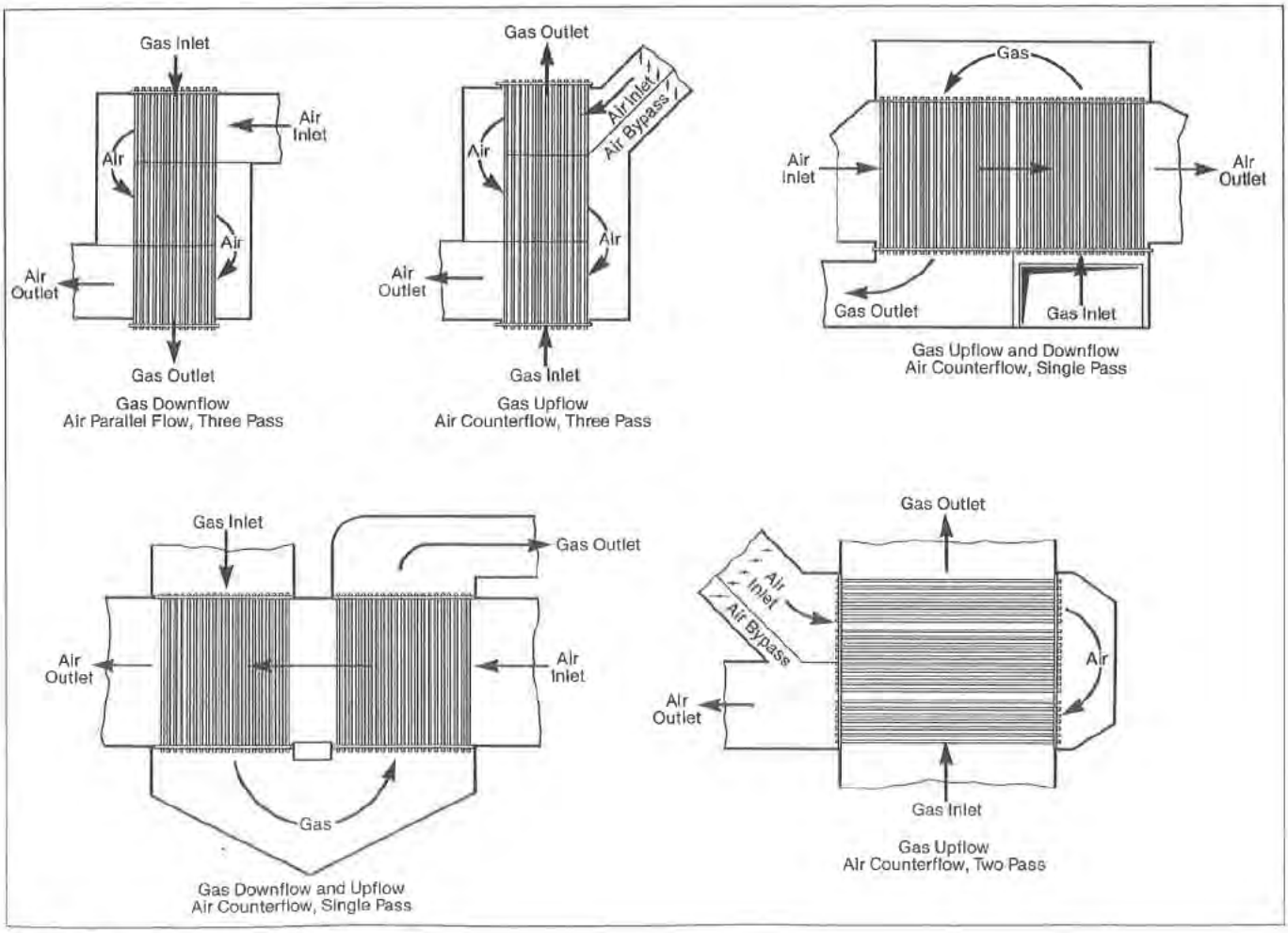


Fig. 14 Various tubular air heater arrangements.

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ing or by a combination of gaskets, springs and external compression of the plate stack. Plate materials and spacing can be varied to accommodate operating requirements and fuel types.

Steel plate air heaters were some of the earliest types used, but their use declined due to plate to plate sealing problems. However, recent sealing developments have prompted increased use in industrial and small utility applications. Plate modules may be combined to make different size air heaters with a variety of flow path arrangements. A single gas pass, two air pass plate air heater is shown in Fig. 15. Modern plate units are somewhat smaller than tubular units for a given capacity and exhibit minimal air to gas leakage.

Steam coil air heaters Steam coil and water coil recuperative air heaters are widely used in utility steam generating plants to preheat combustion air. Air preheating reduces the corrosion and plugging potential in the cold end of the main air heater. Occasionally, they serve as the only source of preheated combustion air. These heaters consist of banks of small diameter, externally finned tubes arranged horizontally or vertically in air ducts between the combustion air fan and main air heater. Combustion air, passing in crossflow outside the tubes, is heated by turbine extraction steam or feedwater flowing inside the tubes. Ethylene glycol is sometimes used as the hot fluid to prevent out of service freezing damage.

Heat pipe A heat pipe is a simple, highly efficient device for transporting thermal energy. The basic thermosyphon type heat pipe used in steam generation unit air heaters consists of an evacuated sealed pipe which has been partially filled with a heat transfer fluid (Fig. 16). The evaporator end of the pipe is exposed to a heat source (hot flue gas) and the other end, the condenser, is placed in a heat sink (cold combustion air). Heat absorbed from the flue gas evaporates the fluid which travels to the combustion air end where heat is released as the fluid condenses. The condensed fluid returns by capillary action and gravity to the evaporator end. Fluid circulation within the tube is continuous as long as there is a temperature difference between evaporator and condenser ends; fluid temperature is nearly isothermal at approximately the average of air and gas temperatures. Large quantities of heat are transferred as the heat of condensation and vaporization are used. Heat pipes operate with the evaporator end lower than the condenser end; they are inclined from the horizontal. Internal surfaces of heat pipes are roughened or grooved to assist fluid circulation and external surfaces are usually finned to increase heat transfer surface area.

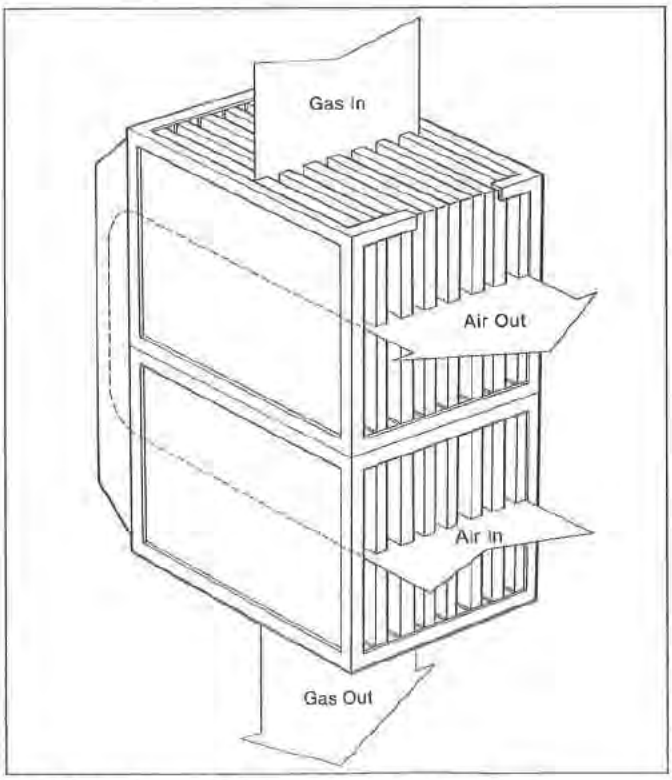


Fig. 15 Single gas pass, two air pass plate air heater.

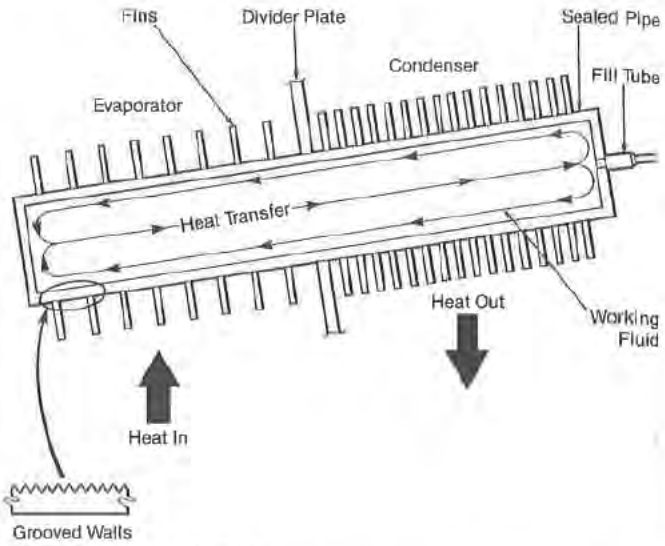


Fig. 16 Heat pipe schematic.

Heat pipe air heaters consist of bundles of parallel heat pipe tubes. About half of the tube length is exposed to flue gas flow and the remaining length is exposed to air flow. A central divider plate separates air and gas and supports the tube bundle. Heat pipe bundles or modules can be combined and enclosed in casing to build air heaters capable of accepting a variety of flow configurations.

Tube bundle configuration is usually in-line for dirty gas applications such as coal and heavy oil and staggered for natural gas and light oil. Typically, 2 in. (51 mm) diameter carbon steel tubes up to 40 ft (12.2 m) long are used. They feature three steel fins per in. (1 per 8.5 mm) on the gas side and up to ten fins per in. (1 per 2.5 mm) on the air side. Corrosion resistant alloy materials can increase cold end corrosion life.

Heat pipe air heaters are smaller than tubular air heaters and air to gas leakage is minimal as in other types of recuperative air heaters. Due to the isothermal behavior of each tube, these units can operate at a lower gas outlet temperature for a specific minimum metal temperature (MMT) compared to a tubular or regenerative air heater. This can permit operation at higher boiler efficiency and can reduce the potential for air heater cold end corrosion.

Of utmost importance is the heat transfer fluid's long term compatibility with tube wall material. Incompatibility can result in internal corrosion which produces noncon-

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densable gases, reduces heat transfer rates and jeopardizes tube pressure integrity. Hydrocarbons and water based fluids, commonly used in carbon steel tubes, are subject to temperature limits of 400 to 800F (204 to 427C). Water based fluids can not be allowed to freeze. High alloy steel or nonferrous tube materials may be used to extend temperature limits or to allow more fluid choices, but their expense precludes use in large heaters.

Heat pipe air heaters have been used successfully in the petrochemical industry. A limited number have also been applied to electric utility steam generation units. More widespread adoption is expected when long term reliability is proven.

Regenerative

In a regenerative air heater, heat is transferred indirectly as a heat storage medium is alternately exposed to hot and cold flow streams. A variety of materials can be used as the medium and periodic exposure to hot and cold flow streams can be accomplished by rotary or valve switching devices. In steam generating plants, tightly packed bundles of corrugated steel plates serve as the storage medium. In these units either the steel plates, or surface elements, rotate through air and gas streams or rotating ducts direct air and gas streams through stationary surface elements.

Regenerative air heaters are relatively small and are the most widely used type for combustion air preheating in electric utility steam generating plants. Their most notable operating feature is that a small but significant amount of air leaks into the gas stream due to the rotary operation.

Ljungström The most prevalent regenerative air heater is the Ljungström type (Fig. 17), which features a cylindrical shell plus a rotor which is packed with bundles of heating surface elements and is rotated through counterflowing air and gas streams. The rotor is enclosed by a stationary housing which has ducts at both ends. Air flows through one half of the rotor and gas flows through the other half. Metallic leaf-type seals minimize air to gas leakage and flow bypass around the rotor. Bearings in upper and lower beam assemblies support and guide the rotor at the central shaft. A rotor speed of one to three rpm is provided by a motor driven pinion engaging a rotor encircling pinrack. Both vertical and horizontal shaft designs are used to accommodate various plant air and gas flow schemes. The vertical shaft design is more common.

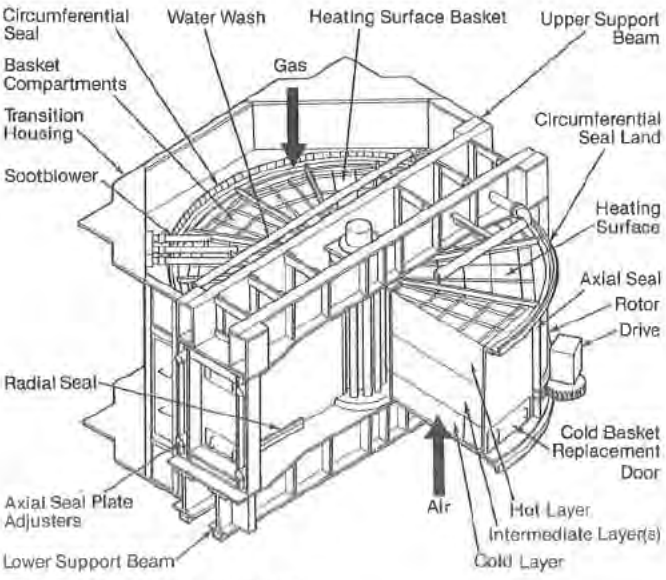


Fig. 17 Ljungström-type air heater.

Rothemühle The Rothemühle-type regenerative air heater uses stationary surface elements and rotating ducts (Fig. 18). The surface elements are supported and contained within a stationary cylindrical shell called the stator. On both sides of the stator, a double wing symmetrical hood rotates synchronously on a common vertical shaft. The central shaft is supported by bearings within the stator and the hoods are driven slowly by a pinion which engages a pinrack encircling the lower hood. Stationary housings surround the hoods. Heat is transferred as flow streams are directed through the heating surface in counterflow fashion, one flow stream inside the hoods and the other outside. Either air or gas may pass through the hoods. However, air is more common because it requires less fan power. Special spring mounted sealing systems employing cast iron seals are used at hood rotating to stationary interfaces to minimize air to gas leakage.

Several features distinguish the Rothemühle from the Ljungström heater. Its relatively low rotating weight (20% of total weight) contributes to high reliability. The stationary stator permits air heater loads to be distributed equally to a number of surrounding points, permitting transfer of significant duct loads through the air

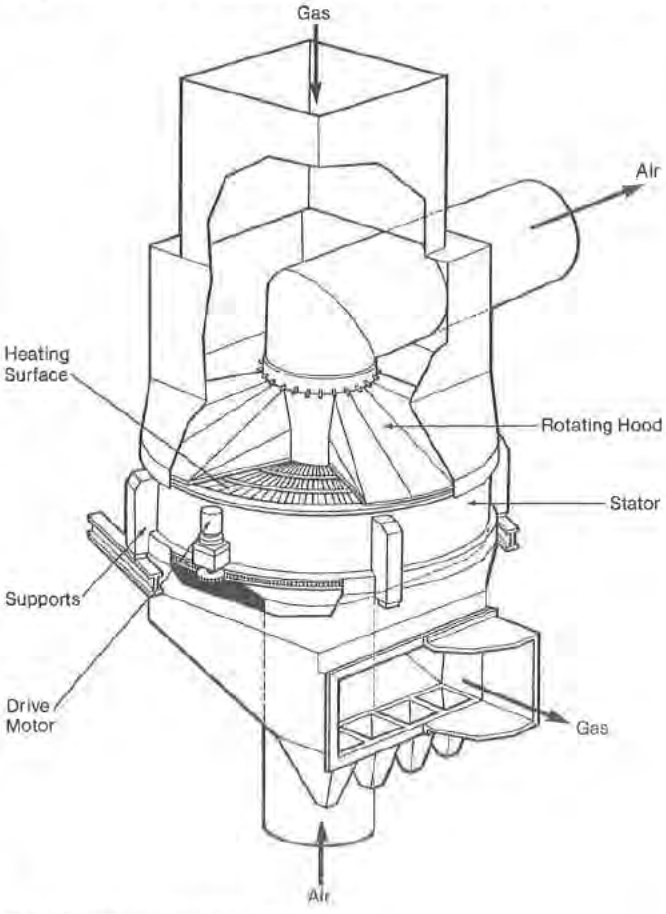


Fig. 18 Rothemühle-type air heater.

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heater into structural steel. The spring mounted hood sealing system, which adapts to the curvature of the stator during operation, allows hot starts without overloading the drive motor. A simple, permanent early warning fire detection system can be embedded in the stator.

Regenerative heating surface Regenerative air heater surface elements are a compact arrangement of two specially formed metal plates. Each element pair consists of a combination of flat, corrugated or undulated plate profiles. The roll formed corrugations and undulations serve to separate the plates to maintain flow paths, increase heating surface area and maximize heat transfer by creating flow turbulence. The steel plates, 26 to 18 gauge thick, are typically spaced 0.2 to 0.4 in. (5 to 10 mm) apart. Closely spaced, highly profiled element pairs exhibit a high heat transfer rate, pressure drop and fouling potential while widely spaced element combinations, where one plate is flat, exhibit a low heat transfer rate, low pressure drop and reduced fouling potential. The combination of plate profile, material and thickness is selected for maximum heat transfer, minimum pressure drop, good cleanability and high corrosion resistance.

Surface elements are stacked and bundled into self-contained baskets and are installed into air heater rotors and stators in two or more layers. The surface layer at the air inlet side, designated the cold layer, is distinguished from other layers by design. Cold layers, which are subject to corrosion and ash fouling, are typically 12 in. (300 mm) deep for economical replacement. Heavy gauge, open profile elements are used for corrosion resistance and cleanability. Practically all cold layer elements are low alloy corrosion resistant steel or, when high corrosion potential exists, porcelain enamel coated steel. Hot and intermediate surface layers are more compact than cold layers and use thinner plates. Figs. 17 and 19 illustrate several heating surface element profiles and air heater surface arrangements.

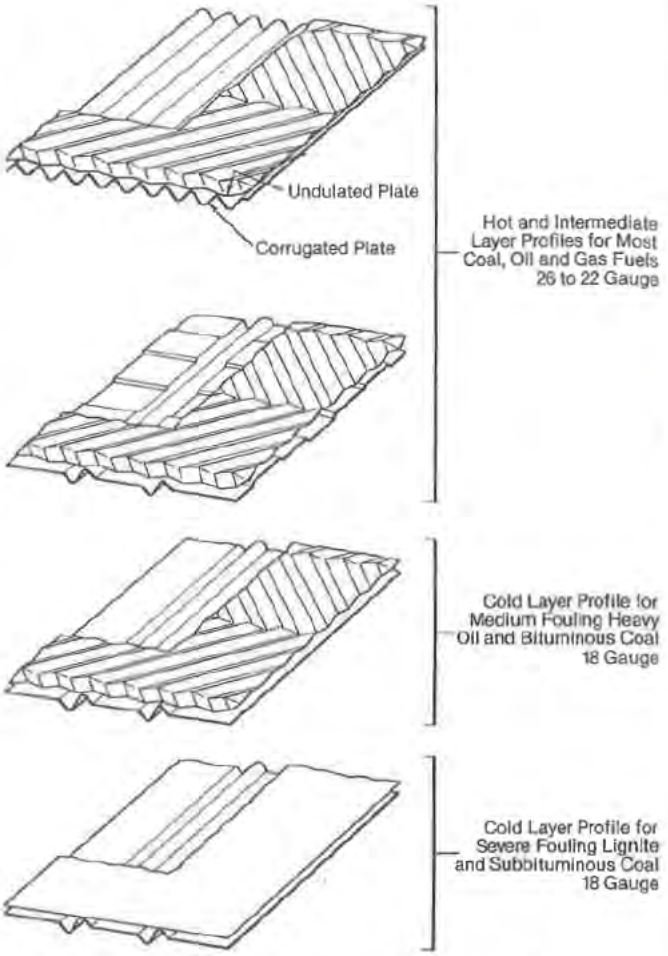


Fig. 19 Regenerative air heater surface element profiles.

Advantages and disadvantages

Many subtle differences exist between air heater designs within a particular type. However, there are some general advantages and disadvantages associated with each type which are listed in Table 1. Note that the recuperative heat pipe air heater is listed separately.

Performance and testing

Air heaters are designed to meet performance requirements in three areas: thermal, leakage and pressure drop. Low performance in any area increases boiler operating costs and may cause unit load curtailment.

Thermal performance

The thermal performance and surface area (A) of a recuperative air heater can be evaluated by:

$$A = Q / (U \quad LMTD \quad F) \qquad (5)$$

where Q is the total thermal load [Btu/h (W)], U is the overall heat transfer coefficient, $LMTD$ is the log mean temperature difference and F is the corresponding geometry correction factor. The performance, U, $LMTD$ and F can be evaluated using the correlations and methodology presented in Chapter 4. The overall U should include convection and radiation components as well as the appropriate gas and air side fouling factors. U typically ranges from 3 to 10 Btu/h ft^2 F (17 to 57 W/m^2 K).

Performance verification Thermal performance is measured by comparing the test gas outlet temperature to its design value. The true outlet temperature is obtained by correcting the measured temperature for air heater leakage and deviations from design conditions.

The ASME Performance Test Code, Section 4.3 (PTC 4.3), provides the following equation which is based on an air heater mass flow heat balance and assumes that the source of all leakage is from the entering air:

Table 1
Advantages and Disadvantages of Air Heater Types

Type	Advantage	Disadvantage
Recuperative	Low leakage No moving parts	Large and heavy Difficult to replace surface
Heat pipe	Low leakage High minimum metal temperatures No moving parts	Difficult to clean Temperature restrictions
Regenerative	Compact Easy to replace surface	Leakage High maintenance Fire hazard

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$$T_2 = T_{2m} + \left(\frac{\%\ lkg}{100}\right)\frac{c_{pa}}{c_{pg}}(T_{2m} - T_1) \qquad (6)$$

where

T_2 = air heater gas outlet temperature corrected for leakage, F (C)
T_{2m} = measured gas temperature leaving air heater, F (C)
$\%\ lkg$ = percent air leakage with respect to inlet gas flow
c_{pa}, c_{pg} = specific heat of air and gas respectively, Btu/lb F (J/kg C)
T_1 = air inlet temperature, F (C)

The measured gas outlet temperature must also be corrected for deviations in various operating parameters such as mass flow rates and operating temperatures in order to accurately assess performance. Suppliers and the ASME Performance Test Codes provide various correction curves and factors for this purpose.

Leakage

Air flow passing from the air side to the gas side is called leakage. It is quantified in pounds per hour (kg/s) but is frequently expressed as a percentage of the gas inlet flow. Leakage is undesirable primarily because it represents fan power wasted in conveying air which bypasses the boiler combustion zone. Leakage can also reduce an air heater's thermal performance.

All air heaters leak. Recuperative units may begin operation with essentially zero leakage, but leakage occurs as time and thermal cycles accumulate. With regular maintenance, leakage can be kept below 3%.

Air heater leakage is inherent with the rotary regenerative design. There are two types of leakage, gap and carryover. Gap leakage occurs as higher pressure air passes to the lower pressure gas side through gaps between rotating and stationary parts. Its rate is given by the following general expression:

$$w_l = KA\ (2g_c\ \Delta P\rho)^{1/2} \qquad (7)$$

where

w_l = leakage flow rate, lb/h (kg/s)
K = discharge coefficient, dimensionless (generally 0.4 to 1.0)
A = flow area, ft² (m²)
g_c = 32.17 lbm ft/lbf s² × (3600 s/h)²
= 4.17 × 10⁸ lbm ft/lbf h² (1 kg m/N s²)
ΔP = pressure differential across gap, lb/ft² (kg/m²)
ρ = density of leaking air, lb/ft³ (kg/m³)

Carryover leakage is the air carried into the gas stream from each rotor (stator) heating surface compartment as the surface passes from the air stream to the gas stream. This leakage is directly proportional to the void volume of the rotor and the rotation speed.

Regenerative air heater design leakage ranges from 5 to 15% but increases over time as seals wear. During recent years effective automatic sealing systems, which nearly eliminate leakage rise due to seal wear, have been applied. These systems monitor and adjust rotating to stationary seals on-line.

Another source of air to gas flow, which appears as air heater leakage, is outside air infiltration into lower pressure gas streams. Infiltration may occur at casing cracks or holes, flue expansion joints and access doors or gaskets. This sometimes neglected source can be significant and difficult to detect if leaks occur under lagging and insulation.

Air heater leakage can be obtained directly as the difference between air or gas side inlet and outlet flows based on velocity measurements. However, because velocity measurements are difficult to obtain accurately in large duct cross-sections, air heater leakage is more accurately based on calculated gas weights using gas analysis, boiler efficiency and fuel analysis data. (See Chapter 9.) Approximate air heater leakage can be determined by the following formulas based on gas inlet and outlet analysis (dry basis).

$$\%\ \text{Leakage} = \frac{\%\ O_2\ \text{Leaving} - \%\ O_2\ \text{Entering}}{21 - \%\ O_2\ \text{Leaving}} \times 90 \qquad (8)$$

Test air heater leakage should be corrected for deviations from design cold end air to gas differential pressure and inlet air temperature before comparison to design leakage.

Pressure drop

In recuperative air heaters, gas or air side pressure drop arises from frictional resistance to flow, inlet and exit shock losses and losses in return bends between flow passes. In regenerative air heaters, the main cause is heating surface frictional flow resistance. In both cases, pressure drop is proportional to the square of the mass flow rate. Typical values at full load flows are 2 to 7 in. wg (0.5 to 1.7 kPa).

Air and gas side pressure drop values are the differences between terminal inlet and outlet static gauge pressures. Correction of measured pressure drops for deviations from design flows and temperatures is necessary before comparison to design values.

Operational concerns

There are several operating conditions and maintenance concerns common to most air heaters. These include corrosion, plugging and cleaning, erosion and fires. Air heaters used with high ash and/or high sulfur content fuels require more attention and maintenance than those firing clean fuels such as natural gas.

Corrosion

Air heaters used on units firing sulfur bearing fuels are subject to cold end corrosion of heating elements and nearby structures. In a boiler, a portion of the sulfur dioxide (SO_2) produced is converted to sulfur trioxide (SO_3) which combines with moisture to form sulfuric acid vapor. This vapor condenses on surfaces at temperatures below its dew point of 250 to 300F (120 to 150C). Because normal air heater cold end metal temperatures are frequently as low as 200F (93C), acid dew point corrosion potential exists. The obvious solution would be to operate at metal temperatures above the acid dew point but this results in unacceptable overall boiler heat losses. Most air heaters are designed to operate at MMTs somewhat below the acid dew point, where the efficiency

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gained more than balances the additional maintenance costs. B&W recommends limiting MMTs to the values in Figs. 20 and 21 when burning sulfur bearing fuels.

When fuel sulfur levels are high, or ambient temperatures or operating loads are low, MMTs may be unacceptably low. These situations dictate the use of active or passive cold end corrosion control methods. Active systems used to raise MMT include: 1) steam- or water-coil air heaters to preheat inlet air, 2) cold air bypass, in which a portion of the inlet air is ducted around the air heater, and 3) hot air recirculation, in which a portion of the hot outlet air is ducted to combustion air fan inlets.

Passive corrosion control methods incorporated in air heater design include: 1) thicker cold end materials, such as 11 or 14 gauge (3 or 2 mm) tubes and 18 gauge (1 mm) regenerative surface elements, 2) low or high alloy cold end surface materials which have at least twice the corrosion life of carbon steel, 3) nonmetallic coating, such as porcelain enamel, teflon, or epoxies on cold elements, 4) nonmetallic cold end surface materials such as extruded ceramic in regeneratives and borosilicate glass tubes in tubulars, and 5) tubular air heater cold end tube arrangements which maximize MMT by providing higher gas flow velocities.

Plugging and cleaning

Plugging is the fouling and eventual closing of heat transfer flow passages by gas-entrained ash and corrosion products. It can occur at the air heater hot end but is most common at the cold end where ash particles adhere to acid moistened surfaces. Plugging increases air heater pressure drop and can limit unit load when fan capacity is reached at less than full load. Air heater deposits are controlled and removed by sootblowing, cold end temperature control, surface design, off-line cleaning and furnace additives depending upon the particular application.

Erosion

Heat transfer surfaces and other air heater parts can suffer erosion damage through impact of high velocity, gas-entrained ash particles. Erosion usually occurs near gas inlets where velocities are highest. However, areas near seals in regenerative air heaters can also be damaged as ash is accelerated through seal gaps. The undesirable effects of erosion are structural weakening, loss of heat transfer surface area and perforation of components which can cause air to gas or infiltration leakage. Erosion rate is a function of velocity, gas stream ash loading, physical nature of ash particles and angle of particle impact. It is controlled by reducing velocities, removing erosive elements from the gas stream, or using sacrificial material.

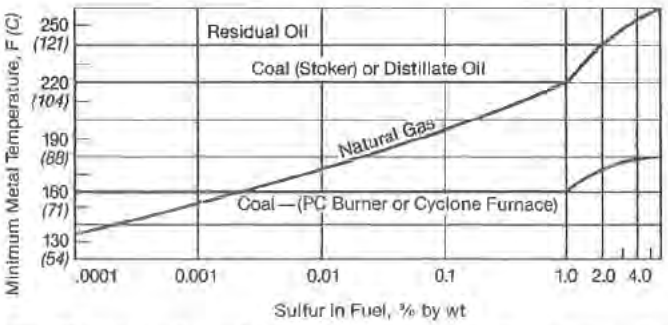


Fig. 20 Recuperative air heater cold end MMT limits when burning sulfur bearing fuels.

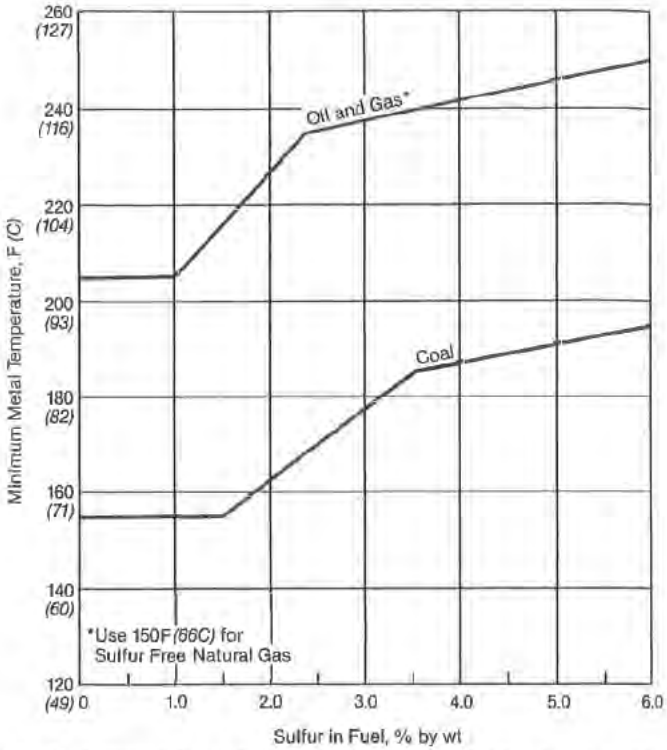


Fig. 21 Regenerative air heater cold end MMT limits when burning sulfur bearing fuels.

In the design stage, air heaters used with fuels containing highly erosive ash can be sized to limit gas inlet velocities to 50 ft/s (15 m/s). Inlet flues can also be designed to evenly distribute gas over the air heater inlet to eliminate local high velocity areas. Dust collectors, or strategically located screens and hoppers, may be used ahead of air heaters to remove some of the ash. In existing problem air heaters, flow distribution baffles may be installed to eliminate local high velocities, sacrificial materials such as abrasion resistant steel or ceramics may be placed over critical areas, or parts can be replaced with thicker materials for longer life.

Erosion in tubular air heaters frequently occurs within about 1 ft (0.3 m) of the gas inlet end due to turbulence as gas enters the tubes. Replaceable sacrificial sleeves may be installed in tube ends or egg crate-type flow straightening grids can be installed at tubular air heater inlets to reduce erosion.

Fires

Air heater fires are rare but do occur, particularly in regenerative units. They may be severe enough to completely destroy an air heater and are detected by thermocouples as well as special early warning systems. Fires usually start near the cold end, which can be fouled with unburned combustible materials. Most fires occur during startup as unburned fuel oil deposited on ash fouled heating surfaces is ignited. Leaking bearing lubrication equipment and heavy accumulations of flyash are also fire hazards. Fires can be avoided by maintaining a clean air heater and proper tuning of boiler firing equipment. Frequent sootblowing during startup and just before shutdown is a strongly recommended fire prevention practice.

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Utility applications

Gas to air recuperative and regenerative air heaters are usually used in utility units, primarily to enhance unit efficiency. Small increments of increased efficiency in large units amount to substantial fuel savings. Utility units generally use multiple air heaters for plant arrangement convenience, type of firing and maximum unit availability.

Pulverized coal-fired units require two streams of hot combustion air, i.e., primary air supplied at high pressure to pulverizers and secondary air supplied at lower pressure directly to burners. Two basic air flow systems are used, hot primary air and cold primary air. Each system uses air heaters. In the hot primary air scheme, used for smaller units, about one third of the combustion air heated in a secondary heater is ducted to hot primary air fans, where it is boosted in pressure and passed to the pulverizers; the remaining two thirds is ducted to the burners. The cold primary air system uses separate air heaters supplied by separate primary and secondary (forced draft) fans. In some units, the primary and secondary air are heated in a single regenerative unit.

If separate regenerative primary and secondary air heaters are used, the primary air heaters, which operate at high air to gas pressure differentials, exhibit twice as much leakage as the secondary units. For this reason, low leakage recuperative air heaters may be used for primary air heating and regeneratives may be used for secondary air heating.

Recuperative and regenerative air heaters are used on oil- and gas-fired units. In general, regardless of fuel type, larger units use regenerative air heaters because of their smaller size and lower initial cost. However, for air to gas pressure differentials above 40 in. wg (10 kPa), in fluidized bed applications for example, recuperative air heaters are usually preferred.

Industrial applications

Industrial units fire a variety of fuels such as wood, municipal refuse, sewage sludge and industrial waste gases as well as coal, oil and natural gas. As a result, many air heater types are used. In the small units, tubular, plate, heat pipe and cast iron heaters are widely used. Fuels fired on stoker grates, such as bituminous coal, wood and refuse, do not require high air temperatures, therefore water- or steam-coil air heaters can be used.

Environmental heat exchanger application

For environmental reasons, emission of certain fossil-fired combustion products may be limited by law. (See Chapter 32.) Systems developed to limit emissions of two objectionable flue gas constituents, NO_x and SO_2, may require the use of specially modified heat exchangers.

NO_x removal

Noncombustion NO_x reduction systems introduce ammonia into flue gas streams. The ammonia (NH_3) reacts with NO_x thermally at high temperatures and in the presence of a catalyst at lower temperatures to form molecular nitrogen (N_2) and water. NH_3 also reacts with some of the SO_3 in the flue gas to form ammonia-sulfur compounds which condense at temperatures below 530F (277C). Because air heaters or gas coolers are usually downstream of NO_x reduction equipment, the heat exchanger surface is subject to rapid fouling and increased corrosion potential, particularly in the 510 to 340F (266 to 171C) range. Regenerative heat exchangers used in these situations are designed with special features to minimize plugging and corrosion. These features include a minimum number of surface layers to minimize plugging between layers; heavy gauge, low alloy, corrosion resistant or enamel coated surface material for long corrosion life; open profile heating surface design for ease of cleaning; and hot and cold end sootblowers. Heat exchangers with these features can operate reliably without off-line water washing for a year or more. As discussed in Chapter 34, some regenerative air heaters have been modified to simultaneously serve as selective catalytic NO_x reduction systems.

SO_2 reduction

When sulfur emission reduction is required, flue gas desulfurization (FGD) systems are frequently used. These systems remove SO_2 from the flue gas by reaction with injected compounds such as limestone. In most cases, the scrubbed flue gas exits the FGD system at a saturation temperature of 120 to 130F (49 to 54C) before entering the stack. In cases where acid dew point corrosion of flues and stack liners is a concern or increased gas buoyancy is needed to improve stack plume dispersal, gas exiting the FGD system is reheated to 180F (82C) or higher. Regenerative and recuperative heat exchangers, similar to those used for air heating, are used for this application.

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Chapter 27
Waste-to-Energy Installations

The disposal of garbage is a problem that has been with us since civilization began. At various times throughout history, composting, animal feed, landfill and incineration have all been popular disposal methods. Today, refuse disposal methods are determined by cost and the effect on our environment.

The most common means of refuse disposal is still landfilling. Even in the late 1970s, nearly all of the refuse generated in North America was landfilled. Incineration with no heat recovery was a popular option that became economically unacceptable with the advent of environmentally responsible air pollution regulations and inexpensive landfill alternatives.

In Europe and Japan, where new landfill sites were less available, incineration continued as a viable option and those plants became the predecessors of today's refuse-to-energy plants. Heat recovery was added in the form of waste heat boilers which were originally hot water boilers and later low pressure and temperature steam boilers. These incinerators with waste heat boilers then evolved into waterwall boilers with integral stokers.

Refuse-fired boiler design parameters and operating characteristics are strongly affected by the components of the refuse, which change with time. The components also vary greatly by location. In North America, typical municipal solid waste (MSW) is high in paper and plastics content (Fig. 1) and typically has a lesser moisture content and greater heating value than that found worldwide. In a less industrialized country the refuse tends to have a greater moisture content and lesser heating value. Table 1 shows representative refuse analyses ranging from 3000 to 6000 Btu/lb (6978 to 13,956 kJ/kg) higher heating value (HHV) basis.

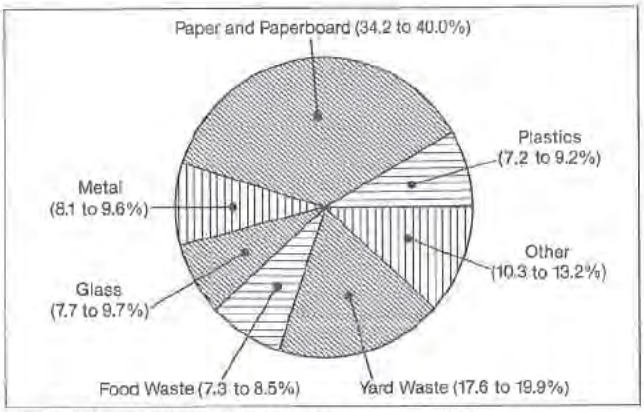


Fig. 1 U.S. municipal solid waste generation — 1990.

In the United States (U.S.) and North America, the refuse characteristics have changed dramatically in a short period of time. With more and more convenience foods, plastics, packaging, containers, and less food scraps due to home garbage disposals, the average refuse heating value has increased and the moisture content has decreased (Table 2). As more recycling programs are implemented, the analysis will continue to change. As glass, aluminum and other metals are recycled the refuse

Table 1
Range Of As-Received Refuse Fuel Analysis

	Weight Percent As-Received						
HHV, Btu/lb (kJ/kg)	3,000 (6,978)	3,500 (8,141)	4,000 (9,304)	4,500 (10,467)	5,000 (11,630)	5,500 (12,793)	6,000 (13,956)
Carbon	16.88	19.69	22.50	25.32	28.13	30.94	33.76
Hydrogen	2.33	2.72	3.10	3.49	3.88	4.27	4.66
Oxygen	12.36	14.42	16.49	18.55	20.62	22.68	24.75
Nitrogen	0.22	0.26	0.30	0.34	0.38	0.42	0.46
Sulfur	0.15	0.18	0.21	0.24	0.27	0.30	0.33
Chlorine	0.34	0.38	0.42	0.46	0.50	0.54	0.58
Moisture	35.72	32.35	29.98	28.60	25.22	22.85	21.46
Ash	32.00	30.00	27.00	23.00	21.00	18.00	14.00
Total	100.00	100.00	100.00	100.00	100.00	100.00	100.00

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Table 2
U.S. Refuse Trends

Increasing heating value per ton of refuse
1960 — 4,200 Btu/lb (9,769 kJ/kg)
1980 — 4,500 Btu/lb (10,467 kJ/kg)
2000* — 5,200 Btu/lb (12,095 kJ/kg)

More paper and paperboard
33% in 1970
41% in 2000

More plastics
2.7% in 1970
9.8% in 2000

Less food wastes
11.5% in 1970
6.8% in 2000

*Estimate

heating value will increase; as paper and plastics are recycled the heating value will decrease.

In 1990 approximately 490,000 t (444,521 t_m) per day of MSW were generated in the U.S. About 13% of that total was recycled, 15% sent to refuse-to-energy facilities, and the balance landfilled. Early in the 21st century this is expected to reach 40% for recycling and 50% for refuse-to-energy facilities. The growth of refuse-to-energy facilities in the U.S. accelerated in the 1980s (Fig. 2) due to the growing disposal costs for landfills and a government-created market for the sale of electric power.

As old landfills closed, new landfills became more difficult and costly to open and tended to be located farther from the source of the refuse, increasing transportation costs. Concerns about ground water contamination resulted in more expensive landfill designs with several containment layers and leachate monitoring and control systems. The passage of the Public Utility Regulatory Policies Act of 1978 (PURPA) required public utilities to purchase the electric power generated by refuse-to-energy plants. This created a revenue flow that helped offset the inherent high capital cost of these plants. These market forces resulted in a proliferation of refuse-to-energy facilities in the northeast U.S. where the costs of landfill and other disposal options were the highest, and selectively throughout North America in response to local environmental or economic factors.

Refuse disposal is a major problem worldwide and there is no single solution. An environmentally sound refuse disposal program includes generating less refuse, recycling components that can be economically reused, combustion of the balance of the refuse (including the efficient generation of electric power), and the landfill of the resulting ash.

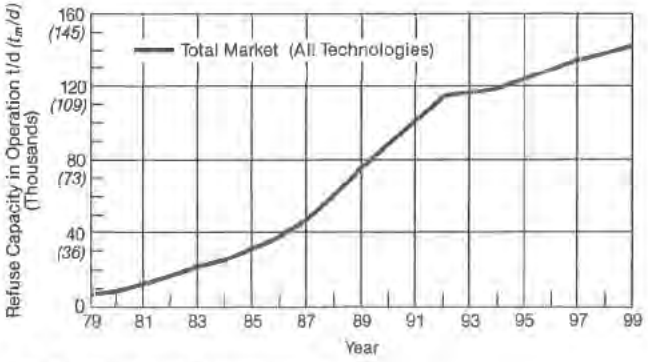


Fig. 2 U.S. refuse-to-energy market.

Refuse combustion alternatives

Two main techniques are used for burning municipal refuse, distinguished by the degree of fuel preparation. The first technique, known as mass burning, uses the refuse in its as-received, unprepared state (Fig. 3). Only large or noncombustible items such as tree stumps, discarded appliances, and other bulky items are removed. Refuse collection vehicles dump the refuse directly into storage pits. Overhead cranes equipped with grapples move the refuse from the pit to the stoker charging hopper. Hydraulic rams move the refuse onto the stoker grates. The combustible portion of the refuse is burned off and the noncombustible portion passes through and drops into the ash pit for reclamation or disposal.

The second burning technique uses prepared refuse, or refuse-derived fuel (RDF), where the as-received refuse is first separated, classified, and reclaimed in various ways to yield salable or otherwise recyclable products (Fig. 4). The remaining material is then moved to the boiler feeders and fed through multiple feeders onto a traveling grate stoker. The RDF is burned, part in suspension and part on a stoker. More finely shredded RDF can also be fired in suspension to supplement conventional fuels in large boilers used for power generation.

Corrosion

Combustion products from municipal refuse are very corrosive. The components that are present in coal, oil and other fuels that contribute to corrosion, as well as to high slagging and high fouling, are all present in refuse (Table 3). Corrosion in refuse-fired boilers is usually caused by the chlorides which deposit on the furnace, superheater and boiler tubes. Several modes of chloride corrosion may occur:

1. corrosion by hydrochlorides (HCl) in the combustion gas,
2. corrosion by NaCl and KCl deposits on tube surfaces,
3. corrosion by low melting point metal chlorides (mainly $ZnCl_2$ and $PbCl_2$), and
4. out of service corrosion by wet salts on the tube surface.

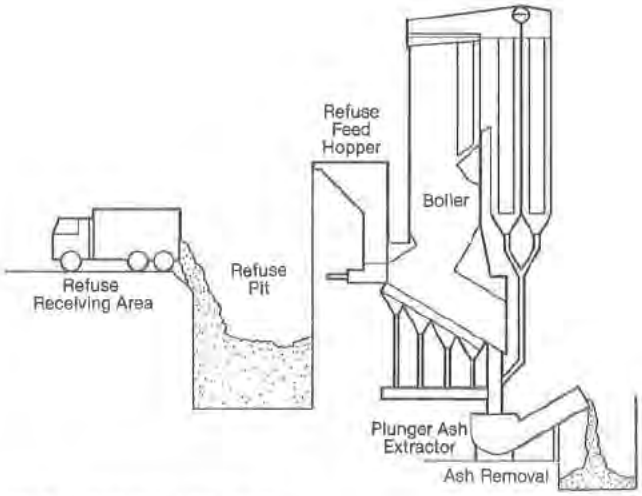


Fig. 3 Mass burning schematic.

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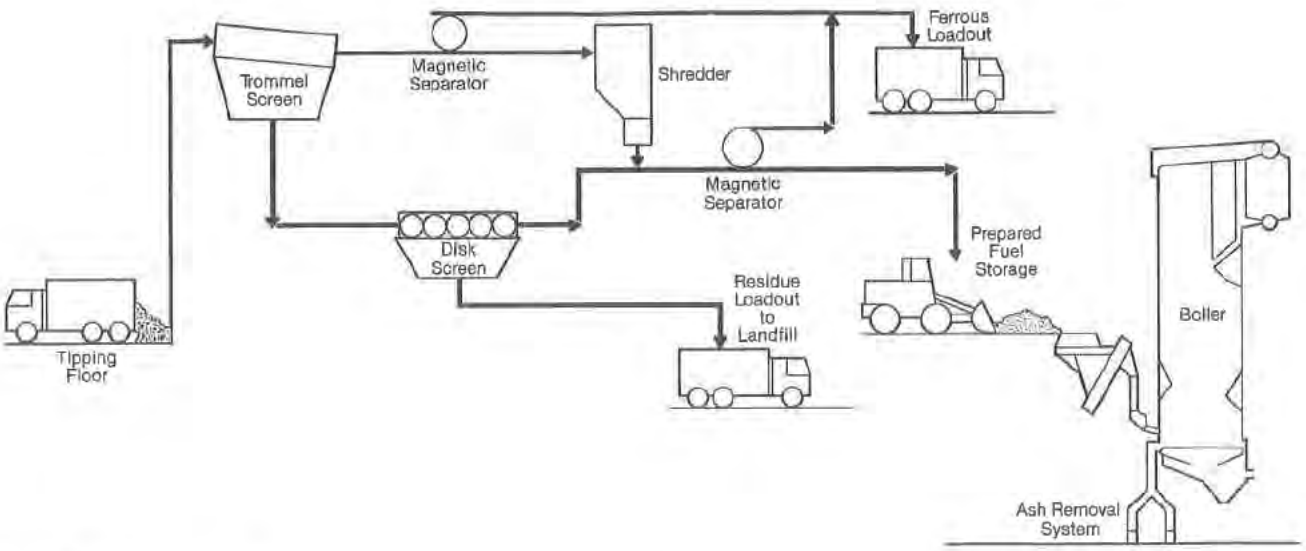


Fig. 4 RDF burning schematic.

The rate of tube metal loss due to corrosion is temperature dependent with high metal temperatures correlating with high rates of metal loss (Fig. 5). Refuse boilers operating at higher steam pressures have higher temperature saturated water in the furnace tubes and, therefore, these furnace tubes have higher metal temperature. Superheater tube metal temperatures are directly related to the steam temperature inside the tubes. In both cases, it is the temperature of the water or steam inside the tube that largely controls the tube metal temperature, rather than the temperature of the flue gas outside of the tube.

Furnace-side corrosion can be aggravated by poor water chemistry control. If water-side deposits are permitted to form, tube wall metal temperatures will rise and furnace corrosion will be accelerated. Standards for feedwater and boiler water quality are based on boiler operating pressures. These standards are no more stringent for refuse-fired units than for other fuels. However, the maintenance of feedwater and boiler water quality, and adherence to those standards, is more critical on refuse-fired boilers due to the highly corrosive nature of the fuel.

Lower furnace corrosion

The lower furnace environment of both mass-fired and RDF-fired units is constantly changing between an oxidizing atmosphere (an excess of O_2 beyond that needed for combustion) and a reducing atmosphere (a deficiency of O_2 below that needed for combustion) which can rapidly accelerate corrosion. Therefore, some form of corrosion protection is needed. Typically, the area of protection will encompass all four walls up to 30 ft (9.1 m) above the grate where there is reasonable assurance that oxidation zones are predominant.

Table 3
Corrosive Constituents in Fuels

Coal	Oil	Refuse	
Sodium	Sodium	Sodium	Chloride
Sulfur	Sulfur	Sulfur	Lead
Potassium	Vanadium	Potassium	Zinc
		Vanadium	

Mass-fired units

Virtually all mass-fired refuse boilers incorporate some type of pin stud and silicon carbide (SiC) refractory to protect the membraned lower furnace walls (Fig. 6). The quality and physical characteristics of the silicon carbide refractory must be maintained through proper application and curing. Lack of control during installation will result in spalling, deterioration and increased maintenance. The refractory material should have high thermal conductivity rates to minimize reducing the effectiveness of the water-cooled surface it is protecting. However, such characteristics may reduce its resistance to erosion as experienced along the grate line due to the scrubbing action of the refuse fuel and ash as it moves along the grate to the ash discharge. Increased erosion-resistant SiC materials are available for these zones. They do, however, have lower thermal conductivities.

A better alternative near the grate line is the use of armour blocks or refractory blocks, rigidly attached to the furnace walls. These blocks extend up the full height of the charging hopper opening which is about 4 ft (1.2 m)

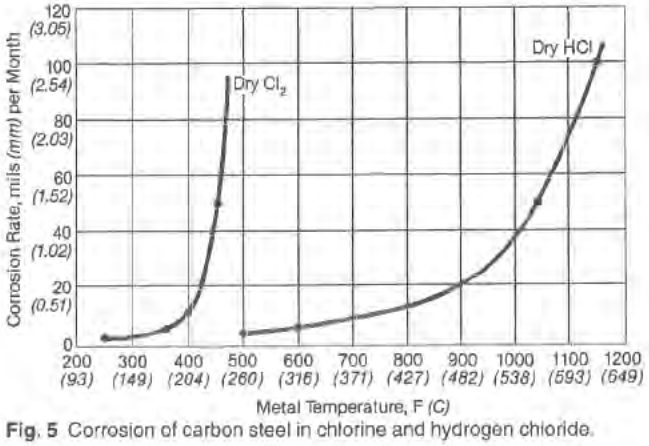


Fig. 5 Corrosion of carbon steel in chlorine and hydrogen chloride.

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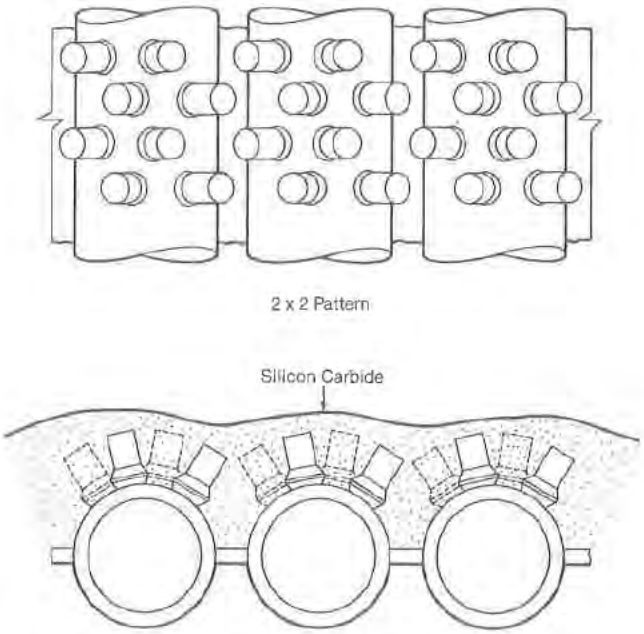


Fig. 6 Lower furnace studs and refractory, mass-fired unit.

high at the front of the furnace and tapers off to about 1 ft (0.3 m) high at the ash discharge end (Fig. 7). These blocks are designed for easy replacement as they wear.

The pin stud pattern, pin stud length and pin stud diameter must be carefully chosen for its ability to hold the refractory in place and to maximize the heat transfer through the stud to the furnace wall tubes. This, in turn, serves two purposes. One is to provide maximum cooling to keep as low a refractory temperature as possible. Maintaining a low refractory surface temperature has a dramatic effect on refractory life, furnace wall fouling and maintenance costs. Secondly, with more heat removed in this lower furnace area, less heating surface is required in the upper furnace to achieve the desired flue gas temperature leaving the furnace.

RDF-fired units

Prior to the late 1980s, RDF boilers were installed with bare carbon steel tubes in the lower furnace and no corrosion protection. It was thought that with the more even combustion with a processed fuel, corrosion would not be a concern in the lower furnace. Early units, operating at low steam pressure and temperature, did not experience corrosion problems. However, as higher pressure and temperature units went into operation, corrosion increased and lower furnace protection was needed.

The same pin stud and refractory design used on mass-fired units has also been tried on RDF units. This solved one problem but created another. Inherent in the RDF combustion process is a high degree of suspension firing and high flame temperatures in the lower furnace. When pin studs and refractory are applied, the lower furnace tubes are insulated, resulting in less heat transfer and hotter flue gas temperatures in the lower furnace. This, in turn, can result in significant slagging on the refractory wall surface. Pin stud and refractory was tried at two RDF facilities. However, increased furnace slagging resulted, and eventually the pin studs and refractory were removed.

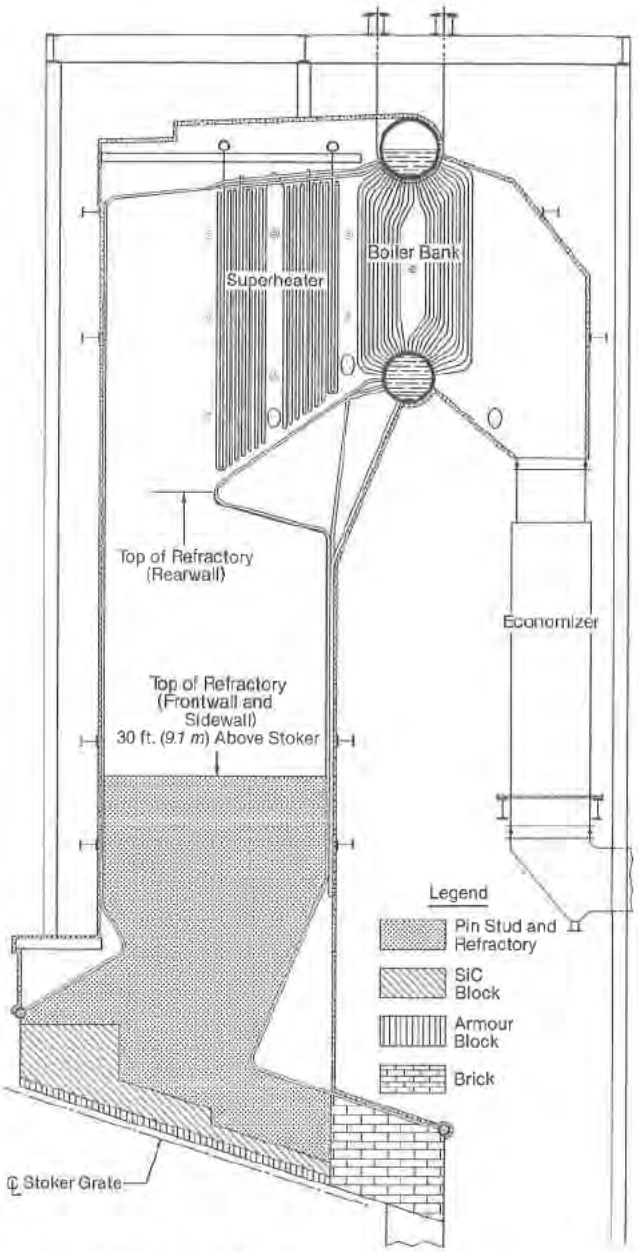


Fig. 7 Refractory type and location, mass-fired unit.

What was needed was a material that was resistant to the chloride corrosion found in refuse boilers while not insulating the lower furnace tubes. Babcock & Wilcox (B&W) pioneered the use of Inconel material as a solution to this lower furnace corrosion problem. In 1986, following rapid corrosion of the bare carbon steel tubes, the lower furnace of the Lawrence, Massachusetts unit was covered with a weld overlay of Inconel material. This overlay proved to be effective in minimizing corrosion in the lower furnace. Based on this early experience the industry followed B&W's lead and Inconel weld overlay was field applied to the lower furnace of a number of operating boilers.

For the RDF boilers supplied as part of the refuse-to-energy plant in Palm Beach County, Florida, the decision was made to add the Inconel protection prior to manufacture. A bimetallic tube construction was used consisting of a carbon steel inner tube co-extruded with an Inconel outer

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tube (Fig. 8). This allows for a more uniform protective coating than does a weld overlay. These boilers went into operation in 1989 and the Inconel bimetallic tubes have experienced no corrosion in their initial years of operation. Today the bimetallic tube is the industry standard for lower furnace corrosion protection in RDF-fired boilers.

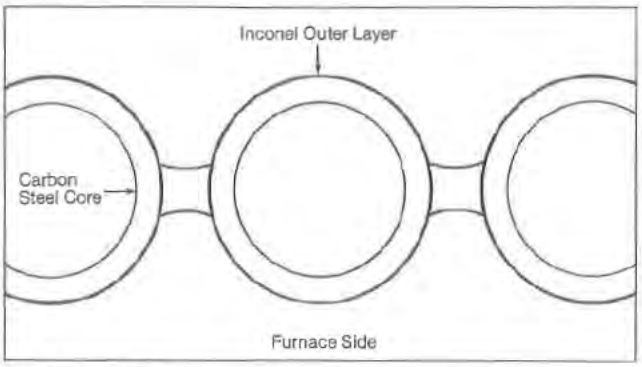


Fig. 8 Bimetallic tube.

Mass burning

Mass burning is the most common refuse combustion technology worldwide. There are more than 1600 mass burn units in operation throughout the world, predominantly in Europe and Japan. When the market for refuse-to-energy facilities expanded rapidly in the U.S. in the early 1980s, many of these refuse plants adopted this well proven mass burning technology (Fig. 9). However, some major differences in the U.S. application resulted in some operational problems. U.S. applications tend to require large units (Fig. 10) to accommodate larger regional facilities instead of the small local and community plants typical of European and Japanese applications. U.S. plants were designed to operate at significantly higher operating pressures and temperatures to take advantage of the economics of production and sale of electric power while typical non-U.S. applications produced hot water and low pressure steam for heating applications. U.S. refuse fuel typically has a higher heating value and lower moisture content. Finally, U.S. units began to be installed at a time when environmental concerns were increasing.

The net result of these characteristics of the market resulted in many early U.S. refuse units experiencing operating problems related to:

1. high rates of slagging in the furnace,
2. higher gas temperature leaving the furnace resulting in overheating of superheaters and excessive fouling in the convection section,
3. tube failures from accelerated corrosion that were metal temperature related, and
4. concerns about the creation of dioxins (Polychlorinated Dibenzoparadioxin, PCDD) and furans (Polychlorinated Dibenzofurans, PCDF) during the combustion process

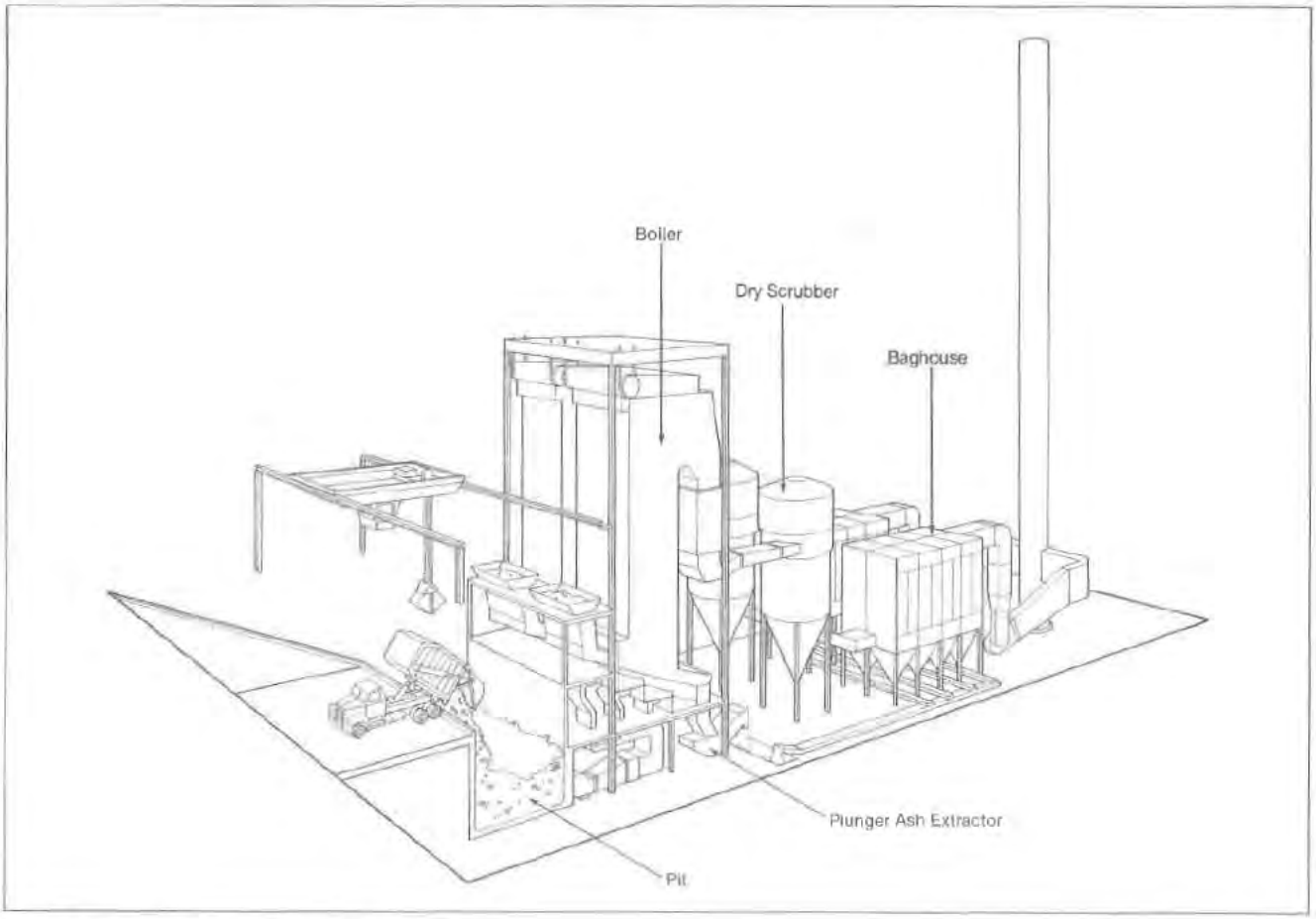


Fig. 9 Typical mass burning refuse-to-energy system.

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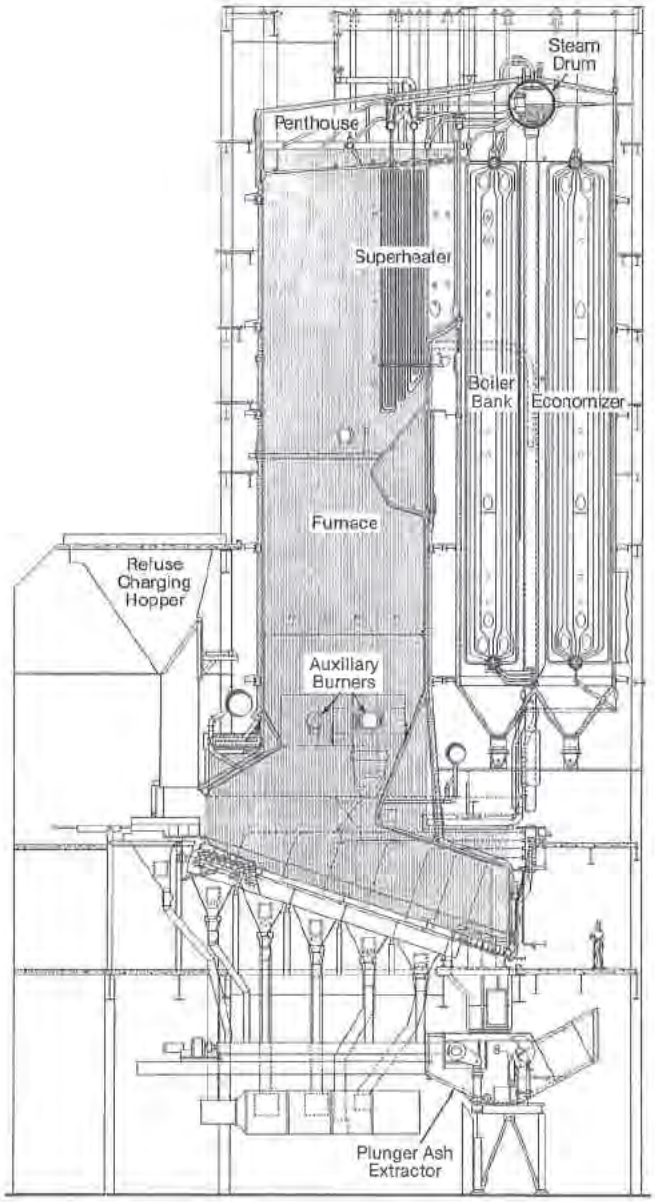


Fig. 10 Typical mass burning unit.

that were related to less than optimum combustion systems, particularly less than optimum turbulence and mixing of fuel and air in the lower furnace.

Boiler plant sizing

A refuse plant must be sized to handle the physical amount of refuse that is delivered to that plant, regardless of the refuse heating value. A refuse boiler, on the other hand, is a heat input device and must be sized for the maximum heat input expected. When designing a refuse boiler you need to know the design tons per day of refuse to be combusted and the typical range of heating values that is expected for the refuse in that location.

The boiler is typically designed for the maximum ton per day input at the maximum refuse heating value. A 1000 t/d (907 t_m/d) refuse plant is actually not the same size plant in all locations. A plant in the Northeast U.S. would typically be designed to handle refuse at heating values as high as 5500 Btu/lb (12,793 kJ/kg), or 458 x 10^6 Btu/h (134 MW_t) total heat input to the boilers. At the other extreme, a plant in a less industrialized country would be designed to handle refuse with a heating value in the range of 3500 Btu/lb (8140 kJ/kg), or 292 x 10^6 Btu/h (85.6 MW_t) total heat input. Both are 1000 t/d (907 t_m/d) plants, but one has refuse boilers that have a 50% larger capacity.

For many of the early U.S. plants, good data were not available on the true range of heating values of the refuse. Refuse boilers were sized for typical heating values of 4500 Btu/lb (10,467 kJ/kg). When the actual heating values were found to be as high as 5200 to 5500 Btu/lb (12,095 to 12,793 kJ/kg), the boilers were actually undersized and could not process the available refuse on a ton per day basis.

Stoker capacity

A refuse stoker has both a heat input limit and a ton per day of refuse limit. If a typical 1000 t/d (907 t_m/d) refuse plant has two 500 t/d (454 t_m/d) boilers, and the design refuse heating value is 5000 Btu/lb (11,630 kJ/kg), each boiler would have a maximum heat input limit of 208.3 x 10^6 Btu/h (61.1 MW_t). If the actual refuse heating value is above 5000 Btu/lb (11,630 kJ/kg), the maximum heat input limit can not be exceeded and therefore the unit's t/d capacity would be reduced below 500 t/d (454 t_m/d). On the other hand, if the actual refuse heating value is below 5000 Btu/lb (11,630 kJ/kg), then the unit could actually process more than 500 t/d (454 t_m/d) of refuse, up to the maximum ton per day limit of that stoker.

The ton per day limit is usually set by a refuse capacity per unit of width, a limit for optimum fuel feed and distribution, or a weight per square foot (m^2), a structural limit. These limits are in the range of 30 t/d (27 t_m/d) per front foot (0.3 m) of width and 65 lb/h ft^2 (2.74 kg/h m^2) of grate area. The grate surface area is set by a grate release rate generally in the range of 300,000 to 350,000 Btu/h ft^2 (946,350 to 1,104,080 W/m^2), but may be lower for low heating value, high moisture fuels. The stoker width and depth are also related to the specific fuel.

A high heating value, low moisture fuel would require a wider, less deep stoker because the fuel will tend to burn more rapidly. A low heating value, high moisture fuel would require a narrow, deeper stoker because more residence time on the stoker is usually needed. The combination of all these criteria will set the maximum ton per day rating of the stoker.

There is also a minimum load that can be effectively handled on a given stoker. This load is also set by both a ton per day limit (minimum fuel inventory on the grate) and a heat input limit (minimum heat input for good combustion).

All of these limits can be incorporated into a capacity diagram, which provides the operator with the boundary limitations around a family of heating value curves. Fig. 11 is such a diagram for a typical 500 t/d (454 t_m/d) boiler burning 5000 Btu/lb (11,630 kJ/kg) refuse.

Stoker design

The combustion of MSW requires a rugged, reliable stoker to successfully convey and burn unsorted refuse. Most stokers use some variation of a reciprocating grate action, with either forward moving or reverse acting grate movement. Some arrangement of moving and stationary grates is used to move the refuse through the furnace and allow time for complete combustion.

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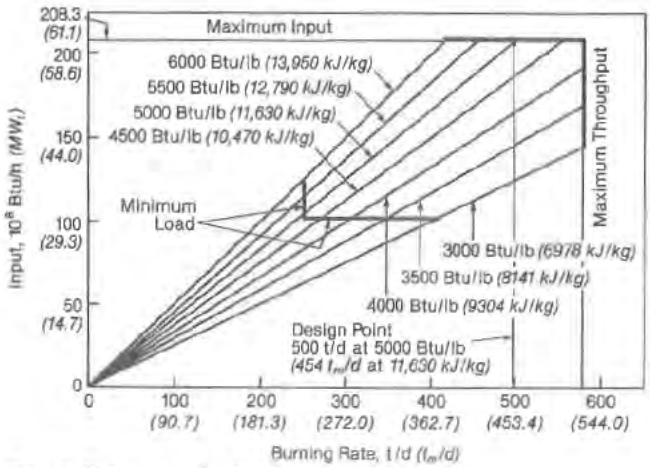


Fig. 11 Stoker capacity diagram.

The stoker illustrated in Fig. 12 is typical of the forward moving reciprocating grate stoker. This grate is designed with alternate moving and stationary rows of grates in a stairstep construction with a downward slope to help move the refuse through the furnace. Each row of grates overlaps the row beneath it and the alternate rows are supported from a moving frame driven by hydraulic cylinders (Fig. 13). These grates move the refuse over the stationary grates, where it is picked up by the next row of moving grates and moved through the furnace. The action of these reciprocating grates rolls and mixes the refuse, constantly exposing new material to the high temperatures in the bed and allowing the combustion air to contact all the burning refuse.

For low heating value, high moisture refuse, drop off steps are often incorporated into the stoker design. The steps are located at the end of each grate module resulting in one to three steps depending on the overall stoker length. These steps promote a tumbling and rolling action as the burning refuse falls off the step. This type of design was used on a number of the early refuse units in the U.S. but was found to be unnecessary with the higher heating value and lower moisture refuse. In fact, it can be a detriment as the tumbling can also result in excursions of high carbon monoxide (CO) emissions.

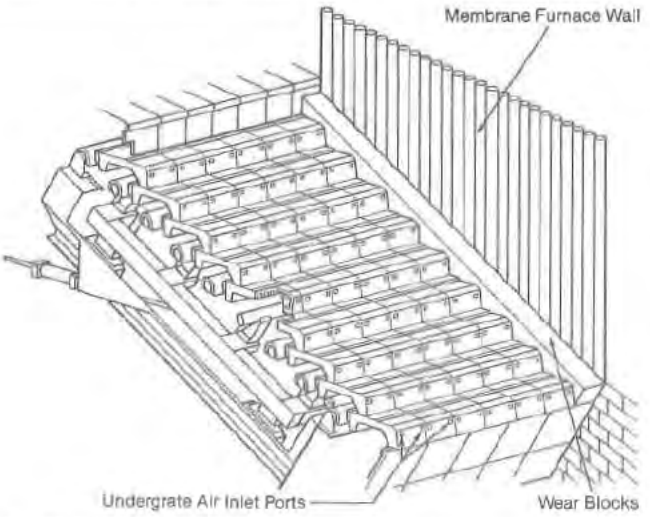


Fig. 12 Forward moving reciprocating grate stoker *(courtesy von Roll Inc.)*.

The grates are usually constructed in a series of standard modules with independent drives and air plenums. This allows the individual grate modules to be factory-assembled to limit field construction time and provide complete duplication of parts for easy maintenance and repair. This modular construction allows any size stoker to be constructed from a small number of standard modules. A typical stoker is usually from two to four modules in length and one to four modules in width. This method of construction also allows for complete zoned undergrate air control to the individual burning areas of the grate and provides complete freedom in the operating speed of the individual grate modules to provide the required feed rate along the grate for complete burnout of the fuel. This ability to control the undergrate air in multiple air zones along the width and depth of the stoker is an important factor in minimizing CO and nitrogen oxides (NO_x) emissions.

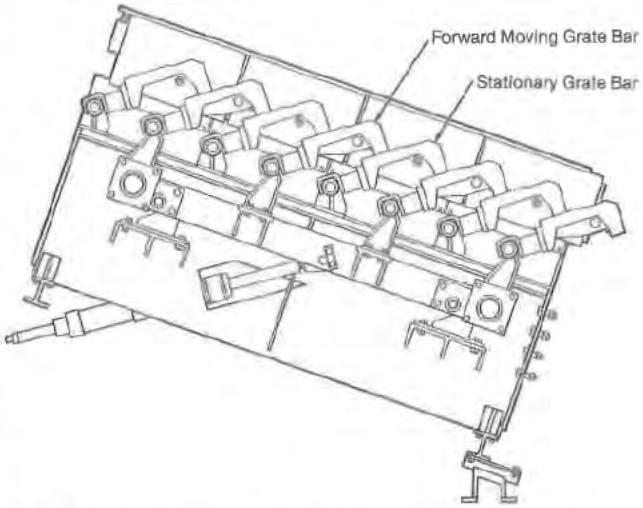


Fig. 13 Reciprocating grate, longitudinal section *(courtesy von Roll Inc.)*.

Fuel handling

The MSW delivered to the refuse plant is generally dumped directly into the storage pit. This large pit also provides a place to mix the fuel. This mixing is done using the crane and grapple to move and restack the refuse as it is dumped into the pit. This produces a fuel, in both composition and heating value, as consistent as possible for the boilers. This is an essential job for the crane operator and any time not required to feed the furnaces is used to mix the fuel. It is not uncommon for the crane operator to mix four grapple loads back into the pit for every one load that goes to the charging hopper.

Fuel feed system

Controlled feed of the fuel is necessary for good combustion to minimize CO and NO_x emissions and to maintain constant steam output. At the bottom of the charging hopper feed chute, a hydraulic ram pushes the fuel into the furnace and onto the stoker grates at a controlled rate. On larger units, multiple charging rams are used across the width of the unit to provide a continuous fuel feed with optimum side to side distribution. The hydraulic rams stroke forward slowly and then retract quickly to provide the positive continuous fuel feed. These rams are simple to

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control with feed rate adjustments made by either the speed of travel or the number of strokes per hour.

Combustion air system

The primary combustion air, or undergrate air, is fed to the individual air plenums beneath each grate module. A control damper at the entrance to each air plenum controls the undergrate air to each section of the grate (Fig. 14). The grate surface is designed to meter the primary combustion air to the burning refuse uniformly over the entire grate area. This is accomplished by providing small air ports or tuyères in the surface of the individual grate bars. These air ports provide openings equal to approximately 3% of the grate area which results in sufficient pressure drop of air resistance across the grate to assure good distribution of the air flow through the grate, regardless of the depth of refuse on the grate. Undergrate air systems are generally designed for 70% of the total air to be undergrate air with expected normal operation of 60%.

Because refuse contains a high percentage of volatiles, a large portion of the total combustion air should enter the furnace as secondary, or overfire, air through the furnace walls. These secondary air ports are located only in the front and rear furnace walls so that the air flow parallels the normal flow pattern through the unit. Older design units generally provided 25 to 30% of the total air as overfire air. With today's emphasis on better combustion and lower emissions, the overfire air systems are designed for 50% of the total air to be overfire air with expected normal operation at 40%.

The basic function of the overfire air is to provide the quantity of air and the turbulence necessary to mix the furnace gases with the combustion air and to provide the oxygen necessary for complete combustion of the volatiles in the lower furnace. Excess air in the furnace is usually maintained in the range of 80 to 100% and complete combustion is demonstrated by a CO value in the furnace of 100 ppm or less.

To aid in the combustion of wet fuels during extended periods of rainy weather, the air system includes steam coil air heaters designed to provide air temperatures in the range of 300 to 350F (149 to 177C) to help dry these fuels and maintain furnace temperature. These steam coil air heaters are commonly used only for the undergrate air because this is the air flow which directly aids in drying wet fuel. These air heaters must be conservatively designed with fin spacing not exceeding 4 to 5 fins/in. (1 fin/6.4 to 5.1 mm). Most plants take combustion air from the storage pit area to help minimize odors. This air is normally contaminated with dust and lint which could plug the steam coil. Some type of cleaning arrangement or filters must therefore be included to keep the steam coil clean.

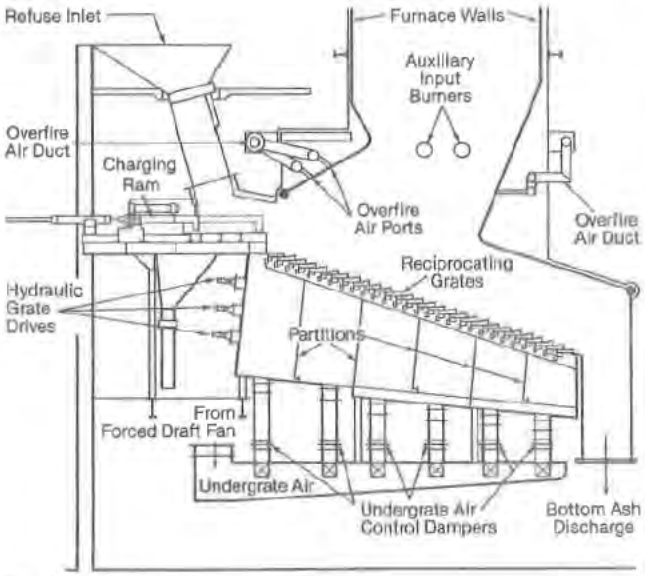


Fig. 14 Combustion air system.

When high moisture fuels are encountered, the first action by the operator is to use the steam coil air heater to provide hot air. However, for very high moisture fuels it may also be necessary to use the auxiliary fuel burners to stabilize combustion in the furnace. These cases are the exception, and in normal operation neither the auxiliary burners nor the steam coil air heater are needed for good combustion.

Ash handling systems

When refuse is burned, the ash takes the form of either light ash, called flyash, or coarse ash, which comes off the stoker. The flyash is entrained in the gas stream until it is removed in the particulate collection device or falls out into the boiler, economizer, or air heater hoppers. The stoker ash consists of ash from the fuel, slag deposits on the grate, ash from the furnace walls and, in some designs, ash from the superheater. The stoker ash is discharged through the stoker discharge chute and from the stoker siftings hoppers.

Plunger ash extractor

The ash from the stoker discharge on mass-fired units may contain large pieces of noncombustible material, in addition to the normal ash from combustion. Ash consistency can vary from fine particles to large and heavy noncombustible objects in the fuel. The ash from the stoker discharge chute falls into a water bath in the plunger ash extractor (Fig. 15) that quenches the ash and controls dusting. After the ash is quenched, a slow moving, hydraulically operated ram cycles forward and back to push and squeeze the accumulated ash up an inclined dewatering section to the discharge of the extractor. The ram cycle continues at a slow speed to push the ash out of the extractor. The dewatered ash has a moisture content of 15 to 20% as a result of the squeezing process on the incline. The lower moisture content can have an economic advantage as the cost of landfilling ash is based on total weight, which includes the weight of water in the ash.

To keep the ash system simple and to minimize costs, ash from the extractor can discharge directly into a truck or bin for final disposal. To move the ash away from the vicinity of the stoker discharge, vibrating and belt-type conveyors are used. A short vibrating conveyor is placed at the discharge of the ash extractor; its metal trough can take the impact of the oversized, noncombustible material falling from the extractor. The vibrating conveyor then transfers the ash to a belt conveyor, minimizing the wear on the belt conveyor that would occur if the extractor discharged directly onto it.

Double gate ash hopper

Double gate ash hoppers can be used (Fig. 16) in lieu of an ash extractor. The hopper has gates on the top and

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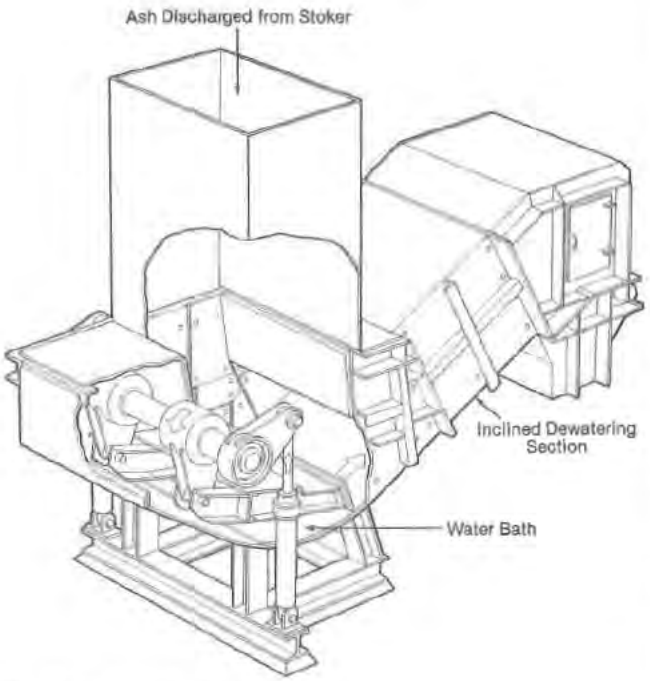


Fig. 15 Plunger ash extractor.

the bottom to control the flow of ash entering and discharging from the hopper. Water spray nozzles help quench the ash in the hopper. The ash can be discharged directly into a truck or onto a conveyor for disposal. This system is simple with low capital cost.

Scrubber, precipitator and baghouse flyash

The flyash collected in the scrubber, precipitator or baghouse hoppers can be handled by dry mechanical screw or chain-type conveyors. These units operate continuously to minimize hopper pluggage problems and they discharge onto a collecting conveyor, which is usually a dry chain type. Because the mechanical conveyors are dust-tight, but not designed to be gas-tight, separate sealing devices, such as rotary seals or double flop valves, are used. The collecting conveyor will collect the flyash discharged by all the conveyors under the rows of hoppers, and move it to a single collection point for ultimate disposal.

RDF firing

RDF technology was developed in North America as an alternative to the mass burning method. Initially, RDF was used as a supplementary fuel for large, usually coal-fired, utility boilers. For this application the RDF was finely processed and sized to 1.5 in. (38.1 mm) maximum size. The resulting RDF was nearly all light plastics and paper.

For supplemental firing, B&W guidelines call for a maximum RDF input of 20% on a heat input basis and no RDF input until the boiler is operating above 50% load. In most cases the RDF is blown into the furnace sidewalls at the pulverized coal burner elevation through an RDF burner with a fuel distribution impeller. Most of the RDF burns in suspension in the high heat input zone of the pulverized coal fire. However, some of the heavier fuel fraction falls out in the lower furnace. Dump grate stokers located in the neck of the ash hopper allow more complete burnout of these heavier pieces before they are discharged into the ash system (Fig. 17). RDF has been successfully co-fired in B&W boilers at Lakeland, Florida, Ames, Iowa, and Madison Gas & Electric in Wisconsin.

RDF has also been successfully co-fired in B&W Cyclone furnaces where the finely processed and sized RDF is injected into the Cyclone secondary air stream moving tangentially inside of the Cyclone barrel. (See Chapter 14.) This method of RDF combustion is used at the Baltimore Gas & Electric Company's Crane station in Maryland.

Dedicated RDF-fired boilers

From this supplemental fuel experience, RDF then became the main fuel for boilers specifically designed to generate full load steam flow when burning RDF (Fig. 18). In some cases where steam flow was required even when refuse was not available, the boiler was designed so that it could also reach full load on wood, coal or natural gas. More commonly there would only be auxiliary gas or oil burners for startup and shutdown.

The first boilers in the world to fire RDF as a dedicated fuel were B&W units which began operation in 1972 in Hamilton, Ontario, Canada. The first of such boilers in the U.S. went into operation in 1979 in Akron, Ohio. The boiler design was highly influenced by the proven technology of wood-fired boilers with respect to their fuel feed system, stoker design, furnace sizing and overfire air system. The transfer of this technology from wood firing to RDF firing was successful in many areas, but in other areas design adjustments were needed to accommodate the unique aspects of RDF.

The operating experience from the first generation designs at Hamilton, Akron and other plants led to second generation designs with improved RDF processing systems, fuel feed systems and boiler design. Specific improvements included the first fuel feeder designed specifically for RDF and the use of alloy weld overlay in the lower furnace for corrosion protection.

The third generation of facilities is essentially today's state-of-the-art design (Fig. 19). This boiler design has a unique lower furnace arch arrangement and an enhanced

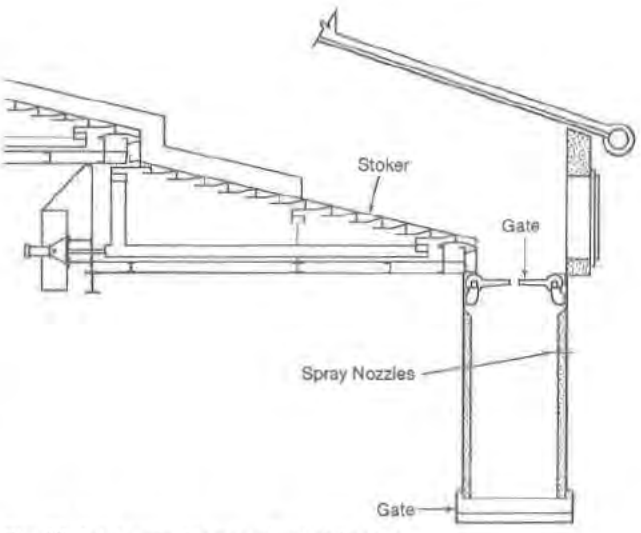


Fig. 16 Ash hopper for mass-fired stoker ash.

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overfire air system to significantly improve combustion efficiency. These third generation designs also incorporate dramatically improved fuel processing systems.

RDF preparation systems

The first generation RDF processing systems were crunch and burn systems. The incoming refuse first went to a hammermill type shredder that produced an RDF with 6 x 6 in. (152 x 152 mm) top size. Ferrous metal was removed by magnetic separators. There was no other material separation and many undesirable components entered the boiler. Shredded particles of glass were embedded in wood and paper resulting in a very abrasive fuel entering the boiler in suspension. Also, the RDF was generally stored in a hopper or bin. RDF is compactible and in nearly every case, significant problems were encountered getting the RDF out of the storage bins.

Second generation RDF processing systems recognized and corrected some of the problems. The shredder for final fuel sizing was moved to the back of the processing system and some type of rough sizing shredder was used as the first piece of equipment in the system. This reduced, but did not eliminate, the problem of abrasive particles embedding in the fuel. Some size separation equipment was introduced, generally removing the small size fraction which is less than 1.5 in. (38.1 mm) composed mostly of broken glass, ceramics and dirt, which was sent to landfill. RDF was stored on the floor rather than in bins or hoppers and was moved by front-end loaders to conveyor belts. This greatly improved the reliability of fuel flow to the boiler.

In third generation RDF processing systems (Fig. 20) the first piece of equipment became a flail mill or similar equipment whose main function was to break open the garbage bags. The refuse was still size separated using a trommel or disk screen with the minus 1.5 in. (38.1 mm) size destined for landfill. Generally, a device such as an air density separator was added to remove the light fraction (paper, plastics, etc.) from this stream to achieve maximum heat recovery. Where it was economically attractive, aluminum separation was added to the plus 1.5 in. (38.1 mm.) minus 6 in. (152 mm) stream.

RDF yield

The ash content of the RDF is directly related to the yield of the processing system or the percentage of RDF produced from a given quantity of MSW. A 70% yield means that 70 t of RDF is produced for every 100 t of incoming MSW. In a processing system with a lower yield, the portion of the

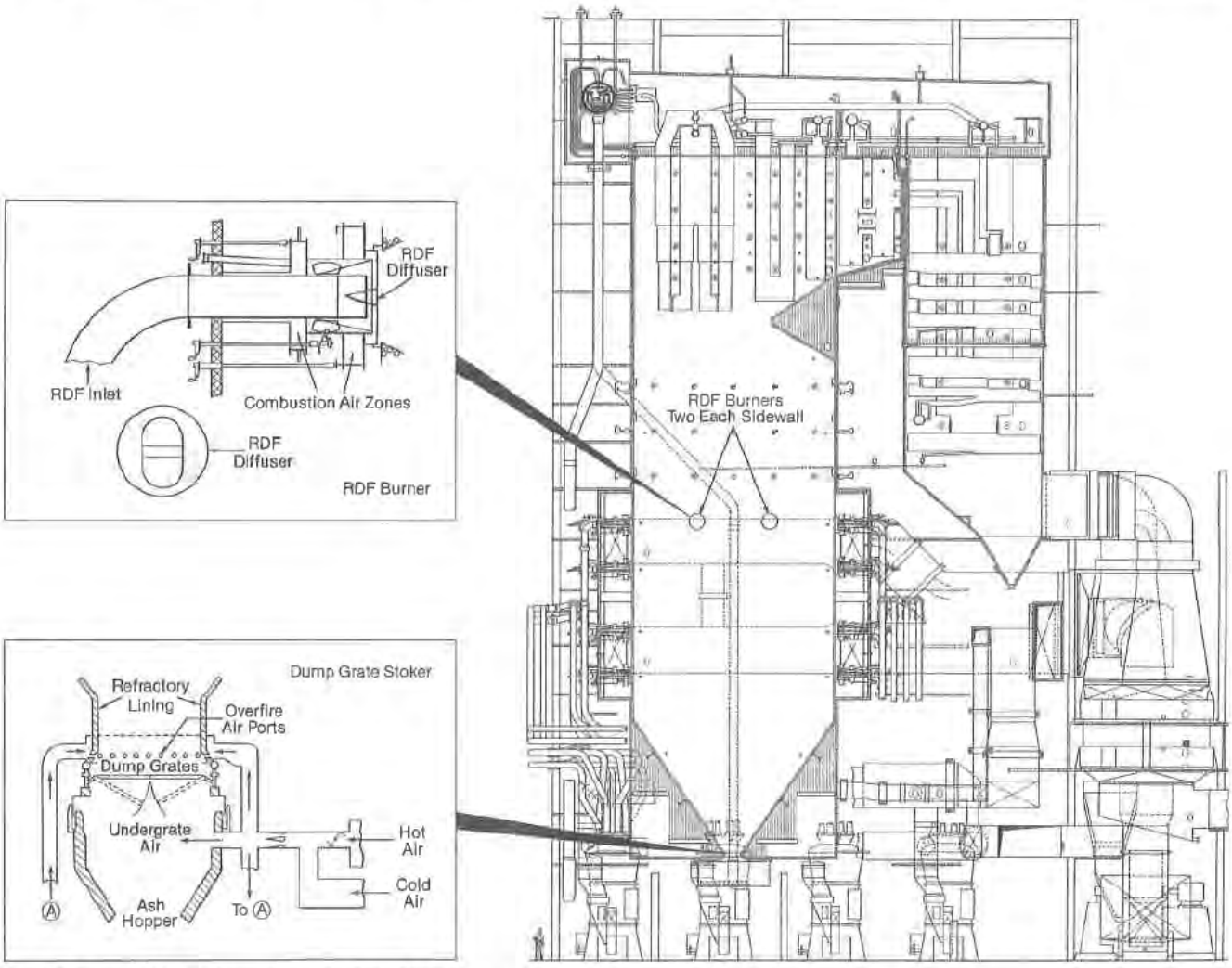


Fig. 17 Typical B&W RB type utility boiler firing RDF as a supplementary fuel

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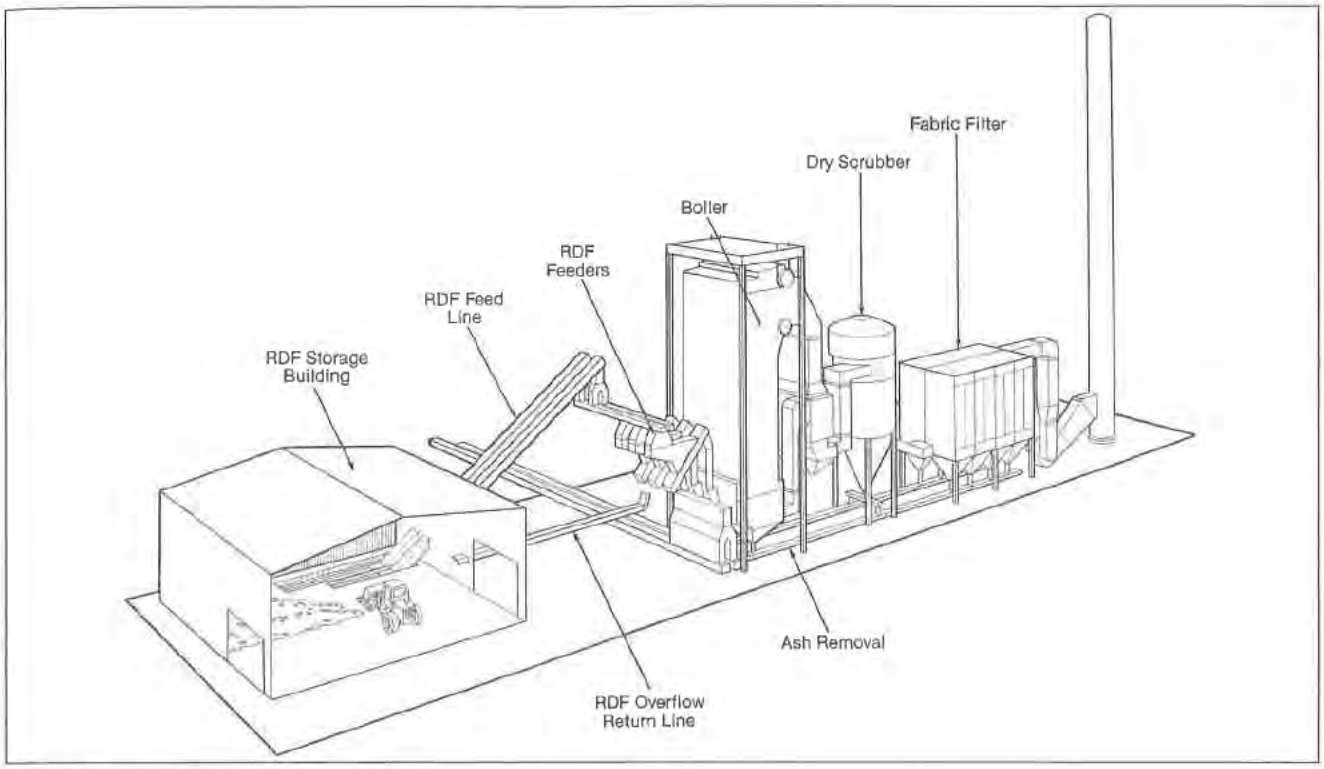


Fig. 18 Typical RDF refuse-to-energy system.

MSW that is rejected is generally high in ash and inerts content and, therefore, the resulting RDF is low in ash content. As the RDF processing system is designed to obtain a higher yield, more of the ash is carried over into the RDF fuel fraction and the ash content is increased.

The RDF heating value is inversely related to the yield; the higher the yield, the lower the heating value. In a high yield system, most of the rejects are ferrous metals and inerts (glass, ceramics, dirt) which have no heating value. While some fuel is also rejected, the quantity is small resulting in a higher net heating value for the RDF. A yield of about 93% represents a crunch and burn type system in which only the ferrous metal is removed. In such a system, the RDF heating value is only marginally higher than the heating value of the incoming MSW.

A typical MSW might have a composition comparable to the reference waste shown in Table 4. The majority of the waste is combustible materials, which have ash contents ranging from approximately 4% for wood to 12% for glossy magazine paper. The glass fraction, yard waste and mixed combustibles may also contain varying quantities of sand, grit and dirt. The predicted composition of the RDF will vary depending on the type of processing system and the resulting yield. Table 5 shows how the ash content in the fuel and the heating value of the fuel will vary as the RDF yield varies with different processing systems. Two cases are considered, one which assumes no front end recycling (curbside recycling or separate recycling facility), and a second case which assumes that such a system is in place in the community.

RDF quality

RDF used to supplement pulverized coal in utility boilers should be low in ash; have minimum ferrous metal, aluminum, and other nonferrous metal; and be small enough in particle size to be fed pneumatically to the boiler. The processing system for such a fuel would generally be a very low yield system, between 40 and 60%.

RDF for dedicated traveling grate stoker boilers should

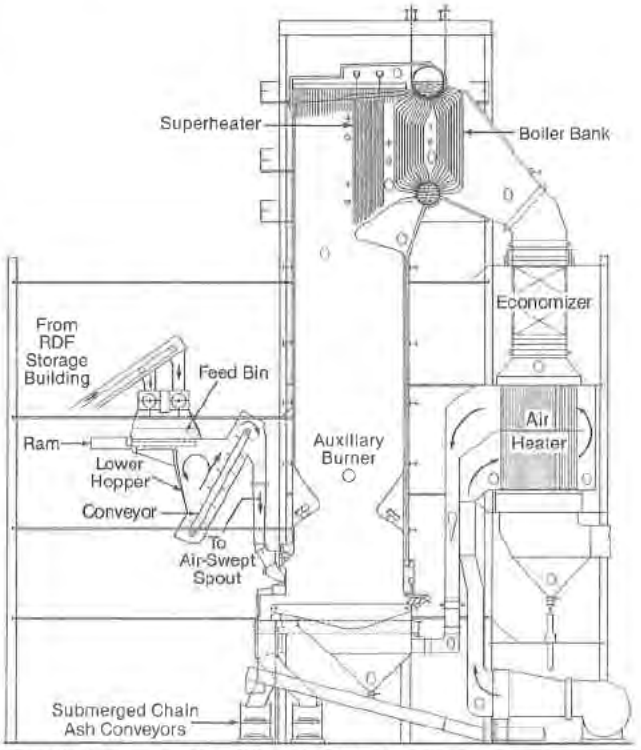


Fig. 19 Third generation RDF unit and fuel feeding system.

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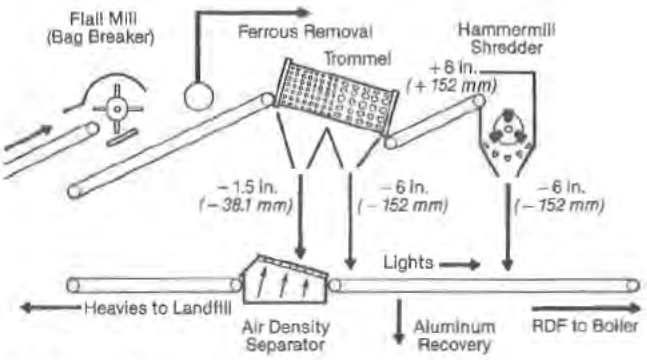


Fig. 20 RDF processing system.

be low in ash, consistent with a high RDF yield from the MSW; as free as possible of ferrous metal, aluminum and other nonferrous metals; and of a particle size distribution that is considerably larger than the particle size of RDF for use in Cyclone or pulverized coal boilers. The processing system for such a fuel will be a higher yield, around 70 to 85%.

RDF produced in a crunch and burn system, in which solid waste is shredded and only the ferrous metal removed, has a yield of about 93%; an inherently high ash content; and contains 100% of the aluminum, other non-ferrous metals, glass, stones and ceramics in the original MSW. While this is a high yield system which is desirable for a dedicated traveling grate stoker boiler, it also contains large quantities of aluminum, glass, and other inerts which results in higher wear on the stoker and lower furnace. Another result is a more conservatively designed stoker and furnace as well as a larger ash handling system.

RDF processing systems

An optimum RDF processing system for a dedicated boiler application achieves the highest yield with the highest heat recovery, while removing ferrous, aluminum and glass before entering the boiler. Such a system (Fig. 21) includes the following:

In-feed conveyors From the tipping floor, the solid waste is fed by front-end loaders to steel pan apron conveyors which feed the flail mill in-feed conveyors.

Initial size reduction The flail mill tears open the plastic garbage bags, coarsely shreds the refuse, and also breaks glass bottles to a size of approximately 1.5 in. (38.1 mm) or less.

Ferrous metal recovery Ferrous metal is extracted from the coarsely shredded MSW in each line by a single-stage overhead magnet. Recovered ferrous metal is moved to a ferrous air classifier where tramp materials such as paper, plastics or textiles are removed, thereby providing a clean ferrous product. A ferrous recovery of 90% is possible.

Size classification and final size reduction After ferrous removal, shredded waste is fed into a rotating trommel screen, a size separating device about 10 ft (3.0 m) in diameter by 60 ft (18.3 m) long. The trommel performs the following functions:

1. removes glass, sand, grit and nonferrous metal less than 1.5 in. (38.1 mm) in size, and
2. removes the minus 6 in. (152 mm), plus 1.5 in. (38.1 mm) fraction which is the proper fuel size without additional processing and contains the bulk of the aluminum cans.

The trommel oversize material, plus 6 in. (152 mm), is then shredded in a horizontal secondary shredder. Because the secondary shredder is a major consumer of energy and has high hammer maintenance costs, the RDF process is specifically designed to reduce the secondary shredder's load by shredding only those combustibles too large for the boiler. Particle size is controlled with a disk screen which recycles oversize material back to the secondary shredder.

Separation of glass, stones, grit and dirt Trommel undersize material, minus 1.5 in. (38.1 mm), passes over an air density separator (ADS) designed to remove dense particles from less dense materials through vibration and air sweeping. This device can efficiently remove glass, stones, grit and dirt, as well as nonferrous metals. The light fraction, which can range from approximately 50 to 90% of the ADS feed, consists essentially of combustibles with high fuel value which are recovered and blended into the main fuel stream.

Aluminum can recovery To optimize aluminum can recovery, an air classifier is provided for the plus 1.5 in. (38.1 mm), minus 6 in. (152 mm) undersize fraction. The air classifier removes the light organic portion of the stream, allowing aluminum cans to be more visible for hand pickers. The air classifier heavy fraction drops onto a conveyor moving at approximately 2.5 ft/s (0.76 m/s) with numerous hand-picking stations on either side of the belt. Cans

Table 4
Typical Reference Refuse

Component Analysis	Reference MSW (% by wt)	Reference RDF (% by wt)
Corrugated board	5.53	—
Newspapers	17.39	—
Magazines	3.49	—
Other paper	19.72	—
Plastics	7.34	—
Rubber, leather	1.97	—
Wood	0.84	—
Textiles	3.11	—
Yard waste	1.12	—
Food waste	3.76	—
Mixed combustibles	17.75	—
Ferrous	5.50	—
Aluminum	0.50	—
Other nonferrous	0.32	—
Glass	11.66	—
Total	100.00	
Ultimate Analysis		
Carbon	26.65	31.00
Hydrogen	3.61	4.17
Sulfur	0.17 (max. 0.30)	0.19 (max. 0.36)
Nitrogen	0.46	0.49
Oxygen	19.61	22.72
Chlorine	0.55 (max. 1.00)	0.66 (max. 1.20)
Water	25.30	27.14
Ash	23.65	13.63
Total	100.00	100.00
Heating value	4,720 Btu/lb (10,979 kJ/kg)	5,500 Btu/lb (12,793 kJ/kg)
Fuel value recovery, % MSW	96	
Mass yield, % RDF/MSW	83	

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Table 5
RDF Yield versus Ash Content and Fuel Heating Value

Mode	RDF Yield %	Ash	Btu/lb (kJ/kg)
Without front-end recycling:			
Mass burn	100%	23.64	4,814 (11,197)
Crunch and burn	93%	19.87	5,146 (11,970)
RDF	83	11.72	5,641 (13,121)
	to 70%	to 8.87	to 5,834 (13,570)
With front-end recycling:			
Mass burn	100%	19.58	5,513 (12,823)
Crunch and burn	93%	17.16	5,898 (13,714)
RDF	85	9.91	6,328 (14,719)
	to 71%	to 6.59	to 6,491 (15,098)

go into hoppers and, by conveyor, to a can flattener. A pneumatic conveyor then transfers the flattened cans into a trailer. An eddy current separator, for the removal of aluminum cans, can replace hand picking if the expected amount of cans is high enough to justify the additional capital cost. Aluminum recovery of 60% is possible with hand picking or the eddy current separator.

Oversized bulky waste (OBW) The OBW shredder is generally a horizontal hammer mill used to shred ferrous metal recovered by the RDF processing lines and preseparated oversized material which includes white goods such as refrigerators and washing machines, furniture and tree limbs. The ferrous metal is magnetically recovered and given a final cleaning by an air scrubber to remove tramp materials. The nonferrous material is integrated into the RDF stream.

Tire shredding line If there is a sufficient supply of tires, a separate tire shredding line can be included. A shear shredder, used specifically for shredding tires, can shred 500 passenger car tires per hour. The shredder includes a rotary screen classifier (trommel) for returning shredded tire chips above 2 in. (51 mm) back to the shredder. A tire chip 2 x 2 in. (51 x 51 mm) or less is the final product which is then blended in with the RDF stream.

RDF storage building RDF from each processing line is conveyed to an RDF storage building. From there, it is either fed directly to the boiler or fed directly to a shuttle conveyor and storage pile. When RDF feed is direct to the boiler, excess RDF from the boiler feed system is returned to the RDF storage building. RDF not being fed directly to the boiler is retrieved from the storage pile by a front-end loader and loaded onto inclined conveyors which transport the RDF to the boiler feed system.

Fuel feed system: metering feeders

A successful RDF metering feeder must meet the following design criteria:

1. controlled metering of fuel to meet heat input demand,
2. homogenization of material to produce even density,
3. liberal access to deal with oversized material problems,
4. maintainability, in place, and
5. fire detection and suppression devices.

A reliable RDF metering feeder (see Fig. 19) is a key feature of the second generation RDF boiler design. One feeder is used for each air-swept fuel distributor spout. Each feeder has an upper feed bin which is kept full at all times by an over-running conveyor to ensure a continuous fuel supply. The fuel in this hopper is transferred to a lower hopper by a hydraulic ram. The ram feed from the upper hopper is controlled by level control switches in the lower hopper. The RDF is fluffed into a uniform density by a variable speed inclined pan conveyor which sets up a churning motion in the lower hopper. The pan conveyor delivers a constant volume of RDF per flight which is carried up the pan conveyor and deposited into the air-swept spout. The rate at which the fuel is deposited into the spout is based on fuel demand.

Air-swept distributor spouts

Air-swept fuel spouts, used extensively in the pulp and paper industry, proved to be equally effective for RDF firing. (See Chapter 26.) Lateral fuel distribution by multiple spouts across the width of the furnace delivers fuel evenly over the grate. Longitudinal distribution is accomplished by continuously varying the pressure of the air sweeping the spout floor. A major feature of this design is its simplicity.

Traveling grate stoker

To date, only traveling grates have been used for spreader-stoker firing of RDF. These grates move from the rear of the furnace to the front, into the direction of

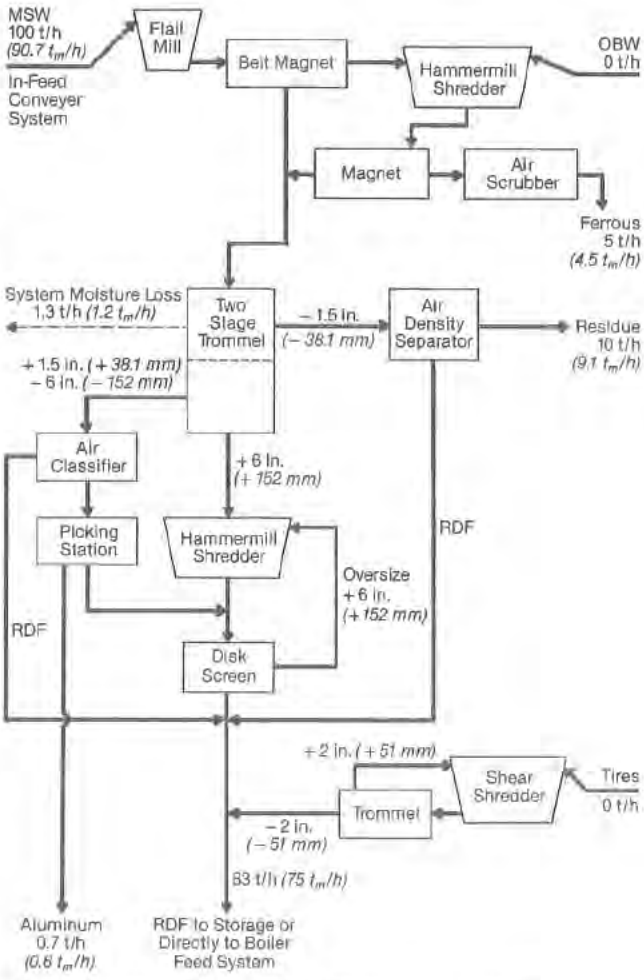


Fig. 21 Components of complete RDF processing system.

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fuel distribution. A single undergrate air plenum is used. There is a wealth of experience worldwide with traveling grate stokers burning a myriad of waste and hard to burn fuels. The parameters for unit design shown in Table 6 were developed from this experience and the uniqueness of the RDF. (See also Chapter 15.)

On mass-fired stokers, a large volume of fuel at the front slowly burns down to a small volume of ash at the back. For an RDF stoker the key is to maintain an even 8 to 10 in. (203 to 254 mm) bed over the entire stoker area. Grate problems are usually due to a shallow ash bed. Operator tendency, when confronted with poor metering and/or fuel distribution, is to run the grates faster. While this technique can minimize bed upset, it will shorten grate life due to higher wear rates and the overheating of the grate bars. With the recommended ash bed thickness, tramp material is minimized, grate temperatures are lowered, wear is reduced and grate life is increased. To achieve this optimum ash bed requires controlled metering of the RDF and proper distribution of the fuel to the grates, as previously described.

A second problem is the accumulation of melted aluminum. The best solution is total removal of the aluminum from the fuel stream. If this is not practical, experience has shown that maintaining proper ash bed thickness will cause the aluminum to solidify in the ash bed rather than on the grate.

Fluidized-bed combustion, both circulating- and bubbling-bed designs, has been considered for RDF. For the high furnace pressure circulating beds, feeding the compressible RDF is a concern. For the bubbling beds, the feed system would be the same as for a stoker unit. In either case, RDF can be properly combusted. For coal firing, the inherent advantage of a fluid bed is the in-bed SO_2 capture. In-bed HCl capture, however, can not be achieved at the required removal efficiencies and therefore a back end scrubber is still required. A fluid bed boiler can achieve slightly lower NO_x emission levels and as emission requirements become more stringent, fluid bed combustion may become a viable alternative for the combustion of RDF.

Lower furnace design configuration

The lower furnace designs of early RDF boilers were largely based upon technology used for wood-fired boilers. This included modest overfire air (OFA) systems with multiple small diameter nozzles designed for 25 to 30% of the total air supply, straight wall furnaces and carbon re-injection systems. The result was less than desired combustion performance due to inadequate turbulent mixing in the furnace. Today's RDF units are designed with fewer, large diameter OFA nozzles designed for 50% of the total air supply with nominal operation at 40%.

In addition, B&W adapted its proven controlled combustion zone (CCZ™) lower furnace design for RDF firing. Developed originally in the early 1970s for very high moisture wood firing, this design consists of twin arches in the lower furnace with the overfire air nozzles directed down into the lower furnace from the arches (Fig. 22). The CCZ™ design for RDF applications without a carbon re-injection system has achieved lower unburned carbon loss than earlier designs which required the use of carbon re-injection systems.

The current B&W state-of-the-art RDF boiler system is exemplified by the Palm Beach County, Florida refuse facility which began operation in 1989. (See Fig. 23.) This design is currently used for all new B&W RDF boilers.

Table 6
Stoker Design Criteria (English units)

Parameters	RDF	Wood
Grate heat release, 10^6 Btu/h ft^2	0.750	1.100
Input per ft of grate width, 10^6 Btu/h	15.5	29
Fuel per in. of distributor width, lb/h	450	1,000
Feeding width as % of grate width	45 to 50	45 to 50
Grate speed, ft/h	25	N/A

Furnace exit gas temperature

Gas temperatures leaving the furnaces of first and second generation designs were higher than anticipated. There were not enough data on RDF firing to accurately predict the relationship between furnace surface area and furnace exit gas temperature. Compounding this problem was a continual increase in the heating value of the RDF due to changes in the composition of the raw refuse and the development of more efficient processing equipment. To achieve the desired furnace exit gas temperatures, the size of the third generation furnace has increased significantly. The furnace width and depth are set by the size of the stoker, therefore the furnace height has increased to achieve the required furnace exit gas temperatures.

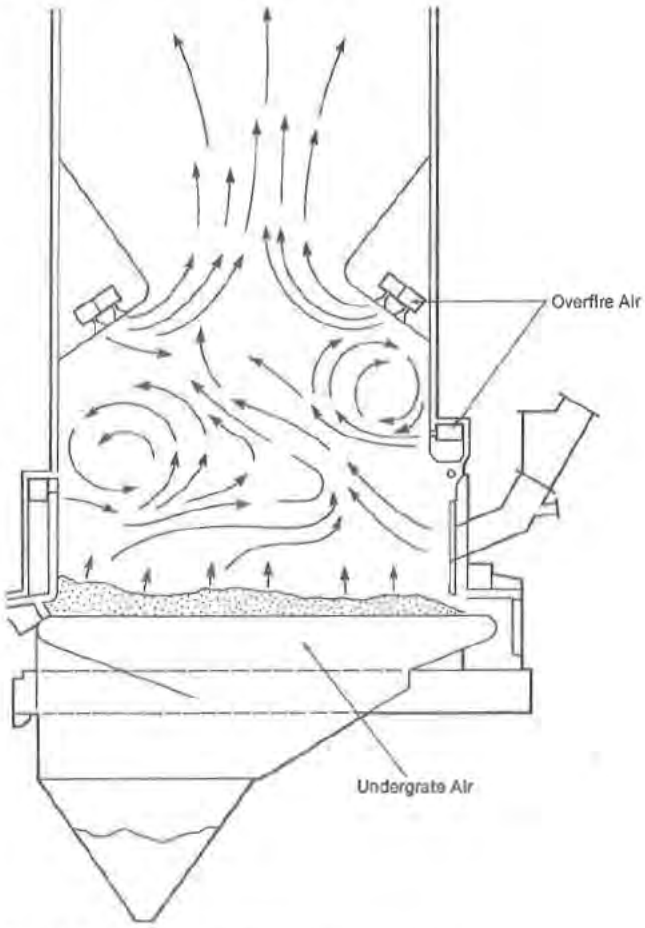


Fig. 22 Controlled combustion zone lower furnace.

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Fig. 23 West Palm Beach waste-to-energy facility.

Ash handling systems

Much of the noncombustible material in the RDF system is removed before it is fed to the boiler. Although systems vary, there is generally some effort made to remove ferrous metals and aluminum, both of which can be troublesome once they reach the stoker grates. Non-combustibles and most of the ash from combustion collect on the traveling grate stoker and discharge off the front into a submerged chain conveyor system.

The submerged chain conveyor (Fig. 24) is a mechanical conveyor that consists of a water filled trough and a dry return trough, with two endless chain strands with flights connected between the strands. The return trough can be either above or below the water filled trough. Ash from the stoker discharge chute drops into the water filled trough. The water absorbs the impact of any larger ash pieces, quenches the ash, and provides a gas-tight seal with the stoker discharge chute.

The chains are usually driven by a variable speed drive to handle varying ash rates. The ash residue is conveyed from the bottom of a water filled trough up an incline section where the ash dewaters and discharges directly into a truck, a storage bin, or onto another type conveyor for final disposal and transport. Because this conveyor uses a dragging action to convey the ash, it is not used on mass-fired units where it can have problems in dragging the large noncombustible items up the incline section.

Fine ash from the boiler siftings hopper, flyash from the boiler, economizer and air heater hoppers, and flyash from the scrubber, baghouse or precipitator hoppers are all handled the same way as previously discussed for mass-fired boilers.

Retrofits to RDF

Most dedicated RDF boilers are new installations. However, it is possible to retrofit existing boilers to become dedicated RDF boilers. To be candidates, the existing boilers must be conservative designs for solid fuels, such as wood or coal. Typically, these are older units which are underutilized or used as standby units. These plants are often located near large metropolitan areas, a source of large quantities of refuse. The conversion of such an older power plant could represent a cost effective solution to that community's refuse disposal problems.

B&W has converted several such boilers from coal-fired to dedicated RDF. Each retrofit is unique in that each of the coal-fired boilers was of different design and originally supplied by different manufacturers. Each retrofit was also the same in that all were designed to the same standards as new RDF units.

The principal modification involves enlarging the furnace to obtain the proper furnace volume for combustion. Although coal-fired boilers have conservatively sized furnaces, refuse firing requires even larger volumes. This is achieved by removing the existing stoker and lower furnace and installing new membrane furnace wall panel extensions (Fig. 25). The new lower furnace is protected from corrosion using either Inconel weld overlay or Inconel bimetallic tubes.

Other pressure part modifications could include:

1. converting the superheater to a counterflow design while adding the proper metals for corrosion protection,

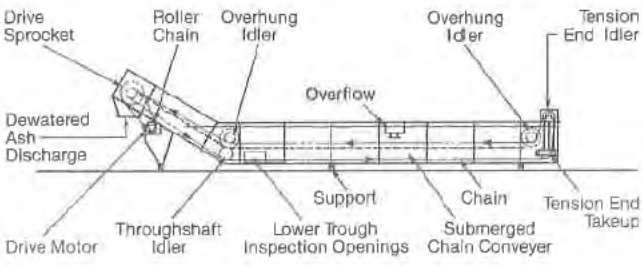


Fig. 24 Submerged chain conveyor for RDF stoker ash.

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2. modifying the boiler, economizer and air heater surface for the proper heating surface distribution and to meet refuse standards for velocities, tube spacing, etc., and
3. possibly adding screen surface to lower the flue gas temperature entering the superheater.

In some cases, the coal stoker can be reused for RDF firing; in other cases it must be replaced. In either case the grate release rate, and other design criteria, must be set to the same design standards as new refuse boilers. Properly executed, the retrofit of an existing boiler to RDF firing will result in an RDF boiler as conservative as a new RDF boiler and capable of operating equally well.

Superheater

Superheater design is critical in both mass burn and RDF-fired refuse boilers because of the highly corrosive nature of the products of combustion. This is compounded in the U.S. by the desire for the highest possible steam temperature and pressure to maximize income from power production sales while still disposing of refuse in an environmentally safe manner. B&W has pioneered the 900 psig (62.1 bar gauge), 830F (443C) high pressure, high temperature steam cycle for refuse boiler application. This 50% increase in pressure and 80F (44C) increase in temperature over the more conventional refuse boiler designs has resulted in a significant improvement in cycle efficiency.

To accomplish this improvement, the superheater must be specifically designed for corrosion protection. Superheater corrosion is a function of many variables including flue gas temperature, flue gas velocity, tube spacing, tube metal temperature, tube metallurgy and ash cleaning equipment. Even a lower steam temperature superheater designed for one or a few of these criteria can experience rapid corrosion. For example, specifying only a low furnace exit gas temperature will not assure long superheater life.

Of these criteria, tube metal temperature and tube material are most critical. To obtain satisfactory refuse boiler superheater performance, two key design features are needed:

1. a parallel flow superheater design as shown in Fig. 26 where the coolest steam conditions are exposed to the hottest gas temperatures and the hottest steam temperatures are matched with the coolest gas temperatures. The result is a design with the lowest maximum superheater metal temperatures.
2. use of Incoloy tube material in the highest tube metal temperature sections of the superheater. Carbon steel is still used in the superheater sections with lower superheater metal temperatures.

B&W refuse boilers were the first to verify successful commercial operation at the 900 psig (62.1 bar gauge), 830F (443C) steam cycle with the first three units going into operation at the Westchester County, New York refuse-to-energy facility in 1984. As of 1992, there were 19 B&W refuse boilers in operation at these steam conditions with more than 75 cumulative years of operating experience. There has been no significant corrosion in any of these superheaters.

B&W is currently extending this leadership position with the first refuse boiler design to use the 1500 psig (103.4 bar gauge) and 930F (499C) steam cycle. This design is based on laboratory corrosion research and full scale superheater test sections installed in operating

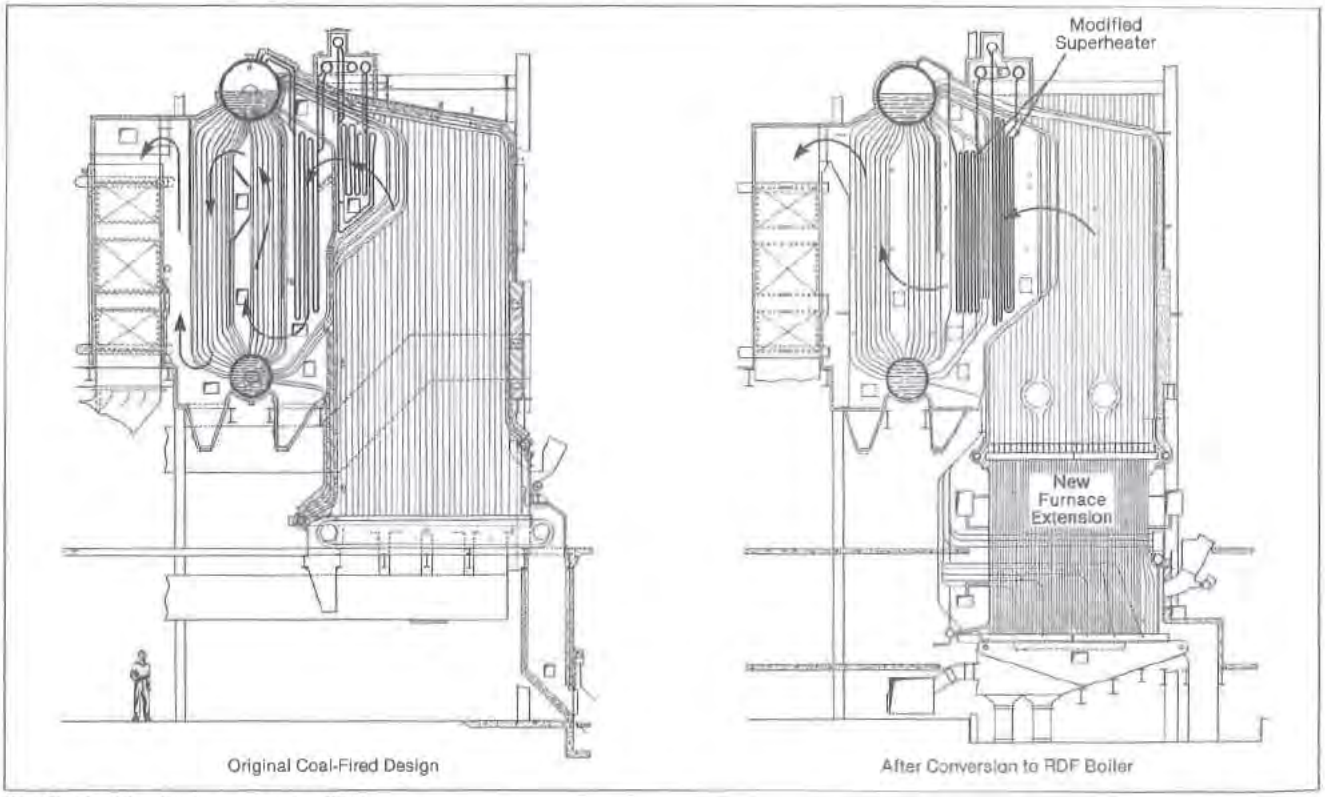


Fig. 25 Coal-fired unit converted to RDF.

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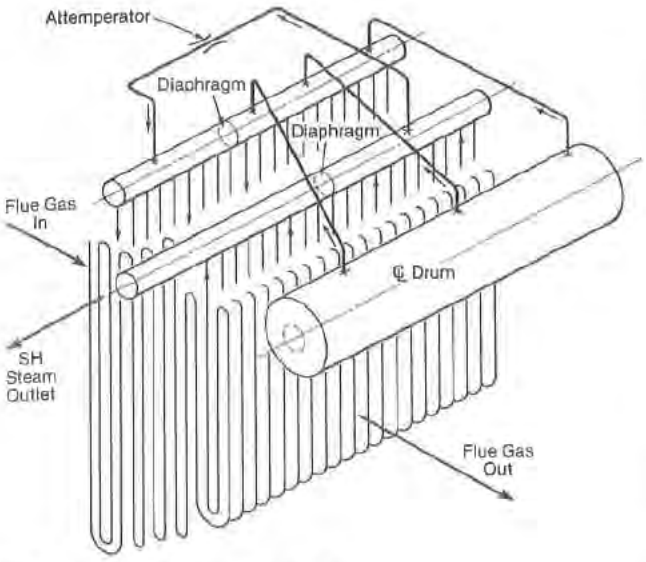


Fig. 26 Parallel flow superheater (SH).

refuse boilers. In one test, two full scale superheater sections, composed of a variety of tube metallurgies, were installed in an operating 900 psig (62.1 bar gauge), 830F (443C) design refuse boiler with steam flow through the sections controlled to simulate 950F (510C) operation. One section was removed after one year of operation and the second after 26 months. This test provided the basis for tube metallurgy selection for the 1500 psig (103.4 bar gauge), 930F (499C) design.

In addition to corrosion concerns, the superheater must be designed to minimize fouling and the potential for erosion due to excessively high flue gas velocities. Maximum design velocity is 30 ft/s (9.1 m/s), but in practice it is usually in the 10 to 15 ft/s (3 to 4.6 m/s) range. Minimum superheater side spacing is 6 in. (152 mm).

Boiler design

The lower furnace design, refuse stokers and refuse feed systems are markedly different for mass-fired and RDF boilers. However, the design requirements for the upper furnace, generating surface and economizer are the same. This is also true for auxiliary equipment such as burners and ash cleaning equipment.

Upper furnace design

The upper furnace must be sized to provide adequate heat transfer surface to reduce the flue gas temperature entering the superheater to an acceptable level. This helps minimize fouling in the superheater and maintain low superheater tube metal temperatures to minimize corrosion. A certain amount of furnace volume is required for complete burnout of the fuel in the furnace and minimum CO emissions. The required volume should be measured from the point where all the combustion air has entered the furnace (the highest level of overfire air ports) to the point where the flue gas enters the first convective heating surface (at the tip of the furnace arch at the bottom of the superheater). Measured in this manner, the required furnace volume per unit of heat input is the same for both mass-fired and RDF boilers.

The furnace must also contain sufficient heating surface to lower the flue gas temperature to help reduce fouling in the first convection section, superheater or boiler bank. These limits are 1600F (871C) entering the superheater and 1400F (760C) entering the boiler bank. As a general rule, the furnace size is set by volumetric requirements in smaller capacity boilers and by maximum gas temperature limits in larger capacity boilers.

Boiler generating bank

Refuse boilers in operation use both the one-drum and two-drum design. In the two-drum design there is both a steam drum (upper drum) and a lower drum, interconnected by the boiler generating bank tubes.

In the one-drum design the steam drum is located outside of the flue gas stream; there is no lower drum. The steam generating bank tubes are shop-assembled modules. These modules may be of either the vertical longflow (Fig. 27) or a vertical crossflow design. Minimum side spacing in the two-drum design and for the generating bank modules used with the one-drum design is 5 in. (127 mm). Maximum design flue gas velocity is set at 30 ft/s (9.1 m/s).

Economizer

The economizers can be either vertical longflow or horizontal crossflow. Economizer side spacing should be no less than 4 in. (102 mm) with a maximum flue gas velocity of 45 ft/s (13.7 m/s).

Air heater

Air heaters may be used for two reasons: 1) to supply preheated air to help dry and ignite the refuse on the stoker, and/or 2) to increase thermal efficiency where high feedwater temperatures preclude designing to lower exit gas temperatures with economizers. RDF-fired units have typically used air heaters to preheat the combustion air to the 300 to 350F (149 to 177C) range. Both tubular and regenerative air heaters have been used successfully. Due to air leakage into the air heater and the potential for fouling, regenerative types have been limited to the outlet side of hot electrostatic precipitators where the flue gases are relatively clean.

When either tubular or regenerative air heaters are used, the design and arrangement should minimize the potential for low-end temperature corrosion. To some extent, the surface arrangement in a tubular air heater will maintain adequate protection. Steam coil air heaters are required at the air inlet, on either type, to preheat the incoming ambient air and maintain temperature above acid dew points. (See also Chapter 19.)

Ash cleaning equipment

To maintain the effectiveness of all convective heating surfaces and to prevent pluggage of gas passages, it is necessary to remove ash and slag deposits from external tube surfaces. Steam or air sootblowers are most commonly used. Saturated steam is preferred for its higher density and better cleaning ability. One disadvantage of sootblowing is that localized erosion and corrosion can occur in areas swept too clean by the blowing medium. This problem can be addressed by installing tube shields on all tubes adjacent to each sootblower for localized protection.

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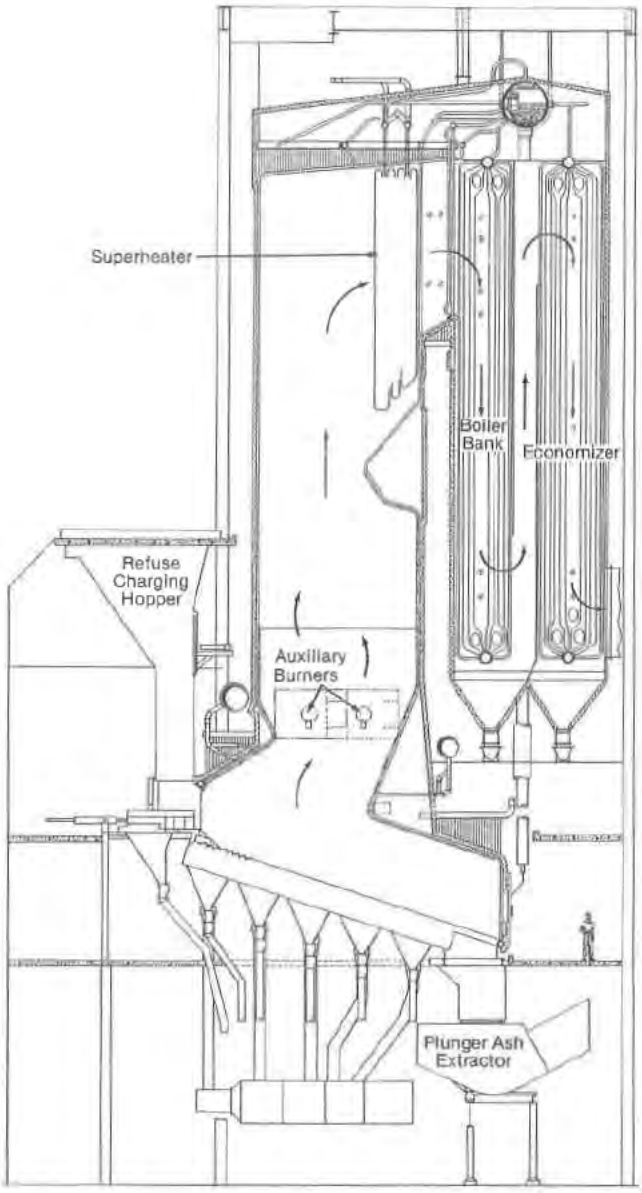


Fig. 27 Typical refuse unit with vertical longflow economizer.

A mechanical rapping system (Fig. 28) can be used to complement the sootblowers. In this system, a number of anvils strike designated pins to impart an acceleration through the superheater tube assembly. The purpose is to remove the bulk of the ash while leaving a light layer of ash on the tubes for corrosion protection. Mechanical rapping systems will not eliminate the need for sootblowers, but will reduce the number of sootblower cleaning cycles required.

Auxiliary input burners

Auxiliary fuel burners are used to maintain furnace temperature during startup, shutdown, and upset conditions since operation at low furnace temperatures could result in the incomplete destruction of volatile organic compounds. In most cases, the auxiliary fuel (oil or gas) burners are designed for only 25 to 30% of the boiler's maximum heat input.

When not in service, the typical gas- or oil-fired burner

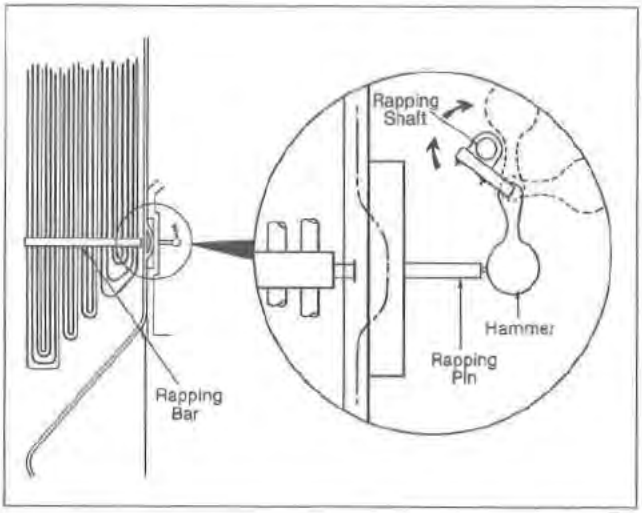


Fig. 28 B&W mechanical rapping system for cleaning superheaters.

requires some amount of air flow through the idle burner for protection against overheating. Because this air leakage represents an efficiency loss, and because these burners are used infrequently, a special design auxiliary input burner (AIB) is used for refuse boilers. The AIB is designed with a retractable burner element which is inserted toward the furnace when in use, and retracted when out of service. There is also a movable refractory block which provides protection against furnace radiation when the burner is out of service. With the burner in service, this refractory block is retracted to one side and the burner is inserted through an opening in the refractory block (Fig. 29).

Upper furnace maintenance platforms

Because refuse is a high fouling fuel, it is necessary to have good access to the convection sections. Maintenance platforms (Fig. 30) are often used to allow access to the superheater area for inspection and maintenance. Either retractable or light weight aluminum support beams are inserted into the furnace from access doors in the front wall to the superheater arch, where the beams are locked in place. Corrugated decking material is then inserted into the furnace through special sidewall access doors, and arranged on top of the support beams. This system provides both a platform for working in the superheater and upper furnace and provides some protection to those working in the lower furnace.

Air pollution control equipment

Various boiler fuels have specific components unique to that fuel. Some of these components, such as sulfur, create specific air pollution emissions that require unique boiler designs or specific air pollution control equipment. These fuels, such as high sulfur coal, are homogenous. This means the fuel will be the same in the future as it is today, and will be the same from one day to the next.

Refuse is a nonhomogenous fuel. It not only changes over the long term, but can change from day to day. Nearly every component of a fuel that can result in an unwanted air pollutant is present in refuse. However, in the early 1980s when the population of refuse-fired boil-

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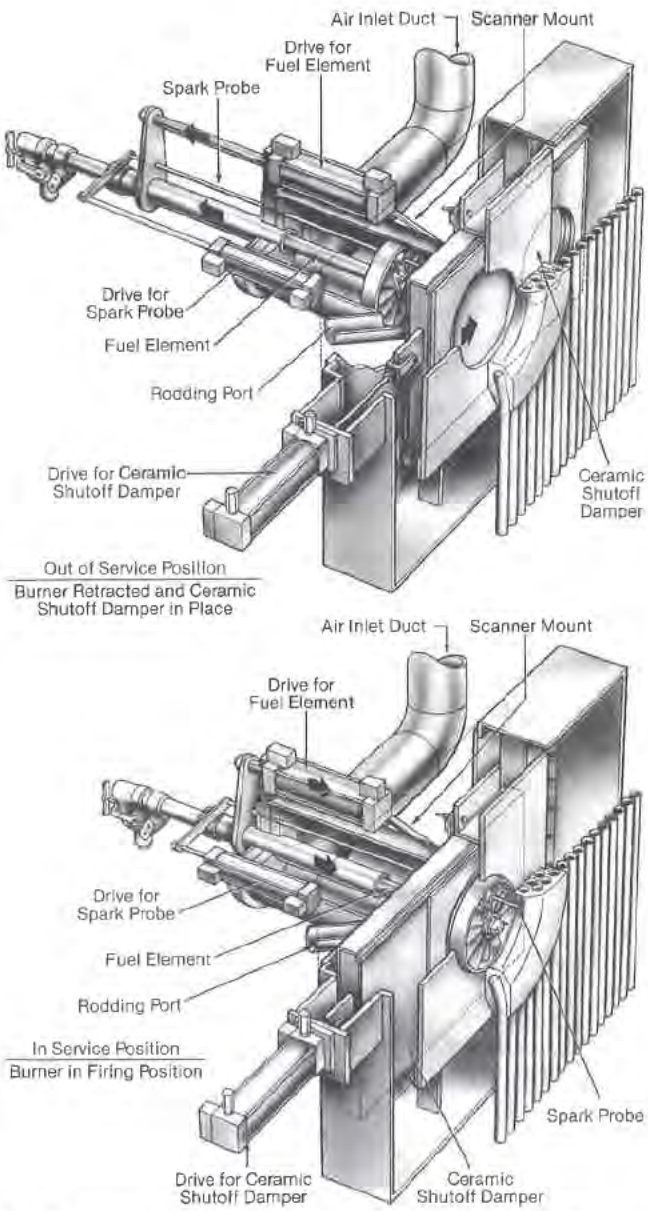


Fig. 29 Auxiliary input burner (out of service and in service positions).

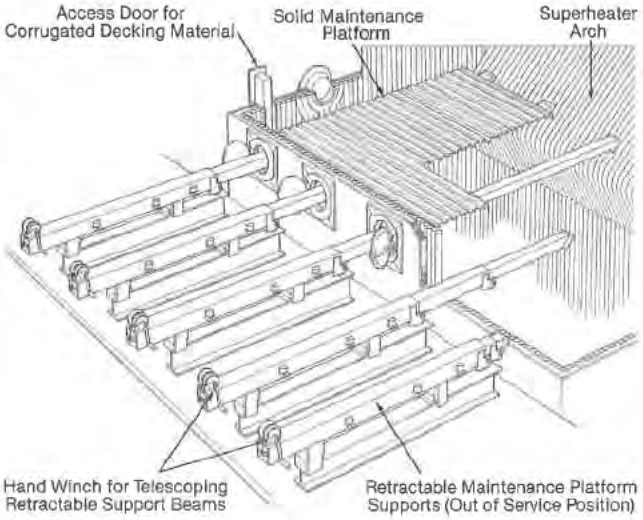


Fig. 30 Upper furnace maintenance platform.

Table 7
Permissible Stack Emissions (1990)

Pollutant	Emission Concentration
NO_x	200 to 350 ppmdv*
CO	20 to 100 ppmdv
VOC	< 10 ppmdv
SO_2	< 35 ppmdv**
HCl	< 20 ppmdv***
Particulate	< 0.01 gr/DSCF
PCDD/PCDF	< 10 ng/Nm³****

Stack emission levels are test data values for units equipped with a dry scrubber and baghouse/precipitator, and may not represent values achievable in all normal daily operations. All stack emission concentrations are corrected to a 7% O_2 reference basis.

* NO_x emissions without add-on NO_x control technology.
** Typical SO_2 emissions with back-end control efficiency in the 70 to 90% range.
*** Typical HCl emissions with back-end control efficiency in the 90 to 98% range.
**** Total dioxin and furan emissions, including all tetra through octa homologies.

ers began to rapidly grow in the U.S., the only emission requirements were on particulates, NO_x and SO_2. Refuse boilers, due to their relatively cool burning systems and the generally low level of fuel bound nitrogen, are low NO_x generators. There are also very low levels of sulfur in refuse. Therefore, early boilers were generally equipped only with an electrostatic precipitator (ESP) for particulate control. As more boilers went into operation and further air emissions data were obtained, additional emission requirements were applied. Initially hydrochlorides were targeted for control. Soon the various state air pollution agencies set regulations for the control of dioxins and furans as well as a long list of heavy metals. Dry scrubbers, used for years to control SO_2 emissions from coal-fired units, were found to be equally effective in controlling HCl emissions from refuse units. These same dry scrubbers were also found to be very effective in controlling dioxin, furan and heavy metal emissions.

With the initial use of dry scrubbers, there was a split in the preferred particulate collection system between the ESP and baghouse. ESPs were used in earlier applications due to their more extensive history of proven performance. However, it has been fairly well documented that the layer of ash and lime that collects on the bags themselves allows improved sorbent utilization for the removal of SO_2 and HCl. This allows better capture of pollutants for the same lime slurry rates, or the same level of pollutant capture at slightly reduced lime slurry rates. Today, the preferred system for nearly all refuse boilers is the dry scrubber/baghouse combination.

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In the late 1980s, lower level NO_x emissions were also being required. In-furnace ammonia or urea injection systems were installed on several refuse boilers and were shown to achieve NO_x reduction efficiencies in the 40% range. This NO_x control technology, termed selective noncatalytic reduction (SNCR), was quickly accepted by the regulatory agencies as best available control technology (BACT). In certain geographical areas, which are nonattainment areas for NO_x emissions, selective catalytic reduction (SCR) systems are also under evaluation for even greater NO_x control. The SCR systems have been demonstrated to achieve up to 90% NO_x reduction on fossil fuel boilers. However, the catalyst itself is fairly easily poisoned, and therefore rendered less effective, by a multitude of substances, all of which are found in refuse to various degrees. At this time it is not clear what the long term life of the SCR catalyst would be on a refuse-fired boiler. (See Chapters 32 through 35.)

During this same time period the emission requirements for CO have also been driven to lower levels. These requirements have been met by a combination of:

1. better overfire air system,
2. more control of undergrate air (more compartments with individual air control),
3. better combustion control systems,
4. larger furnace volumes, and
5. operator training.

The air emissions from refuse-fired boilers are as tightly regulated as those from any combustion system. However, the technology exists today to meet these requirements of the 1990s (Table 7) and will be available to meet future requirements.

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Chapter 32
Environmental Considerations

Since the early 1960s, there has been an increasing worldwide awareness that industrial growth and energy production from fossil fuels are accompanied by the release of potentially harmful pollutants into the environment. Studies to characterize emissions, sources and effects of various pollutants on human health and the environment have led to increasingly stringent legislation to control air emissions, waterway discharges and solids disposal.

Comparable concern for environmental quality has been manifest worldwide. Since the 1970s, countries of the Organization for Economic Cooperation and Development have reduced sulfur dioxide (SO_2) and nitrogen oxides (NO_x) emissions from power plants in relation to energy consumption. In at least the foreseeable future, emission trends are expected to continue downward due to a combination of factors: change in fuel mix to less polluting fuels, use of advanced technologies, and new and more strict regulations. In Japan, the reductions in SO_2 emissions were particularly pronounced due to strong environmental measures taken in the 1970s. In the United States (U.S.), use of flue gas desulfurization systems and low sulfur coal have resulted in an approximate 30% decrease in utility SO_2 emissions between 1970 and 1989, while electricity produced from coal increased approximately 120%.

Environmental control is primarily driven by government legislation and the resulting regulations at the local, national and international levels. These have evolved out of a public consensus that the real costs of environmental protection are worth the tangible and intangible benefits now and in the future. To address this growing awareness, the design philosophy of energy conversion systems such as steam generators has evolved from providing the lowest cost energy into providing low cost energy with an acceptable impact on the environment. Air pollution control with emphasis on particulate, NO_x and SO_2 emissions is perhaps the most significant environmental concern for fired systems and is the subject of Chapters 33, 34 and 35, respectively. However, minimizing aqueous discharges and safely disposing of solid byproducts are also key issues for modern power systems.

Sources of plant emissions and discharges

Fig. 1 identifies most of the significant waste streams from a modern coal-fired power plant. Typical discharge rates for the primary emissions from a new 500 MW coal-fired drum boiler, with and without control equipment, are summarized in Table 1.

Atmospheric emissions arise primarily from the byproducts of the combustion process [SO_2, NO_x, particulate flyash, volatile organic compounds (VOC) and some trace quantities of other materials] and are exhausted from the stack. A second source of particulate is fugitive dust from coal piles and related fuel handling equipment. This is especially significant for highly dusting Western U.S. subbituminous coals. Some low temperature devolatilization of the coal can also emit other organic compounds. A final source of air emissions is the cooling tower and the associated thermal rise plume which contains heat and some trace materials along with the water vapor.

Solid wastes arise primarily from collection of the coal ash from the bottom of the boiler, economizer and air heater hoppers, as well as from the electrostatic precipitators and fabric filters. Pyrite collected in the pulverizers (see Chapter 12) is usually also included. Most of the ash is transported to an ash settling pond where it settles out and is prepared for either landfill or other use. The chemical composition and characteristics of various ashes are discussed in Chapter 20.

The second major source of solids is the byproduct from the flue gas desulfurization (FGD) scrubbing process. Most frequently, this is a mixture containing calcium sulfate or calcium sulfite. After dewatering, processing and treatment, the byproduct may be sold as gypsum or landfilled. Additional sources of solids include the sludge from cooling tower basins, wastes from the water treatment system and wastes from the periodic boiler chemical cleaning.

Aqueous discharges arise from a number of sources. These include once-through cooling water (if used), cooling tower blowdown (if used), sluice water from the ash handling system (via the settling pond), FGD waste water (frequently minimal), coal pile runoff from rainfall, boiler chemical cleaning solutions, gas side water washing waste solutions, as well as a variety of low volume wastes including ion exchange regeneration solutions, evaporator blowdown (if used), boiler blowdown and power plant floor drains. Many of these streams are chemically characterized in Chapter 42. Additional discussions of these systems as well as the controlling regulations are provided in References 1 and 2.

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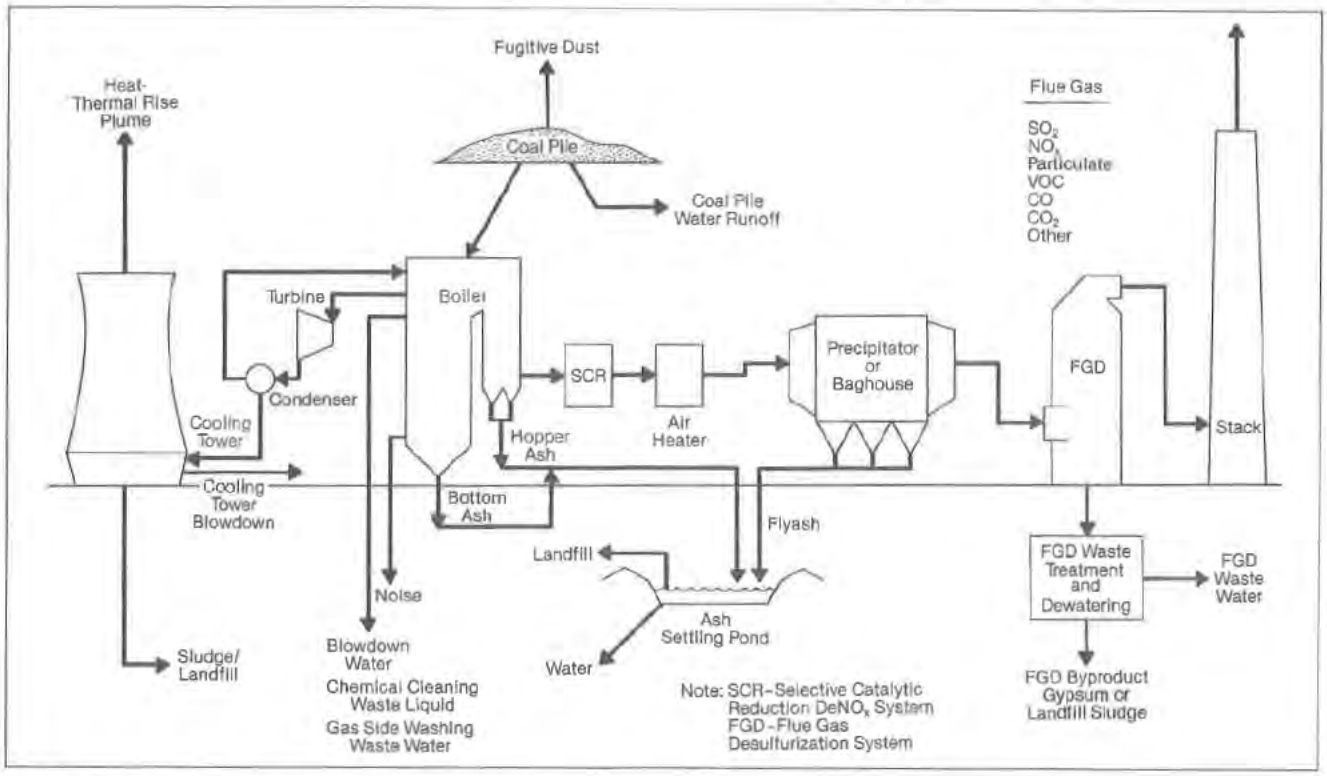


Fig. 1 Typical power plant effluents and emissions.

Air pollution control

U.S. legislation — Clean Air Act

The 1990 Amendments to the Clean Air Act (CAA), signed into law on November 15, 1990, have significantly expanded the original Act. The Amendments represent a significant increase in the U.S. commitment to an improvement in environmental quality. These Amendments build on the framework of the CAA which was originally enacted in 1963 and periodically amended since that time.

The Clean Air Act serves as the national basis for maintaining and improving the air quality through a series of minimum national requirements in a number of areas. The primary objective of the Act is *to protect and enhance the quality of the nation's air resources so as to promote the public health and welfare and the productive capacity of its population.*[3] The legislation generally provides for the U.S. Environmental Protection Agency (EPA) to set the minimum national requirements and to provide regulations, guidelines and guidance to the state and local regulatory agencies for implementation. The state and local government agencies are responsible to develop and implement plans to at least meet the federal requirements. However, they may also adopt more stringent regulations. The Act as amended prior to 1990 addressed the following areas of potential interest to boiler operators.

National Ambient Air Quality Standards These define acceptable air quality levels to protect public health and thereby set objectives for improvement. National Ambient Air Quality Standards have been defined for the six *Criteria Pollutants*: sulfur dioxide (SO_2), nitrogen dioxide (NO_2), carbon monoxide (CO), ozone (O_3), particulate matter and lead.

Table 1
Typical 500 MW Coal-Fired Steam Generator Emissions and Byproducts
[2.5% sulfur, 16% ash, 12,360 Btu/lb (28,749 kJ/kg)]

Emission	Typical Control Equipment	Discharge Rate — t/h (t_m/h) Uncontrolled		Controlled	
SO_x as SO_2	Wet limestone scrubber	9.3	(8.4)	0.9	(0.8)
NO_x as NO_2	Low NO_x burners	2.9	(2.6)	0.7	(0.7)
CO_2	Not applicable	485	(440)	485	(440)
Flyash to air*	Electrostatic precipitator or baghouse	22.9	(20.8)	0.05	(0.04)
Water stream thermal discharge	Natural draft cooling tower	2.8 x 10^9 Btu/h	(821 MW_t)	~0	
Ash to landfill*	Controlled landfill	9.1	(8.3)	32	(29)
Scrubber sludge: gypsum plus water	Controlled landfill or wallboard quality gypsum	0		25	(27.7)

* As flyash emissions to the air decline, ash shipped to landfills increases.

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State Implementation Plans (SIP) These plans are submitted by states to identify how the national standards will be implemented. They divide each state into a number of identifiable areas known as air quality control regions, in which at least the minimum National Ambient Air Quality Standards must be met and control requirements imposed as needed to achieve compliance.

New Source Performance Standards (NSPS) For more than 60 types of new and modified sources, these provide that best demonstrated available control technology must be used to satisfy requirements as set forth in Section III, Title 40 of the U.S. Code of Federal Regulations (CFR).[4]

Air toxics The Act authorized the setting of standards to limit human exposure to air toxics, which include any potentially hazardous non-criteria air pollutants. The standards are risk based and require an ample margin of safety to prevent any adverse effects (essentially zero risk).

New source review This activity is intended to prevent significant environmental deterioration from industrial growth in the form of new plants or expansion of existing plants.

Enforcement The Act provides the EPA and state agencies with the authority to bring civil and criminal actions against violators of the statutes.

Evolution of U.S. clean air legislation and key elements

The Clean Air Act has evolved from the first legislation passed in 1963 to establish authority for setting air quality standards through the Clean Air Act of 1970, which was more comprehensive and provided greater statutory strength for implementing regulatory efforts in the 1970s and 1980s. During the Middle East Oil Embargo of October 1973, the U.S. Congress, in an attempt to ensure an adequate energy supply, modified some air quality requirements through the Energy Supply and Environmental Coordination Act. In 1977 the Clean Air Act was revised to address prevention of significant deterioration of the environment and defined requirements for new sources in areas where National Ambient Air Quality Standards had not been attained.

Elements of the pre-1990 Amendments which have had the most impact on utility and industrial boilers include National Ambient Air Quality Standards and the New Source Performance Standards. National Ambient Air Quality Standards promulgated by the EPA established minimum numerical criteria for the country. Two types of standards have been established, primary standards aimed at prevention of adverse impact on human health and secondary standards to prevent damage to the environment. While ambient air quality standards have been set for the six Criteria Pollutants [CO, SO_2, NO_2, ozone (O_3), particulates and lead], there has been intense controversy in establishing what constitutes safe concentrations, accounting for the combined effects of exposure to multiple pollutants and the limited number of pollutants covered.

Before the 1977 CAA Amendments, FGD systems installed on utility boilers were designed to limit SO_2 emissions to not more than 1.2 lb per million Btu heat input. Burning low sulfur coal with less than 0.7% sulfur required no scrubbing. Coal with 3% sulfur required only 75 to 80% SO_2 removal efficiency.

The Amendments of 1977 required new utility power stations to reduce SO_2 emissions by 90% for high sulfur coals, by 70% for low sulfur coals and by a sliding scale for intermediate sulfur coals. (See Fig. 2.) A key element of the legislation and prior amendments was that a large population of existing operating utility boilers was not required to lower emissions, i.e., grandfathered, unless they were significantly modified or upgraded.

The Amendments of 1990 enhanced the previous legislation in a number of areas which have an impact on both existing and new fossil fuel-fired utility generation capacity.

Nonattainment areas (those areas not achieving National Ambient Air Quality Standards) were given new classifications and deadlines to achieve attainment under Title I of the Act. These deadlines were set on the basis of present pollution levels. This portion of the legislation may have a dramatic but indirect impact on boiler system NO_x emissions in selected nonattainment areas because NO_x is one precursor to ozone formation.

Air toxics were defined in Title III to include a list of 190 hazardous air pollutants, and the EPA was directed to promulgate control standards for sources of these emissions. While air toxics were addressed in the Act of 1970, standards had only been promulgated for seven substances by 1990. The EPA has been directed to impose tight controls through a two-phase strategy. The first phase is technology based and would require application of Maximum Achievable Control Technology (MACT) to selected categories of emission sources which emit more than minimum or threshold quantities of the listed pollutants. The second phase is risk based and may require certain facilities to apply additional controls beyond MACT to reduce remaining residual risk. Utility boilers are initially exempted from this section until the EPA studies the emissions of the air toxics from utility plants and their impact.

Acid rain (discussed below) was addressed under Title IV of the 1990 Amendments in the form of new requirements applying mainly to coal-fired power plants.

Title IV, Acid Deposition Control The 1990 Amendments establish a new market based control methodology which focuses on the control of SO_2 and NO_x emissions from fossil-fired power plants. Annual U.S. emissions of SO_2 are to be reduced by 10 million tons and annual U.S. emissions of NO_x as NO_2 are to be reduced by 2 million tons by the year 2000. This reduction is planned to take place in two phases, with the first beginning in 1995. The market based system requires utilities to use an Allowance (a permit to

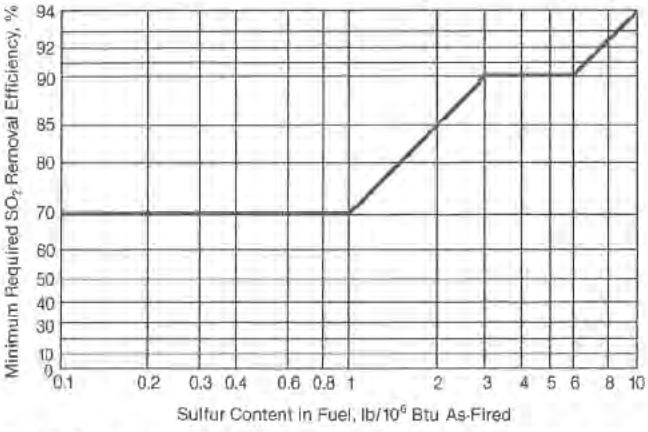


Fig. 2 Effective New Source Performance Standards for coal-fired utility boilers prior to November 15, 1990.

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emit one ton of SO_2 on or after a certain year) for each ton of SO_2 emitted after a certain date. Each existing utility boiler site receives a certain number of Allowances per year. The plant owner then has the option to either: 1) reduce emissions through fuel switching, the application of control technology or other options, or 2) acquire additional Allowances from other plants to cover the actual SO_2 emissions. Any unused Allowances may be sold or kept (*banked*) for future use. The total number of new Allowances issued each year is fixed and new power plants will not receive an Allowance allocation. This effectively caps total utility SO_2 emissions.

The NO_x control section of the acid rain legislation is part of the more traditional command and control approach where specific units or types of equipment will be allowed only certain levels of NO_x emission rates based upon fuel heat input (pounds of NO_x as NO_2 per million Btu input) although averaging will be permitted between plants.

As in prior environmental legislation, local or state environmental regulatory agencies may impose stricter regulations on individual plants, areas or locations.

New Source Performance Standards These regulations establish maximum emission rates for selected pollutants from new boilers. The NSPS were conceived to be a technology — forcing approach to provide an incentive to develop new control technology. Again, these were national maximum limits which could be tightened by the local or state implementing agencies. An example of NSPS requirements for utility boilers with a capacity greater than 250 x 10^6 Btu/h firing coal, oil or gas is provided in Table 2. Fig. 2 provides an overview of the SO_2 reduction and control requirements for coal-fired units. The details for specific applications and other fuels and technologies may be found in Title 40, Part 60, Subparts Ca, D, Da, Db, Dc, E, Ea and BB of the Code of Federal Regulations.[4] The 1990 Amendments to the Act direct the EPA to revise the NSPS.

In December 1987, the EPA issued similar guidelines for smaller industrial units which were updated in December 1989. Selected elements of these rules are summarized in Table 3. Refer to Title 40, Part 60, Subpart Db of the CFR for site specific details and rules.[4]

The NSPS apply not only to new boilers but also to boilers which undergo substantial modifications. However, the extent of the modifications and the change in pollutant emissions which trigger NSPS are currently under review.

Prevention of significant deterioration PSD relates to avoiding a reduction in air quality from industrial growth in areas that already meet air quality standards. In these PSD areas, new major sources and major modifications to existing sources are required to use Best Available Control Technology (BACT). Air quality analyses are also performed to show that the air quality standards will not be exceeded.

Nonattainment Areas that do not meet the air quality standards for a given pollutant are referred to as nonattainment areas for that pollutant. Different rules apply for new sources in these areas. New sources must apply state-of-the-art (SOA) technology and must offset the remaining emissions after application of SOA technology by reducing emissions in existing sources.

In applying the SOA technology, the Lowest Achievable Emission Rate (LAER) is to be attained. This means that the emission rate must meet the most stringent limit in any SIP or the most stringent limitation achieved by plants in the same industrial category, whichever is more stringent.

Table 2
Selected Summary of Federal NSPS for Electric Utility Steam Generators >250 million Btu/h Commencing Construction After September 18, 1978 (Notes 1 and 2)

Fuel	Pollutant	Max. Emissions Rate (lb/10^6 Btu) (Notes 3 and 6)	Req'd. Reduction in Potenial Emissions, % (Note 3)
Coal	SO_2	1.2 or 0.6	90 or 70
	NO_x as NO_2: Subbituminous Bituminous	 0.5 0.6	 (Note 4)
	Particulate	0.03 (Note 5)	99
Oil	SO_2	0.8 or 0.2	90 or 0
	NO_x as NO_2	0.3	(Note 4)
	Particulate	0.03 (Note 5)	70
Gas	SO_2	0.8 or 0.2	90 or 0
	NO_x as NO_2	0.2	(Note 4)
	Particulate	0.03 (Note 5)	—

Notes:
1. Source: 40CFR60 Subpart Da (6/11/79).
2. For reference only: see source for details.
3. *Maximum Emissions Rate* and *Req'd. Reduction in Potential Emissions* must both be met.
4. Selected % reductions are also identified for NO_x but meeting the Maximum Emissions Rate stated constitutes compliance with the % reduction requirements.
5. Separate opacity limit of 20% may be controlling.
6. Approximate SI conversions:[5] mg/Nm^3 per lb/10^6 Btu = 1230 for coal at 6% O_2, 1540 for oil at 3% O_2 and 1590 for gas at 3% O_2.

Best Available Control Technology Classification of a control technology as BACT is determined on a case by case basis by the EPA. It is then up to the owner of the new plant or plant expansion to justify deviations from BACT based on its inapplicability to the proposed plant.

Increments of air quality In addition to the application of BACT, the resulting emissions from the source must meet the available increment of air quality. All areas in the U.S. which meet the National Ambient Air Quality Standards are divided into Class I areas (that are to be kept in a pristine condition) and Class II areas (where industrial growth will be allowed). For each class an allowable degradation increment is determined. The air quality increment for a given area is set so that the resulting air quality after degradation from a new source is cleaner than what is required to satisfy the ambient air quality standard. As might be expected, Class I areas allow almost no degradation. Air quality increments have been set for SO_2, NO_x and particulates.

International regulations — air pollution control

The passage of the Clean Air Act in the U.S. in 1963 marked the first enactment of air pollution control legislation by a major industrial nation. Since that time, air pollution control regulations have become more widespread in industrial and developing nations, particularly in Ja-

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Table 3
Selected Summary of Federal NSPS for Industrial Steam Generators >100 million Btu/h Commencing Construction, Modification or Reconstruction After June 19, 1984 (Notes 1 and 2)

Fuel	Pollutant: Technology	Max. Emissions Rate (lb/10^6 Btu) (Notes 3 and 6)	Req'd. Reduction in Potenial Emissions, % (Note 3)
Coal	SO_2: All	1.2	90 (Note 3)
	NO_x as NO_2:		
	Spreader-stoker	0.6	—
	Mass-feed stoker	0.5	—
	Pulverized coal	0.7	—
	Fluidized bed	0.6	—
	Particulate	0.05 (Note 5)	—
Oil (Resid.)	SO_2	0.8 or 0.5	90 or 0 (Note 3)
	NO_x as NO_2	0.4/0.3 (Note 4)	—
	Particulate	0.10 (Note 5)	—
Gas	SO_2	—	—
	NO_x as NO_2	0.2/0.1 (Note 4)	—
	Particulate	0.10 (Note 5)	

Notes:
1. Source: 40CFR60, Subpart Db (12/18/89).
2. For reference only: see source for details.
3. *Maximum Emissions Rate* and *Req'd. Reduction in Potential Emissions* must both be met.
4. Higher rate for heat release rates >70,000 Btu/h ft^3.
5. Separate opacity limit of 20% may be controlling.
6. Approximate SI conversions: (see Table 2).

pan, Canada and Western Europe. As in the U.S., steam generating plants have been one focus of these regulatory measures, and the primary emissions of concern from combustion processes are SO_2, NO_x and particulates. CO_2 and air toxics have received some recent attention. The detailed regulations continue to evolve rapidly and are quite country specific. However, two trends are widespread:

1. allowable emission limits for controlled pollutants will continue to decline with time, and
2. a wider array of species will be considered for control.

Without attempting to be comprehensive, the following items provide a brief overview of worldwide SO_2 and NO_x regulatory efforts. While these items will probably be quickly out of date, they provide a general indication of the range and application of control measures.

Control approaches One or more of the following of measures have typically been adopted to control emissions:

1. *Emission standards* These limit the mass of SO_2 or NO_x emitted by volume, by heat input, or by unit of time (hourly, daily, annually).
2. *Percent removal requirements* These specify the portion of the uncontrolled emissions which must be removed from the flue gas.
3. *Fuel requirements* Primarily aimed at SO_2 control, these either limit the type of fuel which can be burned or the fuel sulfur content.
4. *Technology requirements* These typically indicate the type of control technology specifically required or indicate the use of the best available control technology or reasonably available control technology at the time of installation. These requirements depend in many cases on some level of economic feasibility.

The most widely used control approach is emission standards, although this is usually combined with one or more of the other approaches. Emissions from new plants are usually more tightly controlled than emissions from existing capacity. Occasionally, older plants are not controlled, although this is changing.

Local, regional or national control of emission standards depends upon the country. In the U.S. and the Netherlands, federal standards provide minimum requirements which local authorities may tighten or apply to a broader range of applications. In Canada and Australia, federal governments can only provide emission control guidelines and local or regional governments set plant limits.

SO_2 control Based upon a recent compilation by the International Energy Agency,[5] Table 4 lists selected SO_2 emission standards for new coal-fired boilers (New Source Performance Standards).

NO_x control As with SO_2 control, NO_x emissions are most frequently regulated through emission standards of mass per unit flue gas volume or per unit heat input. Most specify NO_x emissions as NO_2. Table 5 from Reference 6 provides a summary of selected current and pending NO_x control regulations.

Evolving requirements Korea, Hong Kong, the People's Republic of China, and the countries of Eastern Europe, as well as an array of developing nations, are expected to implement SO_2 and other air pollution control strategies in the near future.[5,6]

Kinds of pollutants, sources and impacts

Air pollutants are contaminants in the atmosphere which, because of their quantity or characteristics, have deleterious effects on human health and/or the environment. The sources of these pollutants are classified as stationary, mobile or fugitive. Stationary sources generally include large individual point sources of emissions such as electric utility power plants and industrial furnaces where emissions are discharged through a stack. Mobile sources are those associated with transportation activities. Fugitive emissions generally include discharges to the atmosphere from pumps, valves, seals and other process points not vented through a stack. They also include emissions from area sources such as coal piles, landfills, ponds and lagoons. They most often consist of particulates and occur in industry related activities in which the emissions are not collected.

The focus of this chapter is stationary emission sources, particularly fired utility and industrial boiler systems. Key pollutants from these sources are SO_2, NO_x, CO and particulate matter. Another class of emissions is called air toxics. These are potentially hazardous pollutants which generally occur in only trace quantities in the effluents from fired processes. However, they are undergoing more intense examination because of their potential health effects.

During the 1980s, concern increased about the potential impact of carbon dioxide (CO_2) emissions from man-made sources on the global climate. CO_2 is one of several so-called *greenhouse* gases which are said to contribute to a potential global warming phenomenon. It is emitted from a variety of naturally occurring and man-made

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Table 4
Selected International SO_2 Emission Limits for New Coal-Fired Plants (Note 1)

Country	Plant Type	Plant Size (Note 6) MW_t	Plant Size (Note 6) 10^6 Btu/h	Emissions Standard mg/Nm³	Emissions Standard lb/10^6 Btu	% Removal
Austria	Boilers	10 to 50	34 to 171	400	0.33	—
	Lignite-fired	>50	>171	400	0.33	—
	Hard coal	>50	>171	200	0.16	—
Belgium		50 to 100	171 to 341	2000	1.6	—
		100 to 300	341 to 1024	1200	1.0	—
		>300	>1024	400 (250 Note 2)	0.33 (0.2)	—
Canada	Utilities	All	All	740 (Note 3)	0.6	—
Denmark	Utilities	>50	>171	860 (Note 3)	0.7	—
Finland		50 to 150	>171 to 512	660	0.54	—
		>150	>512	400	0.33	—
Germany	Conventional	1 to 100	3 to 341	2000	1.6	—
	Boilers	100 to 300	341 to 1024	2000	1.6	60
		>300	>1024	400	0.33	85
	FBC boilers	>1	>3	400	0.33	75
Italy	Utilities	>100	>341	400	0.33	—
Japan	All plants	All	All	----------Plant specific (Note 4) ----------		
Netherlands		50 to 300	171 to 1024	700	0.57	—
		>300	>1024	400	0.33	85
Spain	Utilities	Hard coal	—	2400	2.0	—
		Brown coal	—	9000	7.3	—
	Industry	Hard coal	—	2400	2.0	—
		Brown coal	—	6000	4.9	—
Sweden	Coal-fired	>0.5	>1.7	290	0.24	—
Taiwan		All	—	3150/4000 (Note 5)	2.56/3.25	—
UK		>700	>2388	—	—	90
USA	Utility	>73	>250	1480	1.2	70 to 90
	Industry	>29	>100	1480	1.2	90

Notes:
1. Source: IEA Coal Research, October 1989.
2. Lower limit applies from 1995.
3. Guidelines.
4. Set on a plant by plant basis according to nationally defined formulae.
5. Lower limit applies to imported coal.
6. SI conversion: 8.14×10^{-4} lb/10^6 Btu per mg/Nm³, 3.412×10^6 Btu/MWh, 350 m³ flue gas/GJ input and 30 GJ/tonne.

sources which include the combustion of all fossil and hydrocarbon based fuels. Obviously, improving the power cycle efficiency (more power from less fuel) and the use of fuels with less carbon content are potential methods to address CO_2 emissions from any combustion source. However, there remains disagreement in the scientific discussion of the impact of CO_2 emissions as well as the methods and level of control. Therefore, further discussion of this issue awaits scientific and public clarification and consensus.

Sulfur oxides This category of pollutants includes mainly SO_2 with small quantities of sulfur trioxide (SO_3). The main source of sulfur oxides is from the combustion of coal, with lesser amounts from other fuels such as residual oil. Based on 1985 National Acid Precipitation Assessment Program (NAPAP) data,[7,8] the utility and industrial sectors (smelters, iron and steel mills, refineries) are the largest emitters of sulfur oxides (see Table 6). In the presence of particulate matter, sulfur oxides have been related to irritation of the human respiratory system, reduced visibility, materials corrosion and varying effects on vegetation. The reaction of sulfur oxides with moisture in the atmosphere has been identified as contributing to acid rain.

Nitrogen oxides This category includes numerous species comprised of nitrogen and oxygen, although nitric oxide (NO) and nitrogen dioxide (NO_2) are the most significant in terms of quantity released to the atmosphere. NO is the primary nitrogen compound formed in high temperature combustion processes where nitrogen present in the fuel and/or combustion air combines with oxygen. The quantity of NO_x formed during combustion depends on the quantity of nitrogen and oxygen available, the temperature, the level of mixing and the time for reaction. Control of these parameters has formed the basis for a number of control strategies involving combustion process control and burner design. Based on 1985 NAPAP data, utilities account for 33% of NO_x emitted in the U.S., with the transportation sector emitting 43%. Of the total utility NO_x emissions, 89% comes from coal-fired boilers. The most deleterious effects come from NO_2 which can form from the reaction of NO and oxygen. NO_2 also absorbs the full visible spectrum and can reduce visibility. NO_x has been associated with respiratory disorders, corrosion and degradation of materials, and damage to vegetation. NO_x has also been identified as a precursor to ozone and smog formation.

Carbon monoxide This colorless, odorless gas is formed from incomplete combustion of carbonaceous fuels. CO

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Table 5
Selected International NO_x Emission Limits for Coal-Fired Plants (Note 1)

	New Plants		Existing Plants	
	mg NO_2/Nm^3	lb/10^6 Btu	mg NO_2/Nm^3	lb/10^6 Btu
Austria	200 to 400	0.16 to 0.33	200 to 400	0.16 to 0.33
Belgium	200 to 800	0.16 to 0.65	—	—
Denmark	650 (Note 2)	0.53	(Note 2)	—
European Community (EC)	650 to 1300	0.53 to 1.06	—	—
Finland	200 to 400	0.16 to 0.33	400 to 620	0.33 to 0.50
Germany	200 to 500	0.16 to 0.41	200 to 1300	0.16 to 1.06
Italy	200 to 650	0.16 to 0.53	200 to 650	0.16 to 0.53
Japan	410 to 510	0.33 to 0.41	620 to 720	0.50 to 0.60
Netherlands	400 to 800	0.33 to 0.65	1100	0.90
Sweden	140	0.11	140 to 560	0.11 to 0.46
Switzerland	200 to 500	0.16 to 0.41	200 to 500	0.16 to 0.41
Taiwan	600 to 850	0.49 to 0.69	600 to 850	0.49 to 0.69
UK	650	0.53	—	—
USA	615 to 980	0.50 to 0.80	553 to 614	0.45 to 0.50 (Note 3)

Notes:
1. Source: IEA Coal Research Report, IEA/CR-30, December 1990.
2. In addition to *bubble principle* for utilities.
3. Dry-bottom wall-fired and tangential-fired only; other limits pending.
4. SI conversion: 8.14 x 10^{-4} lb/10^6 Btu per mg/Nm^3, 350 m^3 flue gas/GJ input and 30 GJ/tonne.

emissions from properly designed and operated utility boilers are a relatively small percentage of total U.S. combustion source CO emissions, most of which come from the internal combustion engine in the transportation sector. The primary environmental significance of CO is its effect on human and animal health. It is absorbed by the lungs and reduces the oxygen carrying capacity of the blood. Depending on the concentration and exposure time, it can cause impaired motor skills and physiological stress.

Particulate matter Solid and liquid matter of organic or inorganic composition which is suspended in flue gas or the atmosphere is generally referred to as particulate. Particle sizes from combustion sources are in the 1 to 100 µm range, although particles smaller than 1 µm can occur through condensation processes. Among the effects of particulate emissions are impaired visibility, soiling of surrounding areas, aggravation of adverse effects of SO_2, and human respiratory problems.

Table 6
1985 U.S. Anthropogenic Emissions
NAPAP Data (Version 2)[7,8]

Category	Emissions (10^3 tons) SO_2	NO_x	VOC	TSP
Utility combustion	16,055	6,662	40	570
Industrial combustion	2,679	3,198	97	304
Other combustion	613	790	1,862	1,171
Industrial processes	2,931	926	3,715	1,099
Transportation	864	8,835	8,800	4,195
Other	4	130	7,558	1,044
Total	23,146	20,541	22,072	8,383

Notes: SO_2 — Sulfur dioxide
NO_x — Nitrogen oxides as NO_2
VOC — Volatile organic compounds
TSP — Total suspended particulate

VOC Volatile organic compounds, or more commonly VOC, represent a wide range of organic substances. These compounds consist of molecules containing carbon and hydrogen and include aromatics, olefins and paraffins. A major source is the refining and use of petroleum products. Also included among VOCs are compounds derived from primary hydrocarbons including aldehydes, ketones and halogenated hydrocarbons. The major source of these compounds is the transportation and the commercial/residential combustion sectors. VOCs are environmentally significant because of their role in the formation of photochemical smog through photochemical reactions with NO_x. Control of VOCs has been the primary means of addressing areas of ozone nonattainment. Smog arising from VOC emissions can cause respiratory problems, eye irritation, damage to vegetation and reduced visibility.

Toxic and hazardous materials This is a large category of air pollutants which could have hazardous effects. The EPA had only promulgated standards for arsenic, asbestos, benzene, beryllium, mercury, radionuclides and vinyl chlorides before the passage of the 1990 Amendments to the Clean Air Act. This Act, under Title III, established a list of 190 toxic pollutants for which emissions are to be regulated. The list includes a wide range of simple and complex industrial organic chemicals and a small number of inorganics, particularly heavy metals. The EPA has identified hundreds of categories of air toxics sources, among which are utility boilers, pulp and paper plants, and municipal solid waste incinerators.

Air pollution control technologies

The strategies for control of all emissions from a utility or industrial boiler are formulated by considering design fuels, kind and extent of emission reduction mandated, and economic factors such as boiler design, location, new or existing equipment, age and remaining life.

SO_2 control strategies and technologies SO_2 emissions from coal-fired boilers can be reduced using precombus-

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tion techniques, combustion modifications and postcombustion methods.

Precombustion These techniques include the use of oil or gas in new units or the use of cleaned (beneficiated) coal or fuel switching in existing units. By using gas, sulfur emissions can be reduced to almost zero while the use of low sulfur oil will minimize SO_2 emissions. While the low sulfur content of oil and gas is advantageous, the price volatility and availability of these fuels make them less attractive except where local circumstance dictates. Switching to oil and gas in existing boilers requires attention be given to receiving equipment, storage facilities, combustion equipment including safety systems, boiler design, and unit postcombustion FGD performance. In the case of new systems, oil or gas firing can significantly reduce steam system capital costs. Even switching from one coal to another low sulfur coal can have a dramatic impact on fuel handling, combustion and particulate collection equipment. These effects are explored in more detail in Chapters 20 and 46.

Combustion modifications These techniques are primarily used to reduce NO_x emissions but can also be used to control SO_2 emissions in fluidized-bed combustion where limestone is used as the bed material. The limestone can absorb more than 90% of the sulfur released during the combustion process. (See Chapters 16 and 29.)

Sorbent injection technologies Sorbent injection, while not involving modification of the combustion process, is applied in temperature regions ranging from those just outside the combustion zone in the upper furnace to those at the economizer and duct work following the air heater. Sorbent injection involves adding an alkali compound to the coal combustion gases for eventual reaction with SO_2. Typical calcium sorbents include limestone [calcium carbonate ($CaCO_3$)], lime (CaO), hydrated lime [$Ca(OH)_2$] and modifications of these compounds with special additives. Sodium based compounds are also used. The manner in which injected sorbents react with sulfur oxides and the efficiency of the processes depend on the temperature of injection, sorbent type, sorbent surface area, and molar ratio of sorbent to sulfur. These processes are discussed in Chapter 35.

Wet and dry scrubbing technology Worldwide, wet and dry scrubbing systems are the most commonly used technologies in the coal-fired electric utility industry. 1989 data indicate that in the U.S. approximately 20% of the coal-fired utility capacity used wet scrubbing for SO_2 emission control. Both wet and dry scrubbing use slurries of sorbent and water to react with SO_2 in flue gas, producing wet and dry waste products, respectively.

In the wet scrubbing process, a sorbent slurry consisting of water mixed with lime, limestone, magnesium promoted lime or sodium carbonate (Na_2CO_3) is contacted with flue gas in a reactor vessel. Wet scrubbing is a highly efficient (> 90% at calcium/sulfur molar ratios close to 1.0), well established technology which can produce usable byproducts.

Dry scrubbing involves spraying an aqueous sorbent slurry into a reactor vessel so that the slurry droplets dry as they contact the hot flue gas [~300F (~149C)]. The SO_2 reaction occurs during the drying process and results in a dry particulate containing reaction products and unreacted sorbent entrained in the flue gas, along with flyash. These materials are captured downstream in the particulate control equipment.

Dry scrubbing is a well established technology with considerable operational flexibility, but fouling of downstream duct work is possible. The waste residue is dry.

NO_x control technologies NO_x emissions from fossil fuel-fired industrial and utility boilers arise from the nitrogen compounds in the fuel and molecular nitrogen in the air supplied for combustion. Conversion of molecular and fuel nitrogen into NO_x is promoted by high temperatures and high volumetric heat release rates found in boilers. The main strategies for reducing NO_x emissions take two forms: 1) modification of the combustion process to control fuel and air mixing and reduce flame temperatures, and 2) postcombustion treatment of the flue gas to remove NO_x. (See Chapters 10, 13, 15 and 34.)

Combustion modification This approach to NO_x reduction can include the use of low NO_x burners, combustion staging, gas recirculation or reburning technology.

Low NO_x burners slow and control the rate of fuel and air mixing, thereby reducing oxygen availability in the ignition and main combustion zones. Low NO_x burners can reduce NO_x emissions by 50% or more, depending upon the initial conditions, are relatively low cost and are applicable to new plants as well as retrofits.

Staged combustion uses low excess air levels in the primary combustion zone with the remaining (overfire) air added higher in the furnace to complete combustion. Significant NO_x reductions are possible with staged combustion, although reducing zones and potential for corrosion and slagging exist.

Flue gas recirculation reduces oxygen concentration and combustion temperatures by recirculating some of the flue gas to the furnace without increasing total net gas mass flow. Large NO_x reductions are possible with oil and gas firing while moderate reductions are possible with coal firing. Modifications to the boiler in the form of ducting and an efficiency penalty due to power requirements of the recirculation fans can make the cost of this option higher than some of the other in-furnace NO_x control methods.

Reburning is a technology used to reduce NO_x emissions from Cyclone furnaces (Chapter 14) and other selected applications. In reburning, 75 to 80% of the furnace fuel input is burned in the Cyclone furnace with minimum excess air. The remaining fuel (gas, oil or coal) is added to the furnace above the primary combustion zone. This secondary combustion zone is operated substoichiometrically to generate hydrocarbon radicals which reduce NO formed in the Cyclone to N_2. The combustion process is then completed by adding the balance of the combustion air through overfire air ports in a final burnout zone in the top of the furnace.

Postcombustion The two main postcombustion techniques for NO_x control are selective noncatalytic reduction (SNCR) and selective catalytic reduction (SCR). In SNCR, ammonia or other compounds such as urea (which thermally decomposes to produce ammonia) are injected downstream of the combustion zone in a temperature region of 1400 to 2000F (760 to 1093C). If injected at the optimum temperature, NO_x is removed from the flue gas through reaction with ammonia. SCR is being used worldwide where high NO_x removal efficiencies are required in gas-, oil- or coal-fired industrial and utility boilers. SCR systems remove NO_x from flue gases by reaction with ammonia in the presence of a catalyst.

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Particulate control technologies Particulate emissions from boilers arise from the noncombustible, ash forming mineral matter in the fuel which is released during the combustion process and is carried by the flue gas to the stack. Another source of particulate is the incomplete combustion of the fuel which results in unburned carbon particles. A brief description of the principal options for particulate emissions control in industrial and utility boilers follows while Chapter 33 provides an in depth discussion.

Coal cleaning Historically, physical coal cleaning has been applied to reduce mineral matter, increase energy content and provide a more uniform boiler feed. Although reduction in flue gas particulate loading is one of the potential benefits, coal cleaning has been driven by the many other boiler performance benefits related to improved boiler maintenance and availability and, more recently, the reduction in SO_2 emissions.

Mechanical collectors These are generally cyclone collectors and have been widely used on small boilers when less stringent particulate emission limits applied. Cyclones are low cost, simple, compact and rugged devices. However, conventional cyclones are limited to collection efficiencies of about 90% and are poor at collecting the smallest particles. Improvements in small particle collection are accompanied by high pressure drops.

Fabric filters These filters, also commonly referred to as baghouses, are available in a number of designs (reverse air, shake/deflate and pulse jet), each having advantages and disadvantages in various applications. Applications include industrial and utility power plants firing coal or solid wastes, plants using sorbent injection and dry scrubbing FGD, and fluidized-bed combustors. Collection efficiency can be expected to be at least 99.8% or greater. Fabric filters have the potential for enhancing SO_2 capture in installations downstream of sorbent injection and dry scrubbing systems as discussed in Chapter 35.

Electrostatic precipitators ESPs are available in a broad range of sizes for utility and industrial applications. Collecting efficiency can be expected to be 99.8% or greater of the inlet dust loading. ESPs are considered to be less sensitive to plant upsets than fabric filters because their materials are not as sensitive to maximum temperatures. They also have a very low pressure drop. Power usage of ESPs and fabric filters tend to be similar because the high fan power needed to overcome the higher fabric filter pressure drop is approximately equal to the power consumed in the ESP transformer rectifier sets. ESP performance is sensitive to flyash loading, ash resistivity and coal sulfur content. Lower sulfur concentrations in the flue gas can lead to lower ESP collection efficiency.

Water pollution control

U.S. legislation — Clean Water Act

The objective of the Clean Water Act (CWA) is to *restore, maintain and enhance the chemical, physical and biological integrity of the nation's waters*.[9] The Act does this by setting standards for the quality of bodies of water and limitations on effluents from industrial and municipal activity. The Federal Water Pollution Control Act of 1972 and the amendments included in the Clean Water Act of 1977 authorized the EPA to control toxic pollutants discharged into waterways from point sources. These included pipes, channels and ditches. The 1972 legislation identified 129 priority pollutants which are considered hazardous wastes and which must be limited in discharges into domestic waters. During the 1980s, regulatory activities broadened to include nonpoint sources of toxic pollutants and water quality in lakes and ground water, in addition to previously regulated water quality in rivers and streams.

Requirements under the Act Industries must achieve discharge limits based on *best practicable*, followed by *best available*, treatment technologies as determined by the EPA. Industrial facilities saw an increase in compliance from 36 to 78% between 1972 and 1982. The positive response to the CWA requirements has also had a negative result. Treatment of waste water from point sources results in large quantities of sludge which, in many cases, contain toxic hazardous wastes. The wastes are generally landfilled and have resulted in numerous and large disposal sites throughout the U.S. The effects of complying with both the CWA and the CAA have shifted the burden of ultimate disposal to land based facilities.

Power plant discharge sources

In 1982 the EPA promulgated effluent limits, pretreatment standards and new source standards for 21 major industrial categories which included steam electrical power generation. Key limits in this category for utility power plants are given in Table 7. The EPA surveyed industrial discharges to determine for each discharge which of the 129 priority pollutants were present in sufficient concentration to require national regulation. Most of the organic compound priority pollutants were excluded from the utility plant discharges because of their absence or low concentration. The EPA did not initially set limits for four types of utility plant aqueous wastes, reserving these for future rule making. These are nonchemical metal cleaning wastes, FGD waste water, runoff from materials storage and construction areas (excluding coal piles), and thermal discharges. The following describes the principal aqueous discharge streams from utility power plants.

Once-through cooling water Water from rivers, lakes or oceans is used to absorb heat from the steam condenser. The cooling water exiting the steam condenser is at an elevated temperature and can be returned to the source or pumped to a cooling tower for evaporative cooling before being returned to the steam condenser. In the former case, the cooling water contains significant concentrations of only one principal regulated pollutant, total residual chlorine (TRC), which arises out of chlorine addition for condenser fouling control. The duration of each chlorination event is limited. The concerns over TRC discharge include toxicity to living organisms and the generation of halogenated hydrocarbons.

Cooling tower blowdown When the heated cooling water from the main steam condenser is cooled in an evaporative cooling tower, a buildup in dissolved solids and suspended matter occurs. Most of this buildup is removed from the system by cooling tower blowdown. Some of the suspended matter can settle out in the cooling tower basin and is removed at infrequent intervals. All of the dissolved solids and the remaining suspended solids are removed largely by cooling tower blowdown, although a small amount is discharged by drift in cooling tower ex-

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haust. Blowdown flow is adjusted to keep the concentration of dissolved and suspended solids below the limits required to control condenser tube fouling and corrosion. Blowdown can vary from 3 to 65% of the makeup flow, depending on whether fresh water or salt water is used and whether a portion of the recirculating stream is continuously treated for deposit forming substances. Sources of chemical pollutants in blowdown include chlorine and proprietary organic chemicals for control of biofouling, corrosion inhibitors (consisting of chromate, zinc, polyphosphates, etc.), chemicals for scale control and products of corrosion. Although some of these maintenance chemicals appear on the 126 priority pollutants list, none are permitted to be present in significant levels in cooling tower blowdown (except for chromium and zinc, which are separately regulated).

Ash handling water waste Ash produced from the combustion of fuel, whether oil or coal, is collected at different points in the combustion process. Flyash is the finer size ash collected by particulate collection systems and bottom ash is removed from hoppers at the furnace bottom. Additional hoppers at intermediate points also accumulate ash. In many cases, ash is moved from these points with sluice water, which then goes to a settling pond and can typically contain 5% of suspended solids by weight.

The ash settling pond overflow can contain dissolved and suspended solids, the quantities of which will depend on the source of the ash, the type of combustion process and the point from which it is extracted from the combustion process. In general, oil ash can contain oxides and salts of vanadium, nickel and iron, carbon, organometallic compounds and magnesium compounds when a magnesium oxide additive is used for corrosion control. Oil ash is more soluble than coal ash and settles more slowly because of its particle size. Coal ash, because of the significantly greater quantity of mineral matter in coal compared to oil, is produced in much greater quantities than oil ash (300,000 t/y versus 2000 t/y for a typical 1000 MW utility plant). Bottom ash from coal combustion tends to be a mixture of vitreous metal oxides and silica, is low in solubility and tends to settle quickly because of its coarse size compared to oil ash. Coal flyash can have solubility of several percent and, because of its fineness and the presence of low density hollow cenospheres, tends to settle slowly. Coal ash contains, in addition to the eight or nine major elemental constituents, a large number of trace elements that can appear in pond overflow and which may need to be treated by best practicable technology.

Coal pile runoff Open storage of large quantities of coal is required for an uninterrupted fuel supply to utility plants [on the average, 800 to 2400 yd^3 (611 to 1834 m^3) per megawatt of rated capacity is kept on hand]. The water and oxygen from the air react with the minerals in the coal to produce a leachate contaminated with ferrous sulfate and sulfuric acid. The low pH from the acid accelerates dissolution of many of the metals present in the coal minerals.

FGD blowdown In wet FGD systems, a portion of the absorber slurry, which is sprayed into the flue gas stream to remove SO_2, is removed from the absorber tank for dewatering. In the dewatering process, the solid reaction products are separated from the liquor. The liquor is recycled to the absorber tank where additional sorbent is added. Recycling of the liquor can result in chloride buildup which, in turn, can cause increased sulfate scaling by upsetting the sulfite/sulfate balance. This buildup can be controlled by the loss of liquor retained in the dewatered sludge or by a blowdown. An aqueous blowdown discharge would typically be a saturated solution of calcium sulfate, calcium sulfite and sodium chloride. Also, depending on flyash carryover, traces of metal ions could also be present. In setting effluent limitations in 1982, the EPA reserved regulating FGD aqueous discharge to a future date.

Table 7
1982 Adopted Aqueous Discharge Limits for Steam Power Generating Systems (Note 4)

Source and Pollutant (Note 1)	Effluent Limits, mg/l BAT (Note 2) Maximum	Average
All discharges		
pH	6 to 9	6 to 9
PCBs	No discharge	No discharge
Low volume waste (Note 5), bottom ash and flyash transport water (Note 3):		
TSS	100	30
OG	20	15
Chemical metal cleaning wastes:		
TSS	100	30
OG	20	15
Copper	1.0	—
Iron	1.0	—
Once-through cooling water, total residual chlorine (TRC)	0.2	—
Cooling tower blowdown:		
Free available chlorine	0.5	0.2
Zinc	1.0	1.0
Chromium	0.2	0.2
Other 124 priority pollutants	No detectable amount	
Coal pile runoff:		
TSS (1980)	50	—

Notes:
1. Nomenclature: TSS - total suspended solids; OG - oil and grease; BAT - best available technology.
2. 30 day rolling daily average (Average); maximum any one day (Maximum).
3. Best conventional technology basis.
4. Adapted from Reference 1.
5. Low volume wastes include ion exchange, water treatment, evaporator blowdown, boiler blowdown, lab and floor drains, plus FGD waste water.

Metal cleaning wastes These aqueous wastes can arise from either *chemical* or *nonchemical* cleaning of metal heat transfer surfaces in the boiler.

Chemical metal cleaning uses chemical solvents for water-side cleaning of boiler system components to remove corrosion products. Cleaning intervals are measured in years for large utility boilers, and produce three to four boiler volumes [20,000 to 100,000 gal (75,708 to 378,5401)] of waste water per cleaning. The composition of the waste solvents depends on the construction material of the feedwater system, but largely consists of iron with lesser amounts of copper, nickel, zinc, chromium, calcium and magnesium. The disposal method for the spent solvent

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depends on the type of chemical cleaning solvents used. When hydrochloric acid based solvents are used, spent solvent is treated on-site by neutralization and is discharged subject to the effluent limits in Table 7 or more stringent water quality standards. With approval from appropriate regulating bodies, organic based solvent wastes are often incinerated in other operating boilers at the site. The metals in the chemical cleaning wastes are retained with the normal boiler ash.

Nonchemical cleaning is used to remove fireside deposits by means of high pressure jets of water. The waste water can contain the same metals and pollutants contained in the ash deposits being removed. Because the deposit composition varies with position in the boiler, the wash water composition will depend on the location of the area being cleaned. These waste waters may be classified as either low volume wastes or metal cleaning wastes and are treated according to the corresponding effluent limits.

Low volume wastes These include discharges from ion exchange water treatment, evaporator blowdown, boiler blowdown, cooling tower basin cleaning, laboratory and floor drains, and drains and losses from recirculating house service water systems. FGD blowdown is also included until the EPA develops specific regulations for this stream. By EPA definition, low volume wastes are those from all sources taken collectively as if they were from one source. Excluded are those wastes for which specific effluent limits are established.

Water pollution control technologies

The technologies for waste water treatment used to meet limits for discharge include clarification and filtration. (See also Chapter 42.)

Clarification This is used to settle out larger suspended particles and condition smaller colloidal particles to make them settle and allow filtration for removal. A pond, reservoir or tank is used to allow larger particles to settle in a matter of hours. The finer particles overflow and are made to settle more quickly by the addition of chemical agents, coagulants and polymers that cause agglomeration to sizes large enough to settle out of suspension.

Filtration This uses a porous barrier across flowing liquid to remove suspended materials. Filtration can be used to supplement clarification and permits reducing suspended solids to the parts per billion level. Filter types include sand filters which are generally slow and do not handle fine clay well. Preconditioning with coagulants can improve filtration rates. Dual media filters improve on sand filters by superimposing a coarse, granular material over the fine bed. This allows more of the filter bed to be used, reduces head loss, and provides higher flow rates and longer operating cycles before cleaning.

As required, and with approvals from appropriate regulating bodies, final waste stream pH is controlled by combining various plant streams to provide a neutral pH product. Where needed, acid or alkali addition can be used to achieve the final pH. Other treatments are also available to address other criteria pollutants where concentration warrants.

In selected cases, zero discharge water management is provided which does not return any waste water to water sources. Effectively all water brought into the plant is evaporated through cooling towers, ponds or stack. Residual solids are then sent for disposal.

Solid waste disposal

The Resource Conservation and Recovery Act

The rapid growth of industrial activity and use of consumer goods by our society have resulted in an explosive growth in solid wastes. One of the first attempts to protect health and the environment, reduce waste, conserve natural resources and control hazardous waste production was the Solid Waste Disposal Act of 1965. This was followed in 1976 by the Resource Conservation and Recovery Act (RCRA) which amended the original Act to ensure proper future solid waste management.

The RCRA requires improved solid waste management practices, defines what constitutes hazardous waste and greatly expanded provisions for hazardous waste management. Three notable parts of the Act with far reaching implications are Subtitles C, D and I. Revisions made in 1984 were significant in expanding the regulatory scope of the RCRA through the Hazardous and Solid Waste Amendments.

The definition of a hazardous waste is very broad. It is based on the premise that a material is hazardous when its quantity, concentration or physical/chemical/infectious characteristics *(1) cause, or significantly contribute to, an increase in mortality, or an increase in serious irreversible or incapacitating reversible illness, or (2) pose a substantial present or potential hazard to human health or the environment when improperly treated, stored, transported or disposed of, or otherwise managed.*[10]

Under Subtitle C, hazardous waste can include virtually any type of waste (solid, semisolid, liquid and gaseous) which fits one or more of the following criteria:

1. possesses the characteristics of ignitability, corrosivity, reactivity and toxicity as determined by standard analytical tests and procedures (commonly referred to as a characteristic hazardous waste).
2. is listed as a specific hazardous waste (commonly referred to as a listed hazardous waste).
3. is a mixture of wastes which contains a listed hazardous waste.
4. has not been excluded from RCRA regulation as a hazardous waste.
5. is produced as a byproduct from the treatment of any hazardous waste.

Listed hazardous wastes fall into three categories based on their sources:

1. *Nonspecific source wastes* These may be derived from various industrial processes and are viewed as generic in nature, such as degreasing solvents and electroplating waste water sludges.
2. *Specific source waste* This type of waste is specifically named for and identified in terms of the industrial process from which it is produced, i.e., K047, pink/red water from TNT operations; K050, heat exchanger bundle cleaning sludge from the petroleum refining industry.
3. *Commercial chemical products* These consist of specific chemicals such as organic and inorganic compounds, pesticides and often products which are no longer of use and are identified for disposal.

In the enactment of the RCRA, there were a number of legislative exclusions of waste types which were produced in very large quantities but did not present a danger to

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health or the environment, or that were covered by other environmental regulations. Some of the exclusions are of significance to the power industry and the management of municipal solid wastes. Among the excluded wastes were household wastes, agricultural wastes and wastes from municipal resource recovery plants. In 1980 additional temporary exclusions were made to include wastes from fossil fuel combustion, mining operations, and oil and gas explorations. These exclusions are being examined by the EPA for possible regulation under Subtitle C.

The general trend, whether through federal or state and local action, is toward tighter control and restrictions. Depending on the state and local jurisdiction, high volume waste streams from power plants such as scrubber sludge, flyash and bottom ash are subject to different and highly variable disposal requirements. In some cases, stabilization of the solids is required. In addition, landfills are beginning to use leachate collection systems with single or double linings and extensive monitoring wells.

Kinds of solid wastes

The principal solid waste streams in coal- and oil-fired utility boilers include the following:

1. *Bottom ash* is that portion of fuel ash which falls to the bottom of the furnace or from the stoker discharge. In coal-fired Cyclone furnace boilers, the bottom ash consists of slag which drops from the bottom of the furnace into a slag tank for solidification.
2. *Flyash* is the finer ash material which is borne by the flue gas from the furnace to the back end of the boiler; it drops out in the economizer and air heater hoppers or is collected by particulate control equipment.
3. *Pyrite* is iron sulfide, an impurity which is separated from coal in the pulverizer and which is combined with bottom ash for disposal.

FGD waste characteristics depend on the particular technology used:

1. *Wet scrubbing (calcium based system)* Using natural oxidation produces a wet sludge containing a mixture of calcium sulfite and calcium sulfate reaction products, trace amounts of flyash and unreacted limestone. In a forced oxidation system, the principal difference in the waste is that the reaction product is almost totally in the form of calcium sulfate or gypsum, which is more easily dewatered to a filter cake for landfill or other use.
2. *Dry scrubbing* Waste is dry and contains calcium sulfite, calcium sulfate, flyash and unreacted sorbent (hydrated lime).
3. *Dry lime injection* Waste is dry and contains calcium sulfate, flyash and a large proportion of calcium oxide (CaO).

Solid waste treatment methods

In order to dispose of waste materials from wet collection systems, treatment methods are applied to ultimately produce a solid. These methods include dewatering, stabilization and fixation and are designed to achieve waste volume reduction, stability and better handling, or liquid recovery for reuse.

Dewatering This is used to physically separate water from solids to increase solids content of the product and recover water for further treatment and potential reuse.

A settling pond is the simplest method for dewatering, is not sensitive to inlet solids content, requires low maintenance and is highly reliable. Ponds are often used for ash or limestone scrubber slurries. Sizing provides low flow velocity so that solids can settle undisturbed by gravity. Settling ponds are unpopular with regulatory agencies, require substantial acreage and must be shut down for solids removal.

Thickeners are large cylindrical tanks with a center column that drives radial rakes extending from the bottom of the shaft. The rakes carry plows to stir material on the bottom which slopes toward the center. (See Chapter 35 and 42.) The plows push settled material toward the underflow discharge. Thickeners rely on gravity to separate high specific gravity solids and are often applied to dewatering wet scrubber slurry. Although thickeners are complicated and have high capital and maintenance costs, they have high throughput rates and require less land area than settling ponds.

Cyclone collectors or hydroclones are sometimes used in place of thickeners to remove solids from slurries by centrifugal and liquid shear effects. Cyclones separate and collect particles down to a particular size, with finer particles staying with the liquid overflow. Cyclones do not separate material less than 5 μm effectively and are not efficient with slurries containing more than 15% solids. They are low in cost, have low space requirements and produce low solids content in the liquid fraction and a high liquid content in the solids fraction.

Vacuum filters, either of the drum or belt type, are generally used for second stage dewatering of wet scrubber slurries. They take little space and produce a high solids content product, up to 65% for FGD slurry and 75% for ash slurry. However, vacuum filters are higher in cost and maintenance and are mechanically complex.

Stabilization This increases solids content of waste by adding dry solids such as flyash. Stabilization is applied to impart greater physical stability to the waste, making it easier to place in the landfill and making it less susceptible to future problems. Stabilization and fixation are generally applied to scrubber wastes as the final treatment step after dewatering. Bottom ash and flyash, because of their granular nature, generally dewater easily and do not require stabilization for disposal. For stabilization, a dry solid such as soil or flyash is mixed with the waste slurry and spreads the water in the waste over a larger mass of solids. Also, there is improvement in particle size distribution which leads to closer packing, lower permeability and lower combined volume. Stabilization can be reversible and if the waste is rewetted it may fluidize and fail structurally.

Fixation This involves the addition of an agent such as lime to produce a chemical reaction to bind free water and produce a dry product. Fixation includes a number of processes. Mixing suitable proportions of scrubber slurry with alkaline flyash containing sufficient CaO produces a chemical reaction which results in a material with compressive strength comparable to concrete and with very low permeability. Both characteristics contribute to ease of placement and minimal leaching problems.

When the flyash does not have sufficient alkalinity, lime may be added with the flyash scrubber slurry to produce the cementitious reaction. Four percent addition of lime has produced material with the necessary physi-

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cal properties for disposal or use. The cured material is suitable for structural fill, providing a site which can be used for building construction after completion of the landfill. Comparable fixation reactions with scrubber sludge have been obtained with additions of 5 to 10% of blast furnace slag or Portland cement.

Disposal and utilization methods and requirements

Ultimate disposition of utility plant wastes (ashes and FGD residues) is by disposal (in landfills or impoundments) or by utilization. Where disposal is used, the waste stream is analyzed, and the site is permitted and approved by the appropriate regulatory agencies.

Disposal methods These can be either wet or dry, depending on the physical condition of the material. Wet disposal requires construction of a pond which may be below or above grade with impermeable barriers or dikes provided. Below grade construction may be considered and depends on suitable geology and hydrology at the site. With wet disposal, the waste is placed in slurry or liquid. After placement and settling, the liquid which has separated is collected, treated and either released or recycled. Dry disposal can use a simple method of landfill construction in which the waste is placed and compacted to form an artificial hill. The trend is toward dry disposal because of smaller volumes, more options for site or material reclamation, and the developing interest in dry scrubbing. Baghouse particle collection systems would also encourage the use of dry disposal.

Utilization methods These become more attractive as waste management costs increase. Bottom ash, flyash and boiler slag are used in applications where they can be substituted for sand or gravel. The characteristics of boiler slag and bottom ash also make these materials useful for blasting grit, roofing granules and controlled fills. Flyash, because of its chemistry and physical properties, is applicable in the manufacture of Portland cement and concrete mixes. The value of these materials is so low that the cost of transportation severely limits their use to applications close to the producing power plant.

FGD byproduct use is potentially in the areas of agriculture, metals recovery, sulfur recovery and gypsum. Agricultural use is limited to waste which contains lime; this can be substituted for agricultural lime. However, trace elements from flyash contamination could have an unacceptable impact and make wide use doubtful. Use for metals recovery is limited by undeveloped technology and questionable cost effectiveness. Use for sulfur recovery is limited by incomplete technology development, low market price of sulfur and large amounts of residual byproduct after-sulfur extraction.

Only FGD waste from forced oxidation wet scrubbing systems, which consists of almost pure gypsum, has seen some commercial use for drywall production.

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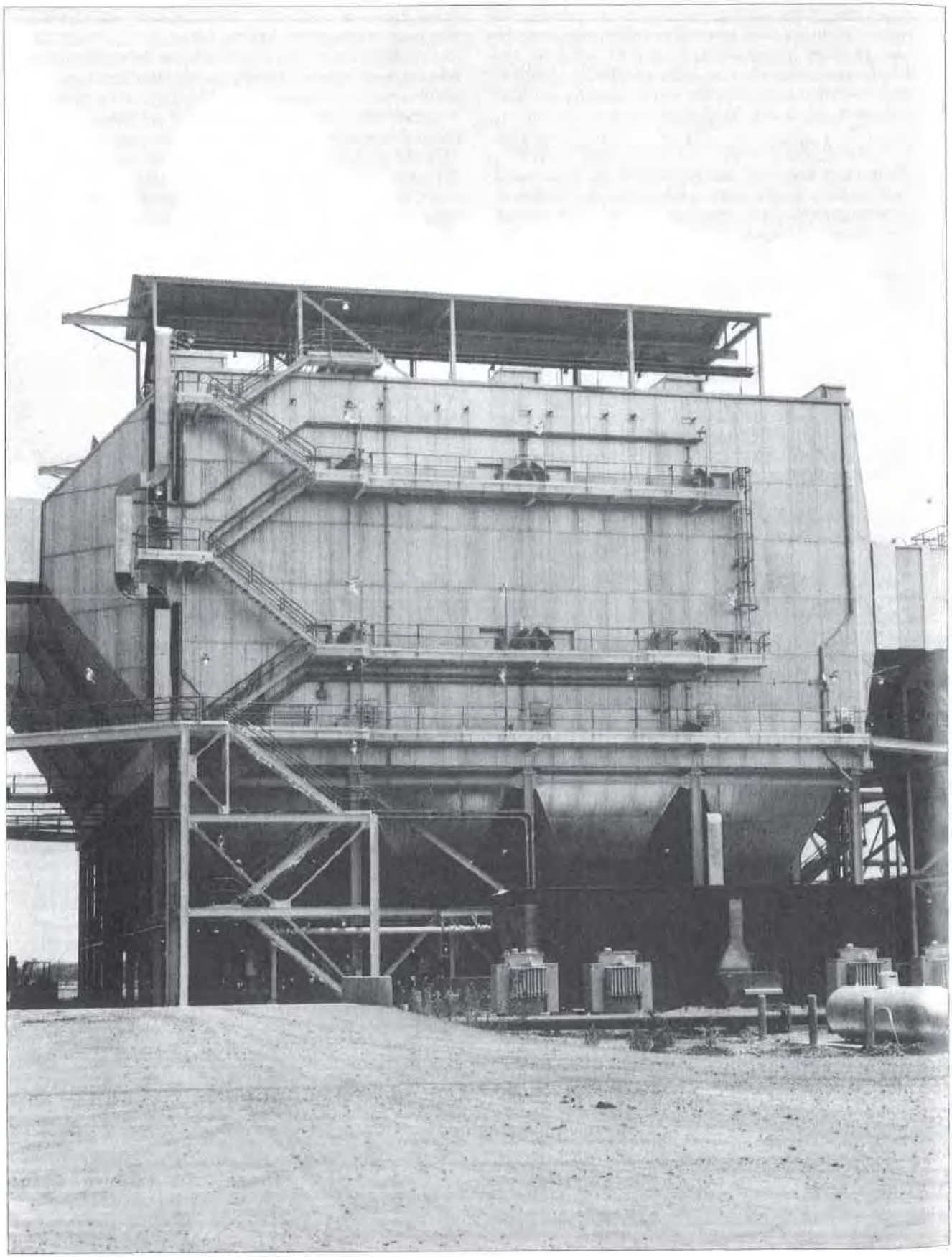
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Large utility electrostatic precipitator installation.

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Chapter 33
Particulate Control

As steam is generated by the combustion of most fossil fuels, the flue gas carries particulate matter or ash from the furnace. Except for natural gas, all other fossil fuels contain some quantity of ash or noncombustibles which form the majority of the particulate. Unburned carbon also appears as particulate. Combustion in most steam generation applications using nonfossil fuel also produces particulate. Control is needed to collect this material and to limit its release to the atmosphere.

All coals contain some amount of ash. Content 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.

When coal is burned in conventional boilers, a portion of the ash drops out of the bottom of the furnace 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 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 30% 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) spectrographic analysis test of a coal sample reveals the major ash components and determines total ash content. Ash constituents 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 8 and 20.

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 smaller yet. 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 mostly 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 and stack opacity to 20% for those units larger than 250 x 10^6 Btu/h (73.3 MW_t) heat input. *Opacity*, measured by 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

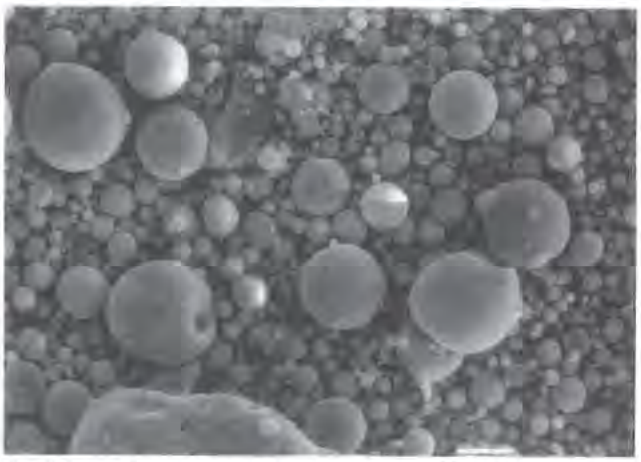


Fig. 1 Flyash from pulverized coal (magnified x 1000).

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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 established stricter regulations than those set by the EPA. There are separate emissions standards for a variety of fired 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 x 10^6 Btu/h (73.3 MW_t), one for the 100 to 250 x 10^6 Btu/h (29.3 to 73.3 MW_t) units and a third for those units less than 100 x 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. (See Chapter 32.)

Particulate control equipment

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, mechanical collectors and venturi scrubbers. Each of these uses a different collection process with different factors affecting the collection performance.

Electrostatic precipitators

An electrostatic precipitator (ESP) electrically charges the ash particles in the flue gas to collect and remove them. The unit is comprised of a series of parallel vertical plates through which the flue gas passes. Centered between the plates are charging electrodes which provide the electric field. Fig. 3 is a plan view of a typical ESP section which indicates the process arrangement.

Charging

The collecting plates are typically electrically grounded and are the positive electrode components. The discharge electrodes in the flue gas stream are connected to a high voltage power source, typically 55 to 75 kV DC, with a negative polarity. An electric field is established between the discharge electrodes and the collecting surface. 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 collection plates 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 collection plates and reduce the likelihood of re-entrainment. In addition, a series of plates and discharge electrodes are necessary to maximize overall particulate collection by increasing the opportunities of the individual ash particles to be charged and collected.

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 particles agglomerate or cling together.

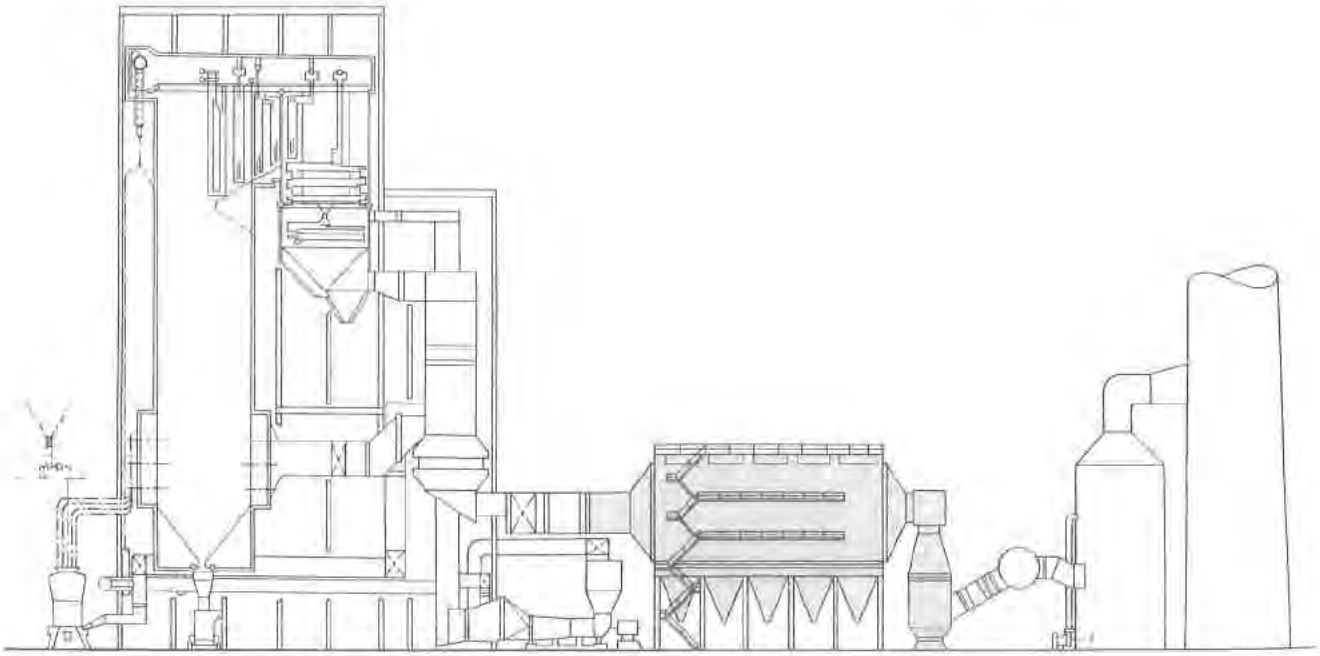


Fig. 2 Particulate control equipment — plant side view.

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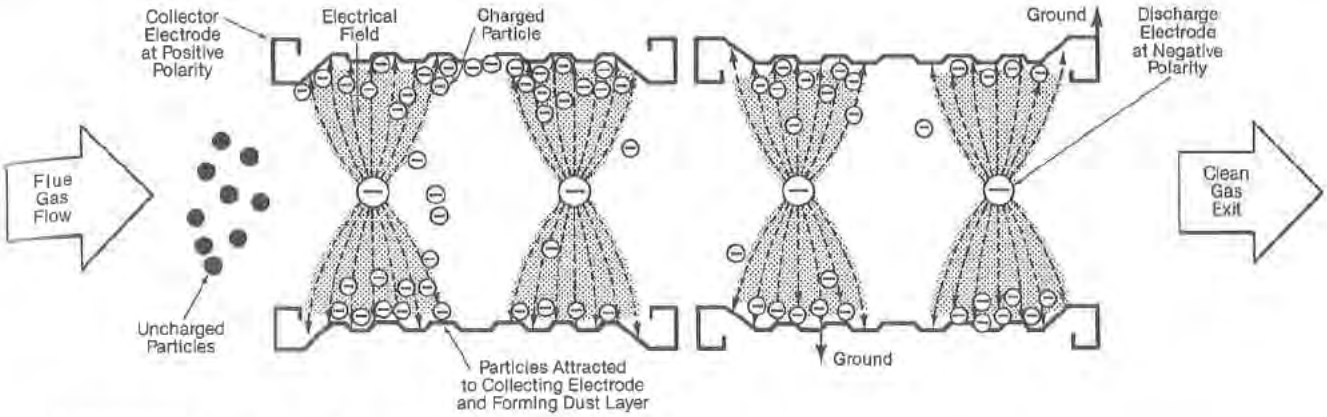


Fig. 3 Particle charging and collection within an ESP.

Rapping

The ash layer must be periodically removed. The most common removal method is rapping which consists of suddenly 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 collection downstream.

While most of the particles are driven to the collection surface, some positively charged particles attach to the discharge electrodes. A separate rapping system is therefore used to clean these electrodes to maintain the maximum charging forces.

Ash removal

The dislodged particulate falls from the collection surface into hoppers. Once the particulate has reached the hopper it is important to ensure that it remains there until the hopper is emptied. (See Chapter 23 for hopper ash removal methods and equipment.)

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]

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

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

or

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

where

E = ESP removal efficiency, %
w = migration velocity, ft/min (m/s)
A = collection surface area, ft^2 (m^2)
V = gas flow, ft^3/min (m^3/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. 4 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.

Fuel and ash characteristics The fuel and ash constituents which reduce resistivity or which 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.

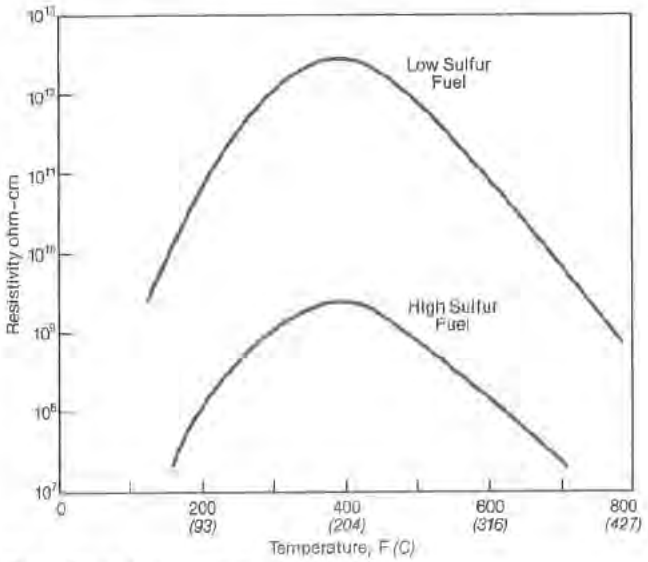


Fig. 4 Typical ash resistivity.

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The fuel and ash constituents and their relative quantities must be reviewed in the sizing 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.

Operating conditions The design operating conditions from the boiler or process also affect precipitator sizing and performance. As indicated on the resistivity curve, 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.

In addition to the quantity of particulate sent to the precipitator, particle size also affects ESP design and performance. A particle size distribution versus collection efficiency curve (Fig. 5) 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.

Flow distribution 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 Industrial Gas Cleaning Institute EP-7.[5]

Precipitator components

All ESPs have several components in common (Fig. 6) although there is some shape and size variation between units.

Discharge electrodes As described in the section on charging, the discharge electrodes, 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 gap of 6 to 8 in. (152 to 203 mm), depending on plate spacing, must be maintained throughout the ESP between the discharge and the grounded components.

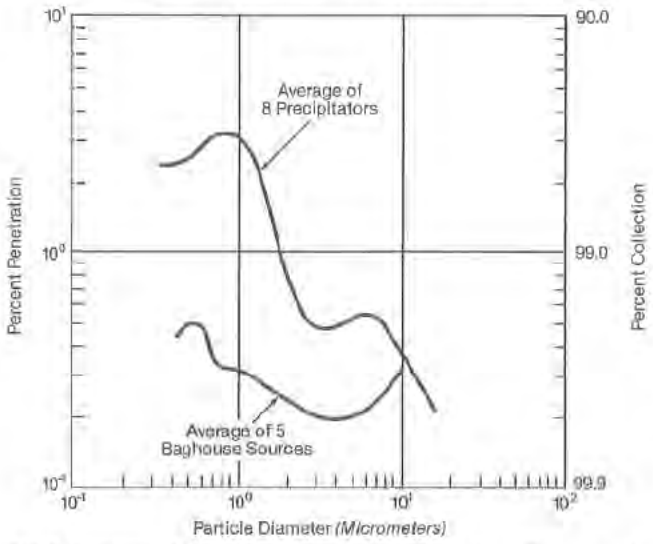


Fig. 5 Summary of fine particulate collection (adapted from Reference 4).

Discharge electrodes are found in several shapes. Common types include the rigid frame, rigid electrode and weighted wire. The rigid frame, shown in Fig. 7, 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 which 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 electrode, or weighted wire design, consists of a round discharge electrode (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, which results in performance deterioration.

Collection surface As previously stated, the collection surface area in the Deutsch-Anderson equation is the total plate area required for particulate collection. Shown in Fig. 6, the collection surface typically consists of a series of roll formed collector plates assembled into a curtain and supported from the top. The curtains are spaced in rows across the width of the precipitator, typically on 12 or 16 in. (305 or 406 mm) centers. In depth, the curtains are arranged into fields. For calculating surface area, the curtain assembly is treated as a plane and includes both sides of the plates. The rolled plates can be up to 50 ft (15.2 m) in length with a shop straightness tolerance of 0.5 in. (12 mm). Collector curtains may also be large flat plates with stiffener bars added to maintain straightness. For optimum performance with a uniform electric field and with no electrical arcing (high current spark), the alignment of collection surface and electrodes must be maintained within tight tolerances.

Rapping systems As shown in Fig. 6, 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 frequency and optimum cleaning. Hammer size is selected to match the application and size of the collector curtain. Although not as effective for many applications, top rapping of the collection surface is also popular. 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.

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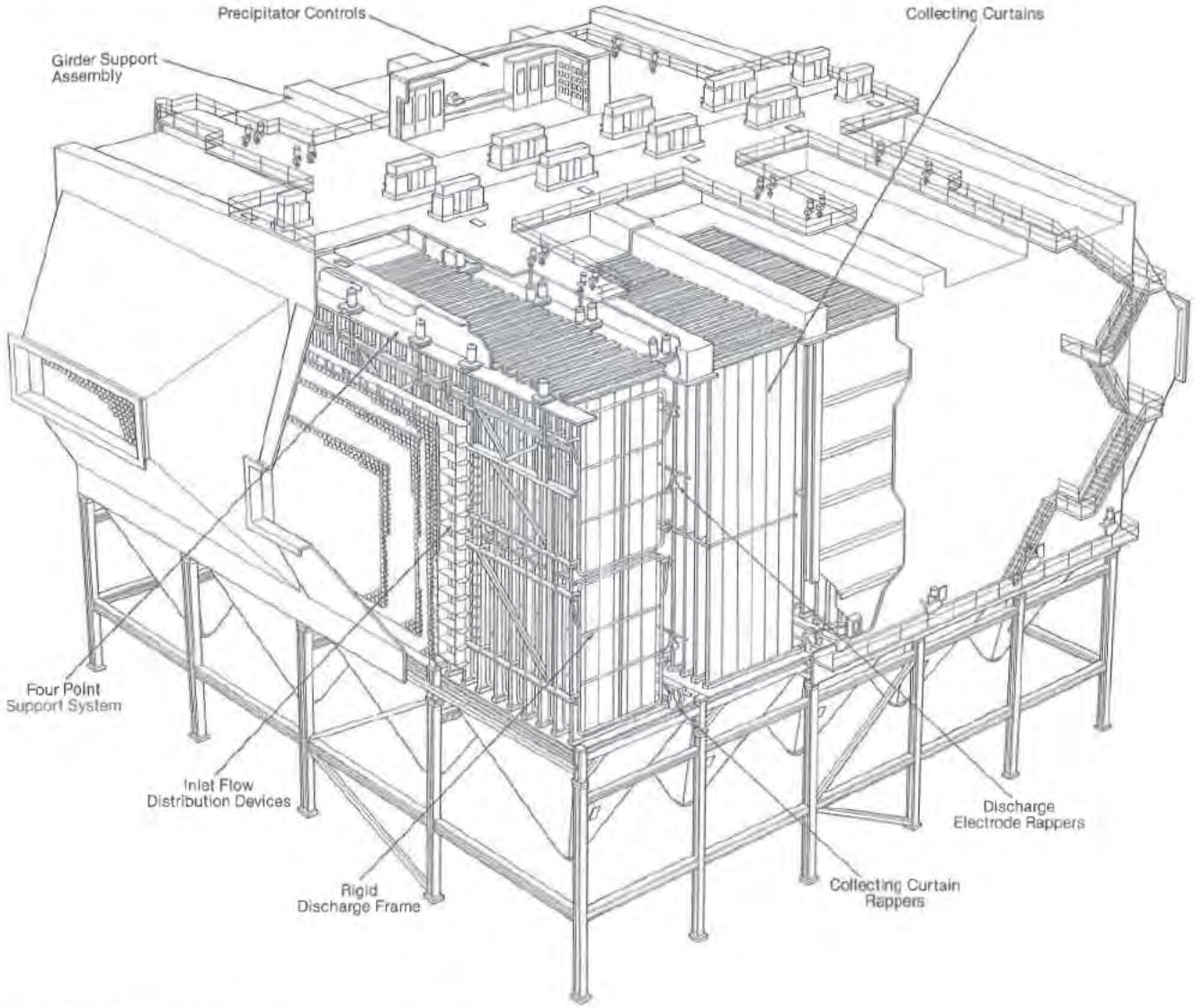


Fig. 6 B&W/Rothemuhle rigid frame electrostatic precipitator.

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's 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. 7. 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.

A rapping system is sometimes used on the flow distribution devices at the precipitator inlet. On those applications where the particulate is sticky or tends to accumulate, a rapping system is needed to keep the surfaces clean and to allow the distribution devices to provide uniform flow to the precipitator inlet at an acceptable pressure loss.

Enclosure As shown in Fig. 6, 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 Metal pyramid shaped hoppers are supported from the lower grid and are made of externally stiffened

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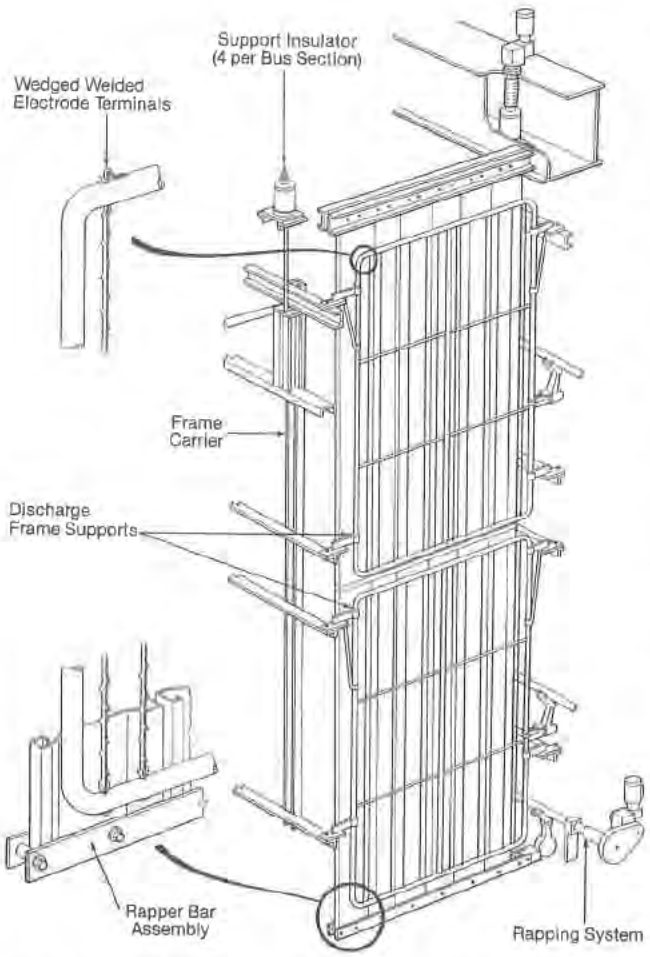


Fig. 7 Rigid frame discharge electrode and rapping system for an ESP.

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

Because many ash removal systems are noncontinu-ous, 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.

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 V AC 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 electronics with quick response times, interface advantages and programming capabilities provide many functions to optimize particulate collection.

Applications and performance

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 which 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 medium and higher ash coals with outlet emissions levels of 0.01 to 0.03 lb/10^6 Btu (12.3 to 36.9 mg/Nm^3 at 6% O_2) heat input common on all coals.

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 resistivity of the ash in the flue gas is typically lower 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.

Pulp and paper In the pulp and paper industry, precipitators are used on power boilers and 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. Refer to Chapter 26 for 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 ash to stick together. The resistivity of the particulate is low so it is collected easily in the ESP. 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 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 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.

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

Performance enhancement

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, gas conditioning, improved flow distribution, 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 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 conditioning 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 1975 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

The collection of acid mists consisting of fine particulate has been accomplished with wet ESPs in some industrial processes. These units differ from the dry, or conventional, ESPs in materials; however, the collection mechanism is basically the same. Typical operation is at the flue gas moisture dew point temperature and particulate loading is low compared to normal coal-fired applications. To withstand the corrosive atmosphere, the construction materials are critical. Typical materials for the wet ESP components contacting the flue gas include:

1. Discharge system — Alloy 276
2. Collection surface — Alloy 276, lead lined, plastic
3. Enclosure — lead lined, acid brick, plastic or plastic coated

Instead of a rapping system, water spray or a water film removes the collected particulate. Most applications are small industrial units; on larger units a modularized concept is used because the configuration and shapes of the components are nonstandard compared to the dry ESP. Collection efficiencies of 99% have been reported with wet ESPs when precipitator sections or modules are placed in series.

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 (Fig. 8) 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 23. 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 control 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.[6] 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. 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.

Two fabric filter design parameters are air/cloth (A/C) ratio and drag. A/C ratio is the gas volumetric flow rate divided by the exposed bag surface area. Industry standards, along with operating experience, establish the design A/C ratios. These values are typically stated with one compartment out of service for cleaning (net condition). The pressure drop includes the drop across the bags, the dustcake and the attachment of the bag to the tubesheet. The calculation of drag for each compartment is useful in evaluating performance.

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 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 the flow from outside to inside (Fig. 9). Note that the tubesheet on the inside-bag filtering is located below the bags and for pulse jet the tubesheet is above the bags.

A reverse air, more correctly termed reverse gas, fil-

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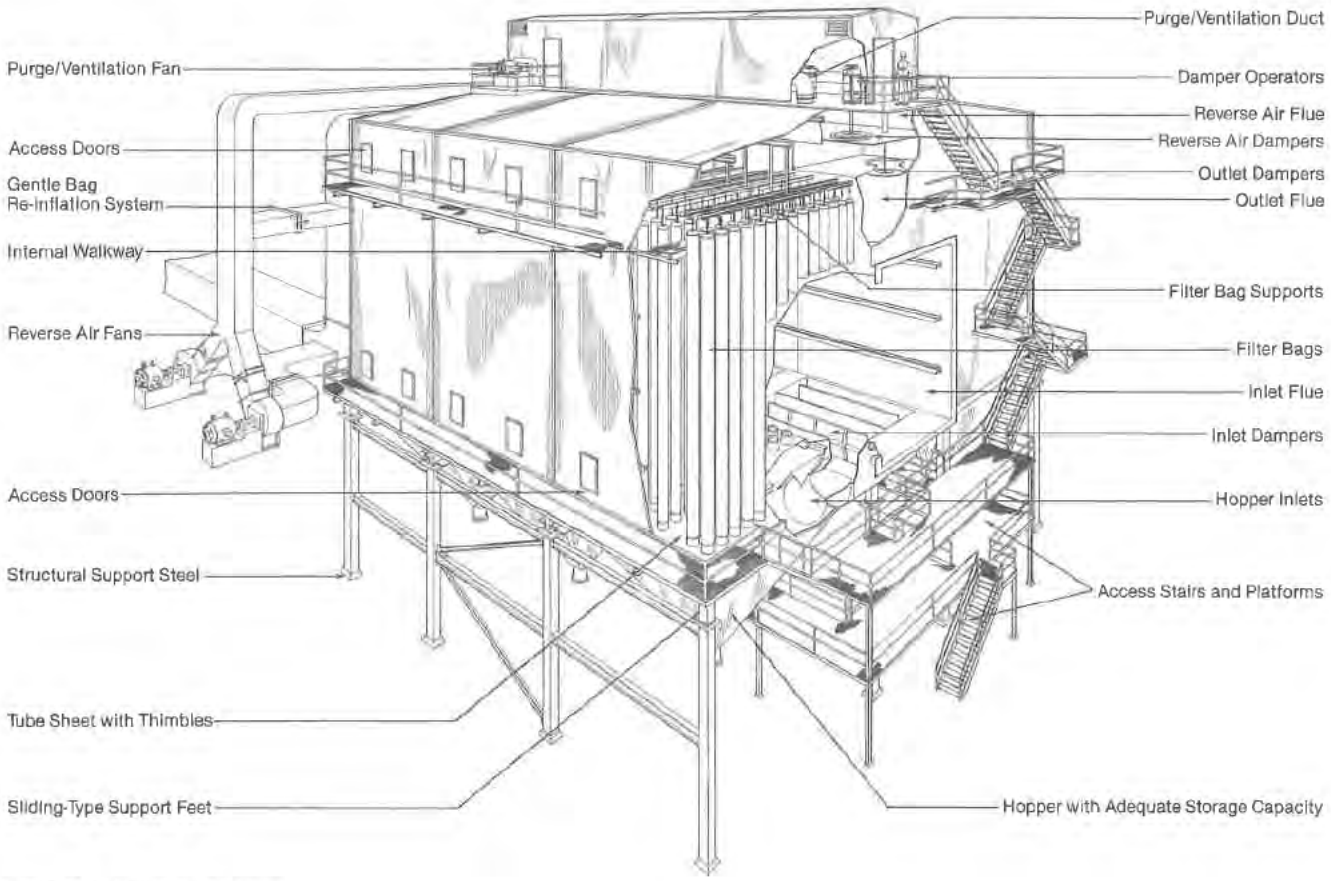


Fig. 8 Fabric filter or baghouse.

ter 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 re-inflated 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. 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 horns to each compartment to assist in the cleaning.

Shake deflate 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. The initial experience with pulse jet filters has been on industrial units, with limited use on utility units to date; however, use of pulse jet units in utility boilers is increasing.

Bag materials and supports Substantial research and development on bags and their materials has taken place to lengthen their life and to select bags for various applications. 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 improve 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 sown into the bag and the bag has a spring loaded support for the reverse air filters. The upper operating temperature limit is 500F (260C) for most fiberglass bags.

In addition to fiberglass, the industrial size filters use synthetic materials with trade names like Nomex®, Ryton®, Gortex® and Hyglass®. Advantages of the synthetic materials include better abrasion resistance and

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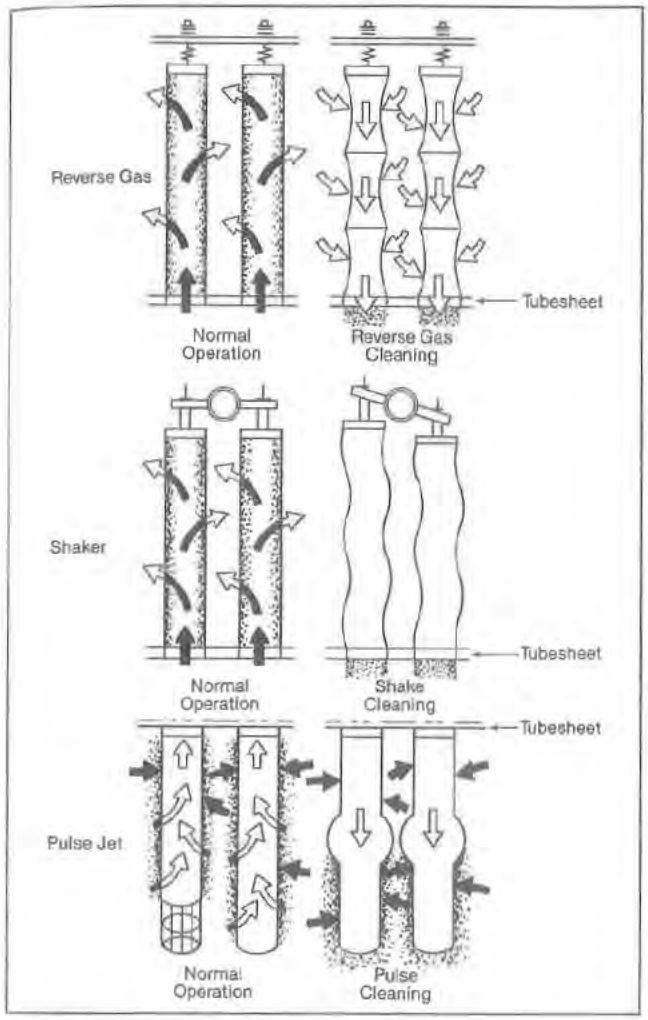


Fig. 9 Fabric filter types.

resistance to acid attack. Disadvantages include higher cost and limited temperature capabilities. For the pulse jet filters, the typical bag size is 5 or 6 in. (130 or 150 mm) diameter with a length of 10 to 20 ft (3 to 6 m).

Enclosure 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. 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.

Performance and applications

The first utility fabric filter in the U.S. was installed with Babcock & Wilcox's participation on an oil-fired Southern California boiler in the mid 1960s.[7] After a few years of limited activity, several utilities had installed fabric filters by the late 1970s. Interest in these systems continues to grow due to the high particulate removal efficiency. They are the preferred collector for some applications. Well designed filters routinely achieve greater than 99.9% particulate removal, meeting all U.S. EPA and local regulations.

Besides the 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 steel industry. A unique advantage with fabric filters in this application 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.)

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%.

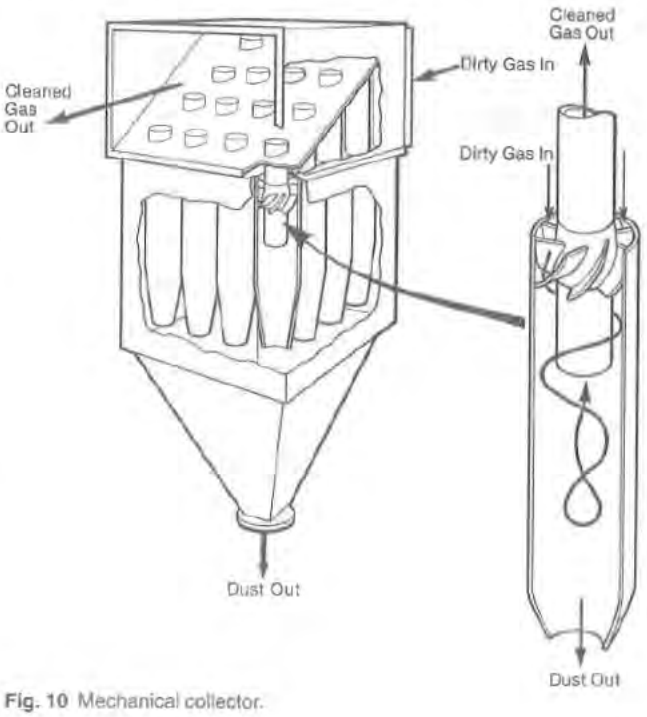


Fig. 10 Mechanical collector.

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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 re-injection 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 Chapters 16 and 29 for more information on fluidized-bed combustion.)

Wet 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 scrubbers are used infrequently as the primary collection device. However, on most coal-fired applications, wet scrubbers are required in series with a high efficiency collector for control of acid gas emissions, so the extra particulate removal is an added benefit.

Other collection devices

Other more specialized particulate control devices include ELECTROSCRUBBER® filters and ceramic tube filters. The ELECTROSCRUBBER® filter combines the technologies of granular filtration and electrostatic collection and uses electrostatic forces on the particulate as the flue gas passes through a recirculating bed of gravel. Charged particles attach to the gravel which is cleaned as the lower bed gravel is recirculated to the top of the bed. The collected dust is sent to disposal and the cleaned gas stream flows to the stack. This collection device is popular on wood-fired units because it aids fire prevention by its handling of glowing embers.

Ceramic tube filters are being developed for high temperature and pressure applications such as coal gasifiers and pressurized fluidized-beds. (See Chapters 17 and 29.)

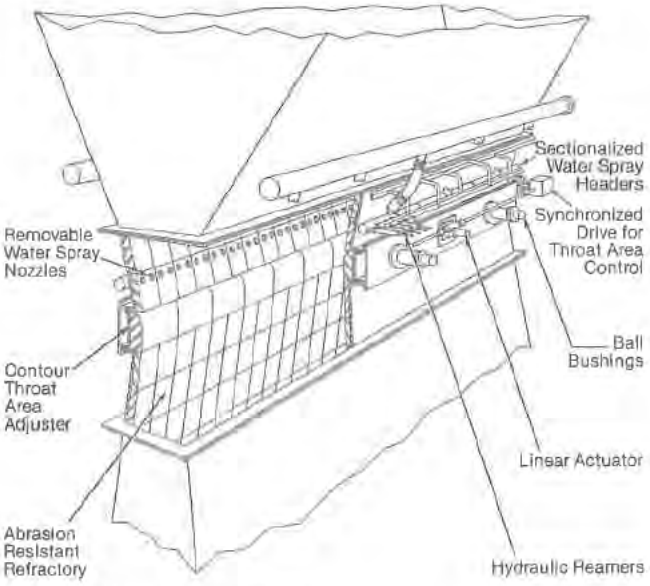


Fig. 11 Venturi-type wet scrubber.

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 which 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. A comparison of overall and particle size collection efficiencies for precipitators and fabric filters is shown in Fig. 5. An application where small particulate dominates would favor a fabric filter for maximum control as long as bag blinding does not occur.

For those applications where an ESP or fabric filter is technically acceptable and high collection efficiencies are required, some general guidelines on capital costs are: 1) on small units, a pulse jet fabric filter is more economical, 2) on large units with medium or high sulfur coal, an ESP is economical, and 3) on low sulfur coal-fired large units, a reverse air fabric filter may be more economical. However, when operating and maintenance costs are also considered, the lowest capital cost may not provide the lowest overall cost. Therefore, it is important to perform a detailed engineering study to quantify all of the variables for a specific site to obtain a true assessment of the real cost.

International particulate control

Particulate control regulations and control equipment usage have been emphasized in Western Europe, Scandinavia, South Africa, Australia, Japan and Canada, as well as the U.S. The traditional high efficiency collection device in all countries has been the ESP for many applications; however, the fabric filter has gained acceptance in Australia and the U.S. for coal-fired units. On new refuse-fired

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units, the combination of spray dryer and fabric filter is popular in Scandinavia, Western Europe and the U.S.

Control challenges for the future

Particulate control will be required for steam generation combustion equipment as long as ash bearing fuels are used. New combustion methods may change the ash characteristics, permitting less costly or more efficient control equipment. Furthermore, new material developments, increased use of electronics and overall equipment size reduction are expected to continue.

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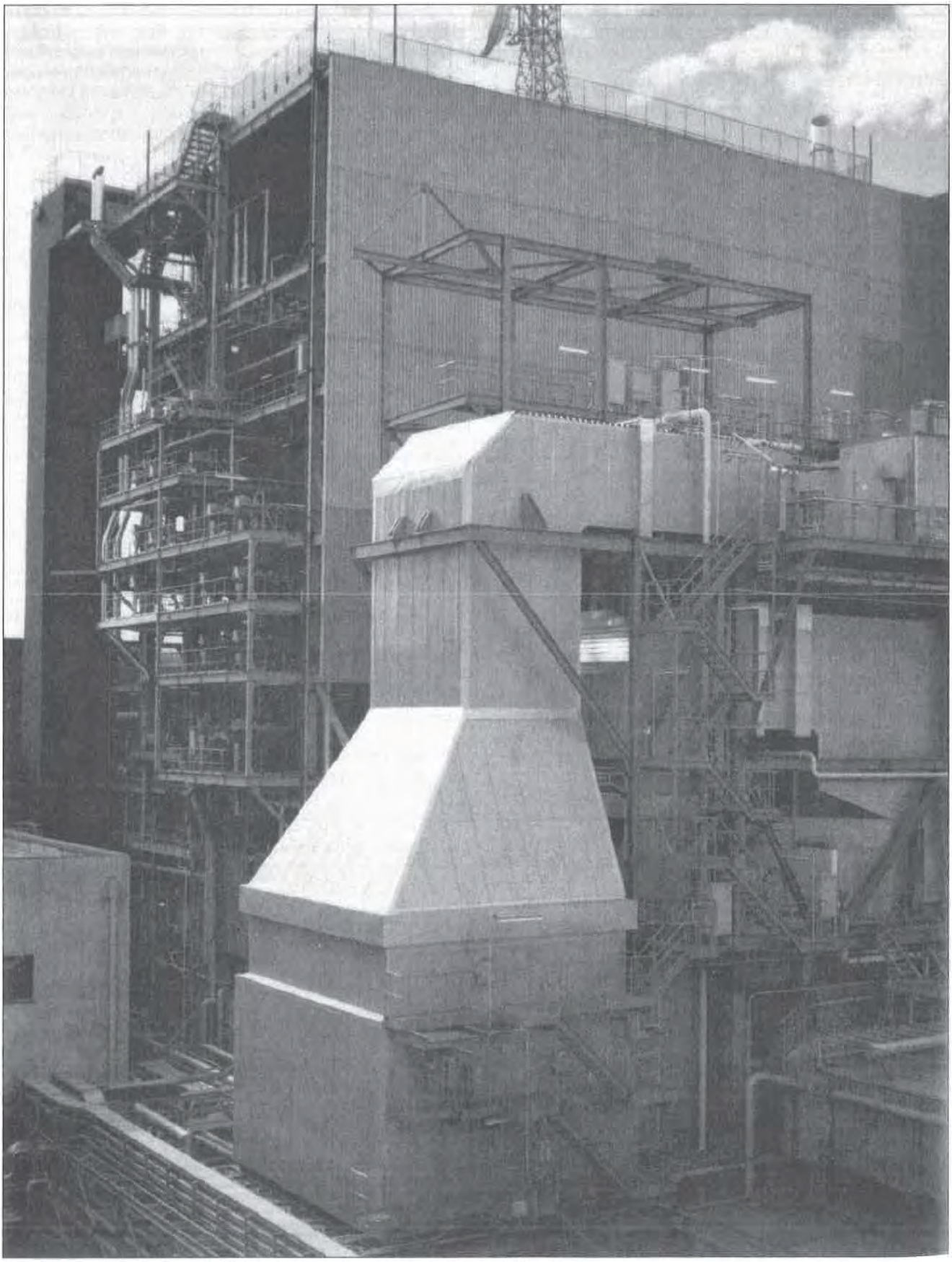
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Selective catalytic reduction (SCR) NO_x emission control system retrofit on a coal-fired boiler.

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Chapter 34
Nitrogen Oxides Control

Nitrogen oxides (NO_x) are one of the primary pollutants emitted during combustion processes, with transportation systems accounting for the largest contribution in the United States. (See Fig. 1.) Along with sulfur oxides (SO_x) and particulate matter, NO_x emissions have been identified as contributors to acid rain and ozone formation, visibility degradation and human health concerns. As a result, NO_x emissions from most combustion sources are regulated and require some level of control as discussed in Chapter 32. A number of approaches can be used to control NO_x emissions.

NO_x formation mechanisms

NO_x refers to the cumulative emissions of nitric oxide (NO), nitrogen dioxide (NO_2) and trace quantities of other species generated during combustion. Combustion of any fossil fuel generates some level of NO_x due to high temperatures and the availability of oxygen and nitrogen from both the air and fuel.

NO_x emissions from fired processes are typically 90 to 95% NO with the balance being NO_2. However, once the flue gas leaves the stack, the bulk of the NO is eventually oxidized in the atmosphere to NO_2. It is the NO_2 in the flue gas which creates the brownish plume often seen in a power plant stack discharge. Once in the atmosphere, the NO_2 is involved in a series of reactions which form secondary pollutants. The NO_2 can react with sunlight and hydrocarbon radicals to produce photochemical (urban) smog and acid rain constituents.

There are two common mechanisms of NO_x formation, thermal NO_x and fuel NO_x.

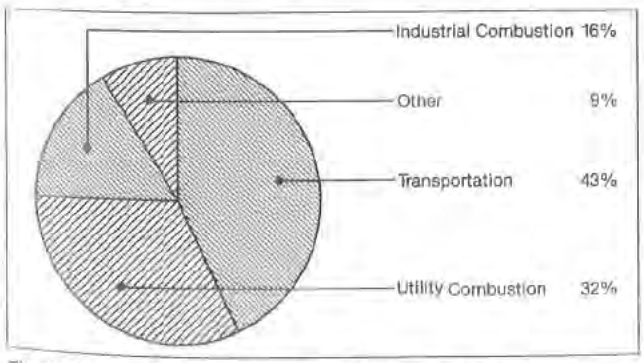


Fig. 1 Sources of United States NO_x emissions — 1985.

Thermal NO_x

Thermal NO_x refers to the NO_x formed through high temperature oxidation of the nitrogen found in the combustion air. The formation rate is a strong function of temperature as well as the residence time at temperature. Significant levels of NO_x are usually formed above 2200F (1204C) under oxidizing conditions with exponential increases as the temperature is increased. At these high temperatures, molecular nitrogen (N_2) and oxygen (O_2) in the combustion air dissociate into their atomic states and participate in a series of reactions. One product of these reactions is NO. The three principal reactions proposed for this process are:[1,2]

$$N_2 + O \rightarrow NO + N \quad (1)$$

$$N + O_2 \rightarrow NO + O \quad (2)$$

$$N + OH \rightarrow NO + H \quad (3)$$

The traditional factors leading to complete combustion (high temperatures, long residence time and high turbulence or mixing) all tend to increase the rate of thermal NO_x formation. Therefore, some compromise between effective combustion and controlled NO_x formation is needed.

Thermal NO_x formation is typically controlled by reducing the peak and average flame temperatures. This can be accomplished through a number of combustion system changes. Controlled mixing burners can be used to reduce the turbulence in the near burner region of the flame and to slow the combustion process. This typically reduces the flame temperature by removing additional energy from the flame before the highest temperature is reached. A second approach is staged combustion where only part of the combustion air is initially added to burn the fuel. The fuel is only partially oxidized and then cooled before the remaining air is added separately to complete the combustion process. A third alternative is to mix some of the flue gas with the combustion air at the burner, referred to as flue gas recirculation. This increases the gas weight which must be heated by the chemical energy in the fuel, thereby reducing the flame temperature.

These technologies have been used effectively with gas, oil and coal firing to reduce NO_x formation. Specific use of each technology or a combination of technologies depends upon the costs, fuel and regulatory requirements. The specific applications of these different combustion

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system modifications are presented in more detail in Chapters 10 and 13.

For fuels which do not contain significant amounts of chemically bound nitrogen, such as natural gas, thermal NO_x is the primary overall contributor to NO_x emissions. In these cases, the technologies described above are particularly effective in NO_x emission control.

Fuel NO_x

The major source of NO_x emissions from nitrogen bearing fuels such as coal and oil is the conversion of fuel bound nitrogen to NO_x during combustion. Laboratory studies indicate that fuel NO_x contributes approximately 50% of the total uncontrolled emissions when firing residual oil, and more than 80% when firing coal. Nitrogen found in fuels such as coal and residual oils is typically bound to the fuel as part of organic compounds. During combustion, the nitrogen is released as a free radical to ultimately form NO or N_2. Although it is a major factor in NO_x emissions, only 20 to 30% of the fuel bound nitrogen is converted to NO.

The majority of NO_x formation from fuel bound nitrogen occurs through a series of reactions which are not fully understood. It does appear, however, that this conversion occurs by two separate paths. These paths, primarily for coal, are represented in Fig 2.

The first path involves the oxidation of volatile nitrogen species during the initial phase of combustion. During the release, and prior to the oxidation of the volatile compounds, nitrogen reacts to form several intermediate compounds in the fuel rich flame regions. These intermediate compounds are then oxidized to NO or reduced to N_2 in the postcombustion zone. The formation of either NO or N_2 is strongly dependent on the local fuel/air stoichiometric ratio. It is estimated that this volatile release mechanism accounts for 60 to 90% of the fuel NO_x contribution.

The second reaction path involves the release of nitrogen radicals during combustion of the char fraction of the fuel. These reactions occur much more slowly than the reactions involving the volatile species.

Conversion of fuel bound nitrogen to NO_x is strongly dependent on the fuel/air stoichiometry but is relatively independent of variations in combustion zone temperature. Therefore, this conversion can be controlled by reducing oxygen availability during the initial stages of combustion. With this reduction, a major portion of the nitrogen released during devolatilization reduces to N_2. This is because hydrocarbon radicals, also released during devolatilization, compete with the nitrogen for the available free oxygen. Techniques such as controlled fuel-air mixing and staged combustion provide a significant reduction in NO_x emissions by controlling stoichiometry in the initial devolatilization zone.

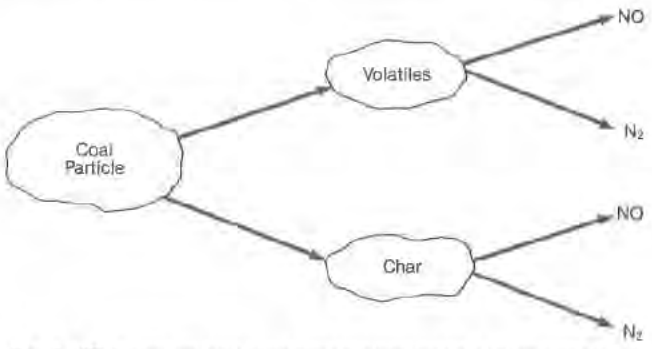


Fig. 2 NO_x formation from nitrogen contained in coal during combustion.

A portion of the NO_x that is formed by oxidation of the fuel bound nitrogen under fuel rich conditions is referred to as *prompt* NO_x, identified through the research of C.P. Fenimore.[3] The name is derived from its formation very early during the combustion process. It was originally identified as the difference between observed and predicted thermal NO emissions.

Prompt NO_x occurs through the formation of intermediate hydrogen cyanide (HCN) species and the reaction between molecular nitrogen and hydrocarbon compounds. This reaction is then followed by the oxidation of HCN to NO.

Although prompt NO_x formation normally has a weak temperature dependence, this dependence can become strong under fuel rich conditions. Modern burners are designed to reduce peak flame temperatures by controlling the rate of fuel and air mixing. Combustion is initiated under fuel rich conditions and this fuel rich zone is where prompt NO_x is formed. Prompt NO_x can contribute from near zero to more than 100 ppmv of NO.

Health and environmental effects of NO_x

Once released into the atmosphere, NO_x becomes a contributor to ozone/photochemical smog, acid rain, particulate matter and potential airborne carcinogens. Each of these pollutants has a significant effect on human health and the environment. However, as discussed in Chapter 32, automobiles and other vehicles are the major source of NO_x with utility power plants contributing less than one third of the total United States (U.S.) emissions.

Ozone/photochemical smog

Naturally occurring ozone, found in the upper atmosphere, is necessary for man's existence. However, man-made ozone, formed in the lower atmosphere, is considered a pollutant. This man-made ozone, formed by a reaction of hydrocarbons and NO_x in the presence of sunlight, is a major component of urban photochemical smog.[4]

Another major component of photochemical smog is NO_2. Once emitted, the NO contained in the flue gas is oxidized to NO_2, which is a yellow-brown gas and is also an oxidant. Urban smog is also made up of a variety of other components.

Acid rain

NO_x and sulfur dioxide (SO_2) contribute to the formation of acid rain,[4] which includes a dilute solution of sulfuric and nitric acids with small amounts of carbonic and other organic acids. The NO_x and SO_2 react with water vapor to form acidic compounds; these acids account for more than 90% of acid rain. Nitric acid is formed through the reaction:

$$NO_2 + OH \rightarrow HNO_3 \qquad (4)$$

Most acid rain control has concentrated on SO_2 contributions instead of those due to NO_x because it is estimated that NO_x contributes less than one third of the acid rain generated.

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Particulate matter

NO_x can also contribute to ambient particulate matter. In the atmosphere NO_x reacts with other airborne chemicals to create nitrates. NO_x also promotes the transformation of SO_2 into sulfate particulate compounds.

Controlling NO_x emissions in steam generating equipment

An understanding of the mechanisms and chemical reactions that produce NO_x emissions has enabled engineers and scientists to develop techniques for reducing emissions of this pollutant. Because NO_x is formed during combustion, research has centered on controlling NO_x at the source. The application of techniques that control fuel-air mixing rates and peak combustion temperatures has been effective in reducing these emissions.

NO_x emissions may be controlled through careful fuel selection and selective fuel switching. Once the fuel is chosen, NO_x emissions are minimized by using low NO_x combustion technology and postcombustion techniques. Typically, the fuel is selected by overall economic considerations. Combustion technology based NO_x control is then typically the lowest overall cost approach provided that it will meet local and federal emissions requirements. Where further control is needed, postcombustion NO_x control is added.

Precombustion fuel effects

Unlike sulfur and particulate constituents in coal, nitrogen species contained in the fuel can not be easily reduced or removed. Therefore, the most common precombustion option for reducing NO_x levels is to switch to a fuel with inherently lower nitrogen content. For boilers capable of firing multiple fuels with minimal impact on the steam cycle, this may be the most cost effective solution.

Coal combustion generally produces the highest NO_x emissions. Oil combustion generates less NO_x while gas firing produces even less. When firing oil fuels, a reduction in fuel nitrogen results in reduced NO_x emissions. However, there does not appear to be a similar correlation between coal nitrogen content and NO_x generation. Other factors in coal chemistry including volatile species, oxygen and moisture content appear to dominate the formation of NO_x during coal combustion. Therefore, reducing coal nitrogen content may not provide a corresponding NO_x reduction.

Combustion techniques

NO_x formation is promoted by rapid fuel-air mixing. This produces high peak flame temperatures and excess available oxygen which, in turn, promotes NO_x emissions. Combustion system developments responsible for reducing NO_x formation include low NO_x burners, such as the coal-fired burner shown in Fig. 3, staged burning techniques, and flue gas recirculation (FGR). The specific NO_x reduction mechanisms include controlling the rate of fuel-air mixing, reducing oxygen availability in the initial combustion zone, and reducing peak flame temperatures. Modern systems can emit less than one third of the NO_x produced by older units.

Additional information regarding the details of the combustion process, combustion techniques and hardware used to control NO_x can be found in Chapters 10, 13, 15, 16, 17 and 29. Information on uncontrolled and controlled NO_x emissions is also presented.

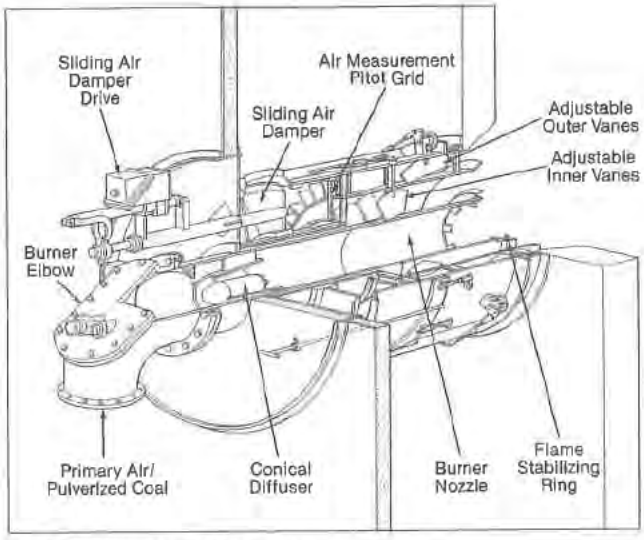


Fig. 3 DRB-XCL™ low NO_x burner for coal firing.

Postcombustion techniques

Many localities require lower NO_x emissions than economically possible by combustion modifications alone. To achieve further reductions, additional techniques are applied downstream of the combustion zone. These postcombustion control systems are referred to as selective noncatalytic reduction (SNCR) and selective catalytic reduction (SCR). In either technology, NO_x is reduced to nitrogen (N_2) and water (H_2O) through a series of reactions with a chemical agent injected into the flue gas.

The most common chemical agents used in commercial applications are ammonia and urea for SNCR and ammonia for SCR systems. Most ammonia-based systems have used anhydrous ammonia (NH_3) as the reducing agent. However, due to the hazards of storing and handling NH_3, many systems use aqueous ammonia at a 25 to 28% concentration. Urea [$(NH_2)_2CO$] can be stored as a solid or mixed with water and stored in solution. An additional byproduct of the urea injection is carbon dioxide (CO_2).

Selective noncatalytic reduction technology

There are currently two basic selective noncatalytic reduction (SNCR) processes available. An ammonia-based system has been developed by Exxon and is called Thermal DeNO$_x$®. A urea-based technology has been developed under the sponsorship of the Electric Power Research Institute (EPRI). Although there are distinct differences to each technology, the overall processes are similar.

Both SNCR technologies inject a reducing agent into NO_x laden flue gas within a specific temperature zone or window. In addition, it is important to properly mix the chemical agent with the flue gas. Finally, the mixture must have adequate residence time for the reduction reactions to occur. The chemical reactions for both processes can be represented by:

Ammonia $$4NO + 4NH_3 + O_2 \rightarrow 4N_2 + 6H_2O \quad (5)$$

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Urea $2NO + (NH_2)_2CO + 1/2\ O_2 \rightarrow 2N_2 + 2H_2O + CO_2$ (6)

The acceptable temperature range for either reaction is 1400 to 2000F (760 to 1093C), although temperatures above 1700F (927C) are preferred. Below 1600F (871C), chemical enhancers, such as hydrogen, are needed to assist the reactions. As the temperature increases within this range, the ammonia or urea may react with available oxygen to form NO_x. This reaction becomes significant at temperatures in excess of 2000F (1093C) and may become dominant as temperatures approach 2200F (1204C).

The application of SNCR technology to a municipal solid waste boiler is shown in Fig. 4. Multiple injection levels are used to maintain NO_x reduction efficiencies at acceptable levels as boiler load changes. Based on operating experience, four or more injection levels may be required for larger units. Multiple levels are required because the flue gas temperature profile changes with boiler load and the injection point for the ammonia or urea must be adjusted accordingly.

The SNCR system consists of storage and handling equipment for the ammonia or urea, equipment for mixing the chemical with the carrier (compressed air, steam or water), and the injection equipment. The key component, the injection system, consists of nozzles generally located at various elevations on the furnace walls to match the expected flue gas operating temperature. The number and location of the nozzles are established by the supplier and are based on obtaining good reagent distribution within the flue gas. Fig. 5 provides a simplified flow diagram for an ammonia-based SNCR system.

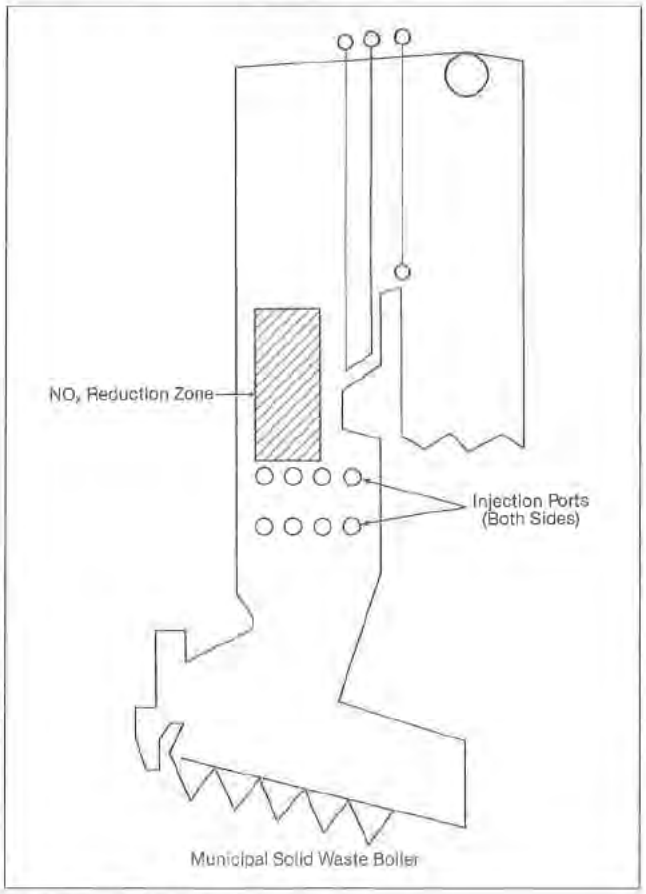


Fig. 4 Typical SNCR application.

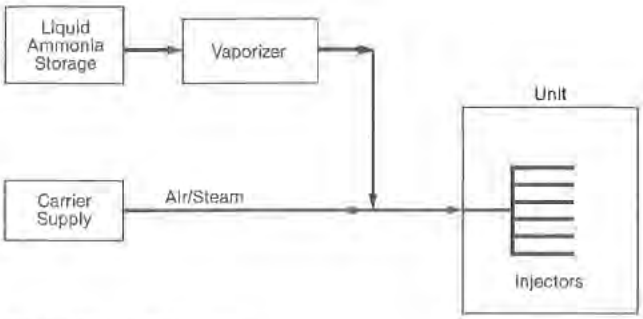


Fig. 5 SNCR deNO$_x$ system.

One major difference between the ammonia and urea based processes is that the ammonia is usually injected into the gas stream in a gaseous state, whereas the urea is injected as an aqueous solution. The urea technology requires a longer residence time for reactions due to the time required to vaporize the liquid droplets once they are in the gas stream. However, as aqueous ammonia systems become more frequently specified, ammonia may also be injected as a liquid with additional residence time required for vaporization.

When injecting ammonia or urea it is important to control the excess unreacted ammonia. As flue gas temperatures are reduced, the excess ammonia can react with other combustion species, primarily sulfur trioxide (SO_3), to form ammonium salts. The major ammonium products formed are ammonium sulfate [$(NH_4)_2SO_4$] and ammonium bisulfate (NH_4HSO_4). Ammonium sulfate is a dry, fine particulate (1 to 3 microns in diameter) that may contribute to plume formation. The ammonium bisulfate is a highly acidic and sticky compound which, when deposited on downstream equipment such as air heaters, contributes to significant fouling and corrosion.

While NO_x reduction levels of 70% are possible under carefully controlled conditions, 30 to 50% reductions in NO_x emissions are more typically used in practice to maintain acceptable levels of reagent consumption and unreacted ammonia carryover.

Most of the current applications for these technologies have been on smaller municipal solid waste or biomass fueled boilers, where the appropriate temperature range is located in the upper furnace, as shown in Fig 4. (See also Chapter 27.) In large utility units, the proper temperature range occurs in the convection pass cavities, making application, especially on retrofits, more challenging. For these applications, overall control may be limited to the 20 to 40% range. New boiler applications may require special design of the boiler convection pass with a dedicated open cavity in an optimal temperature window for the reactions to occur (Fig. 6).

Selective catalytic reduction technology

Fundamental design considerations

Selective catalytic reduction (SCR) systems catalytically reduce flue gas NO_x to N_2 and H_2O using ammonia in a chemical reduction. This technology is the most effective method of reducing NO_x emissions especially where high removal efficiencies (70 to 90%) are required.

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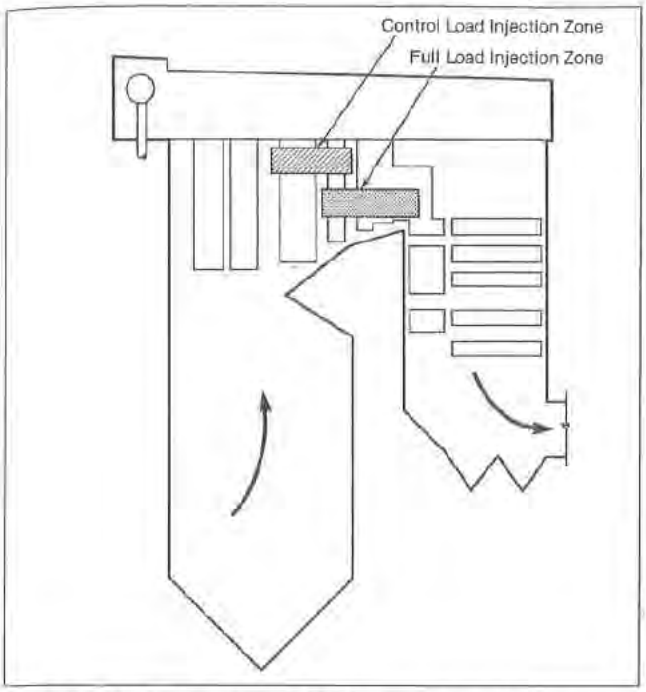


Fig. 6 Coal-fired utility boiler with SNCR temperature window location at full and part load.

Initial development of SCR technology began in Japan around 1963. Early laboratory tests included the evaluation of many catalyst formulations and their life expectancies, as well as process design optimizations. Development tests were followed by several catalytic pilot plant installations during the early 1970s; the first commercial installations occurred in 1978. As mandated by environmental regulations, SCR systems have been installed on utility boilers in Japan, Germany and other European countries. SCR technology has also been applied to U.S. installations primarily for gas turbines, combined cycle systems, and small industrial boilers and process heaters. Legislation passed in areas of California requires the use of this technology on utility installations. Environmental permitting considerations will likely require this technology on many new large installations during the 1990s. Typical examples of new and retrofit applications are shown in Figs. 7 through 9. There are more than 400 SCR systems throughout the world, covering a variety of fuels including natural gas, oil, coal and wood.

The NO_x reduction reactions take place as the flue gas passes through the catalyst chamber. Before entering the catalyst, ammonia is injected into and mixed with the flue gas as shown in Fig. 10. Once the mixture enters the catalyst, the NO_x reactions with NH_3 can be represented as follows:

$$4NO + 4NH_3 + O_2 \rightarrow 4N_2 + 6H_2O \quad (7)$$

$$2NO_2 + 4NH_3 + O_2 \rightarrow 3N_2 + 6H_2O \quad (8)$$

As with SNCR systems, the SCR reactions take place within an optimal temperature range. A variety of catalysts are available. Most can operate within a range of 450 to 840F (232 to 449C), but optimum performance occurs between 675 and 840F (357 and 449C). The minimum temperature varies and is based on fuel, flue gas specifications and catalyst formulation. In addition, this minimum temperature tends to increase with flue gas sulfur dioxide content. This results in a smaller operating range as sulfur content increases in order to eliminate the formation of ammonium sulfate salts in the catalyst bed. Above the recommended temperature range, a number of catalyst materials tend to become less effective.

Catalyst formulation is the key to SCR system performance. Although catalyst formulations are proprietary, material typically falls into one of three categories: base metal, zeolite and precious metal.

Most of the operating experience to date has been with base metal catalysts. These catalysts use titanium oxide with small amounts of vanadium, molybdenum, tungsten or a combination of several other active chemical agents. The base metal catalysts are selective and operate in the specified temperature range. The major drawback of the base metal catalyst is its potential to oxidize SO_2 to SO_3; the degree of oxidation varies based on catalyst chemical formulation. The quantities of SO_3 which are formed can react with the ammonia carryover to form the ammonium sulfate salts as previously discussed. Proper system design and catalyst formulation can minimize this problem.

Zeolite catalysts are relatively new to NO_x emission control. These catalysts are aluminosilicate materials which function similarly to base metal catalysts. One potential advantage of zeolite is its higher operating temperature of 970F (521C). This catalyst may also oxidize SO_2 to SO_3 and must be carefully matched to the flue gas conditions.

Precious metal catalysts are generally manufactured from platinum and rhodium. As with the zeolite units, there is little operating experience with these catalysts. They also require careful consideration of flue gas constituents and operating temperatures. While effective in reducing NO_x, these catalysts can also act as oxidizing catalysts, con-

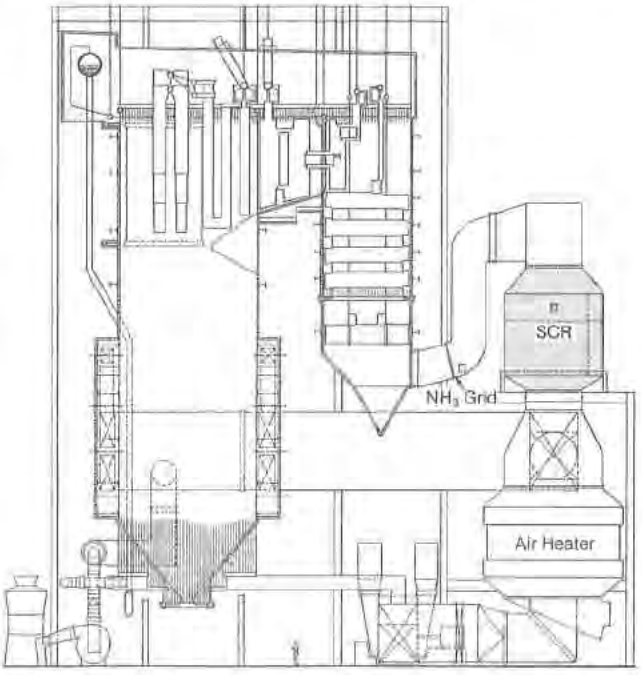


Fig. 7 New utility boiler with SCR.

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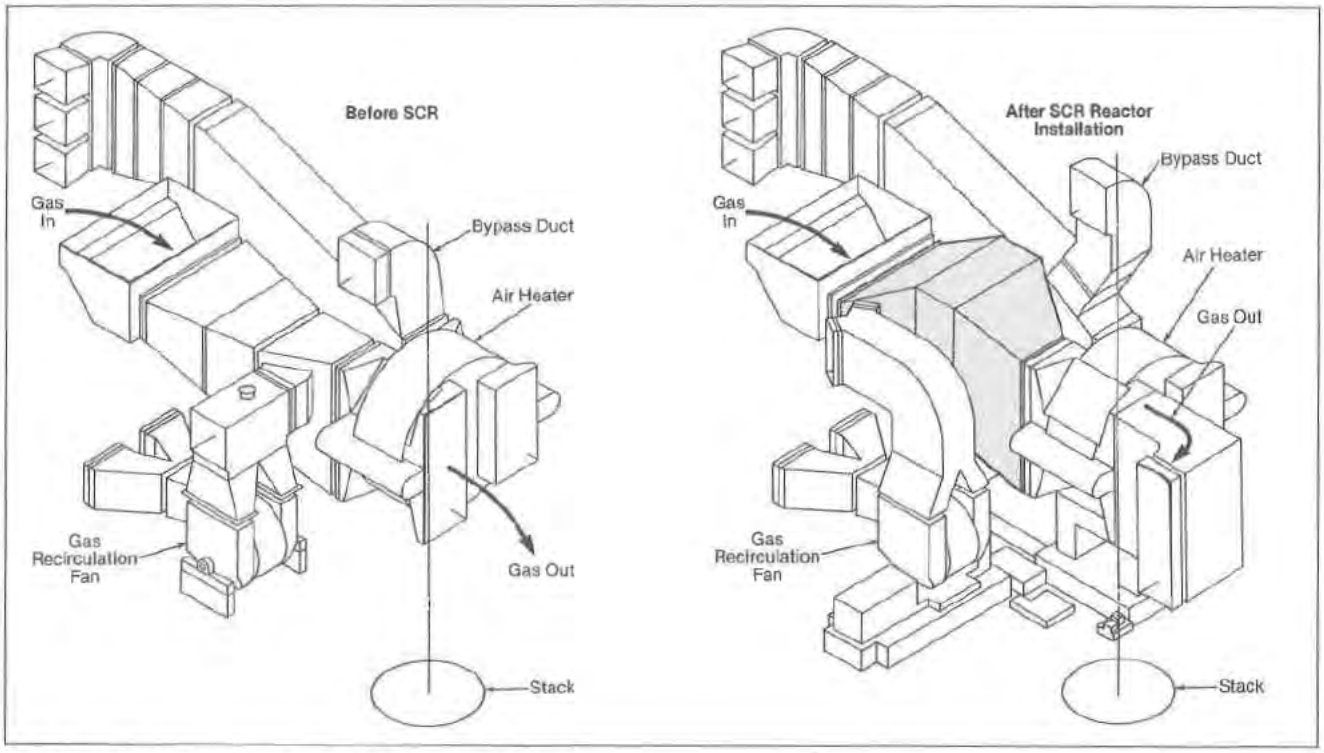


Fig. 8 Retrofit utility boiler application (half of installation shown).

verting CO to CO_2 under proper temperature conditions. However, SO_2 oxidation to SO_3 and high material costs often make precious metal catalysts less attractive.

Early technology used pelletized catalysts in a packed bed configuration. This material was effective for NO_x reduction but could be difficult to handle and frequently added significant pressure drop to the system. As a result, catalyst manufacturing technology evolved to larger, uniform catalyst blocks. Most modern SCR systems use a block type catalyst which is manufactured in the parallel plate or honeycomb configurations shown in Figs. 11 and 12, respectively. For ease of handling and installation, these blocks are fabricated into larger modules.

Each catalyst configuration has its advantages. The plate type unit offers less pressure drop and is less susceptible to plugging and erosion when particulate-laden flue gas is treated in the SCR reactor. The honeycomb configuration often requires less reactor volume for a given overall surface area.

The catalyst is housed in a reactor which is strategically located within the system. This location permits catalyst exposure to proper SCR reaction temperatures. The reactor design includes a sealing system to prevent flue gas bypassing and to provide internal support for the catalyst. The reactor configuration can be vertical or horizontal depending on the fuel used, space available, and upstream and downstream equipment arrangement. Uniform flow distribution is required for maximum performance. Fig. 13 shows a vertical SCR reactor design and Fig. 14 shows a coal-fired boiler installation.

Ammonia and stoichiometry considerations

A diagram of an ammonia control and supply system is shown in Fig. 15. Ammonia can be anhydrous or aqueous, although most commercial experience has been with anhydrous NH_3. However, due to safety concerns, aqueous NH_3 is required for selected SCR system installations.

Ammonia is transported to the site and stored in specially designed tanks. A controlled amount of ammonia



Fig. 9 HRSG with SCR.

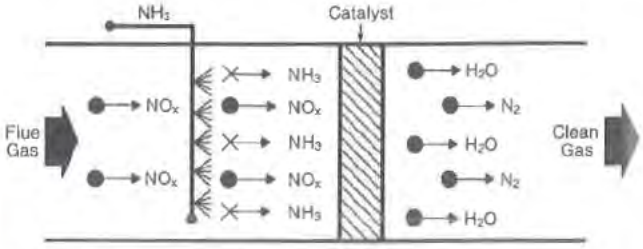


Fig. 10 Principles of NO_x removal process for SCR.

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Fig. 11 Plate-type catalyst block.

is then fed to a vaporizer/mix chamber where it is mixed with air or steam at about a 1:20 volume ratio.

The mixture is then introduced into the flue gas through an injection grid system. The unit-specific grid is designed to provide an even distribution of ammonia throughout the flue gas. The grid configuration is based on the duct size and the distance from the injection grid to the inlet of the catalyst bed. Longer distances require fewer injectors for adequate mixing.

The fundamental process control provides ammonia flow at a constant NH_3/NO_x mole ratio. The product of the inlet NO_x concentration and the boiler flue gas flow yields a NO_x flow signal. Ammonia flow control is then established by multiplying the NO_x flow signal by the NH_3/NO_x mole ratio set point. (See Fig. 15.)

In general, the stoichiometry of NO_x reduction is a 1:1 mole ratio of NH_3 to NO_x. Based on this stoichiometry, for example, a theoretical mole ratio of 0.80 is required for 80% NO_x removal. However, the actual mole ratio required is slightly higher to account for unreacted ammonia carryover from the reactor (NH_3 slip). To account for reaction lag time during abrupt boiler load changes, an ammonia-rich ratio is provided for increasing loads. Similarly, a lean mixture is provided for decreasing loads.

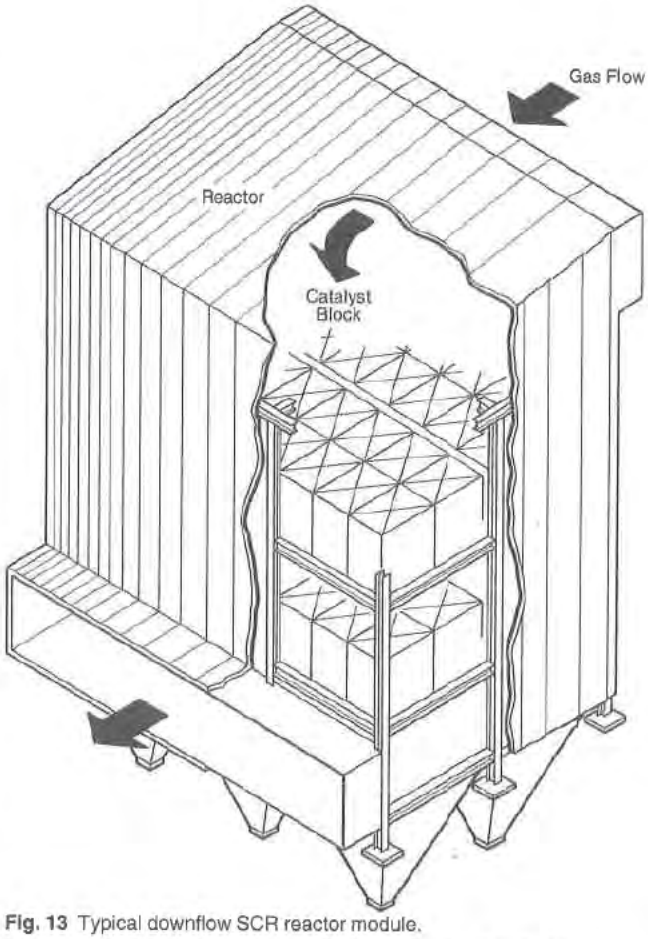


Fig. 13 Typical downflow SCR reactor module.

Related topics

A continuous emissions monitoring (CEM) system is frequently required to monitor all atmospheric pollutants. A detailed discussion of these systems is found in Chapter 36. Data generated from the CEM system can



Fig. 12 Honeycomb catalyst geometry.

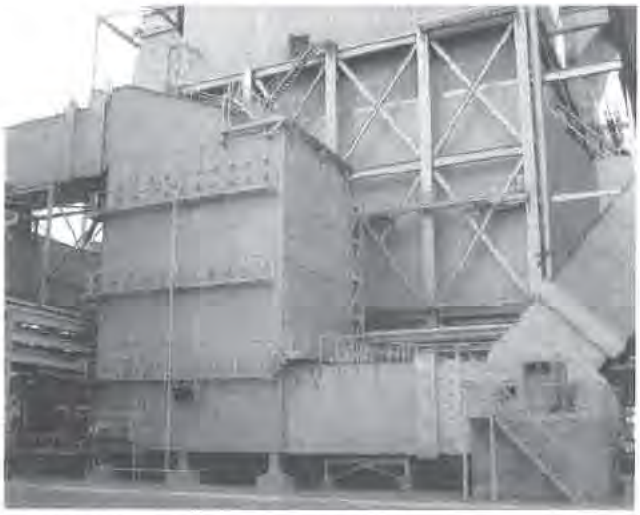


Fig. 14 SCR installation on a coal-fired boiler.

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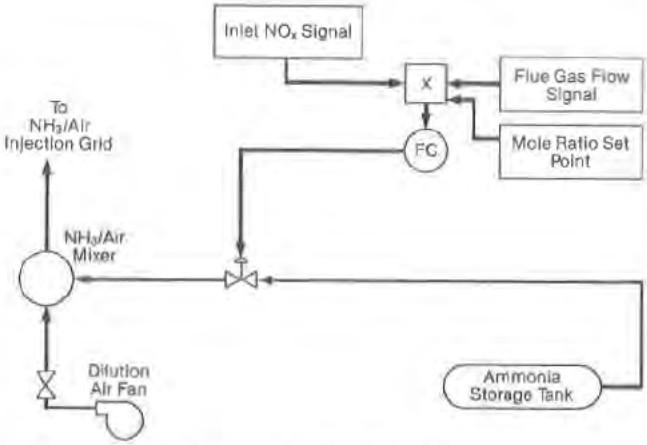


Fig. 15 NH_3/dilution air supply and control system.

be used to control the ammonia flow while achieving the required NO_x emissions level.

The design of each SCR system is unique. The major items to be considered include space constraints, location of existing equipment (retrofit projects), temperature requirements, fuel and cost. Fig. 16 shows several potential options of how the SCR system can fit into the boiler system. For new boilers, or when space is available on a retrofit, the preferred configuration is System (a), which provides the optimum temperature profile. The configuration shown in System (b) was originally developed as a retrofit due to space limitations. Unfortunately, its capital and operating costs are substantially higher than those for System (a) due to the required additional heat exchangers and duct burner (external heat source).

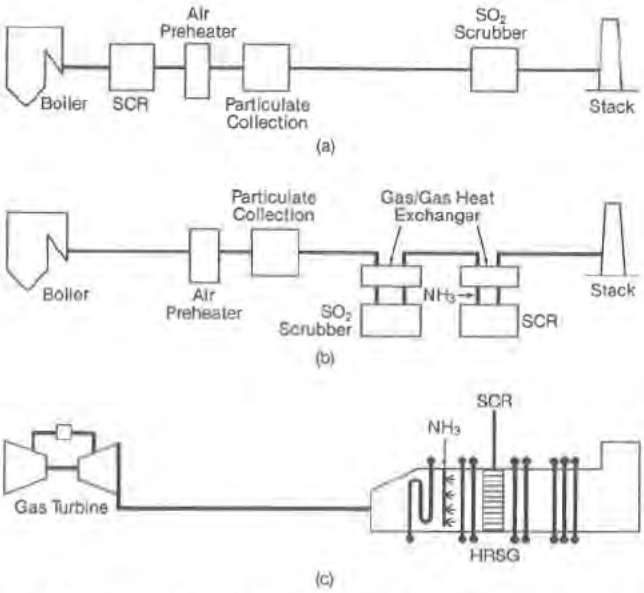


Fig. 16 Examples of SCR deNO$_x$ system configurations: (a) preferred SCR location, (b) possible retrofit configuration, and (c) combined cycle heat recovery steam generator (HRSG) location.

The application of SCR technology to combined cycle systems with heat recovery steam generators (HRSG) is shown as System (c). In this application, the catalyst placement within the HRSG is based on gas temperature requirements.

Selective catalytic reduction must be considered when low NO_x combustion technology provides insufficient NO_x reduction to meet local requirements. The SCR technology has proven to be reliable and is capable of up to 90% NO_x reductions and higher in some selected cases.

Future technology

Selective catalytic and noncatalytic processes have been widely accepted since the mid 1970s. SCR advancements have centered on catalyst developments for improved resistance to erosion, resistance to chemical deactivation, lower SO_2 to SO_3 conversion, and an expanded temperature performance range. Catalyst materials such as zeolite and precious metals are recent innovations.

Most SNCR process experimentation has considered alternatives to ammonia injection. Urea is the likely alternative. Two other alternatives receiving attention include cyanuric acid and ammonium sulfate. In addition, the use of methanol extenders (injected with the NO_x reagent) is being studied. The goal is to extend the useful NO_x reduction temperature range.

A special case of injection technology involves organoamines. For example, methylamine reacts with HNO_2, yielding methanol and nitrogen. This is a low temperature gas phase reaction as opposed to the current high temperature reactions. However, the gas phase concentration of HNO_2 is limited by the oxidation rate of NO to NO_2. This technology, under development, has not been demonstrated on a commercial scale.

A variety of advanced postcombustion NO_x control systems are at various stages of development and demonstration. Many of these are combined NO_x and SO_2 control systems. While the SNCR and SCR systems discussed here basically feature dry reduction of NO_x by ammonia or urea, advanced systems under development offer a variety of options, e.g., wet (aqueous) or dry absorption by solids, absorption plus oxidation by a liquid, and absorption plus reduction by a liquid, among others. However, only the SCR and SNCR systems have achieved widespread commercialization to date. Further development and demonstration of the other systems will be required before significant acceptance is likely and attractive economics are achieved.

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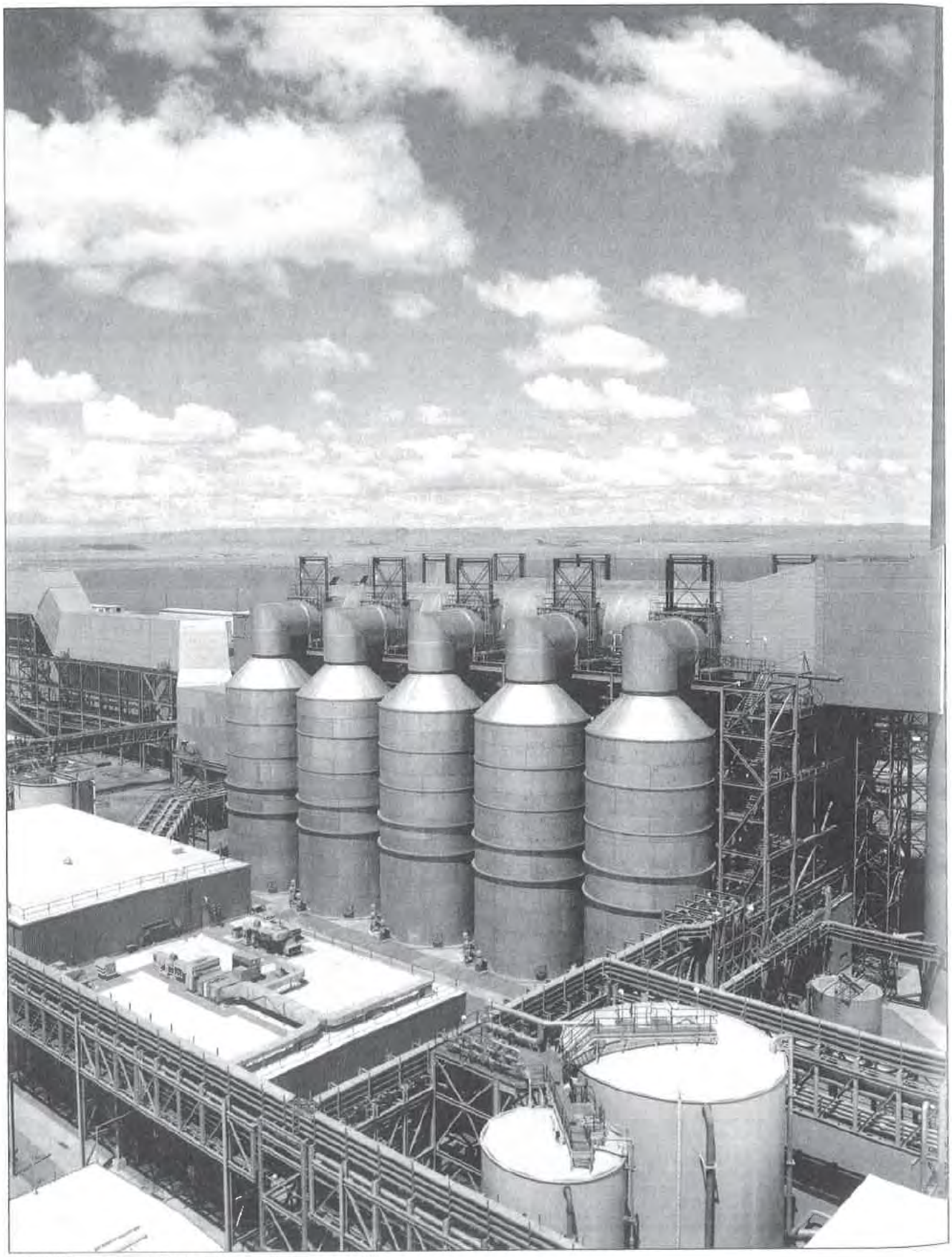
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1600 MW capacity wet SO_2 flue gas desulfurization retrofit.

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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 which 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 portions of 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 estimated to be 180 to 200 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 (g) SO_2 combines with liquid (ℓ) water, it forms a dilute aqueous (aq) 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 in acid rain.) The respective reactions are written:

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

$$O_2(g) + 2H_2SO_3(aq) \rightarrow 2H_2SO_4(aq) \quad (2)$$

The SO_2 can also oxidize in the atmosphere to produce gaseous sulfur trioxide (SO_3). Having a high affinity for water vapor, the sulfur trioxide reactions are written:

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

$$SO_3(g) + H_2O(g) \rightarrow H_2SO_4(\ell) \quad (4)$$

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 is the method used to quantify the acidity of acid rain. The scale is defined by:

$$pH = -\log_{10}[H^+] \quad (5)$$

where $[H^+]$ is the concentration of hydrogen ions in solution. A pH below 7.0 is considered acidic, while a value above 7.0 is alkaline. The pH of pure distilled water is 7.0. If rain water 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 rain water. 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. Rains with pH values of 4.2 have been observed.

SO_2 emissions regulations

Legislative action has been responsible for most industrial SO_2 controls. Major regulations include the Clean Air 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 proposed in many other industrialized nations. Most members of the European Economic Community are regulated and Canadian laws are similar to those in the U.S. A more detailed discussion of regulation is included in Chapter 32.

SO_2 control

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. Of these, the wet SO_2 scrubber has been the dominant worldwide technology for the control of SO_2 from utility power plants. As such, wet flue gas desulfurization (FGD) systems will be the primary focus of this chapter with other systems, especially dry FGD systems, also discussed.

From 1970 to 1988, SO_2 emissions in the U.S. diminished from 28.4 to 20.7 million tons per year. In that same time span SO_2 emissions by electric utilities were reduced from 15.8 to 13.6 million tons per year. In 1970 the average sulfur content of coal consumed by these utilities was about 2.3%. The use of coal by the utilities has grown in this time period from 320 million tons in 1970 to 758 million tons in 1988. If no SO_2 emission controls had been

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implemented in 1970, it is estimated that SO_2 emissions from U.S. electric utilities would have grown to about 34.7 million tons per year by 1988.

The utilities primarily used two strategies for control, switching to low sulfur coal and installing scrubbers. By 1988, the average sulfur content of fuels burned was 1.34% for coal and 1.09% for oil. With fuel switching alone, SO_2 emissions for electric utilities would have been 21.25 million tons in 1988; however, scrubbers, primarily wet, removed 7.65 million tons of SO_2 in 1988. In all, fuel switching accounts for about 64% of the SO_2 control and scrubbers have accounted for 36% of utility SO_2 reductions. (Fuel switching technical considerations are discussed in Chapter 46.)

Wet scrubbers have been used by many utilities since 1970 to meet SO_2 emission regulations. 68,000 MW of FGD capacity were installed in the U.S. between 1970 and 1990.[1] In Japan, 32,000 MW of capacity were added during this period. 48,000 MW of capacity were installed between 1979 and 1990 in Germany. The rest of the world accounts for approximately 4,000 MW. (See Fig. 1).

Wet scrubbers offer the following advantages:

1. As an alternative to fuel switching, the utility can use its normal source of fuel supply.
2. As an alternative to and in contrast to other traditional and emerging flue gas desulfurization methods, wet scrubbers provide high SO_2 removal efficiency and high reagent utilization.

Of the 180,000 MW of worldwide FGD capacity, 85% are wet scrubbers with the balance being predominantly dry scrubbers. Wet scrubbing processes are often categorized by reagent. Table 1 lists several scrubber processes and, of these, the limestone process has been the most widely applied. Within each category several process variations have been developed and commercialized. These are broadly outlined in Table 1 and are discussed later in this chapter. 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 silicate. 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. (See Chapter 11.)

Wet scrubbers

Design

The first FGD wet 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, soon replaced venturi scrubbers. (See Chapter 35.) Fig. 2 displays the typical arrangement.

The dust collector is placed upstream of the wet scrubber. In this arrangement, the induced draft fan is between the dust collector and the scrubber, permitting the fan to operate in a particulate-free, dry flue gas. In Germany, where nearly all FGD systems were retrofits to existing boilers, some induced draft fans did not have sufficient capacity. In some instances, a booster fan is placed downstream of the scrubber to handle the additional scrubber draft. These fans, incorporating a wash system to prevent excessive solid deposition from the scrubber, generally have rubber-lined housings and high alloy rotors.

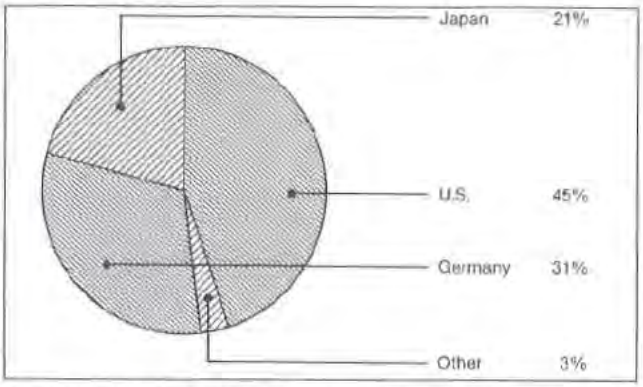


Fig. 1 Wet FGD systems installed throughout the world, on a capacity basis — through 1990.

The maximum size of a wet scrubber depends upon several design factors. Scrubbers which treat more than 6×10^6 lb/h (756 kg/s) of flue gas have been built and no fundamental engineering limits to wet scrubber size have been identified. Prior to 1990, U.S. environmental regulations encouraged utilities to purchase FGD systems which included multiple module designs with spare capacity. Current regulations now provide a framework within which single module scrubber systems can form a part of the compliance strategy. In Germany and Japan, utilities have favored large single modules for each boiler; in some instances, one scrubber treats the flue gas from two or more boilers.

The most popular wet scrubber design is the *spray tower* depicted in Fig. 3. The tower is designed so that, at maximum load, the average superficial gas velocity does not exceed the design gas velocity. For most spray towers, the average gas velocity varies from about 8 to 13 ft/s (2.4 to 4 m/s) based upon scrubber outlet conditions. A typical design velocity for a limestone wet scrubber is about 10 ft/s (3.1 m/s).

Because the flue gas enters the absorber from the side, gas flow nonuniformity in the tower is a potential prob-

Table 1
Wet Scrubber Processes

Limestone —
- with no oxidation inhibition (*natural oxidation*)
- with inhibited oxidation
- with in situ forced oxidation
- with ex situ forced oxidation
- with soluble organic or inorganic buffers

Lime —
- with inorganic buffers (such as magnesium oxide)
- without buffers

Dual alkali —
- sodium carbonate/calcium hydroxide
- sodium carbonate/calcium carbonate

Soda ash —
- with regeneration by steam stripping
- without regeneration

Magnesium oxide with thermal regeneration

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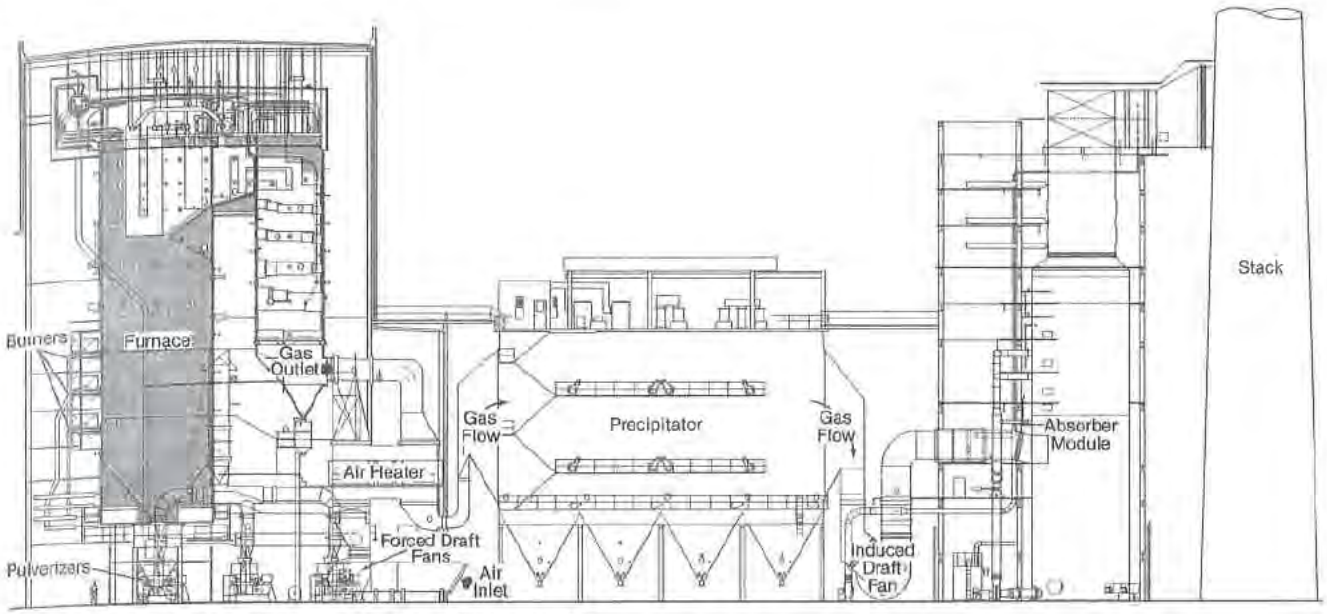


Fig. 2 Typical emission control systems for a coal-fired utility boiler.

lem. This nonuniformity reduces overall SO_2 removal performance and aggravates mist eliminator carryover. The absorber design depicted in Fig. 3 incorporates a sieve or perforated plate tray which reduces flue gas flow maldistribution. The pressure drop across the tray is usually between 1 to 3 in. wg (0.2 to 0.7 kPa). Towers with multiple trays have also been built. The design of the tower is influenced by the reagent (lime or limestone, for example), the desired SO_2 removal level, the tradeoff between fan power and recirculation slurry pump power, and several other factors.

The dominant design variable for all FGD wet scrubbers is the ratio of slurry flow to gas flow in the tower, known as L/G. In the U.S., this term is expressed as gallons per thousand cubic feet of flue gas evaluated at scrubber outlet conditions. In Japan and Germany, the L/G units are l/m^3. A wide range of L/G values have been used on wet scrubbers. Soda scrubbers operate at as low as 10 gal/1000 ft^3 (1.34 l/m^3) and some limestone scrubbers operate at up to 150 gal/1000 ft^3 (20 l/m^3).

Spray nozzles are used in wet scrubbers to control the mixing of slurry with 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. Although plugging could be minimized by using a minimum number of large spray nozzles, flow maldistribution would most likely occur. Therefore, several smaller nozzles are usually preferred. Ceramic nozzles are commonly used.

The large tank at the bottom of the tower is called the *reaction tank* or the *recirculation tank*. The volume of this tank permits several chemical and physical processes to approach completion. Some of these rate processes will be described in more detail later.

Flue gas/scrubber considerations

Flue gas enters the side of 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 spray. The scrubber inlet must be designed to prevent deposition of slurry solids at the wet-dry interface. Because the inlet flue is at the flue gas temperature [for example 300F (149C)] and the shell of the scrubber is at the saturation temperature [typically 125F (52C)], there is a point where the surface temperature abruptly changes. Deposits are most likely to form at this point. Deposition is minimized by a combination of features which prevent periodic slurry splashing on the hot, dry side of the wet-dry line.

Flue gas passes vertically upward through the scrubber. In the unit illustrated in Fig. 3, the gas flow is uniformly distributed by a specially designed perforated plate or sieve tray. This tray serves as a gas-liquid contacting device. Gas-liquid contact is enhanced by a froth of slurry which forms on the tray.

Above the tray, flue gases pass through several spray levels where additional gas-liquid contact is achieved. Each spray level consists of a set of headers and spray nozzles as illustrated in Fig. 3. The spray nozzles produce a relatively coarse spray with mass median droplet diameters of about 2000 to 2500 microns. This suspension of droplets is in countercurrent contact with the flue gas for about one to three seconds. A majority of the SO_2 absorption occurs during this short contact time. The spray zone in combination with the froth on the tray is referred to as the gas-liquid contact zone of the wet scrubber.

A disengagement height is provided above the top spray zone before the flue gases reach the mist eliminators. The purpose is to allow disengagement and return of the largest slurry droplets to the spray zone. For a scrubber operating at an average gas velocity of 10 ft/s (3 m/s), droplets larger than about 600 microns may have sufficient time and mass to fall back to the spray zone.

The mist eliminator design in most wet scrubbers uses *chevrons* to remove additional moisture from the flue gas (see Fig. 3). Chevrons are closely spaced corrugated plates which collect slurry deposits by impaction. They efficiently collect droplets larger than about 20 microns in diameter.

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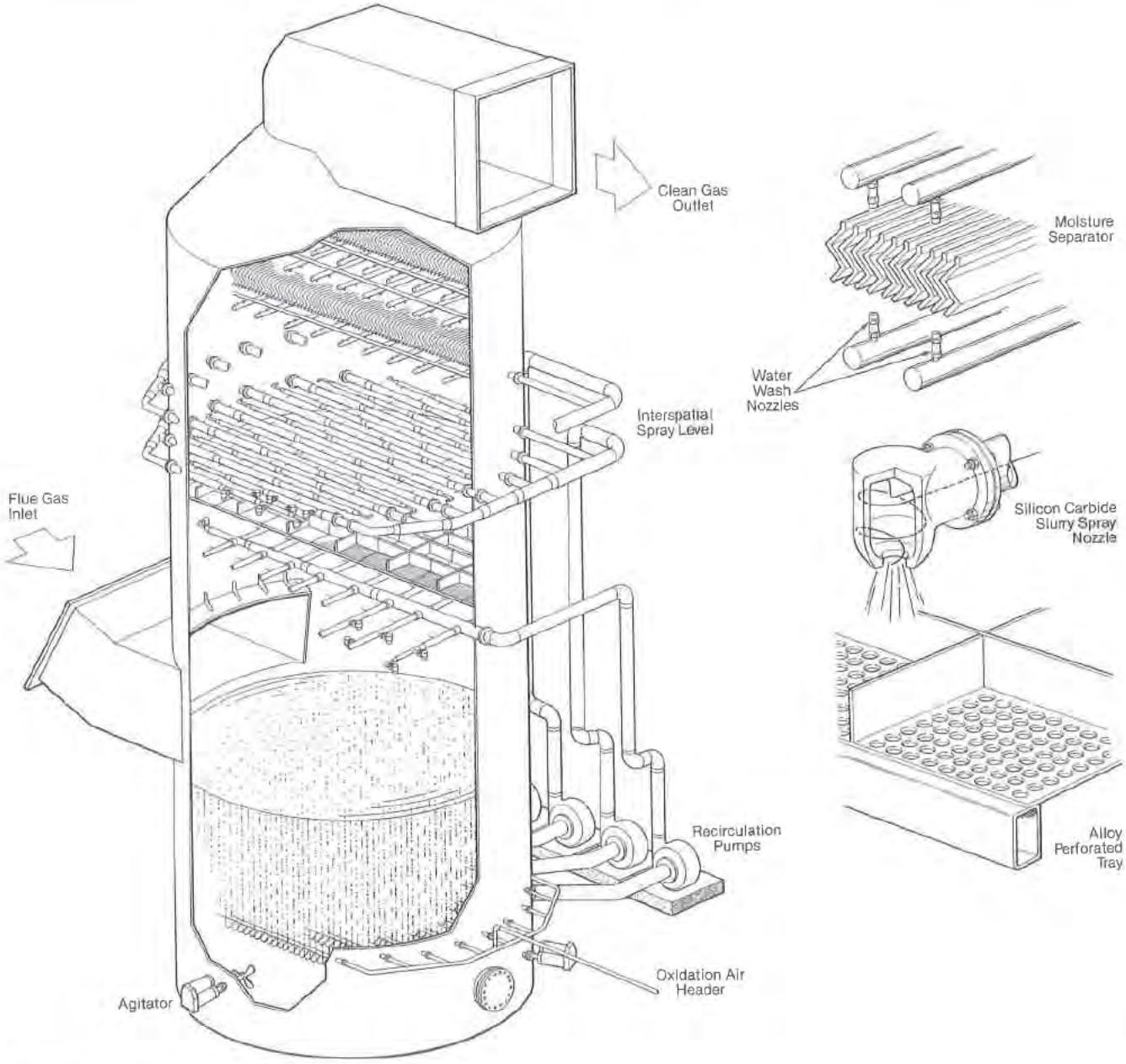


Fig. 3 Wet flue gas desulfurization scrubber module.

The design of the flue 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 a combination of facts. First, the flue gas leaves the mist eliminator saturated with water vapor. Second, 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 some of 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 1.0), and calcium salts can deposit on the walls.

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 more droplets remain. Reheating the flue gas that is leaving the scrubber has been accomplished by various systems:

1. steam coil heaters,
2. mixing with some hot flue gas which is bypassed around the scrubber,
3. mixing with hot air,
4. mixing with hot gases generated by combustion of a clean fuel, and
5. regenerative heat exchangers which transfer heat from the hot flue gas inlet to the cooler flue gas outlet.

Several problems are associated with each of the reheat methods. Deposition and corrosion occur in the heat exchangers. Reheating with bypass gas reduces the over-

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all FGD system effectiveness. Finally, the evaporation of droplets from the scrubber concentrates the corrosive constituents in the slurry. 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 flues from the scrubber to the stack are lined with corrosion resistant materials, and the stack is lined with acid resistant brick or other suitable material. A drainage system is also included to accommodate condensation of water vapor.

Wet scrubber — limestone and lime FGD processes

All of the limestone and lime processes listed in Table 1 are classified as nonregenerable. This means that the reagent is consumed by the process and must therefore be continually replenished. A generic diagram of nonregenerative processes is illustrated in Fig. 4. 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. A description of some of the more frequently used approaches to each process step is presented below followed by examples of two specific processes.

Reagent preparation

Limestone In North America, most limestone wet FGD systems feature on-site wet grinding for slurry preparation. In most cases, the system of choice is a closed loop ball mill. A typical ball mill circuit is shown in Fig. 5. The energy required to achieve a given grind size is estimated by the Bond relationship:

$$W = \frac{10\ W_i}{\sqrt{D_P}} - \frac{10\ W_i}{\sqrt{D_F}} \qquad (6)$$

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

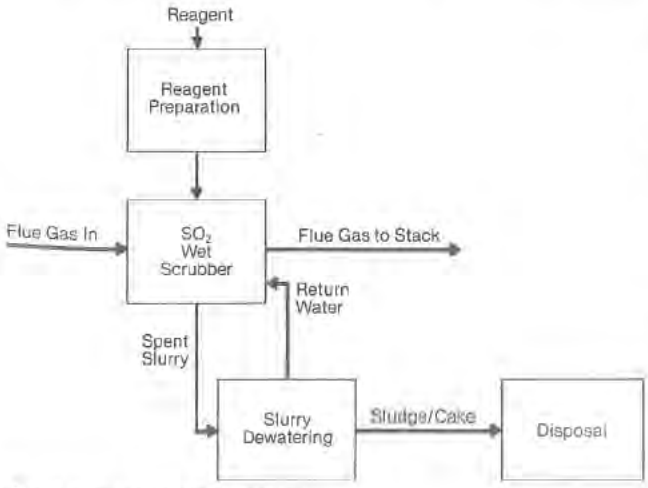


Fig. 4 Wet FGD system diagram.

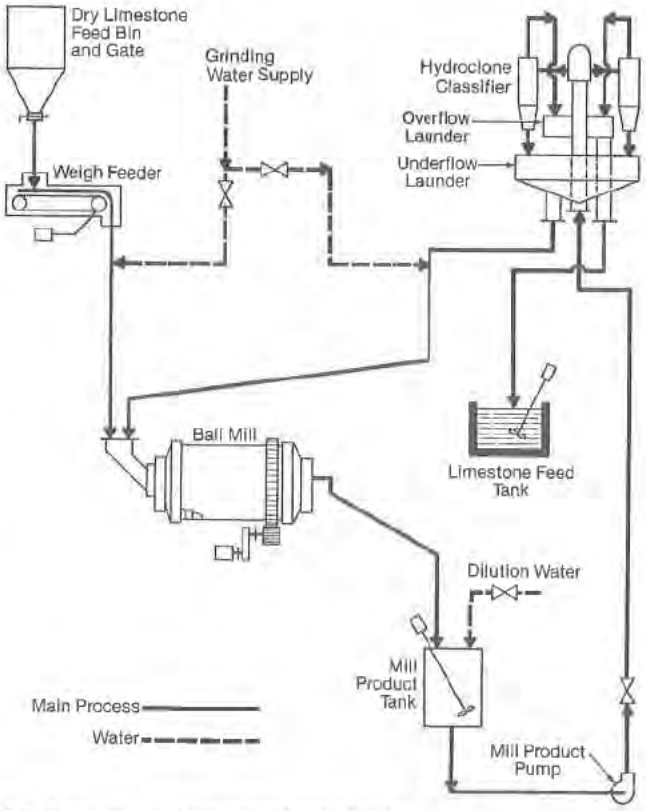


Fig. 5 Limestone reagent preparation system.

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

For on-site grinding systems, the limestone is typically received with a diameter of about 1 in. (25 mm) or less and is fed through a weigh belt feeder to the ball mill. Either fresh or recycled water is added at the feed chute in proportion to the feed rate of the stone. The output from the mill overflows to the mill product tank where it is pumped to a set of hydroclones. These hydroclones separate the fines from the coarse fraction. The coarse fraction (*underflow*) is returned to the ball mill and 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 coarsest limestone grind used in wet FGD systems is one where about 70% passes through a 200 mesh (75 micron) sieve. Fine limestone grinding produces a product where about 95% of the limestone is finer than a 325 mesh (44 micron) sieve. Fine grinding requires a larger ball mill system and higher operating expense. The finer material provides higher limestone utilization, better reactivity with SO_2, and higher SO_2 removal for a given stoichiometry, while permitting a smaller reaction tank for some systems. Stoichiometry is defined as the molar ratio of the reactant, $CaCO_3$ for limestone systems, to the SO_2. It is common practice in the U.S. to define the stoichiometry by using the quantity of SO_2 removed for wet scrubbing systems and by the inlet quantity of SO_2 for other removal systems.

In Germany, off-site grinding is preferred. Limestone is dry ground at a central location and transported by truck to the FGD system. At the site, the pulverized lime-

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stone is pneumatically conveyed to a storage bunker. It is then fed to water-filled slurry preparation tanks. This system requires less space than on-site grinding facilities and reduces some of the FGD operating burden.

Lime Lime based FGD systems are common in the U.S. Three types of slaking systems are used: the detention slaker, the paste slaker and the ball mill slaker. The detention slaker is the simplest but produces the poorest quality slaked lime. The paste slaker, while producing high quality lime, is the most complex system and grit separation can be troublesome. The ball mill slaker produces an intermediate slaked lime quality but requires no postslaking separation of the grit and slaked lime.

A typical slurry preparation system is shown in Fig. 6. This system consists of storage for pebble lime, a weigh belt feeder, slaker, grit removal system, dilution system and slaked lime storage tank. Lime storage is usually maintained at 20 to 25% suspended solids.

Slurry dewatering and disposal

The disposal of reaction products from lime/limestone wet scrubbers has taken many forms. Several approaches, some unique to U.S. units, are used.

Ponding of the spent slurry without dewatering is the simplest disposal method. However, limited site availability and the cost of pond management have restricted its use. The combination of primary dewatering, secondary dewatering and landfilling is the most common approach used in the U.S. Primary dewatering is accomplished using thickeners or hydroclones. A thickener (Fig. 7) is capable of concentrating slurries with particle sizes as small as 5 microns to between 20 and 30% suspended solids. Lime/limestone systems with natural oxidation produce fine crystals and therefore require thickeners.

If the slurry is limited to particles larger than about 15 microns, hydroclones can be used for primary dewatering. These devices, much smaller and simpler than thickeners, are preferred where the particle size distribution is suitable. The gypsum crystals produced by lime/limestone systems with forced oxidation are usually suitable for hydrocyclones.

Secondary dewatering of FGD wastes is accomplished using a number of methods: vacuum drum filters, vacuum

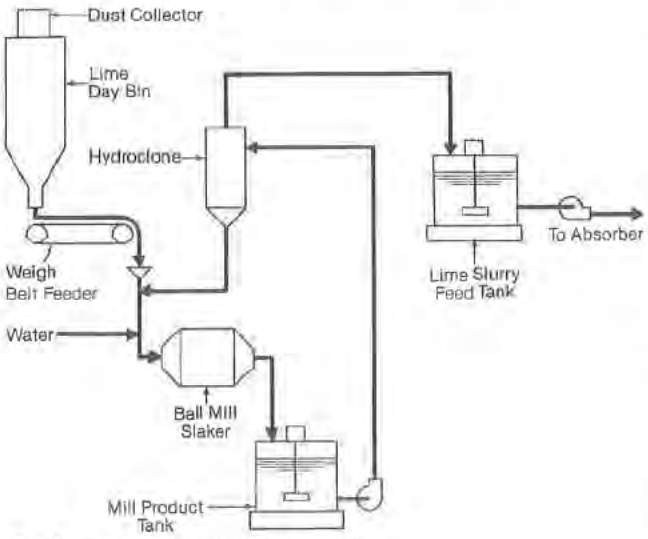


Fig. 6 Lime reagent preparation system.

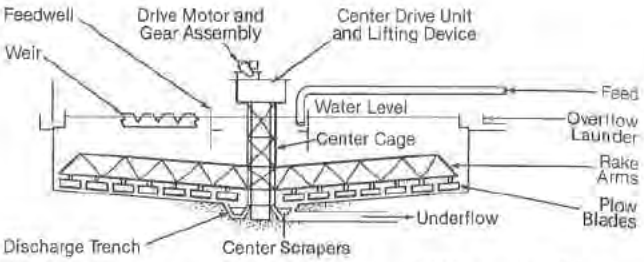


Fig. 7 Thickener system for primary dewatering of FGD scrubber effluent.

belt filters, solid bowl centrifuges and vertical basket centrifuges. Vacuum drum filters are the most common U.S. application, accounting for about 80% of the installations. Germany uses vacuum belt filters on more than 90% of its units.

When landfilling is used for sulfite sludge disposal, additional treatment is required for sludge management at the landfill. *Sludge fixation* is used to increase the compressive strength of the sludge. Sulfite sludge is mixed with flyash and lime to yield a material suitable for landfilling. If gypsum is to be landfilled it needs no special treatment after dewatering.

Overall process flow diagrams and mass balance

All lime/limestone FGD wet scrubbers require fresh water for mist eliminator wash. This fresh water is usually combined with recycle water from the clarifier, thickener and/or vacuum filter to provide the total wash flow. Fresh water is also used for pump seals, reagent preparation and filter cake washing. Water losses from the system include evaporative losses, moisture in the waste solids, waters of crystallization and, in some cases, a controlled water purge. Careful management is required to assure adequate water supplies.

A typical limestone based wet scrubbing process using in situ forced oxidation is depicted in Fig. 8. In this example, the reagent preparation system includes a closed circuit ball mill system producing a slurry of ground limestone to a fineness of 95% minus 325 mesh (44 microns). This means that 95% by weight of the ground limestone will pass through a screen designated as a 325 mesh (44 micron) sieve. This slurry contains 30% by weight of suspended solids. The water used in the milling system is recycled from the spent slurry dewatering system. The feed slurry is pumped to the absorber reaction tank at a controlled rate to maintain the pH of the slurry in this tank at a set point of 5.5. Air is also pumped to this reaction tank and distributed by spargers located near the bottom. The oxygen in the air reacts with any sulfite present in the slurry to produce gypsum (calcium sulfate dihydrate).

The slurry is pumped from the reaction tank to the spray headers shown in Fig. 3. The slurry is sprayed countercurrently into the flue gas where it absorbs the SO_2. The slurry falls to the perforated plate 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 back to the reaction tank.

A small fraction of the slurry being pumped to the spray nozzles is diverted to the dewatering system. The spent slurry typically contains about 15% suspended solids. In this example, a hydroclone is used to concentrate the slurry. The underflow from the hydroclone is

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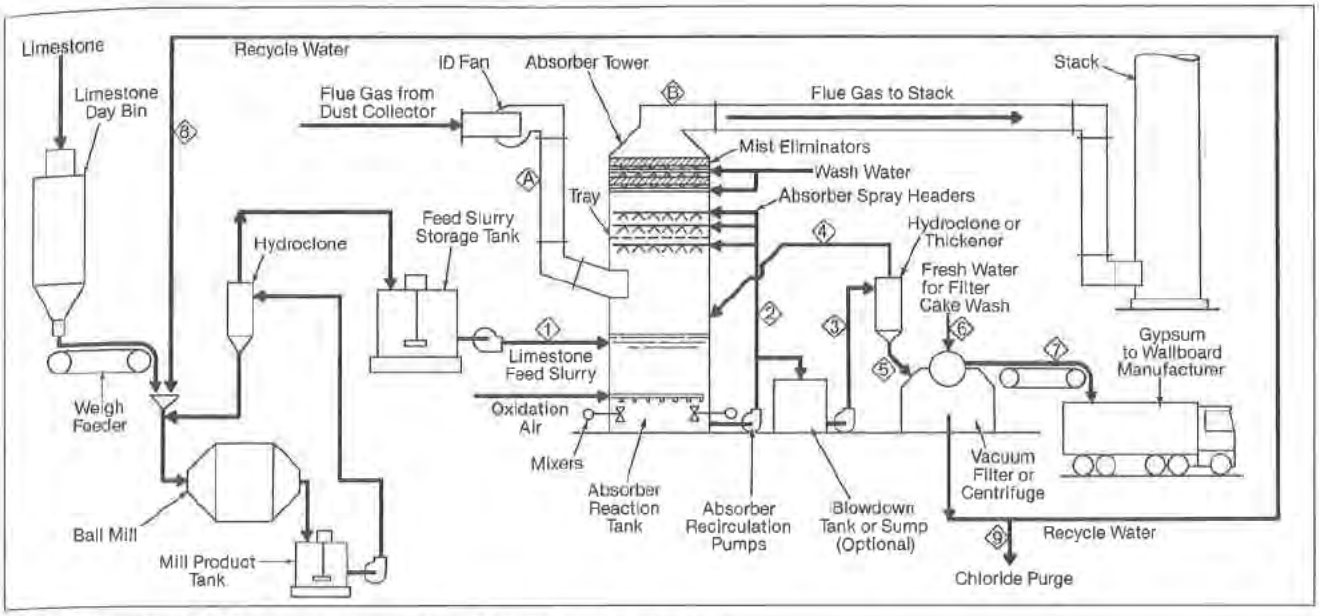


Fig. 8 Wet scrubber FGD system flow diagram (see Table 2 for material balance).

concentrated to 25% solids. The overflow containing 4% suspended solids is sent back to the absorber.

The underflow from the hydroclone is directed to a vacuum filter where the filtered solids are washed with fresh water and concentrated 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 2. A summary of the major power requirements for this limestone to gypsum process is shown in Table 3.

Fig. 9 illustrates a typical flow sheet for an FGD process using a lime produced from a dolomitic limestone. This lime contains about 5% magnesium oxide (MgO), 90% calcium oxide and 5% inerts. Within the process, the MgO is converted to $MgSO_3$, where, because of its relatively high solubility (approximately 10 g/l), the sulfite acts as a buffer to increase the SO_2 capacity of the slurry. The overall reaction is:

$$Ca(OH)_2(s) + SO_2(g) \rightarrow$$

$$CaSO_3 \cdot \tfrac{1}{2} H_2O(s) + \tfrac{1}{2} H_2O(\ell) \qquad (7)$$

In this instance, a closed circuit ball mill system is used to produce a slaked lime:

Table 2
Mass Balance for The Limestone Forced Oxidation System Shown in Fig. 8

Gas side	Inlet	Exit
Stream number	A	B
Flow rate, ACFM	626,724	500,000
Total mass flow rate, lb/h	1,905,400	1,982,091
Mass flow rate of H_2O, lb/h	76,466	157,700
Mass flow rate of SO_2, lb/h	8,565	600
Mass flow rate of HCl, lb/h	344	0
Static pressure, in. wg	6	1
Temperature, F	330	123

180 MW boiler burning coal with 3.2% sulfur
Limestone stoichiometry = 1.03
L/G = 130 gal/1000 ACF
SO_2 efficiency = 93%

Liquid side	Feed Slurry	Recirc	To Hydroclone	Hydroclone Overflow	Hydroclone Underflow	Filter Wash Water	Filter Cake	Recycle Water	Chloride Purge
Stream number	1	2	3	4	5	6	7	8	9
Flow rate, GPM	59	65,260	319	236	83	26	22	51	36
Total flow rate, lb/h	38,819	37,275,000	180,196	125,136	55,060	13,215	24,471	25,732	18,072
Flow rate of sus. solids, lb/h	13,586	5,607,000	27,029	5,005	22,024	0	22,024	0	0
% suspended solids	35	15	15	4	40	0	90	0	0
Chloride conc., ppm in liquid	25,000	25,000	25,000	25,000	25,000	50	100	18,514	18,514
pH	7.8	5.7	5.7	5.7	5.7	7	7	5.7	5.7
Temperature, F	100	123	123	123	123	70	70	70	70

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Table 3
Typical Power Requirements for Limestone Scrubber with Forced Oxidation

Absorber System	Avg. Power (kW)	% Power
Oxidation air blower	375	
Absorber recirc. pump #1	312	
Absorber recirc. pump #2	367	
Absorber recirc. pump #3	380	
Absorber recirc. tank agitators	60	
Mist eliminator wash water pump	19	
Misc. pumps and agitators	24	
Subtotal	1537	60.3%
Dewatering Area	**Avg. Power (kW)**	**% Power**
Vacuum pump for filter	55	
Filter wash water tank heater	16	
Reclaim water pump	14	
Hydroclone overflow pump	15	
Filter feed tank agitator	7	
Clarifier overflow sump pump	6	
Misc. pumps and agitators	13	
Subtotal	126	4.9%
Reagent Preparation	**Avg. Power (kW)**	**% Power**
Ball mill driver	220	
Mill product tank pump	5	
Limestone feed tank agitator	25	
Misc. pumps and agitators	6	
Subtotal	256	10.0%
Other Systems	**Avg. Power (kW)**	**% Power**
General-instrument air	50	2.0%
Differential induced draft fan power	580	22.8%
Subtotal	630	
Total	2549	

Notes:
180 MW boiler burning coal with 2.5% sulfur
Heating value of coal = 12,767 Btu/lb
Absorber L/G = 130 gal/1000 ACF
Total pressure drop = 5 in. wg
Parasitic power = 2.549 MW/180 MW
= 1.42%

$$CaO + H_2O \rightarrow Ca(OH)_2 + Heat \quad (8)$$

$$MgO + H_2O \rightarrow Mg(OH)_2 \quad (9)$$

The absorber towers used for this process are smaller than their limestone counterparts because the L/G required to achieve a comparable level of SO_2 absorption is typically only about 20% of L/G for the limestone units.

The size distribution of the particles in the spent slurry is much smaller than the gypsum particles produced in the limestone forced oxidation process described above. A hydroclone would be incapable of achieving adequate concentration. Therefore, a thickener as illustrated in Fig. 7 is used. By comparison to a hydroclone, a thickener has the advantage of producing a clear overflow. The clear overflow means that this water can be recycled to the process and used for washing mist eliminators. The space requirement for a thickener is many times greater than that required by a hydroclone, however. The underflow from the thickener is directed to a vacuum filter where the unwashed cake is filtered to 45 to 55% solids. This cake is then mixed with lime and flyash to form an aggregate suitable for landfilling.

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. However, the chemical processes involved are complex. SO_2 is a relatively insoluble gas in water. Calcium carbonate ($CaCO_3$) has a low solubility in water. The principal reaction products are calcium sulfite hemihydrate ($CaSO_3 \cdot {}^1/_2H_2O$) and calcium sulfate dihydrate ($CaSO_4 \cdot 2H_2O$), or gypsum. Both of these salts also have low solubilities.

In a limestone system with inhibited oxidation, the following reaction model can be used to describe the process using the chemical species in Table 4.

In the gas-liquid contact zone:

$$SO_2(g) \leftrightarrow SO_2(aq)$$

Dissolving gaseous SO_2 **(10)**

$$SO_2(aq) + H_2O \leftrightarrow HSO_3^- + H^+$$

Hydrolysis of SO_2 **(11)**

$$CaCO_3(s) + H^+ \leftrightarrow Ca^{++} + HCO_3^-$$

Dissolution of limestone **(12)**

$$HCO_3^- + H^+ \leftrightarrow CO_2(aq) + H_2O$$

Acid – base neutralization **(13)**

$$CO_2(aq) \leftrightarrow CO_2(g)$$

CO_2 *stripping* **(14)**

Table 4
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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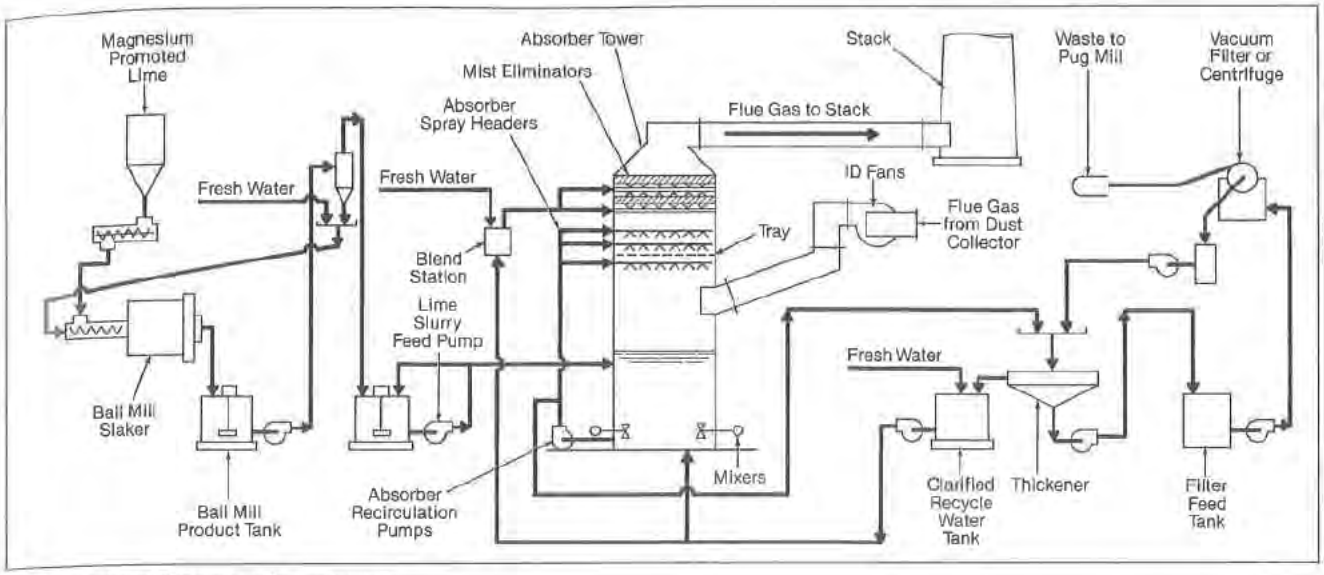


Fig. 9 Wet lime FGD system flow schematic.

$$CaSO_3 \cdot {}^1/_2\, H_2O + H^+ \leftrightarrow Ca^{++} + HSO_3^- + {}^1/_2\, H_2O$$

Dissolution of calcium sulfite **(15)**

In the reaction tank:

$$CaCO_3(s) + H^+ \leftrightarrow Ca^{++} + HCO_3^-$$

Dissolution of limestone **(16)**

$$HCO_3^- + H^+ \leftrightarrow CO_2(aq) + H_2O$$

Acid – base neutralization **(17)**

$$CO_2(aq) \leftrightarrow CO_2(g)$$

CO_2 *stripping* **(18)**

$$Ca^{++} + HSO_3^- + {}^1/_2\, H_2O \leftrightarrow CaSO_3 \cdot {}^1/_2\, H_2O + H^+$$

Precipitation of calcium sulfite **(19)**

Reaction 10 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:

$$\frac{d(Gy)}{dV} = k_g a (y - y^*) \qquad \textbf{(20a)}$$

or

$$N_g = \int \frac{dy}{y - y^*} = \int \frac{k_g a}{G} dV \qquad \textbf{(20b)}$$

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/m^{2}s
a = interfacial surface area, m^2/m^3
y^* = equilibrium SO_2 concentration at the gas-liquid interface
V = volume of the gas-liquid regime, m^3
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 20 can then be integrated to:

$$N_g = -\ell n(1 - E) = k_g a\, V / G \qquad \textbf{(21)}$$

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 spacial distributions, the gas phase residence time (height of spray zone), the liquid residence time, wall effects and the gas flow distribution.

In a limestone based wet scrubber, the rate-limiting reactions in the gas-liquid contact zone are believed to be Reactions 12 and/or 15. The reaction rate for limestone dissolution can be expressed by:

$$\frac{d[CaCO_3]}{dt} = k_c([H^+] - [H^+]_{eq}) S_{pc} [CaCO_3] \qquad \textbf{(22)}$$

where

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

Fig. 10 is a plot of limestone dissolution rates at various pH values and CO_2 partial pressures. The dissolution rate of calcium sulfite can be expressed in a form similar to Equation 22. The calcium sulfite will dissolve only if the H^+ concentration exceeds its equilibrium value in the gas-liquid contact zone.

The equilibrium pH for calcium sulfite is approximately 6.3 at a CO_2 partial pressure of 0.12 atm (12 kPa). Therefore, in order for Reaction 15 to proceed to the right, the pH in the gas-liquid contact zone must be below the

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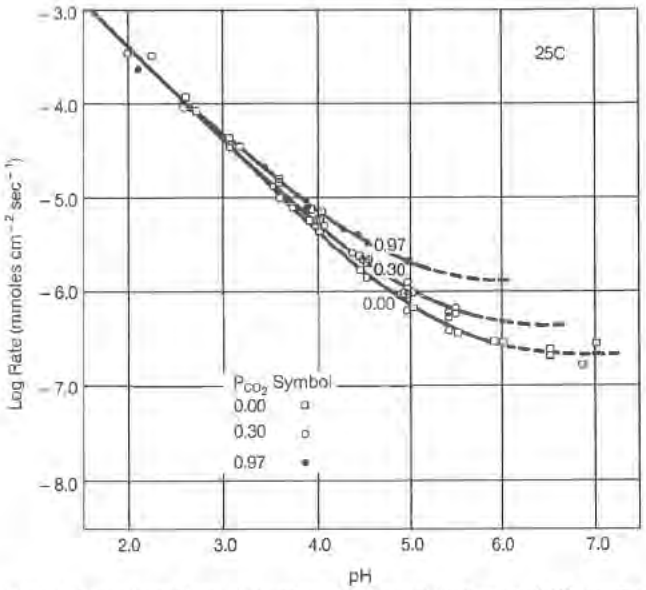


Fig. 10 Limestone dissolution rates at various pH values and CO_2 partial pressures (atm).[2]

equilibrium pH. If the pH were to stay above 6.3, then Reaction 15 would proceed to the left. This is undesirable because scaling due to sulfite precipitation in the scrubber is a likely consequence.

The reaction tank permits Reactions 16 through 19 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:

$$CaCO_3(s) + H^+ + HSO_3^- + \tfrac{1}{2}\,H_2O \rightarrow CaSO_3 \cdot \tfrac{1}{2}H_2O(s) + CO_2(g) \qquad (23)$$

Some oxidation of sulfite to sulfate inevitably occurs in the gas-liquid contact zone of the scrubber:

$$HSO_3^- + \tfrac{1}{2}\,O_2 \rightarrow SO_4^= + H^+ \qquad (24)$$

This is called *natural oxidation* to distinguish it from *forced oxidation* in which air is sparged through the slurry. The extent of natural oxidation is typically 15 to 30%, although it may be as high as 50% or higher.

The sulfate combines with dissolved calcium and water to form calcium sulfate dihydrate, or gypsum. The reaction proceeds as follows:

$$Ca^{++} + SO_4^= + 2H_2O \leftrightarrow CaSO_4 \cdot 2H_2O \qquad (25)$$

The rate of gypsum crystallization can be expressed by:

$$\frac{d[CaSO_4 \cdot 2H_2O]}{dt} = k(R-1)Sp_g\,[CaSO_4 \cdot 2H_2O] \qquad (26)$$

where

$R = A_{Ca^{++}}\,A_{SO_4^=} / K_{Sp}$

$A_{Ca^{++}}$ = 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 in 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 six to ten minutes of residence time based upon the recirculation rate.

Gypsum scaling might be more of a problem in wet scrubbers that use lime or limestone and operate in the natural oxidation mode if it were not for a phenomenon known as *co-precipitation*. The principal reaction product in these wet scrubbers is $CaSO_3 \cdot \tfrac{1}{2}H_2O$. Co-precipitation occurs when the sulfate ion ($SO_4^=$) is substituted for the sulfite ion ($SO_3^=$) in the calcium sulfite crystal lattice. This has the effect of removing sulfate from solution thereby reducing the gypsum supersaturation.

The reaction model for lime based wet scrubbers is similar to that for limestone. Two principal differences exist. First, calcium hydroxide is much more soluble than limestone, approaching 1 g/l at 77F (25C). Secondly, solid $Ca(OH)_2$ has a much higher specific surface area. For example, ground limestone has a specific surface area of about 0.2 to 0.8 m^2/g. By contrast, slaked lime typically varies from about 5 to 15 m^2/g. The equilibrium pH of $Ca(OH)_2$ at 77F (25C) is 12.4 compared to a value of 7.8 for $CaCO_3$. In many respects, lime behaves like a highly reactive limestone. The following lime reactions could be substituted for Reactions 12 and 13.

$$Ca(OH)_2(s) + H^+ \leftrightarrow CaOH^+ + H_2O \qquad (27)$$

$$CaOH^+ + H^+ \leftrightarrow Ca^{++} + H_2O \qquad (28)$$

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 11 and the rate of H^+ consumption by Reactions 12, 13 and 15. As the hydrogen ion concentration increases, i.e., as the pH drops, the equilibrium SO_2 vapor pressure increases. This equilibrium relationship can be expressed by:

$$y^* = \frac{C' \times 10^3 [H^+]}{k_1 (k_2 + [H^+])} \qquad (29)$$

where

k_1 = the equilibrium constant for Reaction 10, moles/l atm

k_2 = the equilibrium constant for Reaction 11, 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

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As the SO_2 vapor pressure rises, the rate of SO_2 absorption diminishes and approaches zero as $y^* \rightarrow y$. (See Equation 20.) 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:

$$H^+ + A^- \leftrightarrow AH \qquad (30)$$

where AH is the generalized acid group. When the pH falls in the gas-liquid contact zone, Reaction 30 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 and L/G ratio (the ratio of slurry flow to gas flow). 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:

$$SO_3^= + H^+ \leftrightarrow HSO_3^- \qquad (31)$$

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. 11.[2] 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.

Additives for scale control

Although many scrubber additives provide performance improvements, others are used specifically for scale control. Scaling, a potential problem in all lime/limestone scrubbers, is primarily due to calcium sulfate ($CaSO_4 \cdot 2H_2O$: gypsum) precipitation onto surfaces within the gas-liquid contact zone. The large reaction tanks used in these scrubbers alleviate this condition but do not eliminate it. Sodium thiosulfate ($Na_2S_2O_3$) is a recognized oxidation inhibitor which, when added to a limestone slurry, reduces the level of natural oxidation. The effectiveness of thiosulfate is proportional to the product of its molar concentration, the calcium ion concentration, and the moles of SO_2 absorbed.

An economic alternative to using $Na_2S_2O_3$ is to add colloidal sulfur to the lime/limestone slurry. The sulfur reacts with the bisulfite in solution as follows:

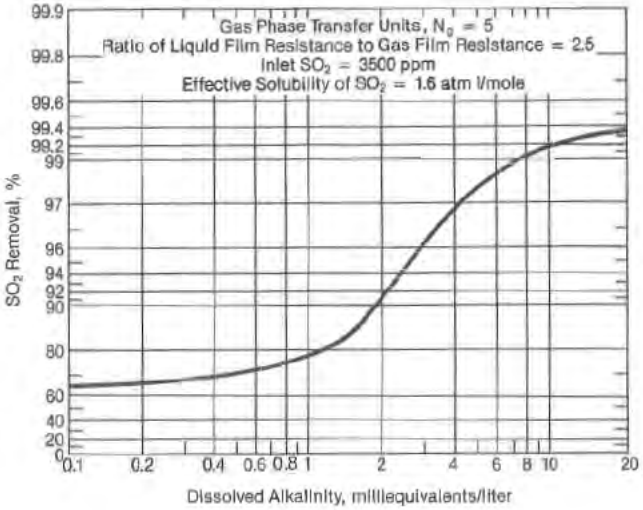


Fig. 11 Influence of dissolved alkalinity on SO_2 performance (1 milliequivalent = 50 ppm $CaCO_3$).

$$S(s) + HSO_3^- \leftrightarrow S_2O_3^= + H^+ \qquad (32)$$

The concentration of sulfur required ranges from 0.004 to 0.006 moles/l.

In addition to reducing scale formation in the scrubber, inhibiting sulfite oxidation permits easier removal of free water from the waste product (dewatering). Dewatering is simplified because the calcium sulfite crystal grows larger in the absence of sulfate ions.

Dry scrubbers

Dry scrubbing is the principal alternative to wet scrubbing for SO_2 control on utility boilers. Since 1980, 7200 MW of dry scrubbers have been installed at U.S. electric utilities. Dry scrubbing is also a popular choice for smaller industrial applications and for combined HCl and SO_2 control on waste-to-energy units. In the U.S., dry scrubbers have mainly been applied to units burning low sulfur fuels. Of the 43 U.S. utility dry scrubber installations, the majority lie west of the Mississippi River.

The advantages of dry scrubbing over wet scrubbing include:

1. less costly construction materials,
2. dry waste products,
3. fewer unit operations, and
4. simplicity of operation.

Dry scrubbing is sometimes referred to as spray absorption, spray drying or semi-wet scrubbing. It involves spraying a highly atomized slurry or aqueous solution of an alkaline reagent into the hot flue gas to absorb the SO_2. Fig. 12 depicts a utility size dry scrubber installation coupled with a baghouse. Unlike a wet scrubber installation, the dry scrubber is positioned before the dust collector. Flue gases leaving the air heater at a temperature of 250 to 350F (121 to 177C) enter the dry scrubber through an array of Turbo-Diffusers® as shown in Fig. 13 for a horizontal flow design. Vertical flow modules such as that shown in Fig. 14 are also supplied.

The quantity of water in the atomized spray is limited so that it completely evaporates in suspension. SO_2 absorp-

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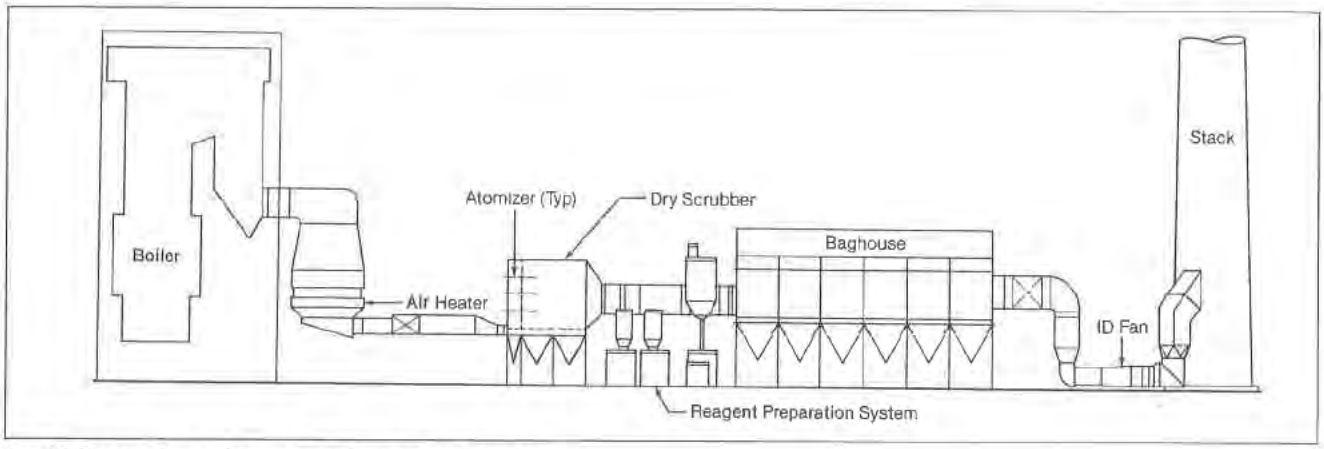


Fig. 12 Dry scrubber emissions control system configuration.

tion takes place primarily while the spray 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*. Reagent stoichiometry and approach temperature are the two primary variables which dictate the scrubber's SO_2 removal efficiency.

The flue gas temperature leaving the dry scrubber may be too low for proper operation of the dust collector. In those instances the gases may be heated before entering the dust collector. The dust collector can be either an electrostatic precipitator (ESP) or a baghouse. The ESP is more forgiving of temperature variation but the baghouse has the advantage of being a better SO_2-lime reactor.

The predominant reagent used in dry scrubbers is slaked lime. The system shown in Fig. 15 consists of storage facilities for pebble lime (CaO), a ball mill slaking system, a system for mixing slaked lime with recycle material from the dust collector, the dry scrubber and the dust collector. Atomization of the slurry is accomplished by pneumatic or rotary atomization. A typical pneumatic atomizer is shown in Fig. 13.

Dry scrubber chemistry

SO_2 absorption in a dry scrubber is similar to that attained by wet scrubbing. The majority 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:

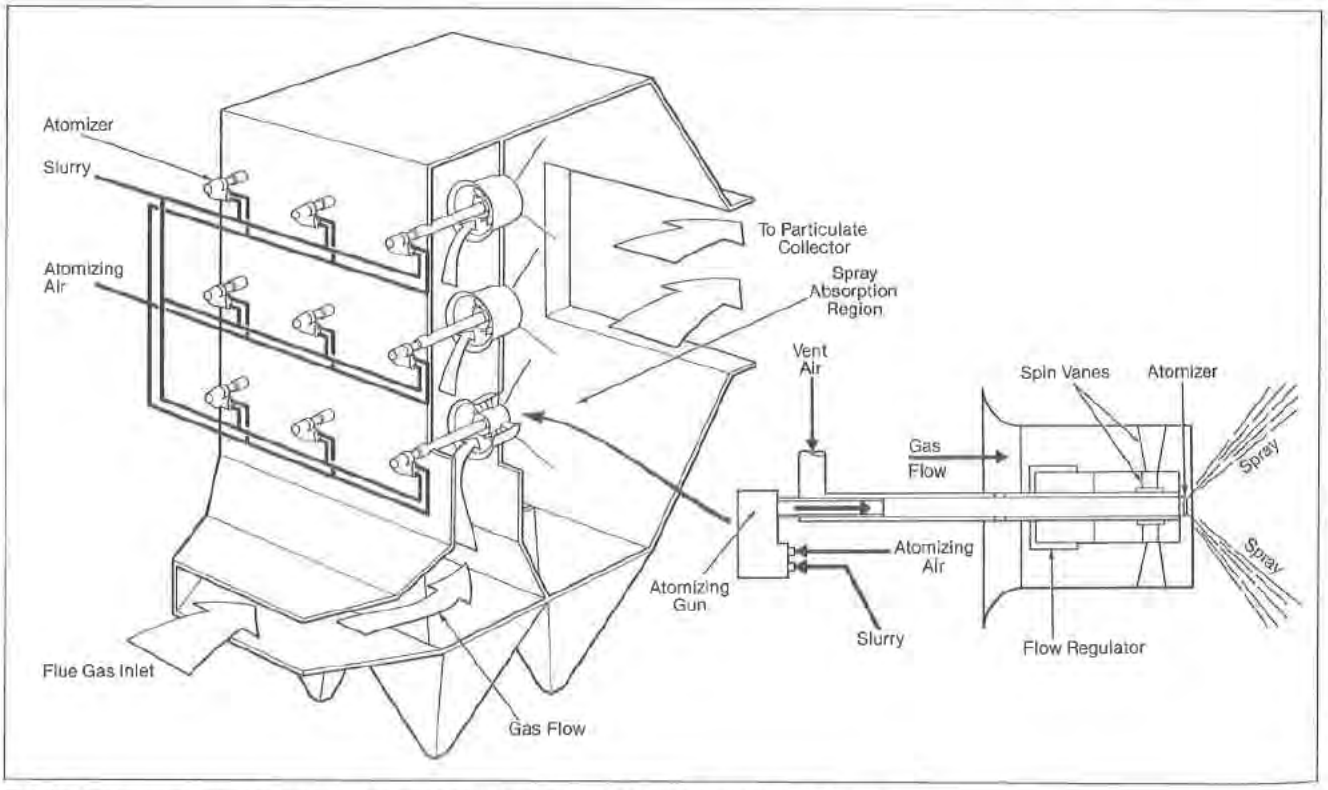


Fig. 13 Dry scrubber reactor module — horizontal flow configuration showing Turbo-Diffuser® insert.

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$$SO_2(g) \leftrightarrow SO_2(aq)$$

Absorption **(33)**

$$Ca(OH)_2(s) \rightarrow Ca^{++} + 2OH^-$$

Dissolution **(34)**

$$SO_2(aq) + H_2O \leftrightarrow HSO_3^- + H^+$$

Hydrolysis **(35)**

$$SO_2(aq) + OH^- \leftrightarrow HSO_3^- \quad \textbf{(36)}$$

$$OH^- + H^+ \leftrightarrow H_2O \quad \textbf{(37)}$$

$$HSO_3^- + OH^- \leftrightarrow SO_3^= + H_2O$$

Neutralization **(38)**

$$Ca^{++} + SO_3^= + \tfrac{1}{2}\, H_2O \rightarrow CaSO_3 \cdot \tfrac{1}{2}\, H_2O\,(s)$$

Precipitation **(39)**

These dry scrubber reactions usually take place in the pH range of 10 to 12.5. The reaction scheme for wet scrubbers depicted by Equations 10 to 19 occurs below a pH of 7.0.

Dry scrubbers are normally sized for a certain gas-phase residence time, which depends on the design approach temperature and the degree of atomization. For example, if the approach temperature is 25F (14C) and the atomization system is designed for a Sauter mean diameter of 50 microns, the reactor residence time would be designed for about ten seconds.

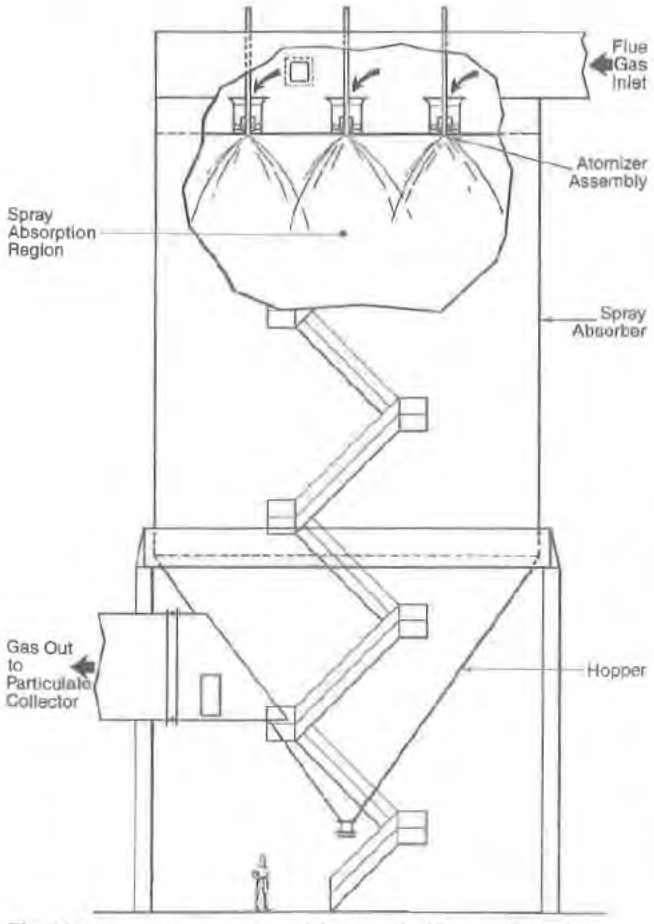


Fig. 14 Dry scrubber reactor module — vertical flow configuration.

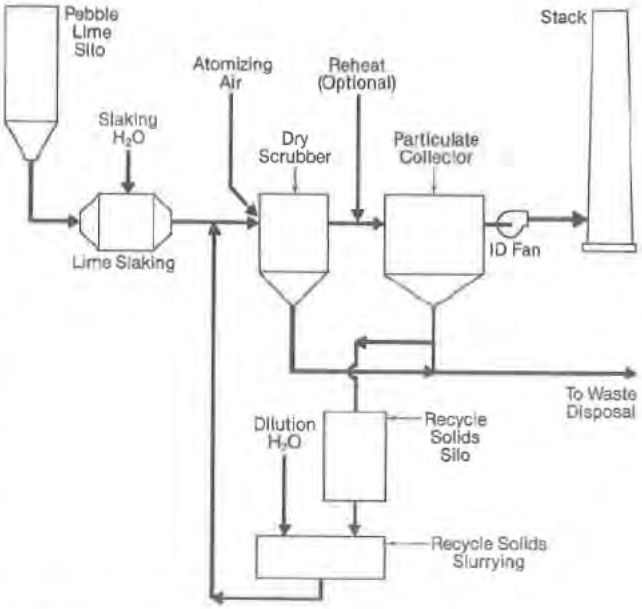


Fig. 15 Dry FGD scrubber system schematic.

The application of dry scrubbers to large electric utility boilers (see Fig. 16) is generally limited to those burning low sulfur coals. This is due primarily to the higher 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.

The primary class of additives used in dry scrubbers is deliquescent salts which, when added to lime, reduce the drying rate and increase the equilibrium moisture content. SO_2 removals above 95% are possible with these additives. Of these salts, calcium chloride is the most popular. Ammonia injection upstream of a dry scrubber also increases SO_2 performance significantly.

Other commercialized technologies

A number of additional approaches and technologies are also available to control SO_2 emissions from fired processes. These range from reducing sulfur in the fuel to special combustion systems to postcombustion technology. Sulfur removal using coal cleaning processes is discussed in Chapter 11. Fluidized beds, illustrating SO_2 control during combustion, are discussed in Chapters 16 and 29. Finally, sulfur removal during coal gasification is covered in Chapter 17.

Other examples of commercialized FGD technology include:

1. furnace sorbent injection,
2. nacholite/trona injection, and
3. activated carbon.

Furnace sorbent injection has developed during the past 20 years. The technology involves the pneumatic in-

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Fig. 16 470 MW utility boiler with dry scrubber and baghouse.

jection of limestone, dolomite or hydrated lime at a gas temperature of 2000 to 2300F (1093 to 1260C). Normally, the injection point is near the nose of the boiler. (See Fig. 17.) Using hydrated lime, 50 to 60% SO_2 capture is achievable with a calcium/sulfur ratio of 2.

Trona and nacholite are naturally occurring forms of sodium carbonate and bicarbonates which react with SO_2 at air heater exit temperatures. A relatively simple injection system for powdered reagent is placed between the air heater and baghouse. SO_2 reactions take place in the flue ahead of the baghouse and on the surface of the bag material. However, sodium carbonates have been observed to catalyze the oxidation of nitric oxide (NO) to nitrogen dioxide (NO_2). Whereas nitric oxide is a colorless gas, NO_2 is a brown gas which creates a visible stack plume.

Limited applications of activated carbon moving beds for SO_2 removal are seen in Germany and Japan. These beds have the added feature of low temperature SCR catalysis for the selective reduction of NO with ammonia.

Although a major worldwide effort has been expended to develop these and other alternatives, more than 97% of all SO_2 emissions control is currently accomplished by conventional wet and dry scrubbing.

Developments and trends

In addition to the commercialized FGD technologies, many new FGD processes are under development at the pilot plant scale. Some of these are at the demonstration stage on operating boilers.

The uncertainty as to which new FGD processes will make it successfully to the marketplace is great. However, it is certain that the successful processes will be those which meet the new regulations in the most economic manner. Processes which minimize capital expenses will be preferred on older power plants nearing retirement and those that are the least costly to build and operate will be preferred on newer power plants.

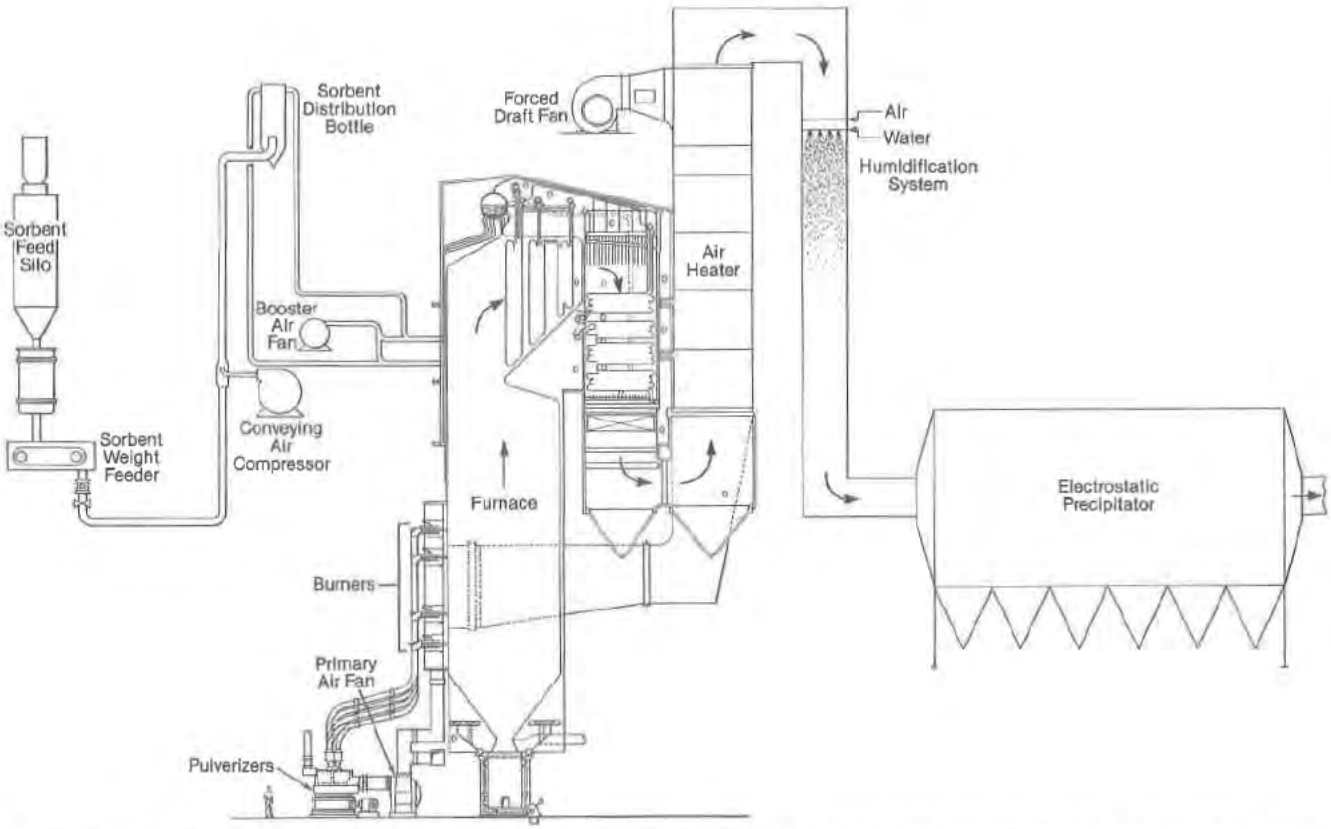


Fig. 17 Furnace sorbent injection system for SO_2 control — includes humidifier system to enhance particulate collection and SO_2 removal efficiency.

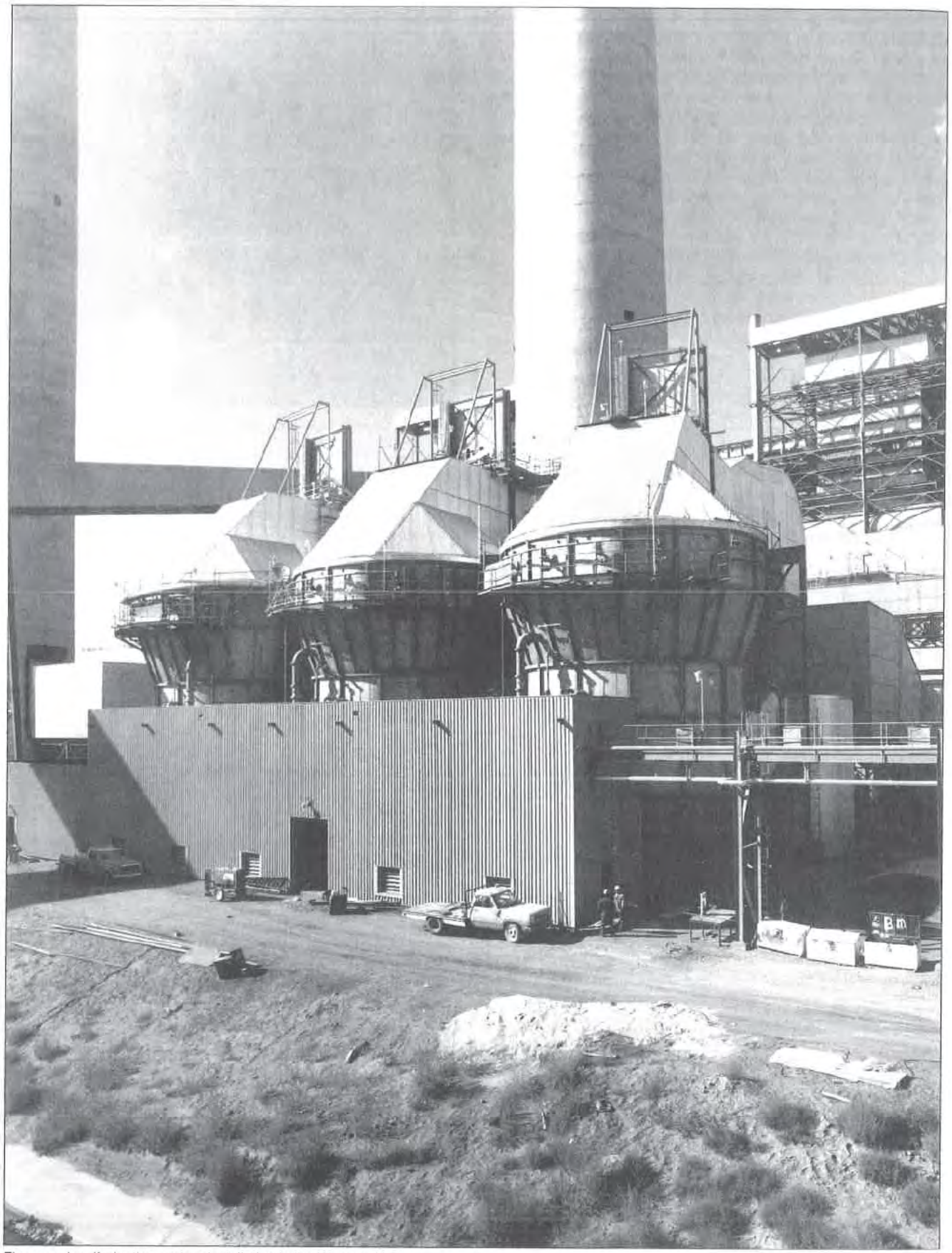
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Flue gas desulfurization system installation in the Western U.S.

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Chapter 36
Environmental Measurement

Environment related measurements at modern power plants are primarily performed to ensure compliance with government regulations. However, measurements are also necessary to monitor and optimize process performance, assess guarantees, calibrate continuous emission monitoring systems (CEMS) and verify design information. As discussed in Chapter 32, fossil fuel power plants emit air, water and solid pollutants. This chapter focuses on measurements for monitoring air pollution systems and, in particular, equipment for controlling sulfur dioxide (SO_2), nitrogen oxides (NO_x) and particulate emissions. Special emphasis is placed on measurements associated with wet SO_2 scrubbing systems which involve perhaps the most complex process.

Postcombustion emissions can be controlled by the use of one or more of the following systems: 1) particulate — electrostatic precipitators or baghouses, 2) NO_x — selective catalytic reduction (SCR) deNO_x systems or selective noncatalytic reduction (SNCR) deNO_x systems, and 3) SO_2 — wet or dry flue gas desulfurization (FGD) scrubbing systems or other forms of sorbent injection. These systems are discussed in Chapters 33, 34 and 35. In each case, the flow rate and composition must be measured for the associated gas, liquid and solid streams. In addition, the overall stack emissions are monitored and recorded.

Instrumentation

Primary measurements taken on environmental equipment include pH, liquid level, solids level, solids concentration (density), continuous emission monitoring for regulated pollutants, static and differential pressure, temperature and flow of liquid, flue gas, air and steam.

Chapter 49 discusses the measurement of pressure, temperature and fluid flow.

pH measurement

The acidic or alkaline nature of a liquid stream is monitored by measuring pH or the potential of hydrogen which is expressed as the negative log of the hydrogen ion concentration ($-\log_{10}[H^+]$). This is the most important parameter in controlling wet FGD systems. pH measurement is also important in water softening, acid neutralization, and other auxiliary processes. The alkalinity, or lack of acidity, is generally proportional to the reagent use in the FGD system. Solutions with a pH less than 7 are considered to be acidic; those having a pH greater than 7 are alkaline.

pH measurements are used to control the feed of fresh reagent slurry to the wet FGD system. (See Chapter 35.) A bleed stream of recirculation slurry is passed through one or two pH sensors. The outputs are typically combined with signals for boiler load, SO_2 level, and fresh reagent density and flow to establish the flow of fresh reagent to the FGD absorber module. An increasing pH reduces the fresh reagent feed flow and a decreasing pH increases the flow.

pH is measured by means of electrodes and a voltmeter. The electrode (Fig. 1) is an electrochemical device comparable to a battery whose voltage changes with pH. In combination electrodes, one of the half cells is the measure-

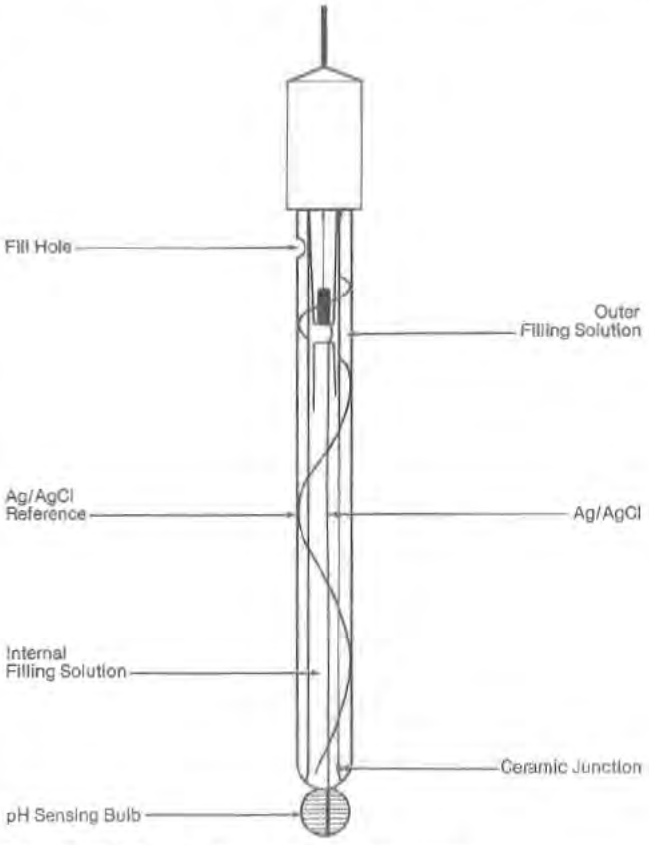


Fig. 1 Combination pH electrode (laboratory version).

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ment electrode or pH sensing bulb and the other is the reference electrode or reference junction. Because the measurement electrode reacts to the pH of the solution, its potential is variable though repeatable. Theoretical electrode output varies from +414 mV at a pH of 0 to -414 mV at a pH of 14. The actual output is sensitive to the reference solution concentration at 77F (25C).

The reference electrode completes the electrical circuit and produces a comparative voltage, so changes in the overall sensor are only functions of the measuring electrode. The reference electrode is comprised of a potassium chloride (KCl) salt solution with a controlled amount of silver chloride (AgCl) dissolved in it. An Ag/AgCl electrode is placed into this electrolyte. The reference potential is a function of the KCl and AgCl concentrations. For example, a 1.0 M (molar) KCl electrolyte produces an offset (E_x), or voltage difference from theoretical, of -8 mV, while 3.3 M KCl produces a -45 mV offset (E_x). The difference, which is consistent across the entire range, is compensated for by standardization or zero adjustment.

The pH sensor is also affected by temperature. The slope (mV per pH unit) changes 1 mV per 5C (see Fig. 2). Temperature is compensated for by slope or temperature adjustment or by automatic temperature compensation.

The pH electrode used in wet FGD processes is subject to fouling and aging. This is compounded by the slurry's abrasive or scaling nature which can cause plugging of the liquid junction. This junction is the point where the KCl reference solution meets the solution being measured. The liquid junction can be a small hole or may be made of fiber, wood, porous ceramic, teflon or kynar. It holds the electrolyte in the probe and enables the electrolyte to form a salt bridge with the measured solution.

There are three types of pH sensors: dip, flow-through and insertion. The dip sensor is inserted into the tank and can be removed for maintenance and calibration. The flow-through sensor is placed in a flowing stream of process solution. The insertion sensor is a variation of the flow-through sensor, but does not require sample lines. It is inserted through packing and a valve into the flowing stream. To maintain or calibrate the insertion sensor, it is retracted through the valve, which is then closed to isolate the process.

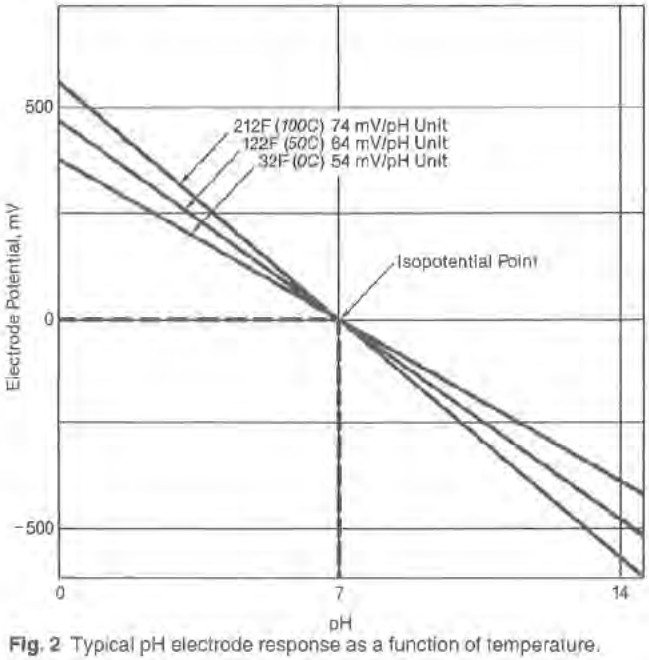


Fig. 2 Typical pH electrode response as a function of temperature.

All sensor types have advantages and disadvantages. The dip sensor is easy to maintain but prone to leakage from scrubber vessels that operate at positive pressure. Sample lines for flow-through sensors can plug. Both flow-through and insertion types are subject to high rates of erosion. All types are subject to scaling, although ultrasonic cleaning devices, which are designed to remove brittle, insoluble coatings, have been used with some success. Ultrasonic cleaners work best if used intermittently, but they can cause probe breakage.

The pH meter is a voltmeter that measures the electrode potential, converts the potential at a given temperature into pH terms, and corrects for nonideal behavior. pH meter operation depends upon proper standardization and calibration. This is best done with stable pH solutions known as buffers. Four of the best buffers, measured at 77F (25C), are pH 4.01, 6.87, 9.18 and 12.45 because they are traceable to the National Bureau of Standards. Two common buffers, pH 7 and 10, are unstable and should be avoided.

Standardization (offset or buffer adjustment) is best done at the isopotential point (pH 7). Standardization adjusts the offset so the meter reads 0.0 mV at pH 7.0. Calibration, or slope adjustment, compensates for alterations in an electrode's response to pH changes. To perform the calibration, the buffer should span the pH of the measured solution. For example, if the expected process pH is 5.4, then a buffer near pH 4 is preferred. The standardization/calibration procedure is summarized below:

1. standardize using a pH 6.87 buffer,
2. calibrate with a buffer that spans the process pH using a different adjustment than that used for standardization (calibration or slope), and
3. use the correct temperature setting, use an automatic temperature compensator or use the same temperature for the two buffers and the process.

Liquid level

Liquid level measurement is needed in any liquid storage tank which is not of the overflow design to control flow out of the tank. Examples in a wet FGD system (Chapter 35) include the tanks for absorber slurry recirculation, reagent storage, thickener underflow and quench recycle. Common liquid level measurement devices are listed in Table 1.

Differential pressure level transmitters are the most common for absorber recirculation tank level control because of their proven reliability. However, they are not the most accurate because of errors due to pressure above the liquid level or due to changes in slurry density.

Other liquid level devices have varying degrees of accuracy due to the tank conditions. For example, most tanks used in FGD systems contain slurry and are therefore agitated, causing waves. The liquid surface level in the absorber tower recirculation tank is also disrupted by: 1) the very high spray flow rates impinging on the surface, 2) foaming due to gas released during the reactions or entrained by the spray, and 3) oxidation air sparged into the tank. Sensors immersed in the slurry also tend to scale. Some level devices use stilling wells to reduce the effect of agitation and to shield the level probe from spray.

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Table 1
Common Liquid Level Measurement Devices for Environmental Control Systems

Sensor Type	Theory of Operation	Environmental Control System Applications	Comments
Differential pressure	Measures differential static pressure with sensors or pressure taps mounted on the wall of the vessel. The sensor contacts the fluid.	Slurry tank level. This technique is often used with a ram or a flush system.	Reliable, subject to plugging in slurry application. Two types are used — flange mounted and extended diaphragm.
Ultrasonic	Emits ultrasonic pulse to the liquid surface. The transit time of the return pulse reflected from surface is converted to the liquid level.	Open tanks containing most fluids or slurries.	Sensors do not contact the liquid. Some problems result from waves, dust, vapor, foam, etc. Background noise can often be filtered electronically.
Electrical capacitance	Detects level by measuring the capacitance between the probe and the tank wall.	High/low level alarm. Discrete level capacitance probes can detect liquid surface.	Some problems with solids buildup or scale.
Float	Float moves up and down with level. The float is on an arm inserted into the vessel or within a housing.	Clean water, mill product tanks (nonscaling): high/low level alarm or constant tank volume.	Magnetic float types are available but are not considered good for scaling applications.
Radio frequency (phase tracking)	Detects an electrical signal which passes through a sensor and is reflected back at both the liquid surface and the bottom of the sensor. The signal phase shift provides a continuous indication of level.	Liquid and solids level.	Relatively new, promising technology. The electric field is much larger than the conductors.
Optical	Detects changes in light intensity.	Sludge level in thickeners.	For sludge level, the change in light intensity is measured at a fixed distance from the surface.

Solids level

Many types of sensors used for measuring liquid level are also used for solids level measurement. However, solids level measurement presents some unique problems due to the often dusty environments, the uneven level caused by bridging or normal angles of repose, or the changes in density or properties due to compaction or aeration. For environmental equipment, solids level measurement is used mainly for storage silos (ash, lime, limestone, gypsum) but is also used for hoppers on precipitators or dry scrubbers. Types of silo measurement devices include electrical capacitance, ultrasonic, and radio frequency (phase tracking) sensors. In addition, load cells or strain gauges, which measure the weight change of the silo as it fills or empties, are common. Another type of solids level device is a cable measurement system in which a weight or float is lowered from the top of the bin or silo. When the float reaches the solids surface, loss of cable tension is detected. The device then measures the distance traveled by counting electronic pulses as the float retracts. The measured level is therefore the silo maximum level less the distance traveled by the probe. This device should not be used when the silo is being filled.

A typical device used to determine high hopper level is a nuclear transmitter. When the ash level rises above the sensor, the strength of the signal from the nuclear source to the receiver is decreased. This signal is converted to a level indication.

Another high/low solids level device is the paddle wheel, which includes a motor that turns a paddle. When material is present, rotation of the paddle stops. The increased torque of the motor as it attempts to turn the stopped paddle triggers a microswitch alarm.

Finally, vibrating probes are sometimes used for single point solids level detection. The vibration of a piezoelectric transducer-driven probe inserted into the vessel is dampened by contact with material. This is detected as a voltage change, which is electronically amplified to yield a level indication.

Solids concentration (density)

The most commonly used devices for density measurement are nuclear absorption meters (see Fig. 3.) These devices measure the absorption of gamma rays from a radioactive source, and the degree of absorption is proportional to the solids content. Nuclear density meters do not contact the slurry; instead, they are installed on the outside of a slurry pipe. Some disadvantages of nuclear density meters are:

1. The signal is not linear with solids content unless a linearizer is used. If accurate chemical analysis of the slurry is available, the device can be precalibrated.
2. A nuclear license is required in the United States (U.S.) and in most other countries.
3. The device does not distinguish between suspended and dissolved solids.
4. Errors can result due to the buildup of solids or scale in the pipe.

Nuclear density meters are typically low maintenance

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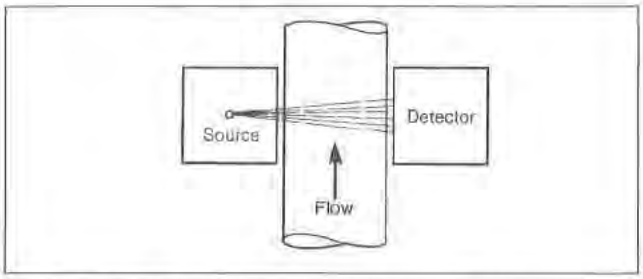


Fig. 3 Nuclear absorption meter for density measurement.

devices. The inaccuracies mentioned above are routinely verified by manual sampling and measurement.

Other devices used for density measurement include differential pressure and ultrasonic devices and vibrating reed instruments. The reed instruments electronically convert the slurry dampening effect on electrically-driven coil vibrations to a density measurement.

Pressure, flow and temperature

Pressure, pressure differential, flow and temperature measurements are common in environmental system applications. They indicate process performance, energy consumption, operational problems and ability to meet design or operating requirements. Theory, instrumentation and application for these devices are discussed in detail in Chapter 40. Selected uses in environmental systems are discussed in Chapters 33 through 35. Comments with particular importance in environmental systems include:

1. Slurries tend to be highly erosive and potentially corrosive, and provision must be made for appropriate environments.
2. The presence of solids in liquid or gas flows can lead to plugging of measuring devices. Purging systems are frequently included to ensure long term operation.
3. Temperature monitoring can be particularly important for freeze protection, crystallization prevention, thermal gradient prevention (thickeners and clarifiers), corrosion prevention (precipitators, fabric filters, dry FGD systems and flues/ducts) and general process control. Elastomeric linings and fiberglass or plastic components must be protected from high temperatures. Unlined stacks must be protected from low temperatures.

Continuous emission monitoring systems

Early power plant emission measurements focused on controlling combustion. The Orsat gas analyzer, which sampled oxygen (O_2), carbon monoxide (CO) and carbon dioxide (CO_2) flue gas concentrations, was used extensively. However, this analyzer is not a continuous monitor.

Largely due to increasing regulation, power plants have been forced to institute continuous emission measurements. In addition, continuous monitoring devices permit process control of the emissions control equipment.

During the 1980s, CEMS were used primarily to measure particulate (opacity), SO_2, NO_x and CO_2. By the early 1990s, CEMS to monitor hydrogen chloride (HCl), CO, ammonia and volatile organic compounds (VOC) were introduced for commercial combustion applications.

The common methods for continuously measuring flue gas constituents are described in Table 2 and are further detailed in Table 3. All of the analyzers are similar in that they use a measuring cell and a reference cell. They are set to zero with air or a calibration standard, the span is set with a calibration standard, and a sample is taken. The result is obtained by comparing the voltages detected across the sample, the zero standard and span standard.

There are three types of analyzers — extraction, dilution extraction and in situ.

In the extraction method (Fig. 4) a sample of flue gas is withdrawn from the process stream and directed to an external analyzer for the constituent concentration determination. This type of analyzer is typically used where there are short distances between the sample point and the analyzer. The probe and sample line are often heated to prevent condensation. Because the gas sample is typically drawn by vacuum to the external analyzer, sample line leaks can compromise the integrity of the sample.

Table 2
CEM Technologies

Technology	Operating Characteristics
Infrared radiation (IR)	An infrared beam passes through a measurement filter and is absorbed by the constitutent gas. A light detector creates a signal which is used to monitor concentrations.
Ultraviolet absorption (UV)	A split beam, with optical filters, phototubes and amplifiers, measures the difference in light beam absorption between the reference and sample.
Chemiluminescence	Ozone (O_3) is injected into the sample to react with NO_x, generating light that is measured by a photocell.
Flame ionization detection	Hydrocarbons are ionized with strong light. The signals are received by the flame ionization detector.
Transmissometer	Light is passed through the stack where it is reflected by a mirror on the opposite side. The quantity of light returning is proportional to particulate matter and aerosols in the flue gas. (See Fig. 5.)
Electrochemical cells	The voltage measured when an O_2 sample is injected into a solution with a strong base is compared to a reference voltage.
Chromatography	A sample, zero and calibration gas are eluted through a column. The output as measured by flame photometric or thermal conductivity detectors is compared.

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Table 3
Continuous Emission Monitoring
(Reference Table 2 for Technologies)

Constituent	CEM Technology
Particulate matter (opacity)	Transmissometer, beta ray absorption
Sulfur dioxide (SO_2)	UV absorption, IR pulsed fluorescence
Nitrogen oxides (NO_x)	Chemiluminescence, UV spectroscopy, IR
Hydrogen chloride (HCl)	IR with gas filter
Carbon monoxide (CO)	IR
Carbon dioxide (CO_2)	IR
Oxygen (O_2)	Electrochemical cell
Volatile organic compounds (VOC)	Flame ionization detection
Other organic air toxics	Chromatography
Ammonia (NH_3)	Same as NO_x*

* NH_3 is converted to NO_x in one of two split streams. Both streams are analyzed for NO_x. NH_3 is determined as the difference.

The dilution-extraction method uses a carrier/dilution media, typically instrument air, which is supplied to the extraction probe in the flue gas process stream. Problems associated with dirty, wet and corrosive flue gases are drastically reduced by the 50 to 200 dilution ratio provided by the sample probe. Precise dilution is assured by accurately controlling the dilution air pressure which yields a constant air flow. The air flow creates a vacuum on the downstream side of a critical sample orifice. The gas sample rate, limited by sonic flow through the orifice, is therefore held constant. Furthermore, because the sample lines from the probe to the analyzer are under positive pressure, minor sample line leakage does not affect the integrity of the sample concentration as it would in a vacuum system. This method has gained wide acceptance in utility and other systems where longer sample lines are required.

In situ analyzers, such as that shown in Fig. 5, provide measurement of flue gas concentrations directly in the process stream. Several types of detection devices are available, and all are based on absorption spectroscopy. Since gas stream particulates can reduce the level of light transmission, ceramic filters are used to exclude these particles. A single in situ probe can measure several gas species and opacity. Problems with in situ analyzers generally occur in the optic or electronic components. Because a separate analyzer is required for each sample point, installations which require multiple sample points frequently find the extractive systems more cost effective.

Continuous analyzers generally require flue gas conditioning. Sample conditioning includes filtration to remove particulate (see Fig. 6), chilling to remove water, heating to maintain process temperature, and dilution (used with extractive monitor and in situ probe).

The need for conditioning coupled with zero and span calibration make extractive systems complicated, costly and vulnerable to plugging, corrosion and leaking. These

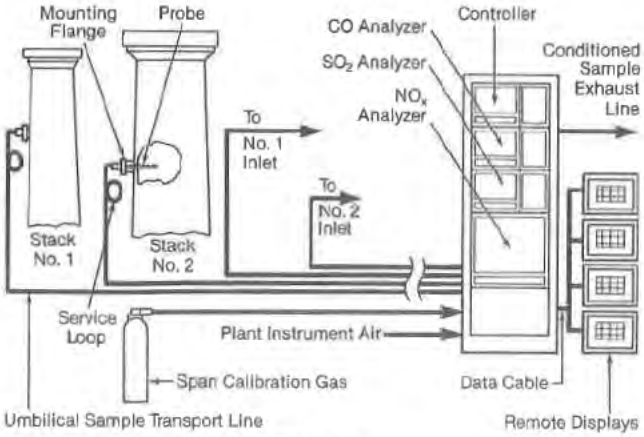


Fig. 4 Extractive stack/inlet CEM system.

systems are also slow to respond due to the time lapse between sampling and analysis. In situ systems use conditioning but do not rely on long sample lines. These systems have faster response, fewer components and better availability than extractive systems. However, they are prone to interference and fouling and are less accessible for service. Extractive systems can be expanded by adding sample points as shown in Fig. 4. It is important to consider the location and ambient conditions in which CEMS equipment is placed. Many of the problems associated with CEMS can be traced to dirty conditions, excessive heat or cold, vibration, humidity, etc.

Monitoring process performance

Environmental control systems (ECS) involve physical and chemical processes. Chemical methods are used to maintain desired removal or emission levels, avoid operating problems, improve reagent utilization, improve process efficiency, troubleshoot equipment, and monitor process changes and catalyst life.

In an FGD system, the quality of the reagent and the recirculating slurry stream is monitored. Analysis is more critical for processes which produce a usable byproduct such as gypsum. In this case, the analysis extends into product purity and feed stream impurities. The quality of the reagent is normally measured extensively when

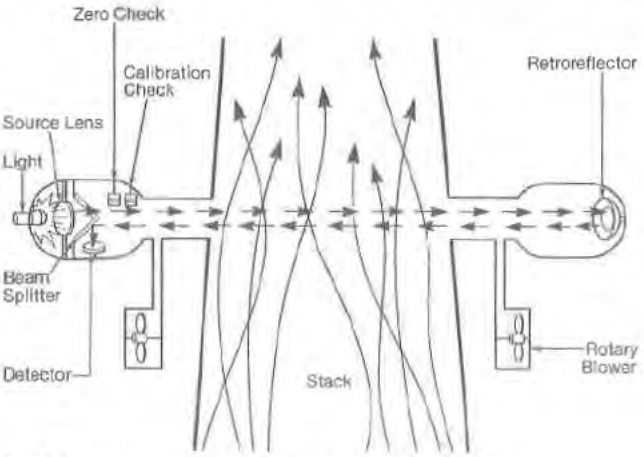


Fig. 5 In situ transmissometer system.

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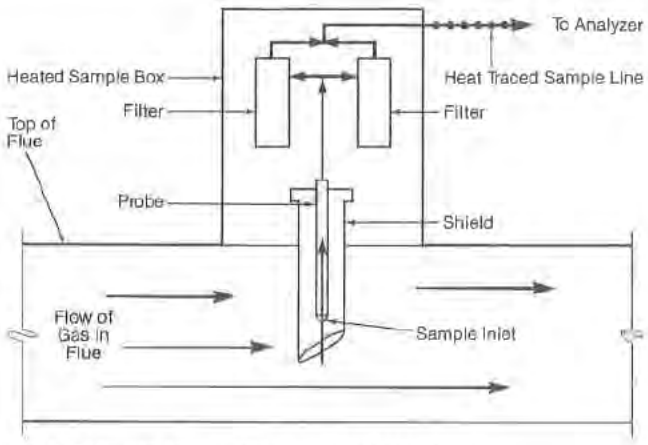


Fig. 6 Extractive sample probe with dust conditioning.

the reagent source is selected. Later it is evaluated if the source is changed or process problems occur.

Tables 4 through 7 list the important analyses for common wet scrubber FGD systems. Additional analyses are required for makeup water streams, sodium-based FGD systems, dual loop systems, regenerative systems, ex situ oxidation processes, water treatment, and sulfuric acid or SO_2 recovery processes. However, the major species and methods would be similar to those presented in Tables 4 through 7; only the concentrations would vary. Other procedures used for design or troubleshooting include relative saturation, which is determined by seed crystal procedures, and ion balance (balance of anions and cations), which is a check of the analysis accuracy.

Performance testing

As with any system, performance tests are conducted on environmental control systems to determine if the units are operating properly. A typical new environmental control system might include low NO_x burners, wet FGD scrubber modules, gypsum processing equipment, and an electrostatic precipitator. An SCR deNO_x system could also be added. The ten most common parameters measured to determine performance for such a system include:

1. system pressure drop,
2. particulate removal efficiency,
3. stack opacity,
4. NO_x emission concentration,
5. SO_2 removal efficiency,
6. makeup water requirements,
7. reagent consumption,
8. gypsum purity,
9. mist eliminator carryover, and
10. electric power consumption.

Other performance parameters that would be measured depend on the type of system. Performance testing is usually extensive. It is common for a new ECS to have two performance tests. The first test is scheduled soon after installation and the final test is performed after one year of operation.

Sample location selection

The first step in the performance test is often done before the ECS is built — selecting the sample locations. Outlet sampling is not difficult if it is done in the stack. However, the inlet and intermediate samples are more difficult to obtain. In Fig. 7, the system inlet at the far left is a section of flue followed by an immediate upward bend. The precipitator outlet is at the induced draft (ID) fan inlet. Fig. 7 also shows test ports in the downward flue section after the absorber tower and at the stack entrance. Test port locations are selected to give the best flow profile within the plant layout limitations.

The number and location of traverse points are selected by Environmental Protection Agency (EPA) Method 1. Application of Method 1 is illustrated in Fig. 8. The distance downstream from the flow disturbance,

Table 4
Lime Quality Analysis

Parameter	Suggested Method*	Suggested Frequency	Use of Data
Slaking rate	ASTM Method C-110-76a	Duplicate analysis on all candidate limes	Temperature rise rate is related to reactivity and lime quality.
Grit	Acid dissolution	Duplicate analysis on all candidate limes	High grit levels indicate poor lime quality.
Weight loss on ignition	Gravimetric	Duplicate analysis on all candidate limes	Weight loss is an indirect measure of lime quality via carbonate and hydroxide content.
Available lime index	Rapid sugar test	Duplicate analysis on all candidate limes	Lime quality
Calcium/ magnesium	Atomic absorption (AA) EDTA titration or Ion Chromatograph (IC)	Duplicate analysis on on all candidate limes then as needed to check reagent quality	Analysis is required for processes producing a usable byproduct or high magnesium lime systems.
Lime slurry solids content, % by wt	Gravimetric	Daily	Check density meter calibration and indicate slaker performance.

* Many methods have been standardized through the American Society of Testing and Materials, Philadelphia, Pennsylvania.

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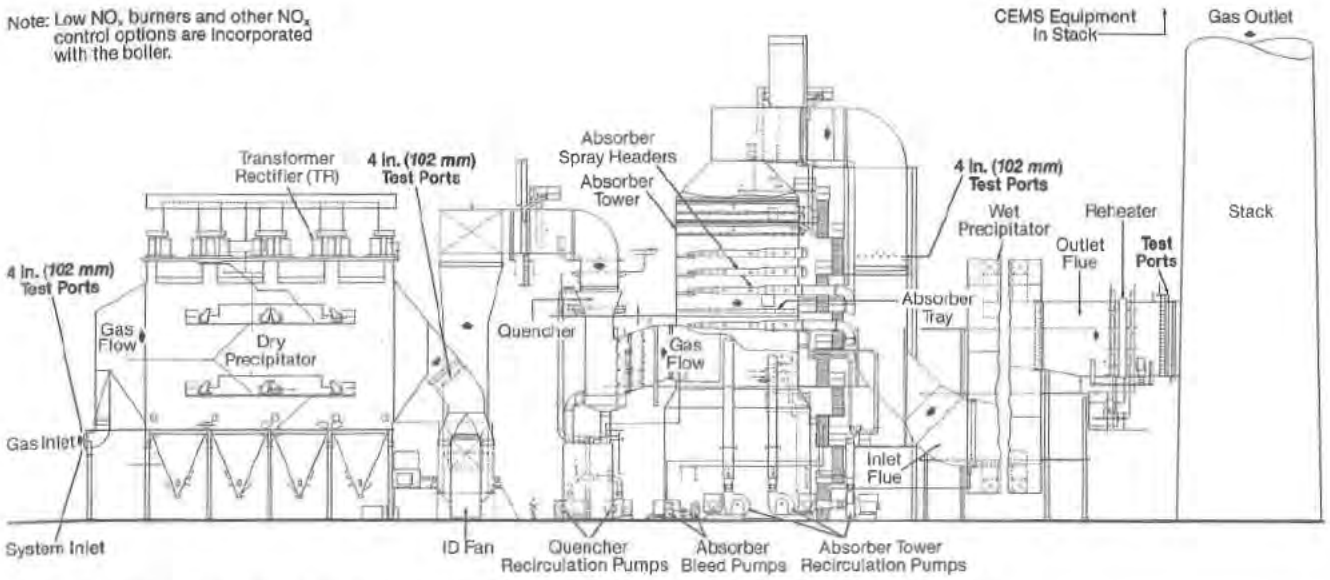


Fig. 7 Typical test port locations indicating compromises made in measurement due to equipment arrangement.

which in this case is the stack inlet, is 138 ft (42.1 m). This is more than eight diameters (138/16 = 8.6). The upstream distance to the stack exit is 5 diameters. Therefore, 12 traverse points are required by Method 1. For rectangular flues, Method 1 is applied by first determining an effective or equivalent diameter which is calculated as follows:

$$DC = \frac{2LW}{(L+W)} \quad (1)$$

where

DC = equivalent diameter
L = length
W = width

Then a matrix layout is used for sample location selection. The cross-section is divided into as many equal rectangular areas as there are traverse points, and the sample is taken at the centroid of each area.

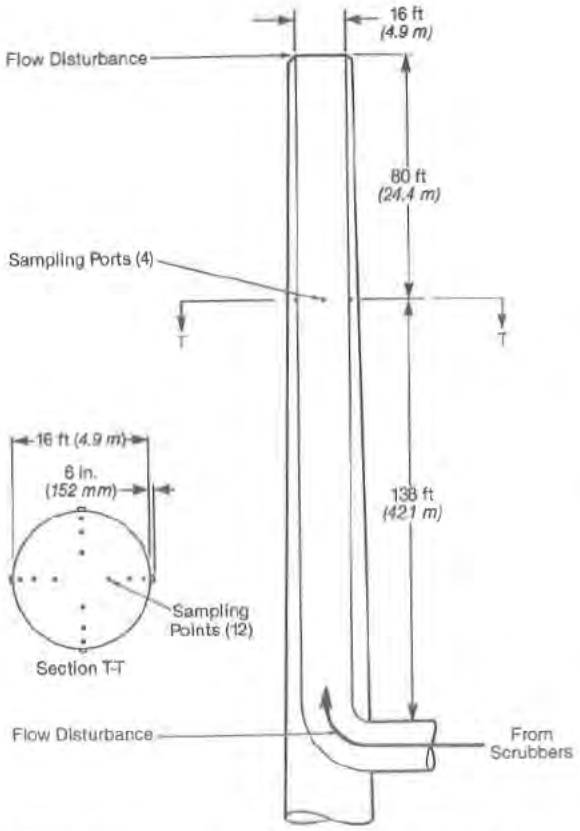


Fig. 8 Stack dimensions and sampling port locations for EPA Method 1.

Gas side measurements

For a performance test of an ECS, a combination of EPA Methods 5 and 8 is the most often used procedure to simultaneously measure particulate, SO_2, and SO_3 or acid mist (H_2SO_4) levels. The Method 5 sampling train, shown in Fig. 9, is used with Method 8 impingers shown in this figure. The data obtained from this traverse also represent Methods 1 through 4. Results include gas velocity by S-type pitot tube (Fig. 10) and CO_2, O_2, moisture, SO_2, particulate, and SO_3 or H_2SO_4 levels. The following calculations are used:

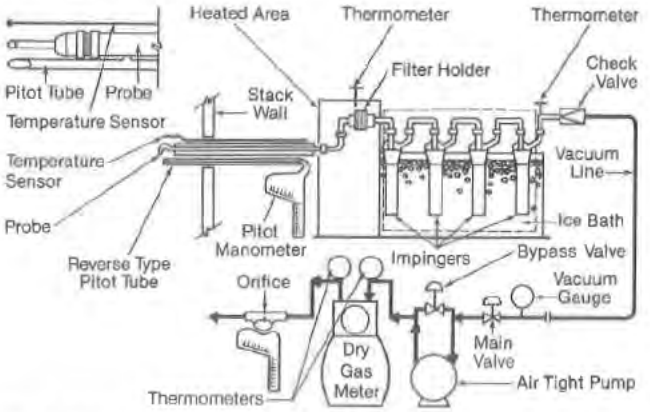


Fig. 9 EPA Method 5 particulate and SO_3 emissions sampling train.

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Determination of moisture content in stack gas

$$B_w = \frac{(MWC)\ (1.339)}{[(MWC)\ (1.339)] + \frac{(DGV)\ (P_m)}{T_m}\ (499.9)} \quad (2)$$

where

B_w = moisture fraction
MWC = impinger weight gain, g
DGV = sampled dry gas volume at meter conditions, ft^3
P_m = pressure at meter, in. Hg
T_m = temperature at meter, R
1.339 = constant for converting grams of H_2O to liters of H_2O vapor at standard conditions, l/g
499.9 = conversion of cubic feet at meter conditions to liters at standard conditions, l R/ft^3/in. Hg

Determination of flue gas molecular weight — dry basis

$$M_d = [(44)(\%\ CO_2\ \text{dry}) + (32)(\%\ O_2\ \text{dry}) + (28)(\%\ N_2\ \text{dry})] \times 0.01 \quad (3)$$

where

M_d = molecular weight of gas on dry basis, lb/lb-mole
44 = molecular weight of CO_2, lb/lb-mole
32 = molecular weight of O_2, lb/lb-mole
28 = molecular weight of N_2, lb/lb-mole

Determination of flue gas molecular weight — wet basis

$$M_w = M_d(1 - B_w) + 18B_w \quad (4)$$

where

M_w = molecular weight of gas on wet basis, lb/lb-mole
M_d = molecular weight of gas on dry basis, lb/lb-mole
B_w = moisture fraction, lb-mole H_2O/lb-mole wet gas
18 = molecular weight of H_2O, lb/lb-mole

Determination of gas humidity

$$H = \frac{18B_w}{M_d(1 - B_w)} \quad (5)$$

where

H = gas humidity, lb H_2O/lb dry gas
B_w = moisture fraction, lb-mole H_2O/lb-mole wet gas
M_d = molecular weight of dry gas, lb/lb-mole
18 = molecular weight of H_2O, lb/lb-mole

Determination of stack gas velocity

$$V_s = 85.48\ C_p \left(\frac{T_s\ \Delta P}{P_s\ M_w} \right)^{1/2} \quad (6)$$

where

85.48 = pitot tube constant in the following units:

$$\frac{\text{ft}}{\text{s}} \left(\frac{(\text{lb/lb-mole})\ (\text{in. Hg})}{(\text{R})\ (\text{in. } H_2O)} \right)^{1/2}$$

V_s = velocity of the flue gas, ft/s
ΔP = average differential pressure measured by S-type pitot tube, in. H_2O
T_s = average gas temperature in duct, R
P_s = absolute duct pressure, in. Hg

Table 5
Limestone Quality Analysis

Parameter	Suggested Method*	Suggested Frequency	Use of Data
Reactivity	pH stat test followed by particle size measurement	Duplicate analysis of candidate limestones	Compared to reference to determine acceptability.
Grindability (Bond Work Index)	Laboratory ball mill test	Duplicate analysis of candidate limestones	Limited by specification of wet milling system design.
Particle size of limestone as delivered	Sieve method	Duplicate analysis of candidate limestones then monitor monthly	Limited by specification of wet milling system design.
Inerts	Acid dissolution	Duplicate analysis of candidate limestones then monitor weekly for gypsum systems	Quality indicator; important for gypsum systems.
Carbonate	CO_2 evolution, $Ba(OH)_2$ absorption, alkalinity titration	Duplicate analysis of candidate limestones then monitor weekly for gypsum systems	Check reagent quality; important for gypsum systems.
Calcium and magnesium	AA, IR, EDTA titration	Duplicate analysis of candidate limestones then monitor weekly for gypsum systems	Check reagent quality.
Limestone slurry solids content, % by wt	Gravimetric	Daily	Check density meter and indicate problems in milling circuit.
Limestone slurry particle size	Wet sieve	Daily	Troubleshooting problems in milling circuit or process such as low utilization or poor gypsum quality.

* Many methods have been standardized through the American Society of Testing and Materials, Philadelphia, Pennsylvania.

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Table 6
Recirculation Slurry/Blowdown Analysis

Parameter	Suggested Method*	Suggested Frequency	Use of Data
pH	2-buffer method calibration or grab sample	Once per shift by grab sample, daily by buffer method	Troubleshoot pH meter and maintain process efficiency.
Slurry suspended solids content	Gravimetric	Weekly	Calibrate density meter and maintain process efficiency.
Cations: Ca^{++} and Mg^{++} primary; K^+ and Na^+ secondary	IC, AA or wet chemistry methods	Daily for Ca^{++} and Mg^{++} in calcium-based systems	To calculate stoichiometry and monitor process chemistry.
Anions: $SO_3^=$, $CO_3^=$, $SO_4^=$ and Cl^- primary; F^- and NO_3^- secondary	IC or wet chemistry methods	Daily for $SO_3^=$, $CO_3^=$, $SO_4^=$, Cl^-	To calculate stoichiometry, measure chloride concentration, and monitor process performance and gypsum production.
Dissolved metals: Fe, Mn, Al, etc.	AA	Infrequently	Process troubleshooting.
Additives: adipic acid, dibasic acid, formate, thiosulfate, scale inhibitors, etc.	IC or wet chemistry methods	Daily	Determine or correct dosage.
TDS (total dissolved solids)	Gravimetric	Weekly	Calibrate density meter for suspended solids; important for mag-lime, high chloride or suspended solids.
Dissolved alkalinity	Acid-base titration	Weekly	Measure alkali loss in filter cake of mag-lime process; important for mag-lime or sodium systems.
Settling tests	Graduated cylinder	Weekly	Predict thickener performance and assist process troubleshooting.
Filter leaf test	Filter and weight	During startup	Predict filter cake properties.
Particle size distribution	Coulter counter	During startup	Determine settling or dewatering characteristics; important for gypsum systems.

* Many methods have been standardized through the American Society of Testing and Materials, Philadelphia, Pennsylvania.

M_w = wet gas molecular weight, lb/lb-mole
C_p = S-type pitot tube correction factor (normally 0.84)

Volumetric flow rate — actual conditions

$$Q_{ac} = 60\, V_s A \qquad (7)$$

where

Q_{ac} = actual volumetric flow rate, ACFM
V_s = velocity of the flue gas, ft/s
A = cross-sectional area of the duct, ft^2
60 = conversion factor, s/min

Volumetric flow rate — dry standard conditions

$$Q_{sd} = \frac{528\ Q_{ac}\ P_s\ (1-B_w)}{29.92\ T_s} \qquad (8)$$

where

Q_{sd} = dry volumetric flow rate at standard conditions, DSCFM
Q_{ac} = actual volumetric flow rate, ACFM
B_w = moisture fraction, lb-mole H_2O/lb-mole wet gas
528 = standard temperature, R
T_s = average gas temperature in duct, R
29.92 = standard pressure, in. Hg
P_s = absolute duct pressure, in. Hg

Total particulate and H_2SO_4 — grains/ACF

$$C_a = \frac{(0.001)\ (15.43)\ (528)\ (1-B_w)\ P_s M_n}{29.92\ V_d\ T_s} \qquad (9)$$

where

C_a = concentration of specie in the flue gas at duct conditions, grains/ACF
M_n = mass of specie collected, mg
V_d = dry gas volume sampled at standard conditions, DSCF
T_s = average gas temperature in duct, R
P_s = absolute duct pressure, in. Hg
B_w = moisture fraction of gas, lb-mole H_2O/lb-mole wet gas
0.001 = conversion of milligrams to grams, g/mg
15.43 = conversion of grams to grains, grains/g
528 = standard temperature, R
29.92 = standard pressure, in. Hg

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Table 7
Dewatering System Analysis

Parameter	Suggested Method*	Suggested Frequency	Use of Data
Thickener or hydroclone underflow solids, %	Gravimetric	Daily	Troubleshoot process and calibrated lime density meters
Clarified water, water analysis (anions, cations, suspended solids)	See Table 6	Weekly	Monitor thickener performance
Supernatent, filtrant solids content, % by wt	Gravimetric	Daily	Monitor filter cloth status
Gypsum Quality Analysis			
Ca^{++}, Mg^{++}, Na^{+}, K^{+}, Fe_2O_3, R_2O_3, SiO_2	AA	Weekly	Product purity and properties
Cl^{-}, $SO_4^{=}$	IC	Daily	Product purity and properties
$SO_3^{=}$, $CaSO_3 \cdot {}^1/_2\ H_2O$	Iodimetric titration	Daily	Product purity and properties
Moisture content	Gravimetric	Daily	Important for salable gypsum and monitoring dewatering system performance.
Surface area	BET analysis (sub-sieve sizer)	Startup or process change	N_2 gas is condensed on surface and absorbed; N_2 corresponds to surface area.
Combined water	Heat to 482F (250C)	Weekly	Gypsum purity
Mean particle size	Coulter counter, Sedigraph plus sieve analysis	Startup or process change	Product purity and properties
Aspect ratio	Scanning electron microscope	Startup or process change as required by end users	Product purity and properties
Total water soluble salts	Sum ($Na^{+} + K^{+} + Mg^{++}$)	As needed	Product purity and properties
Flyash	Determined as acid insoluble	Daily	Determined as acid insolubles along with silica (SiO_2) and other impurities.
pH	Electrode	Daily	Product purity and properties

* Many methods have been standardized through the American Society of Testing and Materials, Philadelphia, Pennsylvania.

Total particulate, H_2SO_4, condensible total particulate and condensible H_2SO_4

$$C_s = \frac{(0.001)\ (15.43)\ M_n}{V_d} \qquad (10)$$

where

C_s = concentration of specie in the flue gas at dry conditions, grains/DSCF
M_n = mass of specie collected, mg
V_d = dry gas volume sampled at standard conditions, DSCF
0.001 = conversion of milligrams to grams, g/mg
15.43 = conversion of grams to grains, grains/g

Total particulate, H_2SO_4, NO_x and SO_2 — lb/h

$$C_m = C_s\,(1.428 \times 10^{-4})(60)(Q_{sd}) \qquad (11)$$

where

C_m = mass flow rate of specie in the flue gas, lb/h
C_s = concentration of specie in the flue gas at dry standard conditions, grains/DSCF
Q_{sd} = dry volumetric flow rate at standard conditions, DSCFM
60 = conversion of minutes to hours, min/h
1.428×10^{-4} = conversion of grains to pounds, lb/grain

Total particulate, H_2SO_4, NO_x and SO_2

$$E = \frac{20.9\ C_m\ F_d}{60\ Q_{sd}\ (20.9 - \%\,O_{2ds})} \qquad (12)$$

where

E = emission rate on a dry basis at 0% O_2, lb/10^6 Btu
C_m = mass flow rate of specie in the flue gas, lb/h
Q_{sd} = dry volumetric gas flow rate at standard conditions, DSCFM
$\%\ O_{2ds}$ = O_2 in the gas on a dry basis, %
F_d = dry F-factor at 0% O_2, determined from fuel analysis, DSCF/10^6 Btu (see Equation 17)
20.9 = O_2 in air, used with % O_2 in the gas to correct to 0% O_2, %
60 = conversion of minutes to hours, min/h

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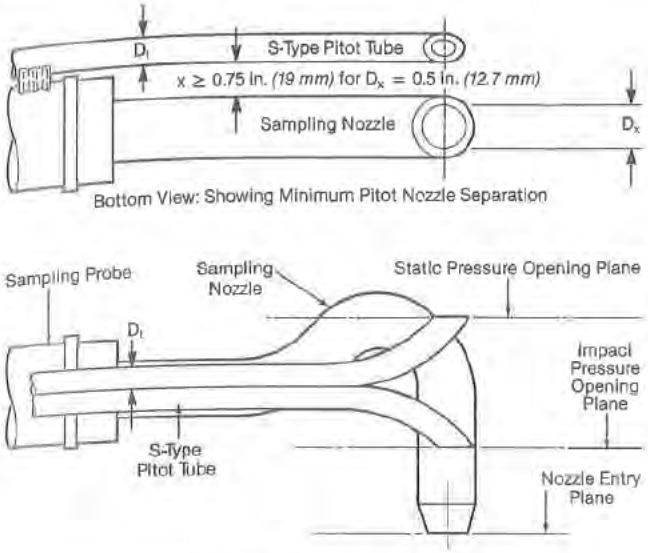


Fig. 10 S-type pitot tube with sampling probe.

SO_2 — parts per million (ppm) wet

$$C_{ppm,w} = \frac{(8.48\times10^{-4})\ (10^6)\ V_s C_i\ (1-B_w)}{V_d} \quad (13)$$

where

$C_{ppm,w}$ = gas SO_2 concentration on a wet basis, ppm wet
V_s = total volume of H_2O_2 impinger solution
C_i = SO_2 concentration in the impinger solution, mg/liter
V_d = gas volume sampled at dry standard conditions, DSCF
B_w = gas moisture fraction, lb-mole H_2O/lb-mole wet gas
8.48×10^{-4} = conversion of millimoles to standard cubic feet, SCF/mM
10^6 = conversion from fraction to ppm

SO_2 — ppm dry

$$C_{ppm,d} = \frac{20.9\ C_{ppm,w}}{(1-B_w)\ (20.9-\%\,O_{2ds})} \quad (14)$$

where

$C_{ppm,d}$ = gas specie (SO_2) concentration on a dry basis, corrected to 0% O_2, ppm dry
$C_{ppm,w}$ = gas specie (SO_2) concentration on a wet basis, ppm wet
B_w = moisture fraction of the gas, lb-mole H_2O/lb-mole wet gas
$\%\,O_{2ds}$ = O_2 in the gas on a dry basis, %
20.9 = O_2 in air, used with % O_2 in the gas to correct to 0% O_2, %

NO_x — ppm dry and ppm wet

NO_x is evaluated directly by monitoring systems as parts per million, dry basis. It can be converted to ppm wet if needed by solving Equation 14 for $C_{ppm,w}$.

Collection/removal efficiency — total particulate, H_2SO_4 and SO_2

$$C_{eff} = \left[(C_{in} - C_{out})/C_{in}\right]\times 100 \quad (15)$$

where

C_{eff} = collection/removal efficiency, %
C_{in} = concentration of matter at inlet, lb/10^6 Btu
C_{out} = concentration of matter at stack, lb/10^6 Btu

Percent isokinetic

$$\%\,I = (V_n / V_s)\times 100 \quad (16)$$

where

$$V_n = \frac{V_m T_s P_m}{60\ t A_n T_m P_s (1-B_w)}$$

and where

$\%\,I$ = % isokinetic
V_s = average flue gas velocity in duct during sampling, ft/s
V_n = average sample gas velocity through the nozzle during sampling, ft/s
V_m = dry gas volume sampled at meter conditions, ft^3
t = elapsed sampling time, min
A_n = cross-sectional area of nozzles, ft^2
B_w = gas moisture fraction, lb-mole H_2O/lb-mole wet gas
T_s = flue gas temperature in the duct, R
T_m = average gas temperature at meter, R
P_m = absolute pressure at the meter, in. Hg
P_s = absolute pressure in the duct, in. Hg
60 = conversion from minutes to seconds, min/s

F-factor

The F-factor used in making emissions rate calculations was developed from chemical and combustion analyses of the fuel. This factor is the ratio of the theoretical volume of dry gases at 0% excess air (O_2) given off by complete combustion of a known amount of fuel, to the gross caloric value of the burned fuel. The data generated from the ultimate analysis of the fuel are used to calculate the F-factor as follows:

$$F_d = (10^6)\frac{3.64(H) + 1.53(C) + 0.57(S) + 0.14(N) - 0.46(O)}{HHV} \quad (17)$$

where

F_d = dry F-factor at 0% O_2, DSCF/10^6 Btu
HHV = higher heating value, Btu/lb
H = hydrogen in the fuel, % by wt
C = carbon in the fuel, % by wt
S = sulfur in the fuel, % by wt
N = nitrogen in the fuel, % by wt
O = oxygen in the fuel, % by wt

F-factors are constant for a given fuel category (within ± 3%). Average values are:

Fuel	DSCF/10^6 Btu
Bituminous coal	9820
Oil	9220
Gas	8740
Wood bark	9640
Wood chips	9280

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Particulate and SO_2 removal efficiencies

The SO_2 and particulate removal efficiencies are determined by inlet and outlet sampling using Equations 12 and 15.

Stack opacity

Stack opacity is determined by the installed transmissometer or by EPA Method 9.

NO_x concentration

NO_x concentration is determined by EPA Method 7. This measurement is taken in the stack on a composite sample. NO_x concentration may be adjusted to design conditions from actual test conditions.

Makeup water requirements

Makeup water requirements are measured at the appropriate terminal point with a certified flow meter. Adjustments from test conditions may be necessary due to variations in gas volume or temperature or other process deviations from design conditions.

Reagent consumption

There are two common ways to measure reagent consumption. The first method is to measure the solid reagent flow (for lime/limestone systems) using a calibrated weight belt feeder. The measurement is time averaged. This mass flow method is cumbersome because the initial and final storage tank volumes and densities must also be measured. This method is best done over a period of at least 24 hours. The second and preferred method is chemical analysis. The reagent consumption is calculated from the stoichiometric ratio, limestone (or reagent) analysis, and gas side measurements by the following equation:

$$LS = \frac{100\ (SR)\ (SO_2)}{64\ (CaCO_3)} \qquad (18)$$

where

LS = limestone consumption, lb/h
SR = stoichiometric ratio (see Chapter 35)
SO_2 = SO_2 removed, lb/h
64 = molecular weight of SO_2, lb/lb-mole
100 = molecular weight of $CaCO_3$, lb/lb-mole
$CaCO_3$ = available $CaCO_3$ in limestone (wt fraction: lb available $CaCO_3$/lb limestone)

Reagent consumption may be adjusted for deviations from SO_2 removal rate and mass flow (or concentration) design conditions.

Gypsum purity

A composite sample of gypsum, taken over the test period, is analyzed as outlined in Table 7. Adjustments may be necessary for flyash contamination, chloride concentration, and inerts if the inlet levels to the FGD system substantially deviate from design.

Other test methods

There are numerous procedures for measuring ECS parameters. The example in this chapter is a formal performance test; other methods could be used. For example, SO_2 can be measured with a CEMS or by other temporary instrument methods. Particulate level is measured by EPA Method 5 or 17 or alternate methods such as the probe shown in Fig. 11.

The future of environmental measurement

Advances in environmental measurement techniques will certainly continue. Technologies such as fiber optics, video imaging and Fourier transform infrared analysis (FTIR), which have been used successfully in the laboratory, are expected to move into the plant. Measurement will be extended to include other air and water pollutants, and air toxics monitoring will become increasingly important.

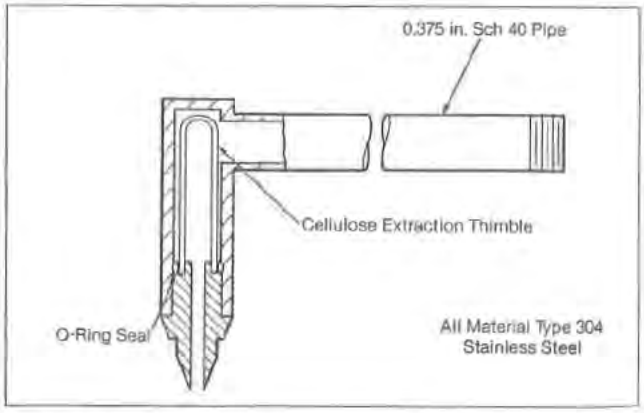


Fig. 11 Probe detail, isokinetic dust sampling.

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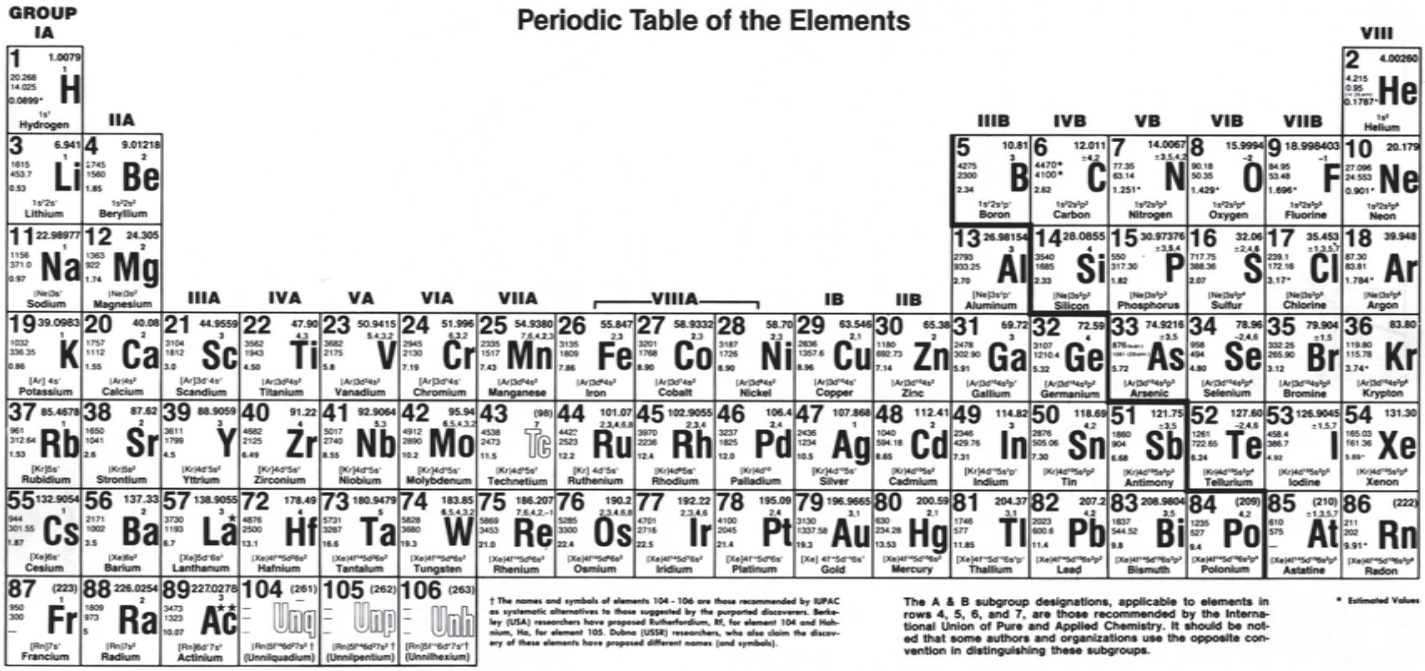
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Table 3.2
Periodic Table of the Elements



† The names and symbols of elements 104 - 106 are those recommended by IUPAC as systematic alternatives to those suggested by the purported discoverers. Berkeley (USA) researchers have proposed Rutherfordium, Rf, for element 104 and Hahnium, Hn, for element 105. Dubna (USSR) researchers, who also claim the discovery of these elements have proposed different names (and symbols).

The A & B subgroup designations, applicable to elements in rows 4, 5, 6, and 7, are those recommended by the International Union of Pure and Applied Chemistry. It should be noted that some authors and organizations use the opposite convention in distinguishing these subgroups.

* Estimated Values

NOTES:
(1) Based upon carbon-12, ( ) indicates most stable or best known isotope.
(2) Entries marked with asterisks refer to the gaseous state at 273 K and 1 atm and are given in units of g/l.

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Table 4.1
Coal and Ash Analysis, An Example of a High Sulfur Bituminous Coal

Proximate Analysis, %		Ultimate Analysis, %		Sulfur Forms, %	
Moisture	12.0	Moisture	12.0	Pyritic	2.0
VM	35.7	Carbon	60.6	Organic	2.0
FC	40.3	Hydrogen	4.1	Sulfate	0.1
Ash	12.0	Nitrogen	1.2		4.1
		Sulfur	4.1		
		Oxygen	6.0		
		Ash	12.0		

Ash Fusion Temperatures, F (Reducing Atmosphere)	
Initial Deformation	1950
Softening	2030
Hemispherical	2100
Fluid	2150

Other

Btu/lb = 10,950 as received
Grindability = 56 HGI
Theoretical air = 7.6 lb/10,000 Btu

Ash Analysis, %	
Si as SiO_2	45.0
Al as Al_2O_3	20.0
Fe as Fe_2O_3	18.0
Ti as TiO_2	1.0
Ca as CaO	7.0
Mg as MgO	1.0
Na as Na_2O	0.6
K as K_2O	1.9
S as SO_3	3.5
P as P_2O_5	0.2

Table 4.2
Mechanical Equipment 500 MW PC-Fired Plant

Steam Generator
Balanced draft, direct-fired, pulverized coal

Pulverizers
Six, roller-mill, 50 t/h each

Main Steam		Reheat Steam	
Flow	3.625×10^6 lb/h	Flow	3.336×10^6 lb/h
Temp	1005F	Temp	1000F
Pressure	2415 psi	Pressure	532 psi

Turbine/Generator
Tandem-compound, four-flow, 3600 rpm, 640,000 kVa at 24 kV, 3 phase, 60 Hz.

Fans
FD: 2 motor-driven, total 4,000,000 lb/h
PA: 2 motor-driven, total 710,000 lb/h
ID: 4 motor-driven, total 5,174,000 lb/h

Fan Horsepower Requirements
FD: 2400 HP based on static rise from -0.5 to 14 in. wg
PA: 1700 HP based on static rise from -0.5 to 16 in. wg
ID: 3800 HP based on static rise from -16 to -1 in. wg and 200F gas inlet.
Fan data assumes 8% air heater leakage and 87% fan/motor efficiency.

Feedwater Pumps
Main: 2 turbine-driven, 50% capacity, 5000 GPM each, TDH: 7200 ft
Booster: 2 turbine-driven, 50% capacity, 4800 GPM each, TDH: 1280 ft
Condensate: 2 motor-driven, 100% capacity, 1400 hp, 7000 GPM each, TDH: 630 ft

Table 4.3
Environmental Equipment 500 MW PC-Fired Plant

Wet Limestone Scrubber (full load)

Limestone usage rate:	30.5 t/h
Fixed sludge waste:	106 t/h
SO_2 at scrubber inlet:	36,568 lb/h
SO_2 at scrubber outlet:	3,657 lb/h
Required SO_2 removal:	90%

Particulate Control

Type:	Electrostatic precipitator
Emissions:	0.03 lb/10^6 Btu
Specific collector area:	350 ft^2/1000 SCFM

Table 4.4
Plant Performance Summary 500 MW PC-Fired Plant

Turbine Steam Conditions	
Pressure	2400psig
Temperature	1000F
Reheat temperature	1000F

Fuel Input
Coal feed 223 t/h
BTU input 4.88×10^9 Btu/h

Ash Flow Rates	t/h
Furnace bottom ash	5.1
Economizer ash	2.2
Fly ash	20.7
Total through unit	28.0
Mill rejects (pyrites)	0.5

Heat Rates at Full Load	
Steam cycle	7880 Btu/kWh
Gross	8985 Btu/kWh
Net	9725 Btu/kWh

Rated Output Power Production	
Total gross	536 MW
Total Net	502 MW

Thermal Discharge	10^6 Btu/h
Cooling towers	2500
Stack	587

Controlled Emissions	lb/h
SO_2	3700
NO_x as NO_2	2900*
Particulate	146

Efficiencies	
Boiler (coal to steam)	88%
Plant (coal to busbar)	35%

Untreated Flue Gas Characteristics

Component	Volume % Dry	Volume % Wet
Nitrogen	81.07	73.47
Carbon dioxide	15.28	13.84
Oxygen	3.26	2.95
Water	---	9.39
Sulfur dioxide	0.39	0.35
	100.00	100.00

Flue Gas Molecular Weight: 30.85 (dry basis), 29.64 (wet basis)

Excess air leaving economizer 17.9%

Mass Flows	lb/10,000 Btu
Actual dry air	8.888
Wet gas from fuel	0.800
Moisture in air	0.116
Wet gas weight	9.804
Water from fuel	0.444
Water in wet gas	0.560
Dry gas weight	9.244
Water in wet gas (as%)	5.711%

Losses	%
Dry gas loss	4.659
Water from fuel	5.056
Moisture in air	0.116
Unburned Combustibles	0.500
Radiation	0.170
Unaccounted for Manufacturer's Margin	1.500
	12.001

*Federal Standard. Current technology can attain lower.

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